

# **Solutions Manual for Analysis, Synthesis, and Design of Chemical Processes**

**Third Edition**

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# Chapter 1

- 1.1 Block Flow Diagram (BFD)  
Process Flow Diagram (PFD)  
Piping and Instrument Diagrams (P&ID)

- (a) PFD
- (b) BFD
- (c) PFD or P&ID
- (d) P&ID
- (e) P&ID

- 1.2 P&ID

- 1.3 It is important for a process engineer to be able to review a 3-dimensional model prior to the construction phase to check for clearance, accessibility, and layout of equipment, piping, and instrumentation.

- 1.4
- (1) Clearance for tube bundle removal on a heat exchanger.
  - (2) NPSH on a pump – affects the vertical separation of feed vessel and pump inlet.
  - (3) Accessibility of an instrument for an operator – must be able to read a PI or change/move a valve.
  - (4) Separation between equipment for safety reasons – reactors and compressors.
  - (5) Crane access for removing equipment.
  - (6) Vertical positioning of equipment to allow for gravity flow of liquid.
  - (7) Hydrostatic head for thermosiphon reboiler – affects height of column skirt.

- 1.5 Plastic models are no longer made because they are too expensive and difficult to change/revise. These models have been replaced with virtual/E-model using 3-D CAD. Both types of model allow revision of critical equipment and instrument placement to ensure access, operability, and safety.

- 1.6 Another reason to elevate the bottom of a tower is to provide enough hydrostatic head driving force to operate a thermosiphon reboiler.

- 1.7 (a) PFD or P&ID  
 (b) PFD  
 (c) PFD  
 (d) P&ID  
 (e) BFD (or all PFDs)

1.8 A pipe rack provides a clear path for piping within and between processes. It keeps piping off the ground to eliminate tripping hazards and elevates it above roads to allow vehicle access.

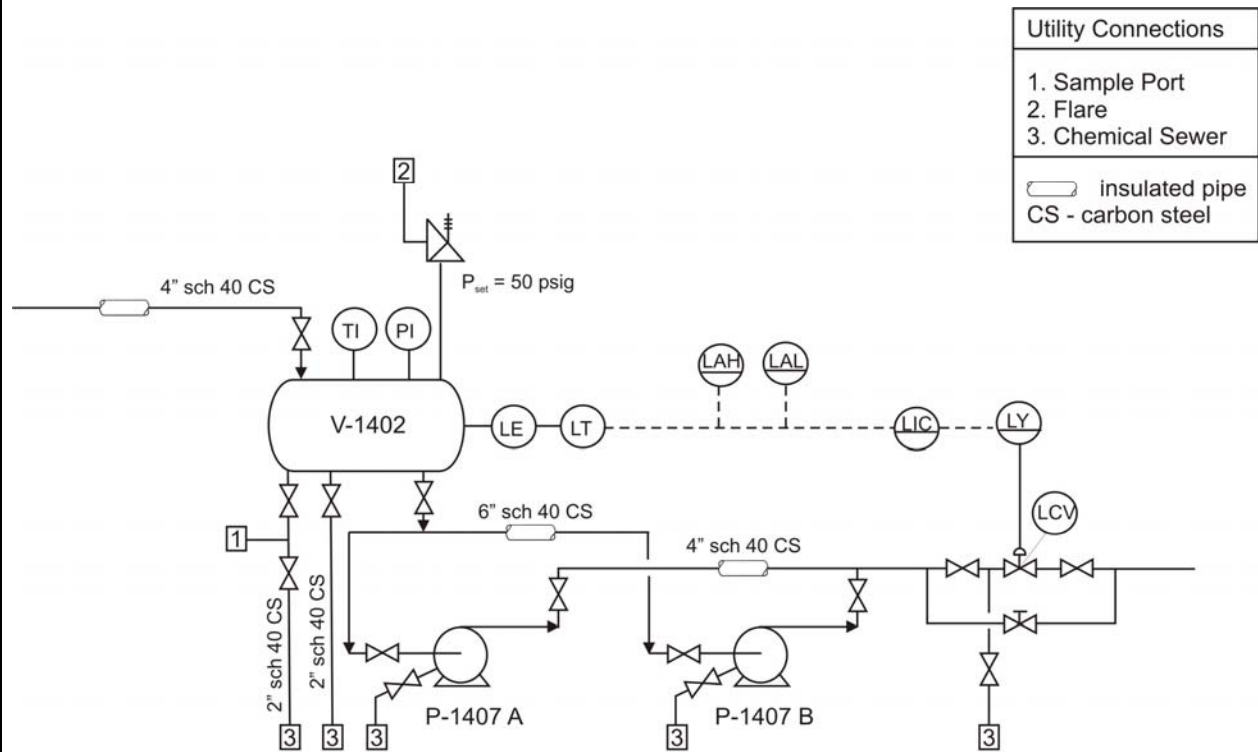
1.9 A structure – mounted vertical plant layout is preferred when land is at a premium and the process must have a small foot print. The disadvantage is that it is more costly because of the additional structural steel.

- 1.10 (a) BFD – No change  
 PFD – Efficiency changed on fired heater, resize any heat exchanger used to extract heat from the flue gas (economizer)  
 P&ID – Resize fuel and combustion air lines and instrumentation for utilities to fired heater. Changes for design changed of economizer (if present)  
 (b) BFD – Change flow of waste stream in overall material balance  
 PFD – Change stream table  
 P&ID – Change pipe size and any instrumentation for this process line  
 (c) BFD – No change  
 PFD – Add a spare drive, e.g. D-301 → D-301 A/B  
 P&ID – Add parallel drive  
 (d) BFD – No change  
 PFD – No change  
 P&ID – Note changes of valves on diagram

- 1.11 (a) A new vessel number need not be used, but it would be good practice to add a letter to denote a new vessel, e.g. V-203 → V-203N. This will enable an engineer to locate the new process vessel sheet and vendor information.  
 (b) P&ID definitely  
 PFD change/add the identifying letter.



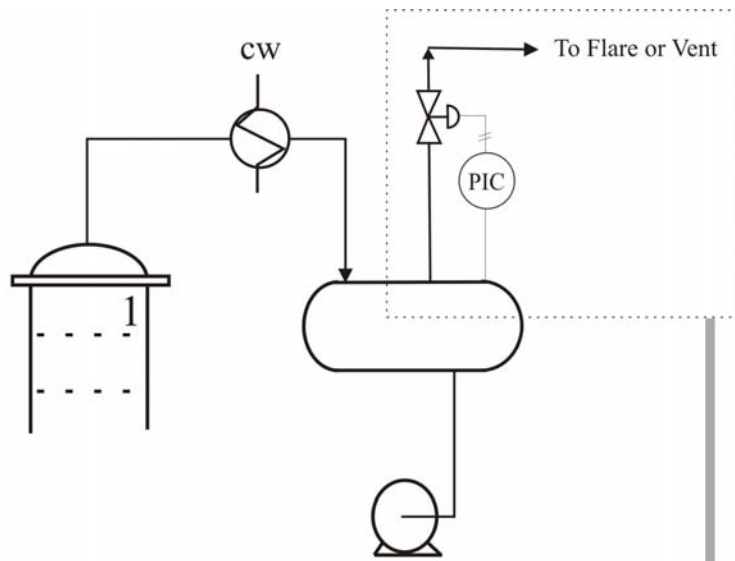
1.12



Solution to Chapter 1 - Problem 12

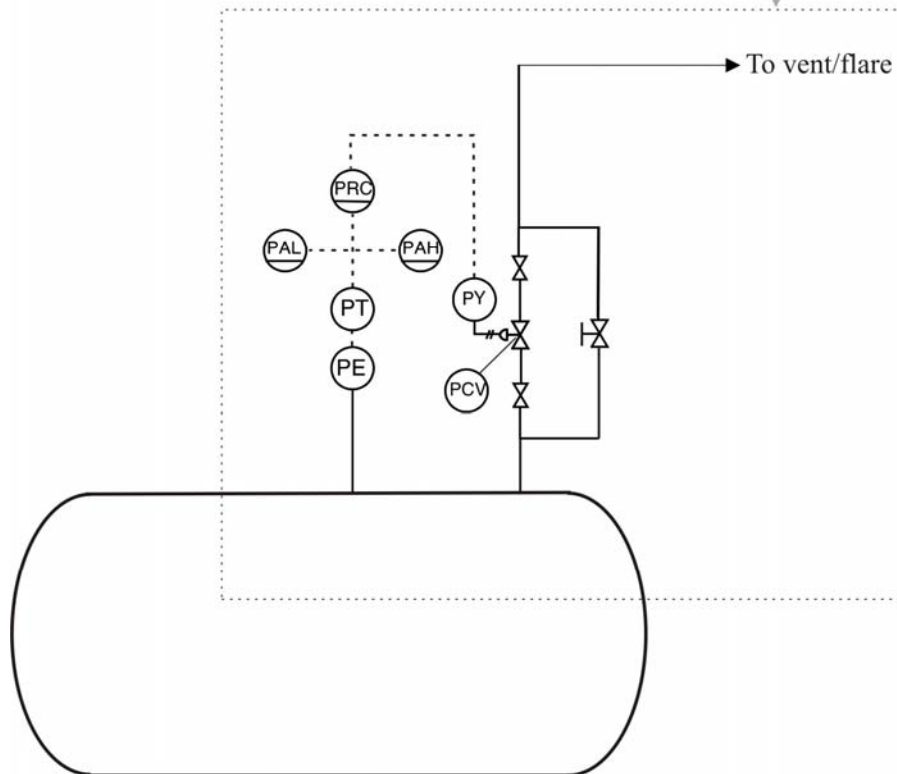
- 1.13 (a) (i) Open globe valve D  
(ii) Shut off gate valves A and C  
(iii) Open gate valve E and drain contents of isolated line to sewer  
(iv) Perform necessary maintenance on control valve B  
(v) Reconnect control valve B and close gate valve E  
(vi) Open gate valves A and C  
(vii) Close globe valve D
- (b) Drain from valve E can go to regular or oily water sewer.
- (c) Replacing valve D with a gate valve would not be a good idea because we lose the ability to control the flow of process fluid during the maintenance operation.
- (d) If valve D is eliminated then the process must be shut down every time maintenance is required on the control valve.

1.14



Solution to Chapter 1 - Problem 14

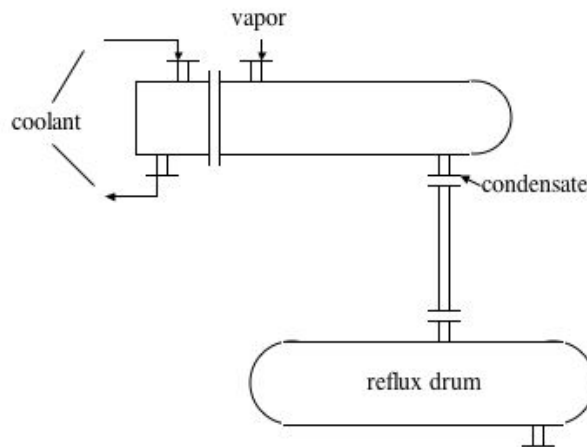
1.15



Solution to Chapter 1 - Problem 15

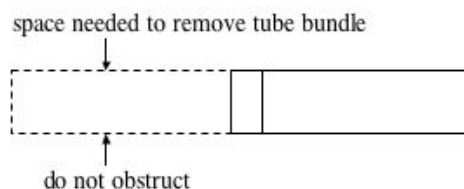
1.16 (a) For a pump with a large NPSH – the vertical distance between the feed vessel and the pump inlet must be large in order to provide the static head required to avoid cavitating the pump.

b) Place the overhead condenser vertically above the reflux drum – the bottom shell outlet on the condenser should feed directly into the vertical drum.



c) Pumps and control valves should always be placed either at ground level (always for pumps) or near a platform (sometimes control valves) to allow access for maintenance.

d) Arrange shell and tube exchangers so that no other equipment or structural steel impedes the removal of the bundle.



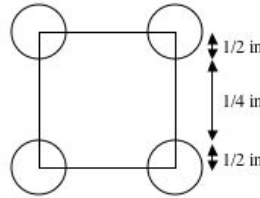
e) This is why we have pipe racks – never have pipe runs on the ground. Always elevate pipes and place on rack.

f) Locate plant to the east of major communities.

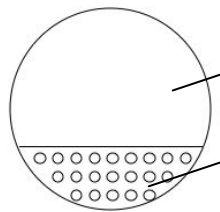
$$1.17 \text{ HT area of 1 tube} = \pi DL = \pi \left( \frac{1}{12} \right) (12 \text{ ft}) = 3.142 \text{ ft}^2$$

$$\text{Number of tubes} = (145 \text{ m}^2) \cdot \left( \frac{3.2808 \text{ ft}}{\text{m}} \right)^2 \left( \frac{1}{3.142 \text{ ft}^2} \right) = 497 \text{ tubes}$$

Use a 1 1/4 inch square pitch  $\Rightarrow$



$$\text{Fractional area of the tubes} = \frac{\pi \left( \frac{1 \text{ m}}{1.25 \text{ in}} \right)^2}{4} = 0.5027 \left( \frac{\text{m}}{\text{in}} \right)^2$$



$$A_{\text{VAP}} = 3 A_{\text{LIQ}} \therefore \text{CSA}_{\text{SHELL}} = 4 A_{\text{LIQ}}$$

$$A_{\text{LIQ}} = \left( \frac{497}{0.5027} \right) \left( \frac{\text{in}}{\text{m}} \right)^2 \left( \frac{\pi}{4} \right) (1 \text{ m})^2 = 777 \text{ in}^2$$

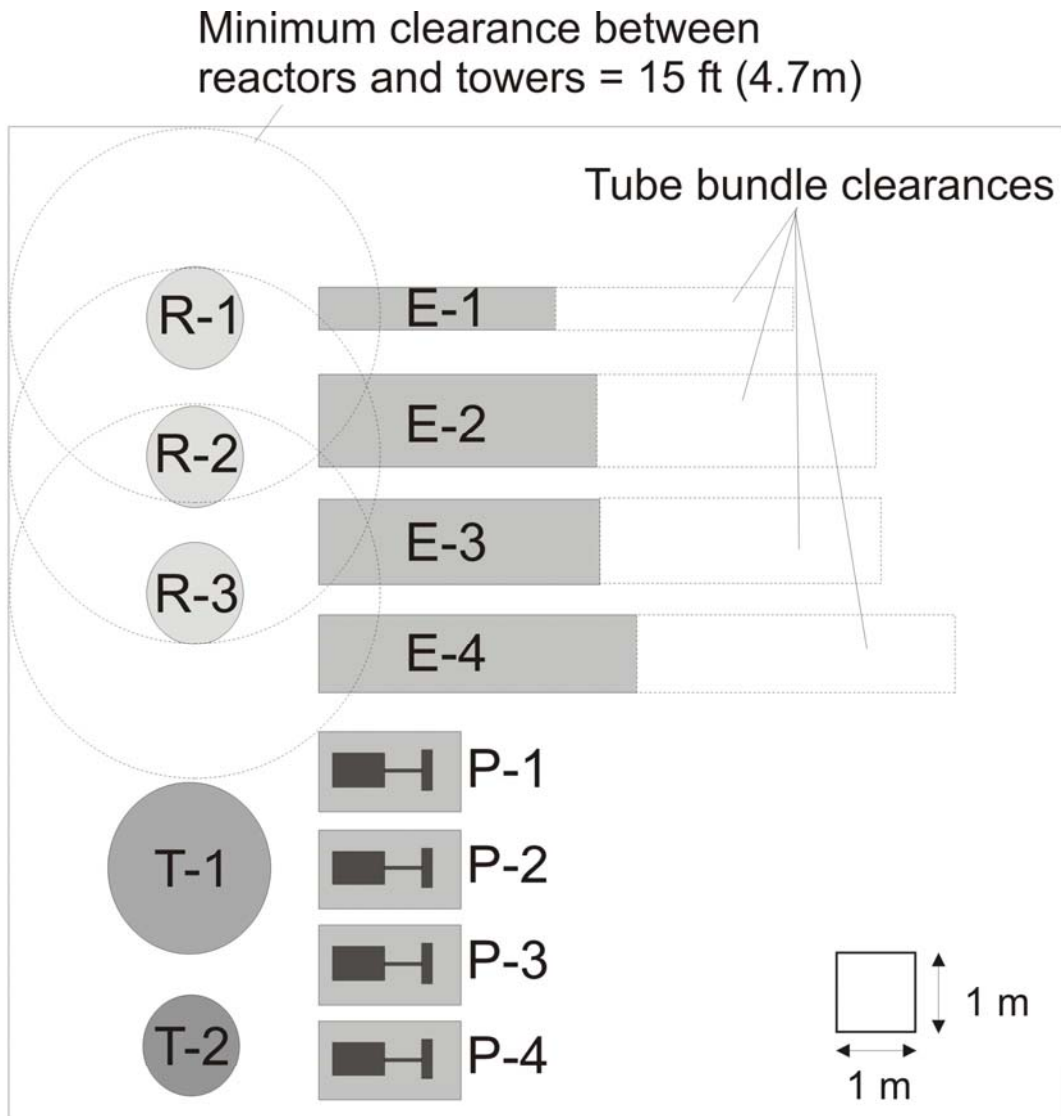
$$\text{CSA}_{\text{SHELL}} = (4)(777) = 3108 \text{ in}^2 \Rightarrow \frac{\pi}{4} D_{\text{SHELL}}^2 = 3108 \text{ in}^2$$

$$D_{\text{SHELL}} = \sqrt{\frac{(4)(3108 \text{ in}^2)}{\pi}} = 62.9 \text{ in} = 1.598 \text{ m}$$

$$\text{Length of Heat Exchanger} = (2 + 12 + 2) \text{ ft} = 16 \text{ ft} = 4.877 \text{ m}$$

$$\text{Foot Print} = 1.598 \times 4.877 \text{ m}$$

- 1.18 From Table 1.11 towers and reactors should have a minimum separation of 15 feet or 4.6 meters. No other restrictions apply. See sketch for details.



Solution to Chapter 1 - Problem 18 - many other variations are possible

1.19

**For shell and tube exchangers**

Assume 12 ft 1.25" tubes on a 1.5" square pitch

Area per tube 3.9270 ft<sup>2</sup> 0.3648 m<sup>2</sup>

fractional area of tubes 0.5454

assume length of shell = tube length + 2 + 2 = 16 ft

**For double pipe exchangers (E-103 and E-105)**

Outer pipe diameter = 8" Sch 20 = 8.125"

inner pipe diameter = 6" sch 40 6.065 inch 0.1541 m

length of pipe 12 ft

HT Area per single run of pipe 1.7702 m<sup>2</sup>

HT per 4 pass unit 7.0806 m<sup>2</sup>

length of 1 unit - 4 stacked Double pipes with U-bends 16 ft

width of multiple units - assume that units are spaced 1 foot apart

|                 | E-101  | E-102  | E-103  | E-104  | E-105   | E-106  |                |
|-----------------|--------|--------|--------|--------|---------|--------|----------------|
| Area            | 36     | 763    | 11     | 35     | 12      | 80     |                |
| number of tubes | 99     | 2092   |        | 96     |         | 220    |                |
| Number of units |        |        | 1.5535 |        | 1.69477 |        |                |
| Shell pass      | 1      | 1      |        | 1      |         | 1      |                |
| Tube pass       | 1      | 2      |        | 2      |         | 1      |                |
| CSA of tubes    | 0.0784 | 0.8282 |        | 0.0380 |         | 0.1742 | m <sup>2</sup> |
| CSA of Shell    | 0.1437 | 1.5184 |        | 0.0697 |         | 0.3194 | m <sup>2</sup> |
| Shell Diameter  | 0.4278 | 1.3904 |        | 0.2979 |         | 0.6377 | m              |
| length          | 16     | 16     | 16     | 16     | 16      | 16     | ft             |
| length          | 4.9    | 4.9    | 4.9    | 4.9    | 4.9     | 4.9    | m              |

**Foot print**

|   |      |      |      |      |      |      |   |
|---|------|------|------|------|------|------|---|
| L | 4.9  | 4.9  | 4.9  | 4.9  | 4.9  | 4.9  | m |
| W | 0.43 | 1.39 | 0.51 | 0.30 | 0.51 | 0.64 | m |

**Vessels, Towers and Reactor**

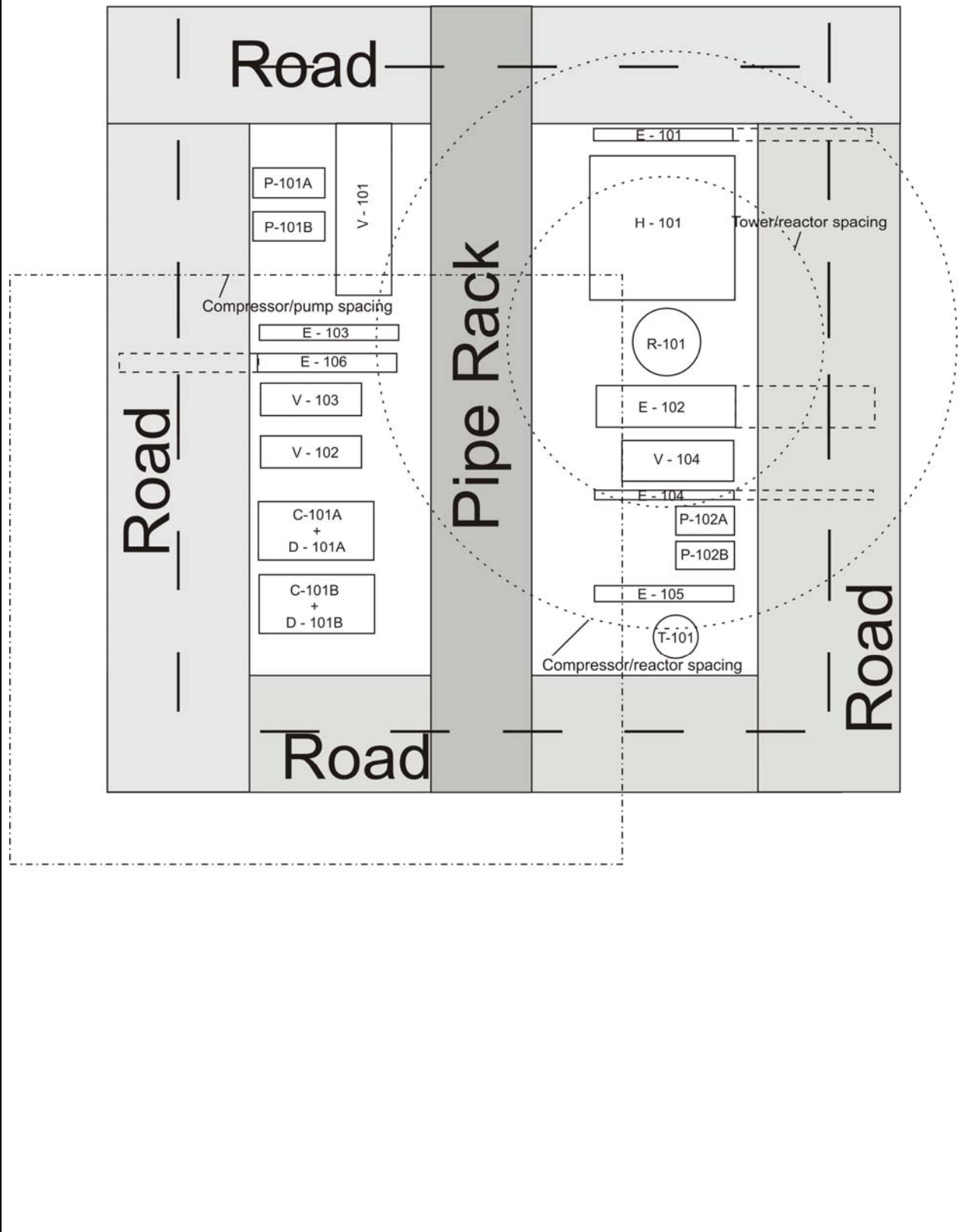
|            | V-101 | V-102 | V-103 | V-104 | T-101 | R-101 |   |
|------------|-------|-------|-------|-------|-------|-------|---|
| Foot print |       |       |       |       |       |       |   |
| L          | 5.9   | 3.5   | 3.5   | 3.9   |       |       | m |
| W          | 1.9   | 1.1   | 1.1   | 1.3   |       |       | m |
| D          |       |       |       |       | 1.5   | 2.3   | m |

**Pumps and Compressors**

|            | P-101 | P-102 | C-101+D-101 |   |  |  |  |
|------------|-------|-------|-------------|---|--|--|--|
| Foot print |       |       |             |   |  |  |  |
| L          | 2.5   | 2     | 4           | m |  |  |  |
| W          | 1     | 1     | 2           | m |  |  |  |

**Fired Heater**

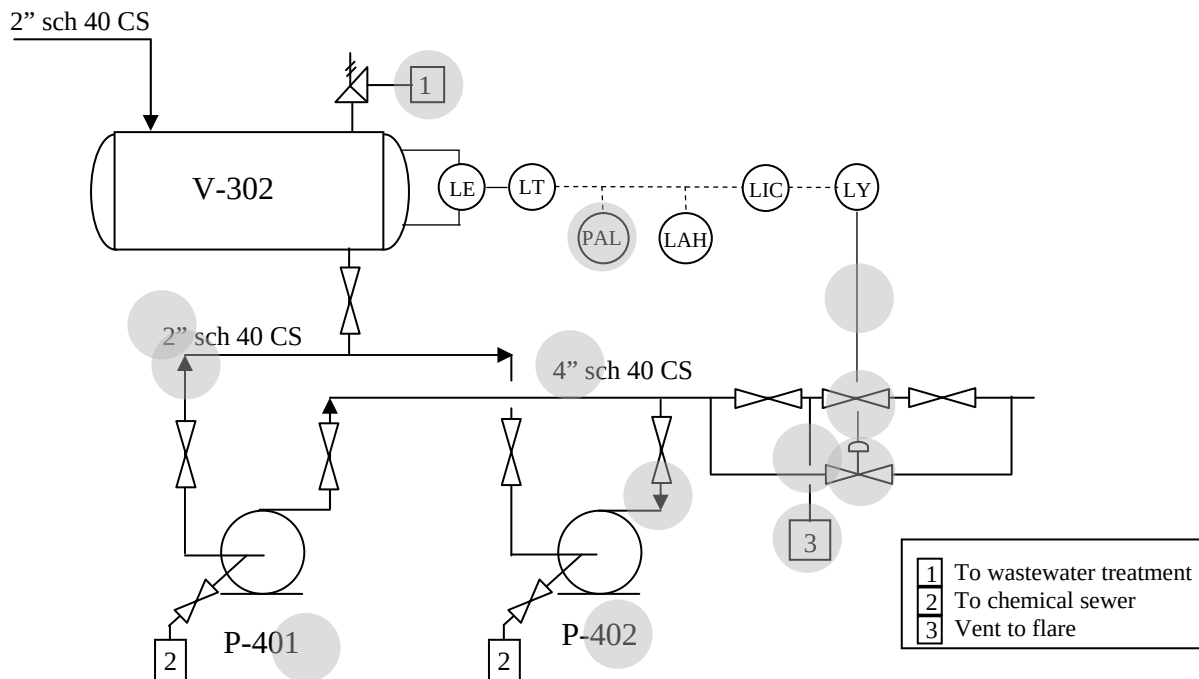
|            |       |   |  |  |  |  |  |
|------------|-------|---|--|--|--|--|--|
| Foot print | H-101 |   |  |  |  |  |  |
| L          | 5     | m |  |  |  |  |  |
| W          | 5     | m |  |  |  |  |  |





- 1.21 (a) A temperature (sensing) element (TE) in the plant is connected via a capillary line to a temperature transmitter (TT) also located in the plant. The TT sends an electrical signal to a temperature indicator controller (TIC) located on the front of a panel in the control room.
- (b) A pressure switch (PS) located in the plant sends an electrical signal to ...
- (c) A pressure control valve (PCV) located in the plant is connected by a pneumatic (air) line to the valve stem.
- (d) A low pressure alarm (PAL) located on the front of a panel in the control room receives an electrical signal from ...
- (e) A high level alarm (LAH) located on the front of a panel in the control room receives a signal via a capillary line.

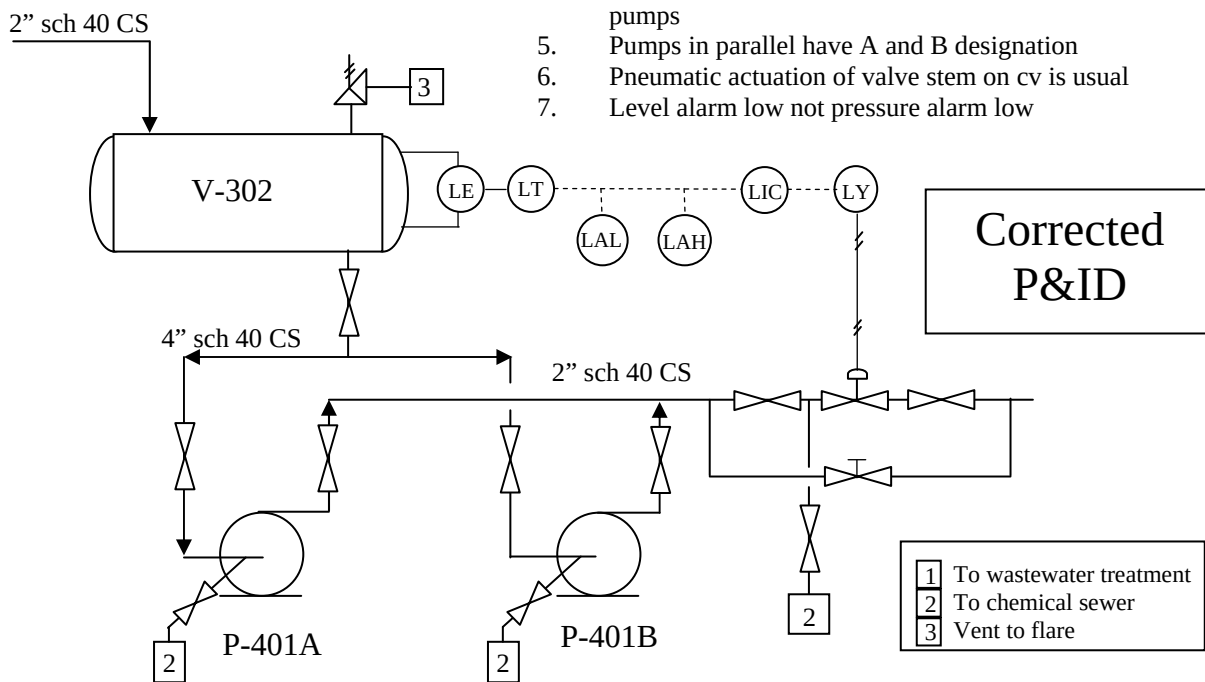
1.22



● = Error

## List of Errors

1. Pipe inlet always larger than pipe outlet due to NPSH issues
2. Drains to chemical sewer and vent to flare
3. Double-block and bleed needed on control valve
4. Arrows must be consistent with flow of liquid through pumps
5. Pumps in parallel have A and B designation
6. Pneumatic actuation of valve stem on cv is usual
7. Level alarm low not pressure alarm low



## Chapter 2

2.1 The five elements of the Hierarchy of Process Design are:

- a. Batch or continuous process
- b. Input – output structure of process
- c. Recycle structure of process
- d. General separation structure of process
- e. Heat-exchanger network/process energy recovery

2.2

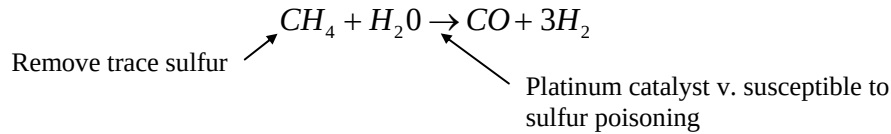
- a. Separate/purify unreacted feed and recycle – use when separation is feasible.
- b. Recycle without separation but with purge – when separation of unused reactants is infeasible/uneconomic. Purge is needed to stop build up of product or inerts.
- c. Recycle without separation or purge – product/byproduct must react further through equilibrium reaction.

2.3 Batch preferred over continuous when: small quantities required, batch-to-batch accountabilities required, seasonal demand for product or feed stock availability, need to produce multiple products using the same equipment, very slow reactions, and high equipment fouling.

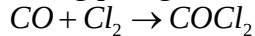
2.4 One example is the addition of steam to a catalytic reaction using hydrocarbon feeds. Examples are given in Appendix B (styrene, acrylic acid.) In the styrene process, superheated steam is added to provide energy for the desired endothermic reaction and to force the equilibrium towards styrene product. In the acrylic acid example, steam is added to the feed of propylene and air to act as thermal ballast (absorb the heat of reaction and regulate the temperature), and it also serves as an anti-coking agent – preventing coking reactions that deactivate the catalyst.

2.5 Reasons for purifying a feed material prior to feeding it to a process include:

- a. If impurity foul or poison a catalyst used in the process.  
e.g. Remove trace sulfur compounds in natural gas prior to sending to the steam reforming reactor to produce hydrogen.

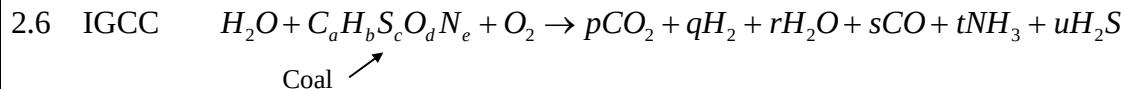


- b. If impurities react to form difficult-to-separate or hazardous products/byproducts.  
e.g. Production of isocyanates using phosgene. Production of phosgene is



The carbon monoxide is formed via steam reforming of  $\text{CH}_4$  to give  $\text{CO} + \text{H}_2$ .  $\text{H}_2$  must be removed from  $\text{CO}$  prior to reaction with  $\text{Cl}_2$  to form  $\text{HCl}$ , which is highly corrosive and causes many problems in the downstream processes.

- c. If the impurity is present in large quantities then it may be better to remove the impurity rather than having to size all the down stream equipment to handle the large flow of inert material.  
e.g. One example is using oxygen rather than air to fire a combustion or gasification processes. Removing nitrogen reduces equipment size and makes the removal of  $\text{CO}_2$  and  $\text{H}_2\text{S}$  much easier because these species are more concentrated.



In modern IGCC plants, coal is partially oxidized (gasified) to produce synthesis gas  $\text{CO} + \text{H}_2$  and other compounds. Prior to combusting the synthesis gas in a turbine, it must be “cleaned” or  $\text{H}_2\text{S}$  and  $\text{CO}_2$  (if carbon capture is to be employed.) Both  $\text{H}_2\text{S}$  and  $\text{CO}_2$  are acid gases that are removed by one of a variety of physical or chemical absorption schemes. By removing nitrogen from the air, the raw synthesis gas stream is much smaller making the acid gas removal much easier. In fact, when  $\text{CO}_2$  removal is required IGCC is the preferred technology, i.e. the cheapest.

## 2.7 Ethylebenzene Process

- a. Single pass conversion of benzene

$$\text{Benzene in reactor feed (stream 3)} = 226.51 \frac{\text{kmol}}{\text{h}}$$

$$\text{Benzene in reactor effluent (stream 14)} = 177.85 \frac{\text{kmol}}{\text{h}}$$

$$X_{sp} = 1 - \frac{177.85 \frac{\text{kmol}}{\text{h}}}{226.51 \frac{\text{kmol}}{\text{h}}} = 21.5\%$$

- b. Single pass conversion of ethylene

$$\text{Ethylene in reactor feed (stream 2)} = 93.0 \frac{\text{kmol}}{\text{h}}$$

$$\text{Ethylene in reactor effluent (stream 14)} = 0.54 \frac{\text{kmol}}{\text{h}}$$

$$X_{sp} = 1 - \frac{0.54 \frac{\text{kmol}}{\text{h}}}{93.0 \frac{\text{kmol}}{\text{h}}} = 99.4\%$$

- c. Overall conversion of benzene

$$\text{Benzene entering process (stream 1)} = 97.0 \frac{\text{kmol}}{\text{h}}$$

$$\text{Benzene leaving process (stream 15 and 19)} = 8.38 + 0.17 \frac{\text{kmol}}{\text{h}}$$

$$X_{ov} = 1 - \frac{8.55 \frac{\text{kmol}}{\text{h}}}{97.0 \frac{\text{kmol}}{\text{h}}} = 91.2\%$$

- d. Overall conversion of ethylene

$$\text{Ethylene entering process (stream 2)} = 93.0 \frac{\text{kmol}}{\text{h}}$$

$$\text{Ethylene leaving process (stream 15 and 19)} = 0.54 + 0 \frac{\text{kmol}}{\text{h}}$$

$$X_{ov} = 1 - \frac{0.54 \frac{\text{kmol}}{\text{h}}}{93.0 \frac{\text{kmol}}{\text{h}}} = 99.4\%$$

2.8 Separation of G from reactor effluent may or may not be difficult. (a) If G reacts to form a heavier (higher molecular weight) compound then separation may be relatively easy using a flash absorber or distillation and recycle can be achieved easily. (b) If process is to be viable then G must be separable from the product. If inerts enter with G or gaseous by-products are formed then separation of G may not be possible but recycling with a purge should be tried. In either case the statement is not true.

2.9 Pharmaceutical products are manufactured using batch process because:

- a. they are usually required in small quantities
- b. batch-to-batch accountability and tracking are required by the Food & Drug Administration (FDA)
- c. usually standardized equipment is used for many pharmaceutical products and campaigns are run to produce each product – this lends itself to batch operation.

2.10 a. Single pass conversion of ethylbenzene

$$\text{Ethylbenzene in reactor feed (stream 9)} = 512.7 \frac{\text{kmol}}{\text{h}}$$

$$\text{Ethylbenzene in reactor effluent (stream 12)} = 336.36 \frac{\text{kmol}}{\text{h}}$$

$$\text{Single pass conversion} = 1 - \frac{336.36 \frac{\text{kmol}}{\text{h}}}{512.7 \frac{\text{kmol}}{\text{h}}} = 34.4\%$$

b. Overall conversion of ethylbenzene

$$\text{Ethylbenzene entering process (stream 1)} = 180 \frac{\text{kmol}}{\text{h}}$$

$$\text{Ethylbenzene leaving process (stream 19, 26, 27 \& 28)} = 3.36 + 0.34 = 3.70 \frac{\text{kmol}}{\text{h}}$$

$$\text{Overall conversion} = 1 - \frac{3.70 \frac{\text{kmol}}{\text{h}}}{180 \frac{\text{kmol}}{\text{h}}} = 97.9\%$$

c. Yield of styrene

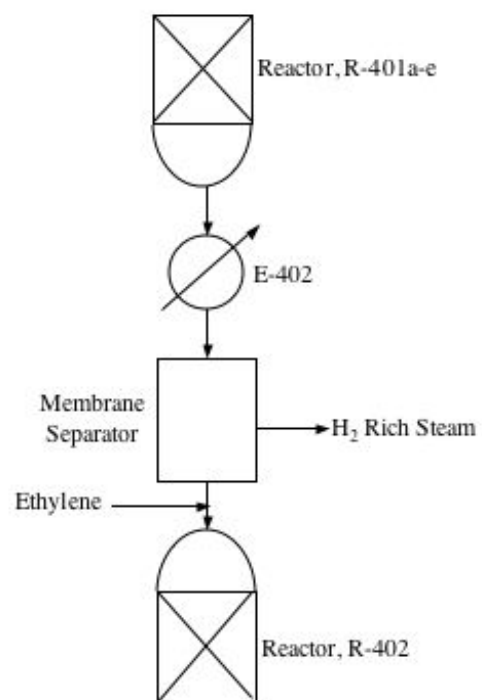
$$\text{Moles of ethylbenzene required to produce styrene} = 119.3 \frac{\text{kmol}}{\text{h}}$$

$$\text{Moles of ethylbenzene fed to process (stream 1)} = 180 \frac{\text{kmol}}{\text{h}}$$

$$\text{Yield} = \frac{119.3 \frac{\text{kmol}}{\text{h}}}{180 \frac{\text{kmol}}{\text{h}}} = 66.3\%$$

Possible strategies to increase the yield of styrene are

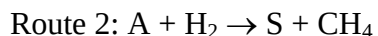
- (i) Increase steam content of reactor feed – this pushes the desired equilibrium reaction to the right.
- (ii) Increasing the temperature also pushes the equilibrium to right but increases benzene and toluene production.
- (iii) Remove hydrogen in effluent from each reactor – this will push the equilibrium of the desired reaction to the right and reduce the production of toluene from the third reaction – use a membrane separator, shown on following page.





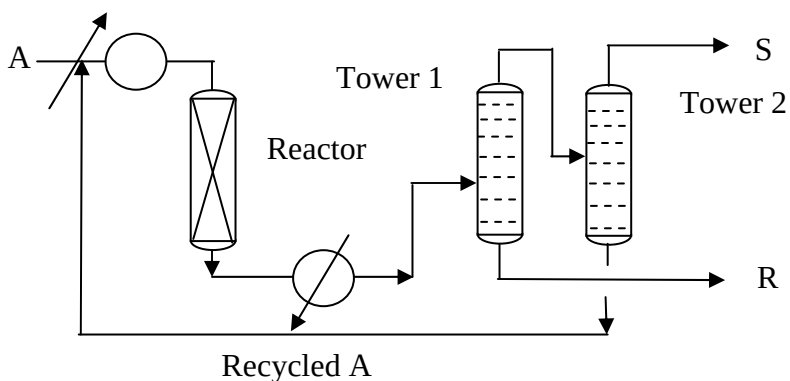
## 2.11 Route 1: $2A \rightarrow S + R$

Key features are that no light components (non-condensables) are formed and only one reactant is used. Therefore, separation of A, R, and S can take place using distillation columns.

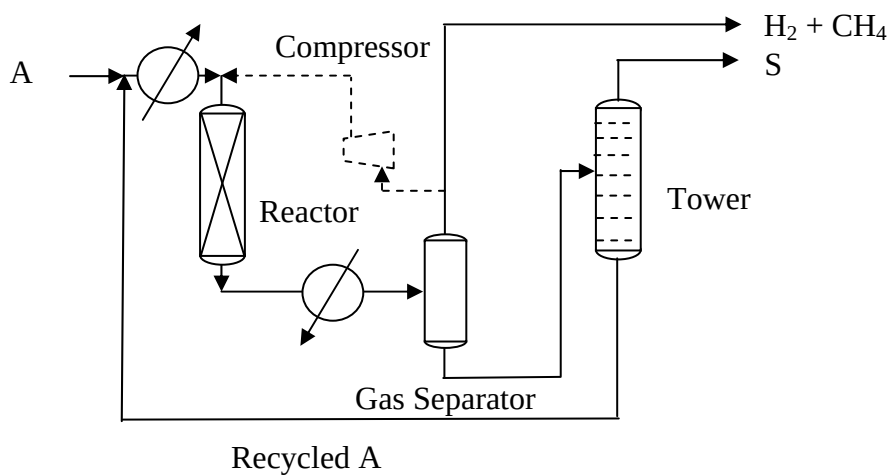


Unlike Route 1, this process route requires separation of the non-condensables from A and S. If hydrogen is used in great excess (as with the toluene HDA process), then a recycle and purge of the light gas stream will be required. Otherwise, if hydrogen conversion is high, the unreacted hydrogen along with the methane may be vented directly to fuel gas.

### Route 1 – PFD sketch



Route 2 – PFD sketch – gas recycle shown dotted since it is only needed if  $H_2$  is used in (considerable) excess and must be recycled.

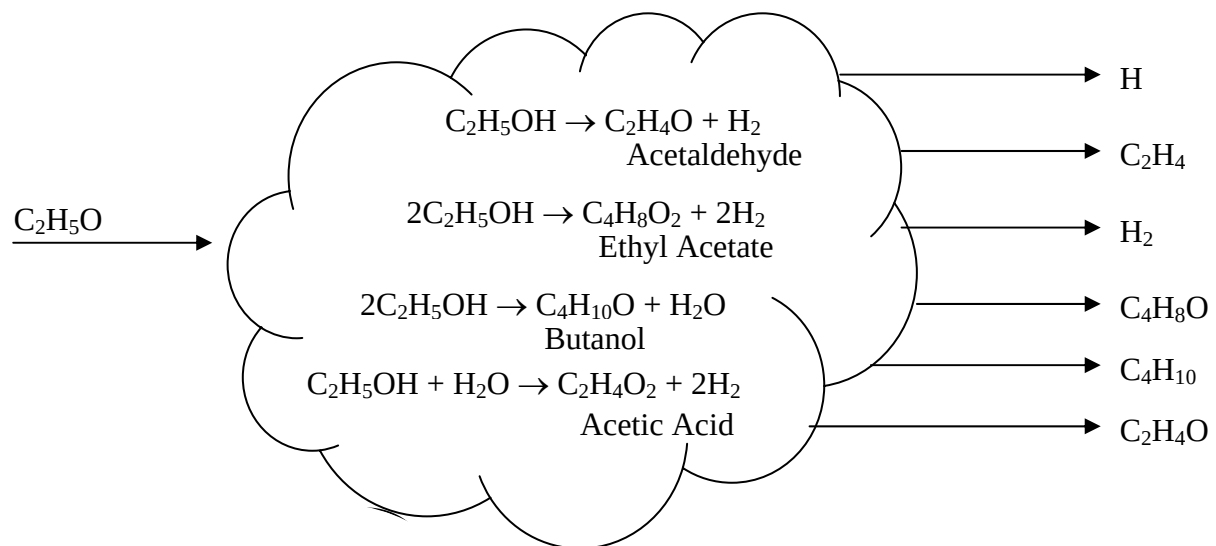


Route 1 is better since:

- Simpler PFD
- No gas recycle (no recycle compressor)
- No build up of inerts ( $\text{CH}_4$ ) so recycle stream is not as large
- All products are valuable – fuel gas in Route 2 has a low value

- 2.12
- a. Good when product(s) and reactant(s) are easily separated and purified (most often by distillation.) Any inerts in the feed or byproducts can be removed by some unit operation and thus recycle does not require a purge.
  - b. When unused reactant(s) and product(s) are not easily separated (for example when both are low boiling point gases) and single pass conversion of reactant is low.
  - c. This is only possible when no significant inerts are present and any byproducts formed will react further or can reach equilibrium.

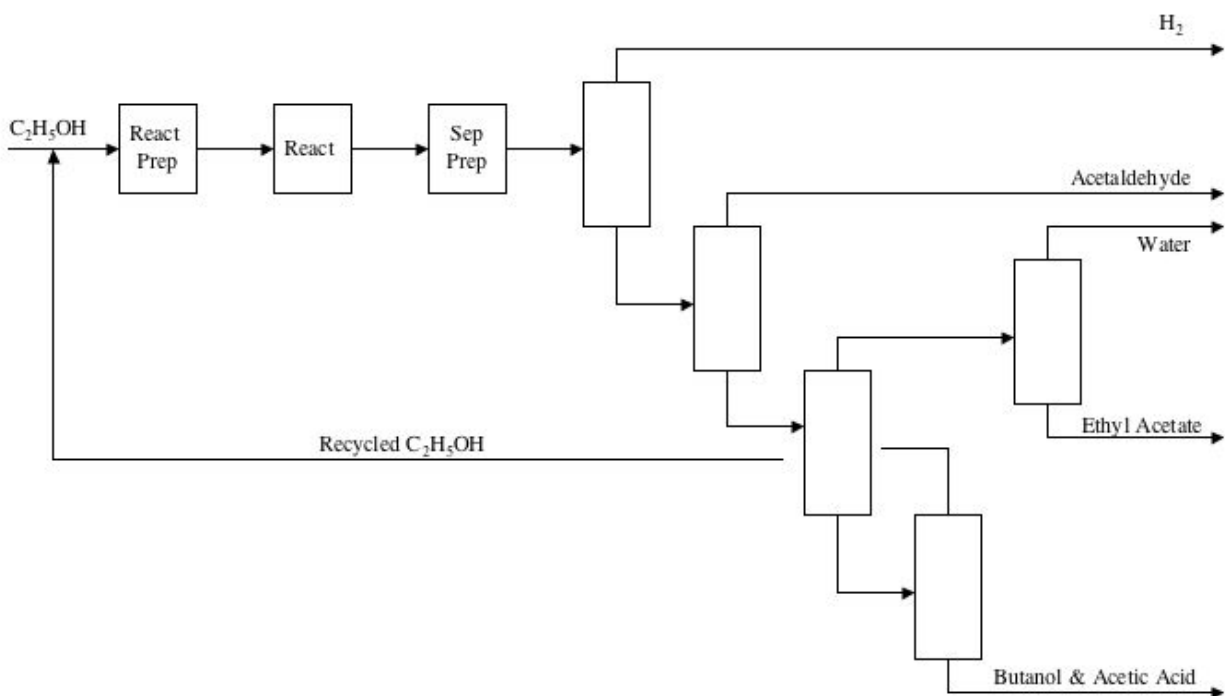
2.13 a.



Order of volatility is acetaldehyde, water, ethyl acetate, ethanol, isobutanol, acetic acid.

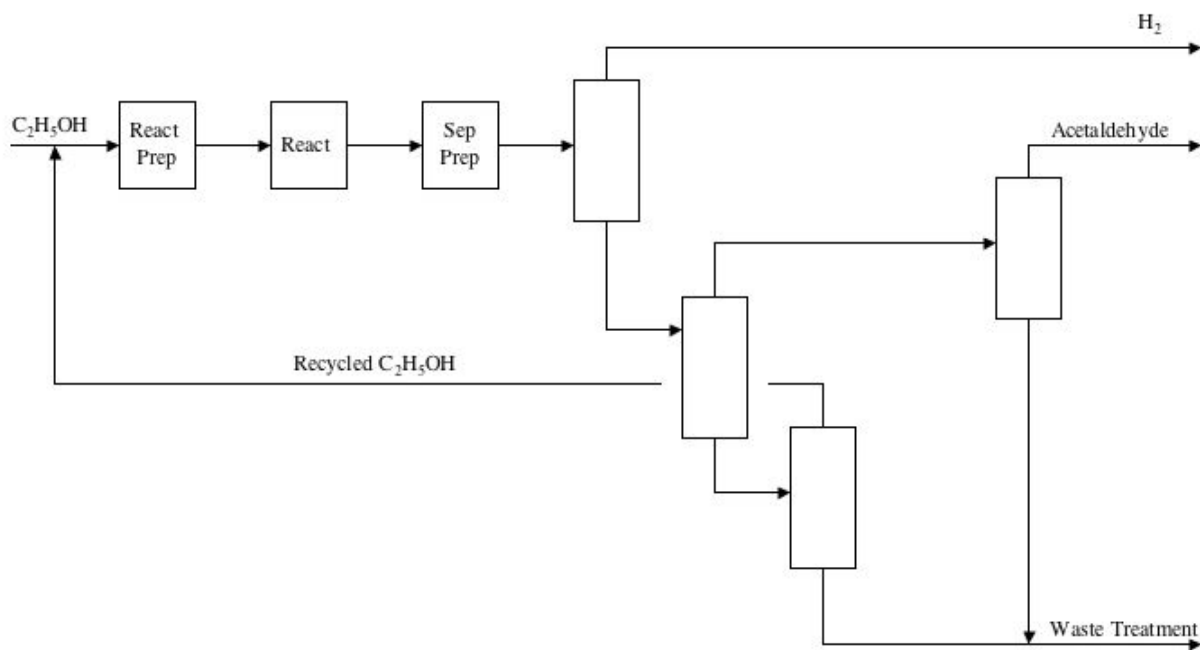
b.

Alternative 1



Alternative 1 assumes butanol and acetic acid can be sold as a mixed product  $\Rightarrow$  very unlikely so probably have to add another column to separate.

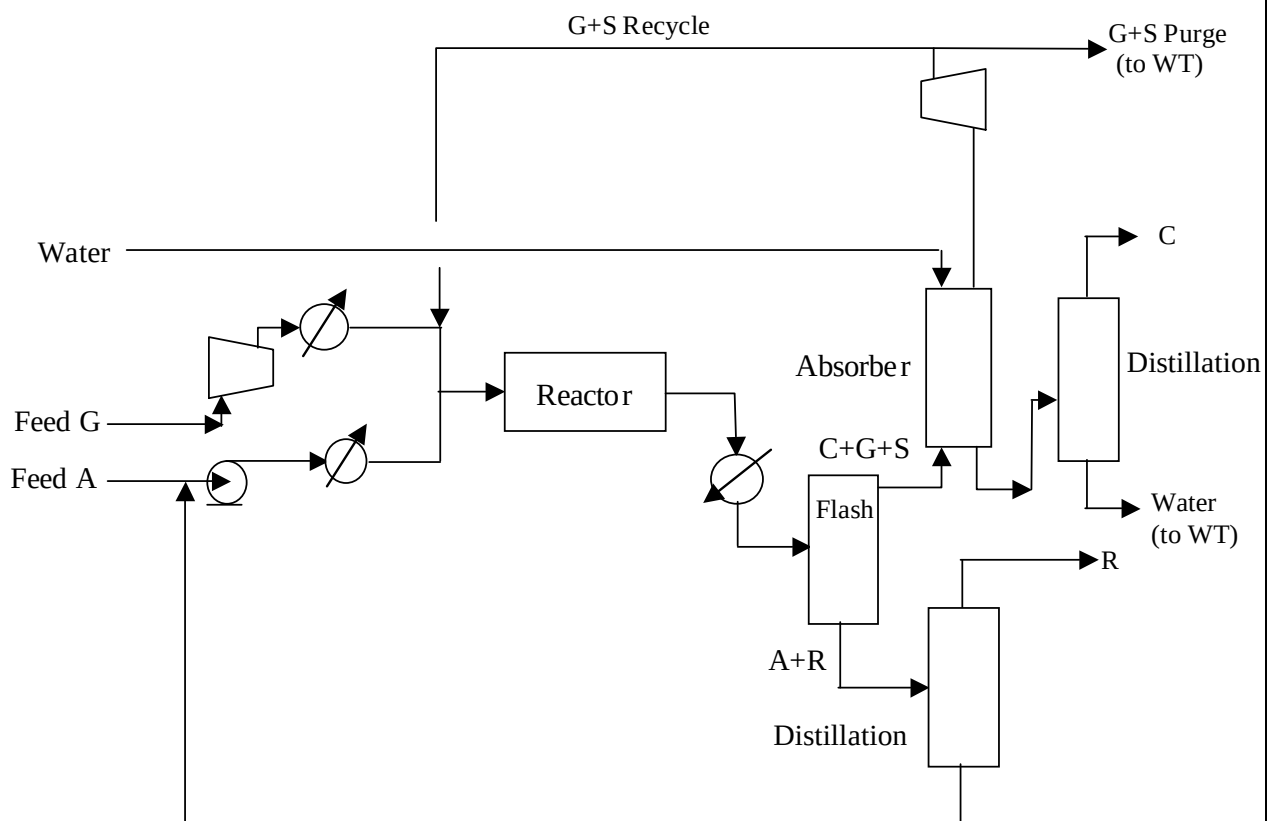
### Alternative 2



This alternative recycles  $C_2H_5OH$  and produces “pure” acetaldehyde – the remaining streams are considered waste – incineration of organics or wastewater treatment are possible ways to remove organics.

2.14

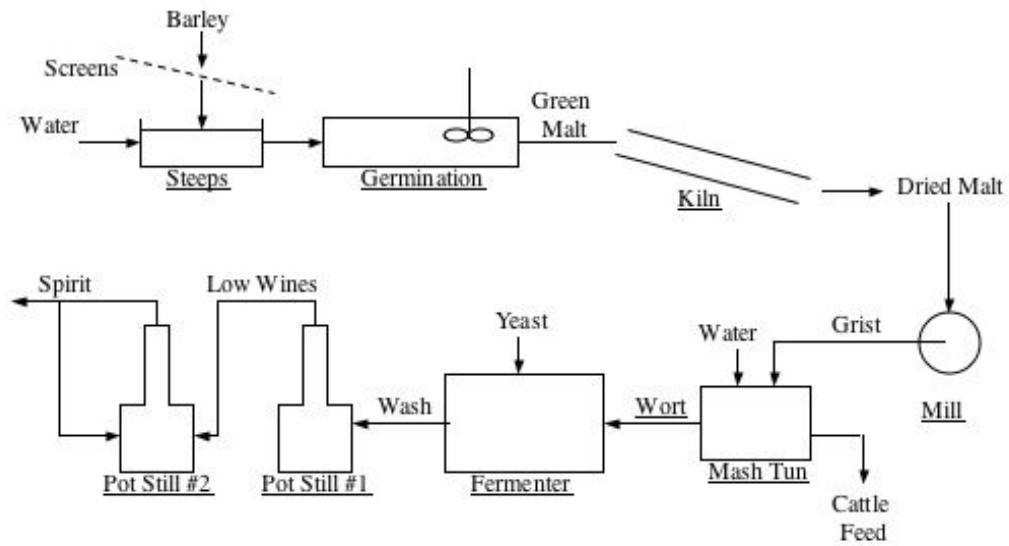
- A and R are both condensable and may be separated via distillation
- C may be separated by absorption into water
- R will be absorbed into water
- G and S cannot be separated except at very high pressure or low temperature
- After reaction, cool and condense A and R from other components.
- Separate A from R using distillation and recycle purified liquid A to the front end of the process
- Treat remaining gas stream in a water absorber to remove product C
- Separate C and from water via distillation
- Recycle unused G containing S – since S does not react further – we must add a purge to prevent accumulation of S in the system. This stream must be recycled as a gas using a recycle gas compressor.



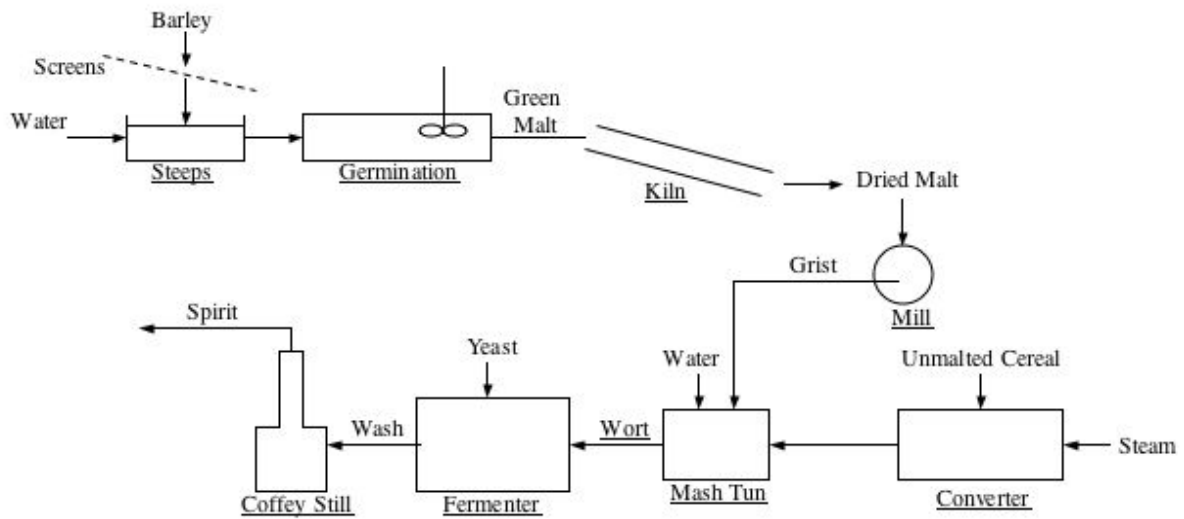
If the value of G was very low, then consider not recycling G (and S.)

2.15

### Malt Whiskey Process



### Grain Whisky Process



## Chapter 3

### 3.1. What is a flowshop plant?

A flowshop plant is a plant in which several batch products are produced using all or a subset of the same equipment and in which the operations for each batch follow the same sequence. Thus the flow of any batch through equipment A, B, C, D ... is always  $A \rightarrow B \rightarrow C \rightarrow D \rightarrow \dots$ . Omissions of equipment are possible but no reversal in direction is allowed.

### 3.2. What is a jobshop plant?

A flowshop plant is a plant in which several batch products are produced using all or a subset of the same equipment but for which the operations of at least one batch product do not follow the same sequence, e.g.,  $A \rightarrow C \rightarrow D \rightarrow B$

### 3.3. What are the two main methods for sequencing multiproduct processes?

Either use multi-product campaigns or multiple single-product campaigns.

### 3.4. Give one advantage and one disadvantage of using single-product campaigns in a multiproduct plant.

Advantage – sequencing of single-product campaigns is relatively simple and repetitive and probably less prone to operator error since the batch recipe remains the same over the entire campaign.

Disadvantage – significant final product storage will be required since all products will not be made all the time and in order to even out supply some inventory of products will have to be maintained in storage. Single-product campaigns may be less efficient than multi-product campaigns.

### 3.5. What is the difference between a zero-wait and a uis process?

A zero-wait process is one in which the batch is transferred immediately from the current piece of equipment to the next piece of equipment in the recipe sequence. This type of process eliminates the need for intermediate storage (storage of unfinished products or intermediates).

A uis (unlimited intermediate storage) process is one in which any amount of any intermediate product may be stored. Such a process maximizes the use of the processing equipment but obviously requires an unlimited amount of storage.



- 3.6 Number of batches of A is twice that for B or C – repeat Example 3.3 with this restriction using a 500 h cycle time.

Table E3.3: Equipment times needed to produce A, B, and C

| Product | Time in Mixer | Time in Reactor | Time in Separator | Time in Packaging |
|---------|---------------|-----------------|-------------------|-------------------|
| A       | 1.5           | 1.5             | 2.5               | 2.5               |
| B       | 1.0           | 2.5             | 4.5               | 1.5               |
| C       | 1.0           | 4.5             | 3.5               | 2.0               |

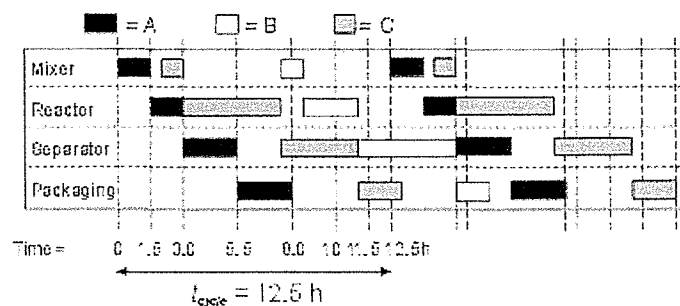
Using Equation (3.6) with  $t_{cycle,A} = 2.5$ ,  $t_{cycle,B} = 4.5$ , and  $t_{cycle,C} = 4.5$

If  $x$  is the number of batches of Products B and C, then  $2x$  is the number of batches of Product A

$$T = 500 = 2x(2.5) + x(4.5 + 4.5) \Rightarrow x = \frac{500}{14} = 35.7$$

Number of batches for each product are A = 70, B=35, C=35

- 3.7 For Examples 3.3 and 3.4, determine the number of batches that can be produced in a month (500 h) using a multi-product campaign strategy with the sequence ACBACBACB. Are there any other sequences for this problem other than the one used in Example 3.4 and the one used here?



The multi-product cycle time = 2.5 + 2.0 + 3.5 + 4.5 = 12.5 h

Number of batches per month = (500)/(12.5) = 40 each of A, B, and C

The only sequences that can be used for multi-product campaigns are ABCABCABC (Example 3.4) and ACBACBACB as used here.

3.8 Consider the multi-product batch plant described in Table P3.8

Table P3.8: Equipment Processing Times for Processes A, B, and C

| Process | Mixer | Reactor | Separator |
|---------|-------|---------|-----------|
| A       | 2.0 h | 5.0 h   | 4.0 h     |
| B       | 3.0 h | 4.0 h   | 3.5 h     |
| C       | 1.0 h | 3.0 h   | 4.5 h     |

It is required to produce the same number of batches of each product. Determine the number of batches that can be produced in a 500 h operating period using the following strategies:

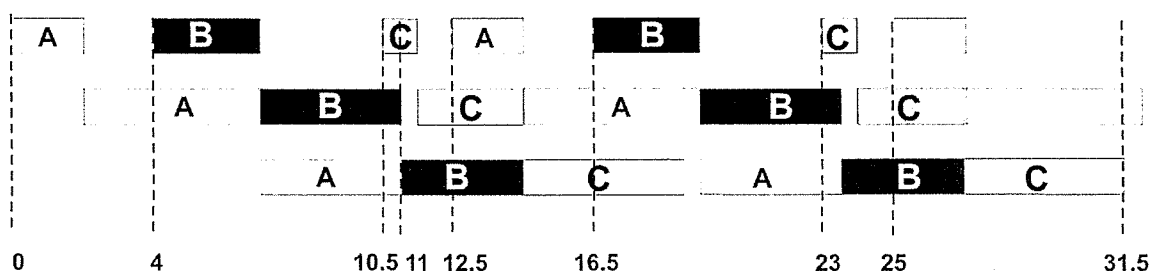
- (a) using single-product campaigns for each product

Using Equation (3.6) with  $t_{cycle,A} = 5.0$ ,  $t_{cycle,B} = 4.0$ , and  $t_{cycle,C} = 4.5$

$$T = 500 = x(5.0 + 4.0 + 4.5) \Rightarrow x = \frac{500}{13.5} = 37.0$$

$x = 37$  batches

- (b) using a multi-product campaign using the sequence ABCABCABC...



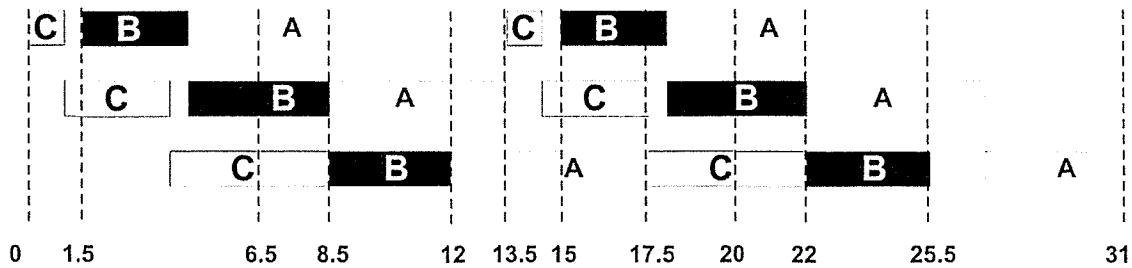
From this diagram we see that the cycle time for the multi-product campaign using the sequence ABC is 12.5 h.

Therefore, the number of batches,  $x$ , of each product that can be made during a 500 h period is given by:

$$T = 500 = 12.5x \Rightarrow x = \frac{500}{12.5} = 40$$

$x = 40$  batches

(c) using a multi-product campaign using the sequence CBACBACBA...



From this diagram we see that the cycle time for the multi-product campaign using the sequence ABC is 13.5 h.

Therefore, the number of batches,  $x$ , of product that can be made during a 500 h period is given by:

$$T = 500 = 13.5x \Rightarrow x = \frac{500}{13.5} = 37.0$$

$x = 37$  batches

- 3.9 Consider the process given in Problem 3.8. Assuming that a single-product campaign strategy is repeated every 500 h operating period and further assuming that the production rate (for a year = 6,000 h) for products A, B, C are 18,000 kg/y, 24,000 kg/y, and 30,000 kg/y, respectively, determine the minimum volume of product storage required. Assume that the product densities of A, B, and C are 1100, 1200, and 1000 kg/m<sup>3</sup>, respectively

The tables below shows the results using data given from Problem 8

| Rate                                                   | Product A                    | Product B                   | Product C                     |
|--------------------------------------------------------|------------------------------|-----------------------------|-------------------------------|
| Volume (m <sup>3</sup> ) of product required per month | 18,000/12/1,100<br>= 1.36    | 24,000/12/1,200<br>= 1.67   | 30,000/12/1,000<br>= 2.5      |
| Cycle time (h)                                         | 5.0                          | 4.0                         | 4.5                           |
| Production rate, $r_p$ (m <sup>3</sup> /h)             | (1.36)/(37)(5)<br>= 0.007371 | (1.67)/(37)(4)<br>= 0.01126 | (2.5)/(37)(4.5)<br>= 0.015015 |
| Demand rate, $r_d$ (m <sup>3</sup> /h)                 | (1.36)/(500)<br>= 0.002727   | 0.003333                    | 0.005                         |

| Product | Campaign time,<br>$t_{camp}$<br>(h) | $r_p - r_d$<br>(m <sup>3</sup> /h) | Minimum volume of<br>product storage, $V_s$<br>(m <sup>3</sup> ) |
|---------|-------------------------------------|------------------------------------|------------------------------------------------------------------|
| A       | (37)(5) = 185                       | 0.007371 – 0.00273 =<br>0.004644   | (0.004644)(185) =<br><b>0.859</b>                                |
| B       | (37)(4) = 148                       | 0.01126 – 0.003333 =<br>0.007928   | (0.007928)(148) =<br><b>1.173</b>                                |
| C       | (37)(4.5) = 166.5                   | 0.015015 – 0.005 =<br>0.010015     | (0.010015)(166.5) =<br><b>1.668</b>                              |

3.10

Table P3.10A: Production rates for A, B, and C

| Product | Yearly production | Production in 500 h |
|---------|-------------------|---------------------|
| A       | 150,000 kg        | 12,500 kg           |
| B       | 210,000 kg        | 17,500 kg           |
| C       | 360,000 kg        | 30,000 kg           |

Table P3.10B: Specific Reactor/Mixer Volumes for Processes A, B, and C

| Process                                  | A      | B      | C      |
|------------------------------------------|--------|--------|--------|
| $v_{react}$ (m <sup>3</sup> /kg-product) | 0.0073 | 0.0095 | 0.0047 |
| $t_{cycle}$ (h)                          | 6.0    | 9.5    | 18.5   |

Let the single-product campaign times for the three products be  $t_A$ ,  $t_B$ , and  $t_C$ , respectively.

Applying Equation (3.6), the following relationship is obtained:

$$t_A + t_B + t_C = 500 \quad (3.9)$$

The number of campaigns per product is then given by  $t_x/t_{cycle,x}$  and

$$\text{batch size (kg/batch)} = \frac{\text{production of } x}{t_x / t_{cycle,x}} \quad (3.10)$$

Furthermore, the volume of a batch is found by multiplying Equation (3.10) by  $v_{react,x}$ , and equating batch volumes for the different products yields:

$$\text{Volume of batch} = \frac{(\text{production of } x)(v_{react,x})}{t_x / t_{cycle,x}} \quad (3.11)$$

$$\frac{(12,500)(.0073)}{t_A / 6.0} = \frac{(17,500)(.0095)}{t_B / 9.5} = \frac{(30,000)(.0047)}{t_C / 18.5} \quad (3.12)$$

Solving Equations (3.9) and (3.12), yields:

$$t_A = 57.8 \text{ h}$$

$$t_B = 166.8 \text{ h}$$

$$t_C = 275.4 \text{ h}$$

$$v_{react,A} = v_{react,B} = v_{react,C} = 9.47 \text{ m}^3$$

$$\text{\#batches per campaign for product A} = t_A / 6.0 = 9.6$$

$$\text{\#batches per campaign for product B} = t_B / 9.5 = 17.6$$

$$\text{\#batches per campaign for product C} = t_C / 18.5 = 14.9$$

Clearly the number of batches should be an integer value. Rounding these numbers yields

For product A

Number of batches = 10

$$t_A = (10)(6.0) = 60 \text{ h}$$

$$V_A = (12,500)(0.0073)/(10) = 9.13 \text{ m}^3$$

For product B

Number of batches = 17

$$t_B = (17)(9.5) = 161.5 \text{ h}$$

$$V_B = (17,500)(0.0095)/(17) = 9.78 \text{ m}^3$$

For product C

Number of batches = 15

$$t_C = (15)(18.5) = 277.5 \text{ h}$$

$$V_C = (30,000)(0.0047)/(15) = 9.40 \text{ m}^3$$

Total time for production cycle = 499 h ~ 500 h

Volume of reactor = 9.78 m<sup>3</sup> (limiting condition for Product B)

3.11

Table P3.11: Batch step times (in hours) for Reactor and Bacteria Filter for Project 8 in Appendix B

| Product         | Reactor* | Precoating of Bacteria Filter | Filtration of Bacteria | Mass produced per batch, kg | Ratio of product, s |
|-----------------|----------|-------------------------------|------------------------|-----------------------------|---------------------|
| L-aspartic Acid | 40       | 25                            | 5                      | 1020                        | 1                   |
| L-phenylalanine | 70       | 25                            | 5                      | 716                         | 1.25                |

\*includes 5 h for filling, cleaning, and heating plus 5 hours for emptying

(a) let

$t_A$  = campaign time for L-aspartic acid

$t_P$  = campaign time for L-phenylalanine

Assuming equal recovery ratios for each amino acid we have

$$t_A + t_P = 8000$$

$$\frac{(716)(t_P)}{(70)} = 1.25 \frac{(1020)(t_A)}{(40)}$$

Solving we get

$$t_A = 1944 \text{ h}$$

$$t_P = 6056 \text{ h}$$

$$\text{yearly production of L-aspartic acid} = (1944)(1020)/(40) = 49,560 \text{ kg}$$

$$\text{yearly production of L-phenylalanine} = (6056)(716)/(70) = 61,950 \text{ kg}$$

$$\text{Number of batches per year for L-aspartic acid} = (1944)/(40) = 48$$

$$\text{Number of batches per year for L-phenylalanine} = (6056)/(70) = 86$$

(b) For each product calculate the average yearly demand and production rate in  $\text{m}^3/\text{h}$  and then find the storage needed for each product

| Rate                                                  | L-aspartic acid                 | L-phenylalanine                 |
|-------------------------------------------------------|---------------------------------|---------------------------------|
| Volume (m <sup>3</sup> ) of product required per year | (49,560)(0.9)/(1,200) = 37.17   | (61,950)(0.9)/(1,200) = 46.46   |
| Cycle time (h)                                        | 40                              | 70                              |
| Campaign time (h)                                     | 1944                            | 6056                            |
| Production rate, $r_p$ (m <sup>3</sup> /h)            | (37.17)/(1944) = 0.019125       | (46.46)/(6056) = 0.0076714      |
| Demand rate, $r_d$ (m <sup>3</sup> /h)                | (37.17)/(8,000) = 0.004646      | (46.46)/(8000) = 0.005808       |
| $r_p - r_d$ (m <sup>3</sup> /h)                       | 0.014479                        | 0.001864                        |
| Storage Volume (m <sup>3</sup> )                      | (0.014479)(1944) = <b>28.14</b> | (0.001864)(6056) = <b>11.29</b> |

(c) Rework part (b) using a 1 month cycle time = 8,000/12 = 666.67 h

Assuming equal recovery ratios for each amino acid we have

$$t_A + t_P = 666.67$$

$$\frac{(716)(t_P)}{(70)} = 1.25 \frac{(1020)(t_A)}{(40)}$$

Solving we get

$$t_A = 162 \text{ h and } t_P = 504.7 \text{ h}$$

$$\text{monthly production of L-aspartic acid} = (4)(1020) = 4,080 \text{ kg}$$

$$\text{monthly production of L-phenylalanine} = (7)(716) = 5,012 \text{ kg}$$

$$\text{Number of batches per month for L-aspartic acid} = (162)/(40) = 4$$

$$\text{Number of batches per month for L-phenylalanine} = (504.7)/(70) = 7$$

Note that these are rounded down so that integer numbers are given per month this gives rise to a slightly lower production rate per year than before.

| Rate                                                   | L-aspartic acid               | L-phenylalanine               |
|--------------------------------------------------------|-------------------------------|-------------------------------|
| Volume (m <sup>3</sup> ) of product required per month | (4,080)(0.9)/(1,200) = 3.06   | (5,012)(0.9)/(1,200) = 3.76   |
| Cycle time (h)                                         | 40                            | 70                            |
| Campaign time (h)                                      | 160                           | 490                           |
| Production rate, $r_p$ (m <sup>3</sup> /h)             | (3.06)/(160) = 0.019125       | (3.76)/(490) = 0.0076714      |
| Demand rate, $r_d$ (m <sup>3</sup> /h)                 | (3.06)/(666.7) = 0.00459      | (3.76)/(666.7) = 0.005638     |
| $r_p - r_d$ (m <sup>3</sup> /h)                        | 0.014535                      | 0.002033                      |
| Storage Volume (m <sup>3</sup> )                       | (0.014535)(160) = <b>2.33</b> | (0.002033)(490) = <b>1.00</b> |

These values are (not surprisingly) approximately 1/12 of the previous results



3. 12

- (a) Referring to Project B.8, Figures B.8.2 and B.8.3 and using batch reaction times for L-aspartic acid and L-phenylalanine of 25 and 55 h, respectively. We get the following information:

Conversion of L-aspartic acid = 42% (84% of equilibrium) (base case = 45%)

Exit concentration of L-phenylalanine = 18.5 kg/m<sup>3</sup> (base case = 21%)

| Product         | Reactor* | Precoating<br>of Bacteria<br>Filter | Filtration<br>of Bacteria | Mass produced<br>per batch, kg | Ratio of<br>products |
|-----------------|----------|-------------------------------------|---------------------------|--------------------------------|----------------------|
| L-aspartic Acid | 35       | 25                                  | 5                         | $(42/45)(1020) = 952$          | 1                    |
| L-phenylalanine | 65       | 25                                  | 5                         | $(18.5/21)(716) = 630.8$       | 1.25                 |

let

$t_A$  = campaign time for L-aspartic acid

$t_P$  = campaign time for L-phenylalanine

Assuming equal recovery ratios for each amino acid we have

$$t_A + t_P = 8000$$

$$\frac{(630.8)(t_P)}{(65)} = 1.25 \frac{(952)(t_A)}{(35)}$$

Solving we get

$$t_A = 1776 \text{ h}$$

$$t_P = 6224 \text{ h}$$

$$\text{yearly production of L-aspartic acid} = (1776)(952)/(35) = 48,316 \text{ kg}$$

$$\text{yearly production of L-phenylalanine} = (6224)(630.8)/(65) = 60,395 \text{ kg}$$

$$\text{Number of batches per year for L-aspartic acid} = (1776)/(35) = \mathbf{50 \text{ or } 51}$$

$$\text{Number of batches per year for L-phenylalanine} = (6224)/(65) = \mathbf{95 \text{ or } 96}$$

Therefore, the number of batches increases but the yearly production decreases

(b) Referring to Project B.8, Figures B.8.2 and B.8.3 and using batch reaction times for L-aspartic acid and L-phenylalanine of 35 and 65 h, respectively. We get the following information:

Conversion of L-aspartic acid = 47% (94% of equilibrium) (base case = 45%)

Exit concentration of L-phenylalanine = 21.5 kg/m<sup>3</sup> (base case = 21%)

| Product         | Reactor* | Precoating<br>of Bacteria<br>Filter | Filtration<br>of Bacteria | Mass produced<br>per batch, kg | Ratio of<br>products |
|-----------------|----------|-------------------------------------|---------------------------|--------------------------------|----------------------|
| L-aspartic Acid | 45       | 25                                  | 5                         | $(47/45)(1020) = 1065$         | 1                    |
| L-phenylalanine | 75       | 25                                  | 5                         | $(21.5/21)(716) = 733$         | 1.25                 |

let

$t_A$  = campaign time for L-aspartic acid

$t_P$  = campaign time for L-phenylalanine

Assuming equal recovery ratios for each amino acid we have

$$t_A + t_P = 8000$$

$$\frac{(733)(t_P)}{(75)} = 1.25 \frac{(1065)(t_A)}{(45)}$$

Solving we get

$$t_A = 1986 \text{ h}$$

$$t_P = 6014 \text{ h}$$

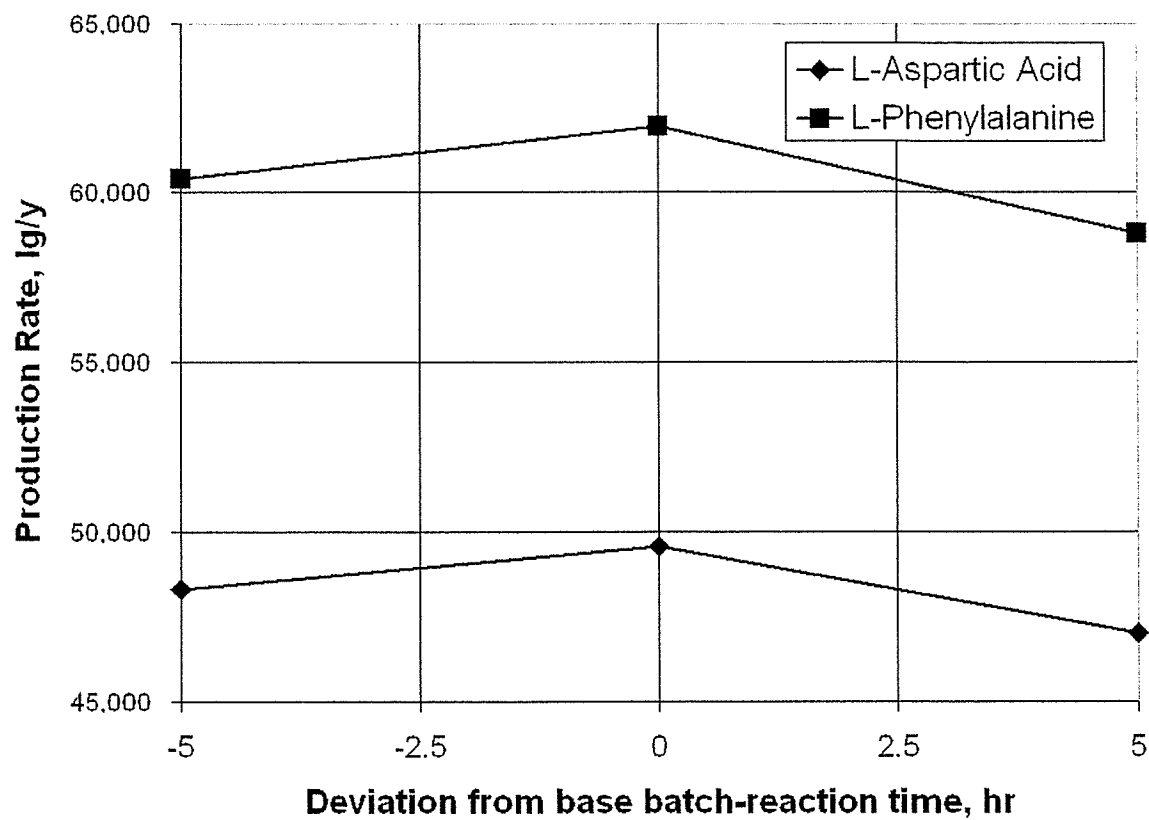
$$\text{yearly production of L-aspartic acid} = (1986)(1065)/(45) = 47,002 \text{ kg}$$

$$\text{yearly production of L-phenylalanine} = (6014)(733)/(75) = 58,777 \text{ kg}$$

$$\text{Number of batches per year for L-aspartic acid} = (1986)/(45) = 44$$

$$\text{Number of batches per year for L-phenylalanine} = (6014)/(75) = 80$$

Therefore, the number of batches decreases and the yearly production decreases – plot the results from problems 11 and 12



This shows that the base case conditions are close to optimal.

## Chapter 5

5.1 For ethylbenzene process in Figure B.2.1

Feeds: benzene, ethylene

Products: ethylbenzene, fuel gas (by-product)

5.2 For styrene process in Figure B.3.1

Feeds: ethylbenzene, steam

Products: styrene, benzene/toluene (by-products), hydrogen (by-product), wastewater (waste stream)

5.3 For drying oil process in Figure B.4.1

Feeds: acetylated castor oil

Products: acetic acid (by-product), drying oil, gum (waste stream)

5.4 For maleic anhydride process in Figure B.5.1

Feeds: benzene, air (note that dibutyl phthalate is not a feed stream)

Products: raw maleic anhydride (Stream 13), off gas (waste stream)

5.5 For ethylene oxide process in Figure B.6.1

Feeds: ethylene, air, process water

Products: fuel gas (by-product), light gases (waste stream), ethylene oxide, waste water (waste stream)

5.6 For formalin process in Figure B.7.1

Feeds: methanol, air, deionized water

Products: off-gas (waste - must be purified to use as a fuel gas), formalin

5.7 The main recycle streams for the styrene process in Figure B.3.1 are:  
ethylbenzene recycle (Stream 29) , reflux streams to T-401 and T-402

5.8 The main recycle streams for the drying oil process in Figure B.4.1 are:  
acetylated castor oil (Stream 14) , reflux streams to T-501 and T-502

5.9 The main recycle streams for the maleic anhydride process in Figure B.5.1 are:  
Dibutyl phthalate (Stream 14), circulating molten salt loop (Stream 15 and 16), and reflux to T-601 and T-602

5.10 Process description for ethylbenzene process in Figure B.2.1

Raw benzene (Stream 1), containing approximately 2% toluene, is supplied to the Benzene Feed Drum, V-301, from storage. Raw benzene and recycled benzene mix in the feed drum and then are pumped by the benzene feed pump, P-301A/B, to the feed heater, H-301, where the benzene is vaporized and heated to 400°C. The vaporized benzene is mixed with feed ethylene (containing 7 mol% ethane) to produce a stream at 383°C that is fed to the first of three reactors in series, R-301. The effluent from this reactor, depleted of ethylene, is mixed with additional feed ethylene and cooled in Reactor Intercooler, E-301, that raises high pressure steam. The cooled stream at 380°C is then fed to the second reactor, R-302, where further reaction takes place. The effluent from this reactor is mixed again with fresh ethylene feed and cooled to 380°C in Reactor Intercooler, E-302, where more high pressure steam is generated. The cooled stream, Stream 11, is fed to the third reactor, R-303. The effluent from R-303, containing significant amounts of unreacted benzene, Stream 12, is mixed with a recycle stream, Stream 13, and then fed to three heat exchangers, E-303 – 305, where the stream is cooled. The energy extracted from the stream is used to generate high- and low-pressure steam in E-303 and E-304, respectively. The final heat exchanger, E-305, cools the stream to 80°C using cooling water.

The cooled reactor effluent is then throttled down to a pressure of 110 kPa and sent to the Liquid Vapor Separator, V-302, where the vapor product is taken off and sent to the fuel gas header and the liquid stream is sent to column, T-301. The top product from T-301 consists of purified benzene that is recycled back to the benzene feed drum. The bottom product containing the ethylbenzene product plus diethylbenzene formed in an unwanted side reaction is fed to a second column, T-302. The top product from this column contains the 99.8 mol% ethylbenzene product. The bottom stream contains diethylbenzene and small amounts of ethylbenzene. This stream is recycled back through the feed heater, H-301, and is mixed with a small amount of recycled benzene to produce a stream at 500°C that is fed to a fourth reactor, R-304. This reactor converts the diethylbenzene back into ethylbenzene. The effluent from this reactor, Stream 13, is mixed with the effluent from reactor R-303.

#### 5.11 Process description of drying oil process in Figure B.4.1

Acetylated castor oil (ACO) is fed to the Recycle Mixing Vessel, V-501, where it is mixed with recycled ACO. This mixture is then pumped via P-501A/B to the Feed Fired Heater, H-501, where the temperature is raised to 380°C. The hot liquid stream, Stream 4, leaving the heater is then fed to the Drying Oil Reactor, R-501, that contains inert packing. The reactor provides residence time for the cracking reaction to take place. The two-phase mixture leaving the reactor is cooled in the Reactor Effluent Cooler, E-501, where low-pressure steam is generated. The liquid stream leaving the exchanger is at a temperature of 175°C and is passed through one of two filter vessels, V-502A/B, that removes any gum produced in the reactor. The filtered liquid, Stream 7, then flows to the ACO Recycle Tower, T-501. The bottom product from this tower contains purified ACO that is cooled in the Recycle Cooler, E-506, that raises low-pressure steam. This stream is then pumped via P-504A/B back to V-501 where it is mixed with fresh ACO. The overhead stream, Stream 9, from T-501 contains the drying oil and acetic acid produced from the cracking of ACO. This stream is fed to the Drying Oil Tower, T-502, where the ACO is taken as the bottom product and the acetic acid is taken as the top product. Both the acetic acid, Stream 11, and the ACO, Stream 12, are cooled (not shown in Figure B.4.1) and sent to storage.

## 5.12 Process description for ethylene oxide process in Figure B.6.1

Ethylene oxide (EO) is formed via the highly exothermic catalytic oxidation of ethylene using air. Feed air is compressed to a pressure of approximately 27 atm using a three stage centrifugal compressor, C-701-3, with intercoolers, E-701 and E-702. The compressed air stream is mixed with ethylene feed and the resulting stream, Stream 10, is further heated to a reaction temperature of 240°C in the Reactor Preheater, E-703. The reactor feed stream is then fed to the first of two reactors, R-701. The feed passes through a bank of catalyst filled tubes submerged in boiler feed water. The resulting exothermic reaction causes the boiler feed water (bfw) to vaporize and the pressure is maintained in the shell of the reactor to enable the production of medium pressure steam. Combustion of the ethylene and ethylene oxide also occur in the reactor. The reactor effluent is cooled in E-704 and is then recompressed to 30.15 bar in C-704 prior to being sent to the EO Absorber, T-701. The EO in the feed stream to the absorber, Stream 14, is scrubbed using water and the bottom product is sent to the EO column, T-703, for purification. The overhead stream from the absorber is heated back to 240°C prior to being fed to a second EO reactor, R-702 that performs the same function as R-701. The effluent from this reactor is cooled and compressed and sent to a second EO absorber, T-702, where the EO is scrubbed using water. The bottom product from this absorber is combined with the bottom product from the first absorber and the combined stream, Stream 29, is further cooled and throttled prior to being fed to the EO column, T-703. The overhead product from the second absorber is split with a purge stream being sent to fuel gas/incineration and the remainder being recycled to recover unused ethylene. The EO column separates the EO as a top product with waste water as the bottom product. The latter stream is sent off-site to water treatment while the EO product is sent to product storage. A small amount of non-condensables are present as dissolved gases in the feed and these accumulate in the overhead reflux drum, V-701, from where they are vented as an off gas.

## Chapter 6

### 6.1 Methods for setting pressure of a distillation column

- Set based on the pressure required to condense the overhead stream using cooling water (minimum of approx. 45°C condenser temperature)
- Set based on highest temperature of bottom product that avoids decomposition or reaction
- Set based on available highest hot utility for reboiler

### 6.2 Run a distillation column above ambient pressure because

- The components to be distilled have very high vapor pressures (very “light” components) and the temperatures at which they can be condensed at or below ambient pressure are  $< T_{\text{cooling water}}$ . Thus, we run the column at a high pressure to avoid the use of refrigeration in the condenser.
- The size of the column (diameter) reduces as the pressure decreases. Although the separation is usually more difficult at higher pressures, it may be more cost effective to run the column at or above atmospheric pressure.

Run a distillation column below ambient pressure because

- The components requiring separation have very low vapor pressures (very “heavy” components) and if distilled at ambient pressure would require excessively high temperatures to reboil.
- For thermally sensitive materials, e.g., foods, pharmaceuticals, and some organics, the bottom temperature in an atmospheric column might cause the material to decompose or react. For such systems, the pressure is reduced below atmospheric to a point where the bottom temperature does not cause thermal degradation of the product.

### 6.3 Run reactions at elevated temperature because

- The rate of reaction rate is faster. This results in a smaller reactor and/or higher single-pass conversion
- The reaction is endothermic and equilibrium limited and increasing the temperature shifts the equilibrium to the right.

Run reactions at elevated pressures because

- The reaction is gas phase and the concentration and hence rate is increased as the pressure is increased. This results in a smaller reactor and/or higher conversion.
- The reaction is equilibrium limited and there are fewer product moles than reactant moles. An increase in pressure shifts the equilibrium to the right.



6.4 Running a process above 250°C is undesirable because

- a. In order to heat the streams to that temperature the use of a fired heater is required, which is expensive.
- b. At temperatures in excess of 400°C process equipment may require more expensive materials of construction.

Examples for doing this for a reactor are: to increase the reaction rate or improve equilibrium for an endothermic reactor. An example for doing this for a distillation column is: to provide a vapor-liquid system for a heavy (high boiling point) component

6.5 A “condition of special concern” is a process condition that deviates from an “ideal” or “low-cost” operating condition. There are many examples given in this chapter. Operating at pressures outside the range of 1 – 10 atmospheres or temperatures outside of the range 45° – 250°/400°C are examples of conditions of special concern. Justification for operating at high temperatures and pressure might be to increase the rate of a desired reaction.

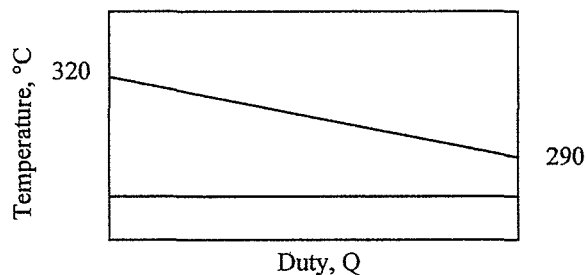
6.6 Many of these products are thermally labile meaning that they degrade at quite low temperatures. The use of vacuum conditions allows vapor-liquid equilibrium and vapor-solid equilibrium (freeze drying or lyophilization) to occur at temperatures below which thermal degradation occurs.

6.7 Distillation of a binary mixture, the effect of an increase in column pressure on:

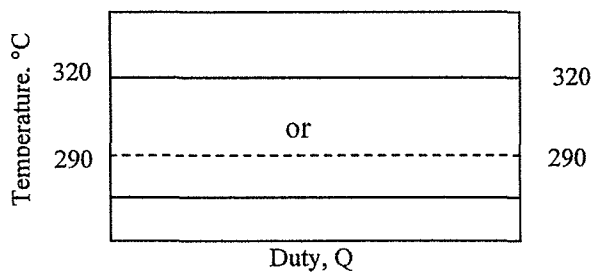
- a. The tendency to flood at a given reflux ratio will decrease because the density of vapor will increase and hence the superficial velocity in the tower will decrease thus moving away from flooding.
- b. For a given top and bottom purity with a fixed number of stages the tendency to flood will increase with pressure. This is because as pressure increases, the separation becomes more difficult and the equilibrium line moves closer to the xy line. The only way to compensate is therefore to increase the reflux ratio that in turn increases the internal flows in the column – hence the vapor flow and velocity will increase and move the column towards flooding.
- c. The number of stages will increase with pressure for the same reason given in (b) above. With  $x_D$ ,  $x_B$ , and  $R$  fixed as the separation gets more difficult, the number of stages must be increased.
- d. As pressure increases, the condenser temperature will increase – this is consistent with Antoine’s equation that as temperature increases so does the vapor pressure

6.8 Required reboiler utility at 290°C or higher. Assume that the exit temperature of utility is fixed at 290°C. Look at the T-Q diagram for the reboiler for both cases:

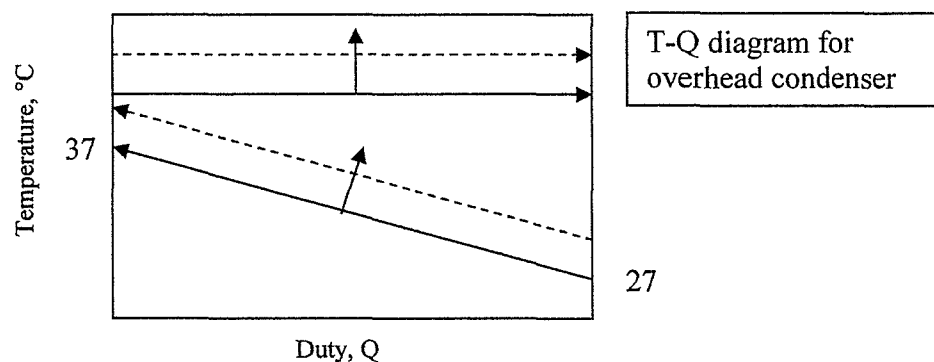
- a. Using high-pressure (42 bar) steam superheated to 320°C. Since hps condenses at 254°C, no condensation of the steam will occur and all heat transfer will be by cooling only. This will lead to a very large heat exchanger (because  $U$  will be limited by the low steam-side heat transfer coefficient) and a large flow of superheated steam.



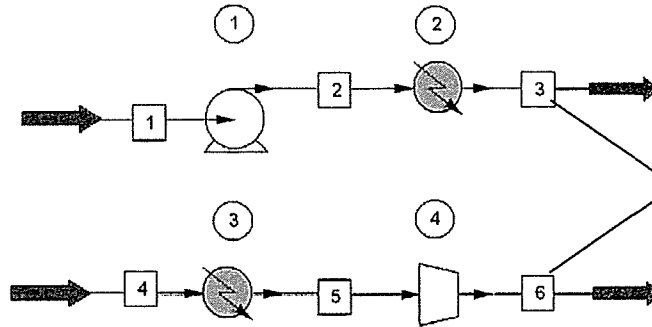
- b. Using saturated steam at 320°C requires a pressure of 112.7 bar. This is very high and would cause the exchanger to be very expensive and possibly requiring special materials of construction. However, the overall heat transfer coefficient  $U$  would be high and the exchanger area would be relatively small. Alternatively we could choose to throttle and desuperheat the steam to a saturation temperature of 290°C and a pressure of 74.4 bar which is still high but somewhat less costly.



- 6.9 If the column is designed to produce a saturated overhead product and reflux then a change in cooling water will affect the column pressure. For example, if the column is working at a pressure  $P_{winter}$  in the winter when the cooling water is available at a temperature of  $27^{\circ}\text{C}$  then as the cw temperature increases, the temperature driving force also drops and not as much vapor can be condensed. This is illustrated below. As less vapor is condensed, vapor will accumulate in the top of the column (because it cannot be condensed) and the pressure increases. As the pressure in the column increases, so does the temperature at which the vapor can condense (dew point). A new equilibrium will be reached when the temperature driving force in the condenser is restored to its original value – thus allowing the correct amount of vapor to be condensed. Thus the column is to some extent self regulating. This swing in pressure (higher in the summer and lower in the winter) occurs quite slowly and would be noticed as a slow drift of days or weeks.



- 6.10 Benzene at 1 atm and 25°C to be vaporized and pressurized to 10 atm and 250°C.
- Pumping a liquid requires less power than compressing a vapor so this suggests that pumping and heating will be better than heating and compressing a vapor.
  - Use a basis of 1,000 kg/h of benzene, simulation results (Chemcad – SRK) are given below



Pump Summary

|                                  |           |
|----------------------------------|-----------|
| Equip. No.                       | 1         |
| Name                             |           |
| Output pressure atm              | 10.3000   |
| Efficiency                       | 0.7500    |
| Calculated power kW              | 0.4001    |
| Calculated Pout atm              | 10.3000   |
| Head m                           | 110.0674  |
| Vol. flow rate m <sup>3</sup> /h | 1.1455    |
| Mass flow rate kg/h              | 1000.0000 |

Heat Exchanger Summary

|                     |          |          |
|---------------------|----------|----------|
| Equip. No.          | 2        | 3        |
| Name                |          |          |
| 1st Stream dp atm   | 0.3000   | 0.3000   |
| 1st Stream T Out °C | 250.0000 | 140.0000 |
| Calc Ht Duty MJ/h   | 741.1040 | 579.4749 |
| LMTD Corr Factor    | 1.0000   | 1.0000   |
| 1st Stream Pout atm | 10.0000  | 0.7000   |

Compressor Summary

|                            |         |
|----------------------------|---------|
| Equip. No.                 | 4       |
| Name                       |         |
| Pressure out atm           | 10.0000 |
| Type of Compressor         | 1       |
| Efficiency                 | 0.7500  |
| Actual power kW            | 45.3997 |
| Cp/Cv                      | 1.0813  |
| Theoretical power kW       | 34.0498 |
| Ideal Cp/Cv                | 1.0763  |
| Calc Pout atm              | 10.0000 |
| Calc. mass flowrate (kg/h) | 1000    |

#### Pump and Heat

$$\text{Cost} = (0.4)(0.06) + (741.1)(.015) = \underline{\$11.14/h}$$

#### Heat and Compress

$$\text{Cost} = (45.4)(0.06) + (579.5)(.015) = \underline{\$11.41/h}$$

Heat and Compression is slightly more expensive than pump and heat. The fact that the two answers are so close is in part due to the low cost of electricity relative to the heating utility used in this problem.

6.11 Production of high purity oxygen via cryogenic distillation.

- a. Normal Boiling Point of  $O_2 = 90.2\text{ K}$ , NBP of  $N_2 = 77.7\text{ K}$
- b. Assuming nearly pure compositions of products, nitrogen at the top ( $\sim 78\text{ K}$ ) and oxygen at the bottom ( $\sim 90\text{ K}$ )
- c. Critical temperatures of  $O_2$  and  $N_2$  are  $155\text{ K}$  and  $126\text{ K}$ , respectively. Thus neither component can be liquefied at  $40^\circ\text{C}$
- d. Obviously, distillation at ambient temperature is impossible because a 2-phase mixture cannot be produced at  $40^\circ\text{C}$ . If distillation is used then it must occur at cryogenic temperatures. Typically these units are run at about 5-6 atm pressure which gives top and bottom temperatures about  $20\text{ K}$  above those in part (a).

6.12 Since the synthesis reaction to produce ammonia is highly exothermic, a high temperature tends to push the reaction to the left (undesirable). The reason for the high temperature must be to increase the kinetics rather than improve the equilibrium conversion (the iron catalyst is only effective above temperatures of about  $400^\circ\text{C}$ ). Since there are fewer product moles than reactant moles, a high pressure pushes the equilibrium to the right (desirable) and also increases the concentration of all species, which in turn increases the kinetic rates.

6.13 The conversion is limited by equilibrium and so it would be increased by using lower temperature and higher pressure. As pointed out in Problem 12, the lower temperature slows the reactions and would lead to much larger and expensive reactors (that are already expensive because of the high pressure and relatively high temperature). Higher pressures could be used but again this will lead to increased costs. Another alternative is to remove ammonia during or after each reactor (using several reactor stages in series) this will significantly increase the single pass conversion per reactor but require a significant amount of additional, cooling, reheating, and separation equipment.

## 6.14 Drying Oil Process

| Equipment | Reactors and Separators<br>Tables 6.1 – 6.3 |          |           |          |                  | Other Equipment<br>Table 6.4 |      |        |       |     |
|-----------|---------------------------------------------|----------|-----------|----------|------------------|------------------------------|------|--------|-------|-----|
|           | High Temp                                   | Low Temp | High Pres | Low Pres | Non-Stoich. Feed | Compr                        | Exch | Heater | Valve | Mix |
| E-501     |                                             |          |           |          |                  |                              |      |        |       |     |
| E-502     |                                             |          |           |          |                  |                              |      |        |       |     |
| E-503     |                                             |          |           |          |                  |                              |      |        |       |     |
| E-504     |                                             |          |           |          |                  |                              |      |        |       |     |
| E-505     |                                             |          |           |          |                  |                              |      |        |       |     |
| E-506     |                                             |          |           |          |                  |                              |      |        |       |     |
| H-501     |                                             |          |           |          |                  |                              |      |        |       |     |
| P-501     |                                             |          |           |          |                  |                              |      |        |       |     |
| P-502     |                                             |          |           |          |                  |                              |      |        |       |     |
| P-503     |                                             |          |           |          |                  |                              |      |        |       |     |
| P-504     |                                             |          |           |          |                  |                              |      |        |       |     |
| R-501     | X                                           |          |           |          |                  |                              |      |        |       |     |
| T-501     | X                                           |          |           |          |                  |                              |      |        |       |     |
| T-502     |                                             |          |           |          |                  |                              |      |        |       |     |
|           |                                             |          |           |          |                  |                              |      |        |       |     |
|           |                                             |          |           |          |                  |                              |      |        |       |     |
|           |                                             |          |           |          |                  |                              |      |        |       |     |

- PCM is shown above
- High temperature in R-501 – need high temperature to initiate cracking reactions. High temperature in T-501 – heavy components (ACO and drying oil) need high temperature to form a two-phase mixture.
- Remedy for high temperature in R-501 – possibly find a catalyst that would promote the cracking reaction at a lower temperature. Note that pressure is not a variable since the cracking reaction occurs in the liquid phase.  
Remedy for the high temperature in T-501 – this may be a problem since reaction (and gum formation) may be occurring in reboiler and would cause plugging and fouling of exchanger surfaces and trays/packing. One remedy would be to operate the tower at vacuum to lower the bottom temperature. For example, at 30 kPa the bottom temperature would drop to about 300°C.

## 6.15 Styrene process

### a. PCM

| Equipment | Reactors and Separators<br>Tables 6.1 – 6.3 |          |           |          |                   | Other Equipment<br>Table 6.4 |      |        |       |     |
|-----------|---------------------------------------------|----------|-----------|----------|-------------------|------------------------------|------|--------|-------|-----|
|           | High Temp                                   | Low Temp | High Pres | Low Pres | Non-Stoich . Feed | Compr                        | Exch | Heater | Valve | Mix |
| C-401     |                                             |          |           |          |                   | X                            |      |        |       |     |
| E-401     |                                             |          |           |          |                   |                              |      |        |       |     |
| E-402     |                                             |          |           |          |                   |                              |      |        |       |     |
| E-403     |                                             |          |           |          |                   |                              | X    |        |       |     |
| E-404     |                                             |          |           |          |                   |                              | X    |        |       |     |
| E-405     |                                             |          |           |          |                   |                              |      |        |       |     |
| E-406     |                                             |          |           |          |                   |                              |      |        |       |     |
| E-407     |                                             |          |           |          |                   |                              |      |        |       |     |
| E-408     |                                             |          |           |          |                   |                              |      |        |       |     |
| E-409     |                                             |          |           |          |                   |                              |      |        |       |     |
| H-401     |                                             |          |           |          |                   |                              |      |        |       |     |
| P-401     |                                             |          |           |          |                   |                              |      |        |       |     |
| P-402     |                                             |          |           |          |                   |                              |      |        |       |     |
| P-403     |                                             |          |           |          |                   |                              |      |        |       |     |
| P-404     |                                             |          |           |          |                   |                              |      |        |       |     |
| P-405     |                                             |          |           |          |                   |                              |      |        |       |     |
| P-406     |                                             |          |           |          |                   |                              |      |        |       |     |
| R-401     | X                                           |          |           |          | X                 |                              |      |        |       |     |
| R-402     | X                                           |          |           |          | X                 |                              |      |        |       |     |
| T-401     |                                             |          |           |          |                   |                              |      |        |       |     |
| T-402     |                                             |          |           |          |                   |                              |      |        |       |     |
| V-401     |                                             |          |           |          |                   |                              |      |        |       |     |
| V-402     |                                             |          |           |          |                   |                              |      |        |       |     |
| V-403     |                                             |          |           |          |                   |                              |      |        |       |     |

- b. High temperatures in R-401 and R-402 – the desired reaction is slightly endothermic and may be equilibrium limited. Therefore, the high temperature may be required to push the equilibrium to the right and/or increase the reaction rate.

Non-stoichiometric feed to R-401 and R-402 – high temperature steam is added to the reactor feed. Steam is not required as a reactant; its purpose is to push the equilibrium to the right by diluting the reaction mixture.

The pressure ratio for C-401 is slightly greater than 3 – since the pressure ratio is so close to 3, it is probably not worth the cost of adding a second stage with intercooling.

The  $\Delta T_{lm}$  for E-403 and E-404 are both greater than 100°C – shows low heat integration but represents a simple, low-cost configuration.

- c. A possible remedy to using such high temperatures in the reactors is to use a lower pressure but since the pressure is already quite low (170 kPa) this would lead to larger reactors and possibly vacuum operations.

The use of steam as a diluent in the reactors, improves equilibrium conversion. If steam is not added, then the reaction would be pushed back to the left – not any viable alternatives to this (higher T and lower P –because MOC or vacuum problems, using say nitrogen as the diluent would mean producing a contaminated hydrogen stream.)

The excessive compression ratio in C-401 could be investigated by looking at a 2<sup>nd</sup> stage with intercooling.

The high  $\Delta T_{lm}$  for E-403 and E-404 could be eliminated with better heat integration.



## Chapter 7

- 7.1 (i) Capacity or size (for heat exchanger this would be heat exchanger area)  
(ii) Operating (or more correctly the design) pressure  
(iii) Materials of construction
- 7.2 CEPCI is used to adjust purchased costs of equipment for different times. It is a measure for the inflation of costs associated with the manufacture of chemical process equipment.
- 7.3 Total module cost represents the all costs associated with the purchase and installation of new equipment for an existing chemical facility. The grass roots cost includes the total module cost plus costs associated with the off sites and utilities needed for a completely new “grass roots” or “green field” facility.
- 7.4 Use a cost exponent of 0.6 to estimate the change in cost associated with a modest change in capacity for a whole chemical process.
- 7.5 The economy of scale refers to the fact that the cost exponent for chemical plant equipment is (usually) less than one. Therefore, as a chemical plant’s capacity increases, the unit cost of equipment (\$/unit of production) decreases.
- 7.6 A Lang factor is a constant (between approximately 3 and 5) that when multiplied by the purchase cost of the equipment gives an estimate of the total installed cost (capital investment) for a chemical process.
- 7.7 Most of the cost of a heat exchanger involves machining and tube costs. The relative change in these costs for an increase in pressure is much smaller than for a process vessel whose purchase price is directly affected by wall thickness and hence operating pressure.

7.8 Actual Cost = \$540 million

For a Class 1 estimate the expected range of accuracy is +6% to -4%. Thus the range of expected cost estimates would be

$$\frac{540}{(1+0.06)} \text{ to } \frac{540}{(1-0.04)} \Rightarrow \boxed{\$509.4 \text{ to } \$562.5 \text{ million}}$$

For a Class 3 estimate, the range of accuracy is 2 to 6 times that of a Class 1 estimate. Use a mid-point of 4 times the accuracy. Thus the range of expected cost estimates would be

$$\frac{540}{(1+(0.06)(4))} \text{ to } \frac{540}{(1-(0.04)(4))} \Rightarrow \boxed{\$435.5 \text{ to } \$642.9 \text{ million}}$$

Similarly, for a Class 5 estimate use a mid-point value of  $\frac{(4+20)}{2}=12$ . Range of estimates

$$\frac{540}{(1+(0.06)(12))} \text{ to } \frac{540}{(1-(0.04)(12))} \Rightarrow \boxed{\$314.0 \text{ to } \$1,038 \text{ million}}$$

7.9 The figures in Appendix A are plotted with the y-axis as the purchased cost per unit of capacity. For a cost exponent of <1 the slope of the curves should be negative. Searching are the figures in Appendix A we find only two instances of positive slopes. These are

Figure A.1 – Rotary compressor  
Figure A.9 – Auto batch separator

These are the only equipment that does not exhibit the “economy of scale” over the whole range of equipment sizes.

7.10

| Year | Capacity (m <sup>3</sup> ) | Cost (£) | CEPCI | Exchange Rate |
|------|----------------------------|----------|-------|---------------|
| 1993 | 75                         | 7,800    | 359   | \$1.40/£      |
| 1998 | 155                        | 13,800   | 390   | \$1.65/£      |
| 2007 | 120                        | ?        | 500   | \$2.00/£      |

$$C_2 = C_1 \left( \frac{I_2}{I_1} \right); \frac{C_a}{C_b} = \left( \frac{A_a}{A_b} \right)^n$$

$$\text{For } 75 \text{ m}^3: C_{2007} = \left( \frac{500}{359} \right) (£7,800) = £10,864$$

$$\text{For } 155 \text{ m}^3: C_{2007} = \left( \frac{500}{390} \right) (£13,800) = £17,692$$

$$n = \frac{\log \left( \frac{10,864}{17,692} \right)}{\log \left( \frac{75 \text{ m}^3}{155 \text{ m}^3} \right)} = 0.6719$$

$$C_{120}^{2007} = \left( \frac{120 \text{ m}^3}{75 \text{ m}^3} \right)^{0.6719} (£10,864) = £14,897 = ($2.00/£)(£14,897) = \boxed{\$29,795}$$

7.11

| Year | Capacity (10 <sup>3</sup> gal) | Purchased Cost (\$) | CEPCI |
|------|--------------------------------|---------------------|-------|
| 1978 | ?                              | 35,400              | 219   |
| 1988 | 105                            | 45,300              | 343   |
| 1998 | 85                             | 45,500              | 390   |

$$C_2 = C_1 \left( \frac{I_2}{I_1} \right); \quad \frac{C_a}{C_b} = \left( \frac{A_a}{A_b} \right)^n$$

$$\text{For 105 gal: } C_{1978} = \left( \frac{219}{343} \right) (\$45,300) = \$28,923$$

$$\text{For 85 gal: } C_{1978} = \left( \frac{219}{390} \right) (\$45,500) = \$25,550$$

$$n = \frac{\log \left( \frac{\$28,923}{\$25,550} \right)}{\log \left( \frac{105 \text{ gal}}{85 \text{ gal}} \right)} = 0.587$$

$$\frac{\$35,400}{\$28,923} = \left( \frac{X}{105} \right)^{0.587} \Rightarrow X = 105 \left( \frac{\$35,400}{\$28,923} \right)^{1/0.587}$$

$$X = 148.2 \text{ gal}$$

7.12

| Year | Capacity (kW) | Purchased Cost (10 <sup>3</sup> \$) | CEPCI |
|------|---------------|-------------------------------------|-------|
| ?    | 1,000         | 645.93                              | ?     |
| 2000 | 500           | 500.00                              | 394   |
| 2007 | 775           | 811.68                              | 500   |

$$C_2 = C_1 \left( \frac{I_2}{I_1} \right)^n; \quad \frac{C_a}{C_b} = \left( \frac{A_a}{A_b} \right)^n$$

In 1986 CEPCI was 318 – use this information to estimate year.

$$\text{For 500 kW: } C_{1986} = \left( \frac{318}{394} \right) (\$500) = \$403.55$$

$$\text{For 775 kW: } C_{1986} = \left( \frac{318}{500} \right) (\$811.68) = \$516.23$$

$$n = \frac{\log \left( \frac{\$403.55}{\$516.23} \right)}{\log \left( \frac{500 \text{ kW}}{775 \text{ kW}} \right)} = 0.5619$$

$$\text{For 1,000 kW: } C_{1986} = (\$403.55) \left( \frac{1,000 \text{ kW}}{500 \text{ kW}} \right)^{0.5619} = \$595.73$$

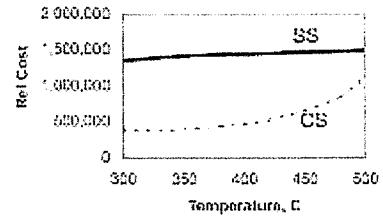
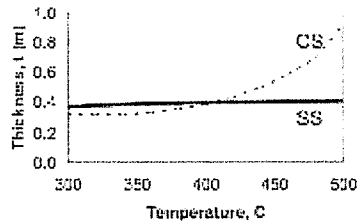
$$CEPCI_{\text{year}} = (318) \left( \frac{\$645.93}{\$595.73} \right) = 344.8$$

CEPCI of 344.8 corresponds to about 1988. The compressor was not purchased in 1976 when the CEPCI was approximately 192.

# 7.13 and 7.14

$$t = \frac{PD}{2SE - 1.2P} + CA = \frac{(150\text{bar})(3.2\text{m})}{2(0.9)S - 1.2(150\text{bar})} + 0.00635$$

Height (H) = 15 m  
 Diameter (D) = 3.2 m  
 P = 150 bar  
 E = 0.9  
 CA = 0.00635 m  
 t = PD/(2SE-1.2P)+CA  
 Weight (W) = πDHtp  
 density, ρ = 8000 kg/m<sup>3</sup>



ASME SA515 - Grade 55 Carbon Steel

| Temperature<br>°C | S<br>bar | t<br>m | Weight<br>kg | Relative cost<br>\$ |
|-------------------|----------|--------|--------------|---------------------|
| 300               | 950      | 0.3201 | 386,130      | 386,130             |
| 320               | 950      | 0.3201 | 386,130      | 386,130             |
| 340               | 950      | 0.3201 | 386,130      | 386,130             |
| 360               | 920      | 0.3316 | 399,976      | 399,976             |
| 380               | 860      | 0.3572 | 430,949      | 430,949             |
| 400               | 800      | 0.3873 | 467,231      | 467,231             |
| 420               | 720      | 0.4365 | 526,530      | 526,530             |
| 440               | 640      | 0.5002 | 603,400      | 603,400             |
| 460               | 560      | 0.5861 | 707,006      | 707,006             |
| 480               | 485      | 0.6890 | 843,243      | 843,243             |
| 500               | 390      | 0.9259 | 1,116,969    | 1,116,969           |

ASME SA-240 - Grade 316 Stainless Steel

| Temperature<br>°C | S<br>bar | t<br>m | Weight<br>kg | Relative cost<br>\$ |
|-------------------|----------|--------|--------------|---------------------|
| 300               | 930      | 0.3716 | 448,344      | 1,345,032           |
| 320               | 915      | 0.3793 | 457,589      | 1,372,768           |
| 340               | 900      | 0.3873 | 467,231      | 1,401,692           |
| 360               | 790      | 0.3928 | 473,891      | 1,421,673           |
| 380               | 785      | 0.3956 | 477,294      | 1,431,882           |
| 400               | 780      | 0.3985 | 480,747      | 1,442,242           |
| 420               | 776      | 0.4008 | 483,547      | 1,450,640           |
| 440               | 772      | 0.4032 | 486,379      | 1,459,130           |
| 460               | 768      | 0.4056 | 489,246      | 1,467,730           |
| 480               | 764      | 0.4080 | 492,147      | 1,476,441           |
| 500               | 760      | 0.4104 | 495,083      | 1,485,250           |

For this temperature range carbon steel is always the best (cheapest) choice of material of construction (MOC).

## 7.15 – 17.19

|                |             |     |
|----------------|-------------|-----|
| Add Equipment  | Unit Number | 100 |
| Edit Equipment | CEPCI       | 500 |

### User Added Equipment

| Exchangers          | Exchanger Type | Shell Pressure (bars) | Tube Pressure (bars) | MOC                               | Area (square meters) | Purchased Equipment Cost | Bare Module Cost |
|---------------------|----------------|-----------------------|----------------------|-----------------------------------|----------------------|--------------------------|------------------|
| Problem 15 → E-101  | Floating Head  | 5                     | 30                   | Stainless Steel / Carbon Steel    | 160                  | \$ 40,100                | \$ 191,000       |
| Problem 16 → E-102  | Floating Head  | 30                    | 5                    | Stainless Steel / Stainless Steel | 160                  | \$ 40,100                | \$ 270,000       |
| Problem 17a → E-103 | Floating Head  | 20                    | 20                   | Carbon Steel / Carbon Steel       | 400                  | \$ 128,000               | \$ 435,000       |
| Problem 17b → E-104 | Floating Head  | 20                    | 20                   | Carbon Steel / Carbon Steel       | 400                  | \$ 74,900                | \$ 256,000       |

| Towers                      | Tower Description                    | Height (meters) | Diameter (meters) | Tower MOC       | Diameter MOC | Pressure (bars) | Purchased Equipment Cost | Bare Module Cost |
|-----------------------------|--------------------------------------|-----------------|-------------------|-----------------|--------------|-----------------|--------------------------|------------------|
| Used for problems 18 and 19 | T-101 20 Stainless Steel Valve Trays | 14              | 2                 | Carbon Steel    |              | 1               | \$ 157,000               | \$ 381,000       |
|                             | T-102 20 Stainless Steel Valve Trays | 14              | 2                 | Stainless Steel |              | 1               | \$ 157,000               | \$ 543,000       |
|                             | T-103 20 Stainless Steel Valve       | 14              | 2                 | Titanium        |              | 1               | \$ 157,000               | \$ 1,020,000     |

### Problem 7.15 and 7.16

Heuristics are consistent with Table 11.11, heuristic 3 – namely the tube side is for corrosive, fouling, scaling, and high pressure fluids. Based on this, it is best to keep the high pressure corrosive fluid in the tubes as given in Problem 15.

This saves approximately \$270k – \$191k = \$79,000

### Problem 7.17

(c) The results clearly indicate the principles of the economy of scale. Much cheaper to use a single shell than multiple shells (\$256k for single shell vs. \$435k for 4 shells) – of course, this is not always possible!

### Problem 7.18

For Carbon Steel,  $C_{BM} = \$381,000$

### Problem 7.19

- (a) 10% nitric acid solution – From Table 7.9 – use stainless steel  $C_{BM} = \$543,000$
- (b) 50% sodium hydroxide solution – From Table 7.9 – use carbon steel  $C_{BM} = \$381,000$
- (c) 10% sulfuric acid solution – From Table 7.9 – use Titanium  $C_{BM} = \$1,020,000$
- (d) 98% sulfuric acid solution – use carbon steel because there is not enough water for ionization  $C_{BM} = \$543,000$  when concentration drops to below about 90% need to use graphite or glass lining

# 7.20 – Toluene Hydrodealkylation Process – Unit 100

|    | A                           | B                              | C                      | D                 | E                                 | F                         | G               | H                      | I                |
|----|-----------------------------|--------------------------------|------------------------|-------------------|-----------------------------------|---------------------------|-----------------|------------------------|------------------|
| 2  |                             | Add Equipment                  |                        | Unit Number       | 100                               |                           |                 |                        |                  |
| 3  |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 4  |                             | Edit Equipment                 |                        | CEPCI             | 500                               |                           |                 |                        |                  |
| 5  |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 6  |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 7  | <b>User Added Equipment</b> |                                |                        |                   |                                   |                           |                 |                        |                  |
| 14 | Compressors                 | Compressor Type                | Power (kilowatts)      | # Spares          | MOC                               |                           |                 | Equipment Cost         | Bare Module Cost |
| 15 | C-101                       | Centrifugal                    | 49.1                   | 0                 | Carbon Steel                      |                           |                 | \$ 189,000             | \$ 517,000       |
| 16 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 17 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 24 | Drives                      | Drive Type                     | Power (kilowatts)      | # Spares          |                                   |                           |                 | Equipment Cost         | Bare Module Cost |
| 25 | D-101                       | Electric - Explosion Proof     | 49.1                   | 1                 |                                   |                           |                 | \$ 77,700              | \$ 117,000       |
| 26 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 27 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 37 | Exchangers                  | Exchanger Type                 | Shell Pressure (barg)  | Pressure (barg)   | MOC                               | (square meters)           |                 | Equipment Cost         | Bare Module Cost |
| 38 | E-101                       | Floating Head                  | 25                     | 41                | Stainless Steel / Stainless Steel | 36                        |                 | \$ 23,900              | \$ 166,000       |
| 39 | E-102                       | Floating Head                  | 23                     | 2                 | Stainless Steel / Stainless Steel | 763                       |                 | \$ 134,000             | \$ 878,000       |
| 40 | E-103                       | Multiple Pipe                  |                        | 5                 | Carbon Steel / Carbon Steel       | 11                        |                 | \$ 5,110               | \$ 16,800        |
| 41 | E-104                       | Floating Head                  | 2                      | 2                 | Carbon Steel / Carbon Steel       | 35                        |                 | \$ 23,800              | \$ 78,300        |
| 42 | E-105                       | Multiple Pipe                  |                        | 2                 | Carbon Steel / Carbon Steel       | 12                        |                 | \$ 5,530               | \$ 18,200        |
| 43 | E-106                       | Floating Head                  | 10                     | 2                 | Carbon Steel / Carbon Steel       | 80                        |                 | \$ 29,200              | \$ 97,000        |
| 44 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 45 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 52 | Fired Heaters               | Type                           | Heat Duty (MM/h)       | Superheat (°C)    | MOC                               | Pressure (barg)           |                 | Equipment Cost         | Bare Module Cost |
| 53 | H-101                       | Process Heater                 | 27000                  |                   | Stainless Steel                   | 25                        |                 | \$ 942,000             | \$ 2,670,000     |
| 54 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 55 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 59 | (with drives)               | Pump Type                      | Power (kilowatts)      | # Spares          | MOC                               | Discharge Pressure (barg) |                 | Equipment Cost         | Bare Module Cost |
| 60 | P-101                       | Reciprocating                  | 14.2                   | 1                 | Carbon Steel                      | 26                        |                 | \$ 62,700              | \$ 272,000       |
| 61 | P-102                       | Centrifugal                    | 3.2                    | 1                 | Carbon Steel                      | 3.4                       |                 | \$ 7,190               | \$ 28,600        |
| 62 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 63 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 73 | Towers                      | Tower Description              | Height (meters)        | Diameter (meters) | Tower MOC                         | Demister MOC              | Pressure (barg) | Equipment Cost         | Bare Module Cost |
| 74 | T-101                       | 42 Stainless Steel Sieve Trays | 29                     | 1.5               | Carbon Steel                      |                           | 2               | \$ 119,000             | \$ 324,000       |
| 75 | T-102                       | 10 meters of 304 Stainless     | 14.2                   | 2.3               | Stainless Steel                   |                           | 24              | \$ 152,000             | \$ 2,030,000     |
| 76 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 77 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 81 | Vessels                     | Orientation                    | Length/Height (meters) | Diameter (meters) | MOC                               | Demister MOC              | Pressure (barg) | Equipment Cost         | Bare Module Cost |
| 82 | V-101                       | Horizontal                     | 5.9                    | 1.9               | Carbon Steel                      |                           | 1               | \$ 17,900              | \$ 54,000        |
| 83 | V-102                       | Vertical                       | 3.5                    | 1.1               | Carbon Steel                      |                           | 23              | \$ 7,260               | \$ 56,100        |
| 84 | V-103                       | Vertical                       | 3.5                    | 1.1               | Carbon Steel                      |                           | 1.5             | \$ 7,260               | \$ 29,500        |
| 85 | V-104                       | Horizontal                     | 3.9                    | 1.3               | Carbon Steel                      |                           | 1.5             | \$ 9,380               | \$ 28,200        |
| 86 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
| 87 |                             |                                |                        |                   |                                   |                           |                 |                        |                  |
|    |                             |                                |                        |                   |                                   |                           |                 | Total Bare Module Cost | \$ 7,380,700     |



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Unit Number 300

CEPCI 500

| Exchangers    | Exchanger Type                 | Shell Pressure (barg)  | Pressure (barg)   | MOC                               | (square meters)           | Equipment Cost  | Bare Module Cost |                  |
|---------------|--------------------------------|------------------------|-------------------|-----------------------------------|---------------------------|-----------------|------------------|------------------|
| E-301         | Floating Head                  | 41                     | 21                | Stainless Steel / Stainless Steel | 62.6                      | \$ 27,000       | \$ 187,000       |                  |
| E-302         | Floating Head                  | 41                     | 21                | Stainless Steel / Stainless Steel | 80.1                      | \$ 29,200       | \$ 203,000       |                  |
| E-303         | Floating Head                  | 41                     | 21                | Stainless Steel / Stainless Steel | 546                       | \$ 97,600       | \$ 679,000       |                  |
| E-304         | Fixed, Sheet, or U-Tube        | 5                      | 21                | Carbon Steel / Carbon Steel       | 1570                      | \$ 165,000      | \$ 551,000       |                  |
| E-305         | Floating Head                  | 5                      | 21                | Carbon Steel / Carbon Steel       | 348                       | \$ 67,100       | \$ 224,000       |                  |
| E-306         | Fixed, Sheet, or U-Tube        | 21                     | 21                | Carbon Steel / Carbon Steel       | 57.8                      | \$ 25,000       | \$ 85,600        |                  |
| E-307         | Floating Head                  | 2                      | 5                 | Carbon Steel / Carbon Steel       | 54.6                      | \$ 26,000       | \$ 85,400        |                  |
| E-308         | Fixed, Sheet, or U-Tube        | 2                      | 41                | Carbon Steel / Carbon Steel       | 22.6                      | \$ 20,600       | \$ 69,500        |                  |
| E-309         | Floating Head                  | 2                      | 5                 | Carbon Steel / Carbon Steel       | 17.5                      | \$ 23,200       | \$ 76,300        |                  |
| Fired Heaters | Type                           | Heat Duty (MW/h)       | Superheat (°C)    | MOC                               | Pressure (barg)           | Equipment Cost  | Bare Module Cost |                  |
| H-301         | Process Heater                 | 35000                  |                   | Stainless Steel                   | 22                        | \$ 1,050,000    | \$ 2,960,000     |                  |
| (with drives) | Pump Type                      | Power (kilowatts)      | # Spares          | MOC                               | Discharge Pressure (barg) | Equipment Cost  | Bare Module Cost |                  |
| P-301         | Positive Displacement          | 15                     | 1                 | Carbon Steel                      | 19                        | \$ 17,200       | \$ 70,500        |                  |
| P-302         | Centrifugal                    | 1                      | 1                 | Carbon Steel                      | 1                         | \$ 6,170        | \$ 24,600        |                  |
| P-303         | Centrifugal                    | 1                      | 1                 | Carbon Steel                      | 1                         | \$ 6,170        | \$ 24,600        |                  |
| P-304         | Centrifugal                    | 1.4                    | 1                 | Carbon Steel                      | 19                        | \$ 6,330        | \$ 29,000        |                  |
| P-305         | Positive Displacement          | 2.7                    | 1                 | Carbon Steel                      | 19                        | \$ 9,190        | \$ 37,600        |                  |
| Towers        | Tower Description              | Height (meters)        | Diameter (meters) | Tower MOC                         | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| T-301         | 45 Stainless Steel Sieve Trays | 27.5                   | 1.7               | Carbon Steel                      |                           | 3               | \$ 147,000       | \$ 408,000       |
| T-302         | 76 Stainless Steel Sieve Trays | 29                     | 1.5               | Carbon Steel                      |                           | 3               | \$ 177,000       | \$ 435,000       |
| Vessels       | Orientation                    | Length/Height (meters) | Diameter (meters) | MOC                               | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| V-301         | Horizontal                     | 6.06                   | 1.21              | Carbon Steel                      |                           | 1.5             | \$ 10,900        | \$ 32,900        |
| V-302         | Vertical                       | 4.86                   | 1.62              | Carbon Steel                      |                           | 1.5             | \$ 14,300        | \$ 58,000        |
| V-303         | Horizontal                     | 4.45                   | 1.48              | Carbon Steel                      |                           | 2               | \$ 11,500        | \$ 34,600        |
| V-304         | Horizontal                     | 4.14                   | 1.38              | Carbon Steel                      |                           | 2               | \$ 10,300        | \$ 31,000        |
| V-305         | Vertical                       | 11                     | 1.72              | Stainles Steel                    |                           | 21              | \$ 27,600        | \$ 702,000       |
| V-306         | Vertical                       | 12                     | 1.85              | Stainles Steel                    |                           | 21              | \$ 33,000        | \$ 889,000       |
| V-307         | Vertical                       | 12                     | 1.97              | Stainles Steel                    |                           | 21              | \$ 36,400        | \$ 1,030,000     |
| V-308         | Vertical                       | 5                      | 0.95              | Stainles Steel                    |                           | 21              | \$ 7,520         | \$ 123,000       |

Reactors R - 301-304

## 7.22 – Styrene Process – Unit 400 - Project B.3

|    |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
|----|-----------------------------|--------------------------------|-----------------------|-------------------|-----------------------------------|---------------------------|-----------------|------------------|------------------|
| 1  |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 2  | Add Equipment               | Unit Number                    | 400                   |                   |                                   |                           |                 |                  |                  |
| 3  |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 4  | Edit Equipment              | CEPCI                          | 500                   |                   |                                   |                           |                 |                  |                  |
| 5  |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 6  |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 7  | <b>User Added Equipment</b> |                                |                       |                   |                                   |                           |                 |                  |                  |
| 14 | Compressors                 | Compressor Type                | Power (kilowatts)     | # Spares          | MOC                               |                           | Equipment Cost  | Bare Module Cost |                  |
| 15 | C-401                       | Centrifugal                    | 364                   | 0                 | Carbon Steel                      |                           | \$ 189,000      | \$ 517,000       |                  |
| 16 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 17 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 24 | Drives                      | Drive Type                     | Power (kilowatts)     | # Spares          |                                   |                           | Equipment Cost  | Bare Module Cost |                  |
| 25 | D-401                       | Electric - Explosion Proof     | 380                   | 1                 |                                   |                           | \$ 212,000      | \$ 318,000       |                  |
| 26 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 27 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 37 | Exchangers                  | Exchanger Type                 | Shell Pressure (barg) | Pressure (barg)   | MOC                               | (square meters)           | Equipment Cost  | Bare Module Cost |                  |
| 38 | E-401                       | Floating Head                  | 1                     | 41                | Carbon Steel / Carbon Steel       | 541                       | \$ 96,800       | \$ 326,000       |                  |
| 39 | E-402                       | Floating Head                  | 4.65                  | 0.6               | Stainless Steel / Stainless Steel | 456                       | \$ 83,500       | \$ 514,000       |                  |
| 40 | E-403                       | Floating Head                  | 41                    | 0.35              | Stainless Steel / Stainless Steel | 2010                      | \$ 354,000      | \$ 2,460,000     |                  |
| 41 | E-404                       | Floating Head                  | 5                     | 1.2               | Carbon Steel / Carbon Steel       | 2130                      | \$ 374,000      | \$ 1,230,000     |                  |
| 42 | E-405                       | Floating Head                  | 5                     | 1.05              | Carbon Steel / Carbon Steel       | 2900                      | \$ 510,000      | \$ 1,680,000     |                  |
| 43 | E-406                       | Floating Head                  | 5                     | 0                 | Carbon Steel / Carbon Steel       | 76.7                      | \$ 28,800       | \$ 94,700        |                  |
| 44 | E-407                       | Floating Head                  | 5                     | 0                 | Carbon Steel / Carbon Steel       | 127                       | \$ 35,600       | \$ 117,000       |                  |
| 45 | E-408                       | Floating Head                  | 0                     | 5                 | Carbon Steel / Carbon Steel       | 902                       | \$ 158,000      | \$ 521,000       |                  |
| 46 | E-409                       | Floating Head                  | 0                     | 5                 | Carbon Steel / Carbon Steel       | 680                       | \$ 120,000      | \$ 393,000       |                  |
| 47 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 48 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 55 | Fired Heaters               | Type                           | Heat Duty (MJ/h)      | Superheat (°C)    | MOC                               | Pressure (barg)           | Equipment Cost  | Bare Module Cost |                  |
| 56 | H-401                       | Process Heater                 | 360000                |                   | Carbon Steel                      | 6                         | \$ 4,850,000    | \$ 10,300,000    |                  |
| 57 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 58 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 62 | (with drives)               | Pump Type                      | Power (kilowatts)     | # Spares          | MOC                               | Discharge Pressure (barg) | Equipment Cost  | Bare Module Cost |                  |
| 63 | P-401                       | Centrifugal                    | 6.4                   | 1                 | Carbon Steel                      | 1                         | \$ 8,580        | \$ 34,200        |                  |
| 64 | P-402                       | Centrifugal                    | 1                     | 1                 | Carbon Steel                      | 0.4                       | \$ 6,170        | \$ 24,600        |                  |
| 65 | P-403                       | Centrifugal                    | 2                     | 1                 | Carbon Steel                      | 1.1                       | \$ 6,610        | \$ 26,300        |                  |
| 66 | P-404                       | Centrifugal                    | 0.38                  | 1                 | Carbon Steel                      | 1                         | \$ 6,170        | \$ 24,600        |                  |
| 67 | P-405                       | Centrifugal                    | 0.75                  | 1                 | Carbon Steel                      | 1                         | \$ 6,170        | \$ 24,600        |                  |
| 68 | P-406                       | Centrifugal                    | 2.65                  | 1                 | Carbon Steel                      | 1.1                       | \$ 6,930        | \$ 27,600        |                  |
| 69 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 70 |                             |                                |                       |                   |                                   |                           |                 |                  |                  |
| 80 | Towers                      | Tower Description              | Height (meters)       | Diameter (meters) | Tower MOC                         | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| 81 | T-401                       | 46 Stainless Steel Sieve Trays | 28                    | 3.15              | Carbon Steel                      |                           | 0.6             | \$ 467,000       | \$ 1,240,000     |
| 82 | T-402                       | 34.5 meters of Ceramic         | 34.5                  | 4.1               | Carbon Steel                      |                           | 0.6             | \$ 357,000       | \$ 2,080,000     |

|     |                                  |                                   |                           |                      |                |              |                    |                        |                     |
|-----|----------------------------------|-----------------------------------|---------------------------|----------------------|----------------|--------------|--------------------|------------------------|---------------------|
| 80  | Towers                           | Tower Description                 | Height<br>(meters)        | Diameter<br>(meters) | Tower MOC      | Demister MOC | Pressure<br>(barq) | Equipment<br>Cost      | Bare Module<br>Cost |
| 81  | T-401                            | 46 Stainless Steel Sieve<br>Trays | 28                        | 3.15                 | Carbon Steel   |              | 0.6                | \$ 467,000             | \$ 1,240,000        |
| 82  | T-402                            | 34.5 meters of Ceramic            | 34.5                      | 4.1                  | Carbon Steel   |              | 0.6                | \$ 357,000             | \$ 2,080,000        |
| 83  |                                  |                                   |                           |                      |                |              |                    |                        |                     |
| 84  |                                  |                                   |                           |                      |                |              |                    |                        |                     |
| 88  | Vessels                          | Orientation                       | Length/Height<br>(meters) | Diameter<br>(meters) | MOC            | Demister MOC | Pressure<br>(barq) | Equipment<br>Cost      | Bare Module<br>Cost |
| 89  | V-401                            | Horizontal                        | 7.35                      | 2.45                 | Carbon Steel   |              | 0.9                | \$ 28,400              | \$ 85,400           |
| 90  | V-402                            | Horizontal                        | 3.86                      | 1.29                 | Carbon Steel   |              | 0.6                | \$ 9,220               | \$ 27,700           |
| 91  | V-403                            | Horizontal                        | 3.86                      | 1.29                 | Carbon Steel   |              | 0.6                | \$ 9,220               | \$ 27,700           |
| 92  | V-404                            | Vertical                          | 20                        | 3.6                  | Stainles Steel |              | 0.7                | \$ 160,000             | \$ 1,270,000        |
| 93  | V-405                            | Vertical                          | 20                        | 3.6                  | Stainles Steel |              | 0.7                | \$ 160,000             | \$ 1,270,000        |
| 94  | V-406                            | Vertical                          | 20                        | 3.6                  | Stainles Steel |              | 0.7                | \$ 160,000             | \$ 1,270,000        |
| 95  | V-407                            | Vertical                          | 20                        | 3.6                  | Stainles Steel |              | 0.7                | \$ 160,000             | \$ 1,270,000        |
| 96  | V-408                            | Vertical                          | 20                        | 3.6                  | Stainles Steel |              | 0.7                | \$ 160,000             | \$ 1,270,000        |
| 97  | V-409                            | Vertical                          | 20                        | 3                    | Stainles Steel |              | 0.7                | \$ 114,000             | \$ 905,000          |
| 98  | V-410                            | Vertical                          | 20                        | 3                    | Stainles Steel |              | 0.7                | \$ 114,000             | \$ 905,000          |
| 99  | V-411                            | Vertical                          | 20                        | 3                    | Stainles Steel |              | 0.7                | \$ 114,000             | \$ 905,000          |
| 100 | V-412                            | Vertical                          | 20                        | 3                    | Stainles Steel |              | 0.7                | \$ 114,000             | \$ 905,000          |
| 101 | V-413                            | Vertical                          | 20                        | 3                    | Stainles Steel |              | 0.7                | \$ 114,000             | \$ 905,000          |
| 102 |                                  |                                   |                           |                      |                |              |                    |                        |                     |
| 103 | Reactors R-401 a-e and R-402 a-e |                                   |                           |                      |                |              |                    | Total Bare Module Cost | \$32,968,400        |

## 7.23 Drying Oil Process – Unit 500 – Project B.4

|    |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
|----|-----------------------------|--------------------------------|-------------------------------|--------------------------|-----------------------------------|----------------------------------|------------------------|-------------------------|-------------------------|--------------|
| 1  |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 2  | Add Equipment               |                                | Unit Number    500            |                          |                                   |                                  |                        |                         |                         |              |
| 3  |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 4  | Edit Equipment              |                                | CEPCI        500              |                          |                                   |                                  |                        |                         |                         |              |
| 5  |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 6  |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 7  | <b>User Added Equipment</b> |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 35 | <b>Exchangers</b>           | <b>Exchanger Type</b>          | <b>Shell Pressure (barg)</b>  | <b>Pressure (barg)</b>   | <b>MOC</b>                        | <b>(square meters)</b>           | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |              |
| 36 | E-501                       | Floating Head                  | 5                             | 1                        | Stainless Steel / Stainless Steel | 26.2                             | \$ 23,200              | \$ 143,000              |                         |              |
| 37 | E-502                       | Floating Head                  | 2                             | 0.36                     | Stainless Steel / Stainless Steel | 57.5                             | \$ 26,300              | \$ 162,000              |                         |              |
| 38 | E-503                       | Floating Head                  | 0.25                          | 5                        | Stainless Steel / Stainless Steel | 2.95                             | \$ 25,000              | \$ 154,000              |                         |              |
| 39 | E-504                       | Floating Head                  | 0.25                          | 41                       | Stainless Steel / Stainless Steel | 64.8                             | \$ 27,200              | \$ 174,000              |                         |              |
| 40 | E-505                       | Floating Head                  | 0.05                          | 5                        | Stainless Steel / Stainless Steel | 0.58                             | \$ 25,000              | \$ 154,000              |                         |              |
| 41 | E-506                       | Floating Head                  | 5                             | 0.25                     | Stainless Steel / Stainless Steel | 919                              | \$ 161,000             | \$ 994,000              |                         |              |
| 42 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 49 | <b>Fired Heaters</b>        | <b>Type</b>                    | <b>Heat Duty (MW/h)</b>       | <b>Superheat (°C)</b>    | <b>MOC</b>                        | <b>Pressure (barg)</b>           | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |              |
| 50 | H-501                       | Process Heater                 | 14400                         |                          | Carbon Steel                      | 13                               | \$ 755,000             | \$ 1,610,000            |                         |              |
| 51 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 52 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 56 | <b>(with drives)</b>        | <b>Pump Type</b>               | <b>Power (kilowatts)</b>      | <b># Spares</b>          | <b>MOC</b>                        | <b>Discharge Pressure (barg)</b> | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |              |
| 57 | P-501                       | Centrifugal                    | 0.9                           | 1                        | Carbon Steel                      | 13                               | \$ 6,170               | \$ 24,600               |                         |              |
| 58 | P-502                       | Centrifugal                    | 1                             | 1                        | Stainless Steel                   | 2                                | \$ 6,170               | \$ 30,700               |                         |              |
| 59 | P-503                       | Centrifugal                    | 0.8                           | 1                        | Stainless Steel                   | 2                                | \$ 6,170               | \$ 30,700               |                         |              |
| 60 | P-504                       | Centrifugal                    | 0.3                           | 1                        | Stainless Steel                   | 1                                | \$ 6,170               | \$ 30,700               |                         |              |
| 61 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 71 | <b>Towers</b>               | <b>Tower Description</b>       | <b>Height (meters)</b>        | <b>Diameter (meters)</b> | <b>Tower MOC</b>                  | <b>Demister MOC</b>              | <b>Pressure (barg)</b> | <b>Equipment Cost</b>   | <b>Bare Module Cost</b> |              |
| 72 | T-501                       | 58 Stainless Steel Sieve Trays | 17                            | 2.1                      | Stainless Clad                    |                                  | 1                      | \$ 212,000              | \$ 579,000              |              |
| 73 | T-502                       | 35 Stainless Steel Sieve Trays | 11                            | 0.45                     | Stainless Steel                   |                                  | 1                      | \$ 41,800               | \$ 122,000              |              |
| 74 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 78 | <b>Vessels</b>              | <b>Orientation</b>             | <b>Length/Height (meters)</b> | <b>Diameter (meters)</b> | <b>MOC</b>                        | <b>Demister MOC</b>              | <b>Pressure (barg)</b> | <b>Equipment Cost</b>   | <b>Bare Module Cost</b> |              |
| 79 | V-501                       | Horizontal                     | 3                             | 1                        | Carbon Steel                      |                                  | 0.1                    | \$ 6,450                | \$ 19,400               |              |
| 80 | V-502                       | Vertical                       | 4.6                           | 0.9                      | Stainless Steel                   |                                  | 0.48                   | \$ 6,760                | \$ 53,500               |              |
| 81 | V-503                       | Horizontal                     | 3                             | 1                        | Stainless Steel                   |                                  | 0.25                   | \$ 6,450                | \$ 40,100               |              |
| 82 | V-504                       | Horizontal                     | 1.5                           | 0.5                      | Stainless Steel                   |                                  | 0.05                   | \$ 3,030                | \$ 18,300               |              |
| 83 | V-505                       | Vertical                       | 5.3                           | 0.53                     | Stainless Steel                   |                                  | 1                      | \$ 4,250                | \$ 33,600               |              |
| 84 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |              |
| 85 |                             |                                |                               |                          |                                   |                                  |                        |                         | Total Bare Module Cost  | \$ 4,374,200 |

Reactor R-501

## 7.24 Maleic Anhydride – Unit 600 – Project B.5

|    |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
|----|-----------------------------|--------------------------------|-------------------------------|--------------------------|-----------------------------------|----------------------------------|------------------------|-------------------------|-------------------------|
| 7  | <b>User Added Equipment</b> |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 14 | <b>Compressors</b>          | <b>Compressor Type</b>         | <b>Power (kilowatts)</b>      | <b># Spares</b>          | <b>MDC</b>                        |                                  | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |
| 15 | C-601                       | Centrifugal                    | 3110                          | 0                        | Carbon Steel                      |                                  | \$ 968,000             | \$ 2,650,000            |                         |
| 16 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 17 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 24 | <b>Drives</b>               | <b>Drive Type</b>              | <b>Power (kilowatts)</b>      | <b># Spares</b>          |                                   |                                  | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |
| 25 | D-601                       | Electric - Open/Drip Proof     | 3108                          | 0                        |                                   |                                  | \$ 265,000             | \$ 398,000              |                         |
| 26 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 27 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 37 | <b>Exchangers</b>           | <b>Exchanger Type</b>          | <b>Shell Pressure (barg)</b>  | <b>Pressure (barg)</b>   | <b>MDC</b>                        | <b>(square meters)</b>           | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |
| 38 | E-601                       | Floating Head                  | 1.8                           | 5                        | Stainless Steel / Stainless Steel | 14.6                             | \$ 23,600              | \$ 145,000              |                         |
| 39 | E-602                       | Floating Head                  | 2                             | 41                       | Stainless Steel / Stainless Steel | 61.6                             | \$ 26,800              | \$ 171,000              |                         |
| 40 | E-603                       | Floating Head                  | 2                             | 41                       | Stainless Steel / Stainless Steel | 1760                             | \$ 309,000             | \$ 1,970,000            |                         |
| 41 | E-604                       | Fixed, Sheet, or U-Tube        | 5                             | 1                        | Stainless Steel / Stainless Steel | 1090                             | \$ 132,000             | \$ 813,000              |                         |
| 42 | E-605                       | Floating Head                  | 2                             | 41                       | Stainless Steel / Stainless Steel | 131                              | \$ 36,100              | \$ 230,000              |                         |
| 43 | E-606                       | Floating Head                  | 0                             | 5                        | Stainless Steel / Stainless Steel | 11.7                             | \$ 24,300              | \$ 150,000              |                         |
| 44 | E-607                       | Floating Head                  | 40                            | 40                       | Stainless Steel / Stainless Steel | 192                              | \$ 44,600              | \$ 309,000              |                         |
| 45 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 46 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 53 | <b>Fired Heaters</b>        | <b>Type</b>                    | <b>Heat Duty (MM/h)</b>       | <b>Superheat (°C)</b>    | <b>MDC</b>                        | <b>Pressure (barg)</b>           | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |
| 54 | H-601                       | Process Heater                 | 32000                         |                          | Carbon Steel                      | 3                                | \$ 1,010,000           | \$ 2,150,000            |                         |
| 55 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 56 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 60 | <b>(with drives)</b>        | <b>Pump Type</b>               | <b>Power (kilowatts)</b>      | <b># Spares</b>          | <b>MDC</b>                        | <b>Discharge Pressure (barg)</b> | <b>Equipment Cost</b>  | <b>Bare Module Cost</b> |                         |
| 61 | P-601                       | Centrifugal                    | 0.3                           | 1                        | Carbon Steel                      | 2                                | \$ 6,170               | \$ 24,600               |                         |
| 62 | P-602                       | Centrifugal                    | 3.8                           | 1                        | Stainless Steel                   | 2                                | \$ 7,470               | \$ 37,100               |                         |
| 63 | P-603                       | Reciprocating                  | 0.1                           | 1                        | Stainless Steel                   | 1                                | \$ 11,900              | \$ 60,400               |                         |
| 64 | P-604                       | Centrifugal                    | 6.75                          | 1                        | Stainless Steel                   | 1                                | \$ 9,720               | \$ 43,300               |                         |
| 65 | P-605                       | Centrifugal                    | 0.7                           | 1                        | Stainless Steel                   | 3                                | \$ 6,170               | \$ 30,700               |                         |
| 66 | P-606                       | Centrifugal                    | 2.4                           | 1                        | Stainless Steel                   | 0.5                              | \$ 6,810               | \$ 33,800               |                         |
| 67 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 68 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 78 | <b>Towers</b>               | <b>Tower Description</b>       | <b>Height (meters)</b>        | <b>Diameter (meters)</b> | <b>Tower MDC</b>                  | <b>Demister MDC</b>              | <b>Pressure (barg)</b> | <b>Equipment Cost</b>   | <b>Bare Module Cost</b> |
| 79 | T-601                       | 14 Stainless Steel Sieve Trays | 10                            | 4.2                      | Stainless Steel                   |                                  | 1.1                    | \$ 492,000              | \$ 1,700,000            |
| 80 | T-602                       | 42 Stainless Steel Sieve Trays | 18                            | 1.05                     | Stainless Steel                   |                                  | 1.1                    | \$ 68,500               | \$ 243,000              |
| 81 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 82 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 86 | <b>Vessels</b>              | <b>Orientation</b>             | <b>Length/Height (meters)</b> | <b>Diameter (meters)</b> | <b>MDC</b>                        | <b>Demister MDC</b>              | <b>Pressure (barg)</b> | <b>Equipment Cost</b>   | <b>Bare Module Cost</b> |
| 87 | V-601                       | Horizontal                     | 3.5                           | 1.17                     | Carbon Steel                      |                                  | 0.1                    | \$ 8,010                | \$ 24,100               |
| 88 | V-602                       | Horizontal                     | 13.2                          | 4.4                      | Stainless Steel                   |                                  | 0.1                    | \$ 119,000              | \$ 740,000              |
| 89 | V-603                       | Horizontal                     | 3.9                           | 1.3                      | Stainless Steel                   |                                  | 0.1                    | \$ 9,380                | \$ 58,300               |
| 90 | V-604                       | Vertical                       | 7                             | 3.8                      | Stainless Steel                   |                                  | 2                      | \$ 68,700               | \$ 764,000              |
| 91 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
| 92 |                             |                                |                               |                          |                                   |                                  |                        |                         |                         |
|    | Total Bare Module Cost      |                                |                               |                          |                                   |                                  |                        | \$ 12,745,300           |                         |

Reactor R-601

## 7.25 Ethylene Oxide – Unit 700 – Project B.6

|                        |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
|------------------------|--------------------------|--------------------------------|------------------------|-------------------|-----------------------------------|---------------------------|-----------------|------------------|------------------|
| 1                      |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 2                      | Add Equipment            | Unit Number                    | 700                    |                   |                                   |                           |                 |                  |                  |
| 3                      |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 4                      | Edit Equipment           | CEPCI                          | 500                    |                   |                                   |                           |                 |                  |                  |
| 5                      |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 6                      |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 7                      | User Added Equipment     |                                |                        |                   |                                   |                           |                 |                  |                  |
| 14                     | Compressors              | Compressor Type                | Power (kilowatts)      | # Spares          | MOC                               |                           | Equipment Cost  | Bare Module Cost |                  |
| 15                     | C-701                    | Centrifugal                    | 19000                  | 0                 | Carbon Steel                      |                           | \$ 4,950,000    | \$ 13,600,000    |                  |
| 16                     | C-702                    | Centrifugal                    | 23000                  | 0                 | Carbon Steel                      |                           | \$ 5,870,000    | \$ 16,100,000    |                  |
| 17                     | C-703                    | Centrifugal                    | 21500                  | 0                 | Carbon Steel                      |                           | \$ 5,620,000    | \$ 15,400,000    |                  |
| 18                     | C-704                    | Centrifugal                    | 5500                   | 0                 | Carbon Steel                      |                           | \$ 1,430,000    | \$ 3,910,000     |                  |
| 19                     | C-705                    | Centrifugal                    | 5500                   | 0                 | Carbon Steel                      |                           | \$ 1,430,000    | \$ 3,910,000     |                  |
| 20                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 21                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 28                     | Drives                   | Drive Type                     | Power (kilowatts)      | # Spares          |                                   |                           | Equipment Cost  | Bare Module Cost |                  |
| 29                     | D-701                    | Electric - Explosion Proof     | 19000                  | 1                 |                                   |                           | \$ 3,210,000    | \$ 4,910,000     |                  |
| 30                     | D-702                    | Electric - Explosion Proof     | 23000                  | 1                 |                                   |                           | \$ 3,660,000    | \$ 5,490,000     |                  |
| 31                     | D-703                    | Electric - Explosion Proof     | 21500                  | 1                 |                                   |                           | \$ 3,610,000    | \$ 5,420,000     |                  |
| 32                     | D-704                    | Electric - Explosion Proof     | 5500                   | 1                 |                                   |                           | \$ 1,130,000    | \$ 1,700,000     |                  |
| 33                     | D-705                    | Electric - Explosion Proof     | 5500                   | 1                 |                                   |                           | \$ 1,130,000    | \$ 1,700,000     |                  |
| 34                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 35                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 45                     | Exchangers               | Exchanger Type                 | Shell Pressure (barg)  | Pressure (barg)   | MOC                               | (Square meters)           | Equipment Cost  | Bare Module Cost |                  |
| 46                     | E-701                    | Floating Head                  | 5                      | 3                 | Carbon Steel / Carbon Steel       | 5550                      | \$ 975,000      | \$ 3,210,000     |                  |
| 47                     | E-702                    | Floating Head                  | 5                      | 3                 | Carbon Steel / Carbon Steel       | 6260                      | \$ 1,100,000    | \$ 3,630,000     |                  |
| 48                     | E-703                    | Floating Head                  | 5                      | 26.8              | Carbon Steel / Carbon Steel       | 12100                     | \$ 2,130,000    | \$ 7,110,000     |                  |
| 49                     | E-704                    | Floating Head                  | 5                      | 25.8              | Carbon Steel / Carbon Steel       | 14100                     | \$ 2,480,000    | \$ 8,290,000     |                  |
| 50                     | E-705                    | Floating Head                  | 41                     | 30                | Carbon Steel / Carbon Steel       | 14100                     | \$ 2,480,000    | \$ 8,870,000     |                  |
| 51                     | E-706                    | Floating Head                  | 5                      | 25.8              | Carbon Steel / Carbon Steel       | 13900                     | \$ 2,450,000    | \$ 8,190,000     |                  |
| 52                     | E-707                    | Floating Head                  | 5                      | 30                | Carbon Steel / Carbon Steel       | 1480                      | \$ 259,000      | \$ 870,000       |                  |
| 53                     | E-708                    | Floating Head                  | 10.5                   | 41                | Stainless Steel / Stainless Steel | 566                       | \$ 101,000      | \$ 701,000       |                  |
| 54                     | E-709                    | Floating Head                  | 10                     | 5                 | Stainless Steel / Stainless Steel | 154                       | \$ 39,300       | \$ 245,000       |                  |
| 55                     | E-710                    | Floating Head                  | 10                     | 30                | Carbon Steel / Carbon Steel       | 22100                     | \$ 3,890,000    | \$ 13,600,000    |                  |
| 56                     | E-711                    | Floating Head                  | 10                     | 30                | Carbon Steel / Carbon Steel       | 19000                     | \$ 3,350,000    | \$ 11,700,000    |                  |
| 57                     | Reactors R-701 and R-702 |                                |                        |                   |                                   |                           |                 |                  |                  |
| 58                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 71                     | [with drives]            | Pump Type                      | Power (kilowatts)      | # Spares          | MOC                               | Discharge Pressure (barg) | Equipment Cost  | Bare Module Cost |                  |
| 72                     | P-701                    | Centrifugal                    | 4                      | 1                 | Stainless Steel                   | 10                        | \$ 7,560        | \$ 37,500        |                  |
| 73                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 74                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 84                     | Towers                   | Tower Description              | Height (meters)        | Diameter (meters) | Tower MOC                         | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| 85                     | T-701                    | 20 Stainless Steel Sieve Trays | 12.2                   | 5.6               | Carbon Steel                      |                           | 30.2            | \$ 1,460,000     | \$ 8,180,000     |
| 86                     | T-702                    | 20 Stainless Steel Sieve Trays | 12.2                   | 5.6               | Carbon Steel                      |                           | 30              | \$ 1,460,000     | \$ 8,150,000     |
| 87                     | T-703                    | 70 Stainless Steel Sieve Trays | 43                     | 8                 | Stainless Steel                   |                           | 10              | \$ 29,700,000    | \$ 99,400,000    |
| 88                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 89                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 93                     | Vessels                  | Orientation                    | Length/Height (meters) | Diameter (meters) | MOC                               | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| 94                     | V-701                    | Horizontal                     | 5.26                   | 1.75              | Stainless Steel                   |                           | 10              | \$ 15,300        | \$ 190,000       |
| 95                     | V-702                    | Vertical                       | 10                     | 5.07              | Carbon Steel                      |                           | 30              | \$ 169,000       | \$ 3,810,000     |
| 96                     | V-703                    | Vertical                       | 10                     | 5.07              | Carbon Steel                      |                           | 30              | \$ 169,000       | \$ 3,810,000     |
| 97                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 98                     |                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| Total Bare Module Cost |                          |                                |                        |                   |                                   |                           |                 | \$ 262,033,500   |                  |

## 7.26 Formalin Process – Unit 800 – Project B.7

|    |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
|----|------------------------------------------|--------------------------------|------------------------|-------------------|-----------------------------------|---------------------------|-----------------|------------------|------------------|
| 1  |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 2  |                                          | Add Equipment                  |                        | Unit Number       | 800                               |                           |                 |                  |                  |
| 3  |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 4  |                                          | Edit Equipment                 |                        | CEPCI             | 500                               |                           |                 |                  |                  |
| 5  |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 6  |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 7  | <b>User Added Equipment</b>              |                                |                        |                   |                                   |                           |                 |                  |                  |
| 14 | Compressors                              | Compressor Type                | Power (kilowatts)      | # Spares          | MOC                               |                           | Equipment Cost  | Bare Module Cost |                  |
| 15 | C-801                                    | Centrifugal                    | 183                    | 0                 | Carbon Steel                      |                           | \$ 189,000      | \$ 517,000       |                  |
| 16 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 17 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 24 | Drives                                   | Drive Type                     | Power (kilowatts)      | # Spares          |                                   |                           | Equipment Cost  | Bare Module Cost |                  |
| 25 | D-801                                    | Electric - Explosion Proof     | 183                    | 0                 |                                   |                           | \$ 70,900       | \$ 106,000       |                  |
| 26 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 27 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 37 | Exchangers                               | Exchanger Type                 | Shell Pressure (barg)  | Pressure (barg)   | MOC                               | (square meters)           | Equipment Cost  | Bare Module Cost |                  |
| 38 | E-801                                    | Floating Head                  | 2.5                    | 10                | Carbon Steel / Carbon Steel       | 405                       | \$ 75,700       | \$ 250,000       |                  |
| 39 | E-802                                    | Floating Head                  | 2.5                    | 41                | Carbon Steel / Carbon Steel       | 4.62                      | \$ 25,000       | \$ 84,400        |                  |
| 40 | E-803                                    | Floating Head                  | 2.5                    | 5                 | Carbon Steel / Carbon Steel       | 28.2                      | \$ 23,300       | \$ 76,600        |                  |
| 41 | E-804                                    | Kettle Reboiler                | 1.5                    | 10                | Stainless Steel / Stainless Steel | 37.3                      | \$ 51,600       | \$ 320,000       |                  |
| 42 | E-805                                    | Floating Head                  | 1.5                    | 5                 | Stainless Steel / Stainless Steel | 269                       | \$ 55,600       | \$ 342,000       |                  |
| 43 | E-806                                    | Floating Head                  | 5                      | 3                 | Stainless Steel / Stainless Steel | 41                        | \$ 24,400       | \$ 150,000       |                  |
| 44 | E-807                                    | Floating Head                  | 10                     | 2.5               | Carbon Steel / Carbon Steel       | 140                       | \$ 37,400       | \$ 124,000       |                  |
| 45 | Reactor R-801 costed as a heat exchanger |                                |                        |                   |                                   |                           |                 |                  |                  |
| 46 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 59 | (with drives)                            | Pump Type                      | Power (kilowatts)      | # Spares          | MOC                               | Discharge Pressure (barg) | Equipment Cost  | Bare Module Cost |                  |
| 60 | P-801                                    | Centrifugal                    | 0.3                    | 1                 | Carbon Steel                      | 3                         | \$ 6,170        | \$ 24,600        |                  |
| 61 | P-802                                    | Centrifugal                    | 1.7                    | 1                 | Stainless Steel                   | 1.5                       | \$ 6,470        | \$ 32,100        |                  |
| 62 | P-803                                    | Centrifugal                    | 0.5                    | 1                 | Stainless Steel                   | 3.5                       | \$ 6,170        | \$ 30,700        |                  |
| 63 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 64 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 74 | Towers                                   | Tower Description              | Height (meters)        | Diameter (meters) | Tower MOC                         | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| 75 | T-801                                    | 10 meters of Ceramic           | 10                     | 0.86              | Carbon Steel                      |                           | 2               | \$ 18,300        | \$ 49,200        |
| 76 | T-802                                    | 31 Stainless Steel Sieve Trays | 19                     | 2.5               | Stainless Steel                   |                           | 2               | \$ 200,000       | \$ 937,000       |
| 77 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 78 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 82 | Vessels                                  | Orientation                    | Length/Height (meters) | Diameter (meters) | MOC                               | Demister MOC              | Pressure (barg) | Equipment Cost   | Bare Module Cost |
| 83 | V-801                                    | Horizontal                     | 4.41                   | 1.1               | Carbon Steel                      |                           | 2               | \$ 8,460         | \$ 25,500        |
| 84 |                                          |                                |                        |                   |                                   |                           |                 |                  |                  |
| 85 | Total Bare Module Cost                   |                                |                        |                   |                                   |                           |                 |                  | \$ 3,069,100     |

## Chapter 8

8.1 The term  $0.18FCI$  has nothing to do with the interest on capital investment. From Table 8.2 we see that the following costs are based on the FCI:

- Maintenance and repairs
- Operating supplies
- Local insurance and taxes
- Plant overhead costs
- Administration costs

8.2 The number of operators per shift is multiplied by approximately 4.5 to get the total number of operators required by the plant because it is assumed that a single operator works 5 shifts a week and 49 weeks per year.

$$\text{Total number of operators} = (\# \text{ operators per shift}) \left( \frac{\left( \frac{\text{days}}{\text{year}} \right) \left( \frac{\text{shifts}}{\text{day}} \right)}{\left( \frac{\text{weeks}}{\text{operator}} \right) \left( \frac{\text{shifts}}{\text{week}} \right)} \right)$$

$$\frac{(365 \text{ days/yr})(3 \text{ shifts/day})}{(49 \text{ weeks/operator})(5 \text{ shifts/week})} \approx 4.5 \text{ operators/year}$$

8.3 Stream Factor (SF) =  $\frac{\# \text{ Days plant operated per year}}{365}$  (Equation 8.5)

8.4  $COM_d = f(\text{FCI, operating labor, raw material costs, utility costs, waste treatment costs})$

8.5 The simple answer is that the cooling water cost does not vary very much with return temperature because the main costs are associated with pumping (electricity), fans (power), cost of water make up and chemical addition.

As  $T_{CW, \text{return}} \uparrow$ ,  $m_{CW} \downarrow \therefore$  pump power  $\downarrow$  but  $W_{BD}$  and  $W_{\text{Make up}} \uparrow$  - so these compensate each other.

Must not exceed  $T_{CW \text{ Return}} = 45^\circ \text{C}$  because of excessive fouling of heat exchanger surfaces above this temperature



8.6 No – this is not the reason for the factor of 1.23 multiplying  $C_{RM}$ . The reason is that in the derivation of Equation (8.2) from the data in Table 8.2 some of the costs are based on the  $COM_D$ . Thus when all terms are added, the factor multiplying  $COM_D$  is  $0.19COM_D$ .

$$COM_D = (C_{RM} + C_{WT} + C_{UT}) + 2.215C_{OL} + 0.146FCI + 0.19COM_D$$

$$0.81COM_D = (C_{RM} + C_{WT} + C_{UT}) + 2.215C_{OL} + 0.146FCI$$

$$\therefore COM_D = \frac{1}{0.81} ((C_{RM} + C_{WT} + C_{UT}) + 2.215C_{OL} + 0.146FCI)$$

$$= 1.23((C_{RM} + C_{WT} + C_{UT}) + 2.215C_{OL} + 0.146FCI)$$

8.7 Yes, the 2.73 multiplier does include costs for supervisory and clerical labor but also includes:

- Laboratory charges
- Plant overhead charges
- Administration costs

8.8 From Table 8.1 we have

- |                    |                                                                                                                                                                                              |
|--------------------|----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|
| • Direct Costs     | Costs that vary with the rate of production e.g. raw materials, utility costs.                                                                                                               |
| • Indirect Costs   | Costs (in the plant) that do not vary with the rate of production, e.g. depreciation, taxes, insurance                                                                                       |
| • General Expenses | Costs associated with management level and administrative functions not directly related to the manufacturing process, e.g. administration, research and development, distribution, selling. |

8.9

- (a) Options (i) and (ii) are not reliable, and option (iii) is the best. If you do not have time to dig up information on the new plant then (i) is probably better than (ii) unless the dye and resin industries are known to have similar costs.
- (b) Use Equation 8.2 when you have no/little information on a particular process or you want to estimate manufacturing costs for a broad range of industries/processes.

8.10 (a) 10,000 kg/h, 58.6 barg – SH = 165°C

(i) Discharge P = 4" Hg absolute

From Table 8.5, theoretical steam required = 3.2 kg steam/kWh

$$\text{Power from Turbine} = \frac{10,000 \left( \frac{\text{kg}}{\text{h}} \right)}{3.20 \left( \frac{\text{kg}}{\text{kWh}} \right)} (0.80) = 2500 \text{ kW} = 2.5 \text{ MW}$$

(ii) Discharge P = 4.82 barg

From Table 8.5, theoretical steam rate = 6.45 kg steam/kWh

$$\text{Power from Turbine} = \frac{10,000 \left( \frac{\text{kg}}{\text{h}} \right)}{6.45 \left( \frac{\text{kg}}{\text{kWh}} \right)} (0.80) = 1240 \text{ kW} = 1.24 \text{ MW}$$

(b) 10,000 kg/h saturated at 10 barg = 4.82 barg

From Table 8.5, theoretical steam rate = 28.80 kg steam/kWh

$$\text{Power from Turbine} = \frac{10,000 \left( \frac{\text{kg}}{\text{h}} \right)}{28.80 \left( \frac{\text{kg}}{\text{kWh}} \right)} (0.80) = 277.8 \text{ kW}$$

(c) a-(i) ⇒ Turbine #1 in Figure 8.6 (HP → surface condenser)

a-(ii) ⇒ Turbine #2 (HP → LP)

b ⇒ Turbine #4 (MP → LP)

8.11 Costs for cooling water are:

- Pumping costs (electric)
- Fan costs (electric)
- Process water make up
- Injection of chemicals for water treatment

$$(a) \text{ Make Up Water (\%)} = \frac{\frac{(517.2)(0.067)}{(1,000)}}{(0.354)} \cdot 100 = 9.8\%$$

$$(b) \text{ Power Cost (\%)} = \frac{(0.06)(2.36 + 1.61)}{(0.354)} \cdot 100 = 67.3\%$$

If power cost doubles:

$$\text{Cost of Cooling Water} = 0.354 + (0.06)(2.36 + 1.61) = \$0.592 / 1,000 \text{ kg (167\%)}$$

$$(c) \text{ New Cost} = 0.354 + \frac{(517.3)(0.067)}{(1,000)} = \$0.389 / 1,000 \text{ kg (109.8\%)}$$

8.12 (a) Using Chemcad and SRK package for k-values and enthalpy calculations, the following data are found: at 45°C vapor pressure of propane = 1556 kPa

(b) At -50°C vapor pressure of propane = 69.47 kPa

(c) For case when there are no  $\Delta P$  in exchangers and condenser and evaporator pressures are set to (a) and (b).

(d) Chemcad simulation gives:

$$Q_{\text{evap}} = 168.7 \text{ MJ/h}$$

$$m_{\text{propane}} = 20 \text{ kmol/h} = 881.9 \text{ kg/h}$$

$$\text{Propane required for 1 GJ} = \frac{10^3 [\text{GJ/MJ}]}{168.7 [\text{MJ/h}]} 881.9 [\text{kg/h}] = 5227.6 \text{ kg} = 118.6 \text{ kmol}$$

Circulating 5227.6 kg/h of propane  $\Rightarrow$  1 GJ/h in condenser

(e) Using 5 kPa  $\Delta P$  in exchangers and a 75% compressor efficiency we get:

$$\text{For } Q_{\text{evap}} = 1 \text{ GJ/h}$$

$$W_{\text{comp}} = 288.7 \text{ kW}$$

$$Q_{\text{cond}} = -2.040 \text{ GJ/h}$$

Cost to produce 1 GJ of refrigeration at -50°C

$$\text{Using propane} = (288.7 \text{ kW})(0.06) + (2.04)(0.354)$$

$$\text{Cost} = \$18.04/\text{GJ}$$

8.13 (a) Repeat problem 8.12 using propylene as refrigerant

Vapor pressure (45°C) = 1,876 kPa

Vapor pressure (-50°C) = 90.6 kPa

$m_{\text{propylene}} = 4,909.3 \text{ kg/h}$  (116.6 kmol/h)

$Q_{\text{cond}} = -2.015 \text{ GJ/h}$

$Q_{\text{evap}} = 1.00 \text{ GJ/h}$

$W_{\text{comp}} = 282.0 \text{ kW}$

$$\text{Cost} = (282)(0.06) + (2.015)(0.354) = \$17.63/\text{GJ}$$

(b) Repeat problem 8.12 using ethane as refrigerant

Vapor pressure (45°C) = No result  $T_{\text{critical}} \approx 32^\circ\text{C}$

Vapor pressure (-50°C) = 554 kPa

We cannot use ethane in a single stage refrigeration cycle because we cannot condense it at a temperature of 45°C – need to go to a 2-stage system to use ethane. It is probably not worth using a 2-stage system for this temperature since other systems (single-stage) can be used.

(c) Repeat problem 8.12 using ammonia as refrigerant

Vapor pressure (45°C) = 1,797 kPa

Vapor pressure (-50°C) = 38.7 kPa

$m_{\text{ammonia}} = 1,012.4 \text{ kg/h}$

$Q_{\text{cond}} = -1.896 \text{ GJ/h}$

$Q_{\text{evap}} = 1 \text{ GJ/h}$

$W_{\text{comp}} = 248.9 \text{ kW}$

$$\text{Cost} = (248.9 \text{ kW})(0.06) + (1.896 \text{ GJ/h})(0.354) = \$15.61/\text{GJ}$$

Note: For problems 12 and 13 only single stage compressors have been used.

