Chemical Process Diagrams
Chapter 1

Department of Chemical Engineering
West Virginia University

3 Levels of Diagram

- Block Flow Diagram (BFD)
- Process Flow Diagram (PFD)
- Piping and Instrumentation Diagram (P&ID) – often referred to as Mechanical Flow Diagram

↓ Complexity increases

↑ Conceptual understanding Increases

As chemical engineers, we are most familiar with BFD and PFD.

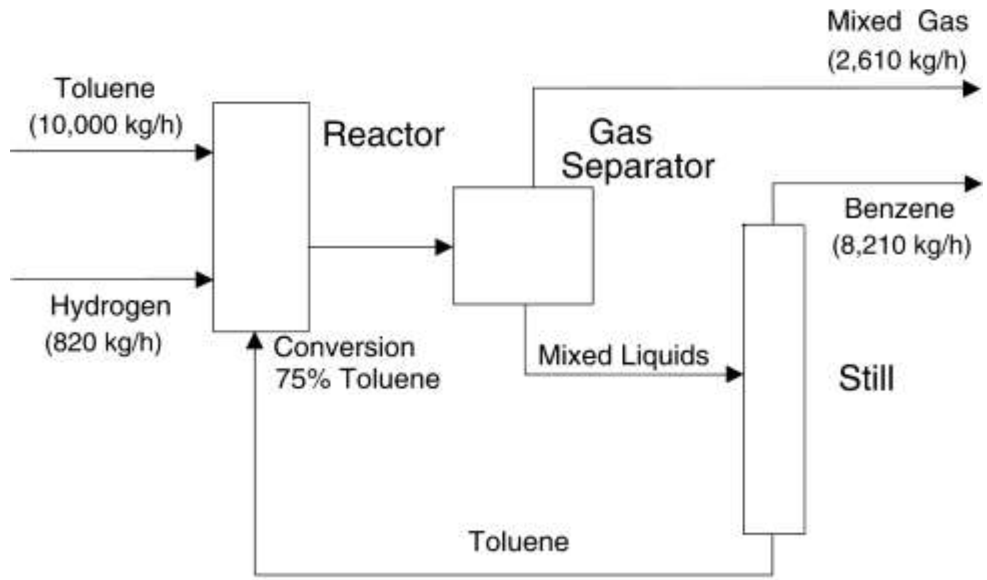
The Block Flow Diagram (BFD)

- BFD shows overall processing picture of a chemical complex
  - Flow of raw materials and products may be included on a BFD
  - BFD is a superficial view of facility – ChE information is missing

Definitions of BFD

- Block Flow Process Diagram
  - Figure 1.1
  - Similar to sketches in material and energy balances
- Block Flow Plant Diagram
  - Figure 1.2
  - Gives a general view of a large complex plant

The Block Flow Process Diagram



Reaction : $C_7H_8 + H_2 = C_6H_6 + CH_4$

Figure 1.1: Block Flow Process Diagram for the Production of Benzene

The Block Flow Plant Diagram

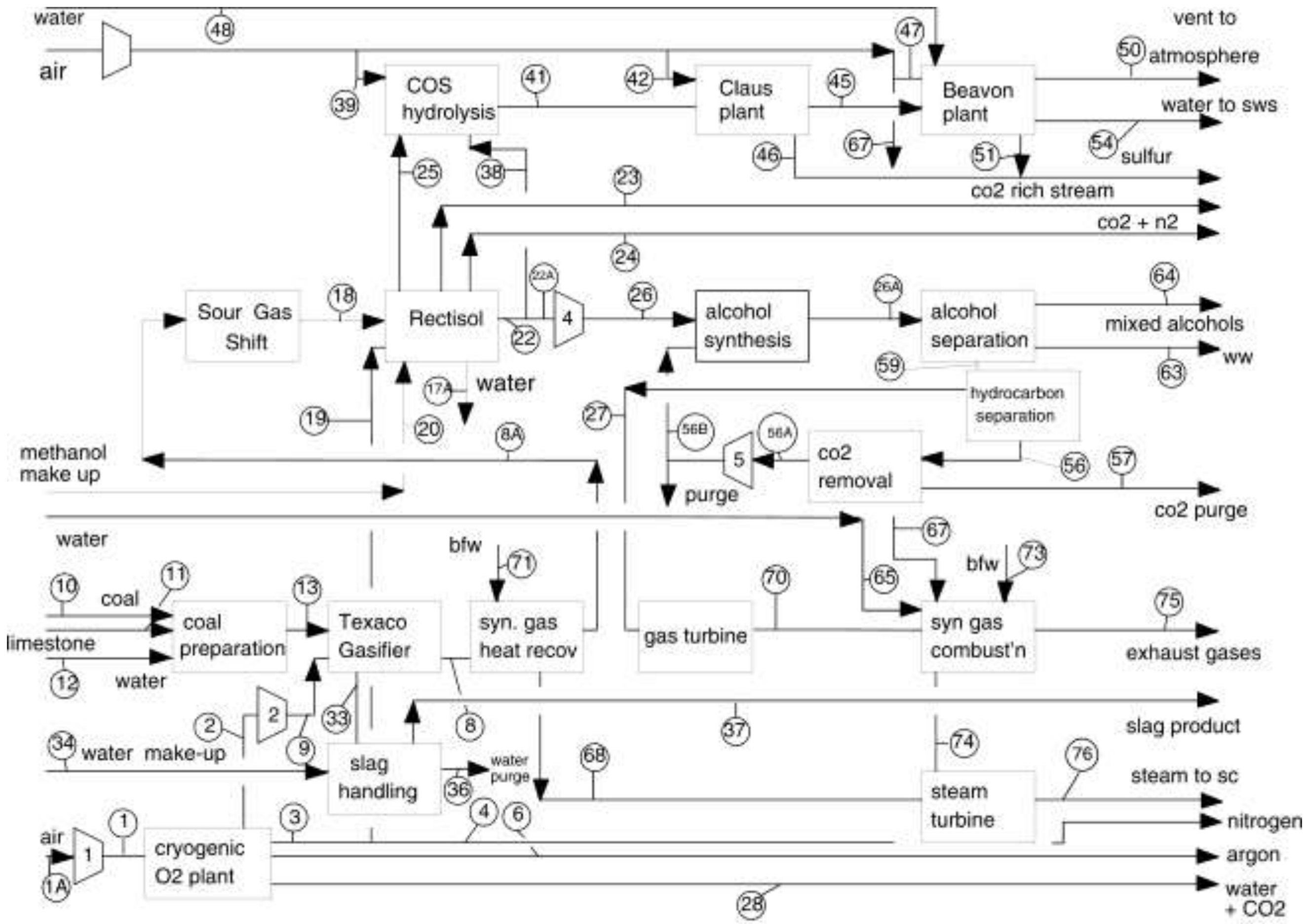


Figure 1.2: Block Flow Plant Diagram of a Coal to Higher Alcohol Fuels Process

The Process Flow Diagram

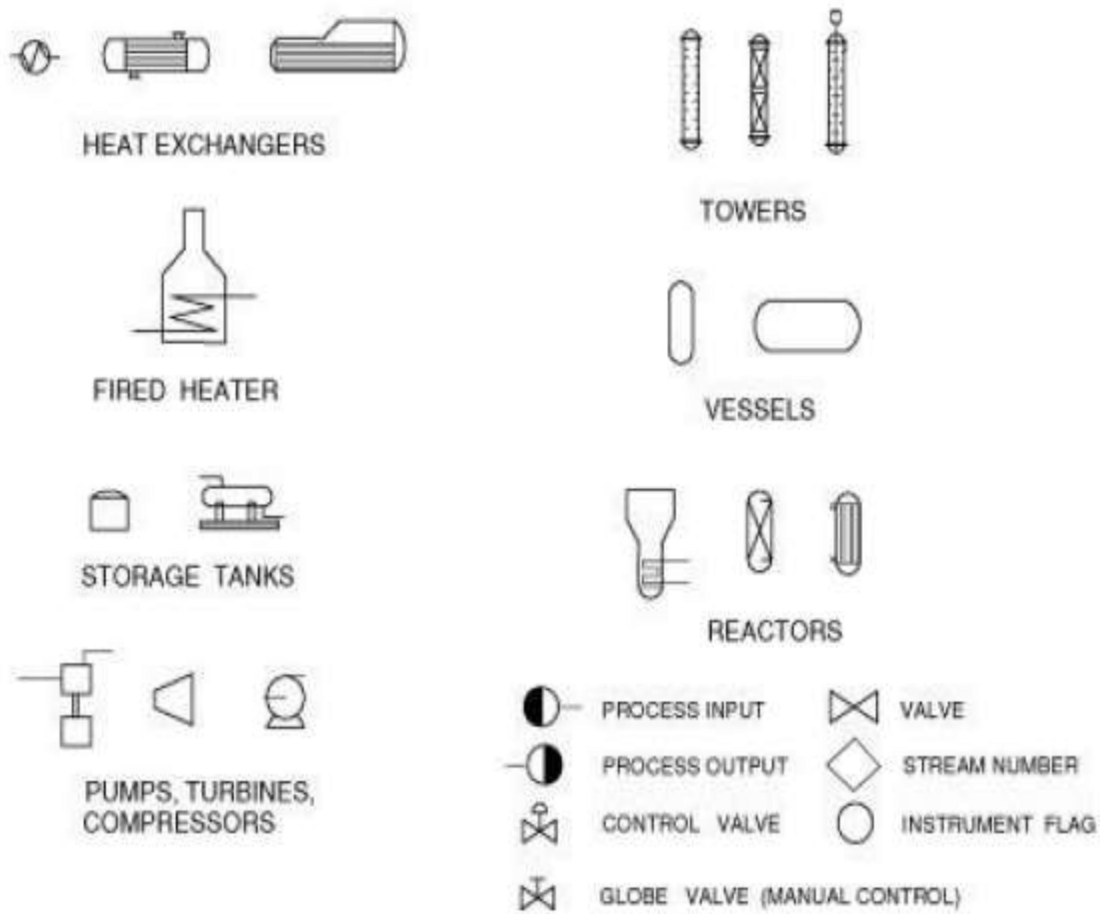
- PFD shows all process engineering information
  - Diagram developed in junior year design projects (especially the 2nd semester)
  - Often PFD is drawn on large paper – textbook breaks down information into 1 diagram and 2 tables

The Process Flow Diagram (cont'd)

- The topology of the process – showing the connectivity of all the streams and the equipment
  - Example for toluene HDA – Figures 1.3 and 1.5
  - Tables 1.2 and 1.4 – list information that should be on the PFD but cannot fit
  - Use appropriate conventions – consistency is important in communication of process information

    ex. Table 1.2

Figure 1.4 Symbols for Drawing Process Flow Diagrams



The Process Flow Diagram (cont'd)

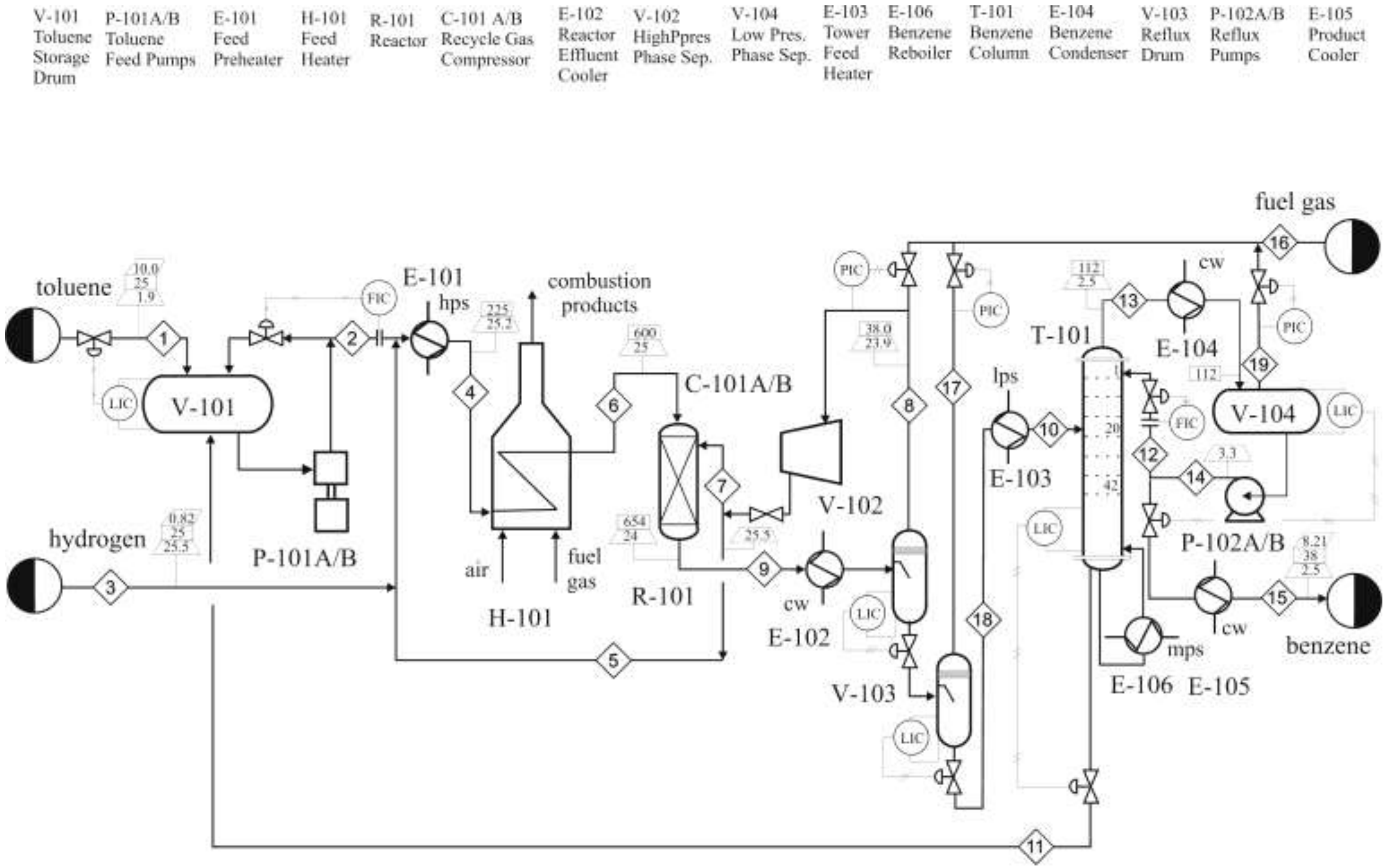


Figure 1.5: Process flow diagram (PFD) for the production of benzene via the hydrodealkylation of toluene

The Process Flow Diagram (cont'd)

Table 1.2 : Conventions Used for Identifying Process Equipment

Process Equipment	General Format **XX-YZZ A/B**
	XX are the identification letters for the equipment classification
	C - Compressor or Turbine
	E - Heat Exchanger
	H - Fired Heater
	P - Pump
	R - Reactor
	T - Tower
	TK - Storage Tank
	V - Vessel
	Y designates an area within the plant
	ZZ are the number designation for each item in an equipment class
	A/B identifies parallel units or backup units not shown on a PFD
Supplemental Information	Additional description of equipment given on top of PFD

Equipment Numbering

- XX-YZZ A/B/…
  - XX represents a 1- or 2-letter designation for the equipment (P = pump)
  - Y is the 1 or 2 digit unit number (1-99)
  - ZZ designates the equipment number for the unit (1-99)
  - A/B/… represents the presence of spare equipment

V-101	P-101A/B	E-101	H-101	R-101	C-101 A/B	E-102	V-102	V-104	E-103	E-106	T-101	E-104	V-103	P-102A/B	E-105
Toluene Storage Drum	Toluene Feed Pumps	Feed Preheater	Feed Heater	Reactor	Recycle Gas Compressor	Reactor Effluent Cooler	HighPpres Phase Sep.	Low Pres. Phase Sep.	Tower Feed Heater	Benzene Reboiler	Benzene Column	Benzene Condenser	Reflux Drum	Reflux Pumps	Product Cooler

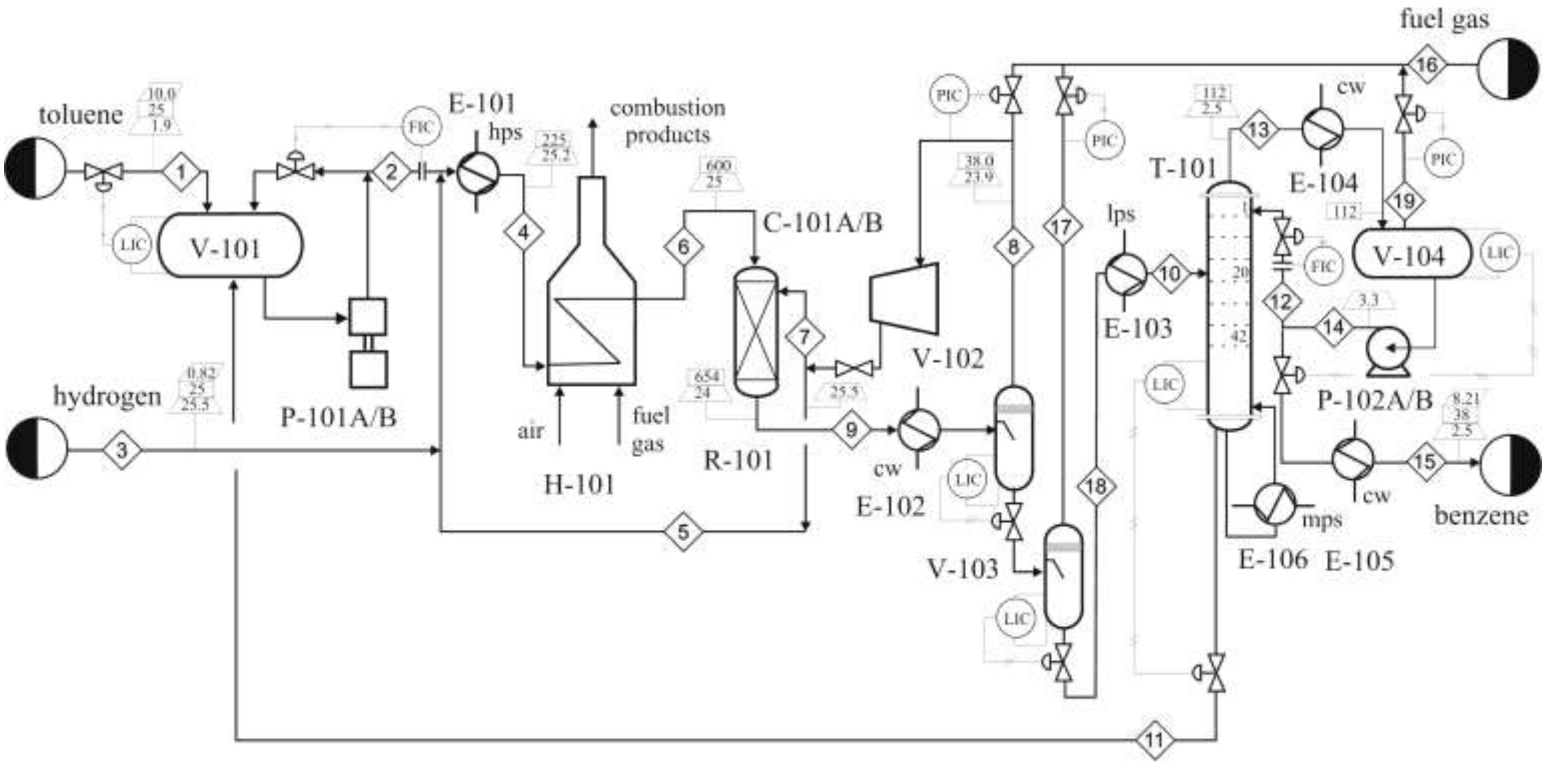


Figure 1.5: Process flow diagram (PFD) for the production of benzene via the hydrodealkylation of toluene

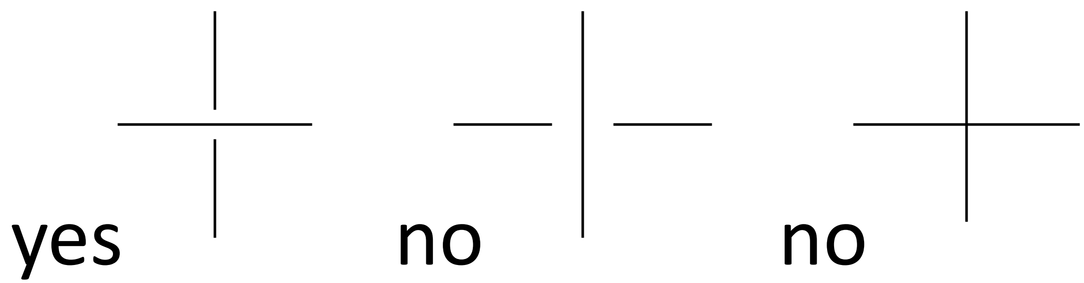
Equipment Numbering (cont'd)

Thus, T-905 is the 5th tower in unit nine hundred P-301 A/B is the 1st Pump in unit three hundred plus a spare

- Use unambiguous letters for new equipment
  - Ex. Turbine use Tb or J not T (for tower)
  - Replace old vessel V-302 with a new one of different design - use V-319 (say) not V-302 – since it may be confused with original V-302

Stream Numbering and Drawing

- Number streams from left to right as much as possible
- Horizontal lines are dominant



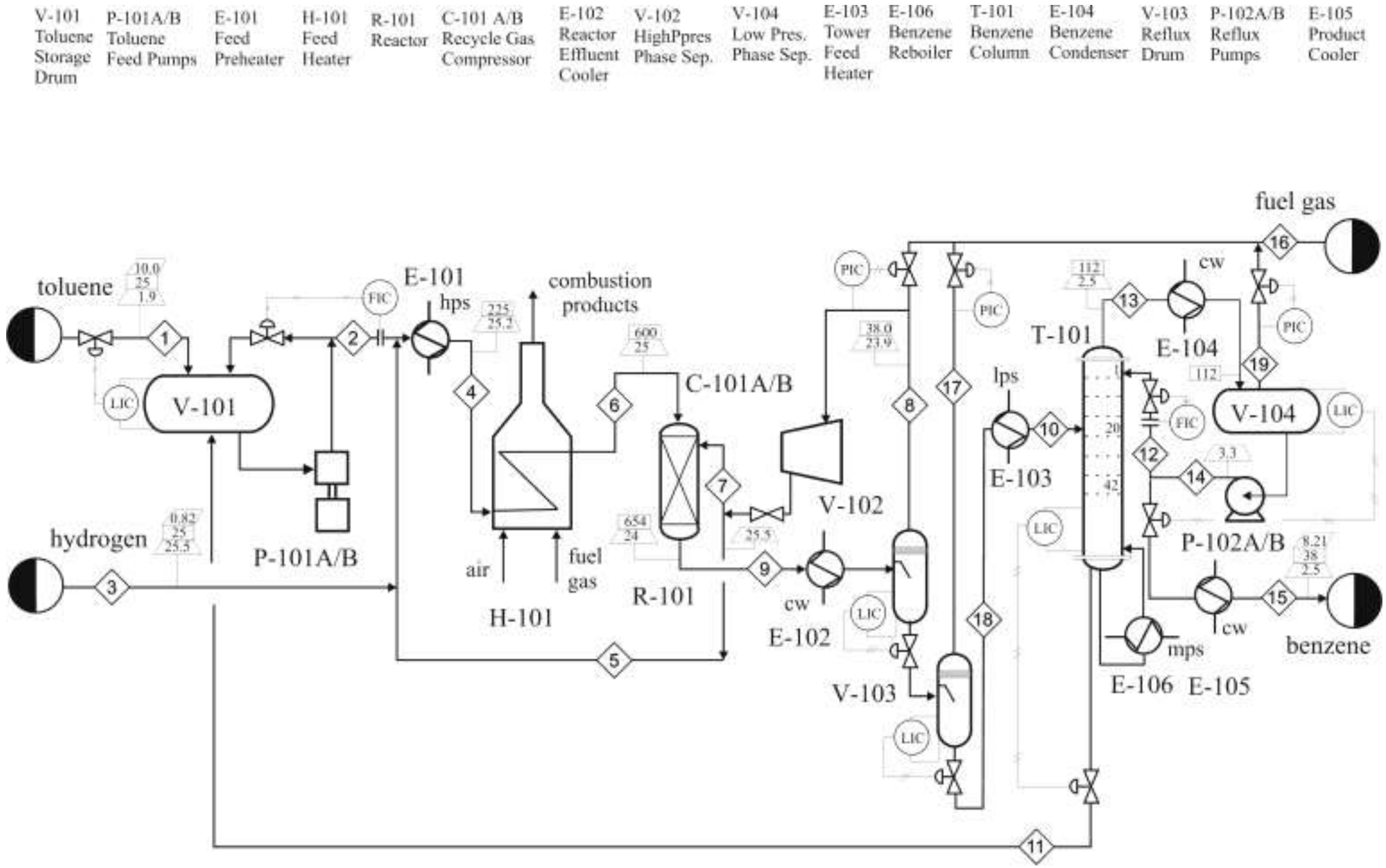


Figure 1.5: Process flow diagram (PFD) for the production of benzene via the hydrodealkylation of toluene

Stream Numbering and Drawing (cont'd)

- Add arrows for
  - Change in direction
  - Inlet of equipment
- Utility streams should use convention given in Table 1.3, lps, cw, fg, etc.

Stream Information

- Since diagrams are small, not much stream information can be included
- Include important data – around reactors and towers, etc.
  - Flags are used – see toluene HDA diagram
  - Full stream data, as indicated in Table 1.4, are included in a separate flow summary table – see Table 1.5

V-101	P-101A/B	E-101	H-101	R-101	C-101 A/B	E-102	V-102	V-104	E-103	E-106	T-101	E-104	V-103	P-102A/B	E-105
Toluene Storage Drum	Toluene Feed Pumps	Feed Preheater	Feed Heater	Reactor	Recycle Gas Compressor	Reactor Effluent Cooler	HighPpres Phase Sep.	Low Pres. Phase Sep.	Tower Feed Heater	Benzene Reboiler	Benzene Column	Benzene Condenser	Reflux Drum	Reflux Pumps	Product Cooler

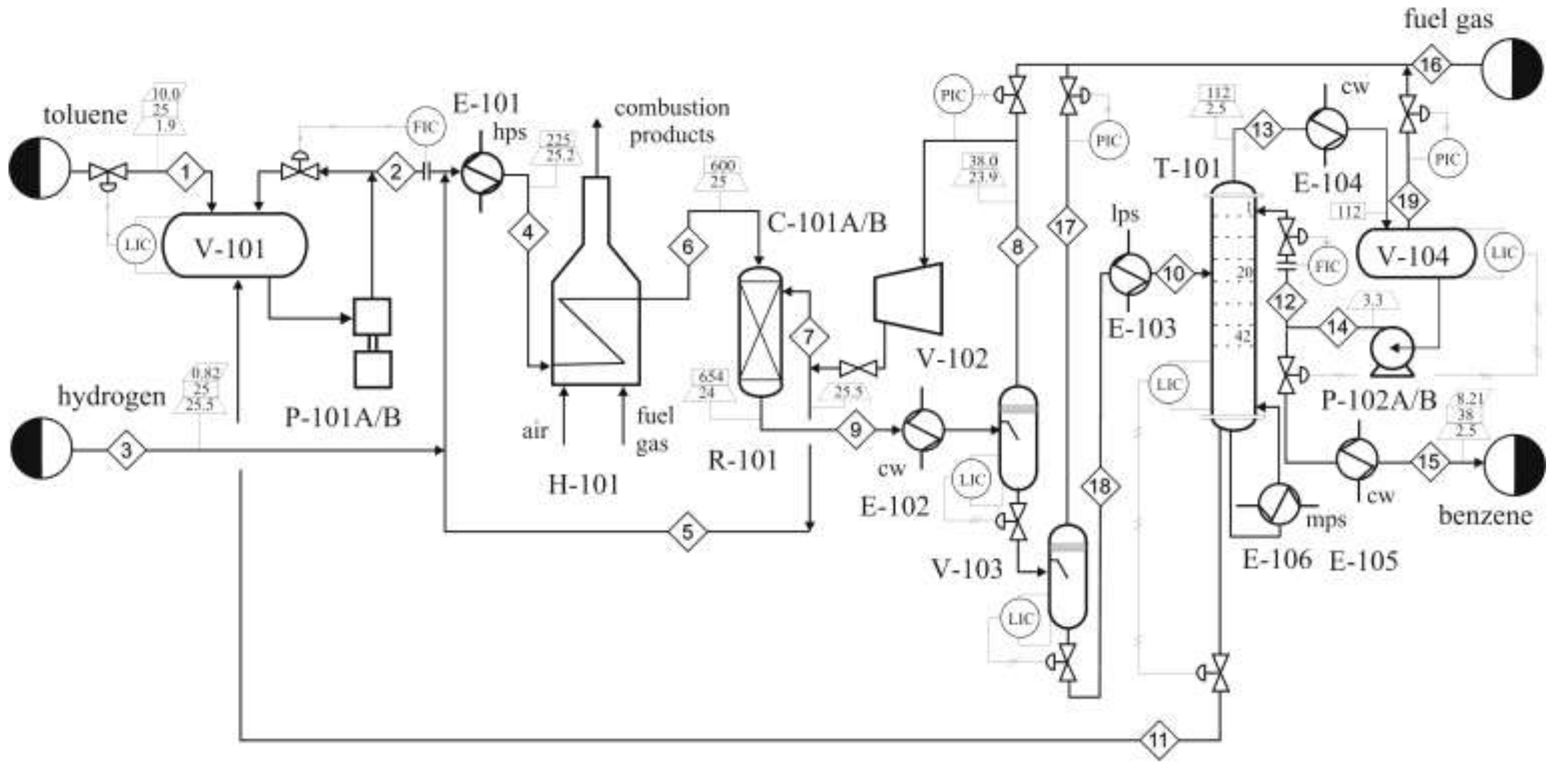
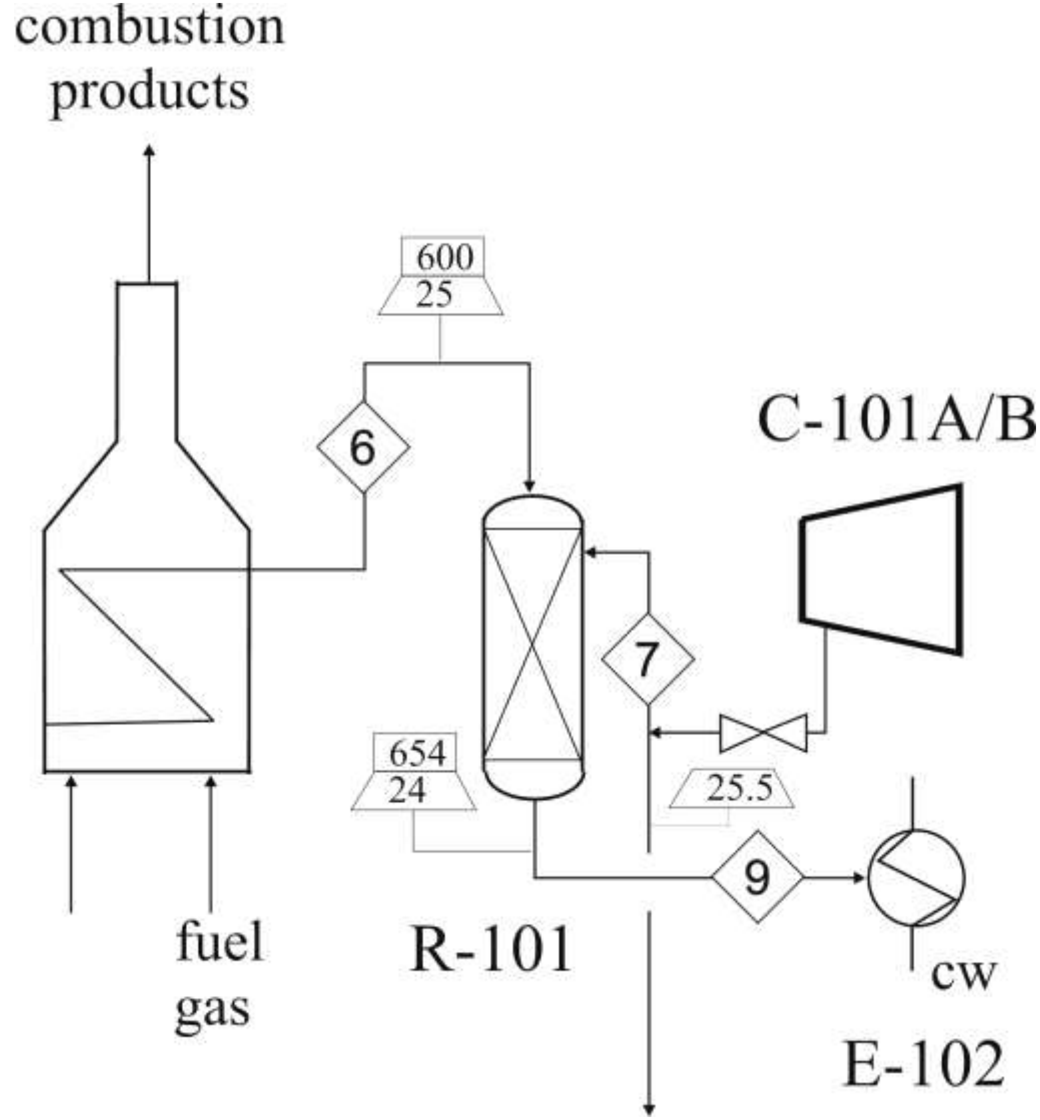


Figure 1.5: Process flow diagram (PFD) for the production of benzene via the hydrodealkylation of toluene

Stream Information - Flags



The Process Flow Diagram (cont'd)

Table 1.4: Information in a Flow Summary

Essential Information
Stream Number
Temperature (°C)
Pressure (bar)
Vapor Fraction
Total Mass Flow Rate (kg/h)
Total Mole Flow Rate (kmol/h)
Individual Component Flow Rates (kmol/h)
Optional Information
Component Mole Fractions
Component Mass Fractions
Individual Component Flow Rates (kg/h)
Volumetric Flow Rates (m^3/h)
Significant Physical Properties
Density
Viscosity
Other
Thermodynamic Data
Heat Capacity
Stream Enthalpy
K-values
Stream Name

The Process Flow Diagram (cont'd)

A Portion of Table 1.5

Stream Number	1	2	3	4	5	6	7	8	9	10
Temperature (°C)	25	59	25	225	41	600	41	38	654	90
Pressure (bar)	1.90	25.8	25.5	25.2	25.5	25.0	25.5	23.9	24.0	2.6
Vapor Fraction	0.0	0.0	1.00	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Mass Flow (tonne/h)	10.0	13.3	0.82	20.5	6.41	20.5	0.36	9.2	20.9	11.6
Mole Flow (kmol/h)	108.7	144.2	301.0	1204.4	758.8	1204.4	42.6	1100.8	1247.0	142.2
Component Mole Flow (kmol/h)										
Hydrogen	0.0	0.0	286.0	735.4	449.4	735.4	25.2	651.9	652.6	0.02
Methane	0.0	0.0	15.0	317.3	302.2	317.3	16.95	438.3	442.3	0.88
Benzene	0.0	1.0	0.0	7.6	6.6	7.6	0.37	9.55	116.0	106.3
Toluene	108.7	143.2	0.0	144.0	0.7	144.0	0.04	1.05	36.0	35.0

Process Streams	
All conventions shown in Table 1.1 apply.	
Diamond symbol located in flow lines.	
Numerical identification (unique for that stream) inserted in diamond.	
Flow direction shown by arrows on flow lines.	
Utility Streams	
lps	Low-Pressure Steam: 3–5 barg (sat) ‡
mps	Medium-Pressure Steam: 10–15 barg (sat) ‡
hps	High-Pressure Steam: 40–50 barg (sat) ‡
htm	Heat Transfer Media (Organic): to 400°C
cw	Cooling Water: From Cooling Tower 30°C Returned at Less Than 45°C†
wr	River Water: From River 25°C Returned at Less Than 35°C
rw	Refrigerated Water: In at 5°C Returned at Less Than 15°C
rb	Refrigerated Brine: In at −45°C Returned at Less Than 0°C
cs	Chemical Waste Water with High COD
ss	Sanitary Waste Water with High BOD, etc.
el	Electric Heat (Specify 220, 440, 660V Service)
ng	Natural Gas
fg	Fuel Gas
fo	Fuel Oil
fw	Fire Water

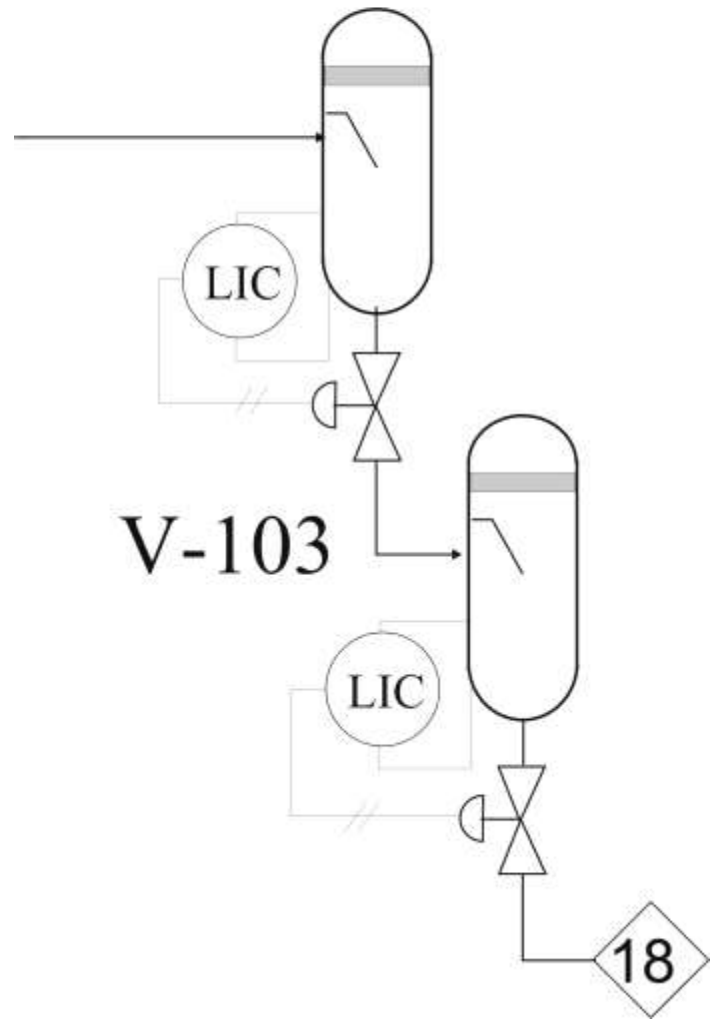
‡These pressures are set during the preliminary design stages and typical values vary within the ranges shown.

