

* Block Flow Diagram. Words $\Rightarrow$ Blocks.

Blocks $\sim$ Represent major equipments.

$\rightarrow$ extracted from ground.

Ex: Sweetening of natural gas (CH_4 C_2H_6 C_3H_8)

* H_2S * $\text{CO}_2 \Rightarrow$ have to remove them?

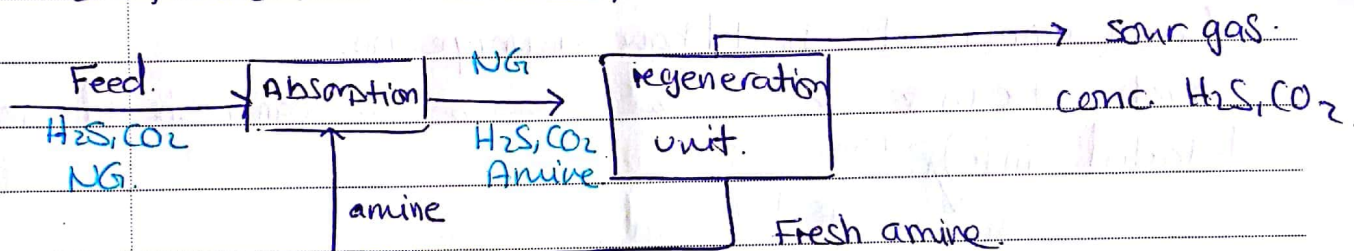
$\rightarrow$ you can soak in water (to remove H_2S) but NOT CO_2 - NOT Feasible.

$\rightarrow$ adsorbent: also NOT Feasible.

$\Rightarrow$ absorbant $\Rightarrow$ using amines (anything containing N).

$\hookrightarrow$ easily to be regenerated.

$\text{H}_2\text{S}, \text{CO}_2 + \text{Amine} \longrightarrow \text{Amine} - \text{H}_2\text{S}, \text{CO}_2$



NOTES- [Main characters].

- Left to right.
- heavy comp $\downarrow$ (Liq) Light $\uparrow$ (gases).
- general outlines.

$\hookrightarrow$ compositions NOT mentioned.

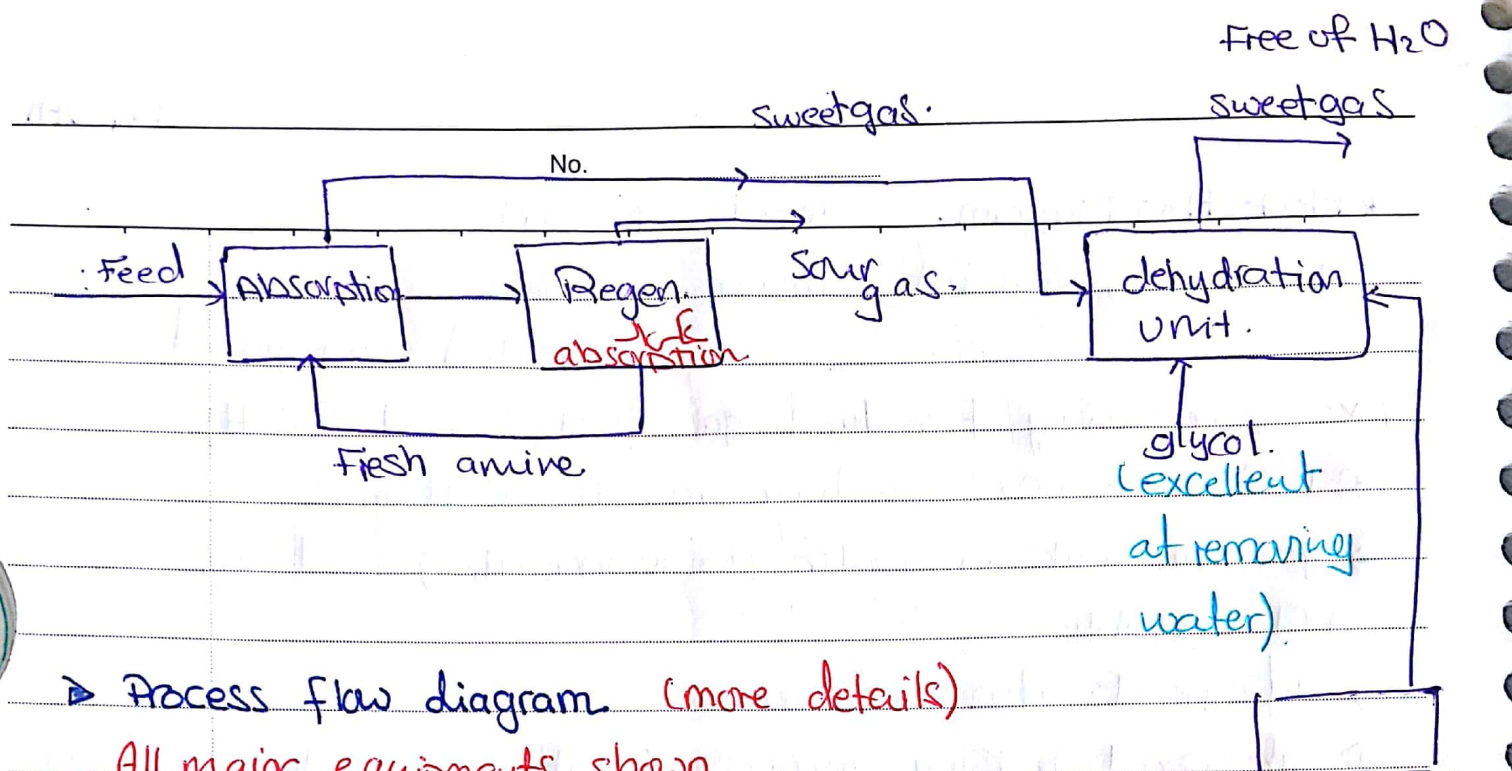
$\hookrightarrow$ critical info should be represented.

- when streams cross: horizontal line: solid vertical line: broken

- natural gas

$\hookrightarrow$ from ground; there's water 40-60 by volume of streams.
it used to freeze in pipes and break equipment.

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► Process flow diagram (more details)

- All major equipments shown.

ex: reactor → CSTR?
→ Batch?
→ BFR?

distillation → trays?
packed?

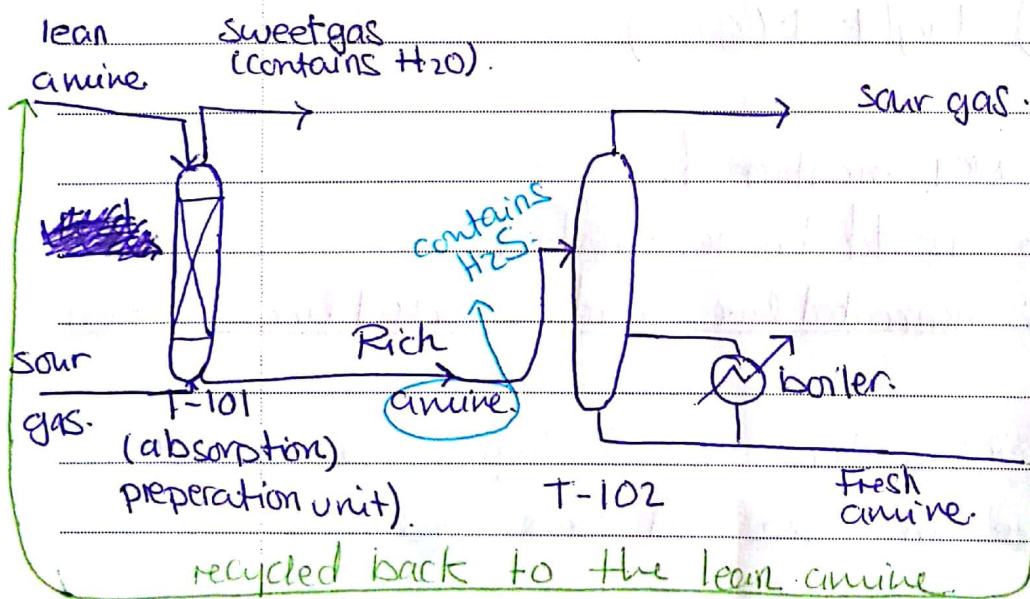
regeneration unit.

✓ each equipment should have a unique no.

- each stream " " " " " (and should provide detailed info) → Temp ✓
→ P ✓
→ composition ✓
Name it!

- all utilities should be provided.

- basic control loops should be provided.



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* Process Flow Diagram:-

Sep 26th
Tuesday

No. _____

① Equipment identification:-

Naming:-

→ Area working

Formula: XX- YZZ A/B → is stand by.

example:-

→ Area 100 / pump no. (1)

Pump P- 1001 A/B.

Tower T- 102

compressor C- 210 A/B → a stand by.

Reactor R-

Storage tank TK-

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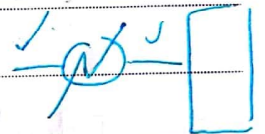
② Utilities:- low pressure steam Lps (3-5) bar.

→ Steam → medium pressure steam mps (10-15) bar

→ power → high " " hps (40-40) bar.

→ cooling water. (CW) 15-30°C.

difference is usually 15°C



③ streams:

(1) stream no.

(2) Temp °C

(3) pressure

(4) Flow rate

(5) vapor quality

④ equipments:

material of constr.

→ heat exchangers (D, L, A, MOC, Type)

→ Reactors (Type, D, V, L)

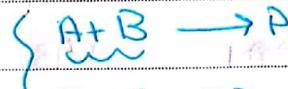
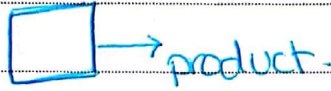
→ towers (L, D, MOC, No. of stages, packing).

☒ struction and synthesis PFD

1 - Define your production rate $\rightarrow$ Batch

continuous

2 Identify input / output structure of the process.



$\rightarrow$ which is cheaper?

* which requires $\uparrow$ energy.

* which has a faster rate of rxn?

* ...

3. define if you have recycle / bypass / purge.

$\rightarrow$ purge:- we don't want the accumulation.

$\Rightarrow$ by pass :- to make the process running

4. Identify the general structure of the separation units.

$\rightarrow$ reactors before separation tools.

$\rightarrow$ light component separated first.

5. Identify whether heat exchanger networking is possible or NOT.

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① Defining production rate:-

BATCH OR CONTINUOUS ?

1. size (production rate) $\rightarrow$ ~~small~~

< 500 ton / year $\Rightarrow$ BATCH

> 5000 ton / year $\Rightarrow$ CONTINUOUS.

processing? → NO FIXED SOLUTION.

Sep 26th, 2017

No. MANY POSSIBILITY.

2- product quality and accountability.

* what is the purity we are needing?

ultra pure ⇒ BATCH

activated
Carbon?
→ stay
→ Remo-
ving
ODAR

3- Operational Flexibility.

⊕ needing to produce different products ⇒

(multiple More than one product ⇒ BATCH Production)

→ Sep 28th, 2018

4- process efficiency

→ continuous process is more efficient.

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5- maintenance and more Labor cost.

→ Batch → each cycle need maintenance // more Labor cost ^{High}
BUT EASILY MAINTAINED

→ continuous →

maintained once a year, BUT HARD MAINTAINCE ~

Batch is preferred when maintenance is required regularly.

6- Feed ^{raw materials} stock availability

Safe side → BATCH

without

* if feed stock available regularly w/o hazard ⇒ Continuous.
if NOT ⇒ Batch.

7- Product demand.

→ Product needs in Seasonal ⇒ BATCH || always needy ⇒ continuous.
Highly demand

~~example:~~

2. if impurities Fault poison the catalyst → separate.
3. " " From hazardous materials → separate.

Example: production of isocyanate:

must be ultra pure. $\text{CO} + \text{Cl}_2 \rightarrow \text{Phosogen (very toxic gas)}$
 Phosogen + Amine → isocyanate. (desired product)

slightly toxic. CO
 Cl_2
 Amine

P₁

$\text{P}_1 = \text{CH}_4 + \text{H}_2\text{O} \rightarrow \text{CO} + 3\text{H}_2 \rightarrow \text{Can react easily with Cl.}$

$\text{CH}_2 + \text{H}_2 = \text{HCl}$ ← *Side reaction

problem:- excess HCL

low conversion phosogen.

∴ High need to separate feed.

