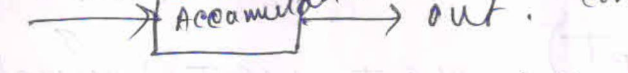
CAPE

④

In - out + generation = Accumulation of mass, momentum, Energy.

$H = U + PV$, U = internal energy = $f(T)$.



Process Engg. Design, operation-control, optimization, intensification

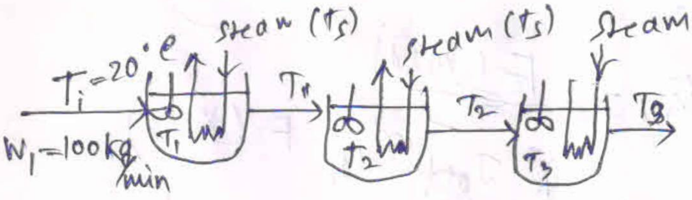
Steadystate Accumulation = 0.

For generation = 0. in = out.

1st
$$\frac{d(MC_pT_1)}{dt} = WC_pT_i - WC_pT_1 + UA(T_{steam} - T_1).$$

2nd
$$\frac{d(MC_pT_2)}{dt} = WC_p(T_1 - T_2) + UA(T_{steam} - T_2)$$

3rd
$$\frac{d(MC_pT_2)}{dt} = WC_p(T_2 - T_3) + UA(T_{steam} - T_3)$$

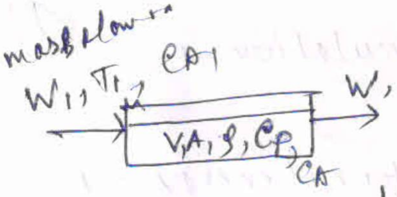


A = Area of heat transfer for coil.

U = overall. heat transfer. coefficient

Himmelblau, D.M., Basic Principles & calculation in chemical Engg. Prentice, Hall.

~~Bruce, A.~~ Finlayson B.A., Introduction to chemical Engineering computing.



$$\frac{d(VA\rho)}{dt} = W_1 - W + \text{generation}$$

$$\frac{d(VA\rho C_p T)}{dt} = W_1 C_p T_1 - W C_p T + \overset{-r_A V}{\text{generation}}$$

For heating system generation $= UA(T_s - T)$.

For reaction " " $=$ Heat of reaction

$= \Delta H_R\,(\cancel{C_A V})\,(-r_A V)$.

$$-r_A = \frac{dC_A}{dt} = -kC_A.$$

Species blance.

$$\frac{d(C_A V\,M_w)}{dt} = \frac{W_1}{\rho} C_{A1} - \frac{W}{\rho} C_A - r_A V.$$

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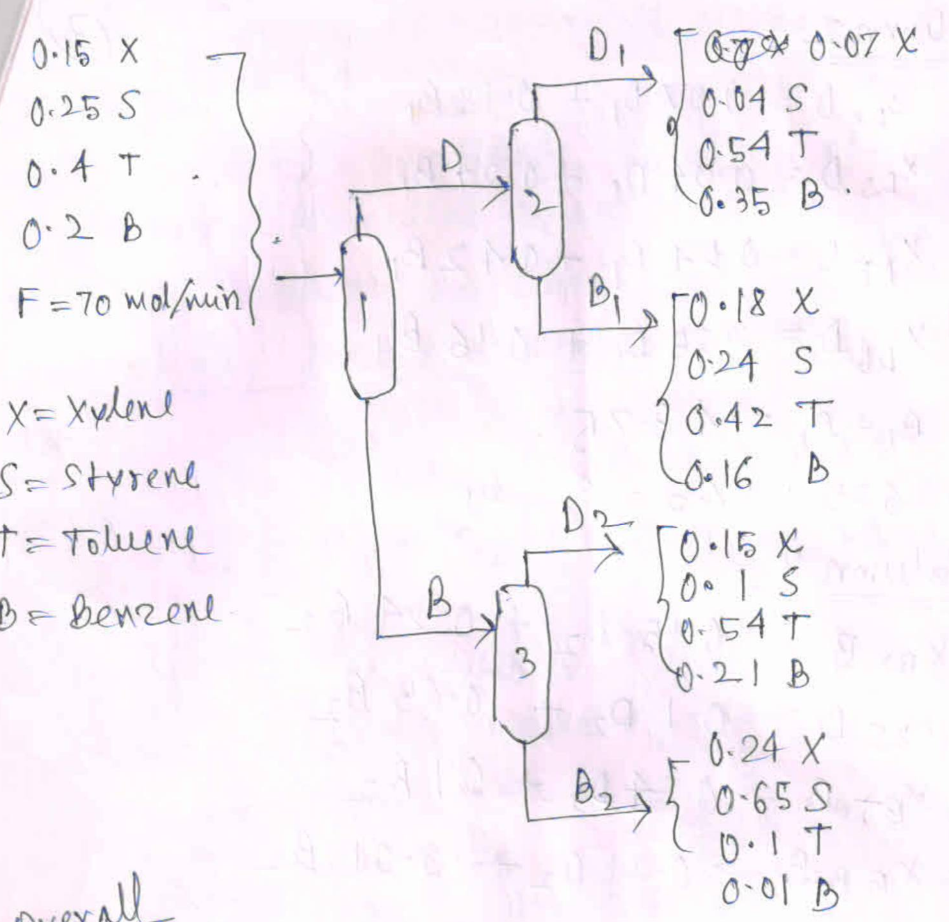
$$\tilde{x}_{new} = \tilde{x}_{old} - J^{-1}(\tilde{x}_{old})\,F(\tilde{x}_{old})$$

$$J = \begin{bmatrix} \frac{\partial F_0}{\partial x_0} & \frac{\partial F_0}{\partial x_1} & \cdots & \frac{\partial F_0}{\partial x_{n-1}} \\ \frac{\partial F_1}{\partial x_0} & \frac{\partial F_1}{\partial x_1} & \cdots & \frac{\partial F_1}{\partial x_{n-1}} \end{bmatrix}$$



X = Xylene
S = Styrene
T = Toluene
B = Benzene

Overall component balance

X: $0.07D_1 + 0.18B_1 + 0.15D_2 + 0.24B_2 = 0.15 \times 70$ — (A1)

S: $0.04D_1 + 0.24B_1 + 0.1D_2 + 0.65B_2 = 0.25 \times 70$ — (A2)

T: $0.54D_1 + 0.42B_1 + 0.54D_2 + 0.1B_2 = 0.4 \times 70$ — (A3)