Because the pressure ratios are all much greater than the recommended maximum of 3~4, multiple stage compressors with inter-cooling should be used. This will reduce the above costs. For example, for ammonia the pressure ratio =  $\frac{1,802}{38.7} = 46.6$ .

Therefore, using 3 stages with inter-cooling with each stage having a  $P_R = 3.6$  we get:

$Q_{\text{cond}} = -1,490 - 24 - 323 = -1,837 \text{ MJ/h}$

$W_{\text{comp}} = 61.5 + 86.6 + 84.5 = 232.6 \text{ kW}$

$Q_{\text{evap}} = 1 \text{ GJ/h}$

$$\text{Cost} = (232.6 \text{ kW})(0.06) + (1.837 \text{ GJ/h})(0.354) = \$14.61/\text{GJ}$$

## 8.14 Operating Costs for the Ethyl Benzene – Project B.2

| Name   | Total Module Cost | Grass Roots Cost | Utility Used        | Efficiency | Actual Usage    | Annual Utility Cost |
|--------|-------------------|------------------|---------------------|------------|-----------------|---------------------|
| E-301  | \$ 221,000        | \$ 266,000       | High-Pressure Steam |            | -1967 MJ/h      | \$ (289,738)        |
| E-302  | \$ 240,000        | \$ 288,000       | High-Pressure Steam |            | -2592 MJ/h      | \$ (381,800)        |
| E-303  | \$ 801,000        | \$ 960,000       | High-Pressure Steam |            | -10080 MJ/h     | \$ (1,484,778)      |
| E-304  | \$ 650,000        | \$ 920,000       | Low-Pressure Steam  |            | -12367.001 MJ/h | \$ (1,366,753)      |
| E-305  | \$ 264,000        | \$ 375,000       | Cooling Water       |            | 4940 MJ/h       | \$ 14,600           |
| E-306  | \$ 101,000        | \$ 142,000       | Low-Pressure Steam  |            | 9110 MJ/h       | \$ 1,006,700        |
| E-307  | \$ 101,000        | \$ 144,000       | Cooling Water       |            | 7280 MJ/h       | \$ 21,400           |
| E-308  | \$ 82,000         | \$ 116,000       | High-Pressure Steam |            | 5280 MJ/h       | \$ 777,900          |
| E-309  | \$ 90,000         | \$ 128,000       | Cooling Water       |            | 5260 MJ/h       | \$ 15,500           |
| H-301  | \$ 3,490,000      | \$ 4,610,000     | Natural Gas         | 0.9        | 24900 MJ/h      | \$ 2,067,000        |
| P-301  | \$ 83,000         | \$ 111,000       | Electricity         | 0.75       | 20 kilowatts    | \$ 10,070           |
| P-302  | \$ 29,000         | \$ 39,000        | Electricity         | 0.75       | 1.33 kilowatts  | \$ 671              |
| P-303  | \$ 29,000         | \$ 39,000        | Electricity         | 0.75       | 1.33 kilowatts  | \$ 671              |
| P-304  | \$ 34,200         | \$ 44,500        | Electricity         | 0.8        | 1.75 kilowatts  | \$ 881              |
| P-305  | \$ 44,400         | \$ 59,300        | Electricity         | 0.75       | 3.6 kilowatts   | \$ 1,810            |
| T-301  | \$ 481,000        | \$ 640,000       | NA                  |            |                 |                     |
| T-302  | \$ 513,000        | \$ 675,000       | NA                  |            |                 |                     |
| V-301  | \$ 38,900         | \$ 55,400        | NA                  |            |                 |                     |
| V-302  | \$ 68,400         | \$ 97,000        | NA                  |            |                 |                     |
| V-303  | \$ 40,900         | \$ 58,200        | NA                  |            |                 |                     |
| V-304  | \$ 36,500         | \$ 52,000        | NA                  |            |                 |                     |
| V-305  | \$ 830,000        | \$ 880,000       | NA                  |            |                 |                     |
| V-306  | \$ 1,050,000      | \$ 1,120,000     | NA                  |            |                 |                     |
| V-307  | \$ 1,220,000      | \$ 1,290,000     | NA                  |            |                 |                     |
| V-308  | \$ 145,000        | \$ 160,000       | NA                  |            |                 |                     |
| Totals | \$ 10,700,000     | \$ 13,300,000    | $FCI = C_{TM}$      |            |                 |                     |
|        |                   |                  | $C_{UT}$            |            |                 |                     |
|        |                   |                  | \$ 390,000          |            |                 |                     |

Note that full credit is given for the steam generated in E-301 – 304 (In CAPCOST, this credit is activated by using a negative duty for the exchanger)

$$N_{OL} = [6.29 + 31.7P^2 + 0.23N_{np}]^{0.5} = [6.29 + 0.23(16)]^{0.5} = 3.16 \frac{\text{operators}}{\text{shift}}$$

$$(N_{np} = 9 \text{ exchangers} + 1 \text{ heater} + 2 \text{ towers} + 4 \text{ reactors} = 16)$$

$$N_{operators} = (4.5)(3.16) = 14.2 \text{ round up to } 15$$

$$C_{OL} = (15)(\$52,900) = \$794,000/\text{yr}$$

$$\text{Number of hr/yr} = 80,000,000/9538.6 = 8387 \text{ h/yr (based on 80,000 tonne/yr of EB)}$$

$$C_{RM, \text{benzene}} = (7761.3)(0.657)(8387) = \$42,770,000/\text{yr}$$

$$C_{RM, \text{ethylene}} = (2819.5)(1.202)(8387) = \$28,420,000/\text{yr}$$

$$FCI \text{ (from Problem 7.21 CAPCOST Output for } C_{TM}) = \$10,700,000$$

$$C_{UT} = \$390,000/\text{yr}$$

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) = 0.180(10.7) + 2.73(0.794) + 1.23(0.390 + 0 + 42.77 + 28.42) = \$92,140,000/\text{yr}$$

$$COM_d = \$92,140,000/\text{yr}$$

$$COM_d = [\$92,140,000/\text{yr}]/[80,000,000 \text{ kg/yr}] = \$1.152/\text{kg EB (slightly higher than in Table 8.4)}$$

### 8.15 Operating Costs for the Styrene Process – Project B.3

| Name   | Total Module Cost | Grass Roots Cost | Utility Used        | Efficiency | Actual Usage     | Annual Utility Cost |
|--------|-------------------|------------------|---------------------|------------|------------------|---------------------|
| C-401  | \$ 610,000        | \$ 869,000       | NA                  |            |                  |                     |
| D-401  | \$ 375,000        | \$ 534,000       | Electricity         | 0.9        | 422 kilowatts    | \$ 213,000          |
| E-401  | \$ 384,871        | \$ 544,000       | High-Pressure Steam |            | 29700 MJ/h       | \$ 4,374,000        |
| E-402  | \$ 607,000        | \$ 744,000       | Low-Pressure Steam  |            | -198460.008 MJ/h | \$ (21,933,039)     |
| E-403  | \$ 2,901,000      | \$ 3,480,000     | High-Pressure Steam |            | -162800.028 MJ/h | \$ (23,980,345)     |
| E-404  | \$ 1,450,000      | \$ 2,070,000     | Low-Pressure Steam  |            | -35160.001 MJ/h  | \$ (3,885,748)      |
| E-405  | \$ 1,980,000      | \$ 2,820,000     | Cooling Water       |            | 396000 MJ/h      | \$ 1,170,000        |
| E-406  | \$ 110,000        | \$ 159,000       | Low-Pressure Steam  |            | 331000 MJ/h      | \$ 36,600,000       |
| E-407  | \$ 138,000        | \$ 197,000       | Cooling Water       |            | 281000 MJ/h      | \$ 830,000          |
| E-408  | \$ 615,000        | \$ 870,000       | Low-Pressure Steam  |            | 341000 MJ/h      | \$ 37,700,000       |
| E-409  | \$ 464,000        | \$ 661,000       | Cooling Water       |            | 342000 MJ/h      | \$ 1,010,000        |
| H-401  | \$ 12,200,000     | \$ 17,400,000    | Natural Gas         | 0.9        | 353000 MJ/h      | \$ 32,610,000       |
| P-401  | \$ 40,300         | \$ 54,200        | Electricity         | 0.8        | 8 kilowatts      | \$ 4,030            |
| P-402  | \$ 29,000         | \$ 39,000        | Electricity         | 0.8        | 1.25 kilowatts   | \$ 629              |
| P-403  | \$ 31,100         | \$ 41,800        | Electricity         | 0.8        | 2.5 kilowatts    | \$ 1,260            |
| P-404  | \$ 29,000         | \$ 39,000        | Electricity         | 0.8        | 0.475 kilowatts  | \$ 239              |
| P-405  | \$ 29,000         | \$ 39,000        | Electricity         | 0.8        | 0.938 kilowatts  | \$ 472              |
| P-406  | \$ 32,600         | \$ 43,800        | Electricity         | 0.8        | 3.31 kilowatts   | \$ 1,670            |
| T-401  | \$ 1,460,000      | \$ 1,960,000     | NA                  |            |                  |                     |
| T-402  | \$ 2,450,000      | \$ 3,180,000     | NA                  |            |                  |                     |
| V-401  | \$ 101,000        | \$ 143,000       | NA                  |            |                  |                     |
| V-402  | \$ 32,700         | \$ 46,600        | NA                  |            |                  |                     |
| V-403  | \$ 32,700         | \$ 46,600        | NA                  |            |                  |                     |
| V-404  | \$ 1,500,000      | \$ 1,820,000     | NA                  |            |                  |                     |
| V-405  | \$ 1,500,000      | \$ 1,820,000     | NA                  |            |                  |                     |
| V-406  | \$ 1,500,000      | \$ 1,820,000     | NA                  |            |                  |                     |
| V-407  | \$ 1,500,000      | \$ 1,820,000     | NA                  |            |                  |                     |
| V-408  | \$ 1,500,000      | \$ 1,820,000     | NA                  |            |                  |                     |
| V-409  | \$ 1,070,000      | \$ 1,300,000     | NA                  |            |                  |                     |
| V-410  | \$ 1,070,000      | \$ 1,300,000     | NA                  |            |                  |                     |
| V-411  | \$ 1,070,000      | \$ 1,300,000     | NA                  |            |                  |                     |
| V-412  | \$ 1,070,000      | \$ 1,300,000     | NA                  |            |                  |                     |
| V-413  | \$ 1,070,000      | \$ 1,300,000     | NA                  |            |                  |                     |
| Totals | \$ 39,000,000     | \$ 51,600,000    | $FCI = C_{TM}$      |            |                  |                     |
|        |                   |                  | $C_{UT}$            |            |                  |                     |
|        |                   |                  | \$ 64,700,000       |            |                  |                     |

Note that full credit for the HPS and LPS in Exchangers E-403 and E-404 has been taken and that credit for the high temperature steam from E-402 has been taken as LPS (which is low).

$$N_{OL} = [6.29 + 31.7P^2 + 0.23N_{np}]^{0.5} = [6.29 + 0.23(24)]^{0.5} = 3.44 \frac{\text{operators}}{\text{shift}}$$

$$N_{np} = (1 \text{ compressor} + 1 \text{ drive} + 9 \text{ exchangers} + 1 \text{ heater} + 2 \text{ towers} + 10 \text{ reactors}) = 24$$

$$N_{operators} = (4.5)(3.44) = 15.5 \text{ round up to } 16$$

$$C_{OL} = (16)(\$52,900) = (16)(\$52,900) = \$846,000/\text{yr}$$

$$\text{Assume } 1 \text{ yr} = 8200 \text{ h}$$

$$C_{RM, \text{ethylbenzene}} = (1.069)(19417)(8200) = \$170,210,000/\text{yr}$$

$$FCI \text{ (from Problem 7.22 CAPCOST Output for } C_{TM}) = \$39,000,000$$

$$C_{UT} = \$64,700,000/\text{yr}$$

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) = 0.180(39.0) + 2.73(0.846) + 1.23(64.7 + 0 + 170.21) = \$298,300,000/\text{yr}$$

$$COM_d = \$298,300,000/\text{yr}$$

$$COM_d = [\$298,300,000/\text{yr}]/[(8200 \text{ h/yr})(12,432 \text{ kg/h})] = \$2.93/\text{kg-styrene}$$

Note that the  $COM_d$  for styrene is \$1.40 greater than the selling prices from Table 8.4. We must greatly improve integration of Styrene and EB processes to make styrene production viable.

## 8.16 Operating Costs for the Drying Oil Process – Project B.4

| Name          | Total Module Cost   | Grass Roots Cost    | Utility Used        | Efficiency | Actual Usage    | Annual Utility Cost |
|---------------|---------------------|---------------------|---------------------|------------|-----------------|---------------------|
| E-501         | \$ 168,511          | \$ 207,000          | Low-Pressure Steam  |            | -6329.002 MJ/h  | \$ (699,457)        |
| E-502         | \$ 191,000          | \$ 235,000          | Unspecified         |            |                 |                     |
| E-503         | \$ 182,000          | \$ 223,000          | Cooling Water       |            | 1030 MJ/h       | \$ 3,030            |
| E-504         | \$ 205,000          | \$ 250,000          | High-Pressure Steam |            | 719 MJ/h        | \$ 105,910          |
| E-505         | \$ 182,000          | \$ 223,000          | Cooling Water       |            | 230 MJ/h        | \$ 680              |
| E-506         | \$ 1,173,000        | \$ 1,440,000        | Low-Pressure Steam  |            | -4961.999 MJ/h  | \$ (548,381)        |
| H-501         | \$ 1,900,000        | \$ 2,700,000        | Natural Gas         | 0.9        | 14700 MJ/h      | \$ 1,221,000        |
| P-501         | \$ 29,000           | \$ 39,000           | Electricity         | 0.8        | 1.125 kilowatts | \$ 566              |
| P-502         | \$ 36,200           | \$ 46,200           | Electricity         | 0.8        | 1.25 kilowatts  | \$ 629              |
| P-503         | \$ 36,200           | \$ 46,200           | Electricity         | 0.8        | 1 kilowatts     | \$ 503              |
| P-504         | \$ 36,200           | \$ 46,200           | Electricity         | 0.8        | 0.375 kilowatts | \$ 189              |
| T-501         | \$ 683,000          | \$ 871,000          | NA                  |            |                 |                     |
| T-502         | \$ 144,000          | \$ 176,000          | NA                  |            |                 |                     |
| V-501         | \$ 22,900           | \$ 32,600           | NA                  |            |                 |                     |
| V-502         | \$ 63,100           | \$ 76,900           | NA                  |            |                 |                     |
| V-503         | \$ 47,300           | \$ 57,000           | NA                  |            |                 |                     |
| V-504         | \$ 22,200           | \$ 26,800           | NA                  |            |                 |                     |
| V-505         | \$ 39,700           | \$ 48,300           | NA                  |            |                 |                     |
| <b>Totals</b> | <b>\$ 5,160,000</b> | <b>\$ 6,740,000</b> |                     |            |                 |                     |

$C_{UT}$

**\$ 84,670**

Note that the utility cost for the Dowtherm exchanger (E-402) is accounted for in the natural gas for the fired heater and lps credit is given for E-501 and E-506.

$$N_{OL} = [6.29 + 31.7P^2 + 0.23N_{np}]^{0.5} = [6.29 + 0.23(10)]^{0.5} = 2.93 \frac{\text{operators}}{\text{shift}}$$

$$N_{np} = (6 \text{ exchangers} + 1 \text{ heater} + 2 \text{ towers} + 1 \text{ reactors}) = 10$$

$$N_{operators} = (4.5)(2.93) = 13.2 \text{ round up to } 14$$

$$C_{OL} = (14)(\$52,900) = \$741,000/\text{yr}$$

Assume 1 yr = 8200 h

$$C_{RM, ACO} = (\$1.764/\text{kg})(1628.7 \text{ kg/h})(8200 \text{ h/yr}) = \$23,560,000/\text{yr}$$

Use value of \$1.764/kg (\$0.80/lb) from <http://www.icis.com/StaticPages/a-e.htm#top> for acetylated castor oil

$$FCI \text{ (from Problem 7.23 CAPCOST Output for } C_{TM}) = \$5,160,000$$

$$C_{UT} = \$84,700/\text{yr}$$

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) = 0.180(5.16) + 2.73(0.741) + 1.23(.085 + 0 + 23.56) = \$32,030,000/\text{yr}$$

$COM_d = \$32,030,000/\text{yr}$

$$COM_d = [\$32,030,000/\text{yr}] / [(8200 \text{ h/yr})(1250.04)] = \$3.13/\text{kg-DO}$$

Note that acetic acid is also produced and so revenue from this byproduct reduces the net COM for Drying Oil (DO).

### 8.17 Operating Costs for the Maleic Anhydride Process – Project B.5

| Name   | Total Module Cost | Grass Roots Cost | Utility Used        | Efficiency | Actual Usage    | Annual Utility Cost |
|--------|-------------------|------------------|---------------------|------------|-----------------|---------------------|
| C-601  | \$ 3,130,000      | \$ 4,460,000     | NA                  |            |                 |                     |
| D-601  | \$ 469,000        | \$ 668,000       | Electricity         | 0.95       | 3270 kilowatts  | \$ 1,650,000        |
| E-601  | \$ 171,273        | \$ 210,000       | Low-Pressure Steam  |            | 1750 MJ/h       | \$ 193,400          |
| E-602  | \$ 202,000        | \$ 246,000       | High-Pressure Steam |            | -16700 MJ/h     | \$ (2,459,900)      |
| E-603  | \$ 2,323,000      | \$ 2,830,000     | High-Pressure Steam |            | -31400.003 MJ/h | \$ (4,625,201)      |
| E-604  | \$ 960,000        | \$ 1,180,000     | Cooling Water       |            | 86900 MJ/h      | \$ 256,000          |
| E-605  | \$ 272,000        | \$ 331,000       | High-Pressure Steam |            | 19200 MJ/h      | \$ 2,821,000        |
| E-606  | \$ 177,000        | \$ 217,000       | Cooling Water       |            | 3050 MJ/h       | \$ 9,000            |
| E-607  | \$ 365,000        | \$ 438,000       | Unspecified         |            |                 |                     |
| H-601  | \$ 2,540,000      | \$ 3,620,000     | Natural Gas         | 0.9        | 29800 MJ/h      | \$ 2,753,000        |
| P-601  | \$ 29,000         | \$ 39,000        | Electricity         | 0.8        | 0.375 kilowatts | \$ 189              |
| P-602  | \$ 43,800         | \$ 55,900        | Electricity         | 0.8        | 4.75 kilowatts  | \$ 2,390            |
| P-603  | \$ 71,300         | \$ 91,000        | Electricity         | 0.8        | 0.125 kilowatts | \$ 63               |
| P-604  | \$ 51,100         | \$ 65,300        | Electricity         | 0.8        | 8.44 kilowatts  | \$ 4,250            |
| P-605  | \$ 36,200         | \$ 46,200        | Electricity         | 0.8        | 0.875 kilowatts | \$ 440              |
| P-606  | \$ 39,900         | \$ 50,900        | Electricity         | 0.8        | 3 kilowatts     | \$ 1,510            |
| T-601  | \$ 2,010,000      | \$ 2,440,000     | NA                  |            |                 |                     |
| T-602  | \$ 286,000        | \$ 350,000       | NA                  |            |                 |                     |
| V-601  | \$ 28,500         | \$ 40,500        | NA                  |            |                 |                     |
| V-602  | \$ 870,000        | \$ 1,050,000     | NA                  |            |                 |                     |
| V-603  | \$ 68,800         | \$ 83,000        | NA                  |            |                 |                     |
| V-604  | \$ 900,000        | \$ 1,040,000     | NA                  |            |                 |                     |
| Totals | \$ 15,000,000     | \$ 19,600,000    | $FCI = C_{TM}$      |            |                 |                     |
|        |                   |                  | $C_{UT}$            |            |                 |                     |
|        |                   |                  | \$ 610,000          |            |                 |                     |

Note that credit is taken for hps in E-602 and E-603

$$N_{OL} = [6.29 + 31.7P^2 + 0.23N_{np}]^{0.5} = [6.29 + 0.23(13)]^{0.5} = 3.05 \frac{\text{operators}}{\text{shift}}$$

$$N_{np} = (1 \text{ compressor} + 1 \text{ drive} + 7 \text{ exchangers} + 1 \text{ heater} + 2 \text{ towers} + 1 \text{ reactor}) = 13$$

$$N_{operators} = (4.5)(3.05) = 13.7 \text{ round up to } 14$$

$$C_{OL} = (14)(\$52,900) = \$741,000/\text{yr}$$

Assume 1 yr = 8200 h

$$C_{RM, \text{benzene}} = (3304)(0.657)(8200) = \$17,800,000/\text{yr}$$

$$FCI \text{ (from Problem 7.24 CAPCOST Output for } C_{TM}) = \$15,000,000$$

$$C_{UT} = \$610,000/\text{yr}$$

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) = 0.180(15.0) + 2.73(0.741) + 1.23(0.61 + 0 + 17.80) = \$27,370,000/\text{yr}$$

$$COM_d = \$27,370,000/\text{yr}$$

$$COM_d = [\$27,370,000/\text{yr}]/[(8200 \text{ h/yr})(2597 \text{ kg/h})] = \$1.29/\text{kg-MA}$$

This is about \$0.30/kg less than value in Table 8.4



### 8.18 Operating Costs for the Ethylene Oxide Process – Project B.6

| Name   | Total Module Cost | Grass Roots Cost | Utility Used          | Efficiency | Actual Usage    | Annual Utility Cos |
|--------|-------------------|------------------|-----------------------|------------|-----------------|--------------------|
| C-701  | \$ 16,000,000     | \$ 22,800,000    | NA                    |            |                 |                    |
| C-702  | \$ 19,000,000     | \$ 27,000,000    | NA                    |            |                 |                    |
| C-703  | \$ 18,200,000     | \$ 25,900,000    | NA                    |            |                 |                    |
| C-704  | \$ 4,610,000      | \$ 6,570,000     | NA                    |            |                 |                    |
| C-705  | \$ 4,610,000      | \$ 6,570,000     | NA                    |            |                 |                    |
| D-701  | \$ 5,680,000      | \$ 8,080,000     | Electricity           | 0.9        | 21100 kilowatts | \$ 10,630,000      |
| D-702  | \$ 6,480,000      | \$ 9,230,000     | Electricity           | 0.9        | 25600 kilowatts | \$ 12,900,000      |
| D-703  | \$ 6,400,000      | \$ 9,110,000     | Electricity           | 0.9        | 23900 kilowatts | \$ 12,000,000      |
| D-704  | \$ 2,010,000      | \$ 2,860,000     | Electricity           | 0.9        | 6110 kilowatts  | \$ 3,080,000       |
| D-705  | \$ 2,010,000      | \$ 2,860,000     | Electricity           | 0.9        | 6110 kilowatts  | \$ 3,080,000       |
| E-701  | \$ 3,784,770      | \$ 5,390,000     | Cooling Water         |            | 58500 MJ/h      | \$ 172,000         |
| E-702  | \$ 4,280,000      | \$ 6,090,000     | Cooling Water         |            | 83200 MJ/h      | \$ 245,000         |
| E-703  | \$ 8,390,000      | \$ 11,900,000    | High-Pressure Steam   |            | 148000 MJ/h     | \$ 21,740,000      |
| E-704  | \$ 9,800,000      | \$ 13,900,000    | Cooling Water         |            | 210000 MJ/h     | \$ 620,000         |
| E-705  | \$ 10,500,000     | \$ 14,500,000    | High-Pressure Steam   |            | 230000 MJ/h     | \$ 33,860,000      |
| E-706  | \$ 9,700,000      | \$ 13,700,000    | Cooling Water         |            | 207000 MJ/h     | \$ 610,000         |
| E-707  | \$ 1,000,000      | \$ 1,450,000     | Cooling Water         |            | 21500 MJ/h      | \$ 63,000          |
| E-708  | \$ 830,000        | \$ 990,000       | High-Pressure Steam   |            | 43800 MJ/h      | \$ 6,458,000       |
| E-709  | \$ 290,000        | \$ 354,000       | Cooling Water         |            | 14200 MJ/h      | \$ 42,000          |
| E-710  | \$ 16,043,000     | \$ 22,400,000    | Medium-Pressure Steam |            | -33101 MJ/h     | \$ (3,908,870)     |
| E-711  | \$ 13,800,000     | \$ 19,300,000    | Medium-Pressure Steam |            | -26179 MJ/h     | \$ (3,091,457)     |
| P-701  | \$ 44,300         | \$ 56,600        | Electricity           | 0.86       | 4.65 kilowatts  | \$ 2,340           |
| T-701  | \$ 9,600,000      | \$ 10,700,000    | NA                    |            |                 |                    |
| T-702  | \$ 9,600,000      | \$ 10,700,000    | NA                    |            |                 |                    |
| T-703  | \$ 117,300,000    | \$ 134,800,000   | NA                    |            |                 |                    |
| V-701  | \$ 224,000        | \$ 247,000       | NA                    |            |                 |                    |
| V-702  | \$ 4,490,000      | \$ 4,840,000     | NA                    |            |                 |                    |
| V-703  | \$ 4,490,000      | \$ 4,840,000     | NA                    |            |                 |                    |
| Totals | \$ 309,200,000    | \$ 397,100,000   |                       |            |                 |                    |

FCI

$C_{UT}$

\$ 98,500,000

Note that credit is taken for hps in R-701 and R-702 that correspond to E-710 and E711. This process looses enormous amounts of useful energy in E-704 and E-706. Heat integration can significantly reduce the utility burden. A turbine in Stream 29 could also reduce the electrical utilities. See note at bottom

$$N_{OL} = \left[ 6.29 + 31.7P^2 + 0.23N_{np} \right]^{0.5} = \left[ 6.29 + 0.23(24) \right]^{0.5} = 3.44 \frac{\text{operators}}{\text{shift}}$$

$$N_{np} = (5 \text{ compressors} + 5 \text{ drives} + 9 \text{ exchangers} + 3 \text{ towers} + 2 \text{ reactors}) = 24$$

$$N_{operators} = (4.5)(3.44) = 15.5 \text{ round up to } 16$$

$$C_{OL} = (16)(\$52,900) = \$846,000/\text{yr}$$

$$\text{Assume } 1 \text{ yr} = 8200 \text{ h}$$

$$C_{RM, \text{ethylene}} = (20,000)(1.202)(8200) = \$197,130,000/\text{yr}$$

$$FCI \text{ (from Problem 7.25 CAPCOST Output for } C_{TM}) = \$309,200,000$$

$$C_{UT} = \$98,500,000/\text{yr}$$

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) = 0.180(309.2) + 2.73(0.846) + 1.23(98.5 + 0 + 197.13) = \$421,600,000/\text{yr}$$

$$COM_d = \$421,600,000/\text{yr}$$

$$COM_d = [\$421,600,000/\text{yr}] / [(8200\text{h}/\text{yr})(20,000 \text{ kg}/\text{h})] = \$2.57/\text{kg}$$

This is about \$0.80/kg more than the selling price – thus we need to improve the energy integration significantly – see note below utility table.

## 8.19 Operating Costs for the Formalin Process – Project B.7

| Name          | Total Module Cost   | Grass Roots Cost    | Utility Used          | Efficiency | Actual Usage    | Annual Utility Cos  |
|---------------|---------------------|---------------------|-----------------------|------------|-----------------|---------------------|
| C-801         | \$ 610,000          | \$ 869,000          | NA                    |            |                 |                     |
| D-801         | \$ 126,000          | \$ 179,000          | Electricity           | 0.65       | 282 kilowatts   | \$ 141,700          |
| E-801         | \$ 295,000          | \$ 420,000          | Medium-Pressure Steam |            | 4110 MJ/h       | \$ 485,500          |
| E-802         | \$ 100,000          | \$ 141,000          | High-Pressure Steam   |            | 76.8 MJ/h       | \$ 11,305           |
| E-803         | \$ 90,000           | \$ 129,000          | Cooling Water         |            | 983 MJ/h        | \$ 2,900            |
| E-804         | \$ 378,000          | \$ 463,000          | Medium-Pressure Steam |            | 37800 MJ/h      | \$ 4,458,000        |
| E-805         | \$ 404,000          | \$ 495,000          | Cooling Water         |            | 32500 MJ/h      | \$ 96,000           |
| E-806         | \$ 177,000          | \$ 218,000          | Cooling Water         |            | 1170 MJ/h       | \$ 3,450            |
| E-807         | \$ 146,000          | \$ 208,000          | Medium-Pressure Steam |            | -8928 MJ/h      | \$ (1,054,300)      |
| P-801         | \$ 29,000           | \$ 39,000           | Electricity           | 0.8        | 0.375 kilowatts | \$ 189              |
| P-802         | \$ 37,900           | \$ 48,400           | Electricity           | 0.8        | 2.13 kilowatts  | \$ 1,070            |
| P-803         | \$ 36,200           | \$ 46,200           | Electricity           | 0.75       | 0.667 kilowatts | \$ 336              |
| T-801         | \$ 58,000           | \$ 78,500           | NA                    |            |                 |                     |
| T-802         | \$ 1,110,000        | \$ 1,330,000        | NA                    |            |                 |                     |
| V-801         | \$ 30,000           | \$ 42,800           | NA                    |            |                 |                     |
| <b>Totals</b> | <b>\$ 3,630,000</b> | <b>\$ 4,710,000</b> |                       |            |                 | <b>\$ 4,150,000</b> |

Note that credit for the mps generated in the reactor is taken account of in E-807, which represents the reactor exchanger.

$$N_{OL} = \left[ 6.29 + 31.7P^2 + 0.23N_{np} \right]^{0.5} = \left[ 6.29 + 0.23(10) \right]^{0.5} = 2.93 \frac{\text{operators}}{\text{shift}}$$

$$N_{np} = (1 \text{ compressor} + 1 \text{ drive} + 5 \text{ exchangers} + 2 \text{ towers} + 1 \text{ reactor}) = 10$$

$$N_{operators} = (4.5)(2.93) = 13.2 \text{ round up to } 14$$

$$C_{OL} = (14)(\$52,900) = \$741,000/\text{yr}$$

$$\text{Assume } 1 \text{ yr} = 8200 \text{ h}$$

$$C_{RM, \text{methanol}} = (2464.75)(0.294)(8200) = \$5,940,000/\text{yr}$$

$$FCI \text{ (from Problem 7.25 CAPCOST Output for } C_{TM}) = \$3,630,000$$

$$C_{UT} = \$4,150,000/\text{yr}$$

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) = 0.180(3.63) + 2.73(0.741) + 1.23(4.15 + 0 + 5.94) = \$15,090,000/\text{yr}$$

$$COM_d = \$15,090,000/\text{yr}$$

$$COM_d = [\$15,090,000/\text{yr}] / [(8200 \text{ h/yr})(3897.06)] = \$0.472/\text{kg-formalin}$$

This is about the same price as the formalin with 6% methanol inhibitor given in Table 8.4

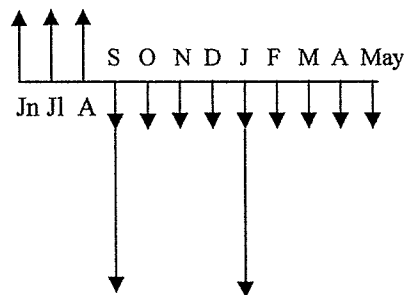
## Chapter 9

### Chapter 9 (Short Answers)

9.1 Simple interest is calculated such that the interest is based on the original principal. Compound interest is calculated such that the interest is based on the accrued principal including previously paid interest, so that there is interest on interest.

9.2 The nominal interest rate is a number based on interest payments once per year; however, if interest is compounded multiple times per year, the interest rate for the period is the nominal rate divided by the number of compounding periods per year. The effective annual interest rate involves using the result of compounding of interest several times per year to calculate an interest rate as if there was one interest payment at the end of the year. These are equal only when there is one compounding period per year.

9.3



9.4 An annuity is a uniform (constant) series of transactions at the same interval. Examples include a loan payment, a fixed monthly deduction from a paycheck into savings, etc.

9.5  $525/485 = 1.0825 \Rightarrow$  so the inflation rate is 8.25%

9.6 Interest is the return on an investment or the charge for borrowing money. Inflation is the increase in cost over time of goods, commodities, and services, which is equivalent to the decrease in purchasing power of money. The time value of money includes the effect of interest in an investment or cost. The time value of money is identical to interest, but it has nothing to do with inflation.

9.7 This term is depreciation

9.8 Depreciation is an accounting procedure that allows a company to reduce their taxable income based on capital expenditures (such as new construction, improvement projects). In a new chemical plant, depreciation reduces the taxable income at the beginning of the project life, so the profitability of the new plant is increased.

9.9 The time value of money suggests that “a dollar today is worth more than a dollar tomorrow.” Therefore, the sooner depreciation occurs, the higher the net cash flow is to the company.

9.10 A company can only use depreciation to offset revenue. Depreciation can never exceed profit. Therefore, a small company, with only one project, may have to defer depreciation if it exceeds their profit from the project. A large company usually has so many profitable projects, that they can use the depreciation in one project that may exceed that project’s profits to offset profits elsewhere in the company.

9.11 After-tax profit is the net profit from the project taking into account all expenses including depreciation. After-tax cash flow is the net cash flow generated by the project after taxes. These two quantities are the same when no depreciation is taken, i.e., in years when the book value of the assets is zero.

9.12  $P = \$1,000$

$$F = \$1,000(1 + i_{eff})$$

(a)  $F = \$1,000 + \$5(52 \text{ weeks}) = \$1,260$

$$\$1,260 = \$1,000(1 + i_{eff})$$

$$i_{eff} = 0.26 = 26\%$$

(b)  $F_n = P(1 + i)^n$

$$\$1,260 = \$1,000 \left( 1 + \frac{i}{12} \right)^{12}$$

$$i = 0.23 = 23\%$$

(c) The interest paid to Paulie is \$5/wk or \$21.67/month (average of 4.33 wks/month). The interest paid to the bank is  $\$1,000(0.07/12) = \$5.83/\text{month}$ . So no matter how long it takes to repay the loan, the interest paid to Paulie will always be greater than that paid to the bank.

9.13

$$(a) \quad \frac{F}{P} = \left(1 + \frac{i}{m}\right)^{mn}$$

$$P = \$1,000$$

$$\text{Daily: } F = \$1,000 \left(1 + \frac{0.095}{365}\right)^{365} = \$1,099.65$$

$$\text{Monthly: } F = \$1,000 \left(1 + \frac{0.10}{12}\right)^{12} = \$1,104.71$$

$$\text{Quarterly: } F = \$1,000 \left(1 + \frac{0.105}{4}\right)^4 = \$1,109.21$$

10.5% p.a. compounded quarterly is the most profitable scheme.

$$(b) \quad F = P(1 + i_{eff})$$

$$\$1,109.21 = \$1,000(1 + i_{eff})$$

$$i_{eff} = .109 = 10.9\%$$

$$(c) \quad i_{eff} = e^{i_{nom}} - 1$$

$$i_{eff} = e^{0.105} - 1$$

$$i_{eff} = 0.111 = 11.1\%$$

$$9.14 \quad F = P(1+i)^n$$

$$F = 2P$$

$$2P = P(1+i)^n$$

$$n = \frac{\ln 2}{\ln P(1+i)} \Rightarrow P = 1$$

| $i$   | $n = 72/i$ | $i_{eff}$ | % error |
|-------|------------|-----------|---------|
| 0.05  | 14.2       | 14.4      | 1.34%   |
| 0.075 | 9.58       | 9.58      | 0.21%   |
| 0.10  | 7.27       | 7.27      | 0.96%   |

The accuracy of the “rule of 72” improves as  $i$  decreases. For today’s investment it is a very accurate approximation.

$$9.15 \quad i_{eff} = \left(1 + \frac{i}{m}\right)^{mn} - 1$$

| $i$  | $i_{eff}$ |
|------|-----------|
| 0.04 | 0.0408    |
| 0.05 | 0.0513    |
| 0.06 | 0.0618    |
| 0.07 | 0.0725    |
| 0.08 | 0.0833    |

$$F = \$5,000(1 + 0.0408)^3(1 + 0.0513)^2(1 + 0.0618)(1 + 0.0725)(1 + 0.0833)$$

$$F = \$7,686.03$$

$$9.16 \quad \frac{F}{P} = \left(1 + \frac{i}{m}\right)^{mn}; F_{cont} = Pe^{in}$$

$$F_{yearly} = \$15,000(1 + 0.09)^{15} = \$54,637.24$$

$$F_{quarterly} = \$15,000\left(1 + \frac{0.09}{4}\right)^{(4)(15)} = \$57,002.02$$

$$F_{monthly} = \$15,000\left(1 + \frac{0.09}{12}\right)^{(12)(15)} = \$57,570.65$$

$$F_{daily} = \$15,000\left(1 + \frac{0.09}{365}\right)^{(365)(15)} = \$57,851.75$$

$$F_{cont.} = e^{(0.09)(15)}(\$15,000) = \$57,861.38$$

$$9.17 \quad \frac{F}{P} = \left(1 + \frac{i}{m}\right)^{mn}$$

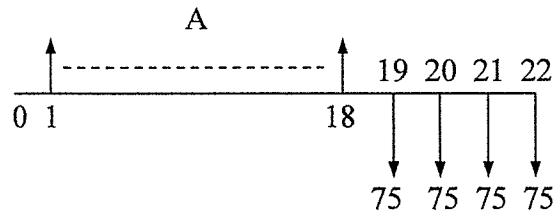
If  $P = \$1$

$$\text{Bank 1: } F = \$1 \left(1 + \frac{0.04}{365}\right)^{365} = \$1.0408$$

$$\text{Bank 2: } F = \$1(1 + 0.041) = \$1.041$$

9.18

(a)



$$(b) \quad \text{For years 1-18: } \frac{F}{A} = \frac{(1+i)^n - 1}{i} \Rightarrow F = A \frac{(1+0.08)^{18} - 1}{0.08}$$

$$\text{For years 19-22: } \frac{P}{A} = \frac{(1+i)^n - 1}{i(1+i)^n} \Rightarrow P = \$75,000 \frac{(1+0.08)^4 - 1}{0.08(1+0.08)^4}$$

$$A \frac{(1.08)^{18} - 1}{0.08} = \$75,000 \frac{1.08^4 - 1}{0.08(1.08)^4}$$

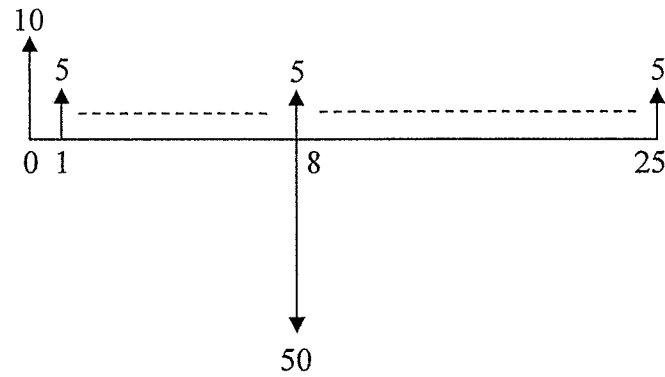
$$A = \$6,633.1 / \text{yr}$$

$$(c) \quad \$5,000 \frac{(1+i)^{18} - 1}{i} = \$75,000 \frac{(1+i)^4 - 1}{i(1+i)^4}$$

$$i = 0.103 = 10.3\%$$

9.19

(a)



(b)  $F_8 = \$10,000(1 + 0.08)^8 + \$5,000 \frac{(1 + 0.08)^8 - 1}{0.08}$

$F_8 = \$71,692$ ; Yes, you will have enough for the down payment.

(c)  $\frac{F}{P} = (1 + i)^n; \frac{F}{A} = \frac{(1 + i)^n - 1}{i}$

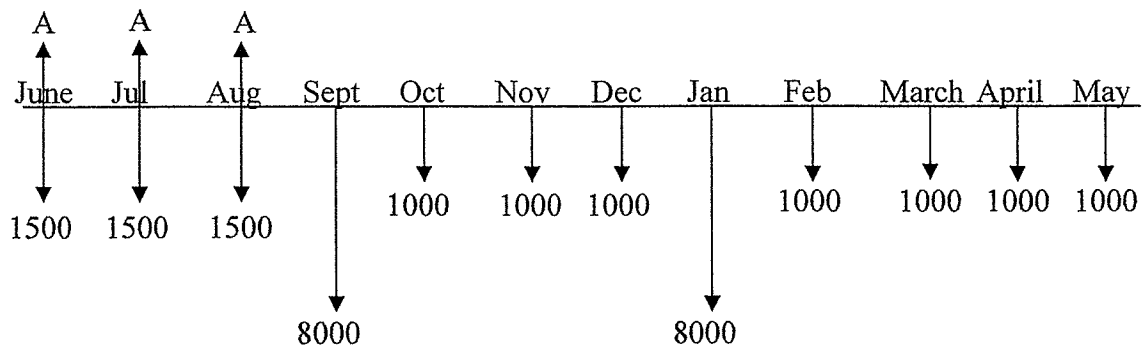
$$F_{25} = \$10,000(1 + 0.08)^{25} + \$5,000 \frac{(1 + 0.08)^{25} - 1}{0.08} - \$50,000(1 + 0.08)^{17}$$

$F_{25} = \$249,000$



9.20

(a)



$$(b) \quad \frac{F}{P} = (1+i)^n; \quad \frac{F}{A} = \frac{(1+i)^n - 1}{i}$$

$$F = (A - \$1,500) \left[ \left(1 + \frac{0.04}{12}\right)^{11} + \left(1 + \frac{0.04}{12}\right)^{10} + \left(1 + \frac{0.04}{12}\right)^9 \right]$$

$$F = \$7,000 \left[ \left(1 + \frac{0.04}{12}\right)^8 + \left(1 + \frac{0.04}{12}\right)^4 \right] + \$1,000 \frac{\left(1 + \frac{0.04}{12}\right)^9 - 1}{0.04}$$

$$(\$4,000 - \$1,500)(1.0373 + 1.0338 + 1.0304) = \$7,000(1.02698 + 1.0033) + \$1,000(9.1209)$$

$$\boxed{A = \$9,046}$$

(c)

$$(\$4,000 - \$1,500)(1.0373 + 1.0338 + 1.0304) + A \left( \frac{\left(1 + \frac{0.04}{12}\right)^9 - 1}{\frac{0.04}{12}} \right) = \$7,000(1.02698 + 1.0033) + \$1,000(9.1209)$$

$$\boxed{A = \$1,716}$$

(d) Finances while in college

9.21

$$(a) \quad F = P \left( 1 + \frac{i}{m} \right)^{nm}$$

$$F = \$1,500(1 + 0.045)^{15} + \$1,000(1 + 0.045)^{14} + \$750(1 + 0.045)^{13} + \\ \$1,000[(1 + 0.045)^{12} + (1 + 0.045)^{11} + (1 + 0.045)^{10} + (1 + 0.045)^9 + (1 + 0.045)^8 + (1 + 0.045)^7] - \\ \$2,000(1 + 0.045)^6 + \$700[(1 + 0.045)^5 + (1 + 0.045)^4] - \$500(1 + 0.045)^3 + \$1,000(1 + 0.045)^2 + \\ \$300(1 + 0.045) - \$2,000$$

$$F = \$10,504$$

$$(b) \quad F_{25} = \$10,504(1 + 0.045)^{10}$$

$$F_{25} = \$16,312$$

$$9.22 \quad \frac{F}{A} = \frac{(1+i)^n - 1}{i}$$

$$(a) \quad \$1,000,000 = \$6,000 \frac{(1+i)^{40} - 1}{i}$$

$$i = 0.0629 = 6.29\%$$

$$(b) \quad \$2,000,000 = \$6,000 \frac{(1+i)^{40} - 1}{i}$$

$$i = 0.0895 = 8.95\%$$

$$(c) \quad F = \$6,000 \frac{(1+0.07)^{40} - 1}{0.07}$$

$$F = \$1,197,811$$

9.23

$$(a) \quad F = \$4,000 \left( \frac{F}{A}, 0.09, 25 \right) \left( \frac{F}{P}, 0.09, 16 \right) + \$5,000 \left( \frac{F}{A}, 0.09, 16 \right)$$

$$F = \$4,000 \left( \frac{(1+0.09)^{25} - 1}{0.09} \right) (1+0.09)^{16} + \$5,000 \left( \frac{(1+0.09)^{16} - 1}{0.09} \right)$$

$$F = \$1,510,171$$

$$(b) \quad \$2,000,000 = \$4,000 \left( \frac{(1+i)^{41} - 1}{i} \right) + \$1,000 \left( \frac{(1+i)^{16} - 1}{i} \right)$$

$$i = 0.1002 = 10.02\%$$

9.24  $\frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$

$$A = \$25,000 \frac{\left( \frac{0.07}{12} \right) \left( 1 + \frac{0.07}{12} \right)^{(12)(3)}}{\left( 1 + \frac{0.07}{12} \right)^{(12)(3)} - 1}$$

$$A = \$772$$

$$A = \$250,000 \frac{\left( \frac{0.07}{12} \right) \left( 1 + \frac{0.07}{12} \right)^{(12)(4)}}{\left( 1 + \frac{0.07}{12} \right)^{(12)(4)} - 1}$$

$$A = \$599$$

$$A = \$250,000 \frac{\left( \frac{0.07}{12} \right) \left( 1 + \frac{0.07}{12} \right)^{(12)(5)}}{\left( 1 + \frac{0.07}{12} \right)^{(12)(5)} - 1}$$

$$A = \$495$$

$$9.25 \quad \frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$$

$$\$500 = \$20,000 \frac{\left(\frac{i}{12}\right) \left(1 + \frac{i}{12}\right)^{(12)(2)}}{\left(1 + \frac{i}{12}\right)^{(12)(2)} - 1}$$

$$i = 0.0924 = 9.24\%$$

$$9.26 \quad \frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$$

$$(a) \quad A = \$200,000 \frac{\left(\frac{0.065}{12}\right) \left(1 + \frac{0.065}{12}\right)^{(12)(30)}}{\left(1 + \frac{0.065}{12}\right)^{(12)(30)} - 1}$$

$$A = \$1,264.06$$

$$(b) \quad A = \$200,000 \frac{\left(\frac{0.065}{12}\right) \left(1 + \frac{0.065}{12}\right)^{(12)(15)}}{\left(1 + \frac{0.065}{12}\right)^{(12)(15)} - 1}$$

$$A = \$1,742.24$$

$$(c) \quad 30 \text{ years: } (360)(\$1,264.14) = \$455,090.40$$

$$15 \text{ years: } (180)(\$1,742.22) = \$313,599.60$$

$$\boxed{\$141,490.80}$$

$$9.27 \quad \frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$$

$$\$1,600 = \$225,000 \frac{\left(\frac{i}{12}\right) \left(1 + \frac{i}{12}\right)^{(12)(25)}}{\left(1 + \frac{i}{12}\right)^{(12)(25)} - 1}$$

$$i = 0.0707 = 7.07\%$$

$$i_{eff} = (1+i)^n - 1$$

$$i_{eff} = \left(1 + \frac{0.0707}{12}\right)^{12} - 1$$

$$i_{eff} = 0.073 = 7.3\%$$

$$9.28 \quad \frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$$

$$\$1,612 = \$250,000 \frac{\left(\frac{0.06}{12}\right) \left(1 + \frac{0.06}{12}\right)^n}{\left(1 + \frac{0.06}{12}\right)^n - 1}$$

$$n = 25 \text{ years}$$

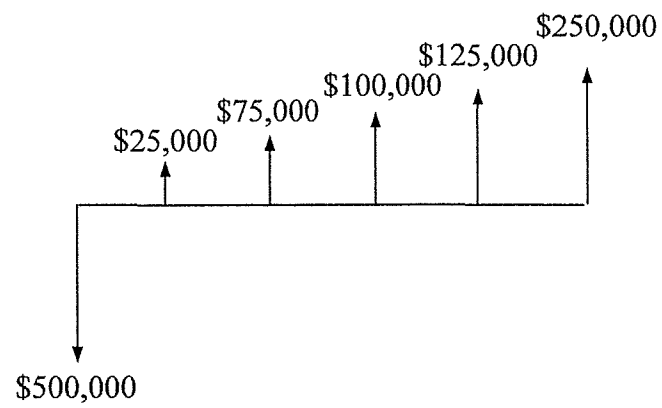
$$9.29 \quad \frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$$

$$A = \$50,000 \frac{\left(\frac{0.07}{12}\right) \left(1 + \frac{0.07}{12}\right)^{(12)(10)}}{\left(1 + \frac{0.07}{12}\right)^{(12)(10)} - 1}$$

$$A = \$580.54 + (0.0005)(\$50,000)$$

$$A = \$605.54$$

9.30



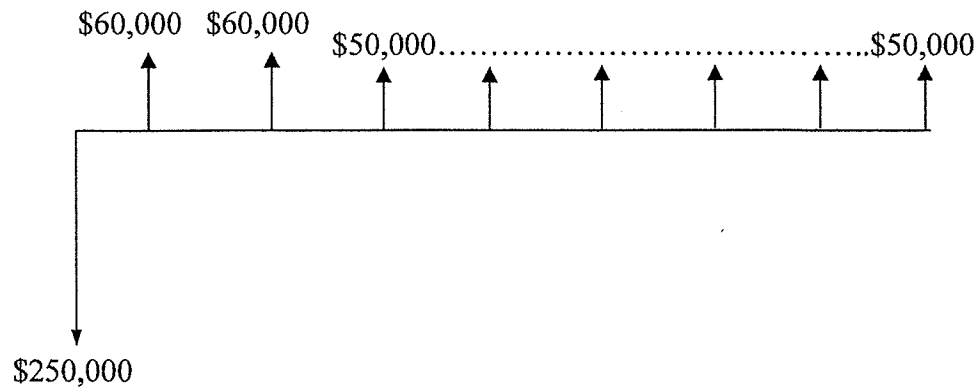
$$F = P(1+i)^n$$

$$F = \$25,000(1+i)^4 + \$75,000(1+i)^3 + \$100,000(1+i)^2 + \$125,000(1+i) +$$

$$\$250,000 = \$500,000(1+i)^5$$

$$i = 0.037 = 3.7\%$$

9.31 (a)



(b) Cumulative, discounted cash flow diagram

| Year | Cash Flow     | Discounted Cash Flow | Cumulative, Discounted Cash Flow |
|------|---------------|----------------------|----------------------------------|
| 0    | -\$250,000.00 | -\$250,000.00        | -\$250,000.00                    |
| 1    | \$60,000.00   | \$55,045.87          | -\$194,954.13                    |
| 2    | \$60,000.00   | \$50,500.80          | -\$144,453.33                    |
| 3    | \$50,000.00   | \$38,609.17          | -\$105,844.15                    |
| 4    | \$50,000.00   | \$35,421.26          | -\$70,422.89                     |
| 5    | \$50,000.00   | \$32,496.57          | -\$37,926.32                     |
| 6    | \$50,000.00   | \$29,813.37          | -\$8,112.96                      |
| 7    | \$50,000.00   | \$27,351.71          | \$19,238.75                      |
| 8    | \$50,000.00   | \$25,093.31          | \$44,332.07                      |

(c)  $F = P(1+i)^n$

$$F = \$44,332.07(1+0.09)^8$$

$$F = \$88,334$$

(d)  $F = \$250,000(1+i)^8 - \$250,000(1+0.09)^8 = \$88,334$

$$i = 0.1125 = 11.25\%$$

$$9.32 \text{ After Tax Cash Flow} = (R - d)(1 - t) + d$$

$$R = \frac{\left( \$60,000 - \frac{\$250,000}{8} \right)}{\left( (1 - 0.45) + \frac{\$250,000}{8} \right)} = \$83,522.73$$

$$R = \frac{\left( \$50,000 - \frac{\$250,000}{8} \right)}{\left( (1 - 0.45) + \frac{\$250,000}{8} \right)} = \$65,340.91$$

Before Tax Profit = Revenue - Depreciation

After Tax Revenue = (Before Tax Profit · 0.55) + Depreciation

Note: All values in millions of dollars.

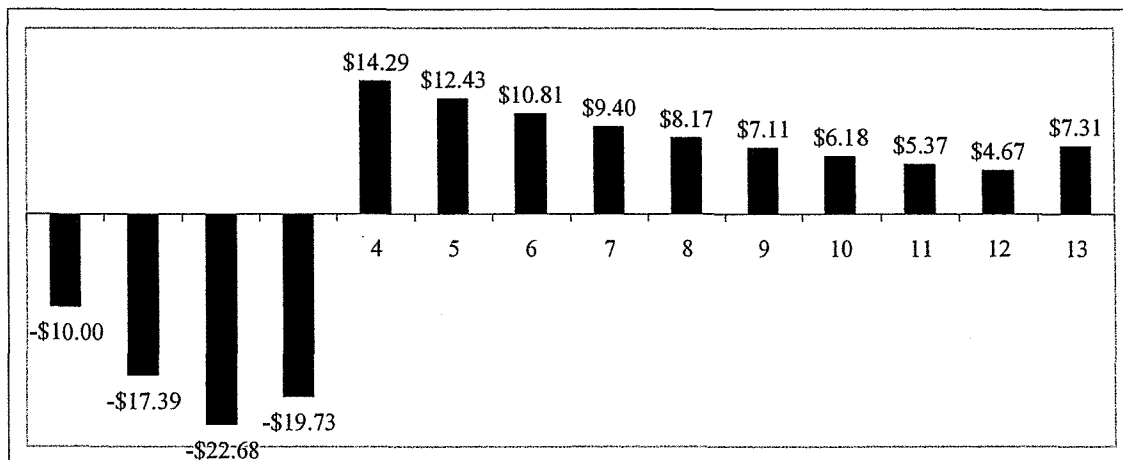
| Year | MACRS  | Depreciation | Revenue     | Before Tax Profit | After Tax Revenue |
|------|--------|--------------|-------------|-------------------|-------------------|
| 1    | 0.2    | \$50,000.00  | \$83,522.73 | \$33,522.73       | \$68,437.50       |
| 2    | 0.32   | \$80,000.00  | \$83,522.73 | \$3,522.73        | \$81,937.50       |
| 3    | 0.192  | \$48,000.00  | \$65,340.91 | \$17,340.91       | \$57,537.50       |
| 4    | 0.1152 | \$28,800.00  | \$65,340.91 | \$36,540.91       | \$48,897.50       |
| 5    | 0.1152 | \$28,800.00  | \$65,340.91 | \$36,540.91       | \$48,897.50       |
| 6    | 0.0576 | \$14,400.00  | \$65,340.91 | \$50,940.91       | \$42,417.50       |
| 7    |        |              | \$65,340.91 | \$65,340.91       | \$35,937.50       |
| 8    |        |              | \$65,340.91 | \$65,340.91       | \$35,937.50       |



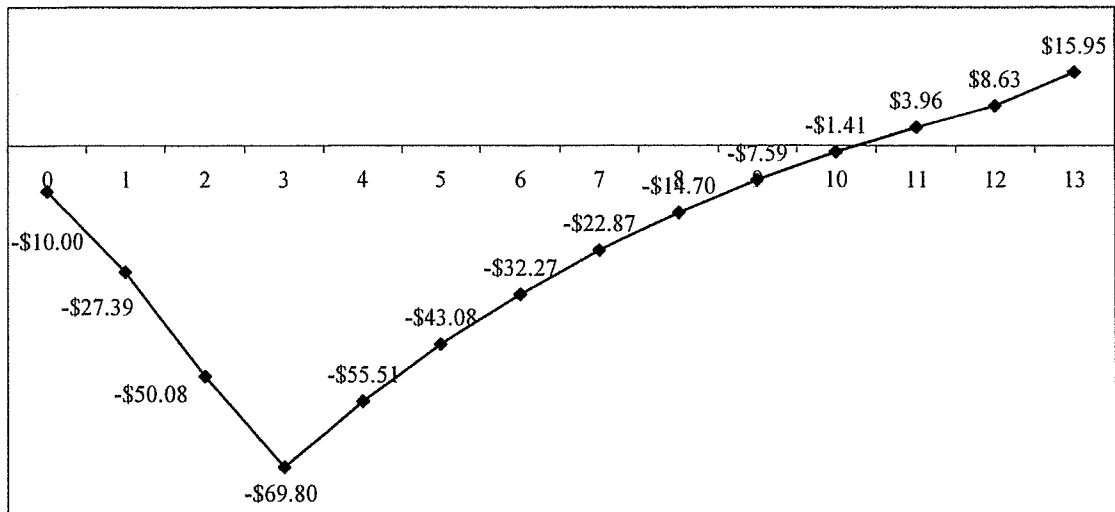
9.33

| Year | Cash Flow | Discounted Cash Flow | Cumulative, Discounted Cash Flow |
|------|-----------|----------------------|----------------------------------|
| 0    | -\$10.00  | -\$10.00             | -\$10.00                         |
| 1    | -\$20.00  | -\$17.39             | -\$27.39                         |
| 2    | -\$30.00  | -\$22.68             | -\$50.08                         |
| 3    | -\$30.00  | -\$19.73             | -\$69.80                         |
| 4    | \$25.00   | \$14.29              | -\$55.51                         |
| 5    | \$25.00   | \$12.43              | -\$43.08                         |
| 6    | \$25.00   | \$10.81              | -\$32.27                         |
| 7    | \$25.00   | \$9.40               | -\$22.87                         |
| 8    | \$25.00   | \$8.17               | -\$14.70                         |
| 9    | \$25.00   | \$7.11               | -\$7.59                          |
| 10   | \$25.00   | \$6.18               | -\$1.41                          |
| 11   | \$25.00   | \$5.37               | \$3.96                           |
| 12   | \$25.00   | \$4.67               | \$8.63                           |
| 13   | \$45.00   | \$7.31               | \$15.95                          |

(a)



(b)



$$(c) \quad NPV = -\$10 - \frac{\$20}{(1+0.15)} - \frac{\$30}{(1+0.15)^2} - \frac{\$30}{(1+0.15)^3} + \frac{\$25((1+0.15)^{10} - 1)}{(0.15)(1+0.15)^{10}} + \frac{\$20}{(1+0.15)^{13}}$$

$$NPV = \$15.9 \text{ million}$$

$$(d) \quad F = \$15.9(1+0.15)^{13}$$

$$F = \$97.8 \text{ million}$$

$$(e) \quad NPV = 0 = -\$10 - \frac{\$20}{(1+i)} - \frac{\$30}{(1+i)^2} - \frac{\$30}{(1+i)^3} + \frac{\$25((1+i)^{10} - 1)}{(0.15)(1+i)^{10}} + \frac{\$20}{(1+i)^{13}}$$

$$i = 0.192 = 19.2\%$$

(f) Proceed

$$9.34 \quad d_k^{SL} = \frac{\text{undepreciated capital}}{\text{remaining time for depreciation}}$$

$$d_k^{DDB} = \frac{2}{5} \left[ 100 - \sum_{j=0}^{k-1} d_j^{DDB} \right]$$

#### Four Year

| Year | DDB   | SL    | MACRS |
|------|-------|-------|-------|
| 1    | 25.00 |       | 25.00 |
| 2    | 37.50 | 21.43 | 37.50 |
| 3    | 18.75 | 15.00 | 18.75 |
| 4    | 9.38  | 12.50 | 12.50 |
| 5    | 2.34  | 6.25  | 6.25  |

#### Six Year

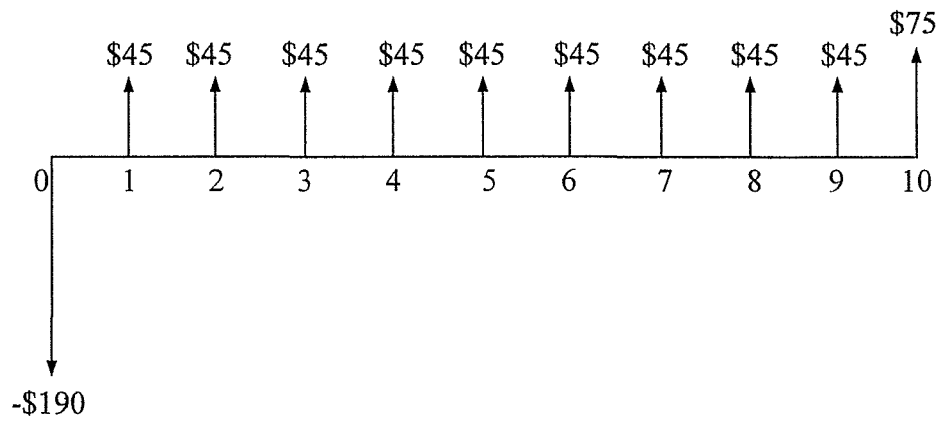
| Year | DDB   | SL    | MACRS |
|------|-------|-------|-------|
| 1    | 16.67 |       | 16.67 |
| 2    | 27.78 | 15.15 | 27.78 |
| 3    | 18.52 | 12.35 | 18.52 |
| 4    | 12.35 | 10.58 | 12.35 |
| 5    | 8.23  | 9.88  | 9.88  |
| 6    | 2.74  | 9.88  | 9.88  |
| 7    | 2.29  | 4.94  | 4.94  |

#### Nine Year

| Year | DDB   | SL    | MACRS |
|------|-------|-------|-------|
| 1    | 11.11 |       | 11.11 |
| 2    | 19.75 | 10.46 | 19.75 |
| 3    | 15.36 | 9.22  | 15.36 |
| 4    | 11.95 | 8.27  | 11.95 |
| 5    | 9.29  | 7.60  | 9.29  |
| 6    | 7.23  | 7.23  | 7.23  |
| 7    | 5.62  | 7.23  | 7.23  |
| 8    | 4.37  | 7.23  | 7.23  |
| 9    | 3.40  | 7.23  | 7.23  |
| 10   | 1.32  | 3.61  | 3.61  |

9.35

(a)

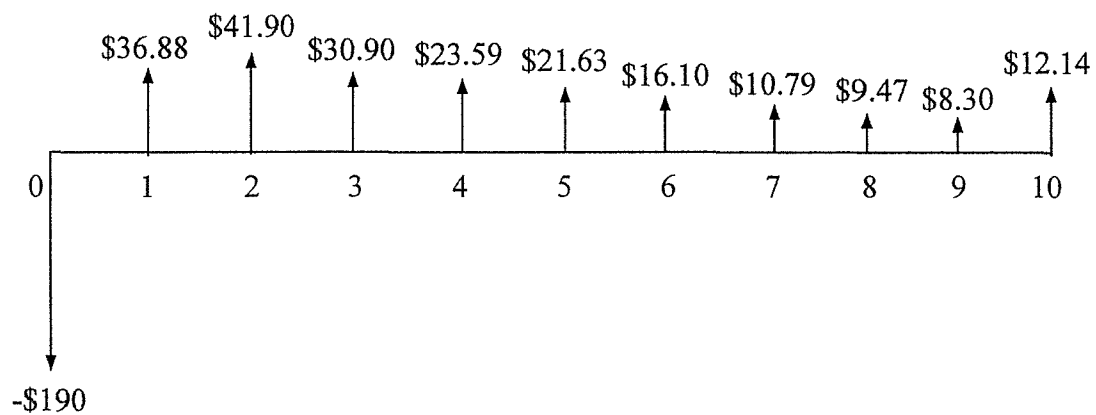


Amounts are in millions of dollars.

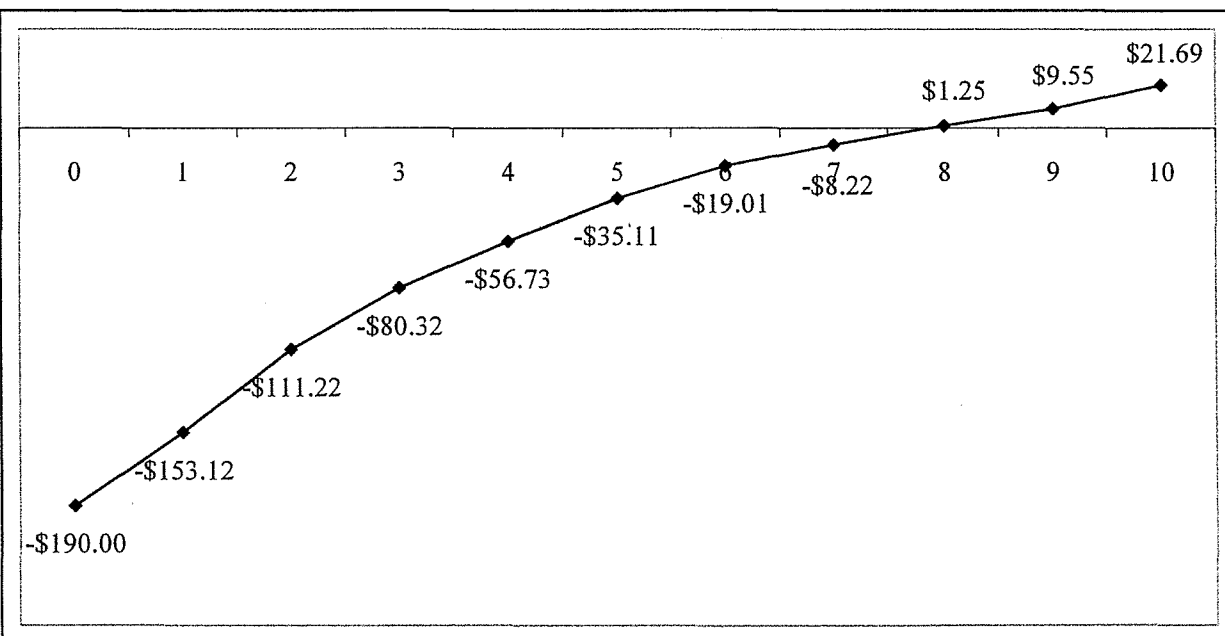
|         |      |
|---------|------|
| Land    | 10   |
| FCI     | 165  |
| WC      | 15   |
| Salvage | 15   |
| $i$     | 0.14 |
| $r$     | 0.4  |
| R       | 70   |
| COM     | 25   |

| Year | Investment | Cash Flow,<br>nondiscounted | R, discounted | COM <sub>d</sub> | MACRS  | (b)<br>$d$ | (c)<br>After Tax<br>Profit | (d) & (c)<br>After Tax<br>Cash Flow | (f)<br>Cumulative,<br>discounted cash flow |
|------|------------|-----------------------------|---------------|------------------|--------|------------|----------------------------|-------------------------------------|--------------------------------------------|
| 0    | -190.00    | -190.00                     |               |                  |        |            | 0.00                       | -190.00                             | -190.00                                    |
| 1    |            | 45.00                       | 61.40         | 21.93            | 0.2    | 33.00      | 3.88                       | 36.88                               | -153.12                                    |
| 2    |            | 45.00                       | 53.86         | 19.24            | 0.32   | 52.80      | -10.90                     | 41.90                               | -111.22                                    |
| 3    |            | 45.00                       | 47.25         | 16.87            | 0.192  | 31.68      | -0.78                      | 30.90                               | -80.32                                     |
| 4    |            | 45.00                       | 41.45         | 14.80            | 0.1152 | 19.01      | 4.58                       | 23.59                               | -56.73                                     |
| 5    |            | 45.00                       | 36.36         | 12.98            | 0.1152 | 19.01      | 2.62                       | 21.63                               | -35.11                                     |
| 6    |            | 45.00                       | 31.89         | 11.39            | 0.0576 | 9.50       | 6.60                       | 16.10                               | -19.01                                     |
| 7    |            | 45.00                       | 27.97         | 9.99             |        |            | 10.79                      | 10.79                               | -8.22                                      |
| 8    |            | 45.00                       | 24.54         | 8.76             |        |            | 9.47                       | 9.47                                | 1.25                                       |
| 9    |            | 45.00                       | 21.53         | 7.69             |        |            | 8.30                       | 8.30                                | 9.55                                       |
| 10   |            | 75.00                       | 26.97         | 6.74             |        |            | 12.14                      | 12.14                               | 21.69                                      |

(e)



(f)



(g) \$21.69 million – after-tax cash flow or cumulative cash flow at the end of year 10.

## Chapter 10

### Chapter 10 (short answers)

10.1  $NPV = 0$  at DCFROR. So, if use hurdle rate to calculate NPV, if  $NPV > 0$ , rate of return exceeds hurdle rate.

10.2 Choose the project with the greatest NPV.

10.3 If improvements are proposed for a process, an incremental economic analysis uses the capital cost of the improvements and the changes in the cost of manufacture (operating cost, revenue, etc.) to evaluate profitability. The entire process is not evaluated.

10.4 Yes. There would be a difference in the operating costs, which would affect the profitability.

10.5 EAOE or Common Denominator or Capitalized Cost methods but not NPV because equipment lives are different.

10.6 Risk cannot be eliminated. Monte Carlo simulation allows the risk to be quantified and analyzed.

10.7 Advantages include quantifying risk and thereby obtaining a more comprehensive picture of the effect of different factors on the profitability. Disadvantages include obtaining predictions of the variability of different parameters from historic data (difficult to predict the future).

10.8 The spreadsheet on the next page shows the cash flow diagrams.

(a) Non-Discounted Cash Flow Diagram shown on next page.

(b) (i) Cumulative cash position (CCP) = \$132 million  
Cumulative cash ratio (CCR) = 1.81

|                                   |
|-----------------------------------|
| CCP = \$132 million<br>CCR = 1.81 |
|-----------------------------------|

(ii) Payback period (PBP) = 4.0 years

|                 |
|-----------------|
| PBP = 4.0 years |
|-----------------|

(iii) Rate of return on investment (ROROI) = 11.00%

|                |
|----------------|
| ROROI = 11.00% |
|----------------|

(c) Discounted Cash Flow Diagram shown on next page.

(d) (i) Net present value (NPV) = \$1.71 million  
Present value ratio (PVR) = 1.01

|                                    |
|------------------------------------|
| NPV = \$1.71 million<br>PVR = 1.01 |
|------------------------------------|

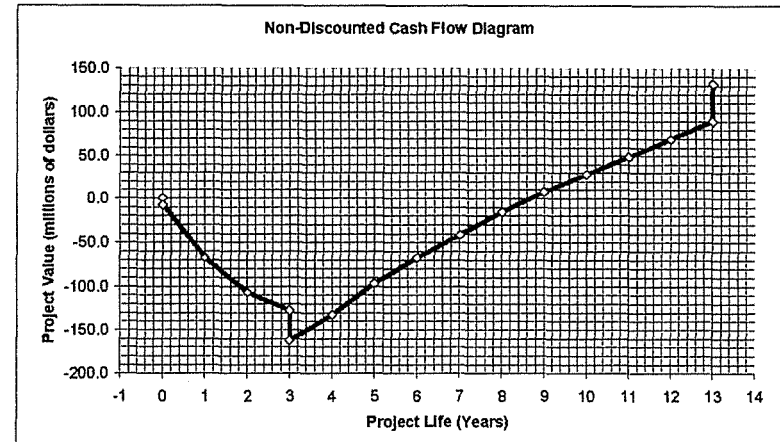
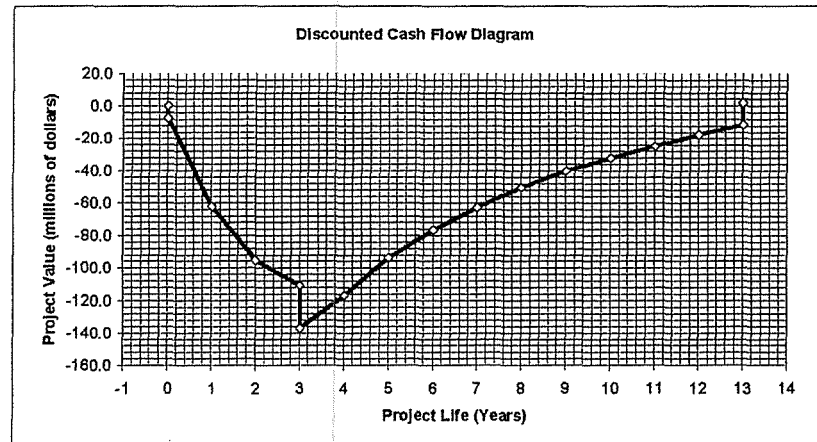
(ii) Discounted payback period (DPBP) = 6.7 years

|                  |
|------------------|
| DPBP = 6.7 years |
|------------------|

(iii) Discounted cash flow rate of return (DCFROR) = 9.73%

|                |
|----------------|
| DCFROR = 9.73% |
|----------------|





Generate CFD

**Discounted Profitability Criterion**

|                                     |       |
|-------------------------------------|-------|
| Net Present Value (millions)        | 1.71  |
| Discounted Cash Flow Rate of Return | 9.73% |
| Discounted Payback Period (years)   | 6.7   |

**Non-Discounted Profitability Criteria**

|                                     |        |
|-------------------------------------|--------|
| Cumulative Cash Position (millions) | 132.00 |
| Rate of Return on Investment        | 11.00% |
| Payback Period (years)              | 4.0    |

| Year | Investment | $d_t$ | $FCI_t - Sd_t$ | R     | $COM_t$ | $(R - COM_t - d_t) / (1 - t) + d_t$ | Cash Flow (discounted) | Cash Flow (non-discounted) | Cumulative Cash Flow (discounted) | Cumulative Cash Flow (non-discounted) |
|------|------------|-------|----------------|-------|---------|-------------------------------------|------------------------|----------------------------|-----------------------------------|---------------------------------------|
| 0    | 0.00       |       | 120.00         |       |         |                                     | 0.00                   | 0.00                       | 0.00                              | 0.00                                  |
| 0    | 7.50       |       | 120.00         |       |         |                                     | (7.50)                 | (7.50)                     | (7.50)                            | (7.50)                                |
| 1    | 60.00      |       | 120.00         |       |         |                                     | (60.00)                | (54.79)                    | (62.29)                           | (67.50)                               |
| 2    | 39.60      |       | 120.00         |       |         |                                     | (39.60)                | (33.03)                    | (95.32)                           | (107.10)                              |
| 3    | 20.40      |       | 120.00         |       |         |                                     | (20.40)                | (15.54)                    | (110.86)                          | (127.50)                              |
| 3    | 35.00      |       | 120.00         |       |         |                                     | (35.00)                | (28.66)                    | (137.52)                          | (162.50)                              |
| 4    |            | 24.00 | 96.00          | 52.00 | 18.00   | 30.00                               | 30.00                  | 20.87                      | (116.85)                          | (132.50)                              |
| 5    |            | 38.40 | 57.60          | 52.00 | 18.00   | 35.76                               | 35.76                  | 22.72                      | (93.93)                           | (96.74)                               |
| 6    |            | 23.04 | 34.56          | 52.00 | 18.00   | 29.62                               | 29.62                  | 17.18                      | (76.75)                           | (67.12)                               |
| 7    |            | 13.80 | 20.76          | 52.00 | 18.00   | 25.92                               | 25.92                  | 13.73                      | (63.02)                           | (41.20)                               |
| 8    |            | 13.80 | 6.96           | 52.00 | 18.00   | 25.92                               | 25.92                  | 12.54                      | (50.48)                           | (15.28)                               |
| 9    |            | 6.96  | .              | 52.00 | 18.00   | 23.18                               | 23.18                  | 10.24                      | (40.24)                           | 7.90                                  |
| 10   |            | .     | .              | 52.00 | 18.00   | 20.40                               | 20.40                  | 8.23                       | (32.01)                           | 28.30                                 |
| 11   |            | .     | .              | 52.00 | 18.00   | 20.40                               | 20.40                  | 7.52                       | (24.49)                           | 48.70                                 |
| 12   |            | .     | .              | 52.00 | 18.00   | 20.40                               | 20.40                  | 6.87                       | (17.62)                           | 69.10                                 |
| 13   |            | .     | .              | 52.00 | 18.00   | 20.40                               | 20.40                  | 6.27                       | (11.35)                           | 89.50                                 |
| 13   |            |       |                |       |         |                                     | 42.50                  | 13.06                      | 1.71                              | 132.00                                |

10.9 The spreadsheet on the next page shows the cash flow diagrams.

- (a) Non-Discounted Cash Flow Diagram shown on next page.
- (b) (i) Cumulative cash position (CCP) = \$132 million  
Cumulative cash ratio (CCR) = 1.81

|                                   |
|-----------------------------------|
| CCP = \$132 million<br>CCR = 1.81 |
|-----------------------------------|

- (iii) Payback period (PBP) = 4.4 years

|                 |
|-----------------|
| PBP = 4.4 years |
|-----------------|

- (iii) Rate of return on investment (ROROI) = 11.00%

|                |
|----------------|
| ROROI = 11.00% |
|----------------|

- (c) Discounted Cash Flow Diagram shown on next page.
- (d) (i) Net present value (NPV) = -\$1.05 million  
Present value ratio (PVR) = 0.99

|                                     |
|-------------------------------------|
| NPV = -\$1.05 million<br>PVR = 0.99 |
|-------------------------------------|

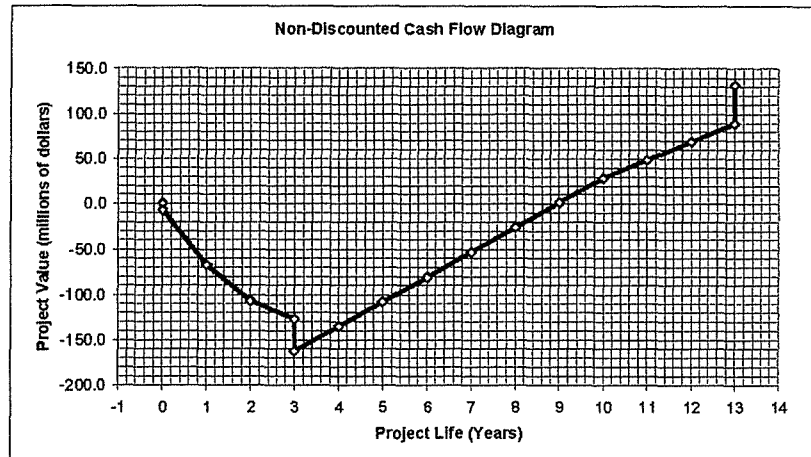
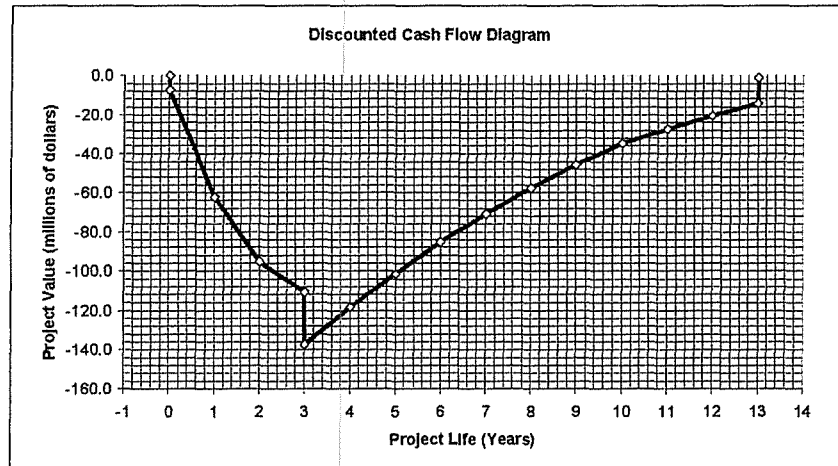
- (iv) Discounted payback period (DPBP) = 7.1 years

|                  |
|------------------|
| DPBP = 7.1 years |
|------------------|

- (v) Discounted cash flow rate of return (DCFROR) = 9.37%

|                |
|----------------|
| DCFROR = 9.37% |
|----------------|

The MACRS depreciation method improves profitability when compared to SL depreciation.



Generate CFD

**Discounted Profitability Criterion**

|                                     |        |
|-------------------------------------|--------|
| Net Present Value (millions)        | (1.05) |
| Discounted Cash Flow Rate of Return | 9.37%  |
| Discounted Payback Period (years)   | 7.1    |

**Non-Discounted Profitability Criteria**

|                                     |        |
|-------------------------------------|--------|
| Cumulative Cash Position (millions) | 132.00 |
| Rate of Return on Investment        | 11.00% |
| Payback Period (years)              | 4.4    |

| Year | Investment | $d_k$ | $FCI_k - Sd_k$ | R     | $COM_k$ | $(R - COM_k - d_k) \cdot (1 - t) + d_k$ | Cash Flow<br>(Non-discounted) | Cash Flow<br>(discounted) | Cumulative Cash Flow<br>(discounted) | Cumulative Cash Flow<br>(Non-discounted) |
|------|------------|-------|----------------|-------|---------|-----------------------------------------|-------------------------------|---------------------------|--------------------------------------|------------------------------------------|
| 0    | 0.00       |       | 120.00         |       |         |                                         | 0.00                          | 0.00                      | 0.00                                 | 0.00                                     |
| 0    | 7.50       |       | 120.00         |       |         |                                         | (7.50)                        | (7.50)                    | (7.50)                               | (7.50)                                   |
| 1    | 60.00      |       | 120.00         |       |         |                                         | (60.00)                       | (54.79)                   | (62.29)                              | (67.50)                                  |
| 2    | 39.60      |       | 120.00         |       |         |                                         | (39.60)                       | (33.03)                   | (95.32)                              | (107.10)                                 |
| 3    | 20.40      |       | 120.00         |       |         |                                         | (20.40)                       | (15.54)                   | (110.86)                             | (127.50)                                 |
| 3    | 35.00      |       | 120.00         |       |         |                                         | (35.00)                       | (26.68)                   | (137.52)                             | (162.50)                                 |
| 4    |            | 17.14 | 102.86         | 52.00 | 18.00   | 27.26                                   | 27.26                         | 18.98                     | (118.56)                             | (135.24)                                 |
| 5    |            | 17.14 | 85.71          | 52.00 | 18.00   | 27.26                                   | 27.26                         | 17.31                     | (101.24)                             | (107.99)                                 |
| 6    |            | 17.14 | 68.57          | 52.00 | 18.00   | 27.26                                   | 27.26                         | 15.81                     | (85.43)                              | (80.73)                                  |
| 7    |            | 17.14 | 51.43          | 52.00 | 18.00   | 27.26                                   | 27.26                         | 14.44                     | (70.99)                              | (53.47)                                  |
| 8    |            | 17.14 | 34.29          | 52.00 | 18.00   | 27.26                                   | 27.26                         | 13.19                     | (57.80)                              | (26.21)                                  |
| 9    |            | 17.14 | 17.14          | 52.00 | 18.00   | 27.26                                   | 27.26                         | 12.04                     | (45.76)                              | 1.04                                     |
| 10   |            | 17.14 | 0.00           | 52.00 | 18.00   | 27.26                                   | 27.26                         | 11.00                     | (34.76)                              | 28.30                                    |
| 11   |            |       | 0.00           | 52.00 | 18.00   | 20.40                                   | 20.40                         | 7.52                      | (27.24)                              | 48.70                                    |
| 12   |            |       | 0.00           | 52.00 | 18.00   | 20.40                                   | 20.40                         | 6.87                      | (20.38)                              | 69.10                                    |
| 13   |            |       | 0.00           | 52.00 | 18.00   | 20.40                                   | 20.40                         | 6.27                      | (14.11)                              | 89.50                                    |
| 13   |            |       |                |       |         |                                         | 42.50                         | 13.06                     | (1.05)                               | 132.00                                   |

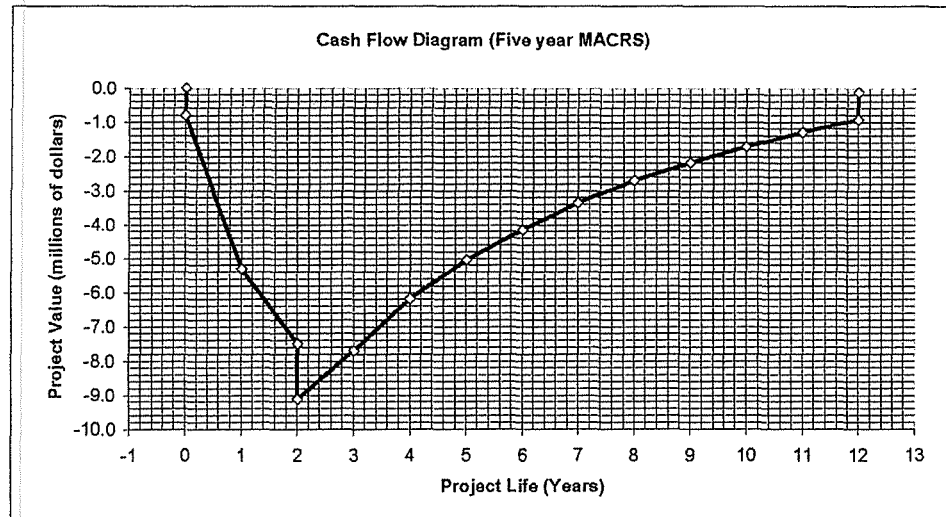
10.10 The spreadsheet and cash flow diagrams for each case are shown on the following pages.

- (a) MACRS method for 5 years

NPV = -\$0.12 million

- (b) Straight Line depreciation over 7 years

NPV = -\$0.33 million



Generate CFD

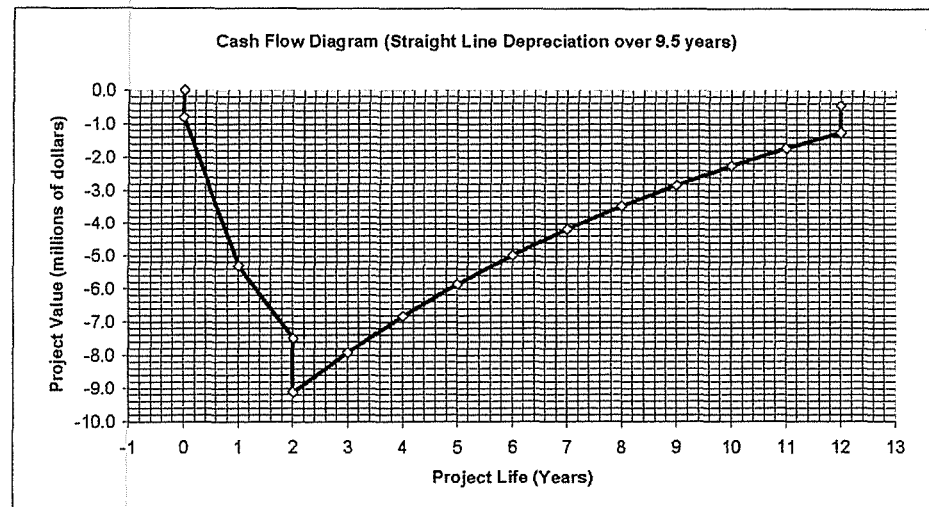
**Discounted Profitability Criterion**

|                                     |        |
|-------------------------------------|--------|
| Net Present Value (millions)        | (0.12) |
| Discounted Cash Flow Rate of Return | 10.72% |
| Discounted Payback Period (years)   | 6.5    |

**Non-Discounted Profitability Criteria**

|                                     |        |
|-------------------------------------|--------|
| Cumulative Cash Position (millions) | 8.58   |
| Rate of Return on Investment        | 11.14% |
| Payback Period (years)              | 3.9    |

| Year | Investment | $d_k$ | $FCI_t - Sd_k$ | R    | $COM_d$ | $R - COM_d - d_k$ | Cash Flow (Non-discounted) | Cash Flow (discounted) | Cumulative Cash Flow (discounted) | Cumulative Cash Flow (Non-discounted) |
|------|------------|-------|----------------|------|---------|-------------------|----------------------------|------------------------|-----------------------------------|---------------------------------------|
| 0    | 0.00       |       | 7.70           |      |         |                   | 0.00                       | 0.00                   | 0.00                              | 0.00                                  |
| 0    | 0.80       |       | 7.70           |      |         |                   | (0.80)                     | (0.80)                 | (0.80)                            | (0.80)                                |
| 1    | 5.01       |       | 7.70           |      |         |                   | (5.01)                     | (4.51)                 | (5.31)                            | (5.81)                                |
| 2    | 2.70       |       | 7.70           |      |         |                   | (2.70)                     | (2.19)                 | (7.50)                            | (8.50)                                |
| 2    | 2.00       |       | 7.70           |      |         |                   | (2.00)                     | (1.62)                 | (9.12)                            | (10.50)                               |
| 3    |            | 1.54  | 6.16           | 4.10 | 1.90    | 1.94              | 1.94                       | 1.42                   | (7.70)                            | (8.56)                                |
| 4    |            | 2.46  | 3.70           | 4.10 | 1.90    | 2.31              | 2.31                       | 1.52                   | (6.19)                            | (6.26)                                |
| 5    |            | 1.48  | 2.22           | 4.10 | 1.90    | 1.91              | 1.91                       | 1.13                   | (5.05)                            | (4.35)                                |
| 6    |            | 0.89  | 1.33           | 4.10 | 1.90    | 1.67              | 1.67                       | 0.90                   | (4.16)                            | (2.67)                                |
| 7    |            | 0.89  | 0.45           | 4.10 | 1.90    | 1.67              | 1.67                       | 0.81                   | (3.35)                            | (1.00)                                |
| 8    |            | 0.45  | -              | 4.10 | 1.90    | 1.50              | 1.50                       | 0.65                   | (2.70)                            | 0.50                                  |
| 9    |            |       | -              | 4.10 | 1.90    | 1.32              | 1.32                       | 0.52                   | (2.18)                            | 1.82                                  |
| 10   |            |       | -              | 4.10 | 1.90    | 1.32              | 1.32                       | 0.46                   | (1.72)                            | 3.14                                  |
| 11   |            |       | -              | 4.10 | 1.90    | 1.32              | 1.32                       | 0.42                   | (1.30)                            | 4.46                                  |
| 12   |            |       | -              | 4.10 | 1.90    | 1.32              | 1.32                       | 0.38                   | (0.92)                            | 5.78                                  |
| 12   |            |       |                |      |         |                   | 2.80                       | 0.80                   | (0.12)                            | 8.58                                  |



Generate CFD

**Discounted Profitability Criterion**

|                                     |        |
|-------------------------------------|--------|
| Net Present Value (millions)        | (0.46) |
| Discounted Cash Flow Rate of Return | 10.03% |
| Discounted Payback Period (years)   | 7.7    |

**Non-Discounted Profitability Criteria**

|                                     |        |
|-------------------------------------|--------|
| Cumulative Cash Position (millions) | 8.74   |
| Rate of Return on Investment        | 11.35% |
| Payback Period (years)              | 4.7    |

| Year | Investment | $d_k$ | $FCI_k - Sd_k$ | R    | $COM_d$ | $R - COM_d - d_k)(1-t) + d$ | Cash Flow<br>(Non-discounted) | Cash Flow<br>(discounted) | Cumulative Cash<br>Flow (discounted) | Cumulative Cash Flow<br>(Non-discounted) |
|------|------------|-------|----------------|------|---------|-----------------------------|-------------------------------|---------------------------|--------------------------------------|------------------------------------------|
| 0    | 0.00       |       | 7.70           |      |         |                             | 0.00                          | 0.00                      | 0.00                                 | 0.00                                     |
| 0    | 0.80       |       | 7.70           |      |         |                             | (0.80)                        | (0.80)                    | (0.80)                               | (0.80)                                   |
| 1    | 5.01       |       | 7.70           |      |         |                             | (5.01)                        | (4.51)                    | (5.31)                               | (5.81)                                   |
| 2    | 2.70       |       | 7.70           |      |         |                             | (2.70)                        | (2.19)                    | (7.50)                               | (8.50)                                   |
| 2    | 2.00       |       | 7.70           |      |         |                             | (2.00)                        | (1.62)                    | (9.12)                               | (10.50)                                  |
| 3    |            | 0.81  | 6.89           | 4.10 | 1.90    | 1.64                        | 1.64                          | 1.20                      | (7.92)                               | (8.86)                                   |
| 4    |            | 0.81  | 6.08           | 4.10 | 1.90    | 1.64                        | 1.64                          | 1.08                      | (6.83)                               | (7.21)                                   |
| 5    |            | 0.81  | 5.27           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.98                      | (5.86)                               | (5.57)                                   |
| 6    |            | 0.81  | 4.46           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.88                      | (4.98)                               | (3.92)                                   |
| 7    |            | 0.81  | 3.65           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.79                      | (4.19)                               | (2.28)                                   |
| 8    |            | 0.81  | 2.84           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.71                      | (3.47)                               | (0.63)                                   |
| 9    |            | 0.81  | 2.03           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.64                      | (2.83)                               | 1.01                                     |
| 10   |            | 0.81  | 1.22           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.58                      | (2.25)                               | 2.65                                     |
| 11   |            | 0.81  | 0.41           | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.52                      | (1.73)                               | 4.30                                     |
| 12   |            | 0.81  | -              | 4.10 | 1.90    | 1.64                        | 1.64                          | 0.47                      | (1.26)                               | 5.94                                     |
| 12   |            |       |                |      |         |                             | 2.80                          | 0.80                      | (0.46)                               | 8.74                                     |

10.11 (a) DCFROR =  $i$ , where NPV = 0

| Year | Nondiscounted<br>Cash Flow (\$10 <sup>6</sup> /yr) | Discounted Cash<br>Flow (\$10 <sup>6</sup> /yr) | Cumulative Discounted<br>Cash Flow (\$10 <sup>6</sup> /yr) |
|------|----------------------------------------------------|-------------------------------------------------|------------------------------------------------------------|
| 1    | -10.000                                            | -9.097                                          | -9.097                                                     |
| 2    | -15.000                                            | -12.412                                         | -21.509                                                    |
| 3    | -15.000                                            | -11.291                                         | -32.800                                                    |
| 4    | 7.015                                              | 4.803                                           | -27.997                                                    |
| 5    | 6.206                                              | 3.866                                           | -24.131                                                    |
| 6    | 6.295                                              | 3.567                                           | -20.564                                                    |
| 7    | 6.852                                              | 3.532                                           | -17.032                                                    |
| 8    | 6.859                                              | 3.216                                           | -13.817                                                    |
| 9    | 7.218                                              | 3.079                                           | -10.738                                                    |
| 10   | 5.954                                              | 2.310                                           | -8.428                                                     |
| 11   | 5.459                                              | 1.927                                           | -6.501                                                     |
| 12   | 5.789                                              | 1.859                                           | -4.643                                                     |
| 13   | 15.898                                             | 4.643                                           | 0.000                                                      |

$$\text{DCFRO}R = i = 0.099 = 9.9\%$$

(b) The DCFRO is very close to 10%; therefore, it is a borderline decision.

10.12 The spreadsheet and cash flow diagrams are shown on the following pages.

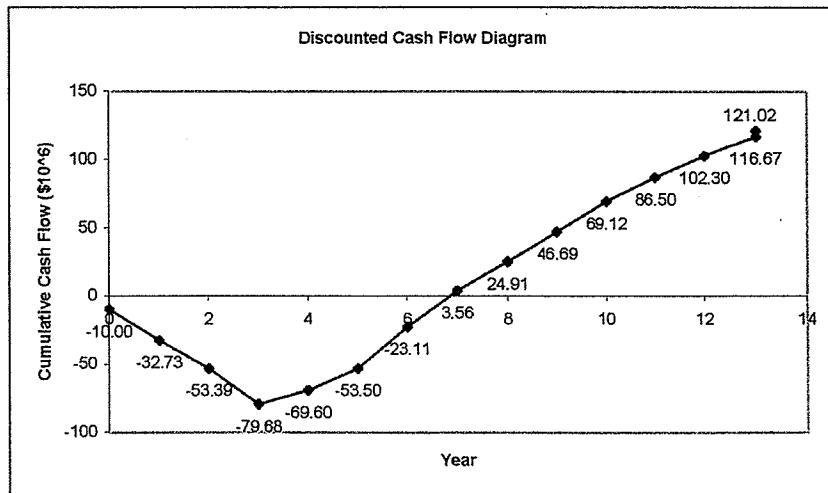
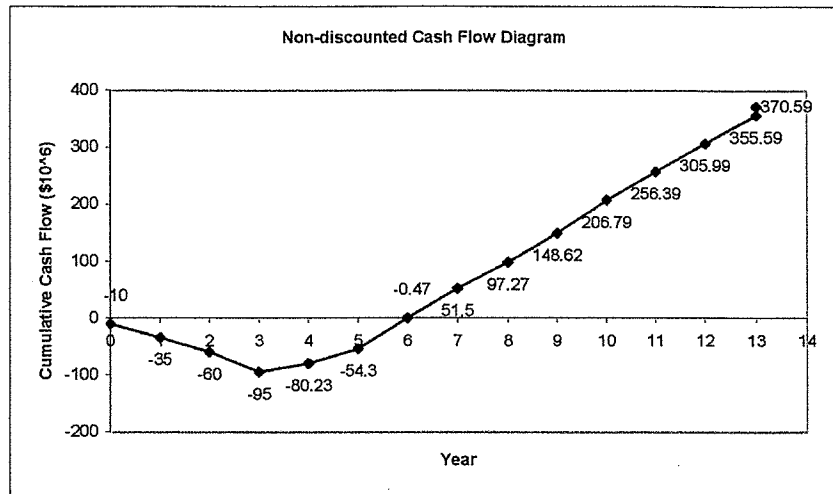
- (a) Payback period (PBP)  $\approx$  2.54 years
- (b) Cumulative cash position (CCP) = \$360.60 million
- (c) Rate of return on investment (RORI) = 78%
- (d) Discounted payback period (DPBP) = 2.47 years
- (e) Net present value (NPV) = \$130.81 million
- (f) Discounted cash flow rate of return (DCFRO) = 32.8%

FCI 80  
i 0.1

| Year | Investment | Depreciation | Revenue  | Total Annual Costs | Net Profit | Income Tax | Net Profit After Tax | After-Tax Cash Flow | Cumulative Cash Flow | Discounted Cash Flow | Disc. Cum. Cash Flow |
|------|------------|--------------|----------|--------------------|------------|------------|----------------------|---------------------|----------------------|----------------------|----------------------|
| 0    | (\$10)     |              |          |                    |            |            |                      | (\$10.00)           | (\$10.00)            | (\$10.00)            | (\$10.00)            |
| 1    | (\$25)     |              |          |                    |            |            |                      | (\$25.00)           | (\$35.00)            | (\$22.73)            | (\$32.73)            |
| 2    | (\$25)     |              |          |                    |            |            |                      | (\$25.00)           | (\$60.00)            | (\$20.66)            | (\$53.39)            |
| 3    | (\$35)     |              |          |                    |            |            |                      | (\$35.00)           | (\$95.00)            | (\$26.30)            | (\$79.68)            |
| 4    |            | \$8.57       | \$60.00  | \$50.00            | \$10.00    | \$3.80     | \$6.20               | \$14.77             | (\$80.23)            | \$10.09              | (\$69.60)            |
| 5    |            | \$8.57       | \$120.00 | \$92.00            | \$28.00    | \$10.64    | \$17.36              | \$25.93             | (\$54.30)            | \$16.10              | (\$53.50)            |
| 6    |            | \$8.57       | \$120.00 | \$47.00            | \$73.00    | \$27.74    | \$45.26              | \$53.83             | (\$0.47)             | \$30.39              | (\$23.11)            |
| 7    |            | \$8.57       | \$120.00 | \$50.00            | \$70.00    | \$26.60    | \$43.40              | \$51.97             | \$51.50              | \$26.67              | \$3.56               |
| 8    |            | \$8.57       | \$120.00 | \$60.00            | \$60.00    | \$22.80    | \$37.20              | \$45.77             | \$97.27              | \$21.35              | \$24.91              |
| 9    |            | \$8.57       | \$120.00 | \$51.00            | \$69.00    | \$26.22    | \$42.78              | \$51.35             | \$148.62             | \$21.78              | \$46.69              |
| 10   |            | \$8.57       | \$120.00 | \$40.00            | \$80.00    | \$30.40    | \$49.60              | \$58.17             | \$206.79             | \$22.43              | \$69.12              |
| 11   |            |              | \$120.00 | \$40.00            | \$80.00    | \$30.40    | \$49.60              | \$49.60             | \$256.39             | \$17.38              | \$86.50              |
| 12   |            |              | \$120.00 | \$40.00            | \$80.00    | \$30.40    | \$49.60              | \$49.60             | \$305.99             | \$15.80              | \$102.30             |
| 13   |            |              | \$120.00 | \$40.00            | \$80.00    | \$30.40    | \$49.60              | \$49.60             | \$355.59             | \$14.37              | \$116.67             |
| 13   | \$15       |              |          |                    |            |            |                      | \$15.00             | \$370.59             | \$4.34               | \$131.02             |

Note: Amounts shown in table are in millions of dollars.





- 10.13 (a) From year 4  
Taxation rate,  $t = \text{income tax} / \text{net profit} = 3.8/10 = 38\%$

$$T = 38\%$$

- (b) Total Fixed Capital Cost ( $FCI_L$ ) =  $80 - 20 = \$60$  million

$$FCI_L = \$60 \text{ million}$$

- (c) Depreciation method corresponds to Straight Line depreciation over 7 years.  
 $d_{SL} = 60/7 = \$8.53$  million

Straight Line depreciation

- (d)  $COM = \text{Total Annual Costs} - \text{Depreciation}$

| Year | Depreciation<br>(\$10 <sup>6</sup> /yr) | Total Annual Costs<br>(\$10 <sup>6</sup> /yr) | COM<br>(\$10 <sup>6</sup> /yr) |
|------|-----------------------------------------|-----------------------------------------------|--------------------------------|
| 0    |                                         |                                               |                                |
| 1    |                                         |                                               |                                |
| 2    |                                         |                                               |                                |
| 3    |                                         |                                               |                                |
| 4    | 8.57                                    | 50.00                                         | 41.43                          |
| 5    | 8.57                                    | 92.00                                         | 83.43                          |
| 6    | 8.57                                    | 47.00                                         | 38.43                          |
| 7    | 8.57                                    | 50.00                                         | 41.43                          |
| 8    | 8.57                                    | 60.00                                         | 51.43                          |
| 9    | 8.57                                    | 51.00                                         | 42.43                          |
| 10   | 8.57                                    | 40.00                                         | 31.43                          |
| 11   |                                         | 40.00                                         | 40.00                          |
| 12   |                                         | 40.00                                         | 40.00                          |
| 13   |                                         | 40.00                                         | 40.00                          |

10.14 (a)  $NPV = -FCI + \sum CF_i(P/F, i, n)$

Process 1

$i = 6\%$

$$NPV = -\$15 + \frac{\$3}{(1.06)^1} + \frac{\$8}{(1.06)^2} + \frac{\$7}{(1.06)^3} + \frac{\$5}{(1.06)^4} + \frac{\$2}{(1.06)^5} = \$6.28 \text{ million}$$

Process 1

$i = 18\%$

$$NPV = -\$15 + \frac{\$3}{(1.18)^1} + \frac{\$8}{(1.18)^2} + \frac{\$7}{(1.18)^3} + \frac{\$5}{(1.18)^4} + \frac{\$2}{(1.18)^5} = \$1.00 \text{ million}$$

Process 2

$i = 6\%$

$$NPV = -\$15 + \frac{\$5}{(1.06)^1} + \frac{\$5}{(1.06)^2} + \frac{\$5}{(1.06)^3} + \frac{\$5}{(1.06)^4} + \frac{\$5}{(1.06)^5} = \$6.06 \text{ million}$$

Process 2

$i = 18\%$

$$NPV = -\$15 + \frac{\$5}{(1.18)^1} + \frac{\$5}{(1.18)^2} + \frac{\$5}{(1.18)^3} + \frac{\$5}{(1.18)^4} + \frac{\$5}{(1.18)^5} = \$0.64 \text{ million}$$

**Recommend Process 1**

(b) set  $NPV = 0$  in the above equations and calculate  $i$

Process 1:  $DCFROR = 21\%$   
 Process 2:  $DCFROR = 19.9\%$   
 Process 1 is better.

(c) Payback Period

Cumulative cash flows.

| Year | Process 1    | Process 2    |
|------|--------------|--------------|
| 1    | 3.00         | 5.00         |
| 2    | <b>11.00</b> | 10.00        |
| 3    | <b>18.00</b> | <b>15.00</b> |
| 4    | 23.00        | 20.00        |
| 5    | 25.00        | 25.00        |

Process 1

$$PBP = 2 + (15 - 11) / (18 - 11) = 2.57 \text{ years}$$

Process 2

$$PBP = 3 \text{ years}$$

$$PBP_1 = 2.57 \text{ years}$$

$$PBP_2 = 3 \text{ years}$$

Process 1 is better.

(d) Parts a, b, and c all concluded with Process 1 being the better choice.

10.15 Compare using  $EAOC$ .

$$EAOC = FCI(A/P, i, n) + YOC$$

$$EAOC_A = 15 \left( \frac{(0.09)(1+0.09)^3}{(1+0.09)^3 - 1} \right) + 4 = \$9.93 \text{ thousand/yr}$$

$$EAOC_B = 25 \left( \frac{(0.09)(1+0.09)^5}{(1+0.09)^5 - 1} \right) + 3 = \$9.43 \text{ thousand/yr}$$

$$EAOC_C = 40 \left( \frac{(0.09)(1+0.09)^7}{(1+0.09)^7 - 1} \right) + 2 = \$9.95 \text{ thousand/yr}$$

Choose Equipment B

10.16 (a)  $EAOC = FCI(A/P, i, n) + YOC$

$$EAOC_A = 5 \left( \frac{(0.15)(1+0.15)^5}{(1+0.15)^5 - 1} \right) + 2 = \$3.49 \text{ thousand/yr}$$

$$EAOC_B = 10 \left( \frac{(0.15)(1+0.15)^7}{(1+0.15)^7 - 1} \right) + 1 = \$3.40 \text{ thousand/yr}$$

Choose Equipment B

(b) When  $EAOC_A - EAOC_B = 0$ ;  $i = 0.1735$

When  $i = 18\%$ ;  $EAOC_A = -\$3.60 \text{ thousand/yr}$

$EAOC_B = -\$3.62 \text{ thousand/yr}$

$i = 18\%$

10.17 Compare using *EAOC*

$$EAOC = FCI(A/P, i, n) + YOC$$

The table below only includes equipment that differs in cost between the two options.

| Option | FCI<br>(\$thousand) | YOC<br>(\$thousand/yr) | Years | A/P  | EAOC<br>(\$thousand/yr) |
|--------|---------------------|------------------------|-------|------|-------------------------|
| 1      | 70                  | 15                     | 10    | 0.15 | 25.38                   |
| 2      | 100                 | 9                      | 15    | 0.13 | 22.09                   |

Choose Option 2

10.18  $CC = P \frac{(1+i)^n}{(1+i)^n - 1}$

Purchased cost of pump = \$35,000

Installation cost =  $(0.75)(\$35,000) = \$26,250$

$P = \$35,000 + \$26,250 = \$61,250$

$i = 10\%$ ,  $n = 3$  years

$$CC = 61,250 \frac{(1+0.1)^3}{(1+0.1)^3 - 1} = \$246,295$$

Capitalized Cost = \$246,295

10.19 Compare using *EAOC*

$$EAOC = FCI(A/P, i, n) + YOC$$

$$EAOC_{RVF} = 15 \left( \frac{(0.11)(1+0.11)^6}{(1+0.11)^6 - 1} \right) + 3 = \$6.55 \text{ thousand/yr}$$

$$EAOC_{FP} = 10 \left( \frac{(0.11)(1+0.11)^8}{(1+0.11)^8 - 1} \right) + 5 = \$6.94 \text{ thousand/yr}$$

$$EAOC_{HC} = 25 \left( \frac{(0.11)(1+0.11)^{10}}{(1+0.11)^{10} - 1} \right) + 2 = \$6.25 \text{ thousand/yr}$$

Choose Hydrocyclone & Centrifuge

10.20 Incremental analysis:  $INPV = -FCI + (P/A, i, n)YS$

$$INPV = -250 + 70 \frac{(1+0.07)^7 - 1}{0.07(1+0.07)^7} = \$127.25 \text{ thousand}$$

$INPV = \$127.25$  thousand  
Purchase and install the baghouse filter.

10.21  $INPV = -FCI + (P/A, i, n)YS$

| Project | FCI<br>(\$ million) | Cash Flow<br>(\$ million) | P/A  | INPV<br>(\$ million) |
|---------|---------------------|---------------------------|------|----------------------|
| A       | 80                  | 11                        | 7.54 | 2.91                 |
| B       | 100                 | 14                        | 7.54 | 5.53                 |
| C       | 120                 | 16                        | 7.54 | 0.60                 |
| A&B     | 180                 | 25                        | 7.54 | 8.44                 |
| A&C     | 200                 | 27                        | 7.54 | 3.52                 |
| B&C     | 220                 | 30                        | 7.54 | 6.13                 |
| ALL     | 300                 | 41                        | 7.54 | 9.04                 |

Projects A and B provide the greatest NPV and meets the capital ceiling for investment of \$250 million.

Choose Projects A and B

10.22  $INPV = -FCI + (P/A, i, n)YS$

| Project    | FCI<br>(\$ thousands) | Yearly Savings<br>(\$ thousands) | Years | P/A  | INPV<br>(\$ thousands) |
|------------|-----------------------|----------------------------------|-------|------|------------------------|
| Solar      | 25                    | 2                                | 15    | 9.40 | -6.19                  |
| Insulation | 5                     | 0.9                              | 15    | 9.40 | 3.46                   |
| Both       | 30                    | 2.5                              | 15    | 9.40 | -6.49                  |

Choose insulation only.

Tax credit needed for solar collector:

When  $INPV_{Solar} = 0$ ,  $FCI_{Solar} = \$18.8$  thousand.

Tax credit =  $1 - (18.8/25) = 0.248$

Tax credit = 25%

10.23 (a)  $ROROI = \text{Incremental Yearly Savings} / \text{Incremental Investment}$

| Case | FCI<br>(\$ million) | Cash Flow<br>(\$ million) | ROROI |
|------|---------------------|---------------------------|-------|
| Base | 75                  | 19                        | 0.253 |
| 1    | 15                  | 3                         | 0.244 |
| 2    | 25                  | 5                         | 0.240 |
| 3    | 30                  | 7                         | 0.248 |

Base case is the best option.

(b)  $INPV = -FCI + (P/A, i, n)YS$

| Case | FCI<br>(\$ million) | Cash Flow<br>(\$ million) | P/A  | INPV<br>(\$ million) |
|------|---------------------|---------------------------|------|----------------------|
| base | 75                  | 19                        | 5.02 | 20.36                |
| 1    | 15                  | 3                         | 5.02 | 20.41                |
| 2    | 25                  | 5                         | 5.02 | 20.45                |
| 3    | 30                  | 7                         | 5.02 | 25.49                |

Case 3 is the best option.

10.24 (a)  $NPV = -FCI + \sum CF_i (P/F, i, n)$

$$NPV = -500 + \frac{100}{1.08} + \frac{90}{1.08^2} + \frac{90}{1.08^3} + \frac{85}{1.08^4} + \frac{80}{1.08^5} + \frac{80}{1.08^6} + \frac{75}{1.08^7}$$

$NPV = -\$47.7$  thousand; Do not invest.

(b) When  $NPV = 0$ ,  $i = 5.01\%$

break-even  $i = 5.01\%$

(c) When  $NPV = 0$ , for profitable investment  $FCI = \$452.3$  thousand

$FCI = \$452.3$  thousand



10.25 Look at incremental cases.

$ROROI = \text{Incremental Yearly Savings} / \text{Incremental Investment}$

$$ROROI_{\text{Base} + 1 - \text{Base}} = (0.77 - 0.75) / (5.1 - 5) = 0.20$$

See table for complete  $ROROI$  calculations.

| Case                       | FCI<br>(\$ million) | Cash Flow<br>(\$ million) | ROROI |
|----------------------------|---------------------|---------------------------|-------|
| Base                       | 5                   | 0.75                      | -     |
| Base + option 1            | 5.1                 | 0.77                      | 0.200 |
| Base + option 1 + option 2 | 5.3                 | 0.79                      | 0.100 |
| Base + option 3            | 5.2                 | 0.782                     | 0.120 |
| Base + option 1 + option 3 | 5.4                 | 0.81                      | 0.133 |

Choose Base case + Option 1

10.26  $EAOC = -FCI(A/P, i, n) + YS$

$$EAOC = (-49.6) \left( \frac{0.12(1 + 0.12)^{12}}{(1 + 0.12)^{12} - 1} \right) + 8 = -\$0.007 \text{ thousand/yr}$$

$EAOC = -\$0.01 \text{ thousand/yr; Borderline investment}$

10.27 (a) Base Case

$$NPV = -FCI + CF_i(P / A, i, n)$$

$$NPV = -21 + (12 - 5.1 - 1.35) \left( \frac{(1 + 0.18)^8 - 1}{(0.18)(1 + 0.18)^8} \right) = \$1.63 \text{ million}$$

$NPV = \$1.63 \text{ million; Recommend construction}$

(b)  $INPV = -FCI + YS(P / A, i, n)$

Alternative 1

$$INPV = -2.15 + (0.08) \left( \frac{(1 + 0.18)^8 - 1}{(0.18)(1 + 0.18)^8} \right) = -\$1.82 \text{ million}$$

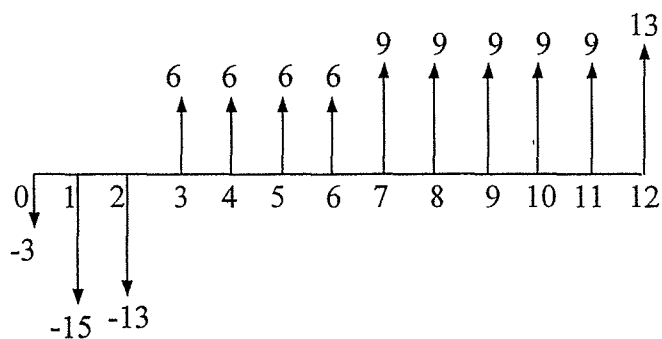
$$INPV = -1.35 + (0.04) \left( \frac{(1 + 0.18)^8 - 1}{(0.18)(1 + 0.18)^8} \right) = -\$1.19 \text{ million}$$

$\text{Do not recommend alternatives}$

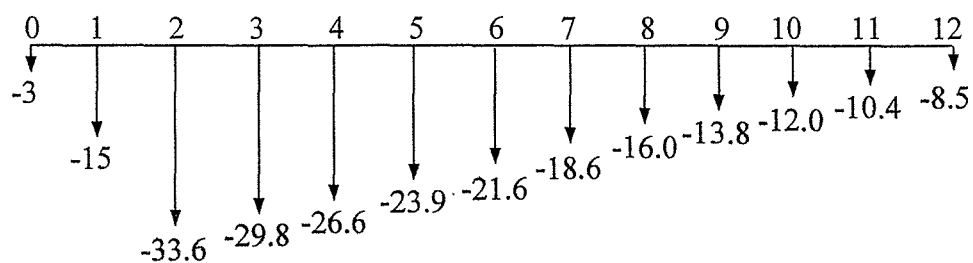
(c)  $INPV = 0 = -3.26 + 4.08YS$

$$YS = 0.80$$

10.28 (a) All values are in \$ millions.



(b) All values are in \$ millions.



(c)  $NPV = -\$8.54$  million  
Do not recommend construction of plant.

$$10.29 \quad INPV = -FCI + YS(P / A, i, n)$$

$$INPV = -1.75 + (YS) \left( \frac{(1+0.18)^8 - 1}{(0.18)(1+0.18)^8} \right) = 0$$

$$YS = \$0.429 \text{ million/yr}$$

$$10.30 \quad INPV = -FCI + YS(P / A, i, n)$$

$$INPV = -FCI + (2.25) \left( \frac{(1+0.15)^6 - 1}{(0.15)(1+0.15)^6} \right) = 0$$

$$FCI = \$8.515 \text{ million}$$

$$10.31 \quad INPV = -FCI + YS(P / A, i, n)$$

$$INPV = -5.1 + (0.9) \left( \frac{(1+i)^9 - 1}{(i)(1+i)^9} \right) = 0$$

$$i < 10.4\%$$

10.32

| Year | Cash Flow<br>(\$ million) | Discounted<br>Cash Flow<br>(\$ million) | Cumulative<br>Cash Flow<br>(\$ million) |
|------|---------------------------|-----------------------------------------|-----------------------------------------|
| 0    | -5.00                     | -5.00                                   | -5.00                                   |
| 1    | -4.00                     | -3.42                                   | -8.42                                   |
| 2    | 2.00                      | 1.46                                    | -6.96                                   |
| 3    | 2.00                      | 1.25                                    | -5.71                                   |
| 4    | 2.00                      | 1.07                                    | -4.64                                   |
| 5    | 2.00                      | 0.91                                    | -3.73                                   |
| 6    | 2.00                      | 0.78                                    | -2.95                                   |
| 7    | 2.00                      | 0.67                                    | -2.28                                   |
| 8    | 2.00                      | 0.57                                    | -1.71                                   |
| 9    | 2.00                      | 0.49                                    | -1.23                                   |
| 10   | 2.00                      | 0.42                                    | -0.81                                   |
| 11   | 2.00                      | 0.36                                    | -0.46                                   |
| 12   | 2.00                      | 0.30                                    | -0.15                                   |
| 13   | 3.00                      | 0.39                                    | 0.24                                    |

$NPV = \$0.24$  million  
Recommend investment

10.33 (a)  $ROROI = \text{Incremental Yearly Savings} / \text{Incremental Investment}$

| Alternative | FCI<br>(\$ million) | Yearly Savings<br>(\$ million) | P/A  | ROROI | INPV<br>(\$ million) |
|-------------|---------------------|--------------------------------|------|-------|----------------------|
| 1           | 2.25                | 0.65                           | 3.50 | 0.29  | 0.02                 |
| 2           | 3.45                | 0.75                           | 3.50 | 0.22  | -0.83                |
| 1&2         | 5.70                | 1.40                           | 3.50 | 0.25  | -0.80                |

All  $ROROI$  values are greater than 18%; therefore, all alternatives are good.

All alternatives will work.

(b)  $INPV = -FCI + YS(P / A, i, n)$

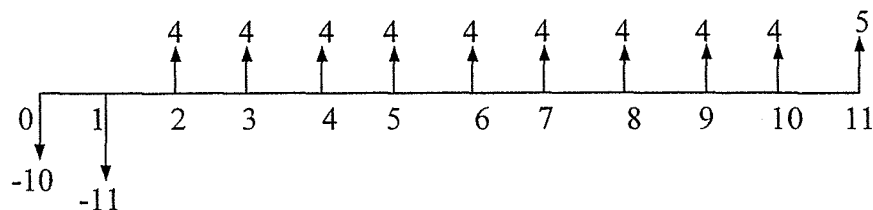
See above table for  $INPV$  values.

Recommend only Alternative 1.

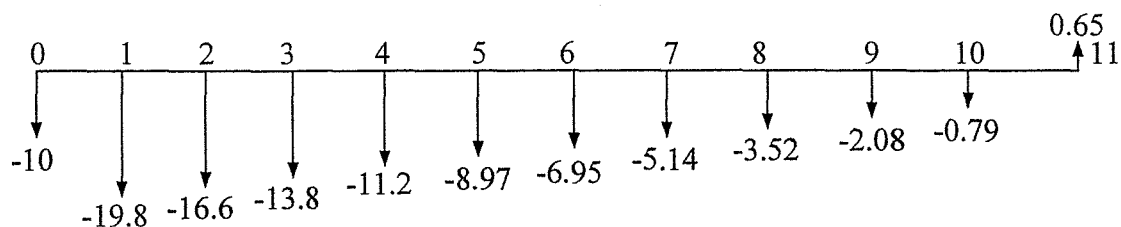
(c) Recommend only Alternative 1 because the  $INPV$  value is positive.

Recommend only Alternative 1.

10.34 (a) All values are in \$ millions.



(b) All values are in \$ millions.



(c)

| Year | Cash Flow<br>(\$ million) | Discounted<br>Cash Flow<br>(\$ million) | Cumulative Disc.<br>Cash Flow<br>(\$ million) |
|------|---------------------------|-----------------------------------------|-----------------------------------------------|
| 0    | -10                       | -10.00                                  | -10.00                                        |
| 1    | -11                       | -9.82                                   | -19.82                                        |
| 2    | 4                         | 3.19                                    | -16.63                                        |
| 3    | 4                         | 2.85                                    | -13.79                                        |
| 4    | 4                         | 2.54                                    | -11.24                                        |
| 5    | 4                         | 2.27                                    | -8.97                                         |
| 6    | 4                         | 2.03                                    | -6.95                                         |
| 7    | 4                         | 1.81                                    | -5.14                                         |
| 8    | 4                         | 1.62                                    | -3.52                                         |
| 9    | 4                         | 1.44                                    | -2.08                                         |
| 10   | 4                         | 1.29                                    | -0.79                                         |
| 11   | 5                         | 1.44                                    | 0.65                                          |

$NPV = \$0.65$  million

(d) For the  $NPV = \$2$  million, profit = \$4.27 million

(e)  $INPV = -FCI + YS(P / A, i, n)$

$$INPV = -3 + YS \left( \frac{(1 + 0.16)^5 - 1}{0.16(1 + 0.16)^5} \right) = 0$$

$YS = 0.916$  million/yr

$$10.35 \quad EAO C = FCI(A/P, i, n) + YOC$$

$$EAO C_A = 5 \left( \frac{0.14(1+0.14)^3}{(1+0.14)^3 - 1} \right) + 5.5 = \$7.65 \text{ thousand/yr}$$

$$EAO C_B = 15 \left( \frac{0.14(1+0.14)^5}{(1+0.14)^5 - 1} \right) + 3.5 = \$7.87 \text{ thousand/yr}$$

$$EAO C_C = 5 \left( \frac{0.14(1+0.14)^9}{(1+0.14)^9 - 1} \right) + 2.5 = \$7.15 \text{ thousand/yr}$$

Choose Alternative C

$$10.36 \quad EAO C = FCI(A/P, i, n) - YS$$

$$EAO C_A = 88 \left( \frac{0.15(1+0.15)^8}{(1+0.15)^8 - 1} \right) - (15 + 25 + 33) = -\$53.39 \text{ thousand/yr}$$

$$EAO C_B = 125 \left( \frac{0.15(1+0.15)^8}{(1+0.15)^8 - 1} \right) - (15 + 45 + 25) = -\$57.14 \text{ thousand/yr}$$

$$EAO C_C = 250 \left( \frac{0.15(1+0.15)^8}{(1+0.15)^8 - 1} \right) - (15 + 75 + 18) = -\$52.29 \text{ thousand/yr}$$

Choose Alternative B



10.37 (a)  $ROROI = \text{Incremental Yearly Savings} / \text{Incremental Investment}$

$$ROROI_A = \frac{(\$32.5 - \$15)}{\$103.7} = \frac{\$17.5}{\$103.7} = 0.17$$

$$ROROI_B = \frac{(\$65 - \$30)}{\$162.4} = \frac{\$35}{\$162.4} = 0.22$$

$$ROROI_C = \frac{(\$74.75 - \$34.5)}{\$216.1} = \frac{\$40.25}{\$216.1} = 0.19$$

From above, choose Case B. However, can also do incremental comparison between cases.

Use Case A as base case because it has the lowest  $FCI$  and evaluate  $ROROI$  going from Case A to Case B.

$$ROROI_{A-B} = \frac{(\$35 - \$17.5)}{(\$162.4 - \$103.7)} = 0.298$$

$ROROI$  is greater than 15%, now use Case B as base case and compare to Case C.

$$ROROI_{B-C} = \frac{(\$40.25 - \$35)}{(\$216.1 - \$162.4)} = 0.098$$

Reject Case C because  $ROROI$  is less than 15%.

Choose Case B

(b)  $EAOC = FCI(A/P, i, n) - YS$

$$EAOC_A = \$103.7 \left( \frac{(0.05)(1+0.05)^7}{(1+0.05)^7 - 1} \right) - \$17.5 = \$0.42 \text{ thousand/yr}$$

$$EAOC_B = \$162.4 \left( \frac{(0.05)(1+0.05)^7}{(1+0.05)^7 - 1} \right) - \$35 = -\$6.93 \text{ thousand/yr}$$

$$EAOC_C = \$216.1 \left( \frac{(0.05)(1+0.05)^7}{(1+0.05)^7 - 1} \right) - \$40.25 = -\$2.90 \text{ thousand/yr}$$

Choose Case B

(c)  $EAOC = FCI(A/P, i, n) - YS$

$$EAOC_A = \$103.7 \left( \frac{(0.05)(1+0.05)^{15}}{(1+0.05)^{15} - 1} \right) - \$17.5 = -\$7.51 \text{ thousand/yr}$$

$$EAOC_B = \$162.4 \left( \frac{(0.05)(1+0.05)^{15}}{(1+0.05)^{15} - 1} \right) - \$35 = -\$19.35 \text{ thousand/yr}$$

$$EAOC_C = \$216.1 \left( \frac{(0.05)(1+0.05)^{15}}{(1+0.05)^{15} - 1} \right) - \$40.25 = -\$19.43 \text{ thousand/yr}$$

Choose Case C

(d)  $EAOC = FCI(A/P, i, n) - YS$

$$EAOC_A = \$103.7 \left( \frac{(0.05)(1+0.05)^5}{(1+0.05)^5 - 1} \right) - \$17.5 = \$6.45 \text{ thousand/yr}$$

$$EAOC_B = \$162.4 \left( \frac{(0.05)(1+0.05)^5}{(1+0.05)^5 - 1} \right) - \$35 = \$2.51 \text{ thousand/yr}$$

$$EAOC_C = \$216.1 \left( \frac{(0.05)(1+0.05)^5}{(1+0.05)^5 - 1} \right) - \$40.25 = \$9.66 \text{ thousand/yr}$$

None of the cases provide an acceptable  $EAOC$ .

### 10.38 Individual solutions may differ slightly because of the Monte Carlo method being used.

Probable Variation of Key Parameters over Plant Life

|                    | Lower Limit | Upper Limit | Base Value     |
|--------------------|-------------|-------------|----------------|
| FCIL               | 0%          | 0%          | \$ 120,000,000 |
| Price of Product   | -13%        | 3%          | \$ 52,000,000  |
| Working Capital    | 0%          | 0%          | \$ 35,000,000  |
| Income Tax Rate*   | 0%          | 0%          | 40%            |
| Interest Rate*     | 0%          | 0%          | 9%             |
| Raw Material Price | -11%        | 19%         | \$ 18,000,000  |
| Salvage Value      | 0%          | 0%          | 1              |

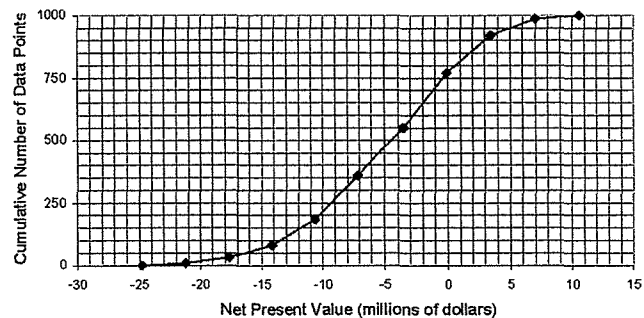
\* Please note that variations for percentages are a percent of a percent.  
For example, a 10% variance on a 12% interest rate would imply a 1.2% uncertainty

Run Economic Analysis

### Net Present Value Data

Low NPV -24.8  
High NPV 10.5

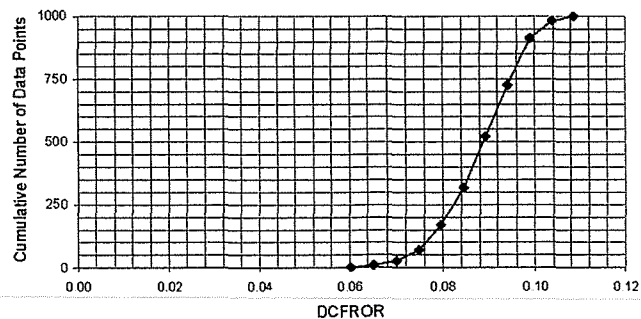
| Bins | Upper Value | # points/bin | Cumulative |
|------|-------------|--------------|------------|
| 0    | -24.8       | 0            | 0          |
| 1    | -21.2       | 8            | 8          |
| 2    | -17.7       | 23           | 31         |
| 3    | -14.2       | 49           | 80         |
| 4    | -10.6       | 104          | 184        |
| 5    | -7.1        | 174          | 358        |
| 6    | -3.6        | 193          | 551        |
| 7    | -0.1        | 218          | 769        |
| 8    | 3.5         | 150          | 919        |
| 9    | 7.0         | 67           | 986        |
| 10   | 10.5        | 14           | 1000       |



### Discounted Cash Flow Rate of Return Data

Low DCFROR 0.06  
High DCFROR 0.11

| Bins | Upper | #/bin | Cumulative |
|------|-------|-------|------------|
| 0    | 0.06  | 0     | 0          |
| 1    | 0.07  | 8     | 8          |
| 2    | 0.07  | 17    | 25         |
| 3    | 0.07  | 44    | 69         |
| 4    | 0.08  | 99    | 168        |
| 5    | 0.08  | 149   | 317        |
| 6    | 0.09  | 205   | 522        |
| 7    | 0.09  | 203   | 725        |
| 8    | 0.10  | 187   | 912        |
| 9    | 0.10  | 69    | 981        |
| 10   | 0.11  | 19    | 1000       |



With at least a 75% chance that the NPV < 0, this might not be a good decision.

### 10.39

Lowest NPV = -\$24.8 million  $\Rightarrow$  approximately 1%  
Highest NPV = \$10.5 million  $\Rightarrow$  approximately 10%

10.40

### Monte Carlo Analysis for (using MACRS 5 yrs)

#### Probable Variation of Key Parameters over Plant Life

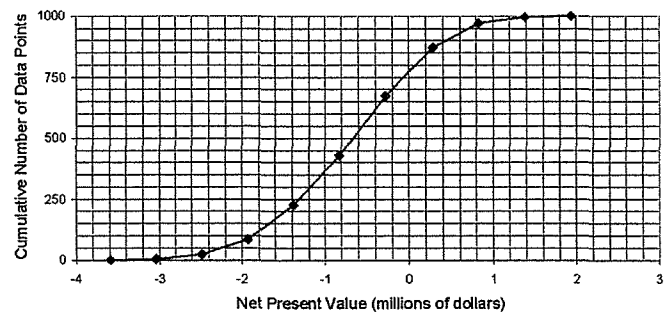
|                    | Lower Limit | Upper Limit | Base Value   |
|--------------------|-------------|-------------|--------------|
| FCIL               | -14%        | 21%         | \$ 7,700,000 |
| Price of Product   | -15%        | 10%         | \$ 4,100,000 |
| Working Capital    | 0%          | 0%          | \$ 2,000,000 |
| Income Tax Rate*   | 0%          | 0%          | 40%          |
| Interest Rate*     | -14%        | 9%          | 11%          |
| Raw Material Price | -11%        | 32%         | \$ 1,900,000 |
| Salvage Value      | 0%          | 0%          | 1            |

\* Please note that variations for percentages are a percent of a percent.  
For example, a 10% variance on a 12% interest rate would imply a 1.2% uncertainty

Run Economic Analysis

### Net Present Value Data

|             |                    |                     |                   |
|-------------|--------------------|---------------------|-------------------|
| Low NPV     | -3.6               |                     |                   |
| High NPV    | 1.9                |                     |                   |
| <u>Bins</u> | <u>Upper Value</u> | <u># points/bin</u> | <u>Cumulative</u> |
| 0           | -3.6               | 0                   | 0                 |
| 1           | -3.0               | 4                   | 4                 |
| 2           | -2.5               | 19                  | 23                |
| 3           | -1.9               | 62                  | 85                |
| 4           | -1.4               | 141                 | 226               |
| 5           | -0.8               | 202                 | 428               |
| 6           | -0.3               | 246                 | 674               |
| 7           | 0.3                | 197                 | 871               |
| 8           | 0.8                | 99                  | 970               |
| 9           | 1.4                | 26                  | 996               |
| 10          | 1.9                | 4                   | 1000              |



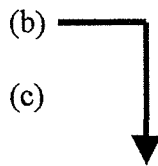
From the above figure, the chance of getting an NPV < 0 is 75-80%. Probably not a wise investment opportunity.

10.41 (a)  $\frac{A}{P} = \frac{i(1+i)^n}{(1+i)^n - 1}$

$$A_{High} = \$250,000 \frac{(0.085)(1+0.085)^{15}}{(1+0.085)^{15} - 1} = \$30,105.12$$

$$A_{Low} = \$250,000 \frac{(0.06)(1+0.06)^{15}}{(1+0.06)^{15} - 1} = \$25,740.69$$

|                                                     |
|-----------------------------------------------------|
| $A_{High} = \$30,105.12$<br>$A_{Low} = \$25,740.69$ |
|-----------------------------------------------------|



### Variable Rate Simulation

| Year | <i>i</i> | Payment<br>(\$/y) | Interest<br>(\$/y) | Principal Paid<br>(\$/y) | Remaining Balance<br>(\$/y) |
|------|----------|-------------------|--------------------|--------------------------|-----------------------------|
| 1    | 0.0765   | 24,795            | 19,114             | 5,681                    | 244,319                     |
| 2    | 0.0664   | 23,010            | 16,230             | 6,779                    | 237,539                     |
| 3    | 0.0727   | 24,078            | 17,271             | 6,807                    | 230,732                     |
| 4    | 0.0757   | 24,577            | 17,471             | 7,106                    | 223,626                     |
| 5    | 0.0694   | 23,583            | 15,526             | 8,057                    | 215,569                     |
| 6    | 0.0720   | 23,968            | 15,521             | 8,447                    | 207,121                     |
| 7    | 0.0734   | 24,164            | 15,196             | 8,968                    | 198,154                     |
| 8    | 0.0611   | 22,528            | 12,108             | 10,420                   | 187,733                     |
| 9    | 0.0600   | 22,397            | 11,271             | 11,126                   | 176,607                     |
| 10   | 0.0836   | 25,170            | 14,761             | 10,409                   | 166,198                     |
| 11   | 0.0802   | 24,788            | 13,325             | 11,463                   | 154,735                     |
| 12   | 0.0717   | 23,925            | 11,100             | 12,825                   | 141,910                     |
| 13   | 0.0670   | 23,488            | 9,505              | 13,983                   | 127,927                     |
| 14   | 0.0783   | 24,432            | 10,022             | 14,410                   | 113,517                     |
| 15   | 0.0623   | 23,252            | 7,071              | 16,180                   | 97,337                      |
| 16   | 0.0807   | 24,421            | 7,850              | 16,570                   | 80,767                      |
| 17   | 0.0786   | 24,312            | 6,352              | 17,960                   | 62,807                      |
| 18   | 0.0704   | 23,951            | 4,423              | 19,528                   | 43,279                      |
| 19   | 0.0727   | 24,028            | 3,147              | 20,880                   | 22,399                      |
| 20   | 0.0660   | 23,878            | 1,479              | 22,399                   | 0                           |

Total Payment: \$478,742

### Fixed Rate Simulation

| Year | $i$   | Payment<br>(\$/y) | Interest<br>(\$/y) | Principal Paid<br>(\$/y) | Remaining Balance<br>(\$/y) |
|------|-------|-------------------|--------------------|--------------------------|-----------------------------|
| 1    | 0.075 | 24,523            | 18,750             | 5,773                    | 244,227                     |
| 2    | 0.075 | 24,523            | 18,317             | 6,206                    | 238,021                     |
| 3    | 0.075 | 24,523            | 17,852             | 6,671                    | 231,349                     |
| 4    | 0.075 | 24,523            | 17,351             | 7,172                    | 224,178                     |
| 5    | 0.075 | 24,523            | 16,813             | 7,710                    | 216,468                     |
| 6    | 0.075 | 24,523            | 16,235             | 8,288                    | 208,180                     |
| 7    | 0.075 | 24,523            | 15,613             | 8,910                    | 199,270                     |
| 8    | 0.075 | 24,523            | 14,945             | 9,578                    | 189,693                     |
| 9    | 0.075 | 24,523            | 14,227             | 10,296                   | 179,396                     |
| 10   | 0.075 | 24,523            | 13,455             | 11,068                   | 168,328                     |
| 11   | 0.075 | 24,523            | 12,625             | 11,898                   | 156,430                     |
| 12   | 0.075 | 24,523            | 11,732             | 12,791                   | 143,639                     |
| 13   | 0.075 | 24,523            | 10,773             | 13,750                   | 129,889                     |
| 14   | 0.075 | 24,523            | 9,742              | 14,781                   | 115,107                     |
| 15   | 0.075 | 24,523            | 8,633              | 15,890                   | 99,217                      |
| 16   | 0.075 | 24,523            | 7,441              | 17,082                   | 82,136                      |
| 17   | 0.075 | 24,523            | 6,160              | 18,363                   | 63,773                      |
| 18   | 0.075 | 24,523            | 4,783              | 19,740                   | 44,033                      |
| 19   | 0.075 | 24,523            | 3,302              | 21,221                   | 22,812                      |
| 20   | 0.075 | 24,523            | 1,711              | 22,812                   | 0                           |

Total Payment: \$490,461

**Monte Carlo Analysis for Example 10.1 (using MACRS 5 yrs)****Probable Variation of Key Parameters over Plant Life**

|                    | <u>Lower Limit</u> | <u>Upper Limit</u> | <u>Base Value</u> |
|--------------------|--------------------|--------------------|-------------------|
| FCIL               | -20%               | 35%                | \$ 150,000,000    |
| Price of Product   | -25%               | 10%                | \$ 75,000,000     |
| Working Capital    | 0%                 | 0%                 | \$ 30,000,000     |
| Income Tax Rate*   | 0%                 | 0%                 | 45%               |
| Interest Rate*     | 0%                 | 0%                 | 10%               |
| Raw Material Price | -10%               | 25%                | \$ 30,000,000     |
| Salvage Value      | 0%                 | 0%                 | 10000000          |

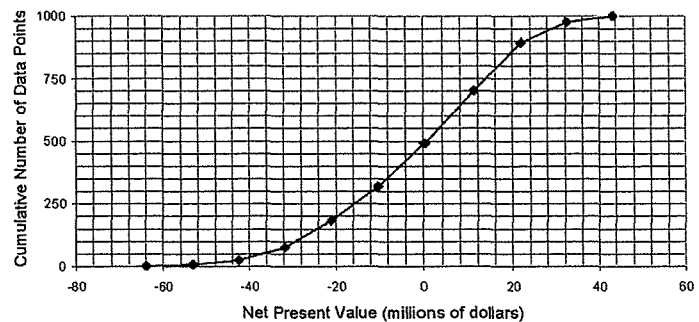
\* Please note that variations for percentages are a percent of a percent.  
For example, a 10% variance on a 12% interest rate would imply a 1.2% uncertainty

Run Economic Analysis

**Net Present Value Data**

Low NPV -63.9  
High NPV 43.3

| <u>Bins</u> | <u>Upper Value</u> | <u># points/bin</u> | <u>Cumulative</u> |
|-------------|--------------------|---------------------|-------------------|
| 0           | -63.9              | 0                   | 0                 |
| 1           | -53.2              | 7                   | 7                 |
| 2           | -42.5              | 16                  | 23                |
| 3           | -31.8              | 51                  | 74                |
| 4           | -21.0              | 108                 | 182               |
| 5           | -10.3              | 137                 | 319               |
| 6           | 0.4                | 172                 | 491               |
| 7           | 11.1               | 211                 | 702               |
| 8           | 21.8               | 190                 | 892               |
| 9           | 32.5               | 84                  | 976               |
| 10          | 43.3               | 24                  | 1000              |



10.43 (a)  $NPV = -FCI + \left(\frac{P}{A}, i, n\right) (R - C_{WT} - C_{RM} - C_{UT})$

$$NPV = -5 + \left(\frac{(1+0.15)^{10} - 1}{(0.15)(1+0.15)^{10}}\right) (4.84 - 0.506 - 2.2 - 1.32) = -\$0.914 \text{ million}$$

This process is not profitable.

(b) When  $NPV = 0$ , number of batches = 26.9 batches/yr

27 batches/yr

(c) When  $NPV = 0$  and number of batches = 26/yr,  $i = 14.1\%$

$i = 14.1\%$

10.44 (a)  $INPV = -FCI + YS(P / A, i, n)$  or  $EAOC = FCI(A/P, i, n) - YS$

$$INPV = -750,000 + (1,321,125 - 950,000) \left(\frac{(1+0.17)^5 - 1}{(0.17)(1+0.17)^5}\right) (1.25) = \$734,196$$

$$EAOC = (750,000) \left(\frac{(0.17)(1+0.17)^5}{(1+0.17)^5 - 1}\right) (1.25) = -\$229,483$$

Recommend the process

(b)  $INPV = -750,000 + (1,321,125 - 950,000) \left(\frac{(1+0.17)^5 - 1}{(0.17)(1+0.17)^5}\right) (x) = \$734,196$

$$x = 1.12$$

Capacity increase of 12%

(c) When  $FCI = \$750,000$ , capacity increase is 25%, and  $NPV = 0$   
 $i = 55\%$

DCFROR is reached



10.45 (a)  $INPV = -FCI + YS(P/A, i, n)$  or  $EAOC = FCI(A/P, i, n) - YS$

$$INPV = -5.1 + (2.75 - 1.5) \left( \frac{(1 + 0.2)^6 - 1}{(0.2)(1 + 0.2)^6} \right) (1.20) = -\$0.11 \text{ million}$$

$$EAOC = (5.1) \left( \frac{(0.2)(1 + 0.2)^6}{(1 + 0.2)^6 - 1} \right) (1.2) = \$0.034 \text{ million}$$

Do not recommend this process.

(b) When  $INPV = 0$  or  $EAOC = 0$ ,  $FCI = \$4.99$  million

$$FCI = \$1.66 \text{ million/tank}$$

(c) When  $INPV = 0$  or  $EAOC = 0$  and  $FCI = \$2$  million/tank,  $i = 0.13$

$$DCF\text{ROR} = 13\%$$

## Chapter 11

- 11.1 For the ethylbenzene process shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics, P-301, V-302, T-302. Comment on any significant differences that you find.

### **P-301 – From Table B.2.3**

P-301 A/B, Carbon steel – positive displacement, Actual power = 15 kW, Efficiency 75%

From Table 11.9, Heuristic 1

Power for pumping liquids:  $\text{kW} = (1.67)[\text{Flow}(\text{m}^3/\text{min})][\Delta P(\text{bar})]/\varepsilon$

$\varepsilon$  = Fractional Efficiency =  $\varepsilon_{sh}$  (see Table 11.5)

density of benzene at 58.5°C = 875 kg/m<sup>3</sup>

mass flow of benzene through pump = 17952.2 kg/h (Stream 3 – Table B.3.1)

Flowrate = 17952.2/875 = 20.5 m<sup>3</sup>/h = 0.342 m<sup>3</sup>/min

$\Delta P = 2000 - 110 = 1890 \text{ kPa} = 18.9 \text{ bar}$

$P_{theoretical} = (1.67)(0.342)(18.9) = 10.8 \text{ kW}$

From Table 11.9 - Heuristic 7

Efficiency 70% at 7.46 kW (10 hp), 85% at 37.3 kW (50 hp)

$$\begin{array}{c} \varepsilon_{sh} \sim 75\% \\ P_{actual} = 10.8/0.75 = 14.4 \text{ kW} \end{array}$$

**Good agreement with heuristics**

### **V-302 - From Table B.2.3**

V-302, Carbon steel with SS demister, vertical,  $L/D = 3$ ,  $V = 10 \text{ m}^3$ , Maximum operating pressure = 250 kPa

From Table 11.6 – Heuristic 4

Optimum length/diameter = 3, but the range 2.5 to 5 is common.

$$\text{Use } L/D = 3$$

From Table 11.6 – Heuristic 5

Holdup time is 5 min for half-full reflux drums and gas/liquid separators, 5–10 min for a product feeding another tower.

Since L/V separator feeds a tower use a hold up of liquid equal to 10 min for half-full drum

Properties of Stream 15

Vapor

Flowrate = 1038 kg/h

Density = 2.169 kg/m<sup>3</sup>Vol flow = 478.6 m<sup>3</sup>/hProperties of Stream 16

Liquid

Flowrate = 24345.9 kg/h

Density = 821.3 kg/m<sup>3</sup>Vol flow = 29.64 m<sup>3</sup>/h

$$\text{Volume of drum, } V = \frac{(29.64)(10)}{(60)(0.5)} = 10 \text{ m}^3$$

**Good agreement with heuristics****T-302 – From TableB.2.3**

T-302, carbon steel, 76 SS sieve trays plus reboiler and total condenser, 45% efficient trays, feed on tray 56, additional feeds ports on 50 and 62, reflux ratio = 0.6608, 15 in tray spacing, column height 28.96 m, diameter = 1.5 m, maximum pressure rating of 300 kPa

**From Tables 11.13 and 11.14– Heuristics**

Table 11.13

Rule 5: Optimum reflux in the range of 1.2 – 1.5  $R_{min}$ Rule 6: Optimum number of stages approximately  $2N_{min}$ Rule 7:  $N_{min} = \ln \{ [x/(1-x)]_{ovhd} / [x/(1-x)]_{bot} \} / \ln \alpha$ Rule 8:  $R_{min} = \{F/D\} / (\alpha - 1)$ 

Rule 9: Use a safety factor of 10% on number of trays

Table 11.14 → Rule 2:  $F_s = u \rho_v^{0.5} = 1.2 \rightarrow 1.5 \text{ m/s(kg/m}^3)^{0.5}$  and Rule 4:  $\epsilon_{tray} = 60 - 90 \%$ **For our case we have**

| Stream              | 18      | 19     | 20     |
|---------------------|---------|--------|--------|
| Temp °C             | 145.4   | 139.0  | 191.1  |
| Pres kPa            | 120.0   | 110.0  | 140.0  |
| Vapor mole fraction | 0.0     | 0.0    | 0.0    |
| Total kmol/h        | 101.1   | 89.9   | 11.3   |
| Total kg/h          | 11024.5 | 9538.6 | 1485.9 |
| Flowrates in kmol/h |         |        |        |
| Ethylene            | 0.00    | 0.00   | 0.00   |
| Ethane              | 0.00    | 0.00   | 0.00   |
| Propylene           | 0.00    | 0.00   | 0.00   |
| Benzene             | 0.17    | 0.17   | 0.00   |
| Toluene             | 0.00    | 0.00   | 0.00   |
| Ethylbenzene        | 90.63   | 89.72  | 0.91   |
| 1,4-DiEthBenzene    | 10.35   | 0.0001 | 10.35  |

Key components are ethylbenzene and 1,4-DiEthylbenzene. The formulae for  $N_{min}$ , etc. should be based on these key components, i.e., use a benzene free basis.

$$x_{ovhd} = 89.72/89.7201 = 0.999999, x_{bot} = 0.91/11.3 = 0.08053, \alpha_{ovhd} = 3.83, \alpha_{bot} = 3.19$$

$$\alpha_{geom\ ave} = (\alpha_{ovhd}\alpha_{bot})^{0.5} = 3.50$$

$$N_{min} = \ln\{ [0.999999/(1 - 0.999999)]/[0.08053/(1 - 0.08053)]\} / \ln(3.50) = 12.9$$

$$R_{min} = \{101.1/89.9\}/(3.5 - 1) = 0.453$$

$$\text{Range of } R = (1.2 \rightarrow 1.5)R_{min} = 0.544 \rightarrow 0.680$$

|                      |              |
|----------------------|--------------|
| $R_{T-302} = 0.6608$ | Within range |
|----------------------|--------------|

$$N_{theoretical} \approx (2)(12.9) = 25.8$$

$$\varepsilon_{tray} = 0.60 - 0.90$$

|                              |                          |
|------------------------------|--------------------------|
| $\varepsilon_{T-302} = 0.45$ | Lower than typical range |
|------------------------------|--------------------------|

$$N_{actual} \approx (25.8/0.45)(1.1) = 63 \text{ trays}$$

|                  |                  |
|------------------|------------------|
| $N_{T-302} = 76$ | High by 13 trays |
|------------------|------------------|

$$\rho_v = 3.546 - 3.879 \rightarrow \text{use } 3.71 \text{ kg/m}^3$$

$$u = (1.2 \rightarrow 1.5)/3.71^{0.5} = 0.62 \rightarrow 0.78 \text{ m/s}$$

$$\text{Vapor flow rate (stream 19)} = 9539 \text{ kg/h}$$

$$\text{Vol. flow rate, } v = 0.714 \text{ m}^3/\text{s}$$

$$D_{tower} = [4v/\pi u]^{0.5} = [(4)(.714)/(3.142)/(0.62 \rightarrow 0.78)]^{0.5} = 1.08 \rightarrow 1.21 \text{ m}$$

|                             |                               |
|-----------------------------|-------------------------------|
| $D_{T-302} = 1.5 \text{ m}$ | Higher than above range for D |
|-----------------------------|-------------------------------|

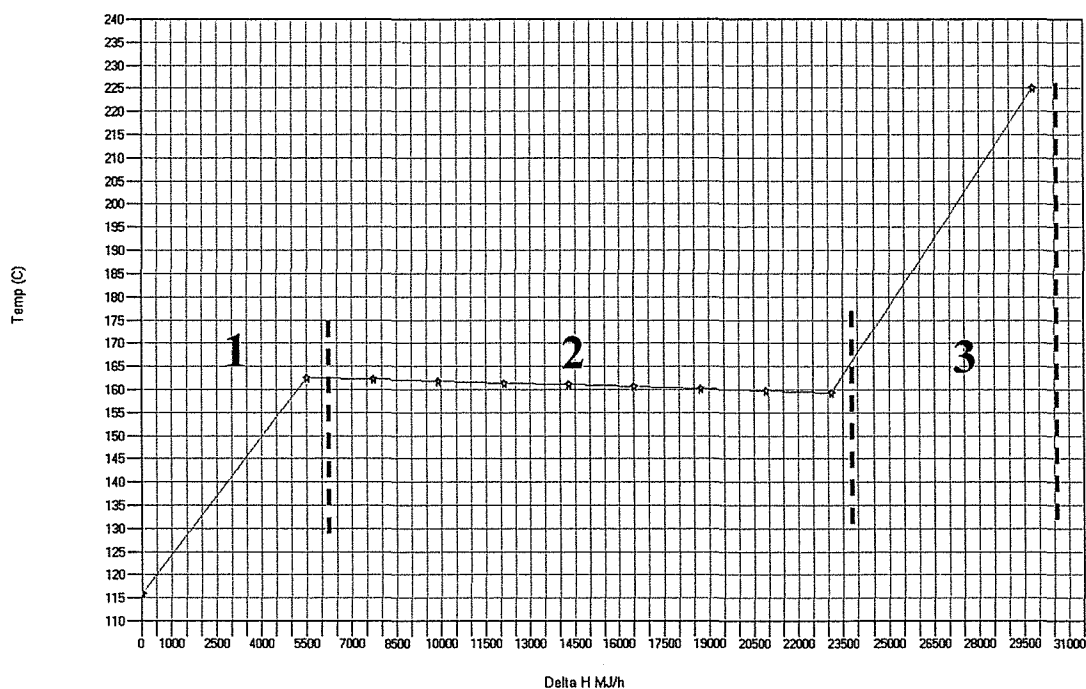
**Overall, for T-302, the agreement with the heuristics is fair. This is probably due, in part, to the use of a fairly low reflux ratio that is at the bottom of the typical range given in the heuristics. This tends to increase the number of trays. In addition, the tray spacing used in T-302 is only 12 or 15" which is significantly lower than the standard spacing of 24". This has the effect of reducing the tray efficiency and requires a larger diameter column to stay away from flooding. These differences probably account for the discrepancies in the above results.**

- 11.2 For the styrene process shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristics, E-401, C-401, T-402. Comment on any significant differences that you find.

**E-401 – From Table B.3.3**

carbon steel,  $A = 541 \text{ m}^2$ , boiling in shell, condensing in tubes, 1 shell – 2 tube passes,  $Q = 29,695 \text{ MJ/h}$

From Table B.3.1 and Figure B.3.1, it can be seen that E-401 heats Stream 2 from  $116^\circ\text{C}$  to  $225^\circ\text{C}$  using high-pressure steam at  $254^\circ\text{C}$ . The T-Q diagram from Chemcad is given below:



This is a 3 zone HX.

$$Q_1 = 5,500 \text{ MJ/h}$$

$$Q_2 = 23,000 - 5,500 = 17,500 \text{ MJ/h}$$

$$Q_3 = 29,700 - 23,000 = 6,700 \text{ MJ/h}$$

**From Table 11.11 – Heuristic 8**

Heat transfer coefficients for estimating purposes,  $\text{W/m}^2\text{C}$ : water to liquid, 850; condensers, 850; liquid to liquid, 280; liquid to gas, 60; gas to gas 30; reboiler 1140. Maximum flux in reboiler  $31.5 \text{ kW/m}^2$ .

So choose  $U_1 = 500$  (liq – cond steam).  $U_2 = 850$  (boiling liq – condensing steam,  $U_3 = 60$  (condensing steam – gas)

$$\Delta T_{lm-1} = \frac{[(254-115)-(254-163)]}{\ln \frac{(254-115)}{(254-163)}} = 113.3^\circ \text{C}$$

$$\Delta T_{lm-2} = (254-163) = 91^\circ \text{C}$$

$$\Delta T_{lm-3} = \frac{[(254-163)-(254-225)]}{\ln \frac{(254-163)}{(254-225)}} = 54.2^\circ \text{C}$$

Using a value of  $F = 1$  – since all zones have a phase change, we get

$$A_1 = \frac{Q_1}{U_1 \Delta T_{lm-1}} = \frac{(5,500 \times 10^6)}{(3600)(500)(113.3)} = 27.0 \text{m}^2$$

$$A_2 = \frac{Q_2}{U_2 \Delta T_{lm-2}} = \frac{(17,500 \times 10^6)}{(3600)(850)(91)} = 62.8 \text{m}^2$$

$$A_3 = \frac{Q_3}{U_3 \Delta T_{lm-3}} = \frac{(6,700 \times 10^6)}{(3600)(60)(54.2)} = 572.3 \text{m}^2$$

$$A_{total} = \sum_{i=1}^3 A_i = 662.1 \text{m}^2$$

|                               |
|-------------------------------|
| $A_{E-401} = 541 \text{ m}^2$ |
|-------------------------------|

**Agreement is within 20% of heuristic - OK**

### **C-401– From Table B.3.3**

carbon steel,  $\dot{W} = 364.2 \text{ kW}$ , 80% adiabatic efficiency

Feed stream = Stream 16

$T = 65^\circ \text{C}$ ,  $P = 0.75 \text{ bar}$ ,  $m = 2682 \text{ kg/h}$ ,  $\text{mw} = 12.4 \text{ kg/kmol}$

$P_{out} = 2.4 \text{ bar}$

### **From Table 11.10 – Heuristic 2**

Theoretical reversible adiabatic power =  $mz_1RT_1[(\{P_2/P_1\}^a - 1)]/a$

where  $T_1$  is inlet temperature,  $R$  = Gas Constant,  $z_1$  = compressibility,  $m$  = molar flow rate,  $a = (k-1)/k$  and  $k = C_p/C_v$ ,  $R = 8.314 \text{ J/mol K}$

assume  $k = 1.4$ ,  $a = 0.2857$

$$P_{rev-adiab} = \frac{2682 \times 10^3}{(12.4)(3600)} \frac{(1)(8.314)(273.2 + 65)}{0.2857} \left[ \left( \frac{2.4}{0.75} \right)^{0.2857} - 1 \right] = 233085 \text{ W} = 233 \text{ kW}$$

From Table 11.10 – Heuristic 8

Efficiencies of large centrifugal compressors is about 76 – 78%

$$P_{actual} = 233/0.77 = 303 \text{ kW}$$

$$\begin{aligned}\epsilon_{C-401} &= 0.80 \\ P_{C-401} &= 364 \text{ kW}\end{aligned}$$

**Compressor appears to be somewhat oversized (~20%), although the estimate is in the ballpark. This may be to allow for future expansion or could be an error in the design calculations.**

**T-402– From Table B.3.3**

carbon steel, total condenser (E-409), feed at location equivalent to tray 36, reflux ratio = 25.8, structured packing,  $C_f = 1$ , diameter = 4.1 m,  $HETP = 0.3$  m, height = 34.5 m

**From Tables 11.13 and 11.14– Heuristics**

Table 11.13

Rule 5: Optimum reflux in the range of  $1.2 - 1.5 R_{min}$

Rule 6: Optimum number of stages approximately  $2N_{min}$

Rule 7:  $N_{min} = \ln\{ [x/(1-x)]_{ovhd}/[x/(1-x)]_{bot} \} / \ln \alpha$

Rule 8:  $R_{min} = \{F/D\}/(\alpha - 1)$

Rule 9: Use a safety factor of 10% on number of trays

Table 11.14

Rule 2:  $F_s = u_p^{0.5} = 1.2 \rightarrow 1.5 \text{ m/s(kg/m}^3)^{0.5}$

Rule 4:  $\epsilon_{tray} = 60 - 90 \%$

**From Table 11.15 – Heuristic 9**

Height equivalent to theoretical stage (HETS) for vapor-liquid contacting is 0.4-0.56 m for 2.5 cm (1 in) pall rings and 0.76–0.9 m. for 5 cm (2 in) pall rings.

Structured packing is generally more efficient than loose packing so an HETP (HETS) = 0.3m is reasonable.

Assuming a headspace of 1.5 m above the packing and a liquid level of 3m at the bottom

$$\text{Number of equivalent theoretical plates} = (34.5 - 4.5)/0.3 = 100 \text{ plates}$$

Using data from Table B.3.3 for the feed and product streams for T-402, we get the following

| Stream No.          | 22     | 23     | 24    |
|---------------------|--------|--------|-------|
| Temperature (°C)    | 119.5  | 105    | 124.5 |
| Pressure (kPa)      | 60     | 210    | 60    |
| Vapor Mole Fraction | 0      | 0      | 0     |
| Total Flow (kg/h)   | 47905  | 35473  | 12432 |
| Total Flow (kmol/h) | 453.9  | 334.2  | 119.7 |
| Component Flows     |        |        |       |
| Water               |        |        |       |
| Ethylbenzene        | 333.0  | 332.66 | 0.34  |
| Styrene             | 120.53 | 1.20   | 119.3 |
| Hydrogen            |        |        |       |
| Benzene             |        |        |       |
| Toluene             | 0.33   | 0.33   |       |
| Ethylene            |        |        |       |
| Methane             |        |        |       |

Using a toluene free basis, we have

$$x_{ovhd} = 332.66/(332.66+1.20) = 0.9964$$

$$x_{bot} = (0.34)/(119.3 + 0.34) = 0.002842$$

From the Chemcad output we get the following values for relative volatility

$$\alpha_{top} = 1.19$$

$$\alpha_{bot} = 1.22$$

$$\alpha_{ave} = [(1.19)(1.22)]^{1/2} = 1.205$$

$$N_{min} = \ln \{ [x/(1-x)]_{ovhd} / [x/(1-x)]_{bot} \} / \ln \alpha$$

$$N_{min} = \frac{\ln \frac{x_{ovhd}(1-x_{bot})}{(1-x_{ovhd})x_{bot}}}{\ln \alpha_{ave}} = \frac{\ln \frac{(0.9964)(1-0.002842)}{(1-0.9964)(0.002842)}}{\ln(1.205)} = 61.7$$

$$R_{min} = \{F/D\}/(\alpha - 1)$$

$$R_{min} = \frac{(453.9)}{(334.2)(1.205-1)} = 6.63$$

Using heuristics we get

$$R = (1.2 - 1.5)R_{min} = 7.95 - 9.94$$

$$N = 2N_{min} = 123$$

$$R_{T-402} = 25.8$$

$$N_{T-402} = 100$$



**The reflux ratio for this column is much higher than the value from the heuristic while the number of trays is lower by ~20%. This column appears to be running inefficiently and should be investigated since the reboiler and condenser duties are probably very high and wasteful in utilities. One possible explanation is that the higher value of  $R$  is chosen to keep the height of the tower down but this seems to be overkill. Tower needs investigating!**

- 11.3 For the drying oil shown in Appendix B, check the design specifications for the following three pieces of equipment against the appropriate heuristic V-501, P-501, H-501. Comment on any significant differences that you find.

**P-501 A/B – From Table B.4.3**

Centrifugal, Carbon steel, Power = 0.9 kW (actual), 80% efficient,  $NPSH_R$  at design flow = 14 ft of liquid

**From Table 11.9**

1. Power for pumping liquids:  $kW = (1.67)[Flow(m^3/min)][\Delta P(bar)]/\epsilon$ ,  $\epsilon$  = Fractional Efficiency =  $\epsilon_{sh}$  (see Table 11.5)
2. Net positive suction head (NPSH) of a pump must be in excess of a certain number, depending upon the kind of pumps and the conditions, if damage is to be avoided.  $NPSH = (\text{pressure at the eye of the impeller} - \text{vapor pressure})/(\rho g)$ . Common range is 1.2–6.1 m of liquid
4. Centrifugal pumps: Single stage for 0.057–18.9  $m^3/min$ , 152 m maximum head; multistage for 0.076–41.6  $m^3/min$ , 1675 m maximum head. Efficiency 45% at 0.378  $m^3/min$ , 70% at 1.89  $m^3/min$ , 80% at 37.8  $m^3/min$ .

**From Table 11.8**

4. Control valves require at least 0.69 bar (10 psi) drop for good control.

Inlet stream is Stream 2 from Figure B.4.1 and Table B.4.1

**Properties of Stream 2**

Liquid

Flowrate = 10,703 kg/h

Density = 793.5  $kg/m^3$

Vol flow = 13.49  $m^3/h$  = 0.2248  $m^3/min$

$P_2$  = 105 kPa = 1.05 bar

$P_3$  = 230 kPa = 2.30 bar (Stream 3)

Note:  $\Delta P_{pump} = P_3 - P_2 + \Delta P_{CV}$

$$P = \frac{1.67V\Delta P}{\epsilon} = \frac{(1.67)(13.49/60)(2.30 - 1.05 + 0.69)}{\epsilon} = \frac{0.728}{\epsilon} \text{ kW}$$

From Table 11.9 – heuristic 4, the efficiency is ~ 45%. This is much lower than the actual efficiency of 80%. Using  $\epsilon$  = 80% we get:

$$P = \frac{0.728}{0.8} = 0.91 \text{ kW}$$

**Excellent agreement with actual power, but the efficiency looks to be too high for such a small pump. This may be a mistake or a high efficiency pump has been specified for this service.**

**H-501 – From Table B.4.3**

total heat duty required = 13219 MJ/h = 3672 kW, design capacity = 4000 kW, Carbon steel tubes, 85% thermal efficiency

**From Table 11.11 – Heuristic 13**

Fired heaters: radiant rate, 37.6 kW/m<sup>2</sup>; convection rate, 12.5 kW/m<sup>2</sup>; cold oil tube velocity = 1.8 m/s approximately equal transfer in the two sections; thermal efficiency 70–90% based on lower heating value; flue gas temperature 140–195°C above feed inlet; stack gas temperature 345–510°C

Only thing to check here is the thermal efficiency which lies in the acceptable range, namely 70%<85%<90% - **good agreement with heuristics.**

**V-501 - From Table B.4.3**

**V-501**, Horizontal, Carbon steel, L/D = 3, V = 2.3 m<sup>3</sup>

**From Table 11.6 – Heuristic 4**

Optimum length/diameter = 3, but the range 2.5 to 5 is common.

Use  $L/D = 3$

**From Table 11.6 – Heuristic 5**

Holdup time is 5 min for half-full reflux drums

**Properties of Stream 2**

Liquid

Flowrate = 10,703 kg/h

Density = 793.5 kg/m<sup>3</sup>

Vol flow = 13.49 m<sup>3</sup>/h

$$\text{Volume of drum, } V = \frac{(13.49)(5)}{(60)(0.5)} = 2.25 \text{ m}^3$$

**Good agreement with heuristics**

## Chapter 12

12.1 When assigning this problem, the students should be told whether a generic answer or a more specific answer is requested. If the more specific answer is required, the problem involves literature searches and involves digging into the process simulator's databank. For some process simulators, the latter is easy, whereas for others it is nearly impossible.

a) The physical property data needed are:

- pure-component heat capacities
- pure-component densities
- equation of state (and its parameters) and/or liquid-state activity coefficient model (and its parameters)
- thermal conductivities
- viscosities
- pure-component vapor pressures

b) Sources of these data are:

- process simulator databank or handbooks
- process simulator databank or handbooks
- process simulator databank or handbooks for critical properties and for some activity coefficient model parameters. Literature data for vapor-liquid equilibria (or liquid-liquid equilibria) from which model parameters can be regressed.
- process simulator databank or handbooks
- process simulator databank or handbooks
- process simulator databank or handbooks

12.2 The U.S. patent literature can be searched at any patent and trademark depository library on-line (1971 to present) or in the hardcopy indices (1790 to present). By Internet, one can use <http://patents.uspto.gov/> to search the front page of U.S. patents from 1976 to the present or <http://patent.womplex.ibm.com/ibm.html> to search U.S. patents 1971 to present (full text and images). Some similar sources are available for patents granted by other countries. Note that many of the useful patents are more than a few years old.

Most such patents include only sketchy experimental data, and then only the bare minimum. For example, the size of the reactor (and catalyst charge), total inlet flowrate and composition, and incomplete productivity data may be presented. For multiple runs, the temperature and/or the inlet composition may have been varied—but maybe not. Space velocities are common.

The process simulator, on the other hand, will require the order with respect to each reactant, the frequency factor (pre-exponential), and the activation energy for each reaction. These data are almost never given in patents. Thus, the student must regress these parameters from the data presented.

12.3 There are many heuristics for reactor design, including:

- Design for heat transfer, not reaction kinetics, as the former is usually controlling.
- Use a CSTR when control of the reactor is essential.
- For exothermic reactions, use a minimum temperature difference between the reactor and the heat-transfer fluid, to increase controllability.
- For endothermic reactions, operate at the lowest possible temperature to minimize heating costs.
- For exothermic reactions, operate at the highest possible temperature to maximize the utility value of the hot stream produced.
- For endothermic reactions, operate at the highest possible temperature to maximize conversion.
- For exothermic reactions, operate at the lowest possible temperature to maximize conversion.
- If the byproduct reaction has a higher activation energy than does the desired reaction, operate at low temperatures.
- If the byproduct reaction has a lower activation energy than does the desired reaction, operate at high temperatures.
- Expect a rapid increase in temperature near the entrance to a tubular reactor for exothermic reactions. Cut the catalyst with inactive support to create a uniform heat release per unit length of reactor.
- Increase selectivity at the expense of yield, as the former can drastically reduce separation costs.
- Design for increased reactor temperature over the life of the catalyst, to compensate for loss of catalyst activity.

Be sure to point out that heuristics are fallible and often mutually inconsistent.

## 12.4

- a) From a capital-costs viewpoint, one wants to minimize the heat-transfer area. This is accomplished if the approach temperature is increased. However, very low temperature utilities can involve higher equipment costs.
- b) From the operating-costs viewpoint, the greater the approach temperature, the smaller the flowrate of cooling medium can be (i.e., the inlet and outlet temperature of the cooling medium can be very different). However, the lower the cooling-medium temperature, the lower is the value of the "cooling" medium as a heat source for heat integration. For example, we might be able to raise high-pressure steam (at high temperature, low approach temperature) or low-pressure steam (at low temperature, high approach temperature). The high-pressure steam is worth more towards lowering operating costs. Thus, there are competing effects on operating costs.
- c) From the operability viewpoint, a smaller approach makes the temperature control more stable. If the approach is one degree, for example, a temperature excursion of one degree would double the heat transfer rate, bringing the reactor back to its design temperature. If the temperature in the reactor suddenly decreased by two degrees, the "cooling" medium would actually heat the reactor back up. On the other hand, if one wanted to change the setpoint for the reactor temperature, a small approach temperature would be a considerable problem. The temperature of the cooling medium would need to be changed.

12.5 The operability of a process involving an exothermic reaction is a serious safety concern. If the cooling system is unable to remove the heat of reaction, the reactor temperature will rise, resulting in an increase in reaction rate. This increased reaction rate will lead to an increased heat release rate. The exponential temperature rise that results can lead to a catastrophic event called a "runaway."

Thus, one tries to design an inherently stable process control system. In this case, the small approach temperature design is preferred. In fact, if there is a sudden failure of the control system, the reactor temperature will remain stable, at least for a little while.