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4. Addition of material to stabilize product

↳ addition of hydrazine to boiler feed water.

so no side rxn occur due to high purity water.

5. Addition of inert materials to control exothermic RXN

↳ production of acrylic acid

$\text{C}_3\text{H}_6 + \frac{3}{2} \text{O}_2 \rightarrow \text{C}_3\text{H}_4\text{O}_2 + \text{H}_2\text{O}$ → only this produce acrylic acid

if it was C_3H_6
 O_2

ultra pure:- $\text{C}_3\text{H}_6 + \frac{5}{2} \text{O}_2 \rightarrow \text{C}_2\text{H}_4\text{O}_2 + \text{CO}_2 + \text{H}_2\text{O}$ ⊗

Side RXN ↗ $\text{C}_3\text{H}_6 + \frac{9}{2} \text{O}_2 \rightarrow 3\text{CO}_2 + 3\text{H}_2\text{O}$ ⊗

↳ we need less purity.

3 reactions in nature.

≡ must introduce inert.

→ the inert is usually something from side products;
(anyways small amounts of it will occur).

↳ introduce steam with the feed.

the inert should NOT react with reactants but should effect the reaction.

⑥ CONCLUSION:-

introducing materials to the feed.

1. controlling Rxn rate → add.
2. Toxics, poisonous, hazardous → DO NOT add.
3. economical and profitable analysis.

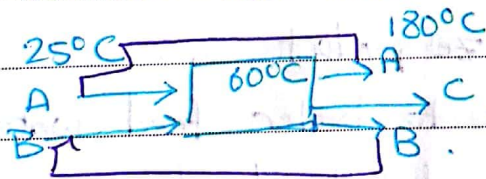
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THIRD: WHEN TO RECYCLE / BURGE / BY PASS ... etc..

THE RECYCLE STRUCTURE OF THE PROCESS:

GOAL:

Recover valuable raw material and energy from the process.



* minimize energy.

to what condition should I recycle A/B?

Reminder: conversion type:-

1. single pass conversion $x = \frac{\text{mol. of reacted species}}{\text{mol of inlet feed}} = \frac{F_{A0} - F_A}{F_{A0}}$

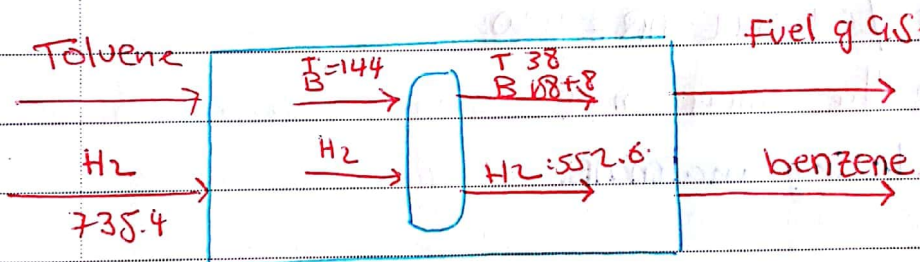
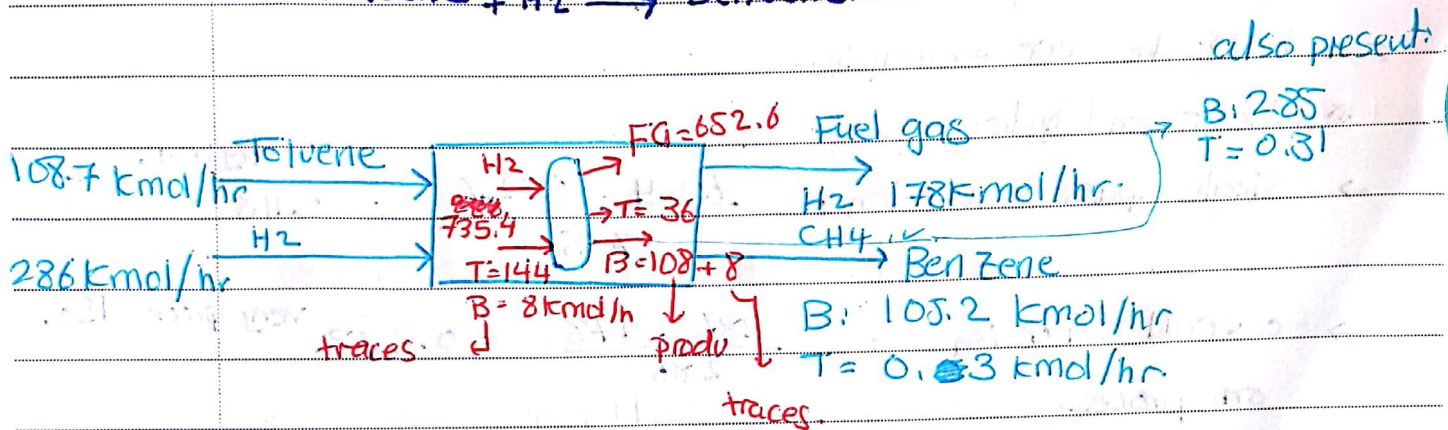
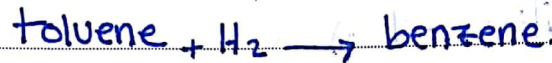
2. over all conversion (process)

≡ $X_{\text{overall}} = \frac{\text{Rxnt consumed}}{\text{Rxnt Fed to process}}$

must be close to 100%.

$$3. \text{ yield} = \frac{\text{mol of desired product}}{\text{mol of limiting reactant Fed}}$$

Example: production of Hydrodealkylation of toluene:



Single pass conversion $X = \frac{F_{T0} - f}{F_{T0}} = \frac{144 - 36}{144} = 0.75$

overall conversion: $X_{\text{overall}} = \frac{108.7 - 0.61}{108.7} = 0.993$

► optimum values:

NOT necessary higher conversion is better.

↳ other things count.

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Conclusion:

1. the lower the single-pass conversion, the greater the recycle must be
2. the lower the single-pass conversion, this leads to higher equipment sizing, and higher utilities cost =).

← back to our example:-

* check on hydrogen:-

→ Single pass conversion = $\frac{735.4 - 652.6}{735.4} = 0.113$ ^{very low.}

→ overall conversion on process = $\frac{286 - 178}{286} = 0.378$ ^{very poor also.}

❖ BOTH SINGLE PASS/OVERALL ARE LOW:-

How to increase the conversion??

* other wise you're losing materials.

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* 3 Ways to Make use of hydrogen.

1. Separate, purify then recycle. $\uparrow$ production rate $\uparrow$ conversion

BP suggestions? $\rightarrow$ advantages? $\rightarrow$ disadvantages?

(1) $\rightarrow$ distillation

$\rightarrow$ bp hydrogen = -253°C need to add amt of utility.

bp $\text{CH}_4 = -161^\circ\text{C}$

NOT FEASIBLE

used for separation

(2) $\rightarrow$ absorption

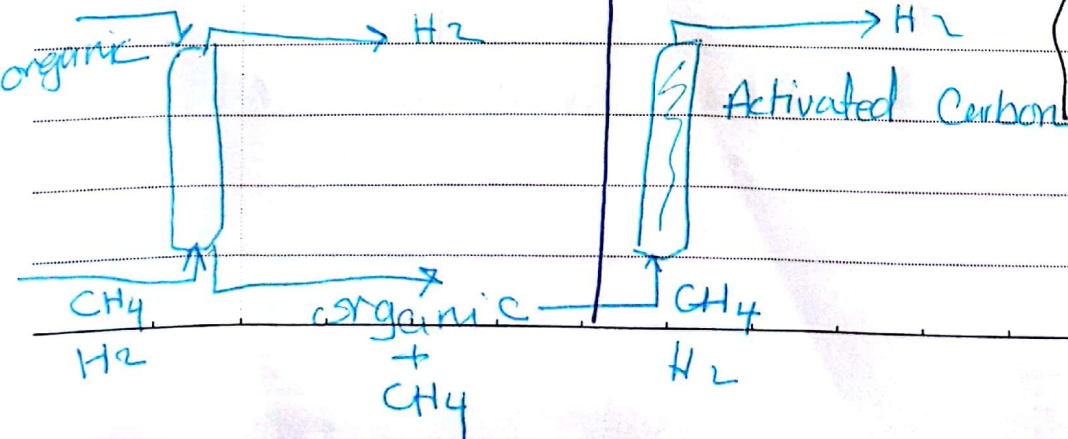
(3) Adsorption

⑥ This process could be

is it Feasible? ✓

cost $\downarrow$
regeneration $\downarrow$

then a regeneration unit.



October 3rd.

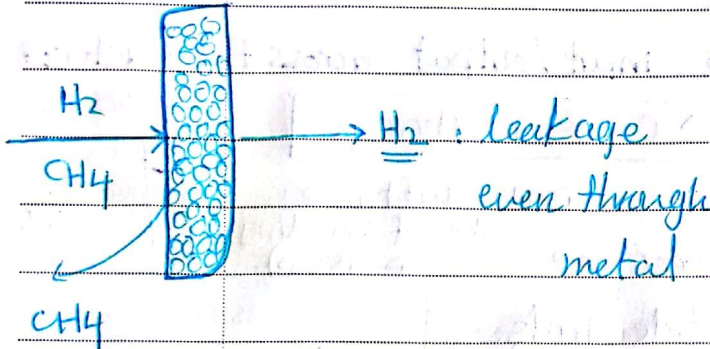
No.

④ membrane separation

it's a Feasible process

however, it may need a

* Large utility cost:



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after all separation.

2. Recycling the feed (unreacted) and product with purging.

Lowly effecting factors:-

1. how much should you purge to prevent building up products in the process [CH₄]



→ require trail/error

3. Recycle the (unreacted) feed and product without purging.



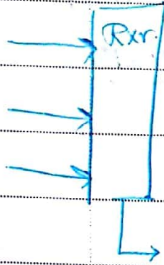
Because I need the desired product, D, I have to max, the concentration of the undesired product, B, in the process to shift the rxn through the D.

fight forever don't let go

* October 5th

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Recycle stream / input, output (multiple)



multiple input/output across the reactor:-



← Rxn shift back (bc of building up Temp)
how to overcome?



Rxn في قاع

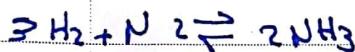
* introduce side feeds

لأنه في قاع درجة الحرارة عالية

it maintain Rxn under control

في القاع تكون هناك زيادة في الضغط

* example:- production of NH_3



* Single pass conv $\Rightarrow$ 15%

* overall conv $\Rightarrow$ 48%

introducing $H_2 + HL \rightarrow T \uparrow$

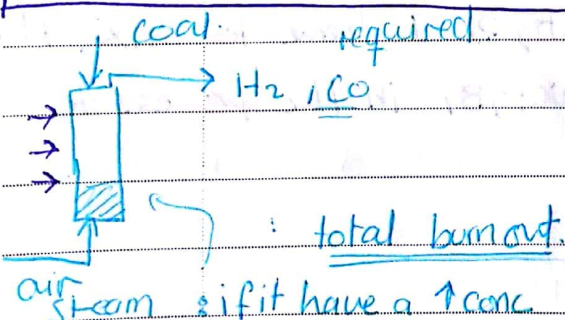
will shift rxn towards reactants.

* introduce one of Rxns at a different location.

2. controlling Reactor Temp

to avoid Rxn temp run away from a set point.

3. concentration controlling

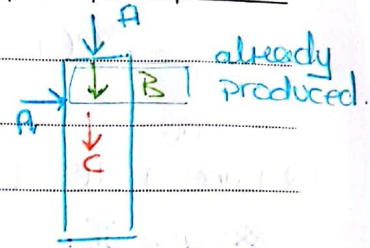


run a multiple feed

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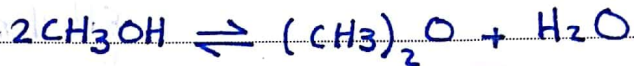
4. optimizing conditions for multiple Rxn

$A \rightleftharpoons B \rightleftharpoons C \rightarrow D$: must enter dose by dose to maintain product.



