Subject Process design Date 31st Jan 2018 No.

Inlet → [process] → outlet

Low cost high cost

A + B → C + D

Desired product

F + G → C ⇐ no need equipment to seperate

Process objective → profetable

→ non profetable

1) Example:- non profetable process

Degredation of organic polutants

organic in water → Filtration ✗

→ ppt ✗

→ coagulation ✗

→ Distillation ✗

→ Evaporation of water ✗

→ Solvent extraction ✗

→ absorption ✓

→ Chemical treatment ✓

↳ oxidation

water + PPm organics $\xrightarrow[\text{photo catalysis (Ultraviolet)}]{TiO_2} CO_2 + H_2O$

UV light

H_2O_2

Feed → [Reactor] → product + slurry TiO_2 → [seperation of gases from by liquid flash drum] → sep. of solid → [sep. of solid] → water

Subject Date No.

2) Profetable process

$A + B \rightarrow C + D$ ← It operate

Desiral product

hetrogeneous with catalyst

or

hemogeneous without

rate

our design specified by:

1- حاجة السوق

2- availability of catalyst

3- availability of raw materials.

To have a profetable process:

1- need of product in market

2- price of the product

3- availability of raw material in market

4- profit

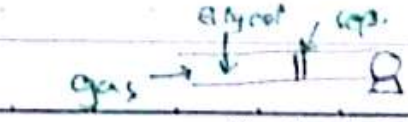
Example: produce of monoethylene Glycol [ME G]

Uses:

- Intermediate in manufacturing polyster
- fabrics
- resins
- A scoolant and heat transfer reagent.
- Anti Freezer
- hydrate inhibitor in gas Industry.

consumptions rate is : 25 million /year

Price $1200 / ton



Production routes:

1) Naphtha → ethylene → ethylene oxide → MEG

2) Syngas (SN_2+H) → dimethyl oxalate → MEG

3) syngas → methanol → olefins (Alkenes C_xH_{2x}) → ethylene → ethylene oxide → MEG

← كيف اختار أي طريق احسن؟

أول اشي لازم نفرق عالبرد سيس وال steps
آخر واحد فيه اكتر Product وكلم بتناسب السوق

* A Primary Analysis any of these processes could be probable (time, market need, complexity, operating condition, etc)
⇒ we should search for route to get the product

* minimize (raw material, equipments, Utilites, Labor cost, etc) also minimize delivery cost and transportations.

Subject Process

Monday
Date 5th Feb, 2013 No.

* One of the process routes are defined it is required to focus on the following :-

1) capacity of the plant
2) one or multi products
3) conversion, selectivity, yield (process efficiency)

Next step:-

1) Define the type of reactor
2) Define the type of separators.
3) Do we have non condensable gases or not.

Next step:-

1) Define the type of field product, etc
2) Do I need Recycle or not.

Next step:- => building the plant <=

Block Flow Diagram (BFD)

written information ⟹ access

: Conversion of world problem into a simple visual BFD

BFD → Block flow process diagram
BFD → Block flow plant diagram

Example:

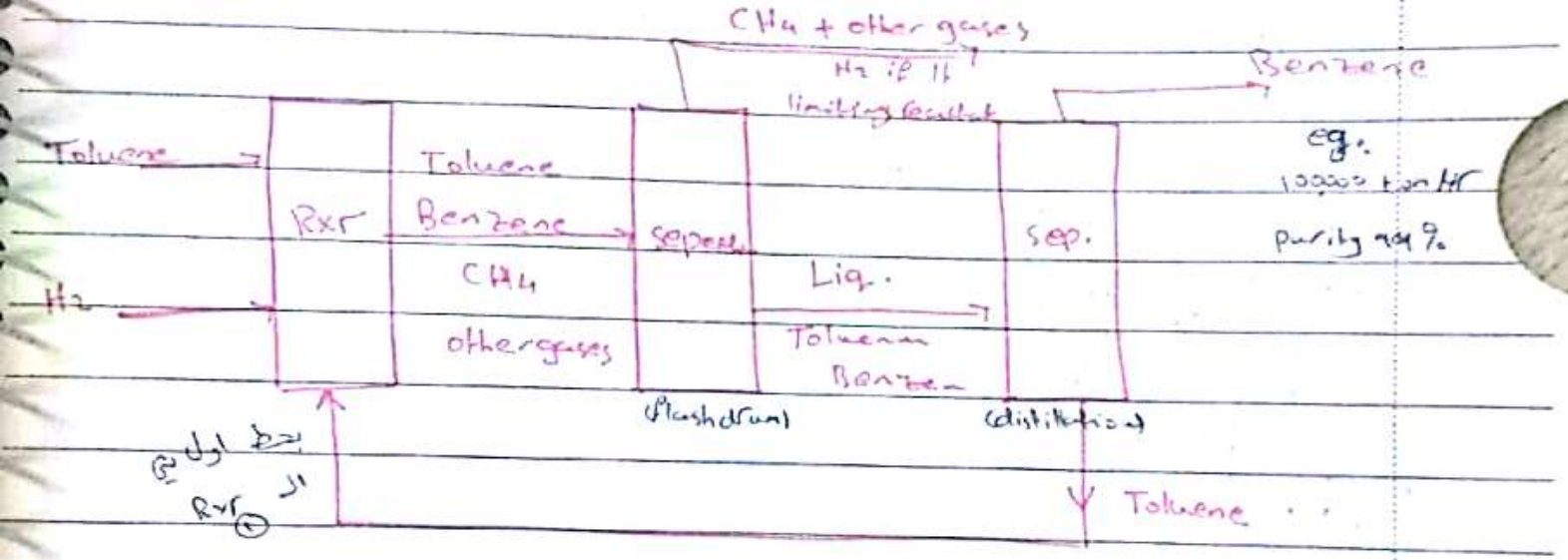
$C_7H_8 + H_2 \rightarrow C_6H_6 + CH_4$

Industrial statement: Toluene and hydrogene are converted in a reactor to produce $C_6H_6 + CH_4$. The Rxn doesnot go to complete and excess toluene should be introduced.

Subject

Date No.

The non condensable gas are tobe seperated and discharge the benzene product and the unreacted toluene are then seperated by any seperation method. The non reacted toluene is recycled back to the R.r then the benzene is tobe removed.



* main characteristics of BFD:.

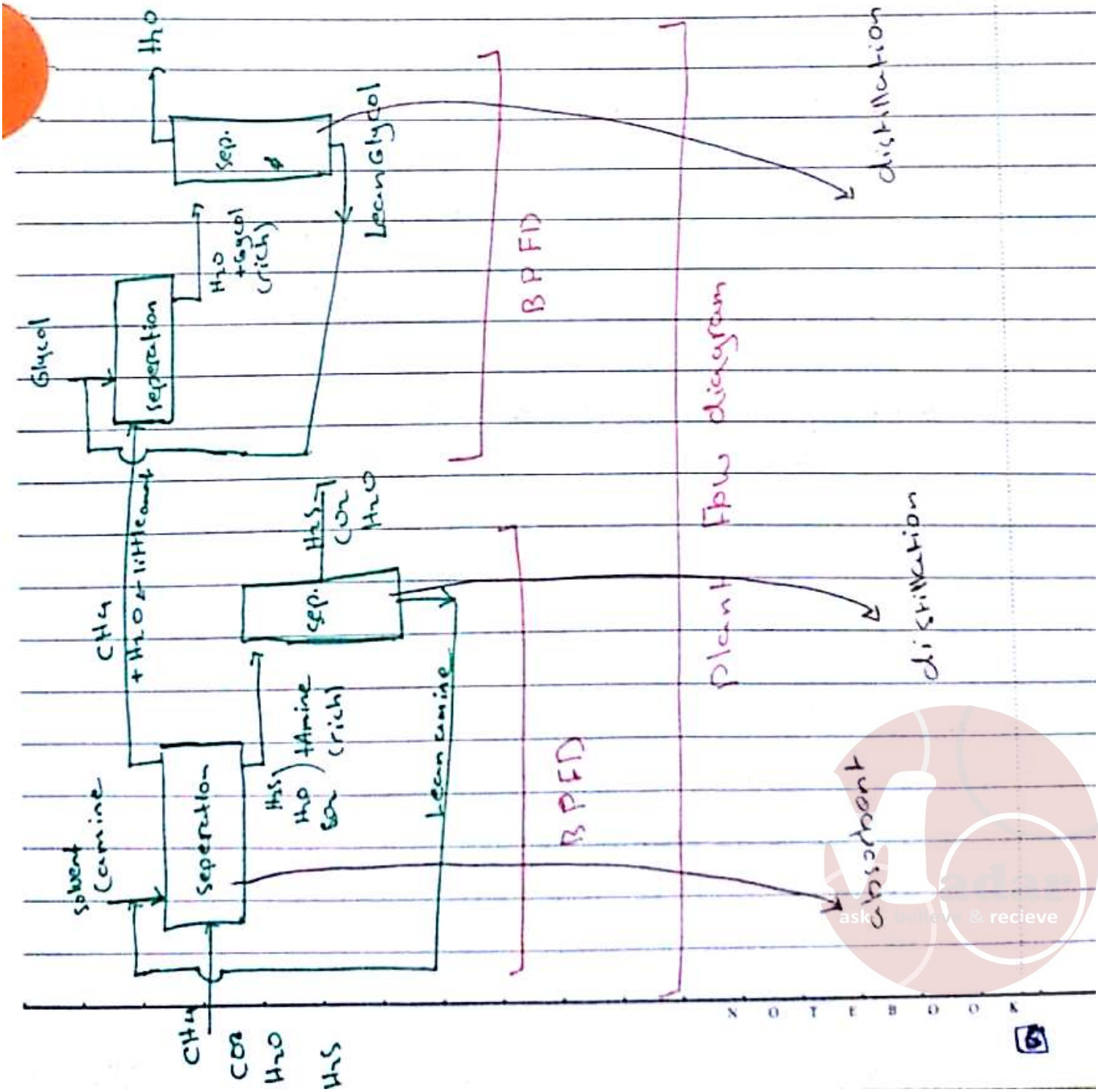
1) operation shown by blocks not by shape of reactors.
2) flow goes from left to right.
3) light stream goes up and the heavy ones goes down.
4) some critical information may be given (production rate, conversion, purity, etc..)
5) if lines cross then horizental line shouldbe solid and the vertical is broken.