12.6 The following are referred to figure numbers.

- a) B1: Stream 13  
B2: Stream 22  
B3: Stream 14  
C8: Stream 11
- b) B1: Stream 16  
B2: Stream 18  
B3: Stream 10  
B4: Streams 9, 12  
C3: Streams 18, 21

12.7 The following are referred to figure numbers.

- a) B2: Streams 5, 8  
B4: five cooling loops on R-501  
C1: Stream 4  
reflux in each distillation column

- b) B1: pump-around of P-201A/B

There are many other examples, but most involve similar pump-arounds for positive-displacement pumps

- c) B1: Streams 16,17  
B2: Streams 15, 18  
B3: Streams 10, 13  
B4: Streams 9,12  
C3: Streams 15, 18, 21
- d) return of condensate to boiler in every flowsheet  
B1: preheater E-202  
C5: bfw/steam between E-703 and E-702  
Use of steam raised in process for the hot utility (several processes)



12.8 Recycles can also be used for:

- Reduction of waste streams
- Mixing of components in a vessel
- Maintaining a consistent temperature in a vessel
- Reduction of humidity in the air (as in the recirculation mode of an air conditioner)
- Seeding of a crystallizer circuit
- Seeding of microorganisms in biological unit operations.
- Control of particle size distribution

12.9 The required information for utilities includes:

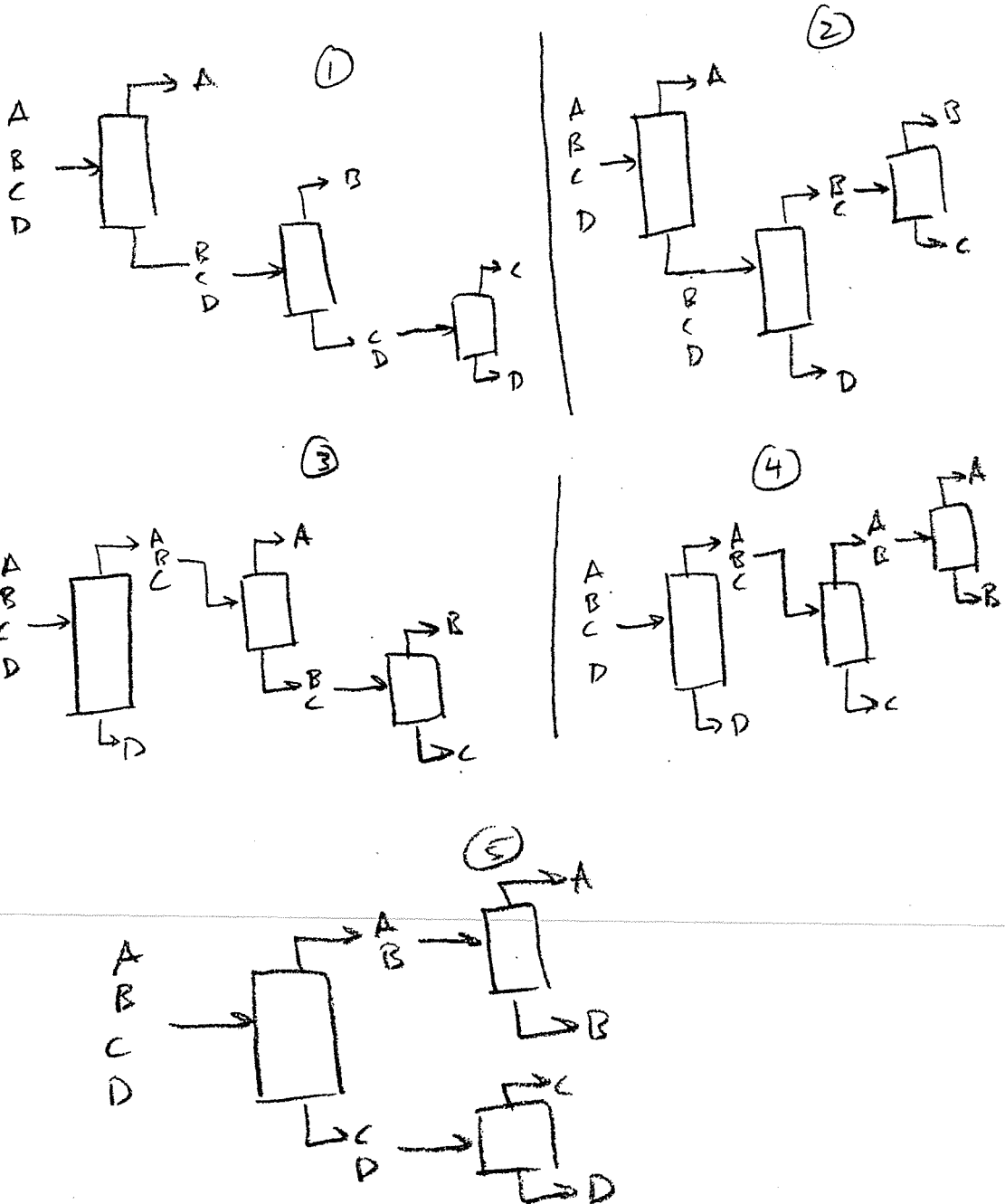
- Cost per unit of energy transferred (e.g., \$/GJ)
- Temperature of utility
- Range of acceptable utility return temperatures
- Pressures of utility
- Maximum flowrate of utility available
- Thermal conductivity, viscosity, and heat capacity of utility or equivalent information to allow for heat-transfer calculations
- Heat of vaporization and vapor pressure for phase-change utility streams
- Cost of pumping of utility
- Availability of utility at site
- Credit for utility returned but unused
- Safety characteristics of the utility
- Operability ranges of the utilities
- Dependability of the utilities (i.e., freedom from interruption, temporal consistency of temperatures and pressures)
- Purity requirements for returned utility stream
- Corrosion properties of utility stream

12.11

Eq. 10, 1

$$S = \frac{[2(N-1)]!}{N!(N-1)!}$$

$$N=4 \quad S = \frac{6!}{4!3!} = \frac{30}{6} = 5$$

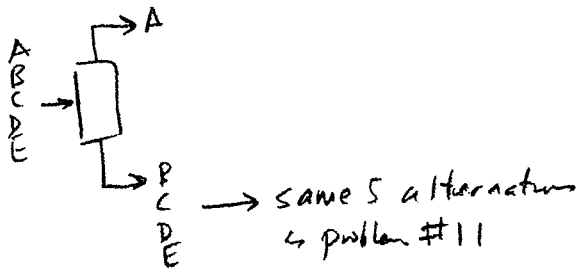


12.12

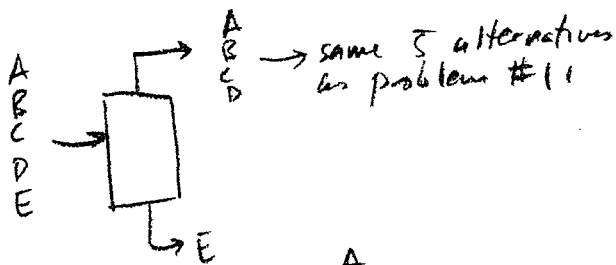
EQ. 10.1

$$S = \frac{[2(N-1)]!}{N!(N-1)!}$$

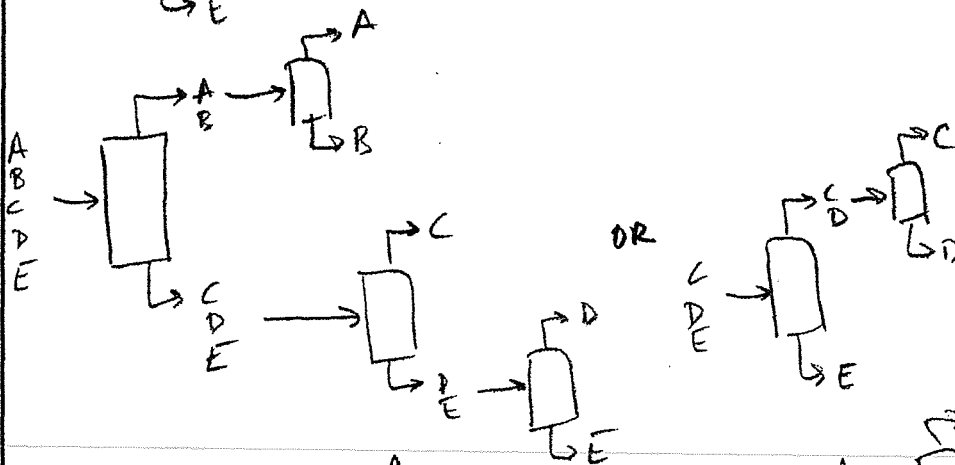
$$N=5 \quad S = \frac{8!}{5!4!} = \frac{8(7)(6)}{4(3)(2)} = 14$$



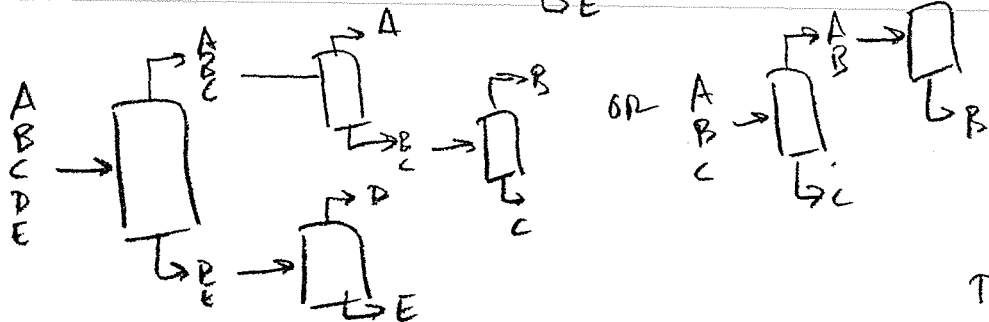
5



5



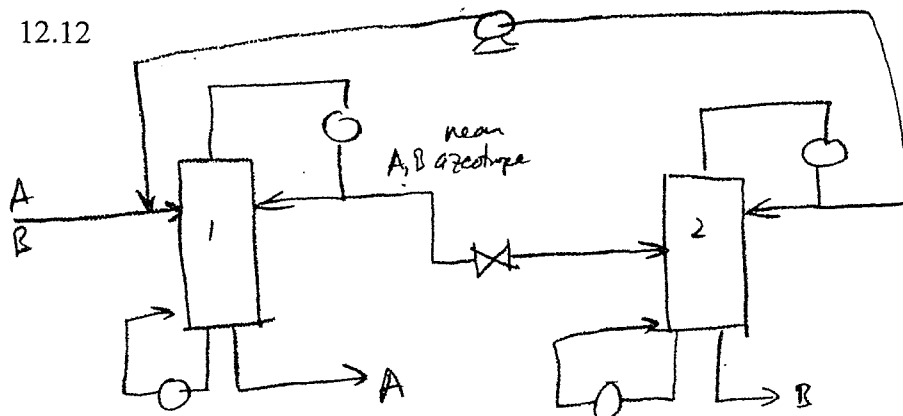
2



2

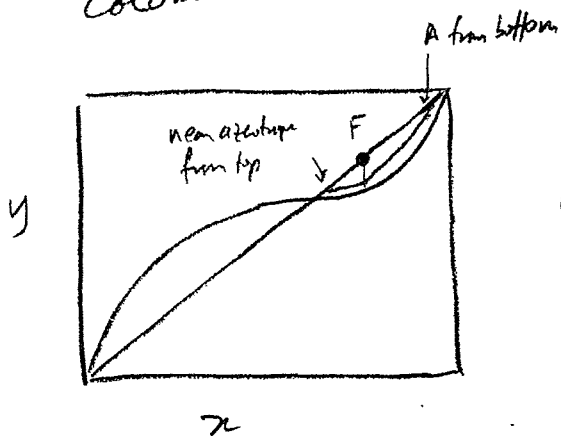
DT = 14

12.12

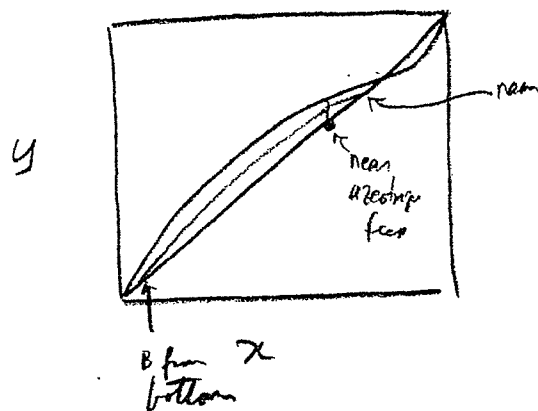


Column 2 not likely at lower P

Column 1



Column 2



lower P - move  
azeotrope to higher  
mole fraction

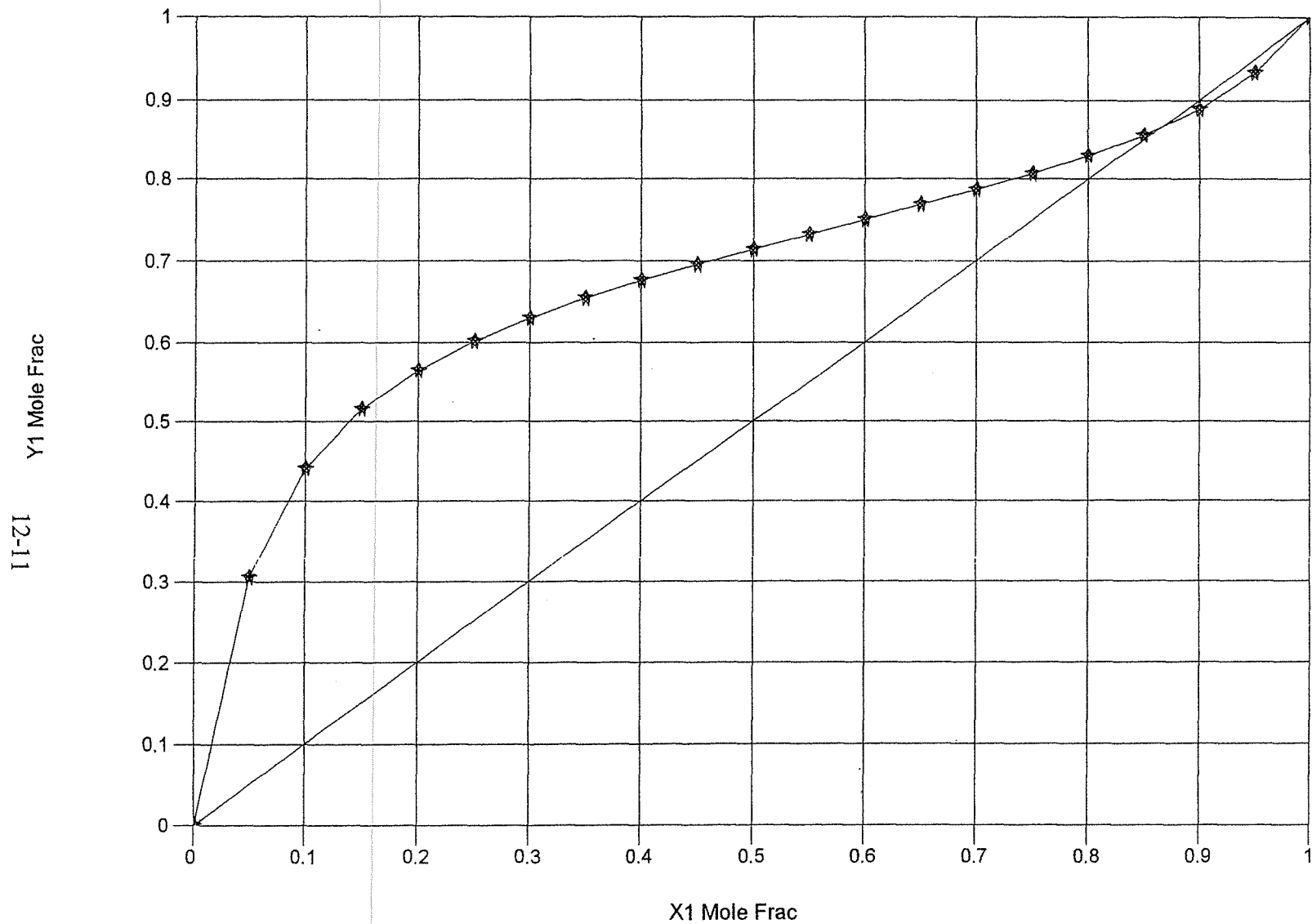
12.14 plots on following pages are from Chemcad

- a) Ideal never predicts azeotrope  
unsuitable to use if existence of azeotrope is known
- b) NRTL predicts pressure-sensitive azeotrope  
at all pressures
- c) UNIFAC does not give hint of predicted  
azeotrope until 1000 kPa
- d) SRK begins to predict azeotrope at 500 kPa

→ Must compare actual data to predicted behaviour  
to validate simulator/thermo model

→ if NRTL is valid → there is a reasonable  
opportunity to do the separation using two  
columns as in Fig 10.4c or in solution to  
problem 10.13 - depending on whether feed  
is above or below azeotropic point

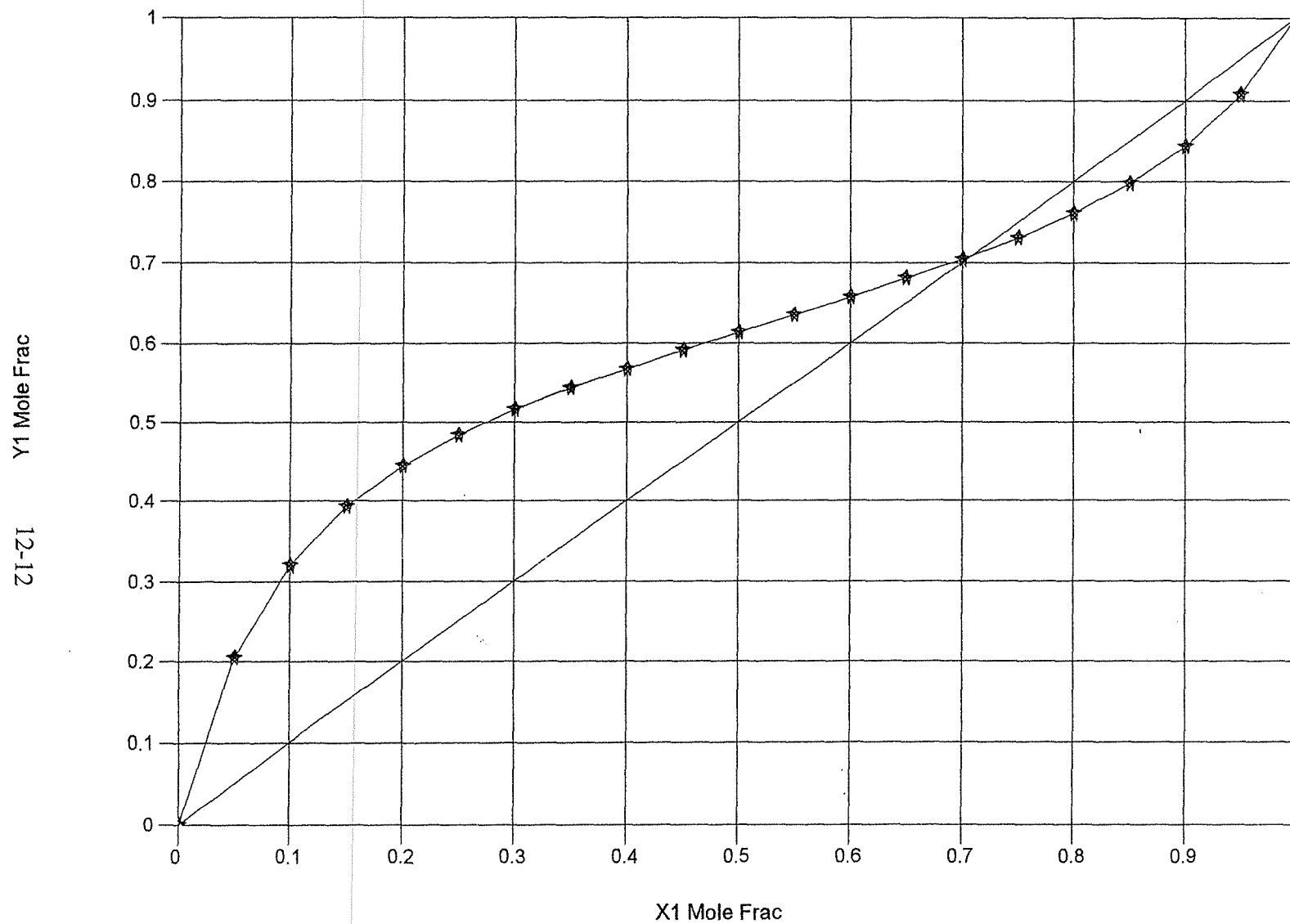
# Methylal / Methanol at 101.32 kPa By NRTL



Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:51:59

★ XY Data

# Methylal / Methanol at 500.00 kPa By NRTL

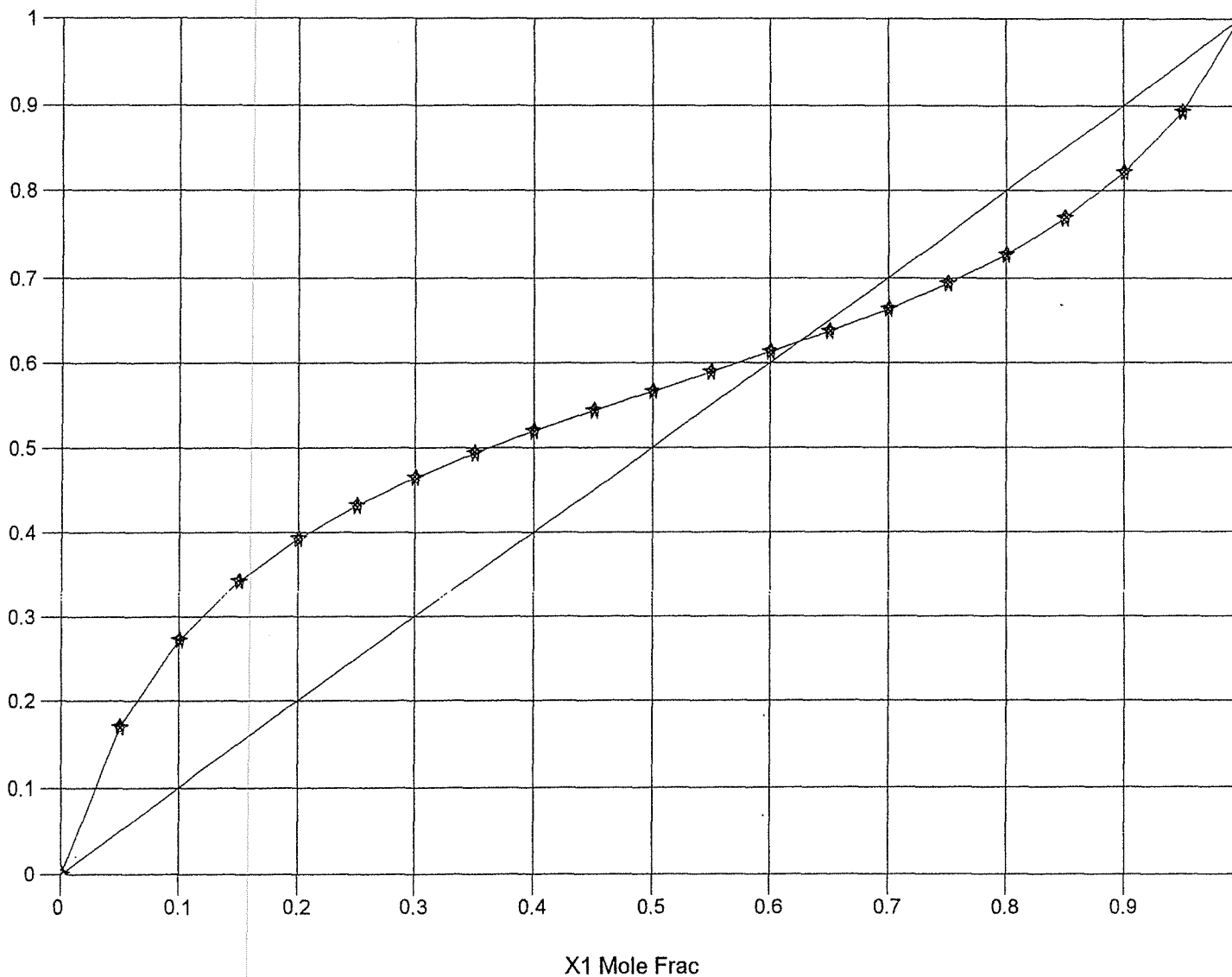


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 11:31:55

★ XY Data

# Methylal / Methanol at 1000.00 kPa By NRTL

12.13  
Y1 Mole Frac

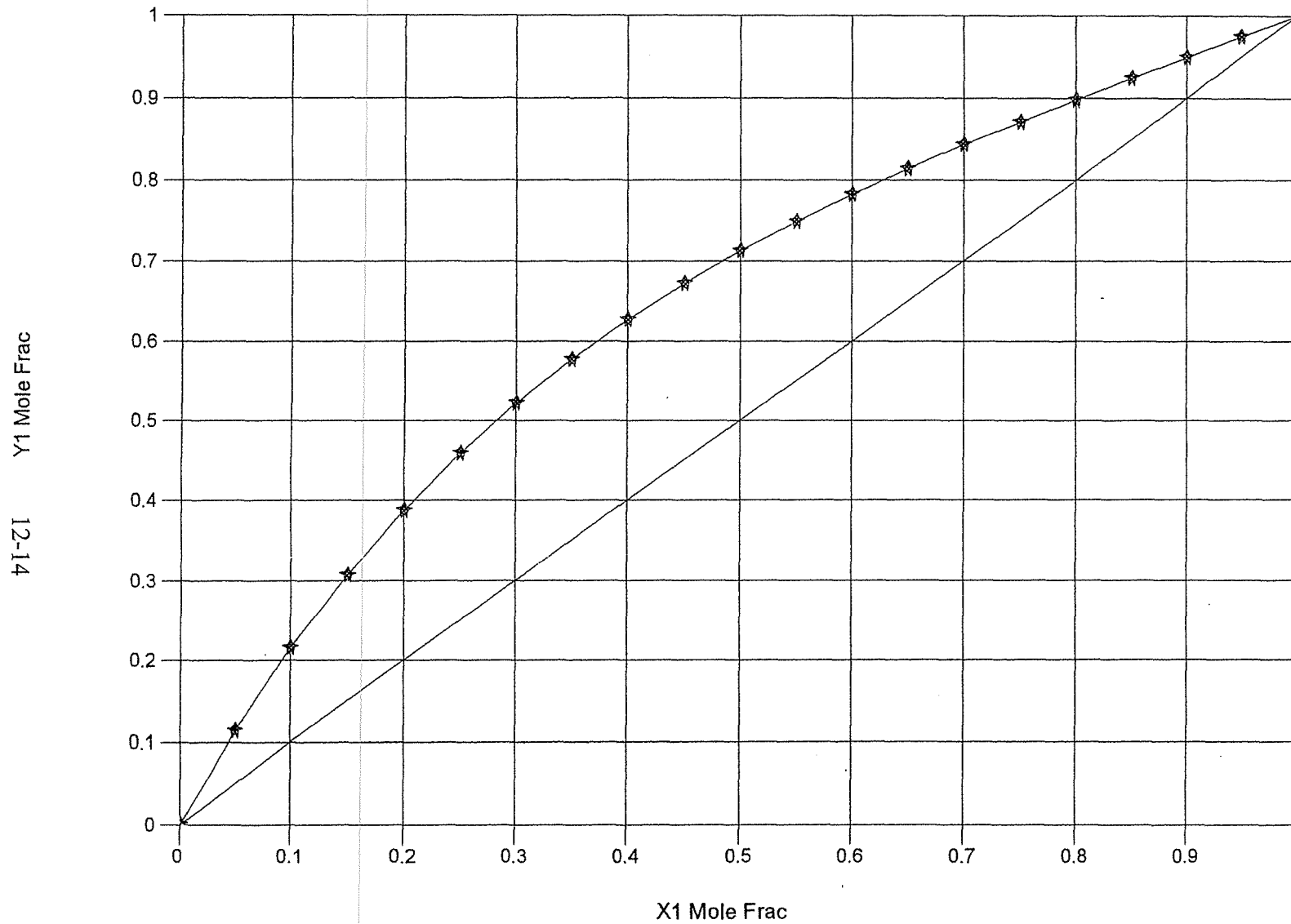


★ XY Data

Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 11:33:01



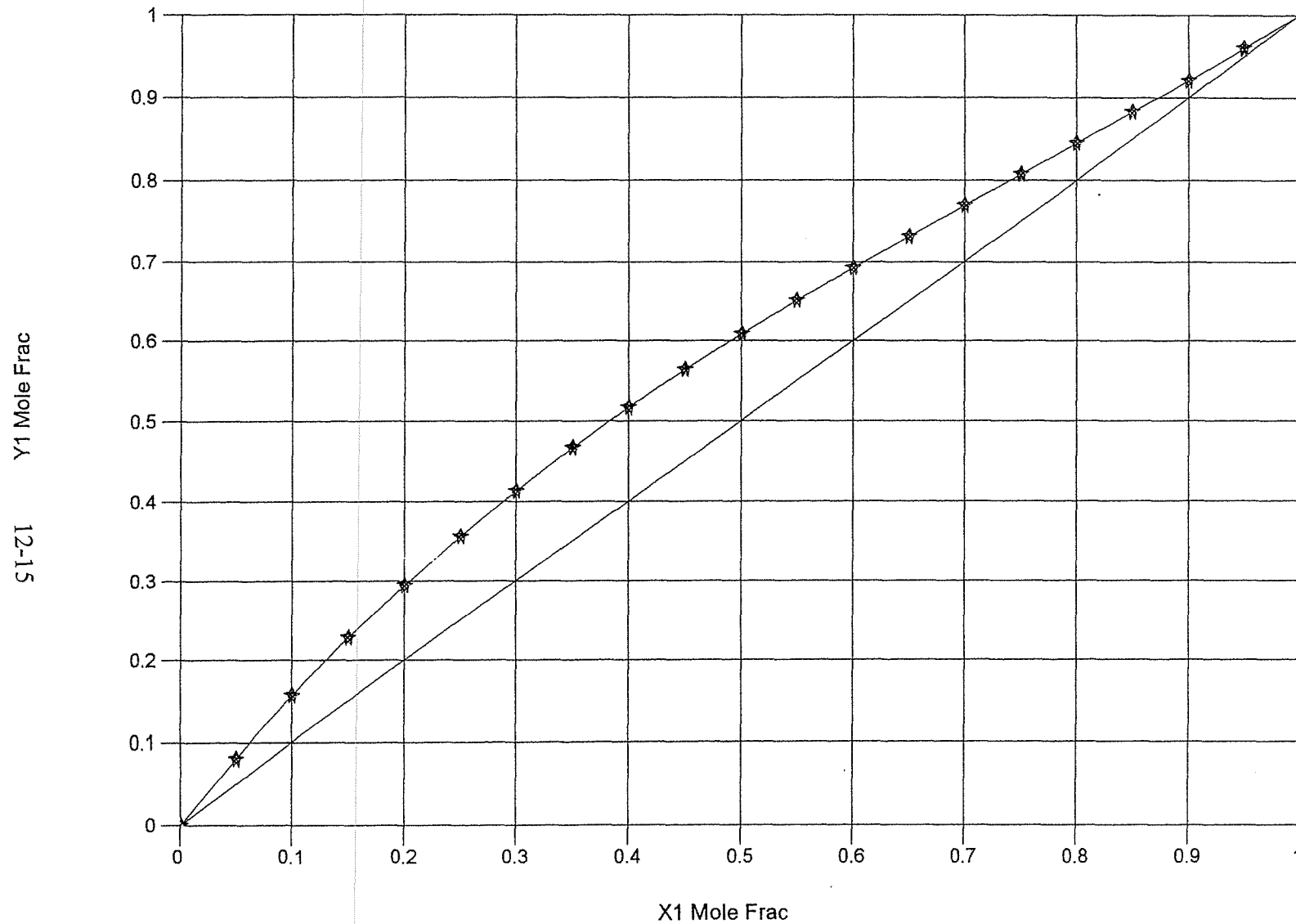
# Methylal / Methanol at 101.32 kPa By UNIF



Job: chapter  
10 binary  
azeotropes  
Date:  
09/12/2002  
Time: 13:57:57

★ XY Data

# Methylal / Methanol at 500.00 kPa By UNIF

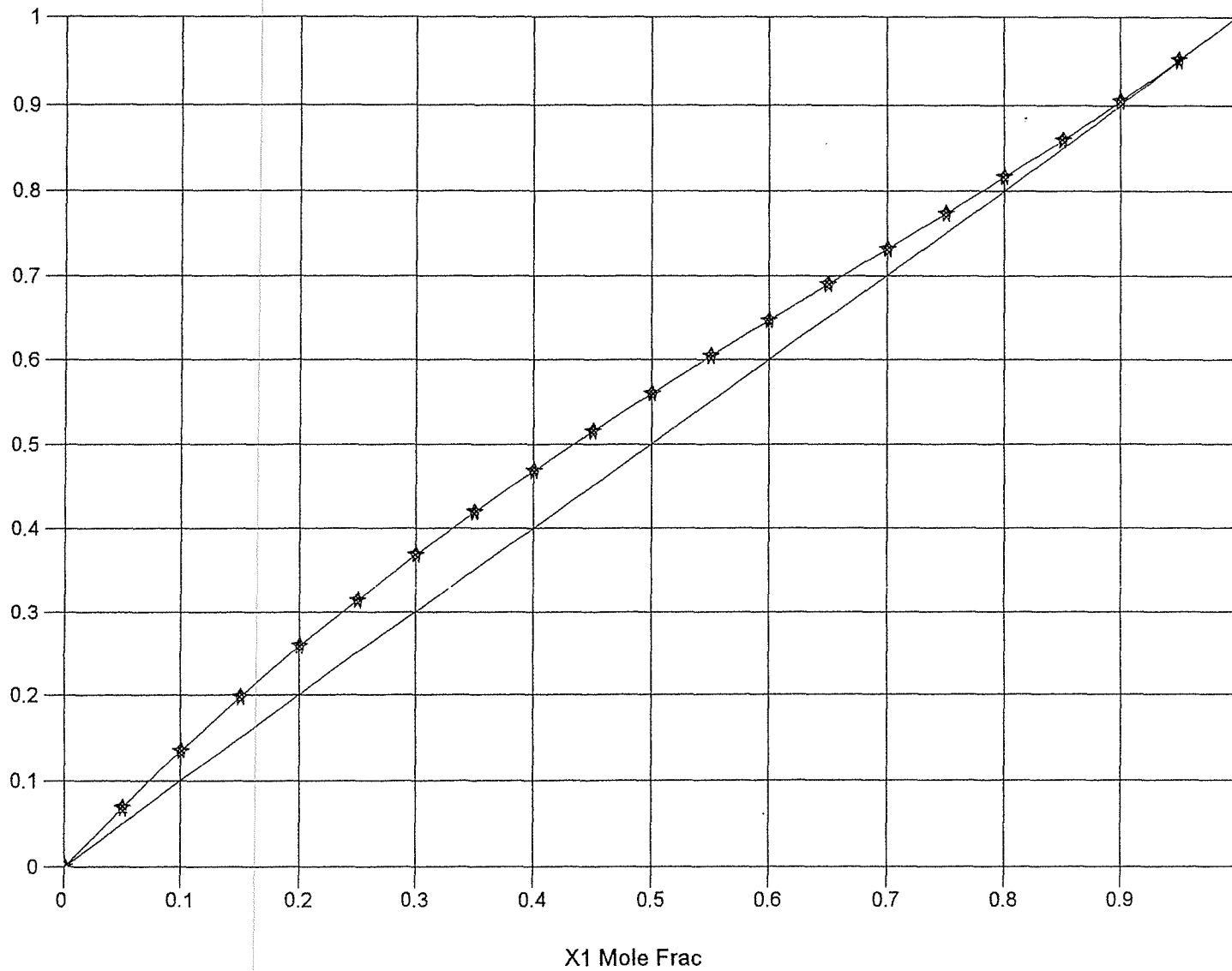


Job: chapter  
10 binary  
azeotropes  
Date:  
09/12/2002  
Time: 13:58:23

★ XY Data

# Methylal / Methanol at 1000.00 kPa By UNIF

91-21 Y1 Mole Frac



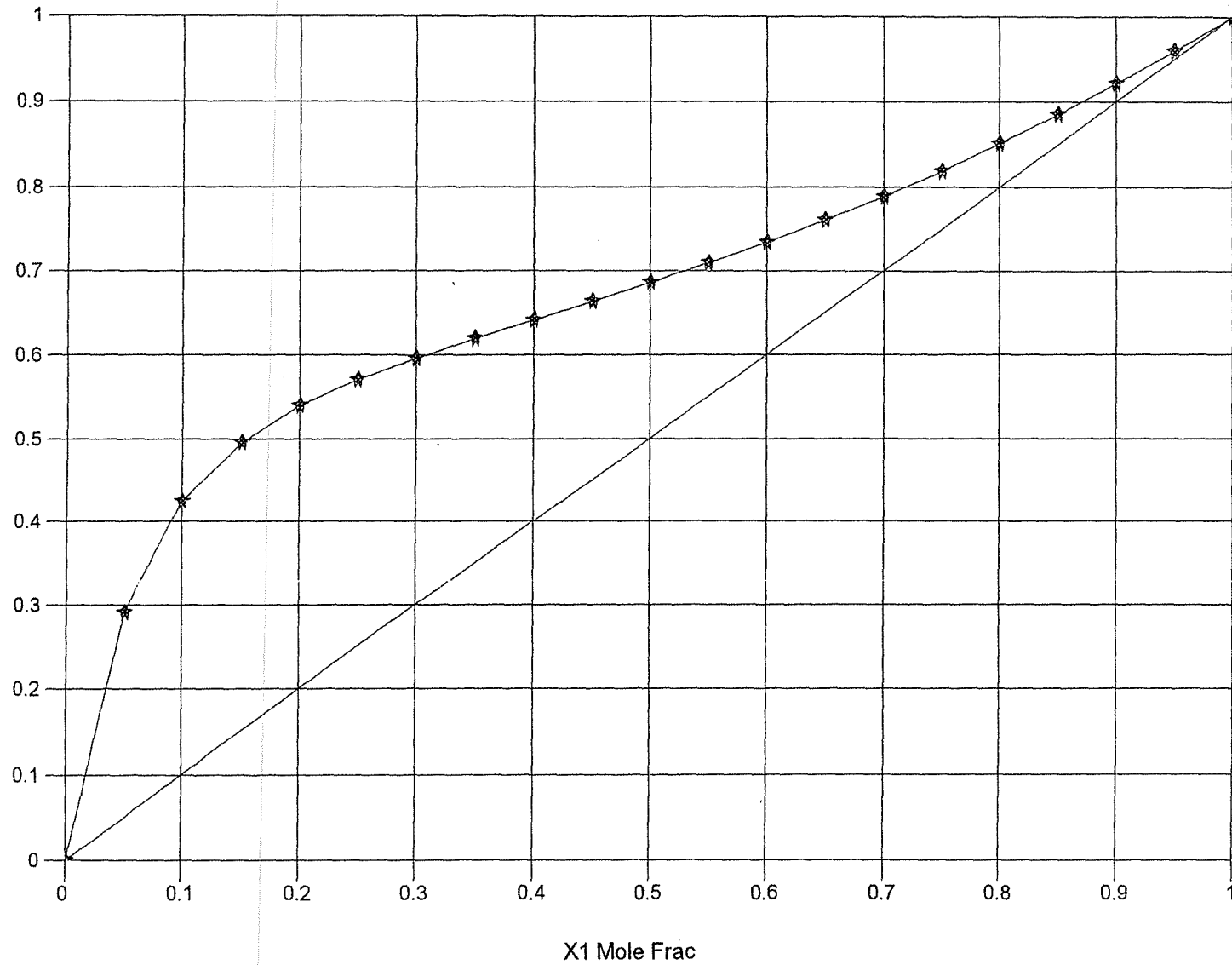
Job: chapter  
10 binary  
azeotropes  
Date:  
09/12/2002  
Time: 13:58:40

★ XY Data

# Methylal / Methanol at 101.32 kPa By SRK

Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:51:32

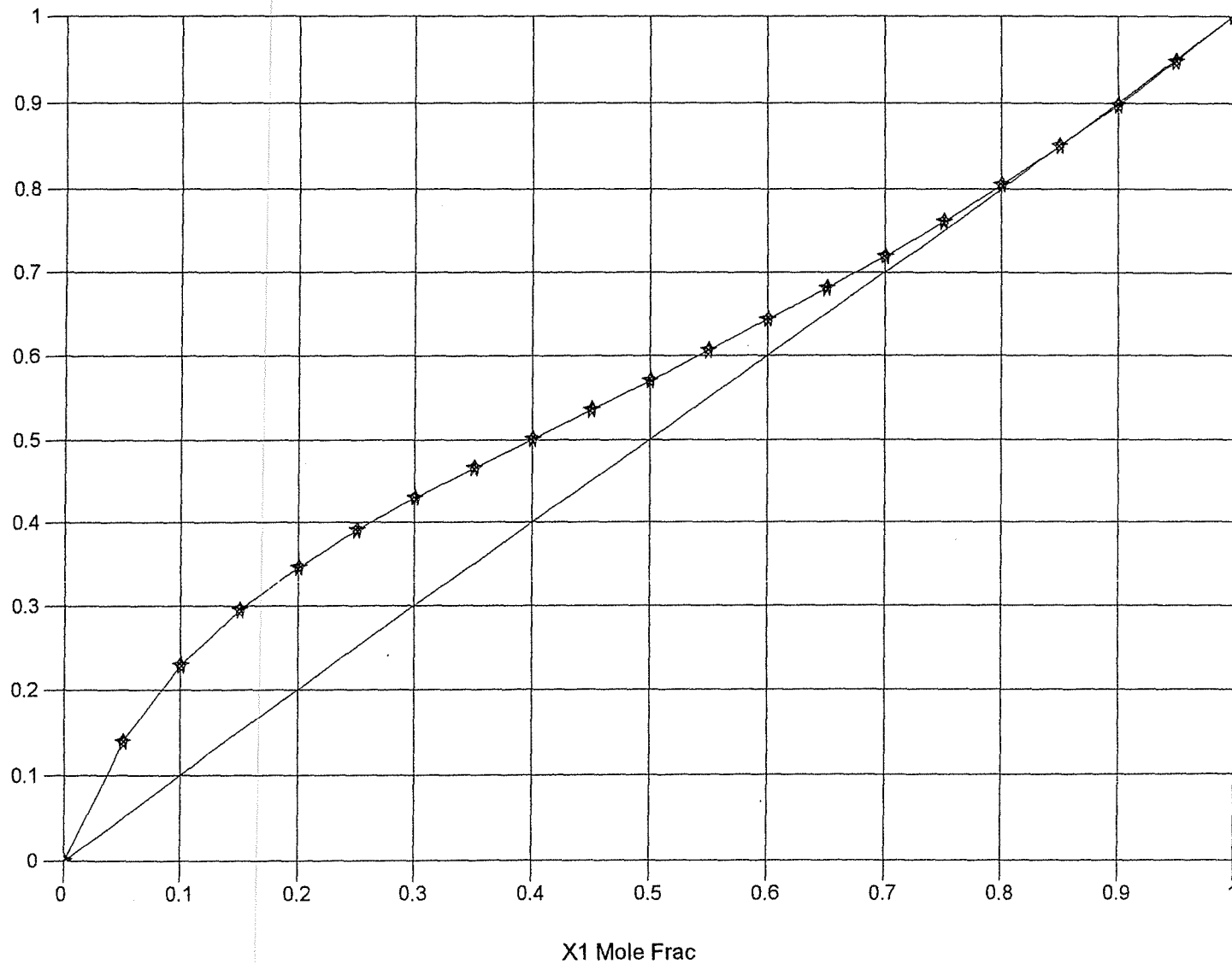
Y1 Mole Frac  
12-17



★ XY Data

# Methylal / Methanol at 500.00 kPa By SRK

81-21 Y1 Mole Frac

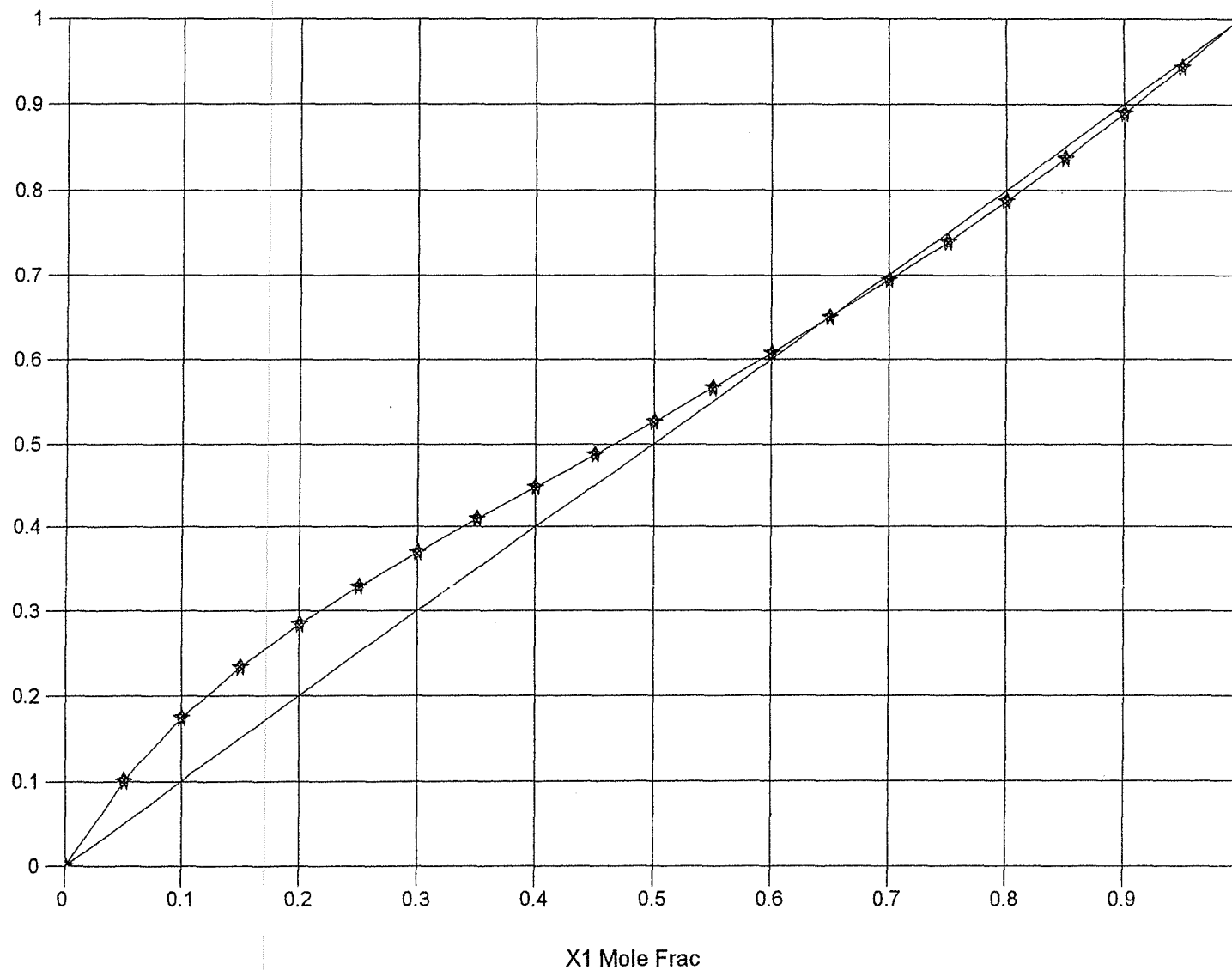


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 11:37:02

★ XY Data

# Methylal / Methanol at 1000.00 kPa By SRK

61-21 Y1 Mole Frac



★ XY Data

Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 11:37:51

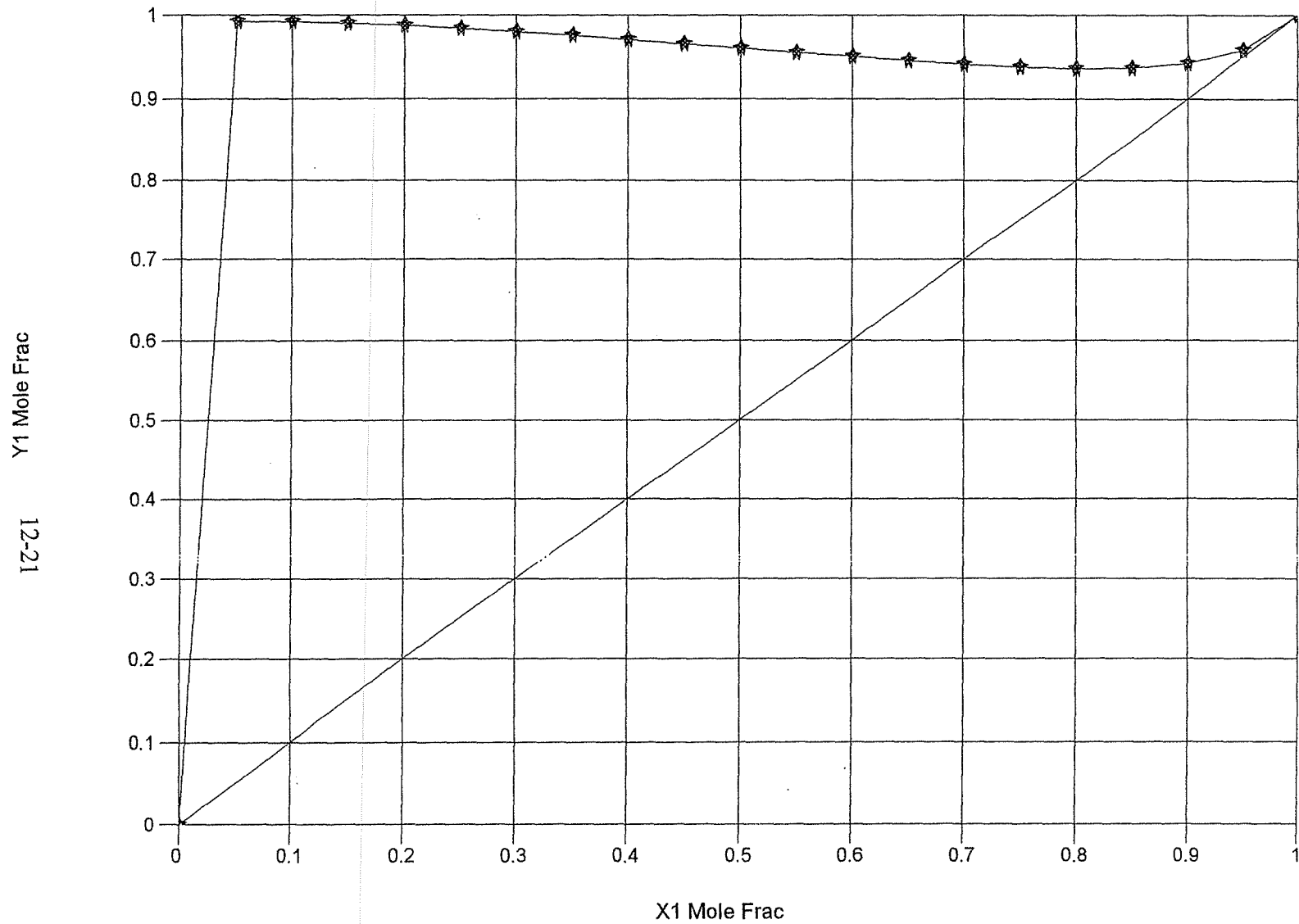
12.15

x-y diagrams from Chemcad on following page

Using UNIQUAC

- at 1500 kPa - distill 50% mixture up to  $\sim 0.7$  mole fr  
then increase P to 5000 kPa - use arrangement in  
Fig 10.4c
  - could reduce P in 1<sup>st</sup> col - say to 500 kPa -  
distill up to 0.8 mole fr  $\rightarrow$  then increase back to  
1500 kPa in second column
  - there are an infinite number of possible pressure  
combinations here
- 
- if only need 99% - may be able to distil at around  
50 kPa in one column (would need lots of trays)
  - in fact - our simulation shows this is possible
- 
- ideal - no azeotropes predicted - unsuitable model
  - NRTL - similar prediction to UNIQUAC
  - UNIFAC - if model were correct (we do not believe it to be)  
could just reduce P and distill to any desired purity
  - SRK - unusual shape of x-y diagrams suggest that  
model is unsuitable - however, if model  
were valid - could work similarly to Fig 10.4c

# Diethyl Ether / Water at 50.00 kPa By UNIQ

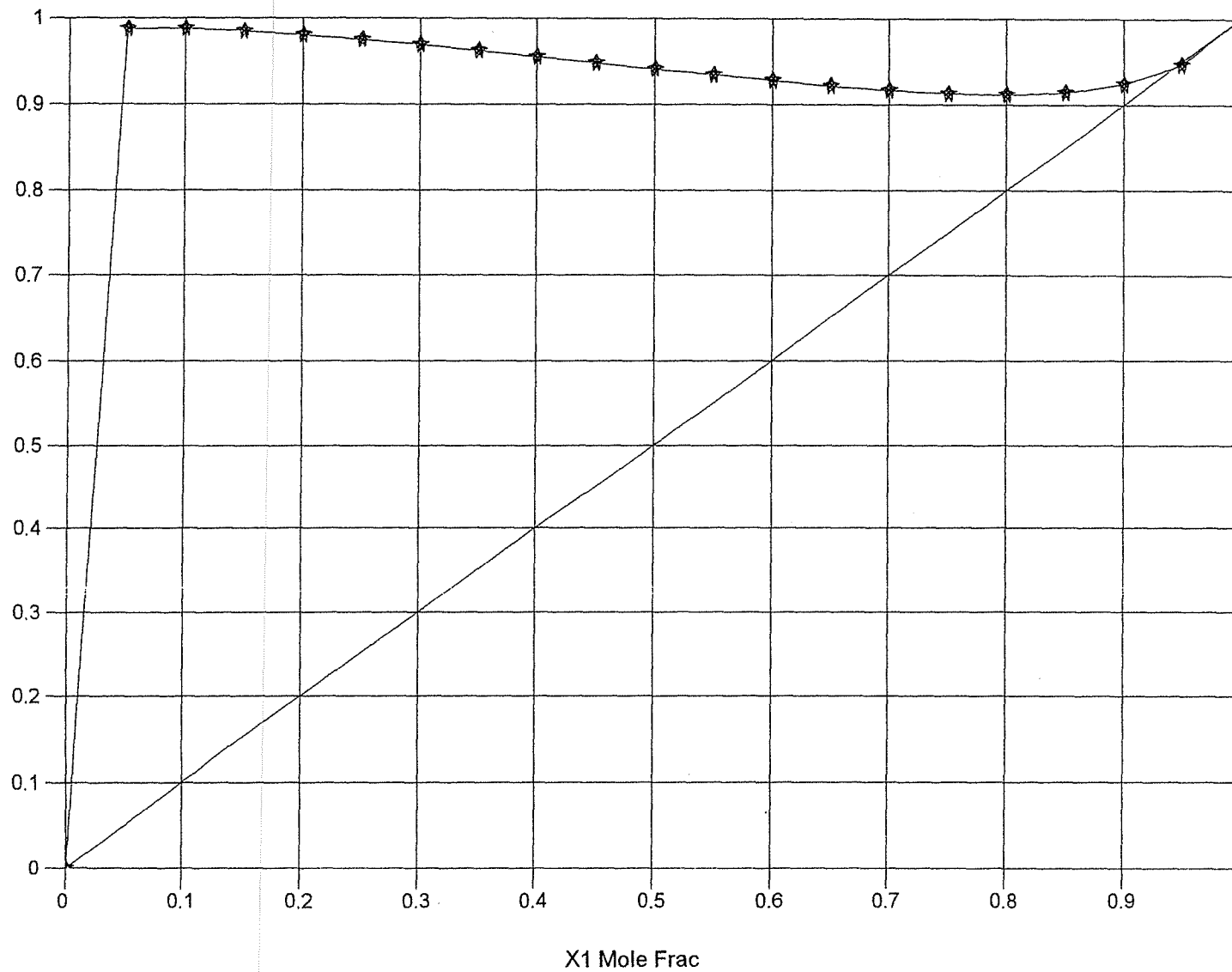


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:45:12

★ XY Data



# Diethyl Ether / Water at 101.32 kPa By UNIQ



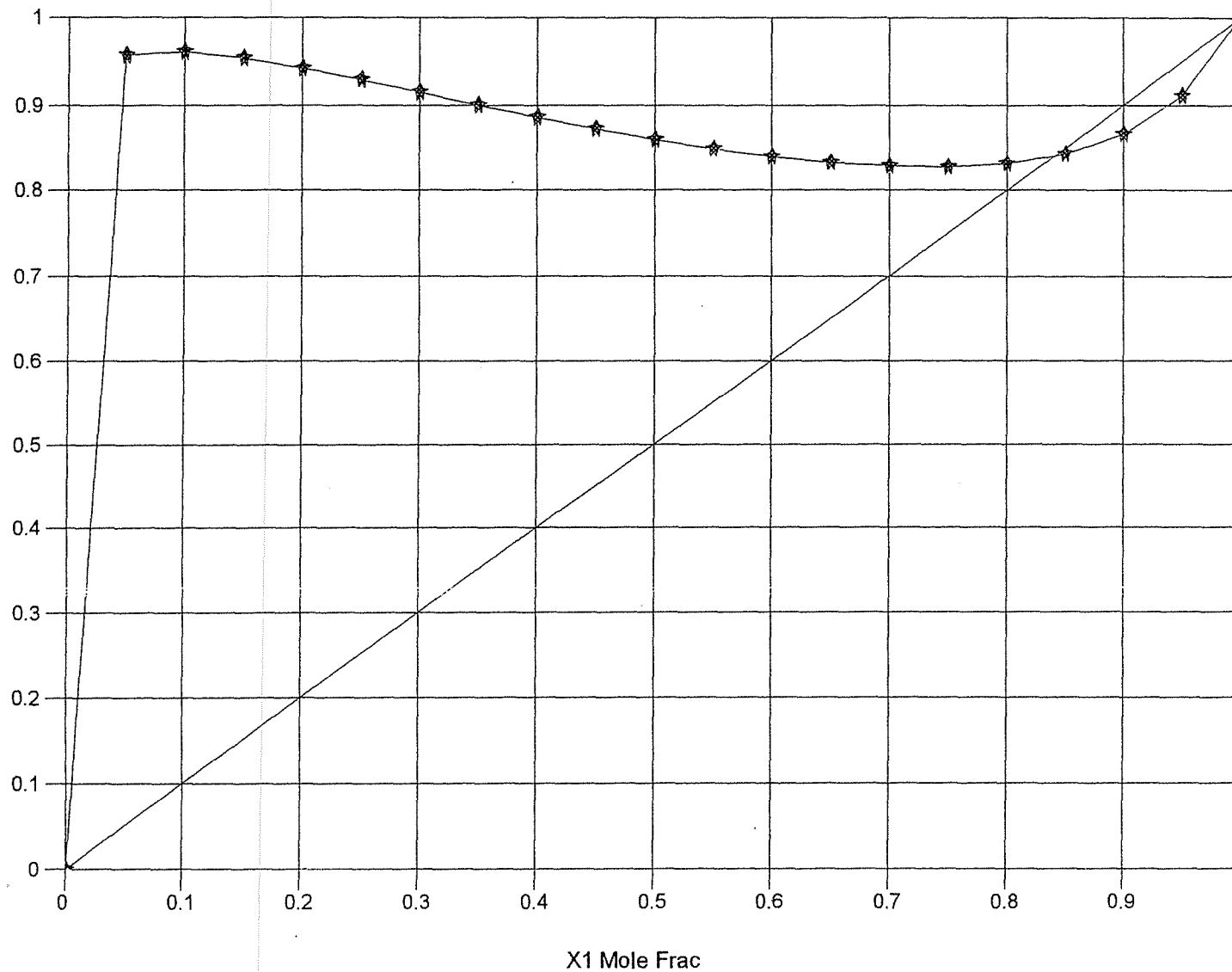
Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:49:02

★ XY Data

# Diethyl Ether / Water at 500.00 kPa By UNIQU

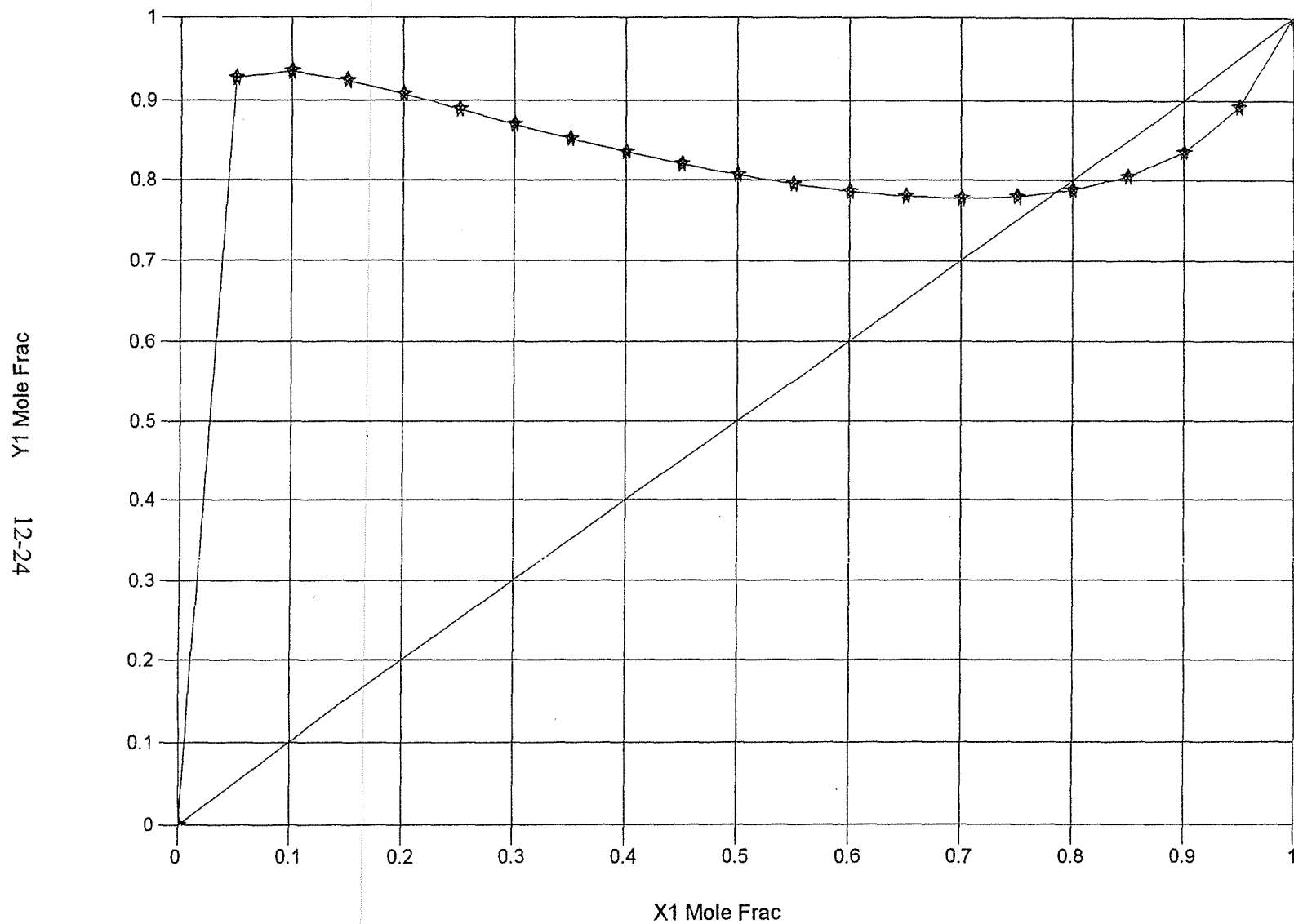
Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:44:36

Y1 Mole Frac  
12-23



★ XY Data

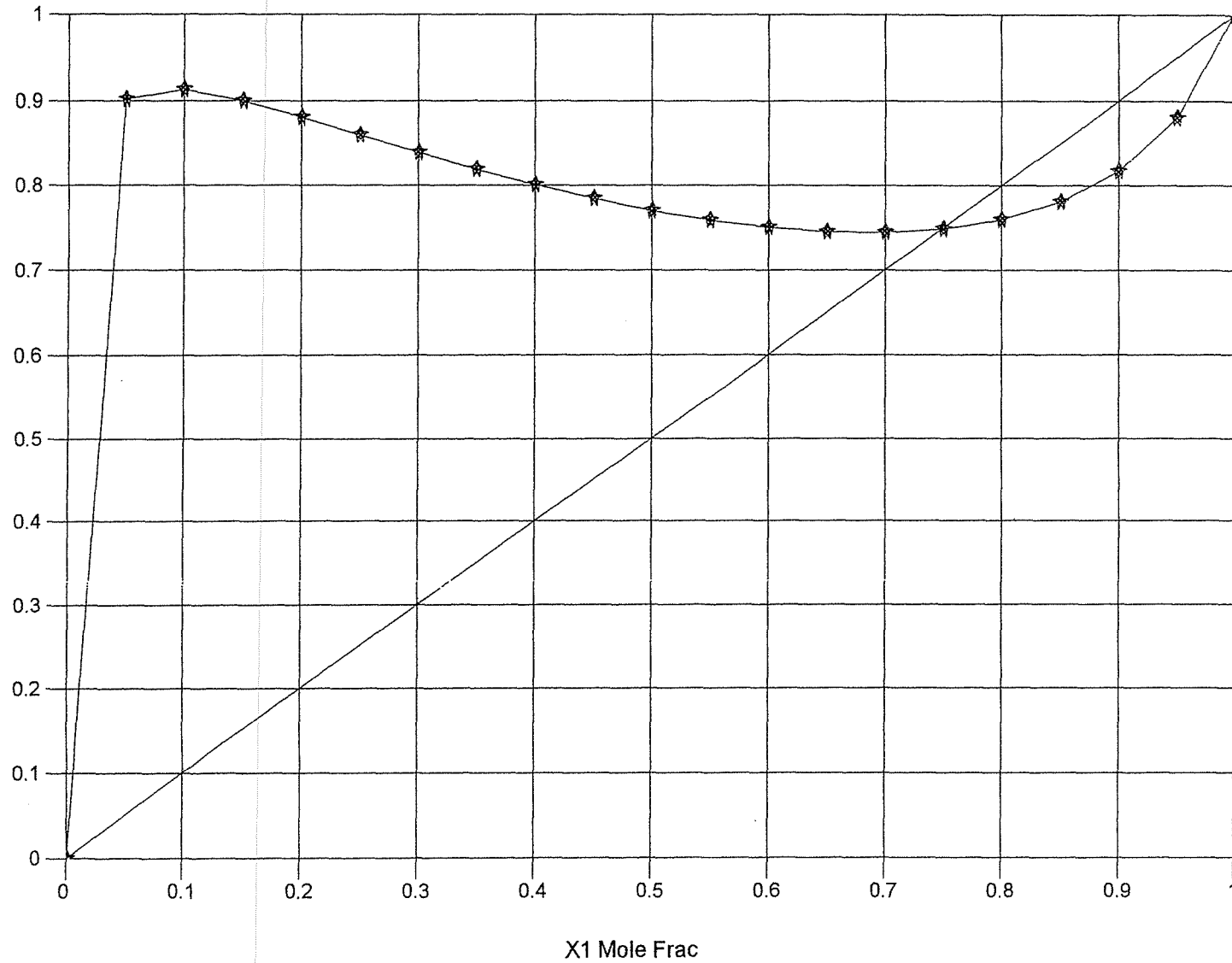
# Diethyl Ether / Water at 1000.00 kPa By UNIQU



Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:44:16

# Diethyl Ether / Water at 1500.00 kPa By UNIQ

52-25 Y1 Mole Frac

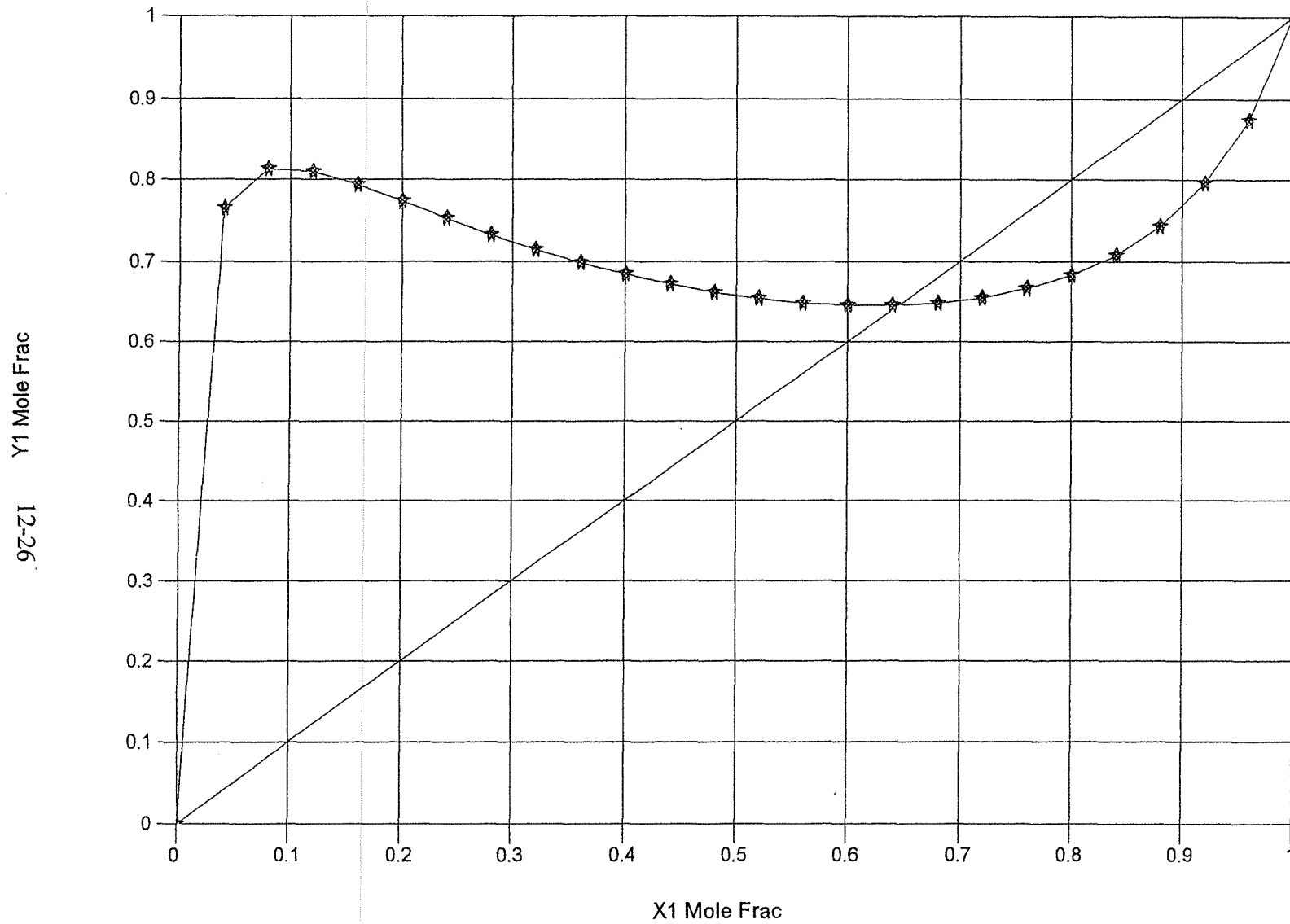


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:43:53

★ XY Data

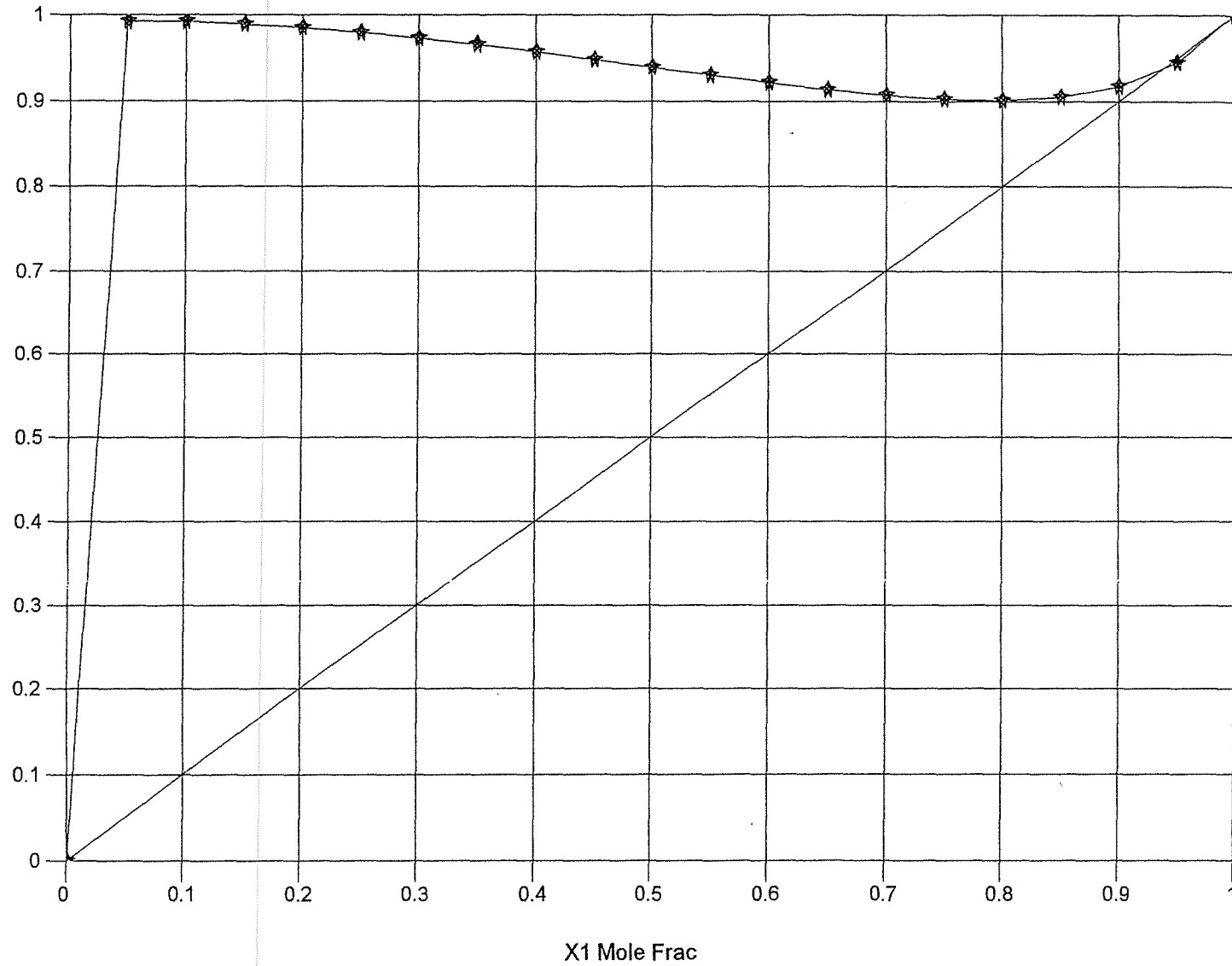
# Diethyl Ether / Water at 5000.00 kPa By UNIQ

Job: ethers  
Date:  
09/12/2002  
Time: 14:10:01



★ XY Data

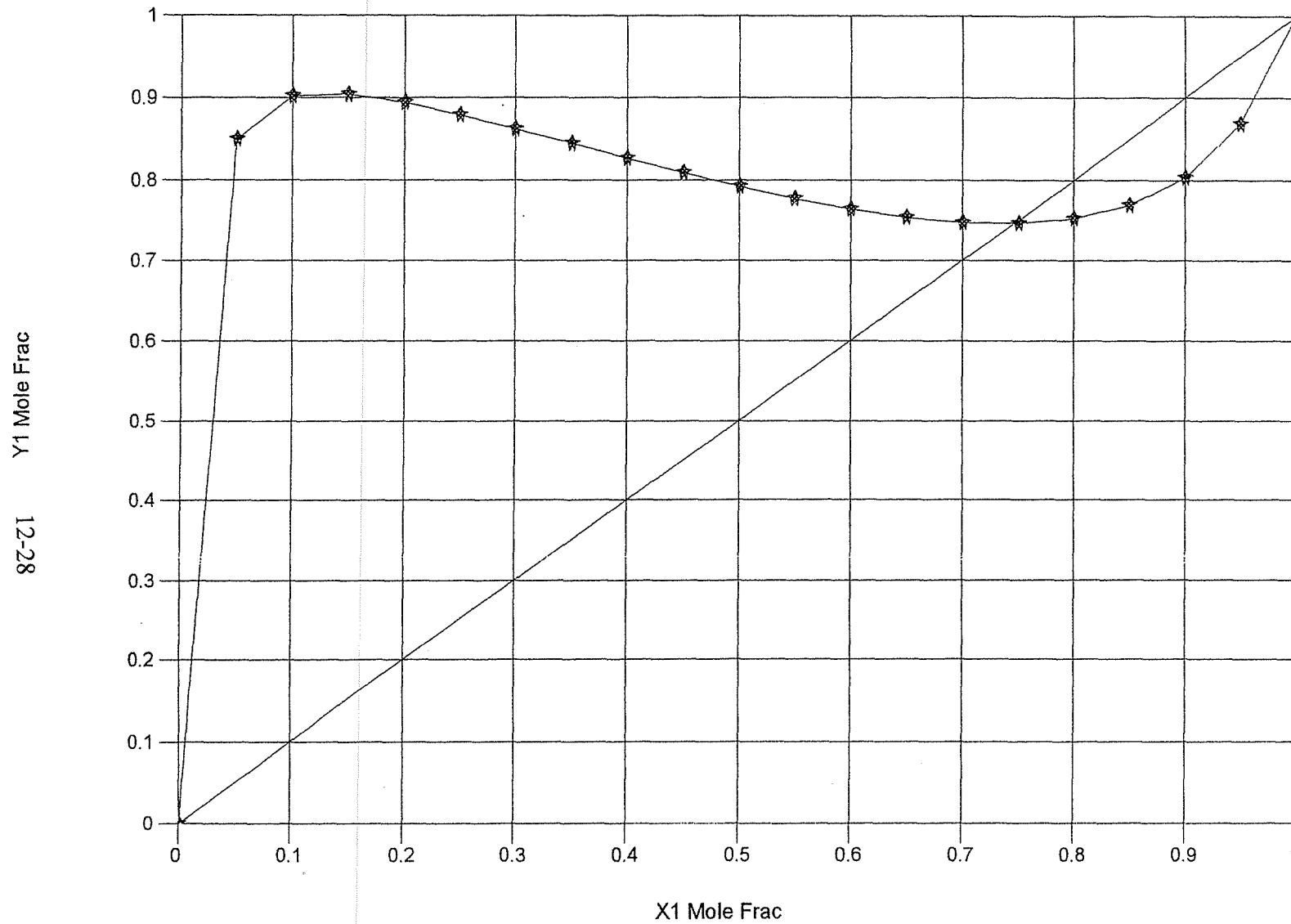
# Diethyl Ether / Water at 101.32 kPa By NRTL



Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:47:59

★ XY Data

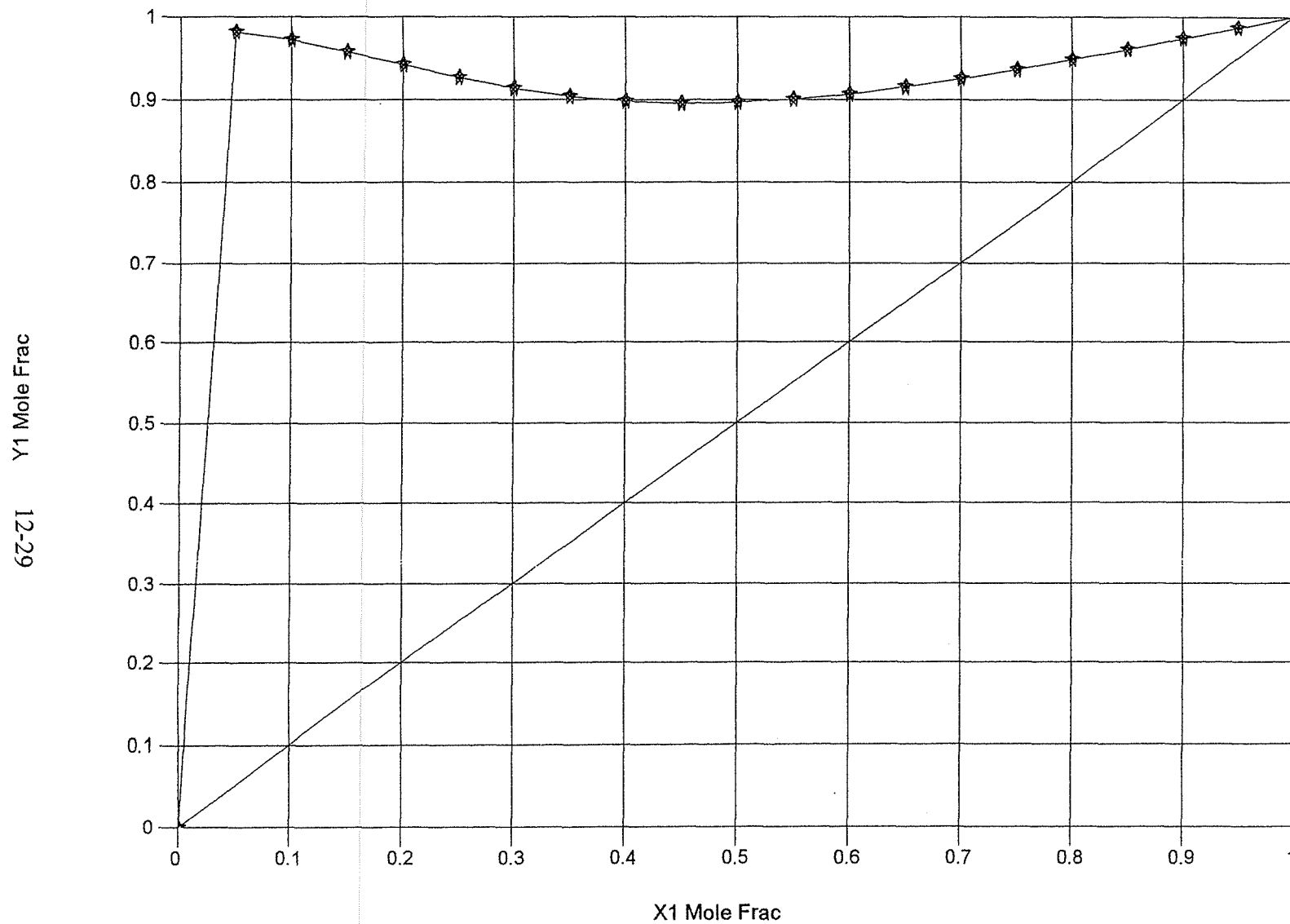
# Diethyl Ether / Water at 1500.00 kPa By NRTL



Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:49:46

★ XY Data

# Diethyl Ether / Water at 101.32 kPa By UNIF

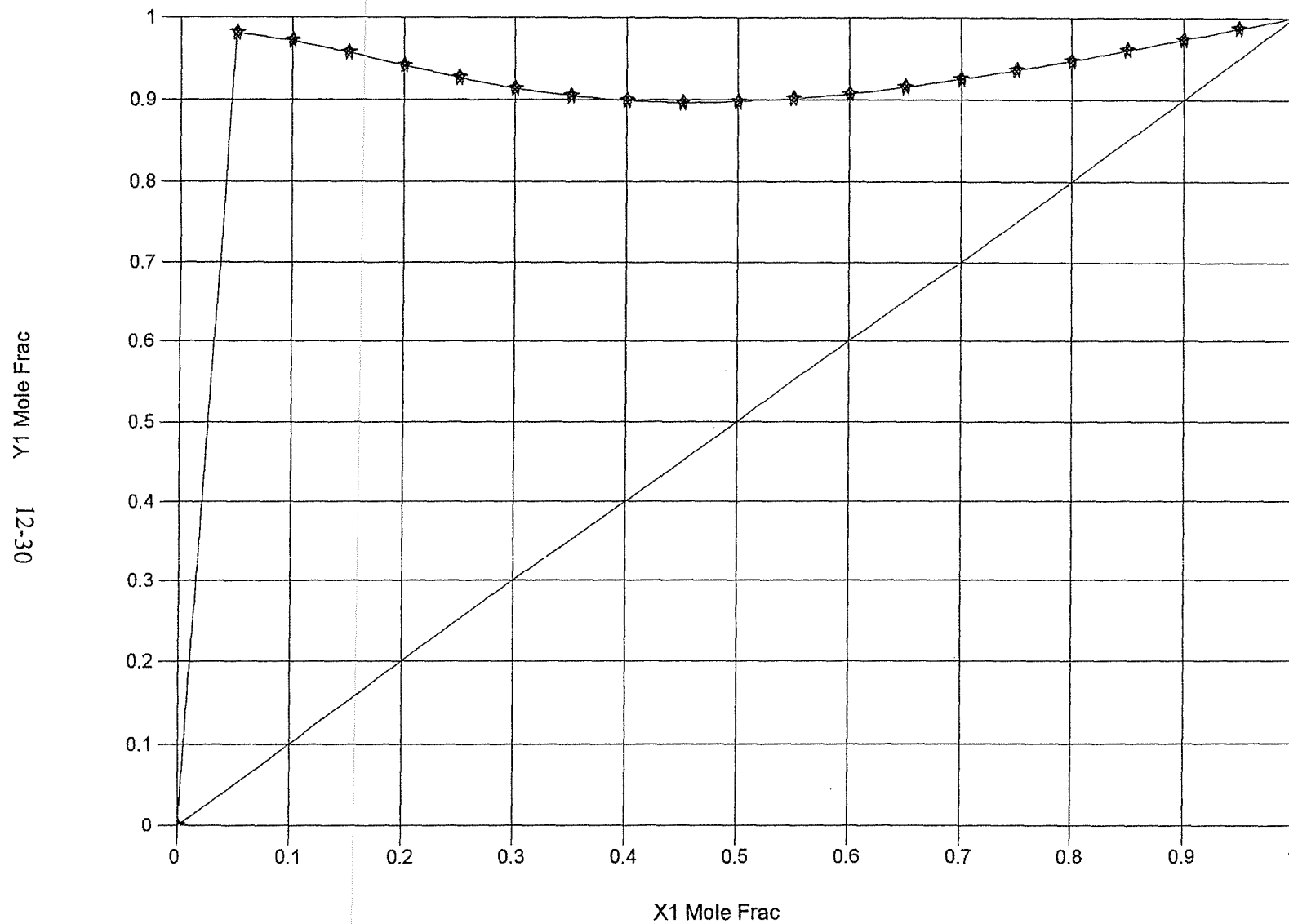


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:48:27

★ XY Data



# Diethyl Ether / Water at 100.00 kPa By UNIF

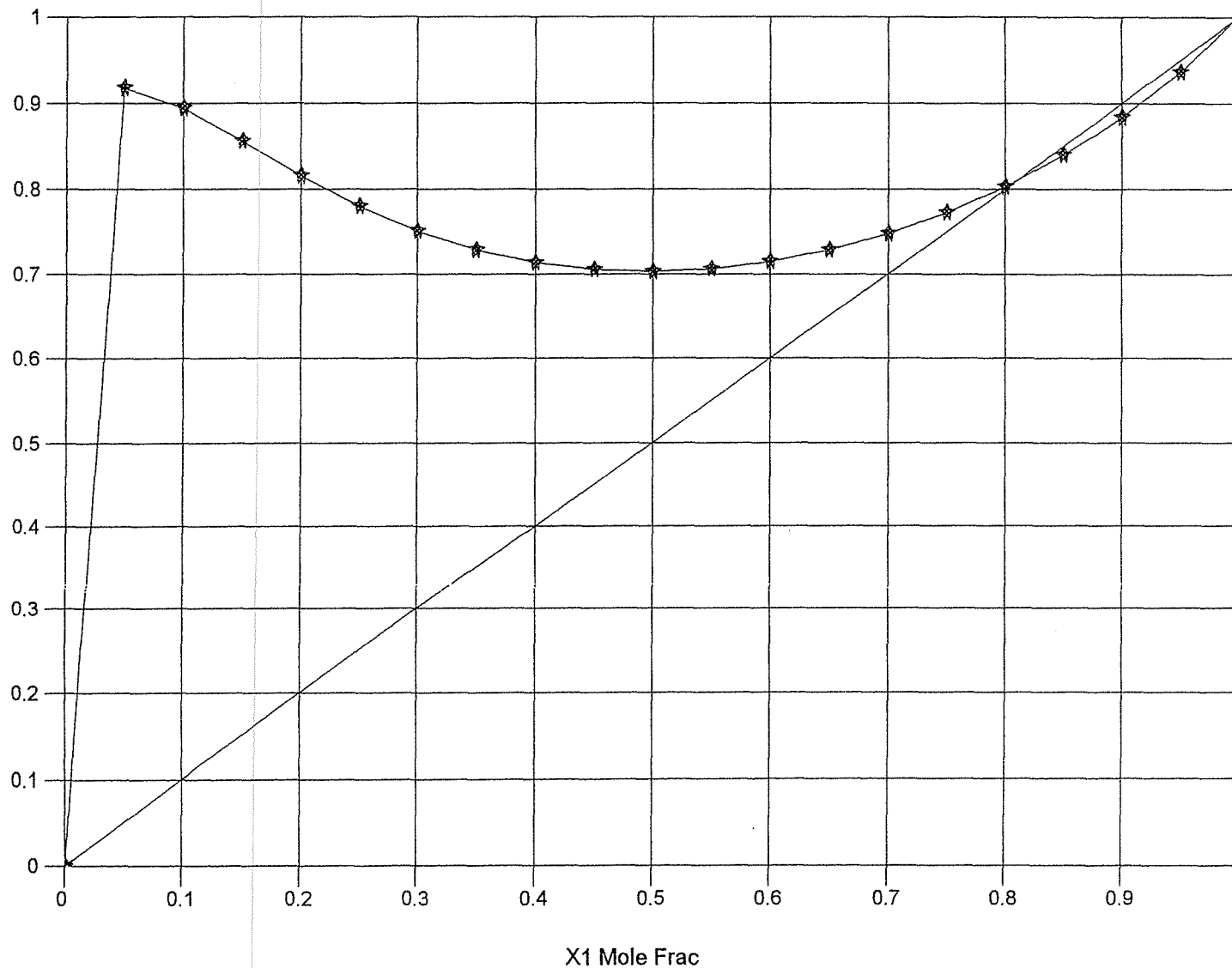


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:48:58

★ XY Data

# Diethyl Ether / Water at 1500.00 kPa By UNIF

Y1 Mole Frac  
12-31

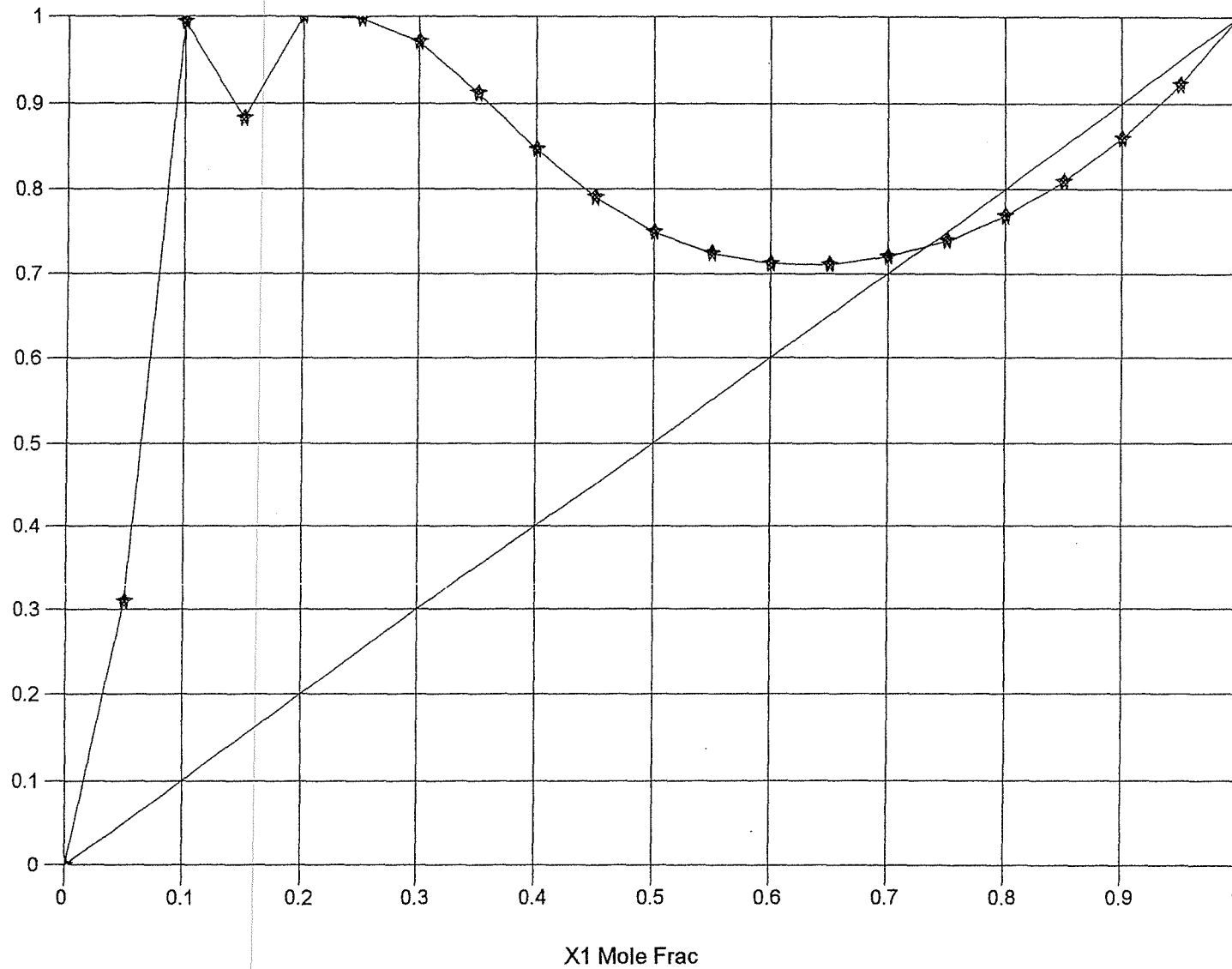


Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:48:13

★ XY Data

# Diethyl Ether / Water at 101.32 kPa By SRK

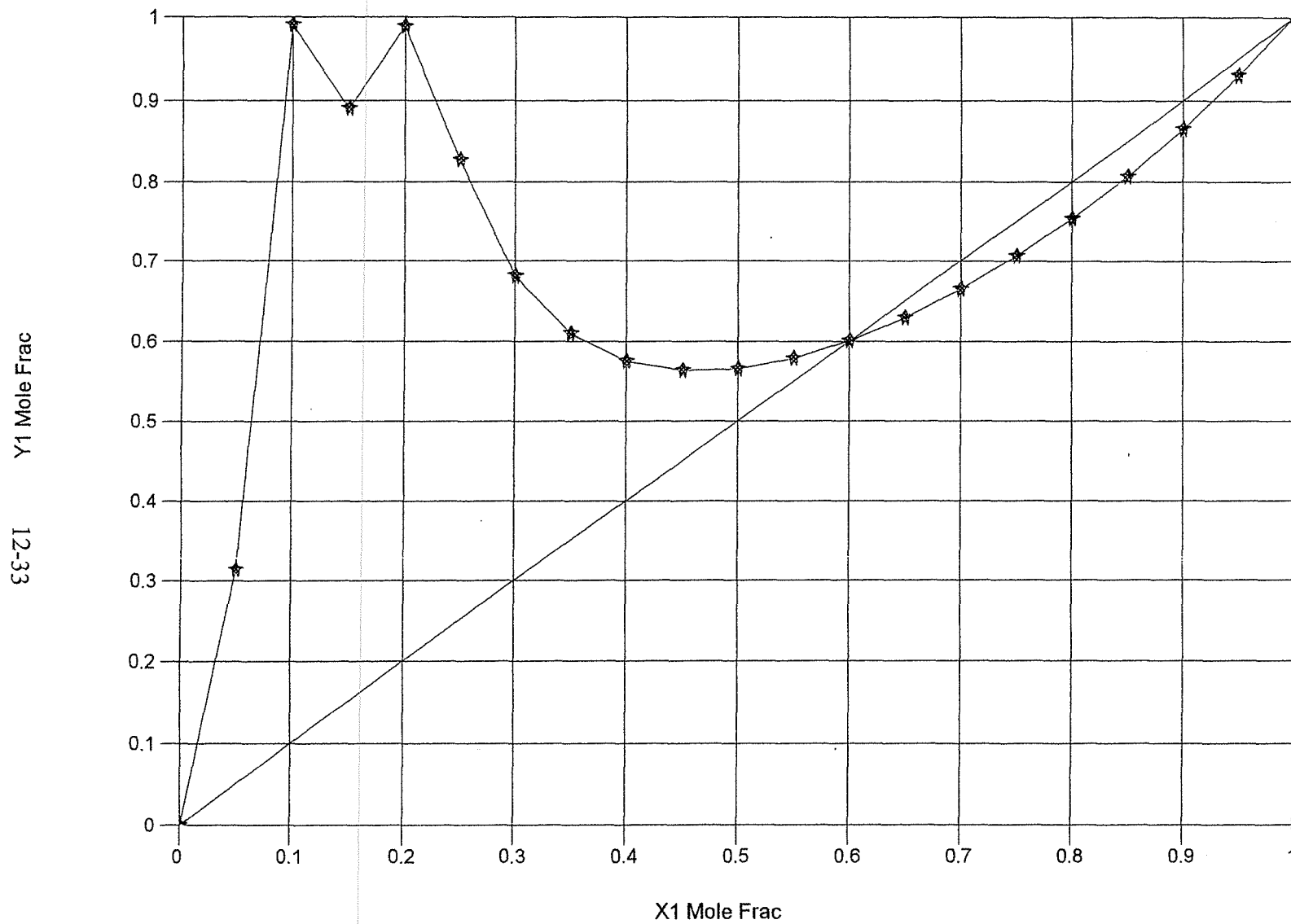
Y1 Mole Frac  
12-32



★ XY Data

Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:47:25

# Diethyl Ether / Water at 1500.00 kPa By SRK



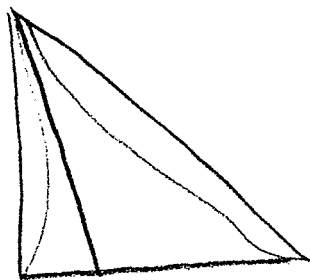
Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 12:47:36

★ XY Data

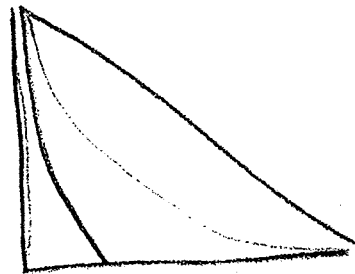
12.16

- plot from Chemcad on following page
- from plot - unclear if boundary is straight or curved

straight



curved

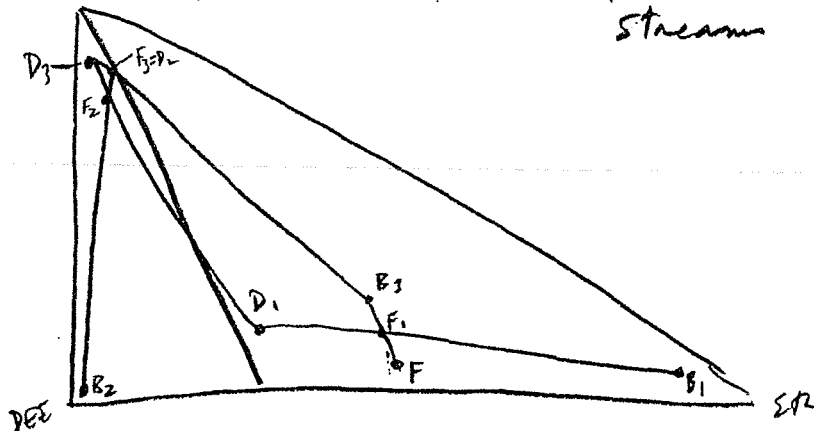


the solution given will work for both cases

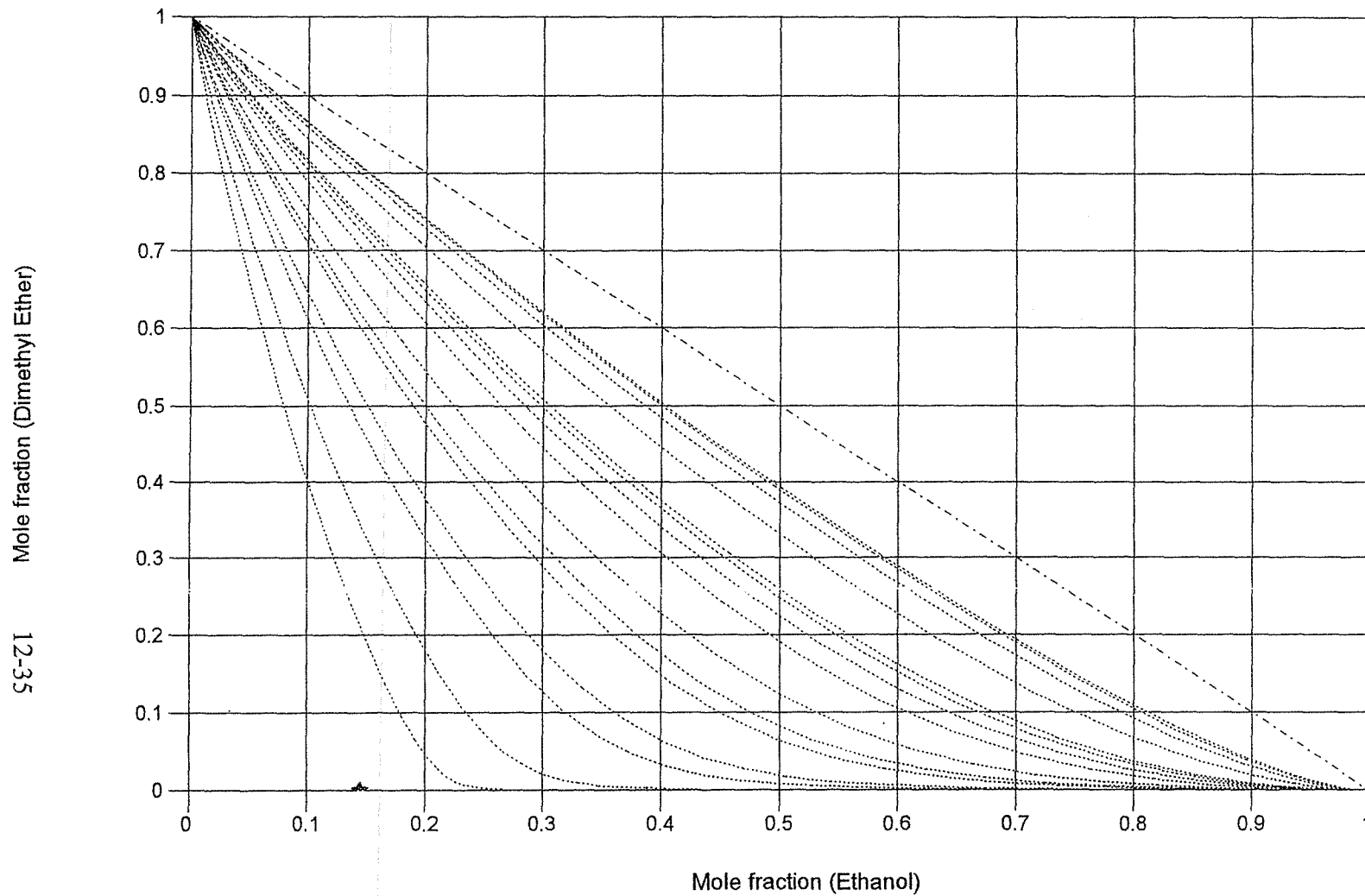
the arrangement is similar to Fig E10.3

where  $B_1$  - ethanol  
 $B_2$  - DEE  
 $D_3$  - DME

note: since mixing is needed  
 cannot produce all 3  
 products in outlet  
 stream



# Residue Curve Map, Ethanol / Dimethyl Ether / Diethyl Ether by UNIQ



Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:05:51

Bubble point at 1500.00 kPa  
Ethanol = 167.697 C  
Dimethyl Ether = 61.416 C  
Diethyl Ether = 140.555 C

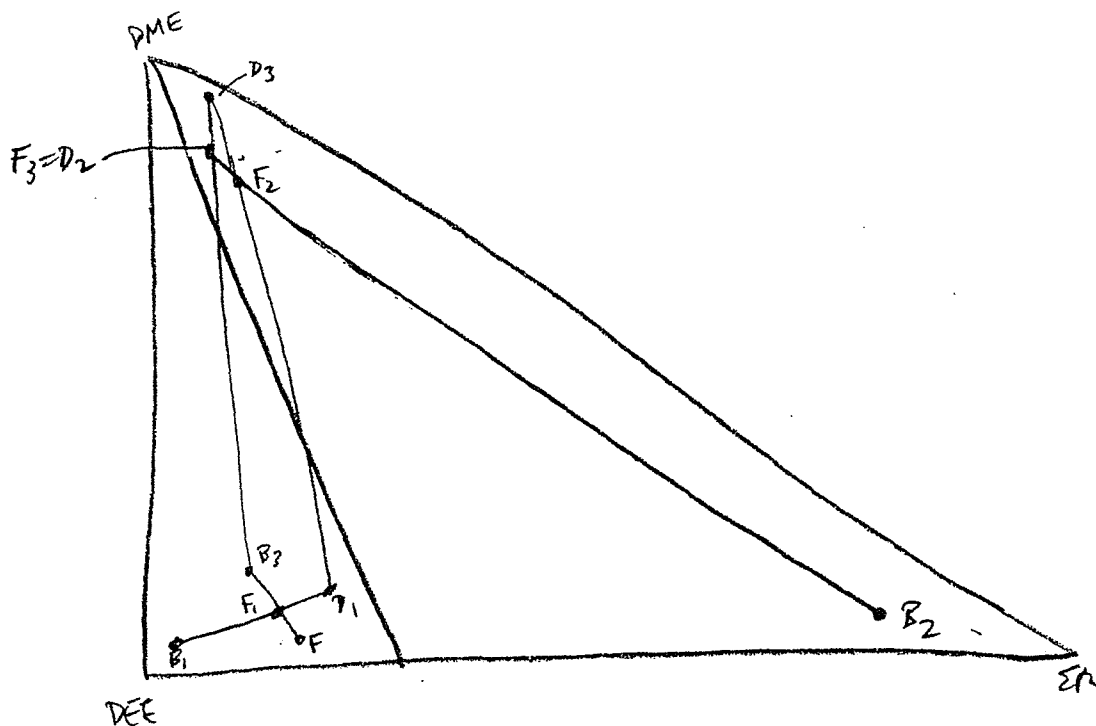
Binary Azeotrope  
(Ethanol,Diethyl Ether) = (0.145,0.855) at 139.331 C

c) Similar arrangement to Fig E10.3 but residue curve maps is different

$B_1 = \text{DEE}$

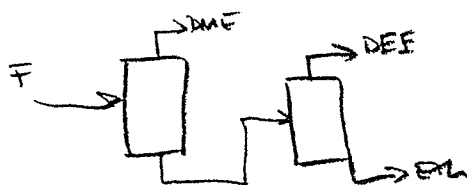
$B_2 = \text{ethanol}$

$D_3 = \text{DME}$



d) if reduce  $P$  to 500 kPa - no azeotrope  
(There is one at 1000 kPa - will disappear between 500 and 1000 kPa)

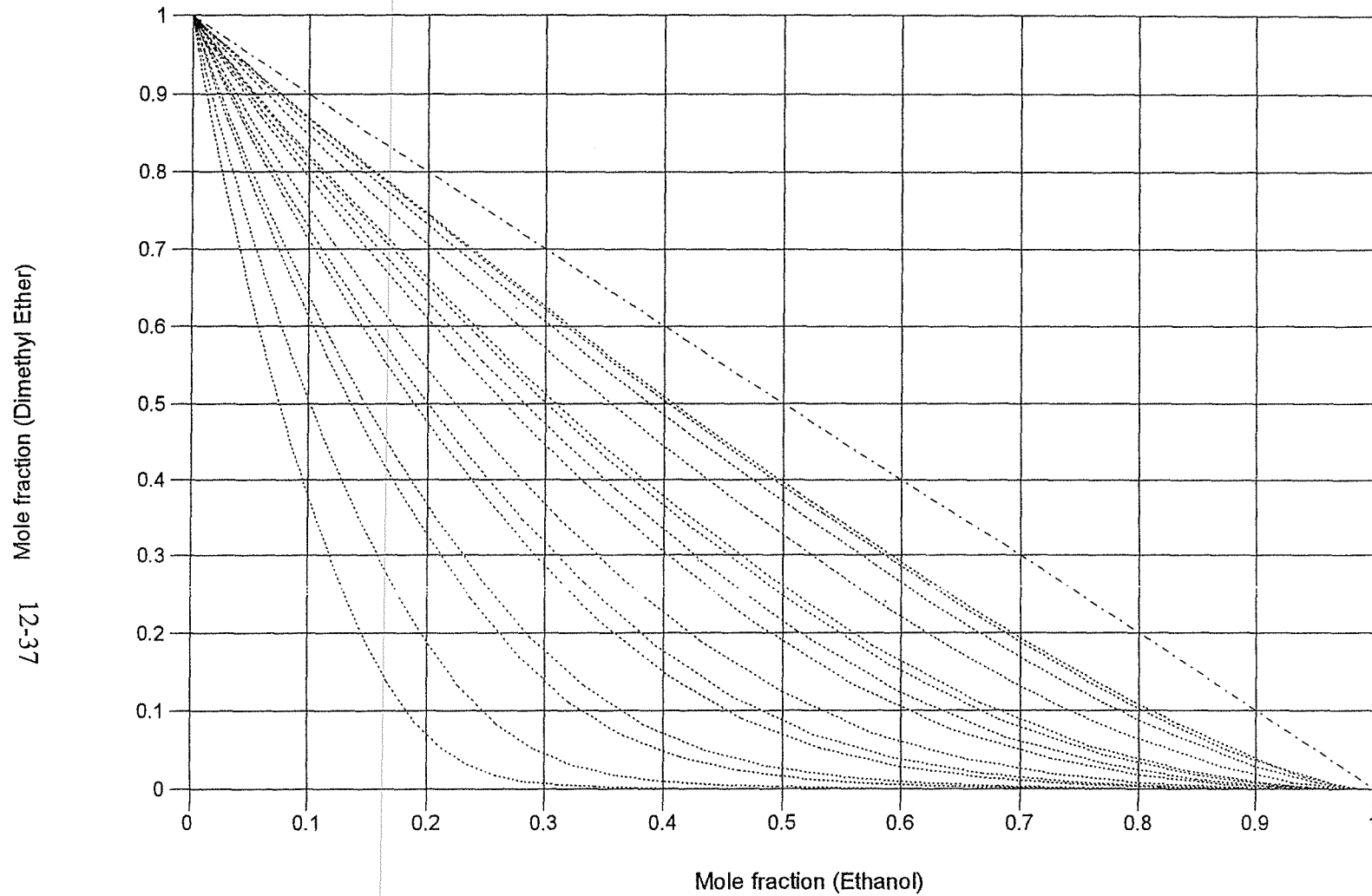
then can do simple separation w/o fear of azeotropes



for example

# Residue Curve Map, Ethanol / Dimethyl Ether / Diethyl Ether by UNIQ

Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:09:09



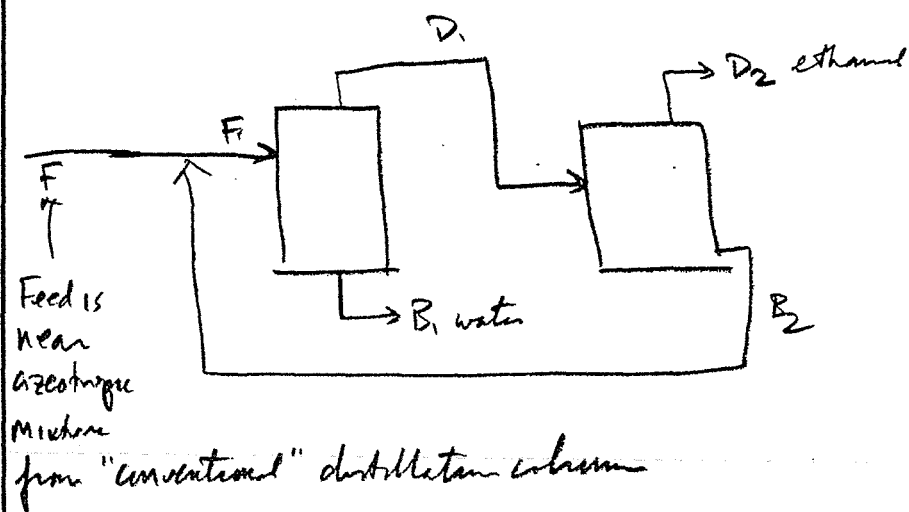
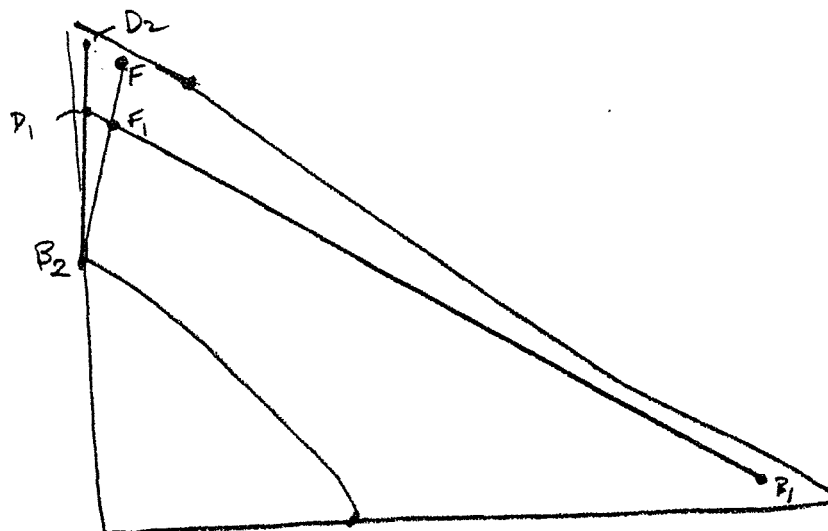
Bubble point at 500.00 kPa  
Ethanol = 125.342 C  
Dimethyl Ether = 19.231 C  
Diethyl Ether = 88.807 C

Binary Azeotrope

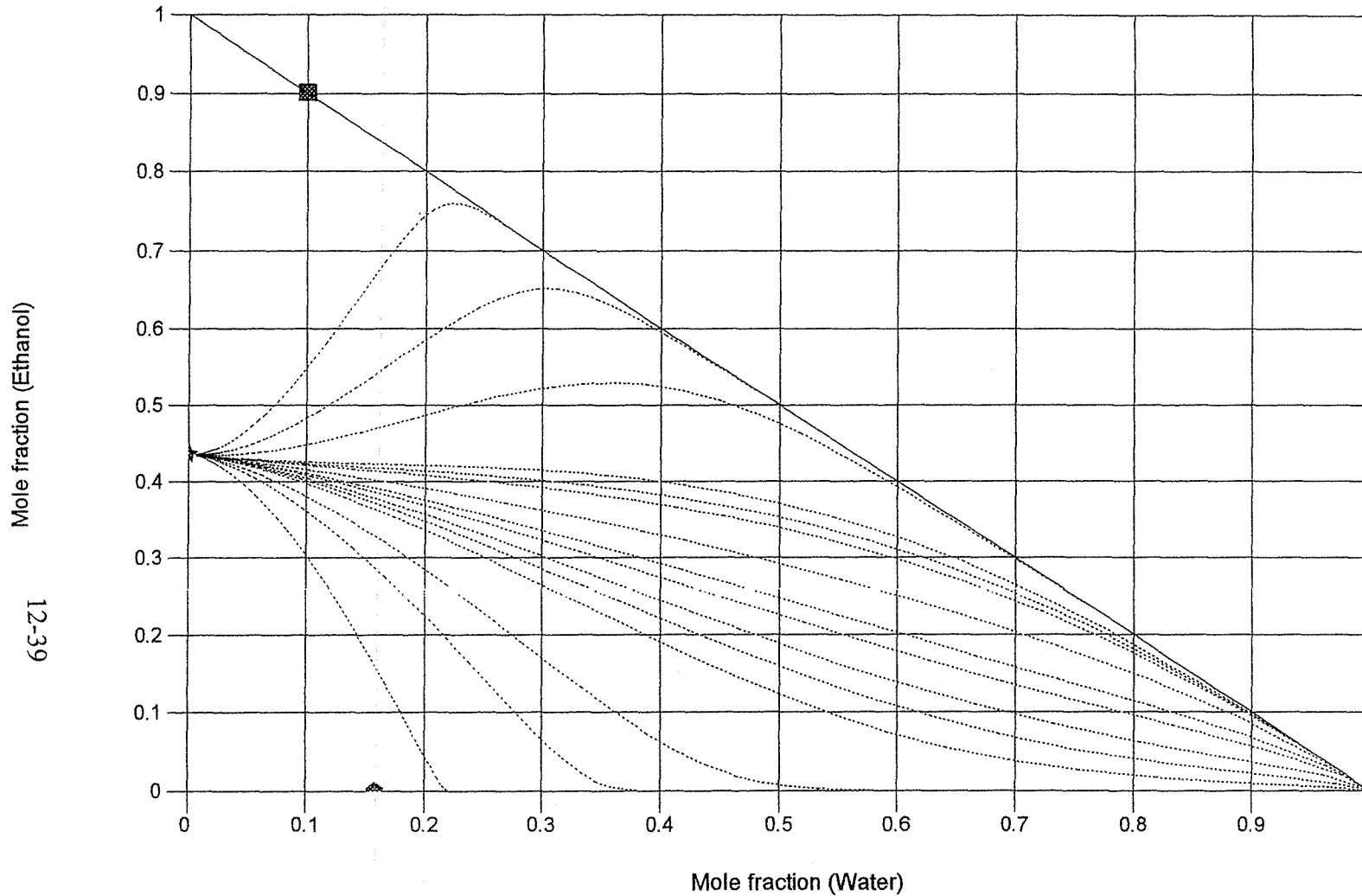


12.17

residue curve map from Chemcad  
on following pages - same conceptual  
result as other processes (1 atm, 5000 kPa)



# Residue Curve Map, Water / Ethanol / Benzene by UNIQU



Job: chapter  
10 binary  
azeotropes  
Date:  
07/12/2001  
Time: 13:28:01

Bubble point at 101.32 kPa  
Water = 100.000 C  
Ethanol = 78.290 C  
Benzene = 80.090 C

Binary Azeotrope  
(Ethanol,Benzene) = (0.436,0.564) at 67.441 C  
(Water,Benzene) = (0.158,0.842) at 79.861 C  
(Water,Ethanol) = (0.100,0.900) at 78.170 C

10.18 Equation 12.1 applies for any “theoretical simple sequence” separation, *i.e.*, where each separation unit makes a perfect split between components. These sequences can be for any separation type, *e.g.*, distillation, filtration, liquid-liquid extraction, leaching, etc. In fact, these separation types can be mixed in the sequence with, for example, one unit being extraction and one being distillation.

10.19 In the derivation of Equation 12.1, it is assumed that each unit is a perfect separator, with any given component only appearing in one of the two outlet streams. The sequence is assumed to contain only such separators, and it is assumed that there is no mixing or splitting of streams between separation units.

10.20 Non-simple separation units include:

- a. Distillation columns with multiple side-draws. Capital costs could go down because the number of columns decreases. Operability and safety may improve because of the reduced number of reboilers, but operability might degrade because of the interconnectedness of a complex column compared with multiple columns with built-in surge capacity (reboiler pots and condenser tanks). Operating costs may decrease through lower heat losses.
- b. Batch columns, in which the overhead is sent to more than one downstream unit on a schedule during the cycle time. These units can decrease capital costs while they increase the complexity of the operation. Operating costs may increase or decrease.
- c. Columns with non-perfect splits, such as the first column in Figure 12.1(c). The advantages of such a column are described on page 400.
- d. A multi-stage filtration column, which has outlets on each stage. Capital costs may decrease because fewer units are required. Operability and safety may be compromised somewhat because the failure of any filter requires the shutdown of the entire column. Operating costs are likely to decrease as solids handling is simplified.

10.21 Equation 12.1 applies to batch and semi-batch processes if and only if each unit is a perfect separator. If a single unit is used to produce three or more outlets, or if the separation is sloppy for any component, the assumptions of the derivation of Equation 12.1 are violated. Note that the multiple “outlets” in the batch process may refer to a single physical outlet port that produces multiple outlet streams scheduled during the cycle time.

10.22 Although this process would result in pure C (in the limit), this is an advantageous strategy only if breaking the A+B azeotrope is not desired. Also, not all of component C is captured in such a process, as C will leave with the azeotropic mixture. The more pure the pot becomes, the more moles of C that are lost. (Note that a batch stripping column with total reflux is not the model for the residue curve map.)

10.23 The composition of a ternary, minimum-boiling azeotrope can be an unstable node or a saddle point. Because compositions close to this ternary azeotrope have higher boiling points, it cannot be a stable node. Whether it is an unstable node or a saddle point depends on the boiling points of the pure components and binary azeotropes.

10.24 Typical heuristics include:

- a. If limited kinetics modeling and/or kinetics data are available, use a batch reactor at the same conditions as the experiment.
- b. If extrapolation/interpolation risk is too great (often the case in pharmaceutical and other biological production), use a batch reactor.
- c. If particles are being grown from seed, try a semi-batch reactor.
- d. If large throughputs are required, try a continuous reactor.
- e. If average properties of the reactor product are important, but variations in properties between small samples of product are not, try a continuous reactor.

10.25 Biological systems tend to have a finite (often short) lifetime before deactivation through age or mutation. Some biological products are only produced during specific time intervals of the organism's life cycle. Materials can slowly poison biological systems or become substrates for cultures that result in undesirable growth characteristics. Biological products often degrade over time, through growth or oxidation that make them undesirable.

- a. Reaction feed preparation must match reaction input schedule, but the time-dependent property changes noted above make the customary surge strategies (e.g., large tanks) unusable.
- b. Similarly, separator feed preparation must often minimize holding times in surge tanks to maintain quality. Biologicals are often extremely sensitive to small temperature changes.
- c. Biological processes often use semi-batch separators such as adsorption or chromatography because separations often involve removal of trace contaminants from streams or capture of products that are trace in reactor outlets.
- d. When either reactors or separators are unsteady state, the recycle may be unsteady. Another factor to consider is the speed of analysis of the stream, the results of which may be needed before recycle commitment.
- e. Environmental control is often unsteady, even when continuous processes are upstream. But environmental control is essentially unsteady when the upstream process is unsteady.

- 10.26 Azeotropic mixtures can be used as solvents to simplify recycle of the solvent from overhead streams (U.S. patent 6,551,978), as refrigerants (U.S. patent 6,221,273), as solvents in other processes such as printing (U.S. patent 6,682,877), and in art restoration (Augerson, C., "The Use of Azeotropes in Treating the Elaborate Polychromy of the Sleigh "The Turtle" (1832), in the Coach Museum, Chateau of Versailles," Wooden Artifacts Group session, 2004 Annual Meeting of the American Institute of Conservation, Portland, Oregon). In each case, the similarity of the boiling/condensing characteristics of azeotropes and pure substances is key to the application. Azeotropes can be produced similarly to the way "pure" components are produced, *i.e.*, they will be the limiting composition in a simple distillation column. A minimum-boiling azeotrope would exit at the top of a column; a maximum-boiling azeotrope would exit at the bottom of a column.
- 10.27 In biological reactor systems, small temperature differences between the reactor and the heating/cooling surfaces are especially important. Large temperature differences can lead to uneven temperature distributions in the reactor, as well as slow response to temperature excursions. Although a large temperature difference can lead to lower costs for the reactor, the extreme temperature sensitivity of most biological systems usually dominates.
- 10.28 Continuous sterilization is often used for process equipment when sufficient cell lysis throughout the process equipment surfaces can be assured. However, batch sterilization is usually preferred, especially for small equipment and for reactor feedstock, where even a very small number of surviving organisms can lead to large cultures and gross contamination.

## Chapter 13

13-1

For all of the problems given in this chapter, the results of computer simulations using the CHEMCAD™ III (Chem Stations, Houston, TX) simulation package are given. Small differences between the data given in Tables 1.5, 1.7, and Appendix B and the results of these simulations are not uncommon. Any significant differences will be noted in the solutions and reasons for the differences will be given. When using other simulators results will differ from those given here. However, these differences should be small. If you cannot get similar answers to those given here be sure to check (1) the thermodynamics package used, (2) the topology of the flowsheet, (3) the equipment specifications. Hint: remove non-condensables from feed when simulating short-cut columns - otherwise you get very low temperatures at the top of these columns. In reality the non-condensable gases would be vented from the overhead reflux drum - but this cannot be simulated using the short-cut methods.

### Toluene Hydrodealkylation Process

The CHEMCAD™ simulation results are given on the following pages. Comparison of the stream compositions shows excellent agreement with the conditions given in Table 1.5 of the text. Differences of 1 or 2°C in temperature and 0.2 kmol/h in flowrates are not unusual. Note that the duties of the heat exchangers from the simulation are all within a few percent, the largest discrepancy being E-105 which is nearly 5% high (in the simulation). The reactor volume is not calculated unless specific kinetics are used (see Problem 13.5). The results for tower, T-101 give 24.5 theoretical stages or  $24.5/0.6 = 40.8$  actual trays - Table 1.7 gives 42 trays. The power of the feed pump, P-101, and the recycle compressor, C-101, are both within a few % of the values in Table 1.7.

|       | <u>Table 1.7</u> | <u>Simulation (following pages)</u> | <u>% diff</u> |
|-------|------------------|-------------------------------------|---------------|
| E-101 | 15,190 MJ/h      | 15,047 MJ/h                         | 0.9           |
| E-102 | 46,660 MJ/h      | 46,650 MJ/h                         | 0.0           |
| E-103 | 1,055 MJ/h       | 1,052 MJ/h                          | 0.3           |
| E-104 | 8,335 MJ/h       | 8,387 MJ/h                          | -0.1          |
| E-105 | 1,085 MJ/h       | 1,138 MJ/h                          | -4.9          |
| E-106 | 9,054 MJ/h       | 9,172 MJ/h                          | -1.3          |
| P-101 | 14.2 kW          | 14.2 kW                             | 0.0           |
| C-101 | 49.1 kW          | 50.2 kW                             | -2.2          |
| T-101 | 42 Trays         | $24.5/0.6 = 40.8$ Trays             | 2.3           |
| H-101 | 27,040 MJ/h      | 27,044 MJ/h                         | 0.0           |

Overall the results of the simulation are all well within the expected accuracy.

Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02

Date: 06-26-97 Time: 22:02

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers |
|-----------|-------|----------------|
| 1         | PUMP  | 91 -2          |
| 2         | HTXR  | 93 -4          |
| 3         | FIRE  | 4 -6           |
| 4         | REAC  | 105 -9         |
| 5         | DIVI  | 8 -101 -19     |
| 6         | FLAS  | 94 -8 -95      |
| 7         | FLAS  | 96 -17 -18     |
| 8         | HTXR  | 18 -97         |
| 9         | SHOR  | 10 -100 -11    |
| 10        | MIXE  | 11 1 -91       |
| 11        | MIXE  | 5 3 -92        |
| 12        | COMP  | 101 -102       |
| 13        | DIVI  | 102 -7 -5      |
| 14        | MIXE  | 7 6 -105       |
| 15        | VALV  | 95 -96         |
| 16        | HTXR  | 9 -94          |
| 17        | MIXE  | 17 21 -99      |
| 18        | VALV  | 19 -21         |
| 19        | MIXE  | 92 2 -93       |
| 20        | CSEP  | 97 -98 -10     |
| 21        | HTXR  | 100 -15        |
| 22        | MIXE  | 98 99 -16      |

## Stream Connections

| Stream | Equipment |    | Stream | Equipment |    | Stream | Equipme |
|--------|-----------|----|--------|-----------|----|--------|---------|
|        | From      | To |        | From      | To |        | From    |
| 1      |           | 10 | 11     | 9         | 10 | 94     | 16      |
| 2      | 1         | 19 | 15     | 21        |    | 95     | 6       |
| 3      |           | 11 | 16     | 22        |    | 96     | 15      |
| 4      | 2         | 3  | 17     | 7         | 17 | 97     | 8       |
| 5      | 13        | 11 | 18     | 7         | 8  | 98     | 20      |
| 6      | 3         | 14 | 19     | 5         | 18 | 99     | 17      |
| 7      | 13        | 14 | 21     | 18        | 17 | 100    | 9       |
| 8      | 6         | 5  | 91     | 10        | 1  | 101    | 5       |
| 9      | 4         | 16 | 92     | 11        | 19 | 102    | 12      |
| 10     | 20        | 9  | 93     | 19        | 2  | 105    | 14      |

Calculation mode : Sequential  
Flash algorithm : Normal

ChemCAD 3.20.16-386 License: WEST VIRGINIA UNIVERSITY Page: 2  
Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 27 2

Equipment Calculation Sequence

|    |    |    |    |   |   |    |   |    |   |    |   |   |   |    |    |    |
|----|----|----|----|---|---|----|---|----|---|----|---|---|---|----|----|----|
| 10 | 1  | 11 | 19 | 2 | 3 | 14 | 4 | 16 | 6 | 15 | 7 | 5 | 8 | 20 | 12 | 13 |
| 18 | 21 | 17 | 22 |   |   |    |   |    |   |    |   |   |   |    |    |    |

Equipment Recycle Sequence

|    |   |    |    |   |   |    |   |    |   |    |   |   |   |    |    |    |
|----|---|----|----|---|---|----|---|----|---|----|---|---|---|----|----|----|
| 10 | 1 | 11 | 19 | 2 | 3 | 14 | 4 | 16 | 6 | 15 | 7 | 5 | 8 | 20 | 12 | 13 |
|----|---|----|----|---|---|----|---|----|---|----|---|---|---|----|----|----|

Recycle Cut Streams

|    |   |   |
|----|---|---|
| 11 | 5 | 7 |
|----|---|---|

Recycle Convergence Method: Direct Substitution

Max. loop iterations 40

Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-004 |
| Temperature | 1.000E-004 |
| Pressure    | 1.000E-004 |
| Enthalpy    | 1.000E-004 |
| Vapor frac. | 1.000E-004 |

Recycle calculation has converged.

COMPONENTS

|   | ID # | Name     |
|---|------|----------|
| 1 | 1    | Hydrogen |
| 2 | 2    | Methane  |
| 3 | 40   | Benzene  |
| 4 | 41   | Toluene  |

THERMODYNAMICS

|                |   |     |
|----------------|---|-----|
| K-value model  | : | SRK |
| Enthalpy model | : | SRK |
| Liquid density | : | API |



Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02

| Overall Mass Balance | kmol/h  |         | kg/h      |           |
|----------------------|---------|---------|-----------|-----------|
|                      | Input   | Output  | Input     | Output    |
| Hydrogen             | 286.000 | 177.964 | 576.519   | 358.739   |
| Methane              | 15.000  | 123.036 | 240.645   | 1973.871  |
| Benzene              | 0.000   | 108.036 | 0.000     | 8439.145  |
| Toluene              | 108.700 | 0.664   | 10015.726 | 61.156    |
| Total                | 409.700 | 409.700 | 10832.890 | 10832.912 |

Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02  
EQUIPMENT SUMMARIES

## Pump Summary

|                     |          |
|---------------------|----------|
| Equip. No.          | 1        |
| Name                |          |
| Output pressure bar | 25.8295  |
| Efficiency          | 0.7500   |
| Calculated power kW | 14.1959  |
| Calculated Pout bar | 25.8295  |
| Head m              | 294.0007 |
| Vol. flow rate m3/h | 16.0060  |

## Heat Exchanger Summary

| Equip. No.          | 2          | 8         | 16          |
|---------------------|------------|-----------|-------------|
| Name                |            |           |             |
| Pressure drop 1 bar | 0.1400     | 0.1400    | 0.1400      |
| T Out Str 1 C       | 225.0000   | 90.0000   | 38.0000     |
| Calc Ht Duty MJ/h   | 15047.3799 | 1052.0774 | -46652.5000 |
| Str1 Pout bar       | 25.3595    | 2.6599    | 23.8600     |

|                     |            |
|---------------------|------------|
| Equip. No.          | 21         |
| Name                |            |
| Pressure drop 1 bar | 0.1400     |
| T Out Str 1 C       | 38.0000    |
| Calc Ht Duty MJ/h   | -1138.0046 |
| Str1 Pout bar       | 2.5199     |

## Fired Heater Summary

|                    |            |
|--------------------|------------|
| Equip. No.         | 3          |
| Name               |            |
| Temperature Out C  | 599.9999   |
| Pressure Drop bar  | 0.3400     |
| Thermal Efficiency | 0.9000     |
| Heat Absorbed MJ/h | 27044.5293 |
| Fuel Usage(SCF)    | 31645.8867 |

13-5

Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02  
EQUIPMENT SUMMARIES

## Reactor Summary

|                               |             |
|-------------------------------|-------------|
| Equip. No.                    | 4           |
| Name                          |             |
| Specify thermal mode:         | 1           |
| C                             | 653.7601    |
| Key Component                 | 4           |
| Frac. Conversion              | 0.7500      |
| Heat of Reaction<br>(kJ/kmol) | -41930.0000 |
| Reactor Pressure bar          | 24.0000     |
| Calc H of Reac.<br>(kJ/kmol)  | -41930.0000 |

## Stoichiometrics:

|          |        |
|----------|--------|
| Hydrogen | -1.000 |
| Methane  | 1.000  |
| Benzene  | 1.000  |
| Toluene  | -1.000 |

## Divider Summary

|                  |        |        |
|------------------|--------|--------|
| Equip. No.       | 5      | 13     |
| Name             |        |        |
| Flow rate units  | 1      | 0      |
| Output stream #1 | 0.7280 | 0.0531 |
| Output stream #2 | 0.2720 | 0.9469 |

## Flash Summary

|                |            |          |
|----------------|------------|----------|
| Equip. No.     | 6          | 7        |
| Name           |            |          |
| Flash Mode     | 2          | 0        |
| Param 1        | 38.0000    |          |
| Param 2        | 23.9000    |          |
| Heat duty MJ/h | -0.5951    |          |
| K values:      |            |          |
| Hydrogen       | 130.395    | 1113.123 |
| Methane        | 15.155     | 123.718  |
| Benzene        | 0.012      | 0.086    |
| Toluene        | 4.020E-003 | 0.027    |

Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02  
EQUIPMENT SUMMARIES

## Shortcut Distillation Summary

|                            |            |
|----------------------------|------------|
| Equip. No.                 | 9          |
| Name                       |            |
| Select mode:               | 2          |
| Column pressure drop (bar) | 0.1700     |
| Light key component        | 3.0000     |
| Light key split            | 0.9900     |
| Heavy key component        | 4.0000     |
| Heavy key split            | 0.0100     |
| Number of stages           | 24.5070    |
| R/Rmin                     | 1.5000     |
| Lower bound R/Rmin         | 1.0100     |
| Upper bound R/Rmin         | 2.0100     |
| Number of stages, minimum  | 13.4679    |
| Feed stage                 | 12.7535    |
| Condenser duty MJ/h        | -8386.7939 |
| Reboiler duty MJ/h         | 9171.5449  |
| Reflux ratio, minimum      | 1.1620     |
| Reflux ratio, calculated   | 1.7430     |

## Mixer Summary

|                     |    |    |         |
|---------------------|----|----|---------|
| Equip. No.          | 10 | 11 | 14      |
| Name                |    |    |         |
| Output Pressure bar |    |    | 25.0000 |
| Equip. No.          | 17 | 19 | 22      |
| Name                |    |    |         |

## Compressor Summary

---

|                      |         |
|----------------------|---------|
| Equip. No.           | 12      |
| Name                 |         |
| Type of Compressor:  | 1       |
| Pressure out bar     | 25.4995 |
| Efficiency           | 0.7500  |
| Actual power kW      | 50.2107 |
| Cp/Cv                | 1.3772  |
| Ideal Cp/Cv          | 1.3405  |
| Calc Pout bar        | 25.4995 |
| Theoretical power kW | 37.6580 |

Job Code: CHP18-1  
EQUIPMENT SUMMARIES

Case Code: CHP18-1

Date: 06-26-97 Time: 22:02

## Valve Summary

|                  |        |        |
|------------------|--------|--------|
| Equip. No.       | 15     | 18     |
| Name             |        |        |
| Pressure out bar | 2.7999 | 2.5000 |

## Component Separator Summary

|                 |        |
|-----------------|--------|
| Equip. No.      | 20     |
| Name            |        |
| Component No. 1 | 1.0000 |
| Component No. 2 | 1.0000 |

Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02  
FLOW SUMMARIES

| Stream No.          | 1        | 2        | 3        |
|---------------------|----------|----------|----------|
| Stream Name         |          |          |          |
| Temp C              | 25.0000  | 60.0250  | 25.0000  |
| Pres bar            | 1.9000   | 25.8295  | 25.4995  |
| Enth MJ/h           | 1296.7   | 2576.9   | -1121.8  |
| Vapor mole fraction | 0.00000  | 0.00000  | 1.0000   |
| Total kmol/h        | 108.7000 | 144.3381 | 301.0000 |
| Flowrates in kmol/h |          |          |          |
| Hydrogen            | 0.0000   | 0.0000   | 286.0000 |
| Methane             | 0.0000   | 0.0000   | 15.0000  |
| Benzene             | 0.0000   | 1.0617   | 0.0000   |
| Toluene             | 108.7000 | 143.2763 | 0.0000   |

| Stream No.          | 4         | 5        | 6         |
|---------------------|-----------|----------|-----------|
| Stream Name         |           |          |           |
| Temp C              | 225.0000  | 45.0450  | 599.9999  |
| Pres bar            | 25.3595   | 25.4995  | 25.0195   |
| Enth MJ/h           | -5098.4   | -21601.  | 21946.    |
| Vapor mole fraction | 1.0000    | 1.0000   | 1.0000    |
| Total kmol/h        | 1204.2986 | 758.9607 | 1204.2986 |
| Flowrates in kmol/h |           |          |           |
| Hydrogen            | 735.3435  | 449.3436 | 735.3435  |
| Methane             | 317.1072  | 302.1072 | 317.1072  |
| Benzene             | 7.8405    | 6.7788   | 7.8405    |
| Toluene             | 144.0074  | 0.7310   | 144.0074  |

| Stream No.          | 7       | 8         | 9         |
|---------------------|---------|-----------|-----------|
| Stream Name         |         |           |           |
| Temp C              | 45.0450 | 38.0000   | 653.7601  |
| Pres bar            | 25.4995 | 23.9000   | 24.0000   |
| Enth MJ/h           | -1211.3 | -31584.   | 20735.    |
| Vapor mole fraction | 1.0000  | 1.0000    | 1.0000    |
| Total kmol/h        | 42.5608 | 1100.9911 | 1246.8594 |
| Flowrates in kmol/h |         |           |           |
| Hydrogen            | 25.1982 | 651.8431  | 652.5054  |
| Methane             | 16.9415 | 438.2538  | 442.0850  |
| Benzene             | 0.3801  | 9.8337    | 116.2570  |
| Toluene             | 0.0410  | 1.0605    | 36.0121   |

| Stream No.          | 10       | 11       | 15       |
|---------------------|----------|----------|----------|
| Stream Name         |          |          |          |
| Temp C              | 90.0000  | 149.6958 | 38.0000  |
| Pres bar            | 2.6599   | 2.8299   | 2.5199   |
| Enth MJ/h           | 6969.7   | 1229.1   | 5387.4   |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000  |
| Total kmol/h        | 141.0997 | 35.6381  | 105.4617 |
| Flowrates in kmol/h |          |          |          |
| Hydrogen            | 0.0000   | 0.0000   | 0.0000   |
| Methane             | 0.0000   | 0.0000   | 0.0000   |
| Benzene             | 106.1741 | 1.0617   | 105.1124 |

Job Code: CHP18-1 Case Code: CHP18-1 Date: 06-26-97 Time: 22:02  
FLOW SUMMARIES

|         |         |         |        |
|---------|---------|---------|--------|
| Toluene | 34.9256 | 34.5763 | 0.3493 |
|---------|---------|---------|--------|

|                     |          |         |          |
|---------------------|----------|---------|----------|
| Stream No.          | 16       | 17      | 18       |
| Stream Name         |          |         |          |
| Temp C              | 35.5823  | 38.3049 | 38.3049  |
| Pres bar            | 2.5000   | 2.7999  | 2.7999   |
| Enth MJ/h           | -8851.6  | -197.45 | 5863.2   |
| Vapor mole fraction | 1.0000   | 1.0000  | 0.00000  |
| Total kmol/h        | 304.2383 | 3.8714  | 141.9969 |
| Flowrates in kmol/h |          |         |          |
| Hydrogen            | 177.9636 | 0.6412  | 0.0211   |
| Methane             | 123.0363 | 2.9551  | 0.8761   |
| Benzene             | 2.9239   | 0.2491  | 106.1741 |
| Toluene             | 0.3145   | 0.0260  | 34.9256  |

|                     |          |          |          |
|---------------------|----------|----------|----------|
| Stream No.          | 19       | 21       | 91       |
| Stream Name         |          |          |          |
| Temp C              | 38.0000  | 35.3601  | 59.0413  |
| Pres bar            | 23.9000  | 2.5000   | 1.9000   |
| Enth MJ/h           | -8590.8  | -8590.7  | 2525.8   |
| Vapor mole fraction | 1.0000   | 1.0000   | 0.00000  |
| Total kmol/h        | 299.4696 | 299.4696 | 144.3381 |
| Flowrates in kmol/h |          |          |          |
| Hydrogen            | 177.3013 | 177.3013 | 0.0000   |
| Methane             | 119.2050 | 119.2050 | 0.0000   |
| Benzene             | 2.6748   | 2.6748   | 1.0617   |
| Toluene             | 0.2885   | 0.2885   | 143.2763 |

|                     |           |           |           |
|---------------------|-----------|-----------|-----------|
| Stream No.          | 92        | 93        | 94        |
| Stream Name         |           |           |           |
| Temp C              | 39.6650   | 48.2868   | 38.0000   |
| Pres bar            | 25.4995   | 25.4995   | 23.8600   |
| Enth MJ/h           | -22723.   | -20146.   | -25917.   |
| Vapor mole fraction | 1.0000    | 0.875653  | 0.883029  |
| Total kmol/h        | 1059.9605 | 1204.2986 | 1246.8594 |
| Flowrates in kmol/h |           |           |           |
| Hydrogen            | 735.3436  | 735.3435  | 652.5054  |
| Methane             | 317.1072  | 317.1072  | 442.0850  |
| Benzene             | 6.7788    | 7.8405    | 116.2570  |
| Toluene             | 0.7310    | 144.0074  | 36.0121   |

Job Code: CHP18-1

Case Code: CHP18-1

Date: 06-26-97

Time: 22:02

## FLOW SUMMARIES

| Stream No.          | 95       | 96        | 97         |
|---------------------|----------|-----------|------------|
| Stream Name         |          |           |            |
| Temp C              | 38.0000  | 38.3049   | 90.0000    |
| Pres bar            | 23.9000  | 2.7999    | 2.6599     |
| Enth MJ/h           | 5665.7   | 5665.7    | 6915.3     |
| Vapor mole fraction | 0.00000  | 0.0265406 | 0.00501726 |
| Total kmol/h        | 145.8684 | 145.8684  | 141.9969   |
| Flowrates in kmol/h |          |           |            |
| Hydrogen            | 0.6623   | 0.6623    | 0.0211     |
| Methane             | 3.8312   | 3.8312    | 0.8761     |
| Benzene             | 106.4233 | 106.4233  | 106.1741   |
| Toluene             | 34.9516  | 34.9516   | 34.9256    |

| Stream No.          | 98      | 99       | 100      |
|---------------------|---------|----------|----------|
| Stream Name         |         |          |          |
| Temp C              | 90.0000 | 35.3990  | 115.1665 |
| Pres bar            | 2.6599  | 2.5000   | 2.6599   |
| Enth MJ/h           | -63.423 | -8788.2  | 6525.4   |
| Vapor mole fraction | 1.0000  | 1.0000   | 0.00000  |
| Total kmol/h        | 0.8972  | 303.3410 | 105.4617 |
| Flowrates in kmol/h |         |          |          |
| Hydrogen            | 0.0211  | 177.9425 | 0.0000   |
| Methane             | 0.8761  | 122.1602 | 0.0000   |
| Benzene             | 0.0000  | 2.9239   | 105.1124 |
| Toluene             | 0.0000  | 0.3145   | 0.3493   |

| Stream No.          | 101      | 102      | 105       |
|---------------------|----------|----------|-----------|
| Stream Name         |          |          |           |
| Temp C              | 38.0000  | 45.0450  | 588.5500  |
| Pres bar            | 23.9000  | 25.4995  | 25.0000   |
| Enth MJ/h           | -22993.  | -22812.  | 20735.    |
| Vapor mole fraction | 1.0000   | 1.0000   | 1.0000    |
| Total kmol/h        | 801.5214 | 801.5214 | 1246.8594 |
| Flowrates in kmol/h |          |          |           |
| Hydrogen            | 474.5417 | 474.5417 | 760.5417  |
| Methane             | 319.0487 | 319.0487 | 334.0487  |
| Benzene             | 7.1589   | 7.1589   | 8.2207    |
| Toluene             | 0.7720   | 0.7720   | 144.0484  |



**13.2** Simulation of the DME process, Appendix B, Figure B.1 and Tables B.1 and B.2  
 Again the simulation results are shown in the output following this page. It is worth noting that due to non-ideal behavior of the water/DME/methanol system a Uniquac/Unifac thermodynamic model was chosen for this simulation. Again the reactor was modeled with a simple stoichiometric reactor module. Therefore no reactor volume is calculated in the output. The required minimum input information for this simulation is given below:

1. Topology (see first page of following simulation)
2. Specification of feed streams - stream 1 (meoh)
3. Equipment specifications
  - P-201 output pressure (efficiency not required for H&M balance)
  - E-201 stream outlet temperature
  - E-202 outlet temperature of feed to reactor.
  - R-201 conversion of key component (no kinetics req'd for H&M balance)
  - E-203 stream outlet temperature
  - T-201 identification of key components, recovery of key components,  $R/R_{min}$
  - T-202 identification of key components, recovery of key components,  $R/R_{min}$
  - E-208 stream outlet temperature
4. Conversion tolerance for recycle loop etc.

note: that for the simulation used here, exchanger E-202 was included. However, to ease the recycle calculation, E-202 could be eliminated if desired (and added back when recycle convergence was obtained). Also note that reflux pumps are not included anywhere in the simulation since the towers are modeled with the reboiler and condenser included and no estimates of the reflux pumps are given

| <u>Equipment</u> | <u>Table B.1</u> | <u>Simulation Results</u> | <u>% diff.</u> |
|------------------|------------------|---------------------------|----------------|
| P-201            | 7.2 kW           | 7.1 kW                    | 1.4            |
| E-201            | 14,400 MJ/h      | 14,361 MJ/h               | 0.3            |
| E-202            | 2,030 MJ/h       | 2021 MJ/h                 | 0.4            |
| E-203            | 12,420 MJ/h      | 12,766 MJ/h               | -2.8           |
| E-204            | 2,490 MJ/h       | 2,833 MJ/h                | -13.8          |
| E-205            | 3,140 MJ/h       | 3,105 MJ/h                | 1.1            |
| E-206            | 5,790 MJ/h       | 5,725 MJ/h                | 1.1            |
| E-207            | 5,960 MJ/h       | 5,899 MJ/h                | 1.0            |
| E-208            | 1,200 MJ/h       | 1,190 MJ/h                | 0.8            |
| T-201            | 22 trays         | 15.2/0.7 = 22 trays       | 0.0            |
| T-201            | 26 trays         | 18.1/0.7 = 26 trays       | 0.0            |

Simulation result for E-204 is somewhat higher than in Table B.1 - but condenser duty (E-205) is OK.

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10:19

Date: 06-27-97 Time: 10:19

## FLOWSHEET SUMMARY

## Equipment Label Stream Numbers

|         |       |    |     |     |    |
|---------|-------|----|-----|-----|----|
| 1 PUMP  | P-201 | 1  | -2  |     |    |
| 2 MIXE  |       | 13 | 2   | -3  |    |
| 3 HTXR  | E-201 | 3  | -4  |     |    |
| 4 HTXR  | E-202 | 4  | 6   | -5  | -7 |
| 5 REAC  | R-201 | 5  | -6  |     |    |
| 6 HTXR  | E-203 | 7  | -8  |     |    |
| 7 VALV  |       | 8  | -9  |     |    |
| 8 SHOR  | T-201 | 9  | -10 | -11 |    |
| 9 VALV  |       | 11 | -12 |     |    |
| 10 SHOR | T-202 | 12 | -13 | -14 |    |
| 11 HTXR | E-208 | 14 | -16 |     |    |
| 12 VALV |       | 16 | -15 |     |    |

## Stream Connections

| Stream | Equipment From | Equipment To | Stream | Equipment From | Equipment To | Stream | Equipment From |
|--------|----------------|--------------|--------|----------------|--------------|--------|----------------|
| 1      |                | 1            | 7      | 4              | 6            | 13     | 10             |
| 2      | 1              | 2            | 8      | 6              | 7            | 14     | 10             |
| 3      | 2              | 3            | 9      | 7              | 8            | 15     | 12             |
| 4      | 3              | 4            | 10     | 8              |              | 16     | 11             |
| 5      | 4              | 5            | 11     | 8              | 9            |        |                |
| 6      | 5              | 4            | 12     | 9              | 10           |        |                |

Calculation mode : Sequential

Flash algorithm : Normal

## Equipment Calculation Sequence

1 2 3 4 5 6 7 8 9 10 11 12

## Equipment Recycle Sequence

2 3 4 5 6 7 8 9 10

## Recycle Cut Streams

13 6

Recycle Convergence Method: Direct Substitution

Max. loop iterations 40

13-13

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10:19

## Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-002 |
| Temperature | 1.000E-002 |
| Pressure    | 1.000E-002 |
| Enthalpy    | 1.000E-002 |
| Vapor frac. | 1.000E-002 |

Recycle calculation has converged.

## COMPONENTS

|   | ID # | Name           |
|---|------|----------------|
| 1 | 117  | Methanol       |
| 2 | 133  | Dimethyl Ether |
| 3 | 62   | Water          |

## THERMODYNAMICS

|                |   |                                      |
|----------------|---|--------------------------------------|
| K-value model  | : | UNQC ( Uniquac with Unifac R and Q ) |
|                |   | LLV Three Phase K Values             |
| Enthalpy model | : | Mixed Model                          |
| Liquid density | : | Library                              |

## UNIQUAC Parameters:

| I | J | Uij-Ujj  | Uji-Uii | Aij   | Aji   |
|---|---|----------|---------|-------|-------|
| 1 | 2 | -218.911 | 582.494 | 0.000 | 0.000 |
| 1 | 3 | -365.948 | 589.393 | 0.000 | 0.000 |
| 2 | 3 | -371.327 | 972.438 | 0.000 | 0.000 |

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10.19

| Overall Mass Balance | kmol/h  |         | kg/h     |          |
|----------------------|---------|---------|----------|----------|
|                      | Input   | Output  | Input    | Output   |
| Methanol             | 259.697 | 1.286   | 8321.220 | 41.193   |
| Dimethyl Ether       | 0.000   | 129.206 | 0.000    | 5952.383 |
| Water                | 2.497   | 131.703 | 44.990   | 2372.633 |
| Total                | 262.195 | 262.195 | 8366.210 | 8366.209 |

13-15

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10:19  
EQUIPMENT SUMMARIES

## Pump Summary

|                     |          |
|---------------------|----------|
| Equip. No.          | 1        |
| Name                | P-201    |
| Output pressure bar | 15.5000  |
| Efficiency          | 0.6000   |
| Calculated power kW | 7.1171   |
| Calculated Pout bar | 15.5000  |
| Head m              | 187.2383 |
| Vol. flow rate m3/h | 10.6017  |

## Mixer Summary

|                     |         |
|---------------------|---------|
| Equip. No.          | 2       |
| Name                |         |
| Output Pressure bar | 15.2000 |

## Heat Exchanger Summary

| Equip. No.          | 3          | 4         | 6           |
|---------------------|------------|-----------|-------------|
| Name                | E-201      | E-202     | E-203       |
| Pressure drop 1 bar | 0.1500     |           | 1.1500      |
| T Out Str 1 C       |            | 250.0000  | 100.0000    |
| VF Out Str 1        | 1.0000     |           |             |
| Calc Ht Duty MJ/h   | 14361.1562 | 2021.1194 | -12765.6055 |
| LMTD (End points) C |            | 121.8691  |             |
| LMTD Corr Factor    |            | 1.0000    |             |
| Str1 Pout bar       | 15.0500    | 15.0500   | 13.9000     |
| Str2 Pout bar       |            | 15.0500   |             |

|                   |            |
|-------------------|------------|
| Equip. No.        | 11         |
| Name              | E-208      |
| T Out Str 1 C     | 50.0000    |
| Calc Ht Duty MJ/h | -1190.7069 |
| Str1 Pout bar     | 7.5000     |

## Reactor Summary

|                       |          |
|-----------------------|----------|
| Equip. No.            | 5        |
| Name                  | R-201    |
| Specify thermal mode: | 1        |
| C                     | 366.9053 |
| Key Component         | 1        |
| Frac. Conversion      | 0.8000   |

13-16

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10:19  
EQUIPMENT SUMMARIES

Heat of Reaction -5273.0000  
(Btu/lbmol)  
Calc H of Reac. -5273.0000  
(Btu/lbmol)

## Stoichiometrics:

Methanol -2.000  
Dimethyl Ether 1.000  
Water 1.000

## Valve Summary

| Equip. No.        | 7       | 9      | 12     |
|-------------------|---------|--------|--------|
| Name              |         |        |        |
| Pressure out bar  | 10.4000 |        | 1.2000 |
| Pressure drop bar |         | 3.4000 |        |

## Shortcut Distillation Summary

| Equip. No.                 | 8          | 10         |
|----------------------------|------------|------------|
| Name                       | T-201      | T-202      |
| Select mode:               | 2          | 2          |
| Column pressure drop (bar) | 0.2000     | 0.3000     |
| Light key component        | 2.0000     | 1.0000     |
| Light key split            | 0.9900     | 0.9900     |
| Heavy key component        | 1.0000     | 3.0000     |
| Heavy key split            | 0.0100     | 0.0100     |
| Number of stages           | 15.1800    | 18.1301    |
| R/Rmin                     | 1.2000     | 1.3000     |
| Lower bound R/Rmin         | 1.0100     |            |
| Upper bound R/Rmin         | 2.0100     |            |
| Number of stages, minimum  | 5.8186     | 8.8926     |
| Feed stage                 | 8.0900     | 9.5651     |
| Condenser duty MJ/h        | -3105.6807 | -5899.2832 |
| Reboiler duty MJ/h         | 2833.1018  | 5725.7480  |
| Reflux ratio, minimum      | 0.2946     | 1.3073     |
| Reflux ratio, calculated   | 0.3535     | 1.6994     |

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10:19  
FLOW SUMMARIES

| Stream No.          | 1         | 2       | 3       |
|---------------------|-----------|---------|---------|
| Stream Name         | meoh feed |         |         |
| Temp C              | 25.0000   | 25.3177 | 45.3236 |
| Pres bar            | 1.0100    | 15.5000 | 15.2000 |
| Enth MJ/h           | -63735.   | -63709. | -78921. |
| Vapor mole fraction | 0.00000   | 0.00000 | 0.00000 |
| Total kmol/h        | 262.19    | 262.19  | 328.15  |
| Flowrates in kmol/h |           |         |         |
| Methanol            | 259.70    | 259.70  | 323.01  |
| Dimethyl Ether      | 0.00      | 0.00    | 1.31    |
| Water               | 2.50      | 2.50    | 3.83    |

| Stream No.          | 4        | 5        | 6        |
|---------------------|----------|----------|----------|
| Stream Name         |          |          |          |
| Temp C              | 153.5873 | 250.0000 | 366.9053 |
| Pres bar            | 15.0500  | 15.0500  | 15.0500  |
| Enth MJ/h           | -64560.  | -62539.  | -62521.  |
| Vapor mole fraction | 1.0000   | 1.0000   | 1.0000   |
| Total kmol/h        | 328.15   | 328.15   | 328.15   |
| Flowrates in kmol/h |          |          |          |
| Methanol            | 323.01   | 323.01   | 64.60    |
| Dimethyl Ether      | 1.31     | 1.31     | 130.51   |
| Water               | 3.83     | 3.83     | 133.03   |

| Stream No.          | 7        | 8         | 9        |
|---------------------|----------|-----------|----------|
| Stream Name         |          |           |          |
| Temp C              | 280.5589 | 100.0000  | 88.0901  |
| Pres bar            | 15.0500  | 13.9000   | 10.4000  |
| Enth MJ/h           | -64542.  | -77308.   | -77310.  |
| Vapor mole fraction | 1.0000   | 0.0371029 | 0.115702 |
| Total kmol/h        | 328.15   | 328.15    | 328.15   |
| Flowrates in kmol/h |          |           |          |
| Methanol            | 64.60    | 64.60     | 64.60    |
| Dimethyl Ether      | 130.51   | 130.51    | 130.51   |
| Water               | 133.03   | 133.03    | 133.03   |

| Stream No.          | 10      | 11       | 12        |
|---------------------|---------|----------|-----------|
| Stream Name         |         |          |           |
| Temp C              | 46.1822 | 153.2090 | 138.6080  |
| Pres bar            | 10.4000 | 10.6000  | 7.2000    |
| Enth MJ/h           | -26161. | -51422.  | -51421.   |
| Vapor mole fraction | 0.00000 | 0.00000  | 0.0451196 |
| Total kmol/h        | 129.85  | 198.29   | 198.29    |
| Flowrates in kmol/h |         |          |           |
| Methanol            | 0.65    | 63.96    | 63.96     |
| Dimethyl Ether      | 129.21  | 1.31     | 1.31      |
| Water               | 0.00    | 133.03   | 133.03    |

Job Code: CHP18-2 Case Code: CHP18-2 Date: 06-27-97 Time: 10:19  
FLOW SUMMARIES

| Stream No.          | 13       | 14       | 15      |
|---------------------|----------|----------|---------|
| Stream Name         |          |          |         |
| Temp C              | 120.2759 | 166.8828 | 50.0000 |
| Pres bar            | 7.2000   | 7.5000   | 1.2000  |
| Enth MJ/h           | -15211.  | -36384.  | -37574. |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000 |
| Total kmol/h        | 65.95    | 132.34   | 132.34  |
| Flowrates in kmol/h |          |          |         |
| Methanol            | 63.32    | 0.64     | 0.64    |
| Dimethyl Ether      | 1.31     | 0.00     | 0.00    |
| Water               | 1.33     | 131.70   | 131.70  |

| Stream No.          | 16      |
|---------------------|---------|
| Stream Name         |         |
| Temp C              | 50.0000 |
| Pres bar            | 7.5000  |
| Enth MJ/h           | -37574. |
| Vapor mole fraction | 0.00000 |
| Total kmol/h        | 132.34  |
| Flowrates in kmol/h |         |
| Methanol            | 0.64    |
| Dimethyl Ether      | 0.00    |
| Water               | 131.70  |



### 13.3 Simulation of the IPA to acetone process, Appendix B, Figure B.10.1 AND TABLES

#### B.10.1 & B.10.2

Again the simulation results are shown in the output following this page. It is worth noting that due to non-ideal behavior of the Acetone/Water/IPA system a Unifac thermodynamic model was chosen for this simulation. Again the reactor was modeled with a simple stoichiometric reactor module. Therefore no reactor volume is calculated in the output. The required minimum input information for this simulation is given below:

1. Topology (see first page of following simulation)
2. Specification of feed streams - Stream 1(IPA), Stream 8 (water)
3. Equipment specifications
  - P-1101 output pressure (efficiency not required for H&M balance)
  - E-1101 stream outlet temperature
  - R-1101 conversion of key component (no kinetics req'd for H&M balance)
  - E-1102 stream outlet temperature
  - E-1103 stream outlet temperature
  - E-1108 stream outlet temperature
  - V-1102 temperature and pressure of flash
  - T-1101 number of trays, location of feeds - this module must be simulated with a rigorous tray-to-tray method because it is an absorber and no short-cut methods exist for this type of unit.
  - T-1102 identification of key components, recovery of key components,  $R/R_{min}$
  - T-1103 identification of key components, recovery of key components,  $R/R_{min}$
4. Conversion tolerance for recycle loop etc.

| <u>Equipment</u> | <u>Table B.10.1</u> | <u>Simulation Results</u> | <u>% diff.</u> |
|------------------|---------------------|---------------------------|----------------|
| P-1101           | 0.43 kW             | 0.42 kW                   | 2.3            |
| E-1101           | 3,550 MJ/h          | 3,550 MJ/h                | 0.0            |
| E-1102           | 3,260 MJ/h          | 3,337 MJ/h                | 2.4            |
| E-1103           | 563 MJ/h            | 506 MJ/h                  | -2.8           |
| E-1104           | 3,095 MJ/h          | 2,897 MJ/h                | 3.5            |
| E-1105           | 3,500 MJ/h          | 3,306 MJ/h                | 5.5            |
| E-1106           | 7,340 MJ/h          | 644 MJ/h                  | -91.2          |
| E-1107           | 7,390 MJ/h          | 689 MJ/h                  | -90.7          |
| E-1108           | 174 MJ/h            | 174 MJ/h                  | 0.0            |
| T-1101           | 2.5 m packing       | 5 theor. stages           | HETP=0.5 m     |
| T-1102           | 66 trays            | 46.7/0.7 = 67 trays       | -1.5           |
| T-1103           | 19 trays            | 12.4/0.7 = 18 trays       | 5.3            |

All the above results are in good agreement except for those associated with column, T-1103. This column was simulated using a short-cut routine in CHEMCAD™ III. This algorithm uses average relative volatilities to make estimates of minimum reflux ratio and minimum number of trays and then applies a correlation to estimate the number of trays required for a given  $R/R_{min}$  value. However, the top product from this column is a water-IPA azeotrope (or close to the azeotrope). This means that at the top of the column the relative volatilities of water and IPA approach unity. Thus the separation will be much more difficult than predicted by this short-cut method. This accounts for the much higher reflux ratio (condenser and reboiler duties) in Table B.10.2

It may be a good exercise to get the students to resimulate this column using a rigorous column package. The results should be much closer than the short-cut results.

Note that the results for T-1101- the absorber seem reasonable giving a HETP of 0.5 m or about 20 inches.

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48

Date: 06-27-97 Time: 11:48

## FLOWSHEET SUMMARY

## Equipment Label Stream Numbers

|    |            |     |      |       |
|----|------------|-----|------|-------|
| 1  | MIXE       | 1   | 14   | -98   |
| 2  | PUMP P-401 | 98  | -100 |       |
| 3  | HTXR E-401 | 4   | -97  |       |
| 4  | REAC R-401 | 97  | -3   |       |
| 5  | HTXR E-402 | 3   | -96  |       |
| 6  | FLAS V-402 | 90  | -5   | -99   |
| 7  | TOWR T-401 | 8   | 5    | -7 -6 |
| 8  | MIXE       | 6   | 99   | -9    |
| 9  | SHOR T-402 | 92  | -11  | -12   |
| 10 | SHOR T-403 | 12  | -14  | -15   |
| 11 | HTXR E-408 | 15  | -94  |       |
| 12 | CSEP       | 9   | -91  | -92   |
| 13 | HTXR E-403 | 96  | -90  |       |
| 14 | VALV       | 100 | -4   |       |

## Stream Connections

| Stream | Equipment From | Equipment To | Stream | Equipment From | Equipment To | Stream | Equipme From |
|--------|----------------|--------------|--------|----------------|--------------|--------|--------------|
| 1      |                | 1            | 9      | 8              | 12           | 92     | 12           |
| 3      | 4              | 5            | 11     | 9              |              | 94     | 11           |
| 4      | 14             | 3            | 12     | 9              | 10           | 96     | 5            |
| 5      | 6              | 7            | 14     | 10             | 1            | 97     | 3            |
| 6      | 7              | 8            | 15     | 10             | 11           | 98     | 1            |
| 7      | 7              |              | 90     | 13             | 6            | 99     | 6            |
| 8      |                | 7            | 91     | 12             |              | 100    | 2            |

Calculation mode : Sequential  
Flash algorithm : Normal

## Equipment Calculation Sequence

1 2 14 3 4 5 13 6 7 8 12 9 10 11

## Equipment Recycle Sequence

1 2 14 3 4 5 13 6 7 8 12 9 10

## Recycle Cut Streams

14

13-22

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48

Recycle Convergence Method: Direct Substitution

Max. loop iterations 40

Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-003 |
| Temperature | 1.000E-003 |
| Pressure    | 1.000E-003 |
| Enthalpy    | 1.000E-003 |
| Vapor frac. | 1.000E-003 |

Recycle calculation has converged.

#### COMPONENTS

|   | ID # | Name        |
|---|------|-------------|
| 1 | 1    | Hydrogen    |
| 2 | 140  | Acetone     |
| 3 | 62   | Water       |
| 4 | 145  | Isopropanol |

#### THERMODYNAMICS

|                |   |             |
|----------------|---|-------------|
| K-value model  | : | UNIFAC      |
| Enthalpy model | : | Latent Heat |
| Liquid density | : | Library     |

\* UNIFAC main groups 62 and 9, No interaction parameters.  
\* UNIFAC main groups 62 and 7, No interaction parameters.  
\* UNIFAC main groups 62 and 5, No interaction parameters.

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48

| Overall Mass Balance | kmol/h |         | kg/h     |          |
|----------------------|--------|---------|----------|----------|
|                      | Input  | Output  | Input    | Output   |
| Hydrogen             | 0.000  | 34.773  | 0.000    | 70.095   |
| Acetone              | 0.000  | 34.773  | 0.000    | 2019.614 |
| Water                | 37.147 | 37.147  | 669.207  | 669.207  |
| Isopropanol          | 34.816 | 0.043   | 2092.278 | 2.562    |
| Total                | 71.963 | 106.736 | 2761.485 | 2761.478 |

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48  
EQUIPMENT SUMMARIES

## Mixer Summary

Equip. No. 1 8  
Name

## Pump Summary

Equip. No. 2  
Name P-401  
Output pressure bar 2.8000  
Efficiency 0.4000  
Calculated power kW 0.4170  
Calculated Pout bar 2.8000  
Head m 22.8841  
Vol. flow rate m3/h 3.3526

## Heat Exchanger Summary

| Equip. No.          | 3         | 5          | 11        |
|---------------------|-----------|------------|-----------|
| Name                | E-401     | E-402      | E-408     |
| Pressure drop 1 bar | 0.1400    | 0.1400     | 0.1400    |
| T Out Str 1 C       | 234.0000  | 45.0000    | 45.0000   |
| Calc Ht Duty MJ/h   | 3549.4170 | -3337.0029 | -174.1102 |
| Str1 Pout bar       | 2.1600    | 2.0200     | 1.2600    |

Equip. No. 13  
Name E-403  
Pressure drop 1 bar 0.1400  
T Out Str 1 C 20.0000  
Calc Ht Duty MJ/h -505.5067  
Str1 Pout bar 1.8800

## Reactor Summary

Equip. No. 4  
Name R-401  
Specify thermal mode: 2  
C 350.0000  
MJ/h 2732.1438  
Key Component 4  
Frac. Conversion 0.9000  
Calc H of Reac. 5.5270e+007  
(J/kmol)

13-25

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48  
EQUIPMENT SUMMARIES

## Stoichiometrics:

|             |        |
|-------------|--------|
| Hydrogen    | 1.000  |
| Acetone     | 1.000  |
| Isopropanol | -1.000 |

## Flash Summary

|                |         |
|----------------|---------|
| Equip. No.     | 6       |
| Name           | V-402   |
| Flash Mode     | 2       |
| Param 1        | 20.0000 |
| Param 2        | 1.6300  |
| Heat duty MJ/h | 23.8612 |

## K values:

|             |           |
|-------------|-----------|
| Hydrogen    | 20410.104 |
| Acetone     | 0.195     |
| Water       | 0.030     |
| Isopropanol | 0.041     |

## Tower Rigorous Distillation Summary

|                          |         |
|--------------------------|---------|
| Equip. No.               | 7       |
| Name                     | T-401   |
| TOP pressure bar         | 1.5000  |
| Colm pressure drop (bar) | 0.1300  |
| No. of stages            | 5       |
| 1st feed stage           | 1       |
| 2nd feed stage           | 5       |
| Est. dist. rate (kmol/h) | 36.9400 |
| Est. stage 1 T C         | 25.0000 |
| Est. bottom T C          | 26.0000 |

## Shortcut Distillation Summary

| Equip. No.               | 9      | 10     |
|--------------------------|--------|--------|
| Name                     | T-402  | T-403  |
| Select mode:             | 2      | 2      |
| Colm pressure bar        | 1.2000 | 1.2000 |
| Colm pressure drop (bar) | 0.2000 | 0.2000 |
| Light key component      | 2.0000 | 4.0000 |
| Light key split          | 0.9950 | 0.9990 |

13-26

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48  
EQUIPMENT SUMMARIES

|                           |            |           |
|---------------------------|------------|-----------|
| Heavy key component       | 4.0000     | 3.0000    |
| Heavy key split           | 0.0060     | 0.0500    |
| Number of stages          | 46.6768    | 12.3589   |
| R/Rmin                    | 1.2000     | 1.5000    |
| Number of stages, minimum | 21.4049    | 6.7886    |
| Feed stage                | 23.4403    | 4.4531    |
| Condenser duty MJ/h       | -2896.6062 | -644.3007 |
| Reboiler duty MJ/h        | 3306.5957  | 688.5560  |
| Reflux ratio, minimum     | 1.7167     | 1.1573    |
| Reflux ratio, calculated  | 2.0601     | 1.7359    |

## Component Separator Summary

|                 |        |
|-----------------|--------|
| Equip. No.      | 12     |
| Name            |        |
| Component No. 1 | 1.0000 |

## Valve Summary

|                   |        |
|-------------------|--------|
| Equip. No.        | 14     |
| Name              |        |
| Pressure drop bar | 0.5000 |



Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48  
FLOW SUMMARIES

| Stream No.          | 1       | 3        | 4       |
|---------------------|---------|----------|---------|
| Stream Name         |         |          |         |
| Temp C              | 25.0000 | 350.0000 | 31.6765 |
| Pres bar            | 1.0100  | 2.1600   | 2.3000  |
| Enth MJ/h           | -15955. | -11399.  | -17697. |
| Vapor mole fraction | 0.00000 | 1.0000   | 0.00000 |
| Total kmol/h        | 51.9628 | 92.6039  | 57.8310 |
| Flowrates in kmol/h |         |          |         |
| Hydrogen            | 0.0000  | 34.7730  | 0.0000  |
| Acetone             | 0.0000  | 34.9339  | 0.1609  |
| Water               | 17.1472 | 19.0334  | 19.0334 |
| Isopropanol         | 34.8156 | 3.8637   | 38.6366 |

| Stream No.          | 5       | 6       | 7       |
|---------------------|---------|---------|---------|
| Stream Name         |         |         |         |
| Temp C              | 20.0000 | 26.7269 | 33.2741 |
| Pres bar            | 1.6300  | 1.6300  | 1.5000  |
| Enth MJ/h           | -1110.0 | -5918.5 | -905.75 |
| Vapor mole fraction | 1.0000  | 0.00000 | 1.0000  |
| Total kmol/h        | 39.7823 | 20.9410 | 38.8413 |
| Flowrates in kmol/h |         |         |         |
| Hydrogen            | 34.7707 | 0.0006  | 34.7702 |
| Acetone             | 4.4803  | 1.7341  | 2.7462  |
| Water               | 0.4146  | 19.1053 | 1.3093  |
| Isopropanol         | 0.1168  | 0.1010  | 0.0157  |

| Stream No.          | 8       | 9       | 11      |
|---------------------|---------|---------|---------|
| Stream Name         |         |         |         |
| Temp C              | 25.0000 | 21.5080 | 61.1912 |
| Pres bar            | 2.0000  | 1.6300  | 1.2000  |
| Enth MJ/h           | -5713.9 | -20026. | -7816.5 |
| Vapor mole fraction | 0.00000 | 0.00000 | 0.00000 |
| Total kmol/h        | 20.0000 | 73.7626 | 32.0499 |
| Flowrates in kmol/h |         |         |         |
| Hydrogen            | 0.0000  | 0.0028  | 0.0000  |
| Acetone             | 0.0000  | 32.1877 | 32.0268 |
| Water               | 20.0000 | 37.7241 | 0.0000  |
| Isopropanol         | 0.0000  | 3.8479  | 0.0231  |

| Stream No.          | 12      | 14      | 15       |
|---------------------|---------|---------|----------|
| Stream Name         |         |         |          |
| Temp C              | 90.3397 | 82.7982 | 109.2365 |
| Pres bar            | 1.4000  | 1.2000  | 1.4000   |
| Enth MJ/h           | -11799. | -1743.1 | -10012.  |
| Vapor mole fraction | 0.00000 | 0.00000 | 0.00000  |
| Total kmol/h        | 41.7099 | 5.8682  | 35.8418  |
| Flowrates in kmol/h |         |         |          |
| Hydrogen            | 0.0000  | 0.0000  | 0.0000   |
| Acetone             | 0.1609  | 0.1609  | 0.0000   |
| Water               | 37.7241 | 1.8862  | 35.8379  |

Job Code: CHP18-3 Case Code: CHP18-3 Date: 06-27-97 Time: 11:48  
FLOW SUMMARIES

Isopropanol 3.8249 3.8210 0.0038

| Stream No.          | 90       | 91           | 92      |
|---------------------|----------|--------------|---------|
| Stream Name         |          |              |         |
| Temp C              | 20.0000  | 21.5080      | 21.5080 |
| Pres bar            | 1.8800   | 1.6300       | 1.6300  |
| Enth MJ/h           | -15241.  | -0.000144186 | -20026. |
| Vapor mole fraction | 0.421707 | 1.0000       | 0.00000 |
| Total kmol/h        | 92.6039  | 0.0028       | 73.7598 |
| Flowrates in kmol/h |          |              |         |
| Hydrogen            | 34.7730  | 0.0028       | 0.0000  |
| Acetone             | 34.9339  | 0.0000       | 32.1877 |
| Water               | 19.0334  | 0.0000       | 37.7241 |
| Isopropanol         | 3.8637   | 0.0000       | 3.8479  |

| Stream No.          | 94      | 96       | 97       |
|---------------------|---------|----------|----------|
| Stream Name         |         |          |          |
| Temp C              | 45.0000 | 45.0000  | 234.0000 |
| Pres bar            | 1.2600  | 2.0200   | 2.1600   |
| Enth MJ/h           | -10186. | -14736.  | -14148.  |
| Vapor mole fraction | 0.00000 | 0.528497 | 1.0000   |
| Total kmol/h        | 35.8418 | 92.6039  | 57.8310  |
| Flowrates in kmol/h |         |          |          |
| Hydrogen            | 0.0000  | 34.7730  | 0.0000   |
| Acetone             | 0.0000  | 34.9339  | 0.1609   |
| Water               | 35.8379 | 19.0334  | 19.0334  |
| Isopropanol         | 0.0038  | 3.8637   | 38.6366  |

| Stream No.          | 98      | 99      | 100     |
|---------------------|---------|---------|---------|
| Stream Name         |         |         |         |
| Temp C              | 31.4815 | 20.0000 | 31.6765 |
| Pres bar            | 1.0100  | 1.6300  | 2.8000  |
| Enth MJ/h           | -17699. | -14107. | -17697. |
| Vapor mole fraction | 0.00000 | 0.00000 | 0.00000 |
| Total kmol/h        | 57.8310 | 52.8216 | 57.8310 |
| Flowrates in kmol/h |         |         |         |
| Hydrogen            | 0.0000  | 0.0023  | 0.0000  |
| Acetone             | 0.1609  | 30.4536 | 0.1609  |
| Water               | 19.0334 | 18.6188 | 19.0334 |
| Isopropanol         | 38.6366 | 3.7469  | 38.6366 |

13.4 Comparison of short-cut and rigorous simulation algorithms for the Benzene Recovery Column, T-101.

The results of the simulations for the two algorithms are given in the attached print outs. Comparison of Streams 1, 2, and 3 (short-cut results) with Streams 11, 12, and 13 (rigorous results) show that the input and output streams for the two simulations are essentially identical. A comparison of other pertinent output is given below - note that the same number of trays are used in both simulations.

| <u>result</u>         | <u>short-cut</u> | <u>rigorous</u> | <u>% diff.</u> |
|-----------------------|------------------|-----------------|----------------|
| reflux (kmol/h)       | 184.6            | 166.457         | 9.8            |
| reflux ratio          | 1.7498           | 1.5782          | 9.8            |
| condenser duty (MJ/h) | 8,397            | 7,874           | 6.2            |
| reboiler duty (MJ/h)  | 9,196            | 8,673           | 5.7            |

Although the differences between the two simulations are not large it is worth noting that this system (benzene - toluene) is often used as the standard example for ideal (constant molal overflow) systems. Even for this system we have some error in this assumption - clearly the more non-ideal the system becomes the greater these differences are likely to become. This is well illustrated in Problem 13.3 above for T-1103.