Example it is desired to produce dimethylether (DME) from methanol. $\Rightarrow$ (in all cases) : Production rate = 50000 ^{ton/}yr

Price :



DME = \$0.95/kg

MeOH = \$0.22/kg

► (1) Alternative : production of DME from

pure methanol. \$0.22

methanol
↓
pure mix alcohol

► (2) " " " " " "

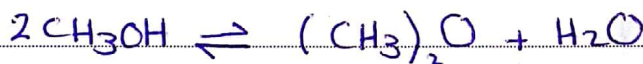
→ methanol : 88%

mix alcohol. \$0.25

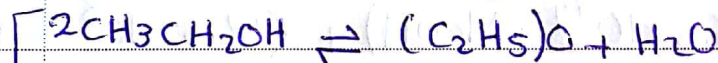
ethanol : 1%

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BUT ethanol will react to give products



DME : \$0.57.



DEE : \$1.27/kg.



Eth. : \$0.57/kg.

NOTES :

[DEE/DME] are easily to separate by one distillation column but may produce isotope.

[DEE/Eth] Save the cost of one distillation column.

* Case: MeOH $\rightarrow$ DME (pure) (production rate)

Desired product: (DME) = 50,000 ton/year.

Mwt (DME) = 46 kg/mol.

$$\text{no of mol} = \frac{50 \times 10^6}{46} = 1.08 \times 10^6 \frac{\text{kmol}}{\text{yr.}}$$

Feed $\Rightarrow$ Cost

Cal $\rightarrow$ feed
Cal $\rightarrow$ output.

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⊗ Required Methanol:-

$$= (2)(1.08 \times 10^6) = 2.174 \times 10^6 \frac{\text{kmol}}{\text{year}}$$

$$n_{\text{DME}} = 2.174 \times 30 = 6.5 \times 10^7 \text{ kg/y.}$$

$$\text{cost MeOH} = 10.22(6.5 \times 10^7) =$$

$$\text{Price DME} = (50 \times 10^6)(0.95) = \$47.5 \text{M}$$

$$\$14.3 \text{M.}$$

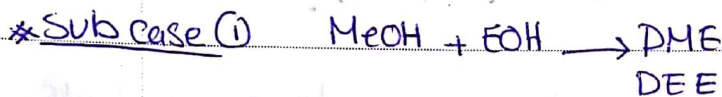
*Case II: mix alcohol

$$\text{profit} = \$33.2 \text{ M/y}$$

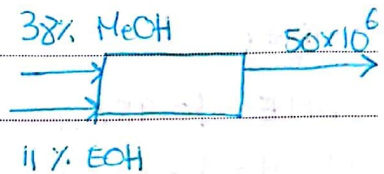


DEE

Ethylene.



DEE.



Cost of Feed: no. of mol. of DME

$$\text{MeOH} = (2)(1.08 \times 10^6 \text{ kmol/y}) = 2.16 \times 10^6 \text{ kmol}$$

88% of feed ← originally represents 88%.

$$\text{EOH} = (2.16 \times 10^6) \times \left(\frac{11\%}{88\%} \right) = 0.27 \times 10^6 \text{ kmol EOH/y}$$

$$= 12.5 \times 10^6 \text{ kg EOH/y}$$

total mass flow rate

$$\text{MeOH} + \text{EOH} = (6.5 \times 10^7 \text{ kg MeOH}) + (12.5 \times 10^6 \text{ kg EOH})$$

test

$$\Rightarrow \text{Cost of Fixed Feed} = (\text{total mass})(0.25) = \$19.4 \text{ M/y}$$

$$\text{Price DME} = \$47.5 \text{M}$$

$$\text{DEE} = (10.1 \times 10^6)(\$1.27)$$

$$\text{price of Feed} = \$12.8 \text{ M/y million.}$$

$$\Rightarrow \text{profit} = 47.5 + 12.8 = \$60.3 \text{M}$$



Assuming complete conv.

to DEE.

mass DEE =

$$\text{mass x production} = \frac{0.27 \times 10^6}{2} = 0.135 \times 10^6$$

$$\Rightarrow \text{profit (MeOH} \rightarrow \text{DME)}$$

(EOH $\rightarrow$ DEE) only

$$= \$60.3 \text{M} - \$19.4 \text{M} = \$40.9 \text{M}$$

Best cost scenario

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*Sub case ② Mixed Feed.

MeOH $\rightarrow$ DME

EOH $\rightarrow$ Ethylene only.

$$\begin{aligned}\text{max ethylene rate} &= 0.2718 \times 10^6 \text{ kmol/y} \\ &= 7.6 \times 10^6 \text{ kg/y}\end{aligned}$$

$$\text{price} = (7.6 \times 10^6)(0.57) = \$4.3 \text{ M/y}$$

total selling price = (DME + Ethylene)

$$= \$47.5 \text{ M} + \$4.3 \text{ M} = \$51.8 \text{ M}$$

$$\Rightarrow \text{profit} = \$51.8 \text{ M} - 19.4 = \boxed{\$32.4 \text{ M}}$$

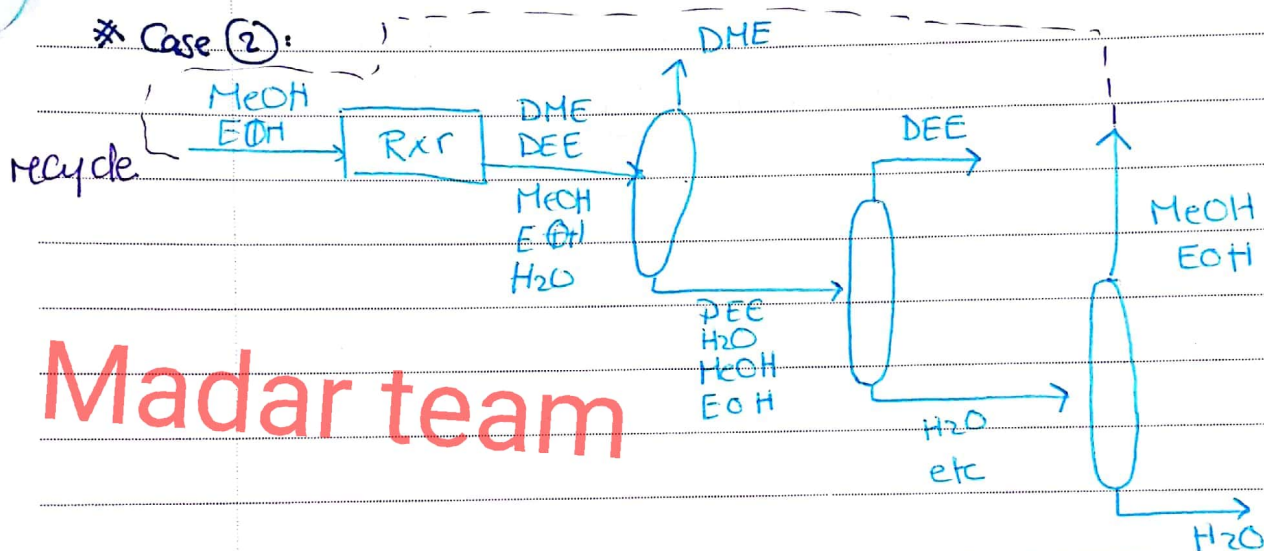
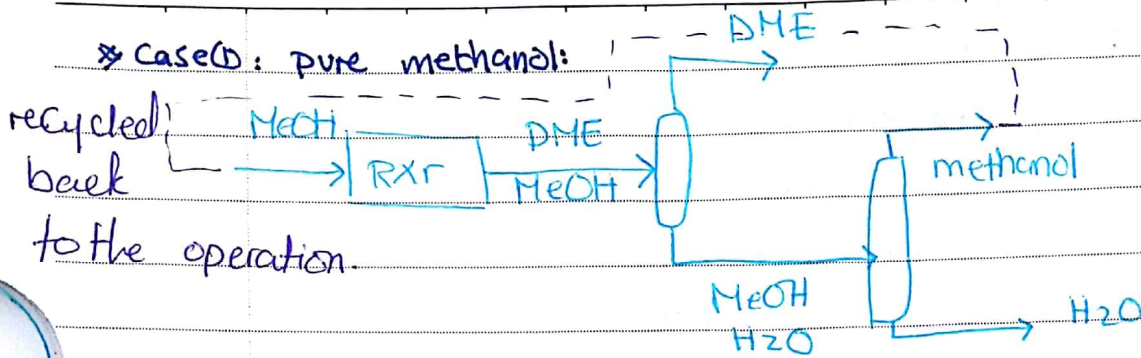
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→ rshawabk@gmail.com.

→ communication

October 10th

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→ the ~~the~~ third case requires a (more) distillation column, gives less products economically
∴ Disgarded.

⇒ suppose that you went with first case:-

Tracing Chemicals Through PFD, P&ID:-

① Reactants: start with Feed and trace chemicals toward Rxn

② products: " " products " " " backward toward Rxn.

Tactics For Tracing chemicals:

→

Simo : start your calculation from the back

miso : " " " "
" beginning

$$n_i = n_i^0 + \sum v_i \xi_i$$
$$2 \text{ MeOH} \rightarrow \text{DME} + \text{H}_2\text{O}$$
$$2 \text{ECH} \rightarrow \text{DEE} + \text{H}_2\text{O}$$
$$2 \text{ EOH} \longrightarrow \text{ETit} + \text{H}_2\text{O}.$$

3

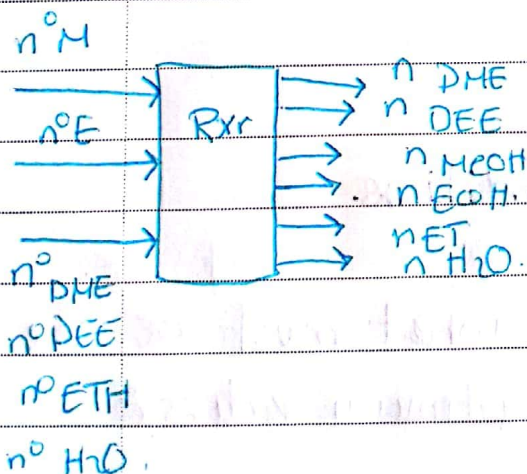
3.

 $\int_{-\infty}^{\infty}$
$$\text{selectivity} = \frac{\text{desired}}{\text{undesired}}$$

* Conversion $\Rightarrow$

$$X_M = \frac{n_M^0 - n_M}{n_M^0}$$

$$X_E = \frac{n_E^0 - n_E}{n_E^0}$$



$$n_{\text{MeOH}} = n^{\circ}_{\text{Meth}} - 2 \xi_1$$

$$n_{\text{EOH}} = n^{\circ}_{\text{EOH}} - 2\sum_2 - 2\sum_3$$

$$n \text{ DME} = n^0 \text{ DME} + \sum_i$$

$$n_{DEE} = n_{DEE}^0 + \int_2$$

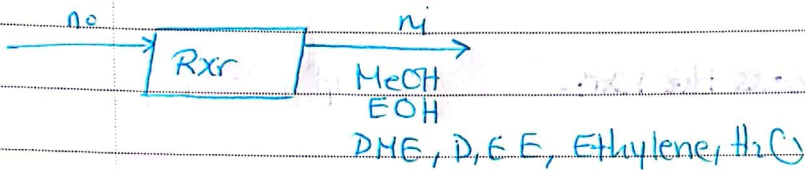
$$n_{H_2O} = n^0_{H_2O} + \sum_1 + \sum_2 + \sum_3$$

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30 40 No. 20 10
 $\Rightarrow$ Midterm + Final + project + HW/Quizes.

cont'd

$$n_i + n_{i^0} + \sum v_i \xi_i$$



$$n_{\text{MeOH}} = n^0_{\text{MeOH}} - 2\xi_1$$

$$n_{\text{EOH}} = n^0_{\text{EOH}} - 2\xi_2 - 2\xi_3$$

$$X = \frac{n^0_{\text{MeOH}} - n_{\text{MeOH}}}{n^0_{\text{MeOH}}}$$

$$S = \frac{\xi_2}{\xi_3} = \frac{\text{mols of desired rxn}}{\text{undesired}}$$

$$n_{\text{Eth}} = n^0_{\text{eth}} + \xi_3$$

$$n_{\text{DEE}} = n^0_{\text{DEE}} + \xi_2$$

$$n_{\text{DME}} = n^0_{\text{DME}} + \xi_1$$

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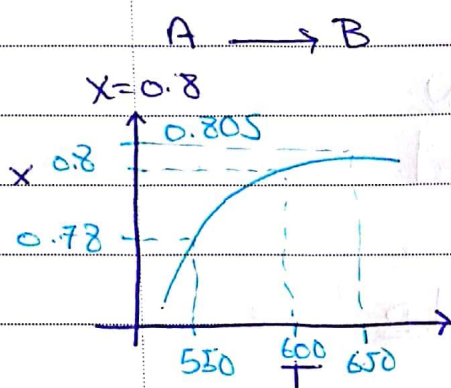
$$n_{\text{H}_2\text{O}} = n^0_{\text{H}_2\text{O}} + \xi_1 + \xi_2 + \xi_3$$

$\rightarrow$ understanding of process conditions.