B: $0.35D_1 + 0.16B_1 + 0.21D_2 + 0.01B_2 = 0.2 \times 70$ → (A4)

Overall material balance

$$D_1 + B_1 + D_2 + B_2 = 70 \quad \text{— (A5)}$$

$D_1 = 26.25$; $B_1 = 17.50$; $D_2 = 8.7499$.
$B_2 = 17.4999$;

On column 2

$X \rightarrow \quad x_{Dx} D = 0.07 D_1 + 0.18 B_1$

$x_{DS} D = 0.04 D_1 + 0.24 B_1$

$x_{DT} D = 0.54 D_1 + 0.42 B_1$

$x_{Db} D = 0.35 D_1 + 0.16 B_1$

$D = B_1 + D_1 = 43.75.$

$= 26.25 + 17.5 = B_1 + D_1$

On column 3

$X \quad x_{BX} B = 0.15 D_2 + 0.24 B_2$

$S \quad x_{BS} B = 0.1 D_2 + 0.65 B_2$

$T \quad x_{BT} B = 0.54 D_2 + 0.1 B_2$

$B \quad x_{BB} B = 0.21 D_2 + 0.01 B_2$

$B = B_2 + D_2 = 8.7499 + 17.5 = 26.25$

$= 26.25$

APE: Review of numerical methods - convergence techniques, Solution of linear and non-linear ~~of~~ algebraic equations, solution of coupled ordinary differential equations. (1)

1. Introduction to chemical Engineering computing, Bruce A Finlayson, John Willy & Sons, Inc.

$$F_1 = \overset{x(1)}{P_1} + \overset{x(2)}{F_2}$$

$F_1 = 100$

$$\overset{x(2)}{F_2} = \overset{x(3)}{P_{2D}} + \overset{x_4}{P_{2B}}$$

$$0.4F_1 = P_1 \times 0.01 + F_2 \overset{x(5)}{X_{F_2}^{T}}$$

$$0.4F_1 = P_1 \times 0.99 + F_2 \overset{x(6)}{X_{F_2}^{B}}$$

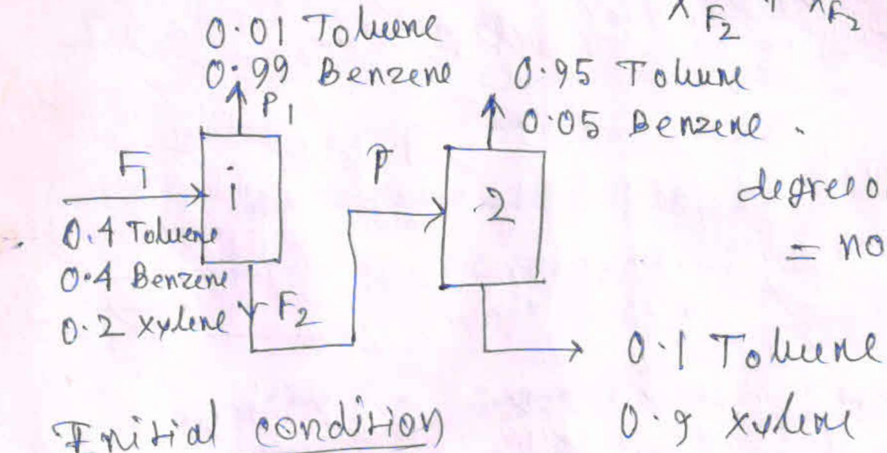
$\overset{x_1}{P_1}, \overset{x_2}{F_2}, \overset{x_3}{P_{2D}}, \overset{x_4}{P_{2B}}$ } variables

$\overset{x_5}{X_{F_2}^{T}}, \overset{x_6}{X_{F_2}^{B}}, X_{F_2}^{X}$ } no of equation

$$F_2 X_{F_2}^{T} = 0.95\, P_{2D} + 0.1\, P_{2B}$$

$$F_2 X_{F_2}^{B} = 0.05\, P_{2D} ~~+ 0.9 P_{2B}~~$$

$$X_{F_2}^{T} + X_{F_2}^{B} + X_{F_2}^{X} = 1.$$



degree of freedom
= no. of variable
− no. of equations.

Initial condition

i.c. $\left\{ \begin{array}{l} F_2 = 500;\ P_{2D} = 250;\ P_{2B} = 250. \\ P_1 = 500;\ X_{F_2}^{T} = 0.~~04~~;\ X_{F_{F_2}}^{B} = 0.01 \end{array} \right.$

With fsolve solver after 8 iteration Solutions are reached fval < 10^{-8}

$$(V-b)\left(P + \frac{a}{V^2}\right) = RT$$

Sol $P_1 = 384.16$ mol/min

$F_2 = 615.83$ "

$P_{2D} = 393.61$ "

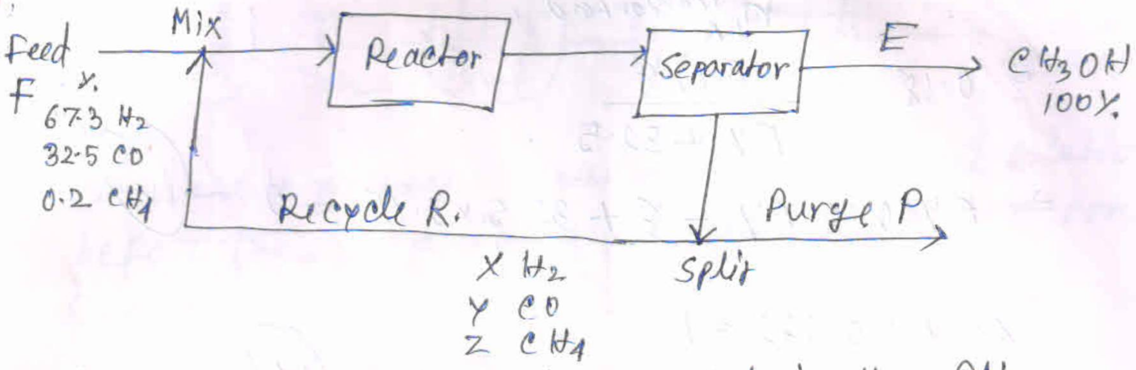
$P_{2B} = 222.22$ mol/min.

$X_{F_2}^{T} = 0.6432$, $X_{F_2}^{B} = 0.03195$

$X_{F_2}^{X} = 0.3247.$

rob Purge-Recycle-Reacting System

Steady state open process $CO + 2H_2 \rightarrow CH_3OH$ The once through conversion of CO in the reactor is 18%.



The purge stream is used to maintain the CH_4 concentration in the exit stream from the separator goin (R & P) at no more than 3.2 mol%.