When using rigorous simulations several other useful options are available such as tray-to-tray profiles of flows and temperatures etc. The liquid and vapor profiles for T-101 are included in the output and it can be seen that the constant molal overflow assumption is a very good one (but not exactly correct even for this "ideal" system).

|                           |                         |
|---------------------------|-------------------------|
| Change in L (above feed)  | 166.21 to 160.91 kmol/h |
| Change in V (above feed)  | 271.93 to 267.21 kmol/h |
| Change in L/V             | 0.611 to 0.602 (1.5 %)  |
| Change in L' (below feed) | 326.68 to 311.79 kmol/h |
| Change in V' (below feed) | 291.05 to 275.72 kmol/h |
| Change in L'/V'           | 1.122 to 1.131 (0.7 %)  |

Job Code: CHP18-4 Case Code: CHP18-4 Date: 06-27-97 Time: 16:30

Date: 06-27-97 Time: 16:30

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers |
|-----------|-------|----------------|
|-----------|-------|----------------|

|   |                   |            |
|---|-------------------|------------|
| 1 | SHOR T-101 shtcut | 1 -2 -3    |
| 2 | TOWR T-101 rigor  | 11 -12 -13 |

## Stream Connections

| Stream | Equipment<br>From | Equipment<br>To | Stream | Equipment<br>From | Equipment<br>To | Stream | Equipment<br>From |
|--------|-------------------|-----------------|--------|-------------------|-----------------|--------|-------------------|
| 1      |                   | 1               | 3      | 1                 |                 | 12     | 2                 |
| 2      | 1                 |                 | 11     |                   | 2               | 13     | 2                 |

Calculation mode : Sequential  
Flash algorithm : Normal

Equipment Calculation Sequence  
1 2

No recycle loops in the flowsheet.

## COMPONENTS

|   | ID # | Name    |
|---|------|---------|
| 1 | 40   | Benzene |
| 2 | 41   | Toluene |

## THERMODYNAMICS

K-value model : SRK  
Enthalpy model : SRK  
Liquid density : Library

Job Code: CHP18-4 Case Code: CHP18-4 Date: 06-27-97 Time: 16:30

| Overall Mass Balance | kmol/h  |         | kg/h      |           |
|----------------------|---------|---------|-----------|-----------|
|                      | Input   | Output  | Input     | Output    |
| Benzene              | 212.340 | 212.340 | 16586.726 | 16586.732 |
| Toluene              | 69.860  | 69.860  | 6436.970  | 6436.968  |
| Total                | 282.200 | 282.200 | 23023.695 | 23023.700 |

Job Code: CHP18-4 Case Code: CHP18-4 Date: 06-27-97 Time: 16:30  
EQUIPMENT SUMMARIES

## Shortcut Distillation Summary

|                             |              |
|-----------------------------|--------------|
| Equip. No.                  | 1            |
| Name                        | T-101 shtcut |
| Select mode:                | 2            |
| Colm pressure drop<br>(bar) | 0.1700       |
| Light key component         | 1.0000       |
| Light key split             | 0.9900       |
| Heavy key component         | 2.0000       |
| Heavy key split             | 0.0100       |
| Number of stages            | 24.5637      |
| R/Rmin                      | 1.5000       |
| Number of stages, minimum   | 13.5053      |
| Feed stage                  | 12.7818      |
| Condenser duty MJ/h         | -8397.1709   |
| Reboiler duty MJ/h          | 9196.0742    |
| Reflux ratio, minimum       | 1.1665       |
| Reflux ratio, calculated    | 1.7498       |

## Tower Rigorous Distillation Summary

|                              |             |
|------------------------------|-------------|
| Equip. No.                   | 2           |
| Name                         | T-101 rigor |
| TOP pressure bar             | 2.7000      |
| Colm pressure drop<br>(bar)  | 0.1700      |
| No. of stages                | 25          |
| 1st feed stage               | 13          |
| Select condenser mode:       | 7           |
| Condenser spec.              | 0.9900      |
| Cond. comp i                 | 1           |
| Select reboiler mode:        | 6           |
| Reboiler spec.               | 0.9702      |
| Rebl. comp i                 | 2           |
| Est. dist. rate<br>(kmol/h)  | 100.0000    |
| Est. reflux rate<br>(kmol/h) | 150.0000    |
| Est. stage 1 T C             | 100.0000    |
| Est. bottom T C              | 130.0000    |
| Calc cond duty MJ/h          | -7874.4390  |
| Calc rebr duty MJ/h          | 8673.1680   |
| Reflux ratio                 | 1.5782      |
| Reflux mole kmol/h           | 166.4570    |
| Reflux mass kg/h             | 13010.6465  |

Job Code: CHP18-4 Case Code: CHP18-4 Date: 06-27-97 Time: 16:30  
FLOW SUMMARIES

| Stream No.          | 1        | 2        | 3        |
|---------------------|----------|----------|----------|
| Stream Name         |          |          |          |
| Temp C              | 90.0000  | 115.7703 | 150.3135 |
| Pres bar            | 2.7000   | 2.7000   | 2.8700   |
| Enth MJ/h           | 6969.6   | 6534.9   | 1233.6   |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000  |
| Total kmol/h        | 141.1000 | 105.4576 | 35.6424  |
| Flowrates in kmol/h |          |          |          |
| Benzene             | 106.1700 | 105.1083 | 1.0617   |
| Toluene             | 34.9300  | 0.3493   | 34.5807  |

| Stream No.          | 11       | 12       | 13       |
|---------------------|----------|----------|----------|
| Stream Name         |          |          |          |
| Temp C              | 90.0000  | 115.7743 | 150.3140 |
| Pres bar            | 2.7000   | 2.7000   | 2.8700   |
| Enth MJ/h           | 6969.6   | 6535.3   | 1233.1   |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000  |
| Total kmol/h        | 141.1000 | 105.4709 | 35.6292  |
| Flowrates in kmol/h |          |          |          |
| Benzene             | 106.1700 | 105.1083 | 1.0617   |
| Toluene             | 34.9300  | 0.3626   | 34.5674  |

Job Code: CHP18-4 Case Code: CHP18-4 Date: 06-27-97 Time: 16:30  
DISTILLATION PROFILE

Unit type : TOWR Unit name: T-101 rigor Eqp # 2

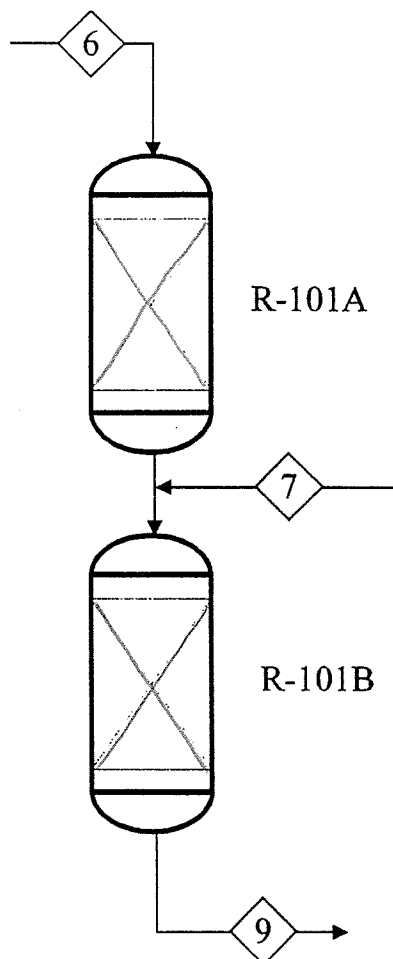
| Stg | Temp<br>C | Pres<br>bar | * Net Flows *    |                 | Feeds<br>kmol/h | Product<br>kmol/h | Du<br>MJ |
|-----|-----------|-------------|------------------|-----------------|-----------------|-------------------|----------|
|     |           |             | Liquid<br>kmol/h | Vapor<br>kmol/h |                 |                   |          |
| 1   | 115.8     | 2.70        | 166.46           |                 |                 | 105.47            |          |
| 2   | 115.9     | 2.70        | 166.21           | 271.93          |                 |                   |          |
| 3   | 116.1     | 2.71        | 166.03           | 271.69          |                 |                   |          |
| 4   | 116.4     | 2.71        | 165.79           | 271.50          |                 |                   |          |
| 5   | 116.7     | 2.72        | 165.48           | 271.26          |                 |                   |          |
| 6   | 117.1     | 2.73        | 165.09           | 270.96          |                 |                   |          |
| 7   | 117.6     | 2.74        | 164.59           | 270.56          |                 |                   |          |
| 8   | 118.2     | 2.74        | 164.00           | 270.06          |                 |                   |          |
| 9   | 118.8     | 2.75        | 163.31           | 269.47          |                 |                   |          |
| 10  | 119.6     | 2.76        | 162.55           | 268.78          |                 |                   |          |
| 11  | 120.4     | 2.77        | 161.74           | 268.02          |                 |                   |          |
| 12  | 121.3     | 2.77        | 160.91           | 267.21          |                 |                   |          |
| 13  | 122.2     | 2.78        | 326.68           | 266.38          | 141.10          |                   |          |
| 14  | 122.6     | 2.79        | 326.24           | 291.05          |                 |                   |          |
| 15  | 123.3     | 2.80        | 325.24           | 290.61          |                 |                   |          |
| 16  | 124.4     | 2.80        | 323.64           | 289.61          |                 |                   |          |
| 17  | 126.1     | 2.81        | 321.12           | 288.01          |                 |                   |          |
| 18  | 128.8     | 2.82        | 318.05           | 285.49          |                 |                   |          |
| 19  | 132.4     | 2.83        | 314.85           | 282.42          |                 |                   |          |
| 20  | 136.6     | 2.83        | 312.42           | 279.22          |                 |                   |          |
| 21  | 140.9     | 2.84        | 311.16           | 276.79          |                 |                   |          |
| 22  | 144.5     | 2.85        | 311.02           | 275.53          |                 |                   |          |
| 23  | 147.2     | 2.86        | 311.35           | 275.39          |                 |                   |          |
| 24  | 149.1     | 2.86        | 311.79           | 275.72          |                 |                   |          |
| 25  | 150.3     | 2.87        |                  | 276.16          |                 | 35.63             |          |

Mole Reflux ratio 1.578

13-35



- 13.5 The simulation of the reactor system is shown in the attached print out. The reactor was simulated as two adiabatic packed beds with an intermediate “cold shot” (Stream 7) at the inlet of the second bed.



The simulation shows that the overall conversion in the reactor is the desired 75%. The temperature profiles for each section of the reactor (R-101A and R-101B) show that the maximum temperatures reached in each section are 653.35°C and 653.74°C, which are close to but do not exceed the 655°C limit. The total combined volumes of each section of the reactor are

$$V_{total} = V_{R101A} + V_{R101B} = 30.11 + 11.45 = 41.56 \text{ m}^3$$

$$\text{from Table 1.7 we have } V_{total} = \pi D^2 L / 4 = \pi (2.3)^2 (10) / 4 = 41.55 \text{ m}^3$$

Thus the simulation and Table 1.7 are in excellent agreement.

Job Code: CHP18-5 Case Code: CHP18-5 Date: 06-27-97 Time: 17:13

Date: 06-27-97 Time: 17:13

## FLOWSHEET SUMMARY

| Equipment | Label       | Stream Numbers |
|-----------|-------------|----------------|
| 1         | KREA R-101A | 6 -98          |
| 2         | MIXE        | 7 98 -99       |
| 3         | KREA R-101B | 99 -9          |

## Stream Connections

| Stream | Equipment<br>From | To |
|--------|-------------------|----|
| 6      |                   | 1  |
| 7      |                   | 2  |
| 9      | 3                 |    |
| 98     | 1                 | 2  |
| 99     | 2                 | 3  |

Calculation mode : Sequential  
Flash algorithm : Normal

Equipment Calculation Sequence  
1 2 3

No recycle loops in the flowsheet.

## COMPONENTS

|   | ID # | Name     |
|---|------|----------|
| 1 | 1    | Hydrogen |
| 2 | 2    | Methane  |
| 3 | 40   | Benzene  |
| 4 | 41   | Toluene  |

## THERMODYNAMICS

K-value model : SRK  
Enthalpy model : SRK  
Liquid density : Library

Job Code: CHP18-5 Case Code: CHP18-5 Date: 06-27-97 Time: 17:13

| Overall Mass Balance | kmol/h   |          | kg/h      |           |
|----------------------|----------|----------|-----------|-----------|
|                      | Input    | Output   | Input     | Output    |
| Hydrogen             | 760.600  | 652.551  | 1533.218  | 1315.412  |
| Methane              | 334.250  | 442.299  | 5362.372  | 7095.798  |
| Benzene              | 7.970    | 116.019  | 622.569   | 9062.695  |
| Toluene              | 144.040  | 35.991   | 13271.990 | 3316.265  |
| Total                | 1246.860 | 1246.860 | 20790.148 | 20790.171 |

Job Code: CHP18-5 Case Code: CHP18-5 Date: 06-27-97 Time: 17:13  
EQUIPMENT SUMMARIES

## Kinetic Reactor Summary

| Equip. No.                  | 1          | 3         |
|-----------------------------|------------|-----------|
| Name                        | R-101A     | R-101B    |
| No. of Reactions            | 1          | 1         |
| Pressure Drop bar           | 1.0000     |           |
| Specify reactor type:       | 2          | 2         |
| Specify reaction phase      | 1          | 1         |
| Specify thermal mode:       | 2          | 2         |
| Tout C                      | 653.3050   | 653.7204  |
| Specify calc. mode:         | 1          | 1         |
| Vol                         | 30.1058    | 11.4480   |
| Key                         | 4          | 4         |
| Conversion                  | 0.6000     | 0.3760    |
| Time Unit:                  | 2          | 2         |
| Volume Unit:                | 1          | 1         |
| Activation E/H of Rxn Unit: |            | 5 5       |
| Molar Flow Unit:            | 1          | 1         |
| Overall Ht of Rxn (MJ/h)    | -3621.1519 | -908.4697 |

## Reaction Stoichiometrics and Parameters for unit no. 1

## Reaction 1

| Rate const = | 2.8330e+007 | Act. E =    | 1.4811e+002 | Hrxn =      | 0.0000e  |
|--------------|-------------|-------------|-------------|-------------|----------|
| Comp         | Stoich.     | Exp.factor  | Adsorb Fac. | Adsorb E    | Adsorb E |
| 1            | -1.00e+000  | 5.0000e-001 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |
| 4            | -1.00e+000  | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |
| 2            | 1.00e+000   | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |
| 3            | 1.00e+000   | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |

## Reaction Stoichiometrics and Parameters for unit no. 3

## Reaction 1

| Rate const = | 2.8330e+007 | Act. E =    | 1.4811e+002 | Hrxn =      | 0.0000e  |
|--------------|-------------|-------------|-------------|-------------|----------|
| Comp         | Stoich.     | Exp.factor  | Adsorb Fac. | Adsorb E    | Adsorb E |
| 1            | -1.00e+000  | 5.0000e-001 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |
| 4            | -1.00e+000  | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |
| 2            | 1.00e+000   | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |
| 3            | 1.00e+000   | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+ |

## Plug Flow Profile for unit no. 1

| Vol  | Temp   | Press | Total   | Mole frac  | Mole fra  |
|------|--------|-------|---------|------------|-----------|
| m3   | C      | bar   | kmol/h  | Hydrogen   | Metha     |
| 0.00 | 600.02 | 25.00 | 1204.29 | 6.107E-001 | 2.635E 0  |
| 1.51 | 602.46 | 24.95 | 1204.29 | 6.074E-001 | 2.667E-00 |
| 3.01 | 604.91 | 24.90 | 1204.29 | 6.040E-001 | 2.701E-00 |

Job Code: CHP18-5 Case Code: CHP18-5 Date: 06-27-97 Time: 17:13  
EQUIPMENT SUMMARIES

|       |        |       |         |            |           |
|-------|--------|-------|---------|------------|-----------|
| 4.52  | 607.46 | 24.85 | 1204.29 | 6.007E-001 | 2.735E-00 |
| 6.02  | 610.02 | 24.80 | 1204.29 | 5.972E-001 | 2.769E-00 |
| 7.53  | 612.63 | 24.75 | 1204.29 | 5.937E-001 | 2.804E-00 |
| 9.03  | 615.24 | 24.70 | 1204.29 | 5.901E-001 | 2.840E-00 |
| 10.54 | 617.91 | 24.65 | 1204.29 | 5.865E-001 | 2.876E-00 |
| 12.04 | 620.63 | 24.60 | 1204.29 | 5.829E-001 | 2.912E-00 |
| 13.55 | 623.35 | 24.55 | 1204.29 | 5.792E-001 | 2.949E-00 |
| 15.05 | 626.13 | 24.50 | 1204.29 | 5.755E-001 | 2.986E-00 |
| 16.56 | 628.91 | 24.45 | 1204.29 | 5.718E-001 | 3.023E-00 |
| 18.06 | 631.68 | 24.40 | 1204.29 | 5.680E-001 | 3.061E-00 |
| 19.57 | 634.46 | 24.35 | 1204.29 | 5.643E-001 | 3.098E-00 |
| 21.07 | 637.24 | 24.30 | 1204.29 | 5.605E-001 | 3.136E-00 |
| 22.58 | 640.02 | 24.25 | 1204.29 | 5.568E-001 | 3.173E-00 |
| 24.08 | 642.74 | 24.20 | 1204.29 | 5.531E-001 | 3.210E-00 |
| 25.59 | 645.46 | 24.15 | 1204.29 | 5.495E-001 | 3.246E-00 |
| 27.09 | 648.13 | 24.10 | 1204.29 | 5.459E-001 | 3.282E-00 |
| 28.60 | 650.74 | 24.05 | 1204.29 | 5.424E-001 | 3.318E-00 |
| 30.11 | 653.29 | 24.00 | 1204.29 | 5.389E-001 | 3.352E-00 |

| Vol<br>m3 | Temp<br>C | Press<br>bar | Total<br>kmol/h | Mole frac<br>Benzene | Mole fra<br>Toluene |
|-----------|-----------|--------------|-----------------|----------------------|---------------------|
| 0.00      | 600.02    | 25.00        | 1204.29         | 6.311E-003           | 1.196E-00           |
| 1.51      | 602.46    | 24.95        | 1204.29         | 9.581E-003           | 1.163E-00           |
| 3.01      | 604.91    | 24.90        | 1204.29         | 1.291E-002           | 1.130E-00           |
| 4.52      | 607.46    | 24.85        | 1204.29         | 1.630E-002           | 1.096E-00           |
| 6.02      | 610.02    | 24.80        | 1204.29         | 1.975E-002           | 1.061E-00           |
| 7.53      | 612.63    | 24.75        | 1204.29         | 2.325E-002           | 1.026E-00           |
| 9.03      | 615.24    | 24.70        | 1204.29         | 2.681E-002           | 9.908E-00           |
| 10.54     | 617.91    | 24.65        | 1204.29         | 3.041E-002           | 9.548E-00           |
| 12.04     | 620.63    | 24.60        | 1204.29         | 3.405E-002           | 9.183E-00           |
| 13.55     | 623.35    | 24.55        | 1204.29         | 3.773E-002           | 8.815E-00           |
| 15.05     | 626.13    | 24.50        | 1204.29         | 4.144E-002           | 8.444E-00           |
| 16.56     | 628.91    | 24.45        | 1204.29         | 4.517E-002           | 8.071E-00           |
| 18.06     | 631.68    | 24.40        | 1204.29         | 4.891E-002           | 7.697E-00           |
| 19.57     | 634.46    | 24.35        | 1204.29         | 5.266E-002           | 7.322E-00           |
| 21.07     | 637.24    | 24.30        | 1204.29         | 5.641E-002           | 6.948E-00           |
| 22.58     | 640.02    | 24.25        | 1204.29         | 6.013E-002           | 6.575E-00           |
| 24.08     | 642.74    | 24.20        | 1204.29         | 6.382E-002           | 6.206E-00           |
| 25.59     | 645.46    | 24.15        | 1204.29         | 6.747E-002           | 5.841E-00           |
| 27.09     | 648.13    | 24.10        | 1204.29         | 7.107E-002           | 5.481E-00           |
| 28.60     | 650.74    | 24.05        | 1204.29         | 7.459E-002           | 5.129E-00           |
| 30.11     | 653.29    | 24.00        | 1204.29         | 7.804E-002           | 4.785E-00           |

Plug Flow Profile for unit no. 3

| Vol<br>m3 | Temp<br>C | Press<br>bar | Total<br>kmol/h | Mole frac<br>Hydrogen | Mole fra<br>Methane |
|-----------|-----------|--------------|-----------------|-----------------------|---------------------|
| 0.00      | 640.68    | 24.00        | 1246.88         | 5.407E-001            | 3.374E-00           |
| 0.57      | 641.41    | 24.00        | 1246.88         | 5.398E-001            | 3.383E-00           |
| 1.14      | 642.07    | 24.00        | 1246.88         | 5.388E-001            | 3.393E-00           |

13-40

Job Code: CHP18-5 Case Code: CHP18-5 Date: 06-27-97 Time: 17:13  
EQUIPMENT SUMMARIES

|       |        |       |         |            |           |
|-------|--------|-------|---------|------------|-----------|
| 1.72  | 642.79 | 24.00 | 1246.88 | 5.379E-001 | 3.402E-00 |
| 2.29  | 643.52 | 24.00 | 1246.88 | 5.370E-001 | 3.411E-00 |
| 2.86  | 644.18 | 24.00 | 1246.88 | 5.360E-001 | 3.420E-00 |
| 3.43  | 644.91 | 24.00 | 1246.88 | 5.351E-001 | 3.430E-00 |
| 4.01  | 645.57 | 24.00 | 1246.88 | 5.342E-001 | 3.439E-00 |
| 4.58  | 646.24 | 24.00 | 1246.88 | 5.333E-001 | 3.448E-00 |
| 5.15  | 646.91 | 24.00 | 1246.88 | 5.324E-001 | 3.457E-00 |
| 5.72  | 647.57 | 24.00 | 1246.88 | 5.315E-001 | 3.465E-00 |
| 6.30  | 648.24 | 24.00 | 1246.88 | 5.307E-001 | 3.474E-00 |
| 6.87  | 648.85 | 24.00 | 1246.88 | 5.298E-001 | 3.483E-00 |
| 7.44  | 649.52 | 24.00 | 1246.88 | 5.290E-001 | 3.491E-00 |
| 8.01  | 650.13 | 24.00 | 1246.88 | 5.281E-001 | 3.499E-00 |
| 8.59  | 650.74 | 24.00 | 1246.88 | 5.273E-001 | 3.508E-00 |
| 9.16  | 651.35 | 24.00 | 1246.88 | 5.265E-001 | 3.516E-00 |
| 9.73  | 651.96 | 24.00 | 1246.88 | 5.257E-001 | 3.524E-00 |
| 10.30 | 652.57 | 24.00 | 1246.88 | 5.249E-001 | 3.532E-00 |
| 10.88 | 653.13 | 24.00 | 1246.88 | 5.241E-001 | 3.540E-00 |
| 11.45 | 653.74 | 24.00 | 1246.88 | 5.234E-001 | 3.547E-00 |

| Vol<br>m3 | Temp<br>C | Press<br>bar | Total<br>kmol/h | Mole frac<br>Benzene | Mole fra<br>Toluene |
|-----------|-----------|--------------|-----------------|----------------------|---------------------|
| 0.00      | 640.68    | 24.00        | 1246.88         | 7.567E-002           | 4.625E-00           |
| 0.57      | 641.41    | 24.00        | 1246.88         | 7.662E-002           | 4.529E-00           |
| 1.14      | 642.07    | 24.00        | 1246.88         | 7.757E-002           | 4.434E-00           |
| 1.72      | 642.79    | 24.00        | 1246.88         | 7.852E-002           | 4.340E-00           |
| 2.29      | 643.52    | 24.00        | 1246.88         | 7.945E-002           | 4.247E-00           |
| 2.86      | 644.18    | 24.00        | 1246.88         | 8.037E-002           | 4.154E-00           |
| 3.43      | 644.91    | 24.00        | 1246.88         | 8.129E-002           | 4.062E-00           |
| 4.01      | 645.57    | 24.00        | 1246.88         | 8.220E-002           | 3.972E-00           |
| 4.58      | 646.24    | 24.00        | 1246.88         | 8.309E-002           | 3.882E-00           |
| 5.15      | 646.91    | 24.00        | 1246.88         | 8.398E-002           | 3.793E-00           |
| 5.72      | 647.57    | 24.00        | 1246.88         | 8.486E-002           | 3.705E-00           |
| 6.30      | 648.24    | 24.00        | 1246.88         | 8.573E-002           | 3.619E-00           |
| 6.87      | 648.85    | 24.00        | 1246.88         | 8.659E-002           | 3.533E-00           |
| 7.44      | 649.52    | 24.00        | 1246.88         | 8.743E-002           | 3.448E-00           |
| 8.01      | 650.13    | 24.00        | 1246.88         | 8.827E-002           | 3.364E-00           |
| 8.59      | 650.74    | 24.00        | 1246.88         | 8.910E-002           | 3.282E-00           |
| 9.16      | 651.35    | 24.00        | 1246.88         | 8.991E-002           | 3.200E-00           |
| 9.73      | 651.96    | 24.00        | 1246.88         | 9.071E-002           | 3.120E-00           |
| 10.30     | 652.57    | 24.00        | 1246.88         | 9.150E-002           | 3.041E-00           |
| 10.88     | 653.13    | 24.00        | 1246.88         | 9.228E-002           | 2.963E-00           |
| 11.45     | 653.74    | 24.00        | 1246.88         | 9.305E-002           | 2.887E-00           |

13-41

Job Code: CHP18-5 Case Code: CHP18-5 Date: 06-27-97 Time: 17:13  
EQUIPMENT SUMMARIES

Mixer Summary

| Equip. No. |   |
|------------|---|
| Name       |   |
|            | 2 |

13-42

Job Code: CHP18-5

Case Code: CHP18-5

Date: 06-27-97

Time: 17:13

## FLOW SUMMARIES

| Stream No.          | 6         | 7       | 9         |
|---------------------|-----------|---------|-----------|
| Stream Name         |           |         |           |
| Temp C              | 600.0000  | 40.8000 | 653.7204  |
| Pres bar            | 25.0000   | 25.0000 | 24.0000   |
| Enth MJ/h           | 21897.    | -1218.7 | 20678.    |
| Vapor mole fraction | 1.0000    | 1.0000  | 1.0000    |
| Total kmol/h        | 1204.2999 | 42.5600 | 1246.8598 |
| Flowrates in kmol/h |           |         |           |
| Hydrogen            | 735.4000  | 25.2000 | 652.5510  |
| Methane             | 317.3000  | 16.9500 | 442.2987  |
| Benzene             | 7.6000    | 0.3700  | 116.0188  |
| Toluene             | 144.0000  | 0.0400  | 35.9912   |

| Stream No.          | 98        | 99        |
|---------------------|-----------|-----------|
| Stream Name         |           |           |
| Temp C              | 653.3050  | 640.6738  |
| Pres bar            | 24.0000   | 24.0000   |
| Enth MJ/h           | 21896.    | 20678.    |
| Vapor mole fraction | 1.0000    | 1.0000    |
| Total kmol/h        | 1204.2999 | 1246.8599 |
| Flowrates in kmol/h |           |           |
| Hydrogen            | 649.0215  | 674.2215  |
| Methane             | 403.6784  | 420.6284  |
| Benzene             | 93.9783   | 94.3483   |
| Toluene             | 57.6216   | 57.6616   |



### 13.6

The results will differ for different simulators and their databases of BIP's. The following is for ChemCAD.

|               | ChemCAD BIPs | BIPS set to zero |
|---------------|--------------|------------------|
| No. of stages | 23.3         | 26.2             |
| Reflux ratio  | 1.745        | 2.147            |

### 13.7

In Appendix B, each of the four processes has a section on the thermodynamic models used. They are summarized below:

| Process         | Model used         | Reason                                     | Another model                                              | Page |
|-----------------|--------------------|--------------------------------------------|------------------------------------------------------------|------|
| DME             | UNIQUAC/<br>UNIFAC | Highly<br>nonideal.<br>VLE Data<br>missing | NRTL, if VLE<br>for all binaries is<br>regressed           | 946  |
| Acrylic<br>Acid | UNIFAC             | LLE Data<br>missing                        | User-input<br>distribution<br>coefficients<br>based on LLE | 957  |
| Acetone         | UNIQUAC            | BIPs<br>available.<br>azeotrope            | NRTL                                                       | 965  |
| Heptenes        | SRK                | Well<br>behaved.                           | PR                                                         | 976  |

13.8

The following results are from ChemCAD.

**NRTL**

Stages required: 17.0

|          | DME                     | Methanol                    | Water                        |
|----------|-------------------------|-----------------------------|------------------------------|
| DME      |                         | 139.995,<br>$\alpha=0.3$    |                              |
| Methanol | 153.78,<br>$\alpha=0.3$ |                             | -24.4933,<br>$\alpha=0.3001$ |
| Water    |                         | 307.166,<br>$\alpha=0.3001$ |                              |

**Wilson**

Stages required: 16.7

|          | DME     | Methanol | Water   |
|----------|---------|----------|---------|
| DME      |         | 54.7165  |         |
| Methanol | 564.363 |          | -52.605 |
| Water    |         | 620.63   |         |

**UNIQUAC**

Stages required: 16.7

|          | DME     | Methanol | Water  |
|----------|---------|----------|--------|
| DME      |         | 361.266  |        |
| Methanol | 32.7318 |          | 95.259 |
| Water    |         | -10.377  |        |

13.9

There are many sets of methanol/water data, 39 of which are given in the *DECHEMA Vapor-Liquid Equilibrium Data Collection*, volume 1, part I, Frankfurt/Main, 1977. Below is a set for 760 Torr from page 48 of this reference.

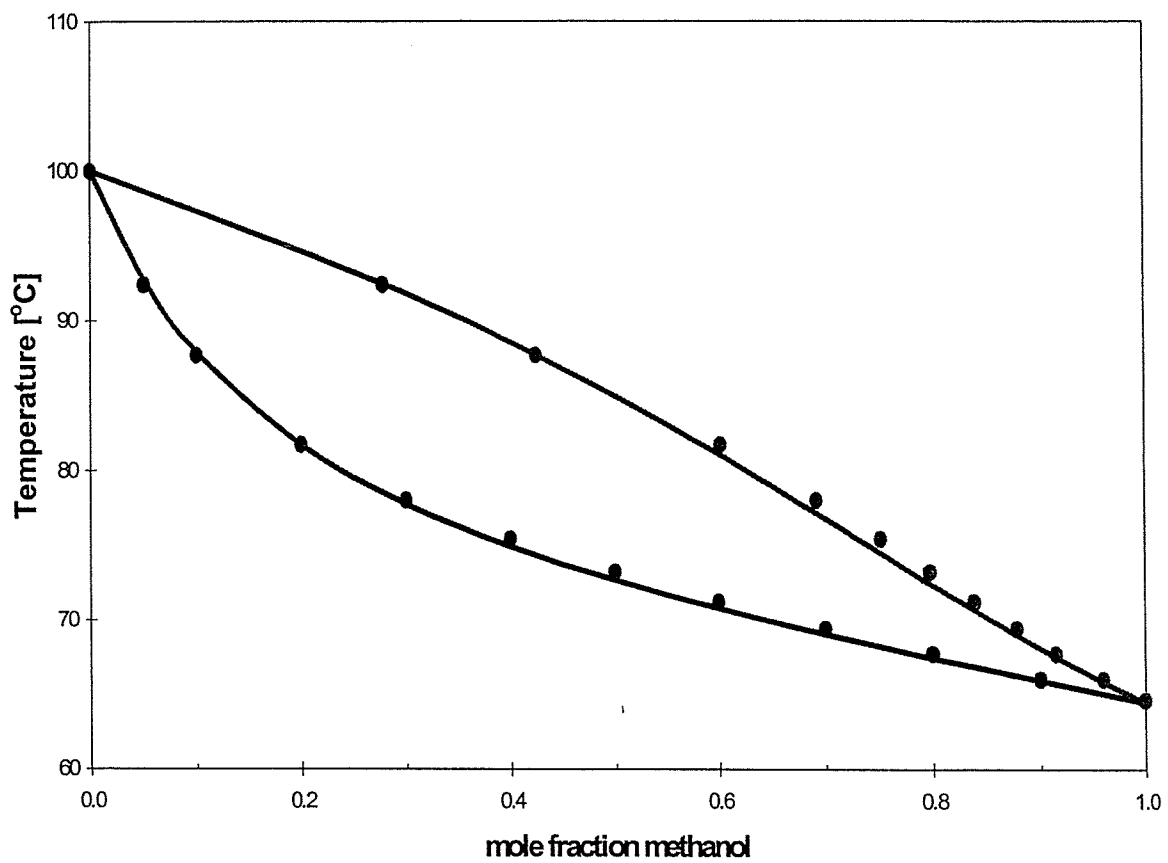
| T [°C] | x[methanol] | y[methanol] |
|--------|-------------|-------------|
| 100.00 | 0           | 0           |
| 92.40  | 0.05        | 0.2780      |
| 87.70  | 0.10        | 0.4250      |
| 81.70  | 0.20        | 0.6020      |
| 78.00  | 0.30        | 0.6920      |
| 75.40  | 0.40        | 0.7520      |
| 73.20  | 0.50        | 0.7980      |
| 71.20  | 0.60        | 0.8380      |
| 69.40  | 0.70        | 0.8780      |
| 67.70  | 0.80        | 0.9150      |
| 66.00  | 0.90        | 0.9600      |
| 64.60  | 1.00        | 1.00        |

The BIPs obtained are:

|                               | Methanol/water | Water/methanol |
|-------------------------------|----------------|----------------|
| UNQUAC<br>ChemCAD<br>database | 95.259         | -10.377        |
| UNQUAC<br>regression          | 30.035         | 40.863         |
| UNIFAC/UNI<br>QAC             | -364.12        | 582.59         |

The results for the data (circles) and the UNQUAC regression are shown in the figure on the following page.

Methanol/Water VLE at 760 Torr



13.10

The following results are from ChemCAD at 760 Torr.

|                            | 25°C  | 35°C  |
|----------------------------|-------|-------|
| Henry's Law                | 816.8 | 676.0 |
| PR with<br>ChemCAD<br>BIPs | 79.0  | 102.1 |

13.11

Different simulators handle supercritical components differently. ChemCAD switches to its database Henry's constants for supercritical components if these Henry's constants are available. Otherwise, it uses the vapor pressure equation, even though it has no meaning physical meaning above the critical point. The vapor pressure equation used is the DIPPR equation, unless only the Antoine's constants are available. If only the Antoine's constants are available and the reduced temperature is greater than 2, the vapor pressure used is the vapor pressure calculated by Antoine's equation at a reduced temperature of 2.

13.12 Simulation of the Ethylbenzene process, Appendix B, Figure B.2.1 and Table B.2.1. The simulation results are shown in the output following this page. The UNIFAC model was used for the phase equilibrium calculations, and the enthalpy calculations used the ChemCAD "Latent Heat" method, which is equivalent to setting heats of mixing to zero. The required minimum input information for this simulation is given below:

1. Flowsheet topology
2. Specification of feed streams (Stream 1 and Stream 2)
3. Equipment specifications
  - H-301 inlet pressures, output temperatures
  - R-301, R-302, R-303, R-304 conversion of key component
  - E-301, E-302, E-303, E-304, E-305 outlet temperatures
  - V-302 pressure
  - T-301, T-302 identification of and recovery of key components
4. Calculational tolerances

13.13 Simulation of the Styrene process, Appendix B, Figure B.3.1 and Table B.3.1. The simulation results are shown in the output following this page. The SRK model was used for the phase equilibrium and the enthalpy calculations. The required minimum input information for this simulation is given below:

1. Flowsheet topology
2. Specification of feed streams (Stream 1 and Stream 4)
3. Equipment specifications
  - H-401 output temperature
  - R-401, R-402 conversion of key component
  - E-401, E-402, E-403, E-404, E-405 outlet temperatures
  - V-401 pressure
  - T-401, T-402 identification of and recovery of key components
  - C-401 outlet pressure, efficiency
4. Calculational tolerances

13.14 Simulation of the Maleic Anhydride process, Appendix B, Figure B.5.1 and Table B.5.1. The simulation results are shown in the output following this page. The "ideal vapor pressure" model (Raoult's Law) was used for the phase equilibrium and "latent heat" model for enthalpies (zero heat of mixing). The required minimum input information for this simulation is given below:

1. Flowsheet topology
2. Specification of feed streams (Stream 1 and Stream 4)
3. Equipment specifications
  - H-601 output temperature
  - R-601 conversion of key component
  - E-601, E-602, E-603 outlet temperatures
  - T-601, T-602 identification of and recovery of key components
  - C-601 outlet pressure, efficiency
4. Calculational tolerances

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers |
|-----------|-------|----------------|
| 1         | FIRE  | 127 -107       |
| 2         | KREA  | 8 -9           |
| 3         | KREA  | 11 -12         |
| 4         | HTXR  | 118 -132       |
| 5         | MIXE  | 7 5 -109       |
| 6         | MIXE  | 9 10 -111      |
| 7         | KREA  | 140 -13        |
| 8         | FLAS  | 119 -126 -145  |
| 9         | SHOR  | 19 -22 -123    |
| 10        | DIVI  | 2 -103 -4      |
| 11        | DIVI  | 103 -5 -10     |
| 12        | MIXE  | 4 107 -6       |
| 13        | VALV  | 135 -119       |
| 14        | PUMP  | 123 -21        |
| 15        | FIRE  | 134 -140       |
| 17        | MIXE  | 12 13 -14      |
| 18        | HTXR  | 111 -11        |
| 19        | CSEP  | 145 -137 -15   |
| 20        | HTXR  | 14 -118        |
| 21        | HTXR  | 132 -135       |
| 22        | DIVI  | 18 -117 -3     |
| 23        | MIXE  | 3 21 -134      |
| 24        | MIXE  | 117 1 -131     |
| 25        | KREA  | 6 -7           |
| 26        | PUMP  | 131 -127       |
| 27        | PUMP  | 17 -18         |
| 28        | MIXE  | 137 126 -16    |
| 29        | TOWR  | 15 -17 -19     |
| 30        | TOWR  | 41 -42 -43     |
| 31        | HTXR  | 109 -8         |

## Stream Connections

| Stream | Equipment |    | Stream | Equipment |    | Stream | Equipment |    |
|--------|-----------|----|--------|-----------|----|--------|-----------|----|
|        | From      | To |        | From      | To |        | From      | To |
| 1      |           | 24 | 15     | 19        | 29 | 117    | 22        | 24 |
| 2      |           | 10 | 16     | 28        |    | 118    | 20        | 4  |
| 3      | 22        | 23 | 17     | 29        | 27 | 119    | 13        | 8  |
| 4      | 10        | 12 | 18     | 27        | 22 | 123    | 9         | 14 |
| 5      | 11        | 5  | 19     | 29        | 9  | 126    | 8         | 28 |
| 6      | 12        | 25 | 21     | 14        | 23 | 127    | 26        | 1  |
| 7      | 25        | 5  | 22     | 9         |    | 131    | 24        | 26 |
| 8      | 31        | 2  | 41     |           | 30 | 132    | 4         | 21 |
| 9      | 2         | 6  | 42     | 30        |    | 134    | 23        | 15 |
| 10     | 11        | 6  | 43     | 30        |    | 135    | 21        | 13 |
| 11     | 18        | 3  | 103    | 10        | 11 | 137    | 19        | 28 |
| 12     | 3         | 17 | 107    | 1         | 12 | 140    | 15        | 7  |
| 13     | 7         | 17 | 109    | 5         | 31 | 145    | 8         | 19 |
| 14     | 17        | 20 | 111    | 6         | 18 |        |           |    |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

Calculation mode : Sequential

Flash algorithm : Normal

Equipment Calculation Sequence

|    |    |    |    |    |    |   |    |    |    |    |    |    |    |   |    |    |   |
|----|----|----|----|----|----|---|----|----|----|----|----|----|----|---|----|----|---|
| 10 | 11 | 30 | 1  | 12 | 25 | 5 | 31 | 2  | 6  | 18 | 3  | 17 | 20 | 4 | 21 | 13 | 8 |
| 19 | 29 | 27 | 22 | 24 | 26 | 9 | 14 | 23 | 15 | 7  | 28 |    |    |   |    |    |   |

Equipment Recycle Sequence

|    |    |    |   |    |    |    |    |   |    |    |   |    |    |   |    |    |    |
|----|----|----|---|----|----|----|----|---|----|----|---|----|----|---|----|----|----|
| 1  | 12 | 25 | 5 | 31 | 2  | 6  | 18 | 3 | 17 | 20 | 4 | 21 | 13 | 8 | 19 | 29 | 27 |
| 22 | 24 | 26 | 9 | 14 | 23 | 15 | 7  |   |    |    |   |    |    |   |    |    |    |

Recycle Cut Streams

127 13

Recycle Convergence Method: Wegstein

|                        |       |                      |      |
|------------------------|-------|----------------------|------|
| Wegstein lower bound   | -5.00 | Wegstein upper bound | 0.00 |
| Acceleration frequency | 4     |                      |      |
| Max. loop iterations   | 200   |                      |      |

Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-004 |
| Temperature | 1.000E-003 |
| Pressure    | 1.000E-003 |
| Enthalpy    | 1.000E-003 |
| Vapor frac. | 1.000E-003 |

Recycle calculation has converged.



Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

| Overall Mass Balance | kmol/h  |         | kg/h      |           |
|----------------------|---------|---------|-----------|-----------|
|                      | Input   | Output  | Input     | Output    |
| Benzene              | 97.169  | 8.683   | 7590.270  | 678.268   |
| Ethylbenzene         | 90.688  | 181.160 | 9628.111  | 19233.250 |
| Ethylene             | 93.000  | 0.502   | 2609.022  | 14.072    |
| 1,4-DiEthBenzene     | 10.424  | 10.437  | 1399.075  | 1400.857  |
| Ethane               | 7.000   | 7.000   | 210.490   | 210.490   |
| Toluene              | 2.000   | 0.000   | 184.283   | 0.003     |
| Propylene            | 0.000   | 2.000   | 0.000     | 84.161    |
| Water                | 0.000   | 0.000   | 0.000     | 0.000     |
| Total                | 300.281 | 209.782 | 21621.249 | 21621.100 |

## COMPONENTS

|   | ID # | Name             | Formula |
|---|------|------------------|---------|
| 1 | 40   | Benzene          | C6H6    |
| 2 | 45   | Ethylbenzene     | C8H10   |
| 3 | 22   | Ethylene         | C2H4    |
| 4 | 383  | 1,4-DiEthBenzene | C10H14  |
| 5 | 3    | Ethane           | C2H6    |
| 6 | 41   | Toluene          | C7H8    |
| 7 | 23   | Propylene        | C3H6    |
| 8 | 62   | Water            | H2O     |

## THERMODYNAMICS

K-value model : UNIFAC  
 No correction for vapor fugacity  
 Enthalpy model : Latent Heat  
 Liquid density : Library

Std vapor rate reference temperature is 0 C.

Atmospheric pressure is 101.3250 kPa.

\* Component ID 22 does not have UNIFAC subgroups.

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

## EQUIPMENT SUMMARIES

## Fired Heater Summary

|                    |            |           |
|--------------------|------------|-----------|
| Equip. No.         | 1          | 15        |
| Name               |            |           |
| Pressure Drop kPa  | 15.0000    | 10.0000   |
| Temperature Out C  | 400.0000   | 500.0000  |
| Heat Absorbed MJ/h | 17295.6367 | 5057.1372 |
| Fuel Usage (SCF)   | 24285.9785 | 7101.0703 |

## Kinetic Reactor Summary

|                             |            |            |          |            |
|-----------------------------|------------|------------|----------|------------|
| Equip. No.                  | 2          | 3          | 7        | 25         |
| Name                        |            |            |          |            |
| Reactor type                | 2          | 2          | 2        | 2          |
| Reaction phase              | 1          | 1          | 1        | 1          |
| Thermal mode                | 2          | 2          | 2        | 2          |
| Pressure Drop kPa           |            |            |          | 50.0000    |
| Tout C                      | 453.8678   | 449.6506   | 497.8943 | 443.4468   |
| Reactor volume m3           | 15.0000    | 18.0000    | 1.0000   | 12.0000    |
| Conversion                  |            |            | 0.9990   |            |
| Key                         | 0          | 0          | 4        | 0          |
| No. of Reactions            | 3          | 3          | 1        | 3          |
| Molar Flow Unit             | 1          | 1          | 1        | 1          |
| Activ. E/H of Rxn Unit      | 7          | 7          | 7        | 7          |
| Volume Unit                 | 1          | 1          | 1        | 1          |
| Time Unit                   | 2          | 2          | 2        | 2          |
| Overall IG Ht of Rxn (MJ/h) | -3438.4460 | -3438.2615 | -10.7212 | -2636.6072 |
| Mass unit                   | 1          | 1          | 1        | 1          |

## Reaction Stoichiometrics and Parameters for unit no. 2

## Reaction 1

|             |             |             |             |             |             |
|-------------|-------------|-------------|-------------|-------------|-------------|
| RateConst = | 1.6700e+006 | Act.E =     | 2.2500e+004 | Hrxn =      | 0.0000e+000 |
| Comp        | Stoich.     | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
| 1           | -1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3           | -1.00e+000  | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2           | 1.00e+000   | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 2

|             |             |             |             |             |             |
|-------------|-------------|-------------|-------------|-------------|-------------|
| RateConst = | 1.0000e+006 | Act.E =     | 2.2500e+004 | Hrxn =      | 0.0000e+000 |
| Comp        | Stoich.     | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
| 2           | -1.00e+000  | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3           | -1.00e+000  | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4           | 1.00e+000   | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 3

|             |             |             |             |             |             |
|-------------|-------------|-------------|-------------|-------------|-------------|
| RateConst = | 3.0000e+008 | Act.E =     | 2.0000e+004 | Hrxn =      | 0.0000e+000 |
| Comp        | Stoich.     | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
| 6           | -1.00e+000  | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3           | -2.00e+000  | 2.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

## EQUIPMENT SUMMARIES

|   |           |             |             |             |             |
|---|-----------|-------------|-------------|-------------|-------------|
| 2 | 1.00e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7 | 1.00e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction Stoichiometrics and Parameters for unit no. 3

## Reaction 1

RateConst = 1.6700e+006 Act.E = 2.2500e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 1    | -1.00e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 2

RateConst = 1.0000e+006 Act.E = 2.2500e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 3

RateConst = 3.0000e+008 Act.E = 2.0000e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | -2.00e+000 | 2.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction Stoichiometrics and Parameters for unit no. 7

## Reaction 1

RateConst = 1.3000e+007 Act.E = 2.5000e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 4    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 1    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction Stoichiometrics and Parameters for unit no. 25

## Reaction 1

RateConst = 1.6700e+006 Act.E = 2.2500e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 1    | -1.00e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 2

RateConst = 1.0000e+006 Act.E = 2.2500e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 3

RateConst = 3.0000e+008 Act.E = 2.0000e+004 Hrxn = 0.0000e+000

| Comp | Stoich. | Exp.factor | AdsorbFac. | AdsorbE | AdsorbExp. |
|------|---------|------------|------------|---------|------------|
|------|---------|------------|------------|---------|------------|

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

## EQUIPMENT SUMMARIES

|   |            |             |             |             |             |
|---|------------|-------------|-------------|-------------|-------------|
| 6 | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3 | -2.00e+000 | 2.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7 | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2 | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Heat Exchanger Summary

| Equip. No.          | 4           | 18         | 20          | 21         |
|---------------------|-------------|------------|-------------|------------|
| Name                |             |            |             |            |
| 1st Stream dp kPa   |             |            | 20.0000     |            |
| 1st Stream T Out C  | 170.0000    | 380.0000   | 280.0000    | 80.0000    |
| Calc Ht Duty MJ/h   | -12355.7969 | -2614.3660 | -10097.7314 | -4939.4751 |
| LMTD Corr Factor    | 1.0000      | 1.0000     | 1.0000      | 1.0000     |
| 1st Stream Pout kPa | 1915.0000   | 1935.0000  | 1915.0000   | 1915.0000  |

|                     |            |
|---------------------|------------|
| Equip. No.          | 31         |
| Name                |            |
| 1st Stream T Out C  | 380.0000   |
| Calc Ht Duty MJ/h   | -1938.1415 |
| LMTD Corr Factor    | 1.0000     |
| 1st Stream Pout kPa | 1935.0000  |

## Mixer Summary

| Equip. No. | 5 | 6 | 12 | 17 |
|------------|---|---|----|----|
| Name       |   |   |    |    |

|            |    |    |    |
|------------|----|----|----|
| Equip. No. | 23 | 24 | 28 |
| Name       |    |    |    |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22  
EQUIPMENT SUMMARIES

## Flash Summary

|                |         |
|----------------|---------|
| Equip. No.     | 8       |
| Name           |         |
| Flash Mode     | 2       |
| Heat duty MJ/h | -0.0003 |

## K values:

|                 |           |
|-----------------|-----------|
| Benzene         | 0.746     |
| Ethylbenzene    | 0.118     |
| Ethylene        | 19521.078 |
| 1,4-DiEthBenzen | 0.019     |
| Ethane          | 77708.516 |
| Toluene         | 0.276     |
| Propylene       | 34.217    |
| Water           | 82.861    |

## Shortcut Distillation Summary

|                          |             |
|--------------------------|-------------|
| Equip. No.               | 9           |
| Name                     |             |
| Mode                     | 2           |
| Light key component      | 2.0000      |
| Light key split          | 0.9900      |
| Heavy key component      | 4.0000      |
| Heavy key split          | 1.0000e-005 |
| R/Rmin                   | 1.2000      |
| Number of stages         | 35.2053     |
| Min. No. of stages       | 14.1985     |
| Feed stage               | 25.4263     |
| Condenser duty MJ/h      | -4973.1953  |
| Reboiler duty MJ/h       | 4975.6084   |
| Colm pressure kPa        | 110.0000    |
| Reflux ratio, minimum    | 0.4662      |
| Calc. Reflux ratio       | 0.5594      |
| Colm pressure drop (kPa) | 10.0000     |

## Divider Summary

|                  |        |        |         |
|------------------|--------|--------|---------|
| Equip. No.       | 10     | 11     | 22      |
| Name             |        |        |         |
| Split based on   | 0      | 0      | 1       |
| Output stream #1 | 0.7000 | 0.5000 |         |
| Output stream #2 | 0.3000 | 0.5000 | 40.0000 |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22

## EQUIPMENT SUMMARIES

## Valve Summary

|                  |          |
|------------------|----------|
| Equip. No.       | 13       |
| Name             |          |
| Pressure out kPa | 110.0000 |

## Pump Summary

|                     |           |            |            |
|---------------------|-----------|------------|------------|
| Equip. No.          | 14        | 26         | 27         |
| Name                |           |            |            |
| Output pressure kPa | 2000.0000 | 2000.0000  | 2000.0000  |
| Efficiency          | 1.0000    | 1.0000     | 1.0000     |
| Calculated power kW | 1.0876    | 11.2529    | 8.6112     |
| Calculated Pout kPa | 2000.0000 | 2000.0000  | 2000.0000  |
| Head m              | 267.8507  | 230.1106   | 237.3644   |
| Vol. flow rate m3/h | 2.0811    | 21.4187    | 16.3906    |
| Mass flow rate kg/h | 1489.5172 | 17938.9434 | 13308.2090 |

## Component Separator Summary

|                 |        |
|-----------------|--------|
| Equip. No.      | 19     |
| Name            |        |
| Component No. 3 | 1.0000 |
| Component No. 5 | 1.0000 |
| Component No. 7 | 1.0000 |

## Tower Rigorous Distillation Summary

|                          |            |            |
|--------------------------|------------|------------|
| Equip. No.               | 29         | 30         |
| Name                     |            |            |
| No. of stages            | 21         | 36         |
| 1st feed stage           | 9          | 26         |
| Top pressure kPa         | 110.0000   | 110.0000   |
| Colm pressure drop (kPa) | 10.0000    | 10.0000    |
| Condenser mode           | 7          | 7          |
| Condenser spec.          | 0.9990     | 0.9500     |
| Cond. comp i             | 1          | 2          |
| Reboiler mode            | 7          | 7          |
| Reboiler spec.           | 0.9900     | 1.0000     |
| Reboiler comp i          | 2          | 4          |
| Iterations               | 100        | 100        |
| Initial flag             | 1          | 1          |
| Calc cond duty MJ/h      | -7256.7793 | -5086.6245 |
| Calc rebr duty MJ/h      | 9129.1543  | 5065.9414  |
| Est. Dist. rate (kmol/h) | 500.0000   | 100.0000   |
| Est. Reflux rate         | 1000.0000  | 200.0000   |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22  
EQUIPMENT SUMMARIES

|                       |           |           |
|-----------------------|-----------|-----------|
| (kmol/h)              |           |           |
| Est. T top C          | 82.0000   | 152.0000  |
| Est. T bottom C       | 145.0000  | 185.0000  |
| Tray type             | 0         | 3         |
| Column diameter m     |           | 1.3716    |
| Tray space m          |           | 0.6096    |
| Thickness (top) m     |           | 0.0016    |
| Thickness (bot) m     |           | 0.0103    |
| No of sections        | 0         | 1         |
| Calc Reflux ratio     | 0.3876    | 0.6617    |
| Calc Reflux mole      | 65.9067   | 57.1168   |
| (kmol/h)              |           |           |
| Calc Reflux mass kg/h | 5158.1973 | 6060.7837 |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22  
FLOW SUMMARIES

| Stream No.          | 1        | 2         | 3         | 4         |
|---------------------|----------|-----------|-----------|-----------|
| Stream Name         |          |           |           |           |
| Temp C              | 25.0000  | 25.0000   | 84.1704   | 25.0000   |
| Pres kPa            | 110.0000 | 2000.0000 | 2000.0000 | 2000.0000 |
| Enth MJ/h           | 4789.5   | 4117.9    | 2292.4    | 1235.4    |
| Vapor mole fraction | 0.00000  | 1.0000    | 0.00000   | 1.0000    |
| Total kmol/h        | 99.0000  | 100.0000  | 40.0000   | 30.0000   |
| Flowrates in kmol/h |          |           |           |           |
| Benzene             | 97.0000  | 0.0000    | 39.7846   | 0.0000    |
| Ethylbenzene        | 0.0000   | 0.0000    | 0.2154    | 0.0000    |
| Ethylene            | 0.0000   | 93.0000   | 0.0000    | 27.9000   |
| 1,4-DiEthBenzene    | 0.0000   | 0.0000    | 0.0000    | 0.0000    |
| Ethane              | 0.0000   | 7.0000    | 0.0000    | 2.1000    |
| Toluene             | 2.0000   | 0.0000    | 0.0000    | 0.0000    |
| Propylene           | 0.0000   | 0.0000    | 0.0000    | 0.0000    |
| Water               | 0.0000   | 0.0000    | 0.0000    | 0.0000    |

| Stream No.          | 5         | 6         | 7         | 8         |
|---------------------|-----------|-----------|-----------|-----------|
| Stream Name         |           |           |           |           |
| Temp C              | 25.0000   | 383.2898  | 443.4468  | 380.0000  |
| Pres kPa            | 2000.0000 | 1985.0000 | 1935.0000 | 1935.0000 |
| Enth MJ/h           | 1441.3    | 30814.    | 30814.    | 30317.    |
| Vapor mole fraction | 1.0000    | 1.0000    | 1.0000    | 1.0000    |
| Total kmol/h        | 35.0000   | 259.0393  | 233.9761  | 268.9761  |
| Flowrates in kmol/h |           |           |           |           |
| Benzene             | 0.0000    | 226.3389  | 203.9239  | 203.9239  |
| Ethylbenzene        | 0.0000    | 0.7004    | 24.1042   | 24.1042   |
| Ethylene            | 32.5500   | 27.9000   | 1.0183    | 33.5683   |
| 1,4-DiEthBenzene    | 0.0000    | 0.0000    | 0.8297    | 0.8297    |
| Ethane              | 2.4500    | 2.1000    | 2.1000    | 4.5500    |
| Toluene             | 0.0000    | 2.0000    | 0.1815    | 0.1815    |
| Propylene           | 0.0000    | 0.0000    | 1.8185    | 1.8185    |
| Water               | 0.0000    | 0.0000    | 0.0000    | 0.0000    |

| Stream No.          | 9         | 10        | 11        | 12        |
|---------------------|-----------|-----------|-----------|-----------|
| Stream Name         |           |           |           |           |
| Temp C              | 453.8678  | 25.0000   | 380.0000  | 449.6506  |
| Pres kPa            | 1935.0000 | 2000.0000 | 1935.0000 | 1935.0000 |
| Enth MJ/h           | 30317.    | 1441.3    | 29144.    | 29144.    |
| Vapor mole fraction | 1.0000    | 1.0000    | 1.0000    | 1.0000    |
| Total kmol/h        | 236.2702  | 35.0000   | 271.2702  | 238.5409  |
| Flowrates in kmol/h |           |           |           |           |
| Benzene             | 174.8614  | 0.0000    | 174.8614  | 148.1625  |
| Ethylbenzene        | 49.8812   | 0.0000    | 49.8812   | 70.5547   |
| Ethylene            | 0.6834    | 32.5500   | 33.2334   | 0.5016    |
| 1,4-DiEthBenzene    | 4.2941    | 0.0000    | 4.2941    | 10.3221   |
| Ethane              | 4.5500    | 2.4500    | 7.0000    | 7.0000    |
| Toluene             | 0.0025    | 0.0000    | 0.0025    | 0.0000    |
| Propylene           | 1.9975    | 0.0000    | 1.9975    | 2.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000    | 0.0000    |



Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22  
FLOW SUMMARIES

| Stream No.          | 13        | 14        | 15       | 16        |
|---------------------|-----------|-----------|----------|-----------|
| Stream Name         |           |           |          |           |
| Temp C              | 497.8943  | 458.4666  | 73.6193  | 73.6193   |
| Pres kPa            | 1990.0000 | 1935.0000 | 110.0000 | 110.0000  |
| Enth MJ/h           | 7082.7    | 36226.    | 8583.7   | 258.23    |
| Vapor mole fraction | 1.0000    | 1.0000    | 0.00000  | 1.0000    |
| Total kmol/h        | 51.2870   | 289.8279  | 271.2577 | 18.5702   |
| Flowrates in kmol/h |           |           |          |           |
| Benzene             | 29.4758   | 177.6382  | 169.2935 | 8.3447    |
| Ethylbenzene        | 21.7397   | 92.2945   | 91.5837  | 0.7108    |
| Ethylene            | 0.0000    | 0.5016    | 0.0000   | 0.5016    |
| 1,4-DiEthBenzene    | 0.0715    | 10.3936   | 10.3804  | 0.0132    |
| Ethane              | 0.0000    | 7.0000    | 0.0000   | 7.0000    |
| Toluene             | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Propylene           | 0.0000    | 2.0000    | 0.0000   | 2.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Stream No.          | 17        | 18        | 19       | 21        |
| Stream Name         |           |           |          |           |
| Temp C              | 82.9686   | 84.1704   | 145.4251 | 185.5938  |
| Pres kPa            | 110.0000  | 2000.0000 | 120.0000 | 2000.0000 |
| Enth MJ/h           | 9714.2    | 9745.2    | 742.05   | -266.83   |
| Vapor mole fraction | 0.00000   | 0.00000   | 0.00000  | 0.00000   |
| Total kmol/h        | 170.0401  | 170.0401  | 101.2176 | 11.2870   |
| Flowrates in kmol/h |           |           |          |           |
| Benzene             | 169.1243  | 169.1243  | 0.1692   | 0.0000    |
| Ethylbenzene        | 0.9158    | 0.9158    | 90.6679  | 0.9067    |
| Ethylene            | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| 1,4-DiEthBenzene    | 0.0000    | 0.0000    | 10.3804  | 10.3803   |
| Ethane              | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Toluene             | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Propylene           | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Stream No.          | 22        | 41        | 42       | 43        |
| Stream Name         |           |           |          |           |
| Temp C              | 139.0379  | 145.4355  | 139.0300 | 170.3402  |
| Pres kPa            | 110.0000  | 120.0000  | 110.0000 | 120.0000  |
| Enth MJ/h           | 1015.2    | 740.82    | 974.96   | -254.89   |
| Vapor mole fraction | 0.00000   | 0.00000   | 0.00000  | 0.00000   |
| Total kmol/h        | 89.9306   | 101.2812  | 86.3230  | 14.9581   |
| Flowrates in kmol/h |           |           |          |           |
| Benzene             | 0.1692    | 0.1691    | 0.1691   | 0.0000    |
| Ethylbenzene        | 89.7613   | 90.6884   | 86.1538  | 4.5345    |
| Ethylene            | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| 1,4-DiEthBenzene    | 0.0001    | 10.4237   | 0.0001   | 10.4236   |
| Ethane              | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Toluene             | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Propylene           | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000   | 0.0000    |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22  
FLOW SUMMARIES

| Stream No.          | 103       | 107       | 109       | 111       |
|---------------------|-----------|-----------|-----------|-----------|
| Stream Name         |           |           |           |           |
| Temp C              | 25.0000   | 400.0000  | 423.2435  | 434.4776  |
| Pres kPa            | 2000.0000 | 1985.0000 | 1935.0000 | 1935.0000 |
| Enth MJ/h           | 2882.5    | 29578.    | 32255.    | 31758.    |
| Vapor mole fraction | 1.0000    | 1.0000    | 1.0000    | 1.0000    |
| Total kmol/h        | 70.0000   | 229.0393  | 268.9761  | 271.2702  |
| Flowrates in kmol/h |           |           |           |           |
| Benzene             | 0.0000    | 226.3389  | 203.9239  | 174.8614  |
| Ethylbenzene        | 0.0000    | 0.7004    | 24.1042   | 49.8812   |
| Ethylene            | 65.1000   | 0.0000    | 33.5683   | 33.2334   |
| 1,4-DiEthBenzene    | 0.0000    | 0.0000    | 0.8297    | 4.2941    |
| Ethane              | 4.9000    | 0.0000    | 4.5500    | 7.0000    |
| Toluene             | 0.0000    | 2.0000    | 0.1815    | 0.0025    |
| Propylene           | 0.0000    | 0.0000    | 1.8185    | 1.9975    |
| Water               | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Stream No.          | 117       | 118       | 119       | 123       |
| Stream Name         |           |           |           |           |
| Temp C              | 84.1704   | 280.0000  | 73.6193   | 184.5116  |
| Pres kPa            | 2000.0000 | 1915.0000 | 110.0000  | 120.0000  |
| Enth MJ/h           | 7452.7    | 26129.    | 8833.4    | -270.75   |
| Vapor mole fraction | 0.00000   | 1.0000    | 0.061950  | 0.00000   |
| Total kmol/h        | 130.0401  | 289.8279  | 289.8279  | 11.2870   |
| Flowrates in kmol/h |           |           |           |           |
| Benzene             | 129.3397  | 177.6382  | 177.6382  | 0.0000    |
| Ethylbenzene        | 0.7004    | 92.2945   | 92.2945   | 0.9067    |
| Ethylene            | 0.0000    | 0.5016    | 0.5016    | 0.0000    |
| 1,4-DiEthBenzene    | 0.0000    | 10.3936   | 10.3936   | 10.3803   |
| Ethane              | 0.0000    | 7.0000    | 7.0000    | 0.0000    |
| Toluene             | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Propylene           | 0.0000    | 2.0000    | 2.0000    | 0.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Stream No.          | 126       | 127       | 131       | 132       |
| Stream Name         |           |           |           |           |
| Temp C              | 73.6193   | 60.5602   | 59.3422   | 170.0000  |
| Pres kPa            | 110.0000  | 2000.0000 | 110.0000  | 1915.0000 |
| Enth MJ/h           | 244.18    | 12283.    | 12242.    | 13773.    |
| Vapor mole fraction | 1.0000    | 0.00000   | 0.00000   | 0.040547  |
| Total kmol/h        | 17.9549   | 229.0401  | 229.0401  | 289.8279  |
| Flowrates in kmol/h |           |           |           |           |
| Benzene             | 8.3447    | 226.3397  | 226.3397  | 177.6382  |
| Ethylbenzene        | 0.7108    | 0.7004    | 0.7004    | 92.2945   |
| Ethylene            | 0.5012    | 0.0000    | 0.0000    | 0.5016    |
| 1,4-DiEthBenzene    | 0.0132    | 0.0000    | 0.0000    | 10.3936   |
| Ethane              | 6.9986    | 0.0000    | 0.0000    | 7.0000    |
| Toluene             | 0.0000    | 2.0000    | 2.0000    | 0.0000    |
| Propylene           | 1.3864    | 0.0000    | 0.0000    | 2.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000    | 0.0000    |

Job Name: Ethylbenzene Date: 12/14/2008 Time: 13:38:22  
FLOW SUMMARIES

|                     |           |           |          |           |
|---------------------|-----------|-----------|----------|-----------|
| Stream No.          | 134       | 135       | 137      | 140       |
| Stream Name         |           |           |          |           |
| Temp C              | 119.8999  | 80.0000   | 73.6193  | 500.0000  |
| Pres kPa            | 2000.0000 | 1915.0000 | 110.0000 | 1990.0000 |
| Enth MJ/h           | 2025.6    | 8833.4    | 14.045   | 7082.7    |
| Vapor mole fraction | 0.00000   | 0.027016  | 1.0000   | 1.0000    |
| Total kmol/h        | 51.2870   | 289.8279  | 0.6153   | 51.2870   |
| Flowrates in kmol/h |           |           |          |           |
| Benzene             | 39.7846   | 177.6382  | 0.0000   | 39.7846   |
| Ethylbenzene        | 1.1221    | 92.2945   | 0.0000   | 1.1221    |
| Ethylene            | 0.0000    | 0.5016    | 0.0004   | 0.0000    |
| 1,4-DiEthBenzene    | 10.3803   | 10.3936   | 0.0000   | 10.3803   |
| Ethane              | 0.0000    | 7.0000    | 0.0014   | 0.0000    |
| Toluene             | 0.0000    | 0.0000    | 0.0000   | 0.0000    |
| Propylene           | 0.0000    | 2.0000    | 0.6135   | 0.0000    |
| Water               | 0.0000    | 0.0000    | 0.0000   | 0.0000    |

|                     |          |
|---------------------|----------|
| Stream No.          | 145      |
| Stream Name         |          |
| Temp C              | 73.6193  |
| Pres kPa            | 110.0000 |
| Enth MJ/h           | 8589.2   |
| Vapor mole fraction | 0.00000  |
| Total kmol/h        | 271.8730 |
| Flowrates in kmol/h |          |
| Benzene             | 169.2935 |
| Ethylbenzene        | 91.5837  |
| Ethylene            | 0.0004   |
| 1,4-DiEthBenzene    | 10.3804  |
| Ethane              | 0.0014   |
| Toluene             | 0.0000   |
| Propylene           | 0.6135   |
| Water               | 0.0000   |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers |
|-----------|-------|----------------|
| 1         | LLVF  | 7 -29 -40 -30  |
| 2         | MIXE  | 25 1 -2        |
| 4         | HTXR  | 4 -5           |
| 5         | HTXR  | 2 38 -3 -39    |
| 6         | VALV  | 5 -8           |
| 7         | MIXE  | 3 8 -9         |
| 8         | KREA  | 9 -10          |
| 9         | HTXR  | 10 6 -11 -31   |
| 10        | KREA  | 11 -12         |
| 11        | HTXR  | 12 32 -13 -33  |
| 12        | HTXR  | 13 34 -14 -35  |
| 13        | HTXR  | 14 36 -15 -37  |
| 14        | CSEP  | 15 -50 -18     |
| 15        | CSEP  | 50 -16 -19     |
| 16        | COMP  | 16 -17         |
| 17        | VALV  | 19 -20         |
| 18        | TOWR  | 20 -21 -22     |
| 19        | TOWR  | 22 -23 -24     |
| 20        | PUMP  | 23 -25         |
| 35        | PUMP  | 21 -26         |
| 36        | PUMP  | 24 -27         |
| 37        | PUMP  | 18 -28         |

## Stream Connections

| Stream | Equipment |    | Stream | Equipment |    | Stream | Equipment |    |
|--------|-----------|----|--------|-----------|----|--------|-----------|----|
|        | From      | To |        | From      | To |        | From      | To |
| 1      |           | 2  | 15     | 13        | 14 | 29     | 1         |    |
| 2      | 2         | 5  | 16     | 15        | 16 | 30     | 1         |    |
| 3      | 5         | 7  | 17     | 16        |    | 31     | 9         |    |
| 4      |           | 4  | 18     | 14        | 37 | 32     |           | 11 |
| 5      | 4         | 6  | 19     | 15        | 17 | 33     | 11        |    |
| 6      |           | 9  | 20     | 17        | 18 | 34     |           | 12 |
| 7      |           | 1  | 21     | 18        | 35 | 35     | 12        |    |
| 8      | 6         | 7  | 22     | 18        | 19 | 36     |           | 13 |
| 9      | 7         | 8  | 23     | 19        | 20 | 37     | 13        |    |
| 10     | 8         | 9  | 24     | 19        | 36 | 38     |           | 5  |
| 11     | 9         | 10 | 25     | 20        | 2  | 39     | 5         |    |
| 12     | 10        | 11 | 26     | 35        |    | 40     | 1         |    |
| 13     | 11        | 12 | 27     | 36        |    | 50     | 14        | 15 |
| 14     | 12        | 13 | 28     | 37        |    |        |           |    |

Calculation mode : Sequential

Flash algorithm : Normal

## Equipment Calculation Sequence

|    |    |    |    |   |   |   |   |    |    |    |    |    |    |    |    |    |    |
|----|----|----|----|---|---|---|---|----|----|----|----|----|----|----|----|----|----|
| 1  | 4  | 6  | 2  | 5 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 17 | 18 | 19 | 20 |
| 16 | 35 | 36 | 37 |   |   |   |   |    |    |    |    |    |    |    |    |    |    |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58

## Equipment Recycle Sequence

2 5 7 8 9 10 11 12 13 14 15 17 18 19 20

## Recycle Cut Streams

25

## Recycle Convergence Method: Wegstein

|                        |       |                      |      |
|------------------------|-------|----------------------|------|
| Wegstein lower bound   | -5.00 | Wegstein upper bound | 0.00 |
| Acceleration frequency | 4     |                      |      |
| Max. loop iterations   | 100   |                      |      |

## Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-003 |
| Temperature | 1.000E-003 |
| Pressure    | 1.000E-003 |
| Enthalpy    | 1.000E-003 |
| Vapor frac. | 1.000E-003 |

Recycle calculation has converged.

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58

| Overall Mass Balance | kmol/h     |            | kg/h         |              |
|----------------------|------------|------------|--------------|--------------|
|                      | Input      | Output     | Input        | Output       |
| Water                | 551241.000 | 551241.000 | 9930598.077  | 9930598.077  |
| Ethylbenzene         | 528.236    | 351.970    | 56081.208    | 37367.607    |
| Styrene              | 121.251    | 240.576    | 12628.489    | 25056.488    |
| Hydrogen             | 89.004     | 177.052    | 179.414      | 356.902      |
| Benzene              | 28.686     | 54.351     | 2240.765     | 4245.613     |
| Toluene              | 34.967     | 66.246     | 3221.900     | 6103.988     |
| Ethylene             | 25.086     | 50.751     | 703.758      | 1423.781     |
| Methane              | 31.035     | 62.314     | 497.900      | 999.709      |
| Total                | 552099.000 | 552244.000 | 10006153.335 | 10006151.521 |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58

## COMPONENTS

|   | ID # | Name         | Formula |
|---|------|--------------|---------|
| 1 | 62   | Water        | H2O     |
| 2 | 45   | Ethylbenzene | C8H10   |
| 3 | 178  | Styrene      | C8H8    |
| 4 | 1    | Hydrogen     | H2      |
| 5 | 40   | Benzene      | C6H6    |
| 6 | 41   | Toluene      | C7H8    |
| 7 | 22   | Ethylene     | C2H4    |
| 8 | 2    | Methane      | CH4     |

## THERMODYNAMICS

K-value model : SRK  
Water/Hydrocarbon immiscible  
Enthalpy model : SRK  
Liquid density : Library

Std vapor rate reference temperature is 0 C.

Atmospheric pressure is 101.3250 kPa.

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
EQUIPMENT SUMMARIES

## Three Phase Flash Summary

|            |         |
|------------|---------|
| Equip. No. | 1       |
| Name       |         |
| Flash Mode | 3       |
| Param1     | 75.0000 |

## Mixer Summary

|                     |          |          |
|---------------------|----------|----------|
| Equip. No.          | 2        | 7        |
| Name                |          |          |
| Output Pressure kPa | 200.0001 | 170.0001 |

## Heat Exchanger Summary

|                     |             |            |            |             |
|---------------------|-------------|------------|------------|-------------|
| Equip. No.          | 4           | 5          | 9          | 11          |
| Name                |             |            |            |             |
| 1st Stream dp kPa   | 35.0000     | 20.0000    | 15.0000    | 15.0000     |
| 1st Stream T Out C  | 802.0000    | 225.0000   | 650.0000   | 270.0000    |
| 2nd Stream T Out C  |             | 253.0000   | 700.0000   |             |
| 2nd Stream VF Out   |             |            |            | 1.0000      |
| Calc Ht Duty MJ/h   | 201698.6250 | 29913.7852 | 19014.2578 | 163175.7813 |
| LMTD (End points) C |             | 69.2964    | 117.4589   | 270.2108    |
| LMTD Corr Factor    | 1.0000      | 1.0000     | 1.0000     | 1.0000      |
| Utility Option:     | 0           | 1          | 1          | 1           |
| 1st Stream Pout kPa | 565.0002    | 180.0001   | 145.0001   | 120.0001    |
| 2nd Stream Pout kPa |             | 4237.0020  | 571.0001   | 4237.0020   |

|                     |            |             |
|---------------------|------------|-------------|
| Equip. No.          | 12         | 13          |
| Name                |            |             |
| 1st Stream dp kPa   | 15.0000    | 15.0000     |
| 1st Stream T Out C  | 180.0000   | 65.0000     |
| 2nd Stream T Out C  |            | 40.0000     |
| 2nd Stream VF Out   | 1.0000     |             |
| Calc Ht Duty MJ/h   | 35230.5078 | 396136.0625 |
| LMTD (End points) C | 100.1799   | 75.7415     |
| LMTD Corr Factor    | 1.0000     | 1.0000      |
| Utility Option:     | 1          | 1           |
| 1st Stream Pout kPa | 105.0001   | 90.0001     |
| 2nd Stream Pout kPa | 600.0002   | 400.0002    |



Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
EQUIPMENT SUMMARIES

## Valve Summary

|                  |          |         |
|------------------|----------|---------|
| Equip. No.       | 6        | 17      |
| Name             |          |         |
| Pressure out kPa | 180.0001 | 60.0000 |

## Kinetic Reactor Summary

|                             |            |           |
|-----------------------------|------------|-----------|
| Equip. No.                  | 8          | 10        |
| Name                        |            |           |
| Reactor type                | 2          | 2         |
| Reaction phase              | 1          | 1         |
| Thermal mode                | 2          | 2         |
| Pressure Drop kPa           | 10.0000    | 10.0000   |
| Tout C                      | 609.9968   | 640.2557  |
| Reactor volume m3           | 1000.0378  | 700.0264  |
| Concentration Flag          | 1          | 1         |
| Conversion                  | 0.5000     | 0.5000    |
| Key                         | 2          | 2         |
| No. of Reactions            | 4          | 4         |
| Molar Flow Unit             | 2          | 2         |
| Activ. E/H of Rxn Unit      | 6          | 6         |
| Volume Unit                 | 1          | 1         |
| Time Unit                   | 2          | 2         |
| Overall IG Ht of Rxn (MJ/h) | 10382.5908 | 4624.7280 |

## Reaction Stoichiometrics and Parameters for unit no. 8

## Reaction 1

RateConst = 4.2400e+006 Act.E = 2.1708e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 2

RateConst = 7.5500e-001 Act.E = -7.8040e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 3    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 3

RateConst = 7.2150e+008 Act.E = 4.9675e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 4

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
EQUIPMENT SUMMARIES

RateConst = 1.7230e+003 Act.E = 2.6857e+004 Hrxn = 0.0000e+000  

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 6    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

#### Reaction Stoichiometrics and Parameters for unit no. 10

Reaction 1  
RateConst = 4.2400e+006 Act.E = 2.1708e+004 Hrxn = 0.0000e+000  

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 3    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

Reaction 2  
RateConst = 7.5500e-001 Act.E = -7.8040e+003 Hrxn = 0.0000e+000  

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 3    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

Reaction 3  
RateConst = 7.2100e+008 Act.E = 4.9675e+004 Hrxn = 0.0000e+000  

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

Reaction 4  
RateConst = 1.7230e+003 Act.E = 2.6857e+004 Hrxn = 0.0000e+000  

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 2    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 6    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

#### Component Separator Summary

| Equip. No.       | 14      | 15      |
|------------------|---------|---------|
| Name             |         |         |
| Top Temp Spec    | 65.0000 | 65.0000 |
| Bottom Temp Spec | 65.0000 | 65.0000 |
| Press Drop kPa   | 15.0000 |         |
| Component No. 1  | 0.0089  | 1.0000  |
| Component No. 2  | 1.0000  |         |
| Component No. 3  | 1.0000  |         |
| Component No. 4  | 1.0000  | 1.0000  |
| Component No. 5  | 1.0000  |         |
| Component No. 6  | 1.0000  |         |
| Component No. 7  | 1.0000  | 1.0000  |
| Component No. 8  | 1.0000  | 1.0000  |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58

Compressor Summary

|                               |          |
|-------------------------------|----------|
| Equip. No.                    | 16       |
| Name                          |          |
| Pressure out kPa              | 240.0001 |
| Type of Compressor            | 1        |
| Efficiency                    | 0.8000   |
| Actual power kW               | 282.9636 |
| Cp/Cv                         | 1.3282   |
| Theoretical power kW          | 226.3709 |
| Ideal Cp/Cv                   | 1.3209   |
| Calc Pout kPa                 | 240.0001 |
| Calc. mass flowrate<br>(kg/h) | 2686     |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
EQUIPMENT SUMMARIES

## Tower Rigorous Distillation Summary

|                           |             |              |
|---------------------------|-------------|--------------|
| Equip. No.                | 18          | 19           |
| Name                      |             |              |
| No. of stages             | 25          | 80           |
| 1st feed stage            | 14          | 25           |
| Top pressure kPa          | 40.0000     | 40.0000      |
| Colm pressure drop (kPa)  | 20.0000     | 20.0000      |
| Condenser mode            | 7           | 7            |
| Condenser spec.           | 0.9900      | 0.9990       |
| Cond. comp i              | 6           | 2            |
| Reboiler mode             | 7           | 7            |
| Reboiler spec.            | 0.9900      | 0.9900       |
| Reboiler comp i           | 2           | 3            |
| Calc cond duty MJ/h       | -28362.5488 | -345656.0625 |
| Calc rebr duty MJ/h       | 33415.6992  | 344720.0000  |
| Est. Dist. rate (kmol/h)  | 300.0000    | 400.0000     |
| Est. Reflux rate (kmol/h) | 780.0000    | 3200.0000    |
| Est. T top C              | 106.0000    | 128.0000     |
| Est. T bottom C           | 146.0000    | 150.0000     |
| Calc Reflux ratio         | 11.5209     | 25.8749      |
| Calc Reflux mole (kmol/h) | 737.8981    | 8741.8809    |
| Calc Reflux mass kg/h     | 64115.9258  | 927916.0000  |

## Pump Summary

|                     |            |           |            |             |
|---------------------|------------|-----------|------------|-------------|
| Equip. No.          | 20         | 35        | 36         | 37          |
| Name                |            |           |            |             |
| Output pressure kPa | 210.0001   | 200.0001  | 200.0001   | 200.0001    |
| Efficiency          | 0.8000     | 0.8000    | 0.8000     | 0.8000      |
| Calculated power kW | 2.6776     | 0.3739    | 0.7455     | 6.3287      |
| Calculated Pout kPa | 210.0001   | 200.0001  | 200.0001   | 200.0001    |
| Head m              | 21.9120    | 19.7192   | 17.5688    | 13.0031     |
| Vol. flow rate m3/h | 45.3298    | 6.7262    | 15.3248    | 145.7092    |
| Mass flow rate kg/h | 35861.6133 | 5565.1875 | 12452.6650 | 142833.7188 |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
FLOW SUMMARIES

| Stream No.          | 1        | 2        | 3        | 4            |
|---------------------|----------|----------|----------|--------------|
| Stream Name         |          |          |          | lps feed     |
| Temp C              | 136.0000 | 116.2373 | 225.0000 | 158.9001     |
| Pres kPa            | 210.0001 | 200.0001 | 180.0001 | 600.0002     |
| Enth MJ/h           | 2007.7   | 3312.3   | 33226.   | -1.9062E+006 |
| Vapor mole fraction | 0.00000  | 0.00000  | 1.0000   | 1.0000       |
| Total kmol/h        | 183.6000 | 521.4581 | 521.4581 | 8000.0000    |
| Flowrates in kmol/h |          |          |          |              |
| Water               | 0.0000   | 0.0000   | 0.0000   | 8000.0000    |
| Ethylbenzene        | 180.0000 | 516.3207 | 516.3207 | 0.0000       |
| Styrene             | 0.0000   | 1.2031   | 1.2031   | 0.0000       |
| Hydrogen            | 0.0000   | 0.0000   | 0.0000   | 0.0000       |
| Benzene             | 1.8000   | 1.8000   | 1.8000   | 0.0000       |
| Toluene             | 1.8000   | 2.1342   | 2.1342   | 0.0000       |
| Ethylene            | 0.0000   | 0.0000   | 0.0000   | 0.0000       |
| Methane             | 0.0000   | 0.0000   | 0.0000   | 0.0000       |

| Stream No.          | 5            | 6            | 7            | 8            |
|---------------------|--------------|--------------|--------------|--------------|
| Stream Name         |              |              |              |              |
| Temp C              | 802.0000     | 800.0000     | 65.0000      | 801.2046     |
| Pres kPa            | 565.0002     | 571.0001     | 90.0001      | 180.0001     |
| Enth MJ/h           | -1.7045E+006 | -9.7846E+005 | -2.2469E+006 | -1.7045E+006 |
| Vapor mole fraction | 1.0000       | 1.0000       | 0.027598     | 1.0000       |
| Total kmol/h        | 8000.0000    | 4590.4279    | 8674.6644    | 8000.0000    |
| Flowrates in kmol/h |              |              |              |              |
| Water               | 8000.0000    | 4590.4279    | 8000.0000    | 8000.0000    |
| Ethylbenzene        | 0.0000       | 0.0000       | 348.2358     | 0.0000       |
| Styrene             | 0.0000       | 0.0000       | 121.2506     | 0.0000       |
| Hydrogen            | 0.0000       | 0.0000       | 89.0037      | 0.0000       |
| Benzene             | 0.0000       | 0.0000       | 26.8858      | 0.0000       |
| Toluene             | 0.0000       | 0.0000       | 33.1671      | 0.0000       |
| Ethylene            | 0.0000       | 0.0000       | 25.0858      | 0.0000       |
| Methane             | 0.0000       | 0.0000       | 31.0353      | 0.0000       |

| Stream No.          | 9            | 10           | 11           | 12           |
|---------------------|--------------|--------------|--------------|--------------|
| Stream Name         |              |              |              |              |
| Temp C              | 632.7648     | 609.9968     | 650.0000     | 640.2558     |
| Pres kPa            | 170.0001     | 160.0001     | 145.0001     | 135.0001     |
| Enth MJ/h           | -1.6713E+006 | -1.6713E+006 | -1.6523E+006 | -1.6523E+006 |
| Vapor mole fraction | 1.0000       | 1.0000       | 1.0000       | 1.0000       |
| Total kmol/h        | 8521.4582    | 8618.8442    | 8618.8442    | 8666.4510    |
| Flowrates in kmol/h |              |              |              |              |
| Water               | 8000.0000    | 8000.0000    | 8000.0000    | 8000.0000    |
| Ethylbenzene        | 516.3207     | 402.1696     | 402.1696     | 340.0488     |
| Styrene             | 1.2031       | 87.1901      | 87.1901      | 120.5307     |
| Hydrogen            | 0.0000       | 69.2212      | 69.2212      | 88.0486      |
| Benzene             | 1.8000       | 13.1984      | 13.1984      | 27.4656      |
| Toluene             | 2.1342       | 18.9001      | 18.9001      | 33.4132      |
| Ethylene            | 0.0000       | 11.3984      | 11.3984      | 25.6656      |
| Methane             | 0.0000       | 16.7659      | 16.7659      | 31.2790      |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
FLOW SUMMARIES

| Stream No.          | 13           | 14           | 15           | 16       |
|---------------------|--------------|--------------|--------------|----------|
| Stream Name         |              |              |              |          |
| Temp C              | 270.0000     | 180.0000     | 65.0000      | 65.0000  |
| Pres kPa            | 120.0001     | 105.0001     | 90.0001      | 75.0001  |
| Enth MJ/h           | -1.8155E+006 | -1.8507E+006 | -2.2468E+006 | -17998.  |
| Vapor mole fraction | 1.0000       | 1.0000       | 0.027633     | 1.0000   |
| Total kmol/h        | 8666.4510    | 8666.4510    | 8666.4510    | 216.3932 |
| Flowrates in kmol/h |              |              |              |          |
| Water               | 8000.0000    | 8000.0000    | 8000.0000    | 71.4000  |
| Ethylbenzene        | 340.0488     | 340.0488     | 340.0488     | 0.0000   |
| Styrene             | 120.5307     | 120.5307     | 120.5307     | 0.0000   |
| Hydrogen            | 88.0486      | 88.0486      | 88.0486      | 88.0486  |
| Benzene             | 27.4656      | 27.4656      | 27.4656      | 0.0000   |
| Toluene             | 33.4132      | 33.4132      | 33.4132      | 0.0000   |
| Ethylene            | 25.6656      | 25.6656      | 25.6656      | 25.6656  |
| Methane             | 31.2790      | 31.2790      | 31.2790      | 31.2790  |

| Stream No.          | 17       | 18           | 19       | 20       |
|---------------------|----------|--------------|----------|----------|
| Stream Name         |          |              |          |          |
| Temp C              | 198.1571 | 65.0000      | 65.0000  | 65.0077  |
| Pres kPa            | 240.0001 | 75.0001      | 75.0001  | 60.0000  |
| Enth MJ/h           | -16980.  | -2.2438E+006 | 14111.   | 14111.   |
| Vapor mole fraction | 1.0000   | 0.00000      | 0.00000  | 0.00000  |
| Total kmol/h        | 216.3932 | 7928.5997    | 521.4584 | 521.4584 |
| Flowrates in kmol/h |          |              |          |          |
| Water               | 71.4000  | 7928.5997    | 0.0000   | 0.0000   |
| Ethylbenzene        | 0.0000   | 0.0000       | 340.0488 | 340.0488 |
| Styrene             | 0.0000   | 0.0000       | 120.5307 | 120.5307 |
| Hydrogen            | 88.0486  | 0.0000       | 0.0000   | 0.0000   |
| Benzene             | 0.0000   | 0.0000       | 27.4656  | 27.4656  |
| Toluene             | 0.0000   | 0.0000       | 33.4132  | 33.4132  |
| Ethylene            | 25.6656  | 0.0000       | 0.0000   | 0.0000   |
| Methane             | 31.2790  | 0.0000       | 0.0000   | 0.0000   |

| Stream No.          | 21      | 22       | 23       | 24       |
|---------------------|---------|----------|----------|----------|
| Stream Name         |         |          |          |          |
| Temp C              | 66.1419 | 119.5297 | 105.1237 | 124.5388 |
| Pres kPa            | 40.0000 | 60.0000  | 40.0000  | 60.0000  |
| Enth MJ/h           | 2115.5  | 17048.   | 1295.0   | 14822.   |
| Vapor mole fraction | 0.00000 | 0.00000  | 0.00000  | 0.00000  |
| Total kmol/h        | 64.0487 | 457.4101 | 337.8520 | 119.5560 |
| Flowrates in kmol/h |         |          |          |          |
| Water               | 0.0000  | 0.0000   | 0.0000   | 0.0000   |
| Ethylbenzene        | 3.4006  | 336.6483 | 336.3146 | 0.3337   |
| Styrene             | 0.1033  | 120.4275 | 1.2031   | 119.2223 |
| Hydrogen            | 0.0000  | 0.0000   | 0.0000   | 0.0000   |
| Benzene             | 27.4657 | 0.0000   | 0.0000   | 0.0000   |
| Toluene             | 33.0791 | 0.3342   | 0.3342   | 0.0000   |
| Ethylene            | 0.0000  | 0.0000   | 0.0000   | 0.0000   |
| Methane             | 0.0000  | 0.0000   | 0.0000   | 0.0000   |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58

## FLOW SUMMARIES

| Stream No.          | 25           | 26           | 27           | 28           |
|---------------------|--------------|--------------|--------------|--------------|
| Stream Name         |              |              |              |              |
| Temp C              | 105.1895     | 66.2044      | 124.5951     | 65.0379      |
| Pres kPa            | 210.0001     | 200.0001     | 200.0001     | 200.0001     |
| Enth MJ/h           | 1304.6       | 2116.8       | 14825.       | -2.2438E+006 |
| Vapor mole fraction | 0.00000      | 0.00000      | 0.00000      | 0.00000      |
| Total kmol/h        | 337.8520     | 64.0487      | 119.5560     | 7928.5997    |
| Flowrates in kmol/h |              |              |              |              |
| Water               | 0.0000       | 0.0000       | 0.0000       | 7928.5997    |
| Ethylbenzene        | 336.3146     | 3.4006       | 0.3337       | 0.0000       |
| Styrene             | 1.2031       | 0.1033       | 119.2223     | 0.0000       |
| Hydrogen            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Benzene             | 0.0000       | 27.4657      | 0.0000       | 0.0000       |
| Toluene             | 0.3342       | 33.0791      | 0.0000       | 0.0000       |
| Ethylene            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Methane             | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Stream No.          | 29           | 30           | 31           | 32           |
| Stream Name         |              |              |              | bfw-hps      |
| Temp C              | 63.4602      | 63.4602      | 700.0000     | 90.0000      |
| Pres kPa            | 75.0000      | 75.0000      | 571.0001     | 4237.0017    |
| Enth MJ/h           | -18276.      | -2.2405E+006 | -9.9747E+005 | -1.0521E+006 |
| Vapor mole fraction | 1.0000       | 0.00000      | 1.0000       | 0.00000      |
| Total kmol/h        | 266.2101     | 7914.5552    | 4590.4279    | 3742.6837    |
| Flowrates in kmol/h |              |              |              |              |
| Water               | 83.0414      | 7913.8349    | 4590.4279    | 3742.6815    |
| Ethylbenzene        | 20.2240      | 0.2604       | 0.0000       | 0.0000       |
| Styrene             | 6.0709       | 0.1590       | 0.0000       | 0.0000       |
| Hydrogen            | 88.9275      | 0.0198       | 0.0000       | 0.0000       |
| Benzene             | 7.9881       | 0.1837       | 0.0000       | 0.0000       |
| Toluene             | 4.3474       | 0.0609       | 0.0000       | 0.0000       |
| Ethylene            | 24.7121      | 0.0268       | 0.0000       | 0.0000       |
| Methane             | 30.8988      | 0.0107       | 0.0000       | 0.0000       |
| Stream No.          | 33           | 34           | 35           | 36           |
| Stream Name         |              | bfw-lps      |              | cw           |
| Temp C              | 253.7945     | 90.0000      | 158.9001     | 30.0000      |
| Pres kPa            | 4237.0017    | 600.0002     | 600.0002     | 400.0002     |
| Enth MJ/h           | -8.8893E+005 | -2.3122E+005 | -1.9599E+005 | -1.4999E+008 |
| Vapor mole fraction | 1.0000       | 0.00000      | 1.0000       | 0.00000      |
| Total kmol/h        | 3742.6837    | 822.5403     | 822.5403     | 525103.0000  |
| Flowrates in kmol/h |              |              |              |              |
| Water               | 3742.6815    | 822.5393     | 822.5393     | 525103.0000  |
| Ethylbenzene        | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Styrene             | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Hydrogen            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Benzene             | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Toluene             | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Ethylene            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Methane             | 0.0000       | 0.0000       | 0.0000       | 0.0000       |

Job Name: styrene\_JAS\_2008 Date: 12/14/2008 Time: 15:25:58  
FLOW SUMMARIES

|                     |              |              |              |          |
|---------------------|--------------|--------------|--------------|----------|
| Stream No.          | 37           | 38           | 39           | 40       |
| Stream Name         |              | hps          |              |          |
| Temp C              | 40.0000      | 253.7945     | 253.0000     | 63.4602  |
| Pres kPa            | 400.0002     | 4237.0017    | 4237.0017    | 75.0000  |
| Enth MJ/h           | -1.4960E+008 | -2.3330E+005 | -2.6321E+005 | 11962.   |
| Vapor mole fraction | 0.00000      | 1.0000       | 0.00000      | 0.00000  |
| Total kmol/h        | 525103.0000  | 982.2584     | 982.2584     | 493.8982 |
| Flowrates in kmol/h |              |              |              |          |
| Water               | 525103.0000  | 982.2587     | 982.2587     | 3.1242   |
| Ethylbenzene        | 0.0000       | 0.0000       | 0.0000       | 327.7514 |
| Styrene             | 0.0000       | 0.0000       | 0.0000       | 115.0207 |
| Hydrogen            | 0.0000       | 0.0000       | 0.0000       | 0.0564   |
| Benzene             | 0.0000       | 0.0000       | 0.0000       | 18.7141  |
| Toluene             | 0.0000       | 0.0000       | 0.0000       | 28.7588  |
| Ethylene            | 0.0000       | 0.0000       | 0.0000       | 0.3469   |
| Methane             | 0.0000       | 0.0000       | 0.0000       | 0.1258   |

|                     |          |
|---------------------|----------|
| Stream No.          | 50       |
| Stream Name         |          |
| Temp C              | 65.0000  |
| Pres kPa            | 75.0001  |
| Enth MJ/h           | -2516.7  |
| Vapor mole fraction | 0.34226  |
| Total kmol/h        | 737.8516 |
| Flowrates in kmol/h |          |
| Water               | 71.4000  |
| Ethylbenzene        | 340.0488 |
| Styrene             | 120.5307 |
| Hydrogen            | 88.0486  |
| Benzene             | 27.4656  |
| Toluene             | 33.4132  |
| Ethylene            | 25.6656  |
| Methane             | 31.2790  |



Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:41

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers |
|-----------|-------|----------------|
| 1         | TOWR  | 9 8 -103 -102  |
| 2         | KREA  | 6 15 -7 -16    |
| 3         | HTXR  | 1 -110         |
| 4         | MIXE  | 5 110 -100     |
| 5         | FIRE  | 100 -6         |
| 6         | HTXR  | 7 -8           |
| 7         | COMP  | 4 -5           |
| 8         | CSEP  | 102 -104 -105  |
| 9         | SHOR  | 11 -13 -14     |
| 10        | MIXE  | 104 103 -12    |
| 11        | REAC  | 105 -11        |
| 12        | MIXE  | 14 10 -9       |
| 13        | KREA  | 91 -93         |
| 14        | MIXE  | 92 93 -3       |
| 15        | DIVI  | 2 -91 -92      |

## Stream Connections

| Stream | Equipment |    | Stream | Equipment |    | Stream | Equipment |    |
|--------|-----------|----|--------|-----------|----|--------|-----------|----|
|        | From      | To |        | From      | To |        | From      | To |
| 1      |           | 3  | 10     |           | 12 | 93     | 13        | 14 |
| 2      |           | 15 | 11     | 11        | 9  | 100    | 4         | 5  |
| 3      | 14        |    | 12     | 10        |    | 102    | 1         | 8  |
| 4      |           | 7  | 13     | 9         |    | 103    | 1         | 10 |
| 5      | 7         | 4  | 14     | 9         | 12 | 104    | 8         | 10 |
| 6      | 5         | 2  | 15     |           | 2  | 105    | 8         | 11 |
| 7      | 2         | 6  | 16     | 2         |    | 110    | 3         | 4  |
| 8      | 6         | 1  | 91     | 15        | 13 |        |           |    |
| 9      | 12        | 1  | 92     | 15        | 14 |        |           |    |

Calculation mode : Sequential

Flash algorithm : Normal

## Equipment Calculation Sequence

3 7 15 4 5 13 14 2 6 1 8 11 9 12 10

## Equipment Recycle Sequence

1 8 11 9 12

## Recycle Cut Streams

9

Recycle Convergence Method: Direct Substitution

Max. loop iterations 150

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:41

### Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-003 |
| Temperature | 1.000E-003 |
| Pressure    | 1.000E-003 |
| Enthalpy    | 1.000E-003 |
| Vapor frac. | 1.000E-003 |

Recycle calculation has converged.

| Overall Mass Balance | kmol/h    |           | kg/h       |            |
|----------------------|-----------|-----------|------------|------------|
|                      | Input     | Output    | Input      | Output     |
| Maleic Anhydride     | 0.000     | 47.447    | 0.000      | 4652.532   |
| DibutylPhthalate     | 0.110     | 0.096     | 30.618     | 26.633     |
| Nitrogen             | 4410.000  | 4410.000  | 123541.739 | 123541.727 |
| Water                | 0.000     | 181.576   | 0.000      | 3271.083   |
| Oxygen               | 1170.000  | 739.222   | 37438.830  | 23654.369  |
| Benzene              | 84.600    | 6.634     | 6608.444   | 518.187    |
| Quinone              | 0.000     | 1.438     | 0.000      | 155.468    |
| Carbon Dioxide       | 0.000     | 265.381   | 0.000      | 11679.422  |
| Maleic Acid          | 0.000     | 1.000     | 0.000      | 116.070    |
| Sodium Nitrite       | 2065.630  | 2065.630  | 142518.161 | 142518.161 |
| Sodium Nitrate       | 2934.369  | 2934.369  | 249406.739 | 249406.739 |
| Total                | 10664.710 | 10652.793 | 559545.000 | 559540.378 |

### COMPONENTS

|    | ID # | Name             | Formula  |
|----|------|------------------|----------|
| 1  | 272  | Maleic Anhydride | C4H2O3   |
| 2  | 910  | DibutylPhthalate | C16H22O4 |
| 3  | 46   | Nitrogen         | N2       |
| 4  | 62   | Water            | H2O      |
| 5  | 47   | Oxygen           | O2       |
| 6  | 40   | Benzene          | C6H6     |
| 7  | 1327 | Quinone          | C6H4O2   |
| 8  | 49   | Carbon Dioxide   | CO2      |
| 9  | 445  | Maleic Acid      | C4H4O4   |
| 10 | 975  | Sodium Nitrite   | NNaO2    |
| 11 | 976  | Sodium Nitrate   | NNaO3    |

### THERMODYNAMICS

K-value model : Ideal Vapor Pressure  
 Enthalpy model : Latent Heat  
 Liquid density : Library

Equip # 1 K-value : VAP  
 Std vapor rate reference temperature is 0 C.  
 Atmospheric pressure is 101.3250 kPa.

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:41  
EQUIPMENT SUMMARIES

## Tower Rigorous Distillation Summary

|                           |             |
|---------------------------|-------------|
| Equip. No.                | 1           |
| Name                      |             |
| No. of stages             | 7           |
| 1st feed stage            | 1           |
| 2nd feed stage            | 7           |
| Top pressure kPa          | 75.0000     |
| Cond pressure drop (kPa)  | 2.0000      |
| Colm pressure drop (kPa)  | 5.0000      |
| Condenser type            | 1           |
| Condenser mode            | 5           |
| Condenser spec.           | 0.5000      |
| Cond. comp i              | 1           |
| Reboiler mode             | 5           |
| Reboiler spec.            | 1.0000      |
| Reboiler comp i           | 4           |
| Iterations                | 200         |
| Calc cond duty MJ/h       | -86766.0000 |
| Calc rebr duty MJ/h       | 18620.3184  |
| Est. Dist. rate (kmol/h)  | 1500.0000   |
| Est. Reflux rate (kmol/h) | 1200.0000   |
| Tolerance                 | 0.0010      |
| Est. T top C              | 100.0000    |
| Est. T bottom C           | 260.0000    |
| Column diameter m         | 4.5720      |
| Tray space m              | 0.6096      |
| No of sections            | 1           |
| Calc Reflux ratio         | 0.1901      |
| Calc Reflux mole (kmol/h) | 531.9780    |
| Calc Reflux mass kg/h     | 140737.7344 |

## Kinetic Reactor Summary

---

|                     |             |             |
|---------------------|-------------|-------------|
| Equip. No.          | 2           | 13          |
| Name                |             |             |
| Reactor type        | 2           | 2           |
| Reaction phase      | 1           | 1           |
| Thermal mode        | 5           | 1           |
| Pressure In kPa     | 250.0000    | 250.0000    |
| Pressure Drop kPa   | 30.0000     | 10.0000     |
| Tout C              | 609.0915    | 600.0000    |
| Q MJ/h              | -72332.6094 | -74636.0547 |
| Reactor volume m3   | 34.5576     | 22.0008     |
| Length of Tubes m   | 3.2000      |             |
| Diameter of Tubes m | 0.0250      |             |
| Number of Tubes     | 22000.0000  |             |

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:41  
EQUIPMENT SUMMARIES

|                             |             |             |
|-----------------------------|-------------|-------------|
| Number of steps             | 1000.0000   |             |
| Tolerance                   | 1.0000e-004 |             |
| Conversion                  | 0.9500      |             |
| Key                         | 7           | 0           |
| No. of Reactions            | 4           | 4           |
| Molar Flow Unit             | 1           | 1           |
| Activ. E/H of Rxn Unit      | 7           | 7           |
| Volume Unit                 | 1           | 1           |
| Time Unit                   | 2           | 2           |
| U W/m2-K                    | 100.0000    |             |
| Util dir                    | 1           | 0           |
| Util T at L C               | 562.0779    |             |
| T ref for HtRcn C           | 25.0000     |             |
| Overall IG Ht of Rxn (MJ/h) | -87301.2031 | -87216.9063 |
| Edit Reaction No.           | -1          | -1          |
| Mass unit                   | 1           | 1           |

#### Reaction Stoichiometrics and Parameters for unit no. 2

##### Reaction 1

RateConst = 7.7000e+006 Act.E = 2.5142e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -4.50e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 1    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

##### Reaction 2

RateConst = 6.3100e+007 Act.E = 2.9850e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -7.50e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 3.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 6.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

##### Reaction 3

RateConst = 2.3300e+004 Act.E = 2.1429e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 1    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -3.00e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 4.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

##### Reaction 4

RateConst = 7.2000e+005 Act.E = 2.7149e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -1.50e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

#### Reaction Stoichiometrics and Parameters for unit no. 13

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:41  
EQUIPMENT SUMMARIES

## Reaction 1

RateConst = 7.7000e+006 Act.E = 2.5143e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -4.50e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 1    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 2

RateConst = 6.3100e+007 Act.E = 2.9850e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -7.50e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 6.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 3.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 3

RateConst = 2.3300e+004 Act.E = 2.1429e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 1    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -3.00e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 8    | 4.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

## Reaction 4

RateConst = 7.2000e+005 Act.E = 2.7149e+004 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE     | AdsorbExp.  |
|------|------------|-------------|-------------|-------------|-------------|
| 6    | -1.00e+000 | 1.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 5    | -1.50e+000 | 1.0000e-010 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 7    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |
| 4    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000 |

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:42  
EQUIPMENT SUMMARIES

## Heat Exchanger Summary

|                     |           |             |
|---------------------|-----------|-------------|
| Equip. No.          | 3         | 6           |
| Name                |           |             |
| 1st Stream dp kPa   | 10.0000   | 5.0000      |
| 1st Stream T Out C  |           | 260.0000    |
| 1st Stream VF Out   | 1.0000    |             |
| Calc Ht Duty MJ/h   | 1737.9625 | -33195.7969 |
| 1st Stream Pout kPa | 250.0000  | 215.0000    |

## Mixer Summary

|                     |          |    |    |    |
|---------------------|----------|----|----|----|
| Equip. No.          | 4        | 10 | 12 | 14 |
| Name                |          |    |    |    |
| Output Pressure kPa | 250.0000 |    |    |    |

## Fired Heater Summary

|                    |            |
|--------------------|------------|
| Equip. No.         | 5          |
| Name               |            |
| Pressure Drop kPa  | 15.0000    |
| Temperature Out C  | 460.0000   |
| Heat Absorbed MJ/h | 28938.4980 |
| Fuel Usage (SCF)   | 40634.5117 |

## Compressor Summary

|                            |           |
|----------------------------|-----------|
| Equip. No.                 | 7         |
| Name                       |           |
| Pressure out kPa           | 250.0000  |
| Type of Compressor         | 1         |
| Efficiency                 | 0.8000    |
| Actual power kW            | 2524.9128 |
| Cp/Cv                      | 1.3987    |
| Theoretical power kW       | 2019.9303 |
| Ideal Cp/Cv                | 1.3974    |
| Calc Pout kPa              | 250.0000  |
| Calc. mass flowrate (kg/h) | 80490     |

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:42  
EQUIPMENT SUMMARIES

## Component Separator Summary

|                 |        |
|-----------------|--------|
| Equip. No.      | 8      |
| Name            |        |
| Component No. 3 | 1.0000 |
| Component No. 5 | 1.0000 |
| Component No. 6 | 1.0000 |
| Component No. 8 | 1.0000 |

## Shortcut Distillation Summary

|                          |             |
|--------------------------|-------------|
| Equip. No.               | 9           |
| Name                     |             |
| Mode                     | 2           |
| Light key component      | 9.0000      |
| Light key split          | 0.9990      |
| Heavy key component      | 2.0000      |
| Heavy key split          | 8.0000e-005 |
| R/Rmin                   | 50.0000     |
| Number of stages         | 16.2351     |
| Min. No. of stages       | 15.9805     |
| Feed stage               | 9.7945      |
| Condenser duty MJ/h      | -84058.4922 |
| Reboiler duty MJ/h       | 136038.0469 |
| Colm pressure kPa        | 70.0000     |
| Reflux ratio, minimum    | 1.2052      |
| Calc. Reflux ratio       | 60.2580     |
| Colm pressure drop (kPa) | 10.0000     |

## Reactor Summary

|                          |              |
|--------------------------|--------------|
| Equip. No.               | 11           |
| Name                     |              |
| Thermal mode             | 1            |
| Temperature C            | 194.1640     |
| Key Component            | 4            |
| Frac. Conversion         | 1.0000       |
| Calc H of Reac. (J/kmol) | -3.5680e+007 |
| Stoichiometrics:         |              |
| Maleic Anhydrid          | -1.000       |
| Water                    | -1.000       |
| Maleic Acid              | 1.000        |

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:42  
EQUIPMENT SUMMARIES

## Divider Summary

|                  |        |
|------------------|--------|
| Equip. No.       | 15     |
| Name             |        |
| Output stream #1 | 0.9000 |
| Output stream #2 | 0.1000 |



Job Name: Maleic Anhydride\_JAS\_2008  
FLOW SUMMARIES

Date: 12/14/2008 Time: 15:53:42

| Stream No.          | 1         | 2         | 3         | 4         |
|---------------------|-----------|-----------|-----------|-----------|
| Stream Name         |           |           |           |           |
| Temp C              | 30.0000   | 460.0000  | 586.1242  | 30.0000   |
| Pres kPa            | 260.0000  | 235.0000  | 235.0000  | 101.0000  |
| Enth MJ/h           | 2107.0    | 42281.    | -32355.   | 408.08    |
| Vapor mole fraction | 0.00000   | 1.0000    | 1.0000    | 1.0000    |
| Total kmol/h        | 42.3000   | 2832.2999 | 2828.4993 | 2790.0000 |
| Flowrates in kmol/h |           |           |           |           |
| Maleic Anhydride    | 0.0000    | 0.0000    | 22.0952   | 0.0000    |
| DibutylPhthalate    | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Nitrogen            | 0.0000    | 2205.0000 | 2205.0000 | 2205.0000 |
| Water               | 0.0000    | 0.0000    | 90.5157   | 0.0000    |
| Oxygen              | 0.0000    | 585.0000  | 370.4695  | 585.0000  |
| Benzene             | 42.3000   | 42.3000   | 4.2890    | 0.0000    |
| Quinone             | 0.0000    | 0.0000    | 0.7110    | 0.0000    |
| Carbon Dioxide      | 0.0000    | 0.0000    | 135.4191  | 0.0000    |
| Maleic Acid         | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Sodium Nitrite      | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Sodium Nitrate      | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Stream No.          | 5         | 6         | 7         | 8         |
| Stream Name         |           |           |           |           |
| Temp C              | 144.7681  | 460.0000  | 609.0915  | 260.0000  |
| Pres kPa            | 250.0000  | 235.0000  | 220.0000  | 215.0000  |
| Enth MJ/h           | 9497.8    | 42281.    | -30051.   | -63247.   |
| Vapor mole fraction | 1.0000    | 1.0000    | 1.0000    | 1.0000    |
| Total kmol/h        | 2790.0000 | 2832.2999 | 2825.1976 | 2825.1976 |
| Flowrates in kmol/h |           |           |           |           |
| Maleic Anhydride    | 0.0000    | 0.0000    | 26.3516   | 26.3516   |
| DibutylPhthalate    | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Nitrogen            | 2205.0000 | 2205.0000 | 2205.0000 | 2205.0000 |
| Water               | 0.0000    | 0.0000    | 92.0598   | 92.0598   |
| Oxygen              | 585.0000  | 585.0000  | 368.7527  | 368.7527  |
| Benzene             | 0.0000    | 42.3000   | 2.3447    | 2.3447    |
| Quinone             | 0.0000    | 0.0000    | 0.7272    | 0.7272    |
| Carbon Dioxide      | 0.0000    | 0.0000    | 129.9619  | 129.9619  |
| Maleic Acid         | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Sodium Nitrite      | 0.0000    | 0.0000    | 0.0000    | 0.0000    |
| Sodium Nitrate      | 0.0000    | 0.0000    | 0.0000    | 0.0000    |

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:42

## FLOW SUMMARIES

| Stream No.          | 9            | 10           | 11           | 12           |
|---------------------|--------------|--------------|--------------|--------------|
| Stream Name         |              |              |              |              |
| Temp C              | 329.1962     | 320.0000     | 194.1640     | 83.6666      |
| Pres kPa            | 80.0000      | 100.0000     | 82.0000      | 75.0000      |
| Enth MJ/h           | -3.2707E+005 | -72.616      | -3.9034E+005 | -68118.      |
| Vapor mole fraction | 0.00000      | 0.00000      | 0.00000      | 1.0000       |
| Total kmol/h        | 501.6278     | 0.1100       | 527.7746     | 2798.0366    |
| Flowrates in kmol/h |              |              |              |              |
| Maleic Anhydride    | 0.0000       | 0.0000       | 24.8517      | 0.4999       |
| DibutylPhthalate    | 501.6268     | 0.1100       | 501.5569     | 0.0556       |
| Nitrogen            | 0.0000       | 0.0000       | 0.0000       | 2205.0000    |
| Water               | 0.0000       | 0.0000       | 0.0000       | 91.0598      |
| Oxygen              | 0.0000       | 0.0000       | 0.0000       | 368.7527     |
| Benzene             | 0.0000       | 0.0000       | 0.0000       | 2.3447       |
| Quinone             | 0.0000       | 0.0000       | 0.3650       | 0.3622       |
| Carbon Dioxide      | 0.0000       | 0.0000       | 0.0000       | 129.9620     |
| Maleic Acid         | 0.0010       | 0.0000       | 1.0010       | 0.0000       |
| Sodium Nitrite      | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Sodium Nitrate      | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Stream No.          | 13           | 14           | 15           | 16           |
| Stream Name         |              |              |              |              |
| Temp C              | 189.2050     | 329.1982     | 418.5000     | 562.0779     |
| Pres kPa            | 70.0000      | 80.0000      | 200.0000     | 200.0000     |
| Enth MJ/h           | -11363.      | -3.2700E+005 | -2.0055E+006 | -1.9331E+006 |
| Vapor mole fraction | 0.00000      | 0.00000      | 0.00000      | 0.00000      |
| Total kmol/h        | 26.2568      | 501.5177     | 5000.0000    | 5000.0000    |
| Flowrates in kmol/h |              |              |              |              |
| Maleic Anhydride    | 24.8517      | 0.0000       | 0.0000       | 0.0000       |
| DibutylPhthalate    | 0.0401       | 501.5167     | 0.0000       | 0.0000       |
| Nitrogen            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Water               | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Oxygen              | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Benzene             | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Quinone             | 0.3650       | 0.0000       | 0.0000       | 0.0000       |
| Carbon Dioxide      | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Maleic Acid         | 1.0000       | 0.0010       | 0.0000       | 0.0000       |
| Sodium Nitrite      | 0.0000       | 0.0000       | 2065.6302    | 2065.6302    |
| Sodium Nitrate      | 0.0000       | 0.0000       | 2934.3694    | 2934.3694    |

Job Name: Maleic Anhydride\_JAS\_2008 Date: 12/14/2008 Time: 15:53:42  
 FLOW SUMMARIES

| Stream No.          | 91           | 92        | 93        | 100          |
|---------------------|--------------|-----------|-----------|--------------|
| Stream Name         |              |           |           |              |
| Temp C              | 460.0000     | 460.0000  | 600.0000  | 143.0405     |
| Pres kPa            | 235.0000     | 235.0000  | 240.0000  | 250.0000     |
| Enth MJ/h           | 38053.       | 4228.1    | -36583.   | 13343.       |
| Vapor mole fraction | 1.0000       | 1.0000    | 1.0000    | 1.0000       |
| Total kmol/h        | 2549.0700    | 283.2300  | 2545.2694 | 2832.2999    |
| Flowrates in kmol/h |              |           |           |              |
| Maleic Anhydride    | 0.0000       | 0.0000    | 22.0952   | 0.0000       |
| DibutylPhthalate    | 0.0000       | 0.0000    | 0.0000    | 0.0000       |
| Nitrogen            | 1984.4998    | 220.5000  | 1984.4998 | 2205.0000    |
| Water               | 0.0000       | 0.0000    | 90.5157   | 0.0000       |
| Oxygen              | 526.5000     | 58.5000   | 311.9694  | 585.0000     |
| Benzene             | 38.0700      | 4.2300    | 0.0590    | 42.3000      |
| Quinone             | 0.0000       | 0.0000    | 0.7110    | 0.0000       |
| Carbon Dioxide      | 0.0000       | 0.0000    | 135.4191  | 0.0000       |
| Maleic Acid         | 0.0000       | 0.0000    | 0.0000    | 0.0000       |
| Sodium Nitrite      | 0.0000       | 0.0000    | 0.0000    | 0.0000       |
| Sodium Nitrate      | 0.0000       | 0.0000    | 0.0000    | 0.0000       |
| Stream No.          | 102          | 103       | 104       | 105          |
| Stream Name         |              |           |           |              |
| Temp C              | 194.0467     | 83.6769   | 194.0467  | 194.0467     |
| Pres kPa            | 82.0000      | 75.0000   | 82.0000   | 82.0000      |
| Enth MJ/h           | -3.9034E+005 | -68121.   | 2.6245    | -3.9034E+005 |
| Vapor mole fraction | 0.00000      | 1.0000    | 1.0000    | 0.00000      |
| Total kmol/h        | 528.8044     | 2798.0069 | 0.0298    | 528.7746     |
| Flowrates in kmol/h |              |           |           |              |
| Maleic Anhydride    | 25.8517      | 0.4999    | 0.0000    | 25.8517      |
| DibutylPhthalate    | 501.5569     | 0.0556    | 0.0000    | 501.5569     |
| Nitrogen            | 0.0025       | 2204.9976 | 0.0025    | 0.0000       |
| Water               | 1.0000       | 91.0598   | 0.0000    | 1.0000       |
| Oxygen              | 0.0005       | 368.7522  | 0.0005    | 0.0000       |
| Benzene             | 0.0266       | 2.3181    | 0.0266    | 0.0000       |
| Quinone             | 0.3650       | 0.3622    | 0.0000    | 0.3650       |
| Carbon Dioxide      | 0.0002       | 129.9618  | 0.0002    | 0.0000       |
| Maleic Acid         | 0.0010       | 0.0000    | 0.0000    | 0.0010       |
| Sodium Nitrite      | 0.0000       | 0.0000    | 0.0000    | 0.0000       |
| Sodium Nitrate      | 0.0000       | 0.0000    | 0.0000    | 0.0000       |

Job Name: Maleic Anhydride\_JAS\_2008    Date: 12/14/2008    Time: 15:53:42  
FLOW SUMMARIES

|                     |          |
|---------------------|----------|
| Stream No.          | 110      |
| Stream Name         |          |
| Temp C              | 112.6301 |
| Pres kPa            | 250.0000 |
| Enth MJ/h           | 3844.9   |
| Vapor mole fraction | 1.0000   |
| Total kmol/h        | 42.3000  |
| Flowrates in kmol/h |          |
| Maleic Anhydride    | 0.0000   |
| DibutylPhthalate    | 0.0000   |
| Nitrogen            | 0.0000   |
| Water               | 0.0000   |
| Oxygen              | 0.0000   |
| Benzene             | 42.3000  |
| Quinone             | 0.0000   |
| Carbon Dioxide      | 0.0000   |
| Maleic Acid         | 0.0000   |
| Sodium Nitrite      | 0.0000   |
| Sodium Nitrate      | 0.0000   |

13.15 Simulation of the Ethylene Oxide process, Appendix B, Figure B.6.1 and Table B.6.1. The simulation results are shown in the output following this page. The PSRK model was used for the phase equilibrium (except for towers T-701 and T-702 in which the UNIFAC model was used), and the SRK model was used for enthalpies. The required minimum input information for this simulation is given below:

1. Flowsheet topology
2. Specification of feed streams (Streams 1, 2, and process water)
3. Equipment specifications
  - R-701, 702 conversion of key component
  - E-701, 702, 703, 704, 705, 706, 707 outlet temperatures
  - T-703 identification of and recovery of key components
  - T-701, 702 number of trays
  - C-701, 702, 703, 704 outlet pressures, efficiencies
4. Computational tolerances

13.16 Simulation of the Formalin process, Appendix B, Figure B.7.1 and Table B.7.1. The simulation results are shown in the output following this page. The Elliot-Suresh-Donahue Equation of State (ESDK) model was used for the phase equilibrium, and the “latent heat” model (zero heat of mixing) was used for enthalpies. The required minimum input information for this simulation is given below:

1. Flowsheet topology
2. Specification of feed streams (Streams 1, 2, and 11)
3. Equipment specifications
  - R-801 conversion of key component, outlet temperature
  - E-801, 802, 803, 806 outlet temperatures
  - T-802 identification of and recovery of key components
  - T-801 number of trays
  - C-801 outlet pressure, efficiency
4. Computational tolerances

13.17 The batch aspects vary by simulator. For Chemcad, these are Batch Reactor, Batch Column, Tank UnitOp, Dynamic Vessel UnitOp, and Scheduling.

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers |
|-----------|-------|----------------|
| 1         | COMP  | 1 -3           |
| 2         | HTXR  | 3 -4           |
| 3         | COMP  | 4 -5           |
| 4         | HTXR  | 5 -6           |
| 5         | COMP  | 6 -7           |
| 6         | MIXE  | 7 9 -10        |
| 7         | VALV  | 2 -8           |
| 8         | HTXR  | 10 -11         |
| 9         | KREA  | 11 -12         |
| 10        | HTXR  | 12 -13         |
| 11        | COMP  | 13 -14         |
| 12        | SCDS  | 15 14 -16 -17  |
| 13        | HTXR  | 16 -18         |
| 14        | VALV  | 18 -19         |
| 15        | KREA  | 19 -20         |
| 16        | HTXR  | 20 -21         |
| 17        | COMP  | 21 -22         |
| 18        | SCDS  | 23 22 -24 -25  |
| 19        | DIVI  | 24 -27 -26     |
| 20        | VALV  | 27 -28         |
| 21        | MIXE  | 8 28 -9        |
| 22        | MIXE  | 17 25 -29      |
| 23        | HTXR  | 29 -30         |
| 24        | VALV  | 102 -31        |
| 25        | SCDS  | 31 -32 -33     |
| 100       | VALV  | 30 -101        |
| 101       | CSEP  | 101 -103 -102  |
| 200       | SHOR  | 90 -91 -92     |
| 201       | SHOR  | 93 -94 -95     |

## Stream Connections

| Stream | Equipment |    | Stream | Equipment |    | Stream | Equipment |     |
|--------|-----------|----|--------|-----------|----|--------|-----------|-----|
|        | From      | To |        | From      | To |        | From      | To  |
| 1      |           | 1  | 15     |           | 12 | 29     | 22        | 23  |
| 2      |           | 7  | 16     | 12        | 13 | 30     | 23        | 100 |
| 3      | 1         | 2  | 17     | 12        | 22 | 31     | 24        | 25  |
| 4      | 2         | 3  | 18     | 13        | 14 | 32     | 25        |     |
| 5      | 3         | 4  | 19     | 14        | 15 | 33     | 25        |     |
| 6      | 4         | 5  | 20     | 15        | 16 | 90     |           | 200 |
| 7      | 5         | 6  | 21     | 16        | 17 | 91     | 200       |     |
| 8      | 7         | 21 | 22     | 17        | 18 | 92     | 200       |     |
| 9      | 21        | 6  | 23     |           | 18 | 93     |           | 201 |
| 10     | 6         | 8  | 24     | 18        | 19 | 94     | 201       |     |
| 11     | 8         | 9  | 25     | 18        | 22 | 95     | 201       |     |
| 12     | 9         | 10 | 26     | 19        |    | 101    | 100       | 101 |
| 13     | 10        | 11 | 27     | 19        | 20 | 102    | 101       | 24  |
| 14     | 11        | 12 | 28     | 20        | 21 | 103    | 101       |     |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10

Calculation mode : Sequential  
Flash algorithm : Normal

#### Equipment Calculation Sequence

|    |    |    |    |    |    |     |     |     |    |    |    |    |    |    |    |    |    |
|----|----|----|----|----|----|-----|-----|-----|----|----|----|----|----|----|----|----|----|
| 1  | 2  | 3  | 4  | 5  | 7  | 200 | 201 | 6   | 8  | 9  | 10 | 11 | 12 | 13 | 14 | 15 | 16 |
| 17 | 18 | 19 | 20 | 21 | 22 | 23  | 100 | 101 | 24 | 25 |    |    |    |    |    |    |    |

#### Equipment Recycle Sequence

|   |   |   |    |    |    |    |    |    |    |    |    |    |    |    |
|---|---|---|----|----|----|----|----|----|----|----|----|----|----|----|
| 6 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 |
|---|---|---|----|----|----|----|----|----|----|----|----|----|----|----|

#### Recycle Cut Streams

9

Recycle Convergence Method: Direct Substitution

Max. loop iterations 40

#### Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-003 |
| Temperature | 1.000E-003 |
| Pressure    | 1.000E-003 |
| Enthalpy    | 1.000E-003 |
| Vapor frac. | 1.000E-003 |

Recycle calculation has converged.

| Overall Mass Balance | kmol/h    |           | kg/h        |             |
|----------------------|-----------|-----------|-------------|-------------|
|                      | Input     | Output    | Input       | Output      |
| Ethylene             | 712.911   | 336.269   | 20000.000   | 9433.694    |
| Ethylene Oxide       | 708.001   | 1068.702  | 31189.564   | 47079.518   |
| Carbon Dioxide       | 0.000     | 31.731    | 0.000       | 1396.474    |
| Oxygen               | 3281.353  | 3052.653  | 105000.002  | 97681.858   |
| Nitrogen             | 14100.092 | 14096.652 | 395000.000  | 394904.000  |
| Water                | 80058.749 | 80090.472 | 1442258.306 | 1442829.798 |
| Total                | 98861.109 | 98676.483 | 1993447.974 | 1993325.051 |

#### COMPONENTS

|   | ID # | Name           | Formula |
|---|------|----------------|---------|
| 1 | 22   | Ethylene       | C2H4    |
| 2 | 129  | Ethylene Oxide | C2H4O   |
| 3 | 49   | Carbon Dioxide | CO2     |
| 4 | 47   | Oxygen         | O2      |
| 5 | 46   | Nitrogen       | N2      |
| 6 | 62   | Water          | H2O     |

#### THERMODYNAMICS

K-value model : PSRK  
IPP flag on  
Enthalpy model : SRK  
Liquid density : Library

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10

Equip # 12 K-value : UNIF

Equip # 18 K-value : UNIF

Std vapor rate reference temperature is 0 C.

Atmospheric pressure is 1.0132 bar.

\* UNIFAC main groups 2 and 1058, No interaction parameters.

\* UNIFAC main groups 52 and 1058, No interaction parameters.

\* UNIFAC main groups 2 and 1058, No interaction parameters.

\* UNIFAC main groups 52 and 1058, No interaction parameters.



Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10  
EQUIPMENT SUMMARIES

## Compressor Summary

| Equip. No.               | 1          | 3          | 5          | 11         |
|--------------------------|------------|------------|------------|------------|
| Name                     |            |            |            |            |
| Pressure out bar         | 3.0000     | 9.0000     | 27.0000    | 30.0000    |
| Type of Compressor       | 1          | 1          | 1          | 1          |
| Efficiency               | 0.8000     | 0.8000     | 0.8000     | 0.8000     |
| Actual power MJ/h        | 68510.1172 | 82629.3125 | 77204.5703 | 19803.6094 |
| Cp/Cv                    | 1.3990     | 1.4008     | 1.4093     | 1.4259     |
| Theoretical power (MJ/h) | 54808.0977 | 66103.4531 | 61763.6602 | 15842.8877 |
| Ideal Cp/Cv              | 1.3966     | 1.3949     | 1.3898     | 1.3663     |
| Calc Pout bar            | 3.0000     | 9.0000     | 27.0000    | 30.0000    |
| Equip. No.               | 17         |            |            |            |
| Name                     |            |            |            |            |
| Pressure out bar         | 30.0000    |            |            |            |
| Type of Compressor       | 1          |            |            |            |
| Efficiency               | 0.8000     |            |            |            |
| Actual power MJ/h        | 19666.7070 |            |            |            |
| Cp/Cv                    | 1.4272     |            |            |            |
| Theoretical power (MJ/h) | 15733.3662 |            |            |            |
| Ideal Cp/Cv              | 1.3679     |            |            |            |
| Calc Pout bar            | 30.0000    |            |            |            |

## Heat Exchanger Summary

| Equip. No.          | 2           | 4            | 8           | 10           |
|---------------------|-------------|--------------|-------------|--------------|
| Name                |             |              |             |              |
| 1st Stream dp bar   | 0.3000      | 0.3000       | 0.3000      | 0.3000       |
| 1st Stream T Out C  | 45.0000     | 45.0000      | 240.0000    | 45.0000      |
| Calc Ht Duty MJ/h   | -58486.8594 | -83202.0703  | 147433.5781 | -215064.4844 |
| 1st Stream Pout bar | 2.7000      | 8.7000       | 26.5000     | 25.4500      |
| Equip. No.          | 13          | 16           | 23          |              |
| Name                |             |              |             |              |
| 1st Stream dp bar   | 0.3000      | 0.3000       | 0.3000      |              |
| 1st Stream T Out C  | 240.0000    | 45.0000      | 45.0000     |              |
| Calc Ht Duty MJ/h   | 229646.0781 | -212483.3281 | -21591.4238 |              |
| 1st Stream Pout bar | 29.7000     | 25.4500      | 89.7000     |              |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10  
EQUIPMENT SUMMARIES

## Mixer Summary

|                     |         |    |         |
|---------------------|---------|----|---------|
| Equip. No.          | 6       | 21 | 22      |
| Name                |         |    |         |
| Output Pressure bar | 26.8000 |    | 90.0000 |

## Valve Summary

|                  |         |         |         |         |
|------------------|---------|---------|---------|---------|
| Equip. No.       | 7       | 14      | 20      | 24      |
| Name             |         |         |         |         |
| Pressure out bar | 27.0000 | 26.5000 | 27.0000 | 10.0000 |

|                  |         |
|------------------|---------|
| Equip. No.       | 100     |
| Name             |         |
| Pressure out bar | 20.0000 |

## Kinetic Reactor Summary

|                             |             |             |
|-----------------------------|-------------|-------------|
| Equip. No.                  | 9           | 15          |
| Name                        |             |             |
| Reactor type                | 2           | 2           |
| Reaction phase              | 1           | 1           |
| Thermal mode                | 1           | 1           |
| Pressure Drop bar           | 0.7500      | 0.7500      |
| Tout C                      | 240.0000    | 240.0000    |
| Q MJ/h                      | -33103.3867 | -26179.0000 |
| Reactor volume m3           | 201.8075    | 201.5797    |
| Concentration Flag          | 1           | 1           |
| Specify calc. mode          | 1           | 1           |
| Conversion                  | 0.2000      | 0.2000      |
| Key                         | 1           | 1           |
| No. of Reactions            | 3           | 3           |
| Molar Flow Unit             | 2           | 2           |
| Activ. E/H of Rxn Unit      | 6           | 6           |
| Volume Unit                 | 1           | 1           |
| Time Unit                   | 2           | 2           |
| Overall IG Ht of Rxn (MJ/h) | -32822.8203 | -26013.8945 |
| Mass unit                   | 1           | 1           |

## Reaction Stoichiometrics and Parameters for unit no. 9

Reaction 1  
RateConst = 1.9600e+000 Act.E = 2.4000e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE      | AdsorbExp.  |
|------|------------|-------------|-------------|--------------|-------------|
| 1    | -1.00e+000 | 1.0000e+000 | 9.8000e-004 | -1.1200e+004 | 1.0000e+000 |
| 4    | -5.00e-001 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10  
EQUIPMENT SUMMARIES

## Reaction 2

RateConst = 9.3600e-002 Act.E = 6.4000e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE      | AdsorbExp.  |
|------|------------|-------------|-------------|--------------|-------------|
| 1    | -1.00e+000 | 1.0000e+000 | 9.8000e-004 | -1.1200e+004 | 1.0000e+000 |
| 4    | -3.00e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 3    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 6    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |

## Reaction 3

RateConst = 4.2768e-001 Act.E = 6.2000e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE      | AdsorbExp.  |
|------|------------|-------------|-------------|--------------|-------------|
| 2    | -1.00e+000 | 2.0000e+000 | 3.3000e-005 | -2.1200e+004 | 2.0000e+000 |
| 4    | -2.50e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 3    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 6    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |

## Reaction Stoichiometrics and Parameters for unit no. 15

## Reaction 1

RateConst = 1.9600e+000 Act.E = 2.4000e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE      | AdsorbExp.  |
|------|------------|-------------|-------------|--------------|-------------|
| 1    | -1.00e+000 | 1.0000e+000 | 9.8000e-004 | -1.1200e+004 | 1.0000e+000 |
| 4    | -5.00e-001 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 2    | 1.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |

## Reaction 2

RateConst = 9.3600e-002 Act.E = 6.4000e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE      | AdsorbExp.  |
|------|------------|-------------|-------------|--------------|-------------|
| 1    | -1.00e+000 | 1.0000e+000 | 9.8000e-004 | -1.1200e+004 | 1.0000e+000 |
| 4    | -3.00e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 3    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 6    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |

## Reaction 3

RateConst = 4.2768e-001 Act.E = 6.2000e+003 Hrxn = 0.0000e+000

| Comp | Stoich.    | Exp.factor  | AdsorbFac.  | AdsorbE      | AdsorbExp.  |
|------|------------|-------------|-------------|--------------|-------------|
| 2    | -1.00e+000 | 2.0000e+000 | 3.3000e-005 | -2.1200e+004 | 2.0000e+000 |
| 4    | -2.50e+000 | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 3    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |
| 6    | 2.00e+000  | 0.0000e+000 | 0.0000e+000 | 0.0000e+000  | 0.0000e+000 |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10

## Scds Rigorous Distillation Summary

|                           |             |             |             |
|---------------------------|-------------|-------------|-------------|
| Equip. No.                | 12          | 18          | 25          |
| Name                      |             |             |             |
| No. of stages             | 5           | 5           | 25          |
| 1st feed stage            | 1           | 1           | 12          |
| 2nd feed stage            | 5           | 5           | 0           |
| Condenser mode            | 0           | 0           | 6           |
| Condenser spec            |             |             | 0.9990      |
| Cond comp i pos.          | 0           | 0           | 2           |
| Reboiler mode             | 0           | 0           | 3           |
| Reboiler spec.            |             |             | 182.3000    |
| Reboiler comp i           | 0           | 0           | 2           |
| Colm press drop bar       |             |             | 0.5000      |
| Est. dist. rate (kmol/h)  | 18000.0000  | 18000.0000  | 284.0000    |
| Est. reflux rate (kmol/h) |             |             | 240.0000    |
| Est. T top C              | 25.0000     | 25.0000     | 92.0000     |
| Est. T bottom C           | 25.0000     | 25.0000     | 182.0000    |
| Top pressure bar          |             |             | 10.0000     |
| Calc Cond duty MJ/h       |             |             | -14398.4043 |
| Calc Reblr duty MJ/h      |             |             | 438665.1250 |
| Calc Reflux mole (kmol/h) | 20059.8223  | 20054.5645  | 313.9723    |
| Calc Reflux ratio         |             |             | 0.8909      |
| Calc Reflux mass kg/h     | 362247.1563 | 362040.4063 | 13823.2471  |
| Tray type                 | 3           | 3           | 3           |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10

## EQUIPMENT SUMMARIES

|                   |        |        |        |
|-------------------|--------|--------|--------|
| Column diameter m | 4.8768 | 4.7244 | 5.4864 |
| Tray space m      | 0.6096 | 0.6096 | 0.6096 |
| Thickness (top) m | 0.0953 | 0.0889 | 0.0333 |
| Thickness (bot) m | 0.0953 | 0.0889 | 0.0333 |
| No of sections    | 1      | 1      | 1      |
| No of passes (S1) | 1      | 1      | 1      |
| Weir side width m | 0.8827 | 0.8446 | 0.9779 |
| Weir height m     | 0.0508 | 0.0508 | 0.0508 |
| System factor     | 1.0000 | 1.0000 | 1.0000 |

## Divider Summary

|                  |        |
|------------------|--------|
| Equip. No.       | 19     |
| Name             |        |
| Output stream #1 | 0.5000 |
| Output stream #2 | 0.5000 |

## Component Separator Summary

|                 |        |
|-----------------|--------|
| Equip. No.      | 101    |
| Name            |        |
| Component No. 1 | 1.0000 |
| Component No. 3 | 1.0000 |
| Component No. 4 | 1.0000 |
| Component No. 5 | 1.0000 |

## Shortcut Distillation Summary

| Equip. No.               | 200         | 201        |
|--------------------------|-------------|------------|
| Name                     |             |            |
| Mode                     | 2           | 2          |
| Light key component      | 2.0000      | 2.0000     |
| Light key split          | 0.9990      | 0.9990     |
| Heavy key component      | 6.0000      | 6.0000     |
| Heavy key split          | 1.0000e-005 | 0.0010     |
| R/Rmin                   | 1.5000      | 1.5000     |
| Number of stages         | 15.7341     | 17.5054    |
| Min. No. of stages       | 8.8737      | 6.8893     |
| Feed stage               | 10.2085     | 9.2527     |
| Condenser duty MJ/h      | -24506.5488 | -9023.3018 |
| Reboiler duty MJ/h       | 448850.5313 | 9181.1094  |
| Colm pressure bar        | 10.0000     | 10.0000    |
| Reflux ratio, minimum    | 1.4672      | 0.1220     |
| Calc. Reflux ratio       | 2.2008      | 0.1829     |
| Number of points         | 0           | 10         |
| Lower bound R/Rmin       |             | 1.1000     |
| Upper bound R/Rmin       |             | 2.0000     |
| Colm pressure drop (bar) | 0.5000      | 0.5000     |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10  
 FLOW SUMMARIES

| Stream No.          | 1          | 2           | 3           | 4           |
|---------------------|------------|-------------|-------------|-------------|
| Stream Name         |            |             |             |             |
| Temp C              | 25.0000    | 25.0000     | 159.1927    | 45.0000     |
| Pres bar            | 1.0132     | 50.0000     | 3.0000      | 2.7000      |
| Enth MJ/h           | -115.73    | 35030.      | 68394.      | 9907.5      |
| Vapor mole fraction | 1.0000     | 1.0000      | 1.0000      | 1.0000      |
| Total kmol/h        | 17381.4452 | 712.9108    | 17381.4452  | 17381.4452  |
| Flowrates in kmol/h |            |             |             |             |
| Ethylene            | 0.0000     | 712.9108    | 0.0000      | 0.0000      |
| Ethylene Oxide      | 0.0000     | 0.0000      | 0.0000      | 0.0000      |
| Carbon Dioxide      | 0.0000     | 0.0000      | 0.0000      | 0.0000      |
| Oxygen              | 3281.3525  | 0.0000      | 3281.3525   | 3281.3525   |
| Nitrogen            | 14100.0925 | 0.0000      | 14100.0925  | 14100.0925  |
| Water               | 0.0000     | 0.0000      | 0.0000      | 0.0000      |
| Stream No.          | 5          | 6           | 7           | 8           |
| Stream Name         |            |             |             |             |
| Temp C              | 206.1111   | 45.0000     | 195.2089    | -6.2991     |
| Pres bar            | 9.0000     | 8.7000      | 27.0000     | 27.0000     |
| Enth MJ/h           | 92537.     | 9334.8      | 86539.      | 35030.      |
| Vapor mole fraction | 1.0000     | 1.0000      | 1.0000      | 1.0000      |
| Total kmol/h        | 17381.4452 | 17381.4452  | 17381.4452  | 712.9108    |
| Flowrates in kmol/h |            |             |             |             |
| Ethylene            | 0.0000     | 0.0000      | 0.0000      | 712.9108    |
| Ethylene Oxide      | 0.0000     | 0.0000      | 0.0000      | 0.0000      |
| Carbon Dioxide      | 0.0000     | 0.0000      | 0.0000      | 0.0000      |
| Oxygen              | 3281.3525  | 3281.3525   | 3281.3525   | 0.0000      |
| Nitrogen            | 14100.0925 | 14100.0925  | 14100.0925  | 0.0000      |
| Water               | 0.0000     | 0.0000      | 0.0000      | 0.0000      |
| Stream No.          | 9          | 10          | 11          | 12          |
| Stream Name         |            |             |             |             |
| Temp C              | 26.3315    | 106.7495    | 240.0000    | 240.0000    |
| Pres bar            | 27.0000    | 26.8000     | 26.5000     | 25.7500     |
| Enth MJ/h           | 31440.     | 1.1798E+005 | 2.6541E+005 | 2.3231E+005 |
| Vapor mole fraction | 1.0000     | 1.0000      | 1.0000      | 1.0000      |
| Total kmol/h        | 18258.1383 | 35635.3116  | 35635.3116  | 35535.1456  |
| Flowrates in kmol/h |            |             |             |             |
| Ethylene            | 1047.9103  | 1047.8481   | 1047.8481   | 838.5940    |
| Ethylene Oxide      | 6.4718     | 6.4621      | 6.4621      | 206.7925    |
| Carbon Dioxide      | 31.7102    | 31.7042     | 31.7042     | 49.5512     |
| Oxygen              | 3049.7688  | 6330.3852   | 6330.3852   | 6203.4489   |
| Nitrogen            | 14091.2997 | 28187.9521  | 28187.9521  | 28187.9521  |
| Water               | 30.9775    | 30.9611     | 30.9611     | 48.8082     |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10  
FLOW SUMMARIES

| Stream No.          | 13         | 14         | 15           | 16         |
|---------------------|------------|------------|--------------|------------|
| Stream Name         |            |            |              |            |
| Temp C              | 45.0000    | 63.7223    | 25.0000      | 30.2942    |
| Pres bar            | 25.4500    | 30.0000    | 30.0000      | 30.0000    |
| Enth MJ/h           | 17246.     | 37050.     | -5.7204E+006 | 6746.0     |
| Vapor mole fraction | 1.0000     | 1.0000     | 0.00000      | 1.0000     |
| Total kmol/h        | 35535.1456 | 35535.1456 | 20000.0003   | 35353.3720 |
| Flowrates in kmol/h |            |            |              |            |
| Ethylene            | 838.5940   | 838.5940   | 0.0000       | 837.8904   |
| Ethylene Oxide      | 206.7925   | 206.7925   | 0.0000       | 15.4413    |
| Carbon Dioxide      | 49.5512    | 49.5512    | 0.0000       | 49.5421    |
| Oxygen              | 6203.4489  | 6203.4489  | 0.0000       | 6202.0000  |
| Nitrogen            | 28187.9521 | 28187.9521 | 0.0000       | 28185.2766 |
| Water               | 48.8082    | 48.8082    | 20000.0003   | 63.2260    |

| Stream No.          | 17           | 18          | 19          | 20          |
|---------------------|--------------|-------------|-------------|-------------|
| Stream Name         |              |             |             |             |
| Temp C              | 51.9139      | 240.0000    | 239.9476    | 240.0000    |
| Pres bar            | 30.0000      | 29.7000     | 26.5000     | 25.7500     |
| Enth MJ/h           | -5.6901E+006 | 2.3639E+005 | 2.3639E+005 | 2.1021E+005 |
| Vapor mole fraction | 0.00000      | 1.0000      | 1.0000      | 1.0000      |
| Total kmol/h        | 20181.7757   | 35353.3685  | 35353.3649  | 35273.1783  |
| Flowrates in kmol/h |              |             |             |             |
| Ethylene            | 0.7036       | 837.8904    | 837.8904    | 670.5651    |
| Ethylene Oxide      | 191.3511     | 15.4427     | 15.4427     | 175.8233    |
| Carbon Dioxide      | 0.0091       | 49.5421     | 49.5421     | 63.4319     |
| Oxygen              | 1.4518       | 6202.0000   | 6202.0000   | 6100.9706   |
| Nitrogen            | 2.6777       | 28185.2748  | 28185.2748  | 28185.2748  |
| Water               | 19985.5828   | 63.2209     | 63.2209     | 77.1108     |

| Stream No.          | 21         | 22         | 23           | 24         |
|---------------------|------------|------------|--------------|------------|
| Stream Name         |            |            |              |            |
| Temp C              | 45.0000    | 63.7760    | 25.0000      | 30.0814    |
| Pres bar            | 25.4500    | 30.0000    | 30.0000      | 30.0000    |
| Enth MJ/h           | -2271.2    | 17396.     | -5.7204E+006 | -7180.0    |
| Vapor mole fraction | 1.0000     | 1.0000     | 0.00000      | 1.0000     |
| Total kmol/h        | 35273.1783 | 35273.1783 | 20000.0003   | 35090.4550 |
| Flowrates in kmol/h |            |            |              |            |
| Ethylene            | 670.5651   | 670.5651   | 0.0000       | 669.9992   |
| Ethylene Oxide      | 175.8233   | 175.8233   | 0.0000       | 12.9436    |
| Carbon Dioxide      | 63.4319    | 63.4319    | 0.0000       | 63.4204    |
| Oxygen              | 6100.9706  | 6100.9706  | 0.0000       | 6099.5377  |
| Nitrogen            | 28185.2748 | 28185.2748 | 0.0000       | 28182.5994 |
| Water               | 77.1108    | 77.1108    | 20000.0003   | 61.9550    |

Job Name: ethylene oxide with recycle Date: 12/14/2008 Time: 14:37:10

## FLOW SUMMARIES

| Stream No.          | 25           | 26           | 27           | 28           |
|---------------------|--------------|--------------|--------------|--------------|
| Stream Name         |              |              |              |              |
| Temp C              | 52.2611      | 30.0814      | 30.0814      | 29.4755      |
| Pres bar            | 30.0000      | 30.0000      | 30.0000      | 27.0000      |
| Enth MJ/h           | -5.6959E+006 | -3590.0      | -3590.0      | -3590.0      |
| Vapor mole fraction | 0.00000      | 1.0000       | 1.0000       | 1.0000       |
| Total kmol/h        | 20182.7218   | 17545.2275   | 17545.2275   | 17545.2275   |
| Flowrates in kmol/h |              |              |              |              |
| Ethylene            | 0.5659       | 334.9996     | 334.9996     | 334.9996     |
| Ethylene Oxide      | 162.8796     | 6.4718       | 6.4718       | 6.4718       |
| Carbon Dioxide      | 0.0115       | 31.7102      | 31.7102      | 31.7102      |
| Oxygen              | 1.4329       | 3049.7688    | 3049.7688    | 3049.7688    |
| Nitrogen            | 2.6751       | 14091.2997   | 14091.2997   | 14091.2997   |
| Water               | 20015.1567   | 30.9775      | 30.9775      | 30.9775      |
| Stream No.          | 29           | 30           | 31           | 32           |
| Stream Name         |              |              |              |              |
| Temp C              | 52.0733      | 45.0000      | 45.0228      | 86.4025      |
| Pres bar            | 90.0000      | 89.7000      | 10.0000      | 10.0000      |
| Enth MJ/h           | -1.1386E+007 | -1.1408E+007 | -1.1408E+007 | -25451.      |
| Vapor mole fraction | 0.00000      | 0.00000      | 0.00000      | 0.00000      |
| Total kmol/h        | 40364.5011   | 40364.5011   | 40355.0000   | 352.4049     |
| Flowrates in kmol/h |              |              |              |              |
| Ethylene            | 1.2695       | 1.2695       | 0.0000       | 0.0000       |
| Ethylene Oxide      | 354.2293     | 354.2293     | 354.2293     | 352.0528     |
| Carbon Dioxide      | 0.0206       | 0.0206       | 0.0000       | 0.0000       |
| Oxygen              | 2.8846       | 2.8846       | 0.0000       | 0.0000       |
| Nitrogen            | 5.3527       | 5.3527       | 0.0000       | 0.0000       |
| Water               | 40000.7448   | 40000.7448   | 40000.7483   | 0.3524       |
| Stream No.          | 33           | 90           | 91           | 92           |
| Stream Name         |              |              |              |              |
| Temp C              | 182.3000     | 45.0228      | 86.4069      | 182.3307     |
| Pres bar            | 10.5000      | 10.0000      | 10.0000      | 10.5000      |
| Enth MJ/h           | -1.0958E+007 | -1.1408E+007 | -25596.      | -1.0958E+007 |
| Vapor mole fraction | 0.00000      | 0.00000      | 0.00000      | 0.00000      |
| Total kmol/h        | 40002.5698   | 40355.0000   | 354.2750     | 40000.7058   |
| Flowrates in kmol/h |              |              |              |              |
| Ethylene            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Ethylene Oxide      | 2.1765       | 354.2293     | 353.8749     | 0.3542       |
| Carbon Dioxide      | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Oxygen              | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Nitrogen            | 0.0000       | 0.0000       | 0.0000       | 0.0000       |
| Water               | 40000.3975   | 40000.7483   | 0.4000       | 40000.3514   |



Job Name: ethylene oxide with recycle      Date: 12/14/2008      Time: 14:37:10  
FLOW SUMMARIES

| Stream No.          | 93       | 94       | 95       | 101          |
|---------------------|----------|----------|----------|--------------|
| Stream Name         |          |          |          |              |
| Temp C              | 92.3400  | 86.3659  | 178.2449 | 45.0200      |
| Pres bar            | 10.0000  | 10.0000  | 10.5000  | 20.0000      |
| Enth MJ/h           | -41538.  | -25468.  | -15912.  | -1.1408E+007 |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000  | 0.00000      |
| Total kmol/h        | 411.7658 | 353.4758 | 58.2900  | 40364.5011   |
| Flowrates in kmol/h |          |          |          |              |
| Ethylene            | 0.0000   | 0.0000   | 0.0000   | 1.2695       |
| Ethylene Oxide      | 353.7715 | 353.4178 | 0.3538   | 354.2293     |
| Carbon Dioxide      | 0.0000   | 0.0000   | 0.0000   | 0.0206       |
| Oxygen              | 0.0000   | 0.0000   | 0.0000   | 2.8846       |
| Nitrogen            | 0.0000   | 0.0000   | 0.0000   | 5.3527       |
| Water               | 57.9943  | 0.0580   | 57.9363  | 40000.7448   |

| Stream No.          | 102          | 103     |
|---------------------|--------------|---------|
| Stream Name         |              |         |
| Temp C              | 45.0200      | 45.0200 |
| Pres bar            | 20.0000      | 20.0000 |
| Enth MJ/h           | -1.1408E+007 | 62.558  |
| Vapor mole fraction | 0.00000      | 1.0000  |
| Total kmol/h        | 40355.0000   | 9.5275  |
| Flowrates in kmol/h |              |         |
| Ethylene            | 0.0000       | 1.2695  |
| Ethylene Oxide      | 354.2293     | 0.0000  |
| Carbon Dioxide      | 0.0000       | 0.0206  |
| Oxygen              | 0.0000       | 2.8846  |
| Nitrogen            | 0.0000       | 5.3527  |
| Water               | 40000.7448   | 0.0000  |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38

## FLOWSHEET SUMMARY

| Equipment | Label | Stream Numbers    |
|-----------|-------|-------------------|
| 1         | PUMP  | 3 -4              |
| 2         | COMP  | 1 -5              |
| 3         | HTXR  | 4 -6              |
| 4         | HTXR  | 5 -7              |
| 5         | MIXE  | 6 7 -8            |
| 6         | REAC  | 8 -68             |
| 7         | REAC  | 68 -69            |
| 8         | HTXR  | 69 -9             |
| 9         | HTXR  | 9 -10             |
| 10        | SCDS  | 11 10 -12 -13     |
| 11        | SHOR  | 70 -14 -15        |
| 12        | PUMP  | 15 -16            |
| 13        | HTXR  | 16 -17            |
| 14        | VALV  | 14 -18            |
| 15        | FLAS  | 21 -22 -23        |
| 16        | FLAS  | 24 -25 -26        |
| 65        | MIXE  | 12 64 -65         |
| 66        | MIXE  | 18 2 -3           |
| 67        | CSEP  | 13 -64 -70        |
| 71        | TOWR  | 72 -73 -74        |
| 75        | SCDS  | 76 -77 -78        |
| 99        | KREA  | 98 -99            |
| 101       | FLAS  | 100 -101 -102     |
| 102       | SCDS  | 103 101 -104 -105 |
| 103       | MIXE  | 105 102 -106      |

## Stream Connections

| Stream | Equipment |    | Stream | Equipment |    | Stream | Equipment |     |
|--------|-----------|----|--------|-----------|----|--------|-----------|-----|
|        | From      | To |        | From      | To |        | From      | To  |
| 1      |           | 2  | 16     | 12        | 13 | 73     | 71        |     |
| 2      |           | 66 | 17     | 13        |    | 74     | 71        |     |
| 3      | 66        | 1  | 18     | 14        | 66 | 76     |           | 75  |
| 4      | 1         | 3  | 21     |           | 15 | 77     | 75        |     |
| 5      | 2         | 4  | 22     | 15        |    | 78     | 75        |     |
| 6      | 3         | 5  | 23     | 15        |    | 98     |           | 99  |
| 7      | 4         | 5  | 24     |           | 16 | 99     | 99        |     |
| 8      | 5         | 6  | 25     | 16        |    | 100    |           | 101 |
| 9      | 8         | 9  | 26     | 16        |    | 101    | 101       | 102 |
| 10     | 9         | 10 | 64     | 67        | 65 | 102    | 101       | 103 |
| 11     |           | 10 | 65     | 65        |    | 103    |           | 102 |
| 12     | 10        | 65 | 68     | 6         | 7  | 104    | 102       |     |
| 13     | 10        | 67 | 69     | 7         | 8  | 105    | 102       | 103 |
| 14     | 11        | 14 | 70     | 67        | 11 | 106    | 103       |     |
| 15     | 11        | 12 | 72     |           | 71 |        |           |     |

Calculation mode : Sequential

Flash algorithm : Normal

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38

Equipment Calculation Sequence

|    |    |    |    |    |    |    |     |     |     |   |   |   |   |   |   |   |    |
|----|----|----|----|----|----|----|-----|-----|-----|---|---|---|---|---|---|---|----|
| 2  | 4  | 15 | 16 | 71 | 75 | 99 | 101 | 102 | 103 | 1 | 3 | 5 | 6 | 7 | 8 | 9 | 10 |
| 67 | 11 | 14 | 66 | 12 | 13 | 65 |     |     |     |   |   |   |   |   |   |   |    |

Equipment Recycle Sequence

|   |   |   |   |   |   |   |    |    |    |    |    |
|---|---|---|---|---|---|---|----|----|----|----|----|
| 1 | 3 | 5 | 6 | 7 | 8 | 9 | 10 | 67 | 11 | 14 | 66 |
|---|---|---|---|---|---|---|----|----|----|----|----|

Recycle Cut Streams

3

Recycle Convergence Method: Direct Substitution

Max. loop iterations 40

Recycle Convergence Tolerance

|             |            |
|-------------|------------|
| Flow rate   | 1.000E-003 |
| Temperature | 1.000E-003 |
| Pressure    | 1.000E-003 |
| Enthalpy    | 1.000E-003 |
| Vapor frac. | 1.000E-003 |

Recycle calculation has converged.

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38

| Overall Mass Balance | kmol/h   |          | kg/h       |            |
|----------------------|----------|----------|------------|------------|
|                      | Input    | Output   | Input      | Output     |
| Methanol             | 256.662  | 109.329  | 8223.975   | 3503.127   |
| Oxygen               | 1630.813 | 1558.081 | 52184.399  | 49857.026  |
| Formaldehyde         | 250.558  | 397.891  | 7523.257   | 11947.076  |
| Water                | 628.299  | 773.764  | 11318.800  | 13939.361  |
| Hydrogen             | 1.655    | 3.523    | 3.336      | 7.101      |
| Nitrogen             | 6249.559 | 6249.560 | 175075.155 | 175075.168 |
| Total                | 9017.547 | 9092.147 | 254329.000 | 254329.000 |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38

## COMPONENTS

|   | ID # | Name         | Formula |
|---|------|--------------|---------|
| 1 | 117  | Methanol     | CH4O    |
| 2 | 47   | Oxygen       | O2      |
| 3 | 114  | Formaldehyde | CH2O    |
| 4 | 62   | Water        | H2O     |
| 5 | 1    | Hydrogen     | H2      |
| 6 | 46   | Nitrogen     | N2      |

## THERMODYNAMICS

K-value model : ESDK  
Enthalpy model : Latent Heat  
Liquid density : Library

Std vapor rate reference temperature is 0 C.  
Atmospheric pressure is 101.3250 kPa.