1. Temp.
2. pressure.
3. composition

$\Rightarrow$ what must be the optimum values?

Introductory example:-



$\rightarrow$ usually runs on 600°C = 0.805.

$\rightarrow$ a designer suggested to run it on 550°C, not great difference on y-axis.

great difference on x-axis.

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Pressure:

1. ALWAYS RUN THE PROCESS OF A PRESSURE RANGE FROM [1-10] atm

Reasons why we need to operate process at higher pressure?

$P > 10$ atm $\rightarrow$ increase density $\rightarrow$ more collision, more production

A $\rightleftharpoons$ B

$\rightarrow$ Faster Rate of RXN.

$\uparrow [A] \rightarrow$ shift rxn toward desired B | Lecha's principle

$\rightarrow$ For light component $\rightarrow$ need it to be in liq phase

$\otimes \uparrow$ pressure to maintain liq phase.

1. Favours equilibrium conversion
2. $\uparrow$ rate of reaction
3. maintain a liquid phase (if needed)

what are the consequences?

1. require thicker wall of the equipment [$\uparrow$ cost]
2. need for expensive compressor and utilities =)

Reasons why we need to operate at

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Low pressure?

$P < 1 \text{ atm}$

Production of crude oil

1. have a favorable equilibrium RXN.
2. maintain a gas phase if needed.
3. Temp Sensitive materials.
4. run the separation at lower bp $\sim \oplus$.

Vacuum distillation.

↳ generate volume (vapor) $\rightarrow$ increase the size.

Sequences?

1. Requires large equipment and special design.
2. air-leakage into equipment.

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► effect of Temperature:

Temp. range from $40 - 260^\circ\text{C}$

→ why do we need to $\uparrow$ Temp?

1. $\uparrow$ Temp will $\downarrow$ the life time of the equipment.

example: the max allowable tensile strength of a typical Carbon steel (CS) and stainless steel at 25°C , 400°C , 450°C as follows:-

Temp.	25	400	450.
CS	1190	970	170
SS	1290	1290	430

Temp range 25-400

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$$CS = \frac{1190 - 970}{1190} = 0.18$$

SS = 0

BUT SS IS more expensive / but safer.

smoother process

Temp range 400°C → 550°C → you have to calculate all values.

$$CS = \frac{970 - 170}{1190} = 0.67$$

$$CS = 0.18 + 0.67 = 85\%$$

1190 initial tensile strength Rupture

$$SS = \frac{1290 - 430}{1290} = 0.67$$

BUT SC → 0 + 0.67 = 67%

⇒ RUNNING AT TEMP ↑ 400°C ⇒ DO NOT USE CARBON STEEL AS A MATERIAL OF CONSTRUCTION.

⇒ Reasons why to go to higher temp? → all reasons (referring to product)

1. conversion
2. rate of RXN
3. improve selectivity.

⇒ consequences?

1. need ↑ utility
2. special material of construction.

⇒ why do we need to run the process at lower temp < 400°C?

1. Favour equilibrium conversion
2. ↑ Rte of RXN
3. ↑ selectivity
4. maintain liq. phase

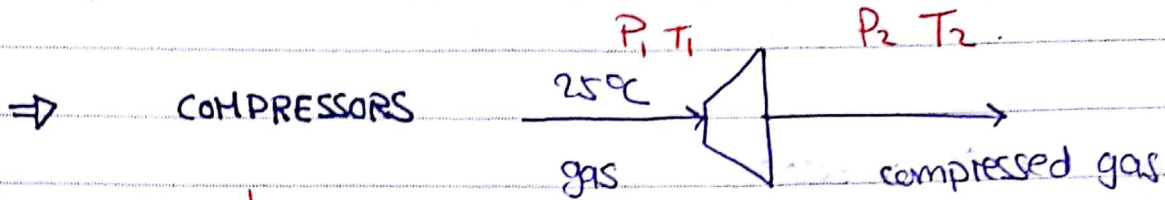
⇒ consequences?

1. ↑ utility cost.
2. Requires special materials of construction.

Madar team

Conditions to run winner

equipments (pumps/ compressors) & heat exchangers, valves, mixers / ..)



$$P_2 \gg P_1$$

$$T_2 \gg T_1$$

1. ALWAYS RUN compressor at $\frac{P_2}{P_1} < 3$

⇒ Guidelines ⇒ if ↑ P is needed?

∴ USE MULTI-STAGE COMPRESSOR with inter



stage cooler.

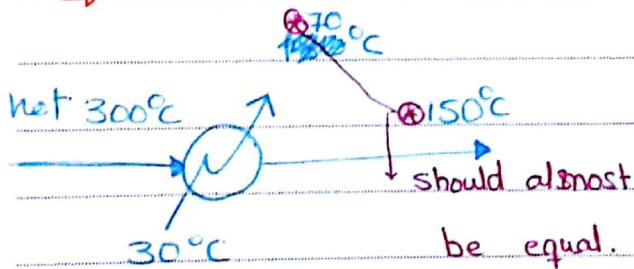
⇒ to keep temp.

* correction factor

⇒ HEAT EXCHANGERS,

log mean temp difference

⇒ Guidelines: ALWAYS RUN HE of LMTD < 100°C



$$LMTD = \frac{(300 - 70) - (150 - 30)}{\ln \left(\frac{300 - 70}{150 - 30} \right)} = 169^\circ\text{C}$$

$$\ln \left(\frac{300 - 70}{150 - 30} \right) = 169^\circ\text{C}$$

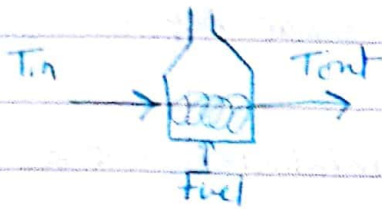
$$\uparrow LMTD = 169^\circ\text{C} > 100^\circ\text{C}$$

⇒ Reason: to save energy.

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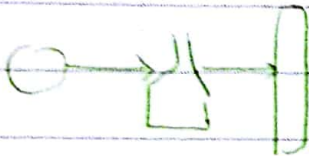
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⇒ PROCESS HEATER.



$T_{out} > T_{steam}$

@ heater always comes before distillation column.



1. to preheat steam
2. keep the process stable

⇒ VALVES:

⊗ to ↑ P

⇒ IF ↑ ΔP is needed ⇒ turbine

⇒ turbine



why?

1. get work

⊗ want control

2. Save energy.

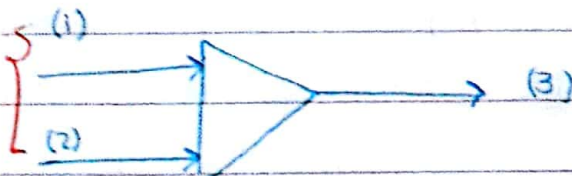
⇒ valve

Special case: IF YOU NEED THE PROCESS TO BE FULLY CONTROLLED

⇒ Valve.

⇒ MIXER:

you might need ^{exchanger} throttling value. ⇒ TO MAINTAIN the



both streams at

close → Temp
P

like mixing

← has minor effects ← compositions.

strong acid / strong base

of ultra purity.

* Mixer? ⇒ Maintain same conditions ⇒ ^{exchanger} throttling values.

October 15th
Today I'm going
to try a little
harder.

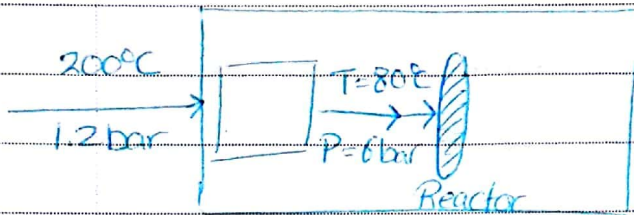
No.

* Example:

Feed stream containing (N_2) at $200^\circ C$ and 1.2 bar is fed to a process and its conditions is required to be $80^\circ C$, 6 bar before entering the Rxx.

Suggest a process conditions and design to minimize heat loss and save equipment life

Assume compressor $\epsilon = 0.7$

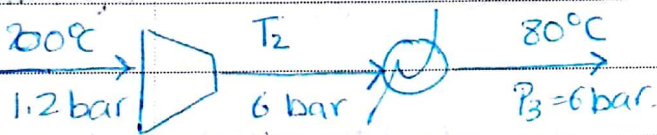


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Assuming long lifetime of equipments

[CASA.I]

introduce the stream as is w/o cooling $\Rightarrow$ Then COMPRESS $\Rightarrow$ then COOL.



Design Equation
For compressor WORK

ratio 3 to maintain
the life

$\Rightarrow$ so assumed

negligible life

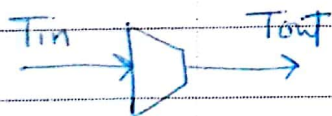
$$W = RT_m \left(\frac{\gamma}{\gamma - 1} \right) \left[\left(\frac{P_{out}}{P_{in}} \right)^{\frac{\gamma}{\gamma - 1}} - 1 \right] / \epsilon$$

ratio $\gamma = \frac{C_p}{C_v} = 1.4$

C_p : Specific heat const. pressure

C_v : " " " volume

$\Rightarrow$ how to calculate T_{out} ?



$$T_{out} = T_{in} \left(1 + \frac{1}{\epsilon} \left[\left(\frac{P_{out}}{P_{in}} \right)^{\frac{\gamma - 1}{\gamma}} - 1 \right] \right)$$

Energy balance for compressor

October 17th

No.

$$q = m C_p \Delta T$$

→ Solution

$$T_{int} = (200^\circ\text{C} + 273) \left(1 + \frac{1}{0.7} \left[\left(\frac{6 \text{ bar}}{1.2 \text{ bar}} \right)^{\frac{1.4-1}{1.4}} - 1 \right] \right) = 867 \text{ K}$$

specific per unit mole/mass

$$= 594^\circ\text{C}$$

$$q = C_p \Delta T$$

$$= (3.5 R) [594 - 50^\circ\text{C}] = 1497 \text{ kJ/mol}$$

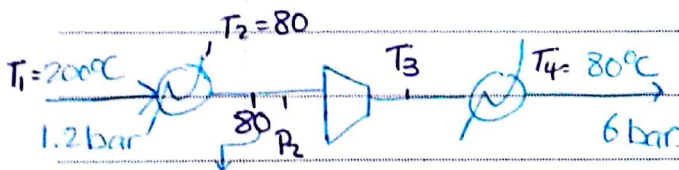
$$W = 11470 \text{ kJ/mol}$$

$$T_2 = 594^\circ\text{C}$$

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CASE II

introduce the stream into a cooler → then compressor → then cooler again



$$q_2 = C_p \Delta T = 3490 \text{ kJ/mol}$$

$$P_2 = P_1 = 1.2 \text{ bar}$$

⊗ NOT a wise choice

to go below zero

around ambient

Assume 80°C

how about T_{int} ? → to find T_m ! Assume it.

↓

$$W = 8560 \text{ kJ/mol}$$

for herostics → valid assumptions

based on experiences.

⊗ try other assumptions

that are below that.

$$T_3 = 374^\circ\text{C} \quad P_3 = 6 \text{ bar}$$

$$q_3 = 8550 \text{ kJ/mol} \quad T_4 = 80^\circ\text{C}$$

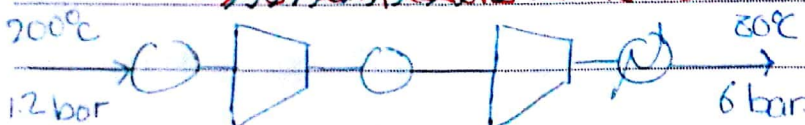
$$q_{\text{total}} = 12040 \text{ kJ/mol}$$

CASE III

cooling → compression → cooling → compression → cooling.

Extrem cooling.