Example :-

Amine treatment unit

Natural gas | H_2S, CO_2 → CO_2, H_2S, H_2O → pure natural gas CH_4

gas → we will go for absorbtion



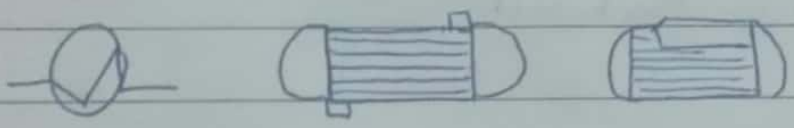
Monday

Subject Date 12th Feb, 2018 No.

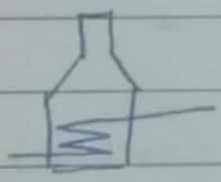
Process flow Diagram :- the more detailed than Block

PFD

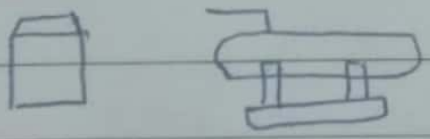
- Flow summery table
  - stream ✱
  - detailed stream information
    P, Composition, T
- Equipment summery table
  - major pieces of equipments with number discription, etc
    ↑ Just the basic
- All Utility streams
  - cooling water
  - steam
- basic control loops
  - Flow
  - Temp
  - pressure



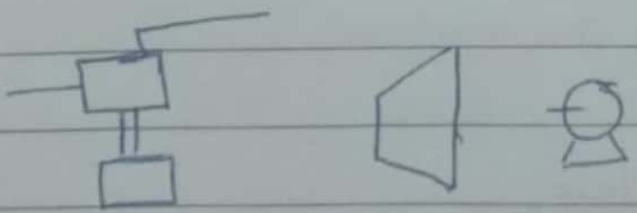
Heat exchangers



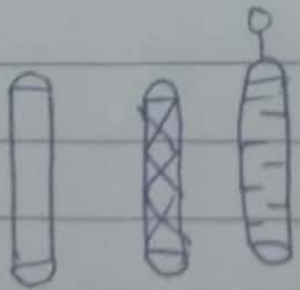
Fired Heater



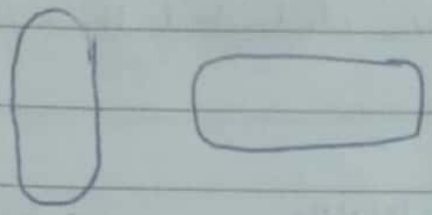
Storage Tanks



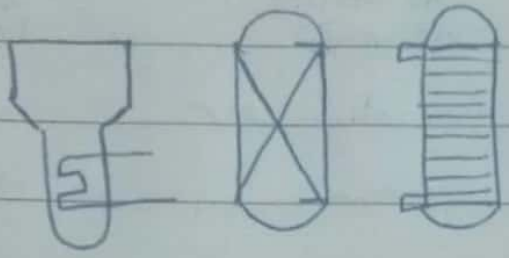
Pumps, Turbines
Compressors



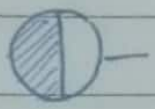
Towers



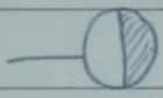
Vessels



Reactors



Process Input



process output

Control Value

Globe Value (Manual Control)

Value

Stream Number

Instrument Flag

Subject Date No.

Process flow diagrams:

- All major equipments
- Each equipment is assigned a number
- All process stream should be presented
- All streams should have a unique No.
- Each stream contains detailed information (Temp., pressure, composition, total flowrate, etc.)
- Utility streams should be provided (CW, LPS, hps, mps, etc.)
  - CW → Cooled water
  - LPS → low pressure stream
  - hps → high pressure stream
  - mps → medium pressure stream
- basic control loop

PFD
- Process topology (الاشكال والاجهزة)
- Stream information
- Equipment information

Process topology ⇒ From software visio.
gas → compressor
Liquid → pump

* if reaction is works at high temp
Later on I have to raise it
fired heater
exchanger

* main process عاوزين الناتج

The Liq. ↓ من فوق

gases ↑ من تحت

* maybe Ineed Jacket to change temp.
* if steady state → Idon't need Fired heater
* Seperation unit work at low temp.
to enhance the sep.

→ seperation at low temp. as absorbtion
* reactor → exchanger → Flash drum
→ non condensable gases Ihave to remove it by flash drum
to enhance the distillation عشان نبدي اباها

Distillation - costy - less safe
Absorbtion - less cost - more safe
How to seperate Unreacted product?
→ if its Light & heavy by distillation
Ican take this part and recycle it back
→ before distillation column, We have to exchanger.

ubject Date No.

Why exchanger or throatling valve?

To : 1- reduce th pressure

2- cooling

3- quality for this stream

partially vapor or partially liquid

toget quality between 0 → 1

* There is electricity Unit → اذا قطعت الكهرباء

V - 101 Vessel # 1

P - 101 pump # 1

pumps and compressors A/B يعني one running one standby

packed bed يعني ←

- stream numbering. From left to right
From top to bottom

- Naming of Equipment

Format XX - YZZ A/B

equipmt name ↙ Area no. ↘ equipment no. → running / standby equipment

bject Date No.

Example:.

1- Pumps area 100, P-101 A/B
2- Compressor C-101 A/B
3- Exchanger E-101
4- Fired heater H-101
5- Reactor R-101
6- Tower T-101
(Absorption, distillation, etc..)
7- Vessels V-101
8- storage TK-101

LPS : Low pressure steam 3-5 bar
mps : medium pressure steam 10-15 bar
hps : high pressure steam 40-50 bar
cw : Cooling water 30-45°C
↙
in hot area
15-35°C
in medium area

- Stream information:.

Stream # (1)
Temp. °C SI units
pressure bar
flowrate
composition & vapor quality } this table

Subject Date No.

- Equipments information ⇐ Table for each equip.

A. Exchanger

1- Naming

2- Type: ① mechanical → shell and tube (Eg.)
→ plate
→ U tube

② process → single, two phase

3- Design ⇒ (Area, No. of tubes, Baffels space, No. of passes, pressure, LMTD, Aspect ratio, material of construction)

B- Tower

1- Design; L, D, N-stages, packing, material of construction

C- Vessels

Design; L, D, horizontal or vertical, P, T; Material of construction

D- pumps : discharge p

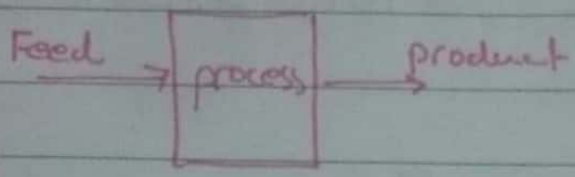
Flow, P_{out}, ΔP, Power

material of construction

Subject

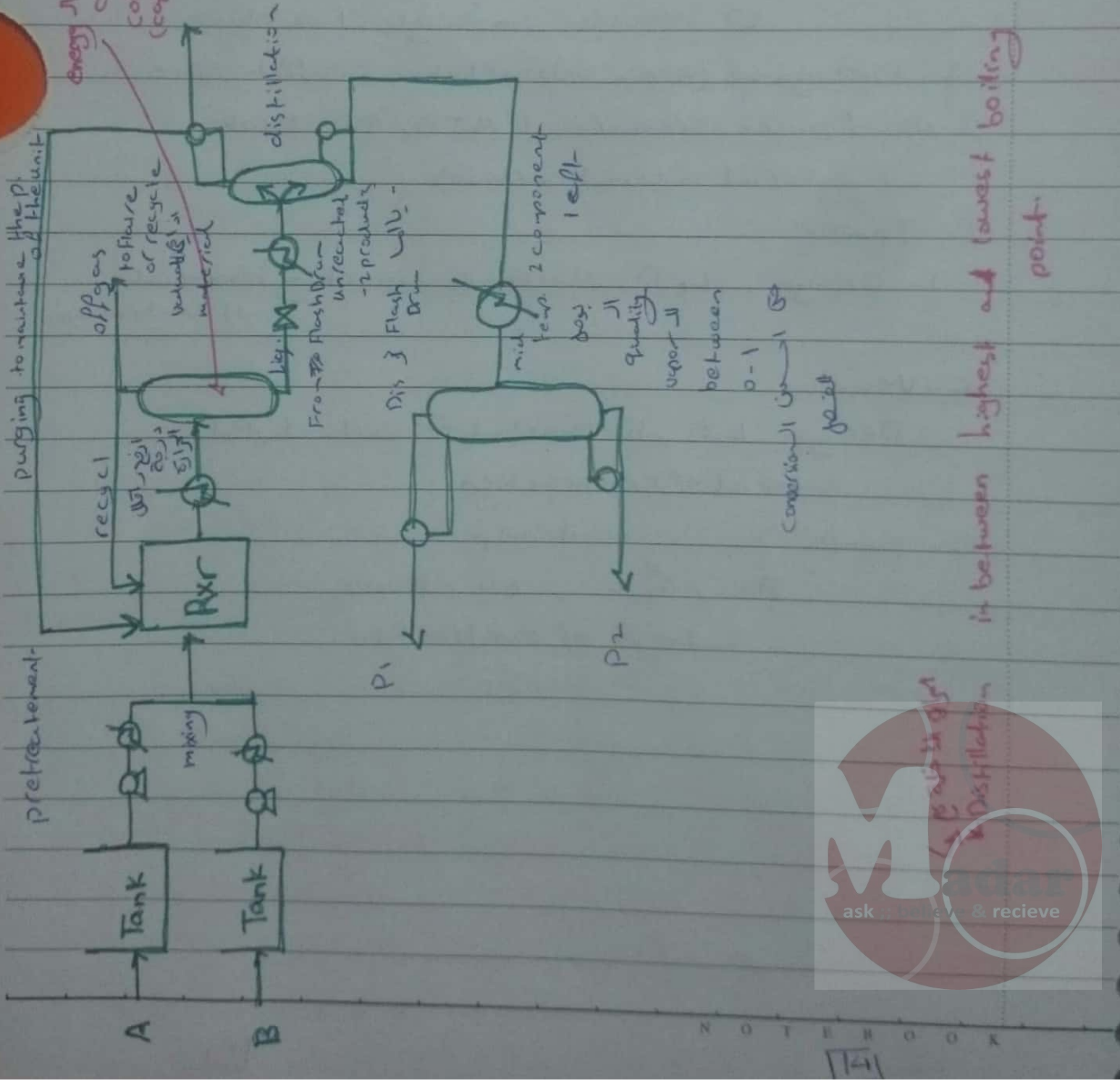
Date 14th Feb, 2013 Wednesday No.