$\therefore$ Assume $Z = 0.032$.

Select a basis $F = 100$ mol/min.

The overall CH_4 balance

$$0.002 \times 100 = 0.032 P$$
$$\Rightarrow P = 6.25 \text{ mol/min}$$

Unknown variables. E, R, X, Y, ξ = extent of reaction.

Component blance for species; is

$$n_i^{out} = n_i^{in} + \nu_i \xi$$

CO: $6.25Y = 0.325 \times 100 + (-1)\xi$

$6.25Y + \xi - 32.5 = 0$ — ⓘ

H_2: $6.25X = 0.673 \times 100 + (-2)\xi$

$6.25X + 2\xi - 67.3 = 0$ —— (ii)

CH_3OH: $E \times 1 = 0 + 1 \times \xi$

$E - \xi = 0$ — (iii)

Fractional conversion $\cancel{\theta} = 0{\cdot}18 =$

$\therefore\ 0{\cdot}18 = \dfrac{-\nu_{LR}\,\xi}{n_{LR}^{\text{reactor feed}}}$ **LR is CO here**

$\Rightarrow 0{\cdot}18 = \dfrac{-(-1)\,\xi}{RY + 32{\cdot}5}$

$\Rightarrow \cancel{RY}\ 0{\cdot}18\,RY - \xi + 32{\cdot}5 \times 0{\cdot}18 = 0$ — (iv)

$x + y + 0{\cdot}032 = 1$

$\Rightarrow x + y + 0{\cdot}032 - 1 = 0;$ — (V)

$\underset{x_1}{E},\ \underset{x_2}{R},\ \underset{x_3}{x},\ \underset{x_4}{y},\ \underset{x_5}{\xi}$, 5 unknowns

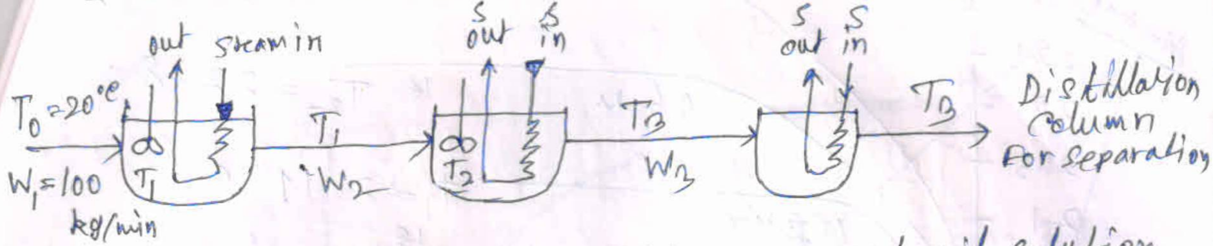
initial condition: 50, 50 0·9 0·1. 100·

With fsolve after 19th iteration

solution is
$E = 31{\cdot}25$. mol/min (x_1)
$R = 705{\cdot}55$ mol/min (x_2)
$x = 0{\cdot}768$ (x_3)
$y = 0{\cdot}2$ (x_4).
$\xi = 31{\cdot}25$ (x_5)

$f_{val} < 10^{-10}$

Series of Strirred tank heaters for oli. heating



It is use to pre heat a multicomponent oil solution before it is fed to a distillation column for Separation.

$$C_p^{oil} = 2.0 \text{ KJ/kg}, \quad W_1 = W_2 = W_3$$

$$Q = UA\,(T_{steam} - T). \quad \underline{\text{at steady State}}$$

$$W_{in} = W_{out}$$

$$UA = 10 \text{ KJ/min/°C} \qquad T_S = 250°C.$$

$$M = \text{mass of oil in each tank} = 1000 \text{ kg}.$$

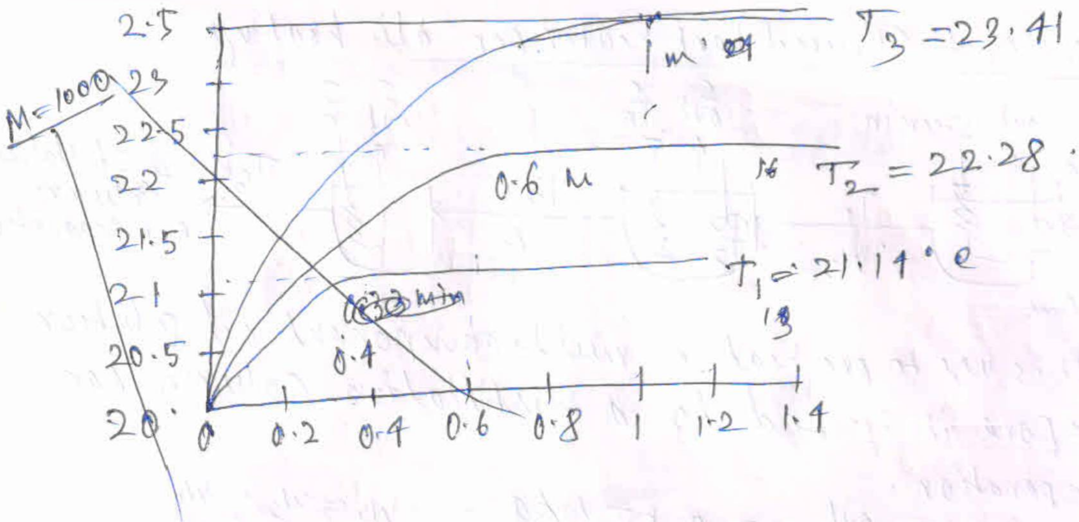
Energy balance equation for three tanks.