†Above 45°C, significant scaling occurs.

Basic Control Loops

- Often the basic control loops (those involving maintaining material balance and reactor controls) are included on the PFD; instrumentation and other control loops are not shown

Basic Control Loops



Equipment Information

- Equipment are identified by number and a label (name) positioned above the equipment on the PFD
- Basic data such as size and key data are included in a separate table (Equipment Summary Table) Table 1.7 (and Table 1.6) in TBWS

Equipment Information

A Section of Table 1.7: Equipment Summary

Vessel	V-101	V-102
Temperature (ºC)	55	38
Pressure (bar)	2.0	24
Orientation	Horizontal	Vertical
MOC	CS	CS
Size		
Height/Length (m)	5.9	3.5
Diameter (m)	1.9	1.1
Internals		s.p. (splash plate)

Table 1.6 Equipment Descriptions for PFD and PIDs

Equipment Type
Description of Equipment
Towers
Size (height and diameter), Pressure, Temperature Number and Type of Trays Height and Type of Packing Materials of Construction
Heat Exchangers
Type: Gas-Gas, Gas-Liquid, Liquid-Liquid, Condenser, Vaporizer

Process: Duty, Area, Temperature, and Pressure for both streams

Number of Shell and Tube Passes

Materials of Construction: Tubes and Shell

Tanks and Vessels

Height, Diameter, Orientation, Pressure, Temperature, Materials of Construction

Pumps

Flow, Discharge Pressure, Temperature, ΔP, Driver Type, Shaft Power, Materials of Construction

Compressors

Actual Inlet Flowrate, Temperature, Pressure, Driver Type, Shaft Power, Materials of Construction

Heaters (Fired)

Type, Tube Pressure, Tube Temperature, Duty, Fuel, Material of Construction

Other

Provide Critical Information

Heat Exchangers	E-101	E-102	E-103	E-104	E-105	E-106
Type	Fl.H.	Fl.H.	MDP	Fl.H.	MDP	Fl.H.
Area (m^2)	36	763	11	35	12	80
Duty (MJ/h)	15,190	46,660	1055	8335	1085	9045
Shell						
Temp. (°C)	225	654	160	112	112	185
Pres. (bar)	26	24	6	3	3	11
Phase	Vap.	Par. Cond.	Cond.	Cond.	l	Cond.
MOC	316SS	316SS	CS	CS	CS	CS
Tube						
Temp. (°C)	258	40	90	40	40	147
Pres. (bar)	42	3	3	3	3	3
Phase	Cond.	l	l	l	l	Vap.
MOC	316SS	316SS	CS	CS	CS	CS
Vessels/Tower/ Reactors	**V-101**	**V-102**	**V-103**	**V-104**	**T-101**	**R-101**
Temperature (°C)	55	38	38	112	147	660
Pressure (bar)	2.0	24	3.0	2.5	3.0	25
Orientation	Horizn'l	Vertical	Vertical	Horizn'l	Vertical	Vertical
MOC	CS	CS	CS	CS	CS	316SS
Size						
Height/Length (m)	5.9	3.5	3.5	3.9	29	14.2
Diameter (m)	1.9	1.1	1.1	1.3	1.5	2.3
Internals		s.p.	s.p.		42 sieve trays 316SS	catalyst packed bed-10m
Pumps/Compressors	**P-101 (A/B)**	**P-102 (A/B)**	**C-101 (A/B)**	**Heater**		**H-101**
Flow (kg/h)	13,000	22,700	6770	Type		Fired
Fluid Density (kg/m^3)	870	880	8.02	MOC		316SS
Power (shaft) (kW)	14.2	3.2	49.1	Duty (MJ/h)		27,040

Pumps/Compressors	P-101 (A/B)	P-102 (A/B)	C-101 (A/B)	Heater	H-101
Type/Drive	Recip./ Electric	Centrf./ Electric	Centrf./ Electric	Radiant Area (m^2)	106.8
Efficiency (Fluid Power/Shaft Power)	0.75	0.50	0.75	Convective Area (m^2)	320.2
MOC	CS	CS	CS	Tube P (bar)	26.0
Temp. (in) (°C)	55	112	38		
Pres. (in) (bar)	1.2	2.2	23.9		
Pres. (out) (bar)	27.0	4.4	25.5		

Key:

MOC	Materials of construction	Par	Partial
316SS	Stainless steel type 316	F.H.	Fixed head
CS	Carbon steel	Fl.H.	Floating head
Vap	Stream being vaporized	Rbl	Reboiler
Cond	Stream being condensed	s.p.	Splash plate
Recipr.	Reciprocating	l	Liquid
Centrf.	Centrifugal	MDP	Multiple double pipe

PFD Summary

- PFD, Equipment Summary Table, and Flow Summary Table represent a “true” PFD
- This information is sufficient for a preliminary estimation of capital investment (Chapter 7) and cost of manufacture (Chapter 8) to be made

The Piping and Instrument Diagram(P&ID)

P&ID – Construction Bible

- Contains: plant construction information (piping, process, instrumentation, and other diagrams)
- P&ID information is explained in Tables 1.8 and 1.9
- Conventions for instrumentation are shown in Figure 1.10

P&ID

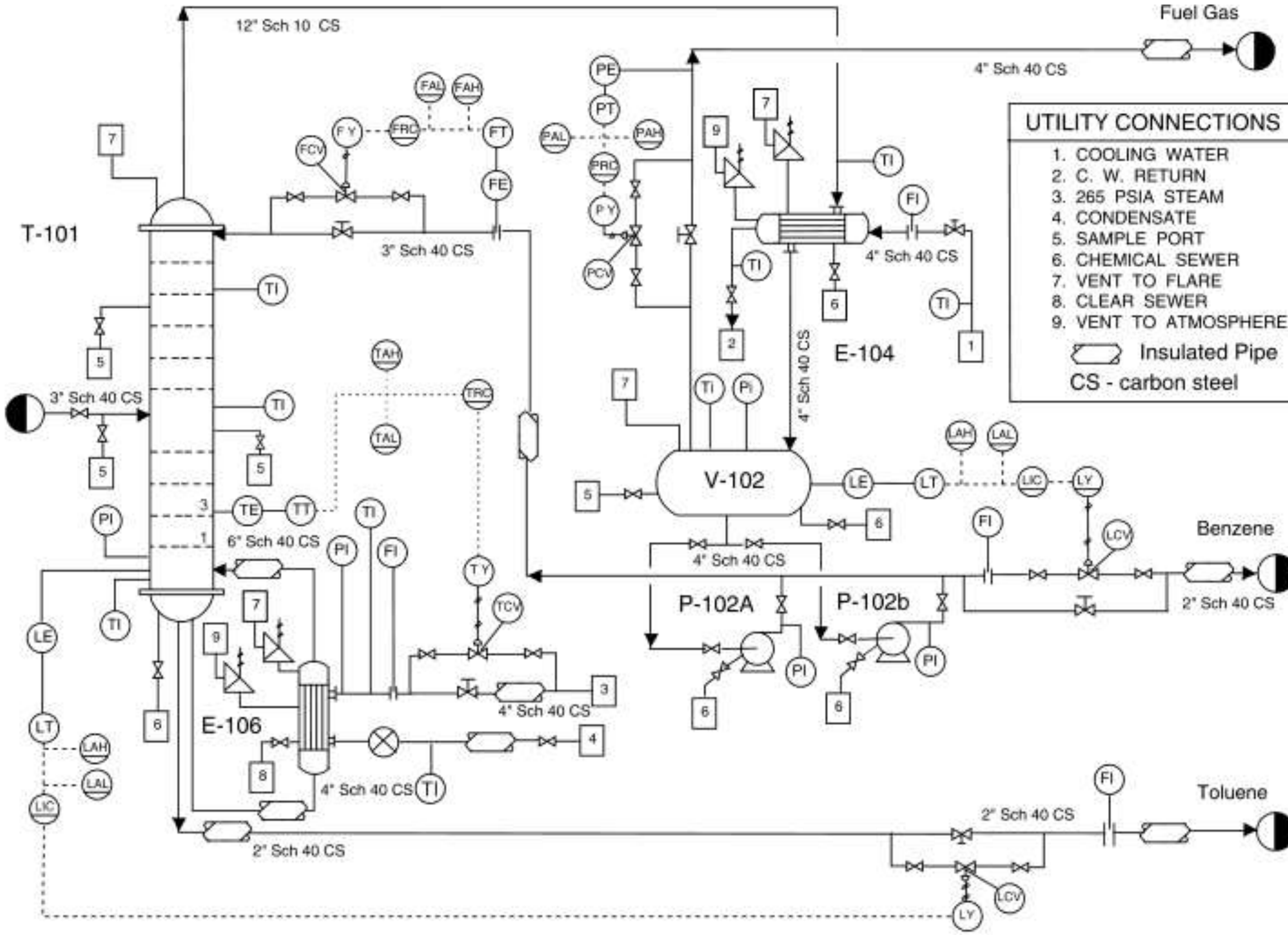


Figure 1.7 : Piping and Instrumentation Diagram for Benzene Distillation (adapted from Kauffman, D, Flow Sheets and Diagrams," AIChE Modular Instruction, Series G: Design of Equipment, series editor J. Beckman, AIChE, New York, 1986, vol 1, Chapter G.1.5, AIChE copyright C 1986 AIChE, all rights reserved)

Look at V-102 on P&ID

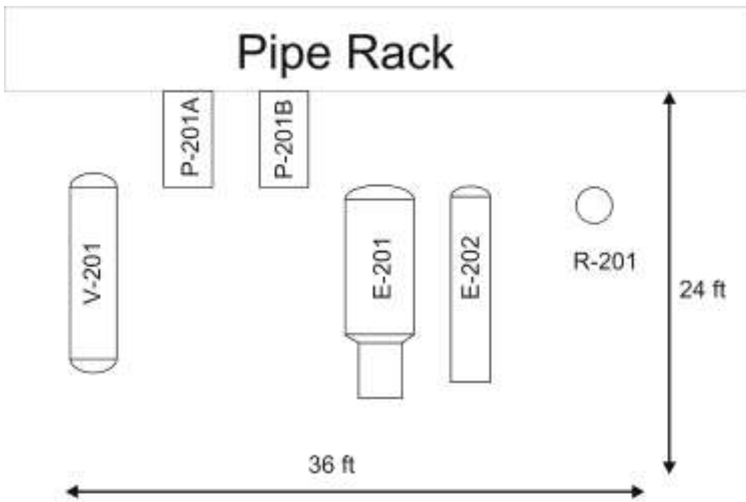
V-102 contains an LE (Level Element)

- LE senses liquid level in separator and adjusts flow rate leaving
- LE opens and closes a valve depending on liquid level
- LE and valve represent a feedback control loop

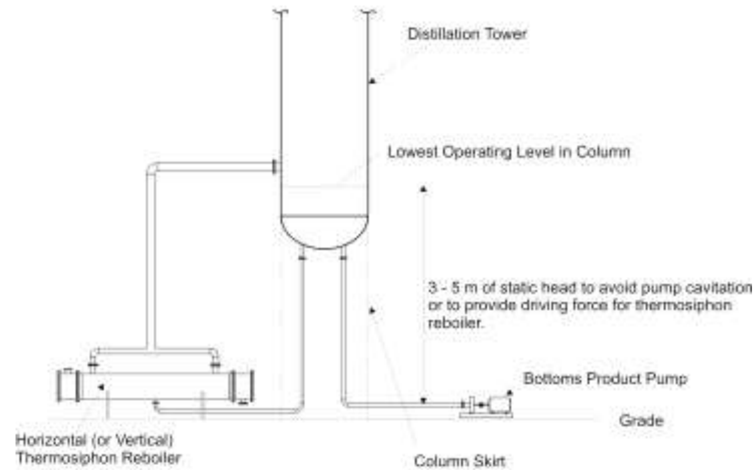
Other Common Diagrams

- Plot Plans – plan or map drawn looking down on plant (drawn to scale with all major equipment identified)
- Elevation Diagrams – show view from side and give information about equipments distance from ground

Other Common Diagrams



Section of Plot Plan



Section of Elevation Diagram

Other Common Diagrams (cont'd)

- Piping Isometrics – show piping in 3-dimensions
- Vessel Sketches – show key dimensions of equipment and locations of inlet and outlet nozzles etc.

Scale Models and Virtual Plants

- 25 years ago physical models were used for review
- Now virtual or electronic models are generated using software (3-d plant diagrams)
- Purpose of Models – catch errors such as
  - Piping clashes
  - Misaligned piping
  - Equipment not easily accessed
  - Sample points not easily reached by operators

OPERATOR AND 3-D IMMERSIVE TRAINING SIMULATORS

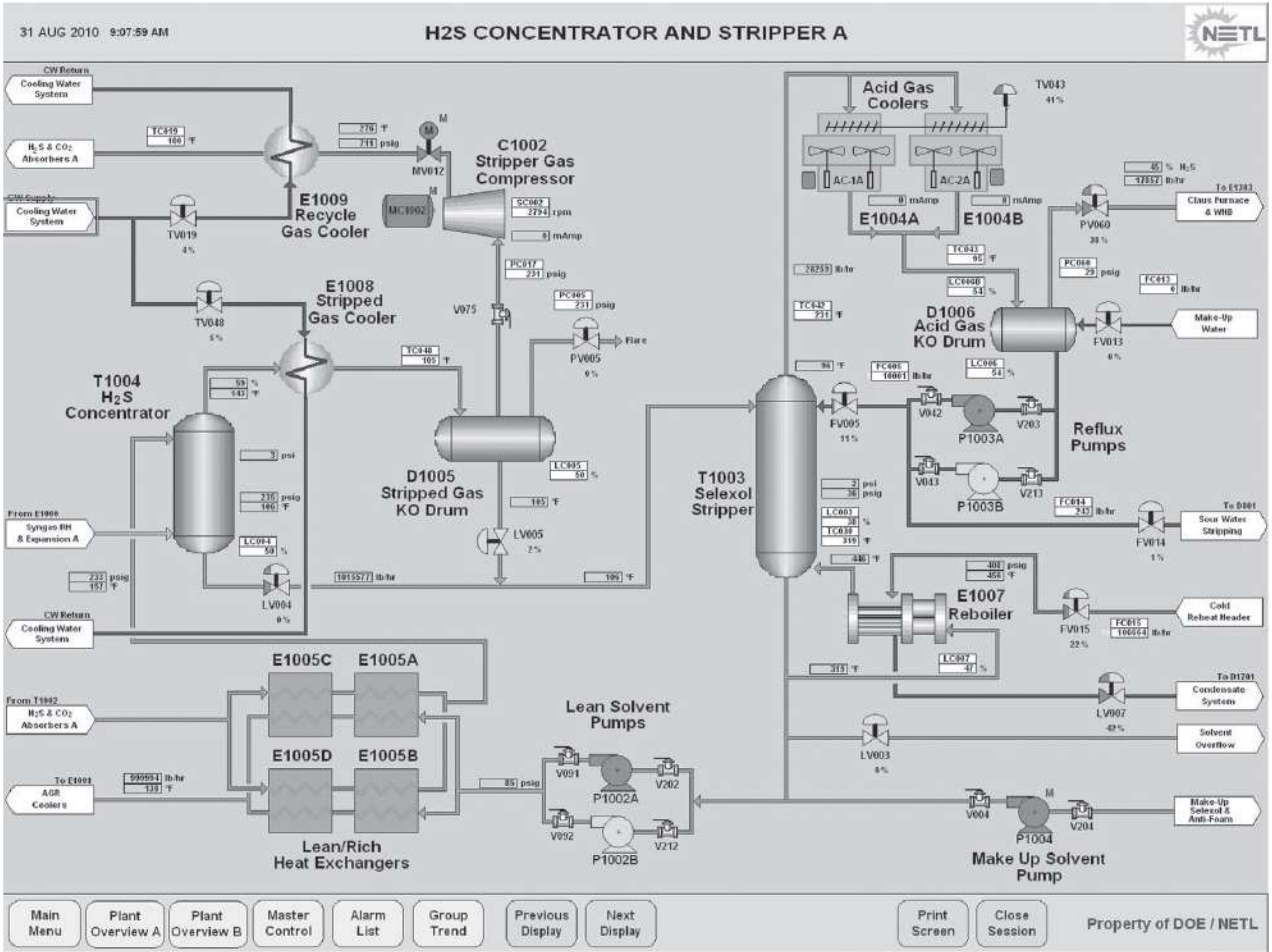


Figure 1.16 Example of an HMI Interface for an OTS (Reproduced by Permission of the DOE's National Energy Technical Laboratory and Invensys Systems Inc., Property and Copyright of Invensys plc, UK)

OPERATOR AND 3-D IMMERSIVE TRAINING SIMULATORS

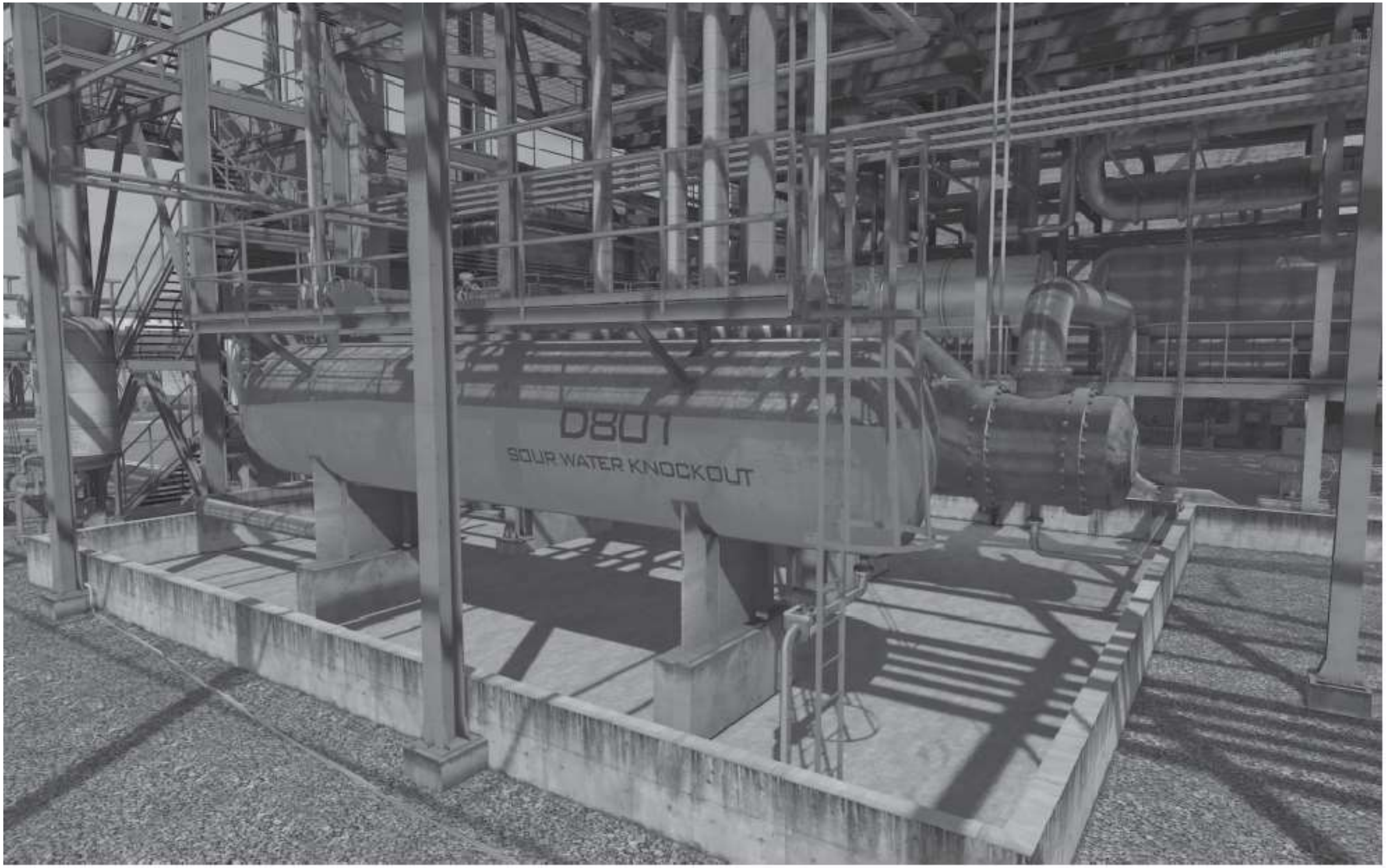


Figure 1.17 An Example of a Computer-Generated Image of a Horizontal Drum (Reproduced by Permission of the DOE's National Energy Technical Laboratory and Invensys Systems Inc., Property and Copyright of Invensys plc, UK)

OPERATOR AND 3-D IMMERSIVE TRAINING SIMULATORS

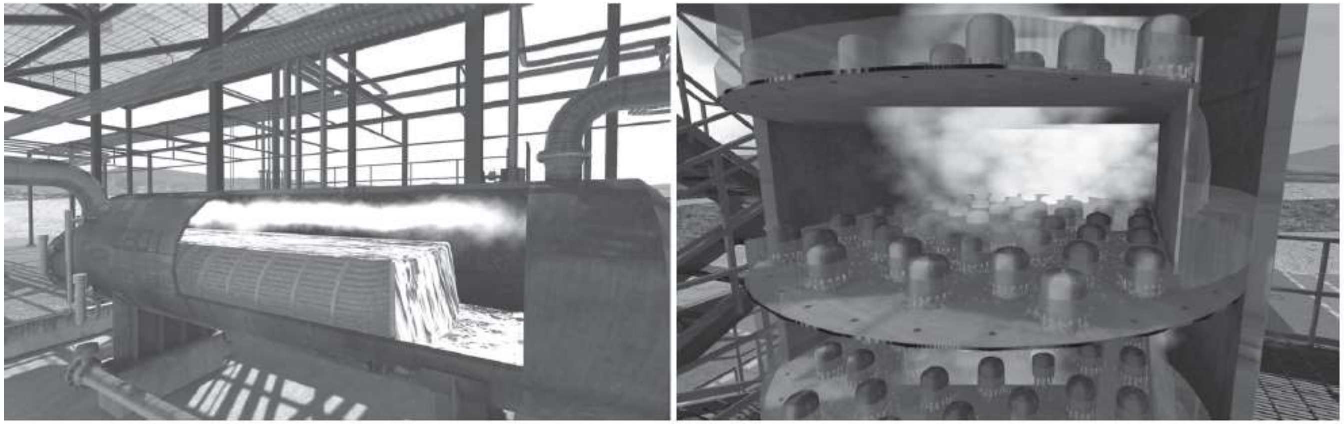


Figure 1.18 Augmented Reality in ITS: (a) Reboiler (b) Bubble-Cap Distillation Column (Reproduced by Permission of the DOE's National Energy Technical Laboratory and Invensys Systems Inc., Property and Copyright of Invensys plc, UK)

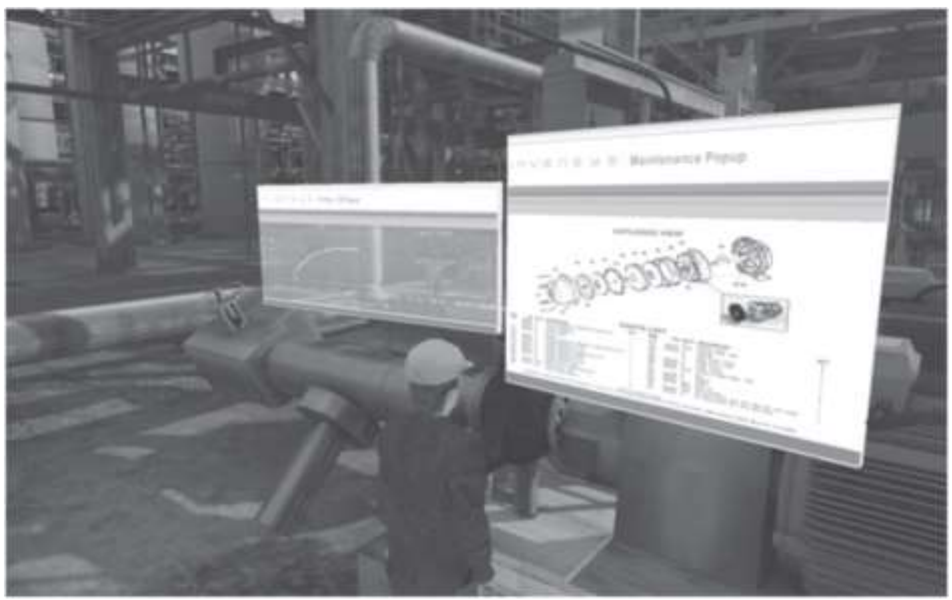
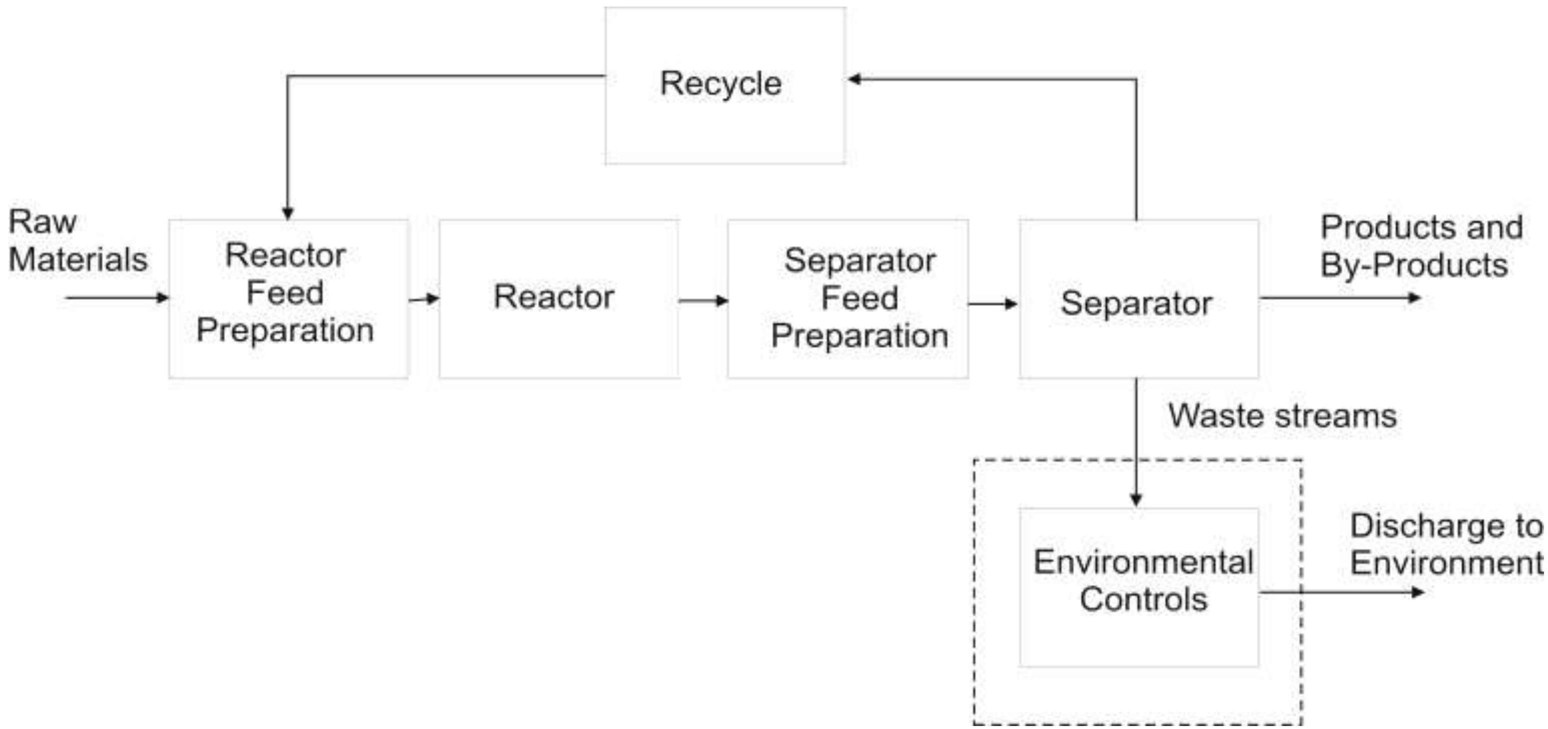


Figure 1.19 An Avatar Can Access Process trends and Observe Equipment Schematics in AR (Reproduced by Permission of Invensys Systems Inc., Property and Copyright of Invensys plc, UK)

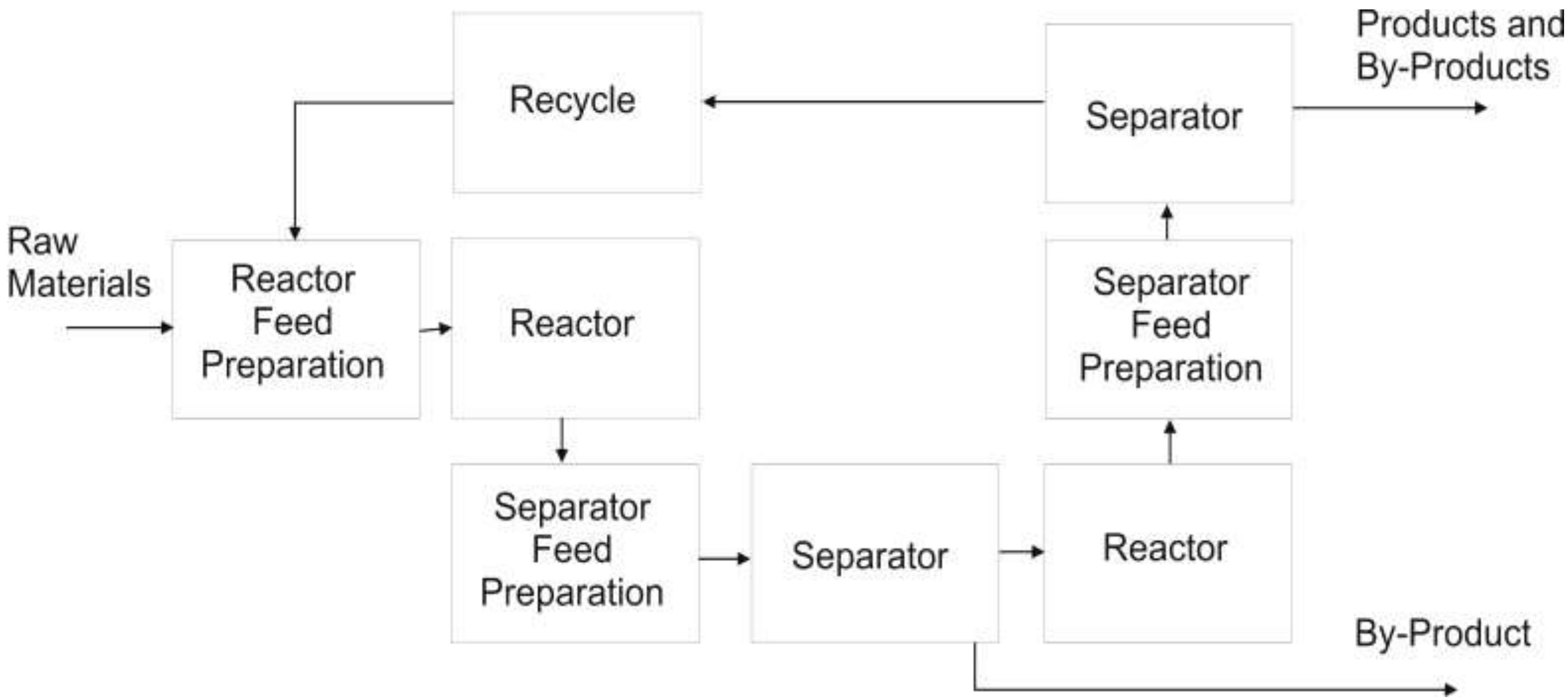
Structure and Synthesis of the Process Flow Diagram

Chemical Engineering Department
West Virginia University

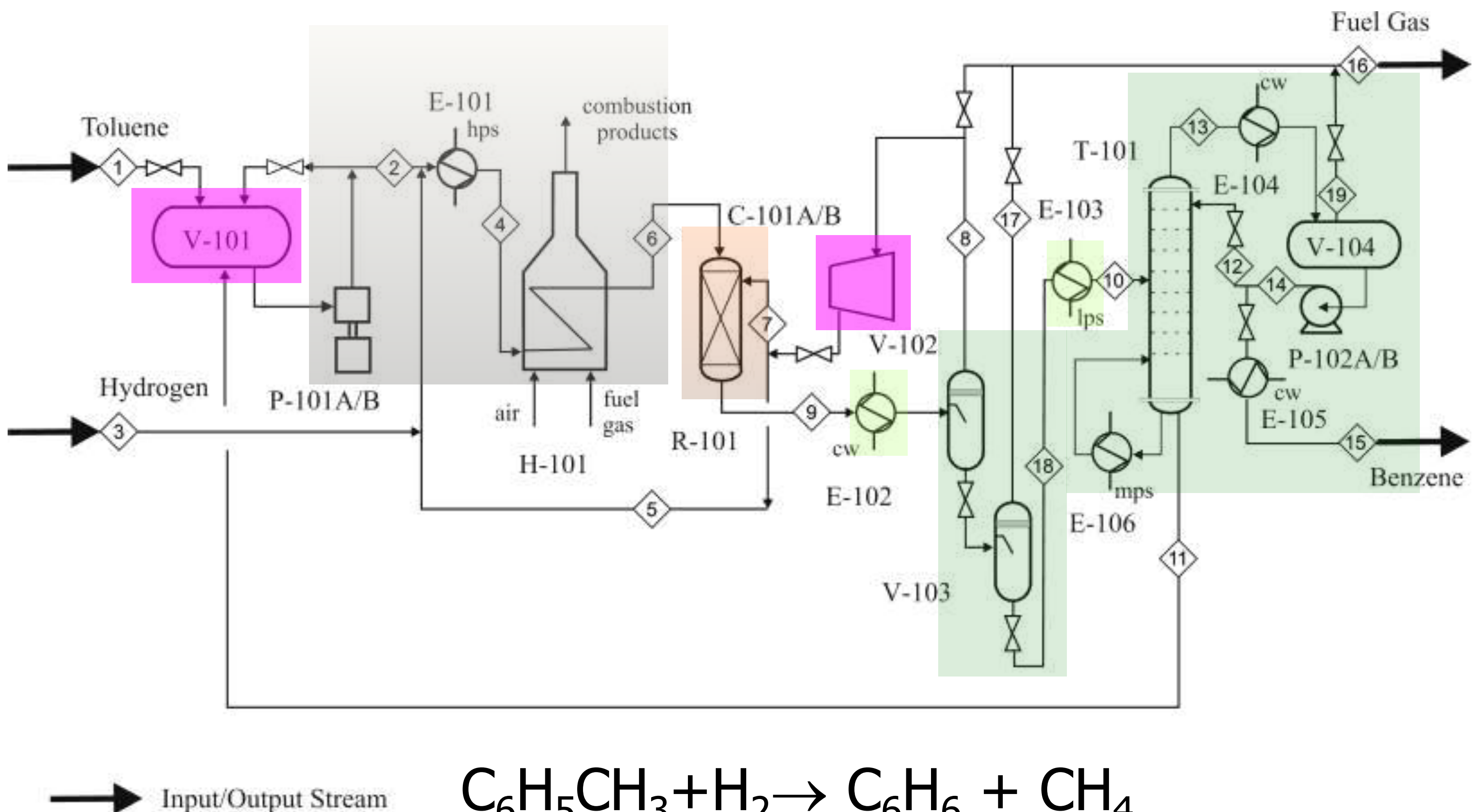
Generic Structure of Process Flow Diagrams



Generic Structure of Process Flow Diagrams



Generic Structure of Process Flow Diagrams



$$C_6H_5CH_3 + H_2 \rightarrow C_6H_6 + CH_4$$

Figure 2.2 Input Output Streams on Toluene Hydrodealkylation PFD

Environmental Control

- End of Pipe vs. Green Approach
  - Most significant changes obtained by changing process chemistry within reactor – eliminate/minimize unwanted by-products
- End of Pipe vs. Common Units
  - Fired Heaters - excess oxygen
    - low sulfur fuel
    - NO_X control
  - Wastewater - biological/sedimentation/ filtration

Approach of Douglas[1]

- Five step process to tackle a conceptual process design
  - Batch vs. continuous
  - Input-output structure
  - Identify and define recycle structure of process
  - Identify and design general structure of separation system
  - Identify and design heat-exchanger network or process energy recovery system

1 – Douglas, J.M., *Conceptual Design of Chemical Processes*, McGraw-Hill, NY, 1988

Batch vs. Continuous

Variables to Consider:

- Size
  - Batch < 500 tonne/yr ~ 1.5 tonne/day (< 2 m^3 of liquid or solid per day)
  - Continuous > 5000 tonne/yr

Batch vs. Continuous(cont.)