PFD



I have → Chemical Rxn $A + B \rightarrow P + Q + R$

Liq. gas Liq.



ubject Date No.

Structure and Synthesis of PFD

1- Decide to have batch or continuous process
   - For Ultrapure
   - For sensetive rxn

2- Identifying input/output structure of the process

3- define the recycle / by-pass / purge streams
   maximizing/minimizing the

4- Identify the general yeild / conversion structure of the seperation process or units

5- Identify whether the process needs heat integration or optimization or not

Batch or continuous process

1- Batch process < 500 ton/year
   continuous 75000 ton/year
   in between you have to decide
   (size) → batch for smaller size
   → continuous is good for large

2- operational flexibility → batch →
   products

Batch: can be used for multiple operation
continuous: for only single or few products
defined from the begining

3- process efficiency → Batch not
Batch: requires strict scheduling and
- requires different cleaning steps
- energy integration is impossible

Subject Date No.

- Utility is higher cycle ک
Utility ک cost
Continuous: does not require all the last items.

4- maintenance and labor
Batch:- more labor and maintenance
process running by hand
Continuous:- less labor and maintenance
* The most important factor → production rate.

5- Feed stock availability
material
Batch:- is favor on limited feed stocks

6- product Demand
Batch:- is ~~favor on limited~~ is used for seasonal demand

7- In term of rate of rxn
Batch:- is used for slow of rxn

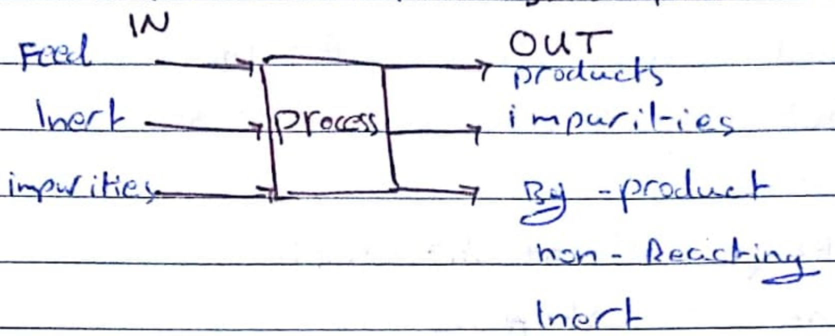
8- Fouling and scaling
Batch:- is used for more fouling process

9- safety Continuous is more safe

10- Controllability :-
Continuous is well-controlled.

Subject   Monday
Date 19th Feb, 2018 No.

INPUT-OUTPUT Structure of the process



[1] INPUTS

Chemicals in → consumed / → non-consumed

chemicals that are not consumed → Requires more operating cost (inert material, make-up catalyst, makeup solvent)
→ this should be introduced

[2] OUTPUTS

Based on the merit

chemicals out → products / → non-Reacting material → merit فقط

[3] UTILITIES & TRACE Components

[4] FEED PURITY ← اذا ما بتأثر ع الReaction so Donot care

a- if these impurities affect the process (Reaction, etc.) I have to purify the feed

b- if these impurities have no effect
If it is Less than 15-20% → introduce it without treatment.

NOTEBOOK

Subject Date No.

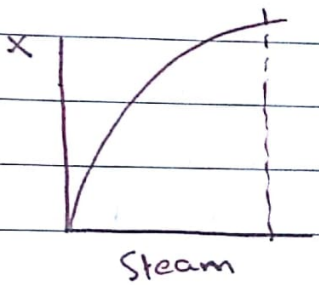
Example

Steam is added to shift the equilibrium of catalytic dehydrogenation of ethyl-benzene to produce styrene.

$C_6H_5CH_2CH_3 \rightleftharpoons C_6H_5CH=CH_2 + H_2$

ethyl-benzene styrene

Rxn takes place $600°T < T < 700°T$, low P



high conversion we need to optimize amount of steam to shift the rxn

IN CONCLUSION:-

1- Economical Analysis (profit)

2- Safety which ← process اختار الـ easy to control.

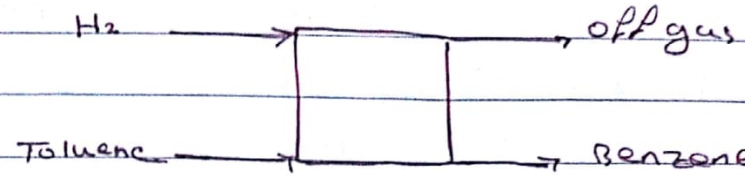
3- Conversion

4- Reaction rate

if the seperation of the impurities is diff. (For azotrope) → don't separate, prior to Feeding.

Subject Date No.

Example: hydrodekalkylation of toluene to produce benzene



methane may be introduced with hydrogen

* either introduce methane / H_2 } more profitable
  more equip. for sep.
  pure H_2 → more pure product
* if impurities fouls or poison the catalyst → separate
* if the impurities react to form hazardous product or difficult to seperate product → purify } From safety point of view

Example: Production of isocyanate

① $CO + Cl_2 \rightarrow COCl_2$ Phosgene
$COCl_2 + Amine \rightarrow isocyanite$

② $CH_4 + H_2O \rightarrow CO + 3H_2$

5. addition of inert materials to control exothermic Rxn or to shift the Rxn ~~or to shift~~ toward products

NOTE BOOK

Toward

ICI

Subject

Date No.

Example: production of Acrylic acid

$C_3H_6 + \frac{3}{2} O_2 \rightarrow C_3H_4O_2 + H_2O$

acrylic

$C_3H_6 + \frac{5}{2} O_2 \rightarrow C_2H_4O_2 + CO_2 + H_2O$

acetic acid

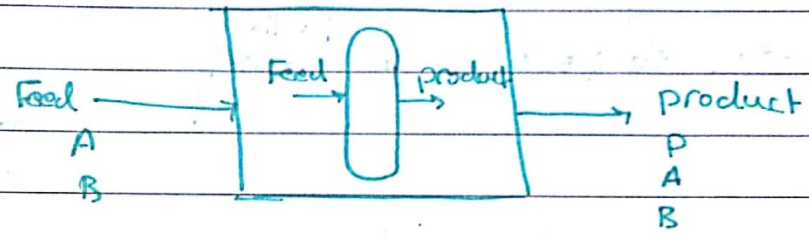
$C_3H_6 + \frac{9}{2} O_2 \rightarrow 3CO_2 + 3H_2O$

Subject Process Date 21th Feb 2018 No.

The recycle structure of the process

Goal : Recovery of material and energy from the process → materials account for 70% of running cost.

Conversion is the criteria



← اذا اكلك في فرق بين conversion الـ in & out

so I must recycle

Conversion

X_B = Reactants consumed in the Rxn / Reactants that introduced into the Rxn

$= \frac{F_{A_0} - F_A}{F_{A_0}} = \frac{C_{A_0} - C_A}{C_{A_0}} = \frac{n_{A_0} - n_A}{n_{A_0}}$

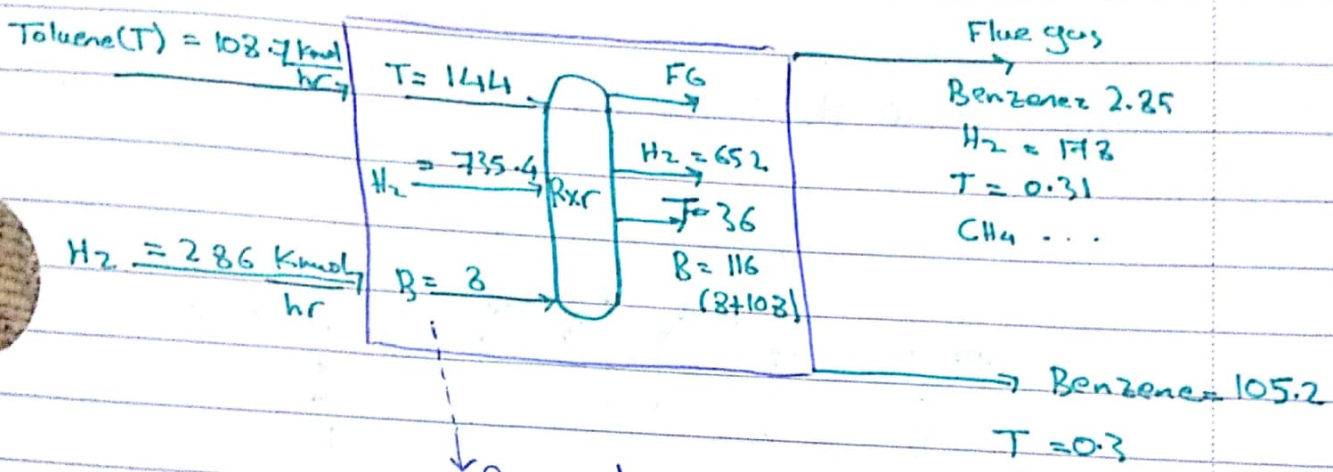
X_p = Reactants that consumed in the process / Reactants that introduced into the process

yeild = no. of moles of the desired products formed / moles of Reactants that Reacted

Subject Date No.