~~3 compressors~~ ~~for 3 stages~~



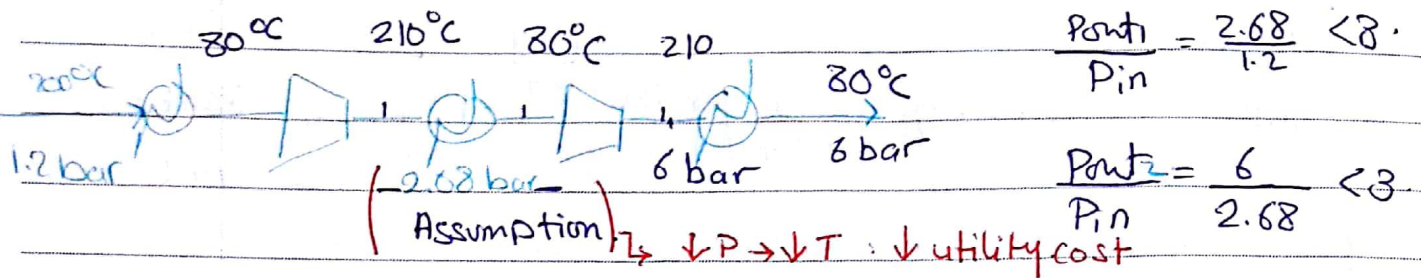
$$\frac{P_{out}}{P_{in}} = \frac{6}{1.2}$$

⇒ after 2nd compressor → one should assume

October 19th

No.

a value



objective: $\downarrow w$ of compressors :) $\downarrow$ damage.

$$w_1 = 3780 \text{ kJ/mol}$$

$$w_2 = 3780 \text{ kJ/mol}$$

$$q_1 = 3480$$

$$q_2 = 3780$$

$$q_3 = 3780$$

$$q_T = 11050 \text{ kJ/mol}$$

$$w_T = 7560 \text{ kJ/mol}$$

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when comparing → no. of equipment
 → Fixed Capital Cost
 → Utility (q, w)

HM on:-

→ Case(3) excel spread sheet:-

cooler compressor cooler compressor cooler

T

P

q/w

→ change values for best results.

- B assume furnace efficiency
- (1) Given that duty Q_T
 - (2) Fuel value
 - (3) Air/Fuel Ratio
 - (4) Temp. of [inlet air, fuel, steam]
 - (5) amount of atomizing steam
 - (6) tube diameter
 - (7) exposed length of tubes.

procedures:-

- [1] Assume tube surface, use chart to get T
- [2] Calculate the heat liberated by fuel.

$$Q_F = \frac{\text{Fuel required duty}}{\eta}$$

- [3] Calculate amount of fuel required

$$m_F = \frac{Q_F}{\text{Fuel value} \rightarrow \text{given.}}$$

- [4] get the air value

$$m_{\text{air}} = \left(\frac{\text{air}}{\text{Fuel}} \right) \cdot m_F$$

- [5] Assume 25% excess air

$$m_{\text{air new}} = 1.25 (m_{\text{air}})$$

- [6] Calculate heat inlet by air

$$q_{\text{air}} = m_{\text{air new}} C_{p_{\text{air}}} (T_{\text{in air}} - T_{\text{ref}})$$

Assume $T_{\text{ref}} = 60^\circ\text{F}$

>> Heat exchangers

1 Area: $10 \text{ m}^2 - 1000 \text{ m}^2$

$< 10 \text{ m}^2 \rightarrow$ double pipe exchanger.

2 tube diameter: $3/4'' (1.9 \text{ cm})$
($1.9 - 2.5 \text{ cm}$) → Range.

3 length: $2.5 \text{ m} \rightarrow 6.5 \text{ m}$

4 pressure: vacuum - 150 bar.

5 Temp: $(-200^\circ\text{C}) - 400^\circ\text{C}$

→ correction Factors

⇒ HEAT EXCHANGER: $q = UA(LMTD)F$

always use $F=0.9$

other wise :- use charts based on LMTD

Standard tube diameter: $3/4'' \Rightarrow$ triangular pitch spacing.

$1'' \Rightarrow$ use 4.9 m Long.

⇒ Shell side: usually start

30 cm → Can accommodate $100 \text{ ft}^2 (9.3 \text{ m}^2)$ ⇒ Area less

60 cm → $400 \text{ ft}^2 (37.2 \text{ m}^2)$ than 100 m^2

90 cm → $1100 \text{ ft}^2 (102 \text{ m}^2)$

above:-

⇒ WHEN TUBE/SHELL SIDE?

1 Tube side.

1. Corrosive. 2. Fouling. 3. Scaling. 4. HP fluid.

Shell side

1. viscous fluid.
2. condensing fluid.

all exchangers $\Rightarrow$ Pressure drop
 $\Rightarrow$ boiling fluids
 other

0.1 bar

0.2 - 0.6 bar $\Rightarrow$ Temp.(1) min. temp. difference = 10°C (liquid)(2) " " " Refrigeration = 5°C (3) air cooler = 22°C $\Rightarrow$ Cooling WATER $\rightarrow$ inlet temp $\approx 20^{\circ}\text{C}$ ($15-25^{\circ}\text{C}$) $\rightarrow$ $T_{\text{out}} - T_{\text{in}} \approx 15^{\circ}\text{C}$ $\star$ initial guess for over all heat transfer coefficient U !1. water-liq 850 $\text{W/m}^2\text{C}$

2. condenser 850 "

3. liq-liq 290

4. liq-gas 60

5. gas-gas 30

6. Reboiler 1140

Madar team

→ window

File | tools | help

Simulation Basis Manager.

1. file → new

case

Template

column

comp fluid package.

Components

(to work on)

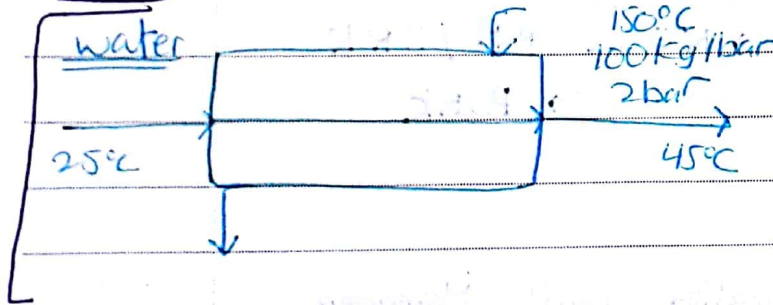
master component list.

Search

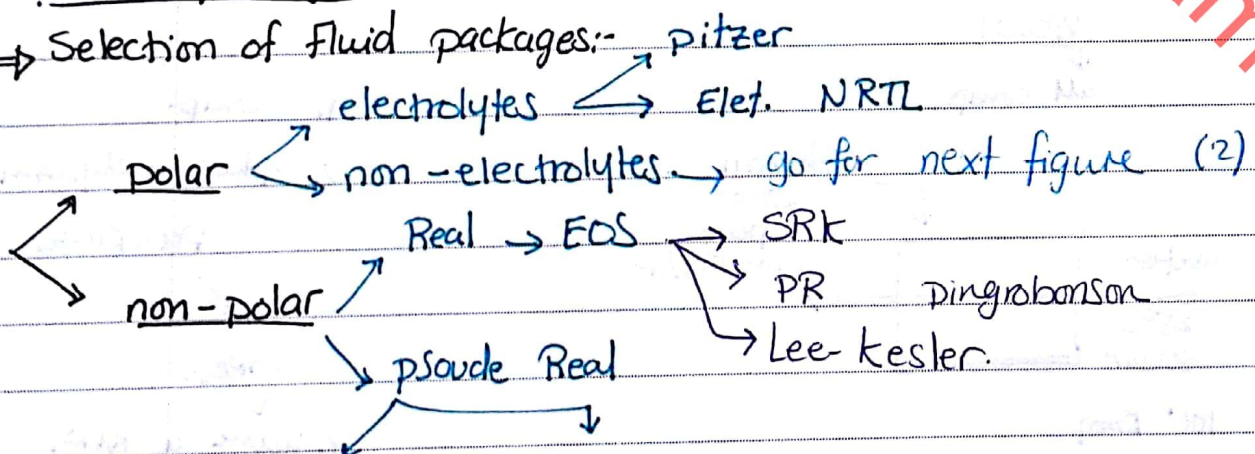
water

add pure comp

Case:-

Fluid Package:-

⇒ Selection of Fluid packages:-



Pressurized:

vacuum: - Brown k-10.

- Chao-Seedler.

- Grayson

- Brown k-10.

* October 24th

No.

polar

Non-electrolyte (Figure 2)

Parameters

Parameter

LL yes

NRTL

UNIQUAC

NO

wilson

NRTL

UNIQUAC

$P < 10 \text{ bar}$

ij

yes available.

parameter unknown

↓

yes → uniuack

LL ↔ NO → unifac.

$P > 10 \text{ bar}$

ij

yes

PR, SRK

NO

PR/SRK

PSRK

[New] → [Case] → Simulation basis Manager

[View]

all comp.

1. comp.

2. thermodynamic

Package

water

25°C
2 bar

n-hexan

150°C
2 bar
100 Kg/hr

100 kmol

Green

3. have a RAN?

use water to cool
cooler

4. Enter simulation

↓ environment.

equipments show.

* hysys:-

*(Duty)

2 Streams in/out

↓ gives you

water stream 1, 2, 3, 4

exist. stream Temp.

WORKSHEET

↓ provides all data.

* Parameters

$$\frac{\sum Q}{x_{cp} A_{cp} f} = 0.173 \left[\left(\frac{T_G}{100} \right)^4 - \left(\frac{T_S}{100} \right)^4 \right] + 7 [T_G - T_S]$$

Assume Furnace efficiency η

→ given that the duty Q_T (1) Fuel values (2)

Air / Fire /

(3)

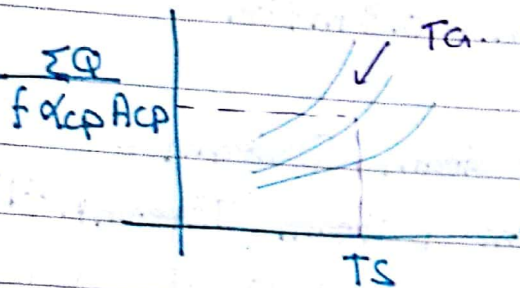
Temp at inlet gas (4)

Fuel (5) steam (6)

tube diameter (7)

expected length of tube (8)

[1] Assume T_S , use chart to get T_G



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[2] Calculate heat liberated by fuel: $Q_F = \frac{\text{Fuel required}}{1}$

[3] " amount of fuel required: $m_F = \frac{Q_F}{\text{Fuel value given}}$

[4] get the air value $m_{air} = \left(\frac{\text{air}}{\text{Fuel}} \right)_{\text{ratio}} \times m_F$

[5] Assume 25% excess air, $m_{air \text{ new}} = 1.25 m_{air}$

$$Q_{air} = m_{air} C_{p \text{ air}} (T_{air \text{ in}} - T_{air \text{ F}})$$

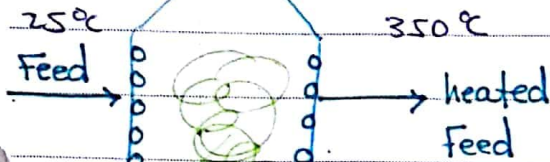
- Furnace Design.

$H_2O \rightarrow R\&N$
 $H_2O \rightarrow \text{Steam}$
 CO_2
 excess pressure

each has it's own partial pressure.

$$Q_{AIR} = m_{AIR} C_{p,AIR} (T_{AIR,IN} - T_{REF})$$

Assum $T_R = 60^\circ F$



amount of atomizing steam

$$= \frac{0.31b}{lb \text{ Fuel}}$$

$$m_{\text{Steam}} = 0.3 m_{\text{Fuel}}$$

high circulation
 Steam Q_s
 + Fuel Q_f
 + Air Q_a

to [increase/decrease heat transfer] heat by steam, $Q_{\text{Steam}} =$

it's usually neglected

$$m_{\text{Steam}} C_{p,\text{Steam}} (T_{\text{Steam},in} - T_R)$$

$$Q_{\text{wall}} = 2\% Q_f$$

$C_{p,avg}$ = Specific heat for reactant

$$Q_{\text{exhaust gas}} = m_{\text{Fuel}} \left(1 + \frac{\text{air}}{\text{Fuel}} \right) C_{p,avg} (T_G - T_R)$$

and non reactant
 AIR



m_f

(inlet mass = outlet mass)

Related to $T_R \rightarrow C_p$

it

$$C_{p,avg} = \sum C_{p,i} X_i$$

$$Q_{\text{net}} = Q_{\text{Fuel}} + Q_{\text{Air}} + Q_{\text{Steam}} - Q_{\text{wall}} - Q_{\text{exhaust gas}}$$

$$\text{number of tubes} = \frac{Q_{\text{total}}}{2\pi r L q_{\text{Flux}}} = N$$

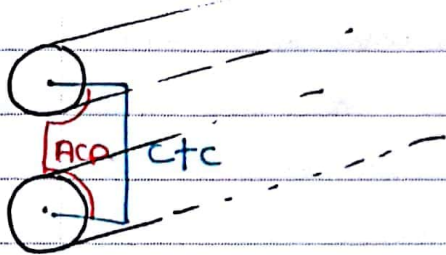
Madar team

October 31th.