- Flexibility
  - Batch can handle many different feeds and products – more flexible
  - Continuous is better for smaller product slate and fewer feeds

Batch vs. Continuous(cont.)

Continuous allows the process to benefit from the "Economy of Scale," but the price is less flexibility

Batch vs. Continuous(cont.)

- Other Issues
  - Accountability and quality control – FDA requires batch accountability
  - Safety – batch is more accident prone
  - Scheduling of equipment – may be most important issue
  - Seasonal demands – *e.g.*, antifreeze, food products

Input – Output Structure

(Process Concept Diagram)

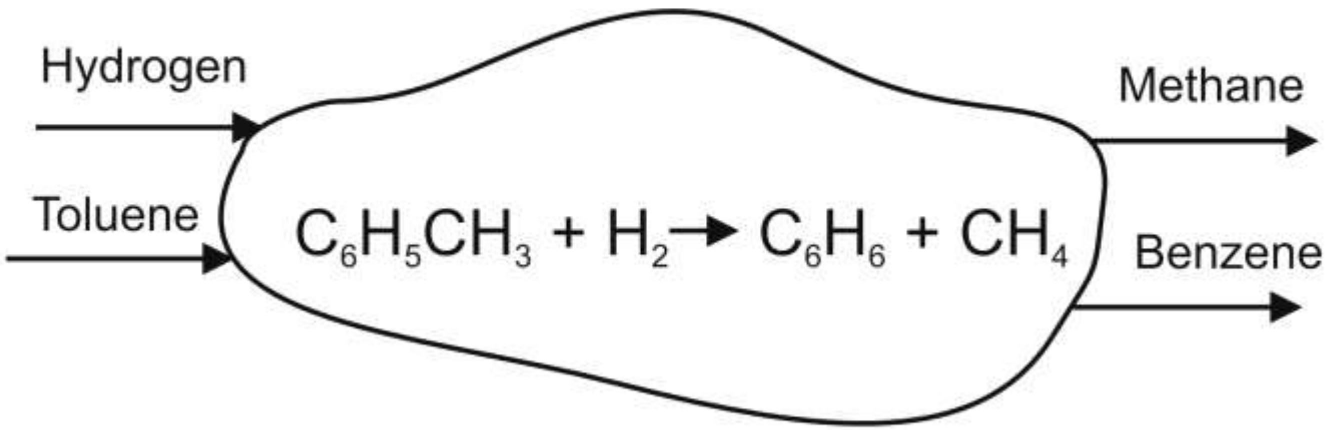


Figure 2.1: Input-Output Structure of Process Concept Diagram for the Toluene Hydrodealkylation Process

Input-Output on PFD

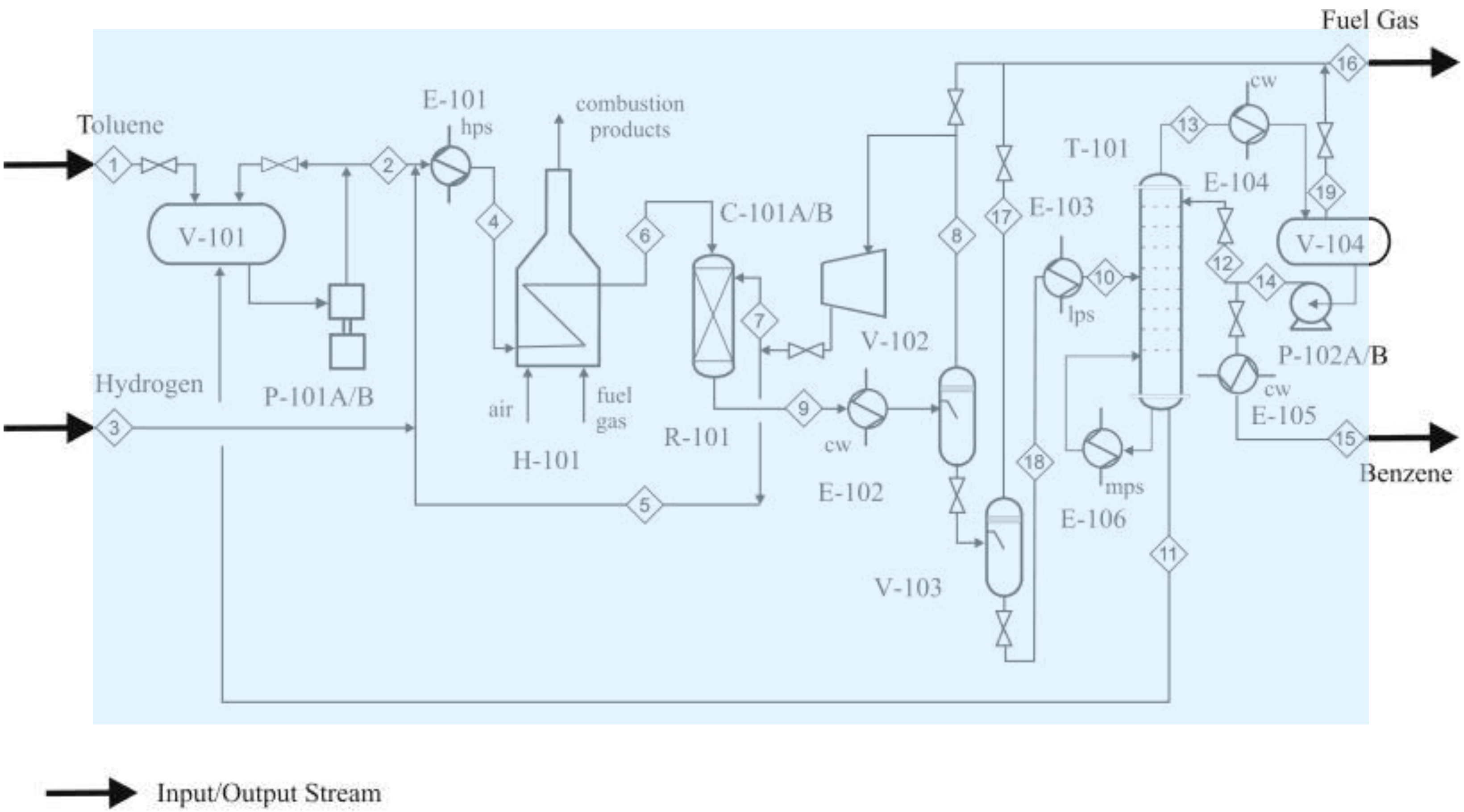


Figure 2.2 Input Output Streams on Toluene Hydrodealkylation PFD

Input-Output – Utility Streams

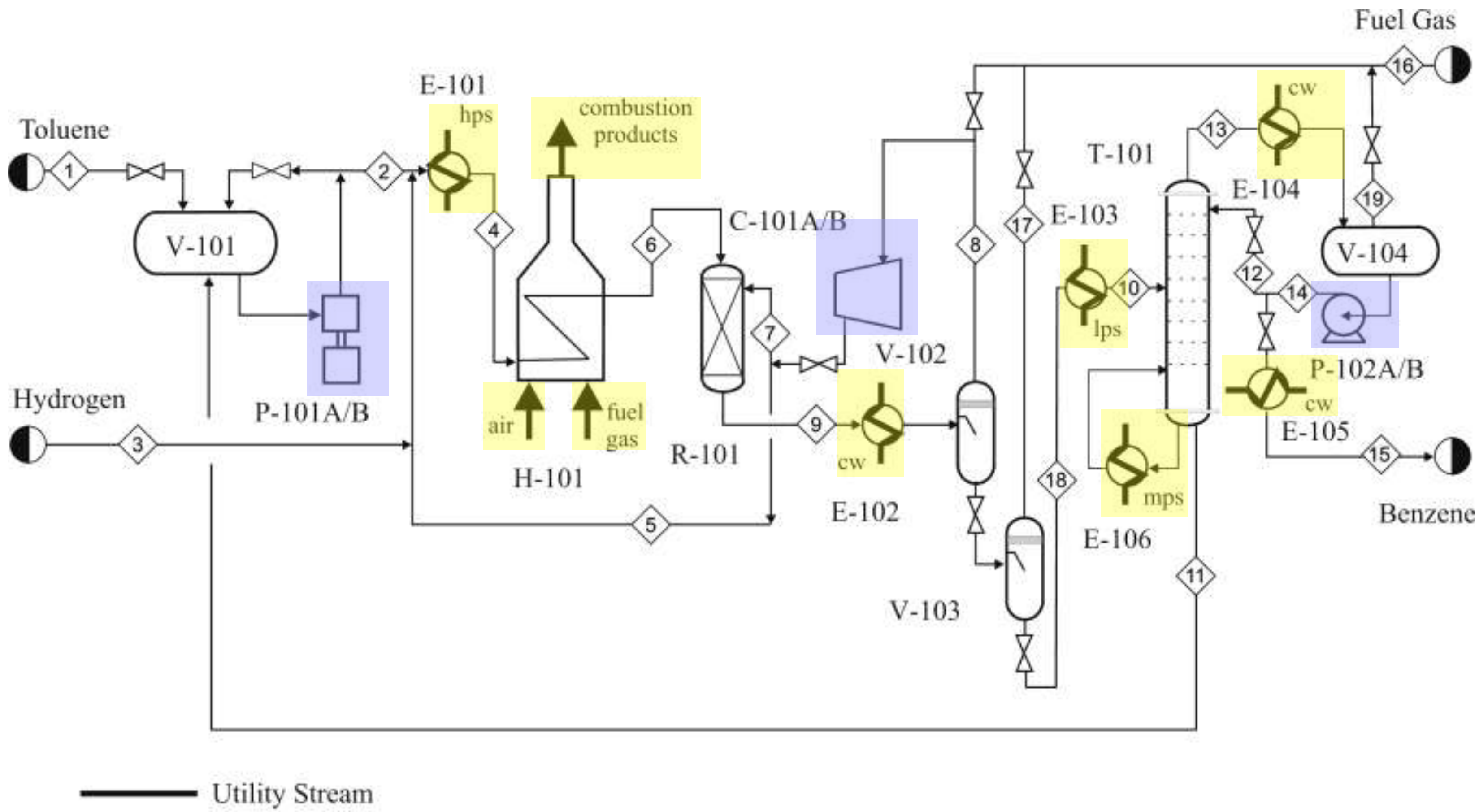
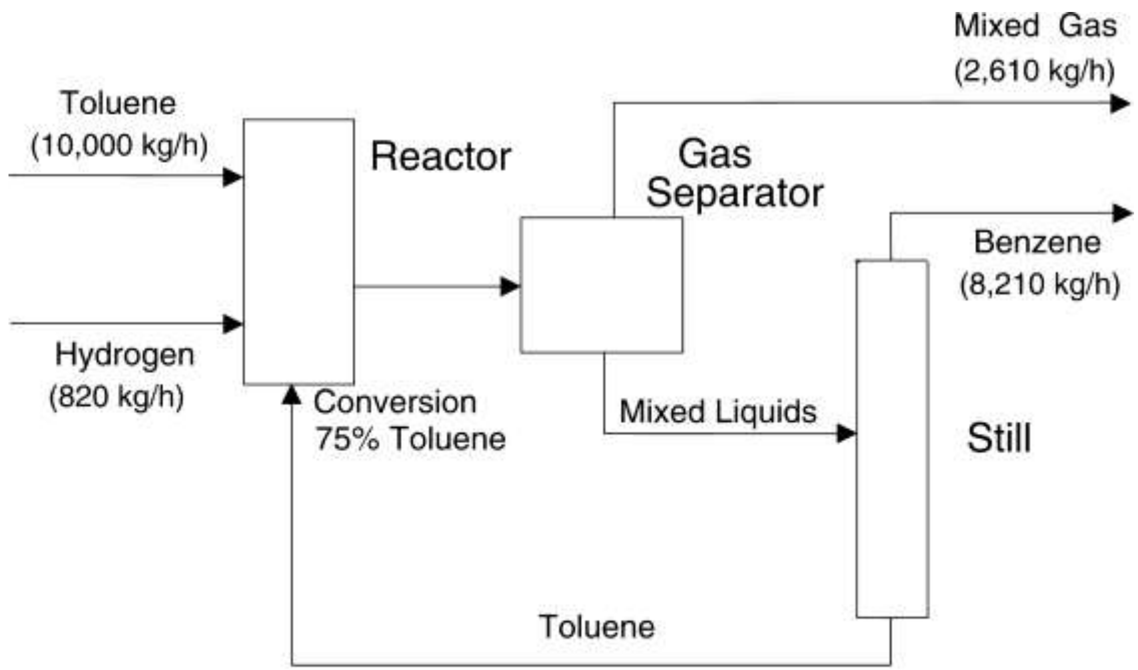


Figure 2.3: Identification of Utility Streams on the Toluene HDA PFD

The Block Flow Process Diagram



Reaction : $C_7H_8 + H_2 = C_6H_6 + CH_4$

$$2C_6H_6 = C_{12}H_{10} + H_2$$

Figure 1.1: Block Flow Process Diagram for the Production of Benzene

Other Input – Output Issues

Purify Feed ?

- Feed purity and trace components
  - Small quantities and "inerts" – do not separate

    Example H_2 in feed contains CH_4

    CH_4 does not react

    so – do not remove

Other Input – Output Issues (cont)

- If separation of impurities is difficult – Do not separate
  - Azeotrope – (water and ethanol)
  - Gases – (requires high *P* and low *T*)

How would you remove CH_4 from H_2?

Other Input – Output Issues (cont)

- If impurities foul or poison catalyst then ***separate***
  - Sulfur – Group VIII Metals (Pt, Pd, Ru, Rh)
  - CO in platinum PEM fuel cells

Note: S and CO may be present in very small amounts (ppm)

Other Input – Output Issues (cont)

- If impurity reacts to form difficult-to-separate material or hazardous product then ***separate***

Phosgene Example

$CH_4 + H_2O \longrightarrow CO + 3H_2$

$CO + Cl_2 \longrightarrow COCl_2$

Any $H_2 \longrightarrow HCl$

Other Input – Output Issues (cont)

- Impurity in large quantities then ***<u>purify</u>*** – why?

  A notable exception is air

Add Materials to Feed

- Stabilize products
- Enable separation/minimize side reactions
  - Anti-oxidants and scavengers
  - Solvents and catalysts

Inert Feeds

- Control exothermic reactions
  - Steam for oxidation reactions (e›
  - Reduces coke formation on catalyst
- Control equilibrium
  - Adding inerts shifts equilibrium to the right

e.g., styrene reaction

$$C_6H_5CH_2CH_3 \rightleftarrows C_6H_5CHCH_2 + H_2$$

Profit Margin

- If \$ Products - \$ Raw Material < 0, then do not bother to pursue this process, but start looking for an alternate route

Toluene HDA vs. Toluene Disproportionation

$C_6H_5CH_3 + H_2 \rightarrow C_6H_6 + CH_4$

Toluene benzene

$2C_6H_5CH_3 \rightarrow C_6H_6 + C_6H_4(CH_3)_2$

Toluene benzene xylene

Toluene used more efficiently

Recycle

- Since raw materials make up from 25 to 75% of total operating costs, should recover as much raw material as possible
- Exception is when raw materials are very cheap

For example, Air Separation

There are basically three ways that unreacted raw materials can be recycled in continuous processes.

1. Separate and purify unreacted feed material from products and then recycle.
2. Recycle feed and product together and use a purge stream.
3. Recycle feed and product together and do not use a purge stream.

Separate and Purify. Through the ingenuity of chemical engineers and chemists, technically feasible separation paths exist for mixtures of nearly all commercially desired chemicals. Therefore, the decision on whether to separate the unreacted raw materials must be made purely from economic considerations. In general, the ease with which a given separation can be made is dependent on two principles.

- First, for the separation process (unit operation) being considered, what conditions (temperature and pressure) are necessary to operate the process?
- Second, for the chemical species requiring separation, are the differences in physical or chemical properties for the species, on which the separation is based, large or small? Examples that illustrate these principles are given below.

3 Basic Recycle Structures

- Separate and purify unreacted feed from products and then recycle, *e.g.*, toluene
- Recycle feed and products together and use a purge stream, *e.g.*, hydrogen with purge as fuel gas
- Recycle feed and products together but do not use a purge stream - must come to Equilibrium

$$2C_6H_6 \rightleftharpoons C_{12}H_{10} + H_2$$

Recycle Structure in PFD

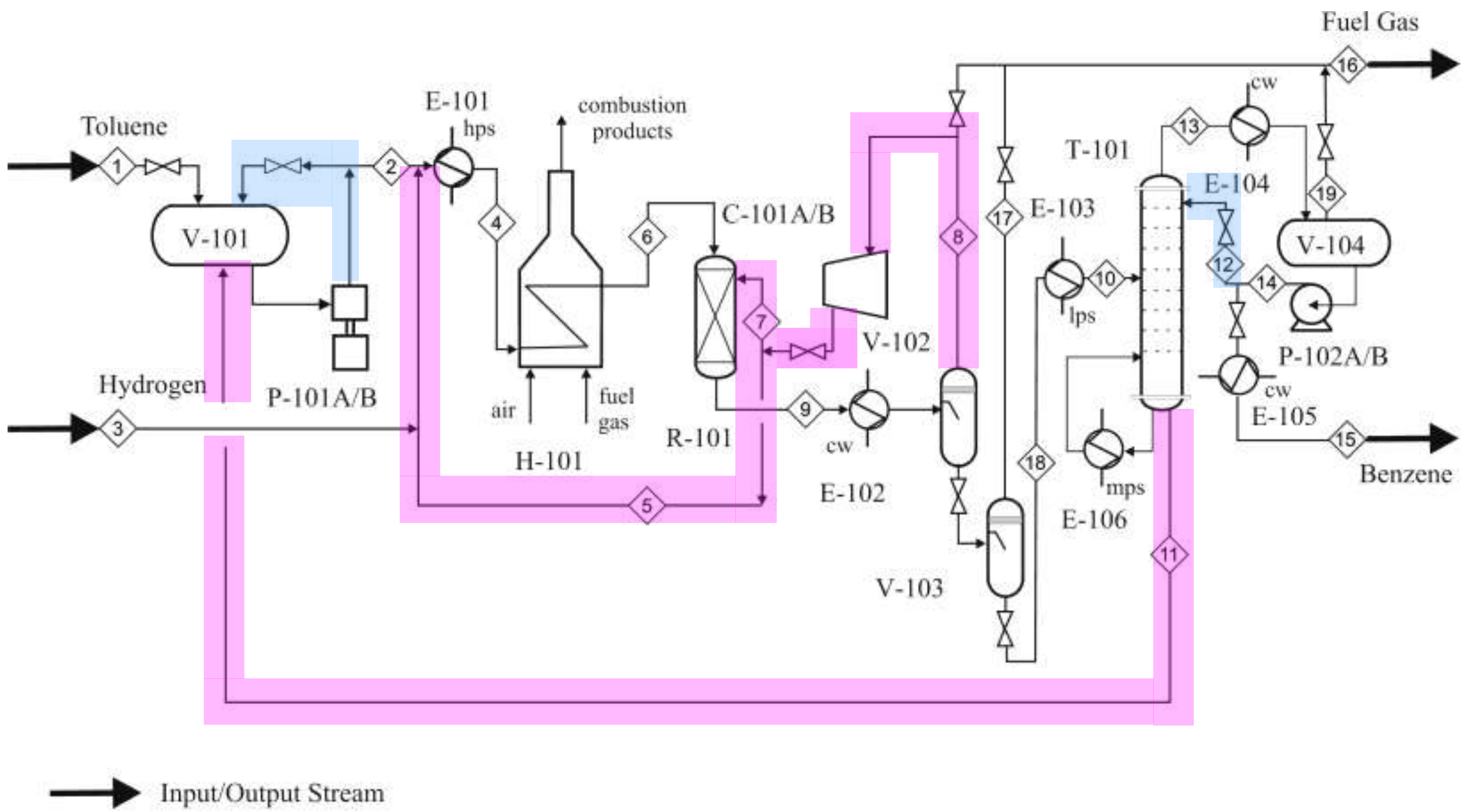


Figure 2.2 Input Output Streams on Toluene Hydrodealkylation PFD

Recycle without separation or purge

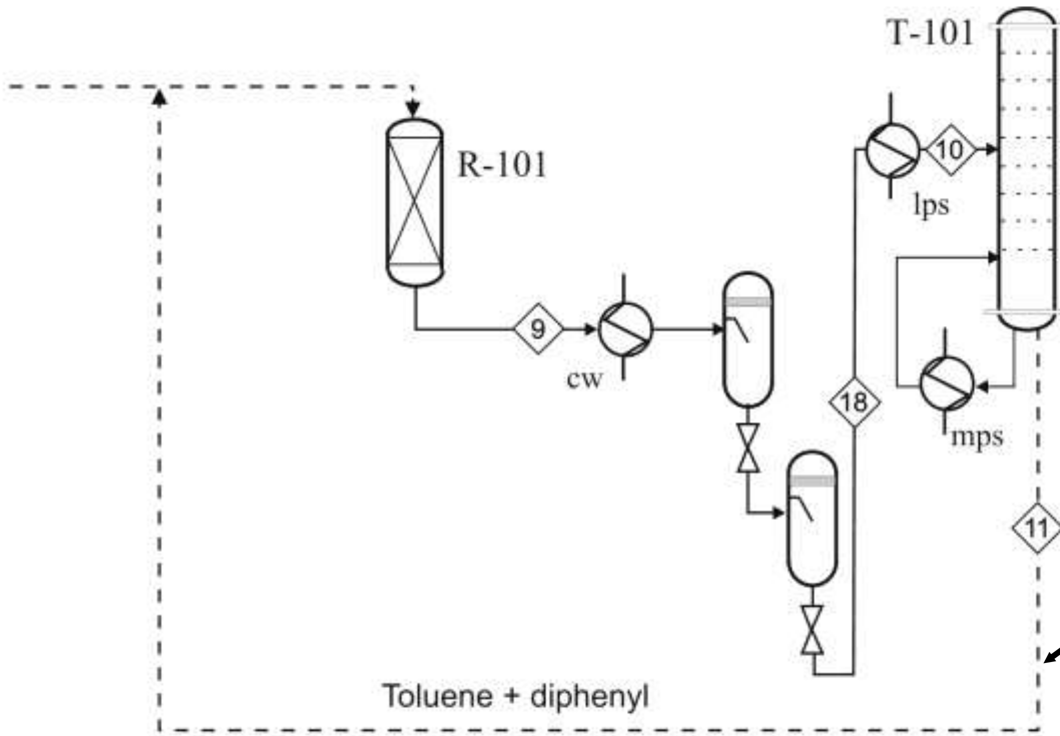


Figure E2.5A: PFD for Alternative A in Example 2.5 - Recycle of Diphenyl without Separation

Recycle increases and equip. and op. costs increase

$$2C_6H_6 \rightleftharpoons C_{12}H_{10} + H_2$$

Recycle with Separation (and Purge)

$$2C_6H_6 \rightleftharpoons C_{12}H_{10} + H_2$$

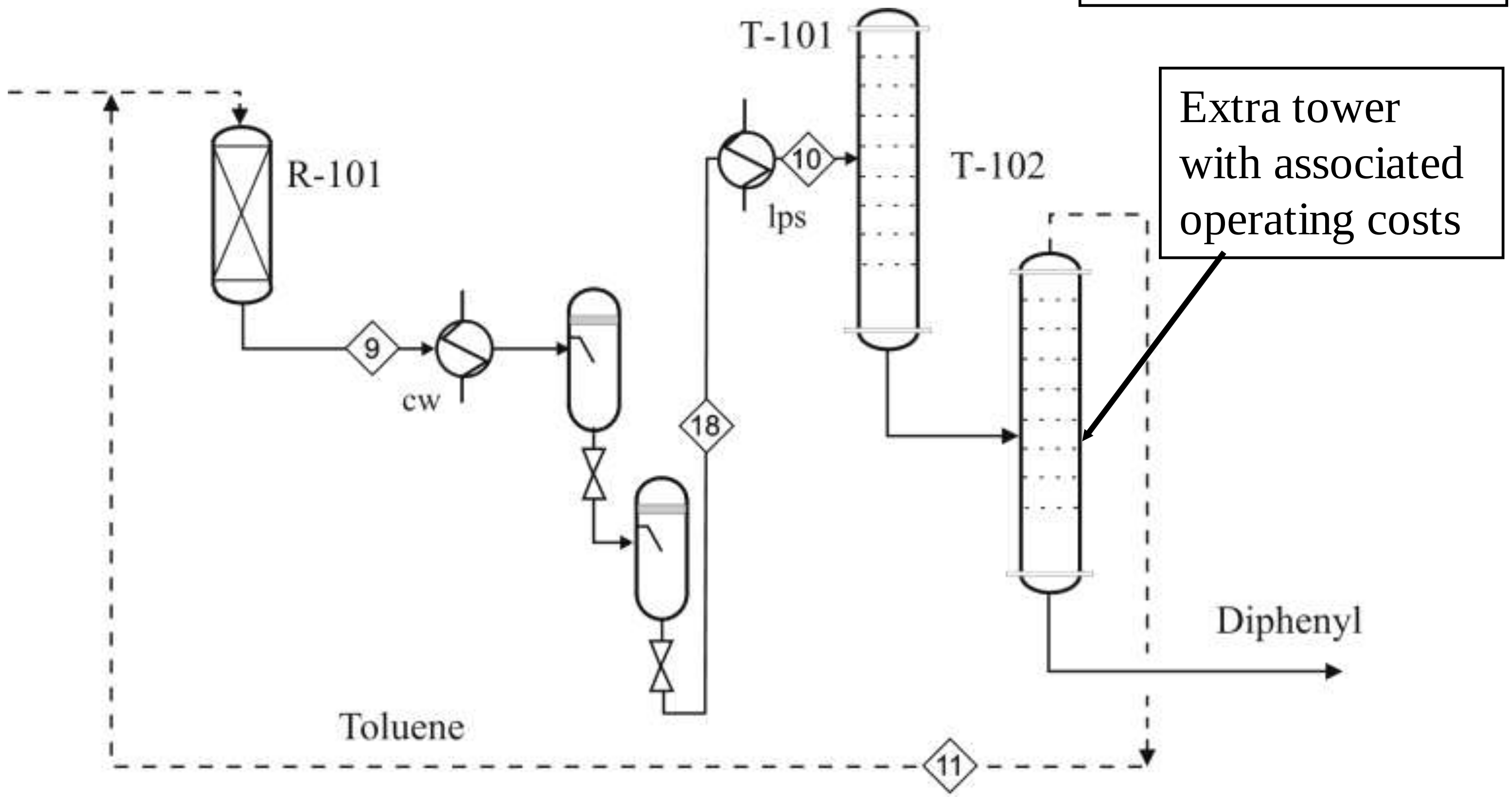


Figure E2.5B: PFD for Alternative B in Example 2.5 - Separation of Diphenyl prior to Recycle of Toluene

Equilibrium Constant
for reaction

$$2C_6H_6 = C_{12}H_{10} + H_2$$

$$\ln K_{eq} = 1.788 - \frac{4135.2}{T(K)}$$

$$x = 1.36 \text{ kmol/h.}$$

The exit conditions of the reactor can be estimated by assuming that the benzene-diphenyl reaction has reached equilibrium, a conservative assumption. Using this assumption and data from Table 1.5 for Stream 9, if x kmol/h of diphenyl is present in the reactor effluent then:

$$K_{eq} = \frac{[C_{10}H_{12}][H_2]}{[C_6H_6]^2} \Rightarrow \exp\left[1.788 - \frac{4135.2}{(654 + 273)}\right] = \frac{(x)(652.6 + x)}{(116 - 2x)^2}$$

Other Issues on Recycle

- Number of recycle streams
- Does excess reactant affect structure
  - Size of Recycle Loop

    H_2 : Toluene = 5 : 1
- Number of Reactors
  - Separate and recycle to different reactors

Other Issues on Recycle (cont.)

- Do we need to purify prior to recycling?
- Is recycling of inerts warranted?
- Can recycling an unwanted inert material push equilibrium to the right?
  – Gasification of coal – CO_2 recycle

Other Issues on Recycle (cont.)

- Can recycling an unwanted inert control reaction
  - CO_2 in Gasifier
- Phase of Recycle Stream?

Chapter 3 – Batch Processing

Department of Chemical Engineering
West Virginia University

Outline

- Design Calculations
- Gantt Charts and Scheduling
- NonOverlapping and Overlapping Operations
- Cycle Times
- Flowshop Plants
- Jobshop Plants

Outline

- Product and Intermediate Storage
- Parallel Process Units
- Equipment Design for Multiproduct Batch Processes

Outline

- Design Calculations
- Gantt Charts and Scheduling
- NonOverlapping and Overlapping Operations
- Cycle Times
- Flowshop Plants
- Jobshop Plants

Design Calculations

- Recipe
- Operations are unsteady state
- Often existing equipment is used – not designed specifically for recipe
- Transfer from equipment and cleaning important
- Recycling and energy integration issues

Recipe

Step 1: 500 kg of reactant A (MW = 100 kg/kmol) is added to 5000 kg of a mixture of organic solvent (MW = 200 kg/kmol) containing 60% excess of a second reactant B (MW = 125 kg/kmol) in a jacketed reaction vessel (R-301), the reactor is sealed, and the mixture is stirred and heated (using steam in the jacket) until the temperature has risen to 95°C. The density of the reacting mixture is 875 kg/m^3 (time taken = 1.5 h).

Step 2: Once the reaction mixture has reached 95°C, a solid catalyst is added, and reaction takes place while the batch of reactants is stirred. The required conversion is 94% (time taken = 2.0 h).

Step 3 ……..

Step 4 ……..

Step 5 ……...

Design Calculations

$$V = \frac{5500[\text{kg}]}{875[\text{kg/m}^3]} = 6.286\, m^3$$

$$V_{required} = \frac{5500[\text{kg}]}{875[\text{kg/m}^3]} \frac{1}{0.6} = 10.48\, m^3 = 2768\, gal$$

60% fill

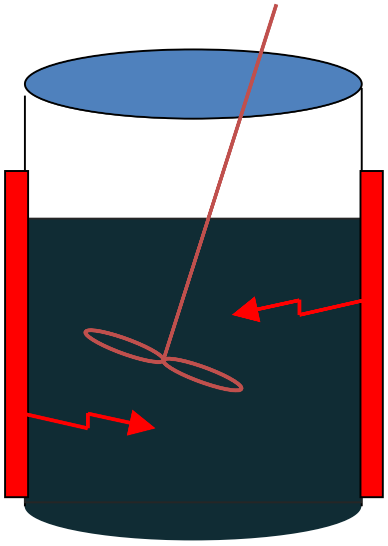
- Because reactors of this sort come in standard sizes, a 3000-gallon (V_{tank}) reactor is selected.