$$\frac{d(MC_pT_1)}{dt} = WC_pT_0 - WC_pT_1 + UA\,(T_S - T_1).$$

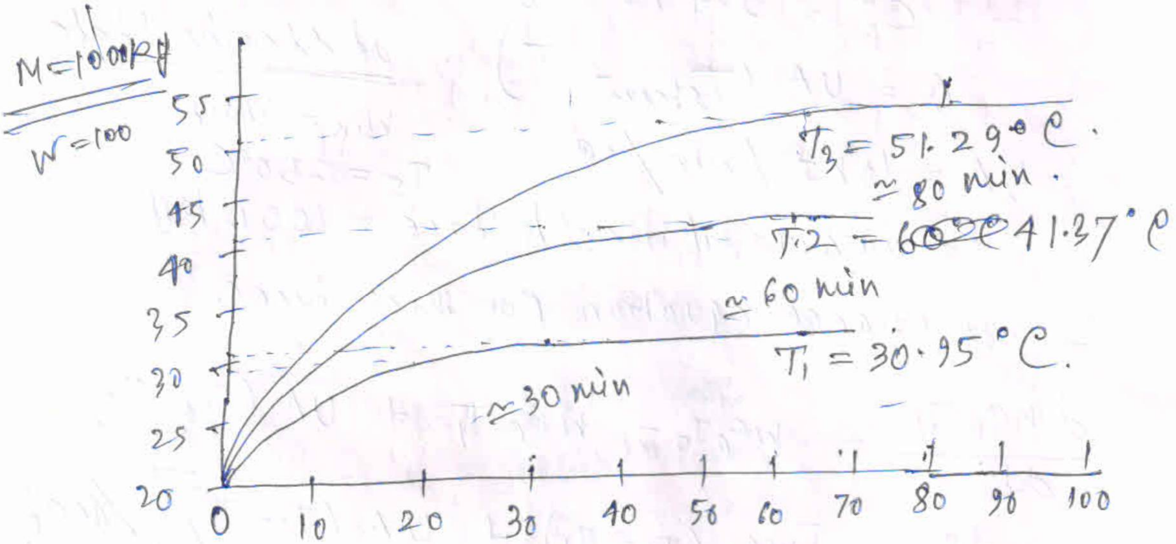
$$\Rightarrow \frac{dT_1}{dt} = \left[WC_p(T_0 - T_1) + UA\,(T_S - T_1)\right] / (MC_p) \quad \text{— ⓘ}$$

$$\frac{dT_2}{dt} = \left[WC_p\,(T_1 - T_2) + UA\,(T_S - T_2)\right] / MC_p \quad \text{— (ii)}$$

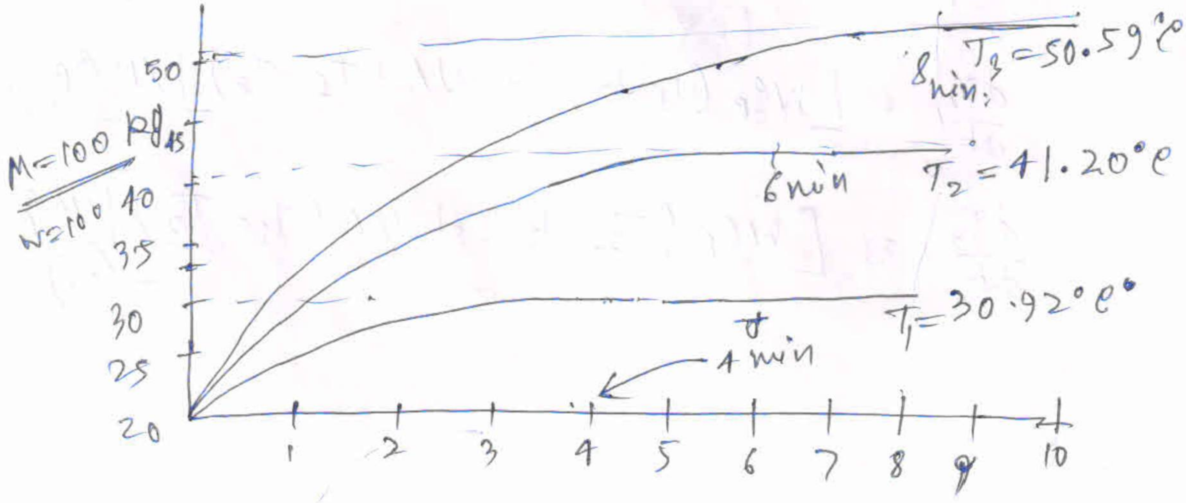
$$\frac{dT_3}{dt} = \left[WC_p\,(T_2 - T_3) + UA\,(T_S - T_3)\right] / MC_p \quad \text{— (iii)}$$



M=1000
2.5
23
22.5
22
21.5
21
20.5
20
0
0.2
0.4
0.6
0.8
1
1.2
1.4
T_3 = 23.41
1m
0.6 m
T_2 = 22.28
T_1 = 21.14 °C
30 min
0.4
13

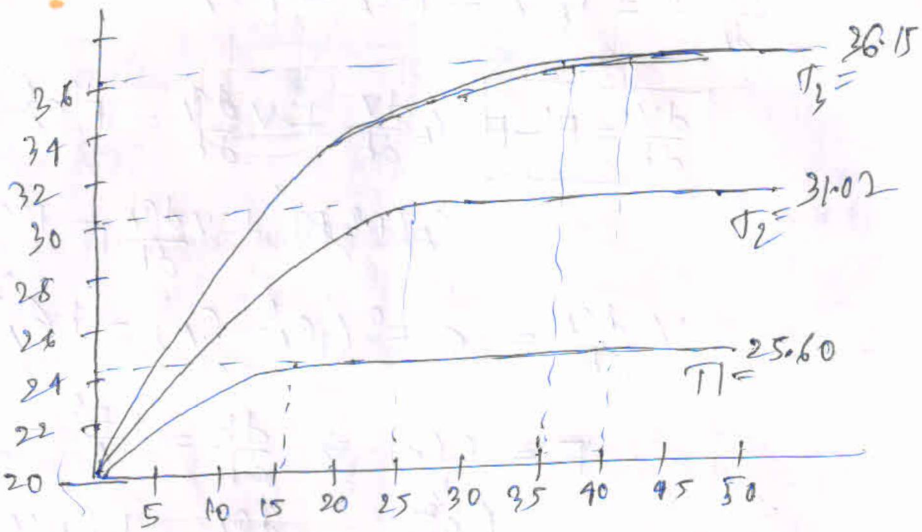


M=100 kg
W=100
55
50
45
40
35
30
25
20
0
10
20
30
40
50
60
70
80
90
100
T_3 = 51.29 °C
≃ 80 min
T_2 = ~~60 °C~~ 41.37 °C
≃ 60 min
T_1 = 30.95 °C
≃ 30 min



M=100 kg
W=100
50
45
40
35
30
25
20
1
2
3
4
5
6
7
8
9
10
T_3 = 50.59 °C
8 min
T_2 = 41.20 °C
6 min
T_1 = 30.92 °C
4 min

W=200
M=1000.



Ode 45 (@functionname, $[0, t_{span}]$, $[T1_0 \cdot T2_0\ T3_0]$)

<u>Runge-kutta 4th order</u> for coupled OdE

$\frac{dY}{dt} = F_1 (Y, W, t)$; $\frac{dW}{dt} = F_2 (Y, W, t)$

with IC Y_0, W_0, t_0.