No.

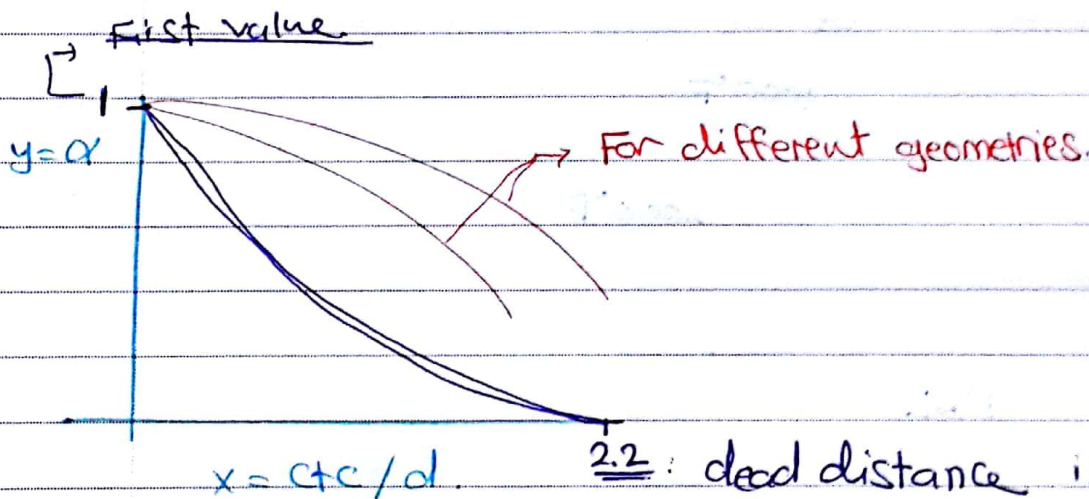
$$\frac{A_{cp}}{\text{cold plane area}} = \frac{C_{tc}}{\text{center to center distance}} \times L \times N$$

⊗ Can't arrange tubes
} Next to each other.
} due to expansion
} in tubes / also
} corrosion.



→ we must multiply using it.

* Calculate α : (fraction) that reduce A_{cp} to real Area.



2.2: dead distance is almost greater than the tubes.

⊗ NOT good to reach.

⊗ Dimension ratio

1-1-1 → 1-1-3

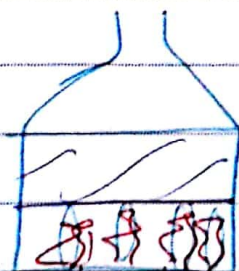
length-width-height.

⊗ mean beam length

$\frac{2}{3} (\text{volume of furnace})^{1/3}$

calculate mean beam length

⊗ what is beam length?



→ maximum gas

(Flame, NO GAS)

} → equation for convection
} → equation for radiation
} → equation for convection

* emissivity *

effect ϵ

boundary

for the gas

October 31

No.

dimension ration

mean beam length

1 - 2 - 1 $\rightarrow$ 1 - 2 - 4

$\frac{2}{3} (V)^{1/3}$

1 - 1 - 4 $\rightarrow$ 1 - 1 - ∞

1 - 2 - 5 $\rightarrow$ 1 - 2 - ∞

1 - 3 - 3 $\rightarrow$ 1 - ∞ - ∞

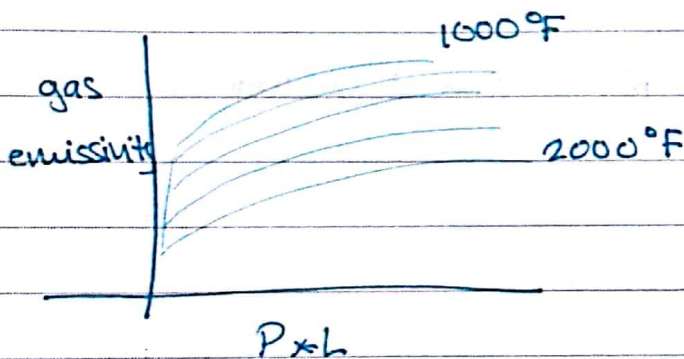
$\Rightarrow$ use smallest dimensions.

$\Rightarrow$ which is (one)

$\Rightarrow$ one is our Ratio.

* Correlation

mean beam length * partial pressure of gas = ϵ



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* after calculating gas $\epsilon \rightarrow$

Calculate the effective Refractive Area: $A_R = A_T - Q/C_p$

* Tubes aren't
near the
wall
there's
a distance
to allow
Reflection

$\Rightarrow$ Obtain

$\frac{\sum Q}{Q/A_c p f}$

$\rightarrow$ need a chart to
apply it

Example: A Furnace is to be designed, for a total duty, of $50,000 \frac{\text{Btu}}{\text{hr}}$.
The overall efficiency is to be 75% oil Fuel with a heating value of ~~17130~~ 17130 btu/lb is used.

excess air is used, with air/fuel ratio of 13.95.

the air is preheated to 400°F

Steam for atomizing the Fuel is used with a ratio of 0.3 lb/lb oil.

the furnace tube are to be $d_o = 5 \text{ in.}$

Center to Center distance = 18.5 in.

with Single row arrangement. The exposed tube length is 38.5 in.

→ the average tube temp. in the radiant section is 800°F

Design the radiant section of a furnace having an average flux of $12,000 \frac{\text{btu}}{\text{hr.ft}^2}$

Assumption will be made in red.

Solution:-

information:

oil is fuel

1. total duty = $50 \times 10^6 \frac{\text{btu}}{\text{hr}}$

2. $\eta = 0.75$

3. Fuel value = $17130 \frac{\text{btu}}{\text{lb}}$

4. air/fuel = 13.95

5. $T_{\text{air}} = 400^\circ\text{F}$

6. Steam = $0.3 \frac{\text{lb}}{\text{lb-fuel}}$

7. $d_o = 5 \text{ in}$

8. C-t-C = 8.5 in

9. tube length = 38.5 in.

10. $T_s = 800^\circ\text{F}$

11. Avg Flux = $12000 \frac{\text{btu}}{\text{hr.ft}^2}$

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No.

Solution:-

Assume = $\frac{\Sigma Q}{A_{cp} \cdot f} = 2 (\text{Avg Flux}) = 2(12000) = 24000$ initial guess

real curve

$\frac{\Sigma Q}{A_{cp} \cdot f} = \frac{24000}{0.57} = 42000$

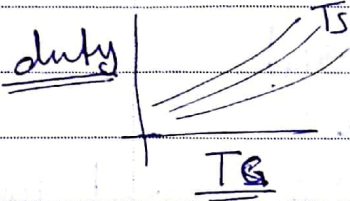
Assume $f = 0.57$

→ go to graph

From Figure

$$T_G = 1700 - 1800$$

$$T_G = 1730^\circ\text{F}$$



→ $Q_{\text{fuel (actual)}} = \frac{50,000,000}{0.75} = 66,670,000 \frac{\text{Btu}}{\text{hr}}$

→ For 100% η

→ $Q_{\text{fuel}} = \frac{50,000,000}{17,130 \times 0.75} =$

$m_F = \frac{66,670,000}{17,130} = 3890 \text{ lb/hr.}$

$m_{\text{air}} = \left(\frac{\text{air}}{\text{Fuel}} \right) m_F = (13.95)(3890) = 54273 \text{ lb/hr.}$

total air flow = $(54273)(1.25) = 67,900 \text{ lb/hr.}$

$Q_{\text{air}} = m c_p \Delta T = (67900)(0.241)[400 - 60] = 5,560,000 \frac{\text{Btu}}{\text{hr}}$

Assume c_p for air = $0.241 \frac{\text{Btu}}{\text{lb}^\circ\text{F}}$

$m_{\text{steam}} = 0.3(3890) = 1167 \frac{\text{lb}}{\text{hr}}$

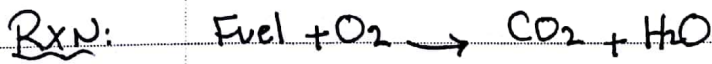
→ $C_p \text{ steam} = 0.46 \text{ Btu/lb}^\circ\text{F @ } 250^\circ\text{F}$

$Q_{\text{steam}} = m_{\text{steam}} C_p \text{ steam } \Delta T$
 $= (1167)(0.46)(250 - 60)$
 $= 101,996 \frac{\text{Btu}}{\text{hr.}}$

$$\leftarrow \dot{Q}_w = \left(\frac{2}{100}\right)(\dot{Q}_F) = 0.02(66,670,000) = 1,333,000$$

November 2nd

$$\dot{Q}_{\text{gas exist}} = m_{\text{fuel}} \left(1 + \frac{\text{air}}{\text{fuel}}\right) C_{p \text{ avg}} (T_G - T_R) \quad \text{No.} \quad 520^\circ \text{R} \rightarrow \text{given in instruction}$$



$$C_{p \text{ avg}} = 0.458 \text{ Btu/lb}^\circ \text{F} @ \text{ } \circledast \text{ avg Temp}$$

$$\text{Avg } C_p = \sum C_{p,i} X_i = 0.458$$

per unit mol
per unit mass

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$$\rightarrow \dot{Q}_{\text{gas exist}} = (13890)(1 + 13.95)(0.458[1730 + 460] - 520) = 44,480,85 \text{ Btu/hr.}$$

$$\rightarrow \dot{Q}_{\text{net}} = \dot{Q}_{\text{Fuel}} + \dot{Q}_{\text{steam}} + \dot{Q}_{\text{Air}} - \dot{Q}_{\text{gas}} - \dot{Q}_{\text{wall}} = 26,517,781 \text{ Btu/hr.}$$

given

$$\dot{q} = 12,000$$

per surface area

⇒ exposed surface area:-

→ defined by design.

$$\Rightarrow \text{total surface area} = 2\pi rL$$

$$\text{for one tube} = 2\pi \left(\frac{5}{12}\right)(385) = 50.4 \text{ ft}^2$$

→ no. of tubes.

$$\dot{Q} = \dot{q} A N$$

$$N = \frac{\dot{Q}}{\dot{q} A} = \frac{26,517,781}{(12,000)(50.4)} = 43.8 \approx 44 \text{ tubes of exposure.}$$

$$A_{\text{cp}} = \text{ctc} \times L \times N$$

$$\left(\frac{8.5}{12}\right) \times 38.5 \times 44$$

$$= 1200 \text{ ft}^2$$

No.

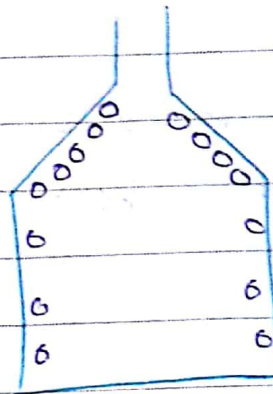
$$A_{cp} \text{ per tube} = \frac{1200}{44} = 27.27 \text{ ft}^2$$

Small
comparing
it

38.5 ft

14.92 ft

40.46 ft



Design based on this ↑

→ Dimension Ratio

$$38.5 : 20.46 : 14.92$$

$$3 : 2 : 1 \text{ approximately.}$$

$$\text{mean beam length} = \frac{2}{3} \sqrt{V \cdot \text{furnace.}}$$

we assumed
one row of
tubes.