Design Calculations

Find the time to heat vessel contents from ambient to 95°C

$$\rho V C_p \frac{dT}{dt} = UA\left(T_s - T\right)$$

$$\ln \frac{\left(T_s - T_{final}\right)}{\left(T_s - T_o\right)} = -\frac{UA\Delta t}{\rho V C_p}$$



Design Calculations

- Transfer from equipment and cleaning important

  Gravity drain or pump

  Cleaning – GMP/documentation
- Recycling and energy integration issues

  Recycling – often use dump tank and accumulate material from several runs

Outline

- Design Calculations
- Gantt Charts and Scheduling
- NonOverlapping and Overlapping Operations
- Cycle Times
- Flowshop Plants
- Jobshop Plants

Gantt Charts and Scheduling

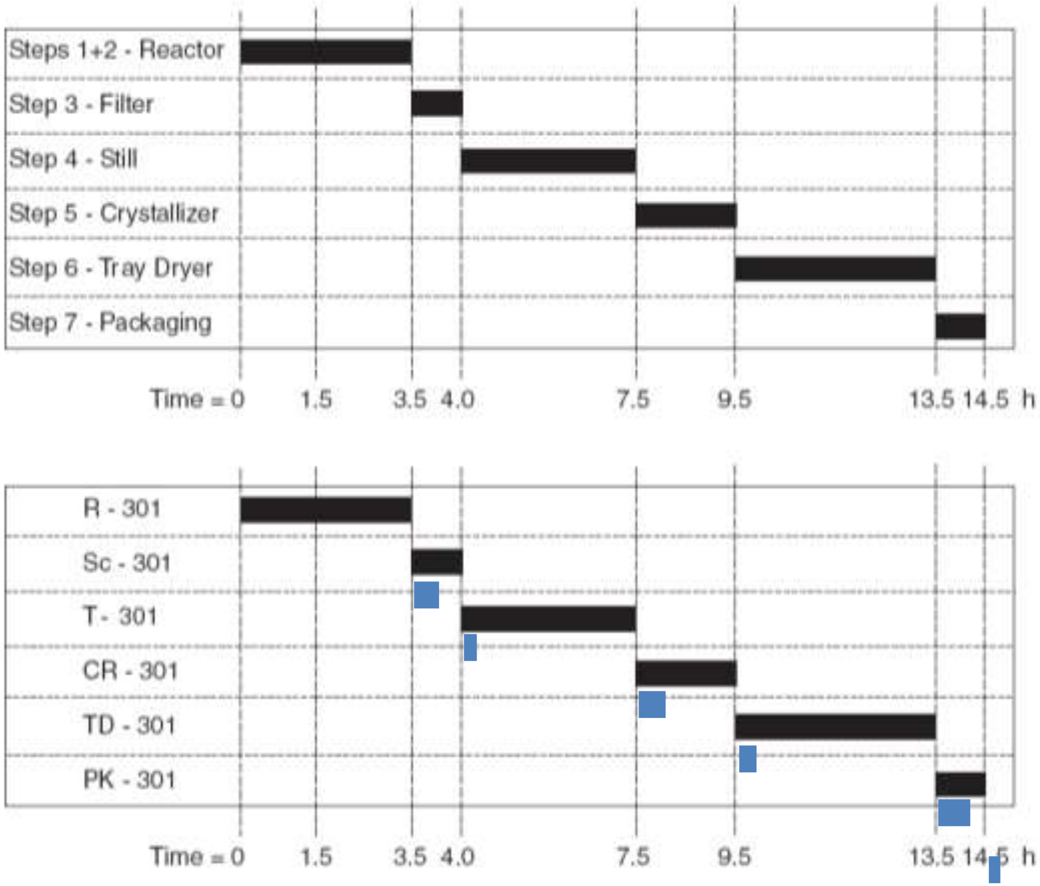


Figure E3.2 Gantt Chart Showing Sequence of Events for the Manufacture of API in Example 3.1

Outline

- Design Calculations
- Gantt Charts and Scheduling
- NonOverlapping and Overlapping Operations
- Cycle Times
- Flowshop Plants
- Jobshop Plants

NonOverlapping and Overlapping Operations

$$T_{NO} = n\sum_{i=1}^{m} t_i$$

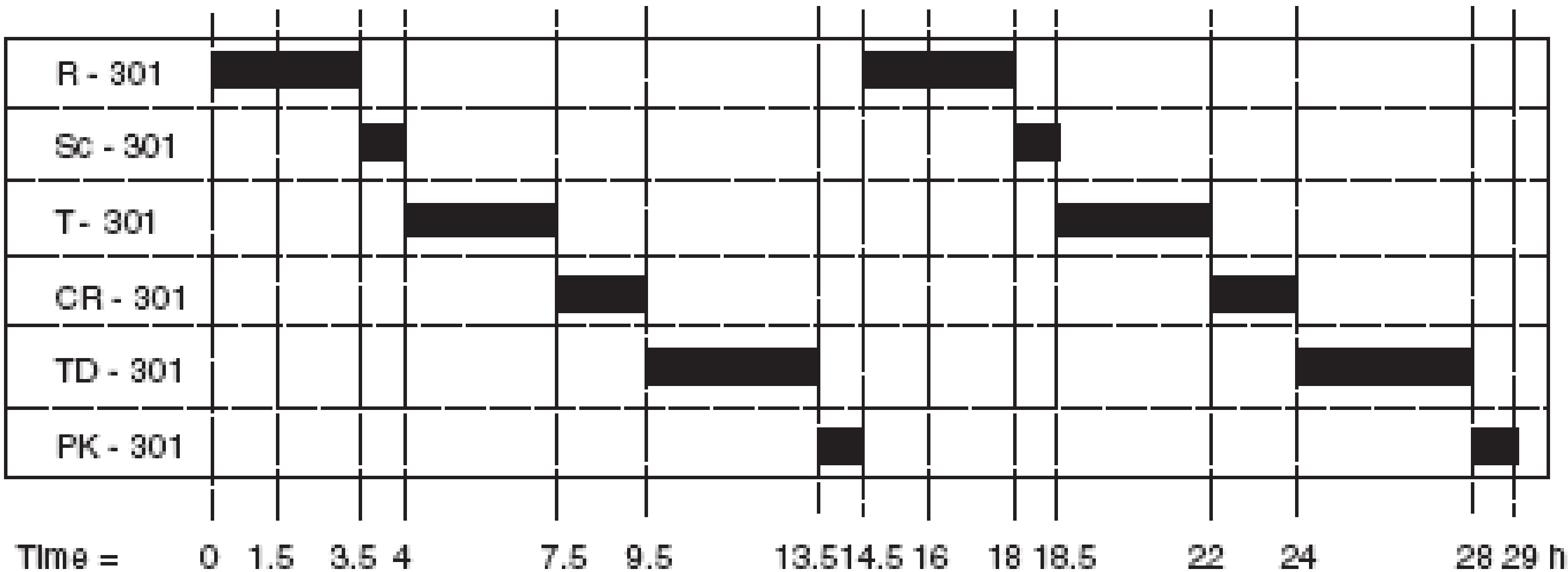


Figure 3.1 Example of a Nonoverlapping Sequence of Batch Operations

NonOverlapping and Overlapping Operations

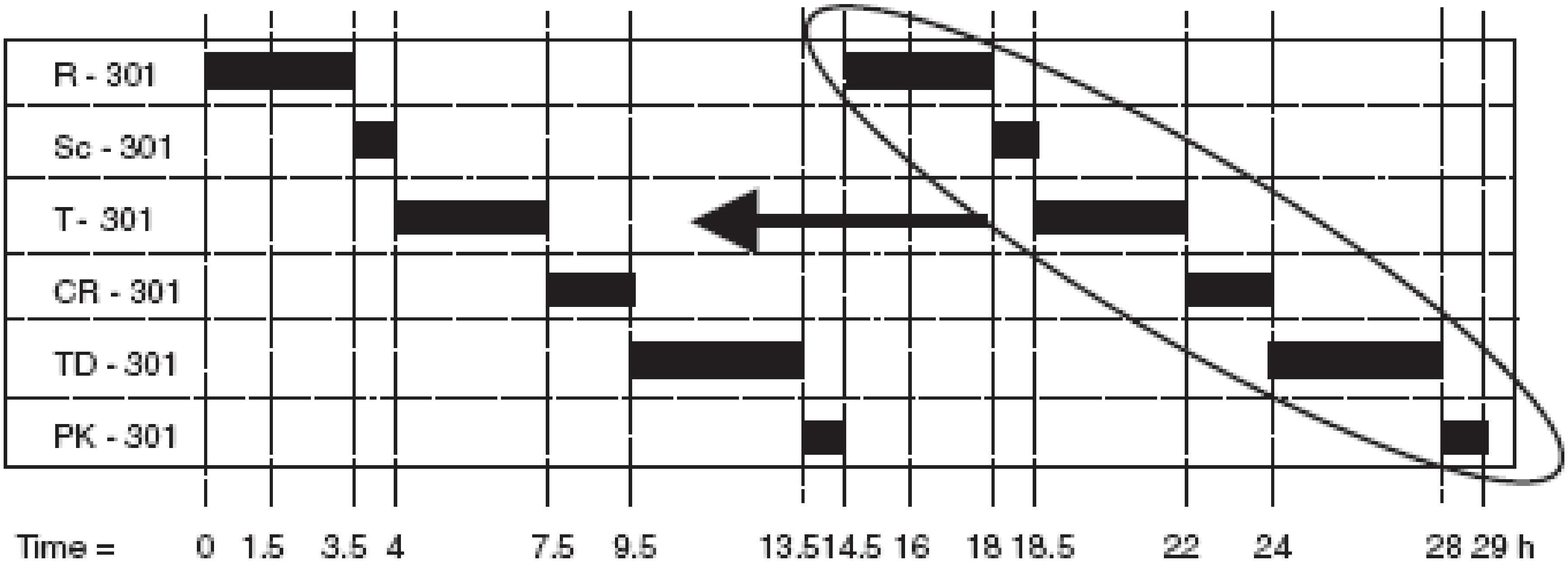


Figure 3.2 Backward Shifting of Batches, Giving Rise to Overlapping Sequencing

NonOverlapping and Overlapping Operations

$$T_O = T = (n-1)\max_{i=1,\dots,m}(t_i) + \sum_{i=1}^{m} t_i$$

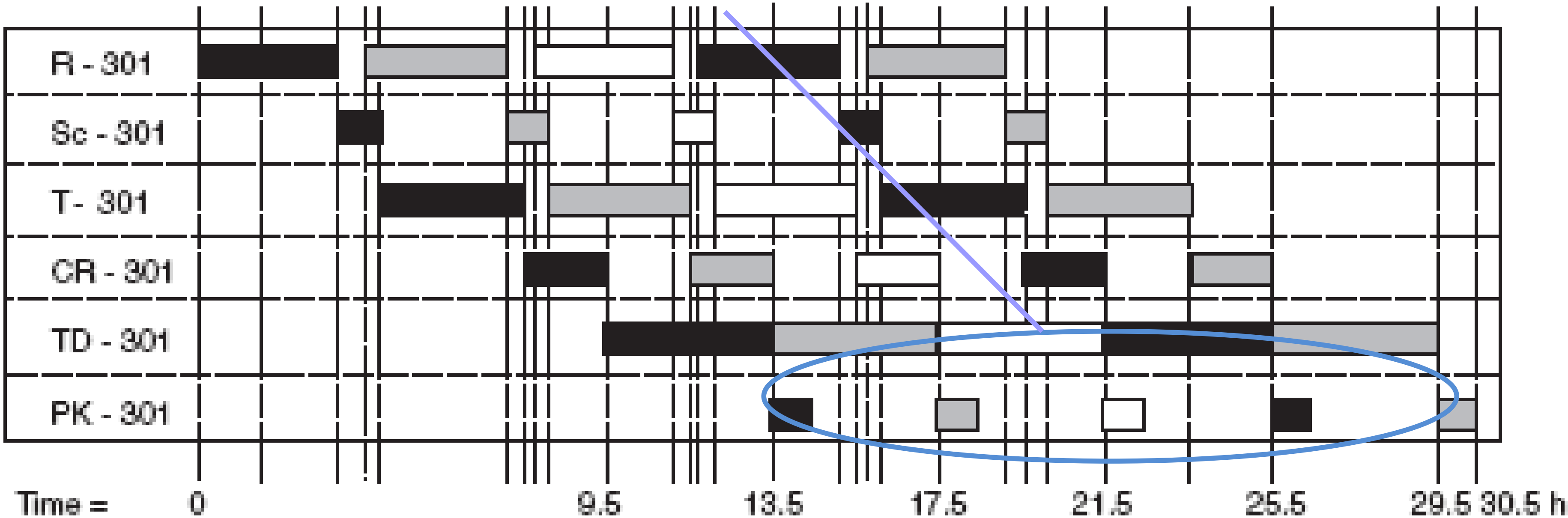


Figure 3.3 The Limiting Case for Overlapping Batch Sequencing

Outline

- Design Calculations
- Gantt Charts and Scheduling
- NonOverlapping and Overlapping Operations
- Cycle Times
- Flowshop Plants
- Jobshop Plants

Cycle Times

t_{cycle} = total time/number of batches

$$t_{cycle,NO} = \frac{T_{NO}}{n} = \frac{n\sum_{i=1}^{m} t_i}{n} = \sum_{i=1}^{m} t_i$$

$$t_{cycle,O} = t_{cycle} = \frac{T}{n} = \frac{(n-1)\max\limits_{i=1,\ldots,m}(t_i) + \sum_{i=1}^{m} t_i}{n}$$

Cycle Times

From the overlapping scheme – when the number of batches (n) to be produced is large,

$$t_{cycle} \cong \max_{i=1,..,m} \{t_i\}$$

Cycle Times

Example

A batch process involves 4 pieces of equipment with the following process times

Heating and Mixing	2.5 h
Reaction	3.5 h
Filtration and Drying	1.5 h
Crystallization and packaging	2.0 h

Determine the cycle times for non-overlapping and overlapping operation.

Cycle Times

Example

T_{NO} = 2.5+3.5+1.5+2.0 = 9.5 h

Outline

- Design Calculations
- Gantt Charts and Scheduling
- NonOverlapping and Overlapping Operations
- Cycle Times
- Flowshop Plants
- Jobshop Plants

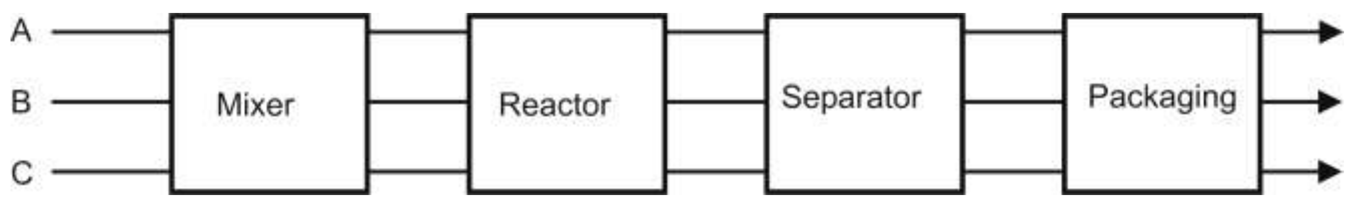
Multiple Products

When we have multiple products that can be produced in the same equipment, the scheduling of multiple batches requires careful planning

The different products may require the same processing steps, or more often will require only a subset of all possible steps. Moreover, the order in which a batch process uses different equipment might also differ from product to product

Flowshop Plants

If all the products use the same equipment in the same order or sequence, but not necessarily for the same lengths of time then the plant is referred to as a **flowshop** plant



Flowshop Plants

Consider 3 products A, B, C each requiring to be run n_A, n_B, and n_C times in a given period.

When we want to produce multiple products using the same equipment then we can either:

- run campaigns of the same product followed by campaigns of the next product followed by……
  i.e., run A - n_A times followed by B - n_B times followed by C - n_C times - AAAAAAAABBBBBBBBCCCCCCC
- Intermingle the products so that we run ABCABCABC… or ACBACBACB…, or AABCAABCAABC

Flowshop Plants

For multiple- (single) product campaigns the total processing time, or production cycle time, is found by adding the operation times for each product.

If the number of batches per campaign is large (for example, >10), then the production cycle time can be approximated by

$$T = \sum_{j=A}^{C} n_j \left\{ t_{cycle} \right\}_j \cong \sum_{j=A}^{C} n_j \left\{ \max_{i=1,..,m} \{t_i\} \right\}_j$$

Flowshop Plants

Example – production of three products A, B, and C

Market demand dictates that equal numbers of batches of the three products be produced over a prolonged period of time.

Determine the total number of batches that can be produced in a production cycle equal to one month of operation of the plant using separate campaigns for each product, assuming that a month of operation is equivalent to 500 h (based on 1/12 of a 6000 h year for a three-shift plant operating five days per week).

Use Single Product Campaigns

Equipment Times (in Hours) Needed to Produce A, B, and C

Product	Time in Mixer	Time in Reactor	Time in Separator	Time in Packaging	Total Time
A	1.5	1.5	2.5	2.5	8.0
B	1.0	2.5	4.5	1.5	9.5
C	1.0	4.5	3.5	2.0	11.0

Flowshop Plants

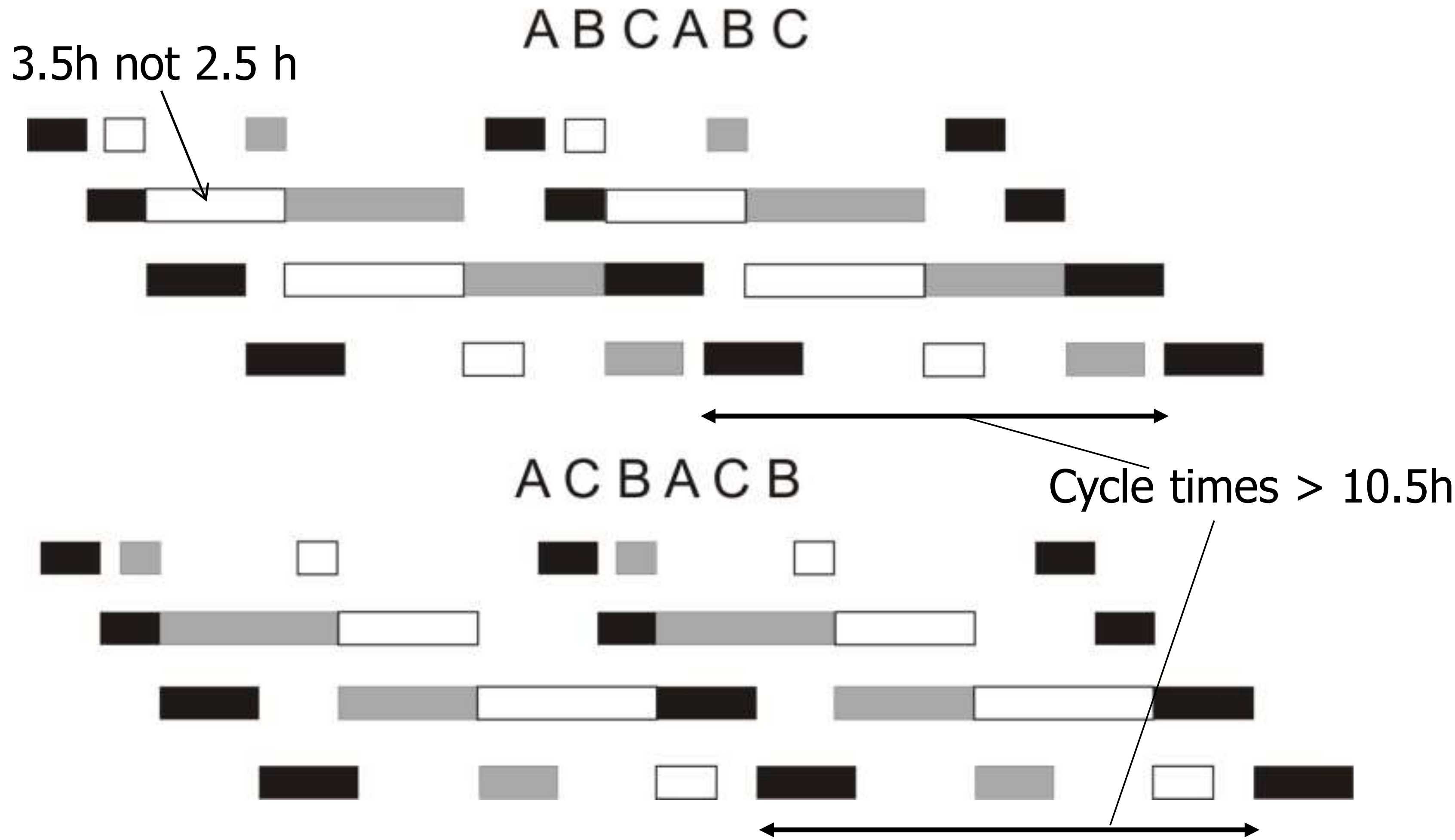
What if we want to run as ABCABCABC. What is the cycle time and how many batches can we produce of each product in 500h?

Equipment Times (in Hours) Needed to Produce A, B, and C

Product	Time in Mixer	Time in Reactor	Time in Separator	Time in Packaging	Total Time
A	1.5	1.5	2.5	2.5	8.0
B	1.0	2.5	4.5	1.5	9.5
C	1.0	4.5	3.5	2.0	11.0

Flowshop Plants

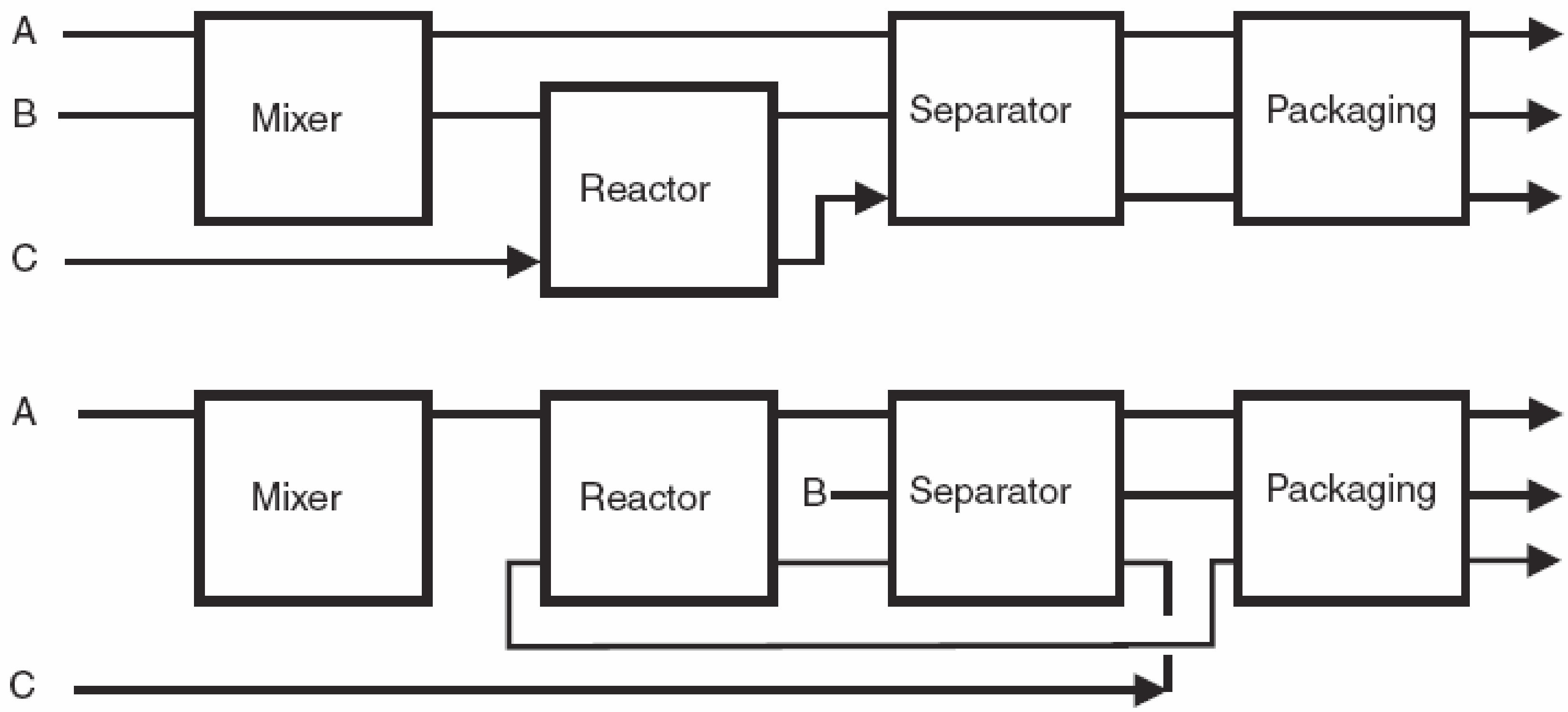
Flowshop Plants



Jobshop Plants

When not all products use the same equipment or the sequence of using the equipment is different for different products, then the plant is referred to as a **jobshop** plant.

Jobshop Plants



Jobshop Plants

For running sequential (multiple) single-product campaigns, the cycle time and number of batches can be determined by the same method used for the flowshop plant.

For sequential operation of batches ABCABC.. or ACBACB… we need to draw the Gantt chart.

Jobshop Plants

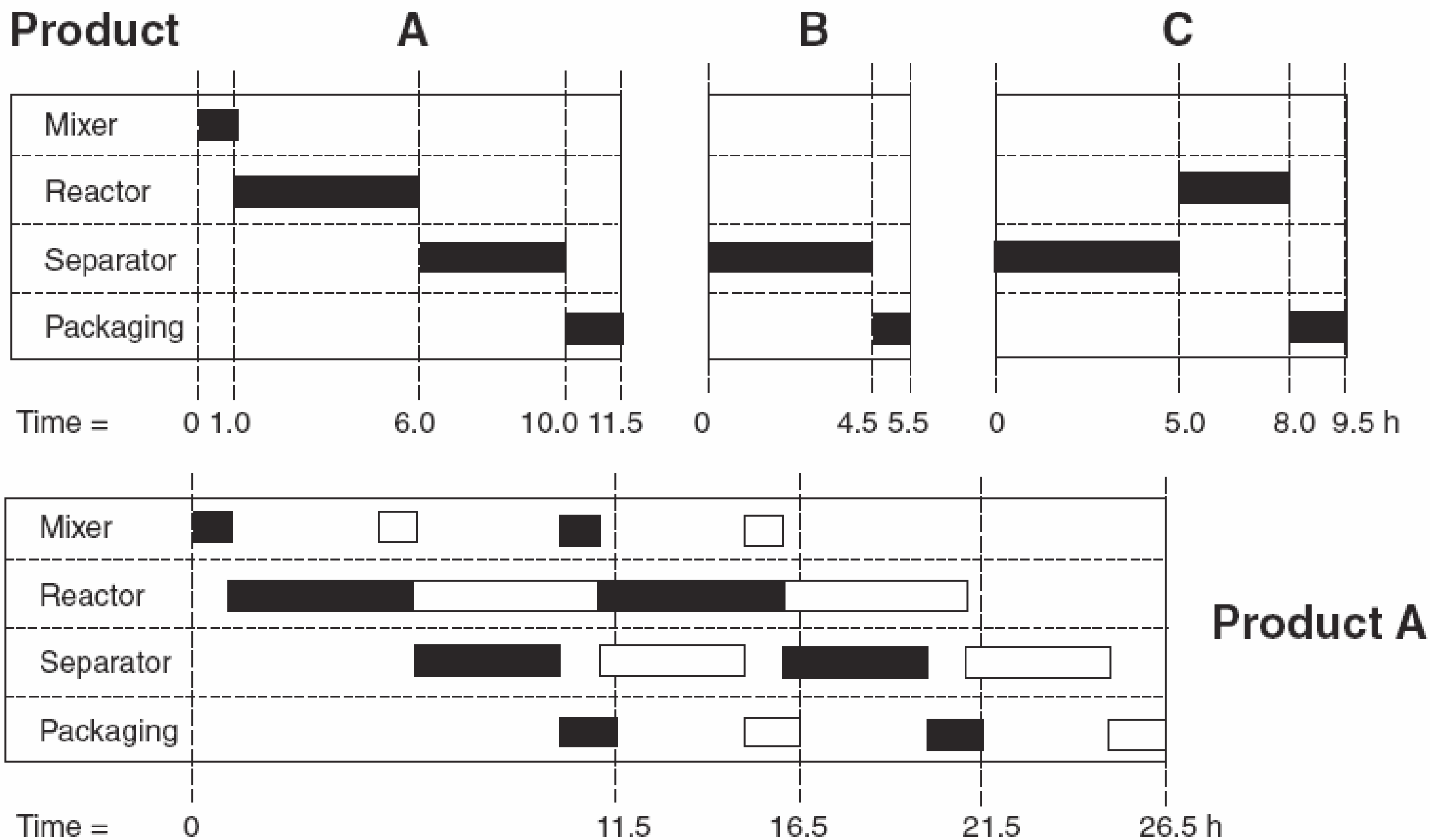
Construct the Gantt charts for overlapping single-product campaigns for products A, B, and C and for the multiproduct campaign with sequence ABCABCABC....

Equipment Processing Times (in Hours) for Processes A, B, and C

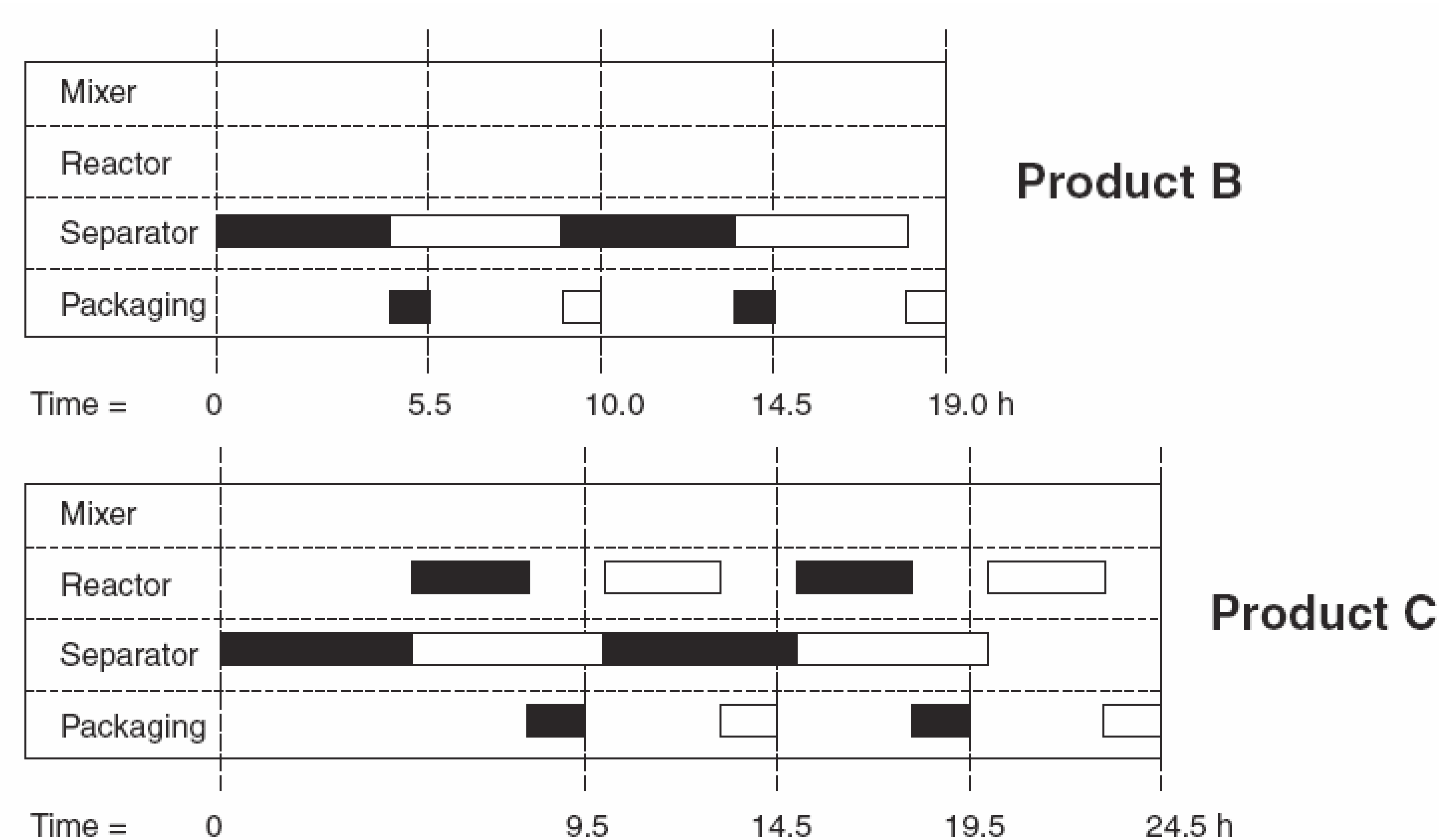
Process	Mixer	Reactor	Separator	Packaging
A	1.0	5.0	4.0	1.5
B	—	—	4.5	1.0
C	—	3.0	5.0	1.5

For equal batches of A, B, and C how many batches can be produced in 500 h?

Jobshop Plants

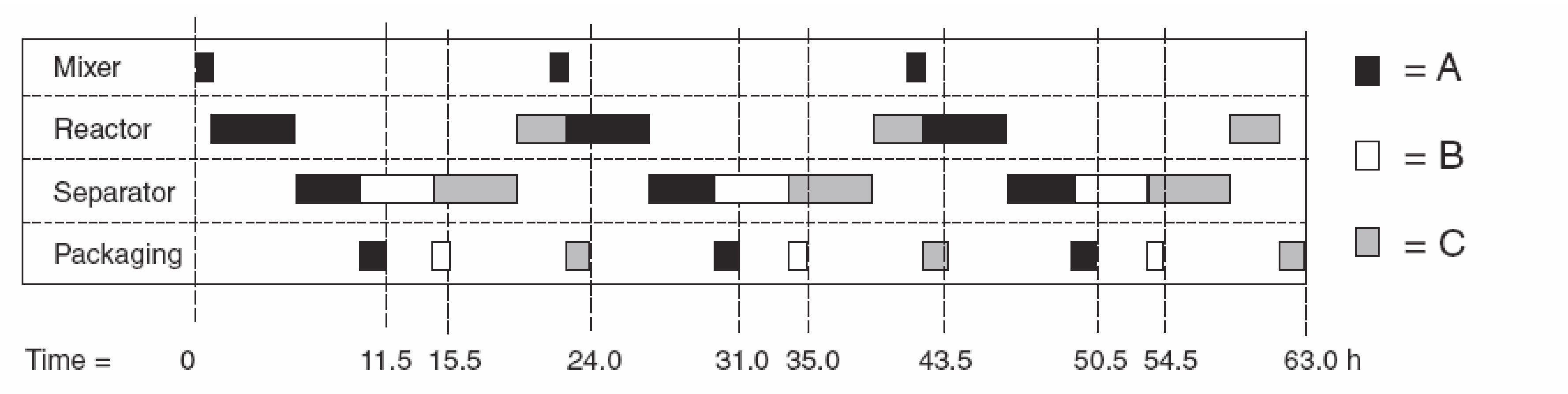


Jobshop Plants



Jobshop Plants

Jobshop Plants



Outline

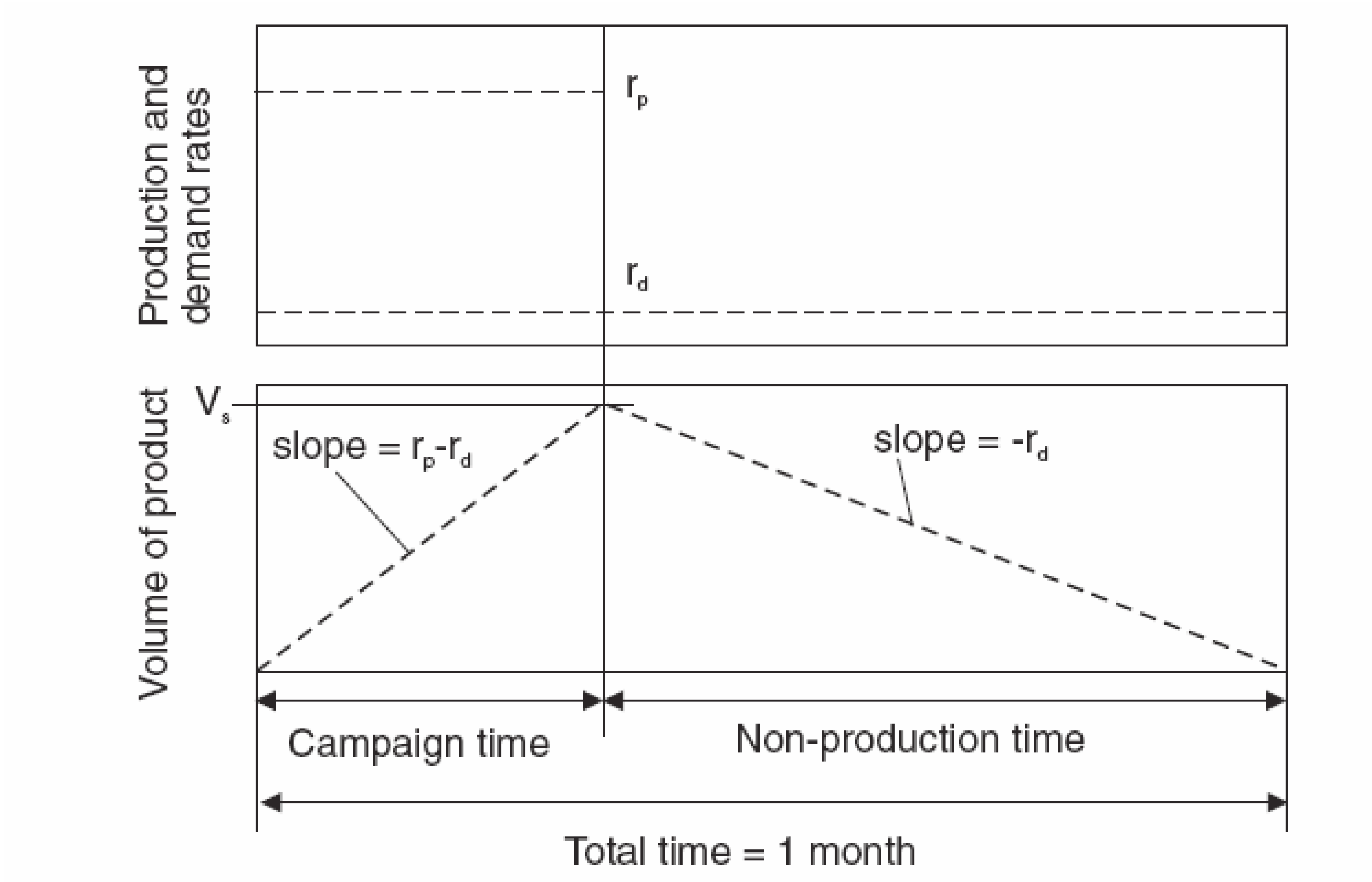
- Product and Intermediate Storage
- Parallel Process Units
- Equipment Design for Multiproduct Batch Processes

Product Storage

Product Storage for Single-Product Campaigns
When using combinations of single-product campaigns in a multi-product plant, it is necessary to store product during the campaign.

The amount of storage is dependent on the rate of production and rate of demand for each product and the cycle time.

Product Storage



Product Storage

Campaign time

$$V_s = \left[r_p - r_d \right] t_{ramp}$$

Rate of production

Rate of demand

Product Storage

Rate	Product A	Product B	Product C
Volume (m^3) of product required per month	10.0	15.0	12.0
Cycle time (h)	2.5*	4.5*	4.5*
Production rate, r_p (m^3/h)	(10)/[(43)(2.5)] = 0.0930	0.07752	0.06202
Demand rate, r_d (m^3/h)	(10)/(500) = 0.020	0.030	0.024

*These are approximate cycle times based on Equation (3.5).

For products A, B, and C, determine the minimum storage capacities for a single-product campaign strategy.

Assume that 1 month = 500 hours. Note that 43 batches of each unit are required for each product per month.

Product Storage

Intermediate Storage

For Multi-product Campaigns

So far, assumed no intermediate product storage available. This type of process is a **zero wait,** or a **zw-process**.

Flow to equipment directly from previous equipment in recipe.

Delay in one step propagates through batch to all steps.

Intermediate Storage

It may be possible to store product in the equipment that has just been used.
e.g. if two feed streams are mixed in a vessel, the mixture could be stored until the next process unit in the production sequence becomes available. This **holding-in-place** method may not work for some unit operations, e.g. in a reactor.