Example :- Hydrodelkylation of Toluene

$$C_6H_5CH_3 + H_2 \longrightarrow C_6H_6 + CH_4$$



Toluen (T) : limiting Reactant

H_2 : excess

single pass conversion

$$X = \frac{144 - 36}{144} = 0.75$$

overall conversion (process)

$$X = \frac{108.7 - 0.61}{108.7} = 0.993$$

$$\text{Yield} = \frac{n.B \text{ produced}}{n_{Tol.} \text{ Reacted}} = \frac{108.5}{108.7 - (0.3 + 0.31)}$$

$$= 0.9995$$

Subject   Date   No.

Conclusion For T

* I have to recycle Because there is differene Between Single & overall conversion.

→ in any process conversion ⫽ 0.995

1- The lower the single pass conversion the greater the recycle must be if $\downarrow x \uparrow$ → Recycle

so there ← size ↑ يزيد الحجم اذا recycle ↓ is optimum Value

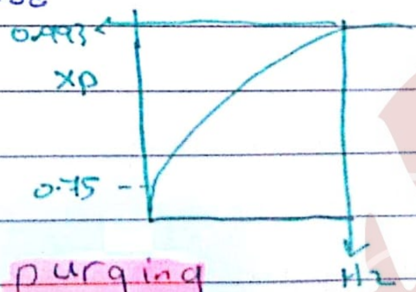
2- The lower the single pass conversion this leads to a higher equipment sizing. (Tubing, pumps, compressors, Towers, reactors)

This should be optimized based on overall process.

Calculations based on hydrogens

single pass $x = \frac{735.4 - 652.6}{735.4} = 0.113$

overall $x = \frac{286 - 178}{286} = 0.378$



مابدي تتراكم ال H_2

so recycle with purging

Subject Date No.

Ineed:

Both single pass and overall conversion of hydrogene are close

→ For Low single conversion:-

Use excess H_2 to enhance the rxn

* But overall low conversion → poor raw material usage (more operating cost↑)

How can we improve this difference?

three ways of improvement

1) seperate, purify, then Recycle.

2) Recycle feed and product together and purge

3) Recycle feed and product w/o without purging

Seperate, purify and Recycle

Selection of the separation unit → How feasible (economical point of view)

Feasibility ← close to ambient condition T, P

large difference in B.P and chemical properities

Subject Date No.

How to recycle H_2

1. Distillation BP H_2 = -252°C

BP CH_4 = -161°C

no way to dis. → large Utility cost

2. Absorption → not feasible @ ambient condition

to be feasible → p↑ & T↓

→ large Utility cost

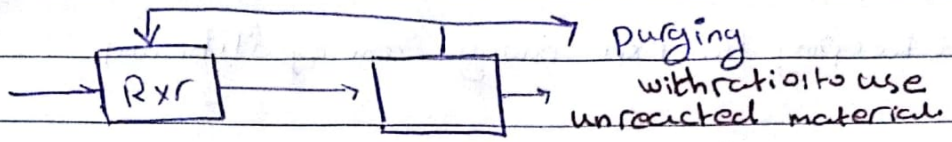
Subject process

Monday
Date 26th Feb 2018 No.

Different ways of Recycling / Separation

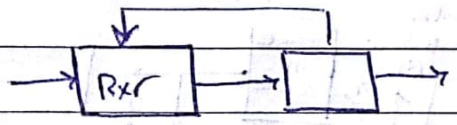
1- Separate / purify → Recycle

2- Recycle the feed / product w/ purging



use this way if separation of products / Reactants is difficult (not feasible) → Recycle + Purge

3. Recycle Feed/product w/o purging



$A + B \rightarrow C$

$C + A \rightarrow D + E$

if there is / are intermediate product(s) that affect Rxn w/o recycling we may not utilies the intermediate to produce the desired product.

Example :-

Rxn of toluene → Benzene

Toluene + H_2 → Benzene

→ $2C_6H_5CH_3 + H_2 \rightarrow C_6H_6$

→ $6C_6H_6 \rightleftharpoons C_{12}H_{10} + H_2$

diphenyle

diphenyl absorbing the Benzene it enhance the Rxn → go for recycle w/o purging

ct

Date No.

Relation ship between Recycle streams and # of Reactors

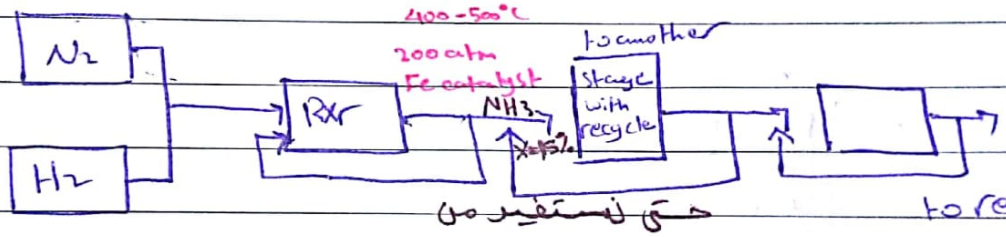
goal ⇒ to enhance the conversion and control the reactor

(keeping the rxn away from equilibrium)

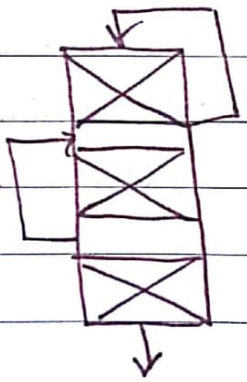
①

Example: production of NH_3

$$N_2 + 3H_2 \longrightarrow 2NH_3 \qquad \Delta H = -92 \frac{kJ}{mol}$$



catalyst ال



interstage cooling

Reactor ... recycle N_2, H_2 to enhance the Rxn.

② controlling Rxr temp. ⇒ highly exothermic

catalyst ال ... damaged with high temperature

highly exothermic Rxn → interstage cooling

use multiple stage rxn.

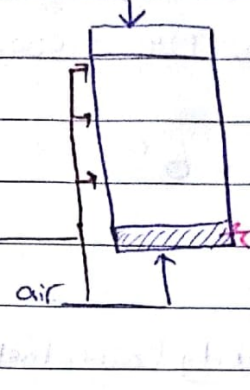
Subject Date No.

③ Concentration Control

gasification rxn

coal, Flyash ← Aowrate الازم ال minimize

$2C+O_2 \rightarrow 2CO$



$C+O_2 \rightarrow CO_2$ Undesired Product

اذا كمية قليلة من O_2 21%

$C/O_2 \uparrow \rightarrow CO$

يجزء مثلا ما يصير احتراق كامل → this enhancing the conversion

④ optimization of conditions of multiple Rxn

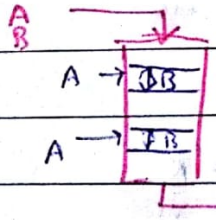
example

$A+B \rightarrow C$

Ihave loadd ← $A \rightleftharpoons B \xrightarrow{\text{equilibrium}} C \rightarrow D$ A or B in excess

A in excess

C: Intermediate



interstage

B: also in excess

Subject Date No.

Example: To input & output structure

It is desired to produce dimethyl ethere (DME)

feasible process { price = $0.95/kg. From methanol (MeOH) = 7 pure
price $0.22/kg production rate of DME = 50000 ton/yr

$$2\,CH_3OH \rightleftharpoons (CH_3)_2O + H_2O$$

DME

As an Alternative

MeOH can be obtained with less purity (83% mol methanol + 11% ethanol + ethere) mixed with price $0.25/kg

Maximum profets with less cost

E-OH ⇌ diethyl ether (DEE)

E-OH → ethylene

Which option should you select?

option [1]

Methanol to dimethyle ether:

desired DME = 50000 ton/yr $= \frac{50000}{46} = 1.08 * 10^6 \frac{kmol}{yr}$

Required methanol as purified

$= (2)(1.08 * 10^6) = 2.16 * 10^6 \frac{kmol}{yr}$

$= 2.176 * 10^6 (32) = 6.9 * 10^7 \frac{kg}{yr}$

Subject Date No.

Cost of MeOH = $(6.9 * 10^7 \frac{kg}{yr})(\$0.22/kg)$

$= \$15$ million (M)

price of DME

$= (50000 \frac{kg}{yr})(\frac{0.95\$}{kg})$

$= \$47.5M$

profit $= \$47.5M - \$15M = \$32.5M$ pure process

Subject Process

Monday
Date 5th March No. 2013

Continue the example: production of DME

selling price ← اضافة على المثال السابق

For DEE = 1.27 $/kg

ethylene = 0.57 $/kg

$2CH_3OH \rightleftharpoons (CH_3)_2O + H_2O$

MeOH DME

$2CH_3CH_2OH \rightleftharpoons (C_2H_5)_2O + H_2O$

EOH DEE

$CH_3CH_2OH \rightarrow C_2H_4 + H_2O$

EOH ethylene

(1) case1, pure MeOH → DME, profits $32.5M

(2) case2, Target to have 50,000 ton/yr DME $(1.08 \times 10^6 \frac{Kmol}{yr})$

→ MeOH → $2.17 * 10^6 \frac{kmol}{yr}$

→ EOH → $= \frac{11\%}{88\%} (2.17 * 10^6) = 12.5 * 10^6 \frac{kg}{yr}$

$= 0.27 * 10^6 \frac{Kmol}{yr}$

Cost of feed $= [(2.17 * 10^6)(32) + (12.5 * 10^6 kg/yr)] (\$0.25/kg)$

$= \$19.4M/yr$

price of DME = $ 47.5M

# of mols of DEE $= \frac{(0.27 * 10^6)}{2} = 0.13 * 10^6 \frac{Kmol}{yr}$

Subject Date No.