After Δt time $t_1 = t_0 + \Delta t$.

$Y_1 = Y_0 + \frac{1}{6} (k_1 + 2k_2 + 2k_3 + k_4)$;

$W_1 = W_0 + \frac{1}{6} (l_1 + 2l_2 + 2l_3 + l_4)$;

$k_1 = F_1 (Y_0, W_0, t_0) \Delta t$, $l_1 = F_2 (Y_0, W_0, t_0) \Delta t$.

$k_2 = F_1 (Y_0 + \frac{k_1}{2}, W_0 + \frac{l_1}{2}, t_0 + \frac{\Delta t}{2}) \Delta t$, $l_2 = F_2 (Y_0 + \frac{k_1}{2}, W_0 + \frac{l_1}{2}, t_0 + \frac{\Delta t}{2}) \Delta t$

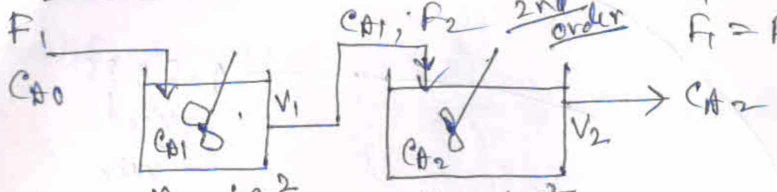
$k_3 = F_1 (Y_0 + \frac{k_2}{2}, W_0 + \frac{l_2}{2}, t_0 + \frac{\Delta t}{2}) \Delta t$, $l_3 = F_2 (Y_0 + \frac{k_2}{2}, W_0 + \frac{l_2}{2}, t_0 + \frac{\Delta t}{2}) \Delta t$

$k_4 = f_1 (Y_0 + k_3, W_0 + l_3, t_0 + \Delta t) \Delta t$, $l_4 = f_2 (Y_0 + k_3, W_0 + l_3, t_0 + \Delta t) \Delta t$

Two CSTR in series $A \rightarrow B$

2nd order

$V_1 = V_2 = V$
$F_1 = F_2 = F$



$-r_{A1} = kC_{A1}^2$ $-r_{A2} = kC_{A2}^2$

$\frac{mol}{s} = k \left(\frac{mol}{lit}\right)^2$

k = rate constant (2nd order).

$k = \frac{mol}{s}\frac{lit^2}{mol^2}$

$= \frac{lit^2}{s\ mol^2}$

$$V_1 \frac{dC_{A1}}{dt} = F_1 C_{A0} - F C_{A1} - kC_{A1}^2 V_1$$

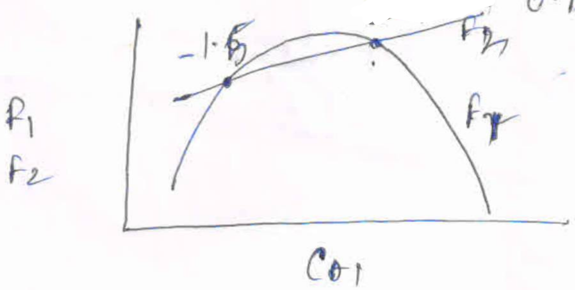
$$\Rightarrow \frac{dC_{A1}}{dt} = \frac{F_1}{V_1}(C_{A0} - C_{A1}) - kC_{A1}^2 \quad \text{— (i)} = f1$$

$$\frac{dC_{A2}}{dt} = \frac{F_2}{V_2}(C_{A1} - C_{A2}) - kC_{A2}^2 \quad \text{— (ii)} = f2$$

$F = 0.05$, $\frac{mol}{s}$ $V = 0.1$, m^3 $k = 1$; $\frac{m^6}{s\ mol^2}$ $C_{A0} = 0.9$. $\frac{mol}{m^3}$.

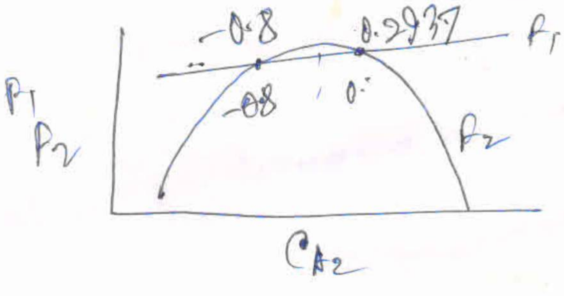
$\frac{F_R}{V_1} C_{A1}$

0.4657. $\frac{mol}{m^3}$



$\frac{f_R}{V_1} - \frac{F_1}{V_1} - 2kC_{A1} - 0$

$\frac{F_2}{V_2} - \frac{F_2}{V_2} - 2kC_{A2}$



$\lambda = \frac{f_R}{V_1} - \frac{f_1}{V_1} - 2kC_{A1}$

$$\frac{d(\rho V)}{dt} = \rho F^0 - \rho F \cdot$$

$$\frac{d(VC_A)}{dt} = F^0 C_A^0 + F C_A - K C_A^2$$

$$\boxed{\frac{dV}{dt} = F^0 - F} \; ; \; C_A \frac{dV}{dt} + V \frac{dC_A}{dt} = F^0 C_A^0 - F C_A - k C_A^2$$

$$C_A(F^0 - \cancel{F}) + V \frac{dC_A}{dt} = F^0 C_A^0 - \cancel{F C_A} - k C_A^2$$

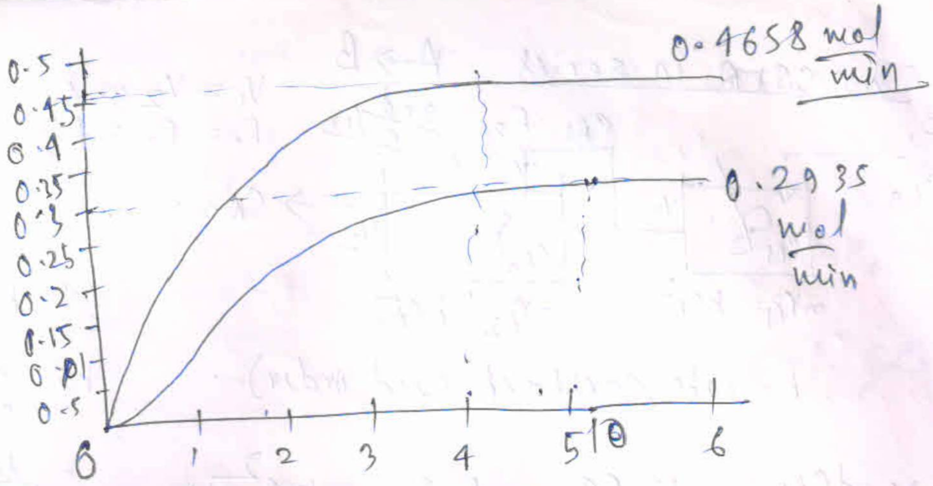
$$V \frac{dC_A}{dt} = \cancel{\rho}\, F^0 (C_A^0 - C_A) - k C_A^2$$