→ to calculate α

$$\frac{C-t-C}{OD} = \frac{3.5}{5} = 1.7 \Rightarrow \text{use the chart}$$

$$\alpha = 0.937$$

$$\alpha A_{cp} = (0.937)(1200) = 1124 \text{ ft}^2$$

total area

for all
tubes.

$$\begin{aligned} \text{Refractory surfaces} &= 2[(20.46)(38.5)] \text{ Top/bottom} \\ &+ [14.92 \times 38.5] \\ &+ [9.79 \times 38.5] \\ &+ 2(20.46 \times 14.92) \end{aligned}$$

$$\text{Refractory area} = 3138 \text{ ft}^2$$

effective Refractory area

November 5th/17

No.

$$AR = AT - A'ACP$$

$$= 3138 - 1129 = 2014$$

$$\frac{AR}{A'ACP} = \frac{2014}{1124} = 1.79 \Rightarrow \text{Ratio} \Rightarrow \text{use it on chart}$$

$$\begin{aligned} \text{Dimension} = 1-2-3 \rightarrow L &= \frac{2}{3} \sqrt[3]{V_{\text{furnace}}} \\ &= \frac{2}{3} \sqrt[3]{38.5 \times 20.46 \times 14.92} \\ &= 15 \text{ ft.} \end{aligned}$$

$\Rightarrow$ to calculate ϵ : partial $P_{O_2} = 0.1084$

pressures $P_{H_2O} = 0.1248$

ϵ

gas out $\Rightarrow$ $\underline{PL_{fm}}$

will be calculated
as an average value.

$$PL_{CO_2} = (0.1084)(15) = 1.63$$

mean length

$$PL_{H_2O} = (0.1248)(15) = 1.87$$

Madar team

← back to our example:-

No. .

November 7th

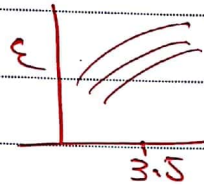
$$P_{CO_2} = 0.1084 \quad P_{H_2O} = 0.1248 \quad L = 15$$

$$P_{CO_2} L = 1.63$$

$$P_{H_2O} L = 1.87$$

$$P_{CO_2} L + P_{H_2O} L = 3.58$$

⇒ Gas emissivity:-



there's the avg.
⇒ using total P

using correlation =
$$\frac{(q_{CO_2} + q_{H_2O})_{T_g} - (q_{CO_2} + q_{H_2O})_{T_s}}{q_b|_{T_g} - q_b|_{T_s}}$$

Flux due difference in temp.

q in absence of gases

multiple with factor

$$\left(\frac{100 - \%}{100} \right)$$

Presena
absence

$$\% = \frac{\text{Partial Pressure } CO_2}{\left(\text{Partial Pressure } CO_2 + \text{Partial pressure } H_2O \right)}$$

$$\% \text{ at } \frac{P_{CO_2}}{P_{CO_2} + P_{H_2O}} = \frac{0.1084}{0.1084 + 0.1248} = 0.464$$

@ $T_g = 173^\circ F$ $q_{CO_2} = 6500$ $q_{H_2O} = 14,500$

@ $T_s = 800^\circ F$ $q_{CO_2} = 650$ $q_{H_2O} = 1950$

* Using correlations: $q_b = 0.173 \epsilon_b \left(\frac{T}{100} \right)^4$ (absence of gases).

$$q_{b_{T_g}} = 0.173 (0.99) \left(\frac{1730 + 460}{100} \right)^4$$

in Rankine.

$$= 39,000$$

$$q_{b_{T_s}} = 4400$$

$$E_G = \frac{(6500 + 14500) - (650 + 1950)}{39000 - 41400} \left(\frac{100 - 8}{100} \right)$$

November 7th

$$= \underline{\underline{0.489}}$$

⇒ USING PROCEDURE GIVEN IN THE HANDOUTS.

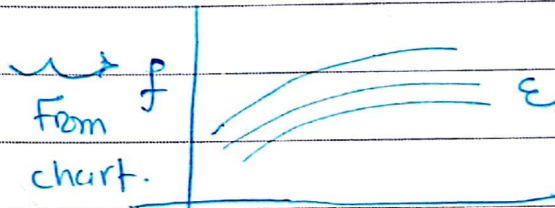
$$E_G \approx 0.5$$

↓ use it to find f .

$$\frac{\sum Q}{\alpha A_{CP} f}$$

↓ then use original chart to find T_G

$$\frac{AR}{\alpha A_{CP}} = 1.79$$



$$f = 0.73$$

↓

$$\frac{\sum Q}{\alpha A_{CP} f} = \frac{26,217,78}{(11,24)(0.73)} = 32,318 \frac{\text{Btu}}{\text{ft}^2} \quad \text{check units.}$$

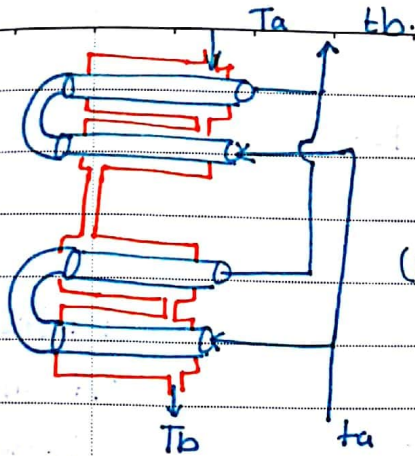
From step 3 use the chart:- $T_G = \underline{\underline{1650^\circ\text{F}}}$

the Assumed value $T_G \approx 1730^\circ\text{F}$

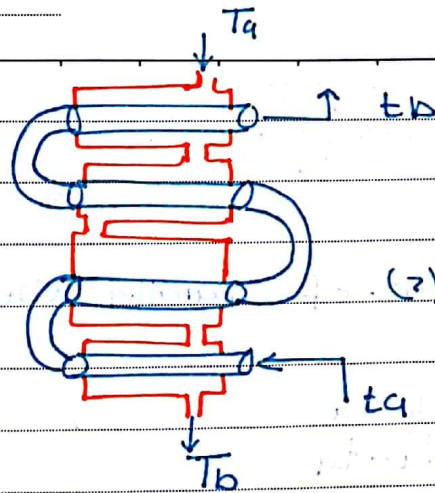
Design of double pipe heat exchanger.

November 12th

No.



(1)



(2)

32"
16"

Two hairpin in.

→ parallel tube

→ Annuli in Series

→ tube side t_a / t_b

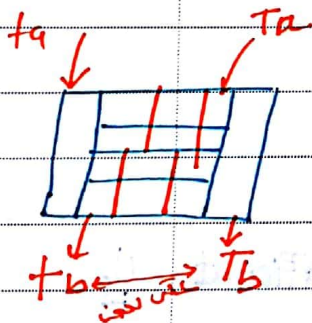
→ shell side T_a / T_b

→ high heat duty

→ high flow rate

$$q = U A \Delta T_m \Rightarrow A = \frac{q}{U \Delta T_m}$$

Re/Pr $\Rightarrow$ Calculated based on tube side / shell side



* double pipe design:

Madar team

mini. ΔP

to decrease pressure drop

Gases تأثير كبير بينهما ان كان هناك

Design equation

$$q = U A \Delta T_m$$

$\Rightarrow$ calculate LMTD

Correction factor Relation:

$$F = \frac{\left[\frac{R-x}{X(R-1)} \right] \ln \frac{(1-P)}{(1-PR)}}{\left[\frac{(R-x)}{R(1-PR)^{1/x}} + \frac{x}{R} \right]}$$

Case (1):

→ Two hair

Pins.

→ Annuli in Series

→ inner pipes in Parallel

$R \neq 1$

⇒ HOW MANY HAIRPINS ARE REQUIRED TO ACHIEVE DUTY

* Correction Factor

* November 12th

No.

$$F = \frac{P(1-X)}{X(1-P) \ln \left[\frac{(1-X)}{(1-P)^{1/X}} + X \right]}$$

R=1

X: no. of parallel tubes.

$$P = \frac{t_b - t_a}{T_a - t_a}$$

$$R = \frac{T_a - T_b}{t_b - t_a}$$

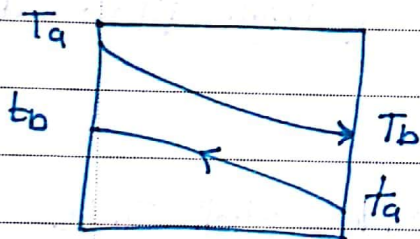
t: tube side.

T: Shell side

b: outlet

a: inlet

⇒ TO CALCULATE THE LMTD



q: known (Duty)

ΔT_m: based on Temp.

U:

• @ counter current.

☒ multiplying correction factor.

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⇒ Calculation of overall heat transfer coefficient U_o

$$h_i = ? \quad h_i = f(Re, Pr)$$

$$Re_f = \frac{4 \dot{m}}{\pi d_i \mu}$$

parallel: Flow rate remains the same

series: Flow rate divide.

⇒ here, you have to assume no. of pipe.

☒ k_f : thermal conductivity for the fluid.

No.

November 12

if $Re < 2100$

$$Nu = \frac{h_i d_i}{k_f} = 1.86 \left[Re Pr \frac{d_i}{L} \right]^{1/3} \left(\frac{\mu}{\mu_w} \right)^{0.14}$$

→ viscosity (center)
→ " (wall)

if Re is transient
(2100 - 4000)

$$Nu = \frac{h_i d_i}{k_f} = 0.116 \left[Re^{2/3} - 125 \right] Pr^{1/3} \times \left(\frac{\mu}{\mu_w} \right)^{0.14} \left[1 + \left(\frac{d_i}{L} \right)^{2/3} \right]$$

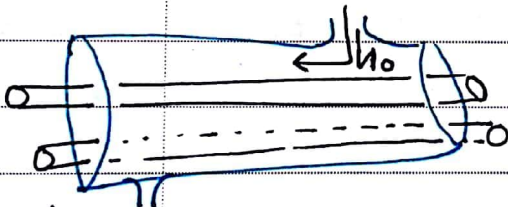
if $Re > 4000$

$$Nu = \frac{h_i d_i}{k_f} Re^{0.8} Pr^{1/3} \left(\frac{\mu}{\mu_w} \right)^{0.14}$$

→ for 1st trial use: $\left(\frac{\mu}{\mu_w} \right)^{0.14} = 1$

☒ SHELL DESIGN:

November 14



☒ All volume except for the pipe.

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D_1 : OD for tubes

D_2 : ID for tubes (Annuli/Shell)

Area of flow $A_f = \frac{\pi}{4} (D_2^2 - N D_1^2)$

☒ Equivalent Diameter:

$$D_e = \frac{D_2^2 - N D_1^2}{D_2 + N D_1}$$

$$Re = \frac{D_e (m / AF)}{\mu}$$

k_f : thermal conductivity

THEN $\Rightarrow$ calculation of Nu number.

(1) Laminar flow $Re < 2100$

local heat

h_o : equivalent transfer diameter

$$Nu = \frac{h_o D_e}{k_f} = 3.66 + 1.2 \left(\frac{D_2}{D_1} \right)^{0.8} + \frac{0.19 [1 + 0.14 \left(\frac{D_2}{D_1} \right)^{0.5} [Re Pr \frac{D_e}{L}]^{0.8}]}{1 + 0.117 [Re Pr \frac{D_e}{L}]^{0.467}}$$

(2) transition / turbulent $\leftarrow$ use previous relation in tube side or the following.

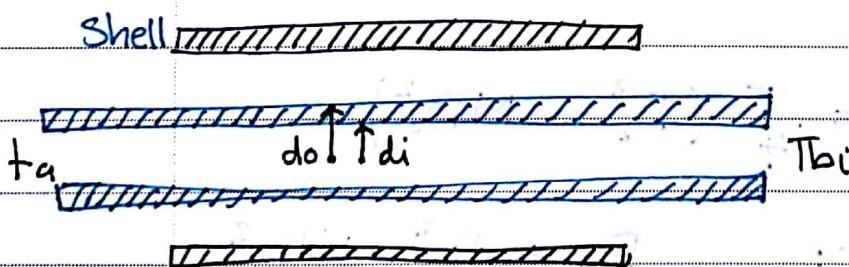
$$Nu = \frac{h_o D_e}{k_f} = \frac{(f/8) [Re - 1000] Pr}{12.7 \sqrt{(f/8)} (Pr^{2/3} - 1)} \left[1 + \left(\frac{D_e}{L} \right)^{1/3} \right]$$

$$f = (0.782 \ln Re - 1.51)^{-2}$$

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$\Rightarrow$ To estimate $M_w = f(T_w)$

$$T_w = \frac{h_i \times T_{avg} + h_o \left(\frac{d_o}{d_i} \right) T_{avg}}{h_i + h_o \left(\frac{d_o}{d_i} \right)}$$



difference (d_o / d_i) is the thickness of the tubes.