## ESDK Parameters:

| I | J | Kij      | Kji      | KijT       | KjiT       |
|---|---|----------|----------|------------|------------|
| 1 | 3 | -0.15000 | -0.15000 | 0.00000    | 0.00000    |
| 1 | 4 | 0.01550  | 0.01550  | 0.00000    | 0.00000    |
| 1 | 5 | -0.08500 | -0.08500 | 0.00000    | 0.00000    |
| 1 | 6 | 0.17270  | 0.17270  | 0.00000    | 0.00000    |
| 2 | 6 | -0.00300 | -0.00300 | 0.00000    | 0.00000    |
| 3 | 4 | 0.34000  | 0.34000  | -111.00000 | -111.00000 |
| 4 | 5 | -0.00700 | -0.00700 | 0.00000    | 0.00000    |
| 4 | 6 | 0.24500  | 0.24500  | 0.00000    | 0.00000    |
| 5 | 6 | -0.02220 | -0.02220 | 0.00000    | 0.00000    |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
EQUIPMENT SUMMARIES

## Pump Summary

|                       |           |           |
|-----------------------|-----------|-----------|
| Equip. No.            | 1         | 12        |
| Name                  |           |           |
| Output pressure kPa   | 300.0000  | 350.0000  |
| Efficiency            | 0.8000    | 0.7500    |
| Calculated power MJ/h | 0.9941    | 1.5183    |
| Calculated Pout kPa   | 300.0000  | 350.0000  |
| Head m                | 25.9730   | 29.7754   |
| Vol. flow rate m3/h   | 4.0002    | 5.6897    |
| Mass flow rate kg/h   | 3120.1714 | 3897.0942 |

## Compressor Summary

|                            |          |
|----------------------------|----------|
| Equip. No.                 | 2        |
| Name                       |          |
| Pressure out kPa           | 300.0000 |
| Type of Compressor         | 1        |
| Efficiency                 | 0.7000   |
| Actual power MJ/h          | 657.5833 |
| Cp/Cv                      | 1.4003   |
| Theoretical power (MJ/h)   | 460.3083 |
| Ideal Cp/Cv                | 1.3989   |
| Calc Pout kPa              | 300.0000 |
| Calc. mass flowrate (kg/h) | 4211     |

## Heat Exchanger Summary

|                     |           |          |            |           |
|---------------------|-----------|----------|------------|-----------|
| Equip. No.          | 3         | 4        | 8          | 9         |
| Name                |           |          |            |           |
| 1st Stream dp kPa   | 35.0000   | 35.0000  | 35.0000    | 35.0000   |
| 1st Stream T Out C  | 150.0000  | 200.0000 | 200.0000   | 100.0000  |
| Calc Ht Duty MJ/h   | 4110.0771 | 76.7449  | -8928.3965 | -983.1221 |
| LMTD Corr Factor    | 1.0000    | 1.0000   | 1.0000     | 1.0000    |
| 1st Stream Pout kPa | 265.0000  | 265.0000 | 185.0000   | 150.0000  |

|                     |            |
|---------------------|------------|
| Equip. No.          | 13         |
| Name                |            |
| 1st Stream dp kPa   | 35.0000    |
| 1st Stream T Out C  | 35.0000    |
| Calc Ht Duty MJ/h   | -1169.7457 |
| LMTD Corr Factor    | 1.0000     |
| 1st Stream Pout kPa | 315.0000   |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
EQUIPMENT SUMMARIES

## Mixer Summary

|                     |          |    |          |     |
|---------------------|----------|----|----------|-----|
| Equip. No.          | 5        | 65 | 66       | 103 |
| Name                |          |    |          |     |
| Output Pressure kPa | 255.0000 |    | 101.3250 |     |

## Reactor Summary

|                              |              |            |
|------------------------------|--------------|------------|
| Equip. No.                   | 6            | 7          |
| Name                         |              |            |
| Thermal mode                 | 1            | 1          |
| Temperature C                | 934.7972     | 923.0992   |
| Key Component                | 2            | 1          |
| Frac. Conversion             | 0.9950       | 0.0500     |
| Reactor Pressure kPa         |              | 220.0000   |
| Calc H of Reac.<br>(kJ/kmol) | -313560.0000 | 85040.0000 |

## Stoichiometrics:

|              |            |            |
|--------------|------------|------------|
| Methanol     | -1.000     | -1.000     |
| Oxygen       | -0.500     | 0.000E+000 |
| Formaldehyde | 1.000      | 1.000      |
| Water        | 1.000      | 0.000E+000 |
| Hydrogen     | 0.000E+000 | 1.000      |

## Scds Rigorous Distillation Summary

|                              |          |             |          |
|------------------------------|----------|-------------|----------|
| Equip. No.                   | 10       | 75          | 102      |
| Name                         |          |             |          |
| No. of stages                | 20       | 24          | 20       |
| 1st feed stage               | 1        | 14          | 1        |
| 2nd feed stage               | 20       | 0           | 20       |
| Condenser mode               | 0        | 7           | 0        |
| Condenser spec               |          | 0.9500      |          |
| Cond comp i pos.             | 0        | 1           | 0        |
| Reboiler mode                | 0        | 7           | 0        |
| Reboiler spec.               |          | 0.9500      |          |
| Reboiler comp i              | 0        | 4           | 0        |
| Colm press drop kPa          | 10.0000  | 20.0000     | 10.0000  |
| Est. dist. rate<br>(kmol/h)  | 320.0000 | 20.0000     | 120.0000 |
| Est. reflux rate<br>(kmol/h) |          | 260.0000    |          |
| Est. T top C                 | 150.0000 | 71.0000     | 50.0000  |
| Est. T bottom C              | 50.0000  | 106.0000    | 50.0000  |
| Top pressure kPa             | 140.0000 | 130.0000    | 105.0000 |
| Calc Cond duty MJ/h          |          | -32465.8008 |          |
| Calc Reblr duty MJ/h         |          | 32748.3359  |          |
| Iterations                   | 0        | 100         | 0        |
| Calc Reflux mole             | 157.4460 | 864.0048    | 71.6312  |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
EQUIPMENT SUMMARIES

(kmol/h)

|                       |           |            |           |
|-----------------------|-----------|------------|-----------|
| Calc Reflux ratio     |           | 37.4236    |           |
| Calc Reflux mass kg/h | 2876.5950 | 24629.4746 | 1302.0605 |
| Column type           | 1         | 0          | 0         |
| Column diameter m     | 0.8718    |            |           |
| Thickness (top) m     | 0.0016    |            |           |
| Thickness (bot) m     | 0.0143    |            |           |
| No of sections        | 1         | 0          | 0         |
| Calc. tolerance       | 0.0012    | 0.0012     | 0.0003    |

### Shortcut Distillation Summary

|                          |            |
|--------------------------|------------|
| Equip. No.               | 11         |
| Name                     |            |
| Mode                     | 3          |
| Light key component      | 1.0000     |
| Light key split          | 0.9500     |
| Heavy key component      | 4.0000     |
| Heavy key split          | 0.0500     |
| R/Rmin                   | 1.3000     |
| Number of stages         | 18.0345    |
| Min. No. of stages       | 9.7624     |
| Feed stage               | 9.6609     |
| Condenser duty MJ/h      | -6779.3984 |
| Reboiler duty MJ/h       | 7062.0737  |
| Colm pressure kPa        | 130.0000   |
| Reflux ratio, minimum    | 5.4205     |
| Calc. Reflux ratio       | 7.0467     |
| Number of points         | 10         |
| Lower bound R/Rmin       | 1.1000     |
| Upper bound R/Rmin       | 2.0000     |
| Colm pressure drop (kPa) | 20.0000    |

### Shortcut Distillation # 11 Case Studies:

| R/Rmin | Reflux ratio | No. of stgs | Feed stg | Qcond<br>MJ/h | Qreb<br>MJ/h |
|--------|--------------|-------------|----------|---------------|--------------|
| 1.10   | 5.96         | 23.1        | 12.2     | -5.866E+003   | 6.149E+003   |
| 1.20   | 6.51         | 19.9        | 10.6     | -6.323E+003   | 6.606E+003   |
| 1.30   | 7.05         | 18.0        | 9.7      | -6.780E+003   | 7.063E+003   |
| 1.40   | 7.59         | 16.8        | 9.0      | -7.236E+003   | 7.518E+003   |
| 1.50   | 8.13         | 15.8        | 8.5      | -7.692E+003   | 7.975E+003   |
| 1.60   | 8.67         | 15.1        | 8.2      | -8.149E+003   | 8.432E+003   |
| 1.70   | 9.22         | 14.6        | 7.9      | -8.606E+003   | 8.889E+003   |
| 1.80   | 9.76         | 14.1        | 7.7      | -9.063E+003   | 9.346E+003   |
| 1.90   | 10.30        | 13.8        | 7.5      | -9.520E+003   | 9.803E+003   |
| 2.00   | 10.84        | 13.5        | 7.3      | -9.977E+003   | 1.026E+004   |



Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
EQUIPMENT SUMMARIES

## Valve Summary

|                  |          |
|------------------|----------|
| Equip. No.       | 14       |
| Name             |          |
| Pressure out kPa | 120.0000 |

## Flash Summary

|                |    |    |            |
|----------------|----|----|------------|
| Equip. No.     | 15 | 16 | 101        |
| Name           |    |    |            |
| Flash Mode     | 0  | 0  | 2          |
| Param 1        |    |    | 50.0000    |
| Param 2        |    |    | 115.0000   |
| Heat duty MJ/h |    |    | -5083.4517 |

## K values:

|              |          |           |           |
|--------------|----------|-----------|-----------|
| Methanol     | 1.184    | 1.325     | 0.226     |
| Oxygen       | 1163.982 | 4151.597  | 5964.569  |
| Formaldehyde | 0.040    | 1.023     | 0.020     |
| Water        | 0.468    | 0.986     | 0.164     |
| Hydrogen     | 5425.908 | 22992.754 | 40958.988 |
| Nitrogen     | 4822.877 | 19785.629 | 30731.762 |

## Component Separator Summary

|                 |        |
|-----------------|--------|
| Equip. No.      | 67     |
| Name            |        |
| Component No. 2 | 1.0000 |
| Component No. 5 | 1.0000 |
| Component No. 6 | 1.0000 |

## Tower Rigorous Distillation Summary

|                          |             |
|--------------------------|-------------|
| Equip. No.               | 71          |
| Name                     |             |
| No. of stages            | 24          |
| 1st feed stage           | 14          |
| Top pressure kPa         | 130.0000    |
| Colm pressure drop (kPa) | 20.0000     |
| Condenser mode           | 7           |
| Condenser spec.          | 0.9500      |
| Cond. comp i             | 1           |
| Reboiler mode            | 7           |
| Reboiler spec.           | 0.9500      |
| Reboiler comp i          | 4           |
| Iterations               | 100         |
| Calc cond duty MJ/h      | -32370.5156 |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
EQUIPMENT SUMMARIES

|                           |            |
|---------------------------|------------|
| Calc rebr duty MJ/h       | 32654.2363 |
| Est. Dist. rate (kmol/h)  | 20.0000    |
| Est. Reflux rate (kmol/h) | 260.0000   |
| Est. T top C              | 70.0000    |
| Est. T bottom C           | 103.0000   |
| Tray type                 | 3          |
| Column diameter m         | 2.2860     |
| Tray space m              | 0.6096     |
| Thickness (top) m         | 0.0016     |
| Thickness (bot) m         | 0.0032     |
| No of sections            | 1          |
| Calc Reflux ratio         | 37.3041    |
| Calc Reflux mole (kmol/h) | 861.2657   |
| Calc Reflux mass kg/h     | 24551.4297 |

## Kinetic Reactor Summary

|                             |             |
|-----------------------------|-------------|
| Equip. No.                  | 99          |
| Name                        |             |
| Reactor type                | 2           |
| Reaction phase              | 1           |
| Thermal mode                | 1           |
| Pressure In kPa             | 250.0000    |
| Tout C                      | 354.2749    |
| Q MJ/h                      | -11896.6875 |
| Reactor volume m3           | 32.4871     |
| Concentration Flag          | 1           |
| Specify calc. mode          | 1           |
| Conversion                  | 0.9000      |
| Key                         | 1           |
| No. of Reactions            | 2           |
| Molar Flow Unit             | 2           |
| Activ. E/H of Rxn Unit      | 6           |
| Volume Unit                 | 1           |
| Overall IG Ht of Rxn (MJ/h) | -13222.2568 |
| Mass unit                   | 2           |
| Partial P unit              | 2           |

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Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
FLOW SUMMARIES

| Stream No.          | 1           | 2        | 3        | 4        |
|---------------------|-------------|----------|----------|----------|
| Stream Name         |             |          |          |          |
| Temp C              | 25.0000     | 30.0000  | 40.6637  | 40.7822  |
| Pres kPa            | 101.3250    | 120.0000 | 101.3250 | 300.0000 |
| Enth MJ/h           | 3.6966E-005 | -18349.  | -24019.  | -24018.  |
| Vapor mole fraction | 1.0000      | 0.00000  | 0.00000  | 0.00000  |
| Total kmol/h        | 145.94      | 76.92    | 99.92    | 99.92    |
| Flowrates in kmol/h |             |          |          |          |
| Methanol            | 0.00        | 76.92    | 94.11    | 94.11    |
| Oxygen              | 30.66       | 0.00     | 0.00     | 0.00     |
| Formaldehyde        | 0.00        | 0.00     | 0.00     | 0.00     |
| Water               | 0.00        | 0.00     | 5.81     | 5.81     |
| Hydrogen            | 0.00        | 0.00     | 0.00     | 0.00     |
| Nitrogen            | 115.28      | 0.00     | 0.00     | 0.00     |
| Stream No.          | 5           | 6        | 7        | 8        |
| Stream Name         |             |          |          |          |
| Temp C              | 183.0128    | 150.0000 | 200.0000 | 171.9390 |
| Pres kPa            | 300.0000    | 265.0000 | 265.0000 | 255.0000 |
| Enth MJ/h           | 657.58      | -19908.  | 734.33   | -19173.  |
| Vapor mole fraction | 1.0000      | 1.0000   | 1.0000   | 1.0000   |
| Total kmol/h        | 145.94      | 99.92    | 145.94   | 245.86   |
| Flowrates in kmol/h |             |          |          |          |
| Methanol            | 0.00        | 94.11    | 0.00     | 94.11    |
| Oxygen              | 30.66       | 0.00     | 30.66    | 30.66    |
| Formaldehyde        | 0.00        | 0.00     | 0.00     | 0.00     |
| Water               | 0.00        | 5.81     | 0.00     | 5.81     |
| Hydrogen            | 0.00        | 0.00     | 0.00     | 0.00     |
| Nitrogen            | 115.28      | 0.00     | 115.28   | 115.28   |
| Stream No.          | 9           | 10       | 11       | 12       |
| Stream Name         |             |          |          |          |
| Temp C              | 200.0000    | 100.0000 | 30.0000  | 84.5653  |
| Pres kPa            | 185.0000    | 150.0000 | 150.0000 | 140.0000 |
| Enth MJ/h           | -28102.     | -29085.  | -40801.  | -24935.  |
| Vapor mole fraction | 1.0000      | 1.0000   | 0.00000  | 1.0000   |
| Total kmol/h        | 278.02      | 278.02   | 143.00   | 224.15   |
| Flowrates in kmol/h |             |          |          |          |
| Methanol            | 31.44       | 31.44    | 0.00     | 13.35    |
| Oxygen              | 0.15        | 0.15     | 0.00     | 0.15     |
| Formaldehyde        | 62.67       | 62.67    | 0.00     | 0.04     |
| Water               | 66.82       | 66.82    | 143.00   | 93.68    |
| Hydrogen            | 1.65        | 1.65     | 0.00     | 1.65     |
| Nitrogen            | 115.28      | 115.28   | 0.00     | 115.28   |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38

## FLOW SUMMARIES

| Stream No.          | 13       | 14        | 15          | 16       |
|---------------------|----------|-----------|-------------|----------|
| Stream Name         |          |           |             |          |
| Temp C              | 89.8453  | 75.4568   | 106.6348    | 106.7102 |
| Pres kPa            | 150.0000 | 130.0000  | 150.0000    | 350.0000 |
| Enth MJ/h           | -44950.  | -5669.7   | -38998.     | -38996.  |
| Vapor mole fraction | 0.00000  | 0.00000   | 0.00000     | 0.00000  |
| Total kmol/h        | 196.87   | 23.00     | 173.87      | 173.87   |
| Flowrates in kmol/h |          |           |             |          |
| Methanol            | 18.10    | 17.19     | 0.90        | 0.90     |
| Oxygen              | 0.00     | 0.00      | 0.00        | 0.00     |
| Formaldehyde        | 62.63    | 0.00      | 62.63       | 62.63    |
| Water               | 116.14   | 5.81      | 110.33      | 110.33   |
| Hydrogen            | 0.00     | 0.00      | 0.00        | 0.00     |
| Nitrogen            | 0.00     | 0.00      | 0.00        | 0.00     |
| Stream No.          | 17       | 18        | 21          | 22       |
| Stream Name         |          |           |             |          |
| Temp C              | 35.0000  | 73.3569   | 75.5718     | 75.5718  |
| Pres kPa            | 315.0000 | 120.0000  | 130.0000    | 130.0000 |
| Enth MJ/h           | -40166.  | -5669.7   | -5682.9     | -0.18970 |
| Vapor mole fraction | 0.00000  | 0.0052328 | 4.0129E-005 | 1.0000   |
| Total kmol/h        | 173.87   | 23.00     | 23.10       | 0.00     |
| Flowrates in kmol/h |          |           |             |          |
| Methanol            | 0.90     | 17.19     | 17.20       | 0.00     |
| Oxygen              | 0.00     | 0.00      | 0.00        | 0.00     |
| Formaldehyde        | 62.63    | 0.00      | 0.09        | 0.00     |
| Water               | 110.33   | 5.81      | 5.81        | 0.00     |
| Hydrogen            | 0.00     | 0.00      | 0.00        | 0.00     |
| Nitrogen            | 0.00     | 0.00      | 0.00        | 0.00     |
| Stream No.          | 23       | 24        | 25          | 26       |
| Stream Name         |          |           |             |          |
| Temp C              | 75.5718  | 106.6357  | 106.6357    | 106.6357 |
| Pres kPa            | 130.0000 | 150.0000  | 150.0000    | 150.0000 |
| Enth MJ/h           | -5682.7  | -33615.   | -33615.     | 0.00000  |
| Vapor mole fraction | 0.00000  | 1.0000    | 1.0000      | 0.00000  |
| Total kmol/h        | 23.09    | 173.77    | 173.77      | 0.00     |
| Flowrates in kmol/h |          |           |             |          |
| Methanol            | 17.20    | 0.91      | 0.91        | 0.00     |
| Oxygen              | 0.00     | 0.00      | 0.00        | 0.00     |
| Formaldehyde        | 0.09     | 62.54     | 62.54       | 0.00     |
| Water               | 5.81     | 110.32    | 110.32      | 0.00     |
| Hydrogen            | 0.00     | 0.00      | 0.00        | 0.00     |
| Nitrogen            | 0.00     | 0.00      | 0.00        | 0.00     |

Job Name: formalin\_JAS\_2008 Date: 12/14/2008 Time: 14:45:38  
FLOW SUMMARIES

| Stream No.          | 64        | 65       | 68       | 69       |
|---------------------|-----------|----------|----------|----------|
| Stream Name         |           |          |          |          |
| Temp C              | 89.8453   | 84.5669  | 934.7972 | 923.0992 |
| Pres kPa            | 150.0000  | 140.0000 | 255.0000 | 220.0000 |
| Enth MJ/h           | 0.0083595 | -24935.  | -19173.  | -19173.  |
| Vapor mole fraction | 1.0000    | 1.0000   | 1.0000   | 1.0000   |
| Total kmol/h        | 0.00      | 224.16   | 276.37   | 278.02   |
| Flowrates in kmol/h |           |          |          |          |
| Methanol            | 0.00      | 13.35    | 33.10    | 31.44    |
| Oxygen              | 0.00      | 0.15     | 0.15     | 0.15     |
| Formaldehyde        | 0.00      | 0.04     | 61.01    | 62.67    |
| Water               | 0.00      | 93.68    | 66.82    | 66.82    |
| Hydrogen            | 0.00      | 1.65     | 0.00     | 1.65     |
| Nitrogen            | 0.00      | 115.28   | 115.28   | 115.28   |

| Stream No.          | 70       | 72       | 73       | 74       |
|---------------------|----------|----------|----------|----------|
| Stream Name         |          |          |          |          |
| Temp C              | 89.8453  | 89.8453  | 75.5764  | 106.6356 |
| Pres kPa            | 150.0000 | 150.0000 | 130.0000 | 150.0000 |
| Enth MJ/h           | -44950.  | -44950.  | -5681.2  | -38985.  |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000  | 0.00000  |
| Total kmol/h        | 196.86   | 196.86   | 23.09    | 173.78   |
| Flowrates in kmol/h |          |          |          |          |
| Methanol            | 18.10    | 18.10    | 17.19    | 0.90     |
| Oxygen              | 0.00     | 0.00     | 0.00     | 0.00     |
| Formaldehyde        | 62.63    | 62.63    | 0.09     | 62.54    |
| Water               | 116.14   | 116.14   | 5.81     | 110.33   |
| Hydrogen            | 0.00     | 0.00     | 0.00     | 0.00     |
| Nitrogen            | 0.00     | 0.00     | 0.00     | 0.00     |

| Stream No.          | 76       | 77       | 78       | 98       |
|---------------------|----------|----------|----------|----------|
| Stream Name         |          |          |          |          |
| Temp C              | 89.8453  | 75.5772  | 106.6358 | 350.0000 |
| Pres kPa            | 150.0000 | 130.0000 | 150.0000 | 250.0000 |
| Enth MJ/h           | -44950.  | -5680.9  | -38986.  | 55571.   |
| Vapor mole fraction | 0.00000  | 0.00000  | 0.00000  | 1.0000   |
| Total kmol/h        | 196.86   | 23.09    | 173.78   | 7713.00  |
| Flowrates in kmol/h |          |          |          |          |
| Methanol            | 18.10    | 17.19    | 0.90     | 94.00    |
| Oxygen              | 0.00     | 0.00     | 0.00     | 1600.00  |
| Formaldehyde        | 62.63    | 0.09     | 62.54    | 0.00     |
| Water               | 116.14   | 5.81     | 110.33   | 0.00     |
| Hydrogen            | 0.00     | 0.00     | 0.00     | 0.00     |
| Nitrogen            | 0.00     | 0.00     | 0.00     | 6019.00  |

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FLOW SUMMARIES

| Stream No.          | 99       | 100      | 101      | 102      |
|---------------------|----------|----------|----------|----------|
| Stream Name         |          |          |          |          |
| Temp C              | 354.2749 | 100.0000 | 50.0000  | 50.0000  |
| Pres kPa            | 250.0000 | 150.0000 | 115.0000 | 115.0000 |
| Enth MJ/h           | 43675.   | -29085.  | -3209.2  | -30959.  |
| Vapor mole fraction | 1.0000   | 1.0000   | 1.0000   | 0.00000  |
| Total kmol/h        | 7755.44  | 278.02   | 132.22   | 145.81   |
| Flowrates in kmol/h |          |          |          |          |
| Methanol            | 9.34     | 31.44    | 5.36     | 26.09    |
| Oxygen              | 1557.77  | 0.15     | 0.15     | 0.00     |
| Formaldehyde        | 84.66    | 62.67    | 1.11     | 61.56    |
| Water               | 84.45    | 66.82    | 8.66     | 58.16    |
| Hydrogen            | 0.21     | 1.65     | 1.65     | 0.00     |
| Nitrogen            | 6019.00  | 115.28   | 115.28   | 0.00     |
| Stream No.          | 103      | 104      | 105      | 106      |
| Stream Name         |          |          |          |          |
| Temp C              | 30.0000  | 42.5463  | 42.3507  | 47.7261  |
| Pres kPa            | 150.0000 | 105.0000 | 115.0000 | 115.0000 |
| Enth MJ/h           | -19992.  | -3029.3  | -20172.  | -51131.  |
| Vapor mole fraction | 0.00000  | 1.0000   | 0.00000  | 0.00000  |
| Total kmol/h        | 70.07    | 130.12   | 72.17    | 217.98   |
| Flowrates in kmol/h |          |          |          |          |
| Methanol            | 0.00     | 1.39     | 3.96     | 30.05    |
| Oxygen              | 0.00     | 0.15     | 0.00     | 0.00     |
| Formaldehyde        | 0.00     | 0.00     | 1.11     | 62.67    |
| Water               | 70.07    | 11.64    | 67.09    | 125.25   |
| Hydrogen            | 0.00     | 1.65     | 0.00     | 0.00     |
| Nitrogen            | 0.00     | 115.27   | 0.00     | 0.01     |

13.18 The Thermodynamic models used are:

| Process          | Enthalpy    | Phase Equ.    | Alternate                             | Notes                          |
|------------------|-------------|---------------|---------------------------------------|--------------------------------|
| DME              | Latent heat | UNIFAC        | UNIQUAC/UNIFAC                        | 2 liquid phases                |
| EB               | Latent heat | UNIFAC        | UNIQUAC/UNIFAC                        | Lack of BIPs                   |
| Styrene          | SRK         | SRK           | PR                                    | HC                             |
| Maleic Anhydride | Latent heat | Raoult's Law  | SRK                                   | Fairly ideal mixtures          |
| Ethylene Oxide   | SRK         | PSRK + UNIFAC | SRK + UNIFAC                          | Simple gases except in columns |
| Formalin         | Latent heat | ESDK          | Polynomial data                       | Hydrogen bonding               |
| Acetone          | Latent heat | UNIFAC        | UNIQUAC with BIPs regressed from data | Azeotrope                      |
| Acrylic Acid     | Latent heat | UNIFAC        | UNIQUAC with BIPs regressed from data | LLE and VLE                    |
| Heptenes         | SRK         | SRK           | PR                                    | HC                             |

13.19 The following results are from Chemcad for vapor-liquid equilibrium of water and air:

| T    | Henry's Law                    | SRK                             |
|------|--------------------------------|---------------------------------|
| 10°C | 11.4 ppm O <sub>2</sub> (mass) | 0.413 ppm O <sub>2</sub> (mass) |
| 27°C | 7.60 ppm O <sub>2</sub> (mass) | 0.659 ppm O <sub>2</sub> (mass) |
|      |                                |                                 |

Note that the Henry's Law results are close to reality, but the SRK results are not only off by more than an order of magnitude and their temperature-dependence is inverse. This is to be expected, since cubic equations of state do poorly for liquid phases and for water.

13.20 The following is from the Chemcad simulation of T-101:

|                  | SRK           | PR            | PR Bips = zero |
|------------------|---------------|---------------|----------------|
| Number of stages | 23.3          | 23.5          | 23.5           |
| Reboiler duty    | 8.78 MMBtu/hr | 8.72 MMBtu/hr | 8.72 MMBtu/hr  |
| Condenser duty   | 5.15 MMBtu/hr | 5.15 MMBtu/hr | 5.15 MMBtu/hr  |
| Reflux Ratio     | 1.75          | 1.77          | 1.77           |

| SRK BIPs | Hydrogen | Methane  | Benzene | Toluene |
|----------|----------|----------|---------|---------|
| Hydrogen | 0        | -0.00900 | 0.32358 | 0.32358 |
| Methane  | -0.00900 | 0        | 0       | 0       |
| Benzene  | 0.32358  | 0        | 0       | 0       |
| Toluene  | 0.32358  | 0        | 0       | 0       |

| PR BIPs  | Hydrogen | Methane | Benzene  | Toluene  |
|----------|----------|---------|----------|----------|
| Hydrogen | 0        | 0.0156  | -0.50000 | -0.50000 |
| Methane  | 0.0156   | 0       | 0.03630  | 0.04000  |
| Benzene  | -0.50000 | 0.03630 | 0        | 0        |
| Toluene  | -0.50000 | 0.04000 | 0        | 0        |

Note that the BIPs are very different for the SKR and PR models, even differences in sign! Note also that there are more binaries represented in the PR database on ChemCAD for this system than for SRK.

However, the major outcome is that it makes no noticeable difference for the column. This is because the important BIPs (between the heavy and light keys, benzene and toluene) are zero in all three cases.

As noted, T-101 was modeled with the shortcut column module (heavy and light key recoveries in distillate of 0.01 and 0.99, R/Rmin of 1.5, partial condenser) followed by a splitter for the distillate. The feed stream to the column was specified, as well as the top and bottom pressures.

The BIPs for PR in ChemCAD can be changed only by creating "cloned" components. The BIPs for the original components in the database cannot be changed.



## Chapter 14

### 14.1 Describe a Pareto analysis. When is it used?

Strictly speaking, a **Pareto analysis** is statistical technique but we use it here in the more common form of the 20-80 principle. Namely, for many things, 80% of the information is associated with only 20% of the issues. Thus when performing an optimization of a process or product, if we itemize all the contributing costs then most often approx. 80% of those costs are associated with only about 20% of the variables. This is very useful to do early in an optimization analysis since it focuses our attention on the important decision variables.

### 14.2 What is the difference between parametric optimization and topological optimization? List one example of each.

**Parametric optimization** focuses on adjusting operating (decision) variables in order to improve the objective function. Examples include, adjusting the T and P at which a reactor operates, or adjusting the surface area of a heat exchanger or number of trays for a distillation column.

**Topological optimization** focuses on adjusting the layout or topology of the flowsheet in order to improve the objective function. Examples include, changing the order in which a separation sequence is implemented, looking at the effect of adding a heat recovery exchanger, or changing a utility (cw to refrigerated water).

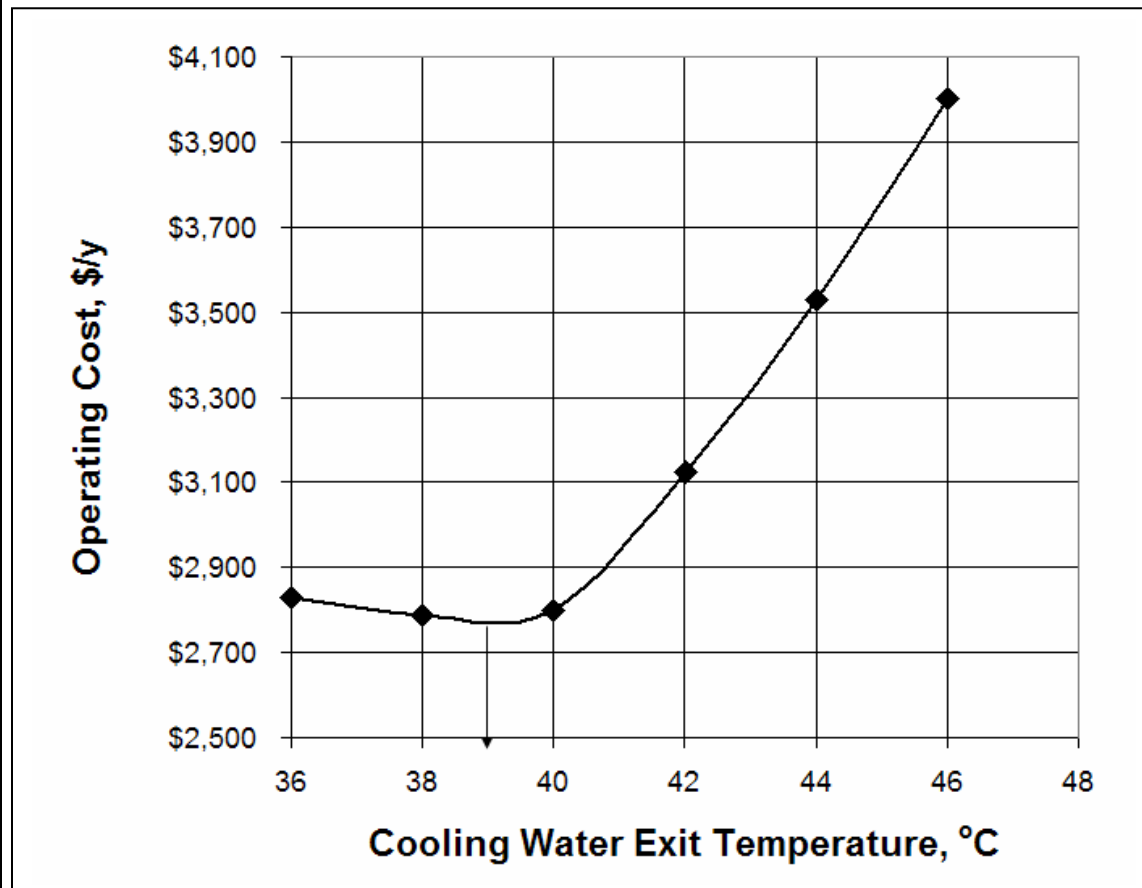
### 14.3 What is an objective function? Give two examples of one.

An **objective function** (OF) is a mathematical relationship that one wishes to minimize (cost) or maximize (profit). For chemical processes, the OF will most often be a function of variables that relate to the economics of the processes. For example, maximizing the net present value (NPV) or minimizing the equivalent annual operating (EAOC) cost are examples of common objective functions. We can also talk about maximizing conversion or yield but such OFs may not lead to the economic optimum that is the commercial goal of all processes.

## 14.4

## Problem 4

|              |      | Cost per cleaning = | \$2,400          |                 |                   |
|--------------|------|---------------------|------------------|-----------------|-------------------|
| $T_{CW,out}$ | $FR$ | Cleanings per year  | Cleaning cost/yr | Pumping Cost/yr | Total yearly cost |
| 36           | 1.11 | 0.9009              | \$ 2,162         | \$ 667          | \$ 2,829          |
| 38           | 1.05 | 0.9524              | \$ 2,286         | \$ 500          | \$ 2,786          |
| 40           | 1.00 | 1.0000              | \$ 2,400         | \$ 400          | \$ 2,800          |
| 42           | 0.86 | 1.1628              | \$ 2,791         | \$ 333          | \$ 3,124          |
| 44           | 0.74 | 1.3514              | \$ 3,243         | \$ 286          | \$ 3,529          |
| 46           | 0.64 | 1.5625              | \$ 3,750         | \$ 250          | \$ 4,000          |



Optimal cooling water exit temperature = 39°C

## 14.5 – Background simulations from CHEMCAD

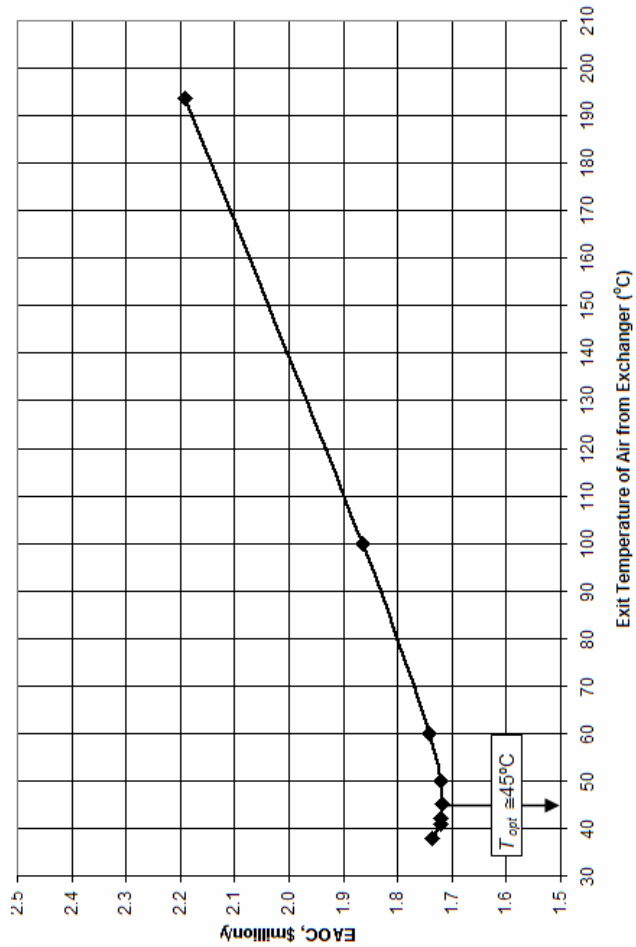
|                                                                                     |           |           |           |                     |           |          |                     |                     |          |           |  |  |  |
|-------------------------------------------------------------------------------------|-----------|-----------|-----------|---------------------|-----------|----------|---------------------|---------------------|----------|-----------|--|--|--|
|  |           |           |           |                     |           |          |                     |                     |          |           |  |  |  |
| Stream No.                                                                          | 3         | 4         | 5         |                     |           |          | Stream No.          | 3                   | 4        | 5         |  |  |  |
| Stream Name                                                                         |           |           |           |                     |           |          | Stream Name         |                     |          |           |  |  |  |
| Temp C                                                                              | 45.0      | 45.0      | 45.0      | 40.0000             | 40.0000   | 40.0000  | Temp C              | 35.0000             | 35.0000  | 35.0000   |  |  |  |
| Pres bar                                                                            | 1.6300    | 1.6300    | 1.6300    | 1.6300              | 1.6300    | 1.6300   | Pres bar            | 1.6300              | 1.6300   | 1.6300    |  |  |  |
| Enth MJ/h                                                                           | -14616    | -4098.0   | -10518    | -14796              | -3128.7   | -11668   | Enth MJ/h           | -14941              | -2403.8  | -12537    |  |  |  |
| Vapor mole fract                                                                    | 0.57557   | 1.0000    | 0.00000   | Vapor mole fraction | 0.52815   | 1.0000   | Vapor mole fraction | 0.49254             | 1.0000   | 0.00000   |  |  |  |
| Total kmol/h                                                                        | 92.6200   | 53.3092   | 39.3108   | Total kmol/h        | 92.6200   | 48.9169  | Total kmol/h        | 92.6200             | 45.6191  | 47.0009   |  |  |  |
| Total kg/h                                                                          | 2674.4008 | 1049.8997 | 1624.5012 | Total kg/h          | 2674.4008 | 824.7096 | Total kg/h          | 2674.4008           | 653.2615 | 2021.1394 |  |  |  |
| Total std L m3/h                                                                    | 4.1787    | 2.2181    | 1.9606    | Total std L m3/h    | 4.1787    | 1.9394   | Total std L m3/h    | 4.1787              | 1.7269   | 2.4518    |  |  |  |
| Total std V m3/h                                                                    | 2075.95   | 1194.85   | 881.10    | Total std V m3/h    | 2075.95   | 1096.41  | Total std V m3/h    | 2075.95             | 1022.49  | 1053.46   |  |  |  |
| Flowrates in kmol/h                                                                 |           |           |           | Flowrates in kmol/h |           |          |                     | Flowrates in kmol/h |          |           |  |  |  |
| Isopropanol                                                                         | 3.8600    | 0.7211    | 3.1389    | Isopropanol         | 3.8600    | 0.4880   | Isopropanol         | 3.8600              | 0.3358   | 3.5242    |  |  |  |
| Water                                                                               | 19.0400   | 2.4423    | 16.5977   | Water               | 19.0400   | 1.6841   | Water               | 19.0400             | 1.1751   | 17.8649   |  |  |  |
| Hydrogen                                                                            | 34.7800   | 34.7798   | 0.0002    | Hydrogen            | 34.7800   | 34.7797  | Hydrogen            | 34.7800             | 34.7797  | 0.0003    |  |  |  |
| Acetone                                                                             | 34.9400   | 15.3660   | 19.5740   | Acetone             | 34.9400   | 11.9651  | Acetone             | 34.9400             | 9.3285   | 25.6115   |  |  |  |
| Q = 0                                                                               | 7,900.992 |           |           | Q                   | 115.65    |          | Q                   | 259.9               |          |           |  |  |  |
|                                                                                     |           |           |           |                     |           |          |                     |                     |          |           |  |  |  |
|                                                                                     |           |           |           |                     |           |          |                     |                     |          |           |  |  |  |
| Stream No.                                                                          | 3         | 4         | 5         |                     |           |          | Stream No.          | 3                   | 4        | 5         |  |  |  |
| Stream Name                                                                         |           |           |           |                     |           |          | Stream Name         |                     |          |           |  |  |  |
| Temp C                                                                              | 30.0000   | 30.0000   | 30.0000   | 25.0000             | 25.0000   | 25.0000  | Temp C              | 20.0000             | 20.0000  | 20.0000   |  |  |  |
| Pres bar                                                                            | 1.6300    | 1.6300    | 1.6300    | 1.6300              | 1.6300    | 1.6300   | Pres bar            | 1.6300              | 1.6300   | 1.6300    |  |  |  |
| Enth MJ/h                                                                           | -15059    | -1856.5   | -13202    | -15158              | -1439.2   | -13719   | Enth MJ/h           | -15243              | -1118.4  | -14124    |  |  |  |
| Vapor mole fraction                                                                 | 0.46557   | 1.0000    | 0.00000   | Vapor mole fraction | 0.44492   | 1.0000   | Vapor mole fraction | 0.42899             | 1.0000   | 0.00000   |  |  |  |
| Total kmol/h                                                                        | 92.6200   | 43.1207   | 49.4993   | Total kmol/h        | 92.6200   | 41.2087  | Total kmol/h        | 92.6200             | 39.7331  | 52.8869   |  |  |  |
| Total kg/h                                                                          | 2674.4008 | 521.9570  | 2152.4440 | Total kg/h          | 2674.4008 | 420.5446 | Total kg/h          | 2674.4008           | 341.6536 | 2332.7472 |  |  |  |
| Total std L m3/h                                                                    | 4.1787    | 1.5640    | 2.6147    | Total std L m3/h    | 4.1787    | 1.4380   | Total std L m3/h    | 4.1787              | 1.3400   | 2.8387    |  |  |  |
| Total std V m3/h                                                                    | 2075.95   | 966.49    | 1109.46   | Total std V m3/h    | 2075.95   | 923.64   | Total std V m3/h    | 2075.95             | 890.56   | 1185.39   |  |  |  |
| Flowrates in kmol/h                                                                 |           |           |           | Flowrates in kmol/h |           |          |                     | Flowrates in kmol/h |          |           |  |  |  |
| Isopropanol                                                                         | 3.8600    | 0.2335    | 3.6265    | Isopropanol         | 3.8600    | 0.1632   | Isopropanol         | 3.8600              | 0.1141   | 3.7459    |  |  |  |
| Water                                                                               | 19.0400   | 0.8256    | 18.2145   | Water               | 19.0400   | 0.5815   | Water               | 19.0400             | 0.4092   | 18.6308   |  |  |  |
| Hydrogen                                                                            | 34.7800   | 34.7796   | 0.0004    | Hydrogen            | 34.7800   | 34.7795  | Hydrogen            | 34.7800             | 34.7795  | 0.0005    |  |  |  |
| Acetone                                                                             | 34.9400   | 7.2821    | 27.6579   | Acetone             | 34.9400   | 5.6845   | Acetone             | 34.9400             | 4.4304   | 30.5096   |  |  |  |
| Q =                                                                                 | 378.1     |           |           | Q =                 | 477.1     |          | Q =                 | 562                 |          |           |  |  |  |

## 14.5 – Background Information from Chemcad

|                     |           |          |           |                     |           |          |           |                     |           |          |           |
|---------------------|-----------|----------|-----------|---------------------|-----------|----------|-----------|---------------------|-----------|----------|-----------|
| Stream No.          | 3         | 4        | 5         | Stream No.          | 3         | 4        | 5         | Stream No.          | 3         | 4        | 5         |
| Stream Name         |           |          |           | Stream Name         |           |          |           | Stream Name         |           |          |           |
| Temp C              | 15.0000   | 15.0000  | 15.0000   | Temp C              | 10.0000   | 10.0000  | 10.0000   | Temp C              | 7.5000    | 7.5000   | 7.5000    |
| Pres bar            | 1.6300    | 1.6300   | 1.6300    | Pres bar            | 1.6300    | 1.6300   | 1.6300    | Pres bar            | 1.6300    | 1.6300   | 1.6300    |
| Enth MJ/h           | -15317    | -870.55  | -14446.   | Enth MJ/h           | -15382    | -678.61  | -14704.   | Enth MJ/h           | -15413.   | -599.56  | -14813.   |
| Vapor mole fraction | 0.41663   | 1.0000   | 0.00000   | Vapor mole fraction | 0.40700   | 1.0000   | 0.00000   | Vapor mole fraction | 0.40301   | 1.0000   | 0.00000   |
| Total kmol/h        | 92.6200   | 38.5880  | 54.0320   | Total kmol/h        | 92.6200   | 37.6963  | 54.9237   | Total kmol/h        | 92.6200   | 37.3272  | 55.2928   |
| Total kg/h          | 2674.4008 | 279.9768 | 2394.4238 | Total kg/h          | 2674.4008 | 231.6270 | 2442.7736 | Total kg/h          | 2674.4008 | 211.5159 | 2462.8850 |
| Total std L m3/h    | 4.1787    | 1.2633   | 2.9154    | Total std L m3/h    | 4.1787    | 1.2031   | 2.9756    | Total std L m3/h    | 4.1787    | 1.1781   | 3.0006    |
| Total std V m3/h    | 2075.95   | 864.90   | 1211.05   | Total std V m3/h    | 2075.95   | 844.91   | 1231.04   | Total std V m3/h    | 2075.95   | 836.64   | 1239.31   |
| Flowrates in kmol/h |           |          |           | Flowrates in kmol/h |           |          |           | Flowrates in kmol/h |           |          |           |
| Isopropanol         | 3.8600    | 0.0795   | 3.7805    | Isopropanol         | 3.8600    | 0.0552   | 3.8048    | Isopropanol         | 3.8600    | 0.0458   | 3.8142    |
| Water               | 19.0400   | 0.2869   | 18.7531   | Water               | 19.0400   | 0.2000   | 18.8400   | Water               | 19.0400   | 0.1665   | 18.8735   |
| Hydrogen            | 34.7800   | 34.7794  | 0.0006    | Hydrogen            | 34.7800   | 34.7793  | 0.0007    | Hydrogen            | 34.7800   | 34.7793  | 0.0007    |
| Acetone             | 34.9400   | 3.4421   | 31.4979   | Acetone             | 34.9400   | 2.6619   | 32.2781   | Acetone             | 34.9400   | 2.3357   | 32.6043   |
| Q =                 | 636       |          |           | Q =                 | 701.6     |          |           | Q =                 | 731.9     |          |           |
|                     |           |          |           |                     |           |          |           |                     |           |          |           |
|                     |           |          |           |                     |           |          |           |                     |           |          |           |
|                     |           |          |           |                     |           |          |           |                     |           |          |           |
| Stream No.          | 3         | 4        | 5         | Stream No.          | 3         | 4        | 5         | Stream No.          |           |          |           |
| Stream Name         |           |          |           | Stream Name         |           |          |           | Stream Name         |           |          |           |
| Temp C              | 6.0000    | 6.0000   | 6.0000    | Temp C              | 5.1000    | 5.1000   | 5.1000    | Temp C              |           |          |           |
| Pres bar            | 1.6300    | 1.6300   | 1.6300    | Pres bar            | 1.6300    | 1.6300   | 1.6300    | Pres bar            |           |          |           |
| Enth MJ/h           | -15430.   | -556.79  | -14873.   | Enth MJ/h           | -15440    | -532.67  | -14908.   | Enth MJ/h           |           |          |           |
| Vapor mole fraction | 0.40085   | 1.0000   | 0.00000   | Vapor mole fraction | 0.39963   | 1.0000   | 0.00000   | Vapor mole fraction |           |          |           |
| Total kmol/h        | 92.6200   | 37.1269  | 55.4931   | Total kmol/h        | 92.6200   | 37.0137  | 55.6063   | Total kmol/h        |           |          |           |
| Total kg/h          | 2674.4008 | 200.5734 | 2473.8272 | Total kg/h          | 2674.4008 | 194.3823 | 2480.0185 | Total kg/h          |           |          |           |
| Total std L m3/h    | 4.1787    | 1.1644   | 3.0143    | Total std L m3/h    | 4.1787    | 1.1567   | 3.0220    | Total std L m3/h    |           |          |           |
| Total std V m3/h    | 2075.95   | 832.15   | 1243.80   | Total std V m3/h    | 2075.95   | 829.61   | 1246.34   | Total std V m3/h    |           |          |           |
| Flowrates in kmol/h |           |          |           | Flowrates in kmol/h |           |          |           | Flowrates in kmol/h |           |          |           |
| Isopropanol         | 3.8600    | 0.0409   | 3.8191    | Isopropanol         | 3.8600    | 0.0382   | 3.8218    | Isopropanol         |           |          |           |
| Water               | 19.0400   | 0.1489   | 18.8911   | Water               | 19.0400   | 0.1393   | 18.9007   | Water               |           |          |           |
| Hydrogen            | 34.7800   | 34.7792  | 0.0008    | Hydrogen            | 34.7800   | 34.7792  | 0.0008    | Hydrogen            |           |          |           |
| Acetone             | 34.9400   | 2.1578   | 32.7822   | Acetone             | 34.9400   | 2.0569   | 32.8830   | Acetone             |           |          |           |
| Q =                 | 749.3     |          |           | Q =                 | 759.6     |          |           | Q =                 |           |          |           |

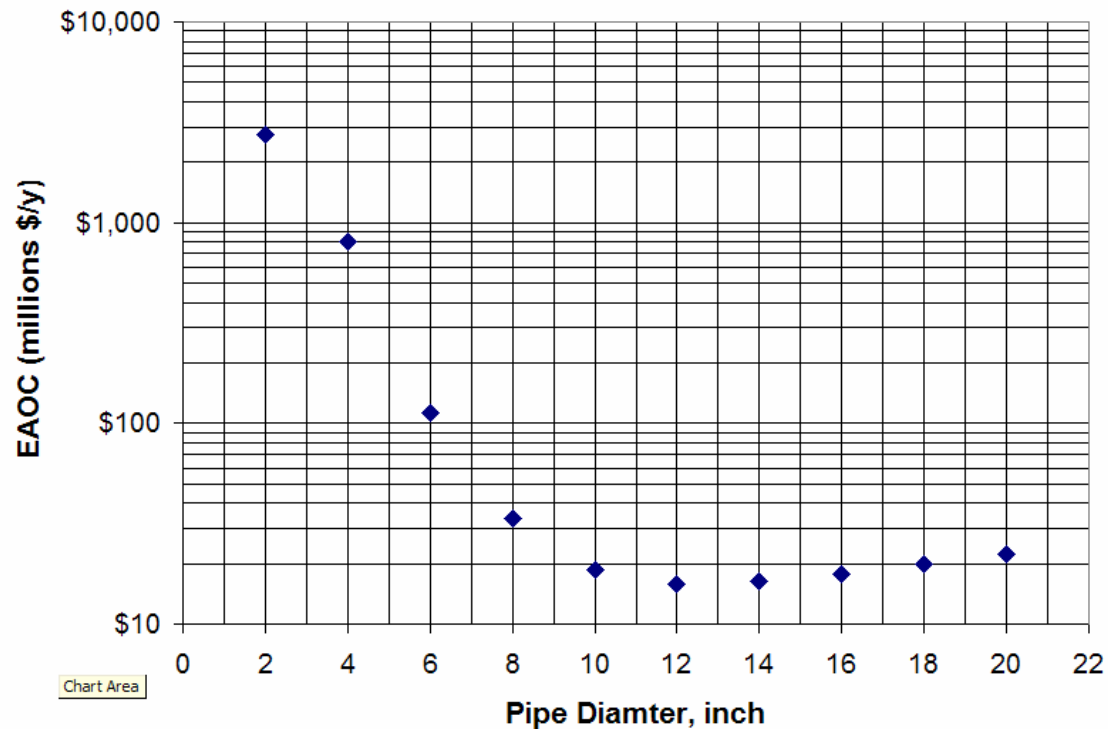


Problem 6

[illegible]

## 14.7 – Optimum Pipe Diameter

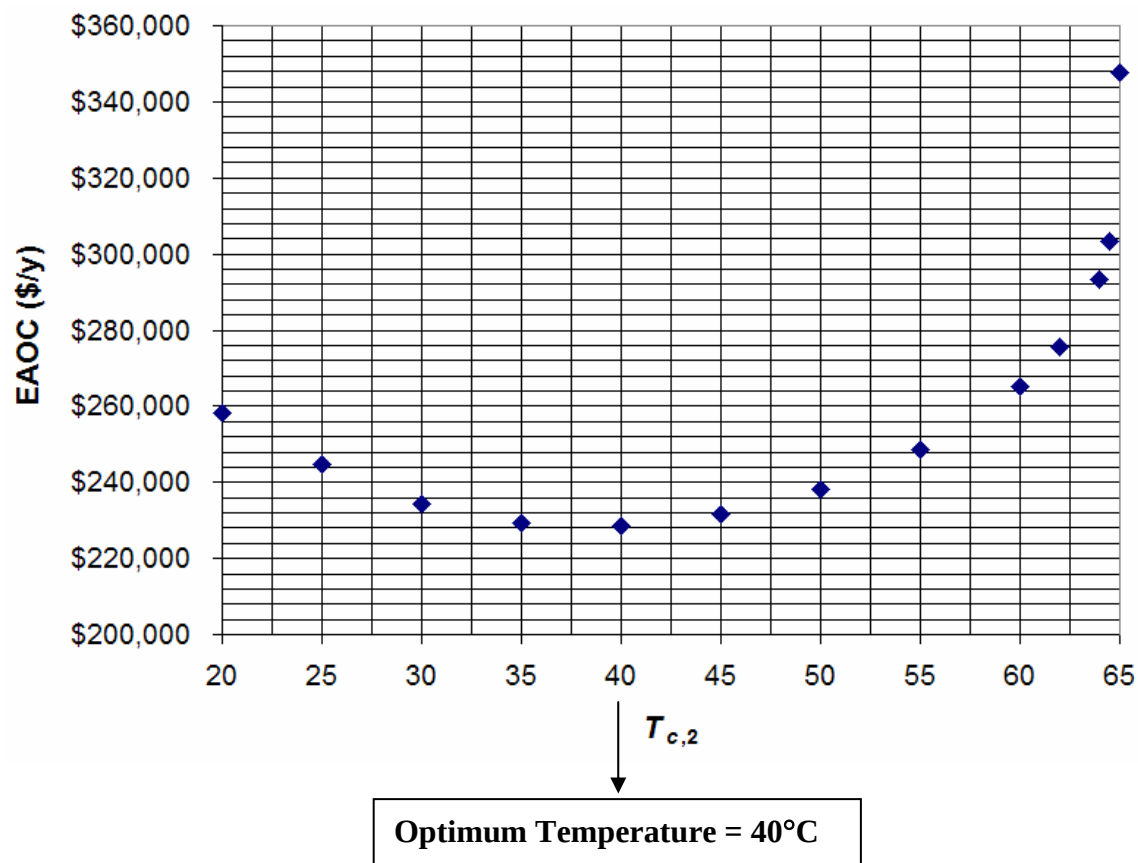
| Problem 7 - Optimum Pipe Diameter |                         |                   |       |                      |        |        |            |                   |                    |                |                      |
|-----------------------------------|-------------------------|-------------------|-------|----------------------|--------|--------|------------|-------------------|--------------------|----------------|----------------------|
| n = 15 y                          |                         | I = 0.09          |       | (A/P, I, n) = 0.1241 |        |        |            |                   |                    |                |                      |
| d <sub>pipe</sub>                 | C <sub>pipe</sub> \$/ft | C <sub>pipe</sub> | v     | Re                   | f      | Q      | W          | C <sub>pump</sub> | C <sub>power</sub> | EAOC           | EAOC                 |
| "                                 | \$/ft                   |                   |       |                      |        |        | kW         |                   | \$/y               | \$/y           | 10 <sup>6</sup> \$/y |
| 2                                 | 25.28                   | \$ 13,346,801     | 90.85 | 429,201              | 0.0050 | 0.1841 | 52494464.8 | \$ 342,892,585    | \$ 27,591,090,674  | 27,635,285,334 | \$ 2,764             |
| 4                                 | 53.93                   | \$ 28,474,411     | 22.71 | 214,601              | 0.0046 | 0.1841 | 1519718.8  | \$ 40,940,008     | \$ 798,764,198     | 807,375,673    | \$ 807               |
| 6                                 | 84.57                   | \$ 44,654,053     | 10.09 | 143,067              | 0.0046 | 0.1841 | 200008.2   | \$ 12,126,029     | \$ 105,124,287     | 112,168,360    | \$ 112               |
| 8                                 | 116.76                  | \$ 61,648,407     | 5.68  | 107,300              | 0.0047 | 0.1841 | 48409.1    | \$ 5,176,622      | \$ 25,443,846      | 33,734,084     | \$ 34                |
| 10                                | 150.24                  | \$ 79,325,521     | 3.63  | 85,840               | 0.0048 | 0.1841 | 16267.8    | \$ 2,690,841      | \$ 8,550,382       | 18,725,240     | \$ 19                |
| 12                                | 184.85                  | \$ 97,598,731     | 2.52  | 71,534               | 0.0050 | 0.1841 | 6709.5     | \$ 1,581,632      | \$ 3,526,530       | 15,830,735     | \$ 16                |
| 14                                | 220.47                  | \$ 116,405,738    | 1.85  | 61,314               | 0.0051 | 0.1841 | 3183.1     | \$ 1,011,111      | \$ 1,673,028       | 16,239,631     | \$ 16                |
| 16                                | 257.01                  | \$ 135,699,094    | 1.42  | 53,650               | 0.0052 | 0.1841 | 1671.8     | \$ 687,067        | \$ 878,693         | 17,798,607     | \$ 18                |
| 18                                | 294.40                  | \$ 155,441,180    | 1.12  | 47,689               | 0.0053 | 0.1841 | 948.6      | \$ 489,030        | \$ 498,576         | 19,843,104     | \$ 20                |
| 20                                | 332.58                  | \$ 175,601,269    | 0.91  | 42,920               | 0.0054 | 0.1841 | 571.9      | \$ 360,986        | \$ 300,601         | 22,130,282     | \$ 22                |



**Optimum pipe diameter = 12"**

## 14.8 – Biological Reactor

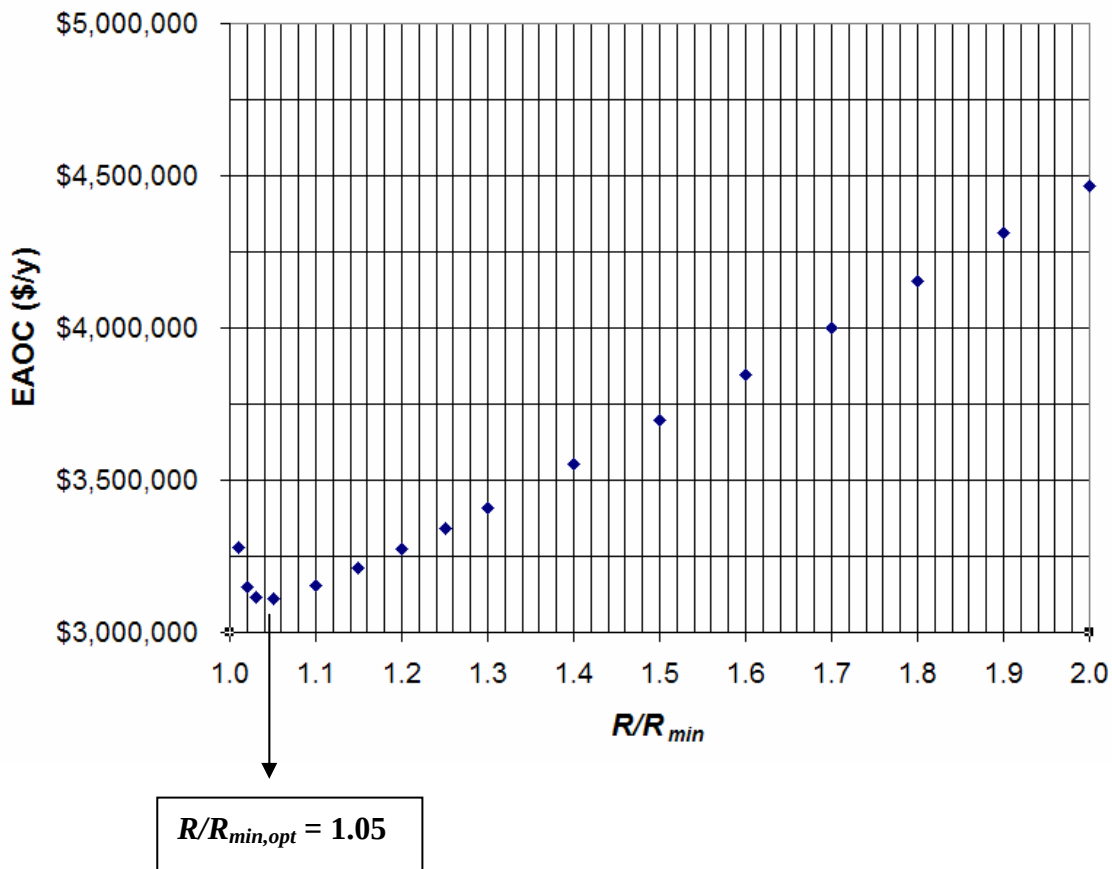
|                  |                           |                            |                |                 |                |                    |                    |            |            |  |
|------------------|---------------------------|----------------------------|----------------|-----------------|----------------|--------------------|--------------------|------------|------------|--|
| ko               | 7.00E+00                  |                            |                |                 |                |                    |                    |            |            |  |
| E/R              | 3300                      |                            |                |                 |                |                    |                    |            |            |  |
| Q                | 5000 gal/h                |                            |                |                 |                |                    |                    |            |            |  |
|                  | 18.9272 m <sup>3</sup> /h | 0.005258 m <sup>3</sup> /s |                |                 |                |                    |                    |            |            |  |
| m                | 18927.20597 kg/h          |                            |                |                 |                |                    |                    |            |            |  |
| Cp               | 4180 J/kg°C               |                            |                |                 |                |                    |                    |            |            |  |
| U                | 400 W/m <sup>2</sup> K    |                            |                |                 |                |                    | A/P,I,n            | 0.1259     |            |  |
| F                | 0.8                       |                            |                |                 |                |                    | h/y                | 8760       |            |  |
| T <sub>c,2</sub> | ΔT <sub>lm</sub>          | Q                          | A              | k               | V              | PC <sub>reac</sub> | PC <sub>exch</sub> | UC         | EAOC       |  |
|                  |                           | kJ/h                       | m <sup>2</sup> | s <sup>-1</sup> | m <sup>3</sup> | \$                 | \$                 | \$/y       | \$/y       |  |
| 20               | 23.3                      | 0                          | 0              | 9.06E-05        | 232.156        | \$ 2,050,896       | \$ -               | \$ -       | 258,211.86 |  |
| 25               | 21.6                      | 395579                     | 15.86772       | 1.09E-04        | 192.230        | \$ 1,746,942       | \$ 58,006          | \$ 17,326  | 244,572.90 |  |
| 30               | 20.0                      | 791157                     | 34.41432       | 1.31E-04        | 160.164        | \$ 1,495,929       | \$ 90,183          | \$ 34,653  | 234,347.28 |  |
| 35               | 18.2                      | 1186736                    | 56.58692       | 1.57E-04        | 134.240        | \$ 1,287,447       | \$ 119,737         | \$ 51,979  | 229,146.31 |  |
| 40               | 16.4                      | 1582314                    | 83.90394       | 1.86E-04        | 113.148        | \$ 1,113,343       | \$ 149,878         | \$ 69,305  | 228,347.32 |  |
| 45               | 14.4                      | 1977893                    | 119.0079       | 2.19E-04        | 95.883         | \$ 967,189         | \$ 182,919         | \$ 86,632  | 231,432.64 |  |
| 50               | 12.3                      | 2373472                    | 167.0764       | 2.58E-04        | 81.670         | \$ 843,888         | \$ 221,944         | \$ 103,958 | 238,148.41 |  |
| 55               | 10.0                      | 2769050                    | 240.3689       | 3.01E-04        | 69.905         | \$ 739,372         | \$ 273,075         | \$ 121,284 | 248,753.50 |  |
| 60               | 7.2                       | 3164629                    | 380.8253       | 3.50E-04        | 60.115         | \$ 650,376         | \$ 354,973         | \$ 138,611 | 265,186.15 |  |
| 62               | 5.8                       | 3322860                    | 496.1103       | 3.71E-04        | 56.665         | \$ 618,515         | \$ 412,725         | \$ 145,541 | 275,376.48 |  |
| 64               | 3.9                       | 3481092                    | 773.1009       | 3.93E-04        | 53.451         | \$ 588,566         | \$ 531,466         | \$ 152,472 | 293,486.11 |  |
| 64.5             | 3.2                       | 3520650                    | 963.7174       | 3.99E-04        | 52.682         | \$ 581,361         | \$ 602,605         | \$ 154,204 | 303,268.13 |  |
| 64.99            | 1.4                       | 3559416                    | 2136.475       | 4.05E-04        | 51.942         | \$ 574,407         | \$ 948,656         | \$ 155,902 | 347,659.11 |  |





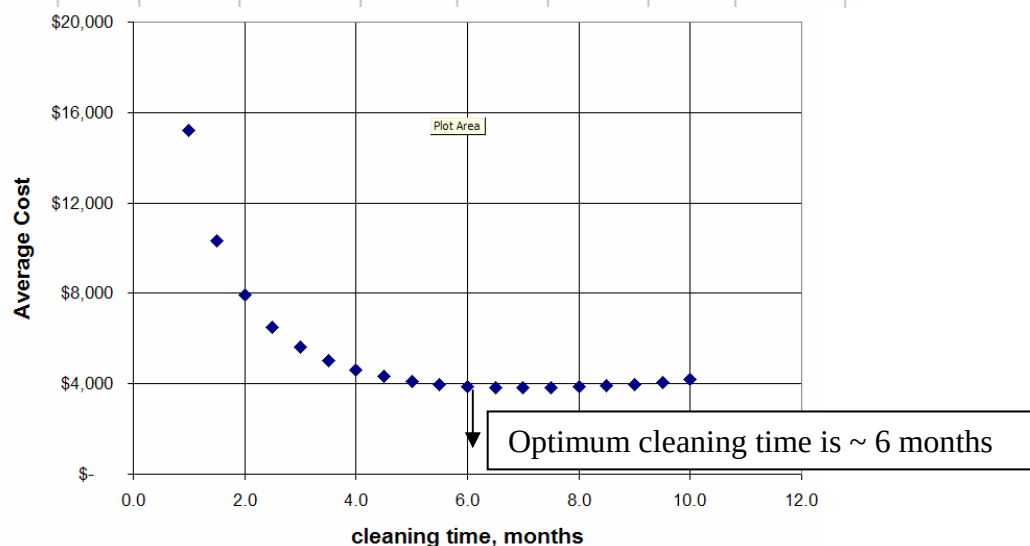
## 14.9 – Distillation Column

| material balance   |           |                        |          |          |       |                         |                            |          |          |                  |                        |      |              |
|--------------------|-----------|------------------------|----------|----------|-------|-------------------------|----------------------------|----------|----------|------------------|------------------------|------|--------------|
|                    | Feed      | top                    | bot      |          |       |                         |                            |          |          |                  |                        |      |              |
| tol in             | 500.0000  | 2.4623                 | 497.5377 |          |       |                         |                            |          |          |                  |                        |      |              |
| benz in            | 500.0000  | 490.0000               | 10.0000  |          |       |                         |                            |          |          |                  |                        |      |              |
| total              | 1000.0000 | 492.4623               | 507.5377 |          |       |                         |                            |          |          |                  |                        |      |              |
| x                  | 0.5000    | 0.9950                 | 0.0197   |          |       |                         |                            |          |          |                  |                        |      |              |
|                    |           |                        |          |          |       |                         |                            |          |          |                  |                        |      |              |
| rel vol, $\alpha$  | 2.3       | (A/P,I,n) = 0.1490     |          |          |       |                         |                            |          |          |                  |                        |      |              |
| R <sub>min</sub>   | 1.5384615 |                        |          |          |       |                         |                            |          |          |                  |                        |      |              |
| N <sub>min</sub>   | 11.046092 | 30.765 kJ/mol 8760 h/y |          |          |       |                         |                            |          |          |                  |                        |      |              |
|                    |           |                        |          |          |       |                         |                            |          |          |                  |                        |      |              |
| R/R <sub>min</sub> | X         | Y                      | N        | V        | lv    | C <sub>ow</sub><br>\$/y | C <sub>steam</sub><br>\$/y | L (m)    | D(m)     | V <sub>col</sub> | C <sub>col</sub><br>\$ | EAOC |              |
| 1.01               | 0.0060241 | 0.766022               | 50.48377 | 1257.673 | 30765 | \$ 119,986.40           | \$ 2,399,727.97            | 50.48377 | 5.319553 | 1121.997         | \$ 3,912,849.14        | \$   | 3,277,783.25 |
| 1.02               | 0.011976  | 0.699438               | 39.07856 | 1265.249 | 30765 | \$ 120,709.21           | \$ 2,414,184.17            | 39.07856 | 5.335551 | 873.7497         | \$ 3,163,581.08        | \$   | 3,147,800.31 |
| 1.03               | 0.017871  | 0.669339               | 35.43038 | 1272.826 | 30765 | \$ 121,432.02           | \$ 2,428,640.36            | 35.43038 | 5.351502 | 796.9244         | \$ 2,925,530.12        | \$   | 3,116,859.71 |
| 1.05               | 0.0294118 | 0.63887                | 32.35666 | 1287.978 | 30765 | \$ 122,877.64           | \$ 2,457,552.74            | 32.35666 | 5.383262 | 736.4524         | \$ 2,735,728.50        | \$   | 3,110,445.87 |
| 1.05               | 0.0294118 | 0.63887                | 32.35666 | 1287.978 | 30765 | \$ 122,877.64           | \$ 2,457,552.74            | 32.35666 | 5.383262 | 736.4524         | \$ 2,735,728.50        | \$   | 3,110,445.87 |
| 1.10               | 0.0571429 | 0.599763               | 29.09741 | 1325.86  | 30765 | \$ 126,491.69           | \$ 2,529,833.71            | 29.09741 | 5.461854 | 681.7488         | \$ 2,562,009.44        | \$   | 3,152,684.84 |
| 1.15               | 0.0833333 | 0.570915               | 27.07392 | 1363.742 | 30765 | \$ 130,105.73           | \$ 2,602,114.67            | 27.07392 | 5.539331 | 652.4628         | \$ 2,468,154.64        | \$   | 3,210,396.57 |
| 1.20               | 0.1081081 | 0.545503               | 25.50421 | 1401.624 | 30765 | \$ 133,719.78           | \$ 2,674,395.63            | 25.50421 | 5.615739 | 631.7069         | \$ 2,401,255.08        | \$   | 3,273,330.57 |
| 1.25               | 0.1315789 | 0.522507               | 24.22778 | 1439.505 | 30765 | \$ 137,333.83           | \$ 2,746,676.60            | 24.22778 | 5.691122 | 616.31           | \$ 2,351,415.34        | \$   | 3,339,569.72 |
| 1.30               | 0.1538462 | 0.501538               | 23.16652 | 1477.387 | 30765 | \$ 140,947.88           | \$ 2,818,957.56            | 23.16652 | 5.765519 | 604.8218         | \$ 2,314,106.43        | \$   | 3,408,236.56 |
| 1.40               | 0.195122  | 0.464669               | 21.50212 | 1553.15  | 30765 | \$ 148,175.97           | \$ 2,963,519.48            | 21.50212 | 5.911504 | 590.1566         | \$ 2,266,324.92        | \$   | 3,550,769.48 |
| 1.50               | 0.2325581 | 0.433271               | 20.25547 | 1628.914 | 30765 | \$ 155,404.07           | \$ 3,108,081.41            | 20.25547 | 6.053971 | 583.0594         | \$ 2,243,137.43        | \$   | 3,698,067.19 |
| 1.60               | 0.2666667 | 0.406171               | 19.28546 | 1704.677 | 30765 | \$ 162,632.17           | \$ 3,252,643.34            | 19.28546 | 6.193161 | 580.9578         | \$ 2,236,263.19        | \$   | 3,848,525.41 |
| 1.70               | 0.2978723 | 0.382505               | 18.50801 | 1780.441 | 30765 | \$ 169,860.26           | \$ 3,397,205.26            | 18.50801 | 6.32929  | 582.3173         | \$ 2,240,710.33        | \$   | 4,001,177.02 |
| 1.80               | 0.3265306 | 0.361629               | 17.87005 | 1856.204 | 30765 | \$ 177,088.36           | \$ 3,541,767.19            | 17.87005 | 6.462553 | 586.1703         | \$ 2,253,306.24        | \$   | 4,155,407.35 |
| 1.90               | 0.3529412 | 0.343052               | 17.33646 | 1931.968 | 30765 | \$ 184,316.46           | \$ 3,686,329.11            | 17.33646 | 6.593123 | 591.8785         | \$ 2,271,944.23        | \$   | 4,310,808.26 |
| 2.00               | 0.3773585 | 0.326398               | 16.88309 | 2007.731 | 30765 | \$ 191,544.55           | \$ 3,830,891.04            | 16.88309 | 6.721157 | 599.0043         | \$ 2,295,172.97        | \$   | 4,467,098.58 |
| 2.50               | 0.4761905 | 0.263381               | 15.35322 | 2386.548 | 30765 | \$ 227,685.03           | \$ 4,553,700.67            | 15.35322 | 7.327846 | 647.5036         | \$ 2,452,199.88        | \$   | 5,256,470.83 |
| 3.00               | 0.5479452 | 0.221298               | 14.46945 | 2765.365 | 30765 | \$ 263,825.52           | \$ 5,276,510.30            | 14.46945 | 7.888011 | 707.0938         | \$ 2,642,746.22        | \$   | 6,052,337.07 |



## 14.10 – Brine Fouling

| Base Case Conditions                                                                                                                                                         |           |           |              |        |          |           |           |              |
|------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|-----------|-----------|--------------|--------|----------|-----------|-----------|--------------|
| T <sub>1</sub>                                                                                                                                                               | 110       | °C        |              |        |          |           |           |              |
| T <sub>2</sub>                                                                                                                                                               | 75        | °C        |              |        |          |           |           |              |
| t <sub>1</sub>                                                                                                                                                               | 40        | °C        |              |        |          |           |           |              |
| t <sub>2</sub>                                                                                                                                                               | 60        | °C        |              |        |          |           |           |              |
| LMTD                                                                                                                                                                         | 26.80     | °C        |              |        |          |           |           |              |
| U                                                                                                                                                                            | 75        | W/m²°C    |              |        |          |           |           |              |
| A                                                                                                                                                                            | 0.00      | m²        |              |        |          |           |           |              |
| C <sub>p,brine</sub>                                                                                                                                                         | \$ 15,000 |           |              |        |          |           |           |              |
| m <sub>brine</sub>                                                                                                                                                           | 14,000    | kg/h      |              |        |          |           |           |              |
| C <sub>p,brine</sub>                                                                                                                                                         | 4,000     | J/kg°C    |              |        |          |           |           |              |
| m <sub>gas</sub>                                                                                                                                                             | 40,000    | kg/h      |              |        |          |           |           |              |
| C <sub>p,gas</sub>                                                                                                                                                           | 800       | J/kg°C    |              |        |          |           |           |              |
|                                                                                                                                                                              |           |           |              |        |          |           |           |              |
| $\frac{Q_2}{Q_1} = \frac{m_{b,2} c_{p,b,2} (T_{b,e,2} - T_{b,i,2})}{m_{b,1} c_{p,b,1} (T_{b,e,1} - T_{b,i,1})} = \frac{(T_b - 40)}{(20)}$                                    |           |           |              |        |          |           |           |              |
| $\frac{Q_2}{Q_1} = \frac{m_{c,k,2} c_{p,c,k,2} (T_{c,k,e,2} - T_{c,k,i,2})}{m_{c,k,1} c_{p,c,k,1} (T_{c,k,e,1} - T_{c,k,i,1})} = \frac{(110 - T_c)}{35}$                     |           |           |              |        |          |           |           |              |
| $\frac{Q_2}{Q_1} = \frac{U_2 A_2 F_2 \Delta T_{lm,2}}{U_1 A_1 F_1 \Delta T_{lm,1}} = (1 - 0.1r) \frac{(110 - T_b) - (T_c - 40)}{42.0551 \ln \frac{(110 - T_b)}{(T_c - 40)}}$ |           |           |              |        |          |           |           |              |
| t                                                                                                                                                                            | Tb        | Tc        | f(Tb)        | Q      | Q(\$)    | Qclean    | ΔQ(\$)    |              |
| 0.0                                                                                                                                                                          | 60        | 75        | -2.34885E-06 |        |          |           |           | \$ 1,000,000 |
| 0.5                                                                                                                                                                          | 59.393    | 76.062329 | 0            | 6.12   | \$ 49    | \$ 15,000 | \$ 49     | \$ 30,098    |
| 1.0                                                                                                                                                                          | 58.7596   | 77.170653 | 0            | 18.62  | \$ 149   | \$ 15,000 | \$ 198    | \$ 15,198    |
| 1.5                                                                                                                                                                          | 58.0984   | 78.327782 | 0            | 31.67  | \$ 253   | \$ 15,000 | \$ 451    | \$ 10,301    |
| 2.0                                                                                                                                                                          | 57.4075   | 79.536814 | 0            | 45.30  | \$ 362   | \$ 15,000 | \$ 814    | \$ 7,907     |
| 2.5                                                                                                                                                                          | 56.6851   | 80.801117 | 0            | 59.55  | \$ 476   | \$ 15,000 | \$ 1,290  | \$ 6,516     |
| 3.0                                                                                                                                                                          | 55.9289   | 82.124354 | 0            | 74.45  | \$ 596   | \$ 15,000 | \$ 1,886  | \$ 5,629     |
| 3.5                                                                                                                                                                          | 55.1368   | 83.510518 | 0            | 90.06  | \$ 720   | \$ 15,000 | \$ 2,606  | \$ 5,030     |
| 4.0                                                                                                                                                                          | 54.3063   | 84.963971 | 0            | 106.41 | \$ 851   | \$ 15,000 | \$ 3,457  | \$ 4,614     |
| 4.5                                                                                                                                                                          | 53.4346   | 86.489486 | 0            | 123.57 | \$ 989   | \$ 15,000 | \$ 4,446  | \$ 4,321     |
| 5.0                                                                                                                                                                          | 52.5187   | 88.092296 | 0            | 141.59 | \$ 1,133 | \$ 15,000 | \$ 5,579  | \$ 4,116     |
| 5.5                                                                                                                                                                          | 51.5553   | 89.778154 | 0            | 160.53 | \$ 1,284 | \$ 15,000 | \$ 6,863  | \$ 3,975     |
| 6.0                                                                                                                                                                          | 50.5409   | 91.553396 | 0            | 180.47 | \$ 1,444 | \$ 15,000 | \$ 8,307  | \$ 3,884     |
| 6.5                                                                                                                                                                          | 49.4714   | 93.42502  | 0            | 201.48 | \$ 1,612 | \$ 15,000 | \$ 9,919  | \$ 3,834     |
| 7.0                                                                                                                                                                          | 48.3424   | 95.400769 | 6.10623E-16  | 223.64 | \$ 1,789 | \$ 15,000 | \$ 11,708 | \$ 3,815     |
| 7.5                                                                                                                                                                          | 47.149    | 97.489241 | 0            | 247.05 | \$ 1,976 | \$ 15,000 | \$ 13,684 | \$ 3,825     |
| 8.0                                                                                                                                                                          | 45.8857   | 99.700004 | 0            | 271.81 | \$ 2,174 | \$ 15,000 | \$ 15,859 | \$ 3,857     |
| 8.5                                                                                                                                                                          | 44.5464   | 102.04374 | 0            | 298.04 | \$ 2,384 | \$ 15,000 | \$ 18,243 | \$ 3,911     |
| 9.0                                                                                                                                                                          | 43.1243   | 104.5324  | 0            | 325.88 | \$ 2,607 | \$ 15,000 | \$ 20,850 | \$ 3,983     |
| 9.5                                                                                                                                                                          | 41.6117   | 107.17944 | -9.74459E-11 | 355.46 | \$ 2,844 | \$ 15,000 | \$ 23,694 | \$ 4,073     |
| 10.0                                                                                                                                                                         | 40        | 110.00004 | -1E-06       | 386.95 | \$ 3,096 | \$ 15,000 | \$ 26,789 | \$ 4,179     |



#### 14.11 – Optimal Cycle Time

$$x = \$350$$

$$y = \$200$$

$$V = 7\text{m}^3$$

$$C_{A0} = 3\text{kmol/m}^3$$

$$C_{\text{clean}} = \$800$$

$$t_{\text{clean}} = 1.0 \text{ h}$$

$$k = 0.153 \text{ h}^{-1}$$

$$A_1 = \$ 6,550$$

$$B_1 = \$11,550$$

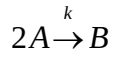
$$C_1 = 1.5 \text{ h}$$

Substituting values into Equation (14.18), we get:

$$(B_1 k C_1 + B_1 k t_{\text{opt}} + B_1) e^{-k t_{\text{opt}}} - A_1 = 0 \quad (14.18)$$

**Solving we get,  $t_{\text{opt}}=10.53 \text{ h}$**

#### 14.12 – Second order reactor – optimum cycle time



A material balance at time,  $t$  gives:  $N_B = \frac{V(C_{A0} - C_A)}{2}$  and for a 2<sup>nd</sup> order reaction we have

$$C_A = \frac{C_{A0}}{1 + C_{A0}kt}$$

$$\therefore N_B = \frac{V}{2} \left[ C_{A0} - \frac{C_{A0}}{1 + C_{A0}kt} \right] = \frac{V}{2} \left[ \frac{C_{A0}^2 kt}{1 + C_{A0}kt} \right]$$

The amount of remaining A is  $N_A = VC_A = \frac{VC_{A0}}{1 + C_{A0}kt}$

The objective function (OF) can now be written as:

$$OF = \frac{xN_B - yN_A - C_{clean}}{\theta} = \frac{\frac{xV}{2} \left[ \frac{C_{A0}^2 kt}{1 + C_{A0}kt} \right] - y \left[ \frac{VC_{A0}}{1 + C_{A0}kt} \right] - C_{clean}}{t + t_{clean}}$$

$$OF = \frac{\frac{xVC_{A0}^2 kt}{2} - yVC_{A0} - C_{clean}(1 + C_{A0}kt)}{(1 + C_{A0}kt)(t + t_{clean})}$$

let

$$A_1 = \frac{xVC_{A0}^2 k}{2}, B_1 = yVC_{A0}, C_1 = C_{clean}, D_1 = C_{clean}C_{A0}k, \text{ and } E_1 = C_{A0}k$$

$$OF = \frac{A_1 t - B_1 - C_1 - D_1 t}{(t + t_{clean})(1 + E_1 t)} = \frac{(A_1 - D_1)t - (B_1 + C_1)}{(t + t_{clean})(1 + E_1 t)} = \frac{(A_1 - D_1)t - (B_1 + C_1)}{E_1 t^2 + (E_1 t_{clean} + 1)t + t_{clean}}$$

Differentiating the OF with respect to  $t$  and setting  $=0$  gives

$$(A_1 - D_1)(E_1 t^2 + (E_1 t_{clean} + 1)t + t_{clean}) - (2E_1 t + E_1 t_{clean} + 1)[(A_1 - D_1)t - (B_1 + C_1)] = 0$$

collecting terms we get

$$t^2 [(A_1 - D_1)E_1 - 2E_1(A_1 - D_1)] + t [(A_1 - D_1)(E_1 t_{clean} + 1) - (A_1 - D_1)(E_1 t_{clean} + 1) + 2E_1(B_1 + C_1)] + (A_1 - D_1)t_{clean} + (E_1 t_{clean} + 1)(B_1 + C_1) = 0$$

$$\therefore \underbrace{E_1(A_1 - D_1)t^2}_a - \underbrace{2E_1(B_1 + C_1)t}_b - \underbrace{[(A_1 - D_1)t_{clean} + (E_1 t_{clean} + 1)(B_1 + C_1)]}_c = 0$$

$$t_{opt} = \frac{-b \pm \sqrt{b^2 - 4ac}}{2a} \quad \text{and} \quad \theta = t_{opt} + t_{clean}$$

# 14.13 – Two product batch sequencing

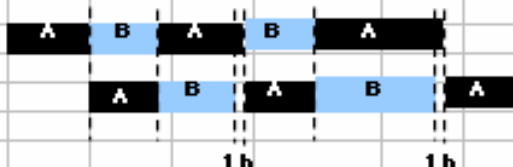
## Chapter 14 - Problem 13

| (a) |                          | <b>Optimum Values</b> |
|-----|--------------------------|-----------------------|
|     | $x_1$                    | 1,037                 |
|     | $x_2$                    | 250                   |
|     | OF = $1.95x_1 + 2.75x_2$ | 2,709.93              |

Subject to

|                      |       |
|----------------------|-------|
| $6x_1 + 8x_2 < 8760$ | 8,223 |
| $7x_1 + 6x_2 < 8760$ | 8,760 |
| $x_1 > 0$            | 1,037 |
| $x_2 > 0$            | 250   |
| $x_2 < 250$          | 250   |

(b) no - this case cannot be implemented without intermediate storage.  
The best we can do is



Use this sequence for 250 cycles to produce the maximum amount of product B this takes  $250(14) = 3500$  h. The remaining time can be used to make product A with a cycle time of 8 h. Thus the additional amount of A that can be produced is  $(8760 - 3500)/8 = 658$

Produce 250 units of B to give a profit of  $(250)(2.75) = 687.5$   
Produce  $(250 + 658)$  units of A to give a profit of  $(908)(1.95) = 1771$   
Profit is **2458**

(c) relax the restriction for the amount of B

|  |                          | <b>Optimum Values</b> |
|--|--------------------------|-----------------------|
|  | $x_1$                    | -                     |
|  | $x_2$                    | 1,095                 |
|  | OF = $1.95x_1 + 2.75x_2$ | 3,011.25              |

Subject to

|                      |       |
|----------------------|-------|
| $6x_1 + 8x_2 < 8760$ | 8,760 |
| $7x_1 + 6x_2 < 8760$ | 6,570 |
| $x_1 > 0$            | -     |
| $x_2 > 0$            | 1,095 |

Optimum Solution is to just make Product B

So the best we can do is to make all B cycle time for B is 8 h (based on reactor/mixer time)  
 $\therefore$  maximum amount of B =  $8760/8 = 1095$  same as above result  
This does not require any intermediate storage - Max profit =  $(1095)(2.75) = 3011.25$

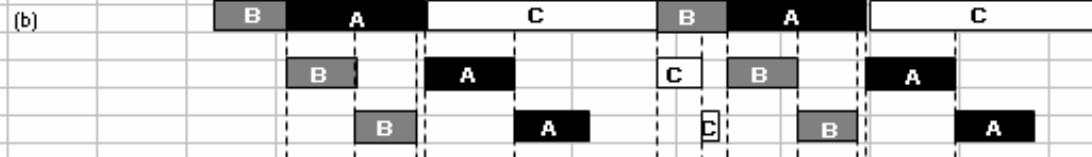
# 14.14 –Three product batch sequencing

| Product | Time in Reactor (hr) | Time in Separator 1 (hr) | Time in Separator 2 (hr) | Value of Product (\$/kg product) | Cost of Feed (\$/kg product) |
|---------|----------------------|--------------------------|--------------------------|----------------------------------|------------------------------|
| A       | 15                   | 10                       | 8                        | 1.35                             | 0.25                         |
| B       | 7                    | 7                        | 6                        | 0.97                             | 0.27                         |
| C       | 25                   | 5                        | 2                        | 1.41                             | 0.23                         |

(a) Let the amount of each product (A, B, C) be equal to  $x_1$ ,  $x_2$ , and  $x_3$ , respectively

| Optimal Values        |                               |       |
|-----------------------|-------------------------------|-------|
| $x_1$                 |                               | 50    |
| $x_2$                 |                               | 857   |
| $x_3$                 |                               | 50    |
| OF                    | $1.1x_1 + 0.7x_2 + 1.18x_3$   | 714   |
| <b>subject to</b>     |                               |       |
| market constraints    | $x_1 > 50$                    | 50    |
|                       | $x_2 > 50$                    | 857   |
|                       | $x_3 > 50$                    | 50    |
| Equipment constraints | $15x_1 + 7x_2 + 25x_3 < 8000$ | 8,000 |
|                       | $10x_1 + 7x_2 + 5x_3 < 8000$  | 6,750 |
|                       | $8x_1 + 6x_2 + 2x_3 < 8000$   | 5,643 |

Optimal mix is shown above, namely make 50 batches of A and C and make 857 batches of B



Using the sequence of BACBACBAC shown above, we can implement the optimal sequence from A.  
Note that multiple B runs can be made to spread out the production of A and C to meet demand.  
Thus a sequence of say BBBBBBBACBBBBBBACBBBBBBBAC could also be used.

#### 14.15 Optimization of T-202

The approach used here follows that developed in the chapter for T-201. The base case conditions are given below:

Feed Temperature = 139°C

Feed Pressure 7.4 bar

Feed Composition

DME = 1.4 kmol/h

MeOH = 64.3 kmol/h

Water = 132.9 kmol/h

##### Utility Costs Used

Cooling Water = 14.8 \$/1000m<sup>3</sup>

MP steam = 13.71 \$/1000 kg

Electricity = 0.06 \$/kWh

FCI = total module cost from Capcost

$R/R_{min} = 1.3$

Recovery of methanol in top product = 0.989

Recovery of water in top product = 0.01

Tray efficiency

Tray spacing = 18 inches

Top temperature = 116°C (in Figure B.1 top temperature is given as 121°C. The top temperature apparently is quite sensitive to the thermodynamic package used for the  $k$  values. The current results use the Uniquac/Unifac option in CHEMCAD and this gives slightly different results from the previous version – hence the different top temperature)

Bottom temperature = 167°C

Spread sheets on the next page give summaries of the results for different pressures and  $R/R_{min}$  values. The equipment considered are T-202, E-206, E-207, E-208, V-202, and P-203A/B. All equipment specifications are based on the following assumptions:

1. Height of column = (number of actual trays times)(tray spacing – 18"). An additional 9 to 15 feet should be added to take into account of the skirt, vapor space at top of tower, and liquid storage at bottom of the tower. This was not considered here.
2. Overall heat transfer coefficient for condenser = 850 W/m<sup>2</sup>°C
3. Overall heat transfer coefficient for reboiler = 1140 W/m<sup>2</sup>°C
4. Overall heat transfer coefficient for E-208 = 280 W/m<sup>2</sup>°C
5. Efficiency of P-203 = 40% and destination pressure = 15.5 bar (to return to front end)
6. Medium pressure steam and cooling water used as utilities.

A graph showing the results is given on the page after the spreadsheets. The optimum reflux ratio is between 1.01 and 1.12 for the three cases considered and the optimum pressure is approximately 7.0 bar. It should be noted for the 5.0 bar case, that the high temperature utility could be changed to low-pressure steam (a topological change). This has the effect of decreasing the steam cost but increasing the size and cost of the reboiler, E-206. The net result is that the 5.0 bar case becomes cheaper than the 7.0 bar case. The full optimization has not been carried out here but should proceed to investigate lower pressures using lps as the heating medium for the column.

# Bivariate Optimization for Column T-202 in DME Process P = 5.0 bar

| R/Rmin | Rmin  | R     | No. Trays | Qcond | Qreboil | N actual | T top | T bot | P power | cooling w | steam  | elect | total utils | E-206 Area |
|--------|-------|-------|-----------|-------|---------|----------|-------|-------|---------|-----------|--------|-------|-------------|------------|
| 1.01   | 1.600 | 1.616 | 37.8      | 5245  | 5043    | 54.0     | 101   | 153   | 5.8     | 15553     | 289957 | 2913  | 308422      | 40         |
| 1.12   | 1.600 | 1.792 | 20.8      | 5564  | 5363    | 29.7     | 101   | 153   | 6.2     | 16499     | 308356 | 3109  | 327963      | 42         |
| 1.29   | 1.600 | 2.064 | 17.5      | 6044  | 5842    | 25.0     | 101   | 153   | 6.8     | 17922     | 335897 | 3411  | 357231      | 46         |
| 1.51   | 1.600 | 2.416 | 15.2      | 6684  | 6481    | 21.7     | 101   | 153   | 7.6     | 19820     | 372637 | 3803  | 396261      | 51         |

\$0.354/GJ \$6.87/GJ \$0.06/kWh

| E-206 P | E-207 DT | E-207 Area | E-207 P | V-202 D | V-202 H | V-202 P | T-202 D | T-202 H | T-202 P | P-203 | E-208 | P E-208 | FCI TM | NPV    | R/Rmin |
|---------|----------|------------|---------|---------|---------|---------|---------|---------|---------|-------|-------|---------|--------|--------|--------|
| 5.3 bar | 65.9     | 26         | 5.0 bar | 0.83    | 2.5     | 5.0 bar | 0.92    | 30      | 5.3 bar | 5.8   | 21.4  | 5.0 bar | 471000 | -3.064 | 1.01   |
| 5.3 bar | 65.9     | 28         | 5.0 bar | 0.84    | 2.5     | 5.0 bar | 0.95    | 16      | 5.3 bar | 6.2   | 21.4  | 5.0 bar | 381000 | -3.017 | 1.12   |
| 5.3 bar | 65.9     | 30         | 5.0 bar | 0.87    | 2.6     | 5.0 bar | 0.99    | 14      | 5.3 bar | 6.8   | 21.4  | 5.0 bar | 375000 | -3.206 | 1.29   |
| 5.3 bar | 65.9     | 33         | 5.0 bar | 0.90    | 2.7     | 5.0 bar | 1.05    | 12      | 5.3 bar | 7.6   | 21.4  | 5.0 bar | 375000 | -3.475 | 1.51   |

P = 7.3 bar

| R/Rmin | Rmin  | R     | No. Trays | Qcond | Qreboil | N actual | T top | T bot | P power | cooling w | steam  | elect | total utils | E-206 Area |
|--------|-------|-------|-----------|-------|---------|----------|-------|-------|---------|-----------|--------|-------|-------------|------------|
| 1.01   | 1.358 | 1.371 | 38.3      | 5310  | 4843    | 54.7     | 115.5 | 167   | 4.2     | 15746     | 278457 | 2121  | 296324      | 69         |
| 1.12   | 1.358 | 1.521 | 21.4      | 5648  | 5180    | 30.6     | 115.5 | 167   | 4.5     | 16748     | 297834 | 2255  | 316836      | 74         |
| 1.29   | 1.358 | 1.751 | 17.9      | 6155  | 5688    | 25.6     | 115.5 | 167   | 4.9     | 18251     | 327042 | 2461  | 347755      | 82         |
| 1.51   | 1.358 | 2.050 | 15.6      | 6830  | 6363    | 22.3     | 115.5 | 167   | 5.4     | 20253     | 365853 | 2728  | 388834      | 91         |

| E-206 P | E-207 DT | E-207 Area | E-207 P | V-202 D | V-202 H | V-202 P | T-202 D | T-202 H | T-202 P | P-203 | E-208 | P E-208 | FCI TM | NPV    | R/Rmin |
|---------|----------|------------|---------|---------|---------|---------|---------|---------|---------|-------|-------|---------|--------|--------|--------|
| 7.6 bar | 80.4     | 22         | 7.3 bar | 0.81    | 2.4     | 7.3 bar | 0.81    | 30      | 7.6 bar | 4.2   | 22.8  | 7.6 bar | 462000 | -2.962 | 1.01   |
| 7.6 bar | 80.4     | 23         | 7.3 bar | 0.82    | 2.5     | 7.3 bar | 0.83    | 17      | 7.6 bar | 4.5   | 22.8  | 7.6 bar | 399000 | -2.977 | 1.12   |
| 7.6 bar | 80.4     | 25         | 7.3 bar | 0.85    | 2.5     | 7.3 bar | 0.87    | 14      | 7.6 bar | 4.9   | 22.8  | 7.6 bar | 377000 | -3.145 | 1.29   |
| 7.6 bar | 80.4     | 28         | 7.3 bar | 0.88    | 2.6     | 7.3 bar | 0.91    | 12      | 7.6 bar | 5.4   | 22.8  | 7.6 bar | 378000 | -3.430 | 1.51   |

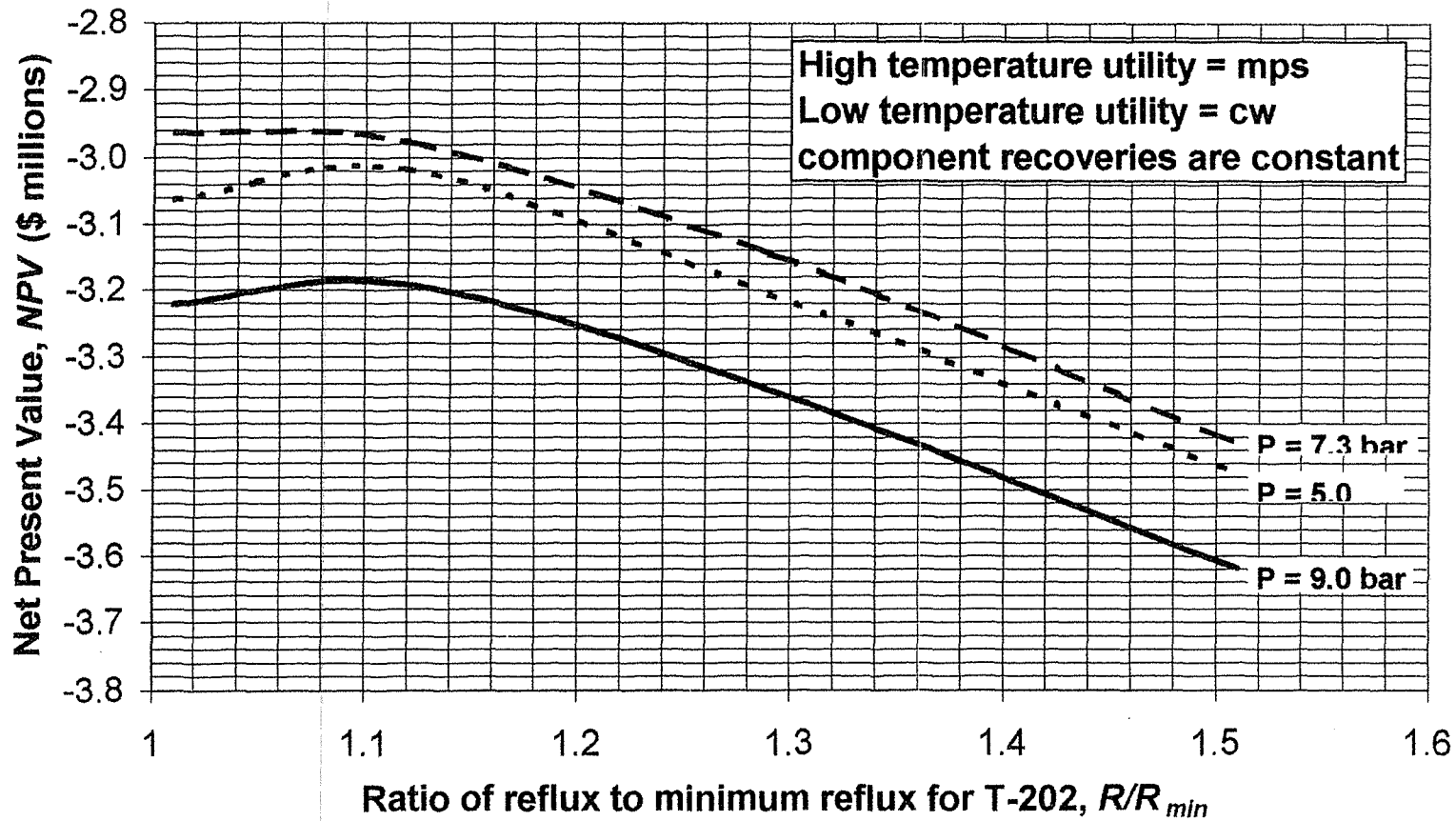
P = 9.0 bar

| R/Rmin | Rmin  | R     | No. Trays | Qcond | Qreboil | N actual | T top | T bot | P power | cooling w | steam  | elect | total utils | E-206 Area |
|--------|-------|-------|-----------|-------|---------|----------|-------|-------|---------|-----------|--------|-------|-------------|------------|
| 1.01   | 1.301 | 1.314 | 39.3      | 5048  | 5281    | 56.1     | 123.4 | 175   | 3.4     | 14969     | 303641 | 1698  | 320308      | 143        |
| 1.12   | 1.301 | 1.457 | 21.8      | 5364  | 5596    | 31.1     | 123.4 | 175   | 3.6     | 15906     | 321752 | 1803  | 339461      | 152        |
| 1.29   | 1.301 | 1.678 | 18.3      | 5837  | 6069    | 26.1     | 123.4 | 175   | 3.9     | 17309     | 348949 | 1965  | 368222      | 164        |
| 1.51   | 1.301 | 1.965 | 15.9      | 6468  | 6700    | 22.7     | 123.4 | 175   | 4.3     | 19180     | 385229 | 2175  | 406584      | 181        |

| E-206 P | E-207 DT | E-207 Area | E-207 P | V-202 D | V-202 H | V-202 P | T-202 D | T-202 H | T-202 P | P-203 | E-208 | P E-208 | FCI TM | NPV      | R/Rmin |
|---------|----------|------------|---------|---------|---------|---------|---------|---------|---------|-------|-------|---------|--------|----------|--------|
| 9.2 bar | 88.3     | 19         | 8.9 bar | 0.76    | 2.3     | 8.9 bar | 0.76    | 31      | 9.2 bar | 3.4   | 23.6  | 9.2 bar | 509000 | -3.22138 | 1.01   |
| 9.2 bar | 88.3     | 20         | 8.9 bar | 0.77    | 2.3     | 8.9 bar | 0.78    | 17      | 9.2 bar | 3.6   | 23.6  | 9.2 bar | 429000 | -3.19258 | 1.12   |
| 9.2 bar | 88.3     | 22         | 8.9 bar | 0.80    | 2.4     | 8.9 bar | 0.81    | 14      | 9.2 bar | 3.9   | 23.6  | 9.2 bar | 409000 | -3.3501  | 1.29   |
| 9.2 bar | 88.3     | 24         | 8.9 bar | 0.82    | 2.5     | 8.9 bar | 0.86    | 12      | 9.2 bar | 4.3   | 23.6  | 9.2 bar | 412000 | -3.6197  | 1.51   |



## Problem 12.3 - Solution

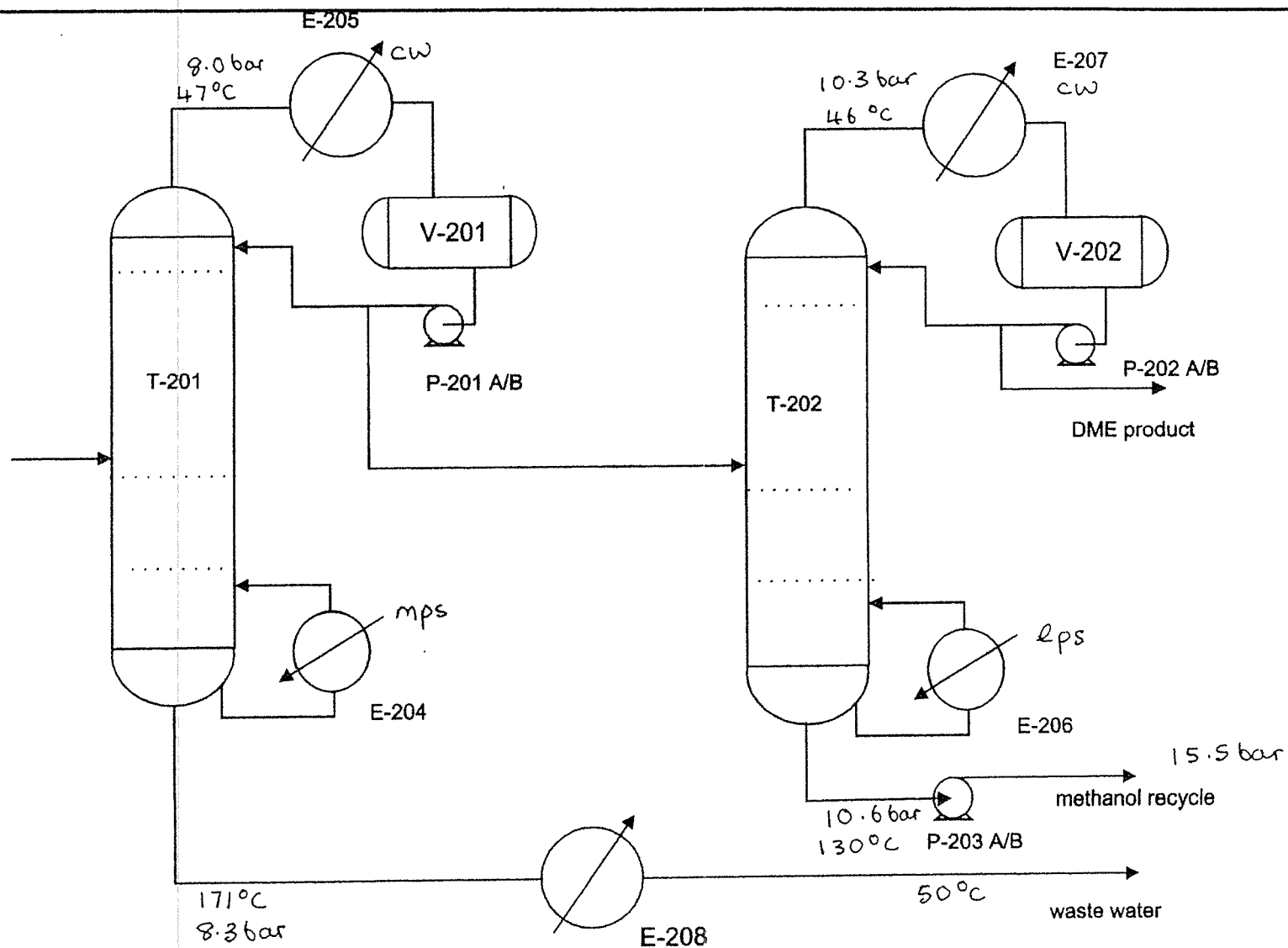


#### 14.16 Optimization of DME separation section with water removal from the first tower.

A PFD for the rearranged separation section of the DME process is shown on the next page. Water is removed as the bottoms product from the first column and DME is removed as the top product from the second column. A base case was taken as a starting point and the operating conditions for the base case are shown in the figure. The costs for the base case for different  $R/R_{min}$  values for both towers are shown in the accompanying spreadsheet.

The results of this preliminary optimization are shown in the figure following the spreadsheet, where NPV is plotted as a function of  $R/R_{min}$ . It should be noted that the reflux ratio of each tower can be varied independently but this is not done for the base case shown here. The lowest NPV for the base case is  $\$-6.411 \times 10^6$  using a reflux ratio of 1.09 times the minimum.

The costs associated with T-202 for Problem 12.3 gave an optimum NPV of  $\$-2.96 \times 10^6$  again for a reflux of 1.1 times the minimum. At this stage, the results given in Table 12.4 should be reworked for current (2002 prices). These can then be used in conjunction with Problem 12.3 to compare the results of Problem 12.4.



Base Case for Problem 12.4 - DME Process with Water removal from first Column, T-201

# Base Case for T-201+ T-202 in DME Process - Problem 12.4

Tower T-201 - water bottom, DME + MeOH top

| R/Rmin | Rmin  | R     | No. Trays | Qcond | Qreboil | N actual | T top | T bot | cooling w | steam  | elect | total utils | E-204<br>Area |
|--------|-------|-------|-----------|-------|---------|----------|-------|-------|-----------|--------|-------|-------------|---------------|
| 1.01   | 0.337 | 0.341 | 60.6      | 7110  | 5446    | 87       | 46.9  | 171   | 21083     | 313128 | 1283  | 335494      | 102           |
| 1.09   | 0.337 | 0.368 | 24        | 7244  | 5580    | 35       | 46.9  | 171   | 21481     | 320833 | 1309  | 343622      | 105           |
| 1.16   | 0.337 | 0.391 | 21.8      | 7378  | 5714    | 32       | 46.9  | 171   | 21878     | 328537 | 1331  | 351746      | 107           |
| 1.24   | 0.337 | 0.418 | 20.5      | 7512  | 5848    | 30       | 46.9  | 171   | 22275     | 336242 | 1357  | 359874      | 110           |
| 1.31   | 0.337 | 0.442 | 19.6      | 7646  | 5982    | 29       | 46.9  | 171   | 22673     | 343946 | 1380  | 367999      | 112           |

E-205

| E-204 P | E-205 DT | Area | E-205 P | V-201 D | V-201 H | V-201 P | T-201 D | T-201 H | T-201 P | P-202 | R/Rmin |
|---------|----------|------|---------|---------|---------|---------|---------|---------|---------|-------|--------|
| 8.3 bar | 11.2     | 208  | 8.0 bar | 1.06    | 3.2     | 8.0 bar | 1.09    | 48      | 8.3 bar | 2.6   | 1.01   |
| 8.3 bar | 11.2     | 212  | 8.0 bar | 1.06    | 3.2     | 8.0 bar | 1.11    | 19      | 8.3 bar | 2.6   | 1.09   |
| 8.3 bar | 11.2     | 216  | 8.0 bar | 1.07    | 3.2     | 8.0 bar | 1.12    | 17      | 8.3 bar | 2.6   | 1.16   |
| 8.3 bar | 11.2     | 220  | 8.0 bar | 1.08    | 3.2     | 8.0 bar | 1.13    | 16      | 8.3 bar | 2.7   | 1.24   |
| 8.3 bar | 11.2     | 224  | 8.0 bar | 1.08    | 3.2     | 8.0 bar | 1.14    | 16      | 8.3 bar | 2.7   | 1.31   |

Tower T-202 - water bottom, DME + MeOH top

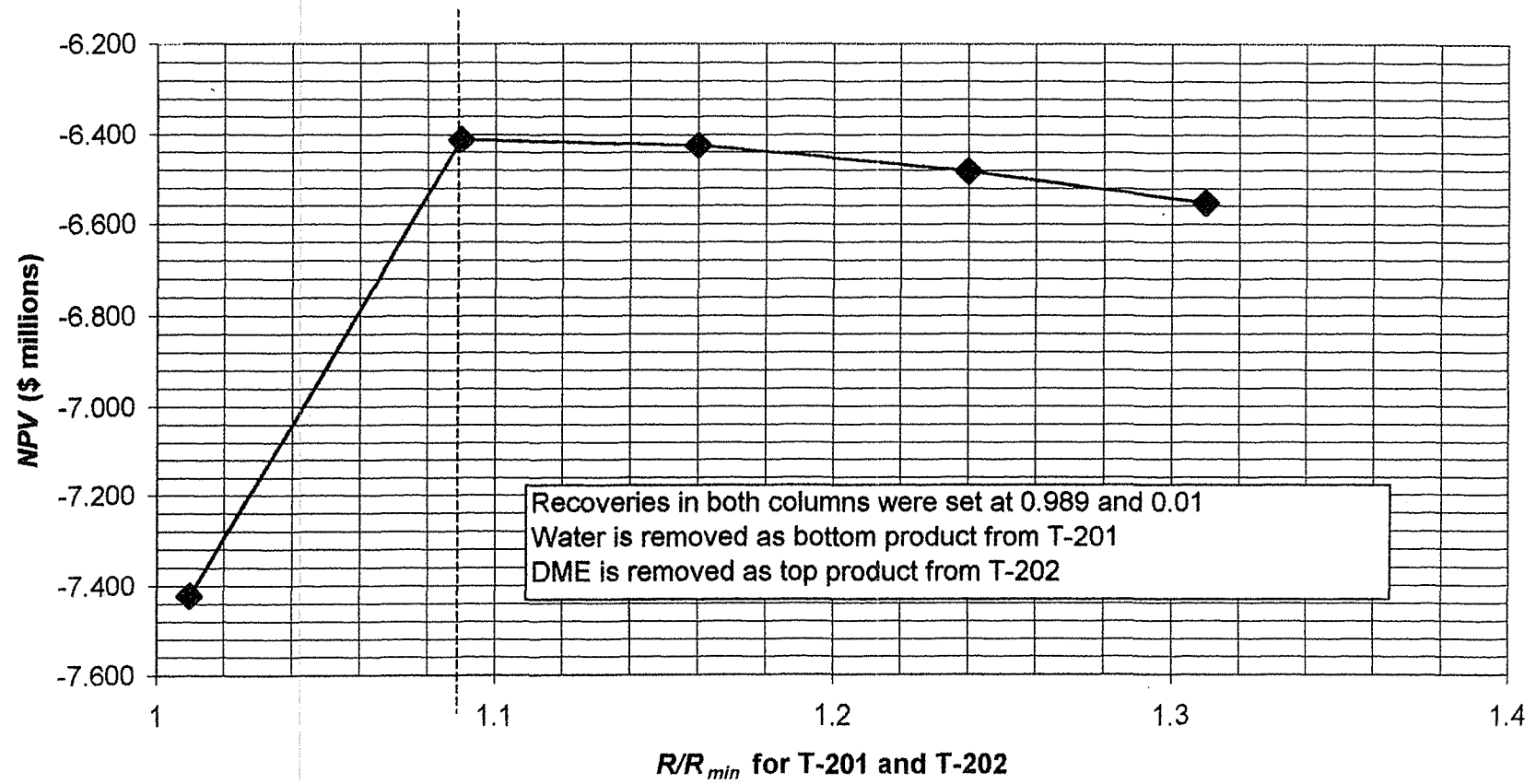
| R/Rmin | Rmin  | R     | No. Trays | Qcond | Qreboil | N actual | T top | T bot | cooling w | steam  | elect | total utils | E-206<br>Area | P-204 |
|--------|-------|-------|-----------|-------|---------|----------|-------|-------|-----------|--------|-------|-------------|---------------|-------|
| 1.01   | 0.122 | 0.123 | 84.6      | 2576  | 3085    | 121      | 45.7  | 130   | 7639      | 327612 | 983   | 336233      | 25            | 1.13  |
| 1.09   | 0.122 | 0.133 | 17.1      | 2598  | 3106    | 25       | 45.7  | 130   | 7704      | 329842 | 986   | 338532      | 25            | 1.13  |
| 1.16   | 0.122 | 0.142 | 14.4      | 2619  | 3127    | 21       | 45.7  | 130   | 7766      | 332072 | 990   | 340828      | 25            | 1.13  |
| 1.24   | 0.122 | 0.151 | 13.4      | 2640  | 3148    | 20       | 45.7  | 130   | 7828      | 334302 | 993   | 343124      | 26            | 1.13  |
| 1.31   | 0.122 | 0.160 | 12.9      | 2661  | 3169    | 19       | 45.7  | 130   | 7891      | 336532 | 996   | 345419      | 26            | 1.13  |

E-207

| E-206 P  | E-207 DT | Area | E-207 P   | V-202 D | V-202 H | V-202 P   | T-202 D | T-202 H | T-202 P  | P-203 | E-208 P  | E-208 A | FCI TM  | NPV    | R/Rmin |
|----------|----------|------|-----------|---------|---------|-----------|---------|---------|----------|-------|----------|---------|---------|--------|--------|
| 10.6 bar | 9.9      | 85   | 10.30 bar | 0.91    | 2.7     | 10.30 bar | 0.60    | 67      | 10.6 bar | 0.8   | 10.6 bar | 20      | 1400000 | -7.422 | 1.01   |
| 10.6 bar | 9.9      | 86   | 10.30 bar | 0.92    | 2.8     | 10.30 bar | 0.60    | 14      | 10.6 bar | 0.8   | 10.6 bar | 20      | 860000  | -6.411 | 1.09   |
| 10.6 bar | 9.9      | 87   | 10.30 bar | 0.92    | 2.8     | 10.30 bar | 0.61    | 12      | 10.6 bar | 0.8   | 10.6 bar | 20      | 831000  | -6.425 | 1.16   |
| 10.6 bar | 9.9      | 87   | 10.30 bar | 0.92    | 2.8     | 10.30 bar | 0.61    | 11      | 10.6 bar | 0.8   | 10.6 bar | 20      | 824000  | -6.482 | 1.24   |
| 10.6 bar | 9.9      | 88   | 10.30 bar | 0.92    | 2.8     | 10.30 bar | 0.61    | 10      | 10.6 bar | 0.9   | 10.6 bar | 20      | 824000  | -6.554 | 1.31   |

millions

Base Case Calculation for Problem 12.4 - T-201 and T-202 reversed



#### 14.17 Use of Low Pressure Steam vs Medium Pressure Steam in E-204.

The data given in Table 12.6 can be used to find the break-even pressure for the switch to lps from mps. The idea is to evaluate the FCI associated with the heat exchanger, E-204, with the cost of the steam. It is assumed here that the overall heat transfer coefficient for the reboiler is fixed at  $1140 \text{ W/m}^2\text{°C}$  for all comparisons. The costs of mps and lps are taken from Table 6.3 and the capital costs of the exchanger are taken from CAPCOST using a floating head exchanger.

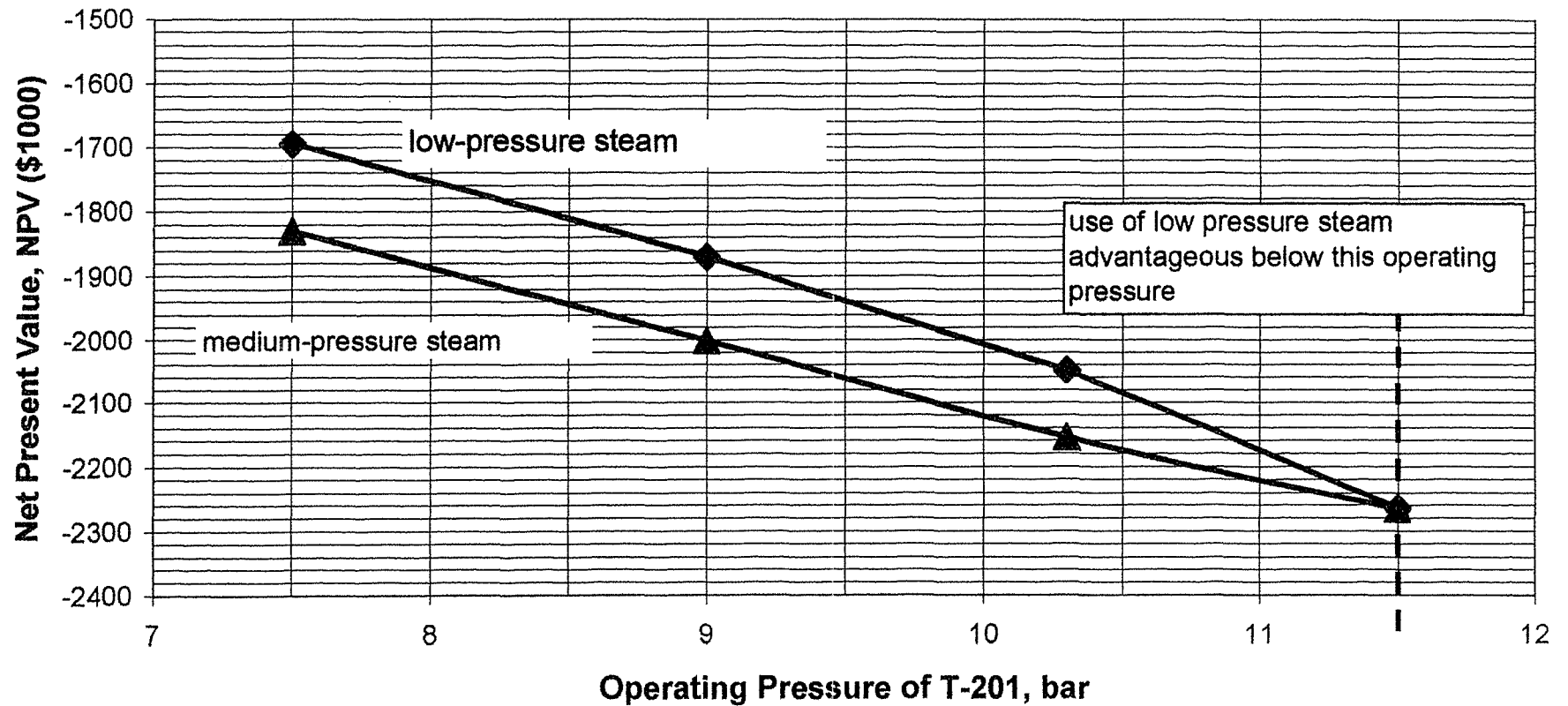
The NPV is evaluated as  $\text{NPV} = -(2.005 \times \text{FCI} + 6.871 \times C_{UT})$ . Since the only things that change are the size and cost of E-204 and the cost of the steam, these are the only costs that should be included in the NPV. A spreadsheet showing the comparisons is given on the next page and a figure showing the two cost curves as a function of T-201 operating pressure is also given. From the Figure, the operating pressure at which the switch to lps should take place is about 11.5 bar.

# Problem 19.11 - Comparison of Low Pressure to Medium Pressure Steam Usage in E-204

| Pres T-201<br>bar | T reboil<br>C | E-205 Duty<br>MJ/h | E-204                |                 |                        |                        |                 |                 |                |                |
|-------------------|---------------|--------------------|----------------------|-----------------|------------------------|------------------------|-----------------|-----------------|----------------|----------------|
|                   |               |                    | Duty cost<br>\$1,000 | cost<br>\$1,000 | Area<br>m <sup>2</sup> | Area<br>m <sup>2</sup> | Cost<br>\$1,000 | Cost<br>\$1,000 | NPV<br>\$1,000 | NPV<br>\$1,000 |
| 11.5              | 157           | 2689               | 285.6                | 308.8           | 218.4                  | 24.3                   | 151             | 72.2            | -2265          | -2266          |
| 10.3              | 153           | 2549               | 270.7                | 292.7           | 88.7                   | 20.0                   | 94.3            | 72.1            | -2049          | -2156          |
| 9                 | 147           | 2352               | 249.8                | 270.1           | 44.1                   | 15.5                   | 77              | 73.2            | -1871          | -2002          |
| 7.5               | 140           | 2127               | 225.9                | 244.2           | 25.9                   | 11.8                   | 71.6            | 75.8            | -1696          | -1830          |

### Comparison of LPS and MPS usage in E-204

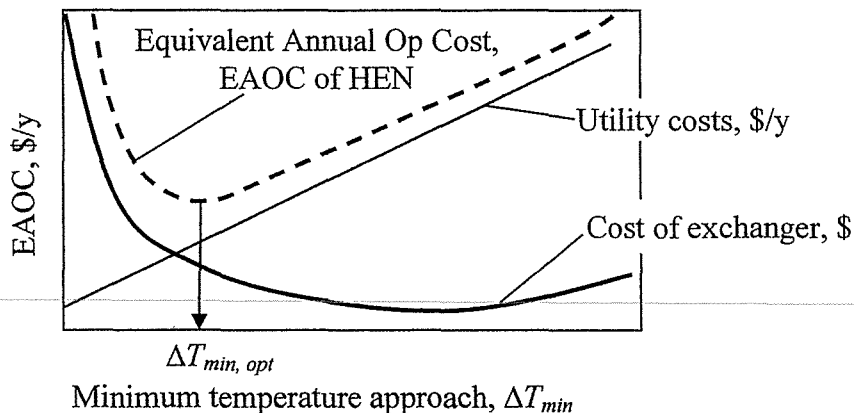
$$NPV = 2.005(FCI) + 6.871(\text{cost of steam})$$





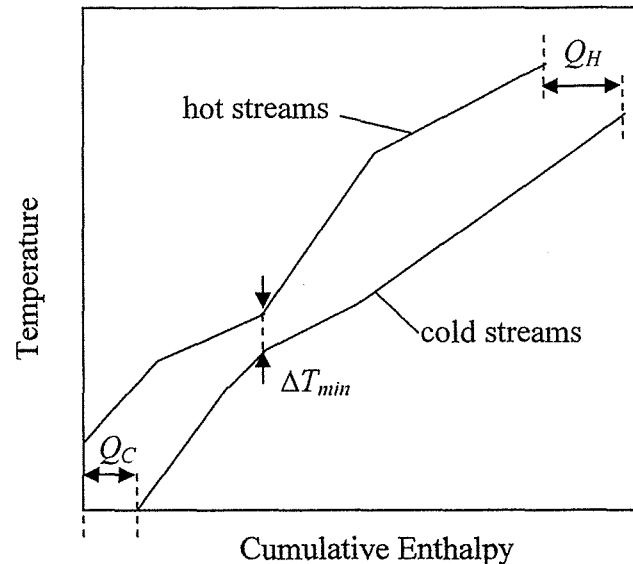
## Chapter 15

- 15.1 Costs affected by changing the min temp approach for a HEN are  
 Cost of utilities – hot and cold  
 Cost of heat exchangers in the network, including process-process and process-utility exchangers
- 15.2 By decreasing the  $\Delta T_{min}$  in a HEN, the process-process heat exchangers at the pinch will require larger areas and will therefore be more expensive. However, the amount and cost of hot and cold utilities for the HEN will be reduced.
- 15.3 (a) For streams requiring an MOC cheaper than CS, the film heat transfer coefficient,  $h$ , for the stream should be increased. This means that smaller heat exchange areas will be computed for these streams and this will lead to lower capital investment. This is equivalent to using a material factor,  $F_M < 1$  (for CS).  
 (b) The reverse is true for streams requiring an MOC more expensive than CS. These  $h$  values should be reduced, leading to larger more expensive heat exchangers. This is equivalent to using a material factor,  $F_M > 1$  (for CS).
- 15.4 As  $\Delta T_{min}$  increases the utility requirements increase (yearly operating costs) but the process-process exchangers become smaller and less expensive. However, the costs of the utility exchangers will increase. The costs for utilities and fixed capital investment (exchangers) are illustrated below. An optimum  $\Delta T_{min}$  will exist.



- 15.5 At the pinch, for streams above the pinch always match streams such that  $\dot{m}c_{p,hot} \leq \dot{m}c_{p,cold}$ . Using this criterion avoids violating the  $\Delta T_{min}$  criterion set at the start of the problem.

15.6 Composite temperature-enthalpy (T-Q) diagram



15.7 A cascade diagram illustrates the amounts of available energy, in excess of that required by process streams, at each temperature level in the temperature interval diagram. The cascade diagram is useful for visualizing how excess energy is cascaded downwards to heat lower temperature streams. It is also useful for identifying where the network pinch occurs.

15.8 Minimum number of exchangers =  $n + m + \# \text{ of utilities}$  (1 – for hot utility above pinch) - 1 =  $n + m$ . The actual number of exchanger may be lower than this if exact matches of energy are possible.

15.9 For streams that change phase, two approaches are used

(i) if the stream is pure and the phase change occurs at a single temperature then we may assume some arbitrary, small  $\Delta T$  of say  $1^\circ\text{C}$  and use an equivalent  $mc_p$  for that stream, where  $(mc_p)_{equiv} = \frac{m[\text{kg/s}]\lambda[\text{kJ/kg}]}{1[^\circ\text{C}]}$  and  $\lambda$  is the latent heat of vaporization.

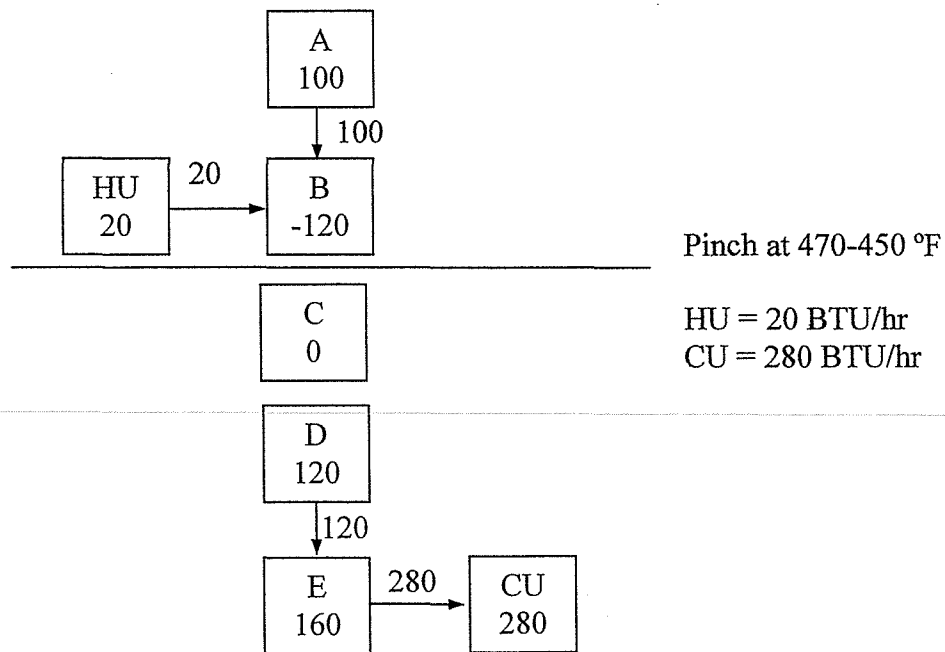
(ii) If the stream has several components and the heating curve is non-linear, then the curve can be approximated by two straight lines, the slopes of which represent equivalent  $mc_p$  values.

15.10 MUMNE = minimum utility, minimum number of exchangers heat exchanger network. It is the HEN that gives the minimum number of exchangers and minimum hot and cold utility duties for a system of process streams that are exchanging energy for a given (chosen) minimum temperature approach for the network.

15.11 (a)

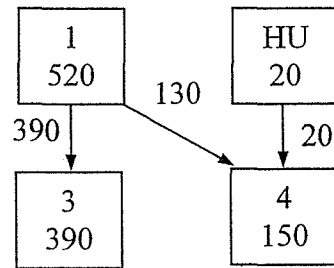
| $\dot{m}C_p$<br>(BTU/hr°F) | 4   | 4   |   | 3   | 5   |                     |               |
|----------------------------|-----|-----|---|-----|-----|---------------------|---------------|
| Stream<br>Number           | 1   | 2   |   | 3   | 4   |                     |               |
| Temperature<br>(°F)        |     |     |   |     |     | Temperature<br>(°F) | Q<br>(BTU/hr) |
| 600                        |     |     |   |     |     | 580                 |               |
|                            | 400 |     | A | 300 |     |                     | 100           |
| 500                        |     |     | B |     |     | 480                 |               |
|                            | 120 |     |   | 90  | 150 |                     | -120          |
| 470                        |     |     | C |     |     | 450                 |               |
|                            | 440 | 440 |   | 330 | 550 |                     | 0             |
| 360                        |     |     | D |     |     | 340                 |               |
|                            | 160 | 160 |   |     | 200 |                     | 120           |
| 320                        |     |     | E |     |     | 300                 |               |
|                            |     | 160 |   |     |     |                     | 160           |
| 280                        |     |     |   |     |     | 260                 |               |
|                            |     |     |   |     |     |                     | 260           |

(b)

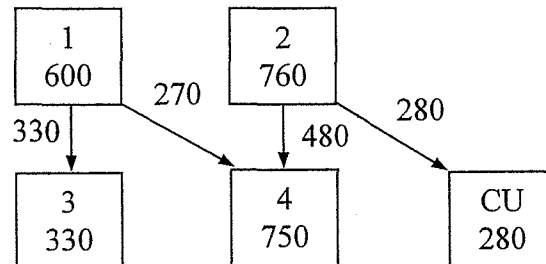


(c)

Above  
3 exchangers



Below  
4 exchangers



(d) Above  $\dot{m}C_{PH} \leq \dot{m}C_{PC} \Rightarrow$

Stream Number  
 $\dot{m}C_p$  (BTU/hr°F)

1

2

3

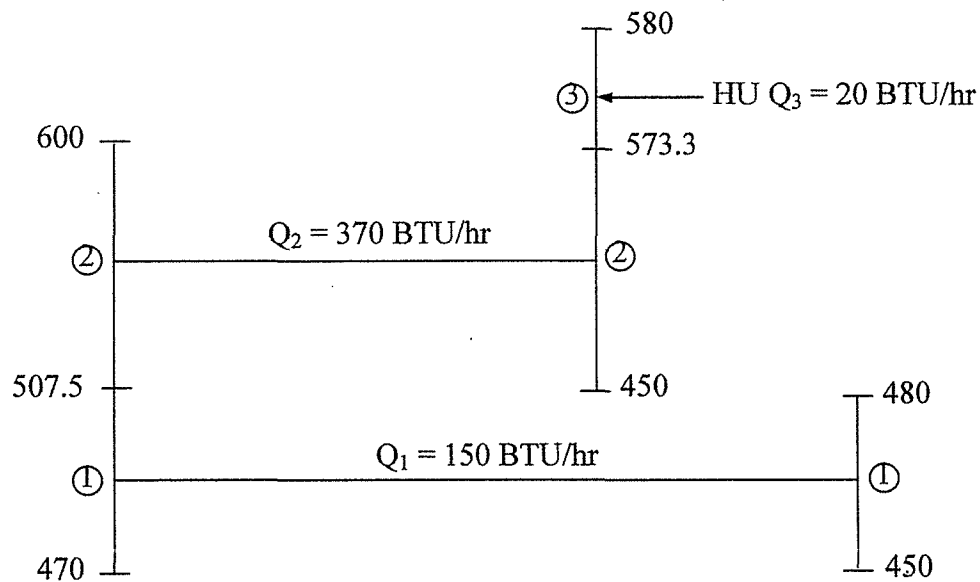
4

4

4

3

5



Or split stream 1

Stream Number  
 $\dot{m}C_p$  (BTU/hr°F)

1

2

3

4

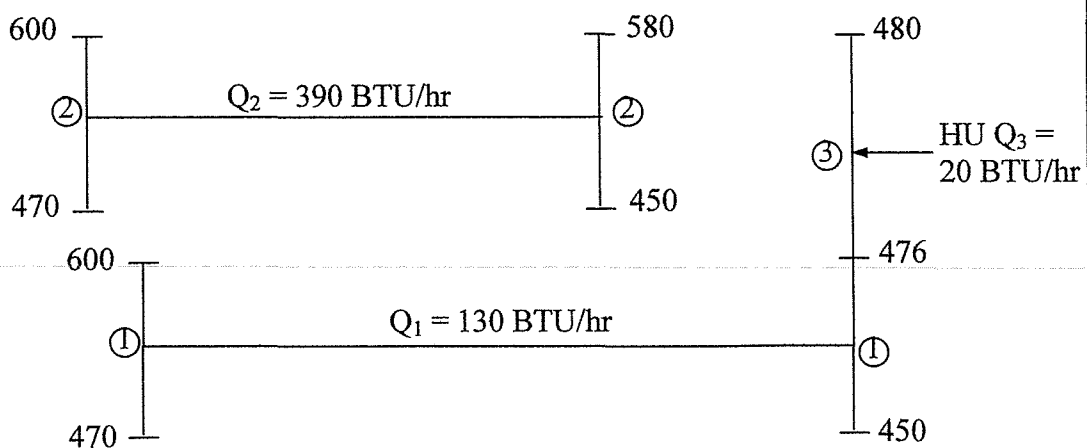
4

4

3

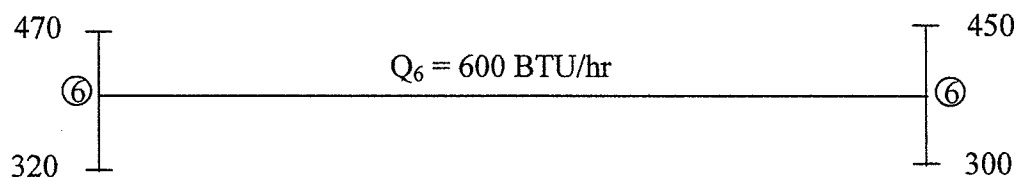
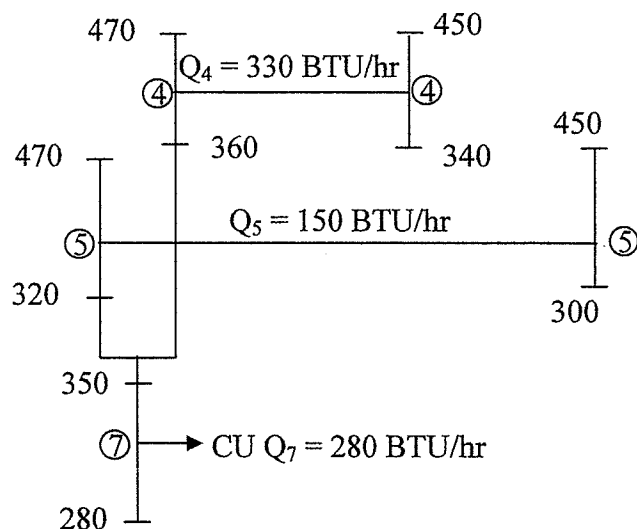
5

3 1

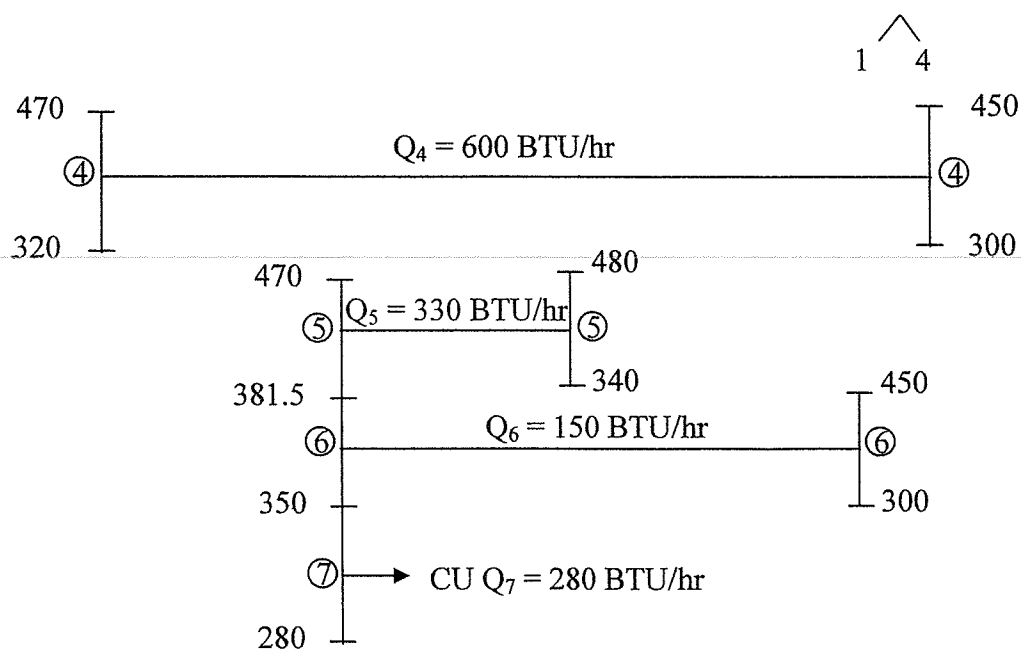


(e) Below pinch  $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

|                         |   |              |   |              |
|-------------------------|---|--------------|---|--------------|
| Stream Number           | 1 | 2            | 3 | 4            |
| $\dot{m}C_p$ (BTU/hr°F) | 4 | 4            | 3 | 5            |
|                         |   | 1 $\wedge$ 3 |   | 1 $\wedge$ 4 |



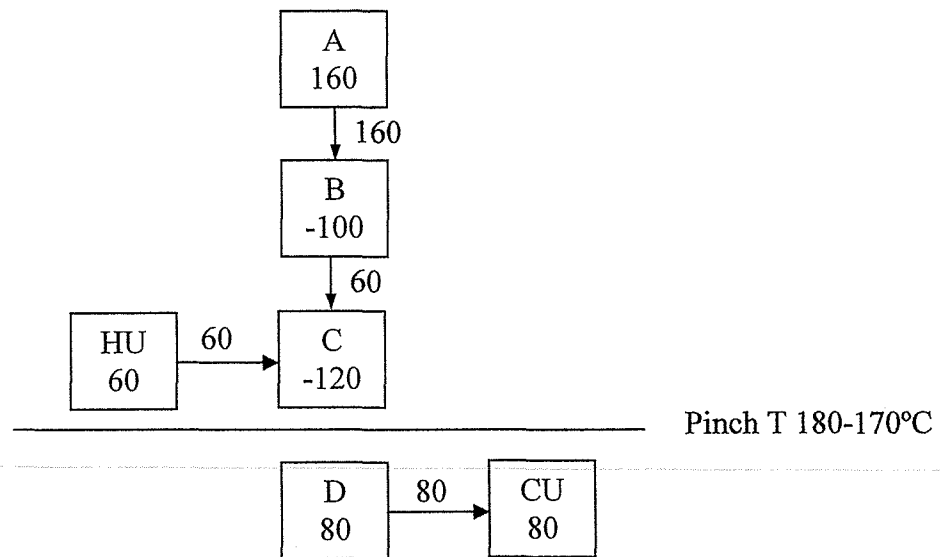
OR



15.12 (a)

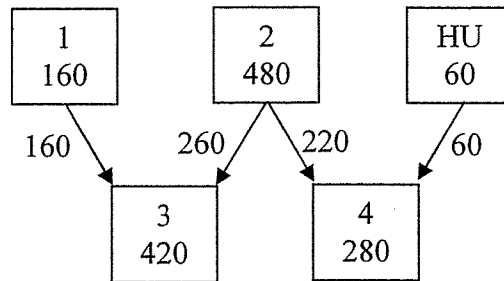
| Stream Number              | 1     | 2     | 3 | 4   |                     |               |
|----------------------------|-------|-------|---|-----|---------------------|---------------|
| $\dot{m}C_p$<br>(BTU/hr°F) | 2     | 4     | 3 | 2   |                     |               |
| Temperature<br>(°F)        |       |       |   |     | Temperature<br>(°F) | Q<br>(BTU/hr) |
| 400                        | ↓ 160 |       | A |     | 390                 | 160           |
| 320                        |       |       |   |     | 310                 | -100          |
| 300                        |       | 480   | C | 360 | 290                 | -120          |
| 180                        |       |       |   |     | 170                 | 80            |
| 100                        |       | ↓ 320 | D | 240 | 90                  |               |
|                            |       |       |   |     |                     | <hr/> 20      |

(b)

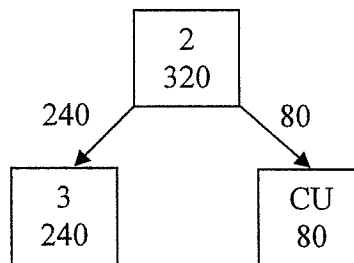


(c)

Above  
4 Exchangers

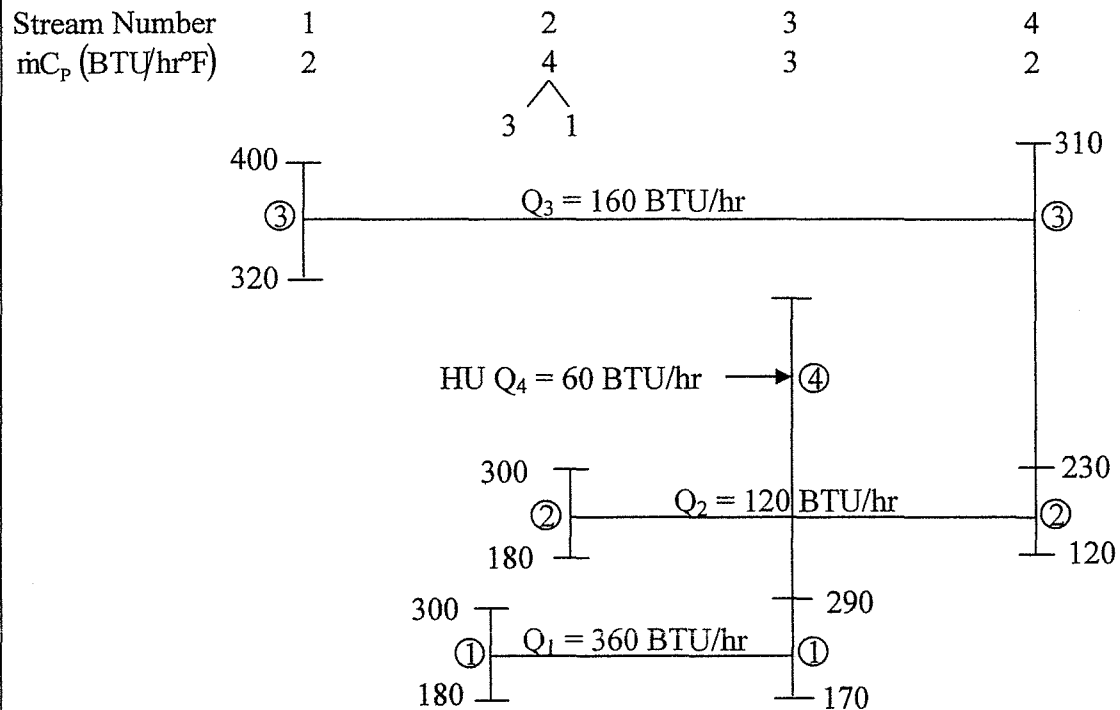


Below  
2 Exchangers

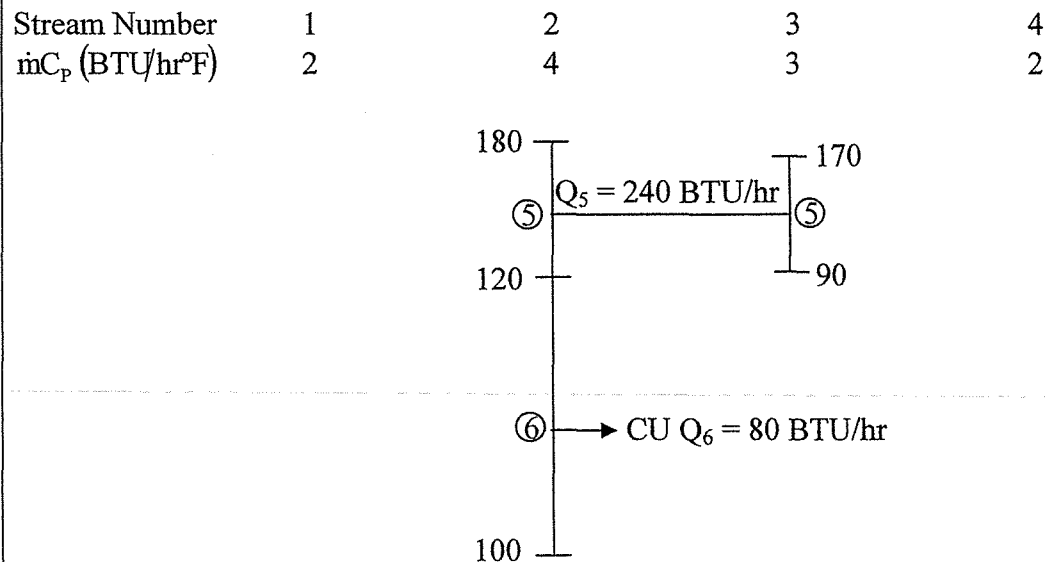




(d) Above pinch  $\dot{m}C_{PH} \leq \dot{m}C_{PC}$



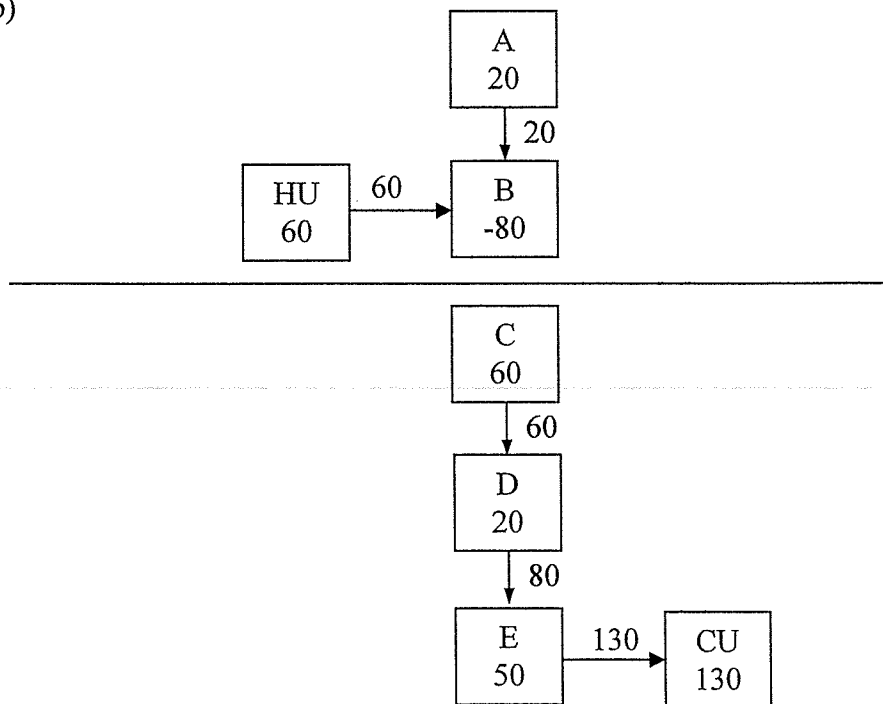
(e) Below pinch  $\dot{m}C_{PH} \geq \dot{m}C_{PC}$



15.13 (a)

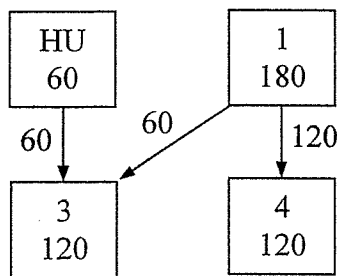
| Stream Number              | 1     | 2    | 3 | 4     |                     |               |
|----------------------------|-------|------|---|-------|---------------------|---------------|
| $\dot{m}C_p$<br>(BTU/hr°F) | 3     | 5    | 3 | 2     |                     |               |
| Temperature<br>(°F)        |       |      |   |       | Temperature<br>(°F) | Q<br>(BTU/hr) |
| 180                        |       |      |   | ↑ 40  | 160                 | 20            |
|                            | 60    |      | A |       |                     |               |
| 160                        | ↓ 120 |      | B | ↑ 120 | 140                 | -80           |
|                            |       |      |   |       |                     |               |
| 120                        |       | 100  | C | 60    | 100                 | 60            |
|                            | ↓ 60  |      |   |       |                     |               |
| 100                        |       | 50   | D | 30    | 80                  | 20            |
|                            |       |      |   |       |                     |               |
| 90                         |       | ↓ 50 | E |       | 70                  | 50            |
|                            |       |      |   |       |                     |               |
| 80                         |       |      |   |       | 60                  |               |
|                            |       |      |   |       |                     |               |
|                            |       |      |   |       |                     | <hr/> 70      |

(b)

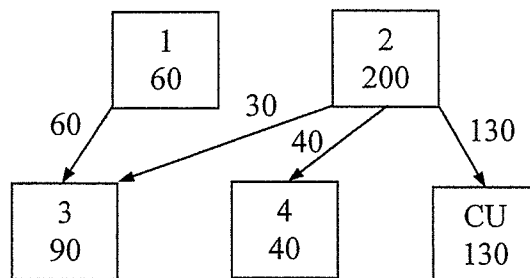


(c)

Above  
3 exchangers

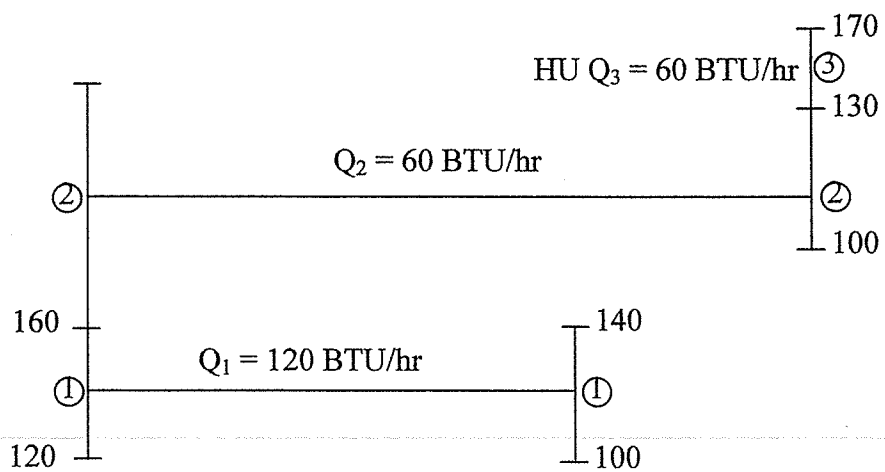


Below  
4 exchangers



(d) Above pinch  $\dot{m}C_{PH} \leq \dot{m}C_{PC}$

|                         |   |   |   |   |
|-------------------------|---|---|---|---|
| Stream Number           | 1 | 2 | 3 | 4 |
| $\dot{m}C_p$ (BTU/hr°F) | 3 | 5 | 3 | 2 |

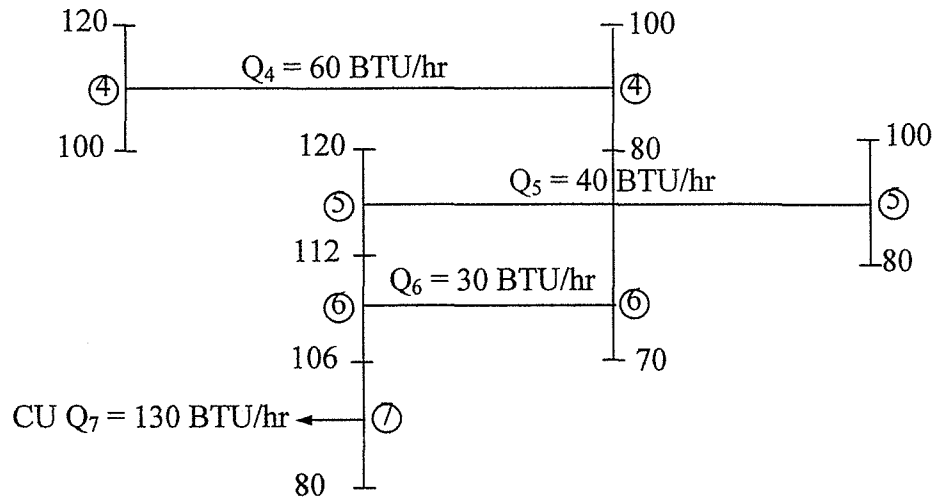


(e)

Below pinch  $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

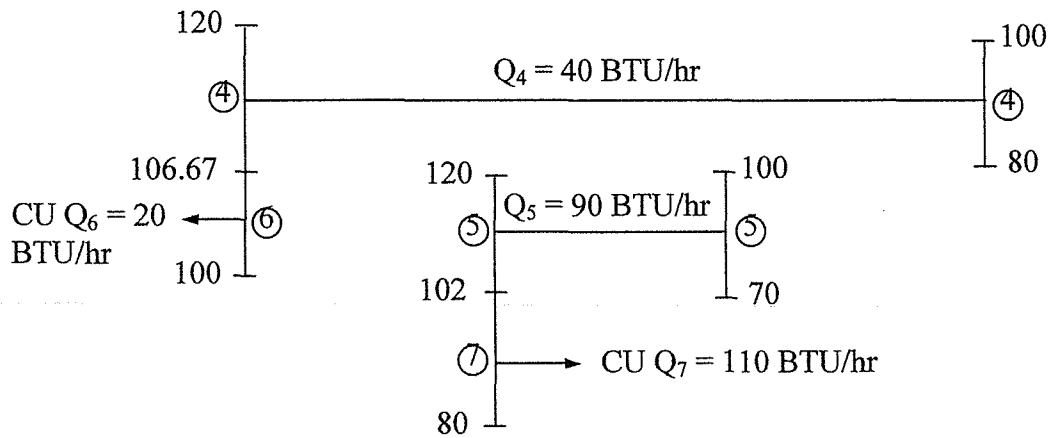
Only one utility stream

|                         |   |   |   |   |
|-------------------------|---|---|---|---|
| Stream Number           | 1 | 2 | 3 | 4 |
| $\dot{m}C_p$ (BTU/hr°F) | 3 | 5 | 3 | 2 |



For two utility streams

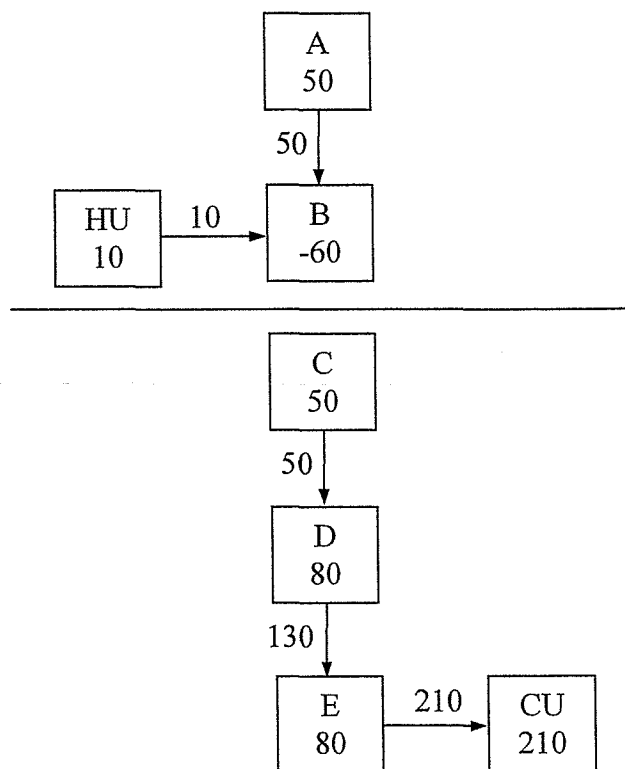
|                         |   |   |   |   |
|-------------------------|---|---|---|---|
| Stream Number           | 1 | 2 | 3 | 4 |
| $\dot{m}C_p$ (BTU/hr°F) | 3 | 5 | 3 | 2 |



15.14 (a)

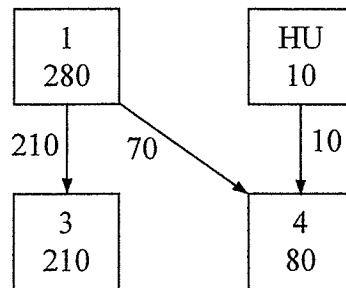
| Stream Number              | 1   | 2   |   | 3   | 4   |                     |               |
|----------------------------|-----|-----|---|-----|-----|---------------------|---------------|
| $\dot{m}C_p$<br>(BTU/hr°F) | 4   | 4   |   | 3   | 4   |                     |               |
| Temperature<br>(°F)        |     |     |   |     |     | Temperature<br>(°F) | Q<br>(BTU/hr) |
| 500                        |     |     |   |     |     | 490                 |               |
|                            | 200 |     | A | 150 |     |                     | 50            |
| 450                        |     |     |   |     |     | 440                 |               |
|                            | 80  |     | B | 60  | 80  |                     | -60           |
| 430                        |     |     |   |     |     | 420                 |               |
|                            | 200 | 200 | C | 150 | 200 |                     | 50            |
| 380                        |     |     |   |     |     | 370                 |               |
|                            | 80  | 80  | D |     | 80  |                     | 80            |
| 360                        |     |     |   |     |     | 350                 |               |
|                            |     | 80  | E |     |     |                     | 80            |
| 340                        |     |     |   |     |     | 330                 |               |
|                            |     |     |   |     |     |                     | 200           |

(b)

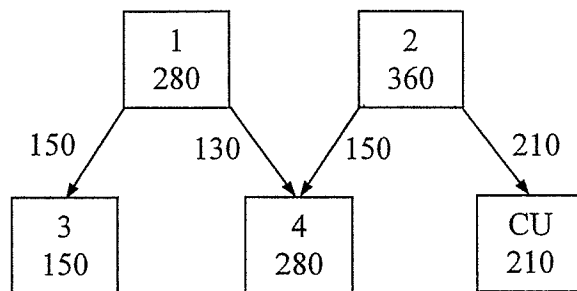


(c)

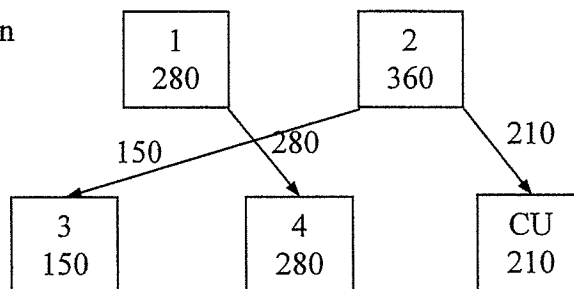
Above pinch  
3 exchangers



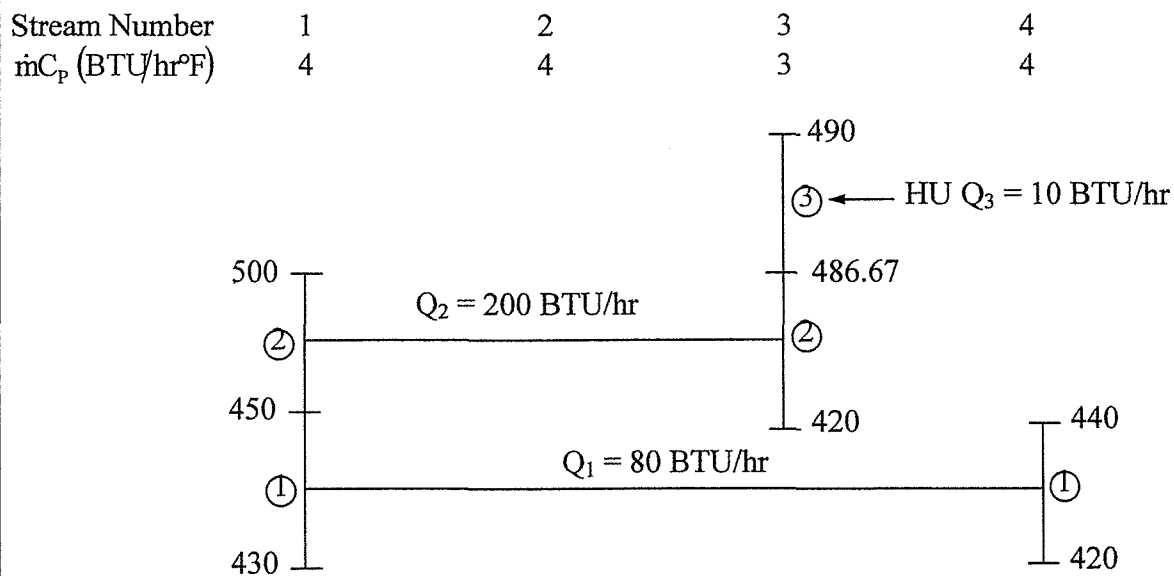
Below pinch  
4 exchangers



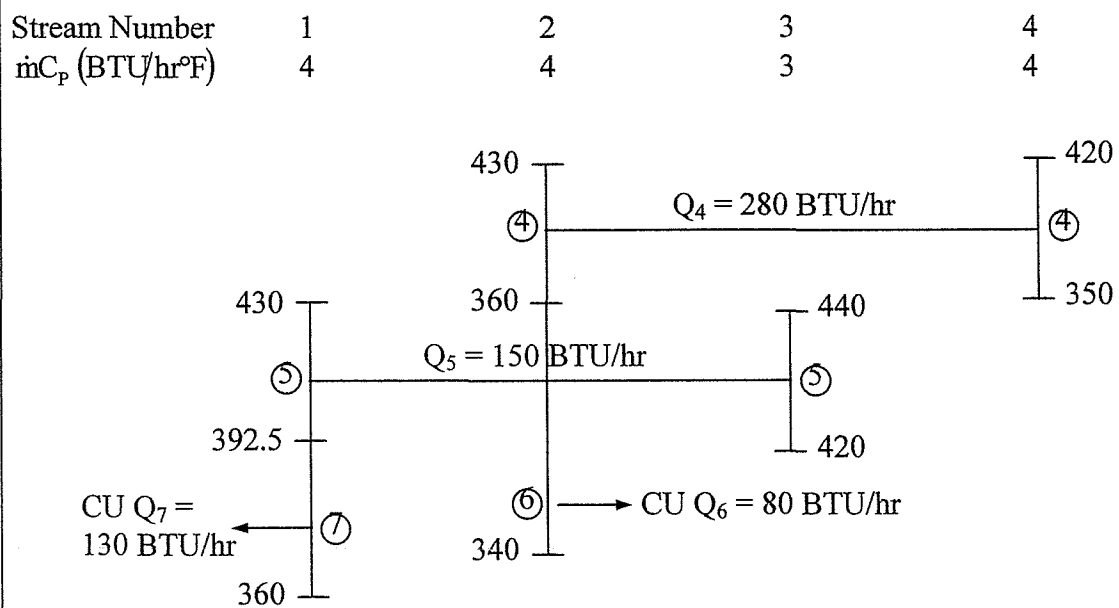
Exact Match Solution  
Below pinch  
3 exchangers



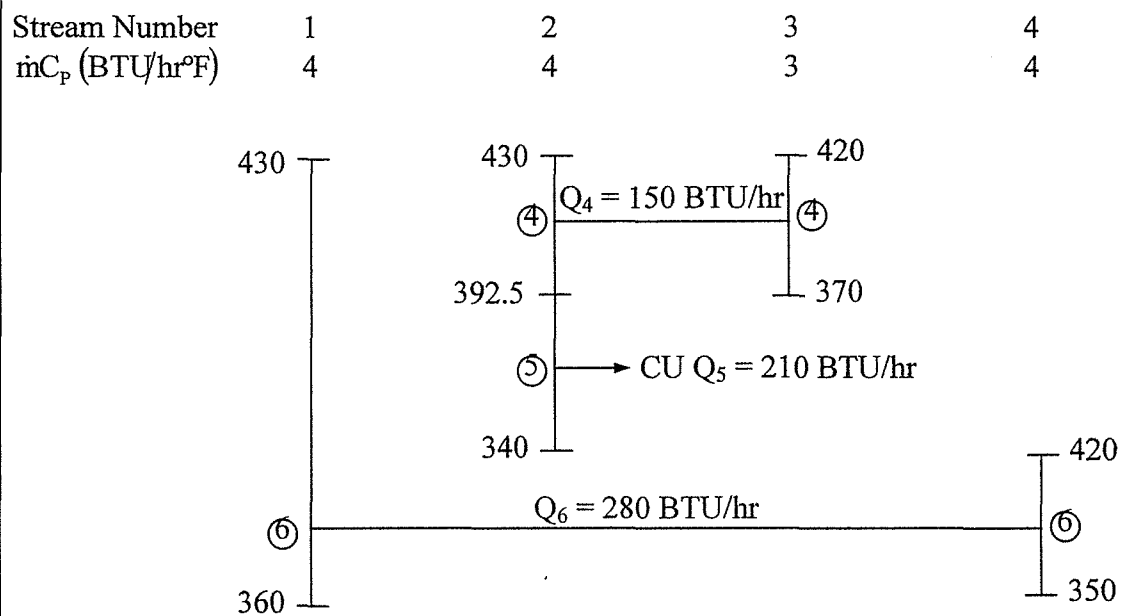
(d) Above pinch  $\dot{m}C_{PH} \leq \dot{m}C_{PC}$



(e) Below pinch  $\dot{m}C_{PH} \geq \dot{m}C_{PC}$



Exact match solution

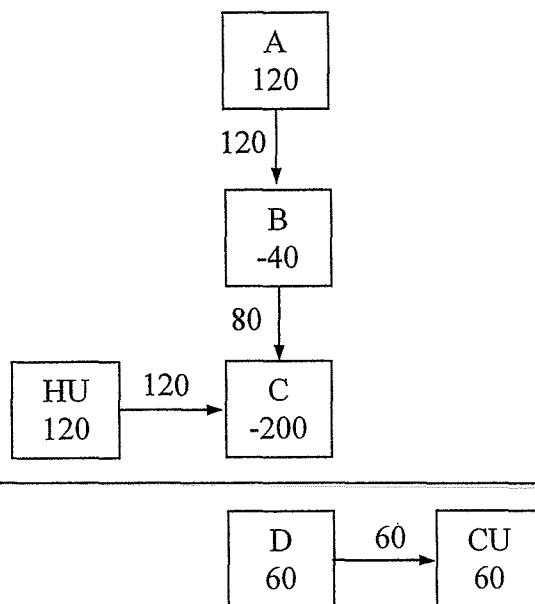




15.15 (a)

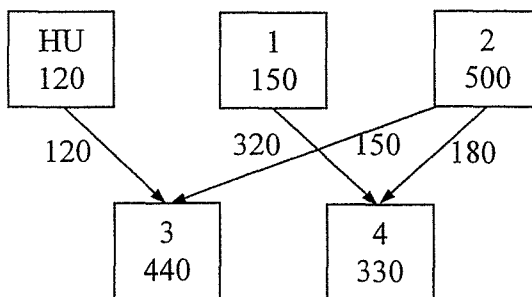
|                            |       |     |   |     |                  |            |
|----------------------------|-------|-----|---|-----|------------------|------------|
| Stream Number              | 1     | 2   | 3 | 4   |                  |            |
| $\dot{m}C_p$<br>(BTU/hr°F) | 3     | 5   | 4 | 3   |                  |            |
| Temperature (°F)           |       |     |   |     | Temperature (°F) | Q (BTU/hr) |
| 250                        | ↓ 120 |     | A |     | 240              | 120        |
| 210                        |       |     |   |     | 200              | -40        |
| 200                        | ↓ 30  | 500 | C | 400 | 190              | -200       |
| 100                        |       |     |   |     | 90               | 60         |
| 40                         | ↓ 300 | D   |   | 240 | 30               |            |
|                            |       |     |   |     |                  |            |
|                            |       |     |   |     |                  | <hr/> -60  |

(b)

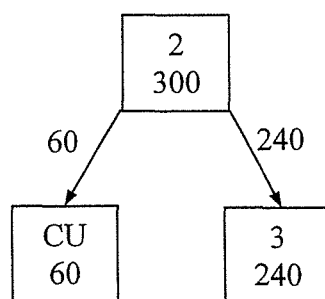


(c)

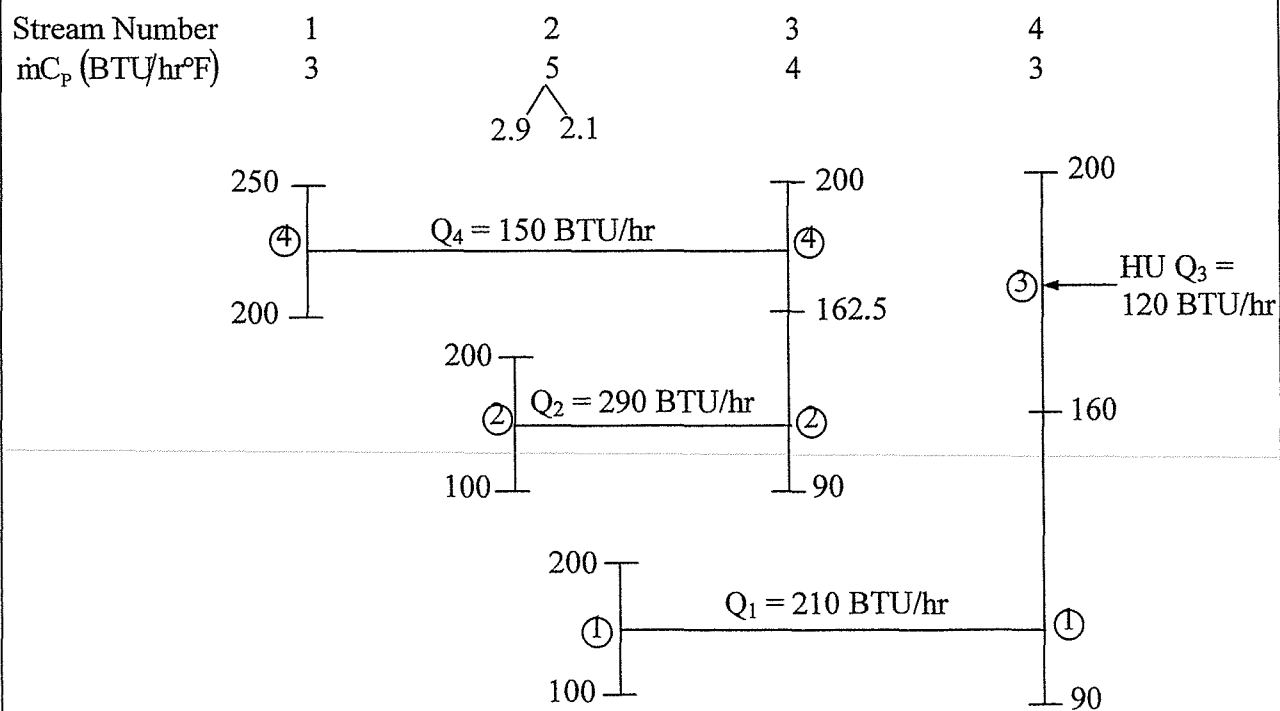
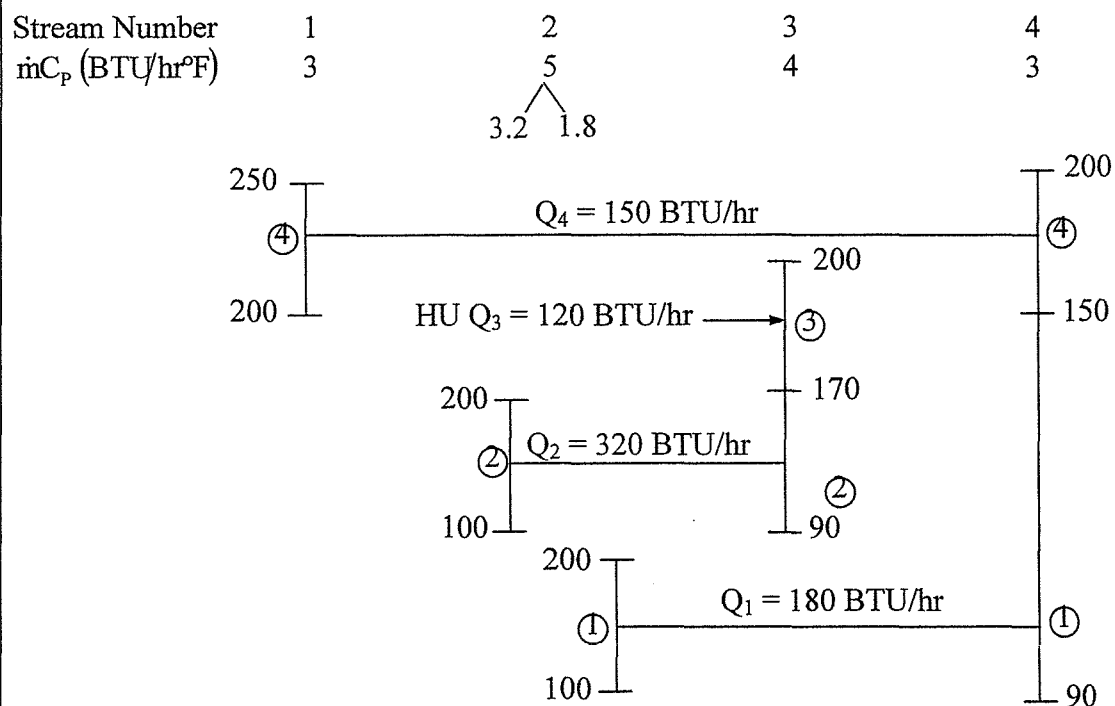
Above pinch  
4 exchangers



Below pinch  
2 exchangers



(d) Above pinch  $\dot{m}C_{PH} \leq \dot{m}C_{PC}$

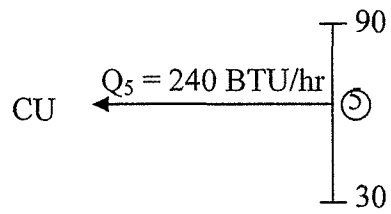


(e) Below pinch  $\dot{m}C_{PH} \geq \dot{m}C_{PC}$

Stream Number  
 $\dot{m}C_p$  (BTU/hr°F)

4

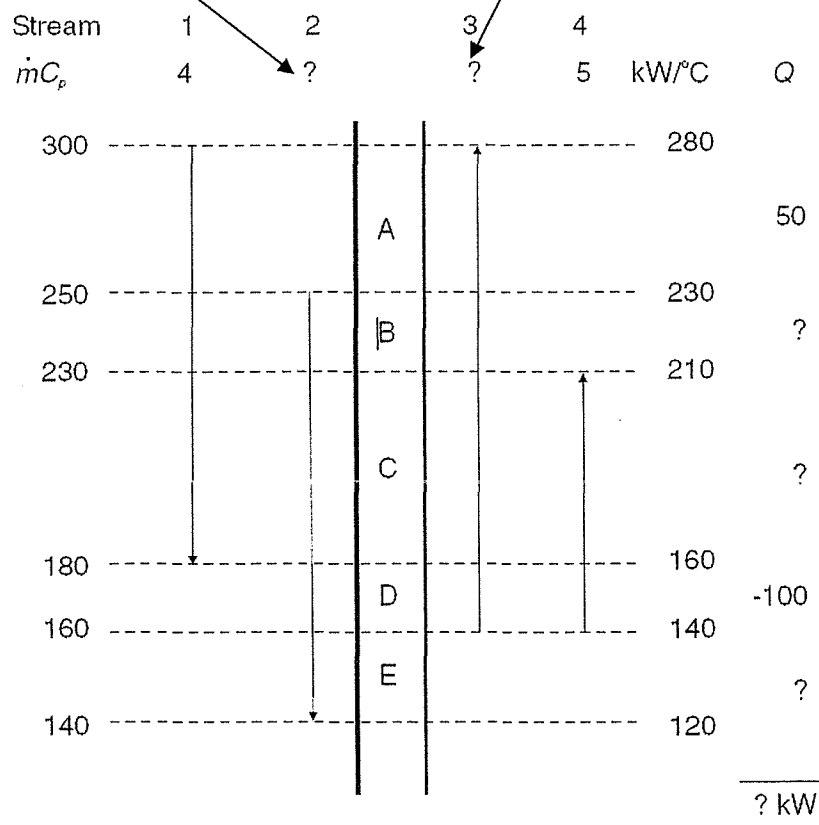
3



15.16 (a)

From interval D  
 $(mc_p)_2(20) - (3+5)(20) = -100$   
 $(mc_p)_2 = 3$

From interval A  
 $(4)(50) - (mc_p)_3(50) = 50 \rightarrow (mc_p)_3 = 3$

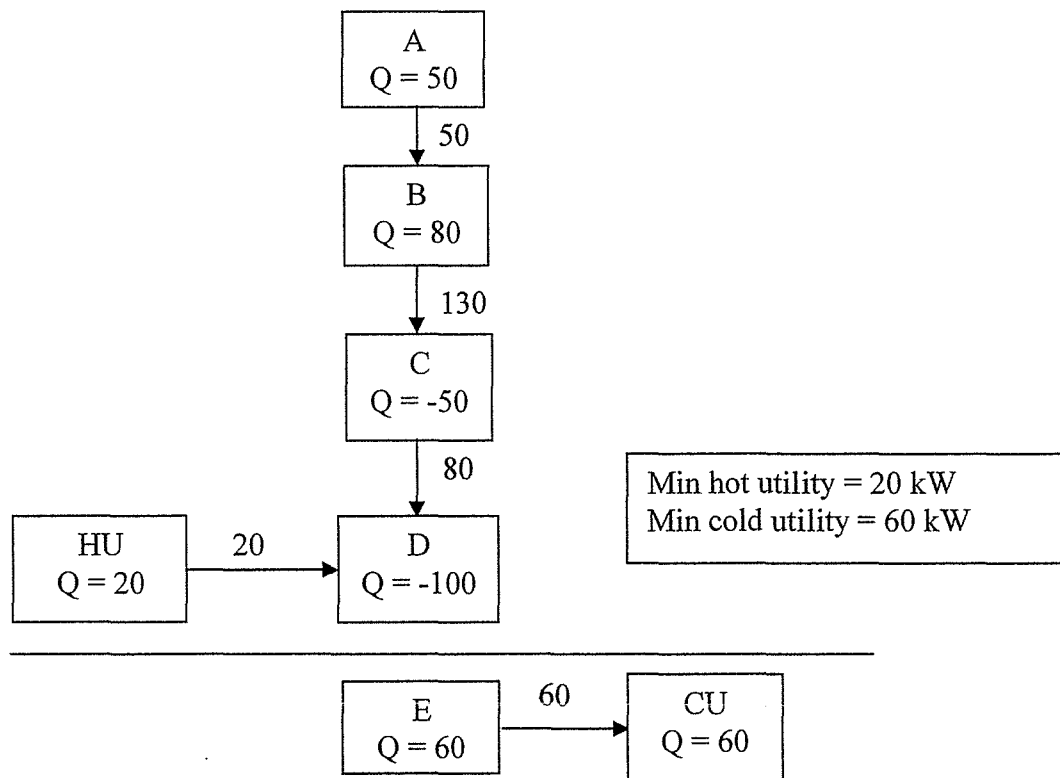


$$Q_B = (4 + 3 - 3)(20) = 80 \text{ kW}$$

$$Q_C = (4 + 3 - 3 - 5)(50) = -50 \text{ kW}$$

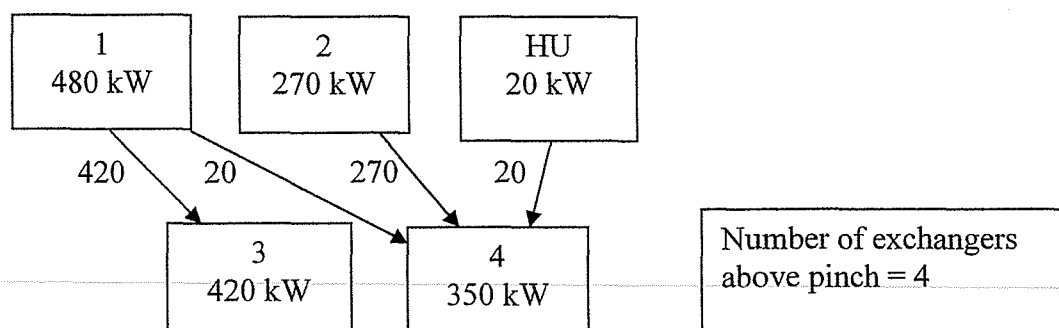
$$Q_E = (3)(20) = 60 \text{ kW}$$

(b)

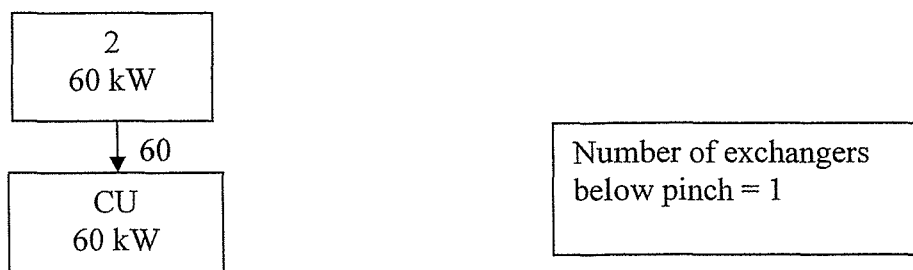


(c)  $T_{\text{pinch, hot}} = 160^{\circ}\text{C}$   
 $T_{\text{pinch, cold}} = 140^{\circ}\text{C}$

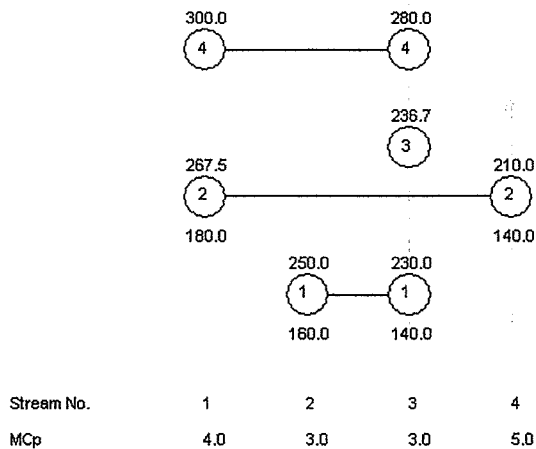
(d) Above Pinch



Below Pinch



### Design above pinch



Number of Exchangers Remaining 0

☒ Process Exchanger (at or away from the Pinch)

☐ Utility Exchanger

| Exchanger | Duty | DT Violation | Area  |
|-----------|------|--------------|-------|
| 1         | 270  | No           | 875   |
| 2         | 350  | No           | 382.9 |
| 3         | 20   | Hot Util Ex  | 28.2  |
| 4         | 130  | No           | 259.7 |

Enter ExchangerDuty  
(positive numbers only)

130

Split Streams

Enter Enthalpy

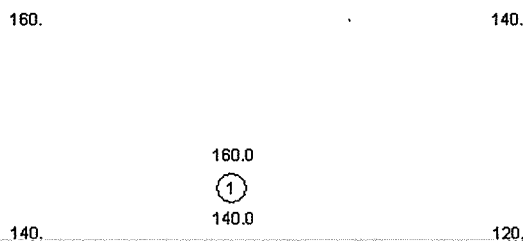
Start Again

Hot Streams - Total  
Enthalpy Remaining

Cold Streams - Total  
Enthalpy Remaining

| Stream  | Enthalpy | Stream | Enthalpy |
|---------|----------|--------|----------|
| Total   | 0        | Total  | 0        |
| Utility | 0        | 3      | 0        |
| 1       | 0        | 4      | 0        |
| 2       | 0        |        |          |
|         |          |        |          |
|         |          |        |          |
|         |          |        |          |
|         |          |        |          |
|         |          |        |          |

### Design below pinch



Number of Exchangers Remaining -1

☐ Process Exchanger (at or away from the Pinch)

☒ Utility Exchanger

| Exchanger | Duty | DT Violation | Area |
|-----------|------|--------------|------|
| 1         | 60   | Cold Util Ex | 79.8 |

Enter ExchangerDuty  
(positive numbers only)

60

Split Streams

Enter Enthalpy

Start Again

Hot Streams - Total  
Enthalpy Remaining

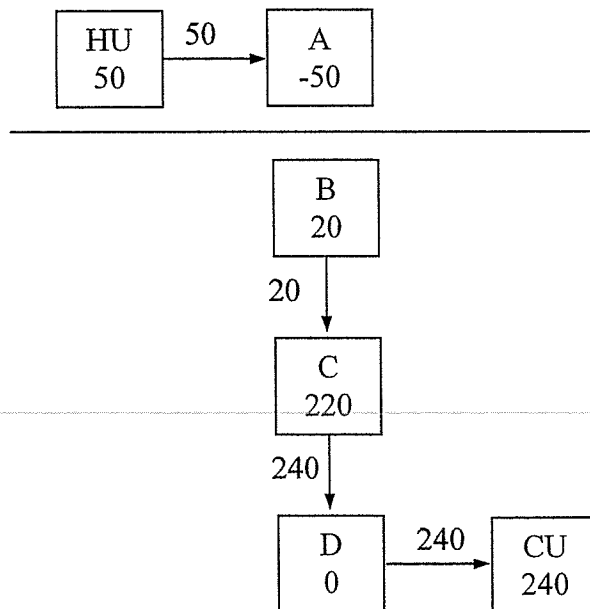
Cold Streams - Total  
Enthalpy Remaining

| Stream | Enthalpy | Stream  | Enthalpy |
|--------|----------|---------|----------|
| Total  | 0        | Total   | 0        |
| 1      | 0        | Utility | 0        |
| 2      | 0        | 3       | 0        |
|        |          | 4       | 0        |
|        |          |         |          |
|        |          |         |          |
|        |          |         |          |
|        |          |         |          |
|        |          |         |          |

15.17 (a)

| Stream Number              | 1   | 2   |   | 3   | 4   |                     |               |
|----------------------------|-----|-----|---|-----|-----|---------------------|---------------|
| $\dot{m}C_p$<br>(BTU/hr°F) | 2   | 4   |   | 5   | 4   |                     |               |
| Temperature<br>(°F)        |     |     |   |     |     | Temperature<br>(°F) | Q<br>(BTU/hr) |
| 400                        |     |     |   |     |     | 380                 |               |
|                            |     | 200 | A | 250 |     |                     | -50           |
| 350                        |     |     |   |     |     | 330                 |               |
|                            | 40  | 80  | B | 100 |     |                     | 20            |
| 330                        |     |     |   |     |     | 310                 |               |
|                            | 220 | 440 | C |     | 440 |                     | 220           |
| 220                        |     |     |   |     |     | 200                 |               |
|                            |     | 400 | D |     | 400 |                     | 0             |
| 120                        |     |     |   |     |     | 100                 |               |
|                            |     |     |   |     |     |                     | 190           |

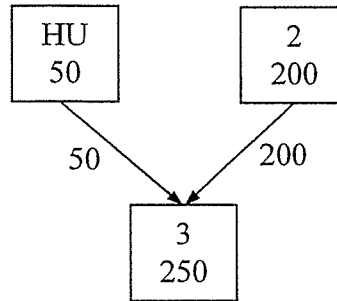
(b)



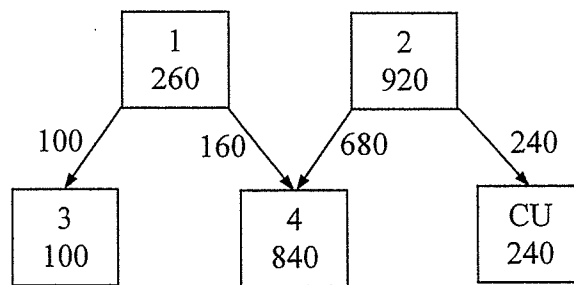


(c)

Above pinch  
2 exchangers

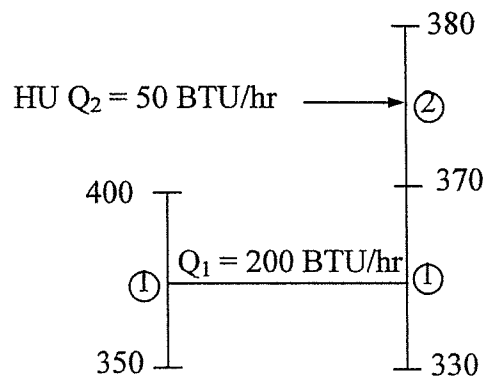


Below pinch  
4 exchangers

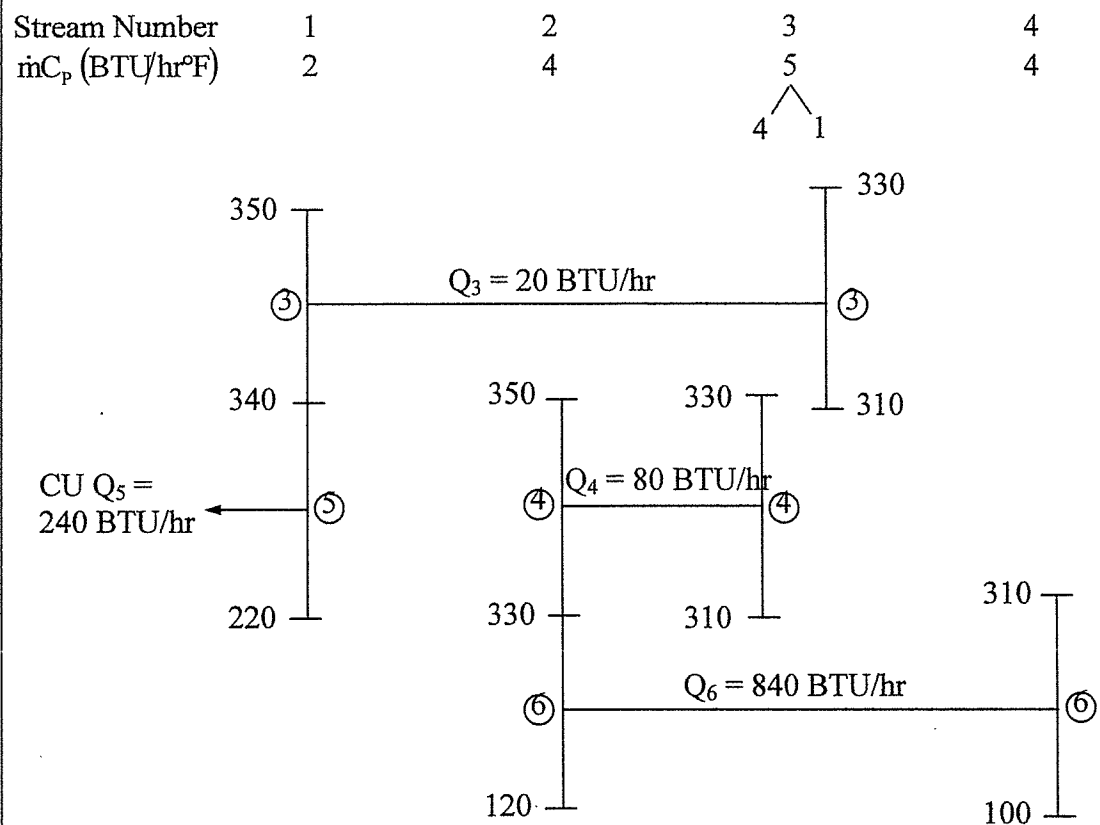


(d) Above pinch  $\dot{m}C_{PH} \leq \dot{m}C_{PC}$

|                         |   |   |   |   |
|-------------------------|---|---|---|---|
| Stream Number           | 1 | 2 | 3 | 4 |
| $\dot{m}C_p$ (BTU/hr°F) | 2 | 4 | 5 | 4 |



(e) Below pinch  $\dot{m}C_{PH} \geq \dot{m}C_{PC}$



# Heat Exchanger Networks - Synthesis, Analysis, and Design (HENSAD)

File Worksheet System Help

## Table of Results

Minimum Temperature Approach = 20°F

## Hot Stream Data

| Mass Flow<br>M-lb/h | Cp<br>Btu/lb°F | Temp In<br>°F | Temp Out<br>°F | Stream Enthalpy<br>MBtu/h | Film Heat Transf. Coef<br>Btu/h/ft <sup>2</sup> /°F |
|---------------------|----------------|---------------|----------------|---------------------------|-----------------------------------------------------|
| 1.000               | 1.000          | 620.0         | 320.0          | 300.0                     | 75.00                                               |
| 6.000               | 1.000          | 420.0         | 120.0          | 1800.                     | 25.00                                               |
| 3.000               | 1.000          | 420.0         | 220.0          | 600.0                     | 10.00                                               |

Cumulative Hot Stream Energy Available = 2700.0 MBtu/h

## Cold Stream Data

| Mass Flow<br>M-lb/h | Cp<br>Btu/lb°F | Temp In<br>°F | Temp Out<br>°F | Stream Enthalpy<br>MBtu/h | Film Heat Transf. Coef<br>Btu/h/ft <sup>2</sup> /°F |
|---------------------|----------------|---------------|----------------|---------------------------|-----------------------------------------------------|
| 5.000               | 1.000          | 400.0         | 600.0          | -1000.                    | 45.00                                               |
| 2.000               | 1.000          | 200.0         | 300.0          | -200.0                    | 30.00                                               |
| 4.000               | 1.000          | 100.0         | 400.0          | -1200.                    | 30.00                                               |

Cumulative Cold Stream Energy Available = -2400.0 MBtu/h

## Data for Generating Temperature Interval Diagram

Number of Temperature Intervals = 4

| Interval | Temperature Range<br>°F |       | Excess Heat<br>MBtu/h | Cumulative Q<br>MBtu/h |
|----------|-------------------------|-------|-----------------------|------------------------|
| A        | 620.0                   | 420.0 | -800.0                | -800.0                 |
| B        | 420.0                   | 320.0 | 600.0                 | -200.0                 |
| C        | 320.0                   | 220.0 | 300.0                 | 100.0                  |
| D        | 220.0                   | 120.0 | 200.0                 | 300.0                  |

Pinch Temperature - Hot = 420°F  
Pinch Temperature - Cold = 400°F

Part (b)

Hot Utility Requirement = 800 MBtu/h  
Cold Utility Requirement = 1100 MBtu/h

Part (a)

Minimum Number of Exchanger Required to Accomplish Minimum Utility Loads  
In Special Circumstances the Minimum Required may be Lower than Indicated Below

Number Above the Pinch = 2  
Number Below the Pinch = 5

Part (c)

## Data for Composite Enthalpy - Temperature Diagram

| Temperature<br>°F | Hot Stream Enthalpy<br>MBtu/h | Temperature<br>°F | Cold Stream Enthalpy<br>MBtu/h |
|-------------------|-------------------------------|-------------------|--------------------------------|
| 120.0             | .0000                         | 100.0             | 1100.                          |
| 220.0             | 600.0                         | 200.0             | 1500.                          |
| 320.0             | 1500.                         | 300.0             | 2100.                          |
| 420.0             | 2500.                         | 400.0             | 2500.                          |
| 620.0             | 2700.                         | 600.0             | 3500.                          |

Heat Transfer Area for Process Exchangers in Network = 2220 ft<sup>2</sup>

(d)

### Network Design Above the Pinch

Help Print Return to Main Menu Exchanger Design Report

Number of Exchangers Remaining 0

☐ Process Exchanger (at or away from the Pinch)

☒ Utility Exchanger

| Exchanger | Duty | DT Violator | Area  |
|-----------|------|-------------|-------|
| 1         | 200  | No          | 122   |
| 2         | 800  | Hot Util Ex | 774.2 |

Enter Exchanger Duty  
(positive numbers only)

800

Split Streams

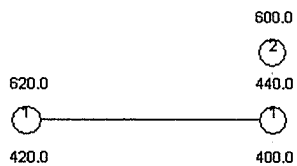
Enter Enthalpy

Start Again

Hot Streams - Total  
Enthalpy Remaining

Cold Streams - Total  
Enthalpy Remaining

| Stream  | Enthalpy | Stream | Enthalpy |
|---------|----------|--------|----------|
| Total   | 0        | Total  | 0        |
| Utility | 0        | 4      | 0        |
| 1       | 0        | 5      | 0        |
| 2       | 0        | 6      | 0        |
| 3       | 0        |        |          |
|         |          |        |          |
|         |          |        |          |
|         |          |        |          |
|         |          |        |          |



| Stream No. | 1   | 2   | 3   | 4   | 5   | 6   |
|------------|-----|-----|-----|-----|-----|-----|
| MCp        | 1.0 | 6.0 | 3.0 | 5.0 | 2.0 | 4.0 |

### Network Design Below the Pinch

Help Print Return to Main Menu Exchanger Design Report

Number of Exchangers Remaining 0

☐ Process Exchanger (at or away from the Pinch)

☒ Utility Exchanger

| Exchanger | Duty | DT Violator  | Area   |
|-----------|------|--------------|--------|
| 1         | 1200 | No           | 1970.9 |
| 2         | 200  | No           | 245.1  |
| 3         | 100  | Cold Util Ex | 70.7   |
| 4         | 400  | Cold Util Ex | 1496.6 |

Enter Exchanger Duty  
(positive numbers only)

600

Split Streams

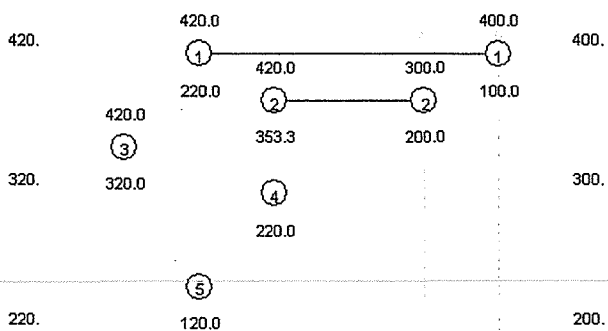
Enter Enthalpy

Start Again

Hot Streams - Total  
Enthalpy Remaining

Cold Streams - Total  
Enthalpy Remaining

| Stream | Enthalpy | Stream  | Enthalpy |
|--------|----------|---------|----------|
| Total  | 0        | Total   | 0        |
| 1      | 0        | Utility | 0        |
| 2      | 0        | 4       | 0        |
| 3      | 0        | 5       | 0        |
|        |          | 6       | 0        |
|        |          |         |          |
|        |          |         |          |
|        |          |         |          |
|        |          |         |          |



| Stream No. | 1   | 2   | 3   | 4   | 5   | 6   |
|------------|-----|-----|-----|-----|-----|-----|
| MCp        | 1.0 | 6.0 | 3.0 | 5.0 | 2.0 | 4.0 |

15.19 Using the heat transfer coefficients given, use the Sensitivity Plot feature of HENSAD to generate the following data and plot

From Problem 15.18, the heat transfer area for the network is 2220 ft<sup>2</sup>

**Input Range and Data for Plots**

Return to Main Menu   Generate Data   Plot Area and Utilities   Plot EAO   Print

Minimum Value of Approach Temperature:  °F

Maximum Value of Approach Temperature:  °F

Number of points (maximum = 100):

Hot Utility Film Coefficient:  Btu/hr/ft<sup>2</sup>/°F

Cold Utility Film Coefficient:  Btu/hr/ft<sup>2</sup>/°F

Temperature Driving Force in Hot Utility Exchangers:  °F

Temperature Driving Force in Cold Utility Exchangers:  °F

Price for Hot Utility:  \$/million Btu

Price for Cold Utility:  \$/million Btu

Time Period for Cost Analysis:  years

Interest Rate:  % p.a.