The upper limit of the intermediate storage concept occurs when there is **unlimited intermediate storage (uis)** available, and this is referred to as a **uis-process.**

Intermediate Storage

Limiting Case

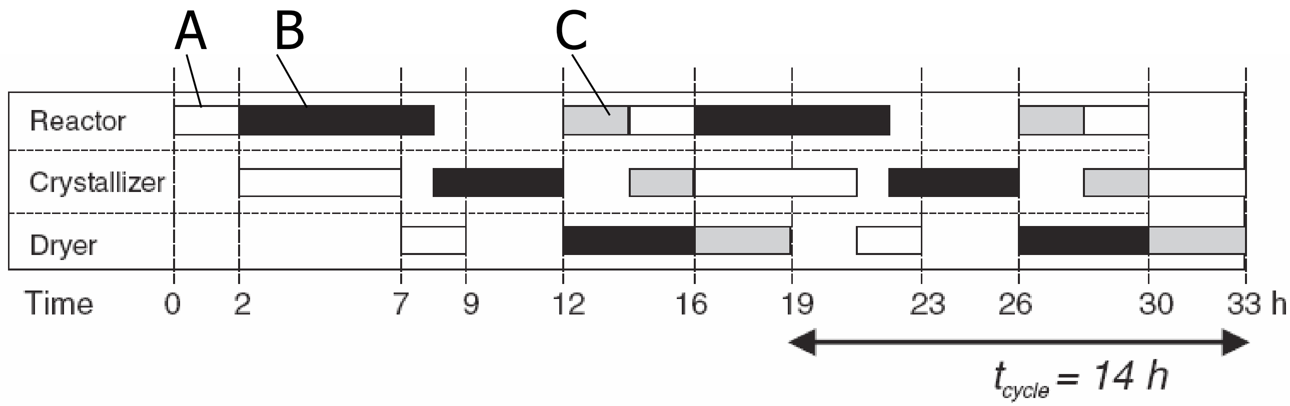
$$t_{cycle,uis} = \max_{j=1,m} \sum_{i=1}^{N} nc_i t_{i,j} \tag{3.8}$$

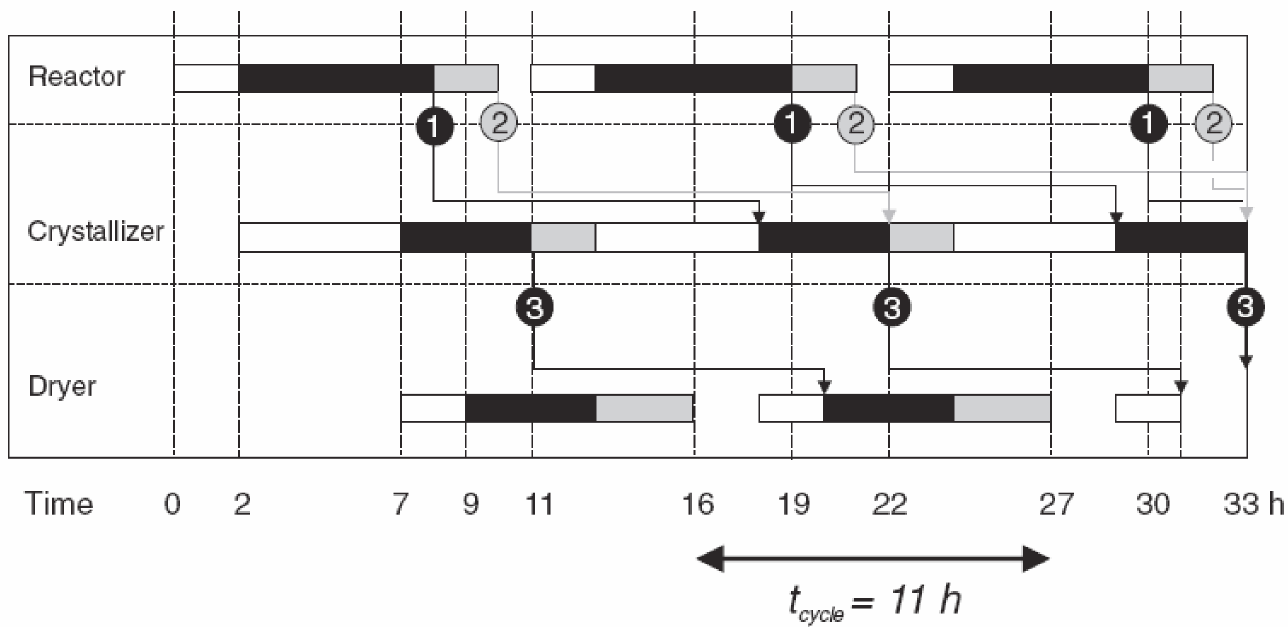
where m is the number of unit operations, N is the number of products, and nc_i is the number of campaigns of product i produced in a single multiproduct sequence. For the case shown in Table 3.2 and Figure 3.7, n =1 (because only one campaign for each product (A, B, and C) is used in the multiproduct sequence), and Equation (3.8) is the maximum value given in the last row of Table 3.2, or 11.0 h.

Table 3.2 Equipment Times (in Hours) Required for Products A, B, and C

Product	Reactor	Crystallizer	Dryer	Total
A	2.0	5.0	2.0	9.0
B	6.0	4.0	4.0	14.0
C	2.0	2.0	3.0	7.0
Total Time per Equipment	10.0	11.0	9.0	

Intermediate Storage





Outline

- Product and Intermediate Storage
- Parallel Process Units
- Equipment Design for Multiproduct Batch Processes

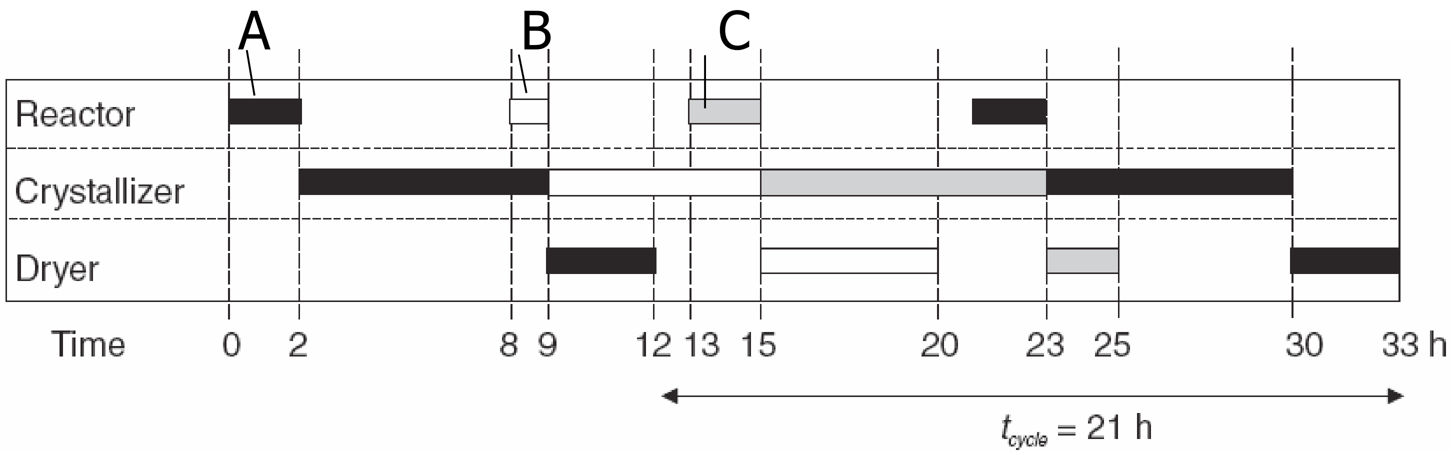
Parallel Process Units

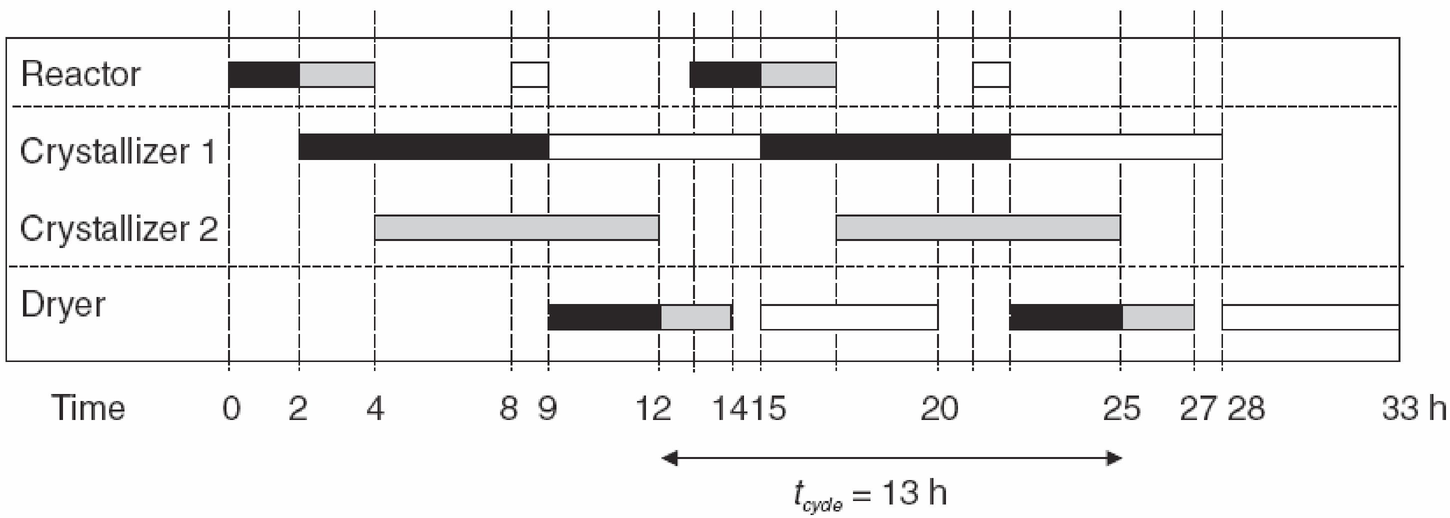
Reduce bottlenecks caused by a single piece of equipment by duplicating the equipment.

Trade-off between added equipment (and maintenance) cost vs. higher throughput.

Makes sense when one equipment dominates cycle time.

Parallel Process Units





Outline

- Product and Intermediate Storage
- Parallel Process Units
- Equipment Design for Multiproduct Batch Processes

Equipment Design for Multi-product Batch Processes

The design of equipment sizes for multiproduct batch processes depends on the

- Production cycle time
- Whether single- or multi-product campaigns are used
- The sequence of products for multi-product campaigns,
- Use of parallel equipment.

Equipment Design for Multi-product Batch Processes

Example

Process	Reactor and Mixer	Filtration	Distillation	Yearly Production	Production in 500 h
A	7.0 h	1.0 h	2.0 h	120,000 kg	10,000 kg
B	9.0 h	1.0 h	1.5 h	180,000 kg	15,000 kg
C	10.0 h	1.0 h	3.0 h	420,000 kg	35,000 kg

Clearly the reactor is the limiting piece of equipment – determine the size of the reactor required.

Equipment Design for Multi-product Batch Processes

From the recipe for each batch, calculations for the specific volume of each reactor can be made – see Chapter 3.

Process	A	B	C
v_{react} (m^3/kg-product)	0.005507	0.007860	0.006103
t_{cycle} (h)	7.0	9.0	10.0

Equations required

total time taken to produce A

Total time for batches $t_A + t_B + t_C = 500$

Reactor volumes $V_A = V_B = V_C$

Chapter 6 – Process Conditions

Chemical Engineering Department
West Virginia University

Heuristics

- Temperature

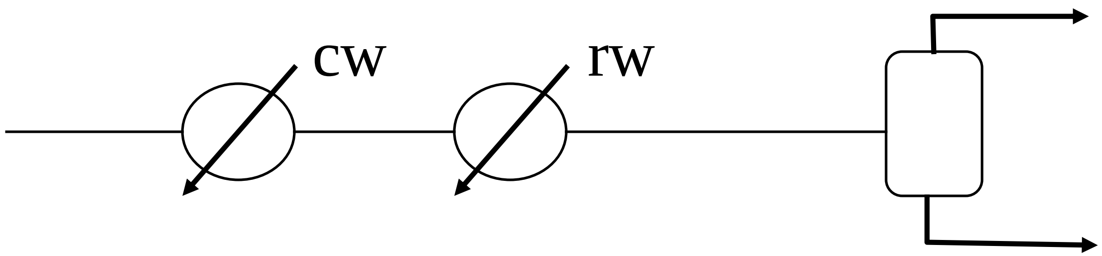
< 40ºC	> 250ºC	> 400ºC
Require Refrigeration	Require Fired Heater	Special M.O.C

- Pressure

< 1 atm	> 10 atm
need Vacuum	Thick Walls - $

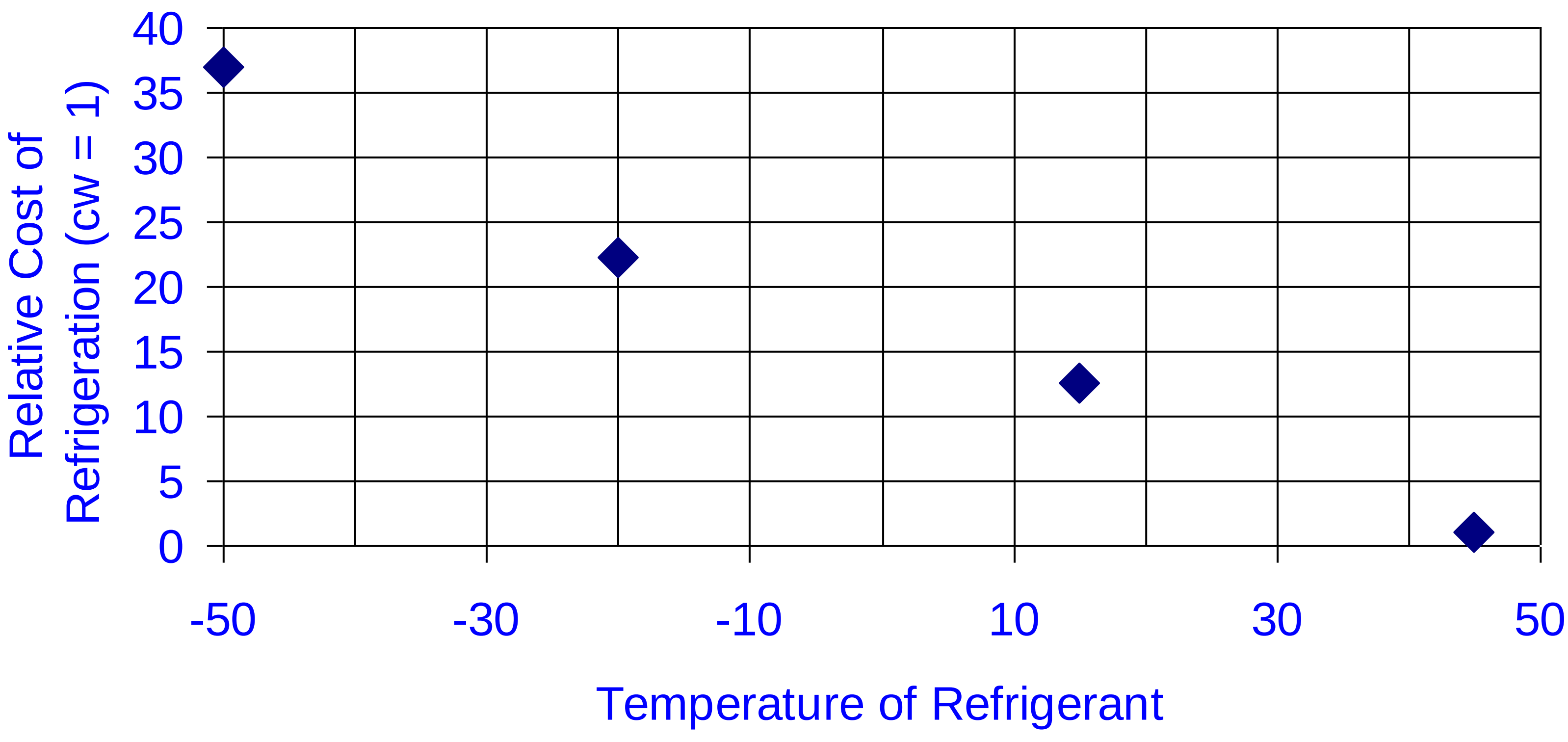
Temperature

- *T* < 40º C – Refrigeration
  - Use as much cooling water as Possible



  - Operating Costs (Table 8.3)
    - Cooling Water (30-40°C) $0.354/ GJ
    - Refrigerated Water (5-15°C) $4.43/ GJ

Temperature



Temperature

- $T > 250°C$ – hp steam @ 260°C (600 psig)
  - Need a molten salt / Dowtherm loop
    - Fired Heaters are very expensive
  - Compare - vaporizer

$$Q = 10{,}000 \text{ kW}$$

$$U = 1000 \text{ W/m}^2\text{K}$$

$$\Delta T = 30°C$$

$$A = \frac{(10 \times 10^6)}{(10^3 \times 30)} = 333 \text{ m}^2$$

Temperature

- C_{BM}
  - Heat Exchanger = $ 1.70×10^5
  - Fired Heater = $ 1.81×10^6
- $T > 400°C$
  - M.O.C. is very Important

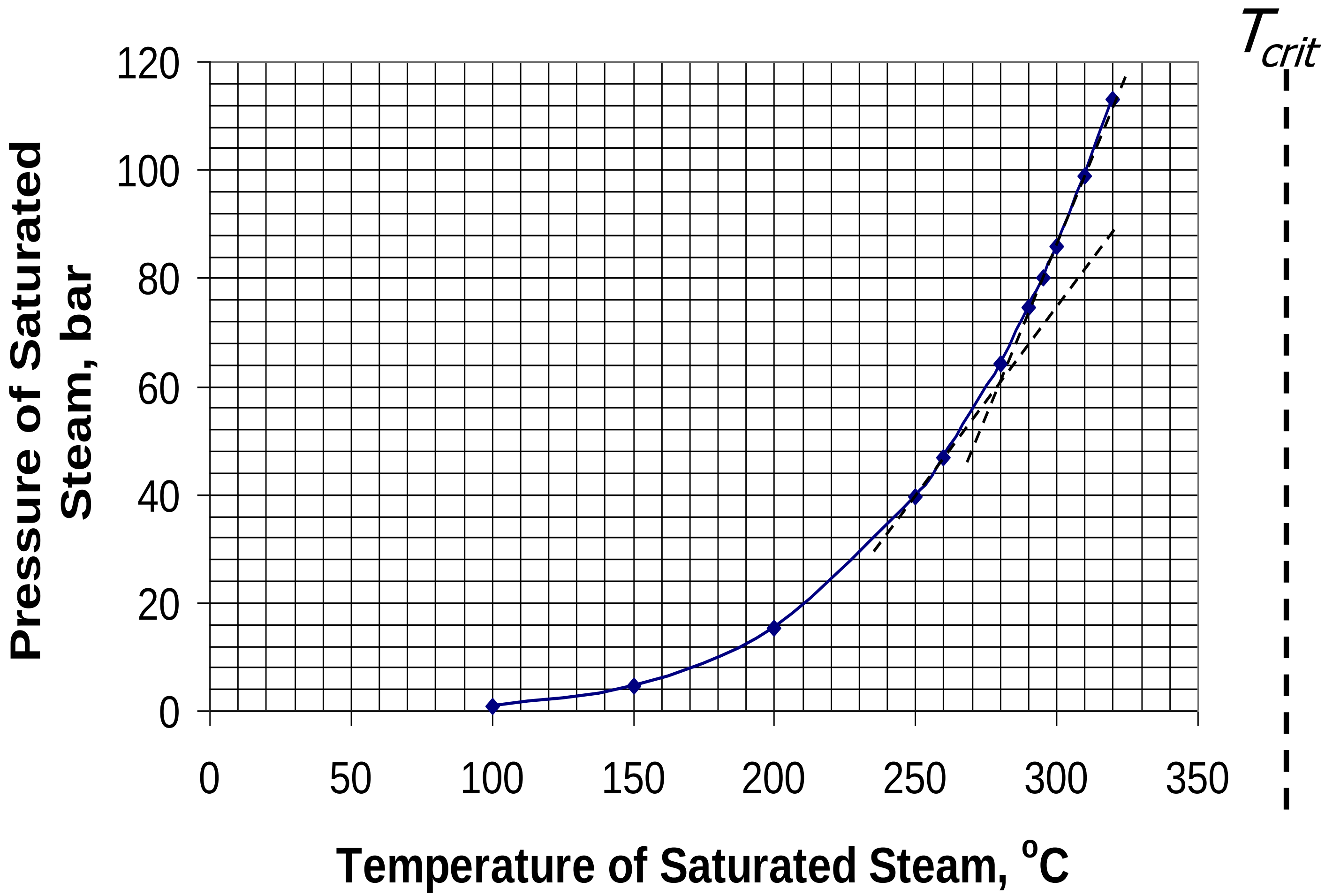
Why not use High-Pressure Steam?

P_{sat} (bar)		T_{sat} (ºC)
15.2		200
39.7		250
46.9		260
64.2		280
86.0		300
74.5		290
80.1		295
98.8		310
113.0		320

Graph of Saturated Steam vs. Pressure

From *Analysis, Synthesis, and Design of Chemical Processes, Fourth Edition,* by Richard Turton, Richard C. Bailie, Wallace B. Whiting, Joseph Shaeiwitz, and Debangsu Bhattacharyya
(ISBN-13: 978-0-13-261812-0)

Why not use High-Pressure Steam?



Pressure

- Vacuum
  - Slightly Higher Cost due to Stiffening Rings
  - Large Equipment
  - Air Leaks
- High Pressure
  - Thick Walls - $
  - H_2 Embrittlement
  - Safety

Minimum Wall Thickness

$$t = \frac{PR}{SE - 0.6P} + CA$$

Wall thickness, *t* = m , design pressure, *P* = bar , vessel radius, *R* = m

S = Design Stress (Max Allowable Working Pressure, bar) this is a function of material and temperature

E = Weld Efficiency(~ 0.9)

CA = Corrosion Allowance (0.00315 to 0.00625 m)

Minimum Wall Thickness

- Look at 36 inch Diameter Vessel with a CA of ¼ in made of CS with *S* = 13,700 psi

P	*t* (m)	*t* /*CA*
14.7 = 1 barg	0.0069	1.09
58.8 = 4 barg	0.0085	1.34
147 = 10 barg	0.0118	1.86

- As *P* > 10 then *t* >*CA*

What About S vs. T ?

- Look at Several Steels in Graph
- For CS $S \downarrow$ as $T > 400$ºC
- Must use Stainless Steel and $\$\uparrow$
- For a given Pressure
  - $t \uparrow$ as $T \uparrow$

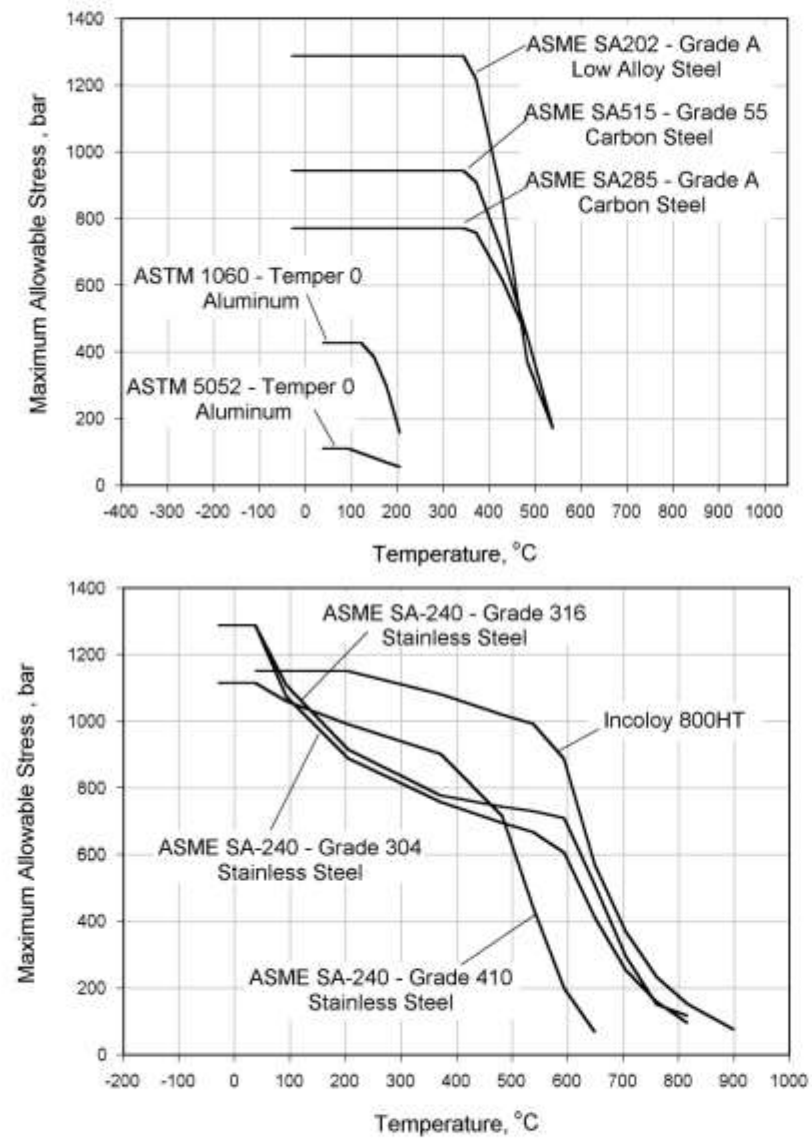
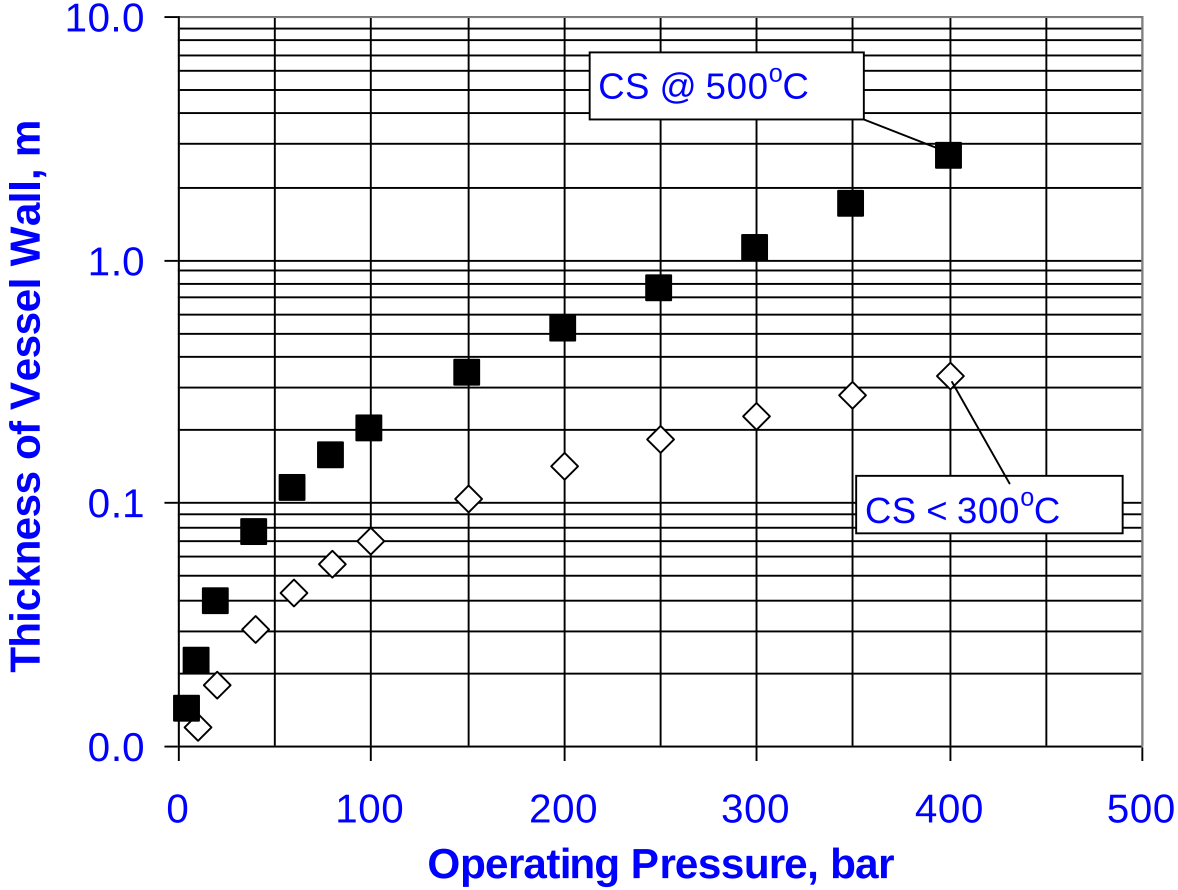


Figure 5.5: Maximum Allowable Stresses for Materials of Construction
as a Function of Operating Temperature
(Data from Perry et al. [3], Chapter 10 and Ref [15])

Material of Construction

1 m Diameter Vessel made of SA285 - Grade A Carbon Steel



From *Analysis, Synthesis, and Design of Chemical Processes, Fourth Edition,* by Richard Turton, Richard C. Bailie, Wallace B. Whiting, Joseph Shaeiwitz, and Debangsu Bhattacharyya (ISBN-13: 978-0-13-261812-0)

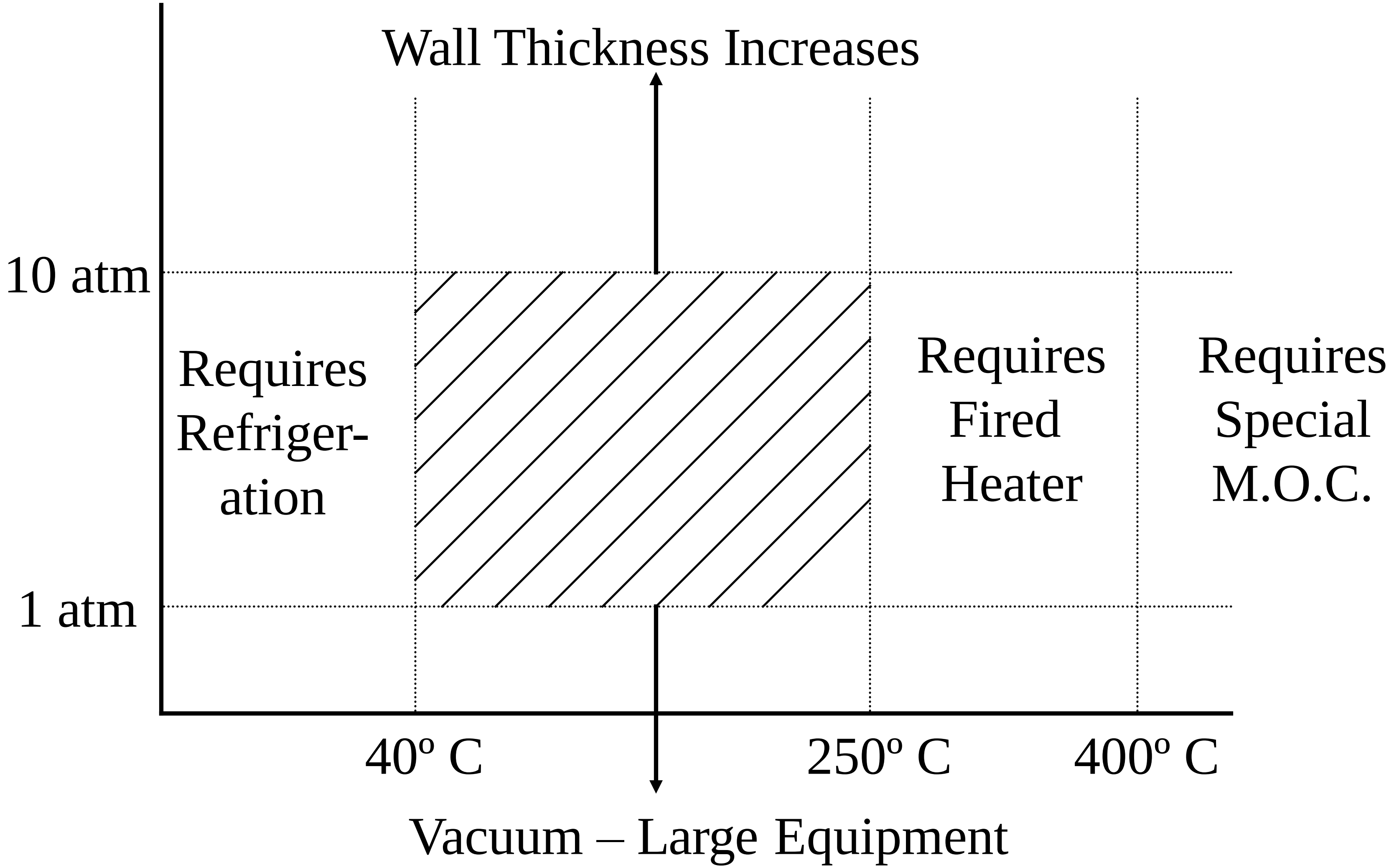
Material of Construction

- Carbon Steel
  - Cheap
- Stainless Steel
  - Expensive
  - Better Chemical/Thermal Resistance
- What About T = 700-900°C?
  - Insulate inside of Pipe
  - Metal – Refraction Lining

Conclusions

- $T < 40°C$ – Refrigeration
- $T > 250°C$ – Fired Heater or Furnace
- $T > 400°C$ – M.O.C. Issues
- $P < 1$ atm – Vacuum and Large Equipment
- $P > 10$ atm – Cost

Operating Conditions



Do we ever operate outside these limits?

- Tables 6.1 – 6.3
  - Reactors and Separators
- Table 6.4
  - Other Equipment

Examples

- Example 1 – Acrylic Acid
  - Appendix B.9
  - Why does T-305 Operate with the top pressure at 0.07 bar?
  - Feed – 86.6 kmol/h Acrylic Acid – nbp =140°C

    6.1 kmol/h Acetic Acid – nbp = 118°C

  Table 6.2 – Reasons for using *P* < 1 atm

  1. Obtain a gas phase for VLE

  2. Temperature sensitive materials

Examples

- Example 2 – Separation of Propane
  - Typical depropanizer operates at 220 psig (16 bar) – why?

  Table 6.2 – reasons for using P > 10 bar

  1. Obtain a liquid phase for VLE

Chapter 15 - Heat Exchange Networks

Chemical Process Design

West Virginia University

Outline

- Heat Integration
- Design Procedure for MUMNE
  - Temperature interval diagram
  - Cascade diagram
  - Temperature-Enthalpy diagram
  - Minimum number of exchangers
  - Design above and below pinch

Heat Integration

- Heat exchange networks
- It saves money to match streams rather than pay to heat one and pay to cool another
- You have already done this on *ad hoc* basis in design projects

Heat Integration

- There is a rigorous methodology
- We will learn MUMNE (***M**inimum **U**tility, **M**inimum **N**umber of **E**xchangers)* method
- Not necessarily (and unlikely to be) economic optimum

Design Procedure

1. Complete energy balance on all streams to determine all temperatures, $\dot{m}C_p$ values, and heat flows.
2. Choose minimum approach temperature. Typically, this is between 5° C and 20° C, but any positive number is valid.
3. Complete temperature interval diagram, Each stream is drawn and labeled. The heat flow in each interval is calculated.

Design Procedure

4. Complete the cascade diagram. The energy excess or deficit is calculated for each interval on the temperature interval diagram.
5. Find the minimum hot and cold utility requirements and identify the pinch temperature.
6. Complete the composite temperature enthalpy diagram. This is a *T-Q* diagram for the entire process.

Design Procedure

7. Determine the minimum number of heat exchangers required above and below the pinch.
8. Design the heat exchanger network.