mass of DEE = (0.13 × 10⁶)(74) = 9.6 × 10⁶ kg/yr

price of DEE = (9.4 × 10⁶)(1.27) = $ 12.8M/yr

Total selling price = $ DME + $ DEE

= $ 60.3M/yr

profit = $ 60.3M − $ 19.4M

= $ 40.8M/yr

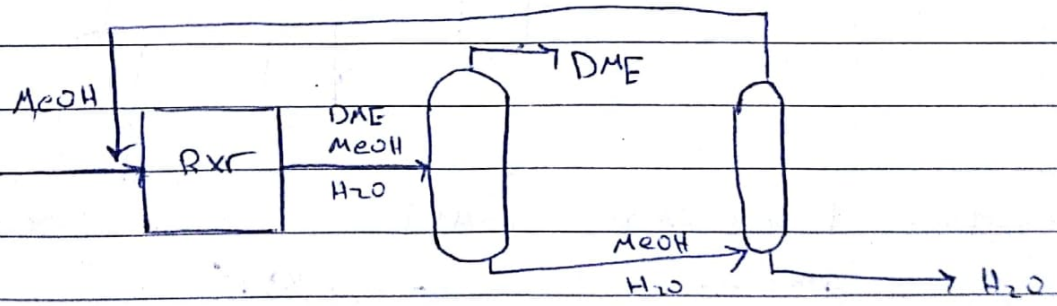
(3) case 3: if all ethanol (EtOH) → ethylene = 7 (100% conversion to ethylene)

Total selling price = $ 51.8 M/yr

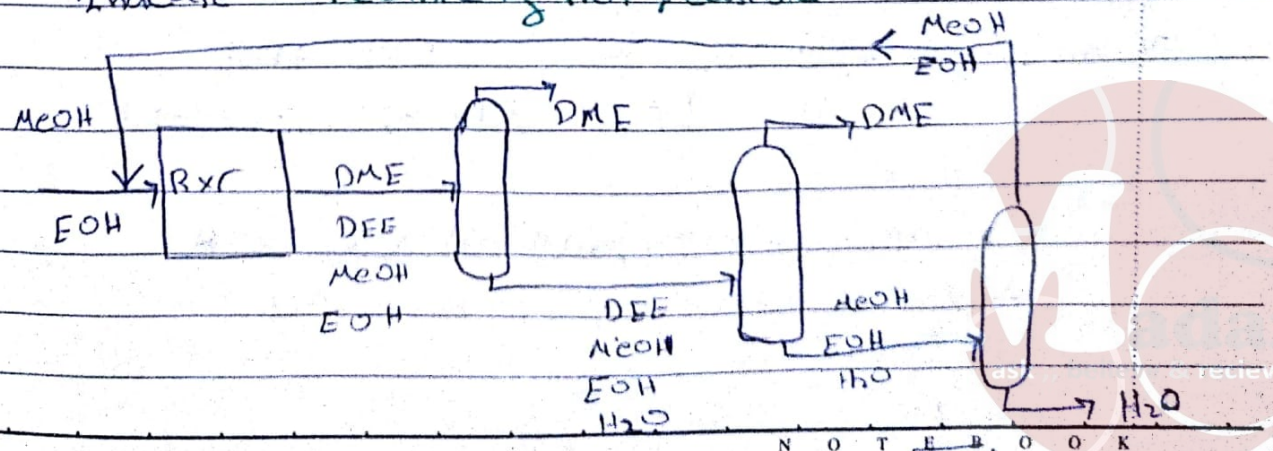
Profit = $ 51.8M − $ 19.4M

= $ 32.4M/yr

PFD 1st case.

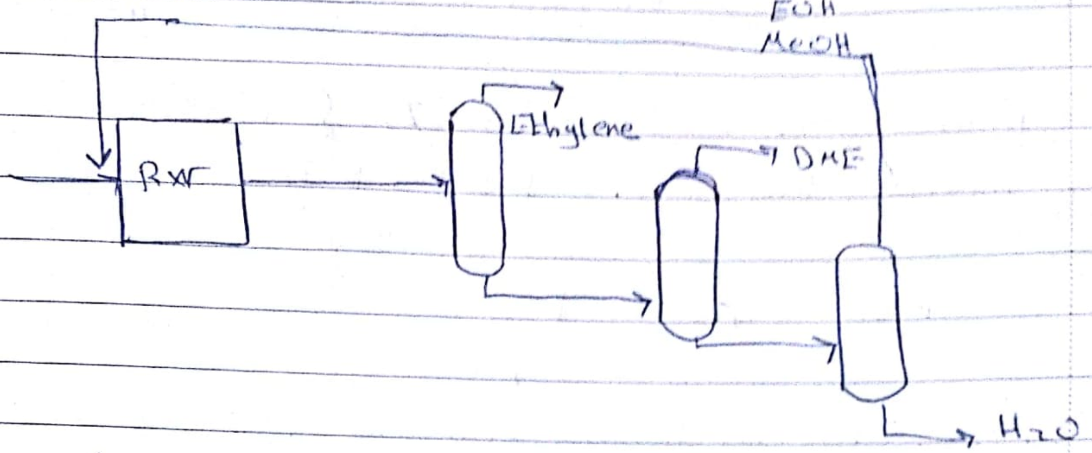


2nd case → technically not feasible

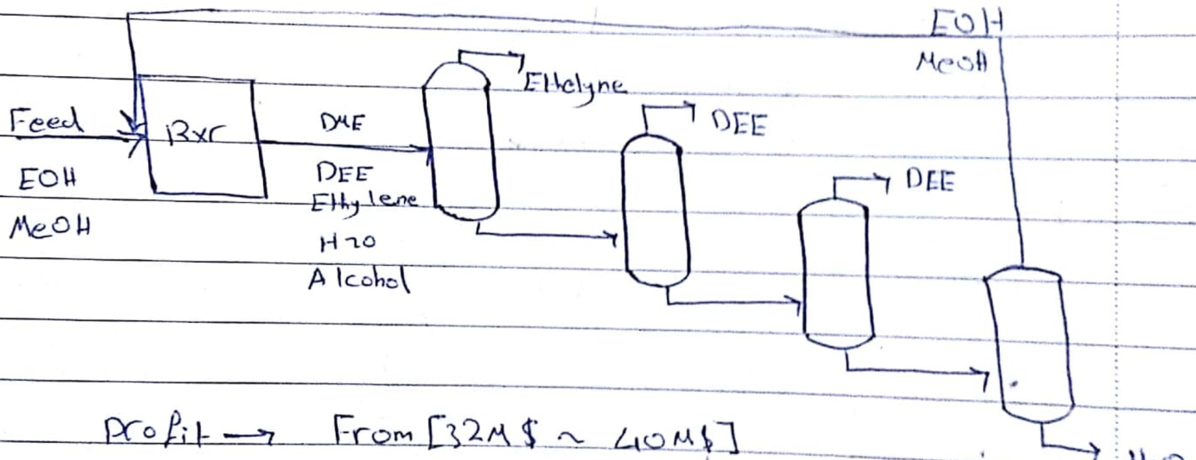


Subject Date No.

3rd case



4th case → technically feasible



Profit → From [32M$ ~ 40M$]

4th case

- Profit ✓
- more than 1 product

بغير الـ condition اذا وقف الطلب

DEE + EOH

Ethylene ← EOH لأنتج DEE ك

↑ From safety of Busniss point of view

* ما بتفرق انا الـfeed process ← %100 pure

∴ Any distarpances will not affect the production

← لازم ابشوف الـ payback period

Subject Process

Date 7th March 2018

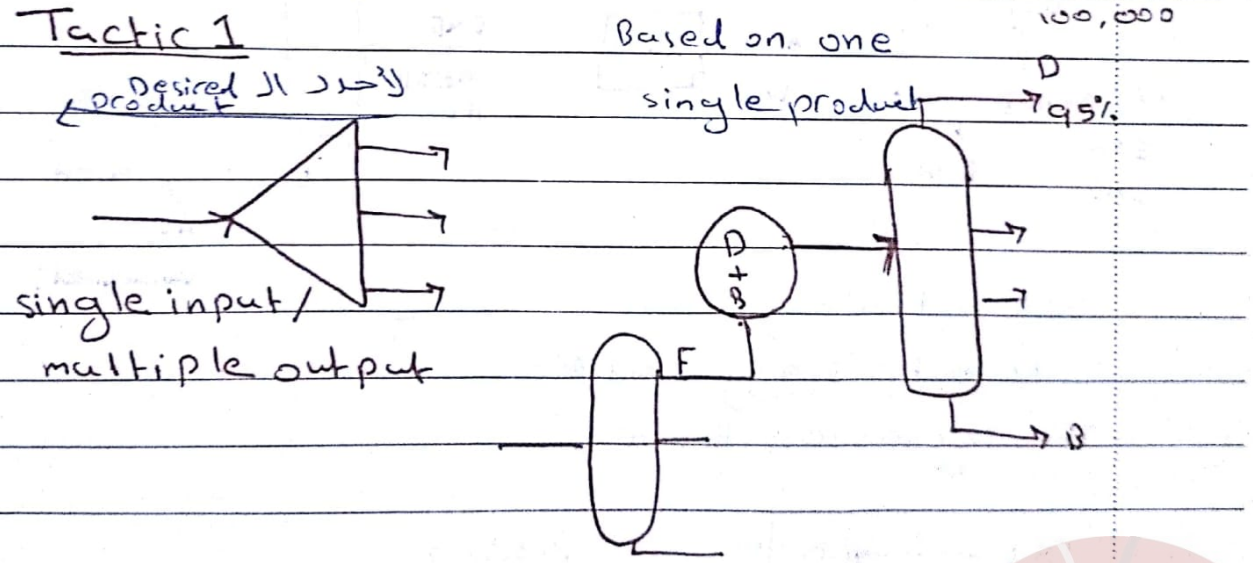
Tracing chemicals through PFD & P&ID

⇒ I must select minimum # of streams To perform a material & Energy balances across the PFD, there are two general guidlines should be followed when tracing chemicals:-

1. Reactants, should start with the feed and trace the chemicals toward the Rxn.
2. products, start from product stream and trace the chemical backward toward the Rxn.

Tactic 1

Based on one

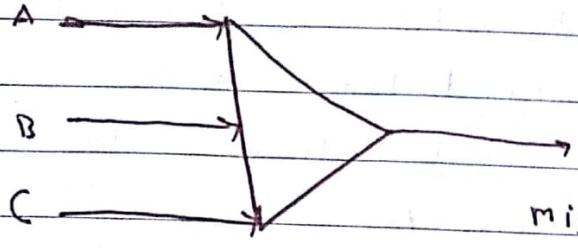


Subject

Date No.