**Exchanger Price**

log(Total Module Cost, \$) = K1 + K2 log(A)

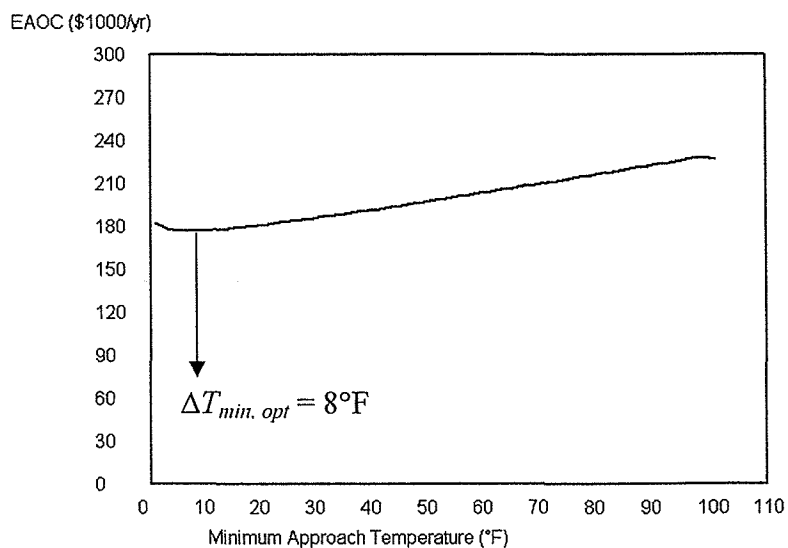
log = log 10   A = Area, ft<sup>2</sup>   K1:    K2:

| Results             |                           |                      |                       |                 |                      |
|---------------------|---------------------------|----------------------|-----------------------|-----------------|----------------------|
| MinTemp Approach °F | Exch Area ft <sup>2</sup> | Hot Util Load MBtu/h | Cold Util Load MBtu/h | EAO (1000)\$/yr | Number of Exchangers |
| 1.000               | 7671.6                    | 705.0                | 1005.                 | 182.5           | 8.000                |
| 3.000               | 6598.5                    | 715.0                | 1015.                 | 177.9           | 8.000                |
| 5.000               | 6126.3                    | 725.0                | 1025.                 | 176.7           | 8.000                |
| 7.000               | 5831.4                    | 735.0                | 1035.                 | 176.4           | 8.000                |
| 9.000               | 5622.7                    | 745.0                | 1045.                 | 176.6           | 8.000                |
| 11.000              | 5464.9                    | 755.0                | 1055.                 | 177.1           | 8.000                |
| 13.000              | 5340.6                    | 765.0                | 1065.                 | 177.9           | 8.000                |
| 15.000              | 5240.2                    | 775.0                | 1075.                 | 178.8           | 8.000                |
| 17.000              | 5157.5                    | 785.0                | 1085.                 | 179.8           | 8.000                |
| 19.000              | 5088.5                    | 795.0                | 1095.                 | 180.8           | 8.000                |
| 21.000              | 5035.6                    | 804.0                | 1104.                 | 181.9           | 8.000                |
| 23.000              | 4994.1                    | 812.0                | 1112.                 | 182.8           | 8.000                |
| 25.000              | 4957.4                    | 820.0                | 1120.                 | 183.8           | 8.000                |
| 27.000              | 4925.1                    | 828.0                | 1128.                 | 184.8           | 8.000                |
| 29.000              | 4896.7                    | 836.0                | 1136.                 | 185.8           | 8.000                |
| 31.000              | 4871.8                    | 844.0                | 1144.                 | 186.9           | 8.000                |
| 33.000              | 4850.0                    | 852.0                | 1152.                 | 188.0           | 8.000                |
| 35.000              | 4831.0                    | 860.0                | 1160.                 | 189.1           | 8.000                |
| 37.000              | 4814.4                    | 868.0                | 1168.                 | 190.2           | 8.000                |
| 39.000              | 4800.1                    | 876.0                | 1176.                 | 191.3           | 8.000                |
| 41.000              | 4787.9                    | 884.0                | 1184.                 | 192.5           | 8.000                |
| 43.000              | 4777.5                    | 892.0                | 1192.                 | 193.6           | 8.000                |

**EAO vs Minimum Temperature Approach**

Print   Return to Previous Screen   Return to Main Menu

EAO - Minimum Temperature Approach Plot



15.20

| $h_o$<br>BTU/hr/ft <sup>2</sup> °F | $F_M$ | $h_{adj}/h_o$<br>Fig 15.16 | $h_{adj}$<br>BTU/hr/ft <sup>2</sup> °F |
|------------------------------------|-------|----------------------------|----------------------------------------|
| 75                                 | 1.00  | 1.00                       | 75.00                                  |
| 25                                 | 2.00  | 0.45                       | 11.25                                  |
| 10                                 | 3.00  | 0.28                       | 2.80                                   |
| 45                                 | 7.00  | 0.09                       | 3.83                                   |
| 30                                 | 2.50  | 0.35                       | 10.50                                  |
| 30                                 | 1.00  | 1.00                       | 30.00                                  |

Using the adjusted heat transfer coefficients from the above table we get the following results.

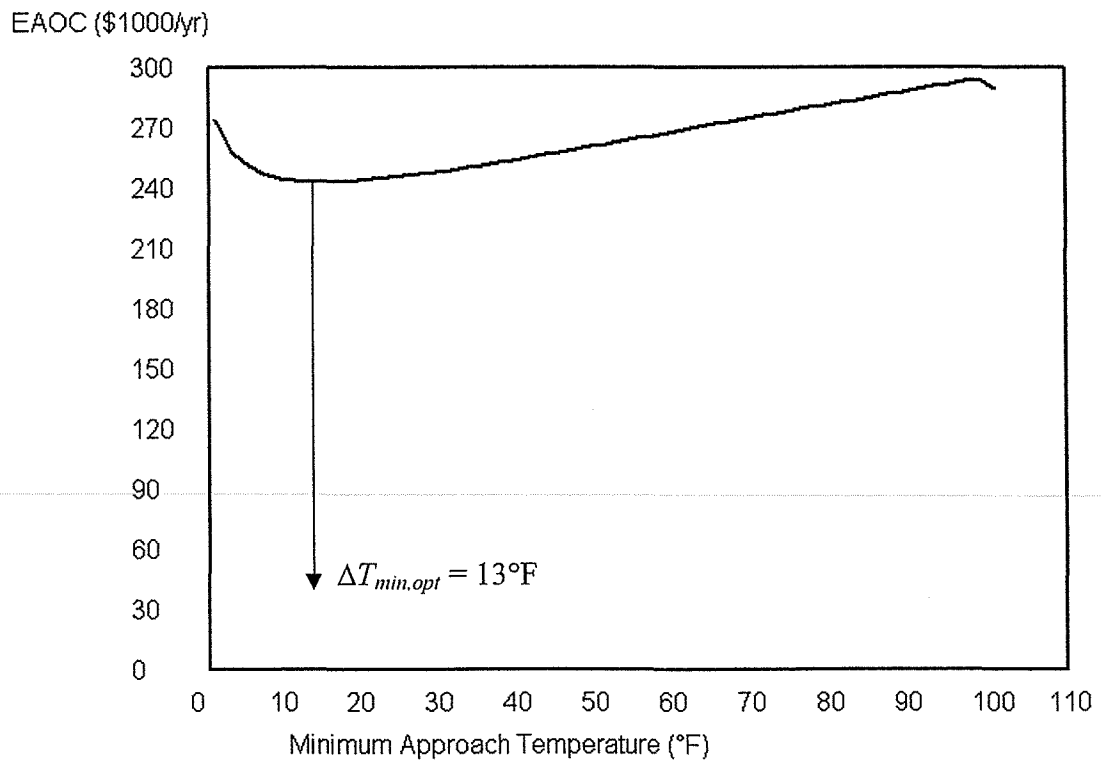
The adjusted heat transfer area for the network is 5765 ft<sup>2</sup>

The EAO and minimum approach temp is shown below:

#### EAOC vs Minimum Temperature Approach

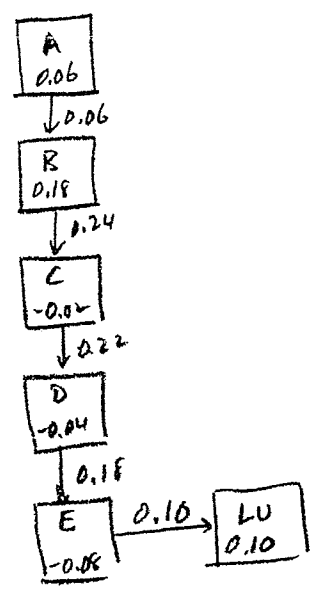
Print Return to Previous Screen Return to Main Menu

EAOC - Minimum Temperature Approach Plot

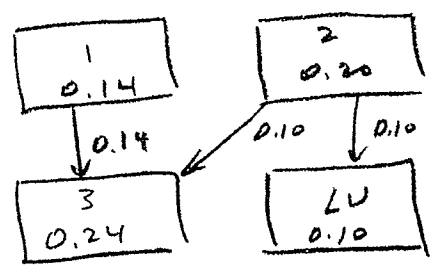


15.21 stream 1 2 3  
 m<sup>3</sup>/s 2 4 4

|      |  |   |  |      | $\sum m(y-x)$ |
|------|--|---|--|------|---------------|
| 0.10 |  |   |  | 0.18 | 0.06          |
|      |  | A |  |      |               |
| 0.07 |  |   |  | 0.12 | 0.18          |
|      |  | B |  |      |               |
| 0.04 |  |   |  | 0.06 | -0.02         |
|      |  | C |  |      |               |
| 0.03 |  |   |  | 0.04 | -0.04         |
|      |  | D |  |      |               |
| 0.02 |  |   |  | 0.02 | -0.08         |
|      |  | E |  |      |               |
| 0.01 |  |   |  | 0.00 | -0.10         |

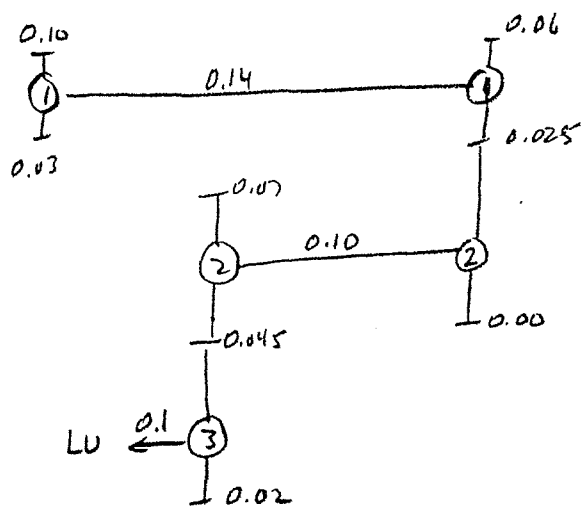


there is no pinch



3 exchangers

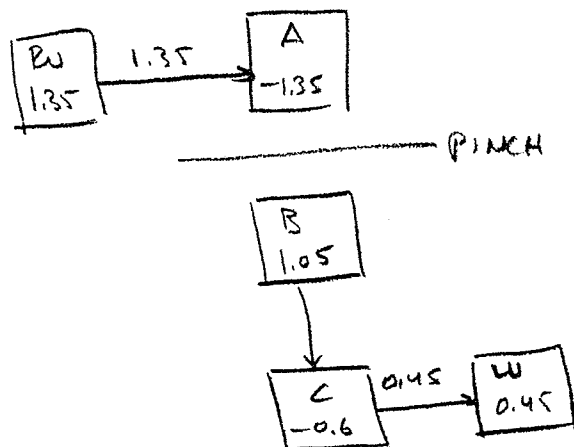
|   |   |   |   |
|---|---|---|---|
|   | 1 | 2 | 3 |
| M | 2 | 4 | 4 |



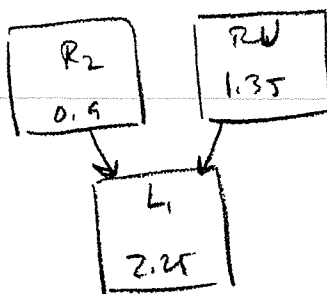


15.22

| stream          | $R_1$ | $R_2$ | $L_1$ | $\Sigma m(b+r)$ |
|-----------------|-------|-------|-------|-----------------|
| $m \frac{b}{h}$ | 60    | 30    | 90    |                 |
| 0.12            |       | 0.9   | 2.25  | 0.09<br>-1.35   |
| 0.09            | 4.2   | 2.1   | 5.25  | 0.065<br>1.05   |
| 0.02            |       |       | 0.6   | 0.00667<br>-0.6 |
| 0.01            |       |       |       | 0<br>-0.9       |

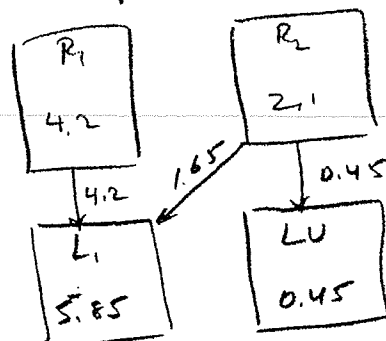


above pinch



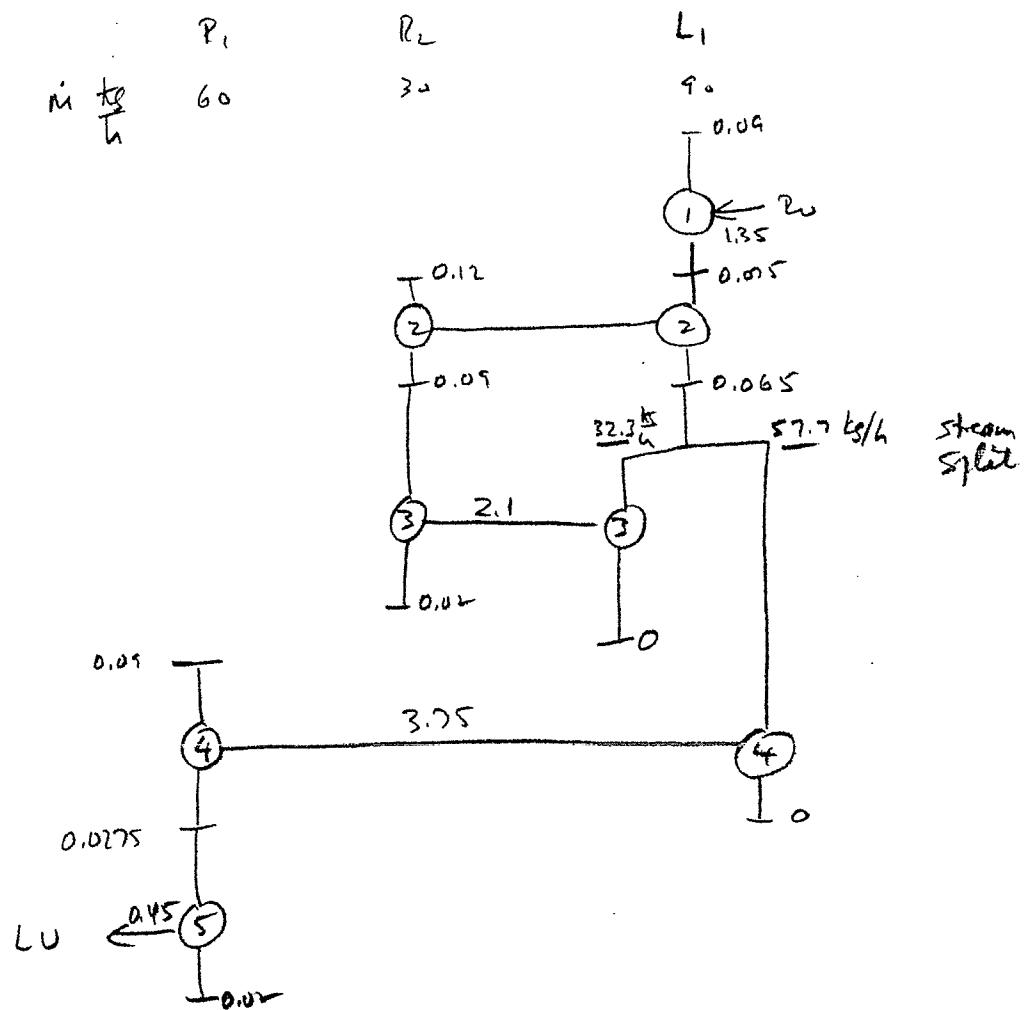
2 exchangers

below pinch



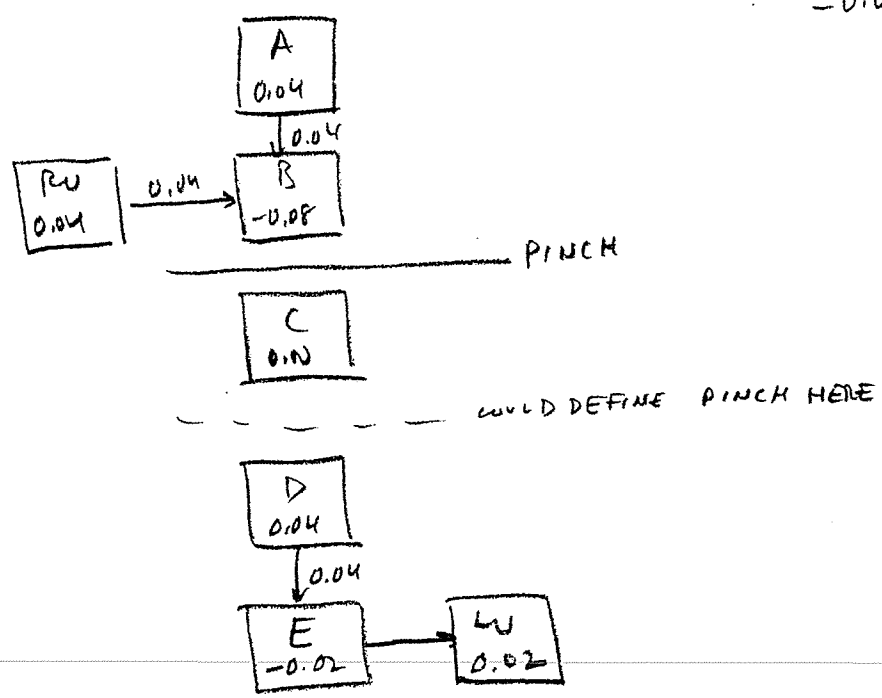
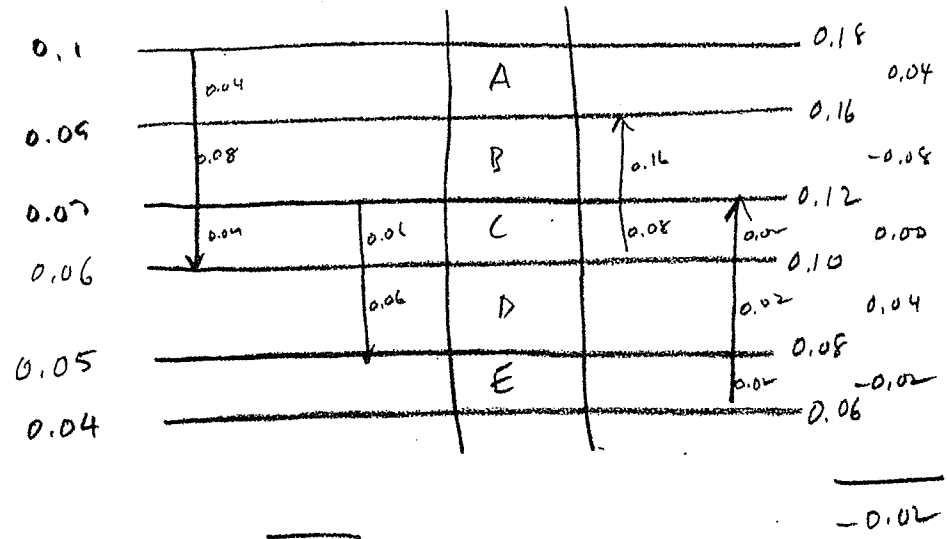
3 exchangers

5 total exchangers

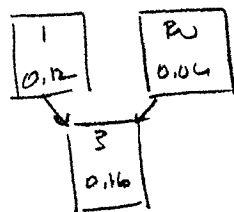


15.23 stream  
 in  $\frac{kg}{s}$

|  | 1 | 2 | 3 | 4 |              |
|--|---|---|---|---|--------------|
|  | 4 | 6 | 4 | 1 | Σ in (17-22) |

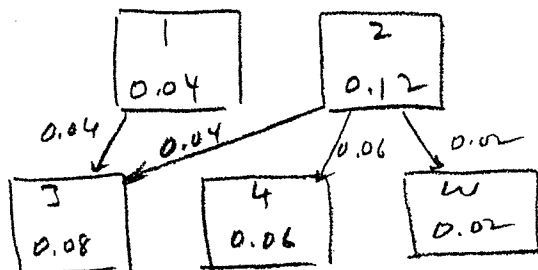


above  
pinch



2 exchangers

below pinch



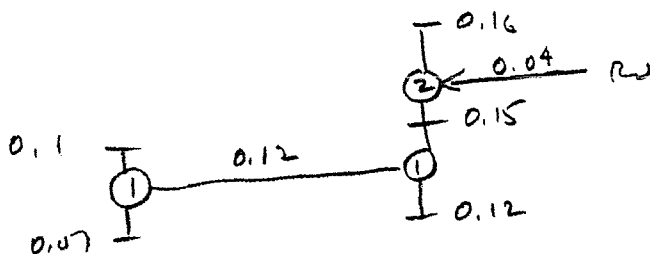
4 exchangers

6 total exchangers

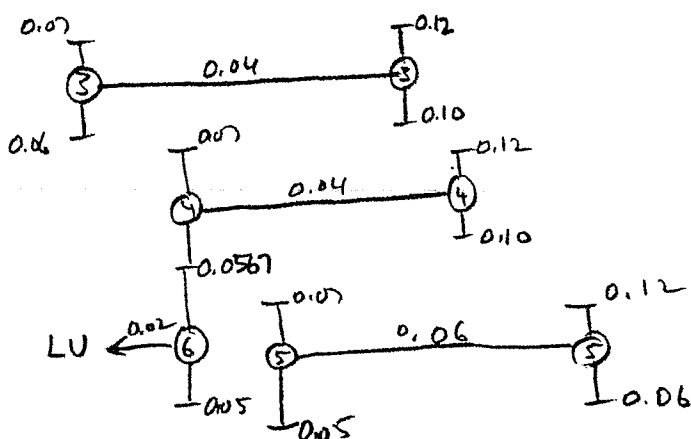
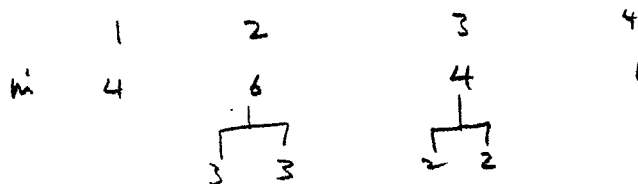
above  
pinch

$P_1 = 1$   
 $M_1 = 4$

$L_1 = 3$   
 $4$

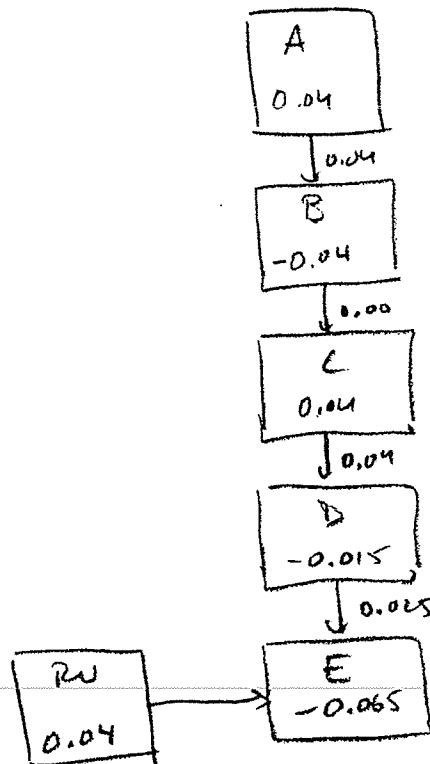


below  
pinch

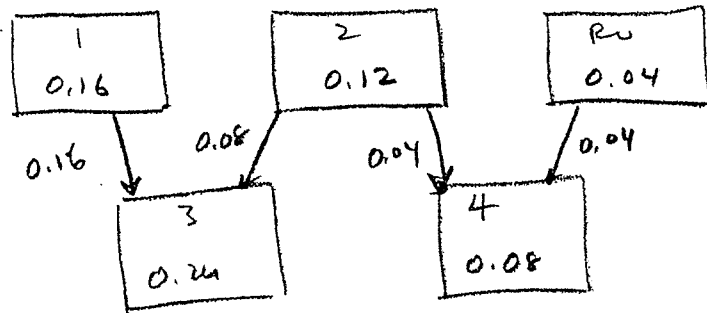


15.24

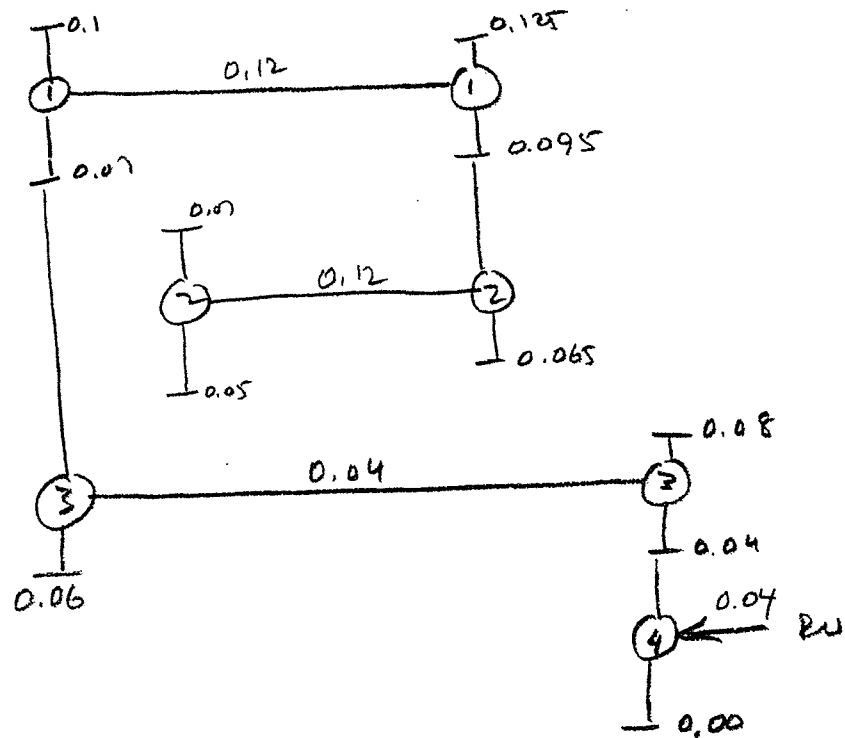
|        | 1    | 2    | 3 | 4     |                               |
|--------|------|------|---|-------|-------------------------------|
| $m$    | 4    | 6    | 4 | 1     | $\sum m(\Delta y - \Delta x)$ |
| 0.10   | 0.04 |      | A |       | 0.14                          |
| 0.09   | 0.04 |      | B | 0.12  | 0.125                         |
| 0.07   | 0.04 | 0.04 | C | 0.06  | 0.095                         |
| 0.06   |      | 0.06 | D | 0.06  | 0.08                          |
| 0.05   |      |      | E | 0.065 | 0.065                         |
| 0.0067 |      |      |   |       | 0                             |
|        |      |      |   |       | -0.04                         |



← not really a pinch point since  
there is only one utility, RW  
true pinch point has RW above  
and CW below

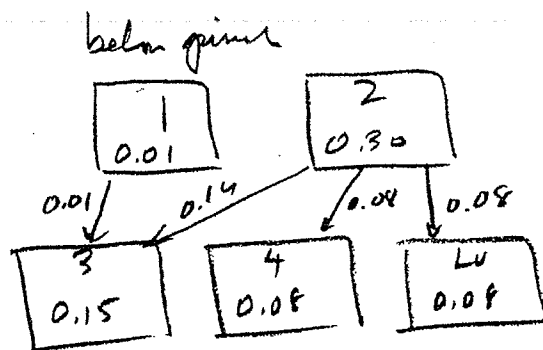
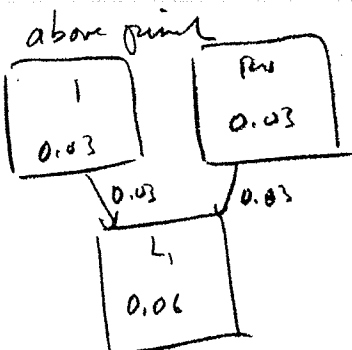
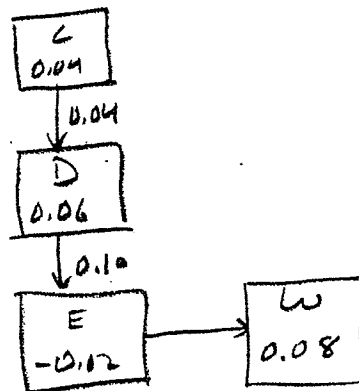
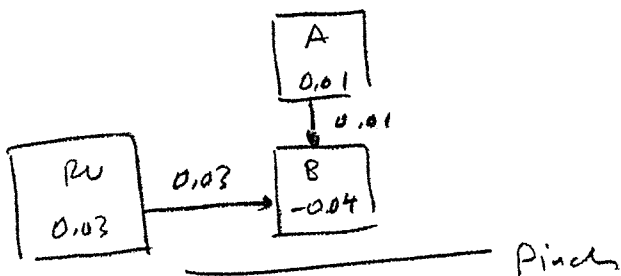


|       |   |   |   |   |
|-------|---|---|---|---|
|       | 1 | 2 | 3 | 4 |
| $m_i$ | 4 | 6 | 4 | 1 |

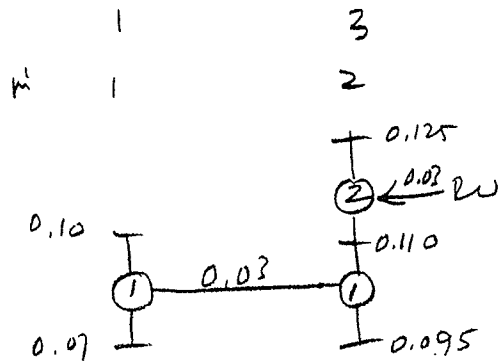


15.25

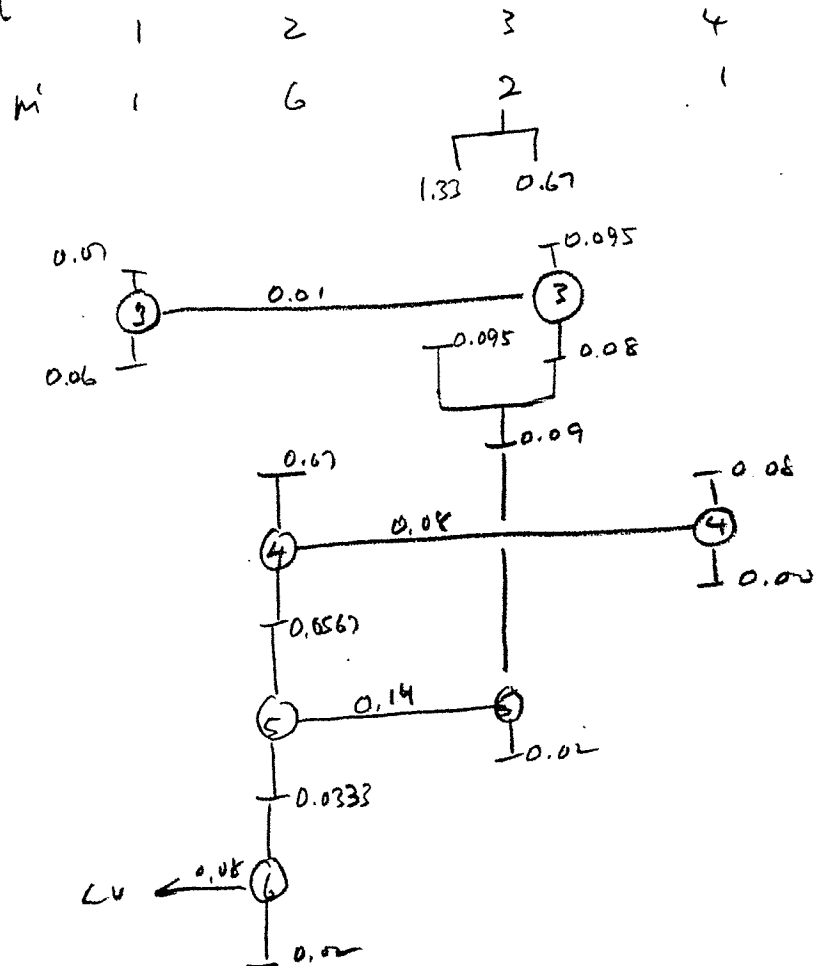
|        | 1    | 2     | 3    | 4     |                             |
|--------|------|-------|------|-------|-----------------------------|
| M      | 1    | 6     | 2    | 1     | $\sum \Delta T (A_H - A_C)$ |
| 0.10   |      |       |      | 0.14  | 0.01                        |
| 0.09   | 0.01 |       |      | 0.125 | -0.04                       |
| 0.07   | 0.02 |       | 0.06 | 0.095 | 0.04                        |
| 0.06   | 0.01 | 0.06  | 0.03 | 0.08  | 0.06                        |
| 0.05   |      | 0.024 | 0.12 | 0.07  | -0.02                       |
| 0.0067 |      |       |      | 0.05  | +0.05                       |



above ground



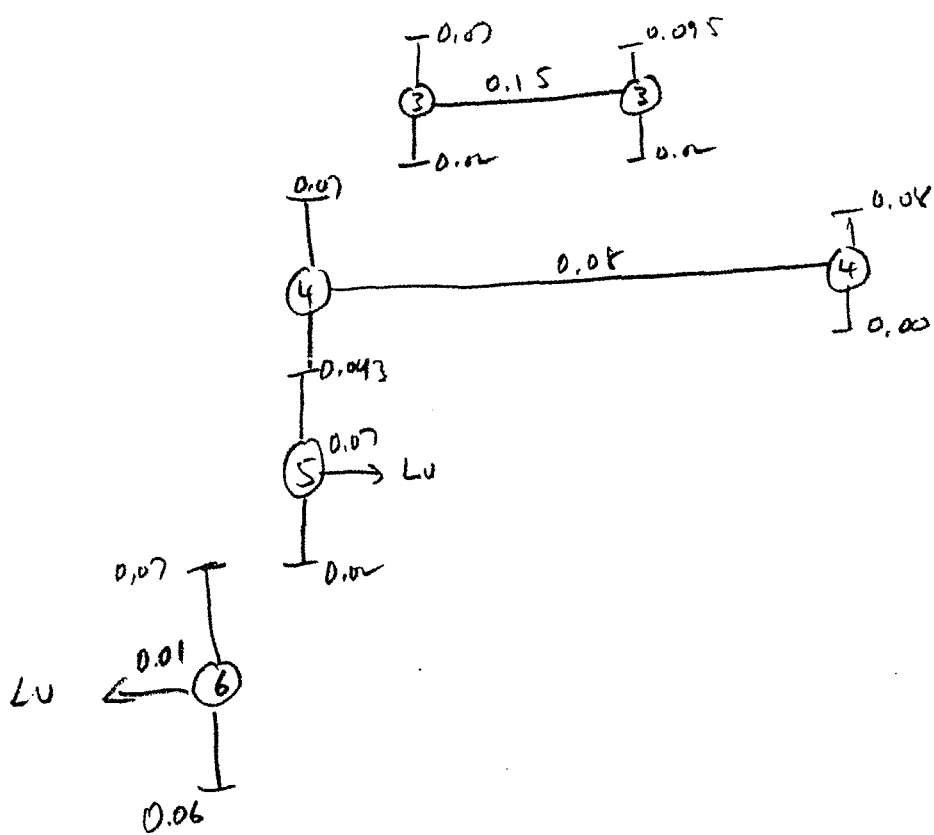
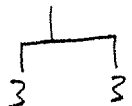
below ground





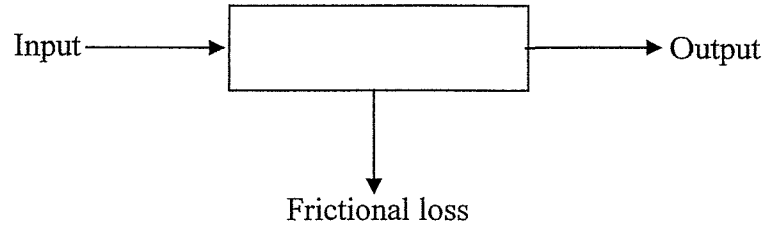
alternative below prime

|       |   |   |   |   |
|-------|---|---|---|---|
|       | 1 | 2 | 3 | 4 |
| $n_i$ | 1 | 6 | 2 | 1 |

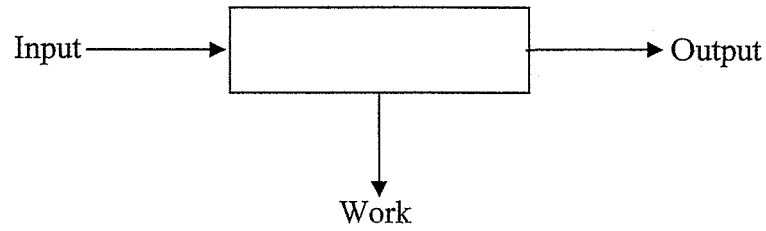


## Chapter 16

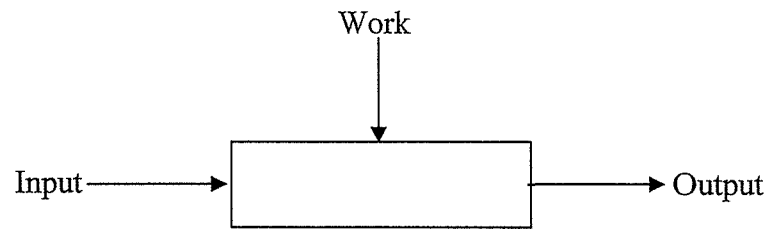
### 16.1 (a) Fluidized Bed



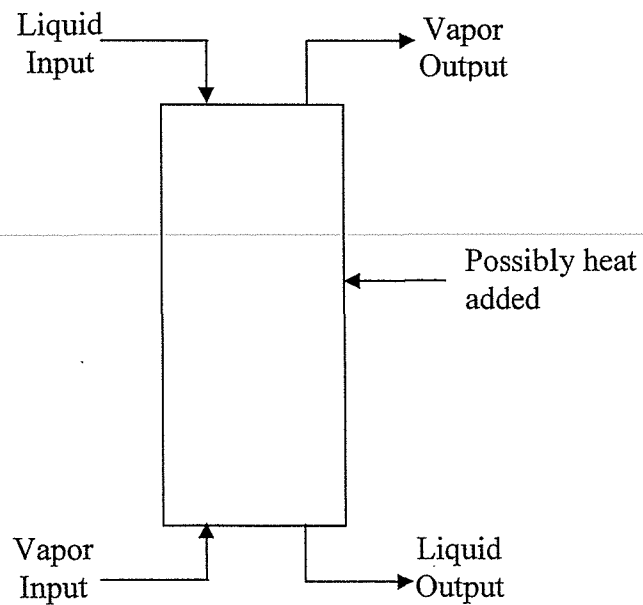
### (b) Turbine



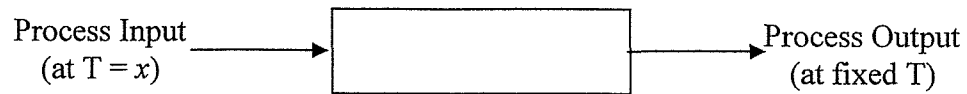
### (c) Pump



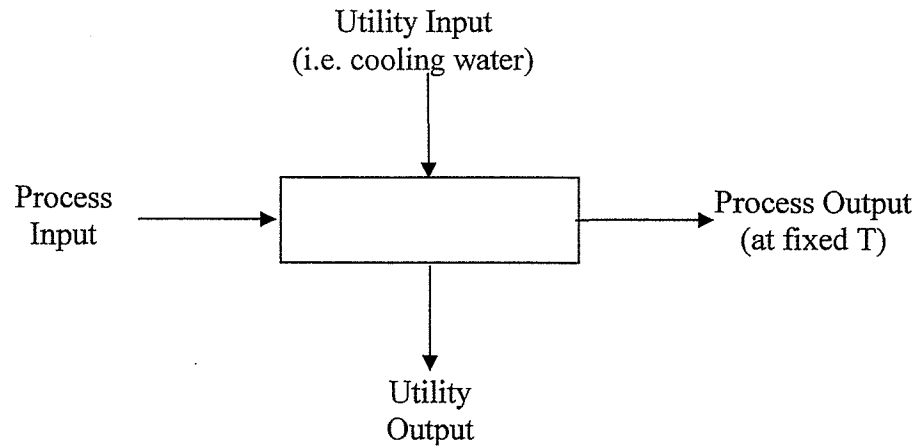
### (d) Stripper



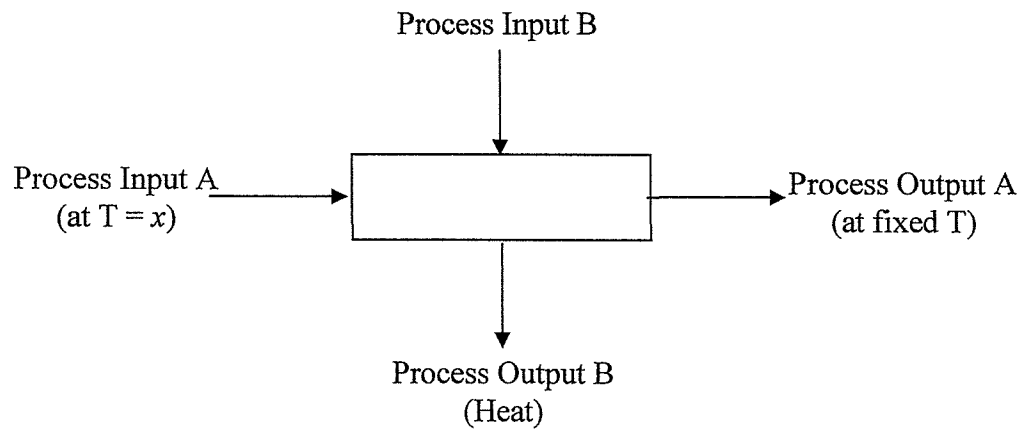
(e) Adiabatic Batch Reactor



(f) Semibatch Reactor with Heat Removal

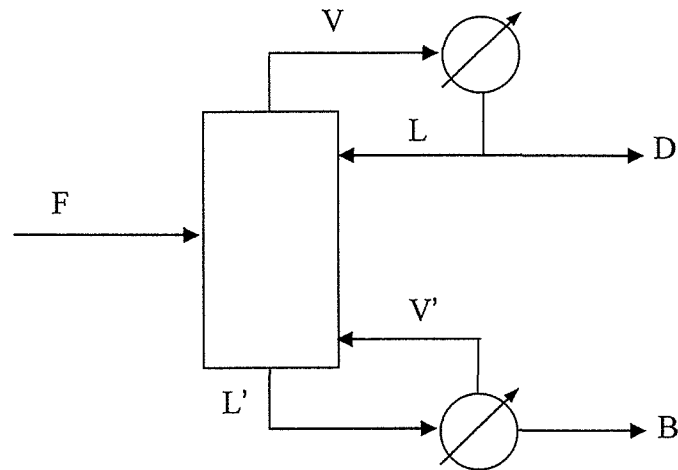


(g) Heat Exchanger between Two Process Streams (no utilities)



- 16.2 (a) Fluidized Bed  
Equipment Independent: mass and energy (if needed) balance  
Equipment Dependent: pressure-drop equation for fluidized bed
- (b) Turbine  
Equipment Independent: energy balance  
Equipment Dependent: none
- (c) Pump  
Equipment Independent: energy balance  
Equipment Dependent: pump curve
- (d) Stripper  
Equipment Independent: mass and energy balance  
Equipment Dependent: flooding relationship
- (e) Adiabatic Batch Reactor  
Equipment Independent: mass and energy balance  
Equipment Dependent: design equations
- (f) Semibatch Reactor with Heat Removal  
Equipment Independent: mass and energy balance  
Equipment Dependent: design equations
- (g) Heat Exchanger between Two Process Streams (no utilities)  
Equipment Independent:  $Q = \dot{m}C_p\Delta T$  without phase change  
 $Q = \dot{m}\lambda$  with phase change  
Equipment Dependent:  $Q = UA\Delta T_{lm}F$

## 16.3 Distillation Column



A = volatile component  
(Assume binary)

Equipment Independent:  $\left[ \begin{array}{l} FZ_{AF} = Dx_{AD} + Bx_{AB} \\ FH_F + Q_R = DH_D + BH_B + Q_C \end{array} \right]$  Column

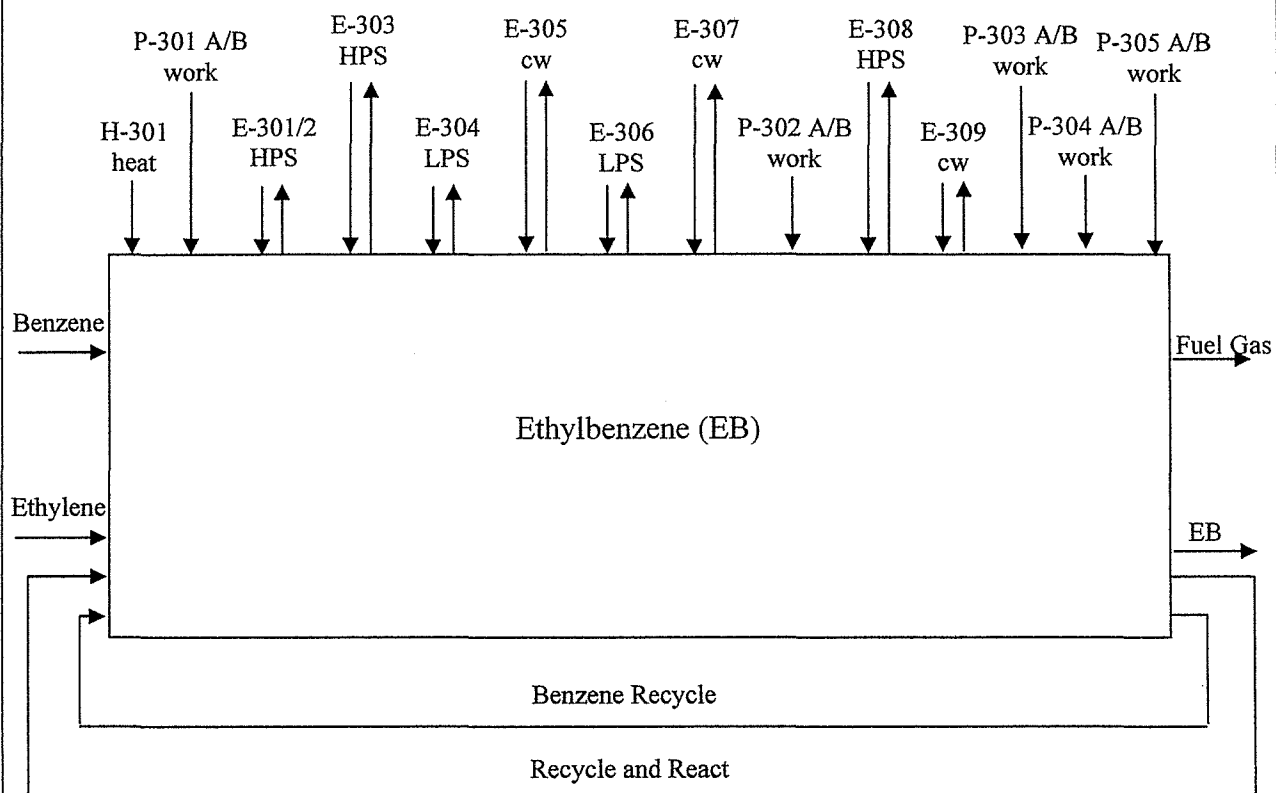
$\left[ \begin{array}{l} V = L + D \\ Q = (\dot{m}C_p\Delta T)_{cw} = (\dot{m}\lambda)_{process} \end{array} \right]$  Condenser

$\left[ \begin{array}{l} L' = B + V' \\ Q = (\dot{m}\lambda)_{steam} = (\dot{m}\lambda)_{process} \end{array} \right]$  Reboiler

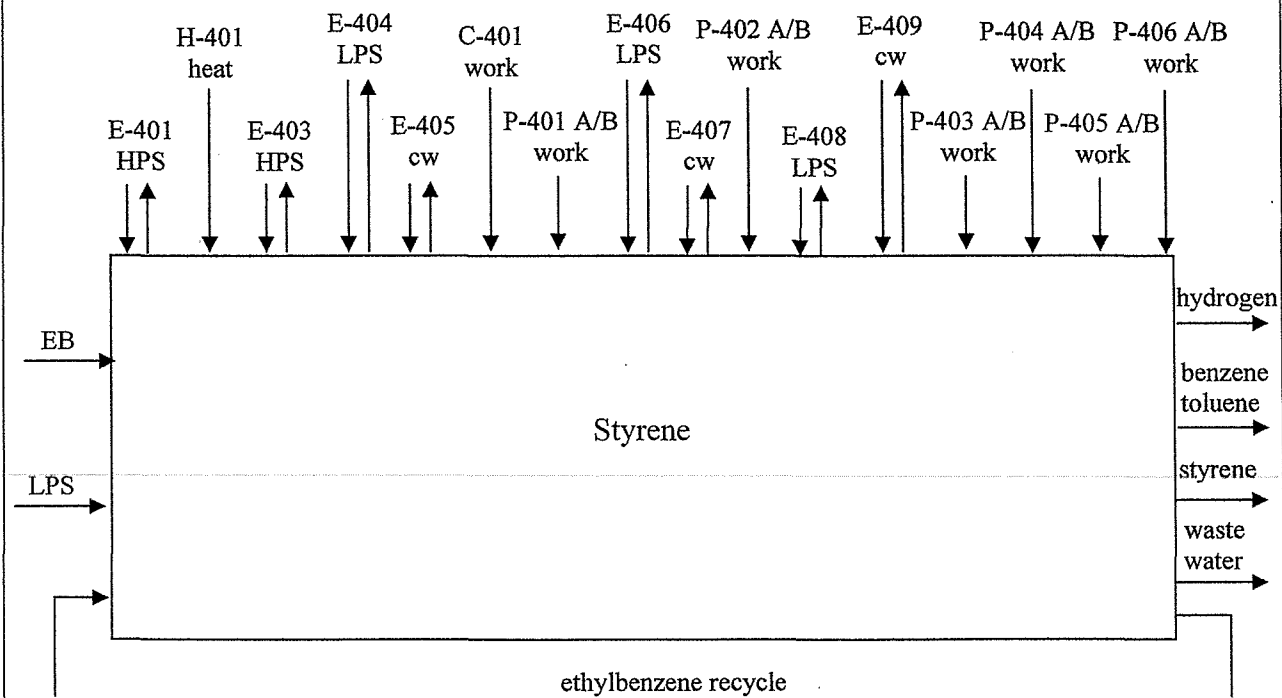
Equipment Dependent:  $\left[ \begin{array}{l} \Delta P \text{ relationship} \\ \text{flooding relationship} \end{array} \right]$  Column

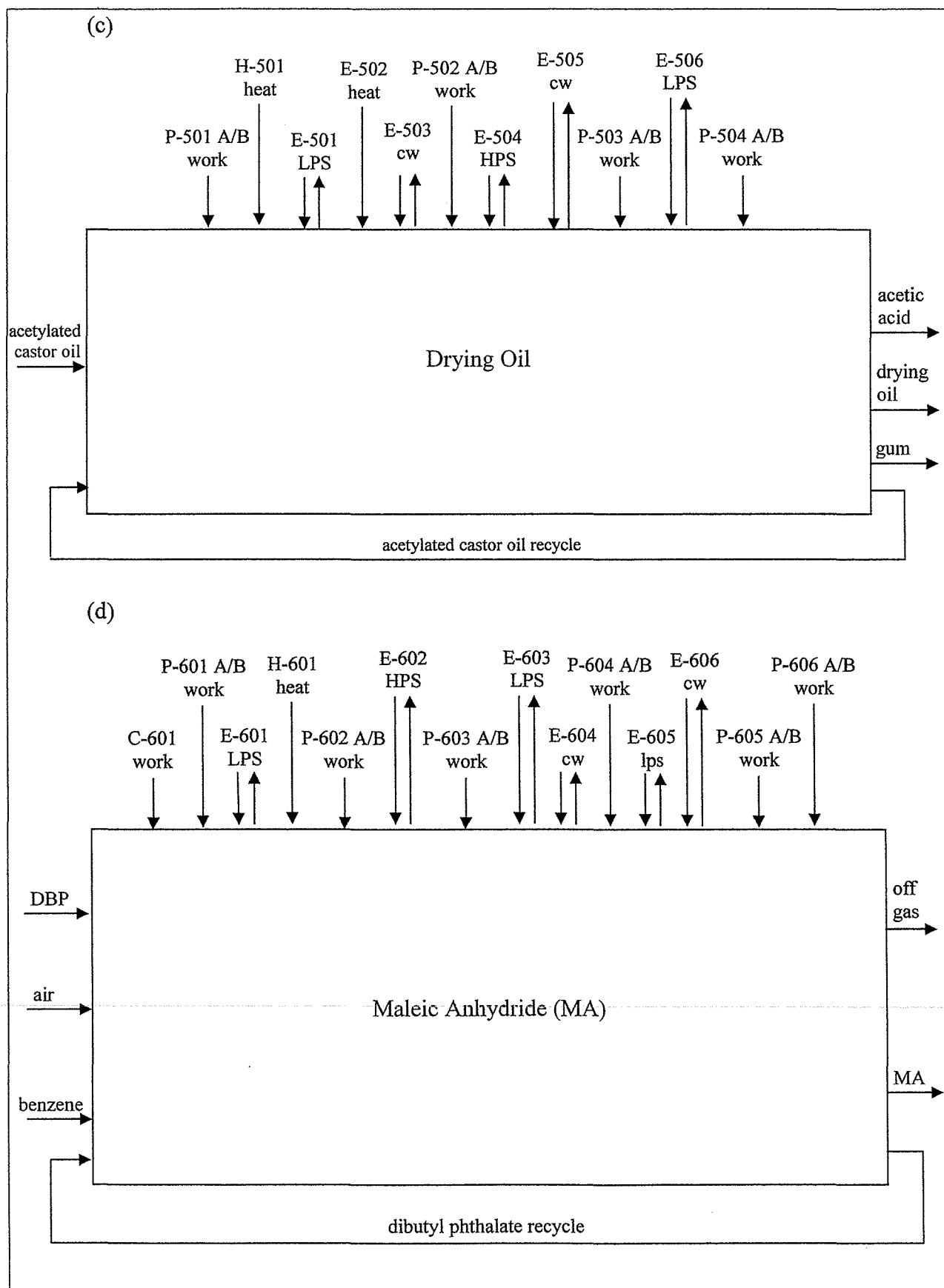
$[Q = UA\Delta T_{LM}]$  Condenser and Reboiler

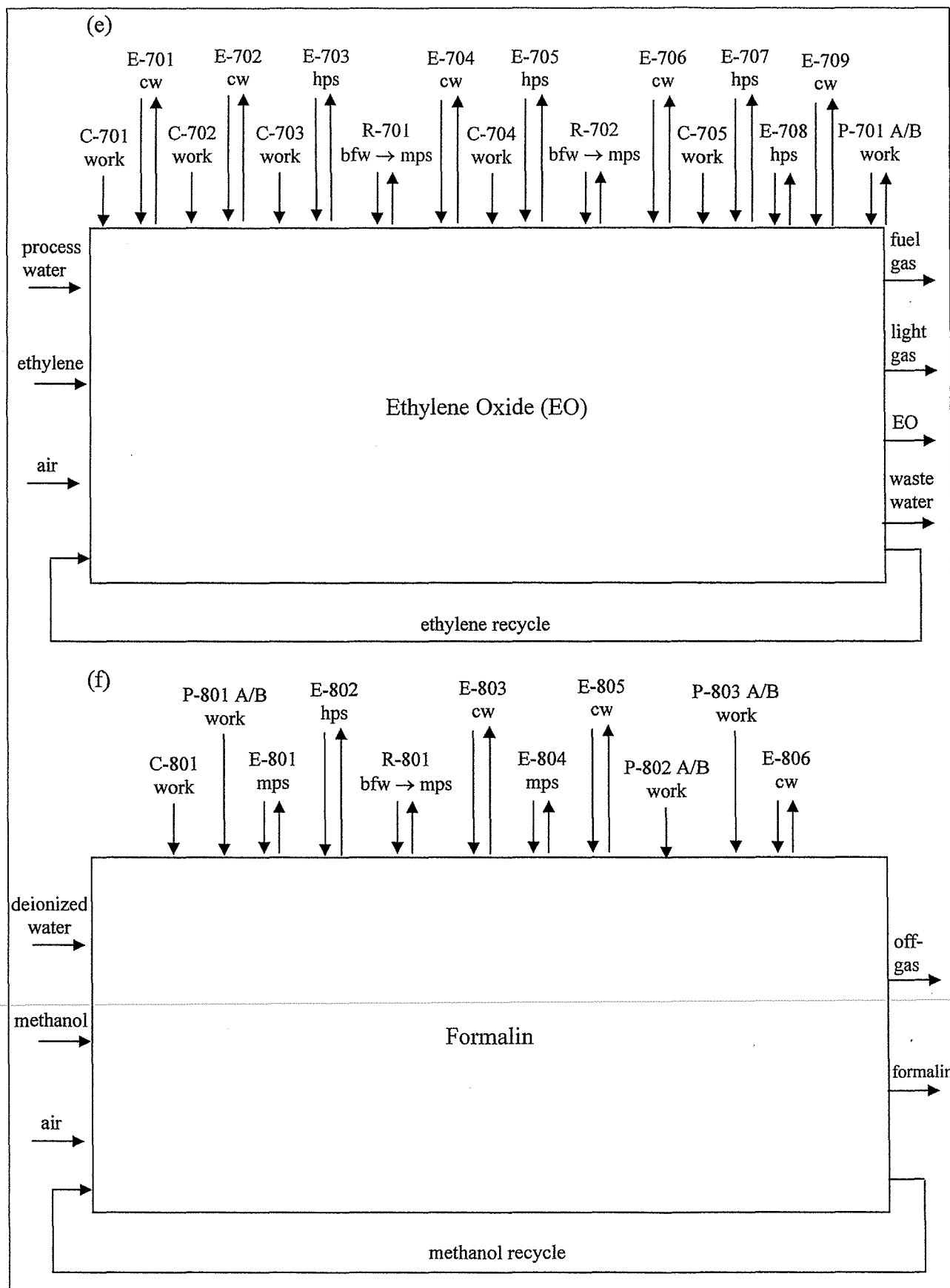
16.4 (a)



(b)









## Chapter 17

17.1 Second order reaction:  $-r = kC_A^2$ ;  $k = Ae^{-E/RT}$

If ideal  $C_i = \frac{P_i}{RT}$  if ideal

$$220^\circ\text{C} = 493\text{K}$$

$$250^\circ\text{C} = 523\text{K}$$

$$\frac{r_2}{r_1} = \frac{A_2}{A_1} e^{-\frac{E}{R}\left(\frac{1}{T_2} - \frac{1}{T_1}\right)} \left(\frac{C_{A2}^2}{C_{A1}^2}\right)$$

$$2 = e^{-\frac{E}{R}\left(\frac{1}{523\text{K}} - \frac{1}{493\text{K}}\right)}$$

$$\frac{E}{R} = 5957.35 \text{ K}$$

$$(a) \quad \frac{r_2}{r_1} = \frac{k_2}{k_1} \left(\frac{C_{A2}^2}{C_{A1}^2}\right)^2 = \left(\frac{A_2}{A_1}\right) e^{-5957.35\text{K}\left(\frac{1}{518\text{K}} - \frac{1}{498\text{K}}\right)} \left(\frac{(P_A/RT)_2^2}{(P_A/RT)_1^2}\right)$$

$$\frac{r_2}{r_1} = 1.587 \left(\frac{P_{A2}}{P_{A1}}\right)^2 \left(\frac{T_1}{T_2}\right)^2 = 1.587 \left(\frac{498\text{K}}{518\text{K}}\right)^2 = 1.47$$

$$\boxed{\frac{r_2}{r_1} = 1.47}$$

$$(b) \quad \frac{r_2}{r_1} = \frac{k_2}{k_1} \left(\frac{C_{A2}^2}{C_{A1}^2}\right)^2 = \left(\frac{P_2}{P_1}\right)^2 = 1.1^2 = 1.21$$

$$\boxed{\frac{r_2}{r_1} = 1.21}$$

$$17.2 \quad P_{\text{tank}} = 3 \text{ atm} = P_1; \quad P_2 = 5 \text{ atm}$$

$$\dot{v}_1 = 12.5 \frac{\text{m}^3}{\text{h}}$$

Assume turbulent flow

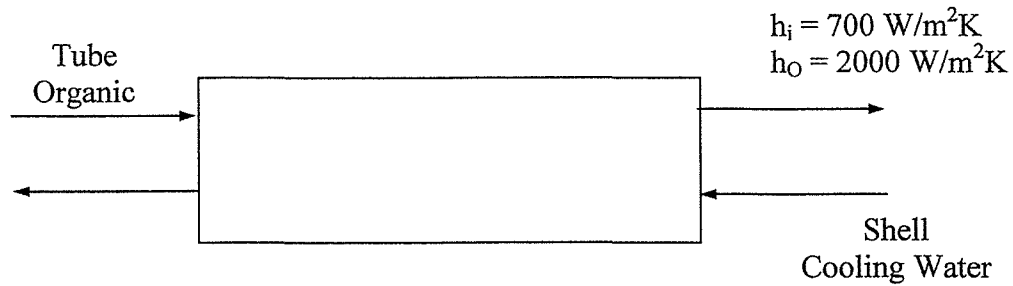
$$\Delta P = \frac{2\rho f L u^2}{D} = \frac{32\rho f L \dot{v}^2}{\pi^2 D^5}$$

$$\frac{\Delta P_2}{\Delta P_1} = \frac{\dot{v}_2^2}{\dot{v}_1^2}$$

$$\frac{5}{3} = \left( \frac{\dot{v}_2^2}{12.5 \text{ [m}^3/\text{h}]} \right)^2$$

$$\dot{v}_2 = 16.1 \frac{\text{m}^3}{\text{h}}$$

17.3



1.25" 16 BWG

 $D_o = 31.75 \text{ mm}$  $R_o = 15.875 \text{ mm}$  $D_i = 28.45 \text{ mm}$  $R_i = 14.225 \text{ mm}$ 

$$\frac{\ln\left(\frac{R_o}{R_i}\right)}{k} = \frac{\ln\left(\frac{15.875 [\text{mm}]}{14.225 [\text{mm}]}\right)}{45 \left[\frac{\text{W}}{\text{mK}}\right]} = 0.002438$$

$$\frac{1}{U_o} = \frac{1}{h_o} + \frac{\ln\left(\frac{R_o}{R_i}\right)}{k} + \frac{R_o}{R_i} \frac{1}{h_i}$$

Change in  $h_i$ :  $\frac{h_{o2}}{h_{o1}} = \left(\frac{v_2}{v_1}\right)^{0.6}$

$$h_{o2} = 2000 \text{ W/m}^2\text{K} (1.15)^{0.6} = 2174.95 \text{ W/m}^2\text{K}$$

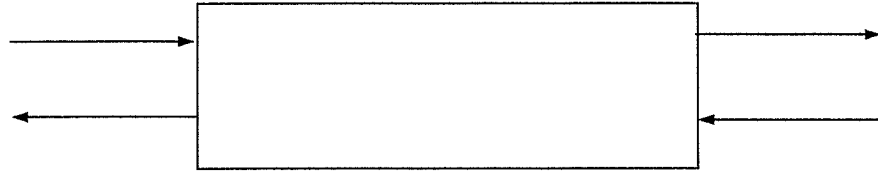
$$\frac{\frac{1}{U_{i1}}}{\frac{1}{U_{i2}}} = \frac{U_{o2}}{U_{o1}} = \frac{\frac{1}{2000} + 0.002438 + \frac{15.875}{14.225} \frac{1}{700}}{\frac{1}{2174.95} + 0.002438 + \frac{15.875}{14.225} \frac{1}{700}}$$

$$\boxed{\frac{U_{o2}}{U_{o1}} = 1.009} \sim 0.9\% \text{ change}$$

17.4

$$h_i = 1000 \text{ W/m}^2\text{K}$$

$$h_o = 500 \text{ W/m}^2\text{K}$$



$$U = 300 \text{ W/m}^2\text{K}$$

Assume thin walls:  $\frac{1}{U_o} = \frac{1}{h_o} + R_{\text{wall \& fouling}} + \frac{1}{h_i}$

$$R_{\text{wall \& fouling}} = \frac{1}{300 \text{ W/m}^2\text{K}} - \frac{1}{1000 \text{ W/m}^2\text{K}} - \frac{1}{500 \text{ W/m}^2\text{K}} = 0.00035 \text{ m}^2\text{K/W}$$

(assumed unchanged)

$$\frac{\frac{1}{U_{o1}}}{\frac{1}{U_{o2}}} = \frac{U_{o2}}{U_{o1}} = \frac{\frac{1}{h_{o1}} + 0.00035 + \frac{1}{h_{i1}}}{\frac{1}{h_{o2}} + 0.00035 + \frac{1}{h_{i2}}}$$

(a)  $h_{o2} = 500(0.8)^{0.6} = 437.34 \text{ W/m}^2\text{K}$

$$\frac{1}{U_{o2}} = \frac{1}{437.34 \text{ W/m}^2\text{K}} + 0.00035 \text{ m}^2\text{K/W} + \frac{1}{1000 \text{ W/m}^2\text{K}}$$

$$U_{o2} = 276.25 \text{ W/m}^2\text{K}$$

(b)  $h_{i2} = 1000(0.8)^{0.8} = 836.5 \text{ W/m}^2\text{K}$

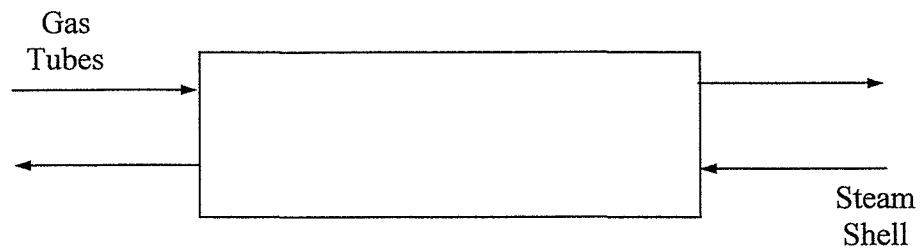
$$\frac{1}{U_{o2}} = \frac{1}{500 \text{ W/m}^2\text{K}} + 0.00035 \text{ m}^2\text{K/W} + \frac{1}{836.5 \text{ W/m}^2\text{K}}$$

$$U_{o2} = 283.38 \text{ W/m}^2\text{K}$$

(c)  $\frac{1}{U_{o2}} = \frac{1}{437.34 \text{ W/m}^2\text{K}} + 0.00035 \text{ m}^2\text{K/W} + \frac{1}{836.5 \text{ W/m}^2\text{K}}$

$$U_{o2} = 262.1 \text{ W/m}^2\text{K}$$

17.5



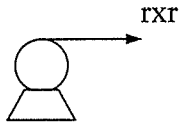
$$\frac{1}{U_o} \approx \frac{1}{h_o} + \frac{1}{h_i}; h_i \propto U^{0.8}$$

$h_o \gg h_i$  assumed since  $h_o$  is condensing and  $h_i$  is process gas

$$\frac{\frac{1}{U_{o1}}}{\frac{1}{U_{o2}}} = \frac{U_{o2}}{U_{o1}} \approx \frac{\cancel{\frac{1}{h_{o1}}} + \frac{1}{h_{i1}}}{\cancel{\frac{1}{h_{o2}}} + \frac{1}{h_{i2}}}$$

$$\frac{U_{o2}}{U_{o1}} = \frac{h_{i2}}{h_{i1}} = \left( \frac{v_{i2}}{v_{i1}} \right)^{0.8}$$

17.6



$$\Delta P = \frac{2\rho f L u^2}{D} = \frac{32 f L \dot{m}^2}{\rho \pi^2 D^5}$$

$$2.5'' \text{ sch } 40 \ D_{i1} = 62.71 \text{ mm}$$

$$3'' \text{ sch } 40 \ D_{i2} = 77.92 \text{ mm}$$

$$\frac{\Delta P_2}{\Delta P_1} = \frac{D_1^5}{D_2^5} = \left( \frac{77.92}{62.71} \right)^5 = 2.96$$

$$\frac{\Delta P_2 - \Delta P_1}{\Delta P_1} = \frac{2.96 - 1}{1} = 1.96$$

$$\dot{W} = \dot{m} \frac{\Delta P}{\rho}$$

$$\frac{W_2}{W_1} = \frac{\cancel{\dot{m}_2}}{\cancel{\dot{m}_1}} \frac{\Delta P_2}{\Delta P_1} \frac{\cancel{\rho_1}}{\cancel{\rho_2}} = 1.96$$

$$\frac{\$}{\text{h}} = (\text{elec. cost})(\dot{W}) \Rightarrow \boxed{\approx \text{double elec. Cost}}$$

$$17.7 \quad \dot{v} = \frac{\pi \Delta P R^4}{8 \mu L}$$

$$(a) \quad \dot{v} \propto \Delta P \quad \uparrow 3x$$

$$(b) \quad \frac{\dot{v}_2}{\dot{v}_1} = \left( \frac{D_2}{D_1} \right)^4 = \left( \frac{52.50}{35.05} \right)^4 = 5.03 \quad \uparrow$$

$$(c) \quad \dot{v} \propto \frac{1}{\mu} \Rightarrow \frac{\dot{v}_2}{\dot{v}_1} = \frac{\mu_1}{\mu_2} = \frac{1}{1.5} = 0.667 \quad \downarrow$$

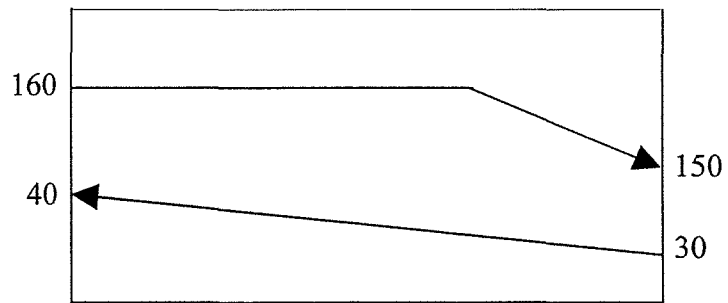
$$(d) \quad \dot{v} \propto \frac{1}{L} \Rightarrow \frac{\dot{v}_2}{\dot{v}_1} = \frac{L_1}{L_2} = \frac{1}{0.1} = 10 \quad \uparrow$$

$$(e) \quad \dot{v} \propto \Delta P R^4 \Rightarrow \frac{\dot{v}_2}{\dot{v}_1} = \frac{\Delta P_2}{\Delta P_1} \left( \frac{D_2}{D_1} \right)^4 = 3(5.03) = 15.1$$

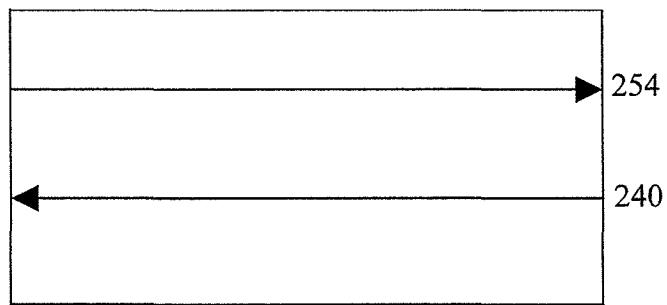
$$(f) \quad \dot{v} \propto \frac{\Delta P}{\mu} \Rightarrow \frac{\dot{v}_2}{\dot{v}_1} = \frac{\Delta P_2}{\Delta P_1} \frac{\mu_1}{\mu_2} = 3(0.667) = 2$$

$$(g) \quad \dot{v} \propto \frac{D^4}{L} = \left( \frac{D_2}{D_1} \right)^4 \left( \frac{L_1}{L_2} \right) = 5.03(10) = 50.3$$

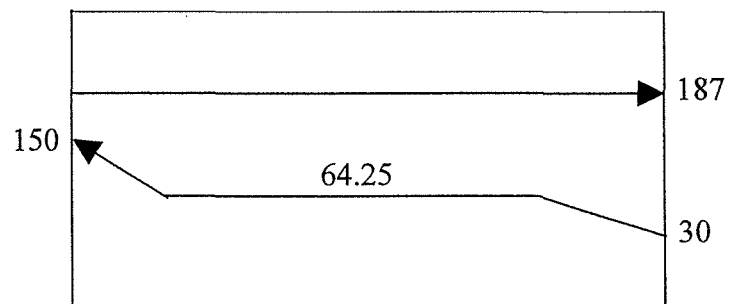
17.8 (a)



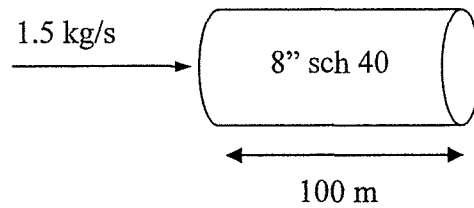
(b)



(c)







$$D_i = 7.981 \text{ in} = 202.7 \text{ mm}$$

$$A = 322.7 \times 10^{-4} \text{ m}^2$$

$$\text{At STP } \rho_{air} = \frac{\text{kmol}}{22.4 \text{ m}^3} \frac{29 \text{ kg}}{\text{kmol}} = 1.295 \frac{\text{kg}}{\text{m}^3}$$

$$\Delta P = \frac{2fL\rho u^2}{D}$$

$$\dot{m} = 1.5 \text{ kg/s} = \rho A u = \left(1.295 \frac{\text{kg}}{\text{m}^3}\right) (0.03227 \text{ m}^2) u$$

$$u = 35.9 \frac{\text{m}}{\text{s}}$$

$$\text{Re} = \frac{D u \rho}{\mu} = \frac{(0.2027 \text{ [m]}) (35.9 \text{ [m/s]}) (1.295 \text{ [kg/m}^3\text{]})}{1.72 \times 10^{-5} \text{ [kg/ms]}} = 547,900$$

$$\varepsilon = 4.6 \times 10^{-5} \text{ m for commercial steel}$$

$$\frac{\varepsilon}{D} = \frac{4.6 \times 10^{-5} \text{ m}}{0.2027 \text{ m}} = 0.00023$$

$$f = 0.00391$$

$$\Delta P = \frac{2(0.00391)(100 \text{ [m]}) \left(1.295 \text{ [kg/m}^3\text{]}\right) u^2}{0.2027 \text{ [m]}} = 4.99 u^2$$

$$\boxed{\Delta P = 4.99 u^2}$$

$$17.10 \quad \frac{dN_A}{dt} = r_A V \quad N_A = N_{AO}(1-x)$$

$$N_{AO} \frac{dx}{dt} = -r_A V$$

$$C_{AO} \frac{dx}{dt} = -r_A V$$

$$\int_0^t dt = C_{AO} \int_0^x \frac{dx}{-r_A}$$

$$t = C_{AO} \int_0^x \frac{dx}{-r_A}$$

$$-r_A = kC_A = kC_{AO}(1-x)$$

$$t = C_{AO} \int_0^x \frac{dx}{kC_{AO}(1-x)}$$

$$t = \frac{1}{k} \ln\left(\frac{1}{1-x}\right)$$

$$(a) \quad \frac{t_2}{t_1} = \frac{\frac{1}{k_2} \ln\left(\frac{1}{1-x_2}\right)}{\frac{1}{k_1} \ln\left(\frac{1}{1-x_1}\right)} = \frac{\ln\left(\frac{1}{1-0.95}\right)}{\ln\left(\frac{1}{1-0.90}\right)} = \frac{\ln(20)}{\ln(10)} = 1.30$$

Increase of 30%

$$(b) \quad \frac{t_2}{t_1} = 1.5 = \frac{\ln\left(\frac{1}{1-x_2}\right)}{\ln\left(\frac{1}{1-0.8}\right)}$$

$$x_2 = 0.91$$

$$17.11 \quad -r_A = \frac{K_1 C_A}{1 + K_2 C_A} = \frac{K_1 C_{AO}(1-x)}{1 + K_2 C_{AO}(1-x)}$$

$$t = C_{AO} \left[ \int_0^x \frac{dx}{K_1 C_{AO}(1-x)} + \int_0^x \frac{K_2}{K_1} dx \right]$$

$$t = \frac{1}{K_1} \ln \left( \frac{1}{1-x} \right) + \frac{K_2}{K_1} C_{AO} x$$

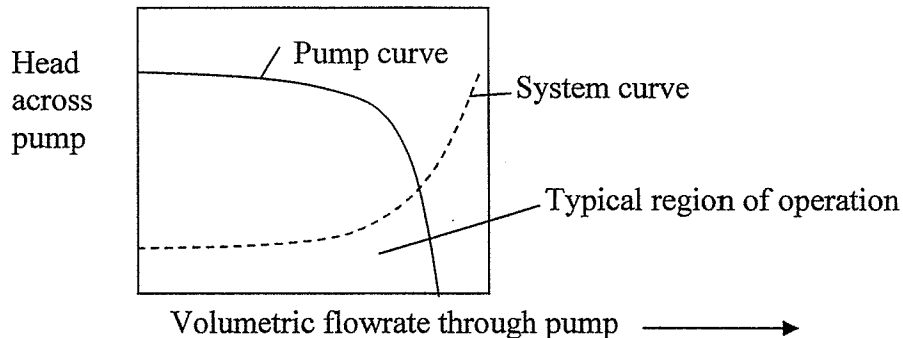
$$\frac{t_2}{t_1} = \frac{\frac{1}{K_1} \ln \left( \frac{1}{1-x_2} \right) + \frac{K_2}{K_1} C_{AO} x_2}{\frac{1}{K_1} \ln \left( \frac{1}{1-x_1} \right) + \frac{K_2}{K_1} C_{AO} x_1}$$

Need more information; limitations of method.

## Chapter 18

- 18.1 A pump curve shows the relationship between pressure head and volumetric flowrate through a pump

For a centrifugal pump

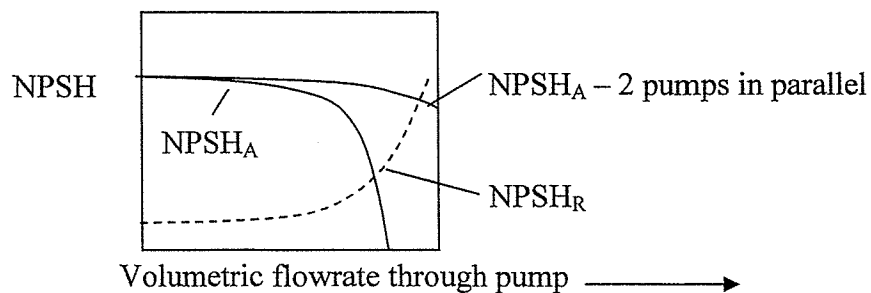


- 18.2 All depends on the system curve, so the statement is false since parallel or series arrangements could give higher scale up. An additional consideration is that when placing two pumps in series, the maximum generated head is twice that for a single pump or 2 pumps in parallel and this may cause overpressure in downstream units.

- 18.3 For the pump and system curve (Figure a) the intersection represents the maximum flow through the pump. Operation to the right of this point is not possible. For the NPSH curve (Figure b), operation to the right of the intersection point may be possible but will cause the pump to cavitate.

- 18.4 The film heat transfer coefficient, for turbulent flow inside a tube, increases with flowrate to the 0.8 power. For shell side flow, for laminar flow, and for phase changes the relationships are different.

- 18.5 The only way to fix the cavitation problem for this case is to raise the  $NPSH_A$  curve. Therefore, by running the pumps in parallel, the flow is split between the pumps and the suction line friction loss will be reduced, thus raising the  $NPSH_A$  curve.



- 18.6 For laminar flow, the flow of liquid is governed by the Hagen-Poiseuille equation given as:

$$-\Delta P_f = \frac{128\mu V L_p}{\pi D_p^4}$$

For turbulent flow the frictional pressure drop is given by

$$-\Delta P_f = \frac{2f_f \rho V^2 L_p}{D}$$

where the friction factor is a weak function of Reynolds number and hence of viscosity. Therefore, the laminar flow case will be much more sensitive to changes in temperature than the turbulent flow case.

- 18.7 The relative change in pressure and hence vapor specific volume in the two columns will be  $10/9.6 = 1.04$  and  $1/0.4 = 2.5$ . The vapor velocity is proportional to the specific volume (1/density) and hence the 1 bar column is much more likely to flood than the 10 bar column.

- 18.8 (a) An increase in column pressure increases the vapor density and decreases the superficial vapor velocity in the tower. Hence the column will have a reduced tendency to flood.
- (b) The increase in column pressure will bring the equilibrium line closer to the x-y line. This makes the separation harder. Hence a higher reflux ratio will be required to obtain the desired separation.
- (c) The number of stages will increase since, again, the separation is harder and the efflux and top and bottom purities are fixed.
- (d) The overhead condenser temperature will increase as predicted by Antoine's equation.

- 18.9 From Equation (18.21), the  $NPSH_A$  is given by

$$NPSH_A = P_{tank} + \rho g h - \frac{2\rho f L u^2}{D} - p^*$$

For this case, the pressure in the tank is the vapor pressure and the 1<sup>st</sup> and last terms on the right hand side cancel. Therefore, changes in ambient temperature will have negligible effect on the tendency of the pump to cavitate. This is in contrast to not saturated liquids whose vapor pressure will change with ambient temperature.

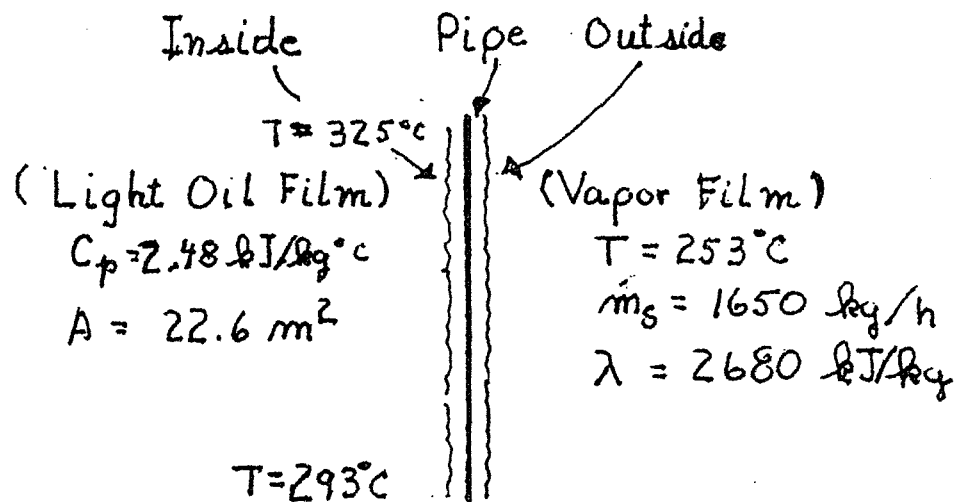
18.10 Assuming turbulent flow, we can write:

$$\frac{\Delta P_2}{\Delta P_1} = \frac{Q_2^2}{Q_1^2} \Rightarrow Q_2^2 = \frac{\Delta P_2}{\Delta P_1} Q_1^2 \Rightarrow Q_2 = Q_1 \sqrt{\frac{\Delta P_2}{\Delta P_1}} = (35) \sqrt{(5-1)/(3-1)} = 49.5 \text{ m}^3/\text{h}$$

18.11 (a) Only the 2-pumps in parallel set up will give you the desired flow. The desired operating point lies to the right of the single and 2-pumps in series set up and hence is not a viable operating point for those arrangements.

(b) There is not enough information to determine if the pumps will cavitate. You would need to look at the NPSH curves.

Problem 18.12



Equations Involved:  $\dot{q} = UA\bar{\Delta T}_{\ln} = \dot{m}_{\text{oil}}C_p\Delta T_{\text{oil}} = \dot{m}_s\lambda$

$$a) \quad \dot{q} = \dot{m}_s\lambda = \frac{1650 \text{ kg} \cdot 2680 \text{ kJ}}{\text{h}} = 4.42 \times 10^6 \text{ kJ/h}$$

$$\Delta T = \frac{(325 - 253) - (293 - 253)}{\ln[(325 - 253)/(293 - 253)]} = 54.4^{\circ}\text{C}$$

$$U = \frac{\dot{q}}{A\bar{\Delta T}_{\ln}} = \frac{4.42 \times 10^6 \text{ kJ/h}}{22 \text{ m}^2 \cdot 54.4^{\circ}\text{C}} = \underline{\underline{3595 \text{ kJ/h m}^2^{\circ}\text{C}}}$$

b) Clean Tubes:

$$R_{\text{ct}} = \frac{1}{U_{\text{ct}}A} = \frac{1 \text{ h m}^2^{\circ}\text{C}}{3595 \text{ kJ} \cdot 22.6 \text{ m}^2} = \underline{\underline{1.23 \times 10^{-5} \text{ h}^{\circ}\text{C/kJ}}}$$

Problem 18.13

Clean tubes:  $R_T^* = R_i^* + R_o^* + R_p^* + R_f^*$  Eq. i

Fig. 1.1 System:  $R_T = R_i + R_o + R_p + R_f$  Eq. ii

Eq. ii - Eq. i:  $(R_T - R_T^*) = (R_i - R_i^*) + (R_o - R_o^*) + (R_p - R_p^*) + (R_f - R_f^*)$  Eq. iii

From Prob 18.1:  $R_T^* = 1.23 \times 10^{-5} \text{ h}^\circ\text{C}/\text{kJ}$

From data in Fig 18.1

$$R_T = \frac{1}{UA} = \frac{1}{22.6 \text{ m}^2 | 2590 \text{ kJ} | \frac{\text{h}^\circ\text{C m}^2}{\text{kJ}}} = 1.71 \times 10^{-5} \text{ h}^\circ\text{C}/\text{kJ}$$

$$(R_T - R_T^*) = (1.71 - 1.23) \times 10^{-5} = 0.48 \times 10^{-5} \text{ h}^\circ\text{C}/\text{kJ}$$

Assume: Little steam side fouling (treated water)  $R_f = R_{f,i}$   
 R boiling water insensitive to flow ( $R_o^* = R_o$ )

For light oil:  $R \sim 1/h \sim 1/m^{0.8}$

For Clean tubes: (see Prob. 18.12)

$$m^* = \frac{q}{C_p \Delta T} = \frac{4.42 \times 10^6 \text{ kJ} | \text{kg}^\circ\text{C} |}{h | 2.48 \text{ kJ} | (325 - 293)^\circ\text{C}} = 55800 \text{ kg/h}$$

For Fig 18.1

$$m = \frac{3.43 \times 10^6 \text{ kJ} | \text{kg}^\circ\text{C} |}{h | 2.48 \text{ kJ} | (325 - 306)^\circ\text{C}} = 55300 \text{ kg/h}$$

$$\frac{R_i^*}{R_i} = \left( \frac{55800}{55300} \right)^{0.8} = 1.007 \Rightarrow R_i^* = 1.007 R_i$$

Eq. iii is reduced to

$$R_f = 0.48 \times 10^{-5} + 0.007 R_i^*$$

In limiting case where oil resistance controls\*  
 $R_i^* = R_T^*$

$$R_f = 0.48 \times 10^{-5} + 0.007 \times 1.23 \times 10^{-5} = 0.49 \text{ h}^\circ\text{C}/\text{kJ}$$

\* Improved estimate obtained by heat transfer ratios to replace this limit (see Prob. 18.12)



Problem 18.14

For clean tubes:

$$R_T = R_o + R_L + R_f + R_p$$

Eq. 1

$$R_f = 0 \text{ (clean tubes)}$$

$$R_p = 0 \text{ (assumed)}$$

$$R / R_i = 1/2$$

Eq. 1 becomes

$$R_T = R_o + 2R_o = 3R_o = 0.123 \times 10^{-5} \text{ h}^\circ\text{C}/\text{kJ}$$

$$R_o = 0.041 \times 10^{-5} \text{ h}^\circ\text{C}/\text{kJ}$$

$$R_i = 2 \times 0.041 \times 10^{-5} = 0.82 \times 10^{-5} \text{ h}^\circ\text{C}/\text{kJ}$$

$$h_o = \frac{1}{R_o A} = \frac{1}{0.041 \times 10^{-5} \text{ h}^\circ\text{C} \mid 22.6 \text{ m}^2} \text{ kJ}$$

$$= 10,800 \text{ kJ/h}^\circ\text{C m}^2$$

$$h_i = \frac{1}{R_i A} = \frac{1}{0.82 \times 10^{-5} \text{ h}^\circ\text{C} \mid 22.6 \text{ m}^2} \text{ kJ}$$

$$= 5,400 \text{ kJ/h}^\circ\text{C m}^2$$

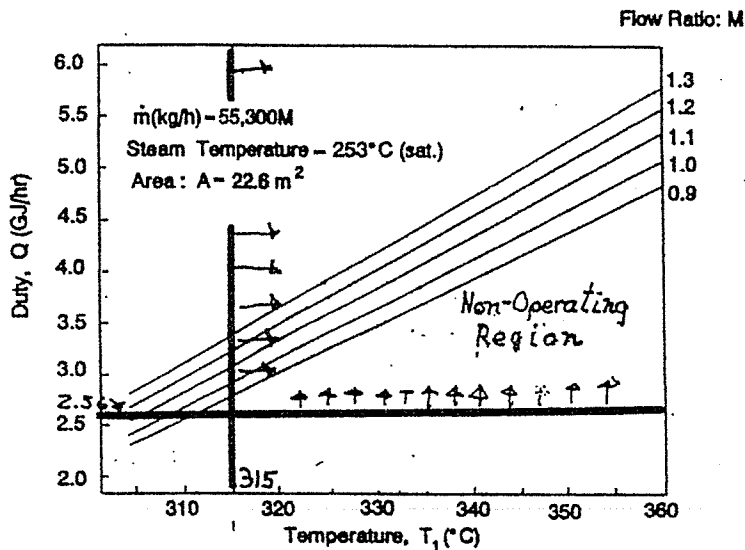
# Problem 18.15

Heat Flux Limitation (Table 9.11, Item 8)  $31.5 \text{ kW/m}^2$

$$\text{Max. } Q = \frac{31.5 \text{ kW}}{\text{m}^2} \times \frac{1 \text{ kJ}}{1 \text{ kW-s}} \times \frac{3600 \text{ s}}{1 \text{ h}} \times 22.6 \text{ m}^2 =$$

$$2.56 \times 10^6 \text{ kJ/h}$$

Temp. Limitation (Table 11.18, Item 14)  $T_{\text{max}} = 315^\circ\text{C}$



Sketch 18.4

Indicate Limits  
for  
 $Q$  & Flux

a) See Sketch: Max.  $T$  in operating =  $315^\circ\text{C}$   
 Max.  $Q = 2.56 \times 10^6 \text{ kJ/h}$

b) Oil temp @ operation exceeds  $315^\circ\text{C}$  (reaches  $325^\circ\text{C}$ )

i. On hot surface polymerization/decomposition expected

ii Exceeds normal flux level for light oil.

c)  $T = 312^\circ\text{C}$ ,  $M = 0.9$  ( $\dot{m} = 0.9 \times 55,300 = 49,800 \text{ kg/h}$ )

# Problem 18.16

From Eq 18.7

$$\frac{Q_{NEW}}{Q_{BC}} = \frac{\{\dot{m} C_{p,oil} (T_{in} - T_B) (1 - 1/K)\}_{NEW}}{\{\dot{m} C_{p,oil} (T_{in} - T_B) (1 - 1/K)\}_{B.C.}} \quad \text{Eq i}$$

$\dot{m}$ ,  $C_{p,oil}$ ,  $T_{in}$  &  $T_B$  are the same for base case  
(BC) & new case (NEW)

Eq i reduces to

$$\frac{Q_{NEW}}{Q_{BC}} = \frac{(1 - 1/K)_{NEW}}{(1 - 1/K)_{B.C.}} \quad \text{Eq ii}$$

From Eq 18.6

$$K = \exp \left[ \frac{U A}{\dot{m} C_{p,oil}} \right]$$

$$U = 2590 \text{ kJ} / \text{h m}^2 \text{ } ^\circ\text{C}$$

$$A_{NEW} = 25.1 \text{ m}^2$$

$$\dot{m} = 55,300 \text{ kg/h}$$

$$A_{BC} = 22.6 \text{ m}^2$$

$$C_p = 2.48 \text{ kJ} / \text{kg } ^\circ\text{C}$$

$$K_{NEW} = \exp \left[ \frac{2590 \times 25.1 \times 1}{55300 \times 2.48} \right] = 1.606$$

$$K_{BC} = \exp \left[ \frac{2590 \times 22.6 \times 1}{55300 \times 2.48} \right] = 1.533$$

Eq ii

$$a) \frac{Q_{NEW}}{Q_{BC}} = \frac{(1 - 1/1.606)}{(1 - 1/1.533)} = \underline{1.09}$$

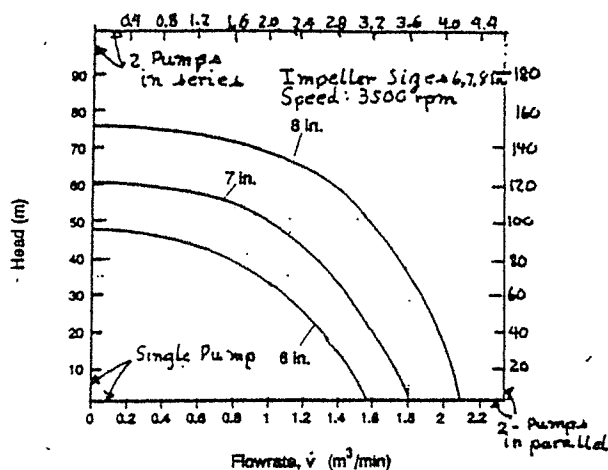
$$b) \frac{A_{NEW}}{A_{BC}} = \frac{25.1}{22.6} = \underline{1.11}$$

Increase in  $Q$  less than increase in  $A$

$$c) \quad Q = \cancel{A} \overset{C}{\downarrow} \overset{T}{\downarrow} \Delta T$$

New case has closer approach temperature  
leading to smaller increase in  $Q$  ratio

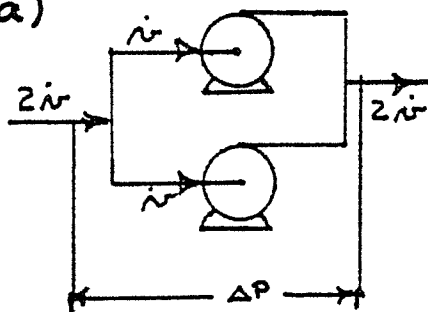
# Problem 18.17



$$\frac{400 \text{ gal}}{\text{min}} \left| \frac{\text{m}^3}{264 \text{ gal}} \right| = 1.5 \frac{\text{m}^3}{\text{min}}$$

$$\frac{800 \text{ gal}}{\text{min}} \left| \frac{\text{m}^3}{264 \text{ gal}} \right| = 3.0 \frac{\text{m}^3}{\text{min}}$$

a)

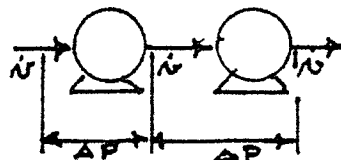


Each pump delivers same flow & pressure drop. Flows are additive

Top scale (flow) is 2x bottom scale

Use top scale & left scale two parallel pumps.

b)



Each pump delivers same ΔP.

Pressure drops are additive.

Left scale is 2x right scale

Use right scale & bottom scale for two series pumps

c)

|    | $\dot{V} = 1.5$ | $\dot{V} = 3.0$ |
|----|-----------------|-----------------|
| 6" | 41 m            | 8 m             |
| 7" | 58 m            | 29 m            |
| 8" | 72 m            | 57 m            |

|    | $\dot{V} = 1.5$ | $\dot{V} = 3.0$ |
|----|-----------------|-----------------|
| 6" | 15 m            | 0               |
| 7" | 50 m            | 0               |
| 8" | 112 m           | 0               |

# Problem 18.18

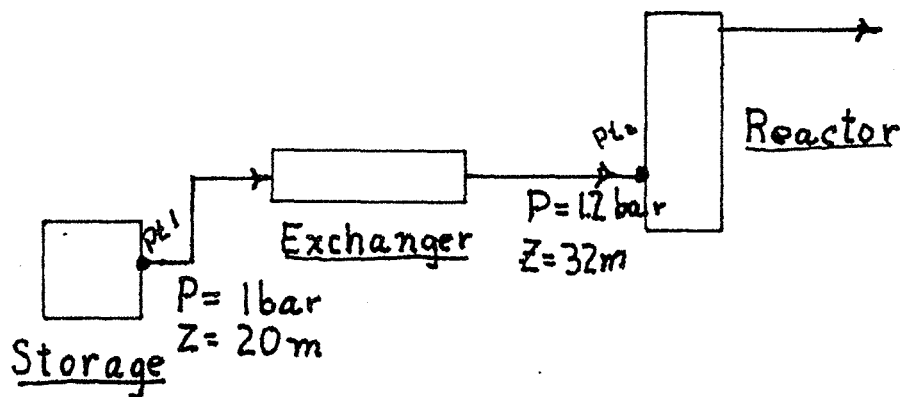


Figure 16.5 provides base case & 12.4 the pump curve

New flow:  $\dot{v} = 1.1 \times 1.5 = 1.65 \text{ m}^3/\text{min}$

Frictional loss:  $h_f = 20.6 \text{ m} \left( \frac{1.65}{1.1} \right)^2 = 46.4 \text{ m}$

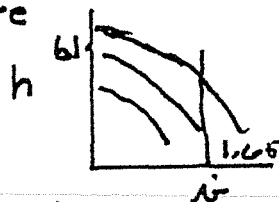
Elevation Change:  $\Delta h_z = 32 - 20 = 12.0 \text{ m}$

Static Pressure Change:  $(1.2 - 1.1) \times 10^5 / (750 \times 9.81) = 2.7 \text{ m}$

a. Head required from pump =  $46.4 + 12.0 + 2.7 = \underline{61.1 \text{ m}}$

b. @  $h = 61.1 \text{ m}$ ,  $\dot{v} = 1.65 \text{ m}^3/\text{min}$

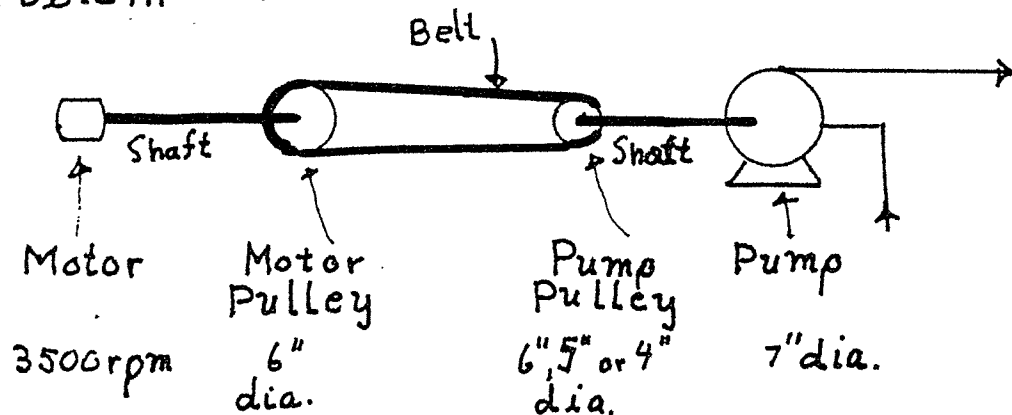
Pump systems shown on Figure 12.4 cannot provide flow & pressure



c. No pump system will operate if placed at point 2. Pressure at inlet required is negative "Not Possible"

Place pump at point 1

# Problem 18.19



Base Case: 6" dia motor & 6" dia. pump pulleys

Pump rpm:  $N_2 = N_1 (D_1/D_2)$

\* identifies values obtained from Fig 12.4  
(7in. impeller)

Head:  $h_2 = h_1^* (D_1/D_2)^2$

Flow:  $v_2 = v_1^* (D_1/D_2)^3$

6" Pulley:  $h_2 = h_1^* (6/6)^2 = 1.00 h_1^*$   
 $v_2 = v_1^* (6/6)^3 = v_1^*$

5" Pulley:  $h_2 = h_1^* (6/5)^2 = 1.44 h_1^*$   
 $v_2 = v_1^* (6/5)^3 = 1.728 v_1^*$

4" Pulley:  $h_2 = h_1^* (6/4)^2 = 2.25 h_1^*$   
 $v_2 = v_1^* (6/4)^3 = 3.375 v_1^*$

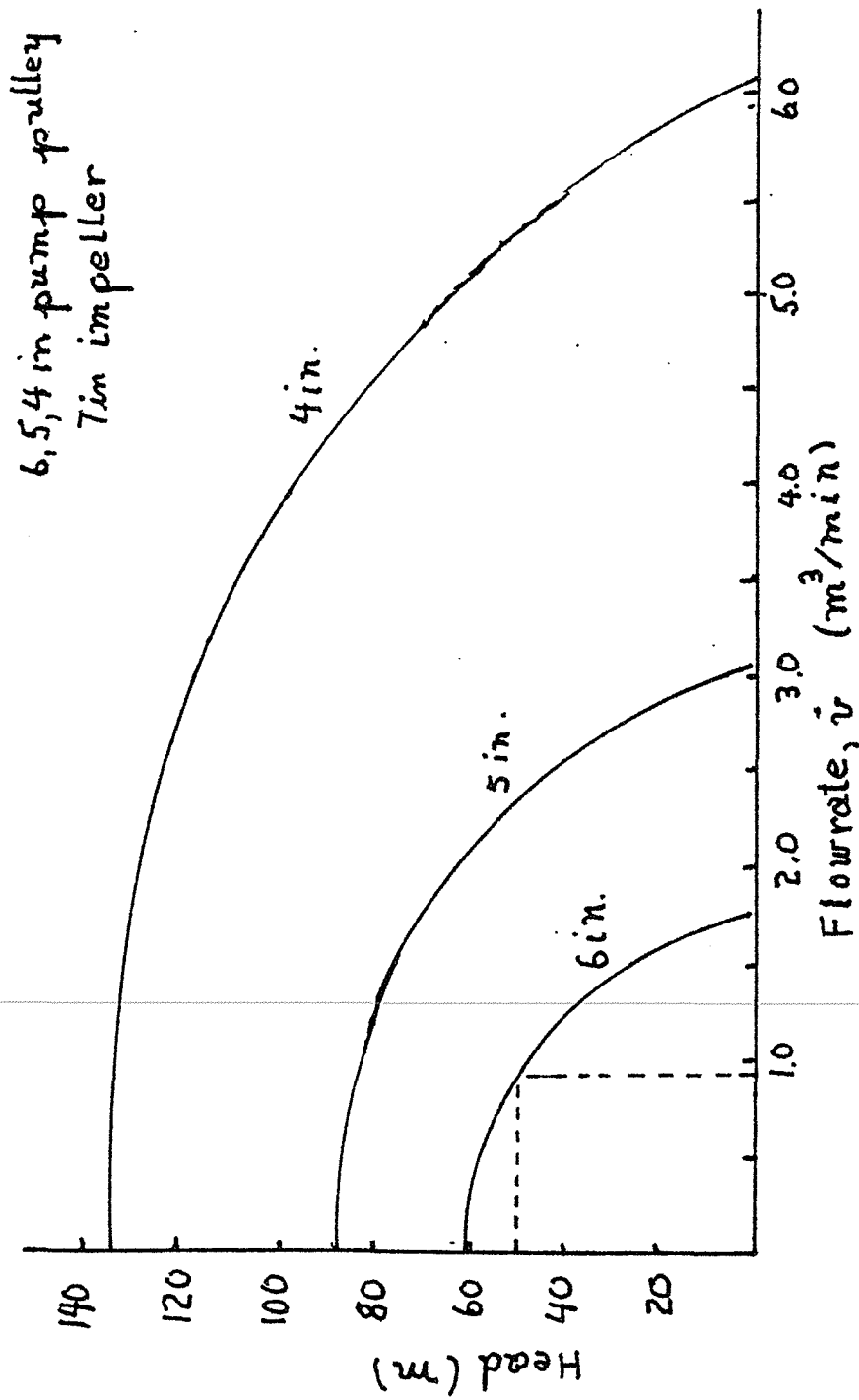
Sample Calculation: (5" Pulley)

From Fig. 16.4 @  $h_1^* = 50 \text{ m}$   $v_1^* = 0.95 \frac{\text{m}^3}{\text{min}}$

$h_2 = 1.44 \times 50 = 72 \text{ m}$

$v_2 = 2.25 \times 0.95 = 3.21 \text{ m}^3/\text{min}$

Problem 18.19 cont.



Performance Curves for Pump at Different Velocities

Problem 18.19

18.20 At 208°C,  $P^* = 78.96 \text{ kPa}$

$$NPSH_A = P_{inlet} - P^* = 79 \text{ kPa} - 78.96 \text{ kPa} = 0.04 \text{ kPa}$$

$$u = \frac{18,500 \text{ kg/h}}{(978.6 \text{ kg/m}^3)(0.001313 \text{ m}^2)(3600 \text{ s/h})} = 4 \text{ m/s}$$

For Figure 18.9  $NPSH_R \approx 0.2 \text{ kPa}$

$$NPSH_A < NPSH_R \Rightarrow \text{Not suitable}$$

Lower  $T$  - 1°C would be enough

$$P^* = 77 \text{ kPa and } NPSH_A = 2 \text{ kPa}$$



18.21 (a)  $1.25 \text{ in sch 80} \Rightarrow D_i = 0.03246 \text{ m}$

$$NPSH_A = P_{\text{tank}} + \rho gh - \frac{2\rho f L_{eq} u^2}{D} - P^*$$

$$NPSH_A = 1.10325 \text{ bar} + 870(9.81)(2)(10^{-5}) \text{ bar} - \frac{2(870)(0.005)(6)(10^{-5})(u^2)}{0.03246} \text{ bar}$$

$$-0.172 \text{ bar}$$

$$NPSH_A = 1.10194 - 0.01608(u^2)$$

to locate A:  $u = \frac{m}{\rho A}$ ;  $u \propto \frac{1}{D^2}$

$$u = 5.73 \text{ m/s} \left( \frac{D_1^2}{D_2^2} \right) = 5.73 \text{ m/s} \left( \frac{0.02664^2}{0.03246^2} \right) = 3.86 \text{ m/s}$$

$$u = 3.86 \text{ m/s}$$

(b)  $T = 40^\circ\text{C}$

$$\ln P^* = 10.97 - \frac{4203.6}{T(\text{K})}$$

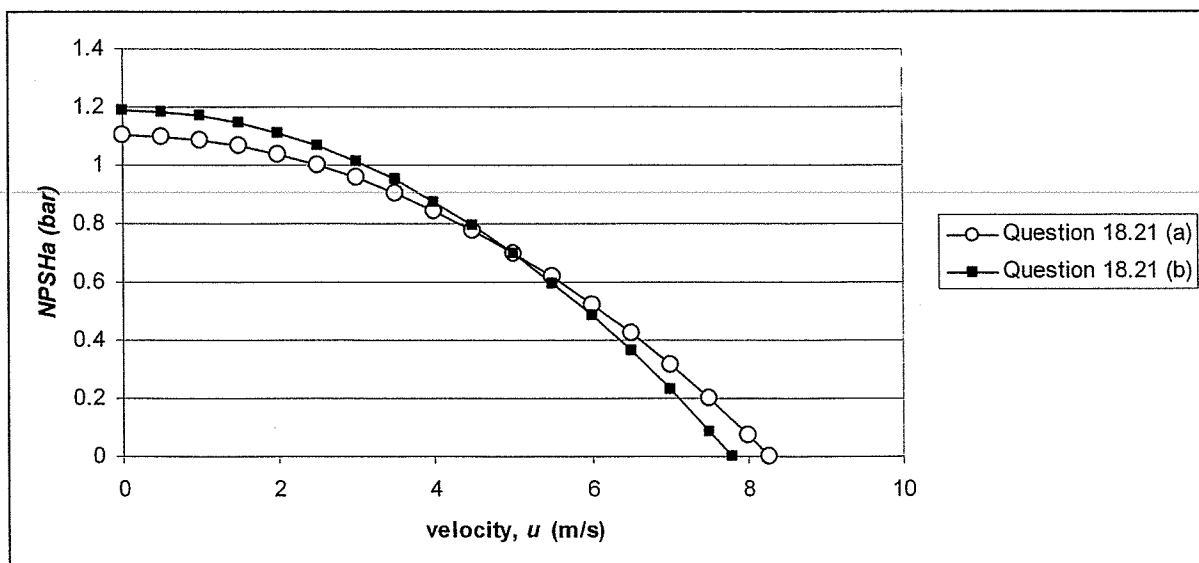
$$\ln P^* = 10.97 - \frac{4203.6}{40 + 273}$$

$$P^* = 0.0854 \text{ bar}$$

$$NPSH_A = 1.10325 \text{ bar} + 870(9.81)(2)(10^{-5}) \text{ bar} - \frac{2(870)(0.005)(6)(10^{-5})(u^2)}{0.02664} \text{ bar}$$

$$-0.0854 \text{ bar}$$

$$NPSH_A = 1.189 - 0.019595u^2$$



18.22 (a) at 125 Mg/h and 3500 rpm

$$P_{in} = 2 \text{ bar}$$

$$P_{out} = ?$$

$$\text{From Figure 18.11 } \frac{P_{out}}{P_{in}} \approx 3.6$$

$$\frac{P_{out}}{P_{in}} = \frac{P_{out}}{2 \text{ bar}} = 3.6$$

$$P_{out} = 7.2 \text{ bar}$$

(b) at 125 Mg/h and 2200 rpm

$$P_{in} = 4 \text{ bar}$$

$$P_{out} = 9 \text{ bar}$$

$$\text{From Figure 18.22 } \frac{P_{out}}{P_{in}} = 1.8$$

$$\frac{P_{out}}{P_{in}} = \frac{P_{out}}{4 \text{ bar}} = 1.8$$

$$P_{out} = 7.2 \text{ bar} \Rightarrow \text{Not possible}$$

(c) at 125 Mg/h

$$P_{in} = 2 \text{ bar}$$

$$P_{out} = 24 \text{ bar}$$

$$\frac{P_{out}}{P_{in}} = 12$$

$$\text{At 3500 rpm: } \frac{P_{out}}{P_{in}} \approx 3.6 \Rightarrow \text{use 2 stages}$$

$$P_{in,1} = 2 \text{ bar}$$

$$P_{out,1} = (2 \text{ bar})(3.6) = 7.2 \text{ bar} = P_{in,2}$$

$$P_{out,2} = (7.2 \text{ bar})(3.6) = 25.92 \text{ bar} \Rightarrow \text{Slightly past 24 bar}$$

$$\text{At 2200 rpm: } \frac{P_{out}}{P_{in}} = 1.8 \Rightarrow \text{use 5 stages}$$

$$P_{in,1} = 2 \text{ bar}$$

$$P_{out,1} = (2 \text{ bar})(1.8) = 3.6 \text{ bar} = P_{in,2}$$

$$P_{out,2} = (3.6 \text{ bar})(1.8) = 6.48 \text{ bar} = P_{in,3}$$

$$P_{out,3} = (6.48 \text{ bar})(1.8) = 11.664 \text{ bar} = P_{in,4}$$

$$P_{out,4} = (11.664 \text{ bar})(1.8) = 21.0 \text{ bar} = P_{in,5}$$

$$P_{out,5} = (21.0 \text{ bar})(1.8) = 37.79 \text{ bar} \Rightarrow \text{Much greater than 24 bar, big waste.}$$

$$\boxed{\text{Use 3500 rpm with 2 stages}}$$

$$18.23 \quad N_{IOG} = 20$$

$$G = 80 \text{ kmol/h}$$

$$L = 20 \text{ kmol/h}$$

$$\frac{L}{G} = 0.25$$

$$\frac{y_{out}}{y_{in}} = \frac{0.006}{0.01} = 0.6$$

$$A = 0.4 = \frac{L}{mG} \Rightarrow m = \frac{20}{(80)(0.4)} = 0.625$$

$$\text{If } L \text{ is at 85\% } A = \frac{L}{mG} = (0.85)(0.4) = 0.34$$

$$\frac{y_{out}}{y_{in}} = 0.56$$

$$y_{out} = 0.56(0.01)$$

$$\boxed{y_{out} = 0.0056}$$

18.24

For a Packed tower, use Eqn (18.23)

$$\frac{y_{A,in} - y_{A,out}}{y_{A,out} - y_{A,in}} = \frac{e^{N_{OG}(1+1/A)} - (1/A)}{1 - (1/A)} \quad (18.23)$$

N = 10  $y_{A,out} = 0$  $y_{A,in} = 0.02$  $y_{A,out} = 0.001$ 

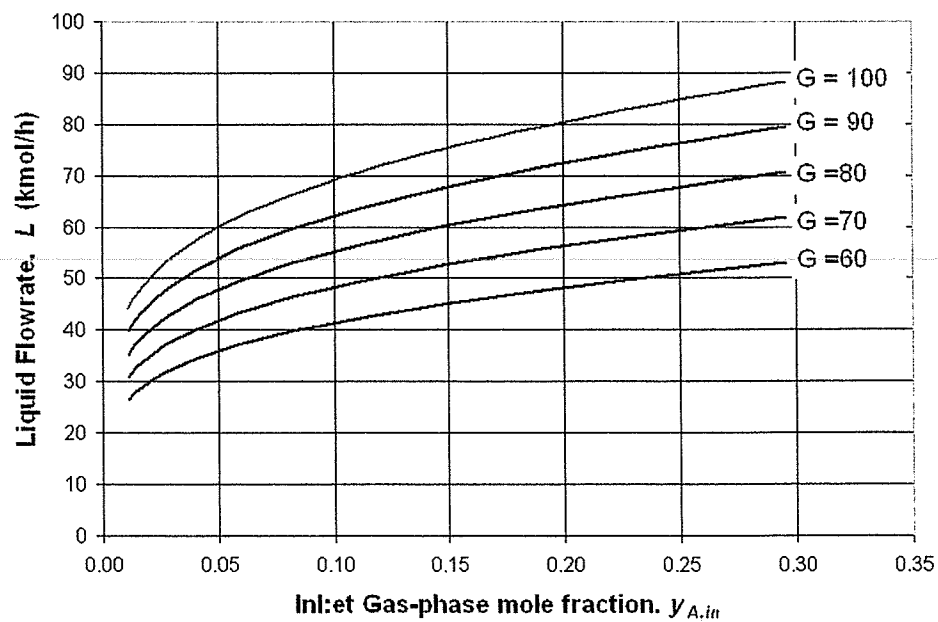
G = 80

L = 40

Solving 18.23 for A we get A = 1.1326

m = L/AG 0.44

| $y_{A,in}$ | A    | G    |      |         |       |       |
|------------|------|------|------|---------|-------|-------|
|            |      | 100  | 90   | 80      | 70    | 60    |
| 0.0110     | 1.00 | 44.1 | 39.7 | 35.317  | 30.9  | 26.49 |
| 0.0138     | 1.05 | 46.4 | 41.7 | 37.0828 | 32.45 | 27.81 |
| 0.0173     | 1.10 | 48.6 | 43.7 | 38.8487 | 33.99 | 29.14 |
| 0.0216     | 1.15 | 50.8 | 45.7 | 40.6145 | 35.54 | 30.46 |
| 0.0268     | 1.20 | 53   | 47.7 | 42.3804 | 37.08 | 31.79 |
| 0.0402     | 1.30 | 57.4 | 51.7 | 45.9121 | 40.17 | 34.43 |
| 0.0584     | 1.40 | 61.8 | 55.6 | 49.4438 | 43.26 | 37.08 |
| 0.0821     | 1.50 | 66.2 | 59.6 | 52.9754 | 46.35 | 39.73 |
| 0.1117     | 1.60 | 70.6 | 63.6 | 56.5071 | 49.44 | 42.38 |
| 0.1477     | 1.70 | 75   | 67.5 | 60.0388 | 52.53 | 45.03 |
| 0.1903     | 1.80 | 79.5 | 71.5 | 63.5705 | 55.62 | 47.68 |
| 0.2397     | 1.90 | 83.9 | 75.5 | 67.1022 | 58.71 | 50.33 |
| 0.2958     | 2.00 | 88.3 | 79.5 | 70.6339 | 61.8  | 52.98 |



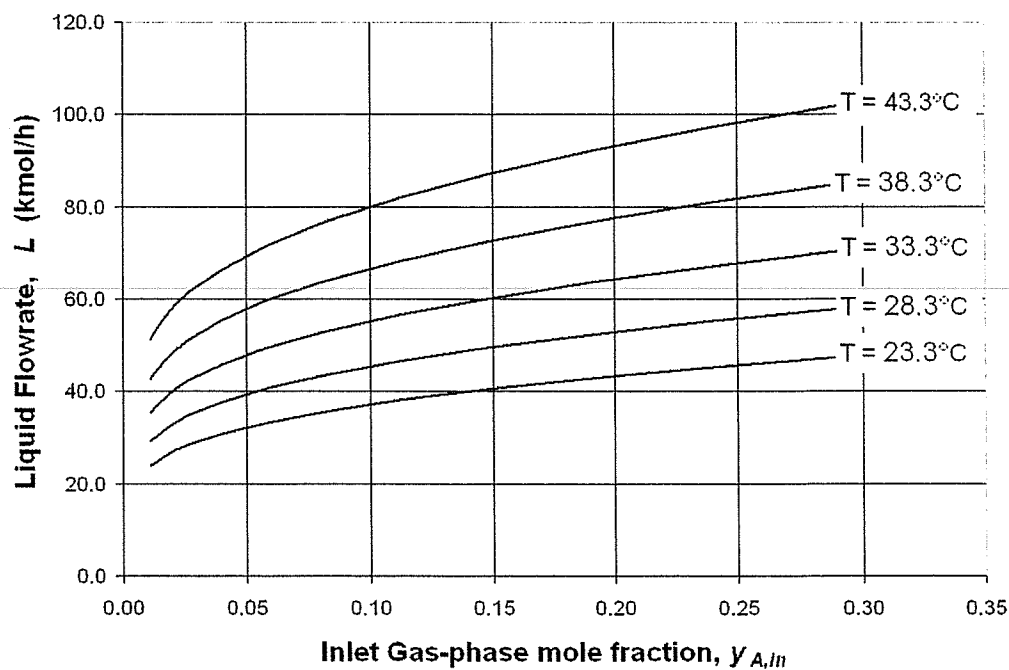
18.25

For a Packed tower, use Eqn (18.23)

$$\frac{y_{A,in} - y_{A,out}}{y_{A,out} - y_{A,eq}} = \frac{e^{N_{OG}(1-A)} - (1-A)}{1 - (1-A)} \quad (18.23)$$

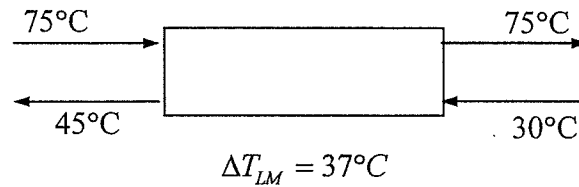
$N = 10$   
 $y_{A,out} = 0$   
 $y_{A,in} = 0.02$   
 $y_{A,eq} = 0.001$   
 $G = 80$  Fixed  
 $L = 40$   
 Solving 18.23 for A we get  $A = 1.1326$   
 $m = L/AG = 0.44$

| $y_{A,in}$ | A    | $m = \exp(10.92 - 3598/T[K])$ $T = 306.5 \text{ K} \quad 33.3^\circ\text{C}$ |      |               |      |      |  |
|------------|------|------------------------------------------------------------------------------|------|---------------|------|------|--|
|            |      | $\leftarrow T$                                                               |      | $\rightarrow$ |      |      |  |
|            |      | 43.3                                                                         | 38.3 | 33.3          | 28.3 | 23.3 |  |
| 0.0110     | 1.00 | 51.2                                                                         | 42.6 | 35.3          | 29.1 | 23.8 |  |
| 0.0138     | 1.05 | 53.7                                                                         | 44.8 | 37.1          | 30.5 | 25.0 |  |
| 0.0173     | 1.10 | 56.3                                                                         | 46.9 | 38.8          | 32.0 | 26.1 |  |
| 0.0216     | 1.15 | 58.8                                                                         | 49.0 | 40.6          | 33.4 | 27.3 |  |
| 0.0268     | 1.20 | 61.4                                                                         | 51.2 | 42.4          | 34.9 | 28.5 |  |
| 0.0402     | 1.30 | 66.5                                                                         | 55.4 | 45.9          | 37.8 | 30.9 |  |
| 0.0584     | 1.40 | 71.6                                                                         | 59.7 | 49.4          | 40.7 | 33.3 |  |
| 0.0821     | 1.50 | 76.8                                                                         | 64.0 | 53.0          | 43.6 | 35.7 |  |
| 0.1117     | 1.60 | 81.9                                                                         | 68.2 | 56.5          | 46.5 | 38.0 |  |
| 0.1477     | 1.70 | 87.0                                                                         | 72.5 | 60.0          | 49.4 | 40.4 |  |
| 0.1903     | 1.80 | 92.1                                                                         | 76.7 | 63.6          | 52.3 | 42.8 |  |
| 0.2397     | 1.90 | 97.2                                                                         | 81.0 | 67.1          | 55.2 | 45.2 |  |
| 0.2958     | 2.00 | 102.3                                                                        | 85.3 | 70.6          | 58.1 | 47.5 |  |



18.26 See Section 22.2 and Example 22.1

18.27 Increase capacity by 25%



$$Q = \dot{m} C_p \Delta T$$

$$Q = \dot{m} \lambda$$

$$Q = UA \Delta T_{LM}$$

Process

$$\text{If } \lambda = \text{constant then } \frac{Q_2}{Q_1} = \frac{\dot{m}_{2,p}}{\dot{m}_{1,p}} = 1.25$$

Cooling Water

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_2 \Delta T_2}{\dot{m}_1 \Delta T_1} = M \left( \frac{T - 30}{15} \right) \Rightarrow 1.25 = M \left( \frac{T - 30}{15} \right)$$

$$\frac{Q_2}{Q_1} = \frac{U_2 \Delta T_{LM,2}}{U_1 \Delta T_{LM,1}}$$

$$1.25 = M^{0.8} \left( \frac{\frac{(75 - 30) - (75 - T)}{\ln \left( \frac{75 - 30}{75 - T} \right)}}{37} \right) \Rightarrow 1.25 = \left( \frac{M^{0.8}}{37} \right) \left( \frac{T - 30}{\ln \left( \frac{45}{75 - T} \right)} \right)$$

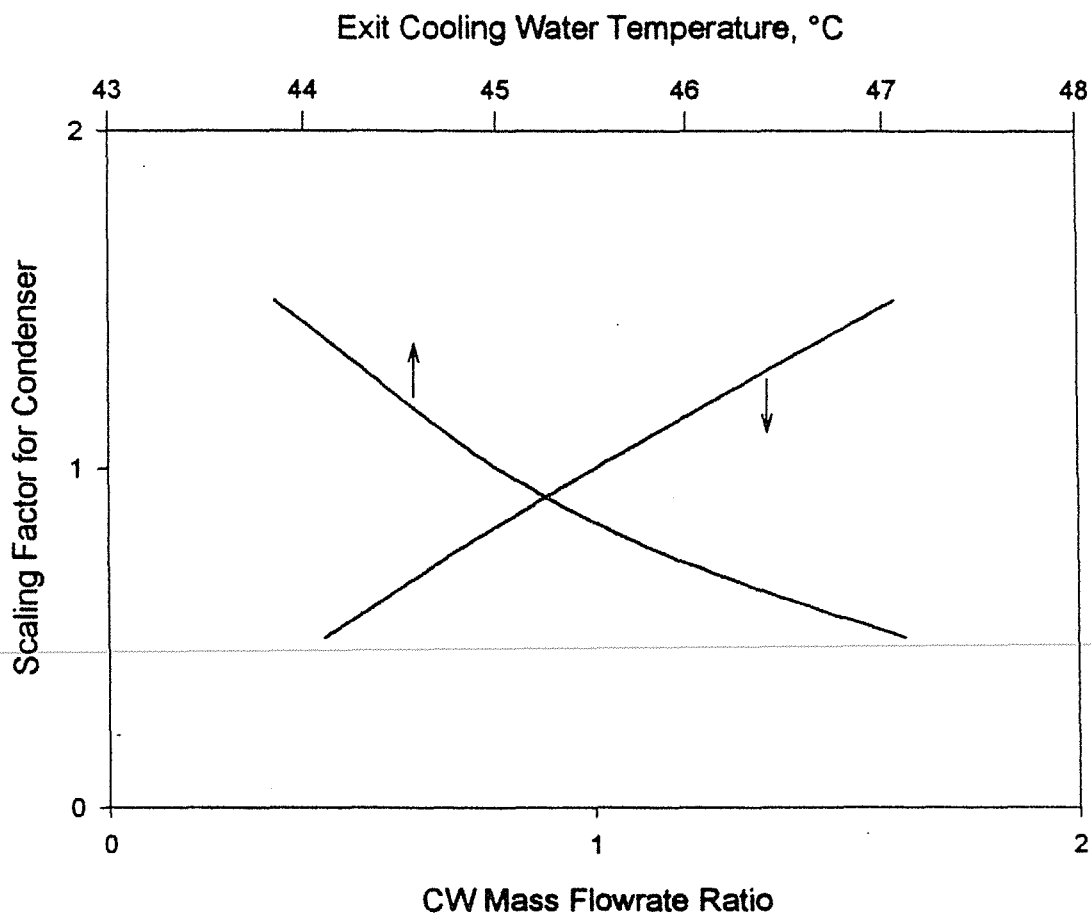
Solve equations (1) and (2) simultaneously

$T = 44.4^\circ\text{C}$  no penalty  
 $M = 1.31$  an increase of 31%

18.28

need to repeat analysis of Problem 18.27  
with  $\frac{Q_2}{Q_1}$  at different values

| $\frac{Q_2}{Q_1}$ | $M$   | $T(^{\circ}\text{C})$ |
|-------------------|-------|-----------------------|
| 0.50              | 0.439 | 47.10                 |
| 0.75              | 0.71  | 45.85                 |
| 1.00              | 1     | 45.00                 |
| 1.25              | 1.31  | 44.40                 |
| 1.50              | 1.62  | 43.85                 |



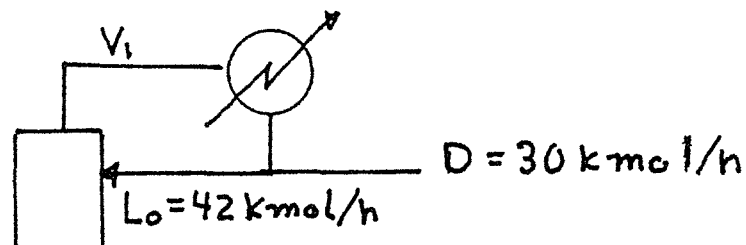
Problem 18.29

Upper velocity limit (flooding):  $N_f = 1.07 \text{ m/s}$

(See page 585)

Low velocity limit (Efficiency  $\epsilon = 0.6$ )

From Figure 12.19 Limit (lower) reflux =  $42 \frac{\text{kmol}}{\text{h}}$



@ Low velocity limit :  $V_{\text{MIN}} = 30 + 42 = 72 \text{ kmol/h}$

@ Upper velocity limit :  $V_{\text{MAX}} = 30 + 107 = 137 \text{ kmol/h}$

Assuming velocity of liquid and vapor flow ( $V$ )

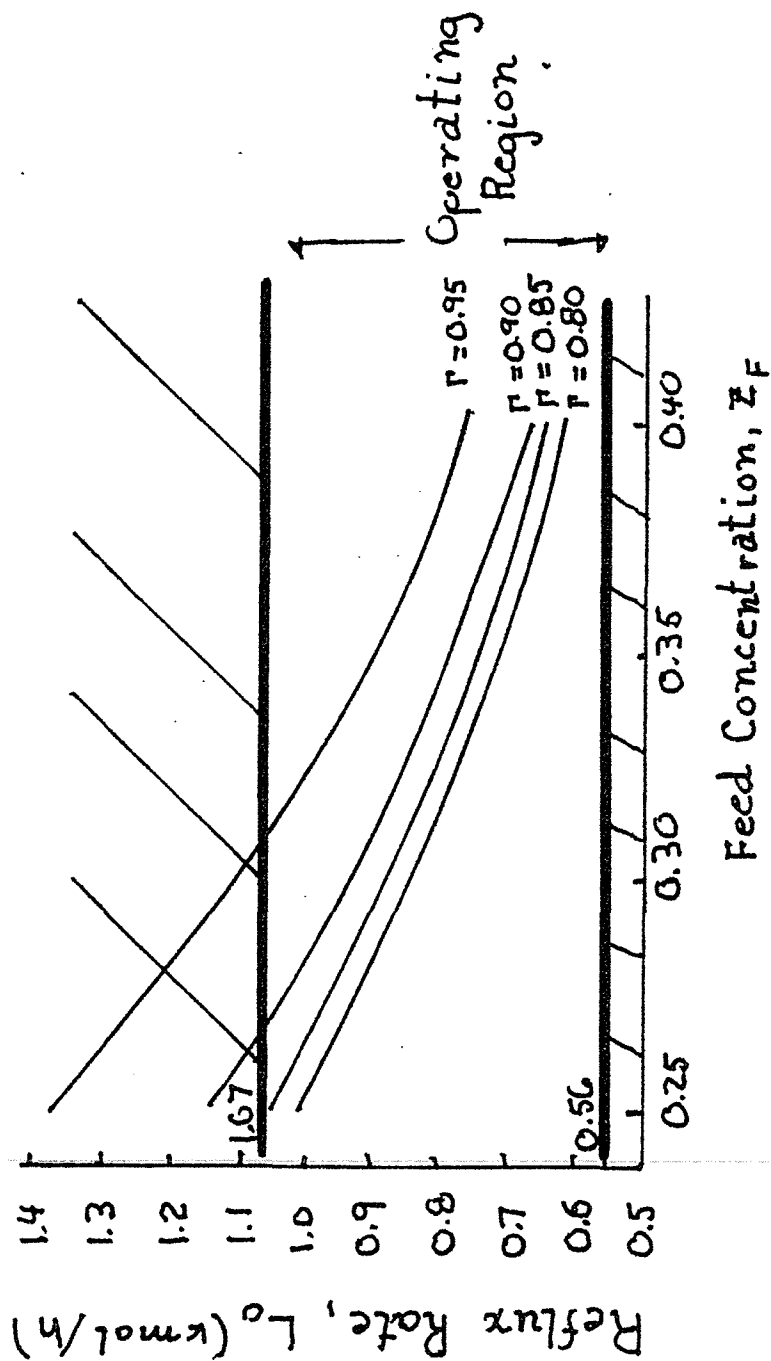
$$\mu_{\text{MIN}} = \mu_{\text{MAX}} \left( \frac{V_{\text{MIN}}}{V_{\text{MAX}}} \right) = 1.07 \left( \frac{72}{137} \right)$$

$$= 0.56 \text{ m/s}$$

Note:  $V_1$  is always a little larger than  $V_{N+1}$   
(see table P18.29) ∴ Flooding more likely  
at top of column



Problem 18.29



Performance Curve for Distillation Tower

Problem 18.30

Assume Figure 1.9 is correct

@ Higher Limit

$$V_{HL} = 107 + 30 = 137 \text{ kmol/h}$$

@ Lower Limit

$$V_{LL} = 42 + 30 = 72 \text{ kmol/h}$$

$$\text{Ratio } V_{LL}/V_{HL} = 72/137 = 0.52$$

Table P16.18 reveals  $V_1(\text{over-head}) > V_{D+1}(\text{boil-up})$

Example @  $P = 0.9$ ,  $z = 0.25$

$$\text{Over-head } V_1 = 144.5; V_{D+1} = 138$$

$$V_1/V_{D+1} = 144.5/138 = 1.05$$

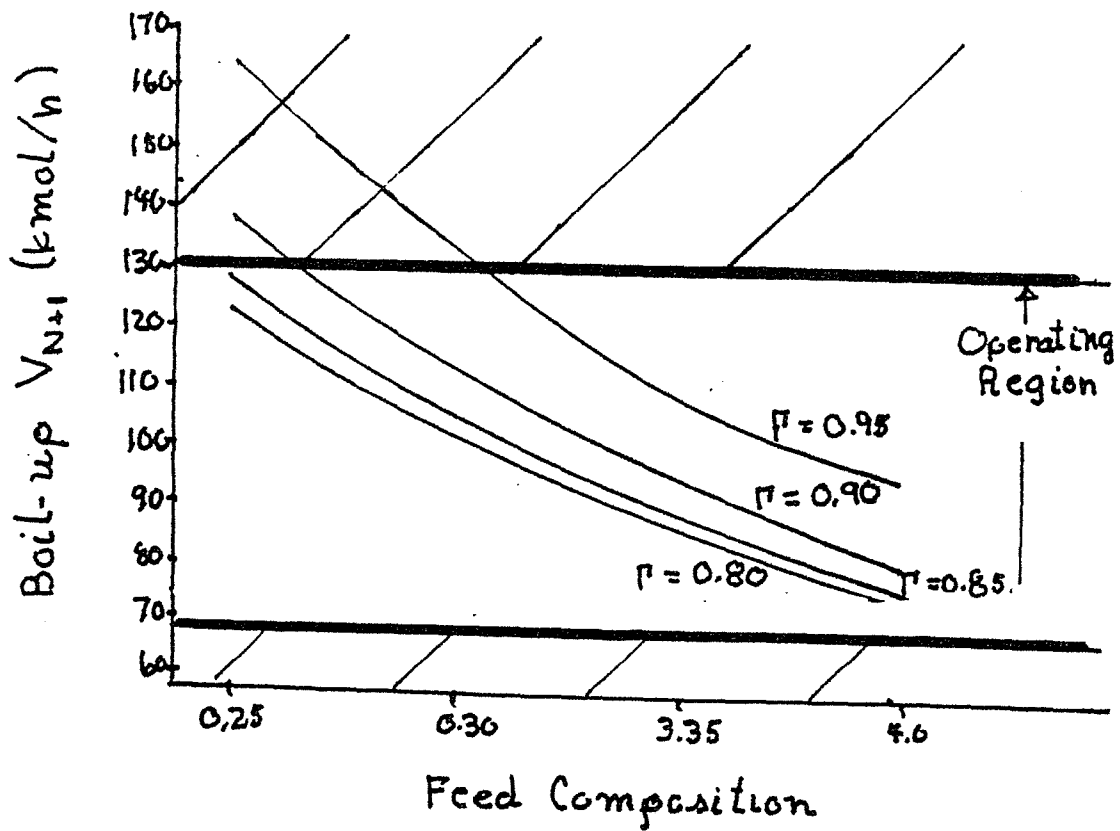
All conditions show ratio to be  $\approx 1.05$

Boil-up Limits:

$$\text{Higher Limit} = 137/1.05 = 130 \text{ kmol/hr}$$

$$\text{Lower Limit} = 76/1.05 = 69 \text{ kmol/hr}$$

# Problem 18.30 (cont.)



Problem 18.30 Performance Diagram for Distillation Tower

Problem 18.31

Assume  $Q_r \propto V_{N+1}$

$$\Delta \dot{H}_r = Q_r / V_{N+1}$$

Select mid-range conditions

$z = 0.3$ ,  $\Gamma = 0.9$  where Table P1.18 gives

$$V_{N+1} = 112.7 \text{ kmol/h}$$

$$Q_r = 3.61 \text{ GJ/h}$$

$$\Delta \dot{H} = 3.61 \text{ GJ/h} / 112.7 \text{ kmol/h} = 0.032 \text{ GJ/kmol}$$

For values of  $(V_{N+1})_{\text{MAX}}$  &  $(V_{N+1})_{\text{MIN}}$  from problem 10.19

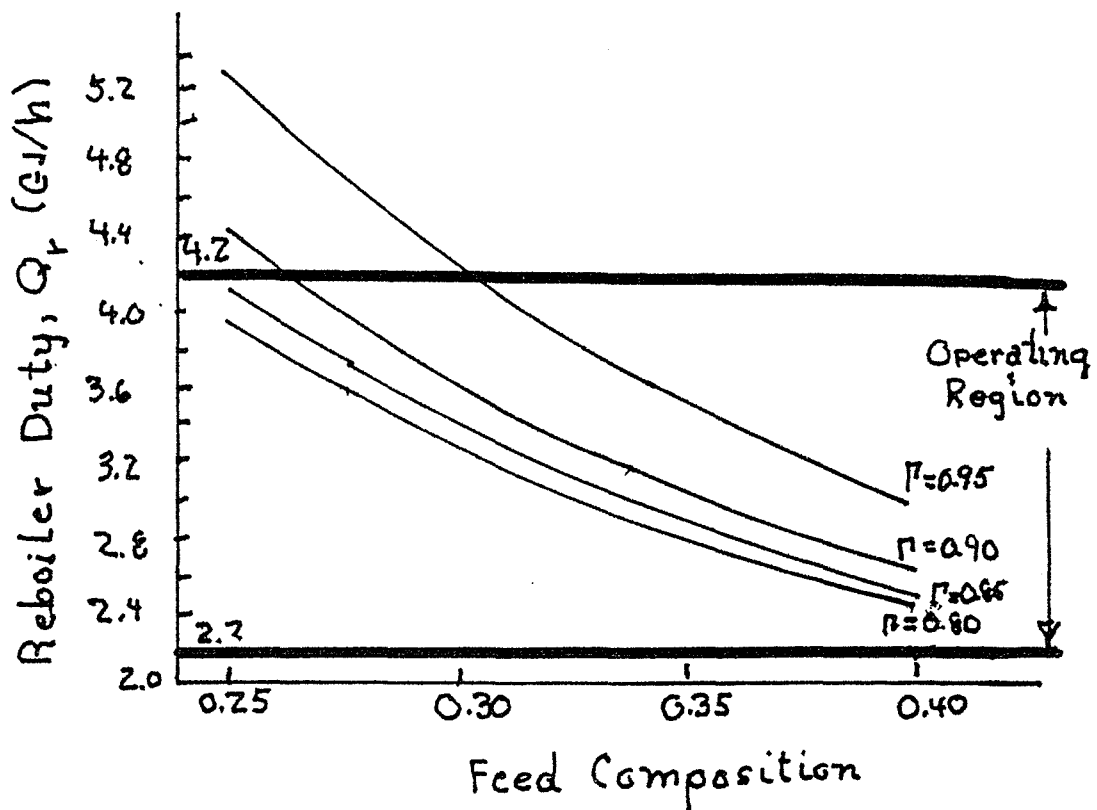
$$\text{Higher Limit: } V_{N+1} = 130 \text{ kmol/hr}$$

$$\text{Lower Limit: } V_{N+1} = 69 \text{ kmol/hr}$$

$$Q_{\text{High}} = 130 \times 0.032 = 4.2 \text{ GJ/h}$$

$$Q_{\text{Low}} = 69 \times 0.032 = 2.2 \text{ GJ/h}$$

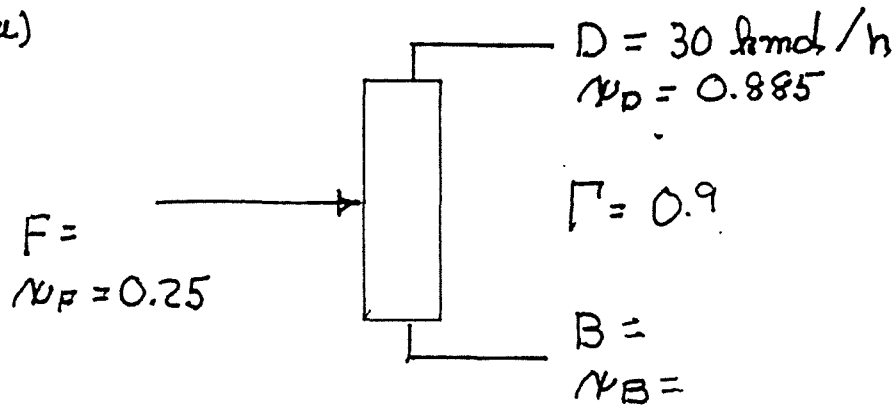
Problem 18.31 cont.



Problem 18.31 Performance Diagram for Distillation Tower

# Problem 18.32

a)



$$\Gamma F w_F = w_D D = (0.885)(30) = 26.55 \text{ kmol/h}$$

$$F = 0.26.55 / (0.25)(0.9) = 118 \text{ kmol/h}$$

$$B = F - D = (118 - 30) = 88 \text{ kmol/h}$$

$$B w_B = F w_F \Gamma = (118)(0.25)(1 - 0.9) = 2.95 \text{ kmol/h}$$

$$w_B = 2.95 / 88 = 0.0335$$

b) From Table P18.29

$$L_0 = 115.4 \text{ kmol/h} ; V_{N+1} = 138 \text{ kmol/h}$$

c) Not recommended: Flows in flooding region

# Problem 18.33

a) Develop curve of  $\Gamma_{\max}$  ( $\Gamma$  at flooding)

Obtained @ intersection of  $F$  with  
maximum reflux  $L_0 = 107$ , see Fig P18.29

| $\Gamma$ | $\eta_F$ | $\Gamma \eta_F$ |
|----------|----------|-----------------|
| 0.95     | 0.304    | 0.288           |
| 0.90     | 0.264    | 0.266           |
| 0.85 *   | 0.245    | 0.208           |
| 0.80 *   | 0.230    | 0.184           |

see plot

\* extrapolated curves

$$@ \eta_F = 0.321$$

$$\Gamma_{\max} = 0.96$$

c)  $F \Gamma \eta_F = D \eta_D = (30)(0.885) = 25.56 \text{ kmol/h}$

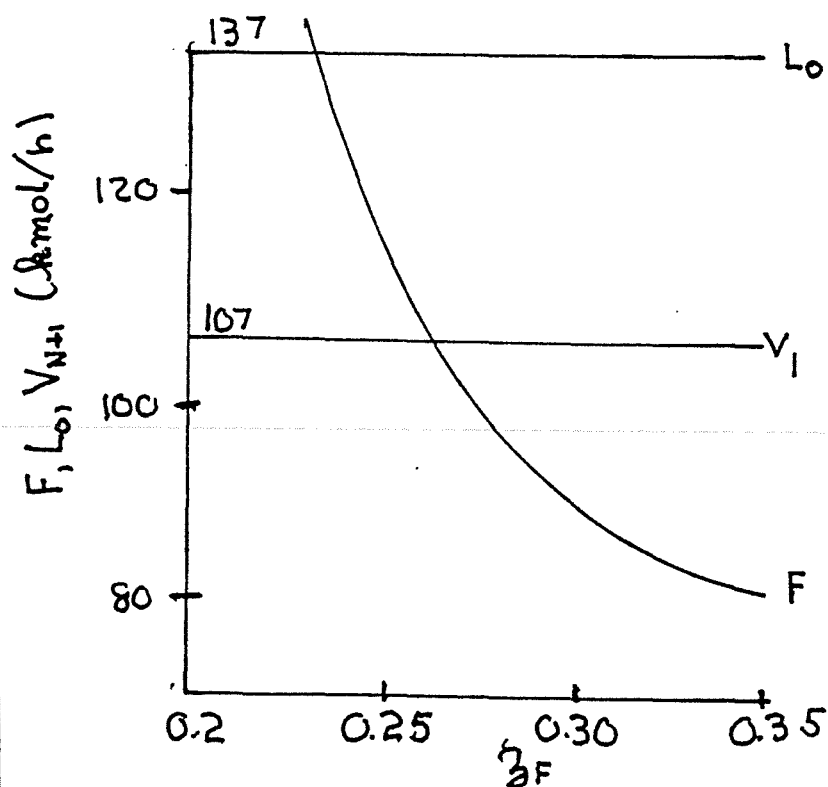
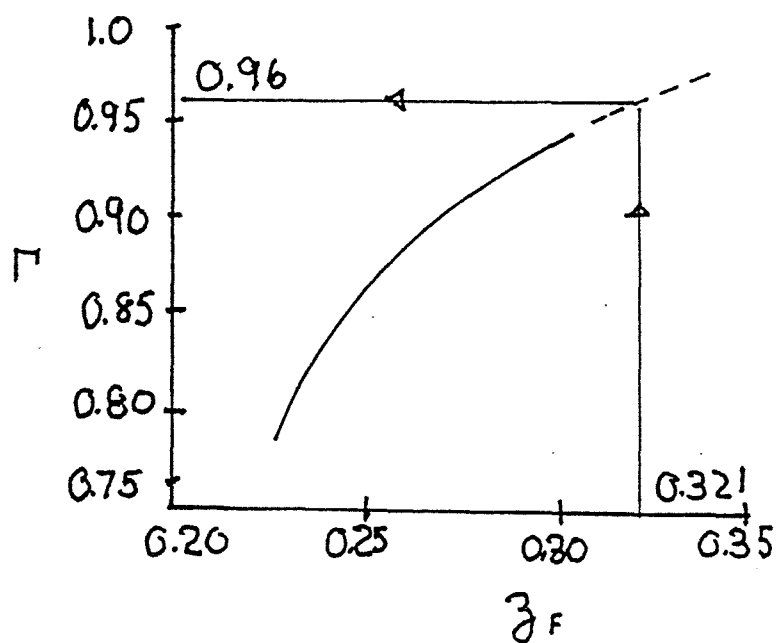
$$F = 25.56 / (\Gamma)(\eta_F)$$

$$V_1 = L_0 + D = 107 + 30 = 137 \text{ kmol/h}$$

| $\eta_F \Gamma$ | F     | $V_1$ | $L_0$ |
|-----------------|-------|-------|-------|
| 0.298           | 88.5  | 137   | 107   |
| 0.264           | 107.6 | ↓     | ↓     |
| 0.245           | 122.7 | ↓     | ↓     |
| 0.230           | 138.9 | ↓     | ↓     |

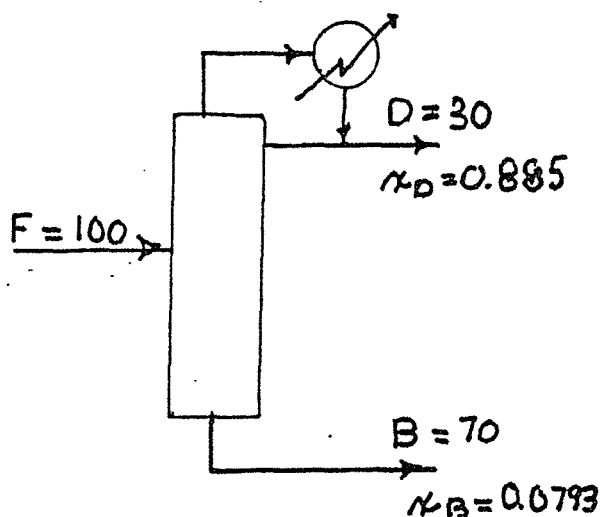
- d)
1. High risk of flooding
  2. High utility costs (Max. reflux used)
  3. Higher duties may not be possible in existing equipment.

# Problem 18.33





# Prob. 18.34 & 18.35 Supplemental Information



| Mole Fractions |         |
|----------------|---------|
| $X_1$          | $Y_1$   |
| 0.00000        | 0.00000 |
| 0.05000        | 0.09915 |
| 0.10000        | 0.18960 |
| 0.15000        | 0.27223 |
| 0.20000        | 0.34783 |
| 0.25000        | 0.41709 |
| 0.30000        | 0.48065 |
| 0.35000        | 0.53907 |
| 0.40000        | 0.59284 |
| 0.45000        | 0.64242 |
| 0.50000        | 0.68821 |
| 0.55000        | 0.73055 |
| 0.60000        | 0.76977 |
| 0.65000        | 0.80616 |
| 0.70000        | 0.83997 |
| 0.75000        | 0.87142 |
| 0.80000        | 0.90074 |
| 0.85000        | 0.92809 |
| 0.90000        | 0.95365 |
| 0.95000        | 0.97758 |
| 1.00000        | 1.00000 |

Operating Line Relationships

Lower:  $y_m = (\bar{L}/\bar{V})x_m + (B/\bar{V})x_B$

Upper:  $y_m = (L/V)x_m + (D/V)x_D$

Fig. 16.18 shows feed is not introduced on plate that gives maximum separation. This figure was obtained using a process simulator (SPR thermal package).

Solutions provided used a McCabe-Thiele diagram and equilibrium data shown above.

For the base case ( $R = 2.38$ )

$$L = (R)(D) = (2.38)(30) = 71.4$$

$$V = V + D = 71.4 + 30 = 101.4$$

$$\bar{V} = V = 101.4$$

$$\bar{L} = L + F = 71.4 + 100 = 171.4$$

$$\alpha = F/D = 100/30 = 3.333$$

Note: Assumptions used in graphical analysis from reading graphs limit the numerical result accuracy. They are valuable in predicting trends and aid in understanding. In this problem the graphical analysis finds 14.2 equilibrium stages & simulator 19.2.

Problem 18.34 (see supplement to Probs. 18.34 & 18.35)

To improve operation feed location is changed to the optimum plate. No changes in any input or output is acceptable. Reflux is the only variable that can be altered.

Re-writing operating lines in terms of  $R$  gives

$$\text{Upper: } y_n = R/(R+1) x_n + 0.885/(R+1)$$

$$\text{Lower: } y_m = \left( \frac{R+d}{R+1} \right) x_{m+1} + \left( \frac{d-1}{R+1} \right) x_D$$

For base case ( $R=2.38$ )

$$\text{Upper: } y_n = 0.704 x_n + 0.2618 ; \text{ Lower: } y_m = 1.69 x_m - 0.547$$

The operating lines for several values of  $R$  plotted along with equilibrium curve. As operating lines approach equilibrium curve and number of stages increase. At  $R=2.05$   $\infty$  number of stages are required.

Solution Strategy:

- i. Make best estimate of  $R$
- ii. Plot diagram
- iii. Step off stages
- iv. Compare to 14.2 stages (current system value)
- v. If not acceptable go to step i. If acceptable terminate

Sample of evaluation

- i.  $R = 2.34$
- ii.  $y_n = (2.34/3.34) x_n + (0.885)/(3.34) = 0.7006 x_n + 0.2650$
- iii.  $y_m = [(2.34+3.33)/3.34] x_m - 0.185/3.34 = 1.699 x_m - 0.534$   
 $\text{STAGES} = 14.2 \quad \text{SEE PLOT}$
- iv. Stages 14.2 calculated 14.2 available
- v. Acceptable

b. Duty  $Q \propto V$

$$\text{New } V = (R+1)D = (2.34+1)(30) = 106.2$$

$$\text{Base Case} = 101.4$$

$$\% \text{ Decrease} = \left( \frac{101.4 - 106.2}{101.4} \right) 100 = 1.2\%$$

Problem 18.34 cont.