Example Problem

Stream Properties for Example Problems

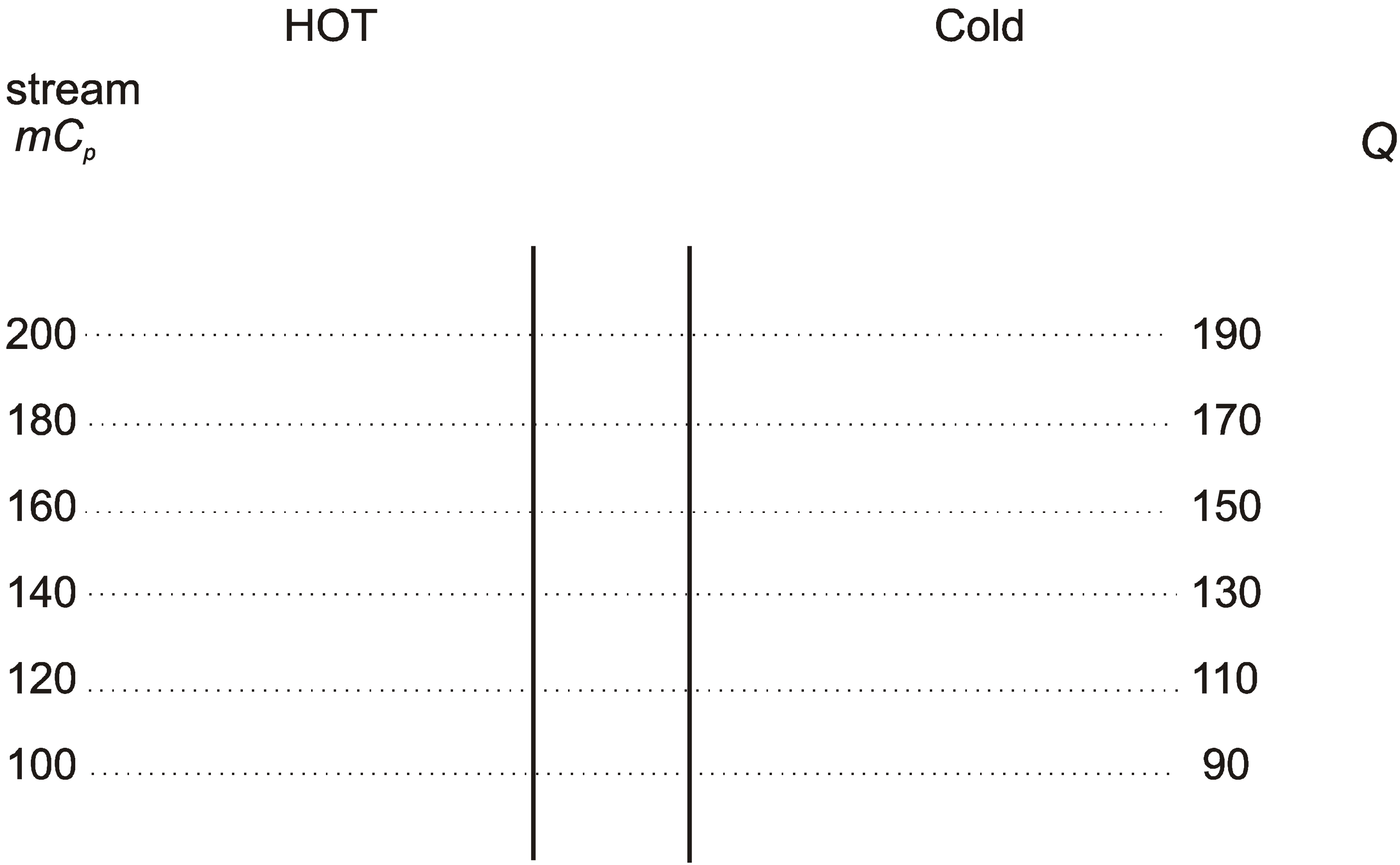
Stream	T_{in}	T_{out}	$\dot{m}C_p$ kW/° C	Q *kW*
1	200	120	3	240
2	140	100	5	200
3	100	170	3	-210
4	110	190	2	-160
Net heat flow				70

Example Problem

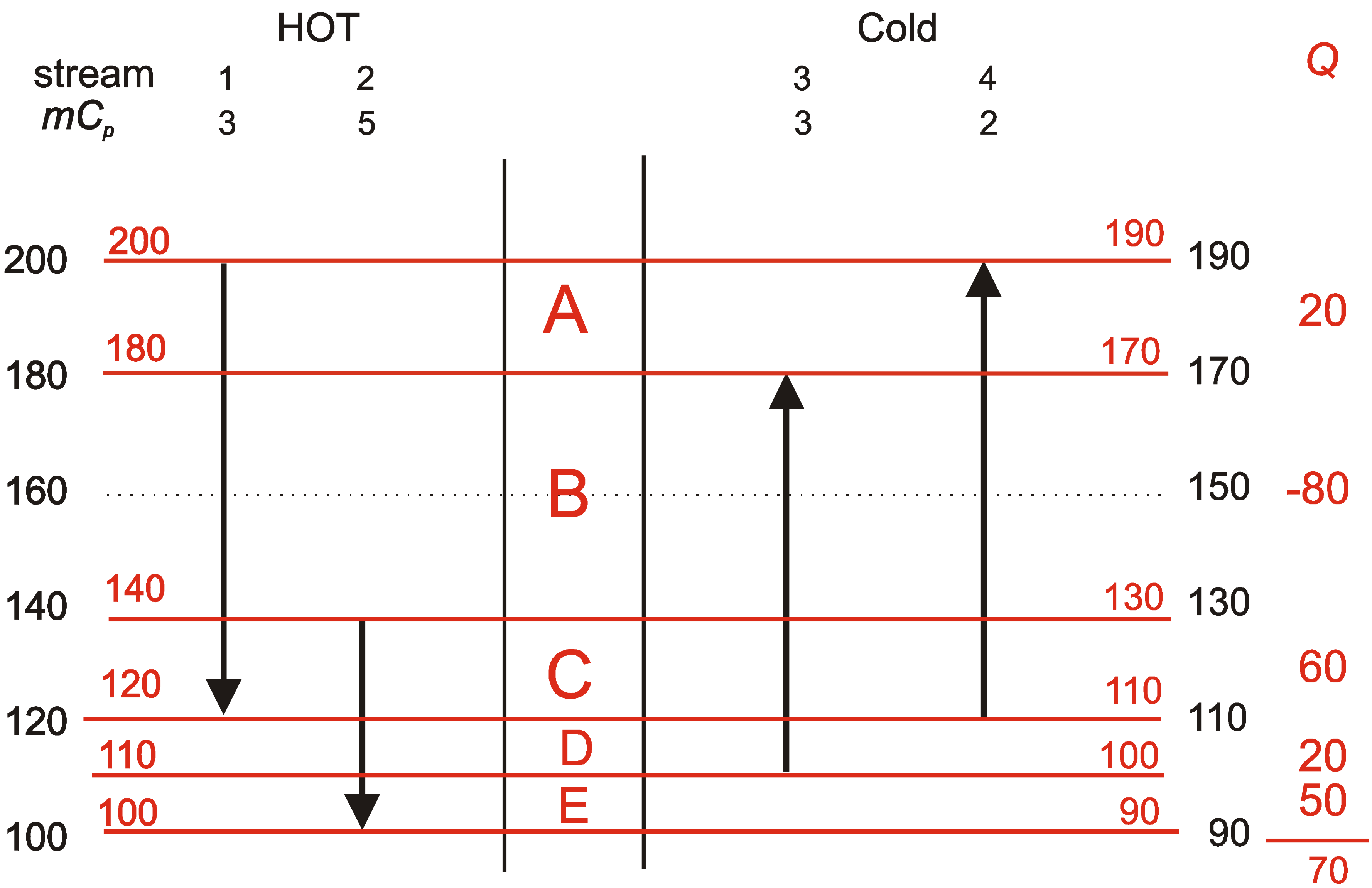
1. The value of Q might not be given in above table, or Q is given and $\dot{m}C_p$ is missing. These are calculated from the energy balance. The sign convention is positive for heat available from a stream and negative for heat needed by a stream.
2. Choose the minimum approach temperature. For this problem, it is 10° C.

Example Problem

3. Draw and label the temperature interval diagram. Label the intervals beginning with "A" for the highest temperature interval. The heat flow for each interval is calculated from, $Q = \sum \dot{m} C_p \Delta T$, where the sum is over all streams existing in that interval.


HOT
Cold
stream
mC_p
Q
200
180
160
140
120
100
190
170
150
130
110
90

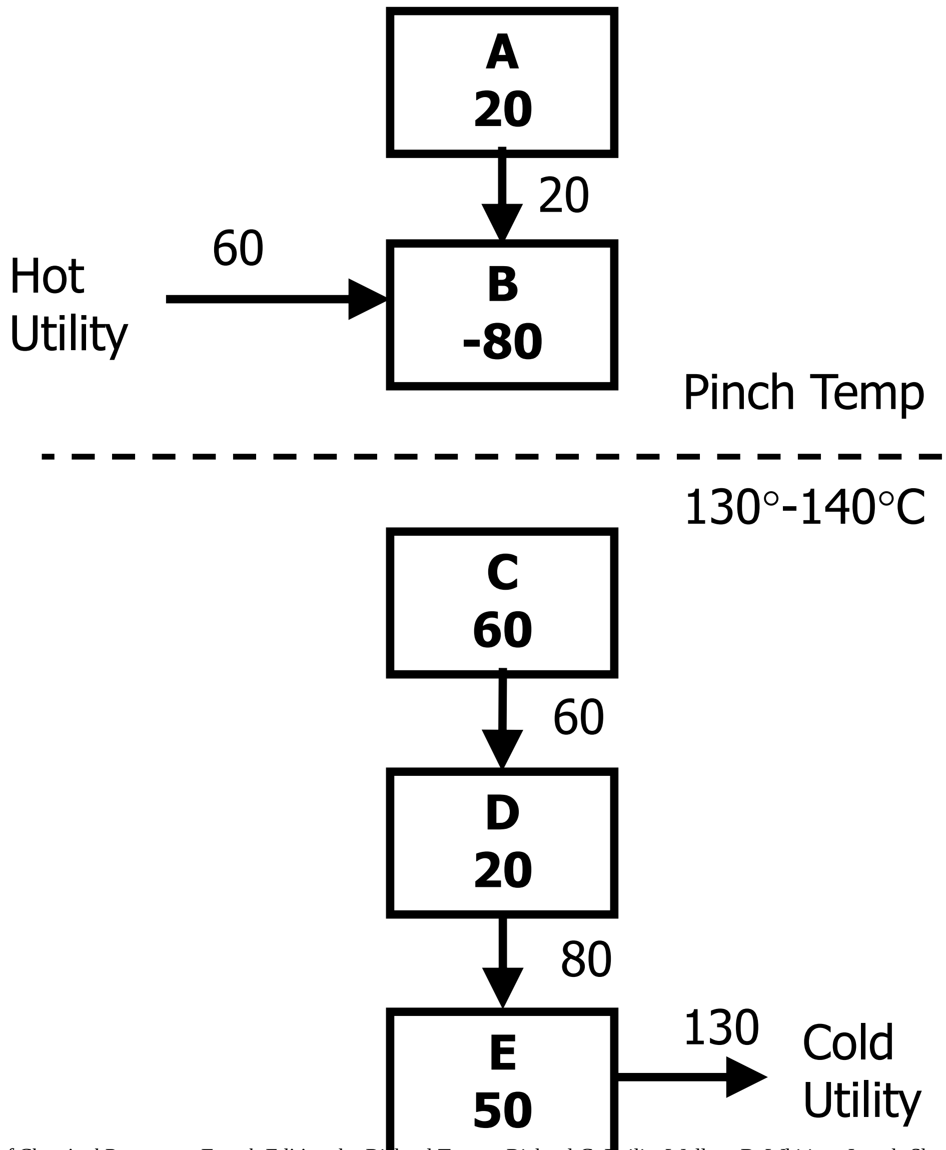
Temperature Interval Diagram



Example Problem

4. Draw the cascade diagram. This represents the cascade of heat flowing down from high to low temperatures. Add utilities where needed. Label the heat flows. The net utility flow should agree with the net heat flow on the earlier table.
5. On the cascade diagram, there will be a location where the heat-flow cascade is not continuous. This represents the pinch temperature

Cascade Diagram



Example Problem

6. Construct the composite temperature enthalpy diagram. This provides useful information, but it is not required to solve the problem.

Example Problem

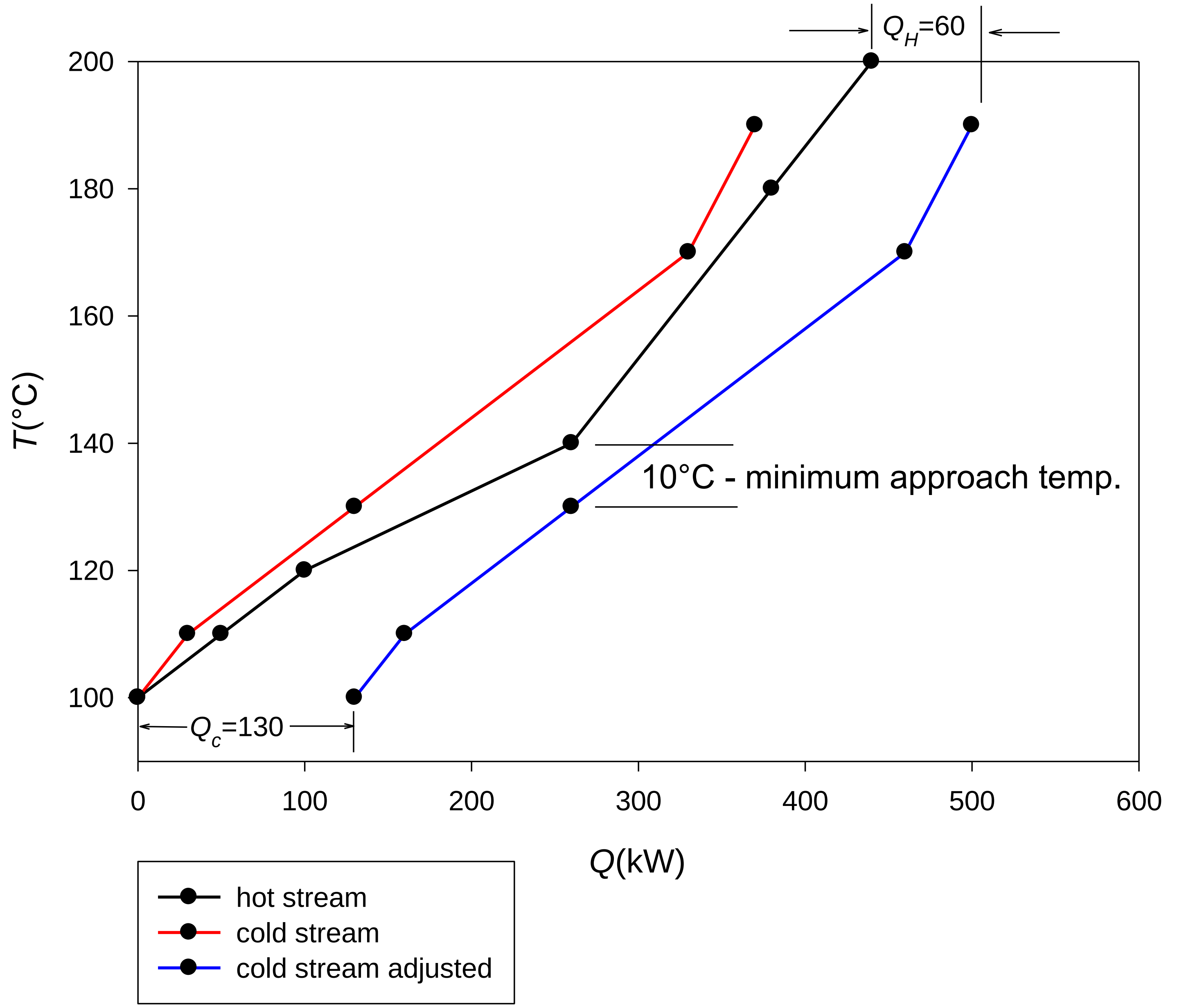
	Hot		Cold	
interval	*T* (°C)	*Q* (kW)	*T* (°C)	*Q* (kW)
E	100		90	
D	110	50	100	
C	120	100	110	30
B	140	260	130	130
A	180	380	170	330
	200	440	190	370

Example Problem

In the table, the temperature shown is at the lower end of the interval. The Q values are obtained by summing all $\dot{m}C_p\Delta T$ existing on the interval and adding it to the previous interval. The temperature difference is for that interval. The $\dot{m}C_p$ value is the sum of all existing streams on that interval.

Example Problem

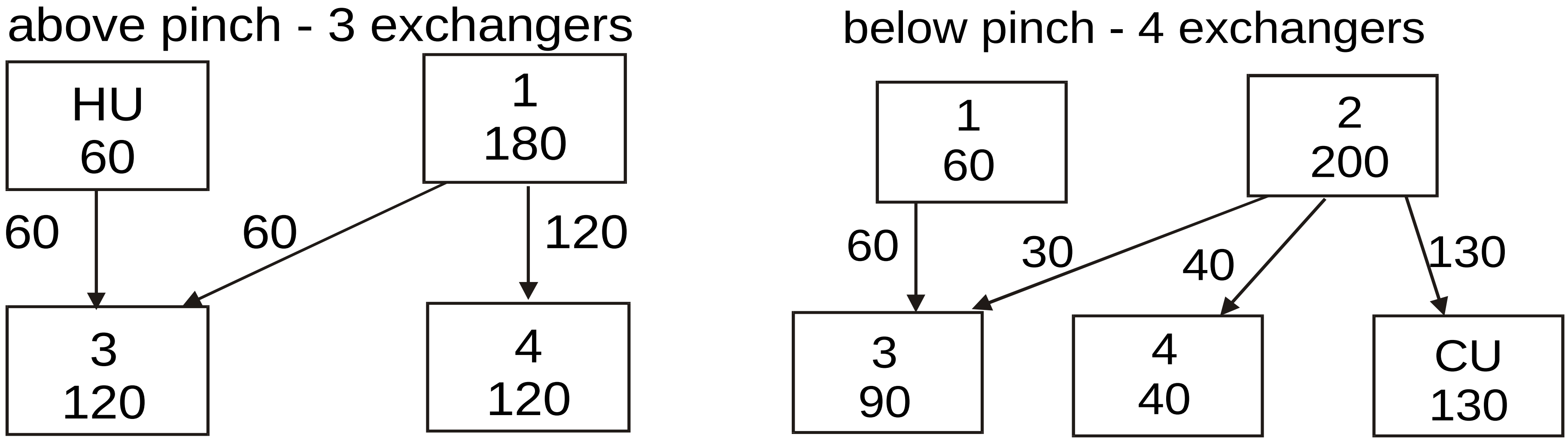
The hot and cold stream lines are plotted, as shown on the following figure. Clearly, there is a temperature cross, so the cold stream line is shifted to the right until the minimum approach temperature of 10° C exists at one point. (It could exist at more than one point by coincidence.) For this problem, all *Q* values for the cold stream must be increased by 130 kW, as shown in the figure. Note how the hot and cold utility requirements are apparent from the diagram.



Composite *T*-*Q* Diagram

Example Problem

7. By representing the heat available in each stream and from the utilities both above and below the pinch, the minimum number of heat exchangers can be determined. This identifies the minimum number, but not necessarily the correct stream matches. The correct number of heat exchangers is the number of process streams + the number of utility streams – 1.



each arrow identifies one heat exchanger,
total number of arrows is total number of hxs.,
but not necessarily the correct stream matches

numbers in boxes are energy in streams
numbers with arrows are energy transferred in
a heat exchanger

Example Problem

7. Note that if a "direct match" is found, *i.e.*, where sets of two streams match heat flows exactly, one fewer exchanger may appear to be possible. However, be careful, the minimum approach temperature may be violated.

Example Problem

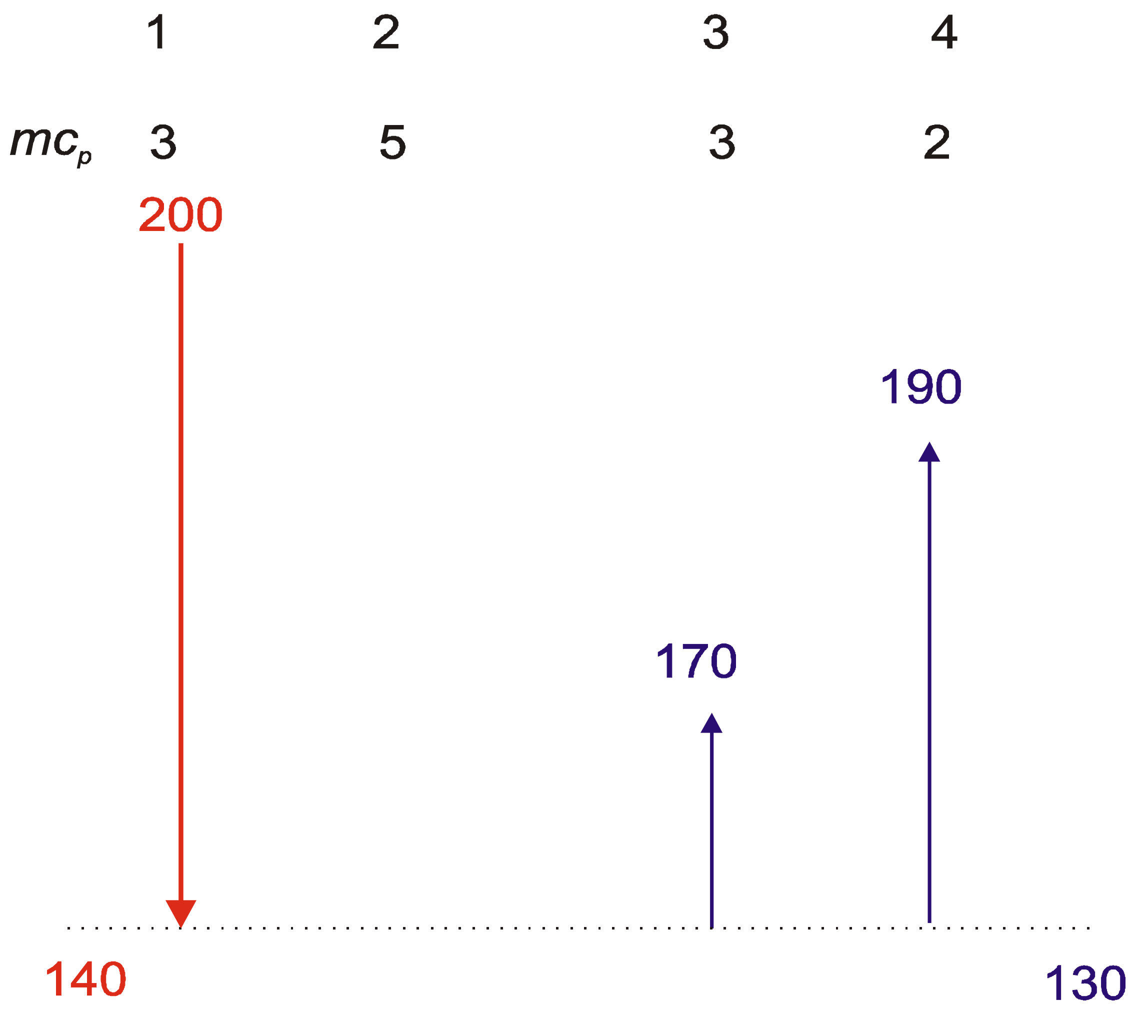
8. Design the heat exchange network. There may not be unique streams here. The design is started at the pinch and you work away from the pinch. Above the pinch, for any streams that exist at the pinch, streams can only be matched such that

$$\dot{m}C_{p_H} \leq \dot{m}C_{pC}$$

Example Problem

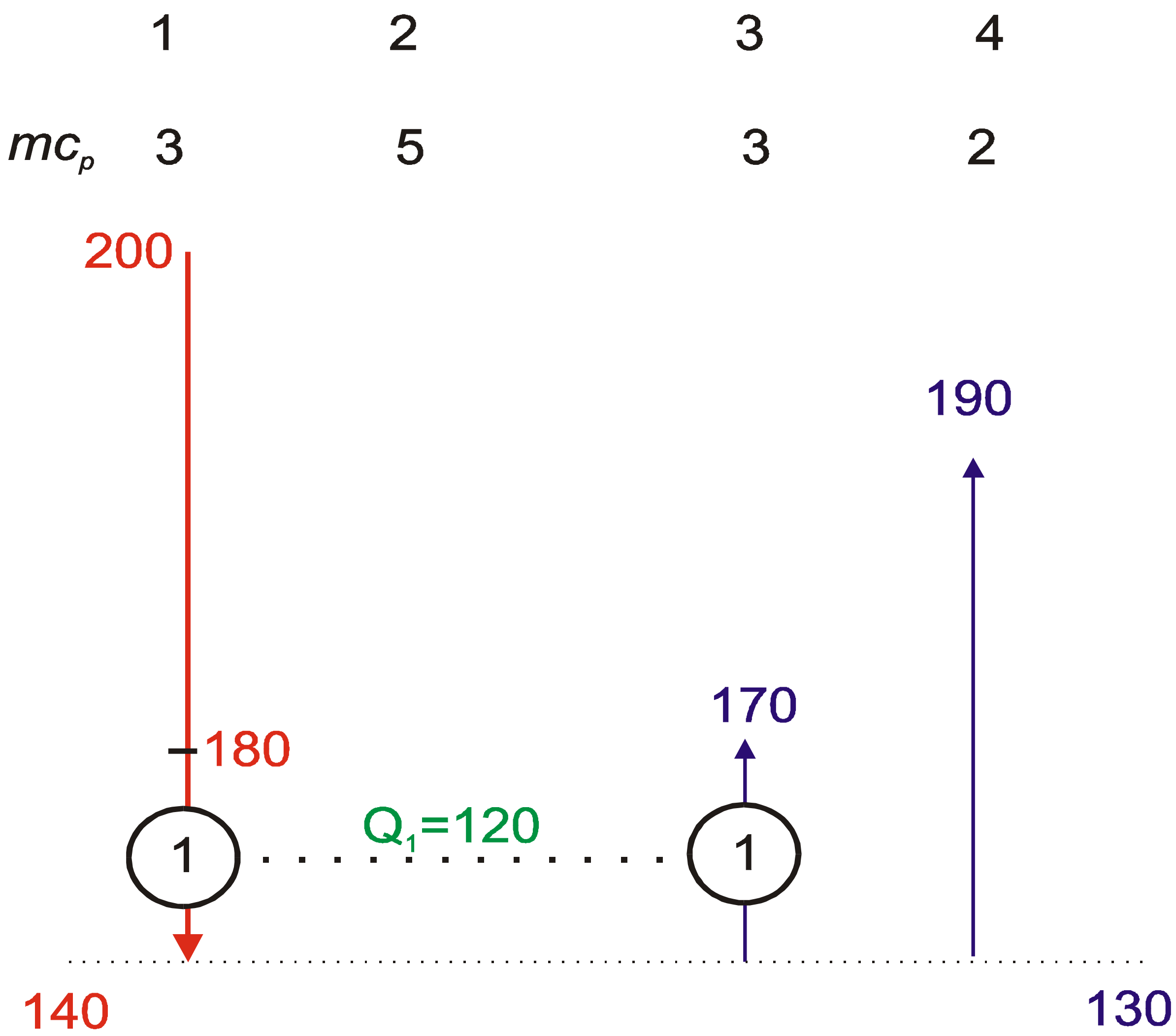
8. When dealing with streams away from the pinch, this criterion is no longer needed. Any streams can be matched as long as the temperatures are valid. If the criterion at the pinch appears impossible to satisfy, streams can be split to satisfy the criterion.

Exchanger Network - Design Above Pinch



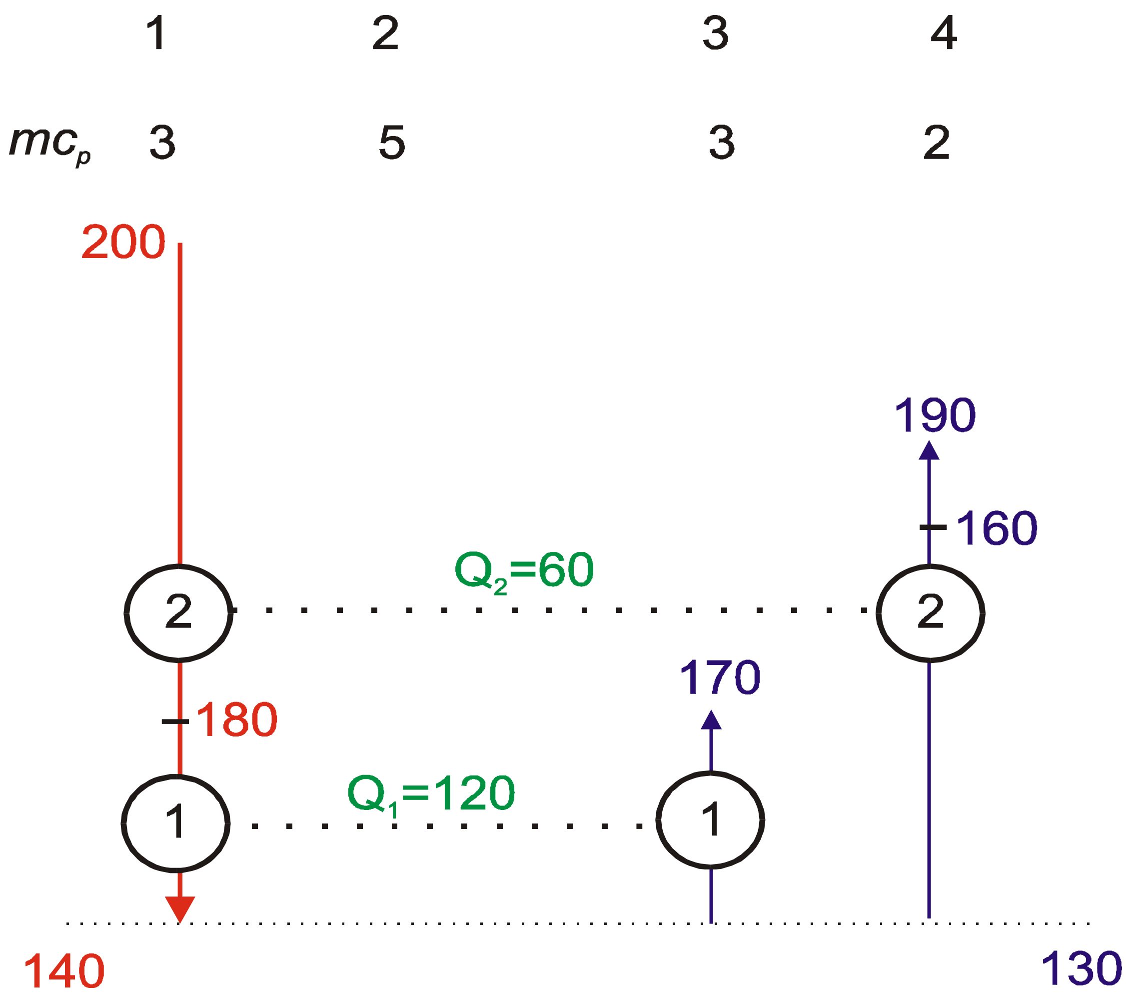
Above Pinch
at pinch match $mc_{pH} \leq mc_{pC}$

Exchanger Network - Design Above Pinch



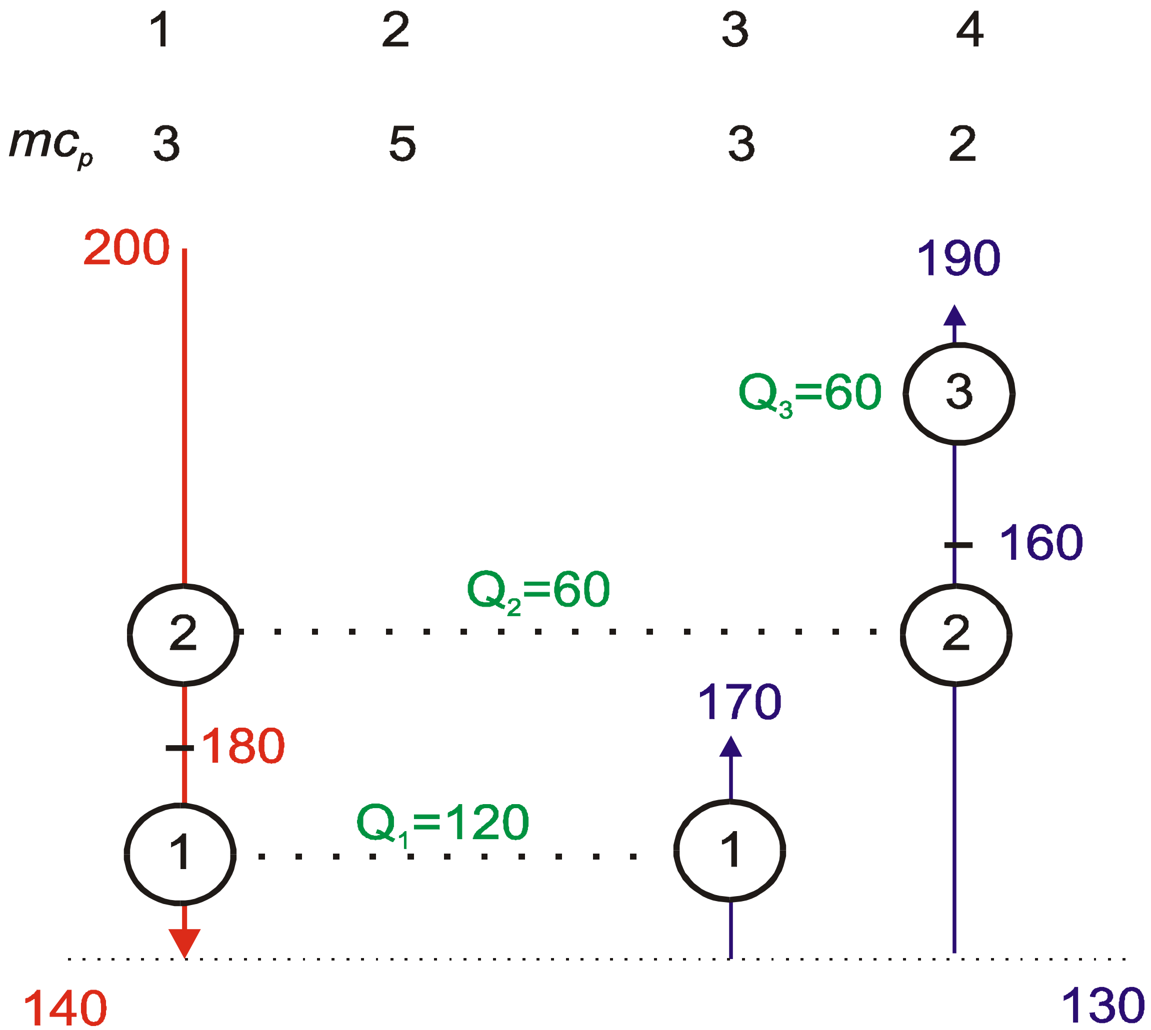
Above Pinch
at pinch match $mc_{pH} \leq mc_{pC}$

Exchanger Network - Design Above Pinch



Above Pinch
at pinch match $mc_{pH} \leq mc_{pC}$

Exchanger Network - Design Above Pinch



Above Pinch
at pinch match $mc_{pH} \leq mc_{pC}$

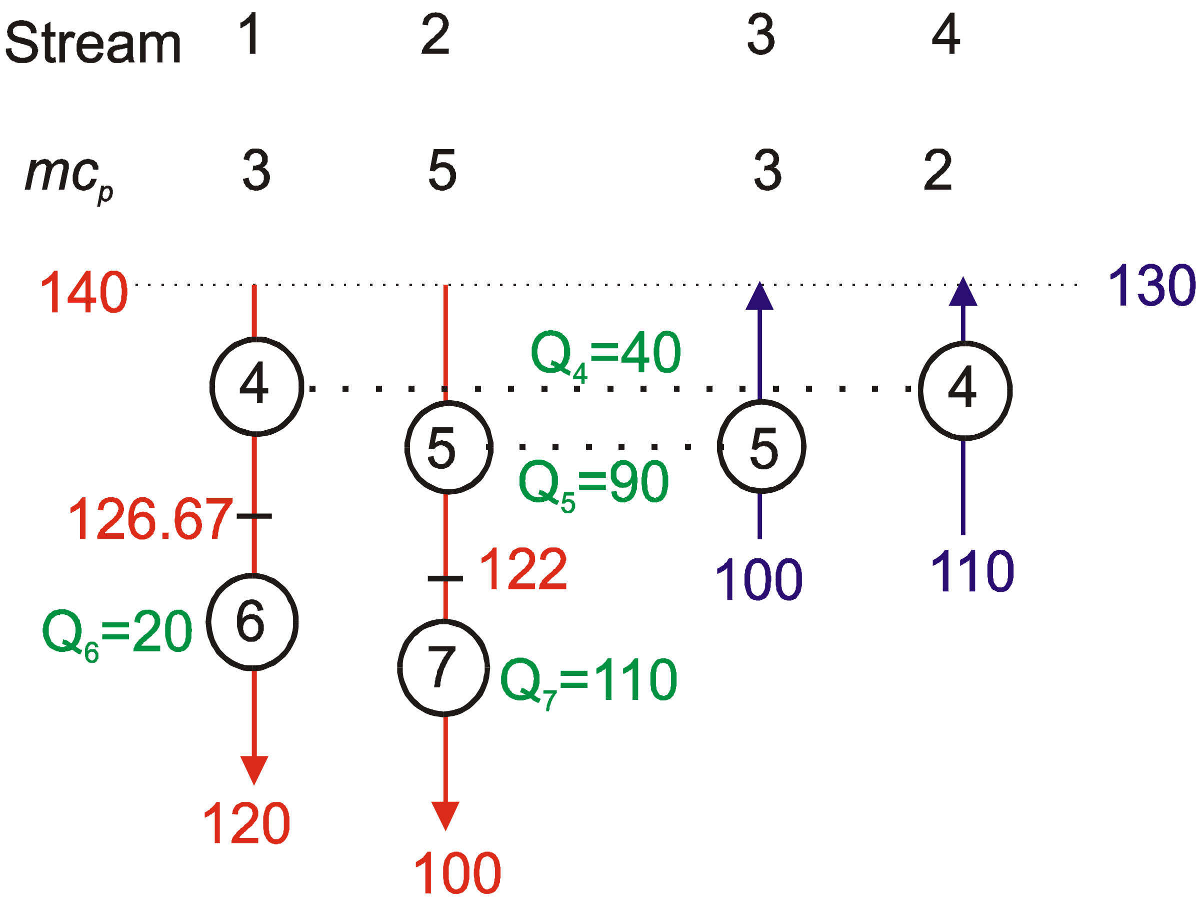
Example Problem

8. The same procedure is done below the pinch, except that the criterion is

$$\dot{m}C_{pH} \geq \dot{m}C_{pC}$$

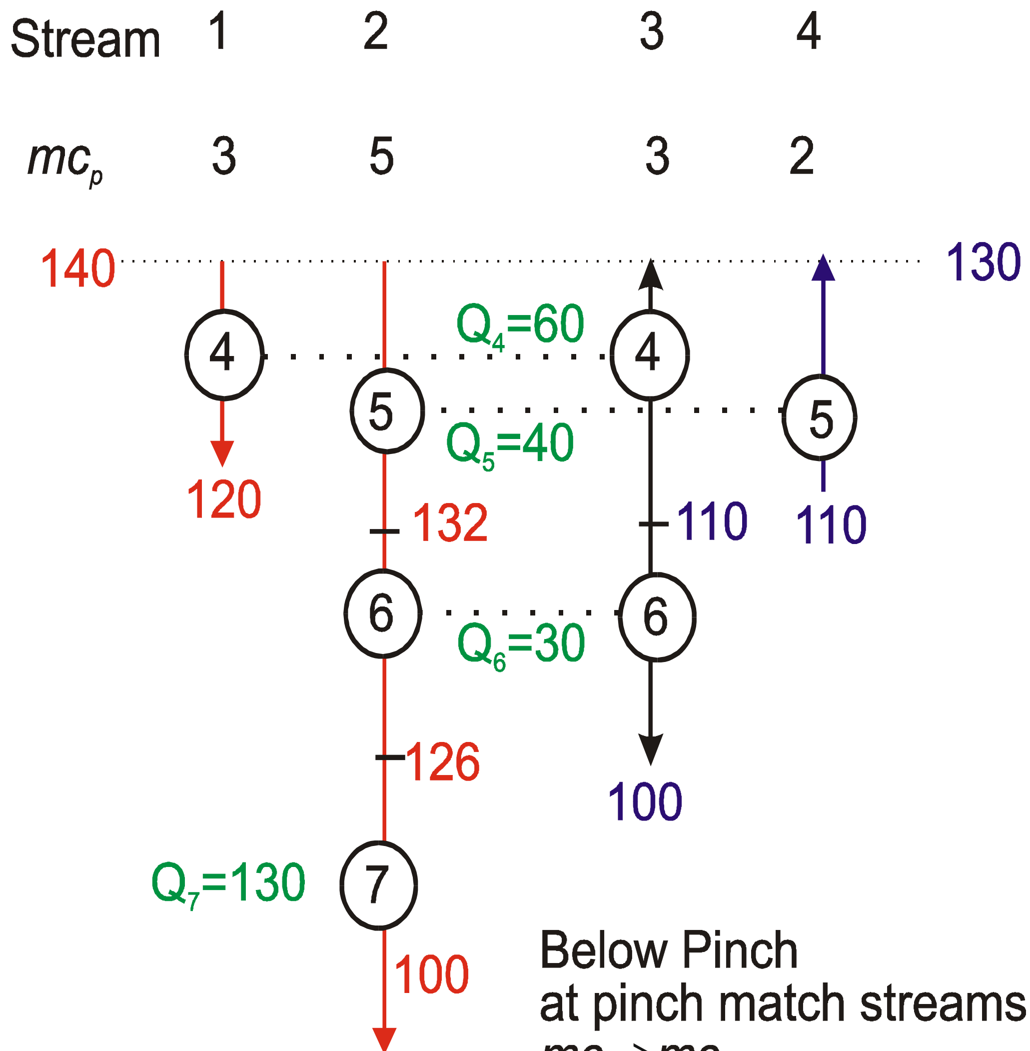
9. Streams are matched and heat exchangers are added until all required heat transfer is accomplished. The entire network, both above and below the pinch, can then be represented on one diagram.