Tactic 2

tracing of chemicals

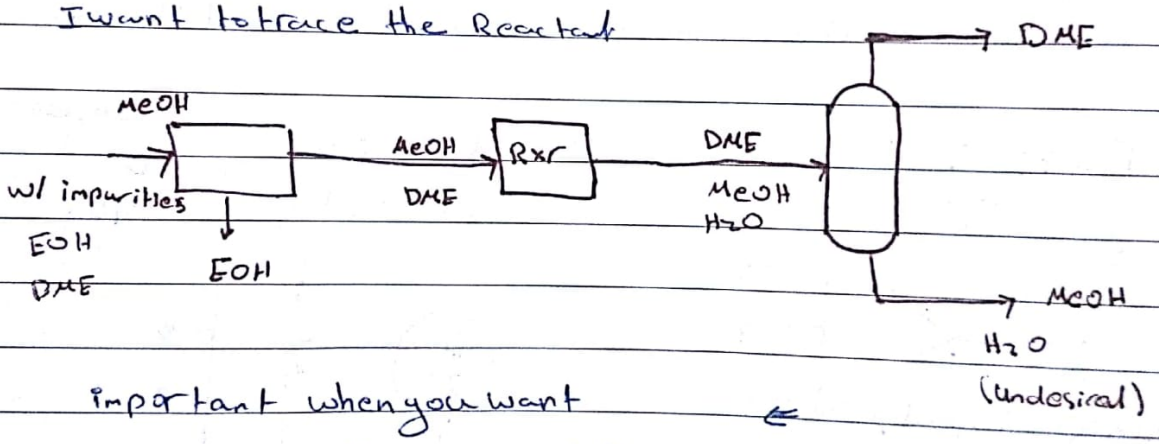


multiple input, single output

Example production of DME

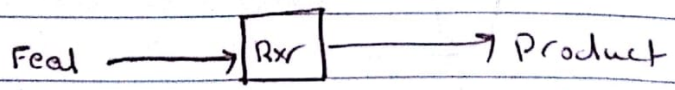
$2\,CH_3OH \rightleftharpoons CH_3OCH_3 + H_2O$

I want to trace the Reactant



important when you want to make Energy and Mass balances

Tracing across the Rxr



n_A
n_B
Unreacted material

$$n_i = n_i^\circ + \sum_{j=1}^{n\text{-Rxns}} \nu_{ij}\,\xi_j$$

ξ_j → extent of Rxn

ν_{ij} → stoich. value

Subject Date No.

Example suppose we have three Rxns

$2MeOH \longrightarrow 1DME + H_2O$ $\{_1$

$2EOH \longrightarrow 1DEE + H_2O$ $\{_2$

$EOH \longrightarrow 1C_2H_4 + H_2O$ $\{_3$

① $n_{MeOH} = n^{\circ}_{MeOH} - 2\{_1$

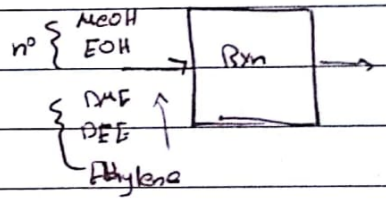
② $n_{DME} = n^{\circ}_{DME} + \{_1$

③ $n_{EOH} = n^{\circ}_{EOH} - 2\{_2 - \{_3$

④ $n_{DEE} = n^{\circ}_{DEE} + \{_2$

⑤ $n_{C_2H_4} = n^{\circ}_{C_2H_4} + \{_3$

recycle



※ of selectivities = n Rxn - 1

conversion specified

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

② $S_1 = \frac{\{_1}{\{_2}$ ← Desired product $\Rightarrow \{_1 = \{_2 S_1$

③ $S_2 = \frac{\{_1}{\{_3} \Rightarrow \{_1 = \{_3 S_2$

multiple Rxn

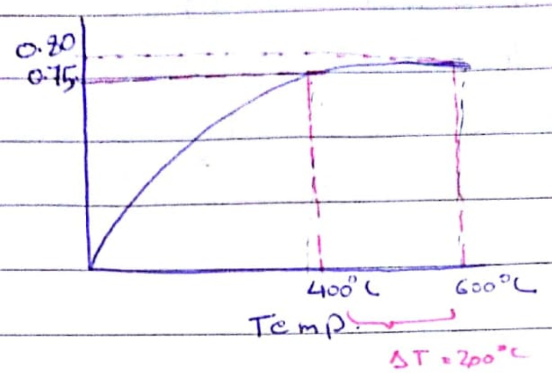
For single Rxn I can use the Rate

Subject

Day: Monday
Date 12th March, 2018 No.

Under standing process conditions

* Reactor



⇒ Why I should go for 400?
* Cost of energy (a lot of utility for 600°C)
* equilibrium of rxn → يتغير الا ردود درجة الحرارة

* Distillation

↑T ↑ΔP Utility↑

advantages For ↑T → conv. ↑

disadvantages → life of process الا يؤثر على

process streams are rarely available at conditions suitable for Rxr and separator Units.
There for, temp., pressure and other streams composition must be adjusted to meet the desired at effective process performance.

N O T E B O O K

37

Subject Date No.

-Effect of pressure

most equipment can operate with pressure up to 10bar without any special material of construction.

P↑ , concentration↑ in gas phase it enhance it toward product.

minimum pressure is 1bar Range 1bar → 10bar

Reasons of high pressure:

(1) Favor equil. conversion P 710bar

(2) Increase the rate of Rxn. (higher concentration in a specific value)

(3) to maintain the liq. phase Rxn.

What is the consequences?

① Increase wall thickness

② Requires expensive compressors

Why do we need to run the process @ low pressure (Vacuum pressure)?

(1) material synsetive to heat

(2) Require gas phase Rxn.

(3) favor the equil. of rxn.

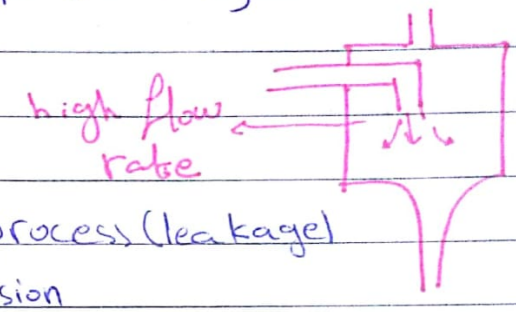
(4) seperation may be easier

Subject Date No.

Consequences?

(1) Larger volume required → high FCI → Investment
Fixed capital

(2) need of vacuum pump or ejectors



(3) air can get into process (leakage)
it can cause explosion

Temperature

ambient condition 25°C → 250°C

T↑ → - damage the equipment
- damage the catalyst
- Increase the utility
- not safe.

Example

The maximum allowable tensile strength for atypical CS (Carbon steel) and SS (Stainless steel) equipments @25°C, 400°C, 550°C are listed below, should we use CS or SS

Temp.	CS	SS
25	1190	1290
400	970	1290
550	170	430

Subject Date No.

Sol.

25°C - 400°C

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

SS: $\frac{1290 - 1290}{1290} = 0$

400°C - 550°C

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

SS: $\frac{1290 - 480}{1290} = 0.67$

25°C - 550°C

CS: 18% + 67% = 85%

SS: 0 + 67% = 67%

go for SS

For 25°C :- 400°C go for CS

CS ثلاث امتحانات الـ SS نعوض الـ

optimum condition not nessarly higher conversion.

optimum الـ temp.

T نثبت الـ P أول

Conditions of special concern for operation of major equipment:-

major equipment :-

(1) Tower

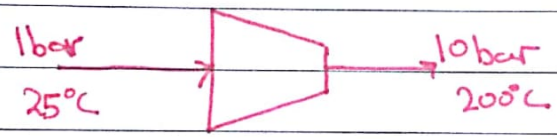
(2) compressor

(3) exchangers

(4) Heaters

(5) Rxr

→ 1. compressors



I need cooling

Ratio should be less than 3:1

$$W = R\,T_{in}\,\frac{\gamma}{\gamma-1}\left[\left(\frac{P_{out}}{P_{in}}\right)^{\frac{\gamma-1}{\gamma}} - 1\right] / \varepsilon$$

ε ← efficiency of compressor

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

P_{in} :- Inlet pressure

P_{out} :- out let pressure

T_{out} ← outlet temp. from compressor

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

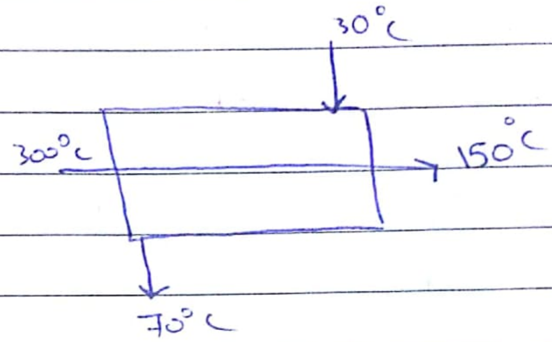
$$Q = \dot{m}\,C_p\,\Delta T$$

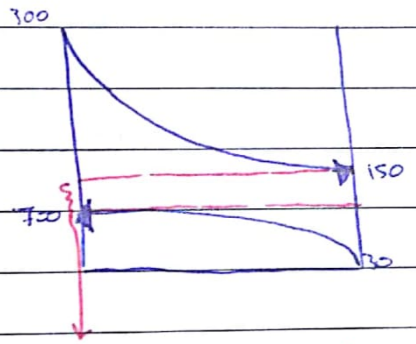
$\Delta T \rightarrow T_{out} - T_{in}$

Optimum ⇒ Wmin. Tmin. So min. Q

بحدد الـ Pout وعلى أثره نحدد W,T
جزء منه بتحول لـ heat

⇒ 2. Exchangers





(Counter-current)

There is a difference so I must increase to opt.