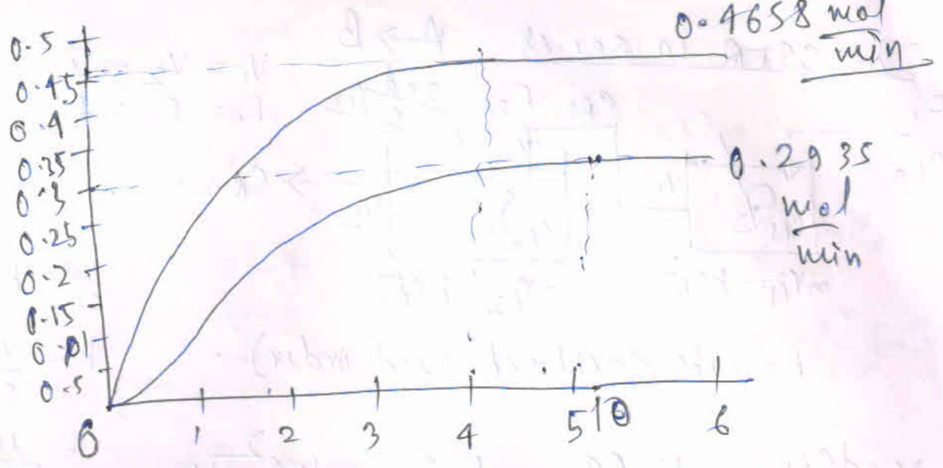
$$F = F(h) \cdot \Rightarrow \frac{dh}{dt} = \frac{F^0}{A} - \frac{R h^{1/2}}{A}$$

$$= R h^{1/2}$$

$$\frac{dC_A}{dt} = \frac{1}{hA}\left[F^0 (C_A^0 - C_A) - k C_A^2\right]$$

$\underline{V = hA}$


0·4658 mol/min
0·2935 mol/min
0·5
0·45
0·4
0·35
0·3
0·25
0·2
0·15
0·01
0·5
0
1
2
3
4
5
θ
6


0.4658 mol/min
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0.35
0.3
0.25
0.2
0.15
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1
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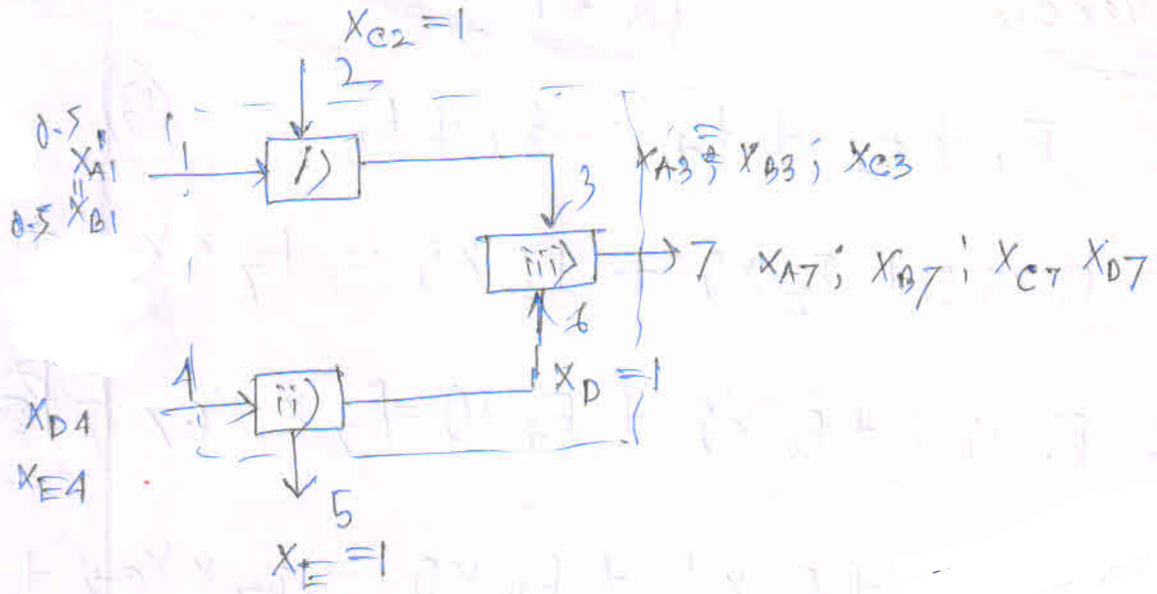
10
6



i) ~~$F_1 + F_2 = F_3$~~ . — ①

A: $F_1 \times \overset{x_{A1}}{0.5} + F_2 \times 0 = F_3 \times x_{A3}$. — ②

B: $F_1 \times \overset{x_{B1}}{0.5} + F_2 \times 0 = F_3 \times x_{B3}$ — ③

C: $F_1 \times 0 + F_2 \times 1 = F_3 \times x_{C3}$. — ④

ii) ~~$F_4 = F_5 + F_6$~~ . — ⑤

D: $F_4 \times x_D = F_5 \times 0 + F_6 \times 1$ — ⑥

E: $F_4 \times x_E = F_5 \times 1 + F_6 \times 0$ — ⑦

iii) ~~$F_3 + F_6 = F_7$~~ .

A: $F_3\, x_{A3} + F_6 \times 0 = F_7 \times x_{A7}$ — ⑧

B: $F_3 \times x_{B3} + F_6 \times 0 = F_7 \times x_{B7}$. — ⑨

C: $F_3\, x_{C3} + F_6 \times 0 = F_7 \times x_{C7}$ — ⑩

D: $F_3 \times 0 + F_6 \times 1 = F_7 \times x_{D7}$. — ⑪

~~$X\ x_{A3} = x_{B3}$~~ — ⑫

Overall 18 equations, 18 variables

~~$F_1 + F_2 + F_4 = F_7 + F_5$~~ — (13)

A: $F_1 \times 0.5 + F_2 \times 0 + F_4 \times 0 = F_7 \times X_{A7} + F_5 \times 0$ — (14)

B: $F_1 \times 0.5 + F_2 \times 0 + F_4 \times 0 = F_7 \times X_{B7} + F_5 \times 0$ — (15)

C: $F_1 \times 0 + F_2 \times 1 + F_4 \times 0 = F_7 \times X_{C7} + F_5 \times 0$ — (16)

D: $F_1 \times 0 + F_2 \times 0 + F_4 X_{D4} = F_7 . X_{D7} + F_5 \times 0$ — (17)

E: $F_1 \times 0 + F_2 \times 0 + F_4 X_{E4} = F_7 \times 0 + F_5 \times 1$ — (18) −5

= (14)

(18) Equations

variables $F_1, F_2, F_3, F_4, F_5, F_6, F_7$ → (Flowrates).