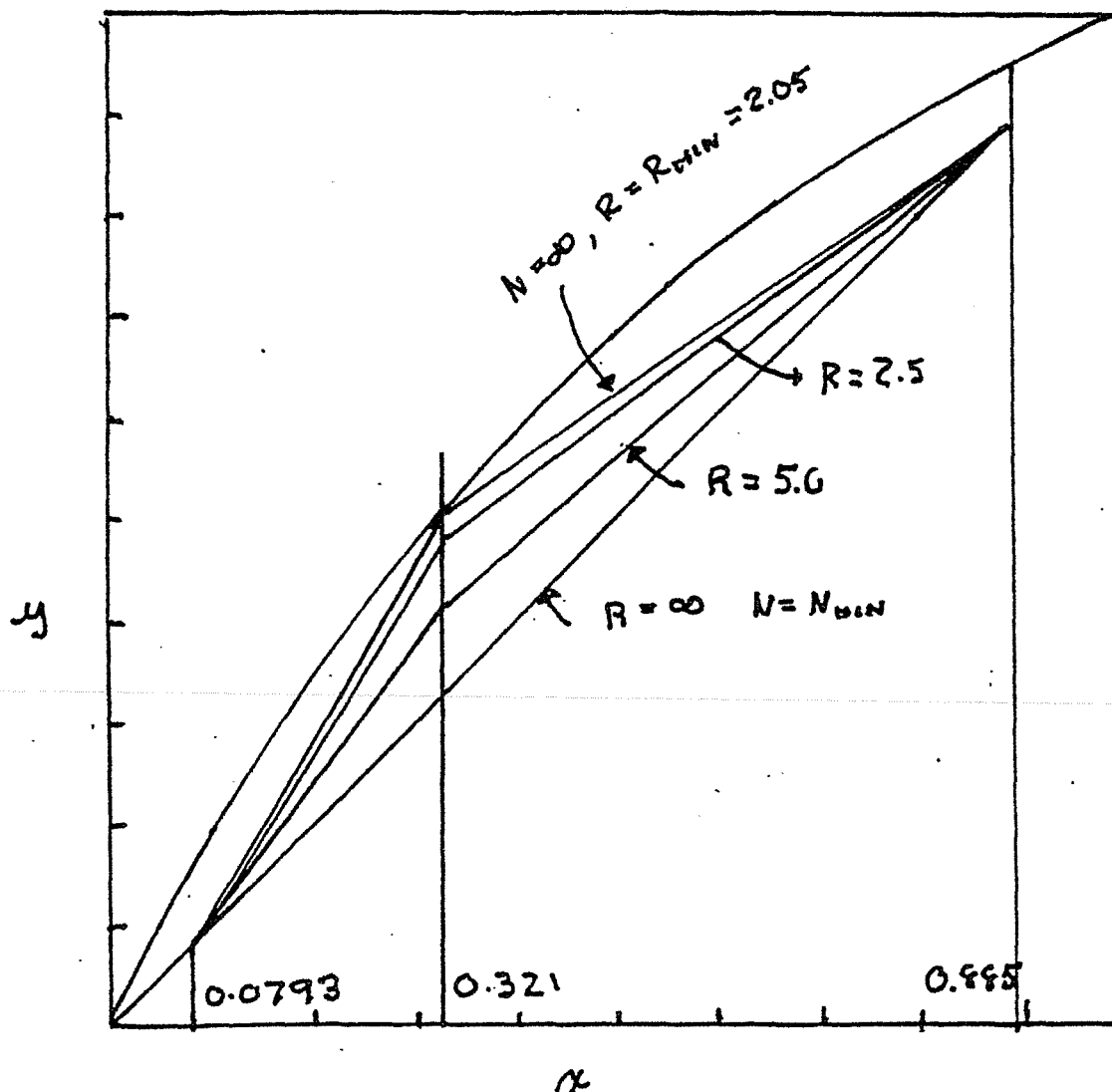
Estimate Current Steam Cost (Low Pressure)

From Table 3.4 Cost = \$ 3.17/GJ

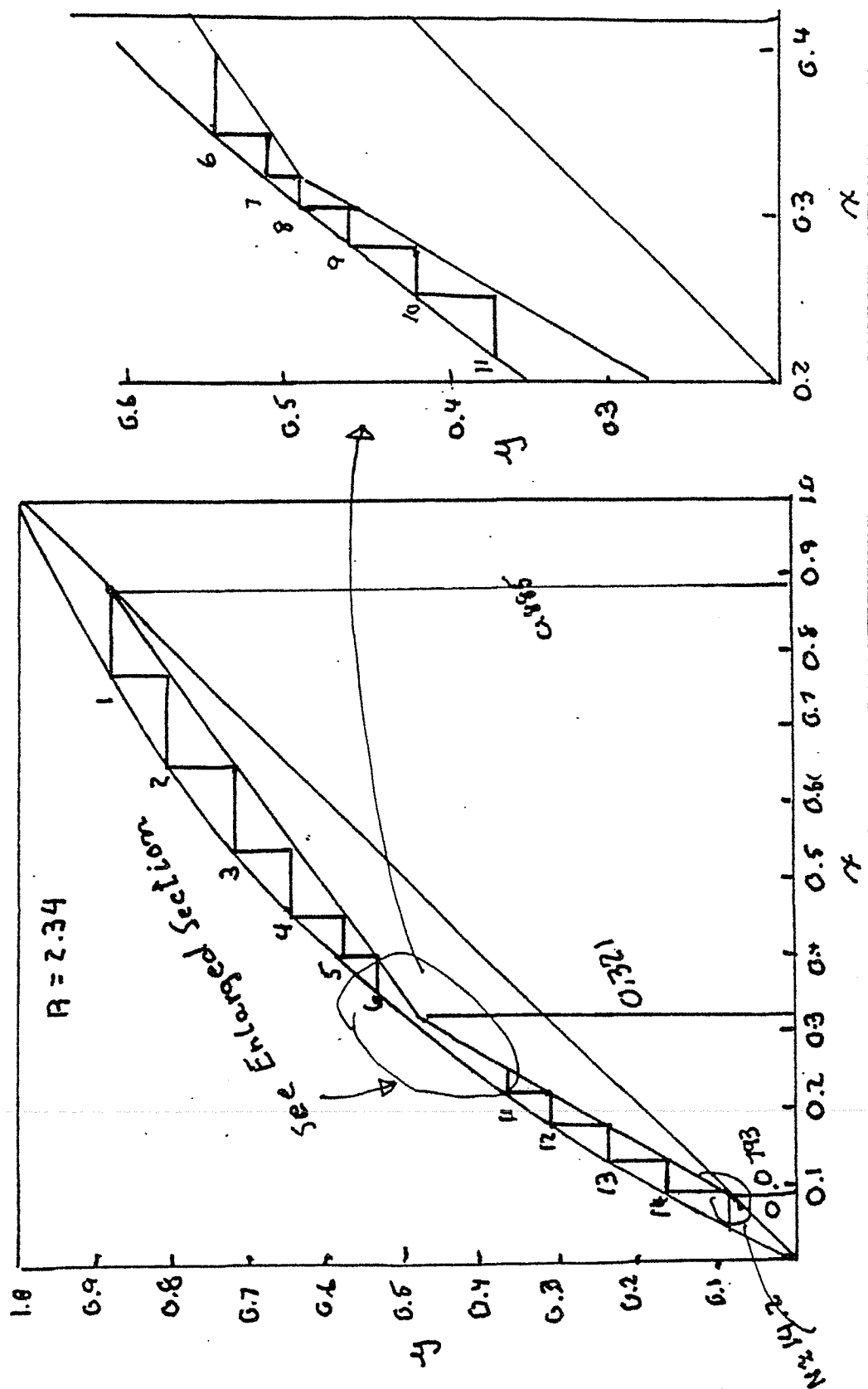
From Table P18.29 Duty = 3.0 GJ/hr

$$\text{Current cost} = \frac{3.0 \text{ GJ}}{\text{Hr}} \times \frac{24 \text{ Hr}}{\text{day}} \times \frac{330 \text{ day}}{\text{yr}} \times \frac{\$ 3.17}{\text{GJ}} = \$75,660/\text{yr}$$

$$\text{Savings} = (0.12)(75,660) = \$9,080/\text{yr}$$



# Problem 18-34



Problem 18.35 (see supplement to Probs 18.34 & 18.35)  
 To improve separation feed moved to optimum location. This alters value of  $x_D$  &  $x_B$ .

Material Balance:

$$Dx_D + Bx_B = Fx_F$$

For  $D=30, B=70, F=100$  &  $x_F=0.321$

$$x_B = 0.4586 - 0.4286x_D$$

Rewriting equation for upper operating line in terms of  $R$  gives

$$y_m = (R/(R+1))x_n + x_D/(R+1)$$

(For  $R=2.38$ )  $y_m = 0.704x_n + x_D/3.38$

The operating lines for several values of  $x_D$  are plotted along with the equilibrium curve. As  $x_D$  increases the operating line approaches the equilibrium line.

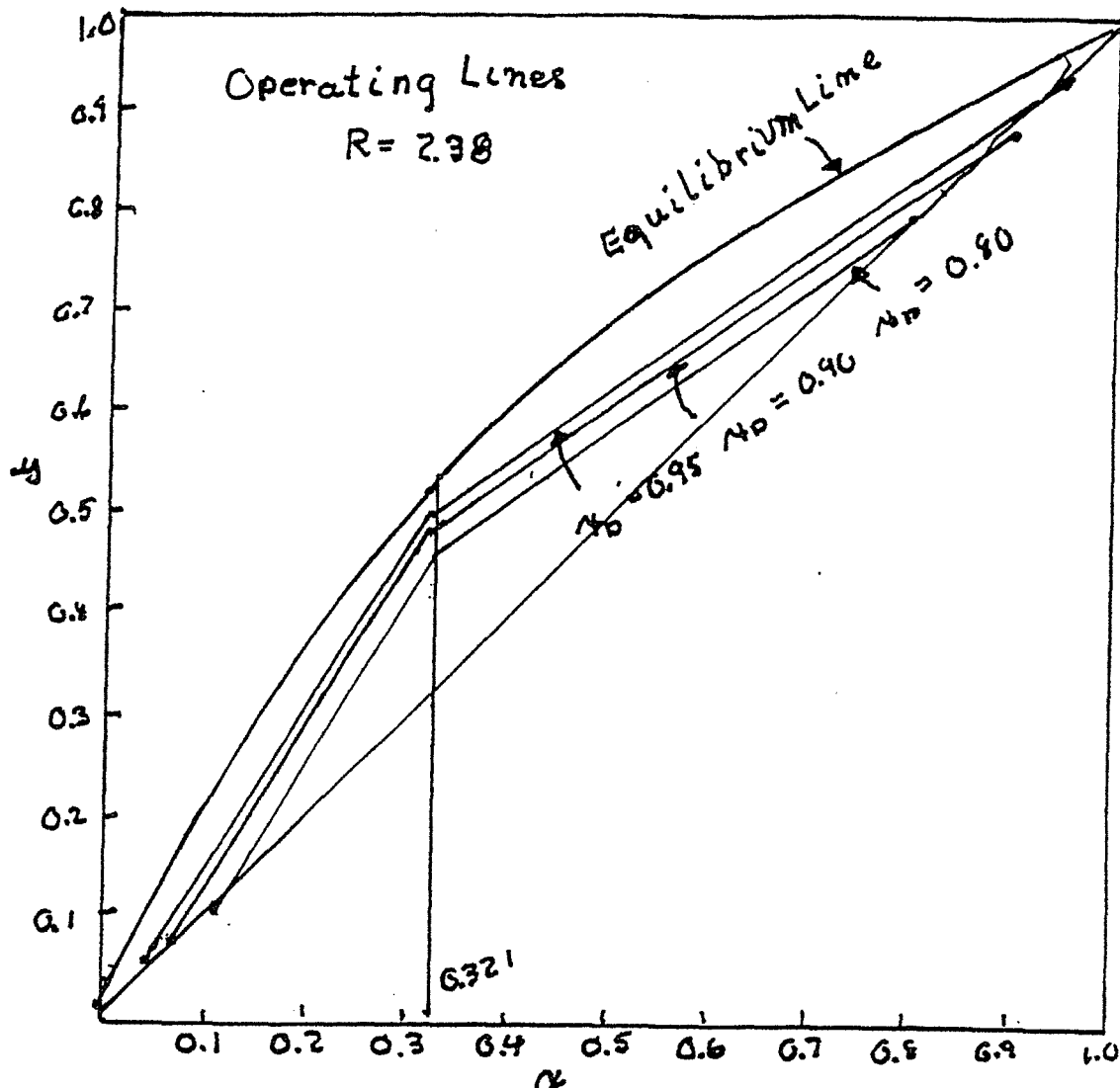
Solution Strategy:

- i. Make best estimate of  $x_D$
- ii. Plot operating line
- iii. Step off stages
- iv. Compare to 14.2
- v. If  $N$  is acceptable terminate. If not return to step i.

Sample Evaluation.

- i.  $x_D = 0.9 \Rightarrow x_B = 0.4586 - 0.4286(0.9) = 0.0729$
- ii. Operating Line:  $y_n = 0.704x_n + 0.9/3.38 = 0.704x_n + 0.279$
- iii. No. of Stages: 18.2 vs 14.2
- iv. Decision: Not acceptable,  $x_D$  is lower

Prob 18.35 cont



↳ Extrapolate to predict  $x_D$

②  $x_D = 0.885$        $N(\text{stages}) = 13.8$

②  $x_D = 0.9$        $N(\text{stages}) = 18.2$

③  $N = 14.2$  using linear extrapolation  $x_D = 0.8864$

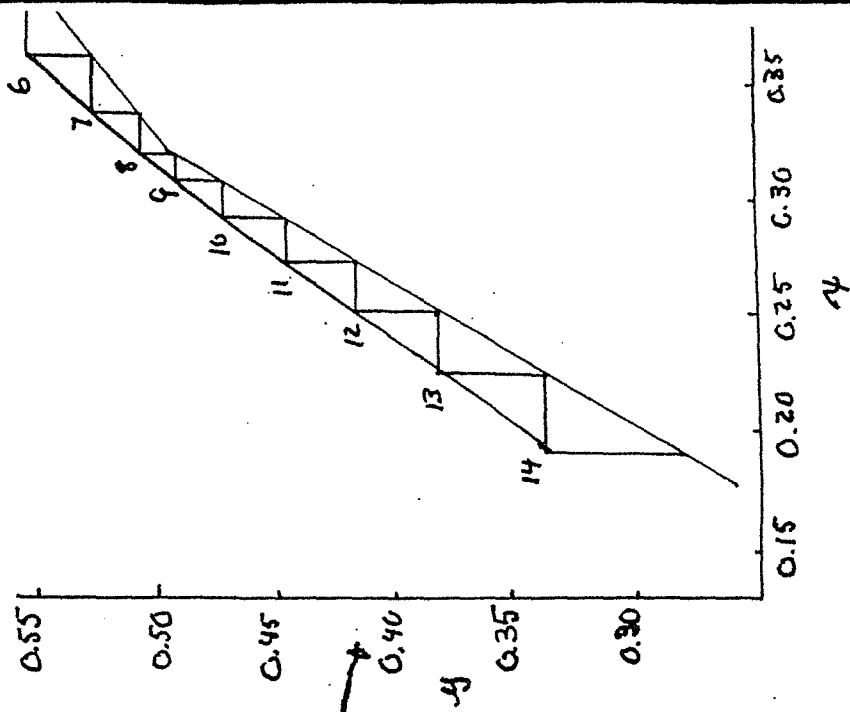
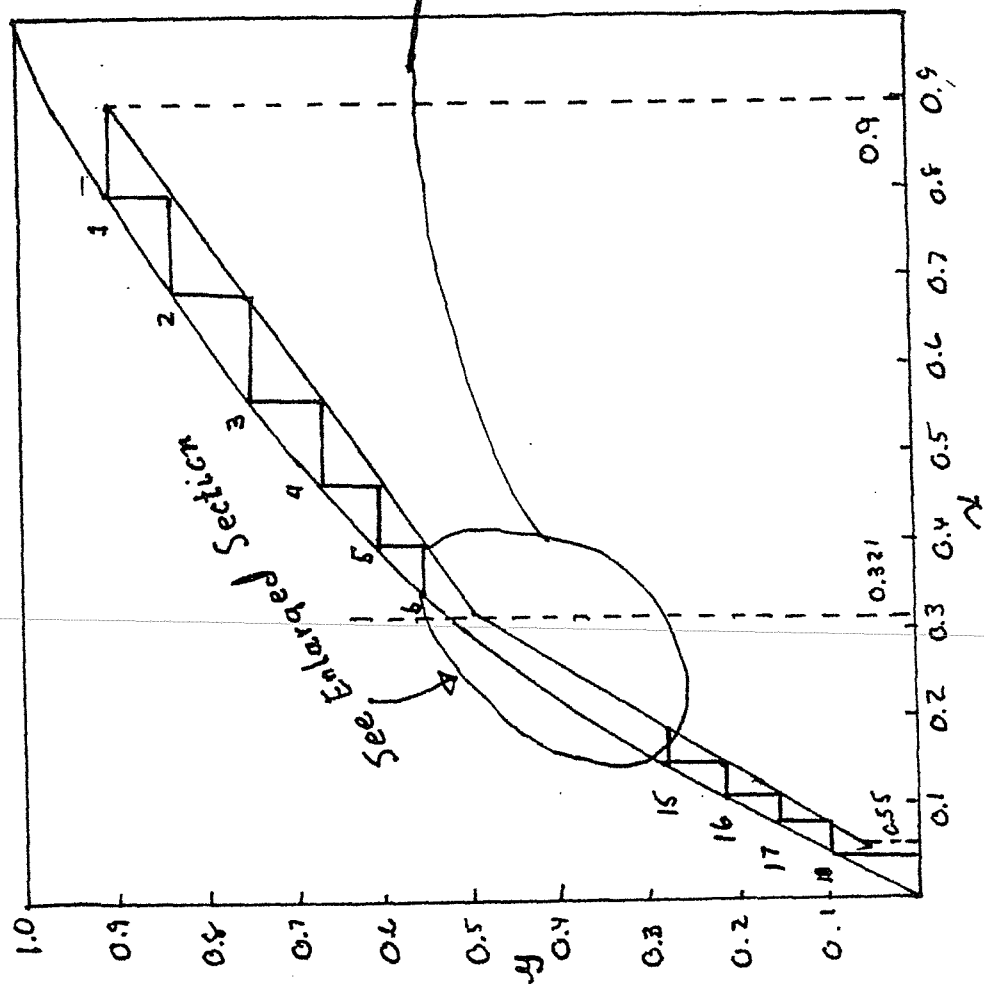
Current revenue =  $\frac{30 \text{ kmol D}}{\text{hr}} \times \frac{0.885 \text{ kmol Benzene}}{\text{kmol D}} \times \frac{78 \text{ kg Benzene}}{1 \text{ kmol Benzene}}$

$\frac{2.2 \text{ lb Benzene}}{1 \text{ kg Benzene}} \times \frac{24 \text{ hr}}{1 \text{ day}} \times \frac{330 \text{ day}}{1 \text{ yr}} \times \frac{\$0.20}{\text{lb Benzene}} = \$727,104/\text{yr}$

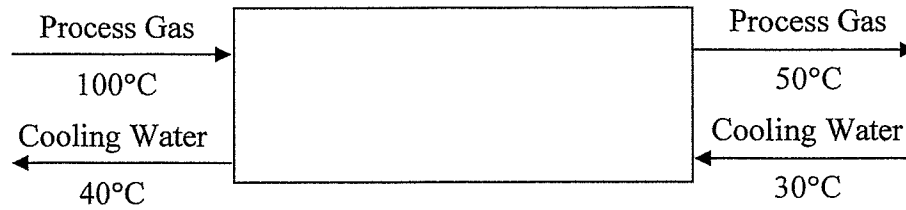
New revenue = (Same calculation with  $x_D = 0.8864$ )

Savings =  $\frac{(\$7.23 - \$7.22) / \text{yr} \times 10^6}{10^6} \approx \$1,066/\text{yr}$

# Problem 18:35



18.36



Design Conditions

$Q_1 = Q_2$  since  $M_{P2} = M_{P1}$  and process temperature remains the same

### Cooling Water Energy Balance

$$M_{CW1} C_{P1} (40^\circ\text{C} - 30^\circ\text{C}) = M_{CW2} C_{P2} (T - 30^\circ\text{C})$$

$$T = 30^\circ\text{C} + \left( \frac{M_{CW1}}{M_{CW2}} \right) (40^\circ\text{C} - 30^\circ\text{C}) = 30^\circ\text{C} + \frac{10}{1.5} = 36.667^\circ\text{C}$$

### Performance Equation

$$\frac{Q_2}{Q_1} = 1 = \frac{U_2 A_2 \Delta T_{LM2}}{U_1 A_1 \Delta T_{LM1}}$$

$$\frac{U_2}{U_1} = \frac{\left( \frac{1}{h_f} + \frac{1}{h_p} \right)^{-1}}{\left( \frac{1}{h_p} \right)^{-1}} = \frac{h_p h_f}{(h_p + h_f)(h_p)} = \frac{h_f/h_p}{1 + h_f/h_p} = \frac{H}{1 + H}; \quad \frac{A_2}{A_1} = 1$$

$$\Delta T_{LM2} = \frac{(100^\circ\text{C} - 36.667^\circ\text{C}) - (50^\circ\text{C} - 30^\circ\text{C})}{\ln \left( \frac{100^\circ\text{C} - 36.667^\circ\text{C}}{50^\circ\text{C} - 30^\circ\text{C}} \right)} = 37.59^\circ\text{C}$$

$$\Delta T_{LM1} = \frac{(100^\circ\text{C} - 40^\circ\text{C}) - (50^\circ\text{C} - 30^\circ\text{C})}{\ln \left( \frac{100^\circ\text{C} - 40^\circ\text{C}}{50^\circ\text{C} - 30^\circ\text{C}} \right)} = 36.41^\circ\text{C}$$

$$1 = \frac{H}{1 + H} \frac{37.59^\circ\text{C}}{36.41^\circ\text{C}} \Rightarrow 0.9685 = \frac{H}{1 + H}$$

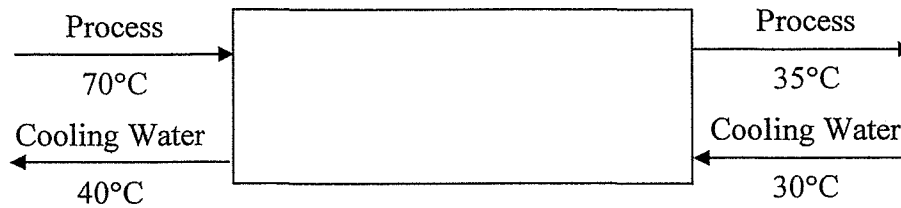
$$H = \frac{0.9685}{1 - 0.9685} = 30.74$$

$$h_f = 30.74 h_p$$

$$\text{if } h_p \approx 30 \text{ W/m}^2\text{K, } h_f \approx 920 \text{ W/m}^2\text{K}$$



18.37



Design Conditions

$$m_{p2} = 0.7m_{p1}$$

Process Stream

$$\frac{Q_2}{Q_1} = \frac{m_{p2}C_{p2}(70^\circ\text{C} - 35^\circ\text{C})}{m_{p1}C_{p1}(70^\circ\text{C} - 35^\circ\text{C})} = \frac{m_{p2}}{m_{p1}} = 0.7$$

Cooling Water

$$\frac{Q_2}{Q_1} = 0.7 = \frac{m_{cw2}C_{pw2}(t_2 - 30^\circ\text{C})}{m_{cw1}C_{pw1}(40^\circ\text{C} - 30^\circ\text{C})}$$

$$0.7 = M_{cw} \frac{t_2 - 30^\circ\text{C}}{10^\circ\text{C}}$$

Performance Equation

$$\frac{Q_2}{Q_1} = 0.7 = \frac{U_2A_2\Delta T_{LM2}}{U_1A_1\Delta T_{LM1}}$$

$$\frac{U_2}{U_1} = (0.7)^{0.6} = 0.8073$$

$$\Delta T_{LM2} = \frac{(70^\circ\text{C} - t_2) - (35^\circ\text{C} - 30^\circ\text{C})}{\ln\left(\frac{70^\circ\text{C} - t_2}{35^\circ\text{C} - 30^\circ\text{C}}\right)} = \frac{65^\circ\text{C} - t_2}{\ln\left(\frac{70^\circ\text{C} - t_2}{5^\circ\text{C}}\right)}$$

$$\Delta T_{LM1} = \frac{(70^\circ\text{C} - 40^\circ\text{C}) - (35^\circ\text{C} - 30^\circ\text{C})}{\ln\left(\frac{70^\circ\text{C} - 40^\circ\text{C}}{35^\circ\text{C} - 30^\circ\text{C}}\right)} = 13.95^\circ\text{C}$$

$$0.7 = \left( \frac{0.8073}{13.95^\circ\text{C}} \right) \frac{65^\circ\text{C} - t_2}{\ln(70^\circ\text{C} - t_2/5^\circ\text{C})}$$

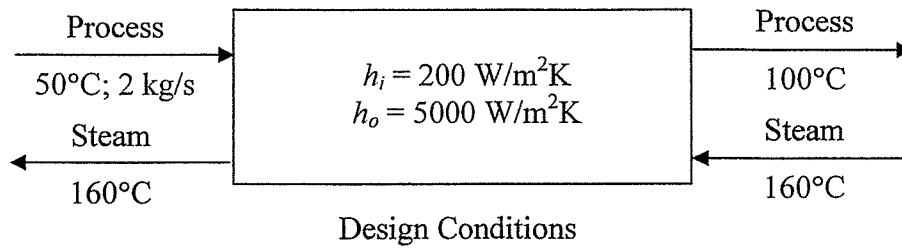
$$t_2 = 46.08^\circ\text{C}$$

$$(a) \quad M_{cw} = \frac{(0.7)(10^\circ\text{C})}{46.08^\circ\text{C} - 30^\circ\text{C}} = 0.435$$

Reduce  $M_{cw}$  to 43.5% of the design flowrate

$$(b) \quad t_2 = 46.08^\circ\text{C}$$

18.38

Steam

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_{steam2} \lambda_2}{\dot{m}_{steam1} \lambda_1} \Rightarrow Q = M_{steam} \quad (\text{Equation 1})$$

Process Stream

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_{p2} C_{p2} (100^\circ\text{C} - 50^\circ\text{C})}{\dot{m}_{p1} C_{p1} (100^\circ\text{C} - 50^\circ\text{C})} \Rightarrow Q = M_p = 1.15 \quad (\text{Equation 2})$$

Performance Equation

$$\frac{Q_2}{Q_1} = \frac{U_2 A_2 / \Delta T_{LM2}}{U_1 A_1 \Delta T_{LM1}}$$

$$\Delta T_{LM2} = \frac{(T - 50^\circ\text{C}) - (T - 100^\circ\text{C})}{\ln\left(\frac{T - 50^\circ\text{C}}{T - 100^\circ\text{C}}\right)}$$

$$\Delta T_{LM1} = \frac{(160^\circ\text{C} - 50^\circ\text{C}) - (160^\circ\text{C} - 100^\circ\text{C})}{\ln\left(\frac{160^\circ\text{C} - 50^\circ\text{C}}{160^\circ\text{C} - 100^\circ\text{C}}\right)} = 82.49^\circ\text{C}$$

$$\frac{\Delta T_{LM2}}{\Delta T_{LM1}} = \frac{(T - 50^\circ\text{C}) - (T - 100^\circ\text{C})}{(82.49^\circ\text{C}) \ln\left(\frac{T - 50^\circ\text{C}}{T - 100^\circ\text{C}}\right)} = \frac{50^\circ\text{C}}{(82.49^\circ\text{C}) \ln\left(\frac{T - 50^\circ\text{C}}{T - 100^\circ\text{C}}\right)}$$

$$\frac{1}{U_1} = \frac{1}{5000 \text{ W/m}^2\text{K}} + \frac{1}{200 \text{ W/m}^2\text{K}} \Rightarrow U_1 = 192.31 \text{ W/m}^2\text{K}$$

$$\frac{1}{U_2} = \frac{1}{5000 \text{ W/m}^2\text{K}} + \frac{1}{200(1.15)^{0.8}} \Rightarrow U_2 = 214.08 \text{ W/m}^2\text{K}$$

( $h_o = \text{constant}$ ,  $h_i = M_p^{0.8}$ )

$$\frac{U_2}{U_1} = \frac{214.08 \text{ W/m}^2\text{K}}{192.31 \text{ W/m}^2\text{K}} = 1.113$$

$$Q = (1.113) \left( \frac{50^\circ\text{C}}{(82.49^\circ\text{C}) \ln(T - 50^\circ\text{C} / T - 100^\circ\text{C})} \right)$$

$$Q = \frac{0.6746}{\ln\left(\frac{T - 50^\circ\text{C}}{T - 100^\circ\text{C}}\right)} \quad (\text{Equation 3})$$

Combine Equation 2 and 3

$$1.15 = \frac{0.6746}{\ln\left(\frac{T - 50^\circ\text{C}}{T - 100^\circ\text{C}}\right)}$$

$$T = 162.7^\circ\text{C}$$

## Chapter 19

- 19.1 From Figure B .10.1, it can be seen that the molten salt is heated by a fired heater in the circulating loop. This implies that heat is added to the reactor, i.e., the reaction is endothermic. At start-up, if the catalyst activity is lower than designed then the amount of reaction taking place will be less and consequently the amount of heat needed to be added to the reactor should increase in order to increase the temperature and the conversion. Therefore, the flow of circulating molten salt should be increased in order to compensate for the higher heat load.

- 19.2 Pressure at the bottom of a column decreases.

$$P_{bot} = P_{top} + \Delta P_{trays}$$

The pressure drop across the column is due to the pressure drop caused by the gas flowing upwards through the layer of liquid on the trays. This pressure drop is due mainly to the weir height which will not change. Thus, the pressure at the top of the column will also decrease.

As pressure decreases both the top and bottom temperatures also decrease in keeping with Antoine's equation.

As the pressure drop decreases, the vapor density will decrease and the superficial velocity will increase – this will tend to increase the tendency to flood. As the temperature at the bottom of the column decreases the driving force in the reboiler will increase causing more vapor to move up the column and again increasing the tendency to flood (although more steam would have to flow into the reboiler).

19.3 given feed is actually  $\approx 7.7$  wt% propane  
 total  $G_s = 4897.516$  kg, which is held  
 constant

|                              |          |          |          |          |          |          |
|------------------------------|----------|----------|----------|----------|----------|----------|
| wt%<br>propane $\rightarrow$ | 5%       | 6%       | 7%       | 8%       | 9%       | 10%      |
| propylene                    | 4652.640 | 4603.665 | 4554.690 | 4505.715 | 4456.740 | 4407.764 |
| propane                      | 244.876  | 243.851  | 342.826  | 391.801  | 440.776  | 489.752  |

above values are kg/h Feed to rxr

Note: This problem is not to be confused with Project 5, App. C  
 and problem 1.1 - there the propane feed impurity is to  
 the process, and the recycle stream affects the  
 rxr feed. In this problem, the reactor is  
 analyzed alone.

Kinetics used - only one reaction



$$r = k C_B C_P$$

$$k = 3500 \exp \left[ \frac{-13,280 \text{ cal/mol}}{R T} \right]$$

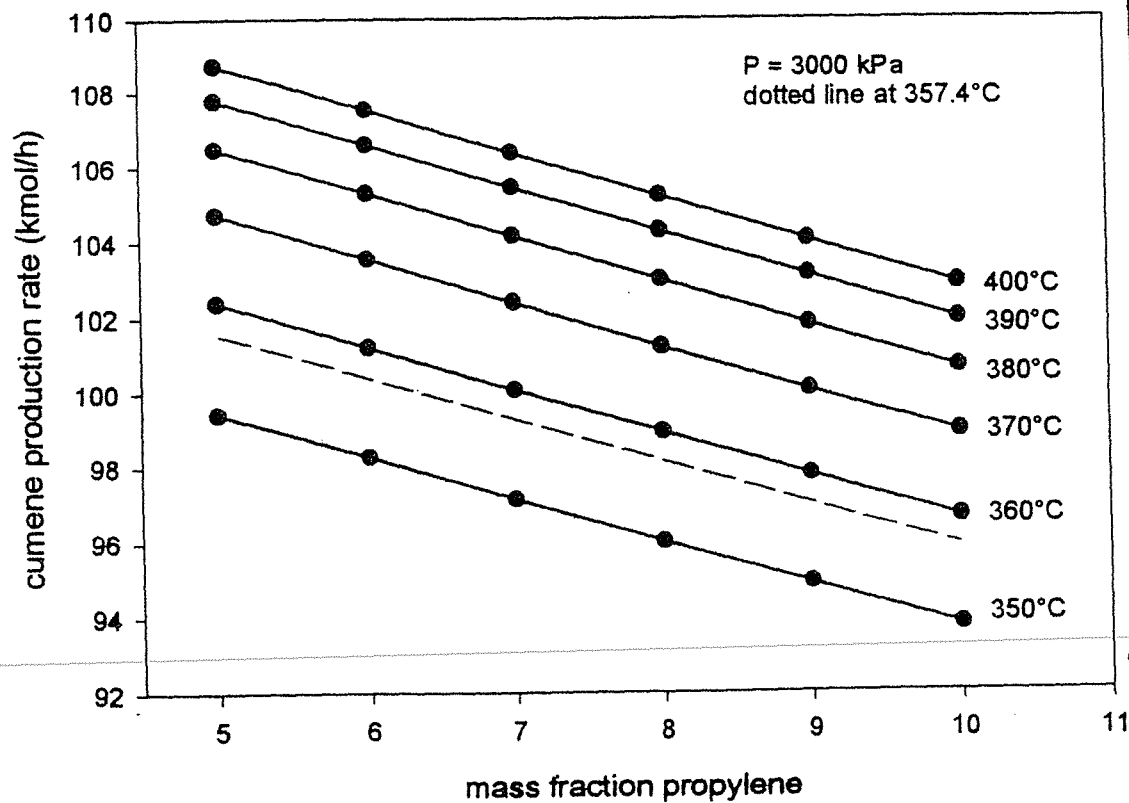
from Example 1.1 - max  $P \approx 4800$  kPa

corresponds to  $T = 357.4^\circ\text{C}$  (by interpolation)

results of Chemical simulation using SPSS Thermo  
cumene production rate kmol/h

wt %  
propylene

|    |       |        |        |        |        |        |
|----|-------|--------|--------|--------|--------|--------|
| 5  | 99.42 | 102.39 | 104.72 | 106.49 | 107.79 | 108.73 |
| 6  | 98.28 | 101.24 | 103.56 | 105.32 | 106.62 | 107.56 |
| 7  | 97.15 | 100.09 | 102.40 | 104.16 | 105.46 | 106.39 |
| 8  | 96.02 | 98.95  | 101.24 | 103.00 | 104.29 | 105.23 |
| 9  | 94.90 | 97.80  | 100.09 | 101.84 | 103.13 | 104.06 |
| 10 | 93.77 | 96.66  | 98.94  | 100.68 | 101.97 | 102.90 |
|    | 350°C | 360°C  | 370°C  | 380°C  | 390°C  | 400°C  |



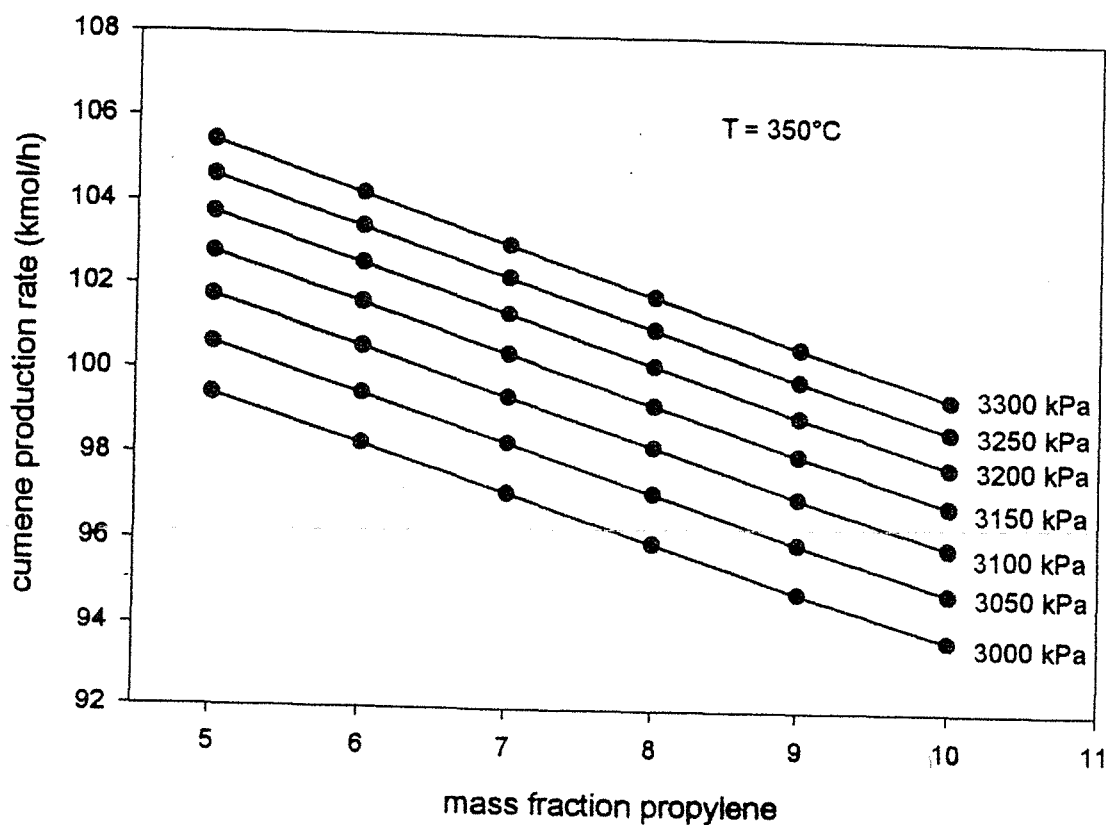
19.4 see discussion with Problem 19.3

cumene production rate kmol/h

|    |       |        |        |        |        |        |        |
|----|-------|--------|--------|--------|--------|--------|--------|
|    |       |        |        |        |        |        |        |
| 5  | 99.42 | 100.62 | 101.74 | 102.79 | 103.75 | 104.63 | 105.44 |
| 6  | 98.28 | 99.47  | 100.58 | 101.61 | 102.56 | 103.44 | 104.24 |
| 7  | 97.15 | 98.33  | 99.42  | 100.44 | 101.38 | 102.25 | 103.04 |
| 8  | 96.02 | 97.18  | 98.27  | 99.27  | 100.21 | 101.06 | 101.84 |
| 9  | 94.90 | 96.04  | 97.11  | 98.11  | 99.03  | 99.88  | 100.65 |
| 10 | 93.77 | 94.91  | 95.97  | 96.95  | 97.86  | 98.70  | 99.47  |

$\uparrow$  3000 3050 3100 3150 3200 3250 3300  
 $P_{\text{H}_2}$  kPa

propene  
at 10





19.5

This problem is solved using the Chemcad process Simulator. SRK thermodynamics is used. The column is assumed to operate at 2.5 bar.

Upon fixing the boil-up rate at 189.53 kmol/h with the feed tray and total number of trays fixed, a 4<sup>th</sup> specification is needed for the scaled-down column. There are several possibilities. The one that was used was  $D = 17.16$  kmol/h, exactly 50% scale-down of the distillate.

Results:  $D = 17.16$  kmol/h  $F = 70.65$  kmol/h  
 $B = 53.49$  kmol/h (exactly 50% of the original feed)

$L_0 = 168.63$  kmol/h  $RR = 9.827$  (reflux ratio)

$V_{N+1} = 189.53$  kmol/h

$L_N = 242.99$  kmol/h

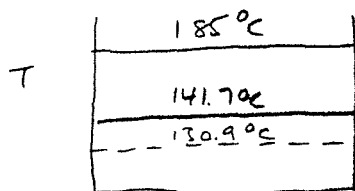
$x_{BD} = 0.999$  benzene mole fr. in top

$x_{BB} = 0.00676$  benzene mole fr. in bottom

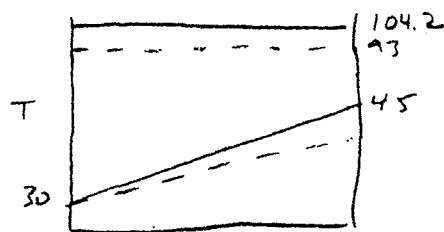
Top  $T = 112.6$  °C } not much change  
Bot  $T = 145.4$  °C } from original

19.6

1 = old ———  
2 = new - - - -



Q  
reboiler



Q  
condenser

$$\frac{1}{U} \approx \frac{1}{h_o} + \frac{1}{h_i} \quad h_o \approx h_i \text{ initially}$$

$$\therefore \frac{U_2}{U_1} = \frac{2}{1 + \frac{1}{M^{0.8}}} \quad M = \frac{\dot{m}_{cw2}}{\dot{m}_{cw1}}$$

reboiler

$$\frac{Q_2}{Q_1} = 1.25 = \frac{U_2}{U_1} \frac{A}{A} \frac{\Delta T_2}{\Delta T_1} = \frac{\Delta T_2}{43.3}$$

$$\Delta T_2 = 54.125^\circ\text{C}$$

$$\Rightarrow T = 130.9^\circ\text{C}$$

could increase steam T  
assume steam T constant here

bottom is toluene  
(assume  $\approx$  pure)

$$P_{\text{bot}} = 174 \text{ kPa} \rightarrow P_{\text{tol}}^* \text{ at } 130.9^\circ\text{C}$$

$$\Delta P = \text{const} = 22.4 \text{ kPa}$$

$$P_{\text{top}} = 151.6 \text{ kPa}$$

vapor pressure  
expression in  
Sect. 13.2

assume top is pure  
benzene

$$T_{\text{top}} = 93^\circ\text{C}$$

$$Q = \dot{m}_{cw} c_p \Delta T \Rightarrow \frac{Q_2}{Q_1} = 1.25 = M \frac{T - 30}{15}$$

$$Q = U A \Delta T_{\text{lm}} \Rightarrow \frac{Q_2}{Q_1} = 1.25 = \frac{2(T - 30)}{\left(1 + \frac{1}{M^{0.8}}\right) (66.42) \ln \frac{63}{93 - T}}$$

$\tau_{\text{old } \Delta T_{\text{lm}}}$

$$M = 2.82$$

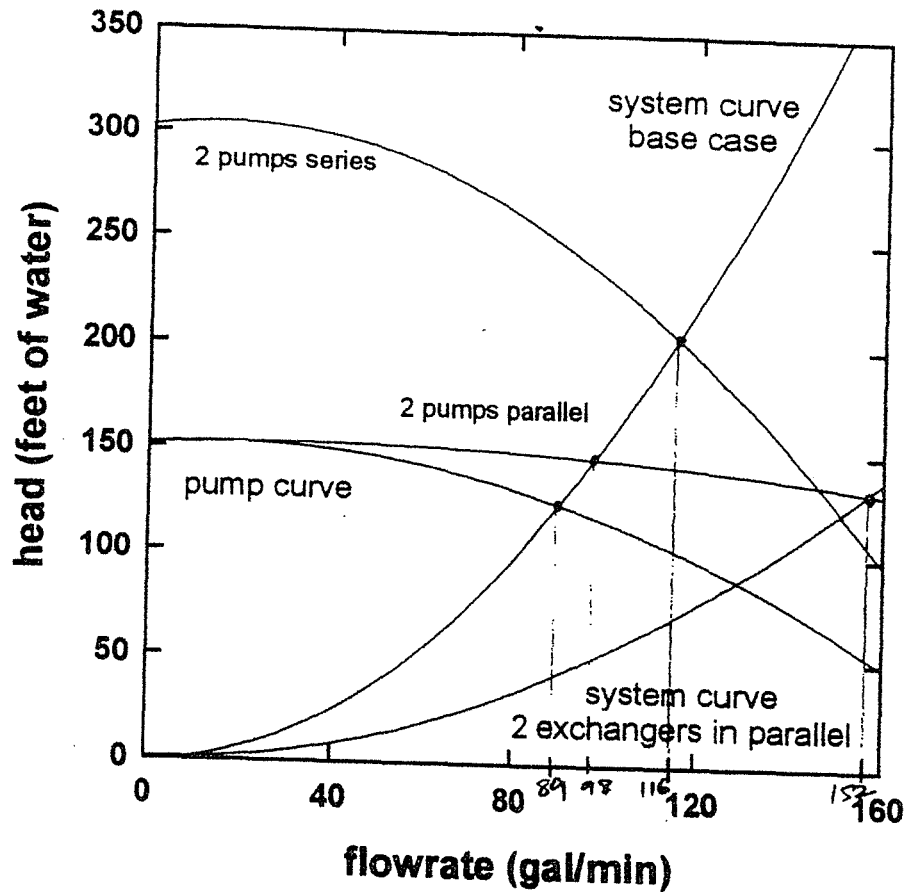
$$T = 36.6^\circ\text{C}$$

Worry about  $v > 10 \text{ m/s}$  and erosion

19.7

Many alternation listed after example in chapter

19.8



| max Flowrates (gal/min) |            | HX configuration |             |
|-------------------------|------------|------------------|-------------|
| (90 scale-up)           |            | base             | parallel    |
| pumps                   | 2 parallel | 98 (15.3%)       | 170 (100%)  |
|                         | 2 series   | 116 (36.5%)      | 158 (85.9%) |

# 19.8 cont'd

results of analyses similar to Example 17.8

Use  $E_{ss}$  FTB.8a  
E 17.8b  
E 17.8c

## Pump configuration

UX  $E_{ss}$   
17.8a  
17.8b  
17.8c  
17.8d

|                          | 2 parallel |             | 2 series |             |
|--------------------------|------------|-------------|----------|-------------|
|                          | HX base    | HX parallel | HX base  | HX parallel |
| $T_2(^{\circ}\text{C})$  | 346.1      | 329.8       | 341.2    | 332.0       |
| $T_4(^{\circ}\text{C})$  | 392.4      | 363.1       | 383.4    | 367.0       |
| flow % scale-up          | 15.3       | 100         | 36.5     | 85.9        |
| $\alpha$ scale-up factor | 1.07       | 1.33        | 1.15     | 1.30        |



19.9 R<sub>xx</sub>  $Q = \dot{m}_D C_p (T_4 - T_3)$  (1)

R<sub>xx</sub> HX  $Q = U A \Delta T_{lm}$  (2)

Cooler HX  $Q = U A \Delta T_{lm}$  (3)

Cooler  $Q = \dot{m}_{cw} C_p (45 - 30)$  (4)

from (1)  $\frac{Q_2}{Q_1} = \frac{M (T_4 - T_3)}{50}$   $M = \frac{\dot{m}_{D2}}{\dot{m}_{D1}}$  Doublen flow rate

from (2)  $\frac{Q_2}{Q_1} = \frac{5 (T_4 - T_3)}{130.44 \left( 4 + \frac{1}{M \cdot 0.8} \right) \ln \frac{510 - T_3}{510 - T_4}}$

from (3)  $\frac{Q_2}{Q_1} = \frac{2 [(T_3 - 30) - (T_4 - 45)]}{337.2 \left[ \frac{1}{M \cdot 0.8} + \frac{1}{M_{cw} \cdot 0.6} \right] \ln \frac{T_3 - 30}{T_4 - 45}}$

Since equal  
resistances  
in CW exchanger

from (4)  $\frac{Q_2}{Q_1} = M_{cw}$  Since maintain  $\Delta T_{cw} = 15^\circ C$   
i.e. supply flow rate to fix outlet  $T = 45^\circ C$

$\frac{1}{U} \approx \frac{1}{h_i} + \frac{1}{h_o}$

$\frac{1}{\frac{U_2}{h_{i2}}} = \frac{U_2}{h_{i2}} = \frac{\frac{1}{h_{i1}} + \frac{1}{h_{o1}}}{\frac{1}{h_{i2}} + \frac{1}{h_{o2}}} = \frac{\frac{2}{h_{i1}}}{\frac{1}{h_{i1} M \cdot 0.8} + \frac{1}{h_{i1} M_{cw} \cdot 0.6}} = \frac{2}{\frac{1}{M \cdot 0.8} + \frac{1}{M_{cw} \cdot 0.6}}$

if  $M = 1.05$

$Q = M_{cw} = 1.02$

$T_3 = 349.0^\circ C$

$T_4 = 397.7^\circ C$

if  $M = 1.588$

$Q = M_{cw} = 1.20$

$T_3 = 340.0^\circ C$

$T_4 = 377.8^\circ C$

alternative soln - fix air rate and let  $T_{cw}$  out vary

19.10

solution strategy similar to 17.6 and 17.7

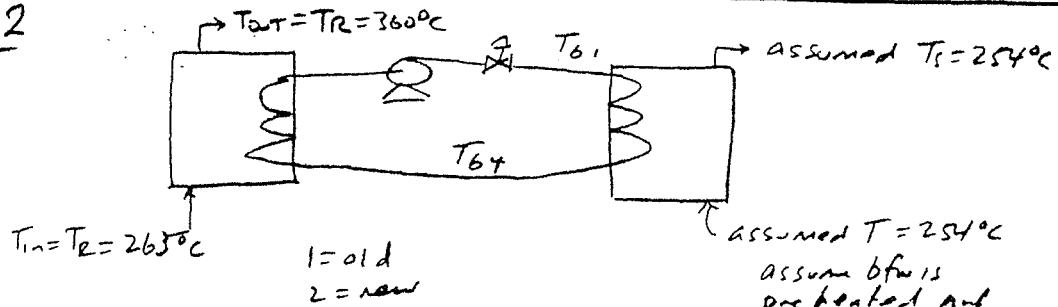
|                         | Pumps 2 parallel |                                                                                    | Pumps 2 series |             |
|-------------------------|------------------|------------------------------------------------------------------------------------|----------------|-------------|
|                         | HX base          | HX parallel                                                                        | HX base        | HX parallel |
| $T_3(^{\circ}\text{C})$ | 343.3            | 327.2                                                                              | 335.5          | 330.6       |
| $T_4(^{\circ}\text{C})$ | 388.4            | 358.4                                                                              | 375.4          | 363.7       |
| flow %<br>Scale-ups     | 15.3             | 100                                                                                | 36.5           | 85.9        |
| Q scale up<br>factor    | 1.05             | 1.25                                                                               | 1.09           | 1.2         |
|                         |                  |  |                |             |

19.11

lower cooling medium T  
use CW or RW and increase flow

lower reaction T, if possible

19/12



rxr

$$Q = \dot{m}_R \left[ (MW) DH_R \right]_{T_R} = \dot{m}_S \lambda$$

↑  
process  
stream
↑  
steam
← neglects sensible heat  
to DH<sub>rxn</sub>

⇒ Q,  $\dot{m}_R$ ,  $\dot{m}_S$  scale identically

$$\frac{Q_2}{Q_1} = M_R = \frac{\dot{m}_{R2}}{\dot{m}_{R1}} = M_S = \frac{\dot{m}_{S2}}{\dot{m}_{S1}}$$

downterm

$$Q = \dot{m} c_p (T_{64} - T_{61}) \Rightarrow \frac{Q_2}{Q_1} = M \frac{T_{64} - T_{61}}{85} \quad (1)$$

$\frac{\dot{m}_2}{\dot{m}_1}$  water salt flow ratio

rxr

$$Q = U_R A_m \frac{(360 - T_{61}) - (360 - T_{64})}{\ln \frac{360 - T_{61}}{360 - T_{64}}} \Rightarrow \frac{Q_2}{Q_1} = \frac{M_R^{0.6}}{37.756} \frac{T_{64} - T_{61}}{\ln \frac{360 - T_{61}}{360 - T_{64}}} \quad (2)$$

$U \propto M_R^{0.6}$  since shell-side controls

steam generator

$$Q = U_S A_s \frac{(T_{64} - 254) - (T_{61} - 254)}{\ln \frac{T_{64} - 254}{T_{61} - 254}} \Rightarrow \frac{Q_2}{Q_1} = \frac{M^{0.8}}{39.23} \frac{T_{64} - T_{61}}{\ln \frac{T_{64} - 254}{T_{61} - 254}} \quad (3)$$

$U_S \propto m^{0.8}$  since salt in tubes  
↳ controlling resistance

19.12 cont'd

pick a value for  $\frac{Q_2}{Q_1}$  which is  $M_R$  and  $M_S$

this is scale factor for process  $\rightarrow$  heat load  $\rightarrow$  steam production

solve ①, ②, and ③ for  $T_{04}$ ,  $T_{01}$ ,  $M$

results

process scale factor

molar salt  
scale factor

| $\frac{Q_2}{Q_1}$ | $T_{01} (^{\circ}\text{C})$ | $T_{04} (^{\circ}\text{C})$ | $M$  |
|-------------------|-----------------------------|-----------------------------|------|
| 0.50              | 356.4                       | 262.0                       | 0.45 |
| 0.75              | 353.3                       | 263.8                       | 0.71 |
| 1.00              | 350.0                       | 265.0                       | 1.00 |
| 1.25              | 346.7                       | 265.9                       | 1.32 |
| 1.50              | 343.5                       | 266.6                       | 1.66 |



19.13

$$a) \quad \dot{m}_1 = 15,500 \frac{\text{kg}}{\text{h}} = 15.5 \text{ Mg/h}$$

$$\Delta P_{\text{pump 1}} = 340 - 0.913(15.5) + 0.0535(15.5)^2 - 0.0101(15.5)^3$$

$$\Delta P = 301 \text{ kPa}$$

$$\text{if } P_1 = 100 \text{ kPa}$$

$$P_2 = 401 \text{ kPa}$$

$$\text{since } P_3 = 225 \text{ kPa}$$

$$\Delta P_{23} = 176 \text{ kPa}$$

$$\text{since } P_3 = P_7 = 225 \text{ kPa}$$

$$P_6 = 250 \text{ kPa}$$

$$P_5 = 375 \text{ kPa}$$

$$P_4 = 100 \text{ kPa}$$

$$\text{ii } \Delta P_{\text{pump 2}} = 275 \text{ kPa}$$

$$275 = 312 - 1.924 \dot{m} + 0.0302 \dot{m}^2 - 0.0112 \dot{m}^3$$

$$\dot{m}' = 11.8 \text{ Mg/h}$$

19.14

$$a) \quad m_7 = 100 \frac{\text{Mg}}{\text{h}} \quad P_{\text{out}} = 101 \text{ kPa}$$

$$\Rightarrow \frac{P_{\text{out}}}{P_{\text{in}}} = 2.7034 \Rightarrow P_{\text{in}} = 37.36 \text{ kPa}$$

$$P(\text{entrance to R-101}) = 77.36 \text{ kPa}$$

$$P(\text{junction of 3+4} \rightarrow 5) = 97.36 \text{ kPa}$$

$$P(\text{TK-101}) = 127.36 \text{ kPa} \Rightarrow P = 127 \text{ kPa}$$

$$b) \quad P(\text{entrance to E-101}) = 97.36 + 70 \left( \frac{m_2}{4} \right)^2$$

$\Delta P \propto m^2$

$$P(\text{exit of P-101}) = 97.36 + (70 + 40) \left( \frac{m_2}{4} \right)^2$$

subtract above from below to get  $\Delta P$  pump 101

$$P(\text{entrance of P-101}) = 101 \text{ kPa}$$

$$\Delta P(P-101) = -3.64 + \frac{110}{16} m_2^2 = 500 + 4.662 m_2^2 - 1.805 m_2^2$$

$\uparrow$  DP from pump curve

$$\Rightarrow m_2 = 7.89 \text{ Mg/h}$$

c)  $P$  at 3+4=5 junction is always 97.36 kPa

$$m_7 = 100 \frac{\text{Mg}}{\text{h}} \text{ always}$$

19.15

a)  $\frac{\Delta P_{\text{new}}}{\Delta P_{\text{old}}} = \left( \frac{\dot{V}_{\text{new}}}{\dot{V}_{\text{old}}} \right)^2$  assuming turbulent flow

$$\Delta P_{\text{new}} = 5 \left( \frac{0.5 \dot{V}_{\text{old}}}{\dot{V}_{\text{old}}} \right)^2 = \underline{1.25 \text{ psi}}$$

b)  $\dot{V}_{\text{new}} = \text{new flow rate}$

$$\Delta P_{\text{fr, piping}} = 5 \left( \frac{\dot{V}_{\text{new}}}{100} \right)^2 \text{ psi}$$

$$\Delta P_{\text{fr, exch}} = 1.25 \text{ psi from part a}$$

$$\Delta P_{\text{pump}} = 32.72 \left[ 1 - \left( \frac{\dot{V}_{\text{new}}}{120} \right)^2 \right] \text{ psi}$$

since valve is wide open,  $\Delta P_{\text{cv}} = 0$  also  $\rho g \Delta h = 0$

$$\therefore \Delta P_{\text{pump}} = \Delta P_{\text{fr, exch}} + \Delta P_{\text{fr, piping}}$$

$$32.72 \left[ 1 - \left( \frac{\dot{V}_{\text{new}}}{120} \right)^2 \right] = 5 \left( \frac{\dot{V}_{\text{new}}}{100} \right)^2 + 1.25$$

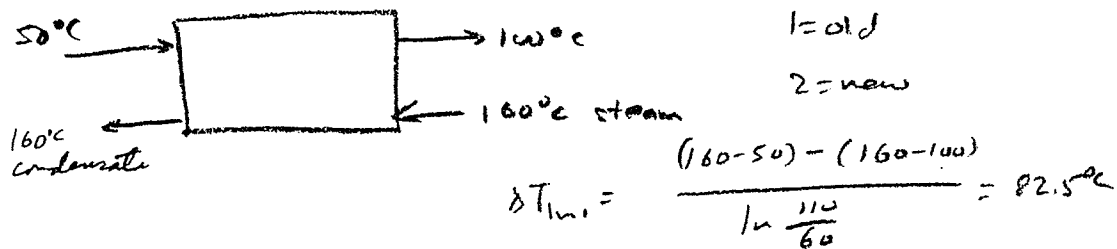
$$\text{Solving } \underline{\dot{V}_{\text{new}} = 106.5 \text{ gpm}}$$

Can be solved graphically - but since eqn for pump curve given - this method is more accurate

Note:

This addition to Problem 19.15 will appear in subsequent printings of this edition.

- c. Assume that all the resistance to heat transfer is on the process fluid side and that the temperature of the condensing steam remains constant. Determine the temperature of the fluid entering Tank 2.



$$\dot{m}_2 = \frac{\dot{m}_1}{2} \text{ for process stream}$$

$$\frac{Q_2}{Q_1} = Q = \frac{\dot{m}_2 c_p}{\dot{m}_1 c_p} \frac{\Delta T_2}{\Delta T_1} \text{ for process stream}$$

$$Q = 0.5 \frac{T - 50}{100} \quad (1)$$

T = outlet process temp  
Now unknown

performance eq

$$\frac{Q_2}{Q_1} = Q = \frac{U_2 A}{U_1 A} \frac{\Delta T_{lm2}}{\Delta T_{lm1}}$$

$$\frac{U_2}{U_1} \propto \left(\frac{\dot{m}_2}{\dot{m}_1}\right)^{0.8} = 0.5^{0.8} \text{ assuming all resistance on process side}$$

$$Q = (0.5)^{0.8} \frac{(160 - 50) - (160 - T)}{82.5 \ln \frac{110}{160 - T}} \quad (2)$$

Solve ① & ② simultaneously for  $T$  &  $Q$

or - this one can be solved analytically by equating  $Q$  in each equation

$$0.5 \frac{\cancel{T_{50}}}{100} = \frac{(0.5)^{0.8}}{82.5} \frac{\cancel{T_{50}}}{\ln \frac{110}{160-T}}$$

$$\frac{110}{160-T} = \exp[0.6962] = 2.006$$

$$\Rightarrow T = 105.2^\circ\text{C}$$

$$\text{also} - Q = 0.552$$

19.16

- a) need system curve  
intercept = 0 since closed loop  
from data provided

$$\Delta P_{fr} = \Delta P_{1-2} + \Delta P_{2-3} + \Delta P_{3-4} + \Delta P_{4-5} + \Delta P_{5-6} = \Delta P_{1-6}$$

$$= 10 + 40 + 10 + 40 + 30 = 130 \text{ kPa}$$

look on pump curve at 130 kPa  $\Rightarrow \dot{V} \approx 1.1 \text{ L/s}$   
answer is on pump curve because there was valve so  
operating at maximum flow rate

- b) for new case -

$$\frac{\Delta P_{exch, \text{new}}}{\Delta P_{exch, \text{old}}} = \left( \frac{\dot{V}_{\text{new}}}{\dot{V}_{\text{old}}} \right)^2 \Rightarrow \Delta P_{exch, \text{new}} = 40 \left( \frac{1}{0.6} \right)^2$$

$$= 111 \text{ kPa}$$

40 kPa  
for case

$$\Rightarrow 222 \text{ kPa } \Delta P \text{ for both exchangers}$$

$$\Delta P_{fr, \text{new}} = 222 + \underbrace{\Delta P_{1-2} + \Delta P_{3-4} + \Delta P_{5-6}}_{\text{all unchanged}} = 272 \text{ kPa}$$

if it was possible - flow would now be through pump at 1.1 L/s  
with  $\Delta P_{fr} = 272 \text{ kPa}$

so this defines new system curve

$$\Delta P_{sys} = A \dot{V}^2 \Rightarrow \Delta P_{sys, \text{new}} = 272 \left( \frac{\dot{V}}{1.1} \right)^2$$

$$272 = A (1.1)^2$$

$$A = 224.79$$

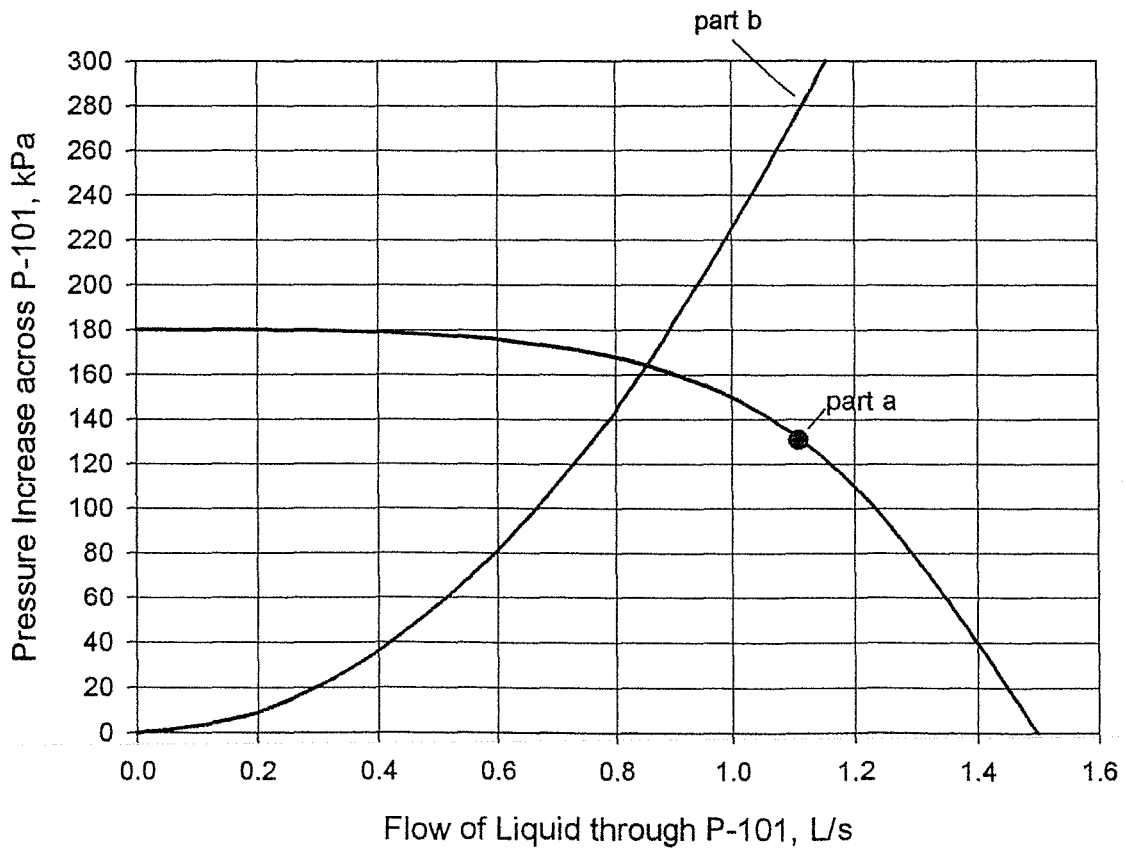
plotted on graph on next page

points on plot

| <u>flow (L/s)</u> | <u><math>\Delta P</math> (kPa)</u> |
|-------------------|------------------------------------|
| 1.0               | 225                                |
| 0.8               | 144                                |
| 0.6               | 81                                 |
| 0.4               | 36                                 |
| 0.2               | 9                                  |
| 0                 | 0                                  |

$$\Rightarrow \dot{V} = 0.85 \text{ L/s}$$

max flow since no value



Solution to Problem 17.16

$$19.17 \quad \Delta P_{out-in} = 20 - 21 = -1 \text{ psi}$$

a)

$$p_{gsh} = 32 - 8 = 24 \text{ ft} \frac{1 \text{ psi}}{2 \text{ ft}} = 12 \text{ psi}$$

$$\Rightarrow \Delta P_{out-in} + p_{gsh} = \text{interception system curve} = 11 \text{ psi}$$

at design flow rate = 50 gpm

$$\Delta P_{pump} = 34 \text{ psi}$$

$$\Delta P_{cu} = 10 \text{ psi}$$

$$\text{so } \Delta P_{pump} = \Delta P_{out-in} + p_{gsh} + \Delta P_{fr} + \Delta P_{cu}$$

$$34 = 11 + \Delta P_{fr} + 10$$

$$\Delta P_{fr} = 13 \text{ psi}$$

$$\text{eqn for system curve} \quad \Delta P_{sy} = 11 + 13 \left( \frac{\dot{V}}{50} \right)^2$$

$$= 11 + 0.0052 \dot{V}^2$$

plot on next page max  $\dot{V} = 60 \text{ gpm}$

$$b) \text{ at } 50 \text{ gpm } NPSH_R = 11 \text{ ft} = 5.5 \text{ psi}$$

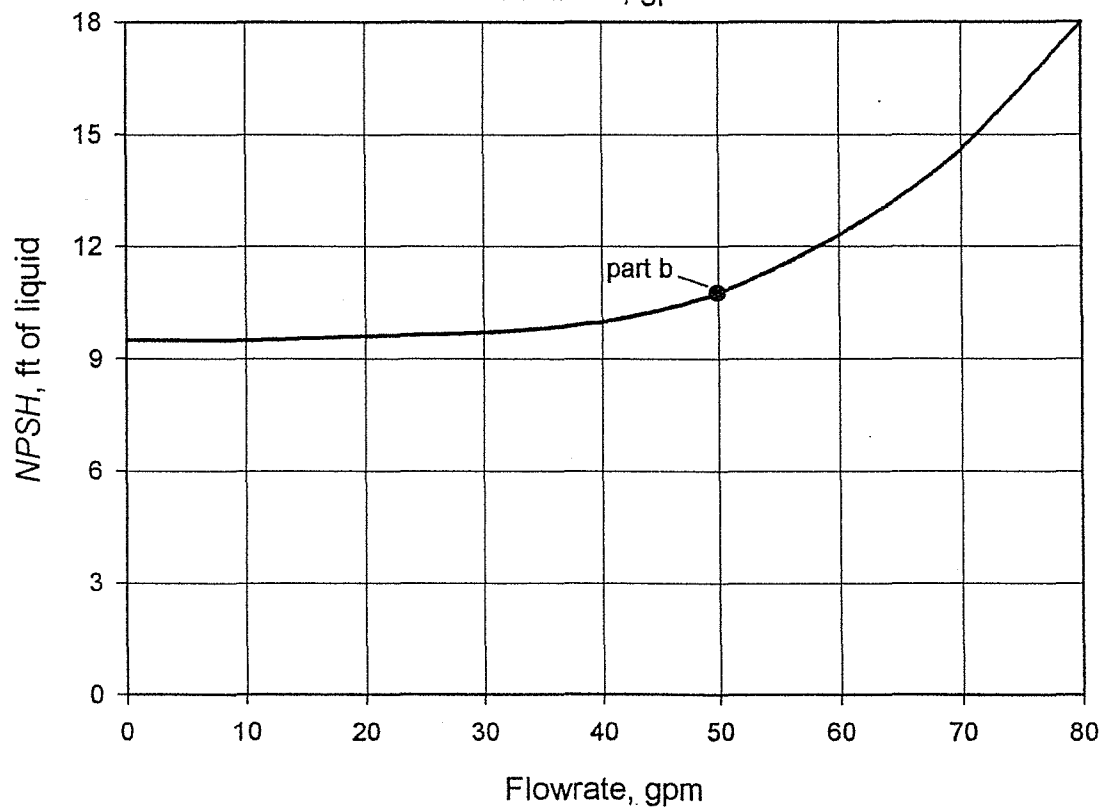
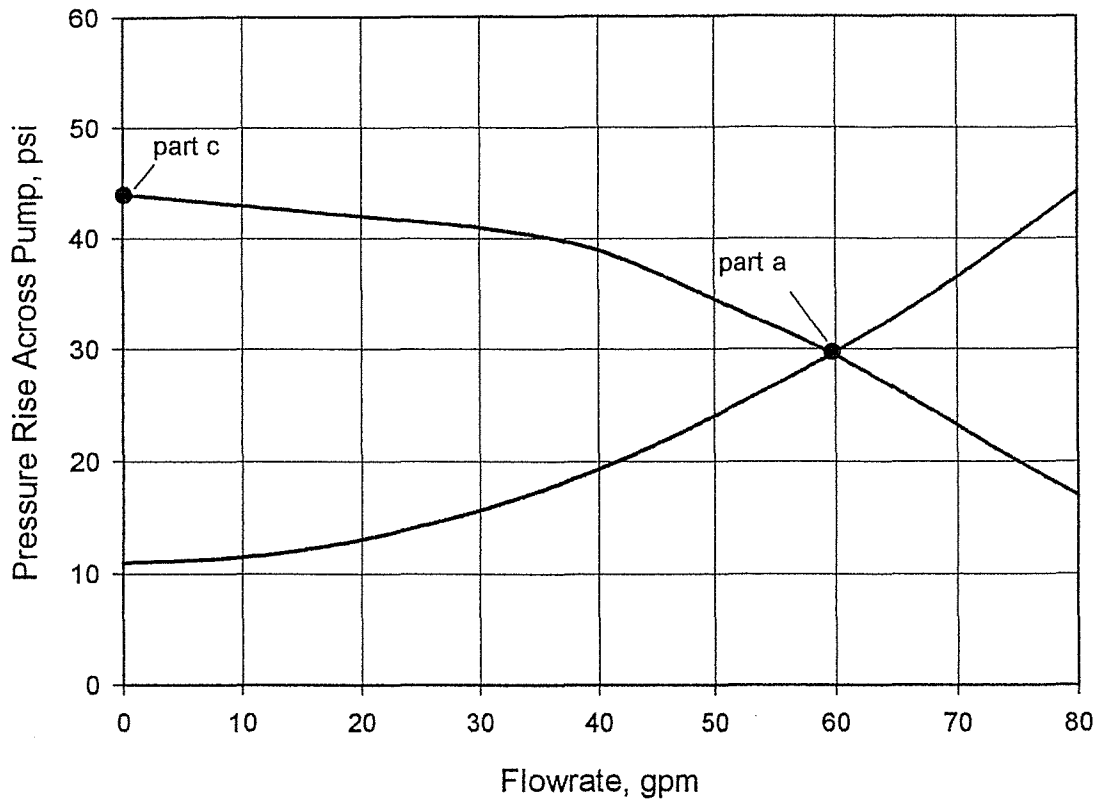
$$NPSH_A = P_{TANK} + p_{gsh} - \Delta P_{fr} - P^*$$

$$= 21 \text{ psi} + \frac{8 \text{ ft}}{2 \frac{\text{ft}}{\text{psi}}} - 1 - P^* = 24 - P^*$$

$$\text{if just cavitating } 24 - P^* = NPSH_R = 5.5 \text{ psi}$$

$$P^* = 18.5 \text{ psi}$$





Solution to Problem 19.17

c) if Cv closed - no flow

from pump curve  $\Delta P_{\text{pump}} = 44 \text{ psi}$

$$\begin{aligned} P_{\text{at E-301 inlet}} &= P_{\text{TANK}} + \Delta P_{\text{static}} + \Delta P_{\text{pump}} \\ &= 21 + \underset{\substack{\uparrow \\ \frac{8 \text{ ft}}{2 \text{ ft}} \\ \text{psi}}}{4} + 44 = 69 \text{ psi} \end{aligned}$$

$$\Rightarrow \underline{P_{\text{des}} \geq 69 \text{ psi}}$$

19.18 (a) Acrylic Acid reboiled at  $89^{\circ}\text{C} = 362\text{ K}$

$$\ln P_{acrylic}^* = 17.31 - \frac{5264.4}{T(\text{K})} = 17.31 - \frac{5264.4}{362\text{K}} = 15.92\text{ kPa}$$

$$P_{acrylic}^* = P_{bottom} = 15.92\text{ kPa}$$

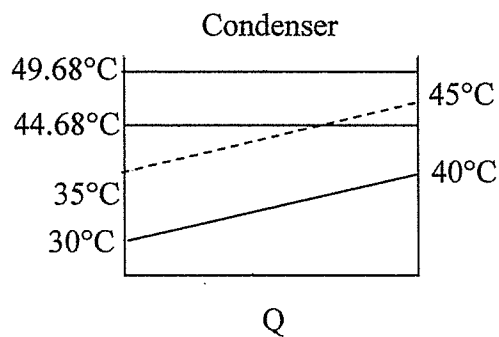
(b)  $P_{top} = P_{bottom} - \Delta P$   
 $P_{top} = 15.92\text{ kPa} - 9.5\text{ kPa} = 6.42\text{ kPa}$

$$\ln P_{acetic}^* = 17.54 - \frac{4981.5}{T(\text{K})}$$

$$\ln(6.42\text{ kPa}) = 17.54 - \frac{4981.5}{T}$$

$$T = 317.68\text{ K} = 44.68^{\circ}\text{C}$$

(c) T-Q Diagrams



$$Q = UA\Delta T_{lm}$$

$$Q = \dot{m}_{cw} C_{P,cw} \Delta T_{cw}$$

$$Q = \dot{m}_p \lambda_p$$

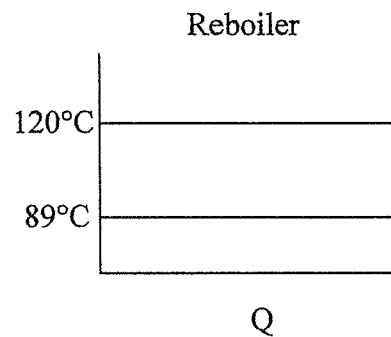
If  $Q$  and  $\dot{m}_{cw}$  are constant

$$\Delta T_{cw} = \text{constant} \Rightarrow T_{in} = 35^{\circ}\text{C}$$

$$T_{out} = 45^{\circ}\text{C}$$

$UA$  is constant  $\Rightarrow \Delta T_{lm}$  is constant

$$T_{cond} = 49.68^{\circ}\text{C} \Rightarrow P_{top} = 8.18\text{ kPa}$$



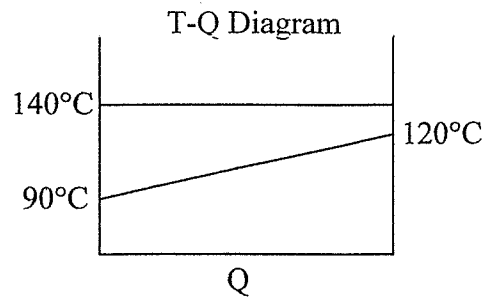
$$P_{bottom} = 8.18\text{ kPa} + 9.5\text{ kPa} = 17.68\text{ kPa}$$

$$T_{bottom} = 91.6^{\circ}\text{C}$$

Since  $Q = UA\Delta T$  and  $Q$  and  $UA$  are constant

$$T_{steam} = 122.6^{\circ}\text{C}$$

19.19 (a)



$$Q = \dot{m}_{steam} \lambda_{steam}$$

$$Q = \dot{m}_p C_p \Delta T$$

$$Q = UA \Delta T_{lm}$$

$$\Delta T_{lm} = \frac{(140^\circ\text{C} - 90^\circ\text{C}) - (140^\circ\text{C} - 120^\circ\text{C})}{\ln\left(\frac{140^\circ\text{C} - 90^\circ\text{C}}{140^\circ\text{C} - 120^\circ\text{C}}\right)} = 32.74^\circ\text{C}$$

$$\frac{Q_2}{Q_1} = Q = \frac{\dot{m}_{steam,2} \lambda_2}{\dot{m}_{steam,1} \lambda_1}$$

$$Q = \frac{\dot{m}_{p,2} C_{p,2} (T - 90^\circ\text{C})}{\dot{m}_{p,1} C_{p,1} (120^\circ\text{C} - 90^\circ\text{C})}$$

$$Q = \frac{U_2 A_2 \Delta T_{lm,2}}{U_1 A_1 \Delta T_{lm,1}} = \frac{M^{0.8}}{32.74^\circ\text{C}} \frac{(140^\circ\text{C} - 90^\circ\text{C}) - (140^\circ\text{C} - T)}{\ln((140^\circ\text{C} - 90^\circ\text{C})/(140^\circ\text{C} - T))}$$

$$\text{All resistance on tube side} \Rightarrow \frac{U_2}{U_1} = \left(\frac{\dot{m}_2}{\dot{m}_1}\right)^{0.8} = M^{0.8}$$

For  $M = 1.15$

$$M \frac{(T - 90^\circ\text{C})}{30^\circ\text{C}} = \left(\frac{M^{0.8}}{32.74^\circ\text{C}}\right) \left(\frac{(T - 90^\circ\text{C})}{\ln((140^\circ\text{C} - 90^\circ\text{C})/(140^\circ\text{C} - T))}\right)$$

$$\frac{1.15}{30^\circ\text{C}} = \left(\frac{1.15^{0.8}}{32.74^\circ\text{C}}\right) \left(\frac{1}{\ln((140^\circ\text{C} - 90^\circ\text{C})/(140^\circ\text{C} - T))}\right)$$

$$\ln \frac{50^\circ\text{C}}{140^\circ\text{C} - T} = 0.891$$

$$T = 119.5^\circ\text{C}$$

(b)  $\Delta P = a + b\dot{v}^2$  to get system curve

$$a = \Delta P_{s-d} + \Delta P_{head}$$

$$\begin{aligned} a &= 250 + \rho gh = 250 \text{ kPa} + (1000 \text{ kg/m}^3)(9.81 \text{ m/s})(-3 \text{ m}) \\ &= 250 \text{ kPa} - 29.43 \text{ kPa} \\ &= 220.57 \text{ kPa} \end{aligned}$$

$$\text{at } \dot{v} = 35 \text{ m}^3/\text{h} \text{ and } \Delta P = 220.57 \text{ kPa} + 20 \text{ kPa} = 240.57 \text{ kPa}$$

$$240.57 = 220.57 + b(35)^2$$

$$b = 0.01633$$

$$\Delta P = 220.57 + 0.01633\kappa^2$$

$$35 \text{ m}^3/\text{h}(1.15) = 40.25 \text{ m}^3/\text{h}$$

$$\text{at } 40 \text{ m}^3/\text{h } \Delta P = 247.03 \text{ kPa} \Rightarrow \text{No problems}$$

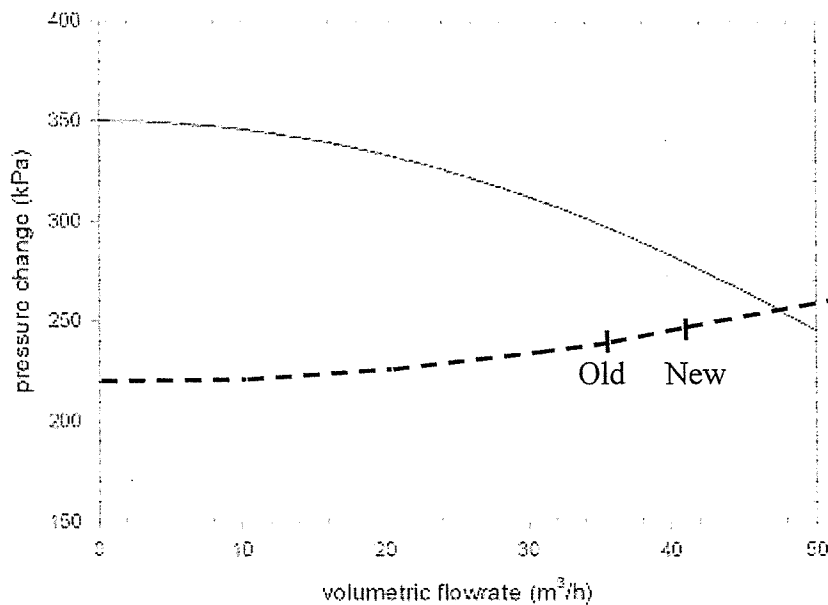


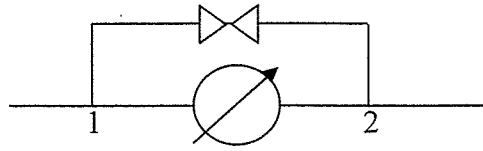
Figure P19.19 Pump Curve for Problem 19

(c) Desuperheater –  $Q = UA\Delta T$

If  $Q$  increases  $\Delta T$  increases  $\Rightarrow$  must raise steam pressure or lower column pressure

(d) Liquid phase reaction  $\Rightarrow$  no effect on pressure  $\Rightarrow$  bad idea

19.20



With valve closed  $U = U_{BC}$ ,  $-\Delta P_{f,BC} = P_1 - P_2$ ,  $V_{OL} \text{ Flow} = Q_{BC}$

With valve open  $U = U_2$ ,  $-\Delta P_{f,2} = Q_2$

- (a) For fully developed turbulent flow,  $-\Delta P_f \propto Q^2$

$$\frac{\Delta P_{f,2}}{\Delta P_{f,BC}} = 1.69 = \left( \frac{Q_2}{Q_{BC}} \right)^2 \Rightarrow Q_2 = Q_{BC} \sqrt{1.69} = 1.3 Q_{BC}$$

Flow increases by 30%

- (b)  $1.69 - 1.3 = 0.39$  or 39% of new flow through bypass

(c)  $U = \left( \frac{1}{h_i} + \frac{1}{h_o} \right)^{-1} \approx \left( \frac{1}{h_o} \right)^{-1} = h_o$

$$\frac{U_2}{U_{BC}} = \frac{h_o}{h_{BC}} = \left( \frac{1.3}{1} \right)^{0.6} = 1.17$$

117% of  $U_{BC}$

19.21 (a)  $\Delta P = \rho gh + \Delta P_f$   
 $\Delta P = (775 \text{ kg/m}^3)(9.81 \text{ m/s}^2)(42 \text{ m} - 6 \text{ m}) + 130,000 \text{ Pa}$   
 $\Delta P = 403420 \text{ Pa} = 403 \text{ kPa}$

Look at pump curve at 7 L/s  $\Rightarrow$  above pump curve – single pump will not work;  
 could try pump in series (parallel will not work).

If  $\Delta P$  is high, maybe pipe diameter is too small.

(b)  $Q = UA\Delta T$   
 $\frac{1}{U} = \frac{1}{5000 \text{ W/m}^2\text{K}} + \frac{1}{3000 \text{ W/m}^2\text{K}} \Rightarrow U = 1875 \text{ W/m}^2\text{K}$   
 $6000 \times 10^3 \text{ W} = (1875 \text{ W/m}^2\text{K})(300 \text{ m}^2)\Delta T$   
 $\Delta T = 10.67^\circ\text{C}$   
 $T_{\text{steam}} = 10.67^\circ\text{C} + 87^\circ\text{C} = 97.7^\circ\text{C}$

|                                         |
|-----------------------------------------|
| $T_{\text{steam}} = 97.7^\circ\text{C}$ |
|-----------------------------------------|

- (c) Flooding – Will column perform this separation without flooding (may need large reflux ratio if  $N$  is much lower than needed?)

$N$  – Are there an efficient number of trays to achieve desired separation without very high reflux ratio to cause flooding?

Feed Location – Does feed location cause trays to be “wasted?” This means that there would be trays that are not doing much separation.

Condenser and Reboiler – Are these sufficient to handle heat transfer?  
 If oversized – will they perform as needed?

19.22 (a)  $T_{steam} = 160^{\circ}\text{C}$   
 $T_{process} = 70^{\circ}\text{C}$   
 $\Delta T = 160^{\circ}\text{C} - 70^{\circ}\text{C} = 90^{\circ}\text{C}$

In order to find  $T_{wall}$  need  $h_{process} \Rightarrow$  Trial and Error

$$Q = UA\Delta T$$

$$\frac{Q}{A} = U\Delta T = h_p (T_{wall} - 70^{\circ}\text{C})$$

$$T_{wall} = 70^{\circ}\text{C} + \frac{U}{h_p} 90^{\circ}\text{C}$$

| $T_{wall}$<br>( $^{\circ}\text{C}$ ) | $\Delta T$<br>( $^{\circ}\text{C}$ ) | $h_{process}$<br>( $\text{W}/\text{m}^2\text{C}$ ) | $h_{steam}$<br>( $\text{W}/\text{m}^2\text{C}$ ) | $U$<br>( $\text{W}/\text{m}^2\text{C}$ ) | $T_{wall}$<br>( $^{\circ}\text{C}$ ) |
|--------------------------------------|--------------------------------------|----------------------------------------------------|--------------------------------------------------|------------------------------------------|--------------------------------------|
| 110                                  | 40                                   | 2500                                               | 3000                                             | 1364                                     | 119.09                               |
| 120                                  | 50                                   | 2300                                               | 3000                                             | 1301                                     | 120.91                               |
| 130                                  | 60                                   | 2160                                               | 3000                                             | 1256                                     | 122.33                               |
| 140                                  | 70                                   | 2100                                               | 3000                                             | 1235                                     | 122.93                               |

$T_{wall}$  is approximately  $120^{\circ}\text{C}$  and  $U = 1301 \text{ W}/\text{m}^2\text{C}$

$$\frac{Q_{new}}{Q_{base}} = \frac{U_{new} \Delta T_{new}}{U_{base} \Delta T_{base}} = \frac{1301 \text{ W}/\text{m}^2\text{C} \cdot 90^{\circ}\text{C}}{1500 \text{ W}/\text{m}^2\text{C} \cdot 40^{\circ}\text{C}} = 1.95$$

$Q_{new} = 95\% Q_{base}$

(b) The reboil rate will increase; therefore, pressure in column will increase.



19.23 (a)  $\Delta P_{12} = P_2 - P_1 = 101 \text{ kPa} - 200 \text{ kPa} = -99 \text{ kPa}$

$$\Delta P_{\text{pump}} = (\Delta P_f + \Delta P_{cv}) - \Delta P_{12}$$

$$\Delta P_{\text{pump}} \text{ from pump curve}$$

$$\Delta P_{cv} = 225 \text{ kPa} - (250 \text{ kPa} - 99 \text{ kPa})$$

$$\Delta P_{cv} = 74 \text{ kPa}$$

(b)

| $\Delta P$ (kPa)  | $P$ (kPa)    | Location/Equipment     |
|-------------------|--------------|------------------------|
|                   | 101          | TK-2901                |
| $100(1.5)^2=225$  |              | Pipe                   |
|                   | 326          |                        |
| $50(1.5)^2=112.5$ |              | E-2905                 |
|                   | <b>438.5</b> | <b>Inlet to E-2905</b> |
| $25(1.5)^2=56.25$ |              | Pipe                   |
|                   | 494.8        |                        |
| $50(1.5)^2=112.5$ |              | E-2904                 |
|                   | <b>607.3</b> | <b>Inlet to E-2904</b> |

$$P_{in} \text{ at E-2904} = 607.3 \text{ kPa}$$

$$P_{in} \text{ at E-2905} = 438.5 \text{ kPa}$$

Note to instructor: There is no part c for this problem. What is labeled part c in the book is the question for part b.

19.24 (a)  $\Delta P_f \propto \dot{m}^2$

$$\therefore \left( \frac{\dot{m}_{1,2}}{\dot{m}_{1,1}} \right)^2 = \left( \frac{\Delta P_{1,2}}{\Delta P_{1,1}} \right) = \left( \frac{52 \text{ kPa}}{40 \text{ kPa}} \right)$$

$$\dot{m}_{1,2} = \dot{m}_{1,1} \sqrt{\frac{52 \text{ kPa}}{40 \text{ kPa}}} = 2 \left( \frac{\text{kg}}{\text{s}} \right) \sqrt{\frac{52 \text{ kPa}}{40 \text{ kPa}}}$$

$$\boxed{\dot{m}_{1,2} = 2.280 \text{ kg/s}}$$

$$\Delta P_f \propto \dot{m}^2$$

$$\therefore \left( \frac{\dot{m}_{2,2}}{\dot{m}_{2,1}} \right)^2 = \left( \frac{\Delta P_{2,2}}{\Delta P_{2,1}} \right) = \left( \frac{52 \text{ kPa}}{40 \text{ kPa}} \right)$$

$$\dot{m}_{2,2} = \dot{m}_{2,1} \sqrt{\frac{52 \text{ kPa}}{25 \text{ kPa}}} = 1.5 \left( \frac{\text{kg}}{\text{s}} \right) \sqrt{\frac{52 \text{ kPa}}{25 \text{ kPa}}}$$

$$\boxed{\dot{m}_{2,2} = 2.163 \text{ kg/s}}$$

(b) Exchanger 1

$$\frac{1}{U_1} = \frac{1}{200 \text{ W/m}^2\text{K}} = \frac{1}{h_i} + \frac{1}{1000 \text{ W/m}^2\text{K}} \Rightarrow h_i = 250 \text{ W/m}^2\text{K}$$

$$U_2 = \left[ \left( \frac{1}{250 M^{0.8}} \right) + \left( \frac{1}{1000} \right) \right]^{-1} = \left[ \left( \frac{1}{250 (2.28/2)^{0.8}} \right) + \left( \frac{1}{1000} \right) \right]^{-1}$$

$$\boxed{U_2 = 217.3 \text{ W/m}^2\text{K}}$$

(c) Exchanger 2

$$\frac{1}{U_1} = \frac{1}{150 \text{ W/m}^2\text{K}} = \frac{1}{h_i} + \frac{1}{1000 \text{ W/m}^2\text{K}} \Rightarrow h_i = 176.5 \text{ W/m}^2\text{K}$$

$$U_2 = \left[ \left( \frac{1}{176.5 M^{0.8}} \right) + \left( \frac{1}{1000} \right) \right]^{-1} = \left[ \left( \frac{1}{176.5 (2.163/1.5)^{0.8}} \right) + \left( \frac{1}{1000} \right) \right]^{-1}$$

$$\boxed{U_2 = 191.3 \text{ W/m}^2\text{K}}$$

(d) Exchanger 1

$$\frac{Q_2}{Q_1} = \frac{U_2 A_2 \Delta T_{lm,2}}{U_1 A_1 \Delta T_{lm,1}} = \frac{\dot{m}_2 C_{p,2} \Delta T_2}{\dot{m}_1 C_{p,1} \Delta T_1} = \frac{2.28}{2} = 1.14$$

$$1.14 = \left( \frac{217.3}{200} \right) \left[ \frac{(T-50) - (T-100)}{72.13} \ln \left( \frac{T-50}{T-100} \right) \right]$$

$$\ln \left( \frac{T-50}{T-100} \right) = \left( \frac{217.3}{200} \right) \left( \frac{50}{72.13} \right) \left( \frac{1}{1.14} \right)$$

$$T = 153.4^\circ\text{C}$$

(e) Exchanger 2

$$\frac{Q_2}{Q_1} = \frac{U_2 A_2 \Delta T_{lm,2}}{U_1 A_1 \Delta T_{lm,1}} = \frac{\dot{m}_2 C_{p,2} \Delta T_2}{\dot{m}_1 C_{p,1} \Delta T_1} = \frac{2.163}{1.5} = 1.442$$

$$1.442 = \left( \frac{191.3}{150} \right) \left[ \frac{(T-50) - (T-100)}{72.13} \ln \left( \frac{T-50}{T-100} \right) \right]$$

$$\ln \left( \frac{T-50}{T-100} \right) = \left( \frac{191.3}{150} \right) \left( \frac{50}{72.13} \right) \left( \frac{1}{1.442} \right)$$

$$T = 159.1^\circ\text{C}$$

(f)  $\Delta P_2 = \Delta P_1$

$$40 \left( \frac{3.1}{2} \right)^2 = 25 \left( \frac{m_2}{1.5} \right)^2$$

$$m_2 = \left( \frac{3.1}{2} \right) (1.5) \sqrt{\frac{40}{25}}$$

$$m_2 = 2.941 \text{ kg/s}$$

## Chapter 20

- 20.1 The main reason is the requirement for large reactors in order to obtain very high conversions. If the reaction rates are very high for the conditions used (for example the combustion of a fuel in air or oxygen) then conversions >99% can be achieved. But for many chemical processes the reaction rates are moderate and as conversion increases so does the volume of reactor required. For example, for a 1<sup>st</sup> order irreversible plug flow reactor the ratio of reactor volumes for conversions of 50%, 90%, 95%, 99%, and 99.9% are as follows:

$$1:3.3:4.3:6.6:10$$

In addition, for high conversions the flow pattern and mixing in the reactor become increasingly important. Even small deviations from ideal plug flow cause large changes in reactor volume at very high conversions. The final decision on what single-pass conversion to use is based on the economics, but we get decreasing returns as the conversion increases.

- 20.2 This statement is not necessarily true. If the reaction is at equilibrium, then according to Le Chatelier's principle - an increase in reaction temperature will lead to a decrease in conversion. However, if the reaction is not at equilibrium then an increase in temperature increases the rates of reaction and this will move the reaction to the right, increasing conversion.

- 20.3 Since the reaction rate doubles for a temperature change from 250°C to 270°C, we may write that

$$\frac{-r_{270}}{-r_{250}} = 2 = \frac{e^{-E/R(273+270)}}{e^{-E/R(273+250)}} \Rightarrow e^{-E/R(\frac{1}{543} - \frac{1}{523})}$$

$$\therefore E/R = -\frac{\ln(2)}{(\frac{1}{543} - \frac{1}{523})} = 9,842$$

- (a) For a change from 250°C to 260°C, we get

$$\frac{-r_{260}}{-r_{250}} = \frac{x}{10} = \frac{e^{-E/R(533)}}{e^{-E/R(523)}}$$

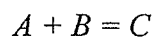
$$\therefore x = 10e^{-9,842[\frac{1}{533} - \frac{1}{523}]} = 14.23 \Rightarrow \text{a 42.3\% increase}$$

- (b) Assuming that the feed composition is constant then the gas phase concentrations increase by 15%

$$\frac{-r_{1.15P}}{-r_P} = \frac{x}{10} = \frac{(1.15)^3}{(1)^3} \Rightarrow x = 1.52 \text{ a 52\% increase}$$

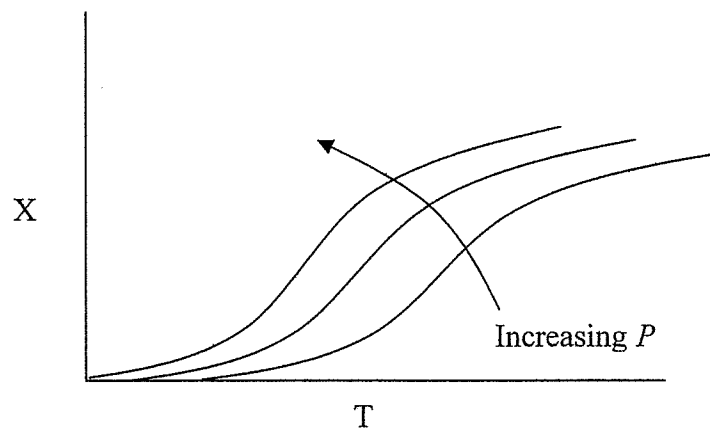
- 20.4 (a) The reaction is exothermic, since heat is removed after the first packed bed section.
- (b) The reaction is run at high temperature because
- (i) a high temperature is needed to obtain high enough reaction rates to give a “reasonably” sized reactor.
  - (ii) the reaction must take place in the vapor phase and the only way to obtain a vapor is to increase the temperature (as opposed to creating a vacuum that would reduce the concentration and hence the reaction rate).

20.5

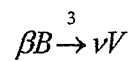
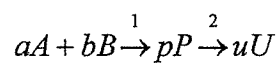


As  $T \uparrow$ ,  $X \uparrow$

As  $P \uparrow$ ,  $X \uparrow$



20.6

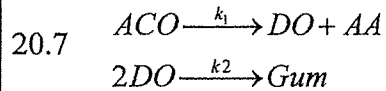


Low  $B$  minimizes  $V$

Low  $x$  (conversion) minimizes  $U$

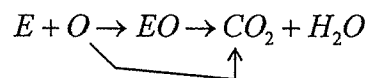
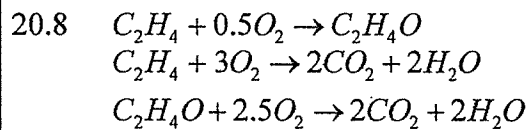
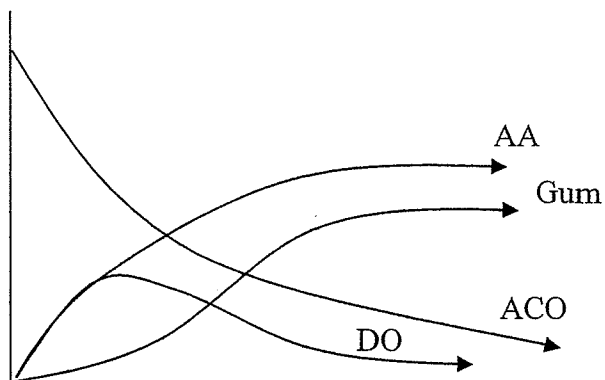
$E_1 > E_2 > E_3 \Rightarrow$  high  $T$  maximizes reaction 1

$$\eta = \frac{r_p}{r_u + r_v} = \frac{k_1 C_A C_B - k_2 C_P}{k_3 C_B^2 + k_2 C_P}$$



$$\eta = \frac{r_{DO}}{r_{gum}} = \frac{k_1 C_{CO} - 2k_2 C_{DO}^2}{k_2 C_{DO}^2} = \frac{k_1}{k_2} \frac{C_{CO}}{C_{DO}^2} - 2$$

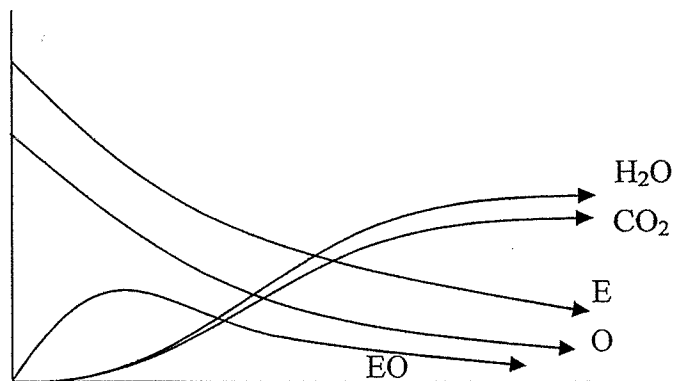
$E_1 < E_2 \Rightarrow$  low T,  $C_{DO}$  low, low conversion



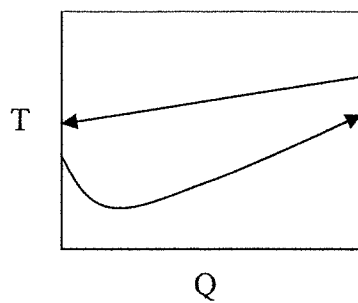
Low conversion

Ethylene excess  $\Rightarrow$  low  $O_2$  – flammable limit issue

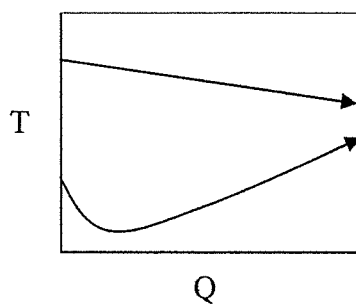
Low temperature



20.9 (a) Countercurrent flow of HTM



(b) Cocurrent flow of HTM



Large  $\Delta T$  desired at entrance where reaction rate highest to keep reaction from quenching.



20.10 (a)  $P + B \rightarrow C$

$$\begin{array}{ccc} \underline{P} & \underline{B} & \underline{C} \\ \frac{1-x}{1-x} & \frac{1-x}{1-0.5x} & \frac{x}{x} \\ \frac{1-x}{1-0.5x} & \frac{1-x}{1-0.5x} & \frac{x}{1-0.5x} \end{array} \quad \varepsilon = 0.5$$

$$-r = kC_{Po}^2 \frac{(1-x)^2}{(1-0.5x)^2}$$

$$C_P = C_{Po} \left( \frac{1-x}{1-0.5x} \right)^2$$

$$C_B = C_{Bo} \left( \frac{1-x}{1-0.5x} \right)^2$$

$$\frac{W}{F_{Po}} = \frac{x}{-r_A}$$

$$\frac{W}{F_{Po}} = \frac{x}{-kC_{Po}^2 \left( \frac{1-x}{1-0.5x} \right)^2}$$

$$\frac{W_2 / F_{Po1}}{W_1 / F_{Po2}} = \left( \frac{x_2}{x_1} \right) \left( \frac{-r_{A1}}{-r_{A2}} \right) = \cancel{x_2} / \cancel{x_1} \cdot \cancel{k_{11}} / \cancel{k_{12}} \cdot \frac{C_{Po1}^2 (1-x_1)^2 (1-0.5x_2)^2}{C_{Po2}^2 (1-x_2)^2 (1-0.5x_1)^2}$$

$$\frac{1}{1.25} = \frac{x_2 (1-x_1)^2 (1-0.5x_2)^2}{x_1 (1-x_2)^2 (1-0.5x_1)^2}$$

$$\frac{1}{1.25} = \frac{x_2 (1-0.68)^2 (1-0.5x_2)^2}{0.68 (1-x_2)^2 (1-(0.5)(0.68))^2}$$

$$2.314 = \frac{x_2 (1-0.5x_2)^2}{(1-x_2)^2}$$

$$\boxed{x_2 = 0.63}$$

$$(b) \quad \frac{W_2}{W_1} \frac{F_{Po1}}{F_{Po2}} = \frac{x_2}{x_1} \frac{-r_{P1}}{-r_{P2}}$$

$$\frac{W_2}{W_1} = \frac{F_{Po2}}{F_{Po1}} = 1.25$$

25% increase

$$(c) \quad \frac{W_2}{W_1} \frac{F_{Po1}}{F_{Po2}} = \frac{x_2}{x_1} \frac{-r_{P1}}{-r_{P2}}$$

$$\frac{1}{1.25} = \frac{k_{11} C_{Po1}^2 (1-x_1)^2 (1-0.5x_2)^2}{k_{12} C_{Po2}^2 (1-x_2)^2 (1-0.5x_1)^2}$$

$$\frac{1}{1.25} = \frac{3.5 \times 10^4 \exp \left[ \frac{-12,530}{573 \text{ K}} \right]}{3.5 \times 10^4 \exp \left[ \frac{-12,530}{T_2} \right]}$$

$$T_2 = 579 \text{ K} = 306^\circ\text{C}$$

$$(d) \quad \frac{W_2}{W_1} \frac{F_{Po1}}{F_{Po2}} = \left( \frac{x_2}{x_1} \right) \left( \frac{-r_{P1}}{-r_{A2P}} \right) = \frac{x_2}{x_1} \frac{k_{11} (P_{Po1}/RT)^2 (1-x_1)^2 (1-0.5x_2)^2}{k_{12} (P_{Po2}/RT)^2 (1-x_2)^2 (1-0.5x_1)^2}$$

$$\frac{1}{1.25} = \left( \frac{y_{A01} (3 \text{ MPa})}{y_{A02} P_2} \right)^2$$

$$P_2 = 3.35 \text{ Mpa}$$

$$20.11 \quad t = N_{Ao} \int_0^x \frac{dx}{-r_A v} = C_{Ao} \int_0^x \frac{dx}{-r_A}$$

$$-r_A = kC_{Ao}(1-x)$$

$$t = C_{Ao} \int_0^x \frac{dx}{kC_{Ao}(1-x)}$$

$$t = \frac{1}{k} \ln(1-x)$$

$$(a) \quad \frac{t_2}{t_1} = \frac{k_1 \ln(1-x_2)}{k_2 \ln(1-x_1)} = \frac{\ln(1-0.75)}{\ln(1-0.50)}$$

$$\boxed{\frac{t_2}{t_1} = 2}$$

$$(b) \quad \frac{t_2}{t_1} = \frac{k_1 \ln(1-x_2)}{k_2 \ln(1-x_1)} = \frac{\ln(1-0.90)}{\ln(1-0.50)}$$

$$\boxed{\frac{t_2}{t_1} = 3.32}$$

$$(c) \quad t = C_{Ao} \int_0^x \frac{dx}{kC_{Ao}^2(1-x)^2} = \frac{1}{kC_{Ao}} \frac{1}{(1-x)} \Big|_0^x = \frac{1}{kC_{Ao}} \left[ \frac{1}{(1-x)} - 1 \right] = \frac{1}{kC_{Ao}} \frac{x}{1-x}$$

$$t = \frac{1}{kC_{Ao}} \frac{x}{1-x}$$

$$\frac{t_2}{t_1} = \frac{x_2}{x_1} \left( \frac{1-x_1}{1-x_2} \right) = \left( \frac{0.75}{0.50} \right) \left( \frac{1-0.5}{1-0.75} \right) = 3$$

$$\boxed{\frac{t_2}{t_1} = 3}$$

$$\frac{t_2}{t_1} = \frac{x_2}{x_1} \left( \frac{1-x_1}{1-x_2} \right) = \left( \frac{0.90}{0.50} \right) \left( \frac{1-0.5}{1-0.90} \right) = 9$$

$$\boxed{\frac{t_2}{t_1} = 9}$$

- (d) Since this is not a linear function, it takes much longer to approach 100% conversion since approach is asymptotic.

# Benzene Production (Prob 20.12-20.16)

## Reactor Simulation Set-up Values

a.) Chemical Reaction:  $C_6H_5CH_3 + H_2 \rightarrow C_6H_6 + CH_4$  \*

I.D. No.                      (1)              (2)              (3)              (4)

b.) Rate Expression

$$-r_1 = k_1 C_1 C_2^*$$

c.) Simulation Values

| Components               | I.D. Number | Stoich. Coefficient | kinetic Exponent |
|--------------------------|-------------|---------------------|------------------|
| Toluene ( $C_6H_5CH_3$ ) | (1) *       | -1 *                | 1.0 *            |
| Hydrogen ( $H_2$ )       | (2) *       | -1 *                | 0.5 *            |
| Benzene ( $C_6H_6$ )     | (3) *       | 1 *                 | - *              |
| Methane ( $CH_4$ )       | (4) *       | 1 *                 | - *              |

d.) Kinetics:  $k_1 = k_0 e^{-[E/RT]}$

$$k_0 [k\text{mol}/m^3] = 3.0 \times 10^{10} *$$

$$E [kcal/kmol] = 5.09 \times 10^4 *$$

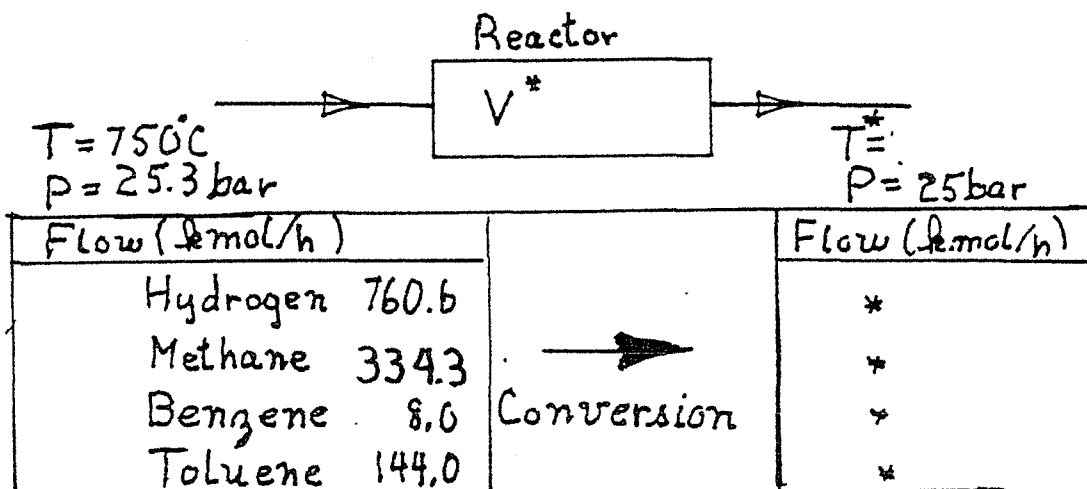
e.) Transport and Thermal Properties:

Enthalpy: SRK\*      Kvalues: SRK\*

f.) Reaction phase: (gas)\*

\* Information Supplied (Reaction Specific)

Prob. 20.12 (Start on previous page)



Reactor Specs.:

Reactor Type:

Thermal Mode: Adiabatic

Simulation: Specify Conversion  $\rightarrow$  Calc. Volume

\* Calculated Values

Sample Simulation (base case):

Set conversion = 0.75

Simulation Results:

$T = 813^\circ\text{C}$

Flows [kmol/h]

Hydrogen 652.6

Methane 442.3

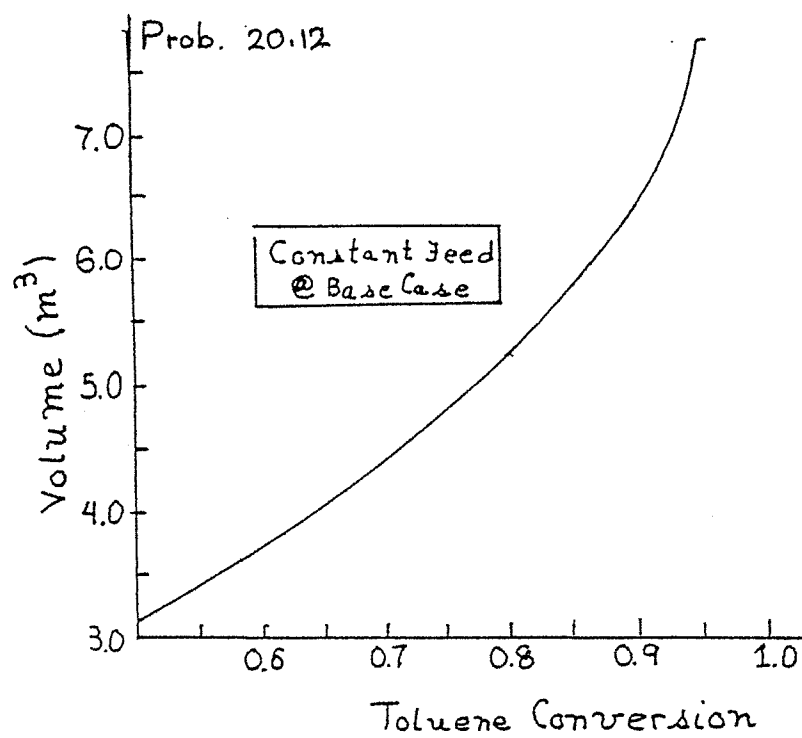
Benzene 116.0

Toluene 36.0

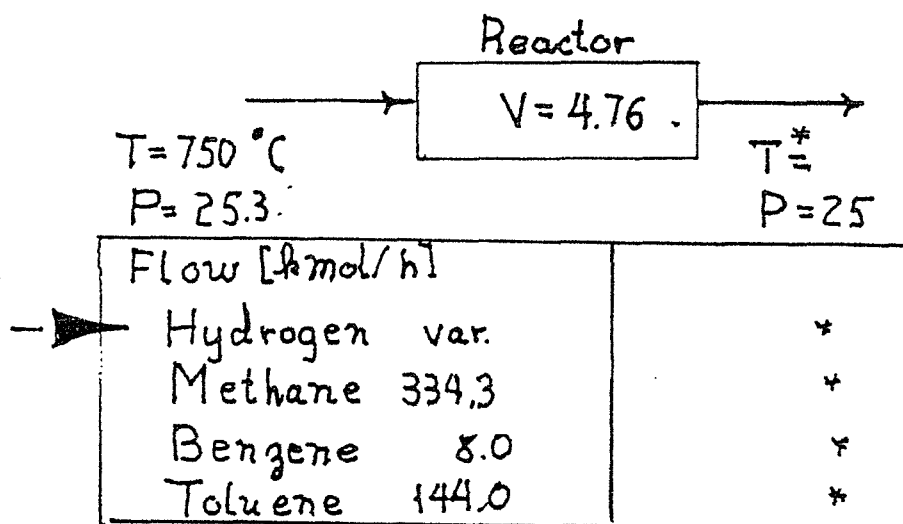
$V = 4.76 \text{ m}^3$

cont.

Prob. 20.12



Prob. 20.13 (start on page preceding Prob 20.12)



Reactor Specs.

Reactor Type: Plug Flow

Thermal Mode: Adiabatic

\* Calculated Values

Simulation: Specify Volume  $\rightarrow$  Calculate Conv.

For various values of Hydrogen flow.  
calculate % xs  $\text{H}_2$  then obtain output stream.

Sample Calculation.

For Hydrogen = 250 kmol/h

$$\text{Excess } \text{H}_2 = [(250 - 144) / 144] \times 100 = 0.736$$

Set  $V = 4.76 \text{ m}^3$

Simulation Results:

$T = 841^\circ\text{C}$

Flows (kmol/h)

Hydrogen 121.5

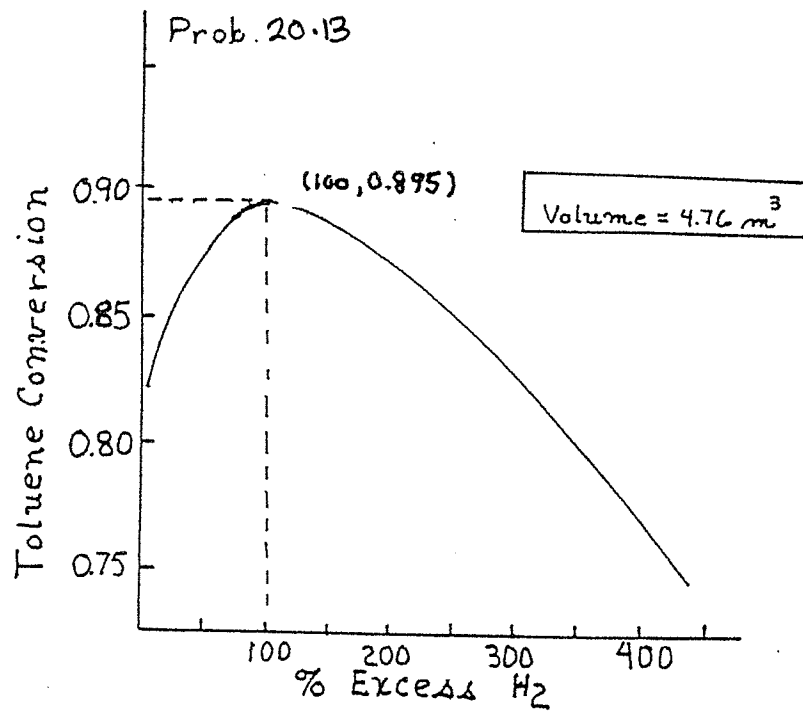
Methane 462.8

Benzene 136.5

Toluene 15.5

SEE PLOT

Conversion =  $(1 - 15.5/144) = 0.89$

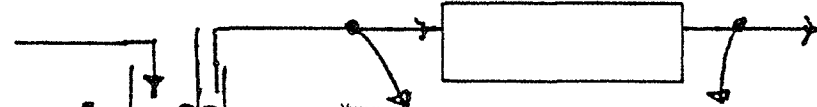




Prob. 20.14 (start on page before Prob. 20.12)

Reactor 1

Reactor 2



|               |        |              |        |
|---------------|--------|--------------|--------|
| T             | 750°C  | *            | *      |
| P             | 25 bar |              |        |
| Flow (kmol/h) |        |              |        |
| Hydrogen      | 760.6  | *            | *      |
| Methane       | 334.3  | *            | *      |
| Benzene       | 8.0    | *            | *      |
| Toluene       | 441.4  | *            | *      |
|               | Input  | Intermediate | Output |

\* Calculated Values

| Reactor Specs. | Reactor 1    | Reactor 2 |
|----------------|--------------|-----------|
| Reactor Type   | Stirred Tank | Plug Flow |
| Thermal Mode   | Adiabatic    | Adiabatic |

Two Step Simulation

Step 1, Reactor 1: Set volume  $\rightarrow$  Calculate

Step 2, Reactor 2: Set conversion  $\rightarrow$  Calc. Volume

Sample Calculation: Reactor Volume = 2.75 m<sup>3</sup>

Simulation Results: Step 1

T = 790 °C      P = 25 bar

Flows (kmol/h)

Hydrogen      691.2

Methane      402.8

Benzene      77.4

Toluene      74.6

Continued

Prob 20.14 Cont

Toluene Conversion in Reactor 1 =  $(144 - 74.6) / 144 = 0.48$

Toluene Conversion needed in Reactor 2 to provide over-all conversion =  $(74.6 - 36) / 74.6 = 0.52$

Simulation Results: Step 2 (Conv. = 0.52)

Temp =  $813^{\circ}\text{C}$       P = 25 bar

Flows = (kmol/h)

Hydrogen      652.6

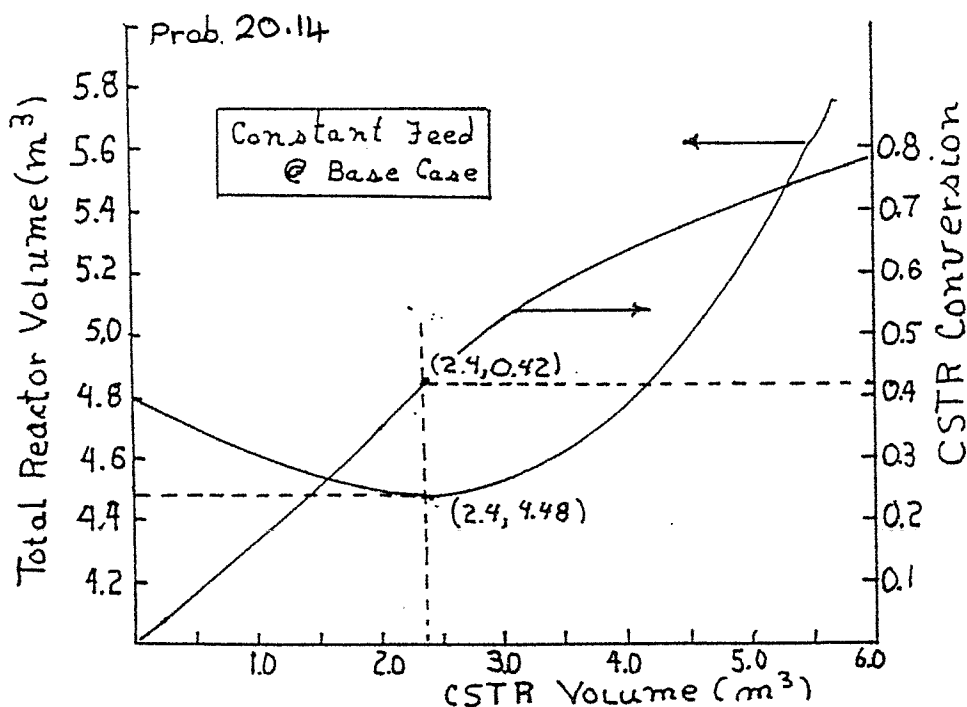
Methane      442

Benzene      116

Toluene      36

$V = 1.72 \text{ m}^3$

Total Volume =  $2.75 + 1.72 = 4.47 \text{ m}^3$



Prob. 20.15 (start on page preceding Prob. 20.12)

| Reactor                                         |          |                                 |
|-------------------------------------------------|----------|---------------------------------|
| $V = 4.76 \text{ m}^3$                          |          |                                 |
| $T = 750^\circ\text{C}$<br>$P = 25 \text{ bar}$ |          | $T = *$<br>$P = 25 \text{ bar}$ |
| Mol fractions                                   |          |                                 |
| Hydrogen                                        | 0.61     | *                               |
| Methane                                         | 0.268    | *                               |
| Benzene                                         | 0.0064   | *                               |
| Toluene                                         | 0.116    | *                               |
| Total Flow                                      | Variable | *                               |

Reactor Specs: \*Calculated Values

Reactor Type: Plug Flow

Thermal Mode: Adiabatic

Simulation: Specify Volume  $\rightarrow$  Calculate Conv.

Sample Calculation:

Total Flow = 750 (kmol/h)

$V = 4.76 \text{ m}^3$

Simulation Results:

$T = 831^\circ\text{C}$

Flows (kmol/hr)

Hydrogen 374.0

Methane 284.5

Benzene 88.3

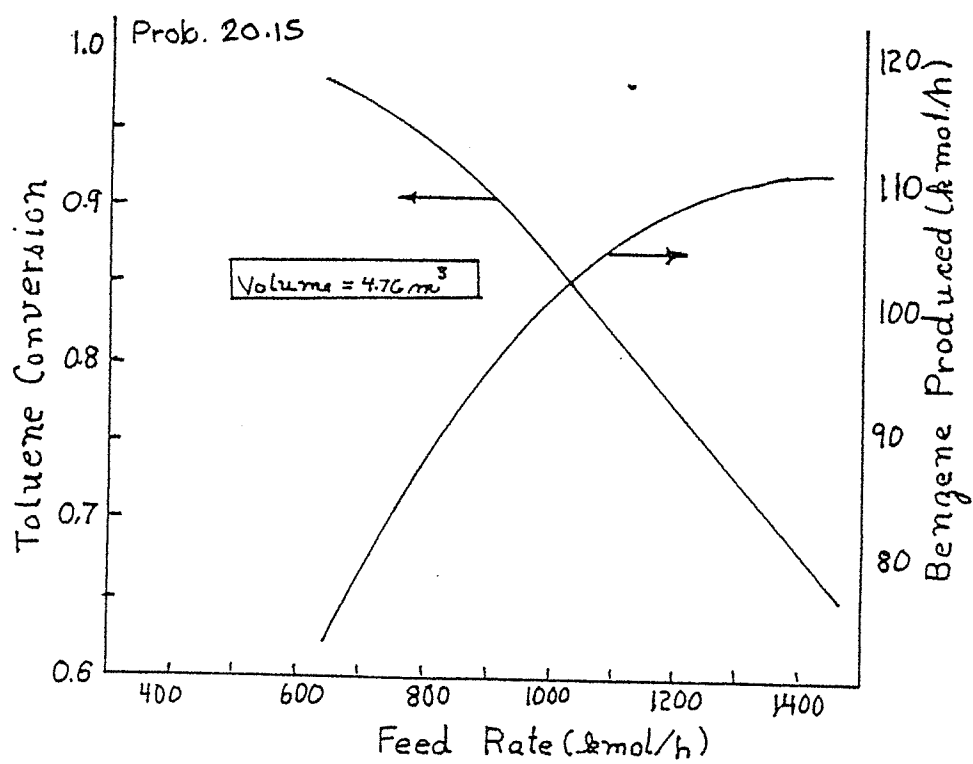
Toluene 3.15

Output

$$\text{Toluene Conversion} = \frac{(0.116)(750) - 3.15}{(0.116)(750)} = 0.963$$

$$\text{Benzene Generated} = (0.116)(750)(0.963) = 838 \left[ \frac{\text{kmol}}{\text{h}} \right] \text{ (cont.)}$$

Prob. 20.15 cont.



Prob. 20.16 (start on page preceding Prob 20.12)

| Inlet                  |       | Reactor     | Outlet           |   |
|------------------------|-------|-------------|------------------|---|
| T = var.<br>P = 25 bar |       | V = 4.76 m³ | T*<br>P = 25 bar |   |
| Flow [kmol/h]          |       |             |                  |   |
| Hydrogen               | 760.6 |             |                  | * |
| Methane                | 334.3 |             |                  | * |
| Benzene                | 8.0   |             |                  | * |
| Toluene                | 144.0 |             |                  | * |

Reactor Specs;

\* calculated

Reactor Type: Plug Flow values

Thermal Mode: Adiabatic

Simulation Specify Volume → Calculate conversion

Sample Calculation:

$$T = 770^{\circ}\text{C}$$

$$V = 4.76 \text{ m}^3$$

Simulation Results:

$$T = 848.5^{\circ}\text{C}$$

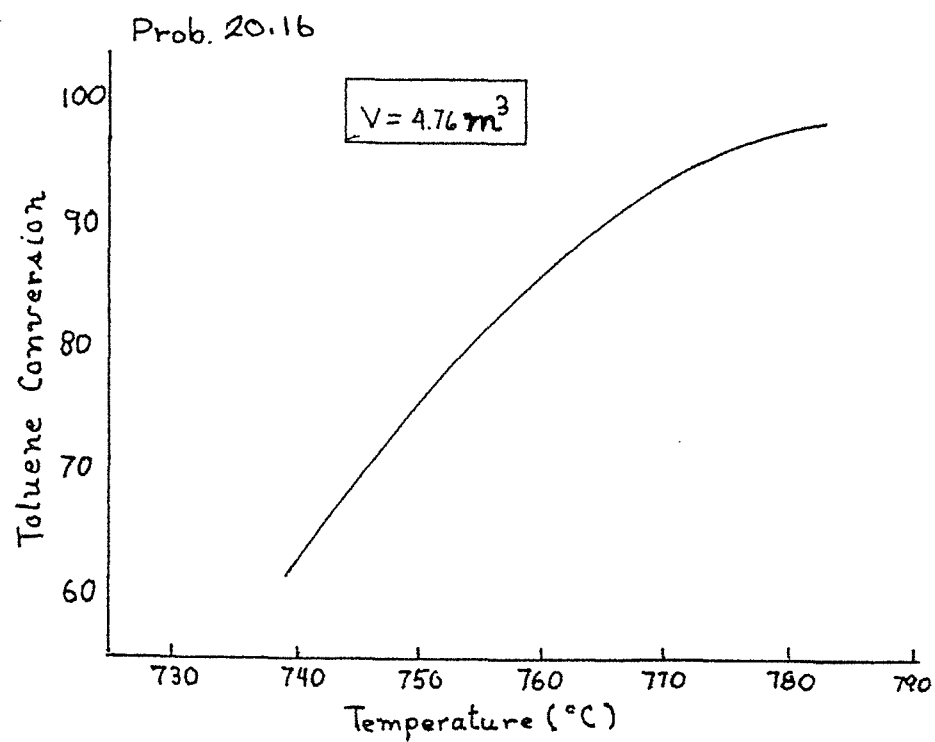
Flows (kmol/h)

|          |       |
|----------|-------|
| Hydrogen | 625.2 |
| Methane  | 469.7 |
| Benzene  | 143.4 |
| Toluene  | 8.63  |

$$\text{Conversion} = 1 - 8.63/144 = 0.94$$

(cont.)

Prob. 20.16 cont.

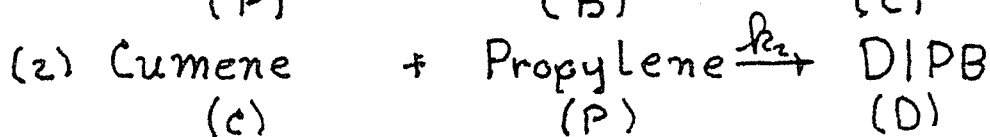
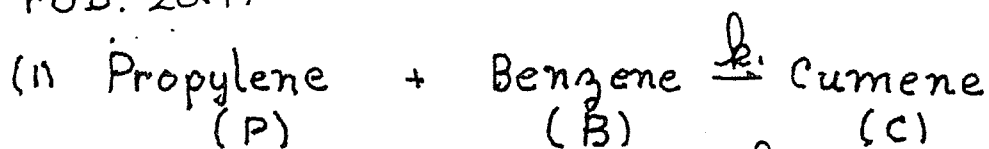


# Cumene Production (Prob 20.17-20.20)

## Reactor Simulation Set-up Values

|                                                                      |                                                                                                                                                                                                          |                     |                  |                                                                                                                                                                       |                     |                  |
|----------------------------------------------------------------------|----------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|---------------------|------------------|-----------------------------------------------------------------------------------------------------------------------------------------------------------------------|---------------------|------------------|
| a) Chemical Reactions:                                               | Reaction 1                                                                                                                                                                                               |                     |                  | Reaction 2                                                                                                                                                            |                     |                  |
|                                                                      | $\text{C}_3\text{H}_6 + \text{C}_6\text{H}_6 \rightarrow \text{C}_9\text{H}_{12}^*$ <div>(1)                (2)                (3)</div> $-r_1 = k_1 C_1 C_2^*$                                          |                     |                  | $\text{C}_3\text{H}_6 + \text{C}_9\text{H}_{12} \rightarrow \text{C}_{12}\text{H}_{18}^*$ <div>(1)                (3)                (4)</div> $-r_2 = k_2 C_1 C_3^*$ |                     |                  |
| c) Simulation Values:                                                |                                                                                                                                                                                                          |                     |                  |                                                                                                                                                                       |                     |                  |
|                                                                      | I.D. Number                                                                                                                                                                                              | Stoich. Coefficient | Kinetic Exponent | I.D. Number                                                                                                                                                           | Stoich. Coefficient | Kinetic Exponent |
| Propylene (C <sub>3</sub> H <sub>6</sub> )                           | (1) *                                                                                                                                                                                                    | -1 *                | 1 *              | (1) *                                                                                                                                                                 | -1 *                | 1 *              |
| Benzene (C <sub>6</sub> H <sub>6</sub> )                             | (2) *                                                                                                                                                                                                    | -1 *                | 1 *              |                                                                                                                                                                       |                     |                  |
| Cumene (C <sub>9</sub> H <sub>12</sub> )                             | (3) *                                                                                                                                                                                                    | +1 *                | - *              | (3) *                                                                                                                                                                 | -1 *                | 1 *              |
| DIPB**                                                               | (4)                                                                                                                                                                                                      |                     |                  | (4) *                                                                                                                                                                 | +1 *                | - *              |
| d) Kinetics:                                                         | $k_1 = k_{0,1} e^{-E/RT}$ <div><math>k_0 [\text{kmol}/\text{m}^3] [\text{s}]</math><br/><math>E [\text{kcal}/\text{kmol}]</math></div> <div><math>2.8 \times 10^7</math><br/><math>24,897^*</math></div> |                     |                  | $k_2 = k_{0,2} e^{-E/RT}$ <div><math>k_0</math><br/><math>2.32 \times 10^9</math><br/><math>35,070^*</math></div>                                                     |                     |                  |
| e) Transport & Thermal Properties                                    | Enthalpy: SRK ; k-values: SRK                                                                                                                                                                            |                     |                  |                                                                                                                                                                       |                     |                  |
| f) Reaction Phase                                                    | gas                                                                                                                                                                                                      |                     |                  |                                                                                                                                                                       |                     |                  |
| * Information Supplied (Reaction Specific)   ** P-Diisopropylbenzene |                                                                                                                                                                                                          |                     |                  |                                                                                                                                                                       |                     |                  |

Prob. 20.17



$$r_1 = k_1 C_P C_B \quad \& \quad r_2 = k_2 C_P C_C$$

$$r_C = r_1 - r_2 = k_1 C_P C_B - k_2 C_P C_C$$

$$r_D = r_2 = k_2 C_P C_C$$

$$\eta (\text{selectivity}) \equiv \frac{r_C}{r_D} = \frac{k_1 C_B C_P - k_2 C_P C_C}{k_2 C_P C_C}$$

$$\eta = \frac{k_1 C_B}{k_2 C_C} - 1$$

For ideal gas  $C_i = n_i/V = p_i/RT = \pi y_i/RT$

$$\eta = \frac{k_1 (\pi y_B/RT)}{k_2 (\pi y_C/RT)} - 1 = k_1 y_B / k_2 y_C - 1$$

$$\frac{k_1}{k_2} = \frac{k_{1,0} e^{-E_1/RT}}{k_{2,0} e^{-E_2/RT}} = \frac{k_{1,0}}{k_{2,0}} e^{(E_2 - E_1)/RT}$$

Let  $k_{1,0}/k_{2,0} = K$ , then

$$\eta = K [e^{(E_2 - E_1)/RT}] [y_B/y_C] - 1$$

$$K = 2.8 \times 10^7 / 2.32 \times 10^9 = 0.012$$

$$E_2 - E_1 = 35,070 - 24,897 = 10,173 \text{ kcal/mol}$$



# Prob. 20.17 (cont.)

Base case flow

$y_P$   
 $y_B$   
 $y_C$   
 $y_D$   
 $y(\text{Propane})$



Differential Reactor

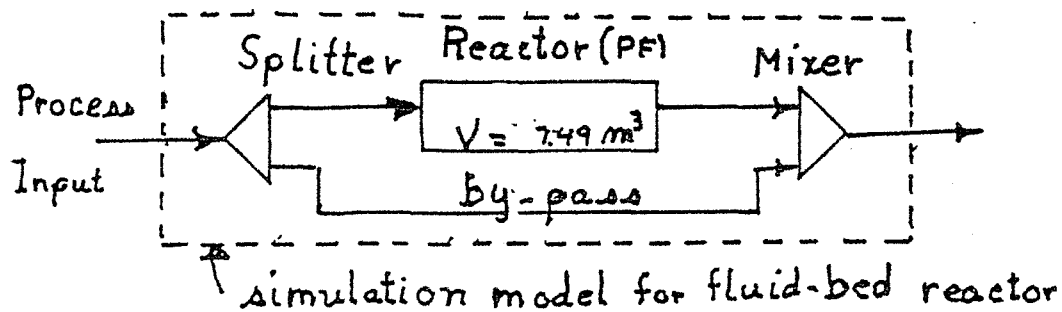
Additions made to adjust reactor mol fractions

$n$  (selectivity)

TREND TABLE FOR CUMULATIVE SELECTIVITY

| Terms in Selectivity Equation | $K$ | $e^{1073/RT}$ | $y_B/y_C$ | $n$ |
|-------------------------------|-----|---------------|-----------|-----|
| Temp ↑                        | ↓   | ↓             | ↓         | ↓   |
| Pressure ↑                    | ↓   | ↓             | ↓         | ↓   |
| $y_B$ ↑                       | ↓   | ↓             | ↑         | ↑   |
| $y_P$ ↑                       | ↓   | ↓             | ↓         | ↓   |
| $y(\text{Propane})$ ↑         | ↓   | ↓             | ↓         | ↓   |
| $y_C$                         | ↓   | ↓             | ↓         | ↓   |
| $y_D$                         | ↓   | ↓             | ↓         | ↓   |

Prob. 20.18 (start on page preceding Prob 20.17)



| Splitter Calculation         |               |              |         |
|------------------------------|---------------|--------------|---------|
| 5% Split fraction to by-pass |               |              |         |
|                              | Process Input | Reactor Feed | By-pass |
| $T^\circ\text{C}$            | 350           | 350          | 350     |
| P kPa                        | 3075          | 3075         | 3075    |
| Benzene                      | 213.6         | 202.9        | 10.68   |
| Propylene                    | 110.5         | 105.0        | 5.53    |
| Propane                      | 5.53          | 5.25         | 0.28    |

Reactor Specs:

Reactor Type: Plug Flow

Thermal Mode: Isothermal

Reactor Simulation: Specify volume  $\rightarrow$  Calculate conversion

| P.F. Reactor Calculation        |              |                   |
|---------------------------------|--------------|-------------------|
| Set Volume = $7.49 \text{ m}^3$ |              |                   |
|                                 | Reactor Feed | Reactor Discharge |
| T $^\circ\text{C}$              | 350          | 350               |
| P kPa                           | 3075         | 3025              |

Prob, 20.18 cont.

|           |       |       |
|-----------|-------|-------|
| Flows     |       |       |
| Benzene   | 202.9 | 121.4 |
| Propylene | 105   | 23.0  |
| Propane   | 5.25  | 5.25  |
| Cumene    | -     | 81.0  |
| DIPB      | -     | 0.49  |

### Mixer Calculation

|           | P.F Reactor<br>Output | By-pass<br>Stream | System<br>Output |
|-----------|-----------------------|-------------------|------------------|
| T °C      | 350                   | 350               | 350              |
| P kPa     | 3025                  | 3025              | 3025             |
| Flows     |                       |                   |                  |
| Benzene   | 121.4                 | 10.68             | 132.1            |
| Propylene | 23.0                  | 5.52              | 28.5             |
| Propane   | 5.25                  | 0.28              | 5.53             |
| Cumene    | 81.0                  | -                 | 81.0             |
| DIPB      | 0.49                  | -                 | 0.49             |

b) Cumene production drops from 91.4 to 81 (11%)

c) DIPB production drops from 2.6 to 0.49 (81%)

Note: For each mole of DIPB produced 2 moles of propylene are consumed & cannot be recovered & recycled

The unreacted propylene can be recovered & recycled & provide higher over-all propylene yields using new catalyst

Prob. 20.19

See Problem 20.18 for

- Flow-sheet of fluid bed reactor
- Calculation strategy
- Calculation for  $V = 7.49 \text{ m}^3$

For  $V = 10 \text{ m}^3$  (point on curve)

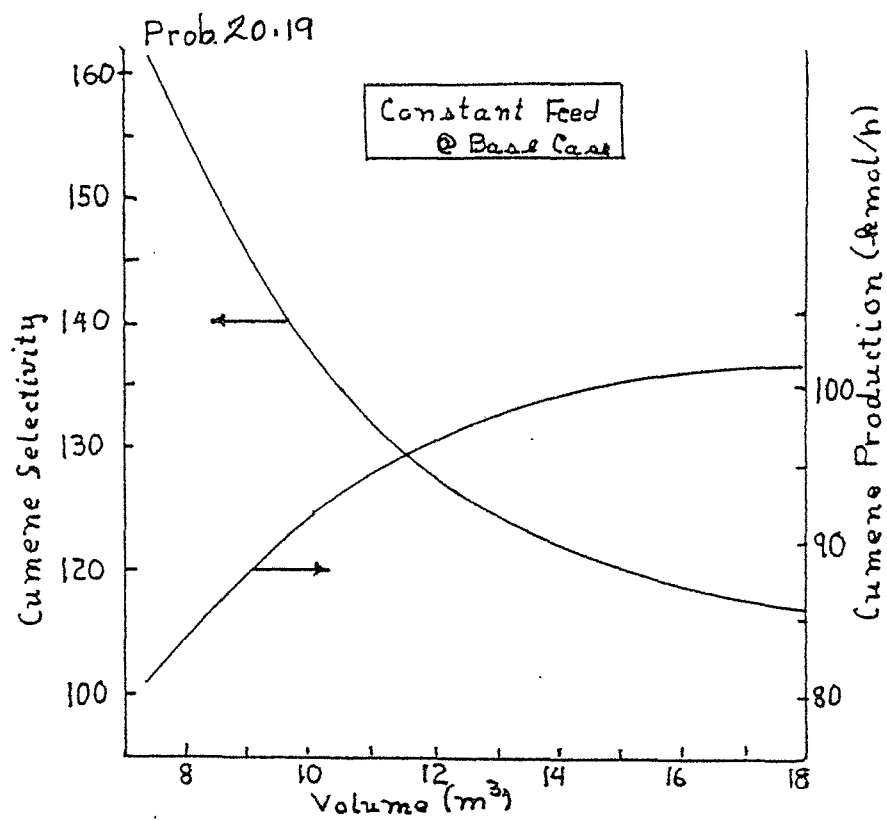
| Stream                      | Input | Output |
|-----------------------------|-------|--------|
| $T \text{ } ^\circ\text{C}$ | 350   | 350    |
| $P \text{ kPa}$             | 3075  | 3025   |
| Flows                       |       |        |
| Benzene                     | 213.6 | 121.6  |
| Propylene                   | 110.5 | 17.9   |
| Propane                     | 5.53  | 5.53   |
|                             | -     | 91.35  |
| DIPB                        | -     | 0.66   |

Cumene Production =  $91.35 \text{ kmol/h}$

Selectivity =  $91.35 / 0.66 = 138$

cont

Prob 20.19 cont.



Prob. 20.20

See Problem 20.18 for

a. Flowsheet of fluid bed reactor

b. Calculation strategy

Assuming ideal gas to maintain constant velocity

$$\text{Total flow (new)} = \text{Total Flow (Base case)} \frac{T_{bc}}{T_{new}}$$

$$\text{Total flow (new)} = \text{Total flow (base case)} \frac{T(\text{base case})}{T(\text{new})}$$

Sample Calculation: @ 400°C

$$\text{Flow (new)} = 329.7 \left( \frac{350+273}{400+273} \right) = 305.2$$

$$\cdot \text{Set Volume} = 7.49 \text{ m}^3$$

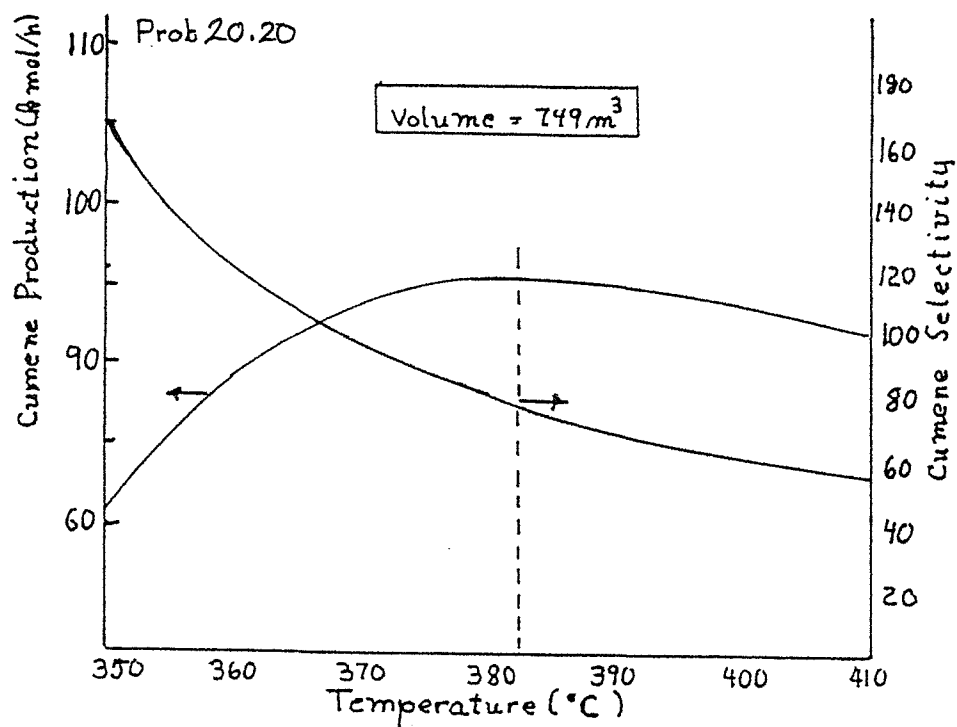
| Stream    | Input | System Output |
|-----------|-------|---------------|
| T °C      | 400   | 400           |
| P kPa     | 3075  | 3025          |
| Flows     |       |               |
| Benzene   | 197.8 | 102.3         |
| Propylene | 102.2 | 5.23          |
| Propane   | 5.19  | 5.19          |
| Cumene    | —     | 94.0          |
| DIPB      | —     | 1.50          |

$$\text{Cumene Production} = 94.0 \text{ kmol/h}$$

$$\text{Selectivity} = 94/1.5 = 62.7$$

Cont.

Prob. 20.20 cont.



Note: Max. occurs @  $380^{\circ}\text{C}$

Prob. 20.21 (Start on preceding page)

This problem confirms value in section 20.53 Option 3

Use Fig. 20.16 as base case

Change from base case

Stream 91 enters at bottom  
Stream 92 exits at top

Reactor Specs.:

Reactor type: Plug Flow

No. Tubes 448 Size L/D = 20 ft / 2 in

Thermal Mode: Utility

Flow: Counter-Current

$U = 60 \text{ W/m}^2 \cdot ^\circ\text{K}$

Calculation Mode:

Set volume  $\rightarrow$  Calculate conversion  
(5.1 m<sup>3</sup>)

Strategy Used (Trial & error):

Vary Feed Rate  $\rightarrow$  Calc. conversion  
until conversion of 0.9 is obtained

| Stream No.    | Process |       | Utility |     |
|---------------|---------|-------|---------|-----|
|               | 2a      | 3     | 91      | 92  |
| T °C          | 234     | 389   | 407     | 342 |
| Flow (kmol/h) |         |       |         |     |
| Hydrogen      | 0       | 40.26 | *       | *   |
| Acetone       | 0.19    | 40.44 | *       | *   |
| Water         | 22.06   | 22.06 | *       | *   |
| IPA           | 44.76   | 4.49  | *       | *   |
| HTM           | *       | *     | 58      | 58  |



Prob. 20.21 (cont.)

a) Increase in feed =  $67 - 57.8 = 9.2 \text{ kmol/h}$

b) % increase =  $[9.2 / 57.8](100) = 15.9\%$

Prob. 20.22 (start on page preceding Prob. 20.21)

Use Fig 20.16 as base case

Change from base case

Increase Stream 91 by 50%

$$\text{Flow} = (581)(1.5) = 872 \text{ mol/h}$$

Reactor Specs.:

Reactor type: Plug Flow

No. of tubes 448 Size L/D = 20 ft/2 in.

Thermal Mode: Utility

Flow: Co-Current

$$U = 60 \text{ W/m}^2 \text{ } ^\circ\text{K}$$

Calculation Mode:

Set volume ( $5.1 \text{ m}^3$ )  $\rightarrow$  Calculation conversion

|                    | Process |       | Utility |     |
|--------------------|---------|-------|---------|-----|
| Stream No.         | 2a      | 3     | 91      | 92  |
| T $^\circ\text{C}$ | 234     | 366   | 407     | 371 |
| Flow               |         |       |         |     |
| Hydrogen           | 0       | 36.18 | *       | *   |
| Acetone            | 0.16    | 36.34 | *       | *   |
| Water              | 19.04   | 19.04 | *       | *   |
| IPA                | 38.64   | 2.46  | *       | *   |
| HTM                | *       | *     | 87      | 87  |
| Total              | 57.84   | 94.02 | 87      | 87  |

a) IPA Conversion =  $[36.18 / 38.64] = 0.936$

b) % Increase in Acetone Produced

$$= \left( \frac{36.18}{34.78} - 1 \right) \times 100 = 4\%$$

## Acetone Production (Prob. 20.21-20.26)

## Reactor Simulation Set-up Values

1) Chemical Reaction:  $(\text{CH}_3)_2\text{CHOH} \rightarrow (\text{CH}_3)_2\text{CO} + \text{H}_2$

I.D. No. (1) (2) (3)

b) Rate Expression:  $-r_1 = k_1 C_1^*$

### c) Simulation Values

| Components                   | I. D.<br>Number | Stoich.<br>Coefficient | Kinetic<br>Exponent |
|------------------------------|-----------------|------------------------|---------------------|
| Isopropynol $[(CH_3)_2CHOH]$ | (1)*            | -1*                    | 1*                  |
| Acetone $[(CH_3)_2CO]$       | (2)*            | 1*                     | -*                  |
| Hydrogen $[H_2]$             | (3)*            | 1*                     | -*                  |

$$d) \quad k_1 = k_0 e^{-[E/RT]}$$

$$k_0 [\Delta]^{-1} = 3.516 \times 10^5 \text{ } ^\circ$$

$$E [\text{kJ/mol}] = 7238 \text{ MJ/kmol}$$

### e) Transport and Thermal Properties

Enthalpy: SRK\*      K-values: SRK

g) React Phase: (gas)

\* Information Supplied (Reaction Specific)

Prob. 20.23 (start on page preceding Prob. 20.21)

Use Fig. 20.16 as base case

Change from base case

Increase Stream 91 by 50%  
i.e. Flow =  $(58)(1.5) = 87 \text{ kmol/h}$

Reactor Specs.

Reactor: Plug Flow

No. of tubes 448 Size L/D = 20 ft / 2 in

Thermal Mode: Utility

Flow: Co-current

$U = 60 \text{ W/m}^2 \cdot \text{K}$

Calculation Mode:

Set volume ( $5.1 \text{ m}^3$ ) → Calculate conversion

Strategy Used (trial & error)

Vary Feed Rate → Calc conversion to obtain 0.9

|               | Process |       | Utility |       |
|---------------|---------|-------|---------|-------|
| Stream No     | 2a      | 3     | 91      | 92    |
| T °C          | 234     | 361   | 407     | 368.5 |
| Flow (kmol/h) |         |       |         |       |
| Hydrogen      | 0       | 39.09 | *       | *     |
| Acetone       | 0.18    | 39.78 | *       | *     |
| Water         | 21.40   | 21.40 | *       | *     |
| IPA           | 43.42   | 4.33  | *       | *     |
| HTM           | *       | *     | 87.0    | 87.0  |
| Total         | 65.00   | 104.1 | 87.0    | 87.0  |

a) Process Feed = 65 kmol/h

b) % increase in acetone produced =  $\left[ \frac{39.09}{39.78} - 1 \right] (100)$   
= 12.4%

Prob. 20.24 (start on page preceding Prob. 20.21)

Use Fig. 20.16 as base case

Changes from base case

Increase Stream 91 (Utility) by  $50^{\circ}\text{C}$

Reactor Specs.:

Reactor: Plug Flow

No. of tubes 488 Size L/D = 20 ft/2 in

Thermal Mode: Utility

Flow: Co-current

$U = 60 \text{ W/m}^2 \cdot ^{\circ}\text{K}$

Calculation Mode:

Set volume ( $5.1 \text{ m}^3$ )  $\rightarrow$  Calculate Conversion

Strategy (trial & error)

i. Select feed rate

ii. Calculate conversion (@ 0.9 solution achieved)

|                          | Process |       | Utility |     |
|--------------------------|---------|-------|---------|-----|
| Stream No.               | 2a      | 3     | 91      | 92  |
| $T^{\circ}\text{C}$      | 234     | 373.7 | 457     | 383 |
| Flow ( $\text{kmol/h}$ ) |         |       |         |     |
| Hydrogen                 | 0       | 51.41 | *       | *   |
| Acetone                  | 0.24    | 51.65 | *       | *   |
| Water                    | 28.15   | 28.15 | *       | *   |
| IPA                      | 57.11   | 5.71  | *       | *   |
| HTM                      | *       | *     | 58      | 58  |
| Total                    | 85.5    | 136.9 | 58      | 58  |

a) Process Feed =  $85.8 \text{ (kmol/h)}$

b) % Increase =  $\left(\frac{51.41}{34.60} - 1\right)(100) = 47.7$

Note: This value in Section 20.3 Option 1 & Fig. 18.12

Prob. 20.25 (start on page preceding Prob 20.21)

Use Fig. 20.16 as base case

Changes from base case

Vary utility flow  
Stream 91

Reactor Specs:

Reactor: Plug Flow

No. of tubes 448 Size  $L/D = 20\text{ft./2in.}$

Thermal Mode: Utility

Flow: Co-current

$U = 60\text{ W/m}^2\text{ }^\circ\text{K}$

Calculation Mode:

Set volume ( $5.1\text{m}^3$ )  $\rightarrow$  Calculate conversion

Strategy Used:

Select several utility flow rates

For each flow

Vary Feed Rate  $\rightarrow$  Calculate to obtain

0.9 conversion (by trial & error)

Example Stream =  $29\text{ kmol/h}$

See Flow Table on next page

Co-ordinates on performance curve

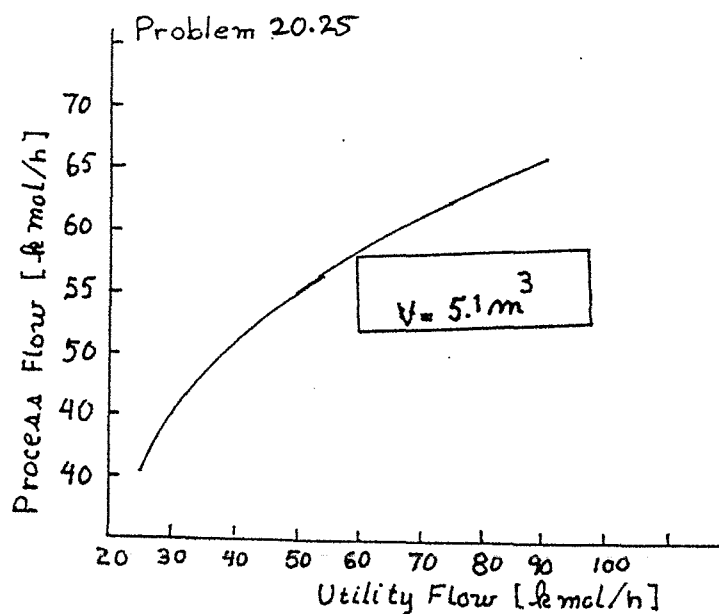
$X = 29\text{ kmol/h}$

$Y = 44.2\text{ kmol/h}$

cont.

Prob. 20.25 cont

| Stream No | Process |       | Utility |       |
|-----------|---------|-------|---------|-------|
|           | 2a      | 3     | 91      | 92    |
| T°C       | 234     | 329   | 407     | 332.2 |
| Flow      |         |       |         |       |
| Hydrogen  | 0       | 26.59 | *       | *     |
| Acetone   | 0.12    | 26.71 | *       | *     |
| Water     | 14.55   | 14.55 | *       | *     |
| IPA       | 29.53   | 2.94  | *       | *     |
| HTM       | *       | *     | 29.0    | 29.0  |
| Total     | 44.2    | 70.8  | 29.0    | 29.0  |



Prob. 20.26 (start on page preceding Prob. 20.21)

Use Fig. 20-1b as a base case

Increase Utility (Stream 91) by  $50^\circ$

$$U = f(\text{Feed rate})$$

Reactor Specs.

Reactor type: Plug Flow

No. of tubes: 448      Size:  $L/D = 20 \text{ ft./2 in.}$

Thermal Mode: Utility

Flow: Co-current

$$U = 60 \text{ W/m}^2 \cdot ^\circ\text{K}$$

Calculation Mode: Set volume ( $5.1 \text{ m}^3$ )  $\rightarrow$  Calculate conversion

Strategy:

Select feed-rate

Evaluate  $U$

Obtain Conversion until 0.9 is obtained

a. Given: 
$$h = h_{bc} \left( \nu / \nu_{bc} \right)^{0.8} \quad (\text{gas-side})$$

Assume:

1. Pressure drop in reactor constant
2. Temperature profile changes modest
3. Ideal gas

Then  $\nu \propto \text{Flowrate}$

$$h = h_{bc} \left( F / F_{bc} \right)^{0.8} = 60 \left( \frac{F}{57.84} \right)^{0.8} \frac{\text{Watt}}{\text{m}^2 \cdot ^\circ\text{K}}$$

Assume

4. Gas (Process side) Film Controls cont.



$$U \approx h_i = 60 (F/57.84)^{0.8} \text{ Watt/m}^2 \cdot \text{K}$$

b) Assume Feed = 93.5 kmol/h

$$U = 60 (93.5/57.84)^{0.8} = 88.1 \text{ kmol/h}$$

Flow Table

|               | Process |        | Utility |        |
|---------------|---------|--------|---------|--------|
|               | Input   | Output | Input   | Output |
| T °C          | 234     | 370.9  | 457     | 376.3  |
| Flow (kmol/h) |         |        |         |        |
| Hydrogen      | -       | 56.18  | *       | *      |
| Acetone       | 0.26    | 56.45  | *       | *      |
| Water         | 30.78   | 30.78  | *       | *      |
| IPA           | 62.46   | 6.27   | *       | *      |
| HTM           | *       | *      | 58      | 58     |
| Total         | 93.5    | 149.68 | 58      | 58     |

$$\text{Conversion} = 1 - (6.27/62.46) = 0.9$$

c) Higher heat transfer coefficient provides greater heat transfer increasing heat available for the endothermic reaction. This results in increased acetone production

$$@ U = 60 \text{ kmol/h ;}$$

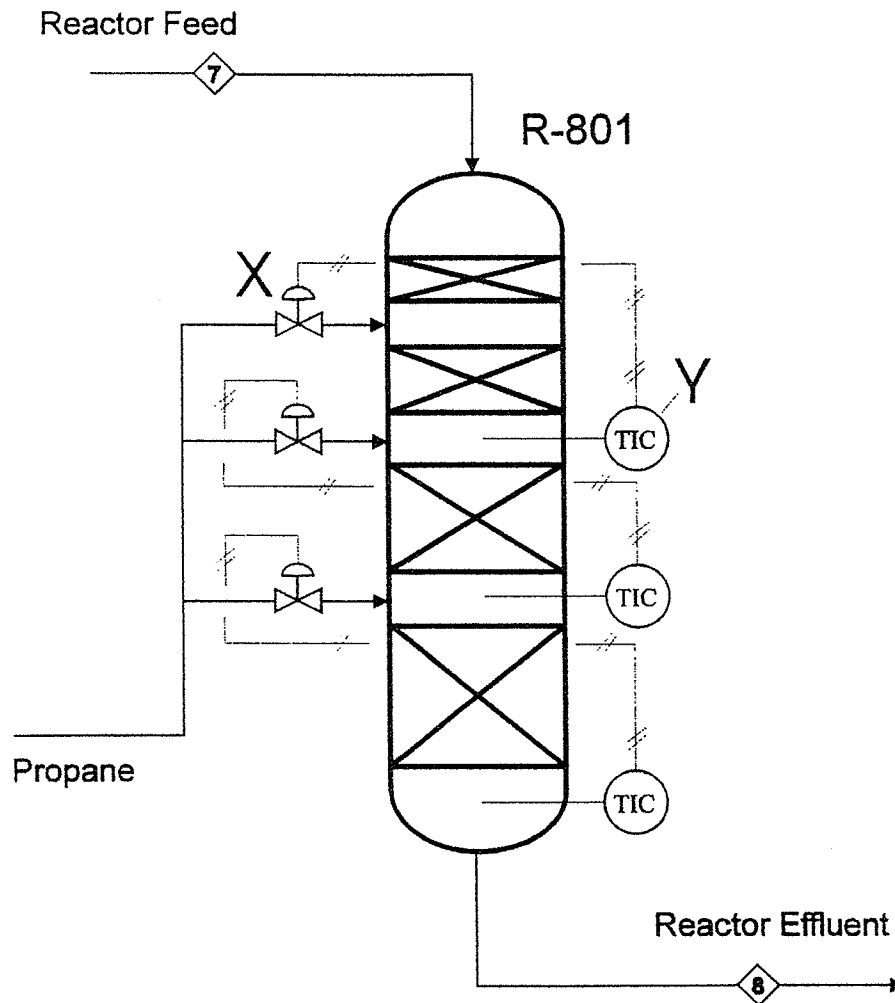
$$\text{Production} = 51.41 \text{ kmol/h [see Prob. 14/18]}$$

$$\% \text{ increase} = (1 - 56.18/51.41)(100) = 9.2\%$$

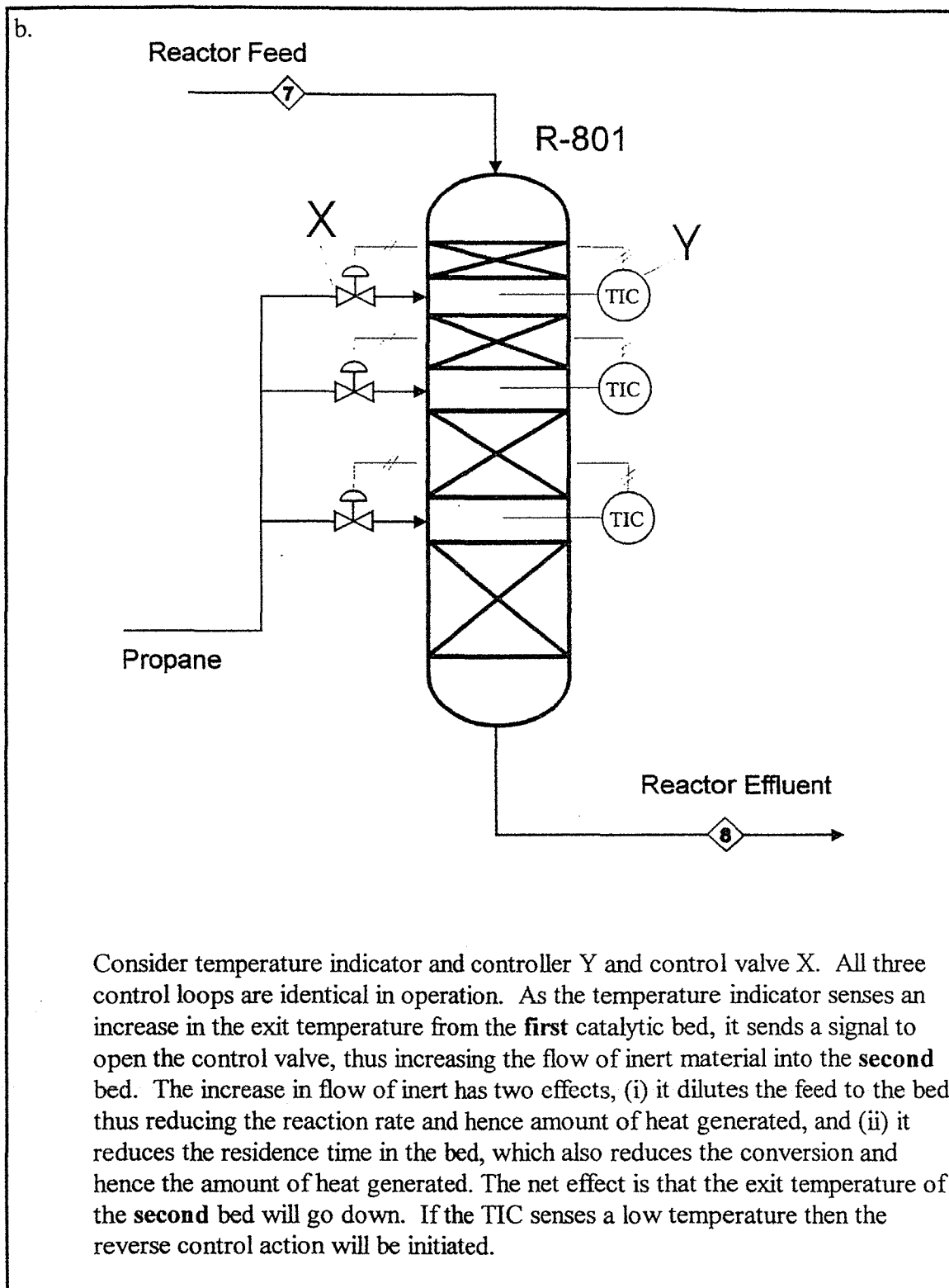
## Chapter 2 |

### 21.1 Cumene Reactor, R-801 - Alternative configuration and control scheme.

a.



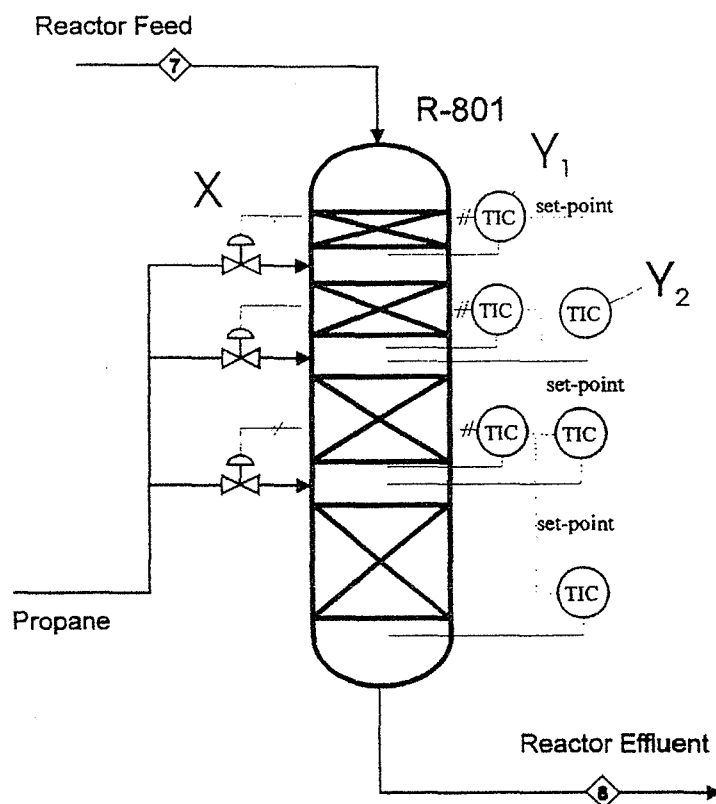
Consider temperature indicator and controller Y and control valve X. All three control loops are identical in operation. As the temperature indicator senses an increase in the exit temperature from the **first** catalytic bed, it sends a signal to open the control valve, thus increasing the flow of inert material into the **first** bed. The increase in flow of inert has two effects, (i) it dilutes the feed to the bed thus reducing the reaction rate and hence amount of heat generated, and (ii) it reduces the residence time in the bed, which also reduces the conversion and hence the amount of heat generated. The net effect is that the exit temperature of the **first** bed will go down. If the TIC senses a low temperature then the reverse control action will be initiated.



- c. The control schemes in parts a. and b. above are both examples of feedback control strategies. Both these control schemes have merit and the better scheme depends to a large extent on the dynamics of the process and the disturbance that it is desired to control. The cumene producing reaction is quite exothermic so that the temperature rise over each bed may be quite large. If the temperature of the catalytic beds is very sensitive to small changes in feed temperature then the feedback scheme of part b. would be preferred, since the bed inlet temperature is monitored and controlled directly. If, on the other hand, the sensitivity to small changes in feed temperature is low then the feedback control scheme of part a. could be used to adjust for changes in catalyst activity occurring during operation. Note that this feedback scheme does not monitor and control directly the feed temperature to each bed. Rather the required feed temperature change is inferred from the exit temperature of the bed.

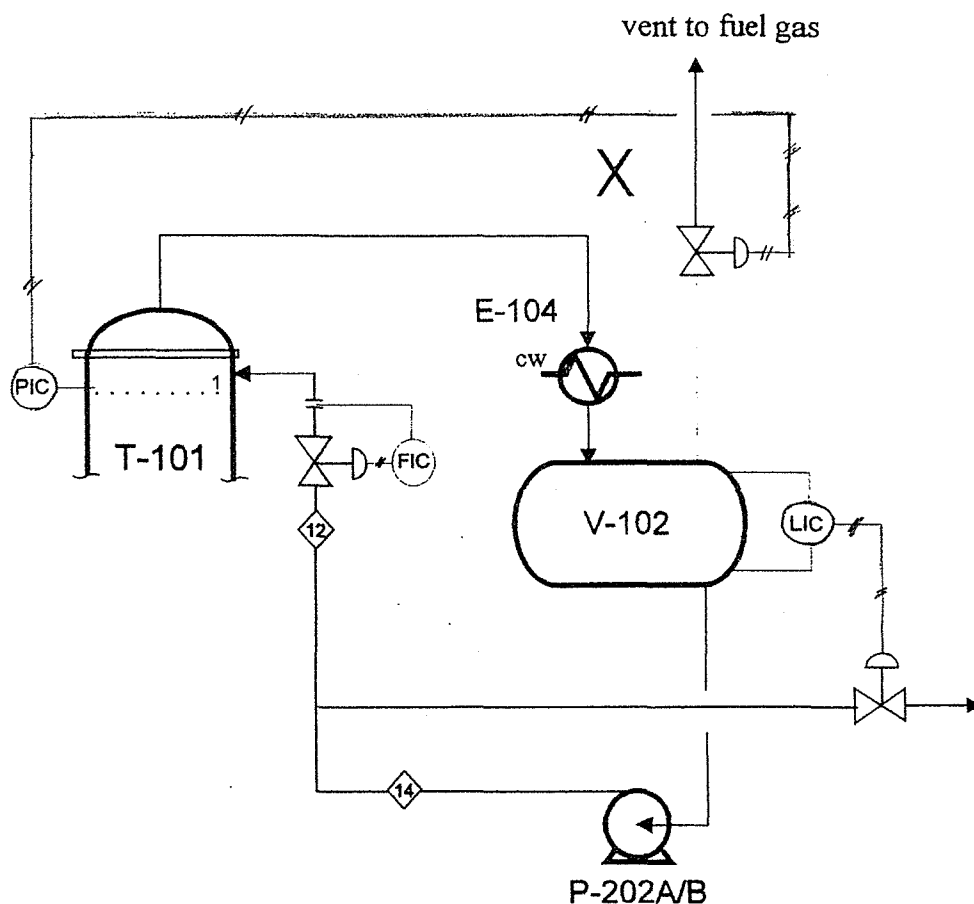
In reality a much more complicated scheme to control the reactor beds would be employed. This scheme would use both the feedback loops considered above cascaded together. In addition, many safety features including emergency shut down procedures would be incorporated into the design of the control system. Most probably, several temperature indicators (thermocouples) would be installed in each catalyst bed so that the axial (and possibly radial) temperature profile could be monitored. For this type of reactor set up, the highest temperature normally occurs at the exit, however, it is always a good idea to monitor the bed temperature profiles. One additional feature that would probably be included is to add a propane feed to the first bed. This is a good idea since the first bed currently has no control loop associated with it and the reaction rates in this bed will be the highest due to the high concentrations of reactants.

d. A combined cascade control system is shown below.



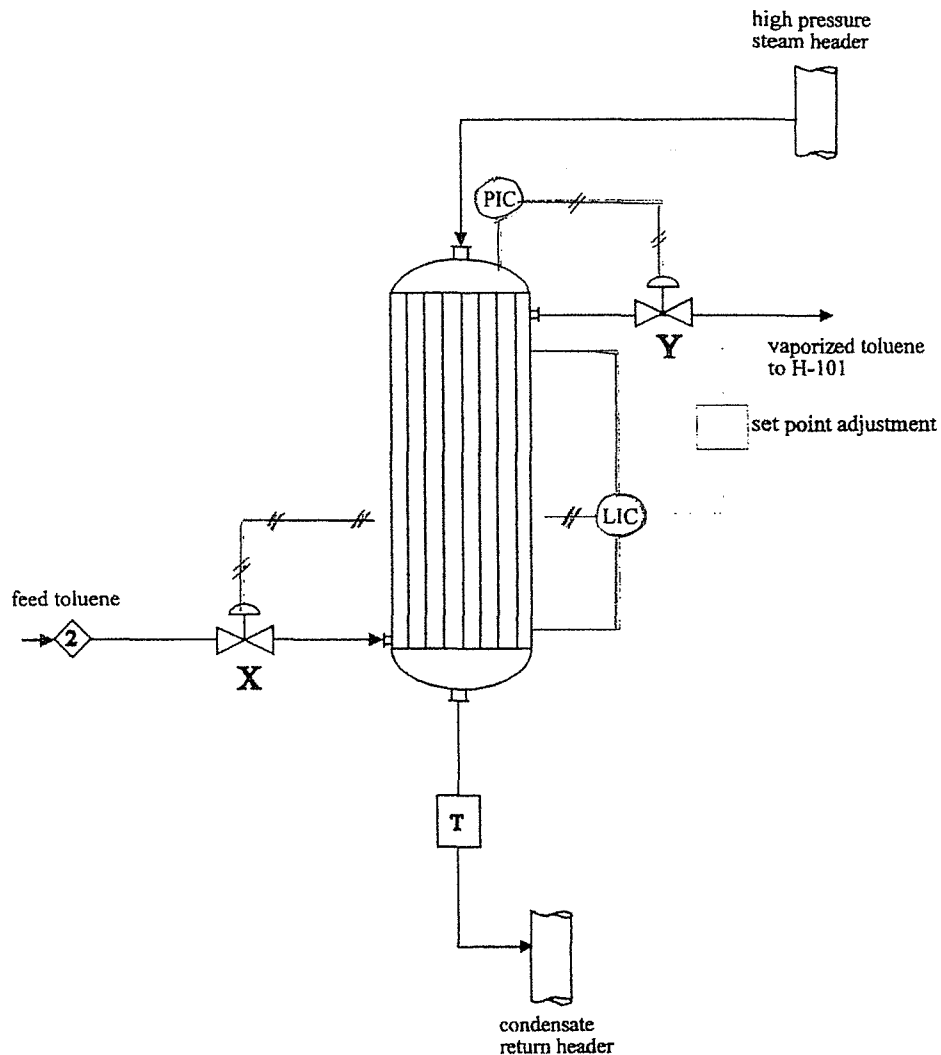
Consider temperature indicator and controllers  $Y_1$  and  $Y_2$  and control valve  $X$ . All three control loops are identical in operation. As the temperature indicator,  $Y_1$ , senses an increase in the exit temperature from the **first** catalytic bed, it sends a signal to open the control valve, thus increasing the flow of inert material into the **second** bed. This control loop essentially monitors and controls the inlet temperature of the second bed. The second temperature indicator,  $Y_2$ , monitors the exit temperature from the second bed. Let us say that  $Y_2$  senses a lower temperature than desired, indicating less conversion in the second bed perhaps due to a decrease in catalyst activity. This control loop will now adjust the set-point of the first controller,  $Y_1$ . In this case it would increase the set-point temperature, causing less propane to be added with the effect of increasing the conversion in the second bed and thus restoring the desired exit temperature (and conversion). Due to the complexity and non-linearity of a system such as this one, the overall control scheme would probably incorporate computer modeling of the reaction system. This would allow the accurate prediction of the changes in the inlet feed stream required to maintain the desired conversion and the whole process would be monitored by computer.