$d_o - d_i = \Delta x$ (Thickness of the inner tube)

November 14th

No.

→ correction for h_i

$$h_{i_{\text{new}}} = h_{i_{\text{old}}} \Phi_i = h_{i_{\text{old}}} \left(\frac{M}{M_w} \right)^{0.14}$$

$$h_{o_{\text{new}}} = h_{o_{\text{old}}} \Phi_o = h_{o_{\text{old}}} \left(\frac{M}{M_w} \right)^{0.14}$$

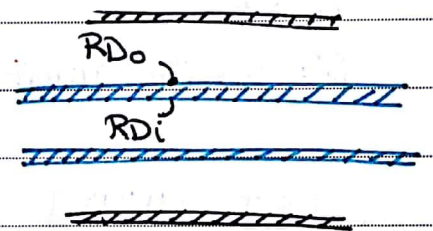
Calculate: over all heat transfer coefficient

$$U_o = \left[\frac{d_o}{d_i h_i} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2 k_w} + \frac{1}{h_o} + \frac{R_{d_o}}{d_i} + R_{D_o} \right]$$

→ R_{D_o}/R_{D_i} : Resistance to scale

: Dirt factor

⇒ obtained from tables.



☒ Design equation

$$q = U_o A (\Delta T_m)$$

$$A = \frac{q}{F * LMTD * U_o}$$

☒ Surface Area ⇒ For one hairpin is know standard A_1

$$N_{\text{hairpin}} = \frac{A}{A_1}$$

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complication ? ⇒ pressure drop !

→ Calculation of ΔP (tube-side).

November 14th

No.

→ [BRITISH UNITS]

$$P_f = \frac{f L G^2}{7.5 \times 10^{12} di S \phi} \text{ viscosity correction factor.}$$

$$f = \frac{64}{Re} : \text{Laminar flow.}$$

$$f = 0.3673 Re^{-0.2314} \text{ turbulent Flow.}$$

units:-

$$G = \rho v = \frac{m}{AF}$$

ΔP : Psi

L : ft

S : specific gravity: unitless

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→ [SI UNITS]

$$\Delta P_f = \frac{f L G^2}{2000 di S \phi} \text{ [Pa]}$$

Entrance: — For Single tube, neglect ΔP .

and exist: — For multiple tubes.

turbulent Flow: $\Delta P = \frac{v^2}{2g}$
[concentration head
velocity]

Laminar Flow: $\Delta P = \frac{\rho v^2}{2g}$

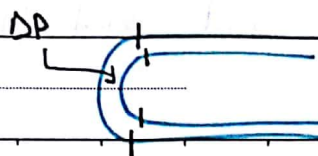
• WHATEVER IT TAKES:

← cont'd ΔP

No.

November 16th

Return bend ΔP , tube-side.



Cases: (1) 180° change in direction.

Rest in $\Delta P = 1.2 \Delta P_v$ (turbulent flow)

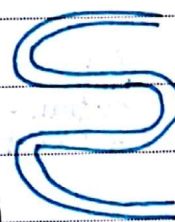
english units $\Delta P_r = f(Re)$ (Laminar flow)

$$\Delta P_r = 1.5 \Delta P_v$$

⇒ for a series of hairpins:

$$\Delta P_r = 1.6 \times 10^{-13} (2 N_{HP} - 1) \frac{G^2}{\text{sp.gr}} \quad (\text{turbulent})$$

(Psi)



$$\Delta P_r = 2 \times 10^{-13} (2 N_{HP} - 1) \frac{G^2}{\text{sp.gr}} \quad (\text{Laminar})$$

⇒ Calculation ΔP in Annuli side (Shell side):-

$$\Delta P_f = \frac{f L G^2}{7.5 \times 10^{12} De (\text{sp.gr}) \phi}$$

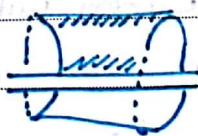
Fluid Flow

f : fanning friction factor

ϕ : viscosity correction factor

Laminar:
Flow

$$f = \frac{64}{Re}$$



: high pressure drop,
on both surfaces.

$$f = \frac{64}{Re} \left[\frac{(1-k^2)}{1+k^2 + \frac{(1-k^2)}{\ln k}} \right];$$

$\ln k = 1.5$

$k = \frac{D_1}{D_2}$: outer diameter
inner diameter.

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Turbulent : $f = 0.3673 Re^{-0.2314}$
Flow

ΔP for entrance and exist fluid.

November 16th.

No:

Entrance : $\Delta P = \Delta P_v = 1.334 \times 10^{-13} \frac{G^2}{\text{Sp.gr}}$
(turbulent)

Exist : $\Delta P_e = 0.5 \Delta P_v = 0.5 \left(1.334 \times 10^{-13} \frac{G^2}{\text{Sp.gr}} \right)$

(Laminar flow
[Entrance + Exist])
 $\Delta P_{\text{entrance + exist}} = 3 \Delta P_v = 3 \left(1.334 \times 10^{-13} \frac{G^2}{\text{Sp.gr}} \right)$

For Return bend (turbulent) $\Delta P_n = 2 \times 10^{-13} N_{HP} \frac{G^2}{\text{Sp.gr}} \rightarrow$ For all hair pens.

Laminar: $\Delta P = 4 \times 10^{-13} N_{HP} \frac{G^2}{\text{Sp.gr}}$
Re ≈ 100

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Example: 10,000 lb/hr of bazeen will be heated from 60°F to 120°F by heat exchanger with (aniline) stream that will be cooled from 150°F to 100 F.

A no. of 16 ft hairpins.

ID tube = 0.115 ft

OD tube = 0.138 ft

ID annuli = 0.172 ft

Find the required no of hairpins and ΔP s.

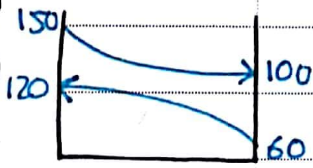
Assume: counter current flow, parallel in tubes with number of parallel branches $X=2$

No.

Fluid property	Benzene @ 90°F	Aniline @ $t_{avg} = 125^\circ\text{F}$
<u>SI</u> μ (cp)	0.55	2.0
<u>english</u> C_p ($\frac{\text{Btu}}{\text{lb}\cdot\text{F}}$)	0.42	0.52
k_f ($\frac{\text{Btu}}{\text{lb}\cdot\text{ft}\cdot\text{F}}$)	9.4	

Solution:

$$q = m C_p \Delta T = (10,000)(0.42)(120 - 60) = 252,000 \frac{\text{Btu}}{\text{hr}}$$



$$m_{\text{aniline}} = \frac{252,000}{(0.52)(150 - 100)} = 9692 \text{ lb/hr.}$$

$$LMTD = \frac{(150 - 120) - (100 - 60)}{\ln\left(\frac{150 - 120}{100 - 60}\right)} = 37.67^\circ\text{F.}$$

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$$\text{correction factor } F = \frac{R - X}{X(R - 1)} \frac{\ln\left(\frac{1 - P}{1 - PR}\right)}{\ln\left[\frac{(R - X)}{R(1 - PR)} + \frac{X}{R}\right]} = 0.86$$

$$\text{tube side: } P = \frac{t_b - t_a}{T_a - t_a} = \frac{100 - 150}{60 - 150} = 0.55$$

$$R = \frac{T_a - T_b}{t_b - t_a} = \frac{60 - 120}{100 - 150} = 1.2$$

NOTE:

tube: Aniline.

Shell: benzene

local heat

$$\text{transfer (shell side) tube } Re = \frac{4m}{\pi d_i \mu} = \frac{4(9692)}{\pi (0.115)(2 \times 2.419)} = 22191 \text{ (turbulent).}$$

correction factor for viscosity.

$$\text{(turbulent): } h_i = \frac{k_f}{d_i} \times 0.023 Re^{0.8} Pr^{1/3} \left(\frac{\mu C_p}{k_f} \right)$$

$$h_i = \frac{0.1}{0.115} \times 0.023 (22191)^{0.8} \left[\frac{(2 \times 2.419)(0.52)}{0.1} \right]^{1/3} = 175.7 \frac{\text{Btu}}{\text{hr ft}^2 \text{F}}$$

To obtain h_o :

first trial:-

$$D_1 = OD = 0.138 \text{ ft}$$

$$D_2 = ID = 0.172 \text{ ft}$$

$$n \text{ hairpin} = 1$$

$$A_f = \frac{\pi}{4} (D_2^2 - n D_1^2)$$

$$Deq = \frac{D_2^2 - n D_1^2}{D_2 + n D_1} = \frac{D_2^2 - D_1^2}{D_2 + D_1}$$

$$Deq = \frac{(D_2 - D_1)(D_2 + D_1)}{D_2 + D_1} = 0.172 - 0.138 = 0.034 \text{ ft}$$

$$= \frac{\pi}{4} (0.172^2 - (1)(0.138)^2)$$

$$= 8.2 \times 10^{-3} \text{ ft}^2$$

$$Re: \quad Res = \frac{De \left(\frac{m}{Af} \right)}{\mu} = \frac{(0.034) \left(\frac{10,000}{8.2 \times 10^{-3}} \right)}{0.55 (2.419)} = 31,165 \quad (\text{turbulent Flow})$$

this correlation is used for both

- h_o
- h_i

$$h_o = \frac{k_f (0.023) Re^{0.8} Pr^{1/3}}{De} = \frac{0.092 (0.023) (31,165)^{0.8} \left[\frac{(0.42)}{(0.55)(2.419)} \right]^{1/3}}{0.034}$$

$$h_o = 445.7 \text{ Btu/hr ft}^2 \text{ } ^\circ\text{F}$$

@ turbulent Flow

→ usually considered the best correlation

Dirt factor: $R_{Di} = R_{Do} = 0.001 \text{ h} \cdot \text{ft}^2 \cdot ^\circ\text{F} / \text{Btu}$ (from tables) *

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→ h_o is multiplied with viscosity.

→ viscosity is effected by Temp.

Final T_{wall} $T_w = \frac{h_i T_{avg} + h_o \frac{d_o}{d_i} T_{avg}}{h_i + h_o \frac{d_o}{d_i}}$

then

h_o/h_i
(again)

$$T_w = \frac{175.7 \left(\frac{100 + 150}{2} \right) + 445.7 \left(\frac{0.138}{0.115} \right) \left(\frac{60 + 120}{2} \right)}{175.7 + 445.7 \left(\frac{0.138}{0.115} \right)}$$

$$T_w = 98.6 \rightarrow \mu (\text{benzene}) = 0.51 \text{ cp} \quad [\text{from tables}]$$

$$\rightarrow \mu (\text{Anni}) = 2.9 \text{ cp}$$

November 19th.

No.

correction factors $\phi_i = \left(\frac{\mu_i}{\mu_w} \right)^{0.14} = \left(\frac{2}{2.9} \right)^{0.14} = 0.95$

$\phi_o = \left(\frac{0.55}{0.51} \right)^{0.14} = 1.01$

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new $\Rightarrow \mu_i = (0.95)(175.7) = 166.9$

values $\Rightarrow h_o = (1.01)(445.7) = 450.4$

$U_D = \left[\frac{d_o}{d_i h_i} + \frac{d_o \ln\left(\frac{d_o}{d_i}\right)}{2 k_w} + \frac{1}{h_o} + \frac{R_{Di} d_o}{d_i} + R_{Do} \right]^{-1}$ tube (exchange) side (on it)

$U_D = \left[\frac{0.138}{0.115 (166.9)} + \frac{0.138 \ln\left(\frac{0.138}{0.115}\right)}{2 (9.4)} + \frac{1}{450.4} + (0.001) \left(\frac{0.138}{0.115} \right) + 0.001 \right]^{-1}$

Finding the area $A = \frac{q}{U * (LMTD) * F} = \frac{252,000}{(77.2)(37.76)(0.86)} \approx 109 \text{ ft}^2$

Area for one hair-pin $\Rightarrow A_1 = 0.435 \text{ ft}^2$

$A_1 = 2\pi r L = \pi D L$



$\Rightarrow$ length 32 ft
for one tube is standard 32 ft

$N = \frac{A}{A_1} = \frac{109}{0.435} = 251$ no. of hair pins.

Finding the length $A = \pi D L_T$ total length $\Rightarrow L$ is used for Δp .

total area

$109 = \pi$

Same

procedure

Same

trial