Exchanger Network - Design Below the Pinch



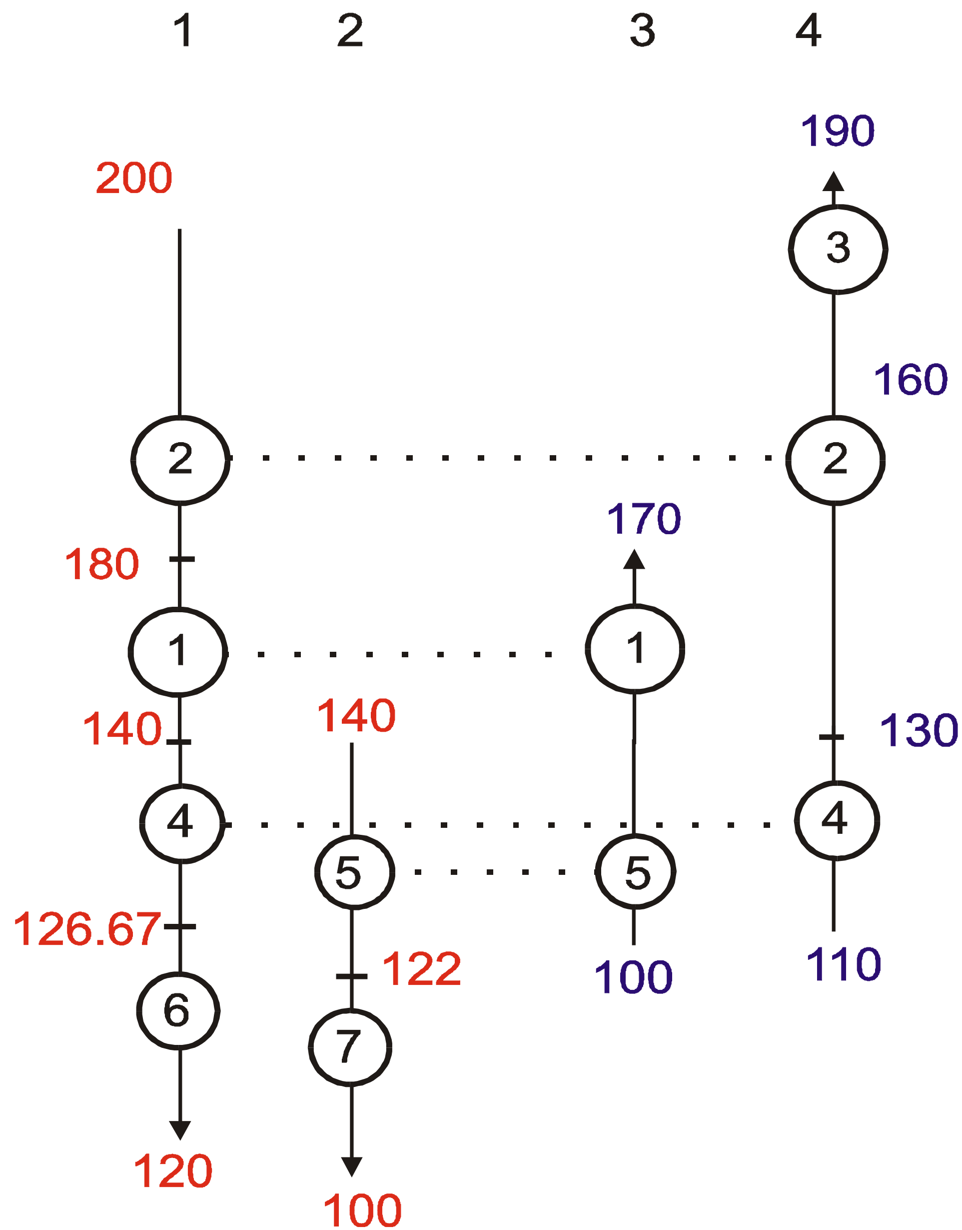
Below Pinch
at pinch match streams at pinch
$mc_{pH} \geq mc_{pC}$

Exchanger Network - Alternate Design Below the Pinch



Below Pinch
at pinch match streams at pinch
$mc_{pH} \geq mc_{pC}$

Overall Network



In-class Example Problem

Stream Properties for In-class Example Problem

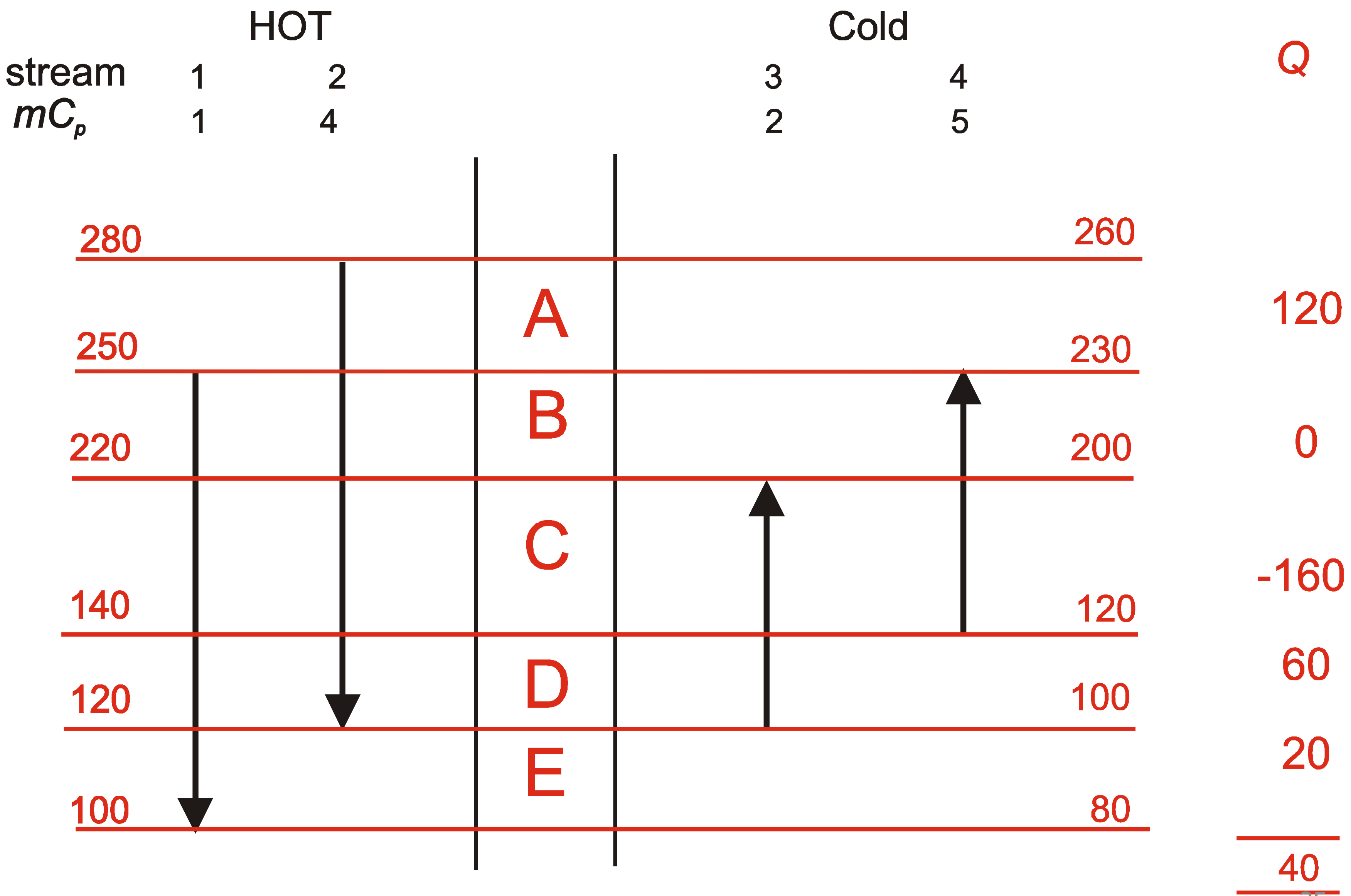
Stream	T_{in}	T_{out}	$\dot{m}$ kg/s	C_p kJ/kg°C
1	250	100	1	1
2	280	120	2	2
3	100	200	1	2
4	120	230	1	5

In-class Example Problem

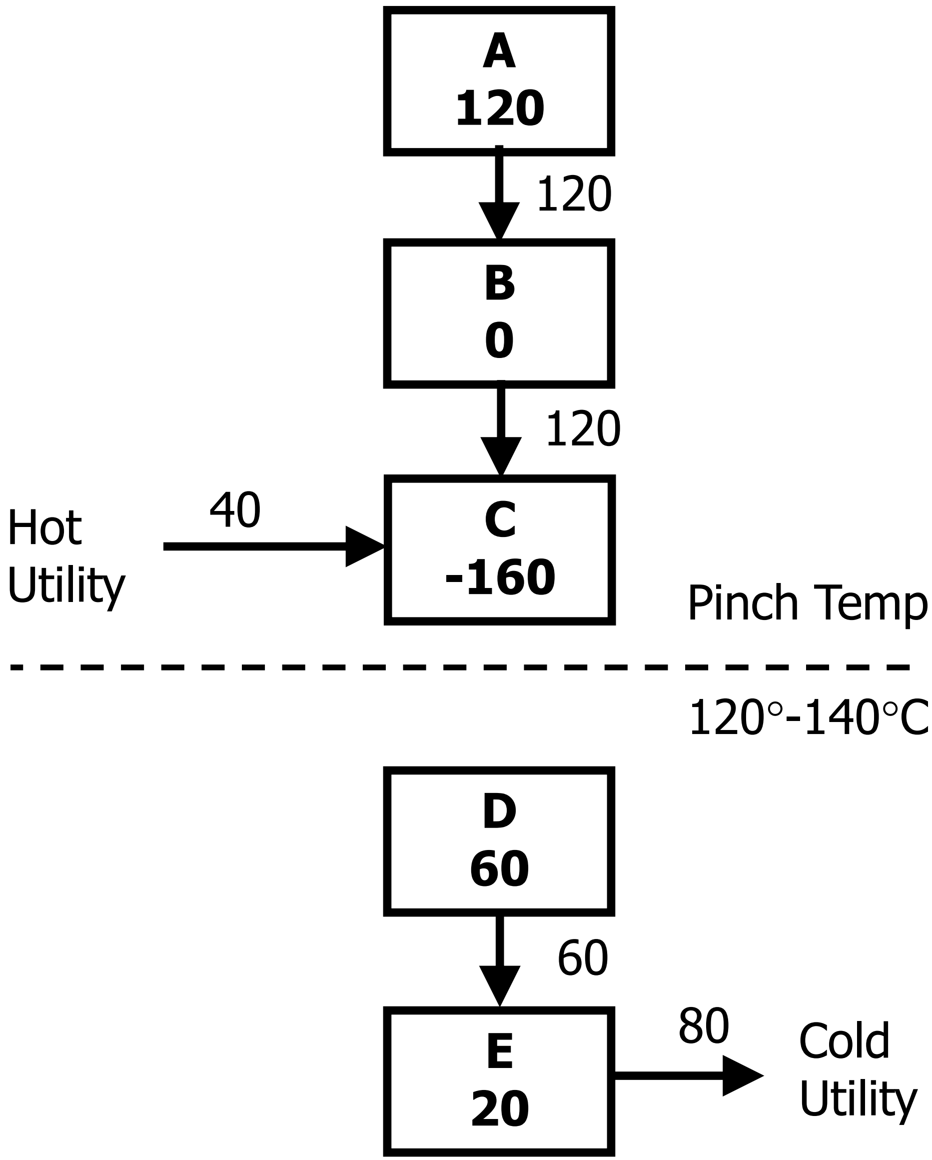
Determine (minimum approach $T = 20°$ C)

a. minimum hot and cold utility consumption
b. pinch temperatures
c. minimum number of heat exchangers required above and below the pinch
d. design of heat exchange network above and below the pinch

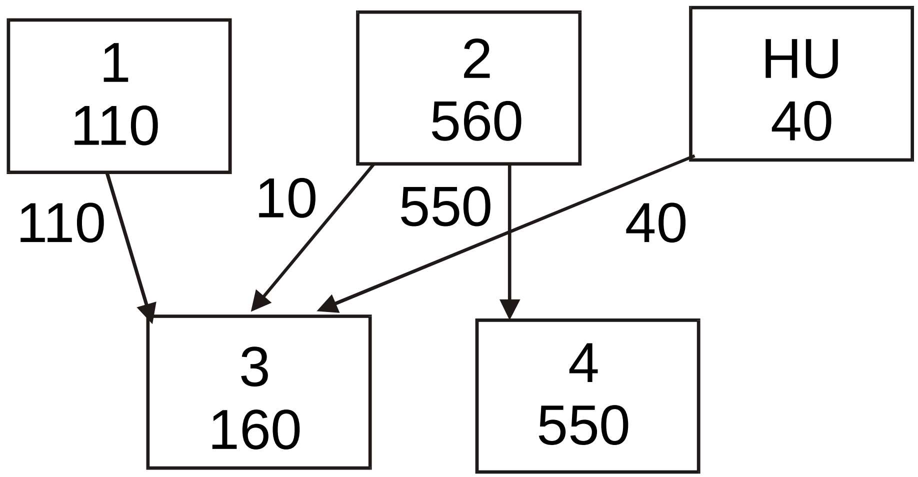
Temperature Interval Diagram



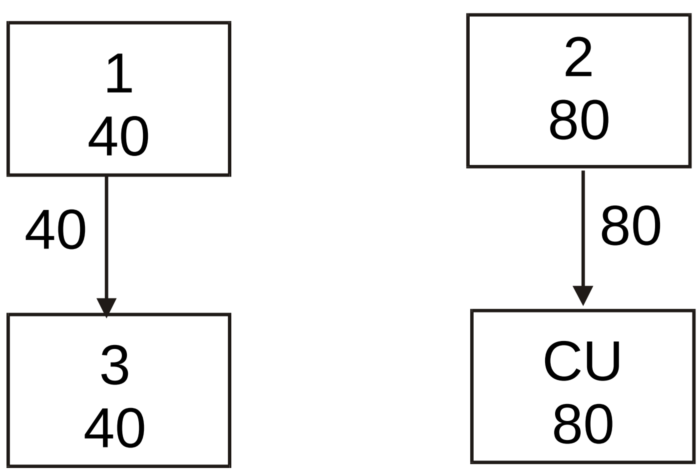
Cascade Diagram



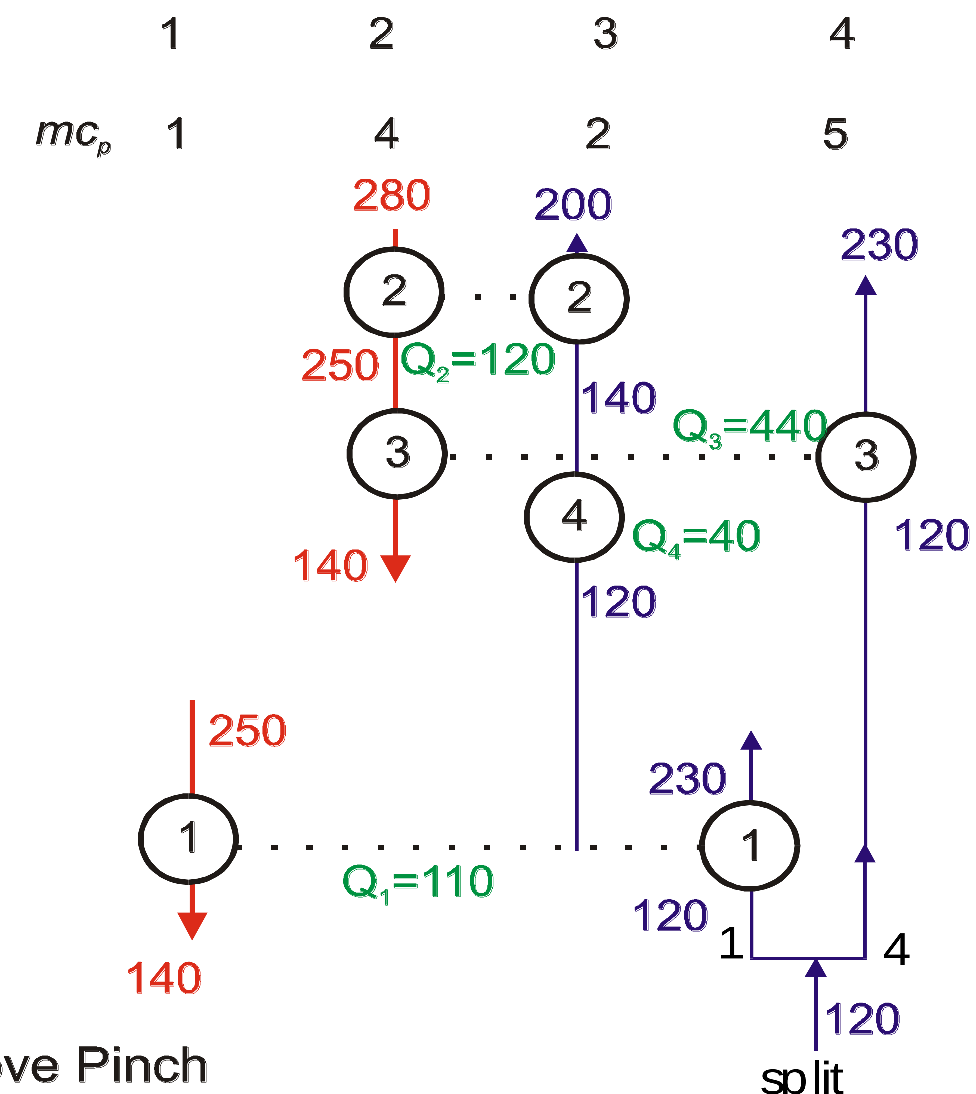
above pinch - 4 exchangers



below pinch - 2 exchangers, if possible

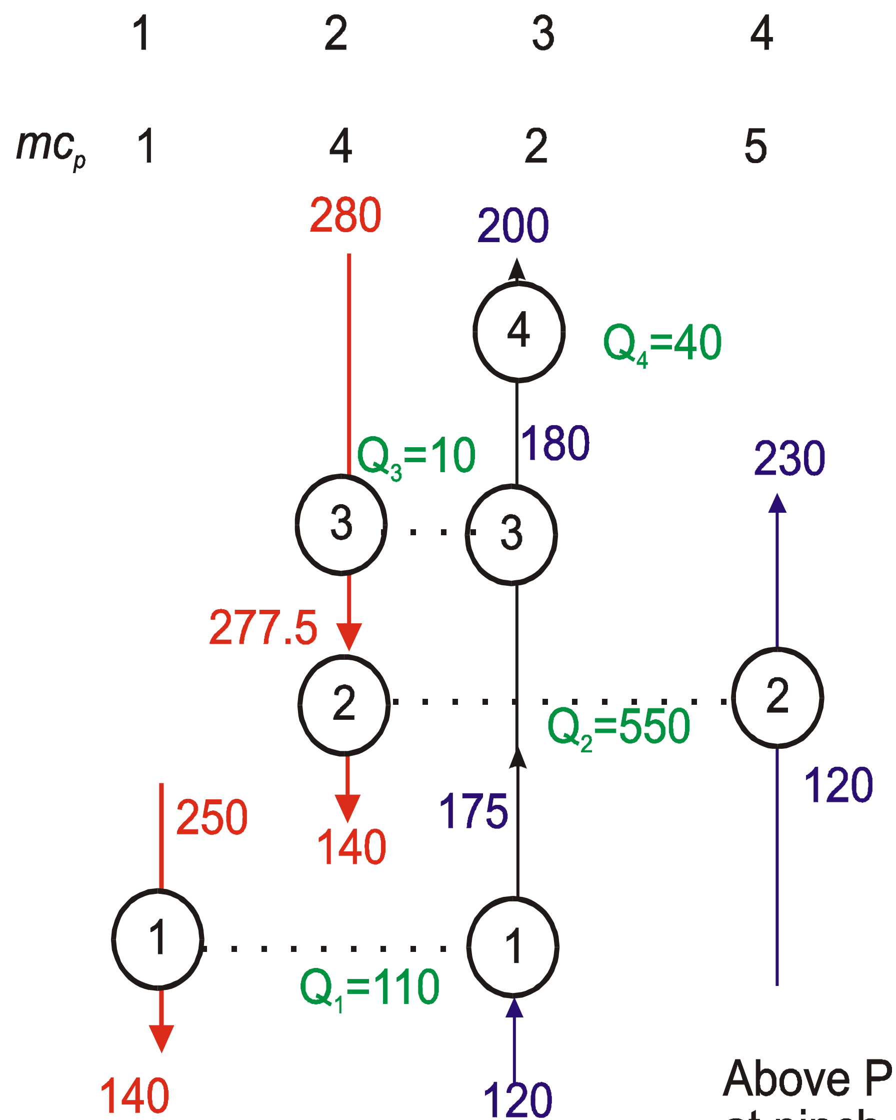


Exchanger Network - Design Above Pinch



Above Pinch
at pinch match $mc_{pH} \leq mc_{pC}$

Exchanger Network - Alternate Design Above Pinch

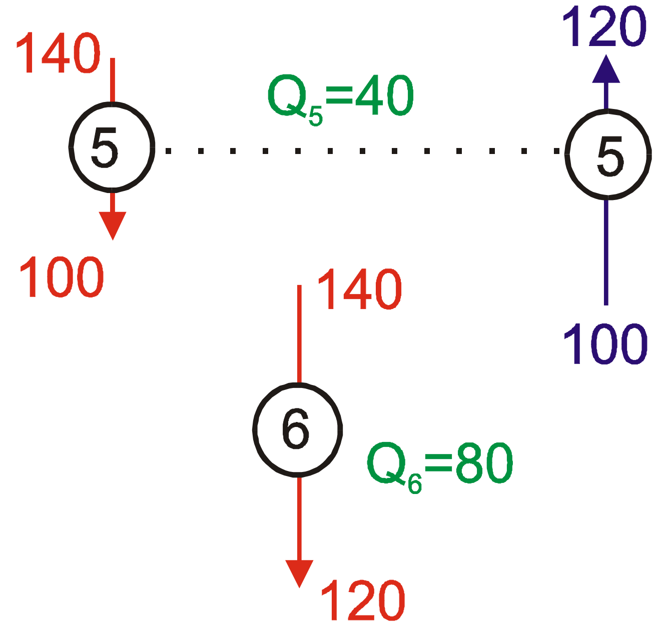


Above Pinch
at pinch match $mc_{pH} \leq mc_{pC}$

Exchanger Network - Design Below the Pinch

incorrect

Stream	1	2	3	4
mc_p	1	4	2	5

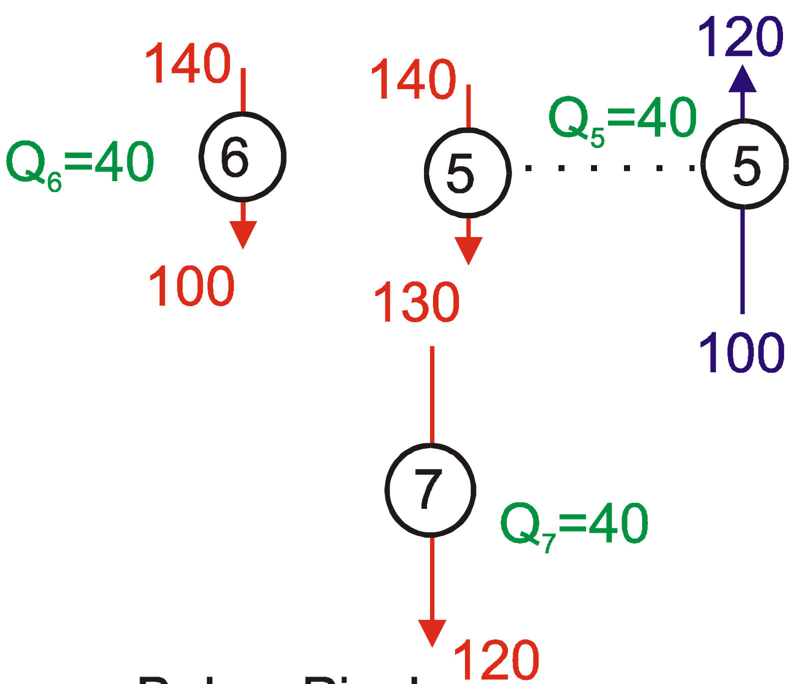


Below Pinch
at pinch match streams at pinch
$mc_{pH} \geq mc_{pC}$

Exchanger Network - Design Below the Pinch

correct

Stream	1	2	3	4
mc_p	1	4	2	5



Below Pinch
at pinch match streams at pinch
$mc_{pH} \geq mc_{pC}$

Summary

- Heat Exchange Networks
- Well-established procedure
- Not necessarily (and unlikely to be) economic optimum but a very good starting point
- Straight forward, but must be careful when matching streams at pinch
- Different correct answers possible

The University of Jordan
School of Engineering
Chemical Engineering Department

Program: B.Sc.
Academic Year: (/)
Semester:______________________

- **CHE 0915571: Chemical Process Design**

- **Course Catalog (2024)***

Evolutionary nature of process design, evolution of conceptual design through process diagrams and flowsheet structure with emphasis on utilities and pipe sizing, synthesis and design of chemical processes by understanding the hierarchy of process design using the onion model; Structure of the chemical process flowsheet; design and optimization of process recycle structure; Sequencing of simple distillation columns; Synthesis and design of heat exchanger networks; Computer applications with emphasis on available flowsheeting packages.

Credit hours	**3**	**Level**	**5**	**Pre-requisite(s)**	**0905421, 0935441**
Instructor **Prof. Menwer Attarakih**		**Office number** **CHE258**		**Office phone** Ext. 22887	**E-Office Hours** Mon & Wed: 10:00-10:30
Course website: https://elearning.ju.edu.jo/login/index.php **Live Streaming Platform:** Microsoft teams		**E-mail** **m.attarakih@ju.edu.jo**		**Lecture room:** Refer to Registration website	

*2024 Curriculum

- **Textbook:**
  1. R. Turton, J. Shaeiwitz, D. Bhattacharyya, W. B. Whiting (2018). Analysis, synthesis and design of chemical processes, 5th Ed., Prentice Hall, PTR, New Jersey.
  2. Instructor Handouts.

- **References:**
  1. Biegler, L. T., Grossmann, I. E. and Westerberg, A. W. (1997). Systematic methods of chemical process design. New Jersy, Prentice-Hall Inc. .
  2. Coulson, J. M. & Richardson, J. F. (2003). Chemical engineering (vol. 6), Pergamon Press, Oxford.
  3. Douglas, J. M. (1988). Conceptual process design of chemical processes, McGraw-Hill Book Co., New York.
  4. Smith, R. (2005). Chemical process design and integration, John Wiley & Sons, New York.
  5. Seider, W. D., Seader, J. D. & Lewin, D. R. (1999). Process design principles, John Wiley & Sons, New York.
  6. CAPE OPEN TO CAPE OPEN Simulation Environment: http://www.cocosimulator.org/

- **Goals**
  1. Understanding new process flowsheet creation and analyzing exist ones.
  2. Understand how to develop process alternatives and generate a base case flowsheet.
  3. Improvement and optimization of the bases case structure through heuristic rules and energy integration.

- **Learning Objectives and Intended Learning Outcomes**

Objectives	Outcomes
1. Understanding of process design through the evolution of chemical process diagrams (O2)	1.1 Understanding, Drawing & reading different types of chemical process flow diagrams (O2) 1.2 Use of process flow diagrams as an evolutionary tool in process design (O2) 1.3 Critical analysis of different types of process flow diagrams (O2)
2. Utilizing Heuristics to confirm the suitability of process design (O1, O2, O4)	2.1 Applying technical heuristics and short-cut methods (O1, O2, O4) 2.2 Using Tables of technical heuristics and guidelines (O1,O4)
3. Understand & Analyze the HDA process as a standard case study with process description (O1)	3.1 Enhance the ability to understand & critically analyze relatively complex flowsheet (O1) 3.2 Ability to use basic rules for writing process description (O1)

▪ Learning Objectives and Intended Learning Outcomes (Continued)

Objectives	Outcomes
4. Apply process flowsheet simulators to carry out complex M & E balances using given case study (O6)	3.1 Developing basic skills to use available commercial and free simulators at single unit and flowsheet levels (e.g. CAPE OPEN TO CAPE OPEN Simulation Environment) (6)
5. Basic understanding of process (system) design through process synthesis as an evolutionary process with the help of shortcut calculations, heuristic tables & process simulators (O1,O2,O4,O7)	5.1 Be able to define design objectives, specifications, process capacity & stream factor (1,4) 5.2 Know how to collect data, major data sources and generate data data banks for process design (7) 5.3 Synthesize and analyze chemical processes using the hierarchical approach of process synthesis through understanding (1,2,4): 5.3.1 Batch versus continuous processes 5.3.2 Input/ Output structure of the process 5.3.3 Reactor section: The heart of the Chemical Process 5.3.4 Recycle Structure of the process. 5.3.5 Structure of the separation system 5.3.6 Energy integration 5.4 Be able to generate different process alternatives in a rapid way using shortcut calculations 5.5 Check the suitability & use of heuristics using process simulators 5.6 Use process simulators to analyze and screen process alternatives 5.7 Be able to identify and design the process recycle structure 5.8 Choose and optimize the operating conditions of individual equipment and its interactions in the process flowsheet
6. Understanding heuristic and analytical approaches for sequencing of simple distillation columns (O1,O2)	6.1 Understand the difference between simple and complex distillation columns (1,2) 6.2 Use heuristics for sequencing simple distillation columns in relation to a complete process flowsheet (1,2) 6.3 Discover the possibility of running into conflict when using heuristic rules for sequencing simple distillation columns (1,2) 6.4 Use the approximate Underwood's equation to develop a quantitative tool for sequencing simple distillation columns (1,2)
7. Understand and apply the basic ideas of heat integration and apply the pinch analysis to design and optimize heat-exchanger networks (O1,O2,O4)	7.1 Understand the difference between the performance of individual heat exchanger and Heat Exchanger Network (HEN) (1,2) 7.2 Understand the basic concept of energy integration and the existence of optimal minimum approach temperature in a HEN (1) 7.3 Be able to identify cold & hot streams and extract its data (1,2) 7.4 Minimize utilities consumption using the TI & Cascade diagrams 7.5 Be able to link the first and second laws of thermodynamics to the pinch analysis (1,2) 7.6 Be able to identify the pinch zone in the HEN (1,2) 7.7 Apply the stream matching at minimum utilities and explore different HEN alternatives (1,2) 7.8 Design and analyze the final HEN and project it back on the process flowsheet (1,2) 7.9 Appreciate the heat integrated flowsheet in terms of savings in the EAOC and in terms of flowsheet complexity (1,2,4)
8. Enhance the ability of students for life-long learning and communication skills (O3, O7)	8.1 Enhance students' skills through intensive use of available data resources and short projects with written and oral presentations (3,7)

Topics Covered

Week	Topics	Reference
1- 3	Evolutionary nature of process design using chemical process flow diagrams	Handouts, Textbook, Chap. 1
4	Utilizing Heuristics to confirm the suitability of process design	Handouts, Textbook, Chap. 11
5	Case Study: The HDA process	Handouts, Textbook, Chap. 1, 5
5- 6	Introduction to process flowsheet simulation	Handouts, Ref. 6
6-11	Synthesis and analysis of chemical processes	Handouts, Textbook, Chap. 2, 3, 6
12-13	Introduction to sequencing of simple distillation columns	Handouts, Chap. 11, Ref. 3, 5
14-16	Introduction to design of heat-exchanger networks	Handouts, Textbook, Chap. 15

Evaluation

Evaluation Tool	Weight	Date
Midterm Exam	30	Will be announced by the department
Short exams	12-15	Will be arranged between the 5th and 16th weeks
Homework & Presentations	5-8	To be arranged one week after the assignment
Final Exam	50	Will be announced by the University

Relationship to Program Outcomes (scale 1 to 5)

New 1 To 7	O1	O2	O3	O4	O5	O6	O7
Percentage of grades	X	X	X	X			
1	Formulate, and solve complex engineering problems by applying principles of engineering, science, and mathematics						
2	Apply engineering design with solutions that meet specified needs with consideration of public health, safety, and welfare, as well as global, cultural, social, environmental, and economic factors						
3	An ability to communicate effectively with a range of audiences						
4	An ability to recognize ethical and professional responsibilities in engineering situations and make informed judgments, which must consider the impact of engineering solutions in global, economic, environmental, and societal contexts						

Relationship to CHE Program Objectives

PEO1	PEO2	PEO3	PEO4	PEO5	PEO6	PEO7	PEO8	PEO9	PEO10	PEO11
√	√	√								

Document Control

Prepared by	Prof. Menwer Attarakih
Last Modified	23.09.2025
Current Version	23.09.2025

1

Nov. 24, 2025

50 kg organic solvent (excess of 2nd reactant B) ρ 1050 $\frac{kg}{m^3}$ $V_{solvent}$

10 kg A (ρ 980 kg/m3) V_A

RxR Mixer (7hr) x 0.94 V 60%

17.5 kg of product is produced

Determine the volume of reaction vessel to produce 1 kg of product (specific volume)

Ques. How to allocate the production of A, B & C during this time horizon.