21.2 The overhead system for T-101 is shown in the sketch below.



- a. As the column pressure begins to drop, a signal is sent to control valve X to close slightly. This in turn causes the flow of non-condensables going to fuel gas to drop. This means that the non-condensables start to build up in the reflux drum, V-102, and the pressure in the drum starts to increase. As the pressure in V-102 increases this causes the pressure in the column, T-101, to increase. Thus the desired control action is achieved.
- b. If the amount of non-condensables fed to the column suddenly increases then there will be an increase in the amount of non-condensables in V-102. With no control action the valve X would not change position and the non-condensables would build up in V-102 causing the pressure to rise in V-102, E-104, and T-101. Because of the control loop, as the pressure in T-101 starts to increase a signal is sent to open valve X, which in turn allows more flow to the fuel gas and hence reduces the pressure in V-102. Thus the desired control action is achieved.

21.3 Toluene vaporizer - control amount of toluene and pressure of toluene leaving the vaporizer.



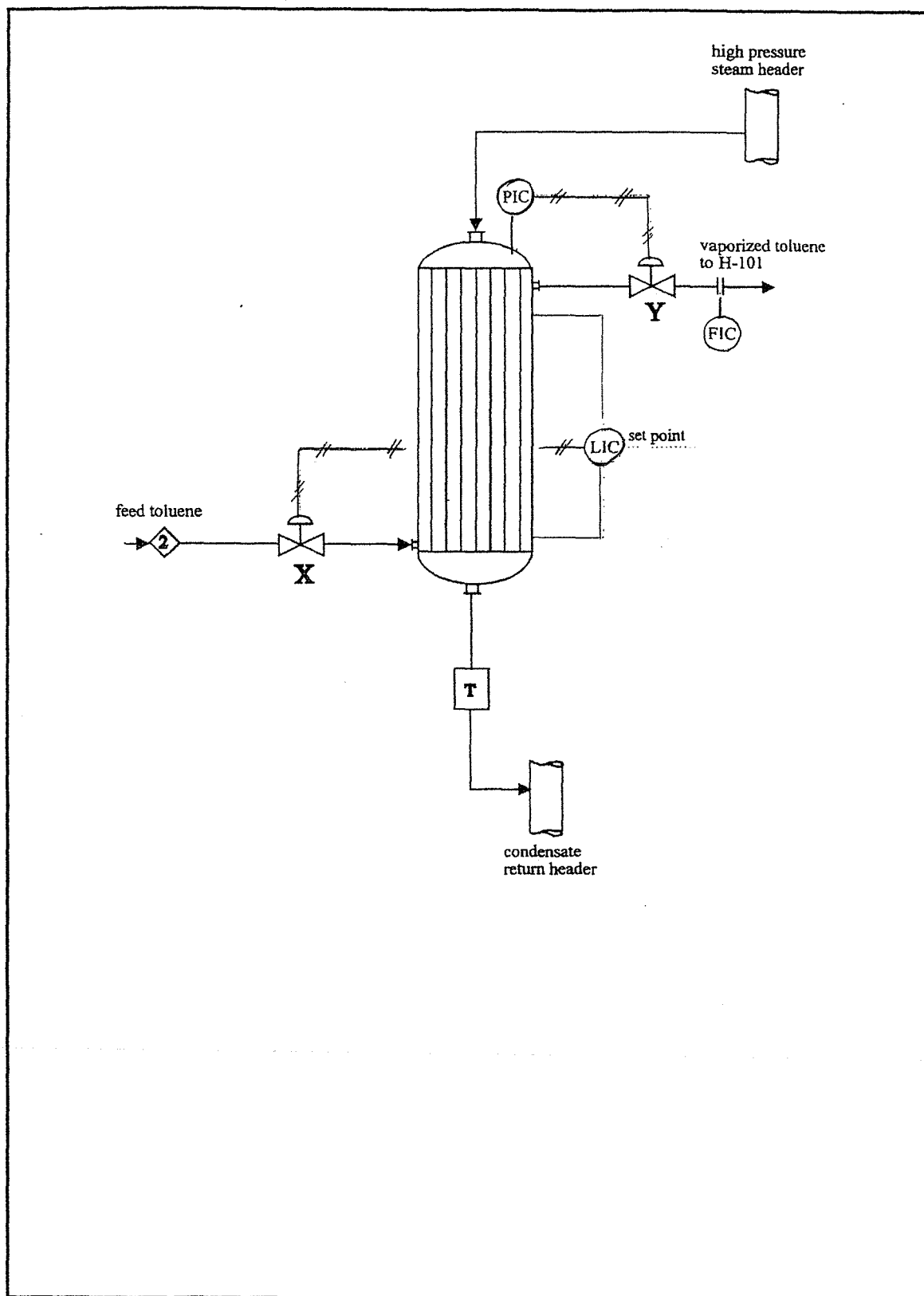
The above control scheme includes two feedback loops. The first loop consists of the level indicator controller (LIC) and the control valve X. This loop controls the amount of flow of vaporized toluene sent to the fired heater. The loop is a simple material balance control which regulates the level of liquid toluene in the vaporizer by opening or closing the valve on the toluene feed line, Stream 2. The second control loop adjusts the exit control valve Y on the toluene vapor line leaving the exchanger to adjust the pressure of the vapor stream. If the pressure at which the toluene is to be vaporized is changed, then it will be necessary to adjust the level of the liquid toluene in the exchanger. An automatic set point adjustment scheme is shown as the square box linking the electrical signal (dotted line) between the PIC and the LIC.

The reason that the set point adjustment is required is best illustrated by considering the case when the required pressure (and hence temperature) of the toluene vapor is to be increased (or decreased). In order to increase the pressure in the shell side of the exchanger, we would increase the set point for the PIC which in turn would cause control valve Y to close. Without changing the set point on the LIC the flow of vaporized toluene would decrease. This is because if the level of liquid in the exchanger is kept constant, then from the exchanger performance equation we have

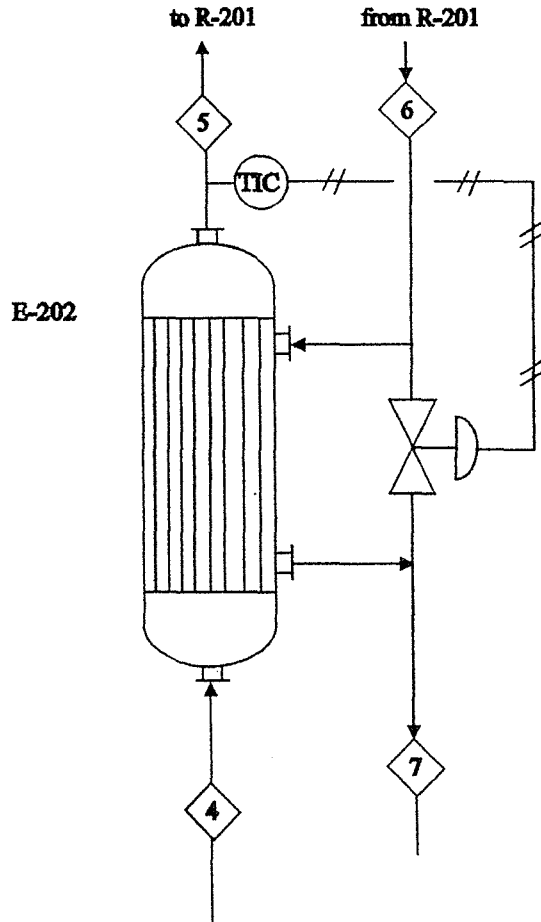
$$Q = UA\Delta T_{lm}$$

By maintaining the level of liquid constant, the effective  $A$  and  $U$  remain constant. Note that the majority of the heat transfer occurs in the liquid covered portion of the tubes, i.e., the vapor phase heat transfer above the liquid level is assumed to be negligible due to the very low gas side film coefficient. Since the pressure of the shell side of the exchanger has increased the temperature at which the toluene boils (vaporizes) increases (recall Antoine's equation). The temperature driving force,  $\Delta T_{lm}$ , thus decreases and consequently  $Q$  will also decrease - this results in less toluene being vaporized. We can remedy this situation by increasing the value of  $A$  (the portion of the tubes exposed to boiling toluene) - thus we adjust the set point of the LIC so that the level increases. This adjustment in level will be small for small changes in pressure and could be done manually. Alternatively, there is no direct flow control on the vapor leaving the exchanger, an alternative scheme would be to use a cascade control system with a flow element on the toluene exit stream adjusting the set point on the LIC as shown on the next page.





21.4 DME reactor feed exchanger, E-202



- a. If the temperature of stream 5 increases then the TIC would respond by opening up the control valve. This would cause less of Stream 6 to flow through E-202 and hence reduce the amount of heat exchanged in E-202 and thus the exit temperature, Stream 5, would decrease. For a decrease in the temperature of stream 5, the reverse control action would occur.
- b. Fouling in the heat exchanger - The net effect of exchanger fouling is to decrease the overall heat transfer coefficient  $U$ . The amount of heat that can be exchanged is given by  $Q = UA\Delta T_{lm}F$ . Since  $A$  is constant and  $F$  will change very little, the only way to maintain  $Q$  as  $U \downarrow$  is to increase  $\Delta T_{lm}$ . This can be accomplished by increasing the flow of Stream 6 through the exchanger, i.e., closing the control valve. This will result in a higher exit temperature of the shell side fluid leaving E-202 and hence will increase the temperature driving force in the heat exchanger.

If fouling becomes excessive then the control valve may close shut and further control action will be lost.

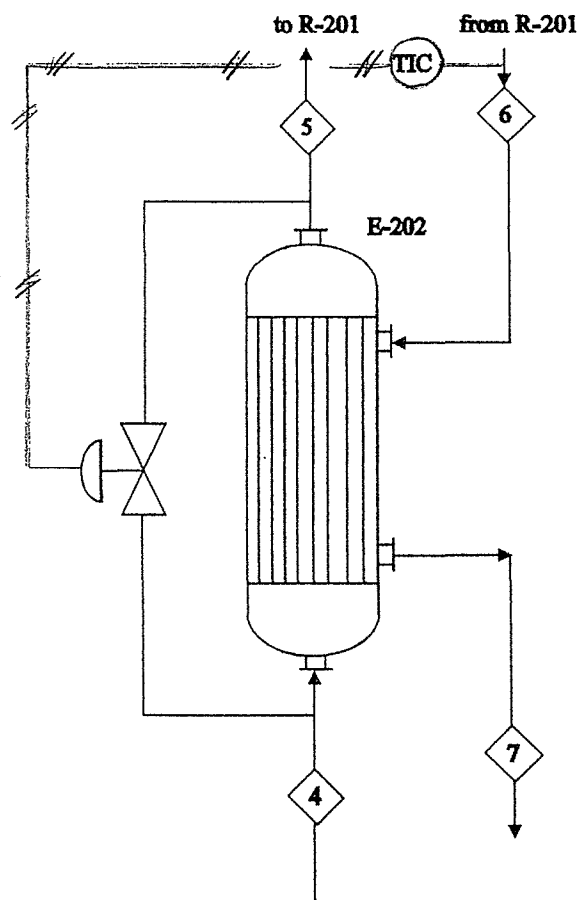
Loss in catalyst activity

The loss in catalyst activity will result in a decrease in conversion and hence the exit temperature (temperature of Stream 6) will slowly drop. This will result in a drop in the temperature of Stream 5, and will initiate the TIC to slowly close the control valve. The net effect of this is to force more material through the shell side of the exchanger and this will increase  $U$  and increase the exit temperature of the material leaving the shell side of E-202. Both these actions will compensate for the decrease in the  $\Delta T_{lm}$  which occurs due to the decrease in temperature of Stream 6.

Unlike the fouling of the heat exchanger considered above, the loss in activity has a major impact on the down stream processing. This is because Stream 7 now contains less DME and more methanol and columns, T-201 and T-202, will now be affected.

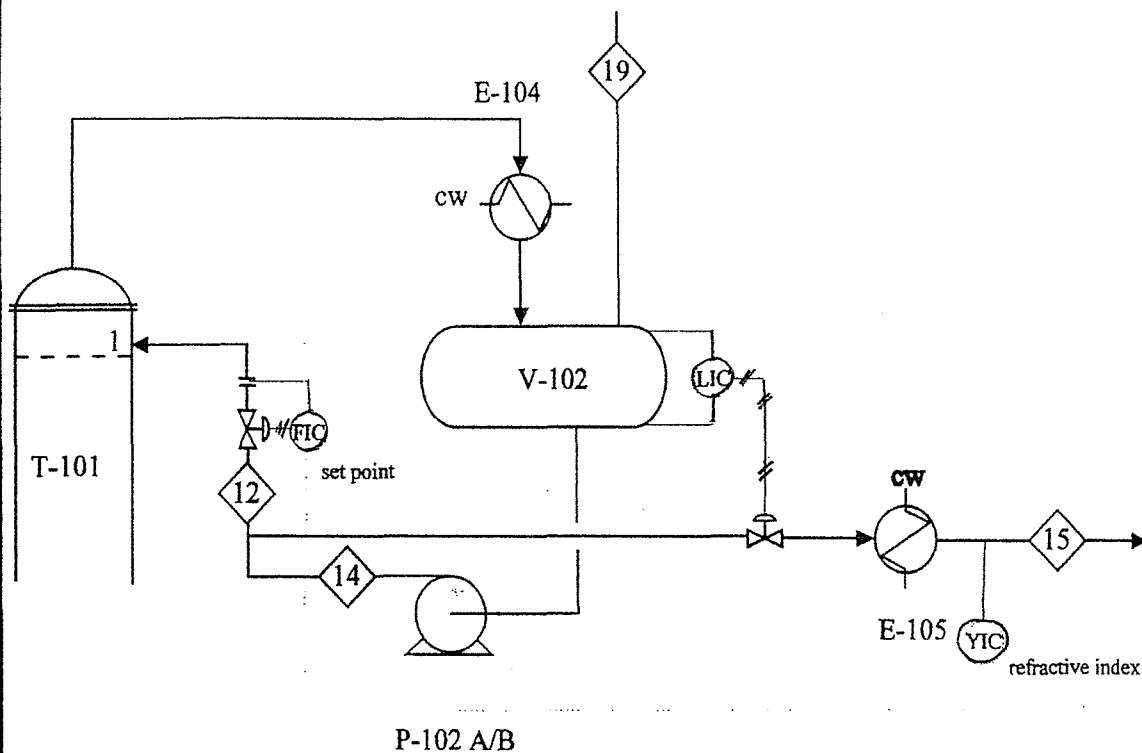
- c. The strategy used is feedback control. It is feedback because the control action is initiated by changing an input variable (amount of flow of Stream 6 through E-202) upon measuring a deviation in the output variable (the temperature of Stream 5 leaving E-202).

- d. There are several ways to do this. The most direct way is to switch the bypass to the inlet side of the exchanger as shown below.



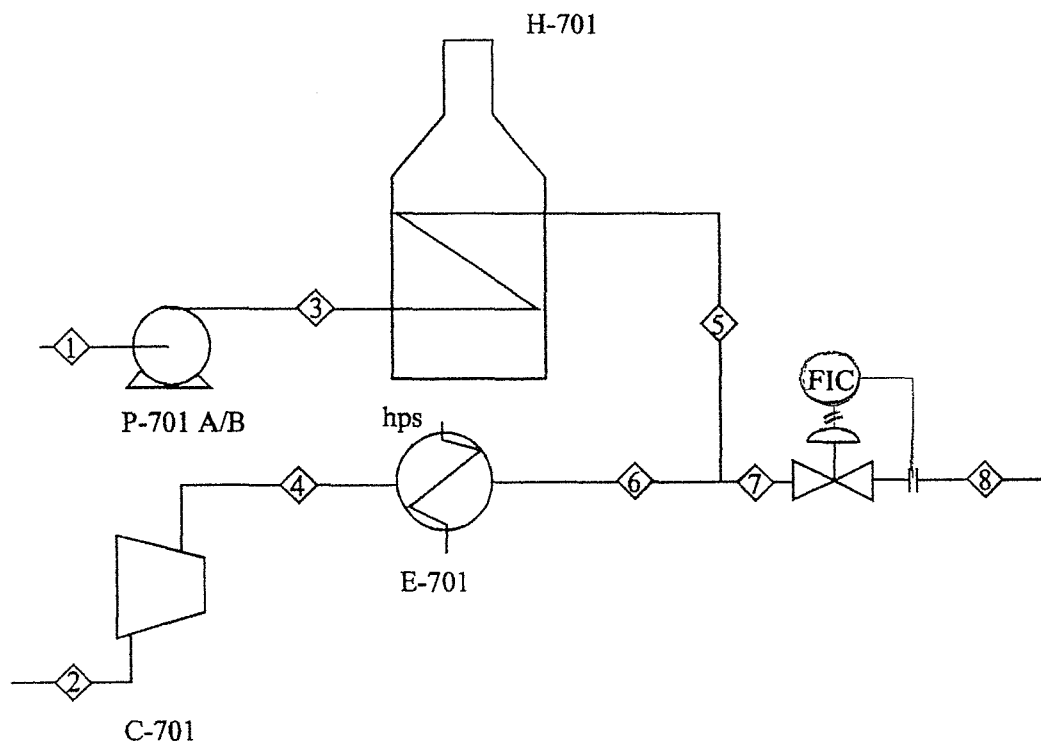
Note that for this set up we can manipulate the amount of feed material passing through E-202 in order to maintain the exit reactor temperature. For the case when the catalyst activity changes, this strategy would allow us to control directly the conversion in the reactor, since the temperature of Stream 6 is a direct indication of the conversion.

### 21.5 Regulation of overhead product purity of Benzene Tower, T-101



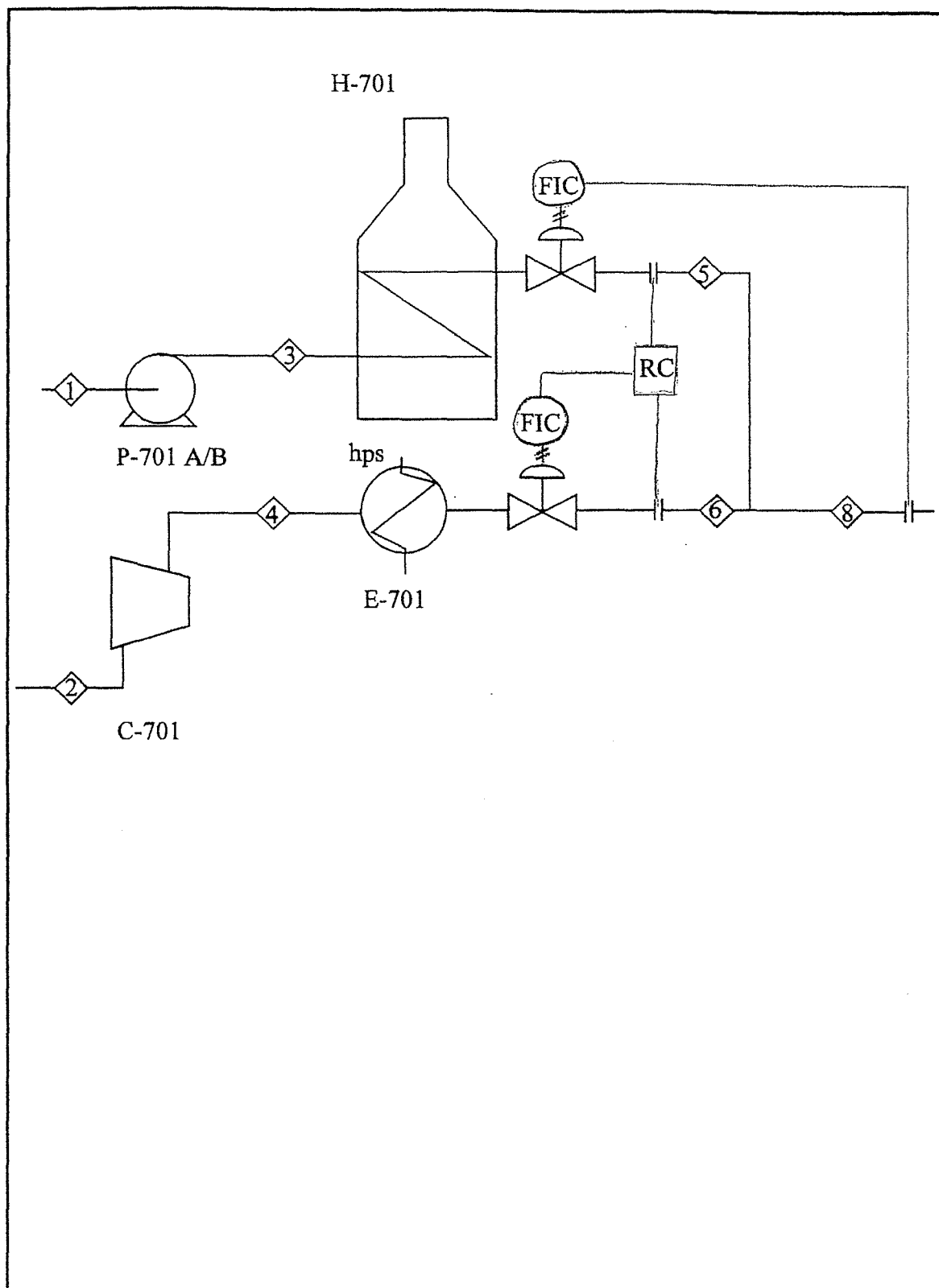
- The purity of the overhead product stream is controlled by a refractive index monitor which adjusts the set point of the flow controller on the reflux line. Thus the reflux ratio is adjusted (up or down) depending on the purity of Stream 15, as inferred by the refractive index of this stream.
- This type of control system is an example of a cascade control system.

21.6 Currently the control scheme looks like the following.



The current control system only regulates the total flow of material going to the reactor, in Stream 8. If the control valve closes slightly then the pressure in Stream 7 changes and thus the pressure drops in the upper line 1-3-5 and the lower line 2-4-6 must also change. However, the air flow and the naphthalene flow are determined by the characteristics of C-701 and P-701 respectively. The pump and compressor curves will have different head vs flowrate relationships. The result is that as the total flow through the control valve changes the ratio of air to naphthalene also changes.

The obvious remedy to this situation is to place separate flow controllers on Streams 5 and 6 as shown below. In addition, a ratio controller is added which maintains the desired ratio of air to naphthalene in Stream 7. Finally, the total flow rate to the reactor is maintained by adjusting the set point of the naphthalene valve. This strategy allows the total flow to the reactor to be controlled while maintaining the desired air to naphthalene ratio.



## 21.7 Driving an automobile

A responsible person drives an automobile using a combination of feedforward and feedback control strategies with each cascading on the other. Consider driving on a road in a suburban area. As we move down the street, we see a stop sign 50 yards ahead of us. We gauge the required braking needed to stop smoothly at the stop sign and depress the brake pedal. As the car begins to slow we continually monitor the pressure on the brake pedal and may increase or decrease the pressure depending on the rate at which the car slows. All of a sudden we see out of the corner of our eye a small child on a bicycle in a driveway just a few yards ahead. We brake much harder and perhaps come to a fairly quick stop. Having avoided an accident with the child, we continue to the stop sign.

As mentioned earlier, we automatically use feed forward (anticipatory) and feedback (reflective) control strategies. The sighting of the stop sign and the child are feedforward actions while the monitoring of the amount of braking is a feedback strategy.



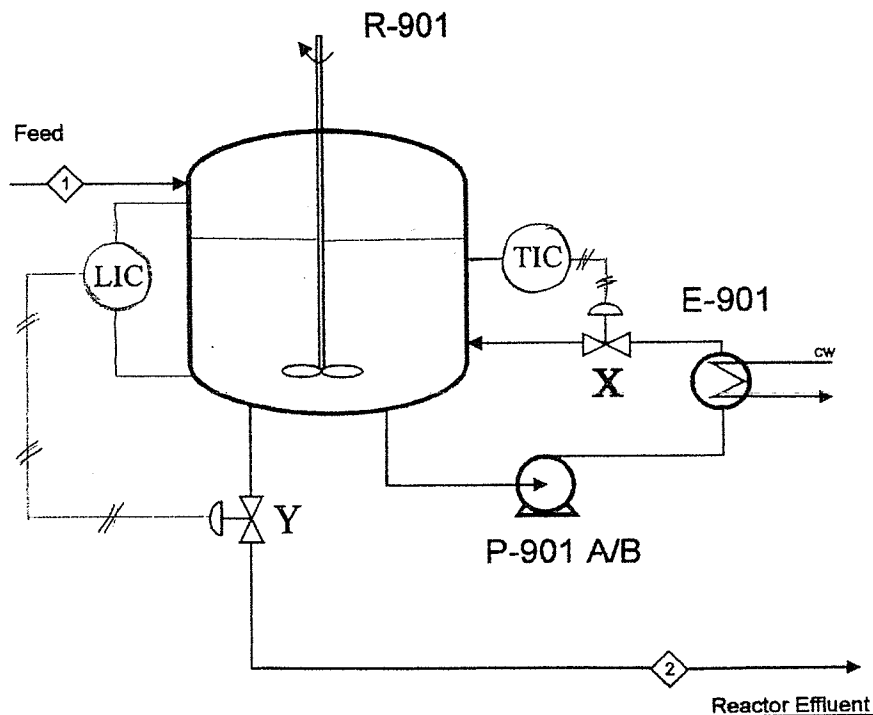
21.8 a. Regulate the temperature of the reactor

This is easily accomplished by placing a control valve (X) on the discharge of pump, P-901, and using a temperature indicator controller (TIC) placed either in the reactor or in Stream 2 to adjust the setting of the valve. For example, if the temperature in the reactor was too high, then the TIC would cause the control valve X to open causing more fluid to circulate through E-901 and hence increasing the heat removal from the reactor - causing the reactor temperature to drop. This control loop is shown in the diagram below.

b. Regulate the inventory of the reactor

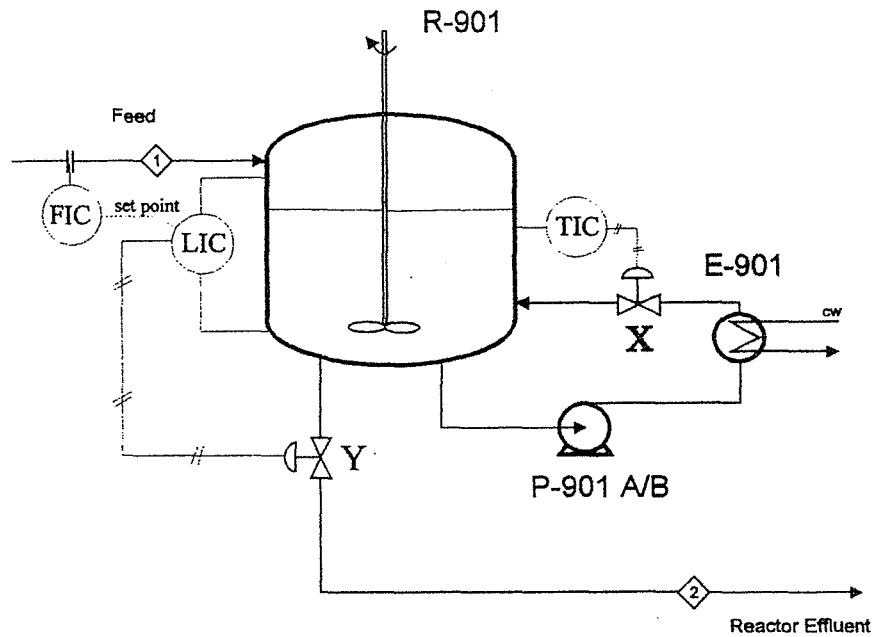
Again this can be easily accomplished by using a liquid level controller (LIC) in the reactor to adjust the setting of a control valve (Y) in the exit line from the reactor, Stream 2. If a change in flowrate of Stream 1 were to take place, say a decrease in flow. Then the level in the reactor would start to drop, the LIC would sense this change and close the control valve Y. This in turn would reduce the flow of liquid from the reactor thus allowing the level in the reactor to increase back to its original level. This control loop is shown in the diagram below.

Note that the rate at which fluid is circulated through the external heat removal loop (P-901 and E-901) does not affect the level in the reactor.



### Maintaining constant conversion/residence time

One simple way of maintaining a constant residence time in the reactor would be to use a feed forward signal from a flow element placed in the feed line, Stream 1, to adjust the set point of the level controller. Thus if the flow of Stream 1 were to drop, the level would be lowered to compensate and if the flow were to increase then the reverse action would be initiated.

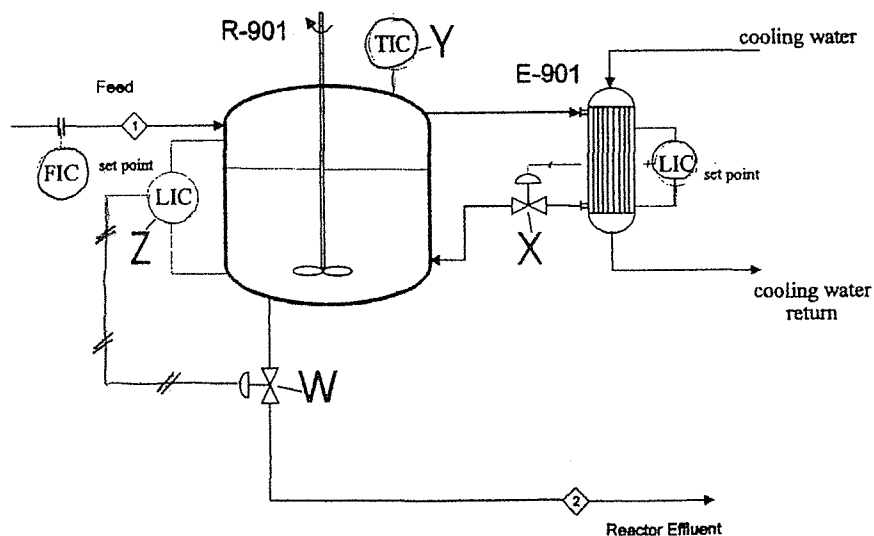


## 21.9 Control of the temperature in the reactor

We will assume that the external heat exchanger, E-901, condenses the vapor in the shell side of the heat exchanger and that the level of condensate in the exchanger can be measured. The control strategy is to adjust the level of condensate in E-901 to expose more or less heat transfer area to the vapor. Since it is necessary to control the level of condensate, a material balance control loop is required. The control valve X is adjusted with the signal from the level controller (LIC) on the condensate level in E-901. A temperature control then adjusts the set point of the level indicator to respond to changes in the reactor temperature. For example, if the reactor temperature were to increase, then a signal would be sent from the temperature sensing element, Y, to the set point of the level controller. The level set point would be decreased so that more exchanger surface area is exposed to condensing vapor and hence more condensation can take place. This would cause the temperature of the reactor to drop. This control scheme is shown in the diagram below.

### b. Control of the reactor residence time

In order to maintain a constant inventory in the reactor a feedback material balance control loop is provided for the reactor. A level indicator controller, Z, controls the position of the control valve, W. In order to maintain a constant residence time the level in the reactor must change in sympathy with the changing feed flow rate. Therefore, a feedforward control loop is added which senses the flow of Stream 1 and adjusts the level in the reactor by changing the level set point, Z. These loops are shown in the diagram below.



c. Explain how this system responds to an increase in flow of Stream 1

As Stream 1 increases, the flowrate controller sends a signal to the level indicator to increase the level set point. Initially, the control valve (W) on the liquid exit line from the reactor will not change position and the level slowly rises. When the level in the reactor rises above the new set point, the material balance control loop comes into play and the control valve (W) starts to open and eventually a new steady state will be reached. Note that at steady state the liquid feed into and out of the reactor must be the same - all that has happened is that the level in the reactor, i.e., the liquid inventory, has increased thus maintaining the residence time of the liquid in the reactor.

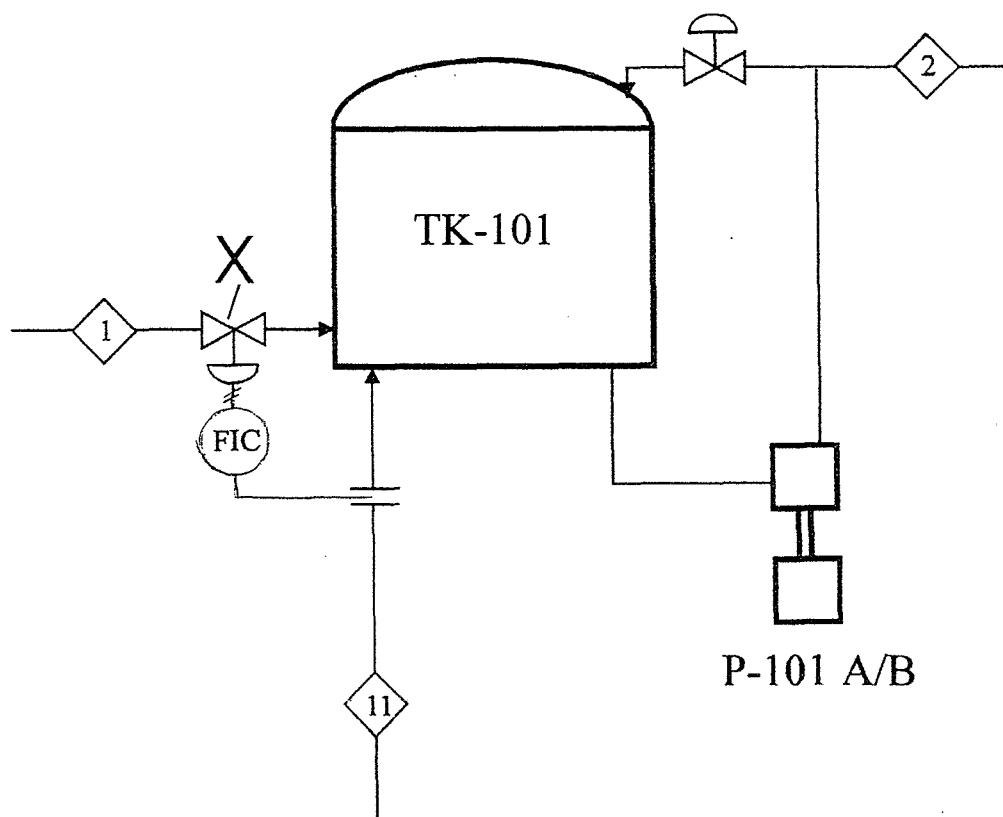
d. Explain how this system responds to fouling of the tubes on the cw side of E-901

As the heat exchange tubes in E-901 begin to foul the amount of heat transfer taking place will decrease (for a given condensate level in the exchanger). This means that the temperature and pressure in the reactor will start to rise since not enough heat is being removed from the reactor system. As the temperature in the reactor rises the TIC (Y) sends a signal to adjust the set point of the condensate level in the exchanger. The level set point is decreased. This means that more tube area is exposed to condensing vapor and more condensation can occur. As more condensation occurs the temperature and pressure in the reactor decreases.

21.10

A feedforward control loop is implemented for this system in the diagram below. The idea is to measure the flowrate of the recycle stream, Stream 11, and adjust the control valve (X) on the toluene feed, Stream 1. Thus if the recycle stream increases a signal is sent to the control valve X to close and vice-versa.

The downside of this feedforward system is that the “model” (control algorithm), describing how the valve X should change with changes in Stream 11, is prone to be not exactly correct. The consequence of this type of error can be very serious. For example if the valve on Stream 1 did not open quite enough when the flow of Stream 11 decreased then eventually the feed tank would empty. If feedforward loops are used then care should be taken to ensure that material flows in the plant will be maintained. It is usually safer to incorporate (cascade) the feedforward control strategy with some form of feedback loop (as was done in several of the preceding problems).



## Chapter 22

22.1

a)  $\frac{L}{MG} = 1.48$

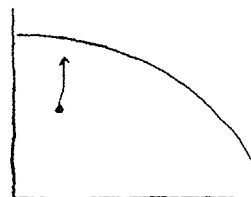
in theory - if 6.12  $L+G \uparrow 20\%$

$\frac{L}{MG}$  is constant  $\rightarrow$  OK

however: beware flooding  $f(G^2)$

y-axis  $\uparrow$  by 1.44 here

x-axis constant



if designed for 80% of flooding  
would be  $\approx 96\%$  of flooding

$$\frac{L}{G} \sqrt{\frac{P^*}{P}}$$

b) need to reduce  $m$  by 20% to compensate  
for  $G \uparrow$  by 20% (flooding may still be a problem)

$$m = 0.8 / 0.337 = 0.270 = \frac{P^*}{P}$$

$$P^* = 0.2723 \text{ if } P = 1.01 \text{ atm} \Rightarrow T = 21.4^\circ\text{C}$$

$$P = \frac{0.341 \leftarrow P^* \text{ at } 27^\circ\text{C}}{0.270} = 1.26 \text{ atm}$$

either raise  $P$  to 1.26 atm or lower  $T$  to  $21.4^\circ\text{C}$   
(or intermediate combination of both)

22.2

a) new  $\frac{L}{MG} = 1.2$   
 b)

$L \downarrow$  16176 mol/h

$G \uparrow$  49458 mol/h

$m \uparrow \begin{cases} P \downarrow & 0.8 \text{ atm} \\ T \uparrow & 32^\circ\text{C} \end{cases}$

$y_{in} \uparrow$  0.04

$x_{in} \neq 0$  0.0031235

incorrect measurement

other impurities in feed

system leaks in absorber

c) assume cannot change disturbance or feed of G

if  $L \downarrow$  or  $G \uparrow$  is cause  $\rightarrow T = 22^\circ\text{C}$  will compensate

if  $y_{in} = 0.04$   $\rightarrow P = 1.23 \text{ atm}$   
 ( $y_{out}/y_{in} = 0.05$ )  
 $A = 1.2$

if  $T \uparrow$  increase  $P = 1.24 \text{ atm}$

if  $P \downarrow$  decrease  $T = 17^\circ\text{C}$

if  $x_{in} \neq 0$  cannot achieve desired separation  
 (desired Y-axis on Colburn plot is  $< 0$ )

22.3

a)  $\frac{L}{MG} = 1.35$  (new) increase L by 10%  
decrease T = 24.6°C  
increase P = 1.11 atm  
↗  
new m = 0.307

b)  $\frac{L}{MG} = 1.58$  L ↑ by 6.75%  
 $\frac{y_{out}}{y_{in}} = 0.04$  a need m = 0.316  
T = 25.2°C ~ P = 1.08 atm

c)  $\frac{y_{out}}{y_{in}} = 0.0375$  L ↑ by 8.1%  
 $\frac{L}{MG} = 1.6$  a need m = 0.3125  
T = 25°C ~ P = 1.09 atm

22.4

bad catalyst

bad extractor performance

new extraction solvent contaminating product

bad inhibitor or bad inhibitor performance

instrumentation malfunction - T too high at bottom of column and not observed



22.5

suppose CW T increased

$\Rightarrow$  T at top of column increased

$\Rightarrow$  P at top of column increased

$\Rightarrow$  P at bottom of column increased

$\Rightarrow$  T at bottom of column increased  
could be above  $90^{\circ}\text{C}$

remedy: need rw available for these situations  
(either to use or to cool CW (latter requires HX))

or

add less heat in reboiler

(lower steam T by throttling  
or spray in water to re-saturate)

22.6

startup transient from upstream

feed T  $\uparrow$

bad relief valve or corroded open  
(not a transient)

transient P  $\uparrow$  in steam header

steam leak into system

need additional information  
to decide on these

22.7

steam condenses due to outside  $T \downarrow$   
closed vessel  $\Rightarrow P \downarrow$  if steam condenses

Wisconsin nights cooler than Southern California  
 $\Rightarrow$  more condenses,  $P \downarrow$  more

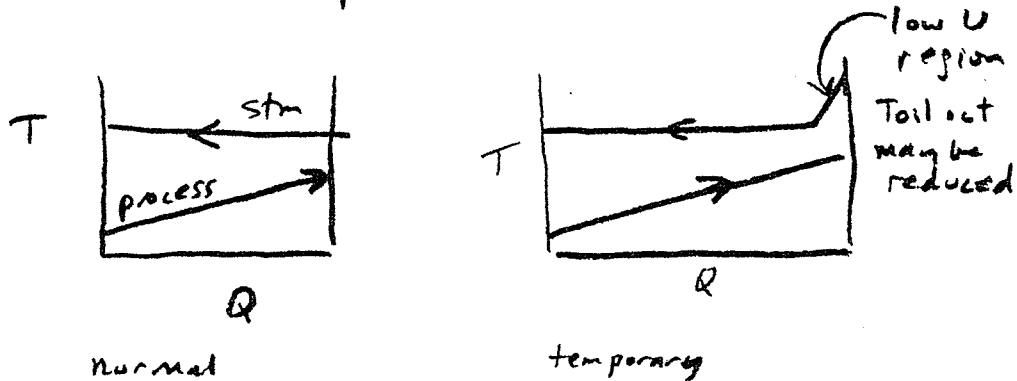
rupture caused by implosion - vacuum in vessel

Soln: design for vacuum  
add support rings to existing vessel

22.8

safety issue if regulator fails

steam is superheated if add regulator



$\mu = \text{strong } f(T)$  for viscous oil

Since  $T \downarrow$  as  $\mu \uparrow \rightarrow$  frictional  $\Delta P \uparrow$

$P$  at pump inlet  $\downarrow \rightarrow$  could cause cavitation

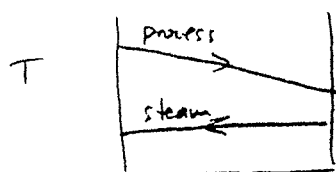
use desuperheater - spray in water

22.9

temporary problem

$U$  higher than design since little initial fouling

$$Q = UA \Delta T_{lm}$$



design



operation

$$U \uparrow \text{ so } Q \uparrow \Rightarrow T_{out} \downarrow \Delta T_{lm} \downarrow$$

$$\text{Since } Q \uparrow \dot{m}_{\text{steam}} \uparrow \Rightarrow \Delta P_{\text{steam line}} \uparrow$$

↑ Since steam header is at fixed P  
get back-pressure  
 $P_{\text{steam drum}} \uparrow$  and valve opens

$$\text{as } P_{\text{steam}} \uparrow T_{\text{steam}} \uparrow \rightarrow \text{come to operating conditions at } T_{\text{process}} \downarrow, T_{\text{steam}} \uparrow Q \uparrow$$

if not temporary

Solution

1.  $A_{HX}$  oversized,  $\dot{m}_{\text{steam}} \uparrow$

increase steam line diameter

2. undersized line from V-101 to steam header  
 $\Delta P_{\text{friction}} \uparrow$

increase steam line diameter

3. P set too low on relief valve

increase P setting  
(must be careful -  
must be done properly)

4. relief valve stuck + leaking steam

replace relief valve

22.10

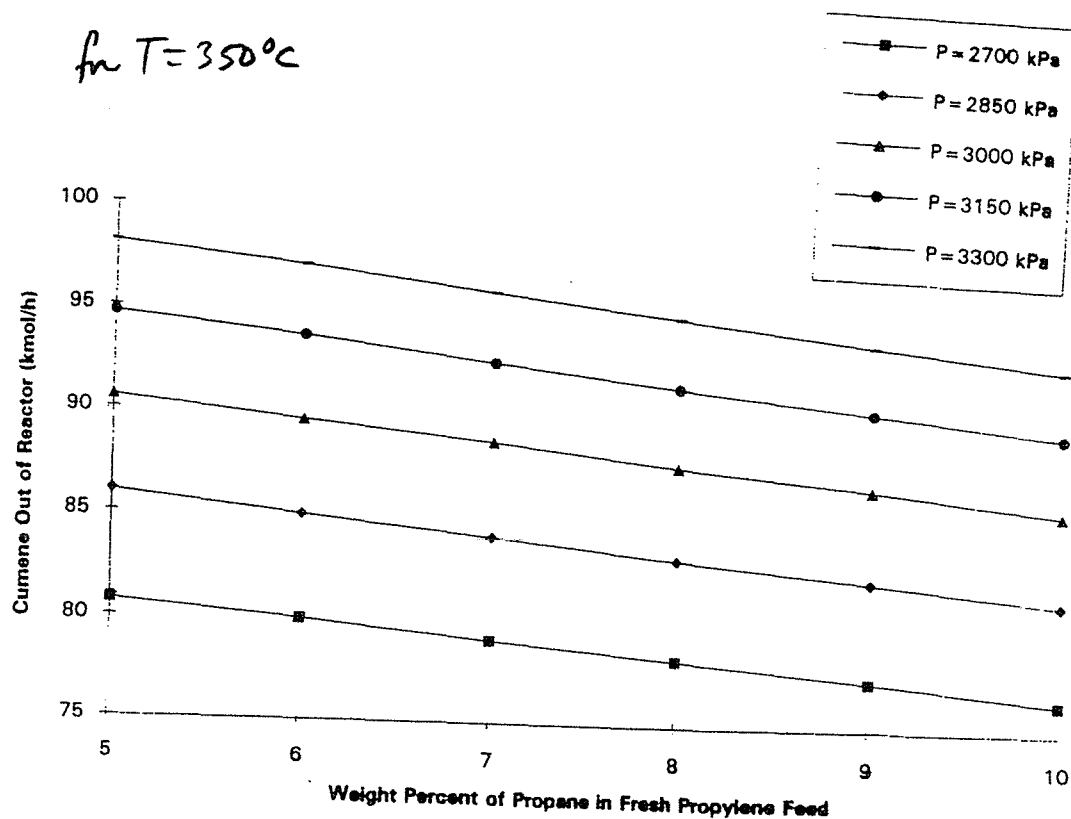
This problem is not to be confused with 19.3 and 19.4.

Here, the entire Flowsheet with kinetics (identical) for both reactions must be simulated. It is assumed that both reactions have the same kinetics (this is how Project 5, Appendix C is obtained), which are those given in Chapter 13 for this problem.

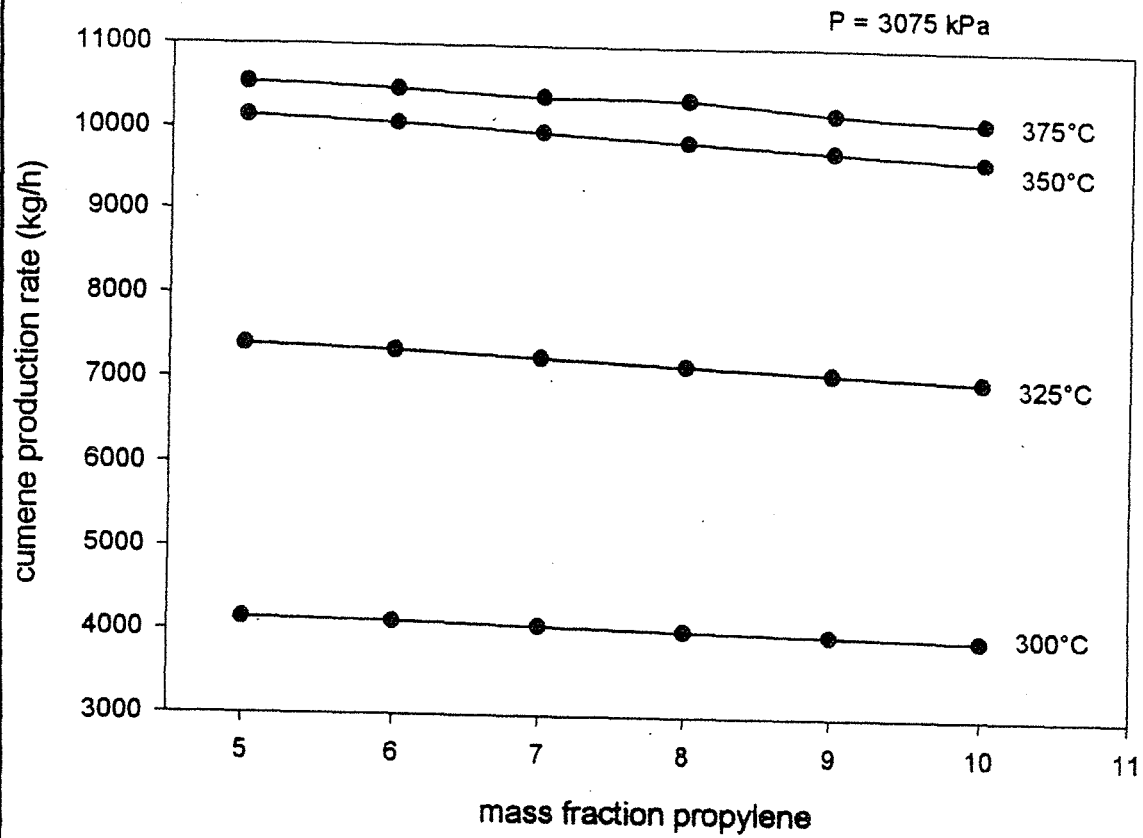
The propane impurity is that of the feed. The reactor feed is more dilute due to the recycle stream.

The results given are in the range of the requested solution, but the parameters differ slightly.  
(Note: SPRK is used in chemical process simulation)

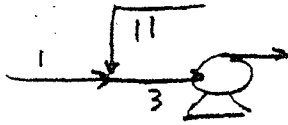
for  $T = 350^\circ\text{C}$



22.10 (cont'd)



22.11



current conditions

|                    | 1      | 11     | 3      |
|--------------------|--------|--------|--------|
| T (°C)             | 25     | 57     | 41     |
| $\dot{m}$ (kmol/h) | 105.00 | 106.89 | 211.89 |

$$5.5 \times 10^{-7} \frac{\text{m}^3}{\text{s}} = 55 \frac{\text{L}}{\text{s}} \quad \text{NPSH}_R = 6.1 \text{ ft}$$

$$\text{NPSH}_A = h_{\text{supply}} + h_{\text{static}} - h_{\text{fr}} - h_{\text{vp}} \quad 1 \text{ kPa} = 0.402 \text{ ft benzene}$$

let's say stream 11 T ↑ to make stream 3 T = 50°C

$$p^* = 94 \text{ kPa} = 37.8 \text{ ft}$$

$$\text{NPSH}_A = 11.2 - 8.1 = 2.9 \text{ ft}$$

will cavitate

Note: refer to calculations given with Project 5 in Appendix C

what is cut-off T for cavitation

$$11.2 - x = 6.1 \text{ ft}$$

$$x = 5.1 \text{ ft} = 12.69 \text{ kPa}$$

$$\frac{29.7}{0.407} + 12.7 = 86.6 \text{ kPa}$$

$$\ln p^* = 12.8888 - \frac{2695.05}{T}$$

$$T = 319.86 \text{ K} = 47^\circ\text{C}$$

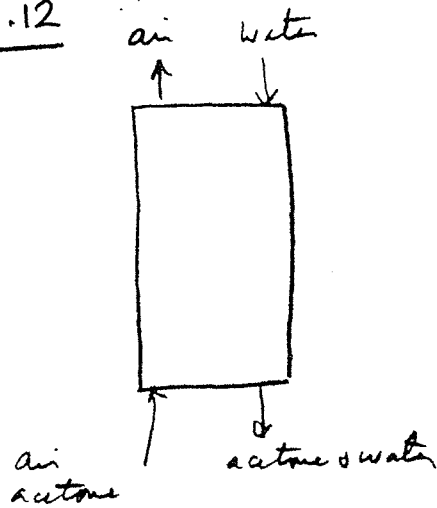
so if do mixing pt mass balance

$$105(25) + x(57) = 212(47)$$

$$x = 129 \text{ mol/h} \leftarrow \text{if get this flow of recirc at same T} \Rightarrow \text{set pump cavitation}$$

if T of stream 11 ↑ → can set cavitation at lower flow

22.12



typical design - 75% of flooding  
→ maximum possible scale-up  
is 33% on flows

the flooding limit will  
likely be a bottleneck.

if need to debottleneck -

try increasing  $P$  - pump up  $H_{2O}$  & feed

decreasing  $P$  - assume isothermal

22.13

a) only valve for Feed section is in Stream 7  
 $\Rightarrow$  cannot control air & naphthalene feeds independently

Will reactor remain fluidized?

Will surtch condenser performance be affected?

How will vacuum stills be affected?

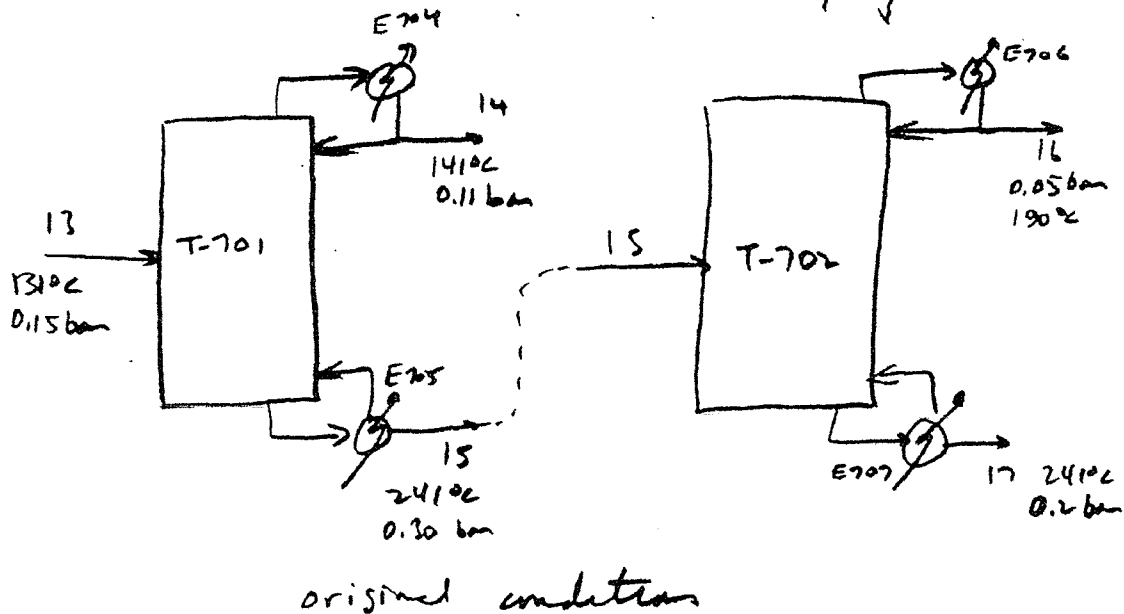
b) see section 22.4

|           |                    |           |                   |
|-----------|--------------------|-----------|-------------------|
| air old   | 151.47 Mg/h        | air new   | 43.8 Mg/h         |
| naph2 old | 12.82 Mg/h         | naph2 new | 6.41 Mg/h         |
|           | <u>164.29 Mg/h</u> |           | <u>50.21 Mg/h</u> |

ratio = 0.3  $\leftarrow$  feed to rx is only 30% of original

if  $\frac{U}{U_{mf}} < 3.33$  can have fluidization problem

at 50% scale-down in column - need to worry about weeping

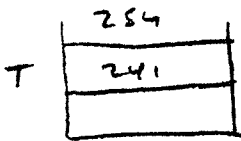




22.13 (cont'd)

T-701

E-705



$\mu_{process} \downarrow 50\%$

$\Rightarrow \mu_{steam} \downarrow 50\%$

$U, A \approx \text{constant}$

$\mu \downarrow 50\% \Rightarrow \Delta T \downarrow 50\%$

$$Q = \dot{m}_s \lambda$$

if keep  $T = 254^\circ\text{C}$  for steam

$$Q = \dot{m}_p \lambda$$

$$T_{1s} = 247.5^\circ\text{C}$$

$$Q = U A \Delta T$$

$$\Rightarrow P_{1s} = 37 \text{ kPa}$$

$$\text{at top } P = 37 - 19 = 18 \text{ kPa}$$

$\Delta P \approx \text{constant}$   
based on weir height

$$\text{top } Pr_{ratio} = \frac{18}{11} = 1.64$$

$$\text{bottom } Pr_{ratio} = \frac{37}{30} = 1.23$$

significant increase in  
absolute  $P$

$\Rightarrow \rho(\text{density})$  increase

significantly -

could cause weeping

E-704



at 18 kPa  $T = 154^\circ\text{C}$

at 11 kPa  $T = 141^\circ\text{C}$

$$\Delta T_{in,old} = 103.31^\circ\text{C}$$

$$Q = \dot{m}_p \lambda$$

$$Q = \dot{m}_w c_p \Delta T$$

$$Q = U A \Delta T_{lm}$$

1 = old

2 = new

$$\frac{Q_2}{Q_1} = 0.5 = M$$

$\mu_{out}/\mu_{in}$

$$\frac{T-30}{15}$$

$$0.5 = M^{0.8} \frac{(154-30) - (154-T)}{\ln \frac{124}{154-T}}$$

$$103.31$$

$$T = 50.1^\circ\text{C}$$

as outlet  $T$   
increased

$$M = 0.373$$

as flow reduction  
more than 50%

$$103.31 \ln \frac{124}{154-T}$$

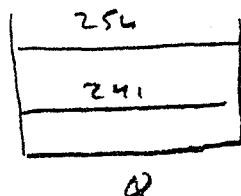
22.13 cont'd

can also keep internal flows constant by increasing reflux - would avoid weeping

T-702

E-707

T



as before

$$T = 247.5^\circ\text{C}$$

$$\text{now } P = 24 \text{ kPa}$$

$$\text{at top } \frac{a}{5} = 1.8$$

$$\text{at bottom } \frac{24}{20} = 1.2$$

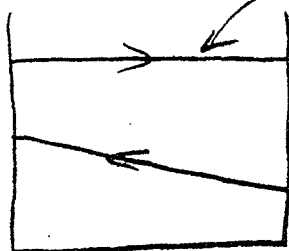
$$\text{top } P = 24 - 15 = 9 \text{ kPa}$$

$$\text{top } T \approx 205^\circ\text{C}$$

absolute P increases increases density,  $\Rightarrow$  worry about weeping

E-706

T



at 9 kPa  $205^\circ\text{C}$

at 5 kPa  $190^\circ\text{C}$

$$\Delta T_{\text{pin, old}} = 152.38^\circ\text{C}$$

Q

$$0.5 = M \frac{T-30}{15}$$

$$M = \frac{x_{1, \text{old}}}{x_{1, \text{new}}} \quad \begin{matrix} 1 = \text{old} \\ 2 = \text{new} \end{matrix}$$

$$0.5 = \frac{M^{0.8}}{152.38} \ln \frac{T-30}{205-T}$$

$$T = 49.7^\circ\text{C}$$

$$M = 0.381$$

22.14

- Increased flow increases  $NPSH_R$
- H-801 load  $\uparrow$
- propylene more dilute in reactor - need to increase  
+ benzene T, P to maintain  
conversion (See Part 10)
- need more air in E-802
- should be no bottlenecks after flash

for rxr - see Problem 10

$NPSH$  - Stream 2 original  $105(42) = 4620 \text{ kg/h propylene}$   
 $517(44) = 231.88 \text{ kg/h propane}$   
 $\sim 4.8 \text{ wt\% propane}$

$$\text{if } 10 \text{ wt\% propane } \frac{x}{4620+x} = 0.1 \quad x = 513.33 \frac{\text{kg}}{\text{h}}$$

original total feed =  $4852 \text{ kg/h}$   $\Rightarrow$  6% increase  
new total feed =  $5133 \text{ kg/h}$

if look at  $NPSH$  curve in Appendix C, Project 5  
new  $NPSH_R = 6.5 \text{ ft}$

$$\Delta P_{f, \text{new}} = 1.06^2(1) = 1.12 \text{ ft} \quad NPSH \text{ not a problem}$$
$$NPSH_A = 6.88 \text{ ft}$$

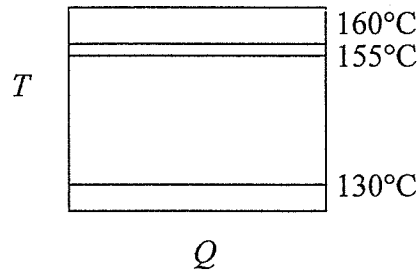
i.e. from Problem 10 - rxr is bottleneck since cannot  
handle 10% propane impurity

22.15 (a) Pump and system curves:

$$\dot{v}_{\max} \approx 73 \text{ m}^3/\text{h} \Rightarrow \approx \frac{73 \text{ m}^3/\text{h}}{50 \text{ m}^3/\text{h}} = 1.46$$

46% scale-up

Heat Exchanger:



$$Q = \dot{m}_s \lambda_s$$

$$Q = \dot{m} \lambda$$

$$Q = UA \Delta T_{lm}$$

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_{s2} \cancel{\lambda_{s2}}}{\dot{m}_{s1} \lambda_{s1}}$$

$$\frac{Q_2}{Q_1} = \frac{\dot{m}_2 \cancel{\lambda_2}}{\dot{m}_1 \lambda_1}$$

$$\frac{Q_2}{Q_1} = \frac{\cancel{\psi_2} \cancel{A_2} \Delta T_{lm2}}{\cancel{\psi_1} \cancel{A_1} \Delta T_{lm1}}$$

$$\frac{U_2}{U_1} \approx 1$$

$$\frac{\dot{m}_2}{\dot{m}_1} = \frac{\Delta T_2}{\Delta T_1} = \frac{160^\circ\text{C} - 130^\circ\text{C}}{155^\circ\text{C} - 130^\circ\text{C}} = \frac{30^\circ\text{C}}{25^\circ\text{C}} = 1.2$$

20% scale-up

Vaporizer limits scale-up since  $T_{\max}$  for steam is  $160^\circ\text{C}$ .

(b)  $NPSH_A = P_{\text{tank}} + \rho gh + \Delta P_f - P^*$

If stored as liquid in equilibrium with vapor  $P_{\text{tank}} = P^*$

$$NPSH_A = \rho gh \Rightarrow \text{should not cavitate}$$

## 22.16 Drying Oil Facility

### Trouble Shooting

#### (a) Noisy pump

We can do calculations, but these components are not volatile, so the most likely cause is some form of mechanical problem –worn bearings, etc. A remedy might be to switch to the back up pump and see if whining stops. Then, the original pump can be fixed.

#### (b) Problem with T-501 and the increased Dowtherm A flowrate.

Although a drop in ambient temperature might have an effect, it would be small. As temperature drops, heat loss increases and inlet cooling water temperature drops.

Detailed calculations would show that the issue is not with the change in ambient temperature in October. Look at heat loss:

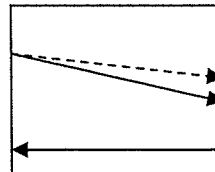
$$Q = UA\Delta T_{LM}$$

$$Q = (\dot{m}c_p\Delta T)_{Dowtherm}$$

Since Dowtherm has been increased to compensate to keep purity (specifications) unchanged, assume that  $Q$  has not changed.

$$Q(c) = \dot{m}(\uparrow)c_p(c)\Delta T(\downarrow)$$

$$Q(c) = U(\text{must } \downarrow)A(c)\Delta T_{LM}(\uparrow)$$



$\Delta T$  decreases  
 $\Delta T_{LM}$  increases

But  $U$  will increase because the mass flow of Dowtherm has increased therefore something in  $U$  must have decreased to compensate!

$$\frac{1}{U_i} = \frac{1}{h_i} + \frac{1}{h_{fi}} + \frac{d_{LM}}{k_{wall}} + \frac{d_i}{d_o h_{fo}} + \frac{d_i}{d_o h_o} \approx \frac{1}{h_i} + \frac{1}{h_o}$$

Conclusion – there must be fouling!!

Could be on Dowtherm side, but since Dowtherm is a heat transfer fluid, it is unlikely that it fouls (unless temperature is too hot – max temp is 400°C at higher temps Dowtherm carbonizes!).

The problem is more likely on process side, since the temperature is 342°C, and reaction will occur at this temperature. The hottest point in reboiler is on tube

walls where reaction will occur. Note that there is not much DO here, so gum formation will be slow, but only a small amount is needed for film heat transfer to decrease.

Alternatively, the filter may not be working properly, so the gum content in Stream 7 should be checked.

### ***How to solve the problem?***

- a) If filter is a problem – recommend regular samples of Stream 7 to identify when “gum breakthrough” occurs.
- b) Recommend to clean reboiler – but requires shut down and the fouling is on the shell side.
- c) To resolve the problem permanently, we must reduce the bottom temperature by reducing column pressure.
  - As pressure decreases, the column moves towards flooding, because vapor density drops and vapor velocity increases.
  - But as the pressure drops, separation becomes easier since equilibrium line moves away from  $xy$ -line. Thus fewer stages are needed for a given separation or a lower reflux is needed with a fixed number of stages. With lower reflux, we move away from flooding and therefore compensate for lower pressure – at least qualitatively. The only way to be sure is to do the calculations.

### **(c) ACO Feed problem**

There are two issues:

First, the control scheme is wrong. The control valve should be using the level control of V-501 as its signal. As it is now, the flow into V-501 will fluctuate based on the level in the storage tank, probably being overridden by operators to maintain flow.

Second, another probable cause is that ambient temperature has dropped causing the ACO feed's viscosity to increase.

Thus, feed problems may be a combination of colder weather and operator error.

If this switch in control scheme is made, we will run into a bottleneck when the viscosity of the oil is so high, that when the CV is wide open, the design flow rate cannot be maintained.

A solution is to heat feed tank and also heat trace the supply line. The feed tank may have a steam heater that no one has turned on or maybe it is malfunctioning. The line is already insulated, but for extended periods of time this will not help.

The supply line could be retro-fitted with larger diameter pipe.

(d) Steam venting from E-506

Most probable cause is the addition of the extra pipe and elbows connecting the exchanger to the low-pressure steam header. This extra pipe causes additional frictional loss and requires a higher pressure in the shell of E-506 in order to deliver the steam to the header.

If all the steam were to flow to the header then the friction pressure loss in the line would increase in proportion to the added equivalent length:

A solution is to reset PSV-519 to 60 kPa above header. This can be done because the lps header is at 607 kPa, and this would allow the shell to operate at 667 kPa, which is less than 90% of the rated pressure of 750 kPa for E-506. It is also possible to re-pipe the connection with 3" pipe, but this is more expensive.

## 22.17 Scale Down

The equipment that must be considered is as follows:

Exchangers – what are new outlet temperatures

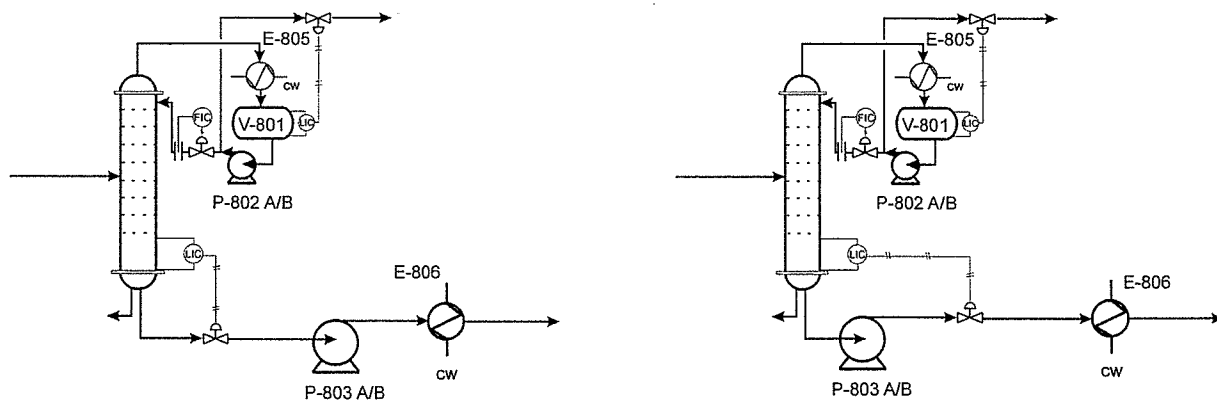
Pressurizing of system – reactor inlet pressure drops

Conversion in reactor – how is it affected by longer space times

Tower flooding/weeping

- (a) By replacing the 1.5 in pipe with 1 in pipe, the pressure drop probably induces cavitation in the pump. Calculations would support this. Also, having the valve before the pump lowers the pressure, further inducing cavitation (left figure). The figure on the right is a better design.

So, the obvious solution is to retrofit to 1.5 in pipe. Since it worked that way before, the valve issue is less important. However, in general, it is poor design to put the valve before the pump.



- (b) Something is happening right after plant start up. Two possibilities come to mind. If the heat exchanger were designed to include a fouling factor, the tubes would probably not be fouled upon start up. Another possibility is that the catalyst deactivates with time, but is at full activity at start up. In either case, there would be better heat transfer, more boiler feed water consumption, and more steam produced. The downstream steam header is at a constant pressure, so the increase in steam produced results in an increased pressure upstream, possibly opening the pressure-relief valve. Calculations show that a fraction of a percent is sufficient to open the pressure relief valve. This problem should not be ignored, since lost steam is lost money; although, calculations show that the lost steam value is only hundreds or thousands per year. Perhaps the pressure-relief valve should be set to a higher pressure. This requires internal permits for obvious safety reasons.
- (c) During the summer, warmer cooling water affects several units, specifically the absorber, the distillation column, and the product cooler. The increased fuel gas rate supports this because if the temperature in the absorber increases, absorption equilibrium favors the vapor phase.

Similarly, in the distillation column, the log-mean temperature difference must stay the same, since all flows stay the same, so increasing the cooling water temperature requires that the condensation temperature in the column condenser increase. This increases the column pressure, thereby increasing the temperature of the reboiled vapor. Therefore, the temperature into E-806 increases. So, the outlet temperature

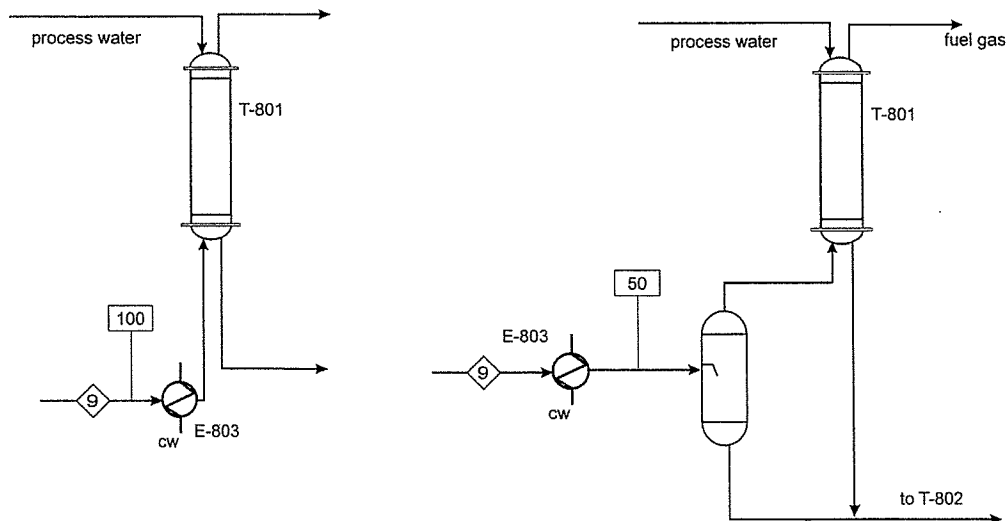


would increase for two reasons. First of all, the inlet product is at a higher temperature. Secondly, the cooling water is at a higher temperature.

Finally, in the calculations, the log-mean temperature correction factor ( $F$ ) has been omitted. This turns out to be a lower number than desired, so E-806 is significantly undersized to begin with. Clearly, this problem is exacerbated in the summer months.

- (d) Additionally, the following process improvement might be a good idea. The value of methanol in Stream 12 is about \$2 million/yr. Methanol recovery should be a high priority. Even under normal operating conditions, the absorber temperature may be higher than necessary. The absorber pressure could be raised by pressurizing the entire process (rather than adding a compressor right before the absorber). Additional cooling and/or separation equipment might be worthwhile.

Two options for better methanol recovery might be:



22.19

- (a) Fuel gas has increased in order to maintain Stream 2 at 900°F. This is probably due to a lower HT coefficient on tube side in furnace. During temporary upset, oil would sit inside tube and radiant furnace walls may have caused coking on the inside surface of tubes. This means tube walls are hotter now and so this may be a problem in the long term.

Solution: Stop oil flow temporarily and inject steam through tubes to “burn off” coke.

- (b) Problem same as (a) except coke furnace is greater causing a significant  $\Delta P$  increase across furnace tubes and resulting in reduction in oil flow.

Solution: May need to isolate furnace and burn coke off slowly with steam (same as part (a) but will take longer.)

- (c) Same problem as (a) and (b) except a tube may have ruptured. This would cause a lower product flow and now the oil is supplying fuel to the furnace causing the O<sub>2</sub> level in the stack to drop so air/fuel ratio must be increased.

Solution: Must shut down furnace and replace ruptured tube(s).

## Chapter 23

Problems 1-6 of this chapter are ethics questions for which students should be able to find several competing rights and responsibilities. Some common issues are given for the individual problems below. There are several effective ways to use these problems:

- Assign them as regular homework assignments, for which the student prepares a written response to the questions asked. The typical length of a student response is one single-spaced printed page, which should include a description of the ethical issues involved, an analysis of which issues are most important, and a conclusion as to the best solution.
- Assign students to read and formulate answers to several questions for homework. Then, in class, students are chosen at random to present their solutions to the class and to lead a class discussion of the problem.
- Assign them as small-group homework assignments. Then, in class, the group explains the issues and their solution and leads the class discussion.

For additional case studies and decisions regarding ethical dilemmas see the web page for the Board of Ethical Review (BER) of the National Society of Professional Engineers (NSPE) at:

<http://www.cwru.edu/affil/wwwethics/engcases.html>

23.1 Issues that students raise include (but certainly are not limited to) the right of the company to trade secrets, the responsibility of the employee to employer, the desire to help one's colleagues professional development, and the improbability that any discussion in class will put the company at a competitive disadvantage. The students should be able to analyze these competing issues and interests and to form a rational basis for a response. The problem provides a vehicle to expose students to confidentiality agreements and to codes of ethics. The AIChE Code of Ethics and (especially) the NSPE Code of Ethics deal with this issue. Students should be able to find the correct sections and use these in their analysis. Classroom discussion should include what ethical and legal limitations employers can put on employees divulging information. Finally, students should realize that a call to their former employer is likely to clear up the ambiguity. If students choose to not divulge the information, even if it would be allowed, this is an ethical issue that should also be discussed.

23.2 This is a common problem for students, and they often will have very strongly held competing views. This problem is a vehicle to getting students involved in the issues and to develop a sense of moral autonomy. It is helpful in the discussion to have different students look at the problem from the different viewpoints. Would they view the issues differently if they were the interviewer rather than the interviewee, for example?

|      |                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                     |
|------|---------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|
| 23.3 | The issues here at first seem many moral rather than ethical, but students often realize that this kind of an issue can arise throughout one's career—as they move from job to job within an organization or when they learn more about the organization they work for. After students have discussed the anti-personnel mines (often with different concerns during wartime and during peace), they should be encouraged to look at potentially less obvious yet similar issues, such as: the company's investment policy, other more benign products that one believes "should not be manufactured." To simulate discussion and to broaden the problem, one could ask if the student should quit after working for the company for many years, when the employee realizes that the company produces such devices. |
| 23.4 | This problem involves many competing rights and responsibilities: to one's employer, employees, government, society, etc. The discussion often finally centers on the responsibility of the government to set reasonable standards and the responsibility of the company to meet those standards. However, the facts seem to suggest that employee health may not be adequately protected by the apparent standard. Students then often realize that the problem is more complex. This is a good time to mention the "general duty clause" of the OSHA Act, as well as the fact that a no-cost or low-cost solution may be available.                                                                                                                                                                               |
| 23.5 | This problem is a good vehicle for "reflection in action," since a the problem is given in detail and the responses of some chemical engineers are available. Students should be able to find numerous rights and responsibilities from the code of ethics that affect this scenario.                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                                               |
| 23.6 | The "falsified data strike back" scenario can be used in a classroom immediately after the "falsified data" case (Section 20.1.2) is discussed. The point here is to see what would have happened if the data were originally falsified. Once those consequences are discussed, it is interesting to discuss the outcome if the original data weren't falsified. The discussion then leads to exactly how should one have written the first report, and what should be done now, if the data had been falsified. When students suggest that Jay should contact his former boss, they should be asked to play through that scene to see whether this strategy is more likely to clarify the issue or to confuse it.                                                                                                  |
| 23.7 | It is best if the instructor first contacts the Board to get this information. Most boards are very helpful in this regard. Students are likely to immediately question whether all faculty in the department are registered and why not. They will likely ask for direct advice as to whether they should start the registration process and why.                                                                                                                                                                                                                                                                                                                                                                                                                                                                  |

|       |                                                                                                                                                                                                                                                                                                                                                                                                                                |
|-------|--------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|
| 23.8  | Codes of conduct can be obtained from companies, but often only if the company name is removed. The instructor should obtain a couple from companies that have recently hired graduates from the school. These codes vary very widely. Some will include codes of ethics, others will not. Some describe anti-discrimination and general personnel policies, some do not. Students will be somewhat uneasy with the omissions. |
| 23.9  | There are many such heuristics. The chapter does not focus heavily on heuristics related to family responsibilities. Students often are eager to develop guidelines in this realm. Whistleblowing is another area of interest in which students often generate several heuristics.                                                                                                                                             |
| 23.10 | If students need suggestions for an ethical dilemma, suggest a classical scenario in which a student observes another student cheating on an exam or on a project.                                                                                                                                                                                                                                                             |
| 23.11 | The Process Safety Management regulation is a bridge between ethics, professional responsibility, and safety. At the end of the regulation is a good synopsis.                                                                                                                                                                                                                                                                 |
|       |                                                                                                                                                                                                                                                                                                                                                                                                                                |

## Chapter 24

24.1 Total employee hours worked per year:

$$40 \text{ h/wk}(47 \text{ wk/yr})(30,000 \text{ employees}) = 56.4 \times 10^6$$

- a. Fatalities per year =  $(56.4 \times 10^6 \text{ h})(4 \text{ deaths}/10^8 \text{ h}) = \underline{2}$
- b. Injuries per year =  $(56.4 \times 10^6 \text{ h})(0.49 \text{ incidents}/2 \times 10^5 \text{ h}) - 2 \text{ deaths} = \underline{136}$

- a. Fatalities per year = 2
- b. Injuries per year = 136

24.2 Students can be taken on a tour of the steam plant or expected to find it and ask those in charge appropriate questions.

- a. Explosion of the main steam drum from overpressure and explosion of the furnace after a buildup of unburnt fuel are probably the most obvious, but students can find numerous variants of these or accident scenarios that arise from problems with raw materials (fuel, water) or product distribution (steam, power).
- b. Students should be able to identify many such safeguards, including: relief devices, various control-system loops and interlocks, operator training, flame detection, etc.

24.3 All of these MSDS's are readily available from suppliers or the web sites given in Section 21.2.1.

24.4 Students should be able to find many such regulations, especially through the Federal Register and Code of Federal Regulations, which can be searched by keyword (see Table 21.2). These range from EPA regulations for benzene in motor gasoline to the OSHA PEL. In some states, additional regulations can be found.

24.5 Although all 13 section of the PSM standard are relevant to a unit operations laboratory, the most fruitful items for the students' analyses are: process safety information, operating procedures, training, and compliance safety audit.

24.6

- a. The volatile components in paints (both oil- and water-based) tend to evaporate slowly. In the closed-cup flash point determination, one looks at the equilibrium vapor concentration. In the open-cup determination, one looks at the steady-state vapor concentration when the paint is exposed to air. If the vapor pressures of the components is high enough at room temperature, the equilibrium vapor concentration could be above the LFL. However, if the rate of evaporation is slow compared to the rate of diffusion of the component from the area immediately above the liquid to the bulk atmosphere, the steady-state vapor concentration could be below the LFL.
- b. The open-cup flash point is the more useful property, since it is more representative of the potential vapor concentrations generated as paints are used.

24.7 This mixture has an octane number of 87 by definition. Using Raoult's Law, one can calculate the equilibrium concentration of the iso-octane and n-heptane.

- a.  $P_{\text{sat}}^* (298 \text{ K, iso-octane}) = 6530 \text{ Pa}$        $\text{LFL} = 0.008$        $\text{UFL} = 0.070$   
 $P_{\text{sat}}^* (298 \text{ K, n-heptane}) = 6050 \text{ Pa}$        $\text{LFL} = 0.010$        $\text{UFL} = 0.084$   
 $P = 101,325 \text{ Pa}$   
 $x(\text{iso-octane}) = 0.117$        $x(\text{n-heptane}) = 0.883$  (from molar volumes)  
 $y_i = (P_{\text{sat},i}^* x_i) / P$   
 $\therefore y(\text{iso-octane}) = 0.00754$        $y(\text{n-heptane}) = 0.0527$

Thus, at the equilibrium, the vapor would be within the flammability limits and would support combustion.

- b. The air-gasoline mixture will be within its flammable limits near the surface of the liquid. Because the concentration is so much above the LFL, the flammable vapor cloud will extend a significant distance above the gasoline pool, creating a serious flammability hazard. Note that, if the ambient temperature is just slightly above 25°C, the vapor immediately above the gasoline pool will be above the UFL, and the vapor will not support combustion. However, as the gasoline vapors mix with the air, the vapor will be within the flammability limits somewhere in the atmosphere, and a flame could propagate towards the liquid. These properties make the vapor space in a heated gasoline tank non-flammable, while making it very unsafe to throw a lit match towards a pool of vaporizing gasoline.

24.8 Many of the options for each of the HAZOP keywords lead to hazardous conditions for this reactor, when applied to flowrates and temperatures.

24.9 This problem is similar to Problem 24.7

PEL(benzene) = 1 ppm TWA; 5 ppm STEL

$P_{\text{sat}}$  (benzene, 298 K) = 12,700 Pa

$P = 1.01325$  bar

$x(\text{benzene}) \approx 0.0025$  (assuming all components have approximately the same molecular weight)

$y(\text{benzene}) = (P_{\text{sat}} * x) / P = 313$  ppm

The benzene concentration is well above the PEL at 298 K. As the temperature rises, the concentration increases because of the increase in the vapor pressure of benzene.



## Chapter 25

### General Comments:

In the first printing of the third edition, “the ethylbenzene process” appears twice. It will be removed in subsequent printings.

The suggestions that follow are just a few ideas. The overall themes are:

- a. Heat integration saves energy, and is green.
- b. Reducing the reactant flowrate is usually green, because most of these reactants come from oil or coal. In a life-cycle analysis, the pollution associated with refining oil or processing coal is assigned to the raw material. Therefore, anything that increases overall conversion is green.
- c. If a solvent is introduced, it should be recycled.
- d. Anything that increases selectivity is green. This follows from #2.
- e. Most “greening” suggestions are also good ideas from an economic standpoint. However, the maximum selectivity may not correspond exactly to the economic optimum.

### 25.1 ethylbenzene

reduce benzene loss – recover from fuel-gas stream – benzene is made from toluene, which is isolated from a refinery cut

recover more ethylbenzene from fuel gas

heat integration – preheat feed to fired heater with hot reactor effluent – this is probably greener than producing steam

adjust reactor conditions to increase selectivity for ethylbenzene

|      |                                                                                                                                                                                                                                                                                                                                                                         |
|------|-------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------------|
| 25.2 | styrene                                                                                                                                                                                                                                                                                                                                                                 |
|      | <p>recycle waste water, purify it (necessary because it might contain some organic residues), and use it to make steam for the process feed</p> <p>heat integration where possible</p> <p>maximize selectivity</p> <p>recycle the waste benzene and toluene to the companion ethylbenzene plant – virtually all styrene plants are coupled with ethylbenzene plants</p> |
| 25.3 | maleic anhydride                                                                                                                                                                                                                                                                                                                                                        |
|      | <p>heat integration – hot streams: molten salt, reactor effluent, tower condenser; cold streams – benzene feed or air/benzene mixed feed stream, reboiler</p> <p>maximize selectivity</p>                                                                                                                                                                               |
| 25.4 | ethylene oxide                                                                                                                                                                                                                                                                                                                                                          |
|      | <p>recycle waste water to absorber feeds</p> <p>maximize selectivity by keeping with low conversion – this is a classic <math>A \rightarrow B \rightarrow C</math> reaction where B is the desired product (<math>A \rightarrow C</math> is also occurring), so low conversion is also economical</p>                                                                   |
| 25.5 | formalin                                                                                                                                                                                                                                                                                                                                                                |
|      | <p>heat integration – use Stream 9 to preheat fired-heater feed to reduce energy use and emissions from products of combustion</p> <p>recovery of methanol from off-gas stream – there is significant loss – millions of dollars per year</p>                                                                                                                           |
| 25.6 | DME                                                                                                                                                                                                                                                                                                                                                                     |
|      | <p>selectivity</p> <p>heat integration</p>                                                                                                                                                                                                                                                                                                                              |

## 25.7 acetone

recycle waste water to absorber feed

There is no need to purify the IPA/water stream for recycle in T-403 to the level shown. There are several stages doing very little separation. The reflux ratio could be reduced, since there are too many stages. This saves energy and fossil fuels. Alternatively, fewer stages column could be used. Or, if the reflux ratio is reduced, a smaller diameter column is needed. This saves on materials, for which manufacture is energy intensive.

heat integration – between reactor effluent and reactor feed – reduces energy requirement for E-407

## Projects on CD

### Review of Project C.1 – Allyl Chloride Production

There are two separate problems to be addressed. First, we must increase the production rate of allyl chloride by as much as possible and as soon as possible. Second, we must look at the longterm profitability of the plant and determine what can be done and what additional equipment will be required. Let us look at these two problems separately.

#### 1. Increase the Production of Allyl Chloride

One key thing to remember is that we do not want to buy new equipment. Since the Alabama plant is only down for a short while any capital expenditures will be wasted. One could argue that buying a new pump (off the shelf), for example, could be justified due to its low cost. However, if production is to be increased immediately, the installation of the new equipment, no matter how temporary the installation, will take time since safety codes etc. must be followed – so it is probably better to concentrate on changes in the plant which can be done without new physical plant. Secondly, the existing equipment has limitations and that it is essential to determine what are these limitations. Finally, we must be aware that changes in one section of the process may/will impact other areas.

#### 1.1 Look at Required Pressure Profile through the process for Increased Throughput

As we process more raw materials, the pressure drops over all the equipment will increase. The feed pressures are determined by the saturation vapor pressures of the offsite storage, since both propylene and chlorine are stored as saturated liquids at 25°C. The pressure information is confirmed in Table C.1 and students should realize that both these materials would be stored as liquids. The feed pressures of the raw materials (Streams 1 and 2) are much larger than required and the flows are regulated through control valves with fairly large  $\Delta P$ s. However, as the flows increase, and since the downstream pressure of Stream 5 must not exceed 2.1 bar, the pressure of propylene into the fired heater and chlorine into the reactor must increase to overcome the increased pressure drop through the equipment. Consider the pressure profile at design conditions in the process.

#### At Current Conditions

|               | Pressure at inlet to equipment<br>(bar) | $\Delta P$ across equipment<br>(bar) |
|---------------|-----------------------------------------|--------------------------------------|
| Battery limit | 2.09                                    |                                      |
| E-602         | 2.43                                    | 0.34                                 |
| E-603         | 2.77                                    | 0.34                                 |
| R-601         | 3.04                                    | 0.27                                 |
| J-601         | 3.24                                    | 0.20                                 |
| H-601         | 3.58                                    | 0.34                                 |

Now as the flow increases so will the pressure drop, except for  $\Delta P$  across the fluidized bed reactor which will be approximately constant (as stated in the problem). We recall that for a fluid, the pressure drop across a pipe is given by

$$-\Delta P_f = 2\rho v^2 f \frac{L_{eq}}{d}$$

For small changes in absolute pressure we can say that  $\Delta P_f \approx v^2 = (\text{mass flow})^2$ , i.e., gas behaves as an incompressible fluid. Thus we can scale up using  $\Delta P \propto (\text{mass flow})^2$ .

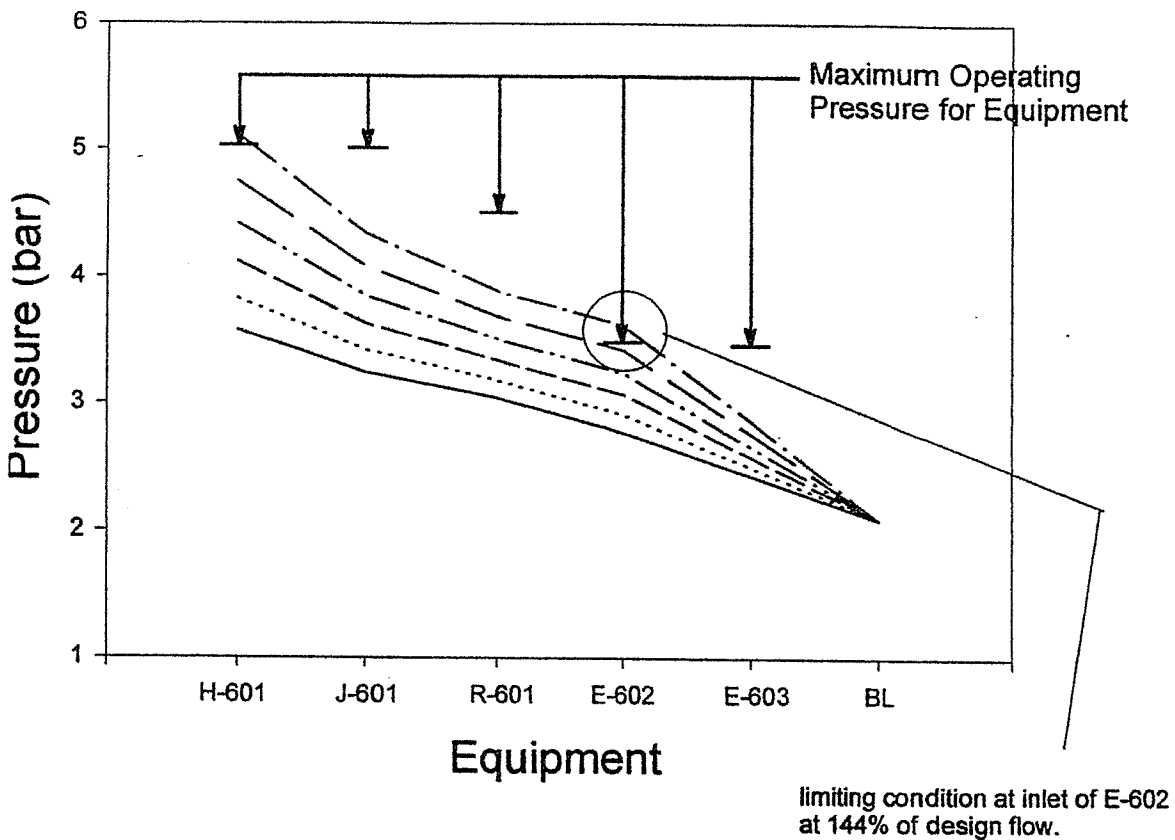
|               | @ Design<br>Pressure | $\Delta P$ | @1.2×Design<br>Pressure | $\Delta P$                     | @1.44×Design<br>Pressure | $\Delta P$ | $P_{\max}$  |
|---------------|----------------------|------------|-------------------------|--------------------------------|--------------------------|------------|-------------|
| Battery Limit | 2.09 bar             |            | 2.09*                   |                                | 2.09*                    |            |             |
|               |                      | 0.34       |                         | (0.34)(1.2) <sup>2</sup> =0.49 |                          | 0.71       |             |
| E-603         | 2.43                 | 0.34       | 2.58                    | 0.49                           | 2.80                     | 0.71       | 3.50        |
| E-602         | 2.43                 | 0.27       | 3.07                    | 0.27                           | <b>3.51</b>              | 0.27       | <b>3.50</b> |
| R-601         | 2.77                 | 0.20       | 3.34                    | 0.29                           | 3.78                     | 0.41       | 4.50        |
| J-601         | 3.04                 | 0.34       | 3.63                    | 0.49                           | 4.19                     | 0.71       | 5.00        |
| H-601         | 3.24                 |            | 4.12                    |                                | 4.90                     |            | 5.00        |

When we scale-up by 44%, using this analysis, we find that the inlet pressure to E-602 equals the maximum design pressure (see Table C.2) and we reach a bottleneck. Clearly H-601 is also very close to its maximum operating pressure. Thus maximum scale-up based on design pressure is 44% of design. The pressure profiles through the process for different scale-up ratios are shown in the figure on the next page.

**Maximum Scale-up = 1.44 or 144%**

\*note: we have assumed that the battery limit pressure is constant at 2.09 bar (as given in the problem statement) despite the increase in flow. This is possible if Stream 5 flows to a pressure sink, such as a distillation column or flash drum where the system pressure is controlled by the condensation temperature. If this were not the case and the battery limit pressure increased with flowrate then the maximum scale-up would be less than 44% based on the analysis of maximum operating pressure. Note also that both raw material streams must be vaporized prior to entering Unit 600. It is assumed that the vaporizers for the propylene and chlorine can handle the increased demand – since there is no information given on this point we need only make note of this assumption in our analysis.

### Pressure Profiles through Unit 600 - Scale-Up assuming Incompressible Flow



In the above analysis we have assumed that the density of the fluid (gas) does not vary with changes in absolute pressure – clearly this assumption is not particularly good for this problem.

We can go back and repeat the analysis assuming that  $\Delta P \propto (\text{mass flow})^2 / \rho$ . If we assume ideal gas behavior, then we can scale the frictional pressure drop using the following equation:  $\Delta P \propto (m_2/m_1)^2 (P_1/P_2)$ , where  $m$  is mass flowrate,  $P$  is absolute pressure and 1 and 2 denote design and scaled-up conditions.

The solution algorithm involves trial and error and an example for a scale-up factor of 55% is illustrated below:

|               | @ Design<br>Pressure | $\Delta P$ | @1.55×Design<br>Pressure | $\Delta P$ | @1.55×Design<br>$\Delta P$ | @1.55×Design<br>Pressure | $P_{max}$ |
|---------------|----------------------|------------|--------------------------|------------|----------------------------|--------------------------|-----------|
| Battery Limit | 2.09 bar             |            | 2.09*                    |            | 2.09*                      | 2.09*                    |           |
| E-603         | 2.43                 | 0.34       | 2.91                     | 0.82       | 0.74                       | 2.83                     | 3.50      |
| E-602         | 2.43                 | 0.34       | 3.73                     | 0.82       | 0.64                       | 3.47                     | 3.50      |
| R-601         | 2.77                 | 0.27       | 4.00                     | 0.27       | 0.27                       | 3.74                     | 4.50      |
| J-601         | 3.04                 | 0.20       | 4.48                     | 0.48       | 0.36                       | 4.10                     | 5.00      |
| H-601         | 3.24                 | 0.34       | 5.30                     | 0.82       | 0.57                       | 4.67                     | 5.00      |

$\longleftrightarrow$   
1<sup>st</sup> iteration

$\longleftrightarrow$   
2<sup>nd</sup> iteration

$\longleftrightarrow$   
3<sup>rd</sup> iteration

$\Delta P_2 = (0.34)(1.55)^2 [\frac{1}{2}(2.09+2.43) / \frac{1}{2}(2.09+2.91)] = 0.74$

Use ratio of  $[P_{ave} \text{ from design} / P_{ave} \text{ from previous iteration}]$

solution converges after 5 iterations

again the pinch pressure is at the inlet of E-602, i.e., 3.5 bar

**Maximum Scale-Up = 1.55 or 155%**

As we will see when we consider other factors, the maximum scale-up is less than 44% so that the added effort of using the trial and error solution to account for compressibility effects is not justified.

**1.2. Look At Individual Equipment**

Assume that the reactor conditions are to stay the same, namely use a temperature of 510°C and assume that small changes in pressure will not affect the selectivity.

**1.2.1 Fired Heater, H-601**

Design Duty = 4000 MJ/h  
 Maximum Duty = 5400 MJ/h

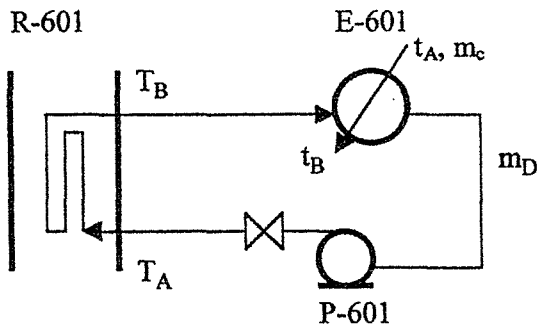
Scale-up =  $5400/4000 = 1.35$  or 135% of design. The reason for the under utilization of the fired heater at design conditions is not clear from the problem statement. One possible reason is that the unit was oversized in order for start-up of this unit or another unit. Current operation probably takes place with one burner or a bank of burners not operating. It should be an easy matter to bring the new burners on line and achieve this level of scale-up. We assume that the unit can be scaled linearly, i.e. 35% more flow can be processed through the heater with the same exit temperature of 545°C.

**Maximum Scale-Up = 1.35 or 135%**

### 1.2.2 Jet Mixer, J-601

This mixer is nothing more than a restriction in the line to provide added pressure drop and hence ensure good mixing. We need to control the ratio of propylene and chlorine feeds to the reactor in the same ratio as in the design case. The only restriction for this unit will be the maximum operating pressure, and this has already been accounted for in the analysis given above.

### 1.2.2 Reactor, R-601 and the Dowtherm Cooling Loop



#### Two Important Issues

- (i) What are input variables & what are output/response variables?
- (ii) How do we reconcile current operation with design calculations?

- (i) The flow system (exchangers, pumps, valves and pipes) exist. Therefore, all attributes associated with equipment, e.g., pump characteristics, pipe sizes, heat transfer areas, etc. are **fixed**. In an operating system, such as this one, the only variables that we can control are the flowrates of Dowtherm (using the control valve) and the flow of cooling water through E-601. Thus the only fixed inputs are the temperature of the cooling water into E-601 (30°C) and the equipment specifications. The variables shown on the diagram on the previous page can be categorized as follows.

Fixed (input) -  $t_A$  (30°C) + equipment parameters

Variable (input – specified by us) -  $m_D, m_{cw}$

Output (to be calculated) -  $T_A, T_B, t_B$ , and  $q$  (heat exchanged in E-601 and R-601)

C.1-5



ii. How do we reconcile current operation with the design calculations? The following points are noted from the problem statement.

(a) Current operations are close to design and have been checked recently (Table C.1, Table C.2, and the PFD are accurate reflections of current operating conditions)

(b) The heat transfer area for E-601 as built (2.6 m<sup>2</sup>) is significantly different from that used in design calculations (2.1 m<sup>2</sup>).

The important point to note here is that if the design calculations are correct and the equipment was built to the specifications given in Table C.2, then the current operations could not be close to design. In practice, a more rigorous set of calculations would have been performed prior to the construction of the plant and the actual plant, as it was built, reflects the results of these calculations. The utility of the design calculations provided in the problem statement is, therefore, not in predicting absolute values of heat transfer coefficients but rather in identifying and quantifying the relative magnitude of the heat transfer resistances. This will be illustrated further below. From this discussion it should be clear that the Base Case conditions should be chosen as the current operation.

#### Scale-Up Calculations for Dowtherm Loop

We use a ratio analysis to compare the scaled-up case (subscript 2) with the base case (subscript 1). We define the following terms in the analysis:

$Q$  = ratio of the heat removed from the reactor in scaled-up case ( $q_2$ ) to the heat removed in the base case ( $q_1$ ).

$M_D$  = ratio of Dowtherm A flowrate in scaled-up case ( $m_{D2}$ ) to Dowtherm A flowrate in base case ( $m_{D1}$ ).

$M_{cw}$  = ratio of cooling water flowrate in scaled-up case ( $m_{cw2}$ ) to cooling water flowrate in base case ( $m_{cw1}$ ).

We further assume that the specific heats of cooling water and Dowtherm A are the same for the two cases (this will be a good assumption since temperatures do not change greatly). Finally,  $q$  in the reactor must equal  $q$  in E-601 for steady state operation.

#### R-601

$$Q = \frac{q_2}{q_1} = \frac{m_{D2} C_{PD2} (T_B - T_A)}{m_{D1} C_{PD1} (400 - 350)} \Rightarrow Q = M_D \frac{(T_B - T_A)}{50} \quad \text{ENERGY BALANCE} \quad (1)$$

$$Q = \frac{U_2 A_2 \Delta T_{LM2}}{U_1 A_1 \Delta T_{LM1}} = \frac{\left[ \frac{1}{267} + \frac{1.14}{2500} + \frac{1.14}{1750 M_D^{0.8}} \right]^{-1} (T_B - T_A)}{\frac{1}{\ln \left\{ \frac{511 - T_A}{511 - T_B} \right\}} 133.4}$$

$$Q = \frac{(4.2013 \times 10^{-3} + 6.5143 \times 10^{-4} M_D^{-0.8})^{-1} (T_B - T_A)}{(206)(133.4) \ln \left( \frac{511 - T_A}{511 - T_B} \right)} \quad \text{PERFORMANCE EQN} \quad (2)$$

### E-601

#### ENERGY BALANCE

$$Q = \frac{M_{CW2} C_{PCW2} (t_B - 30)}{M_{CW1} C_{PCW1} (40 - 30)} \Rightarrow Q = M_{CW} \frac{(t_B - 30)}{10} \quad (3)$$

#### PERFORMANCE EQUATION

$$Q = \frac{U_2' A_2' / \Delta T_{LM2}}{U_2' A_2' / \Delta T_{LM1}} = \frac{\left( \frac{1}{1700 M_D^{0.8}} + \frac{1}{1700 M_{CW}^{0.8}} \right)^{-1}}{850} \frac{(T_B - t_B - T_A + 30)}{\ln \left( \frac{T_B - t_B}{T_A - 30} \right) 340}$$

$$Q = \frac{\left( \frac{1}{2 M_D^{0.8}} + \frac{1}{2 M_{CW}^{0.8}} \right)^{-1}}{340} \frac{(T_B - t_B - T_A + 30)}{\ln \left( \frac{T_B - t_B}{T_A - 30} \right)} \quad (4)$$

From the above we see that we have 4 equations in 6 unknowns ( $Q$ ,  $M_D$ ,  $M_{CW}$ ,  $T_B$ ,  $T_A$ , and  $t_B$ ). It is important to realize that we can manipulate or control only two of these variables in the plant - namely  $M_D$  and  $M_{CW}$ . Intuitively, we should set these as high as possible to obtain the maximum cooling possible. What are the maximum values possible for these flows (flow ratios)?

#### Cooling Water Flowrate, $M_{CW}$

From the problem statement we know that the current cooling water velocity is 2.0 m/s and that the maximum cooling water flowrate through any piece of equipment is 3.5 m/s (this limit is to avoid erosion). Thus we can increase the cooling water flowrate by a factor of 3.5/2.0, that is the maximum value of  $M_{CW}$  is 1.75.

$$M_{CW} = 1.75 \quad (5)$$

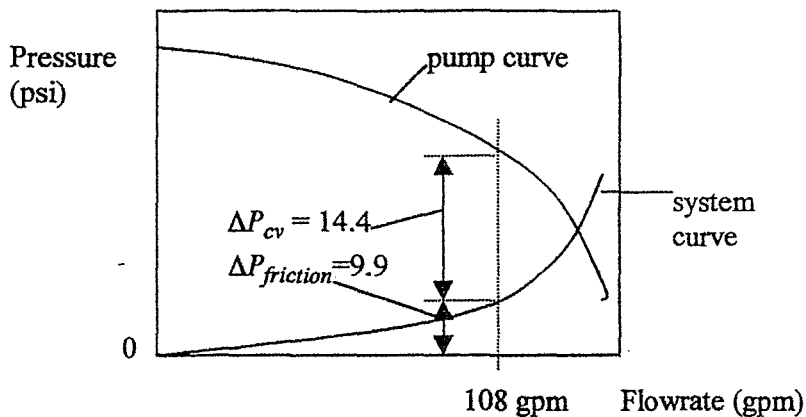
#### Dowtherm A Flowrate, $M_D$

In order to determine the maximum increase in Dowtherm A flowrate it is necessary to use the system curve with the pump curve provided (Figure C.2). First, locate the current operating condition on the pump curve. (review Chapter 12, section 12.2 for more details).

At base case conditions Dowtherm A flowrate = 4.62 kg/s = 108 gpm. From design calculations

|                                         |   |                                                                                    |
|-----------------------------------------|---|------------------------------------------------------------------------------------|
| $\Delta P_{cv} = 12.4 \text{ psi}$      | } | Total $\Delta P = 22.3 \text{ psi}$ - this point lies on the pump curve at 108 gpm |
| $\Delta P_{friction} = 9.9 \text{ psi}$ |   |                                                                                    |

We note that at a flow of 108 gpm the frictional pressure drop = 9.9 psi. Also for a closed loop there are no hydrostatic head differences so that system curve looks like the following.



We consider three possible cases

- (i) Operate single pump with control valve wide open
- (ii) Operate two pumps in series
- (iii) Operate two pumps in parallel

We can also reconfigure the reactor tubes in parallel which shifts the system curve downwards. For this case the flowrate through each bank of tubes would be 1/3 of the flowrate through the pump, and the equivalent length of pipe would be 1/3 of the original reactor coil.

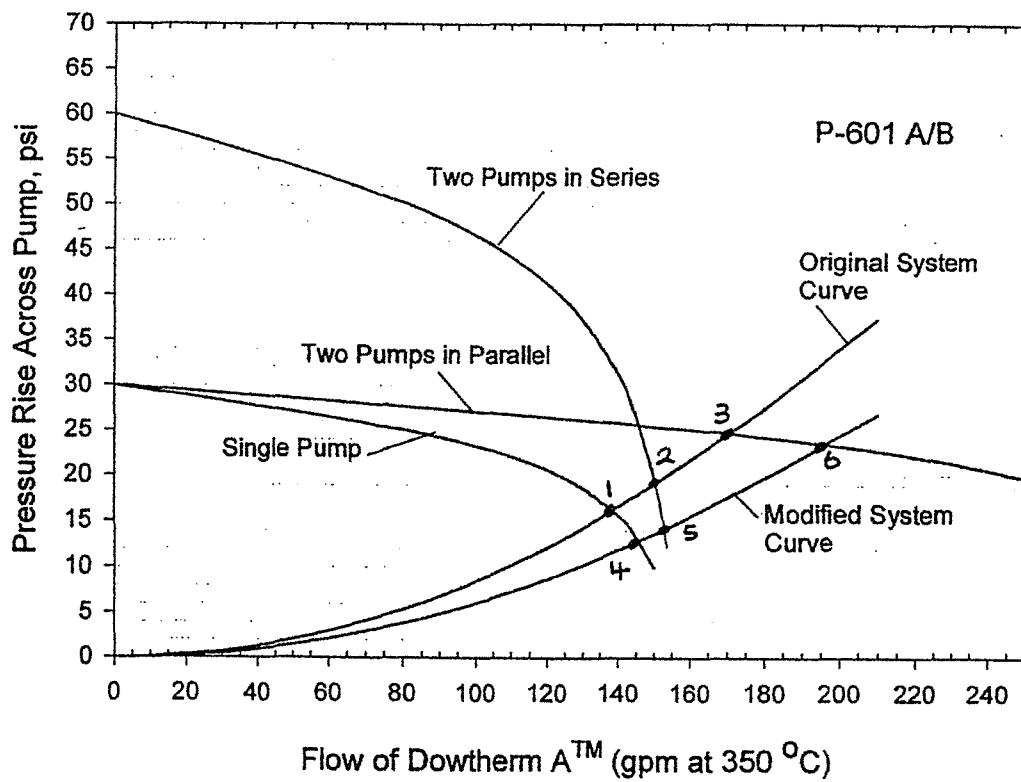
Now  $\Delta P_{friction, R-601, 2} = \Delta P_{friction, R-601, 1} (1/3)^2 (1/3) = \Delta P_{friction, R-601, 1} (1/27) = 2.9/27 = 0.107$  psi

For this case the base case frictional pressure drop =  $7 + 0.107 = 7.107$  psi, down from 9.9 psi when reactor tubes piped in series.

The pump curves and system curves for all the cases given above are shown in the figure on the next page. The maximum flows of Dowtherm A for all combinations of cases are given by the intersections of the appropriate system and pump curve – which corresponds to the case when the control valve is wide open. The scale up factors for each case are listed below:

| <u>Case</u>                      | <u>Point on Graph</u> | <u>Flowrate</u> | <u><math>M_D</math></u> |
|----------------------------------|-----------------------|-----------------|-------------------------|
| <u>Reactor Tubes in Series</u>   | (next page)           | (gpm)           |                         |
| 1 pump                           | 1                     | 137             | 1.27                    |
| 2 pumps in series                | 2                     | 150             | 1.39                    |
| 2 pumps in parallel              | 3                     | 170             | 1.57                    |
| <u>Reactor Tubes in Parallel</u> | (next page)           | (gpm)           |                         |
| 1 pump                           | 4                     | 146             | 1.35                    |
| 2 pumps in series                | 5                     | 153             | 1.42                    |
| 2 pumps in parallel              | 6                     | 196             | 1.81                    |

Clearly if we want to maximize flow through the loop we should operate both pumps in parallel. The valving will already exist since spared pumps are piped in parallel (see the P&ID in Figure



1.7). Clearly running both pumps in parallel with the control valves wide open leaves the operators with nothing to control reactor upsets with. Although this situation would not be acceptable for long term operations the necessity to increase production as much as possible might provide the incentive to operate like this for a short time. Even if this operation were deemed unsafe it does allow a maximum scale-up level to be assessed.

#### Maximum Scale-up for Reactor Loop

For 2 pumps running in parallel and the reactor tubes piped in series we put  $M_D = 1.57$  and  $M_{cw} = 1.75$  into equations (1) – (4) and solve to give  $Q = 1.34$ . Thus a scale-up of approx. 134% is possible. If we rearrange the reactor tubes in parallel then the form of equation (2) must be modified to take account of the reduced flow through each tube bank as follows:

$$Q = \frac{(4.2013 \times 10^{-3} + 6.5143 \times 10^{-4} (M_D/3)^{-0.8})^{-1} (T_B - T_A)}{(206) (133.4) \ln \left\{ \frac{511 - T_A}{511 - T_B} \right\}} \quad (2A)$$

Solving equations (1), (3), and (4) along with the modified equation (2) above gives us  $Q = 1.337$ . Thus a scale-up of 134% is possible for this configuration. From this we see that the maximum scale-up for the reactor cooling loop is 134%.

**Maximum Scale-up = 1.34 or 134%**

#### 1.2.3 Waste heat Boiler, E-602, and the Crude Allyl Chloride Cooler, E-603

We see from the PFD that there is a level controller on the BFW level in E-602. However, in the problem statement it states that all the heat transfer tubes are covered by BFW so that there is no immediate fix that we can implement. Note: that if the level in the exchanger were say only covering 90% of the tubes we could have immediately increased the amount of heat removal by 11% by simply covering all the tubes.

In addition, another “quick fix” would be to reduce the pressure in the steam header causing a lower boiling temperature and hence increasing the  $\Delta T$  driving force in E-602. However, since this steam is probably being used elsewhere in the plant, a lower steam pressure would probably cause problems elsewhere in the plant.

Having eliminate some of the “easy” fixes we must concentrate on the heat transfer analysis.

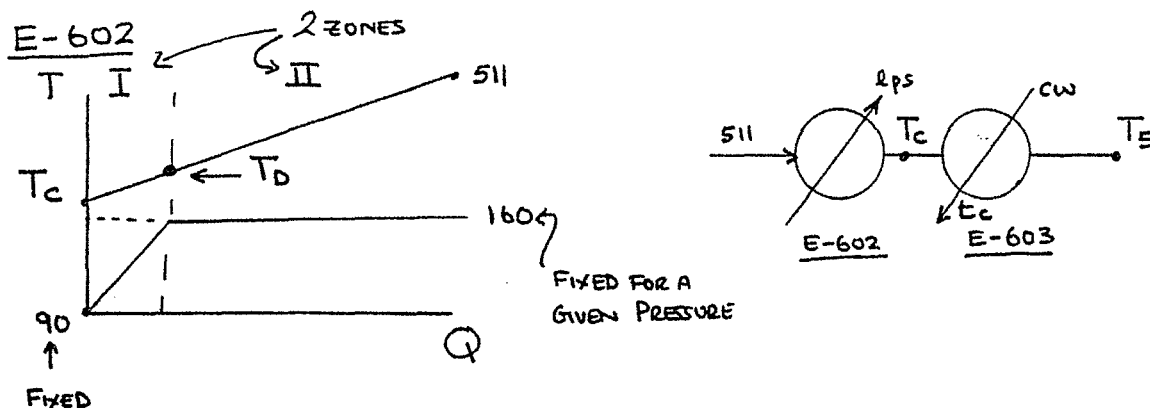
#### Analysis of E-602

$Q = UA\Delta T_{lm}$  this applies to both zones (desubcooling and boiling)

Note that since all the heat transfer resistance is on the tubeside (process stream is a gas) we can scale  $U$  by  $(m_{process})^{0.8}$ . Since we are currently operating at design conditions, that is the exit temperature from E-602 is 200°C, we conclude that as the process flow increases so will the exit temperature.

Of course, as the exit temperature from E-602 increases this will directly impact the performance of E-603 (also currently operating under design conditions, that is exit temperature is 50°C). However, for E-603 we can increase the cooling water flowrate, although this does not change the heat transfer coefficient (again we assume all the resistance is on the process gas side), it does change the  $\Delta T$  driving force since the exit temperature of the cw decreases.

The equations describing the performance of E-602 and E-603 are given below.



USE SUBSCRIPT 2 FOR SCALED-UP CASE & SUBSCRIPT 1 FOR BASE (DESIGN) CASE.  
 RATIO OF STEAM FLOWS  $M_{S2}/M_{S1} = M_S$ . RATIO OF PROCESS FLOWS  $M_{P2}/M_{P1} = M_P$ . RATIO OF HEAT TRANSFERRED,  $Q_2/Q_1 = Q$ .

ENERGY BALANCE ON E-602

$$Q = \frac{Q_2}{Q_1} = \frac{M_{S2}}{M_{S1}} = \frac{M_{P2}}{M_{P1}} \frac{(511 - T_c)}{(511 - 200)} \frac{C_{P2}}{C_{P1}} \Rightarrow M_S = M_P \frac{(511 - T_c)}{311} \quad (6)$$

INTERMEDIATE TEMPERATURE  $T_D$  (AT INTERSECTION OF ZONES)

$$\frac{T_D - T_c}{h_{fs} - h_f} = \frac{511 - T_D}{h_{fg}} \Rightarrow T_D = 63.8 + 0.875 T_c \quad (7)$$

NOTE IF T

PERFORMANCE EQUATION FOR ZONE II (Boiling)

$$Q = \frac{Q_{II2}}{Q_{II1}} = M_S = \left( \frac{M_{P2}}{M_{P1}} \right)^{0.8} \left( \frac{A_{II2}}{A_{II1}} \right) \frac{(511 - T_D)}{\ln \left( \frac{511 - 160}{T_D - 160} \right)} \frac{1}{182.4} \quad (8)$$

$M_P^{0.8}$        $A_{II1} = 44.54 \text{ m}^2$

PERFORMANCE EQUATION FOR ZONE I (DESUBCOOLING)

$$Q = \frac{q_{I2}}{q_{I1}} = M_s = \left( \frac{M_{P2}}{M_{P1}} \right)^{0.8} \left( \frac{A_{I2}}{A_{I1}} \right) \frac{(T_c - T_D + 70)}{\ln \frac{(T_c - 90)}{(T_D - 160)}} \frac{1}{93.6} \quad \text{--- (9)}$$

$\downarrow$   
 $M_P^{0.8}$

$\uparrow$   
 $A_{I1} = 12.46 \text{ m}^2$

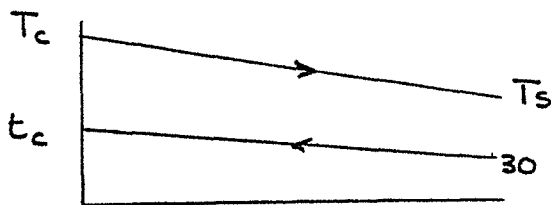
AREA BALANCE

$$A_{II2} + A_{I2} = 57.0 \quad \text{--- (10)}$$

UNKNOWNNS  $M_P, M_s, A_{II2}, A_{I2}, T_c, T_D$  & 5 EQUATIONS

$\therefore$  GIVEN  $M_P$  (SCALE-UP FACTOR) WE CAN SOLVE.

E-603



ENERGY BALANCE

$$Q = \frac{q_2}{q_1} = \frac{M_{P2} C_{P2} (T_c - T_s)}{M_{P1} C_{P1} (200 - 50)} = \frac{M_{CW2} C_{PW2} (t_c - 30)}{M_{CW1} C_{PW1} (40 - 30)}$$

$$M_P \frac{(T_c - T_s)}{150} = M_{CW} \frac{(t_c - 30)}{10} \quad \text{--- (11)}$$

PERFORMANCE EQN

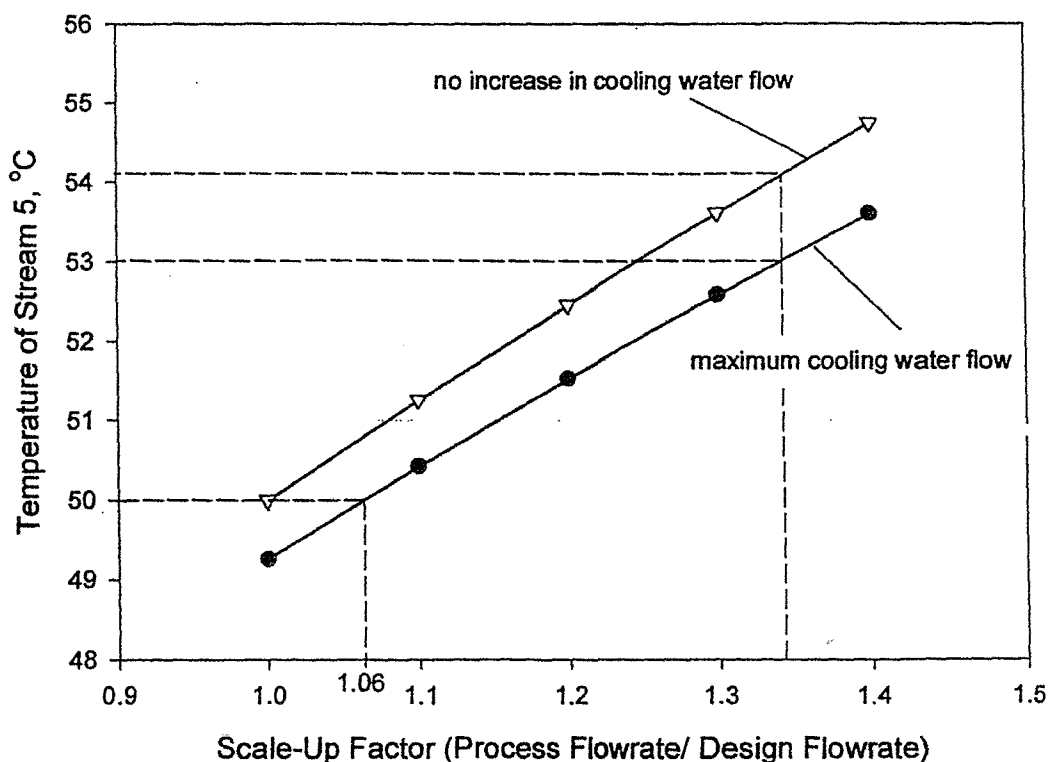
$$Q = \frac{q_2}{q_1} = M_P \frac{(T_c - T_s)}{150} = \left( \frac{M_{P2}}{M_{P1}} \right)^{0.8} \frac{(T_c - t_c - T_s + 30)}{\ln \frac{(T_c - t_c)}{(T_s - 30)}} \frac{1}{67.33} \quad \text{--- (12)}$$

IF WE INCLUDE INFO. FROM E-602 THEN FOR A GIVEN  $M_P$  WE KNOW  $T_c$   $\therefore$  REMAINING UNKNOWNNS ARE  $T_s, t_c$  &  $M_{CW}$   $\begin{cases} \text{SET THIS TO 1.75} \\ \text{\& SOLVE} \end{cases}$

By solving the equations for E-602 and E-603 we can estimate the value of the exit temperature from E-603 as a function of scale-up. Assuming that we can increase the cooling water flow by 1.75 (same as in reactor case) we find that we can increase the process flow by **1.06 or 6%** and still maintain the desired exit temperature of 50°C. A plot of exit temperature from E-603 as a function of scale-up is shown below. For a 34% increase in process flow the exit temperature increases to 53.0°C. If this temperature change can be accommodated down stream then 34% scale-up is possible. Otherwise only a 6% scale-up can be achieved.

**Scale-up = 1.06 or 106%**  
[if temperature of Stream 5 must = 50°C]

Exit Temperature from E-603 as a Function of Scale-Up



We conclude that scale-up is limited to 106% of current operation if  $T_5$  is limited to 50°C but can be increased to 134% (our previous maximum scale-up) if  $T_5$  can be increased slightly.



## 2. Long Term Economic Improvement

Look at variable operating costs.

| <u>Raw Materials</u>                                          | <u>Unit Cost</u> | <u>Yearly Cost</u>                           |
|---------------------------------------------------------------|------------------|----------------------------------------------|
| Propylene                                                     | \$ 0.47/kg       | $(3187)(8000)(0.47) = \$ 11.98 \times 10^6$  |
| Chlorine                                                      | \$ 0.18/kg       | $(1399)(8000)(0.18) = \$ 2.01 \times 10^6$   |
| Propylene Recycle<br>(100% recovery of propylene in Stream 5) | \$ 0.47/kg       | $(2439)(8000)(0.47) = (\$ 9.17 \times 10^6)$ |
| Total                                                         |                  | \$ 4.82 $\times 10^6$                        |

The raw material costs are usually the biggest cost for a process. The above estimate is optimistic since it is assumed that all the propylene in Stream 5 is recovered and recycled. If this were not the case then a recommendation to improve separation of propylene would be appropriate. The overall conversion of both raw materials is very high. However, the selectivity to allyl chloride is relatively poor with 87.3 % of the propylene going to allyl chloride and 79.0% of the chlorine going to allyl chloride. These yields might be improved significantly by using a catalyst and this would be an excellent suggestion (note additional kinetic information would be required to quantify the benefits of changing to a catalytic process).

| <u>Utilities</u> |              |                                                                         |
|------------------|--------------|-------------------------------------------------------------------------|
| Natural Gas      | \$ 2.5/GJ    | $(4000 \times 10^6)(8000)(2.5 \times 10^{-9})/0.9 = \$88,900$           |
| CW               | \$ 0.16/GJ   | $((2188+1025) \times 10^6)(8000)$<br>$(0.16 \times 10^{-9}) = \$ 4,100$ |
| Electricity      | \$ 0.06/kW-h | $(2.5)(0.06)(8000) = \$ 1,200$                                          |
| LPS credit       | \$ 3.17/GJ*  | $(2850 \times 10^6)(8000)(3.17 \times 10^{-9}) = (\$72,300)$            |

\* the cost of BFW is not included since it is assumed that all the steam produced in E-602 will be returned as condensate .

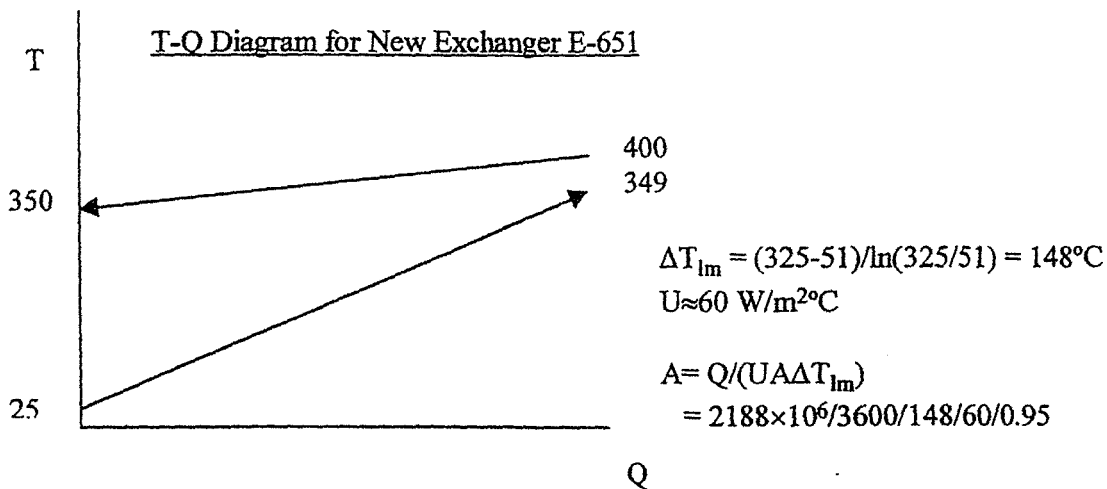
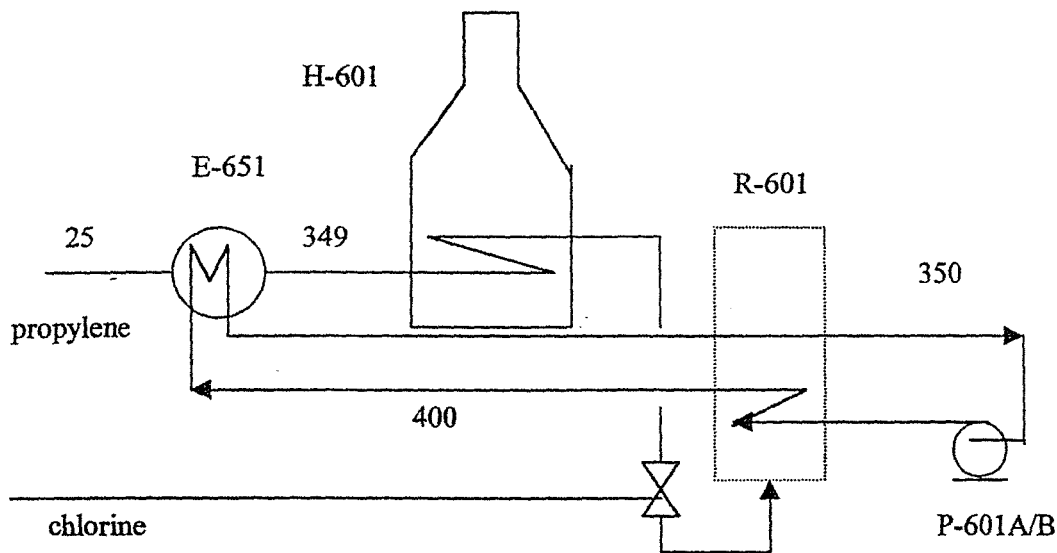
Although the total utility costs (\$94,200/yr) are relatively small compared to the raw material costs, significant savings may still be obtained.

### 2.1 Ways to Reduce Utility Costs

#### 2.1.1 Heat Integration

(a) Make more lps, mps, or hps in Dowtherm A loop instead of using cooling water in E-601. This is a good idea so long as we can sell the extra steam. If we can sell it, then it is unlikely that we will be able to get anything more than the equivalent fuel credit.

(b) Reduce the natural gas usage in H-601 by preheating propylene.



$A = 72.0 \text{ m}^2$  using area we can estimate the fixed capital investment  
 MOC = SS, Shell and Tube floating head, Operating Pressure = 11.0 bar.

$FCI = C_{TM} = \$ 88,000$

Savings in natural gas =  $(2188 \times 10^6)(8000)(2.5 \times 10^{-9}) / 0.9 = \$ 48,600$

$EAOC = C_{TM} (0.15)(1.15)^5 / (1.15^5 - 1) - 48600 = \$ -22,350/\text{yr}$

A negative EAOC means a positive savings – implement this scheme.

Many other heat savings schemes possible but this may be easiest to implement.

#### 2.1.2 Replace control valves with turbines/expanders

We can not get much work out of streams and turbines are expensive. Propylene flow is too small and the pressure drops are too small.

### 3. Observations from Plant

- 3.1 Pump makes a whining noise
- 3.2 Insulation is falling off
- 3.3 Relief valve on E-602 is leaking steam

Students should address these points if only to bring them to someone's attention. Often we find students do not make any comment about these observations – when asked they say that they noticed them but couldn't figure out what the reasons could be so they ignored them!! Must report these potential problems to your supervisor – at the very least!

#### 3.1 Pump makes a whining noise

- Probably a bad bearing. Recommend maintenance crew takes it off line and performs inspection and needed repairs.
- These repairs should be carried out before scale-up starts since both pumps may be required to operate in parallel.
- The noise is almost certainly not caused by cavitation

The inlet to the pump is at 350°C at which Dowtherm A has a vapor pressure of 5.2 bar. The minimum pressure of the stream entering P-601 is about 10.9 bar which gives a  $NPSH_A = 10.9 - 5.2 = 5.7$  bar. No  $NPSH_R$  is given for this pump but typical values are given in Chapter 9 and none are greater than about 0.6 bar. Therefore, very unlikely that pump is cavitating.

#### 3.2 Insulation is falling off line from reactor to E-602.

- We have a small additional heat loss from the system and a very small loss in steam production from E-602
- It is possible that insulation was taken off deliberately. If pipe had a long horizontal run then the metal could creep and the pipe might begin to sag. By removing the insulation the wall temperature of the pipe would drop significantly and the sagging would cease!
- There is no information given as to whether the pipe is sagging but it is worth checking before we try to reinsulate.
- Personnel safety may also be an issue. However, this is only a concern if an operator could touch the pipe – this should also be checked.

#### 3.3 Steam leaking from safety relief valve (RV) on E-602

Possible causes include:

- (i) Some dirt is stuck in the valve seat causing a leak. Remedy - check and fix.
- (ii) RV is set to too low a pressure. Remedy – check and reset to correct pressure.
- (iii) Shell side of E-602 is at 8 bar (max. op. press.) – two possible ways that this could occur:
  - (a) valve on steam line is closed and BFW pump is pressurizing E-602. Remedy – check and fix.
  - (b) LPS header is at >8 bar and is pressurizing E-602. Remedy – check and fix.

Since no information is given on the steam at the plant, i.e., pressure or flowrates have been fluctuating, (i) and/or (ii) are the most likely causes.

## Project C.2 – New 20,000 metric tons per year Allyl Chloride Facility

Look at the base case process, which should be taken as the existing Beaumont facility scaled-up to 20,000 metric tons per year. The economic breakdown of the facility is as follows:

Total utility cost,  $C_{UT} = \$ 3.42 \times 10^6$  per year

Cost of operating labor,  $C_{OL} = \$ 1.08 \times 10^6$  per year

Raw material costs,  $C_{RM} = \$ 9.43 \times 10^6$  per year

Waste treatment cost,  $C_{WT} = \$ 0.26 \times 10^6$  per year

Gross roots capital cost of plant,  $C_{GR} = \$ 15.39 \times 10^6$  per year

Annual Revenue,  $R = \$ 37.98 \times 10^6$  per year (34.3 from allyl chloride, and 3.2 from HCl)

Assuming the following:

Working capital =  $\$ 3.08 \times 10^6$

Cost of Manufacture,  $COM_d = \$ 21.83 \times 10^6$

$i=10\%$ ,  $t=0.15$ , start-up = 2 years, plant life = 10 years

Depreciation = MACRS (20, 32, 19.2, 11.52, 11.52, 5.76)

We get a NPV =  $\$ 56.3 \times 10^6$  (note:  $NPV \approx \$30 \times 10^6$  when  $t=0.4$ ) see attached spreadsheet

Let us analyze the economics of the plant – start with the biggest costs and work our way down. This follows the Pareto analysis discussed in Chapter 19.

Raw Material Costs ( $\$ 9.43 \times 10^6$  per year)

Overall conversion of chlorine  $\approx 100\%$

Overall conversion of propylene  $\approx 100\%$

} Clearly small losses will occur

Despite the fact that the raw material cost is the highest operating cost there is nothing we can do to improve the overall utilization of the raw materials, i.e., conversion cannot be increased. However, the raw material usage could be improved by improving selectivity. From problem statement we have little control over this element of the design and thus the raw material usage is already optimized. However, suggestions regarding finding alternative reaction pathways (e.g. catalytic) are to be encouraged.

Utility Costs ( $\$ 3.42 \times 10^6$  per year)

Largest users of utilities are

Refrigeration Compressors =  $\$ 2.66 \times 10^6$  per year (78% of  $C_{UT}$ )

Propylene Recycle Compressors =  $\$ 0.21 \times 10^6$  per year (6% of  $C_{UT}$ )

Fired Heater =  $\$ 0.17 \times 10^6$  per year (5% of  $C_{UT}$ )

⇒ Look at refrigeration loop

C.2-1

**Base Case for 20,000 tonnes/yr Allyl Chloride Facility (scaled-up Beaumont design)**

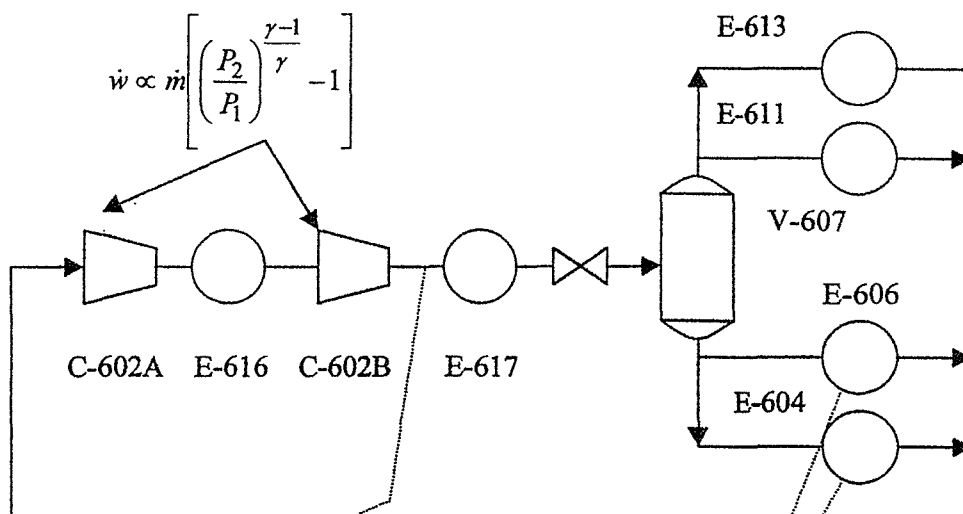
| <u>Year</u> | <u>FCI + WC</u> | <u>Revenue</u> | <u>COMd</u> | <u>d</u> | <u>(R-COMd-d)(1-t)+d</u> | <u>CF</u> | <u>CCF</u> | <u>DCF</u> | <u>CDCF</u> |
|-------------|-----------------|----------------|-------------|----------|--------------------------|-----------|------------|------------|-------------|
| 0           |                 |                |             |          |                          |           |            |            |             |
| 1           | -7.695          |                |             |          |                          | -7.70     | -7.70      | -7.00      | -7.00       |
| 2           | -10.773         |                |             |          |                          | -10.77    | -18.47     | -8.90      | -15.90      |
| 3           |                 | 37.98          | 21.83       | 3.08     | 14.19                    | 14.19     | -4.28      | 10.66      | -5.24       |
| 4           |                 | 37.98          | 21.83       | 4.92     | 14.47                    | 14.47     | 10.19      | 9.88       | 4.64        |
| 5           |                 | 37.98          | 21.83       | 2.95     | 14.17                    | 14.17     | 24.36      | 8.80       | 13.44       |
| 6           |                 | 37.98          | 21.83       | 1.77     | 13.99                    | 13.99     | 38.35      | 7.90       | 21.34       |
| 7           |                 | 37.98          | 21.83       | 1.77     | 13.99                    | 13.99     | 52.35      | 7.18       | 28.52       |
| 8           |                 | 37.98          | 21.83       | 0.89     | 13.86                    | 13.86     | 66.21      | 6.47       | 34.99       |
| 9           |                 | 37.98          | 21.83       |          | 13.73                    | 13.73     | 79.93      | 5.82       | 40.81       |
| 10          |                 | 37.98          | 21.83       |          | 13.73                    | 13.73     | 93.66      | 5.29       | 46.10       |
| 11          |                 | 37.98          | 21.83       |          | 13.73                    | 13.73     | 107.39     | 4.81       | 50.91       |
| 12          | 3.078           | 37.98          | 21.83       |          | 13.73                    | 16.81     | 124.19     | 5.35       | 56.27       |

**NPV = 56.27**

FCI = 15.39  
 taxation rate, t= 15%  
 utilities cost 3.42  
 raw materials cost = 9.43  
 waste treatment cost = 0.26  
 revenue = 37.98  
 discount rate, i = 10%  
 depreciation = MACRS (20,32,19.2, 11.52, 11.52, 5.76)

C2-2

## Interesting Points about the Refrigeration Loop



$$\dot{w} \propto \dot{m} \left[ \left( \frac{P_2}{P_1} \right)^{\frac{\gamma-1}{\gamma}} - 1 \right]$$

Why is this set at 20 bar?

Because this is the lowest pressure at which we can condense propylene with cooling water.

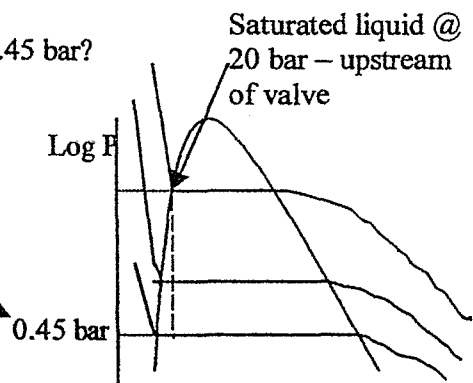
Why do we not vaporize all the propylene in throttling process?

Because these exchangers have large duties and are used to condense overhead products in distillation columns. Thus it is better to vaporize cold liquid propylene (at constant temperature) than to use cold propylene vapor.

Why do we use a 5°C temperature approach for E-604 and E-606? Because there is a trade-off between compression costs (to cool the propylene below current low temperature) and heat exchanger costs (heat transfer area). Since the electric utility costs for compression are so high, the use of a 5°C temperature approach (instead of the normal 10°C minimum value) is justified.

Why is the pressure at the inlet to C-602A set at 0.45 bar?

Setting this pressure sets the temperature



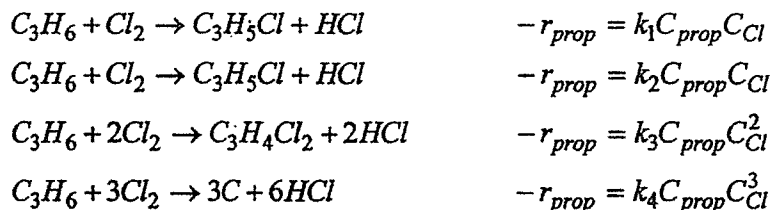
C12-3

We can optimize the refrigeration loop a little bit but we can't reduce the power requirement by much. What we need to do is to increase the temperature in the process in order to reduce the load on the refrigeration loop. If we look at the PFD for the process we see that we need a low temperature in the distillation column condensers (namely in T-601 and T-603) and a low inlet temperature to T-601. The reason for this is so that we can form an L-V mixture. If we refer back to Chapter 8 and look under the tables of special concerns for separators, we see that either low temperature or high pressure may be required to obtain an L-V mixture suitable for distillation. Therefore, pressurizing the columns should have a similar effect as using a low temperature in the condensers. We can pressurize the columns in several ways:

- (1) The conservative method is to compress Stream 5.
- (2) Alternatively we may pressurize the whole front end of the process by pumping the feed streams (propylene and chlorine) to a high pressure.

Alternative (1) is conservative because we do not change the pressure of the reaction section of the process and hence the selectivity will not change.

Alternative (2) is potentially cheaper in utility costs since we have eliminated the compressor but the front-end equipment, e.g., R-601, H-601, E-602 and E-603, must be designed to withstand higher pressures and will be more costly. Additionally the pressure will change considerably and the selectivity may change. Although no reaction kinetics are given consider the effect of increasing pressure on the competing reactions assuming elementary kinetic expressions.



$$\text{Selectivity} = \frac{k_1 C_{prop} C_{Cl}}{k_1 C_{prop} C_{Cl} + k_2 C_{prop} C_{Cl} + k_3 C_{prop} C_{Cl}^2 + k_4 C_{prop} C_{Cl}^3}$$

The effect of increasing the pressure can be seen in the effect that the pressure has on the gas phase concentration. Assuming ideal gas behavior we have  $C \propto P/RT$ , therefore as pressure increases the concentration increases proportionally. Looking at the expression for selectivity we can see that the denominator increases more with increasing pressure than does the numerator. Thus the selectivity is expected to decrease with increasing pressure. Thus Alternative (1) may be better.

#### Why compress Stream 5 and by how much?

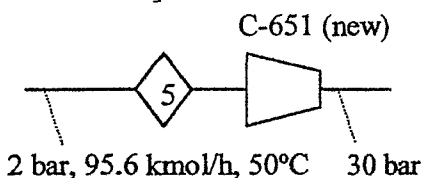
It can be argued that little is to be gained by pressurizing Stream 5 since we are replacing one set of compressors (refrigeration compressors C-602A&B) with another compressor(s) in Stream 5.

C.2-4

Also if we do add compressors to what pressure should we compress Stream 5? This is the subject of a parametric optimization (see Chapter 19) but let us say that for comparison we will choose to compress Stream 5 to a pressure which will allow the refrigeration loop to be eliminated, i.e., so that we can use cooling water in the condensers. This pressure is approximately 30 bar.

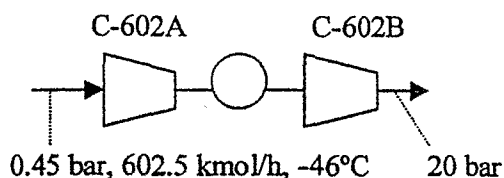
Look at two compressor schemes:

#### New Compressor in Stream 5



$$P_2/P_1 = 15$$

#### Refrigeration Loop



$$P_2/P_1 = 44.4$$

If we compare the theoretical power for C-651 to that of C-602 we get approximately the following relationship (assuming a single stage of compression):

$$\frac{w_2}{w_1} = \frac{m_2 T_2 [(P_2/P_1)_2^a - 1]}{m_1 T_1 [(P_2/P_1)_1^a - 1]} = \frac{(95.6)(323) [(15)^{0.286} - 1]}{(602.5)(227) [(44.4)^{0.286} - 1]} = 0.135$$

Clearly the option of placing a compressor(s) in Stream 5 to replace the refrigeration compressors is a good idea.

#### What are the Consequences of Increasing the Pressure in Stream 5 to 30 Bar on the Separation ?

Obviously we have chosen 30 bar pressure so that we can eliminate the costly refrigeration loop and use cooling water in the overhead condensers. However, as the top temperature in the columns increase so do the bottoms temperatures. For the first column, T-601, operating at 30 bar the bottom temperature is approximately 200°C compared to 45°C previously. The question now arises as to whether allyl chloride is stable at this temperature. The following is a quote from the Kirk-Othmer encyclopedia, Vol 2, p106, "Allyl compounds are normally stable at room temperature even in the absence of inhibitors. However, prolonged exposure to air, especially in the presence of light or heat, can cause peroxide formation. Contamination with either peroxides, or traces of catalytically active transition metals, can present hazards on heating as a result of rapid and strong exothermic polymerization." The reason that the Beaumont facility operated at very low temperatures was to avoid the problems associated with this uncontrolled polymerization!!

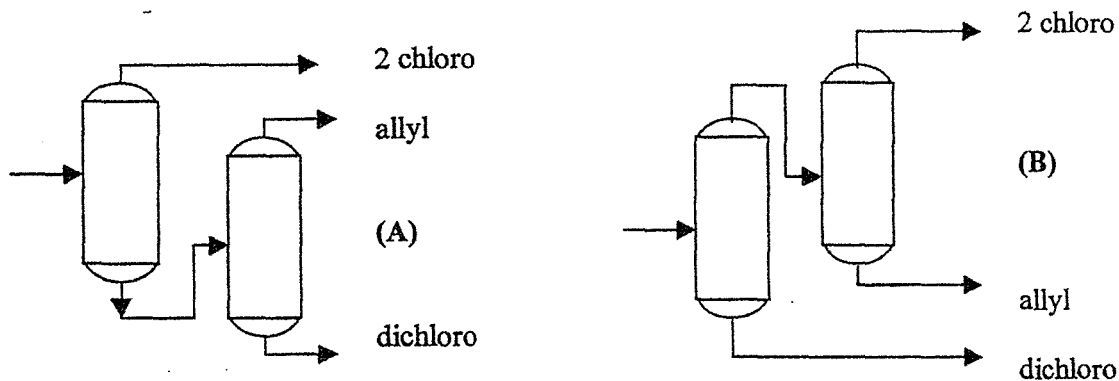
The lesson here is that there are large savings to be made in the operating costs of the plant by increasing the pressure of the separations section. However, whenever we change the process conditions by a large amount we must be sure that we are still in a safe operating region and that we take into account all the potential effects that these changes may have on the process.



### Other Issues

Some other issues which should be considered in the optimized design, but which have less impact on the overall economics are listed below:

- (1) Changing the sequence of the columns. Allyl chloride has the largest flowrate but is intermediate in volatility (2 chloro -allyl -dichloro). We could change the sequencing as follows:



There appears to be little advantage of (B) over (A) so the current sequence is probably OK.

- (2) The first two columns, T-601 and T-603, appear to be doing the same separation. This is wasteful since we only require a sharp split between the propylene/HCl and the organic chlorides. We should just make the first column a little taller (more stages) and improve the separation thus eliminating the need for T-603 (this is a topological optimization).
- (3) Since we take the overhead product from T-601 as a liquid then immediately vaporize it in E-607 it would seem a good idea to use a partial condenser for T-601 and save on utilities.
- (4) Why do we recycle propylene as a liquid? This is a nice option to have when shutting down the process since we can liquefy the recycle and sent it to storage. However, during regular operations we just vaporize the recycle. We may want to eliminate C-601A and B or not operate these compressors during normal operation. Note that if we operate the separations section at above 20bar (T-601 and T-602) we can recycle and liquefy the propylene without the use of compressors!
- (5) In the front end of the process it is worth looking at heat integration
- (i) Make high pressure steam in Dowtherm A cooler.
  - (ii) Drop the temperature into the fluidized bed. Since the reactor is essentially well mixed, (at least the solids are) we can feed the reactants into the reactor at temperatures

Ci 2-6

significantly below the reactor operating temperature. This also reduces the duty on the Dowtherm A cooling loop since the heat of reaction is now used in heating the incoming feed.

(iii) The heat from the reactor effluent stream can be utilized more efficiently.

There are probably many more cost saving measures that can be implemented, but the major ones have been addressed above.

### Project 3

This project is concerned with determining whether 50% scale-down is possible for this process, and how it can be accomplished. One possible solution that some students have considered as an alternative, but have never staked an entire project grade on is to construct a tank or tanks to store 50% of the product, and to continue to operate the plant at full capacity for one-half year. An atmospheric pressure, carbon steel tank to hold 40,000 metric tons of phthalic anhydride (about 40,000 m<sup>3</sup>) would cost between 2 and 3 million dollars.

Parts of this problem are given as end-of-chapter problems in the text. One challenge for students is to recognize that determining the conditions for 50% scale-down is equivalent to defining and solving several of these end-of-chapter problems. A second challenge is to recognize that, in the context of a chemical process, these end-of-chapter problems are coupled. The solution to one affects another, and in many cases, simultaneous solution is required.

Problem 22-13 asks the students to determine potential bottlenecks to the scale down. Detailed calculations not presented elsewhere are presented in that problem.

One bottleneck is the feed section. This is partly due to poor placement of the only control valve after the mixing point. We call this a caricature problem because it illustrates the point that it is far better (and standard practice) to place one control valve after the pump and one either before or after the compressor. The result of having only one control valve is that only one air flowrate is possible for 50% reduction in naphthalene. This assumes that there is only one impeller and shaft available on this compressor, i.e., it only operates at one rpm. This is discussed and analyzed in Section 19.4. Because the total flow to the reactor is reduced to about 1/3 of the original flow, maintenance of fluidization in the reactor is an unknown, and another potential bottleneck. Insufficient information is given in the problem, and what we want is for students to raise the question of maintenance of fluidization.

Another bottleneck is the distillation columns. They must be scaled down by 50%. The coupled heat exchanger-column temperature and pressure analysis, as shown in Section 19.2, is presented with Problem 22.13. Since sieve trays are involved, weeping may be an issue. A method for avoiding weeping is to increase the reflux ratio in each tower. Another issue is the low pressure of each tower. Because of the vacuum conditions, the required pressure increases are significant on an absolute scale. This results in a significant density increase which could also promote weeping, since the column diameter would be much larger than it would need to be.

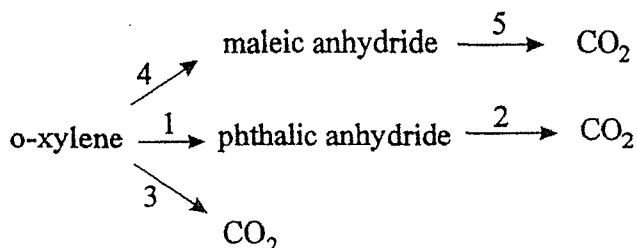
It is also necessary to analyze the molten salt loop to determine how much reduction in molten salt is required. It is not true that the molten salt flowrate reduces by 50%. In fact, the reduction is to 45% of the original value. This solution is presented in Problem

19.12. In this solution, the energy balance and heat exchanger equations for the reactor and for the molten salt cooler must be solved simultaneously using the methods developed in Section 19.3. This is a good example of how analysis of a process involves coupled end-of-chapter problems. Analysis of either the reactor or the molten salt cooler involves solution of two simultaneous, non-linear equations. In the context of the molten salt loop, these two problems become coupled and the result is a set of five non-linear equations, three of which are coupled.

When we assign this problem to students, the best projects analyze one or two of the above bottlenecks well but include nothing about the others. Since analysis of the molten salt loop is specifically requested, all students attempt to solve this problem. As might be expected, there is often difficulty in correctly analyzing the simultaneous equations involved. One common error is not understanding that the fluidized bed can be assumed to be at a constant temperature and not accounting for this in the heat transfer analysis for the reactor. This results in using the reactor feed temperature in the expression for  $\Delta T_{lm}$ .

#### Project 4

Let us begin the analysis of this problem by examining the reaction network:



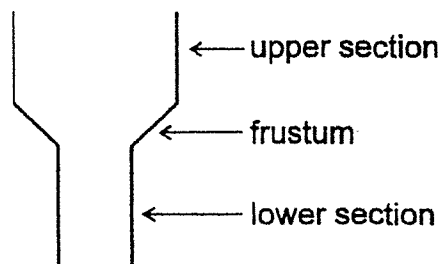
Intuitive examination of the activation energies suggests that the lowest allowable temperature will maximize the selectivity for reaction 1 over reactions 3 and 4. Once reaction 1 is selected, maximization of the selectivity for phthalic anhydride occurs at low conversions because reactions 1 and 2 are of the form  $A \rightarrow B \rightarrow C$ . Figure E14.3 illustrates this concept. More details are presented in Problem 14.3. The idea that a low conversion of reactant is a desirable result is always disconcerting to students. In fact, when we assigned this project, the students who arrived at the conclusion that low conversion was an optimum solution did so because of exhaustive optimization not because of the above analysis! The optimum conversion of o-xylene found was in the 30% to 40% range.

Obtaining the above optimum depends upon other topological choices having been made. One involves whether to operate above or below the flammability limit for o-xylene. Due to the enormous cost of the air compressor, operation above the flammability limit is chosen to minimize the amount of air needed.

The large compressor cost also affects the reactor design. The lowest operating pressure for the reactor without requiring additional compression before the switch condensers is optimal. It is also a good idea to arrange multiple reactor units in parallel (rather than have one large unit) in order to minimize the pressure drop and save on compression costs.

Because of the low conversion in the reactor, an additional distillation column is needed to separate and recycle the unreacted o-xylene. In this sequence, the order of decreasing volatility is o-xylene, maleic anhydride, phthalic anhydride. The question of distillation sequencing arises, and the optimum, by a very slight margin, is removal of o-xylene for recycle first, followed by the maleic/phthalic separation. However, this is such a close call that slight changes in constraints could alter this result. In terms of distillation column design, vacuum conditions suggest the possibility of a design commonly used in practice, but which seems unusual to students. Because of the large changes in absolute pressure in the column, the gas density is much higher at the bottom where the pressure is higher. Therefore, the design diameter is smaller at the bottom. A

typical vacuum column for this situation has a larger diameter upper section and a smaller diameter lower section with the two sections swaged using a frustum.



Another possibility is to have the same diameter for the entire column but to have less fractional active tray area in the lower section.

Heat integration is always a good way to reduce costs. However, we always raise the question of process start-up when a solution with significant heat integration is presented. The only significant heat integration possible is preheating the boiler feed water with the reactor effluent. Other heat transfer considerations include placement of the fired heater. Some solutions find it more economical to place it after the mixing point in order to save on steam in the other feed heat exchanger. Other solutions eliminate all other feed heat exchangers and only use a fired heater after the mixing point.

## Project 5

Most of the solution to this problem is in Section 22.4. For the information of the instructor, the observed "off-spec" conditions in the problem were obtained by running a simulation with 10 wt% propane impurity at the same total  $C_3$  feed mass flowrate, the conditions that would exist if the excess impurity were not observed.

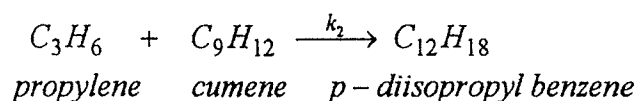
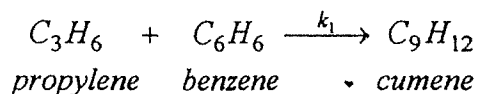
In terms of the troubleshooting aspects, most students can come up with one reason for the observations. However, most students are happy to stop once they have one possible solution. Our goal is to force students to consider as many alternatives as possible. The solution in Section 22.4 includes numerous alternatives, most of which cannot be distinguished without rigorous process simulation and/or more process information. Therefore, students must begin to learn to analyze problems with incomplete information.

Students are also asked to suggest temporary short-term modifications to return the process to normal production. A common suggestion is to increase the reaction temperature. However, as the analysis in Section 22.4 shows, significant temperature increases may not be possible. Increasing reaction pressure, which is a definite possibility, is not often suggested because students forget that increasing pressure for a gas-phase reaction also increases concentration and reaction rate. Given the limitations on increasing the temperature, some combination of increasing pressure and temperature may be the best alternative.

The problem with pump cavitation is another "caricature" problem. From the analysis in Section 22.2.2, it is seen that the propylene feed pump cannot be cavitating. The observed noise is most likely due to a worn bearing. However, the benzene feed pump could possibly cavitate if the recycle stream temperature were increased slightly by altering process conditions to compensate for other problems. This analysis is presented in Problem 22.11.

## Project 6

We begin by analyzing the reaction network:



As shown in Problem 14.4, low temperature and excess benzene improve the selectivity for cumene over p-diisopropyl benzene. The actual values for these parameters require further optimization.

At low temperatures, the reactor increases in size and cost. However, the fired heater capital and operating costs decrease; and, since a second distillation column is not needed, there are additional savings in the separation section. However, at the lowest possible temperature, the reactor size begins to dominate. Therefore, the optimum conditions appear to be a reactor feed temperature in the 330°C range, with a benzene/propylene mole ratio just below 2. The temperature of 330°C is obtained if the reactor is treated as isothermal (inlet  $T$  = outlet  $T$ , but with a temperature spike in reactor). An even better solution is to feed to the reactor as low as 300°C, to take a temperature rise in the reactor, and to use the reactor effluent to preheat the reactor feed, thereby reducing the load on the fired heater. An even more sophisticated solution is to use staged adiabatic packed beds with intercooling by the feed stream before entering the fired heater. This also reduces the load on the fired heater. At various times, all of the above alternatives have been developed by different students and student groups.

While the heat integration discussed above can save significant capital and operating costs, the issue of process start-up is a concern. We always ask students about this when solutions with significant heat integration are presented. Since a fired heater may be required for start-up, if one is installed, it would then might more economical to use it for part of the reactor pre-heating rather than let it sit idle and install an "integrated" heat exchanger for the reactor pre-heating. Analyzing the economics of this situation is beyond the scope of the economic analysis presented in Section 1.

One time when this problem was assigned, a 20°C maximum temperature rise in the reactor was a constraint. The result was operation at temperatures above 350°C to increase the temperature driving force with the 254°C boiling water and increased excess benzene to maintain a high enough selectivity for cumene to eliminate the need for a second distillation column.



One feature that students often do not consider is the benzene leaving in the fuel gas stream from the flash vessel. Optimization of the temperature and pressure in the flash vessel is an important component of the optimization. However, optimum conditions vary too much to suggest a range of solutions. In most all cases when cooling water is used in the heat exchanger prior to the flash vessel, over \$1 million of benzene is wasted as fuel gas. Addition of a second heat exchanger/flash with refrigerated water can recover on the order of 75% of this benzene. Use of refrigerated water is usually minimized due to its high cost. In this case, however, very little is needed and the benzene savings far outweighs the cost of the equipment and the refrigerated water.

**Base Case for 20,000 tonnes/yr Allyl Chloride Facility (scaled-up Beaumont design)**

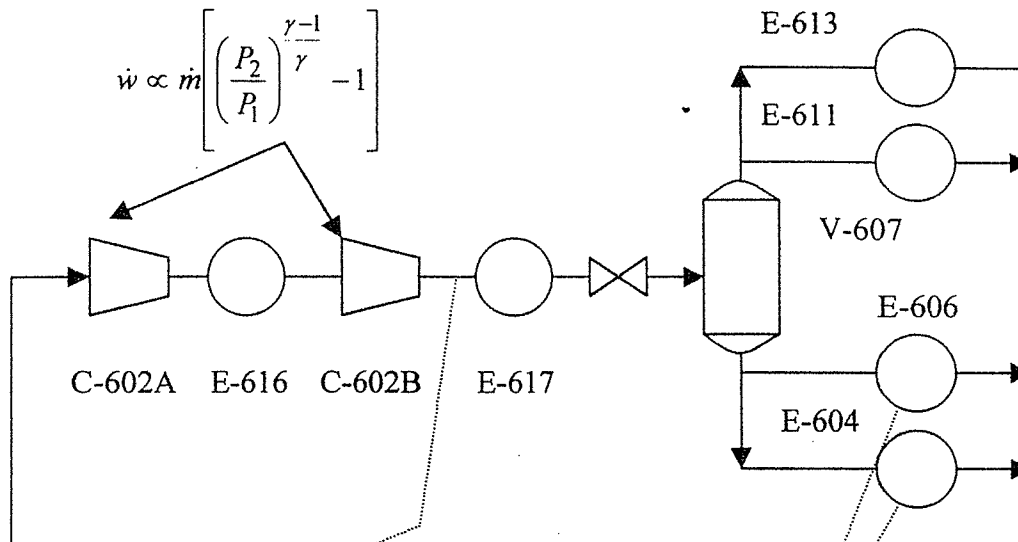
| <u>Year</u> | <u>FCI + WC</u> | <u>Revenue</u> | <u>COMd</u> | <u>d</u> | <u>(R-COMd-d)(1-t)+d</u> | <u>CF</u> | <u>CCF</u> | <u>DCF</u> | <u>CDCF</u> |
|-------------|-----------------|----------------|-------------|----------|--------------------------|-----------|------------|------------|-------------|
| 0           |                 |                |             |          |                          |           |            |            |             |
| 1           | -7.695          |                |             |          |                          | -7.70     | -7.70      | -7.00      | -7.00       |
| 2           | -10.773         |                |             |          |                          | -10.77    | -18.47     | -8.90      | -15.90      |
| 3           |                 | 37.98          | 21.83       | 3.08     | 14.19                    | 14.19     | -4.28      | 10.66      | -5.24       |
| 4           |                 | 37.98          | 21.83       | 4.92     | 14.47                    | 14.47     | 10.19      | 9.88       | 4.64        |
| 5           |                 | 37.98          | 21.83       | 2.95     | 14.17                    | 14.17     | 24.36      | 8.80       | 13.44       |
| 6           |                 | 37.98          | 21.83       | 1.77     | 13.99                    | 13.99     | 38.35      | 7.90       | 21.34       |
| 7           |                 | 37.98          | 21.83       | 1.77     | 13.99                    | 13.99     | 52.35      | 7.18       | 28.52       |
| 8           |                 | 37.98          | 21.83       | 0.89     | 13.86                    | 13.86     | 66.21      | 6.47       | 34.99       |
| 9           |                 | 37.98          | 21.83       |          | 13.73                    | 13.73     | 79.93      | 5.82       | 40.81       |
| 10          |                 | 37.98          | 21.83       |          | 13.73                    | 13.73     | 93.66      | 5.29       | 46.10       |
| 11          |                 | 37.98          | 21.83       |          | 13.73                    | 13.73     | 107.39     | 4.81       | 50.91       |
| 12          | 3.078           | 37.98          | 21.83       |          | 13.73                    | 16.81     | 124.19     | 5.35       | 56.27       |

**NPV = 56.27**

FCI = 15.39  
 taxation rate, t= 15%  
 utilities cost 3.42  
 raw materials cost = 9.43  
 waste treatment cost = 0.26  
 revenue = 37.98  
 discount rate, i = 10%  
 depreciation = MACRS (20,32,19.2, 11.52, 11.52, 5.76)

C.6-3

## Interesting Points about the Refrigeration Loop



Why is this set at 20 bar?

Because this is the lowest pressure at which we can condense propylene with cooling water.

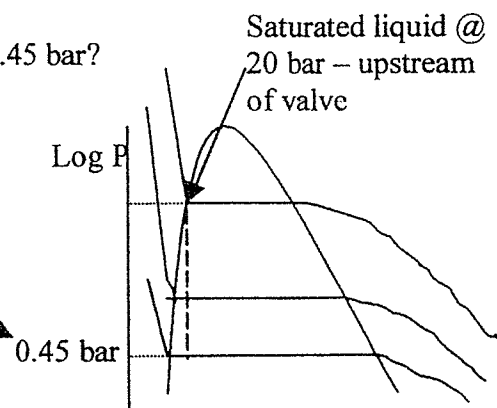
Why do we not vaporize all the propylene in throttling process?

Because these exchangers have large duties and are used to condense overhead products in distillation columns. Thus it is better to vaporize cold liquid propylene (at constant temperature) than to use cold propylene vapor.

Why do we use a 5°C temperature approach for E-604 and E-606? Because there is a trade-off between compression costs (to cool the propylene below current low temperature) and heat exchanger costs (heat transfer area). Since the electric utility costs for compression are so high, the use of a 5°C temperature approach (instead of the normal 10°C minimum value) is justified.

Why is the pressure at the inlet to C-602A set at 0.45 bar?

Setting this pressure sets the temperature



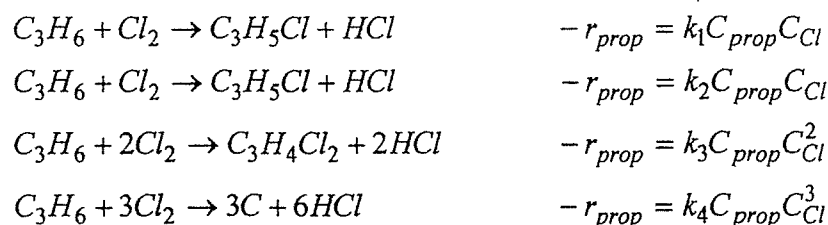
C.6-4

We can optimize the refrigeration loop a little bit but we can't reduce the power requirement by much. What we need to do is to increase the temperature in the process in order to reduce the load on the refrigeration loop. If we look at the PFD for the process we see that we need a low temperature in the distillation column condensers (namely in T-601 and T-603) and a low inlet temperature to T-601. The reason for this is so that we can form an L-V mixture. If we refer back to Chapter 8 and look under the tables of special concerns for separators, we see that either low temperature or high pressure may be required to obtain an L-V mixture suitable for distillation. Therefore, pressurizing the columns should have a similar effect as using a low temperature in the condensers. We can pressurize the columns in several ways:

- (1) The conservative method is to compress Stream 5.
- (2) Alternatively we may pressurize the whole front end of the process by pumping the feed streams (propylene and chlorine) to a high pressure.

Alternative (1) is conservative because we do not change the pressure of the reaction section of the process and hence the selectivity will not change.

Alternative (2) is potentially cheaper in utility costs since we have eliminated the compressor but the front-end equipment, e.g., R-601, H-601, E-602 and E-603, must be designed to withstand higher pressures and will be more costly. Additionally the pressure will change considerably and the selectivity may change. Although no reaction kinetics are given consider the effect of increasing pressure on the competing reactions assuming elementary kinetic expressions.



$$\text{Selectivity} = \frac{k_1 C_{prop} C_{Cl}}{k_1 C_{prop} C_{Cl} + k_2 C_{prop} C_{Cl} + k_3 C_{prop} C_{Cl}^2 + k_4 C_{prop} C_{Cl}^3}$$

The effect of increasing the pressure can be seen in the effect that the pressure has on the gas phase concentration. Assuming ideal gas behavior we have  $C \propto P/RT$ , therefore as pressure increases the concentration increases proportionally. Looking at the expression for selectivity we can see that the denominator increases more with increasing pressure than does the numerator. Thus the selectivity is expected to decrease with increasing pressure. Thus Alternative (1) may be better.

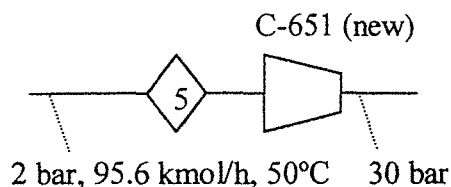
#### Why compress Stream 5 and by how much?

It can be argued that little is to be gained by pressurizing Stream 5 since we are replacing one set of compressors (refrigeration compressors C-602A&B) with another compressor(s) in Stream 5.

Also if we do add compressors to what pressure should we compress Stream 5? This is the subject of a parametric optimization (see Chapter 19) but let us say that for comparison we will choose to compress Stream 5 to a pressure which will allow the refrigeration loop to be eliminated, i.e., so that we can use cooling water in the condensers. This pressure is approximately 30 bar.

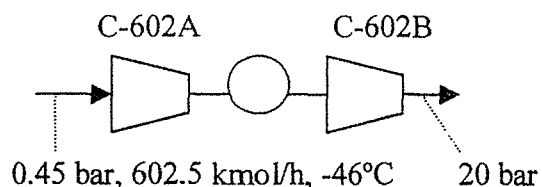
Look at two compressor schemes:

#### New Compressor in Stream 5



$$P_2/P_1 = 15$$

#### Refrigeration Loop



$$P_2/P_1 = 44.4$$

If we compare the theoretical power for C-651 to that of C-602 we get approximately the following relationship (assuming a single stage of compression):

$$\frac{w_2}{w_1} = \frac{m_2 T_2 [(P_2/P_1)_2^a - 1]}{m_1 T_1 [(P_2/P_1)_1^a - 1]} = \frac{(95.6)(323) [(15)^{0.286} - 1]}{(602.5)(227) [(44.4)^{0.286} - 1]} = 0.135$$

Clearly the option of placing a compressor(s) in Stream 5 to replace the refrigeration compressors is a good idea.

#### What are the Consequences of Increasing the Pressure in Stream 5 to 30 Bar on the Separation ?

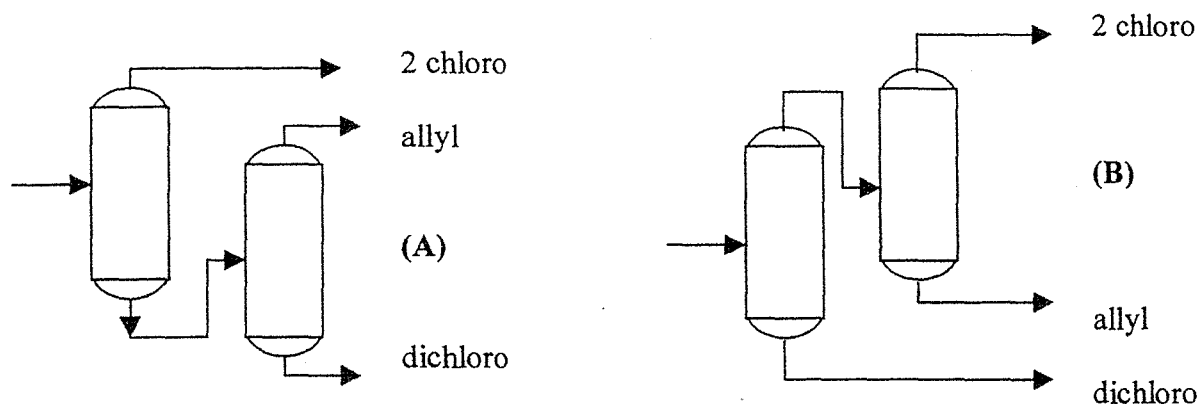
Obviously we have chosen 30 bar pressure so that we can eliminate the costly refrigeration loop and use cooling water in the overhead condensers. However, as the top temperature in the columns increase so do the bottoms temperatures. For the first column, T-601, operating at 30 bar the bottom temperature is approximately 200°C compared to 45°C previously. The question now arises as to whether allyl chloride is stable at this temperature. The following is a quote from the Kirk-Othmer encyclopedia, Vol 2, p106, "Allyl compounds are normally stable at room temperature even in the absence of inhibitors. However, prolonged exposure to air, especially in the presence of light or heat, can cause peroxide formation. Contamination with either peroxides, or traces of catalytically active transition metals, can present hazards on heating as a result of rapid and strong exothermic polymerization." The reason that the Beaumont facility operated at very low temperatures was to avoid the problems associated with this uncontrolled polymerization!!

The lesson here is that there are large savings to be made in the operating costs of the plant by increasing the pressure of the separations section. However, whenever we change the process conditions by a large amount we must be sure that we are still in a safe operating region and that we take into account all the potential effects that these changes may have on the process.

### Other Issues

Some other issues which should be considered in the optimized design, but which have less impact on the overall economics are listed below:

- (1) Changing the sequence of the columns. Allyl chloride has the largest flowrate but is intermediate in volatility (2 chloro –allyl –dichloro). We could change the sequencing as follows:



There appears to be little advantage of (B) over (A) so the current sequence is probably OK.

- (2) The first two columns, T-601 and T-603, appear to be doing the same separation. This is wasteful since we only require a sharp split between the propylene/HCl and the organic chlorides. We should just make the first column a little taller (more stages) and improve the separation thus eliminating the need for T-603 (this is a topological optimization).
- (3) Since we take the overhead product from T-601 as a liquid then immediately vaporize it in E-607 it would seem a good idea to use a partial condenser for T-601 and save on utilities.
- (4) Why do we recycle propylene as a liquid? This is a nice option to have when shutting down the process since we can liquefy the recycle and sent it to storage. However, during regular operations we just vaporize the recycle. We may want to eliminate C-601A and B or not operate these compressors during normal operation. Note that if we operate the separations section at above 20bar (T-601 and T-602) we can recycle and liquefy the propylene without the use of compressors!
- (5) In the front end of the process it is worth looking at heat integration
  - (i) Make high pressure steam in Dowtherm A cooler.
  - (ii) Drop the temperature into the fluidized bed. Since the reactor is essentially well mixed, (at least the solids are) we can feed the reactants into the reactor at temperatures