16

$X_{A3}, \overset{2}{X_{B3}}, X_{C3}, X_{A7}; \overset{7}{X_{B7}}; X_{C7}: X_{D7}$

$X_{D4}, X_{E4}, X_{A1}, X_{B1}$ → 11

compositions. → 18 variables

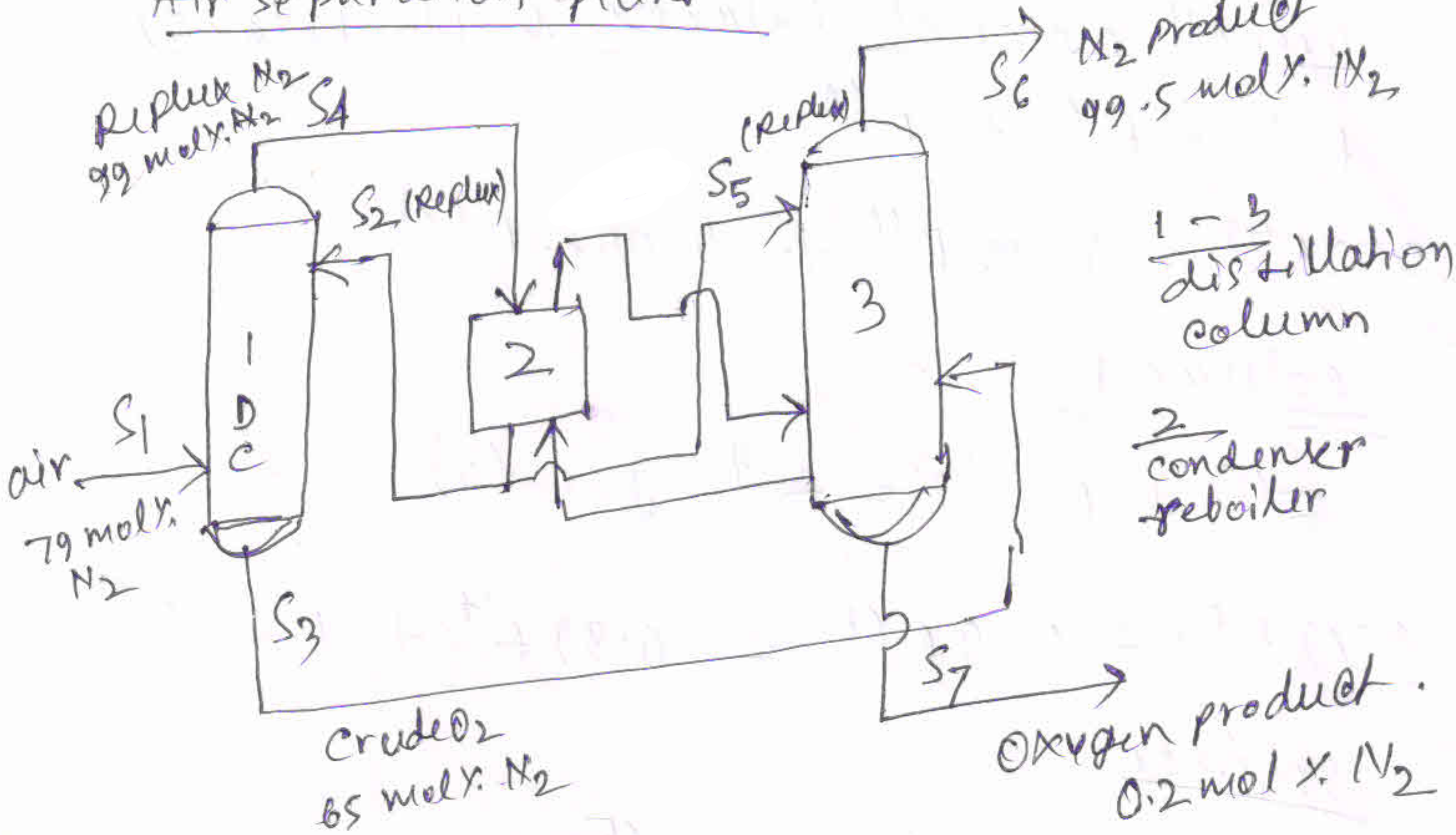
~~$X_{A1} + X_{B1} = 1$~~ — (15)

$X_{A3} + X_{B3} + X_{C3} = 1$ —

(17) $X_{A7} + X_{B7} + X_{C7} + X_{D7} = 1$ — (17)

$X_{D4} + X_{E4} = 1$ — (18)

Air separation plant



$x_2^{(1)} = 0.21$

$x_2^{(3)} = 0.35$

$x_2^{(4)} = 0.01 = x_2^{(2)} = x_2^{(5)}$

$x_2^{(6)} = 0.005, \quad x_2^{(7)} = 0.998$

$F^{(5)} = 3F^{(2)}$ given

1 → N_2
2 → O_2

$F_7 = \underline{\underline{226.44}}$ mol.

Reflux to the 2nd column is three times that of the first column.