$50+10 = 60.0$
-17.5
$\boxed{42.5}$

$S_{A,rx}$

$$\boxed{V = V_A + V_S}$$

$V_{A\,60\%} = \frac{10\,kg}{980\,kg/m^3}$ | $V_S = \frac{50\,kg}{1050\,kg/m^3}$

V 60%

$$V_{60\%} = \frac{10}{980} + \frac{50}{1050} = 0.010204 + 0.047619$$

$$\boxed{V_{60\%} = 0.05782\ m^3}$$

0.6 V_{60} .. 60%

1 V_{100} → x

$$\boxed{V_{A,Rx}} = \boxed{V_{100\%} = 0.05782} \left| \frac{100}{60} \right. = 0.0937\ m^3$$

B ≡ Batch size

$$\text{specific volume} = \frac{\text{Volume}}{\text{Batch size}} \Rightarrow \frac{V_{A,Rx}}{B_A} = S_{A,Rx}$$

B = 17.5 kg of product A

$$S_{A,Rx} = \frac{0.0937\ [m^3]}{17.5\ [kg]\ \text{of product A}} = 0.005507 \left[\frac{m^3}{kg\ prod}\right]$$

$S_{B,RX} = 0.007860 \frac{[m^3]}{[kg]\ \text{product}}$ | $S_{C,RX} = 0.006103 \left[\frac{m^3}{kg\ pr}\right]$

$$\boxed{V = S \cdot B}$$

$Q_A = 10{,}000$ kg | $Q_B = 15{,}000$ kg, $Q_C = 35{,}000$ kg | TH = 500 kg

$t_{cycle,A} = 7.0$ hr | $t_{cycle,B} = 9.0$ h | $t_{cycle,C} = 10.0$ hr

$B_A = \frac{Q_A}{n_A ?}$ | $B_B = \frac{Q_B}{n_B}$ | $B_C^? = \frac{Q_C}{n_C ?}$

$T_A = n_A\, t_{cycle,A}$ | $T_B = n_B\, t_{cycle,B}$ | $T_C = n_C\, t_{cycle,C}$

$$T_A + T_B + T_C = 500$$

$V_A = S_{A,Rx}\, B_A$ | $V_B = S_{B,Rx}\, B_B$ | $V_C = S_{C,Rx}\, B_C$

$$V_{Rx} = \max(V_A, V_B, V_C)$$

2

Nov. 24/2025

$\pi_A \quad | \quad \pi_B \quad | \quad \pi_C$

0 $\qquad\qquad\qquad\qquad$ 600

$(\pi_A + \pi_B + \pi_C = 500)$ — (1)

$$\left[\begin{array}{l} B_A = \dfrac{Q_A}{n_A} \\ \pi_A = n_A \cdot t_{cycl,A} \end{array}\right] \Rightarrow B_A = \frac{Q_A}{\pi_A / t_{cycl,A}} \quad \Big| \quad B_B = \frac{Q_B}{\pi_B / t_{cyc,A}} \quad \Big| \quad B_C = \frac{Q_C}{\pi_B / t_{cyc,C}}$$

Use, heuretic that the batch ~~sizes~~ sizes are the same for each product

$$\begin{array}{ll} \pi_A + \pi_B + \pi_C = 500 & -1 \\ ?\ B_A = Q_A / \pi_A / t_{cyc,A} & -2 \\ ?\ B_B = Q_B / \pi_B / t_{cy,A} & -3 \\ ?\ B_C = Q_C / \pi_C / t_{cycl,C} & -4 \end{array}$$

unknowns:

$\pi_A,\ B_A,\ \pi_B,\ B_B,\ \pi_C,\ B_C$

(6)

4 eqs

$DF = 6 - 4 = 2$ more equation

$$\left|\begin{array}{l} B_A = B_B \\ B_B = B_C \end{array}\right| \longrightarrow Q_A / \pi_A / t_{cycl,A} = Q_B / \pi_B / t_{cycl,B} \quad *$$

$$Q_B / \pi_B / t_{cycl,B} = \frac{Q_C}{\pi_C / t_{cyc,C}} \quad **$$

system:

$\pi_A + \pi_B + \pi_C = 500$ — (1)

$$\left[\frac{10,000}{\pi_A / 7}\right] = \left[\frac{15,000}{\pi_B / 9}\right] \qquad (10,000 \times 9)\ \pi_B = (15,000 \times 9)\ \pi_A$$

(2)

$(15,000 \times 9)\ \pi_A - (10,000 \times 7)\ \pi_B = 0$

$$\left[\frac{35,000}{\pi_C / 10}\right] = \frac{15,000}{\pi_B / 9} \qquad (35,000 \times 10)\ \pi_B - (15,000 \times 9)\ \pi_C = 0$$

(3)

$$\begin{bmatrix} 1 & 1 & 1 \\ 15\times 10^3 \times 9 & -10\times10^3 \times 7 & 0 \\ 0 & 35\times10^3 \times 10 & -15\times10^3 \times 9 \end{bmatrix} \begin{bmatrix} \pi_A \\ \pi_B \\ \pi_C \end{bmatrix} = \begin{bmatrix} 500 \\ 0 \\ 0 \end{bmatrix} \Rightarrow \begin{array}{l} \pi_A = 63 \\ \pi_B = 122 \\ \pi_C = 315 \end{array}$$

check:

$$B_A = \frac{Q_A}{\pi_A / t_{cycl,A}} = \frac{10\,000}{63/7} = 1110 \text{ Kg}$$

$$B_B = \frac{15\,000}{122/9} = 1110 \text{ Kg}$$

$$B_C = \frac{35\,000}{315/10} = 1110 \text{ Kg}$$

3

Nov. 24/2025

$$\left[\frac{m^3}{kg\,prod}\right] \cdot [kg\,product]$$

$$V_A = S_A \cdot B_A = 0.005507 \times 1110 = 6.11\ m^3$$

$$V_B = S_B \cdot B_B = 0.007866 \times 1110 = 8.73\ m^3$$

$$V_C = S_C \cdot B_C = 0.006103 \times 1110 = 6.77\ m^3$$

$$V_{RX} = \max\,(V_A, V_B, V_C)$$

$$V_{RX} = \max\,(6.11,\ 8.73,\ 6.77)$$

$$\boxed{V_R = 8.73\ m^3}$$

\# of batches:

$$\boxed{B_i = \frac{Q_i}{n_i}} \qquad \boxed{B_i = 1110\ \text{kg product}}$$

A: $n_A = \dfrac{10{,}000}{1110} = 9$ batches

B: $n_B = \dfrac{15{,}000}{1110} \cong 14$ batch

C: $n_C = \dfrac{35{,}000}{1110} = 32$ batches

$T_A = n_A\, t_{cycle,A}$

$n_A = \dfrac{63}{7} = 9$

$n_B = \dfrac{122}{9} = 14$

$n_C = \dfrac{315}{10} = 32$

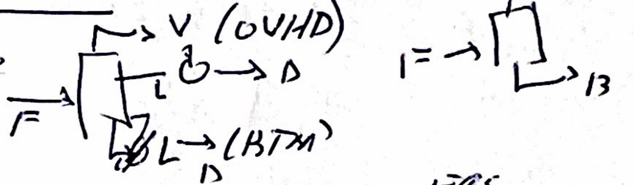
Batch size of product i

$$B_i\,[kg] = \frac{\text{Production quantity } i}{\text{\# of batches } i} = \frac{Q_i}{n_i}$$

$$B_i = \frac{Q_i}{T_i / t_{cycle,i}}$$

Heuristics For Towers (Table 11.13)

(1) Determine LK & HK volatilities @ the



$$\alpha_{top} = \frac{P^*_{LK}(T_{top})}{P^*_{HK}(T_{top})} \qquad \ln P^*(T) = A - \frac{B}{T+C} \quad \text{Antoine EQs}$$

$$\alpha_{BTM} = \frac{P^*_{LK}(T_{BTM})}{P^*_{HK}(T_{BTM})} \qquad \boxed{\alpha = \sqrt{\alpha_{top} \cdot \alpha_{BTM}}}$$

(2). minimum Reflux Ratio

$$R = \frac{L}{D}$$ D overhead product

$$R_{min} = \left(\frac{F}{D}\right)\left(\frac{\alpha}{\alpha-1}\right), \text{ when feed @ bubble point temp.}$$

$$R_{min} = \left(\frac{F}{D}\right)\left(\frac{\alpha}{\alpha-1}\right) - 1, \text{ when feed @ Dew point}$$

$(N \to \infty)\ R_{min} < R < \infty\ (D=0)$

(3). Optimum Reflux ratio: $\boxed{R_{opt} = (1.2 - 1.5)\, R_{min}}$

(4). ~~Optimum~~ Minimum number of theoretical trays ~~N_{opt}~~ N_{min}:

$$N_{min} = \frac{\ln\left[\left(\frac{x}{1-x}\right)_{OVHD} \Big/ \left(\frac{x}{1-x}\right)_{btm}\right]}{\ln\alpha}$$ Fenske-Underwood eqn.

(5.) Optimum number of theoretical trays N:

$$\boxed{N_{opt} = 2\, N_{min}} \Rightarrow \boxed{N = N_{opt}(1 + 0.1)}$$

(6). Actual —"— of trays

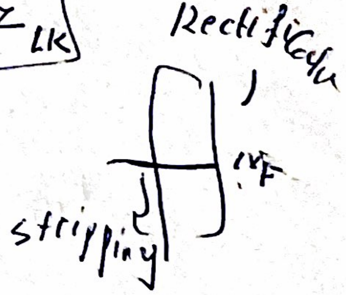
$$N_{actual} = \frac{N}{\underset{\text{tray efficiency}}{\varepsilon}} = \frac{N_{opt}(1+0.1)}{\varepsilon}$$

(7) Feed tray position, (N_F) $\quad \boxed{x = x_F = z_{LK}}$

$\left(\frac{x}{1-x}\right)_{feed}$

$$N^{Rect}_{min} = \frac{\ln\left[\left(\frac{x}{1-x}\right)_{OVHD} \Big/ \left(\frac{x}{1-x}\right)_{feed}\right]}{\ln(\alpha)}$$

$(x)_{LK}, (1-x)_{HK}$



Dec. 11, 2023

8) Theoretical # of trays $N^{Rect.}$

$$\frac{N_{min}^{Rect}}{N_{min}} = \frac{N^{Rect.}}{N} \Rightarrow \boxed{\text{Find } N^{Rect}}$$

Feed Location

N^{Rect}

9) From Table (11.4), Calculate Tower Diameter (D)

Use: vapor velocity u:

$$u = \frac{F_s}{\sqrt{\rho_V}} \qquad F_s \in (1.2 - 1.5) \left[\frac{m}{s} / \left[\frac{kg}{m^3}\right]^{1/2}\right]$$

$F_s \equiv$ Vapor factor

Based on $F_s \Rightarrow D$ can be determined

$$u = \frac{(1.2 - 1.5)}{\sqrt{\rho_V}} \qquad \rho_V =$$

(OVHD) 112 °C

137 °C, 117 °C

$$\boxed{\frac{1}{\rho_V}} = \left[\frac{1}{\rho} = \frac{x_{LK}}{\rho_{LK}} + \frac{x_{HK}}{\rho_{HK}}\right] \text{ at OVHD} \quad \Bigg\} \quad \rho = \left(\frac{m}{V}\right) = \frac{PM}{RT}$$

V mass Flow rate of OVHD $\left[\frac{kg}{h}\right]$ P tower pressure

$\dot{V}$ volumetric Flow rate of (OVHD) T — temp. @ OVHD

$$\dot{V} = Au \Rightarrow A = \left(\frac{\pi}{4}\right) D^2 \left[\frac{m^3}{s}\right]$$

$$\boxed{D_{tower} = \sqrt{\frac{4\dot{V}}{\pi u}}}$$

10) Tower pressure Drop: $\boxed{\Delta P_{tower} = (N_{actual})(\Delta P_{tray})}$

$\boxed{\Delta P_{tray} \quad 0.007 \text{ bar}}$

Feed tray position N_F

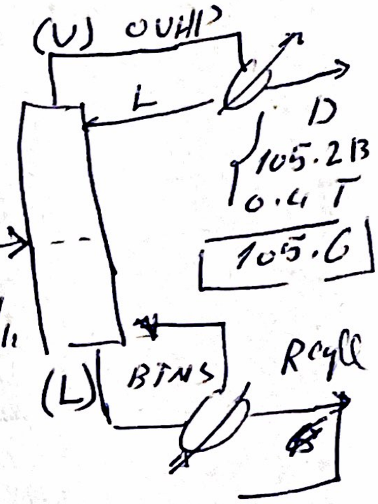
$(x_B)_{Feed} =$

$$N_{min}^{Rect} = \ln\left[\left(\frac{0.996}{1-0.996}\right)/\right.$$

T-101 Hydrodealkylation of Toulene — Dec. 11, 2023

3

$$C_6H_5CH_3 + H_2 = C_6H_6 + CH_4$$



Data

column pressure ~ 3 bar

T_{top} ~~142~~ $^\circ C$, T_{BTM} 147 $^\circ C$

T 1.22 kmol/h

X H_2 0.02 kmol/h

B 289.46 kmol/h

$\Sigma F_i =$ ~~289.48~~ 290.7

LK : B

HK : T

$x_B = \frac{106.3}{142.2}$ 0.748

B 106.3 kmol/h

T 35 kmol/h

H_2 0.02 —"—

CH_4 0.88

$\Sigma F_i = 142.2$ kmol/h

BTM

B 1.1 kmol/h

T ~~34.7~~ kmol/h

T 34.6

$\Sigma F_i =$ ~~102.8~~ 35.7 kmol/h

(OVHD) composition for LK

OVHD

$$x_B = \frac{289.46}{\cancel{289.48}\ 290.7} = 0.996$$

1 ~~x~~ 2

$x_T = 1 - 0.996$

BTM

~~x_B~~

$$x_B = \frac{1.1}{35.7} = 0.0308$$

$1 - x_B = x_T = 1 - 0.0308$

$$\alpha_{OVHD} = \alpha^*_{top} = \frac{P^*_{LK}}{} \frac{P^*_B(T_{top})}{P^*_T(T_{top})} \qquad T_{top} = 122\,^\circ C$$

$$\alpha_{BTM} = \alpha^*_{BTM} = \frac{P^*_B(T_{BTM})}{P^*_T(T_{BTM})} \qquad T_{BTM} = 147\,^\circ C$$

$$\alpha = \sqrt{\alpha_{OVHD} \cdot \alpha_{BTM}} = \sqrt{2.41 * 2.13} = 2.28$$

V = 22 700 kg/h

a) bubble-point

$$R_{min} = \frac{F}{D}\left(\frac{1}{\alpha - 1}\right) = \frac{142.2}{105.6}\left(\frac{1}{2.28-1}\right) = 1.05 \quad R_{min}$$

$$R_{opt} = (1.2 - 1.5) R_{min} = (1.2 - 1.5) \times 1.05 = 1.26 \to 1.58$$

Range of R

$$N_{min} = \frac{\ln\left[\left(\frac{0.996}{1-0.996}\right) / \left(\frac{0.0308}{1-0.0308}\right)\right]}{\ln(2.28)} = 10.9 \quad N_{min}$$

$N_{opt} = 2 N_{min} = 2 \times 10.9 = 21.8$, $\varepsilon = 0.6$

$N = N_{opt}(1 + 0.1) = 21.8 \times 1.1 = 23.98 \sim 24$

$$N_{actual} = \frac{N}{\varepsilon} = \frac{24}{0.6} \simeq 40 \text{ trays}$$

Dec. 11, 2022

× To determine tower diameter D_{tower}

F_s vapor factor (1.2 – 1.5)

$\rho_V \Rightarrow \frac{1}{\rho_V} = \frac{x_B}{\rho_B} + \frac{x_T}{\rho_T}$ $\boxed{u = \frac{F_s}{\sqrt{\rho}}}$

$\rho_B = \frac{P M_B}{R T_{top}} = \frac{3 \text{ bar} \cdot 78 \text{ g}}{\text{mol}} \Big| \frac{\text{mol} \cdot \text{K}}{0.0834 \text{ L} \cdot \text{bar}} \Big| \frac{1}{(273+112) \text{K}}$

x_B 0.996

$\rho_B = 7.3$ kg/m3

$\rho_T = \frac{3 \times 92}{0.0834 \; \underbrace{(273+112)}_{385}} = 8.60$ kg/m3

x_T (1 – 0.996) 0.004

$\frac{1}{\rho_V} = \underbrace{\frac{0.996}{7.3}}_{0.1364} + \underbrace{\frac{(1-0.996)}{8.6}}_{0.000465} = 7.3$ kg/m3

$\rho_V = 0.996 \times 7.3 + (1-0.996) \times 8.6$

$u = \frac{1.2 - 1.5}{\sqrt{7.3}} = 0.44 \rightarrow 0.555$ ~0.6 m/s

OVHD = Vapor Flow rate: 22700 kg/h $\dot{V} = \frac{\dot{m}}{\rho_V}$

$\dot{V} = \frac{22700 \text{ kg/h}}{7.3 \text{ kg/m}^3} = 3109 \frac{\text{m}^3}{\text{h}} \Big| \frac{1 \text{h}}{3600 \text{s}}$

$\dot{V} = 0.864$ m³/s

$D_{tower} = \sqrt{\frac{4V}{\pi u}} = \sqrt{\frac{4 \times 0.864}{\pi \times 0.44}}$ (0.6)

$D_{tower} = 1.83 \text{ m} — 1.35 \text{ m}$ $\underbrace{(1.35 — 1.83)\text{m}}_{D_{tower}}$

$\Delta P_{tower} = \frac{N_{actual}}{\Delta P_{tray}} = \frac{40}{.007} = 0.28$ bar

$N_{min}^{Rect} = \frac{\ln \left[\left[\frac{0.996}{1-0.996} \right] \Big/ \frac{0.748}{1-0.748} \right]}{\ln(2.28)} = \frac{4.43}{0.824} = \boxed{5.4 \; N_{min}^{Rect}}$

249 2.96

$\frac{5.4}{10.9} = \frac{N^{Rect}}{N} = \frac{N^{Rect}}{24} \Rightarrow N^{Rect} = 24 \times \frac{5.4}{10.9} \cong 11$

$N_c^{strip} = 24 - 11 = 13$ plates

$N_{act}^{Rect} = \frac{11}{0.6} = 18$ ← Feed tray

$N_{str} = 40 - 18 = 22$

Gas velocity

2

Dec. 27 / 2025

$u = K\sqrt{\frac{\rho_L}{\rho_V} - 1}$

$\dot m_V = \rho_V \dot V_V \qquad \dot V_V = uA$

$\dot m_V = \rho_V \cdot u A$?

$A = \frac{\pi}{4} D^2$ cross-sectional area

$\boxed{n = \frac{m}{M}}$

$\dot m_V = \rho_V \cdot u \cdot \frac{\pi}{4} D^2$

$\boxed{D^2 = \frac{4}{\pi}\frac{\dot m_V}{\rho_V \cdot u}}$

$\dot m_V = 9200\ kg/h = \frac{9200}{3600}\left[\frac{kg}{s}\right] = 2.56\ \frac{kg}{s}$

$u = 0.235\ m/s = 0.75\,u = 0.75 \times 0.313$

$\rho_V = 8\ kg/m^3$

$D^2 = \frac{4}{\pi}\,\frac{9200}{3600}\cdot\frac{1}{8}\cdot\frac{1}{0.235} = 4.683$

$D = \sqrt{\;} = 1.326\ m = 1.33\ m$

D from vapor

Now $\frac{1}{2}V = V_{1/2}$ $\qquad V = AL$ vessel volume

$\boxed{V_{\frac{1}{2}} = AL = \frac{1}{2}\frac{\pi}{4}D^2 \cdot L = \frac{\pi}{4\times 2} \times (1.33)^2 \cdot L}$ — *

$V_{\frac{1}{2}}$ is calculated as follows:

$\dot m_L = \rho_L \dot V_L = \rho_L\left(\frac{V_{1/2}}{t}\right)$ | ρ_L 850 kg/m³, $\dot m_L$ 11570 kg/h, $t = 5\ min = \frac{5}{60}\ h$

$V_{\frac{1}{2}} = \frac{\dot m_L \cdot t}{\rho_L} = \frac{11570\ \dot m_L \times 5}{850 \times 60}$

$\boxed{V_{1/2} = 1.134\ m^3}$

from * $\quad 1.134 = \frac{\pi}{4} \times (1.33)^2 \cdot \frac{L}{2}$

$L = \frac{V_{1/2} \times 4 \times 2}{\pi\ (1.33)^2} = \frac{1.134 \times 4 \times 2}{\pi\ (1.33)^2}$

$\boxed{L = 1.633\ m}$

Average molar ~~weight~~ molecular of m

$\cancel{M_1}$ $M_{mix} = \sum x_i \cdot M_i$ (mol fract)

$M_{mix} = 0.\cancel{141} \times 2 + 0.758 \times 16 + 0.081 \times 78 + 0.011 \times 92 = 19.740$

			x_i	M_i
H_2	651.9	Kmol/h	0.5922	2
CH_4	438.3		0.3982	16
C_6H_6	9.55		0.0087	78
Tol	1.05		9.5×10^{-4}	92
	1100.8 Kmol/h			

$M = 0.5922 \times 2 + 0.3982 \times 16 + 0.0087 \times 78 \simeq 7.56$ 8.23

$$\boxed{PV = nRT \Rightarrow \frac{m}{V} = \rho = \frac{PM}{RT}}$$

$$\rho = \frac{23.9\ \text{bar} \times 7.56\ \text{kg/kmol}}{0.08314\left(\frac{\text{L.bar}}{\text{mol.K}}\right)(38+273)\text{K}\left(\frac{10^3\ \text{mol}}{\text{kmol}}\right)}$$ 8.23

$$\rho = \frac{23.9 \times 7.56\ \text{kg}}{0.08314\ \text{L} \times 311 \times 10^3} = \frac{7}{10^3}\frac{\text{kg}}{\text{L}}$$ 8.23

$$\rho = \frac{7\ \text{kg}}{\text{L}} \times \frac{10^3\ \text{L}}{\text{m}^3 \times 10^3}$$ 8.23

$$\boxed{\rho = 7\ \frac{\text{kg}}{\text{m}^3}}$$ = 8.23 kg/m3

Table 9.6 Heuristics for Process Vessels (Drums) (Adapted from Walas S. M., *Chemical Process Equipment: Selection and Design*, Butterworths, Stoneham, MA, 1988, copyright © 1988 by Butterworth Publishers, adapted by permission of Butterworth Publishers, Stoneham, MA, all rights reserved)

1. Drums are relatively small vessels that provide surge capacity or separation of entrained phases.
2. Liquid drums are usually horizontal.
3. Gas/liquid phase separators are usually vertical.
4. Optimum length/diameter = 3, but the range 2.5 to 5 is common.
5. Holdup time is 5 min for half-full reflux drums and gas/liquid separators, 5-10 min for a product feeding another tower.
6. In drums feeding a furnace, 30 min for half-full drum is allowed.
7. Knockout drums placed ahead of compressors should hold no less than 10 times the liquid volume passing per minute.
8. Liquid/liquid separations are designed for settling velocity of 0.085–0.127 cm/s (2–3 in/min)
9. Gas velocity in gas/liquid separators, $u = k\sqrt{\rho_l/\rho_v - 1}$ m/s (ft/sec) $k = 0.11$ (0.35) for systems with mesh deentrainer and $k = 0.0305$ (0.1) without mesh deentrainer.
10. Entrainment removal of 99% is attained with 10.2–30.5 cm (4–12 in) mesh pad thickness; 15.25 cm (6 in) thickness is popular.
11. For vertical pads, the value of the coefficient in Step 9 is reduced by a factor of 2/3.
12. Good performance can be expected at velocities of 30–100% of those calculated with the given k; 75% is popular.
13. Disengaging spaces of 15.2–45.7 cm (6–18 in) ahead of the pad and 30.5 cm (12 in) above the pad are suitable.
14. Cyclone separators can be designed for 95% collection at 5 µm particles, but usually only droplets greater than 50 µm need be removed.

$L = 2.5\,D$

$\frac{L}{D} = 3$ opt

$\frac{1.63}{2.33} = 1.2$

determine L once D is done

D → determine m/s ft/s

m/s ft/s

$u_{actual} = 0.75\,u$

$\frac{L}{D} = 3$

$L = 3D$

$u = K\sqrt{\frac{\rho_l}{\rho_v} - 1} = 0.23\ \frac{m}{s}$

$\frac{1}{2}$ volume of liquid for 5 minutes

($\dot m_v = \rho_v \dot V_v$, $\dot V = uA$)

$D = 1.33\ m$

$\boxed{2}$ $V_{1/2} = \frac{1}{2}\cdot\frac{\pi}{4}D^2 L = (0.695\,L)\ m^3$

$\boxed{1}$

$\dot m = \rho_v u A$?

$\dot V \Rightarrow \dot m_v = \rho_v u \frac{\pi}{4}D^2$?

$\dot m_L = 11570\ kg/h$

$\rho_L = 850\ kg/m^3$

how to calc $V_{1/2}$

$\dot m_v = 9200\ \frac{kg}{h} = \frac{9200}{3600}\ \frac{kg}{s}$

$\dot V = \frac{V_{1/2}}{t}$

$\dot V_L = \frac{11570}{850}\ \frac{m^3}{h} = \frac{11570}{850 \cdot 3600}\ \left[\frac{m^3}{s}\right]$

$\frac{9200}{3600} = 8 \times 0.23\ [\frac{m}{s}]\ \frac{\pi}{4}D^2$ (kg/m³)

$t = 5$ min $= 5$ min $\left|\frac{60\,s}{1\,min}\right.$

$V_{1/2} = \frac{\dot V_L}{t}$

$D^2 = \frac{9200}{3600} \cdot \frac{4}{8 \times 0.23 \cdot \pi}$

$\dot V_L = \left[\frac{11570}{850 \cdot 3600}\right]\left[\frac{m^3}{s}\right]$ volumetric flowrate

$t = 5\ min \cdot \frac{60\,s}{min}$

$t = (5 \times 60)\,s$

$D = 1.33\ m$

$V = \dot V_L \cdot t = \frac{11570}{850} \cdot \frac{5 \times 60}{3600}$

$L = 1.63\ m$ $= 0.695\,L$ $= (V = 1.13\ m^3)$

$= 0.695\,L \Rightarrow L = \frac{V}{0.695} = \frac{1.13}{0.695}$

Table 1.5 Flow Summary Table for the Benzene Process Shown in Figure 1.3 (and Figure 1.5)

Stream Number	1	2	3	4	5	6	7	8	9	1
Temperature (°C)	25	59	25	225	41	600	41	38	654	90
Pressure (bar)	1.90	25.8	25.5	25.2	25.5	25.0	25.5	23.9	24.0	2.6
Vapor Fraction	0.0	0.0	1.00	1.0	1.0	1.0	1.0	1.0	1.0	0.
Mass Flow (tonne/h)	10.0	13.3	0.82	20.5	6.41	20.5	0.36	9.2	20.9	11.
Mole Flow (kmol/h)	108.7	144.2	301.0	1204.4	758.8	1204.4	42.6	1100.8	1247.0	14
Component Flowrates (kmol/h)										
Hydrogen	0.0	0.0	286.0	735.4	449.4	735.4	25.2	651.9	652.6	0.
Methane	0.0	0.0	15.0	317.3	302.2	317.3	16.95	438.3	442.3	0.8
Benzene	0.0	1.0	0.0	7.6	6.6	7.6	0.37	9.55	116.0	10
Toluene	108.7	143.2	0.0	144.0	0.7	144.0	0.04	1.05	36.0	35.

	9	10	11	12	13	14	15	16	17	18	19
	654	90	147	112	112	112	38	38	38	38	112
	24.0	2.6	2.8	3.3	2.5	3.3	2.3	2.5	2.8	2.9	2.5
	1.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	1.0	0.0	1.0
	20.9	11.6	3.27	14.0	22.7	22.7	8.21	2.61	0.07	11.5	0.01
	1247.0	142.2	35.7	185.2	290.7	290.7	105.6	304.2	4.06	142.2	0.90
	652.6	0.02	0.0	0.0	0.02	0.0	0.0	178.0	0.67	0.02	0.02
	442.3	0.88	0.0	0.0	0.88	0.0	0.0	123.05	3.10	0.88	0.88
	116.0	106.3	1.1	184.3	289.46	289.46	105.2	2.85	0.26	106.3	0.0
	36.0	35.0	34.6	0.88	1.22	1.22	0.4	0.31	0.03	35.0	0.0

Table 1.7 Equipment Summary for Toluene Hydrodealkylation PFD

Heat Exchangers	**E-101**	**E-102**	**E-103**	**E-104**	**E-105**	**E-106**
Type	Fl.H.	Fl.H.	MDP	Fl.H.	MDP	Fl.H.
Area (m^2)	36	763	11	35	12	80
Duty (MJ/h)	15,190	46,660	1055	8335	1085	9045
Shell						
Temp. (°C)	225	654	160	112	112	185
Pres. (bar)	26	24	6	3	3	11
Phase	Vap.	Par. Cond.	Cond.	Cond.	l	Cond.
MOC	316SS	316SS	CS	CS	CS	CS
Tube						
Temp. (°C)	258	40	90	40	40	147
Pres. (bar)	42	3	3	3	3	3
Phase	Cond.	l	l	l	l	Vap.
MOC	316SS	316SS	CS	CS	CS	CS
Vessels/Tower/Reactors	**V-101**	**V-102**	**V-103**	**V-104**	**T-101**	**R-101**
Temperature (°C)	55	38	38	112	147	660
Pressure (bar)	2.0	24	3.0	2.5	3.0	25
Orientation	Horizontal	Vertical	Vertical	Horizontal	Vertical	Vertical
MOC	CS	CS	CS	CS	CS	316SS
Size						
Height/Length (m)	5.9	3.5	3.5	3.9	29	14.2
Diameter (m)	1.9	1.1	1.1	1.3	1.5	2.3
Internals		s.p.	s.p.		42 sieve trays 316SS	Catalyst packed bed-10m
Pumps/Compressors	**P-101 (A/B)**	**P-102 (A/B)**	**C-101 (A/B)**	**Heater**		**H-101**
Flow (kg/h)	13,000	22,700	6770	Type		Fired
Fluid Density (kg/m^3)	870	880	8.02	MOC		316SS
Power (shaft) (kW)	14.2	3.2	49.1	Duty (MJ/h)		27,040
Type/Drive	Recip./Electric	Centrf./Electric	Centrf./Electric	Radiant Area (m^2)		106.8
Efficiency (Fluid Power/Shaft Power)	0.75	0.50	0.75	Convective Area (m^2)		320.2
Pumps/Compressors	**P-101 (A/B)**	**P-102 (A/B)**	**C-101 (A/B)**	**Heater**		**H-101**
MOC	CS	CS	CS	Tube P (bar)		26.0
Temp. (in) (°C)	55	112	38			
Pres. (in) (bar)	1.2	2.2	23.9			
Pres. (out) (bar)	27.0	4.4	25.5			

Key:

	Materials of construction		
MOC	Stainless steel type	Par	Partial
316SS	316	F.H.	Fixed head
CS	Carbon steel	Fl.H.	Floating head
Vap	Stream being vaporized	Rbl	Reboiler
Cond	Stream being condensed	s.p.	Splash plate
Recipr.	Reciprocating	l	Liquid
Centrf.	Centrifugal	MDP	Multiple double pipe

Table 2.2 Common Refrigerants and Their Ranges of Cooling (Data from References [12] and [13])

Refrigerant	Typical Operating Temperature Range (°C)	Vapor Pressure at 45°C (bar)	Critical Pressure (bar)	Critical Temperature (°C)
Methane	−129 to −184	749	46.0	−82.5
Ethane	−59 to −115	1453	48.8	32.3
Ethylene	−59 to −115	2164	50.3	9.3
Propane	4 to −46	15.3	42.5	96.7
Propylene	4 to −46	18.45	46.1	91.6
N-Butane	16 to −12	4.35	38.0	152.0
Ammonia	27 to −32	17.8	112.8	132.5
Carbon Dioxide	4 to −50	787	73.8	31.1
Methylene Chloride	4 to −12	1.21	60.8	236.9
Methyl Chloride	4 to −62	9.84	66.8	143.1
R-134a (1,1,1,2-tetrafluoro-ethane)	4 to −50	11.6	40.6	101.0
R-152a (1,1-difluoro-ethane)	4 to −50	10.4	45.0	113.5

Table 6.1 Possible Reasons for Operating Reactors and Separators Outside the Temperature Ranges of Special Concern

Stream Condition	Process Justification for Operating at This Condition	Penalty for Operating at This Condition
High Temperature ($T > 250°C$)	**Reactors** • Favorable equilibrium conversion for endothermic reactions • Increase reaction rates • Maintain a gas phase • Improve selectivity • Other reasons	• Use of special process heaters • $T > 400°C$ requires special materials of construction
	Separators • Obtain a gas phase required for vapor-liquid equilibrium • Other reasons	
Low Temperature ($T < 40°C$)	**Reactors** • Favorable equilibrium conversion for exothermic reactions • Temperature-sensitive materials • Improved selectivity • Maintain a liquid phase • Other reasons	• Uses expensive refrigerant • May require special materials of construction for very low temperatures
	Separators • Obtain a liquid phase required for vapor-liquid or liquid-liquid equilibrium • Obtain a solid phase for crystallization • Temperature-sensitive materials • Other reasons	

Table 6.2 Possible Reasons for Operating Reactors and Separators Outside the Pressure Range of Special Concern

Stream Condition	Process Justification for Operating at This Condition	Penalty for Operating at This Condition
High Pressure ($P > 10$ bar)	**Reactors** • Favorable equilibrium conversion • Increased reaction rates for gas-phase reactions (due to higher concentration) • Maintain a liquid phase • Other reasons	• Requires thicker-walled equipment • Requires expensive compressors if gas streams must be compressed
	Separators • Obtain a liquid phase for vapor-liquid or liquid-liquid equilibrium • Other reasons	
Low Pressure ($P < 1$ bar)	**Reactors** • Favorable equilibrium conversion • Maintain a gas phase • Other reasons	• Requires large equipment • Special design for vacuum operation • Air leaks into equipment that may be dangerous and expensive to prevent and costly to remove
	Separators • Obtain a gas phase for vapor-liquid equilibrium • Temperature-sensitive materials • Other reasons	

Table 6.3 Possible Reasons for Non-Stoichiometric Reactor Feed Compositions of Special Concern

Stream Condition	Process Justification for Operating at This Condition	Penalty for Operating at This Condition
Inert Material in Feed to Reactor	• Acts as a diluent to control the rate of reaction and/or to ensure that the reaction mixture is outside the explosive limits (exothermic reactions) • Inhibits unwanted side reactions • Other reasons	• Causes reactor and downstream equipment to be larger since inert takes up space • Requires separation equipment to remove inert material • May cause side reactions (material is no longer inert) • Decreases equilibrium conversion
Excess Reactant	• Increases the equilibrium conversion of the limiting reactant • Inhibits unwanted side reactions • Other reasons	• Requires separation equipment to remove excess reactant • Requires recycle • Added feed material costs (due to losses in separation and/or no recycle)
Product Present in Feed to Reactor	• Product cannot easily be separated from recycled feed material • Recycled product retards the formation of unwanted by-products formed from side reactions • Product acts as a diluent to control the rate of reaction and/or to ensure that the reaction mixture is outside the explosive limits, for exothermic reactions • Other reasons	• Causes reactor and downstream equipment to be larger • Requires larger recycle loop • Decreases equilibrium conversion • Decreases selectivity

Table 6.4 Changes in Process Conditions That Are of Special Concern for a Stream Passing through a Single Piece of Equipment

Type of Equipment	Change in Stream Condition Causing Concern	Justification or Remedy	Penalty for Operating Equipment in This Manner
1. Compressors	$P_{out}/P_{in} > 3$	**Remedy:** Use multiple stages and intercoolers.	High theoretical work requirement due to large temperature rise of gas stream.
	High-temperature inlet gas	**Remedy:** Cool the gas before compression.	High theoretical work requirement and special construction materials required.
2. Heat Exchangers	$\Delta T_{lm} >$ 100°C	**Remedy:** Integrate heat better within process (see Chapter 15).	Large temperature driving force means that valuable high-temperature energy is wasted.
		Justification: Heat integration not possible or not profitable.	
3. Process Heaters	$T_{out} < T_{steam\ available}$	**Remedy:** Use high-pressure steam to heat process stream.	Process heaters are expensive and unnecessary if heating can be accomplished by using an available utility.
		Justification: Heater may be needed during start-up.	
4. Valves	Large ΔP across valve	**Remedy:** For gas streams install a turbine to recover lost work.	Wasteful expenditure of energy due to throttling.
		Justification: 1. Valve used for control purposes. 2. Installation of turbine not profitable. 3. Liquid is being throttled.	
5. Mixers (Streams Mixing)	Streams of greatly differing temperatures mix	**Remedy:** Bring temperatures of streams closer together using heat integration.	Wasteful expenditure of high-temperature energy.

		Justification: Heater may be needed during start-up.	
4. Valves	Large ΔP across valve	**Remedy:** For gas streams install a turbine to recover lost work.	Wasteful expenditure of energy due to throttling.
		Justification: 1. Valve used for control purposes. 2. Installation of turbine not profitable. 3. Liquid is being throttled.	
5. Mixers (Streams Mixing)	Streams of greatly differing temperatures mix	**Remedy:** Bring temperatures of streams closer together using heat integration.	Wasteful expenditure of high-temperature energy.
	Streams of greatly differing composition mix	**Justification:** 1. Quenching of reaction products. 2. Provides driving force for mass transfer.	Causes extra separation equipment and cost.

Table 6.5 Process Conditions Matrix for the PFD of the Toluene Hydrodealkylation Process Shown in Figure 1.5

Equipment	Reactors and Separators Tables 6.1–6.3					Other Equipment Table 6.4				
	High Temp	*Low Temp*	*High Pres.*	*Low Pres.*	*Non-Stoich. Feed*	*Comp*	*Exch.*	*Htr.*	*Valve*	*Mix*
R-101	X		X		X					
V-101										
V-102			X							
V-103										
V-104										
T-101										
H-101										
E-101							X			
E-102							X			
E-103										
E-104										
E-105										
E-106										
C-101										
P-101										
P-102										
PCV on Stream 8									X	
PCV on Stream from V-101 to V-103									X	

Suggested Problems for Chapter 1

1, 2, 7, 10, 11, 12, 14, 21, and 22

Suggested Problems for Chapter 2

1, 2, 3, 4, 5

7, 9 10, 11, 12, 13, 14

Suggested problems

Chapter 15 – Heat Exchanger Networks

2

11

12

13

14