$$LMTD = \frac{(300-70)-(150-30)}{\log_e \frac{300-70}{150-30}} = 169°C$$

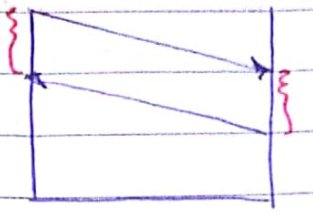
not 150°C

I am wasting energy → max LMTD should be 100°C

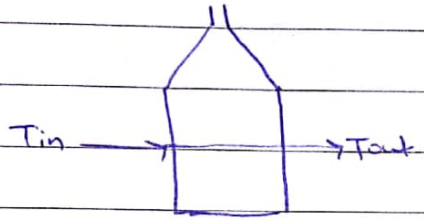
we should run the equip. @ LMTD ≤ 100°C (surface area)

Subject Date No.

difference نفس الـ
so باخد اي Δ فيهم
LMTD دول حسابي



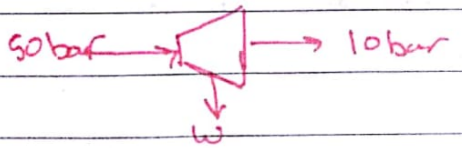
→ 3. Process heater



For Fired heater the
Tout ⩾ Tsat
otherwise → use heat exchanger
only use F.H For start up condition.
Lit T ⩽ Tsat

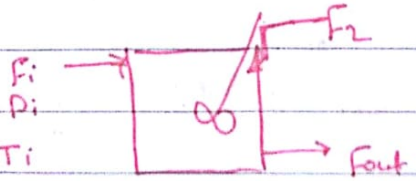
→ 4. Values
should be used at low ΔP

ΔP = 40 bar this is not feasible
Instead use Turbine



Control الـ Turbine اسهل *
so go for Valves

→ 5. mixers



→ if P2 = 30 bar P1 = 2 bar not feasible from it
* you should mix streams of close temp. & pressure or Conc..

* if there is large difference in P & T, capositions this will requir extra equip. cost.

Example:

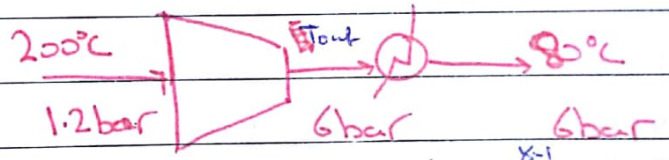
Feed stream (N_2) @ 200°C, 1.2 bar is feed to a process. and its condition is required to be 80°C, 6bar, before entering the Rxr suggest a process condition and design to minimize heat loss and maximize equipment life

Assume compressor efficiency, $\varepsilon = 0.7$

$C_p = 3.5R$, $R = 8.314$

Solution:

* Case 1



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

$$= 200\left[1 + \frac{1}{0.7}\left(\left(\frac{6}{1.2}\right)^{\frac{1.4-1}{1.4}} - 1\right)\right]$$

$$q = mc_p\Delta T = (1)(3.5R)(594-200) = 14917 \ \frac{KJ}{mol}$$

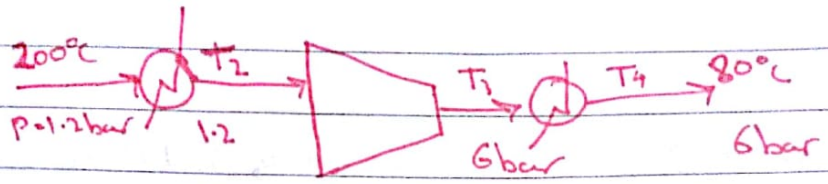
$$W = \frac{RT_{in}}{K}\frac{\gamma}{\gamma-1}\left(\left(\frac{P_{out}}{P_{in}}\right)^{\frac{\gamma-1}{\gamma}} - 1\right)/\varepsilon$$

$$= \frac{R(200+273)}{1.4-1}\frac{1.4}{1.4}\left(\left(\frac{6}{1.2}\right)^{\frac{1.4-1}{1.4}} - 1\right)/0.7$$

$$= 11470 \ KJ/mol$$

Subject Date No.

Case2



$T_1 = 200°C$ assumed

$P_1 = 1.2$ bar

$T_2 = 80°C$

$T_3 = 374°C$ calculated

$P_3 = 6$ bar

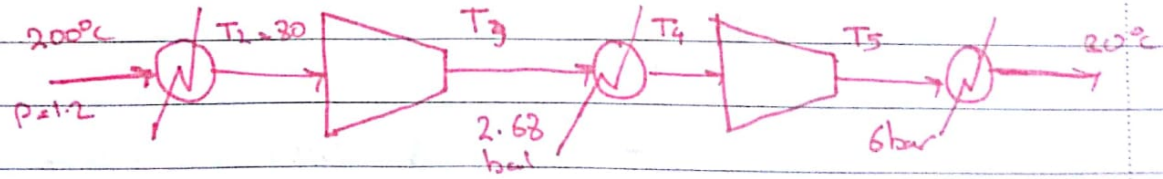
$T_4 = 80$ (should be)

$W = 8560 \frac{KJ}{mol}$

$q_1 = 3490$

$q_2 = 8550$ } $q_{total} =$

Case3



T_2 is assumed $= 80°C$

T_3 is calc. $= 210°C$

T_4 is assumed $= 80°C$

Subject Date No.

T_5 is calc. / $T_5 = 210°C$

T_6 is required $= 80°C$

$W_1 = 3780$ kJ/mol

$W_2 = 3780$ kJ/mol

$q_1 = 3490$ kJ/mol

$q_2 = 3780$ kJ/mol

$q_{total} = 11050$

$W_{total} = 7560$ kJ/mol

Case 1 ⇒ not possible ⇒ error in Ratio.

H.W

⇒ make other trials

Subject Process Date 4th Apr, 2018 No.

Heuristics for Distillation

1. Distillation is the most economical separation if multi component sep. is needed
2. pressure of the column is 1bar above atm p. → more economical separation
   ⇒ Defined by cooling media ⇒ water
   $35 < \Delta T < 65$
3. Ideal mix, $\alpha = \frac{P_1^{\circ}}{P_2^{\circ}}$ ← light / ← heavy
4. optimum Reflux ratio → $(1.2 - 1.5) R_{min}$
5. optimum % of actual trays? ← usually stages are going
   $= 2 N_{min}$
   = 40one
6. $N_{min} = \frac{Ln \frac{(x/(1-x))_{top}}{(x/(1-x))_{bottom}}}{Ln \alpha}$
7. For binary system $x_D \approx 1$
   $R_{min} = \left(\frac{F}{D}\right)\left(\frac{1}{\alpha - 1}\right)$ ⇒ Feed at bubble point Temp.

   $(R_{min} + 1) = \left(\frac{F}{D}\right)\left(\frac{\alpha}{\alpha - 1}\right)$ → Feed at dew point

Subject Date No.

8- For actual # of stages → add 10% as safety factor $N_{actual} = 1.1\ N_{calculated}$

9- Reflux pump is made at least 10% over size

10- For absorption column → Absorption factor

$$\Rightarrow A = \frac{L}{mV}$$

$$1.25 < A < 2.0$$

11- Reflux Drum → Horizental 

more than one phases

12- if we have two phases in the Reflux Drum → diameter 7 0.4m

13- If the tower is 0.9 m in Diameter → then added 1.2 m at the top added 1.8 at the bottom

14- $20 < \text{Tower } L/D < 30$

$\frac{L}{D} \approx 10$ can be used for small diameter

15- Tray spacing (0.5 - 0.6m), up to 0.9 m For cleaning purpose ←

16- max. tray efficiency, depends on vapor factor

0.6 - 0.9

$F = u \rho^{0.5}$

$1.2 < F < 1.5$ based on this values

$u = 0.6$ m/s low P

$u = 1.8$ m/s high P

17- ΔP across tray = 0.007 bar

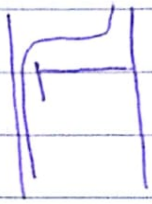
18- tray types
1. sieve tray → hole Diameter 0.6 - 0.7 cm
2. bubble

not for high ΔP

3. valve tray → 3.8 cm

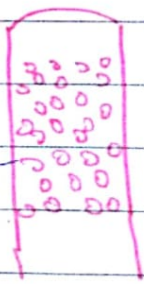
19- weir height = 5cm

Length



Heuristic for packed bed column

1- Random packing are used for columns less than 0.9 m in diameter when low pressure is required



2- low gas rate = 14.2 m^3/min → use 2.5 cm packing

high gas rate = 56.6 m^3/min ⇒ use 5 cm packing

3- L/D > 15 كل ما قل D زاد الطول

زاد ال ΔP

4- if plastic is used length = (6 - 7.6) m

Reason to avoid packing Deformation

5- Packed tower should be operated near 70% of flooding conditions

Subject Date No.

6- Height equivalent to theoritical stages (HETP)

(0.4 - 0.56) → for 2.5 cm pallring

(0.76 - 0.9) → For 5 cm pall ring

7- Δp across packed column

- absorber, regenerator → 0.8 - 2.2 cm H_2O / m of packing

10 m H_2O = 1 atm

- Atmospheric p column

use 3.3 - 6.7 cm H_2O / m of packing

- vacuum tower → 0.8 - 3.3 cm H_2O / m of packing

max. value for all should no exceed

8.37 cm H_2O / 1m packing

Subject Process Date 16th Apr 2012 No.

Heuristics for Rxr Design

1) Batch Rxr

a) Production rate < 1,000,000 lb/year
~ 454,000 kg/year

b) Rxn rate is long

2) CSTR

a) Liquids, slurry Reactants & product

b) Power for CSTR-mixer 0.1-0.3 KW/m3
if heat is transferred, multiply the above value by 3.

c) Ideal CSTR behavior is achieved when the mean RTD (5-10) times Required for homogeneous mixing.

d) Revolution 500-2000 RPM

e) Slow Rxns of liquids and slurries should be conducted to a CSTR.

f) No. of CSTRs in series should be
4-5 ≡ performance of PFR
minimize the volume of CSTR and increase cost & recieve
No. of CSTRs in series

NOTEBOOK

Subject Date No.

PFR → Perfectly non mixed
CSTR → Perfectly mixed

infinite No. of CSTR = 1 PFR

PFR

1) used for high production rate → PFR RTD is changed gradually → more conversion

2) packed bed Rxr (size of catalyst)
- Fluidized bed of (0.1 mm)
- slurry bed (1mm)
- packed bed (2-5mm)

3) Importance of catalyst
(1) change the selectivity
(2) Increase the rate of rxn (Enhance)

then the ← No.1 ← selectivity ← Rate

why? Because the selectivity gives the desired product.

Subject Date No.

4) Effect of Temperature

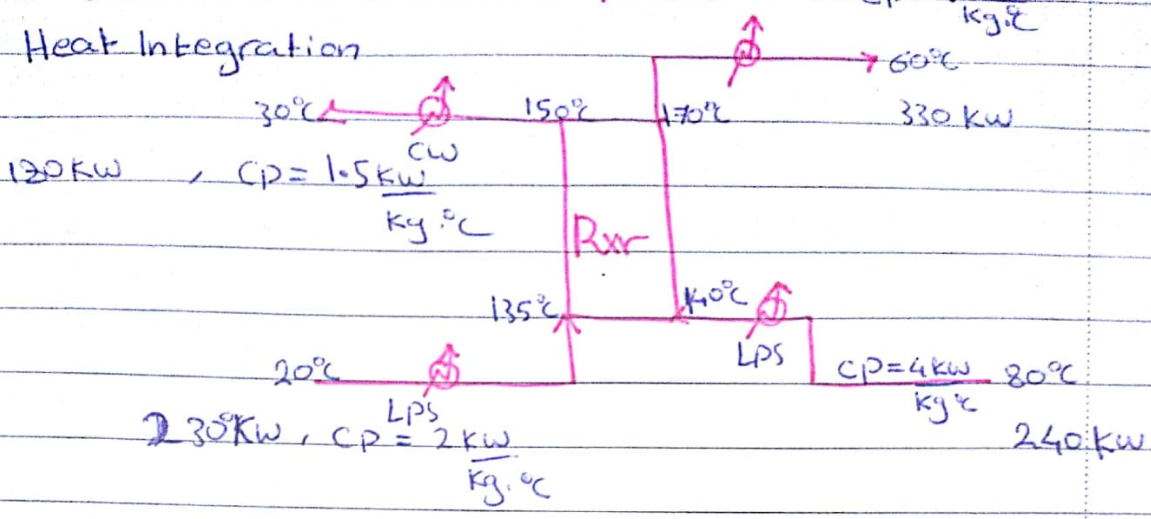
↑ by 10°C ⇒ Double the Rate of Rxn

Subject Process

Day W
Date 18th Apr. 2018 No.

Optimization of chemical process

Heat Integration



$Q = m C_p \Delta T$

$\Sigma Q_H = 470 KW$

$\Sigma Q_C = 510 KW$

$\Sigma Q = 980 KW$

Stream	Ts°C	Tp°C	Cp	Ts	Tp
cold 1	20	135	2	25	140
hot 1	170	60	3	165	55
cold 2	80	140	4	85	145
hot 2	150	30	1.5	145	25

$\Delta T_{min} = 10°C$

Rule 1 :- Add $\frac{\Delta T_{min}}{2}$ → Cold stream

Substract $\frac{\Delta T_{min}}{2}$ → hot stream

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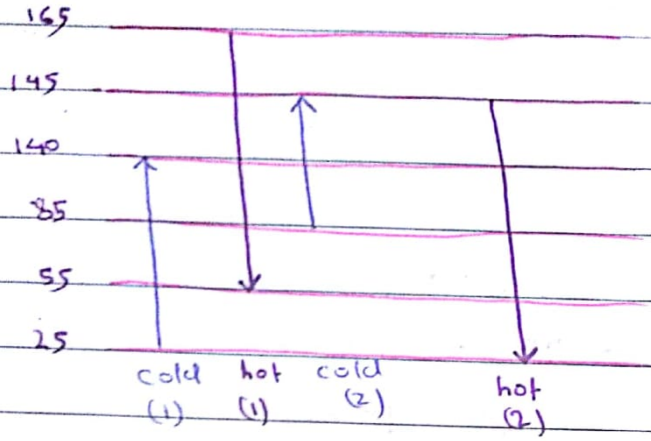
Subject Date No.

$\frac{\Delta T_{min}}{2} = \frac{10}{2} = 5$



Cp	2	3	4	1.5

AFTER INTEGRATION

T	ΔT	$\Sigma mCpc - \Sigma mCph$	ΔHnet (kw)
165	20	-3	-60
145	5	-0.5	-2.5
140	55	1.5	82.5
85	30	-2.5	-75
55	30	0.5	15
25			

حرارة زائدة

Surpulas / Deficite

S
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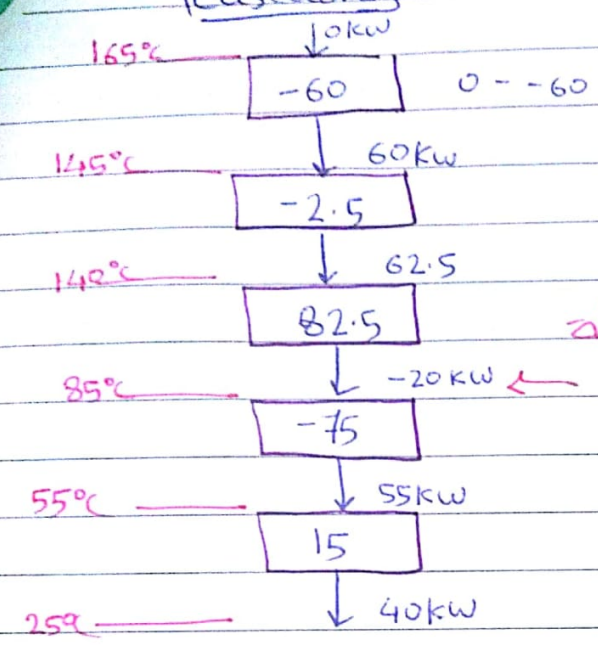
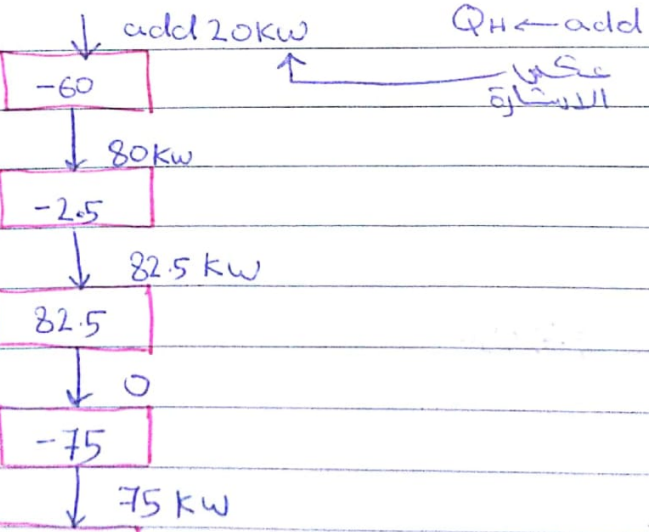
Subject Date No.

Cascade





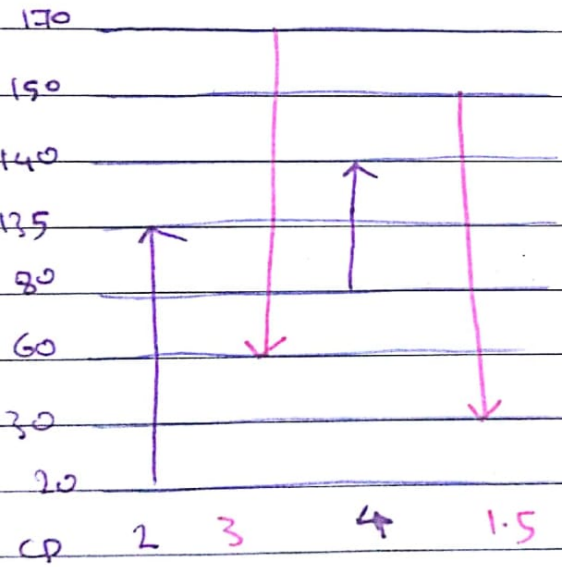
after heat integration

NOTEBOOK

Subject Date No.

Actual interval table

I	ΔT	Σmcpн	Σ Hnet hot	|Σ ΔHnet|
170	20	-3	-60	60
150	10	-4.5	-45	45
140	5	-4.5	-22.5	22.5
135	55	-4.5	-247.5	247.5
80	20	-4.5	-90	90
60	30	-1.5	-45	45
30	10	0	0	0



Σmcpc	Σ ΔHcalc net
0	0
0	0
4	20
6	330
2	40
2	60
2	20

abs بصيف الـ Q بتعت الـ Qc

acc. ΔHh	acc. ΔHc
510	|
450	530
405	510
382.2	180
135	140
45	80
0	60

↑ بيبدأ من الـ lowest ...

لل highest temp.

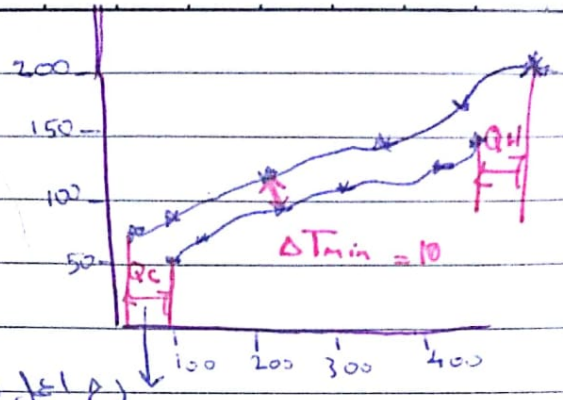
لأنه يعطي duty

أعلى

x axis	y axis	
T	Acc. ΔHh	Acc. ΔHc
170	510	–
150	450	–
140	405	530
135	382.5	510
80	45	180
60	0	140
30	–	80
20		60

NOTEBOOK

Subject Date No.



وهاد اعلى نقطة

لأحد الـ curves

$\Delta T_{min} = 10$ بحيث تكون