Overall material balance & N_2 balance

$$F^{(1)} = F^{(6)} + F^{(7)}.$$

$$0.79\,F^{(1)} = 0.995\,F^{(6)} + 0.002\,F^{(7)}$$

column 1

$$F^{(1)} + F^{(2)} = F^{(4)} + F^{(3)}.$$

$$0.79\,F^{(1)} + 0.99\,F^{(2)} = 0.99\,F^{(4)} + 0.65\,F^{(3)}$$

splitter

$$F^{(4)} = F^{(2)} + F^{(5)}$$

$$F^{(5)} = \cancel{2}\,3F^{(2)}.$$

$$F^{(1)} \ldots F^{(7)}.$$

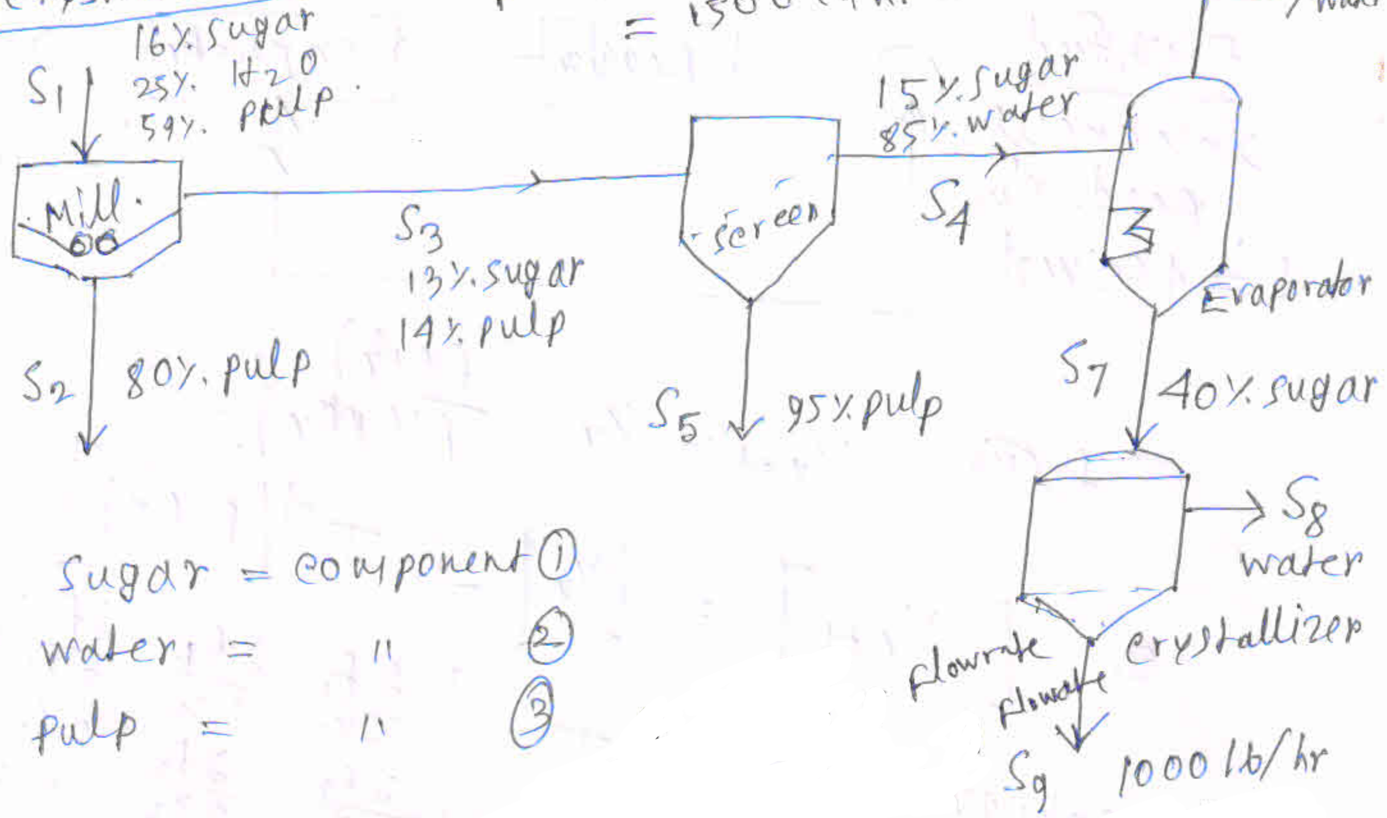
$F^{(4)} = \frac{\overset{8}{\cancel{40}}}{\underset{3}{\cancel{15}}} \times 2500$

$F^{(7)} = 2500$ lb/hr

$F^{(8)} = 250\cancel{0} \times \frac{06}{1\cancel{0}}$

$= 1500$ lb/hr.

Sugar crystalization



sugar = component ①

water = " ②

pulp = " ③

$$\begin{cases} 0.16 F^{(1)} = \omega_1^{(2)} F^{(2)} + 0.13 F^{(3)} \\ 0.25 F^{(1)} = \omega_2^{(2)} F^{(2)} + 0.73 F^{(3)} \\ 0.59 F^{(1)} = 0.8 F^{(2)} + 0.14 F^{(3)} \\ \omega_1^{(2)} + \omega_2^{(2)} + 0.8 = 1 \end{cases}$$

mill, sugar.
" , water
" , pulp
Stream 2

$$\begin{cases} 0.13 F^{(3)} = 0.15 F^{(4)} + \omega_1^{(5)} F^{(5)} . \\ 0.73 F^{(3)} = 0.85 F^{(4)} + \omega_2^{(5)} F^{5} . \\ 0.14 F^{(3)} = 0.95 F^{(5)} . \\ \omega_1^{(5)} + \omega_2^{(5)} + 0.95 = 1 . \end{cases}$$

Screen. sugar
Screen, water
Screen, pulp
Stream 5

$$\begin{cases} 0.85 F^{(4)} = F^{(6)} + 0.6 F^{(7)} ; \\ 0.15 F^{(4)} = 0.4 F^{(7)} \end{cases}$$

Evaporator, water
" , sugar

$$\begin{cases} 0.6 F^{(7)} = F^{(8)} . \\ 0.4 F^{7} = 1000 ; \end{cases}$$

$F^{(7)} = \frac{\overset{500}{\cancel{1000}}}{0.\underset{2}{\cancel{4}}} \times \cancel{10}\,5 = 2500 .$

Crystallizer , water
" , suger

Single reaction - idealsolution

outlet $\rightarrow N_i^{o} = N_i^{(F)} + \frac{\upsilon_i}{-\upsilon_A} x_A N_A^{(F)}$ $\left[\xi_i = \frac{x_A N_A^{F}}{-\upsilon_A} \right.$

A is limiting reactant.

x_A = is conversion.

$$Q + \sum_{\substack{\text{all species} \\ \text{at inlet}}} N_i^{(F)} H_i(T^F, P^F)$$

$$= \sum_{\substack{\text{all species} \\ \text{at outlet}}} \left[N_i^{(F)} + \frac{\upsilon_i}{-\upsilon_A} x_A N_A^{F} \right] H_i(T, P)$$

$$Q + \frac{x_A N_A^{(F)}}{-\upsilon_A} \cancel{\Theta} \left[-\sum \upsilon_i H_i(T, P) \right]$$

$$= \sum N_i^{(F)} \left[H_i(T, P) - H_i(T^F, P^F) \right]$$

$N^{i}_{out} = N^{i}_{in} + \xi \upsilon_i$

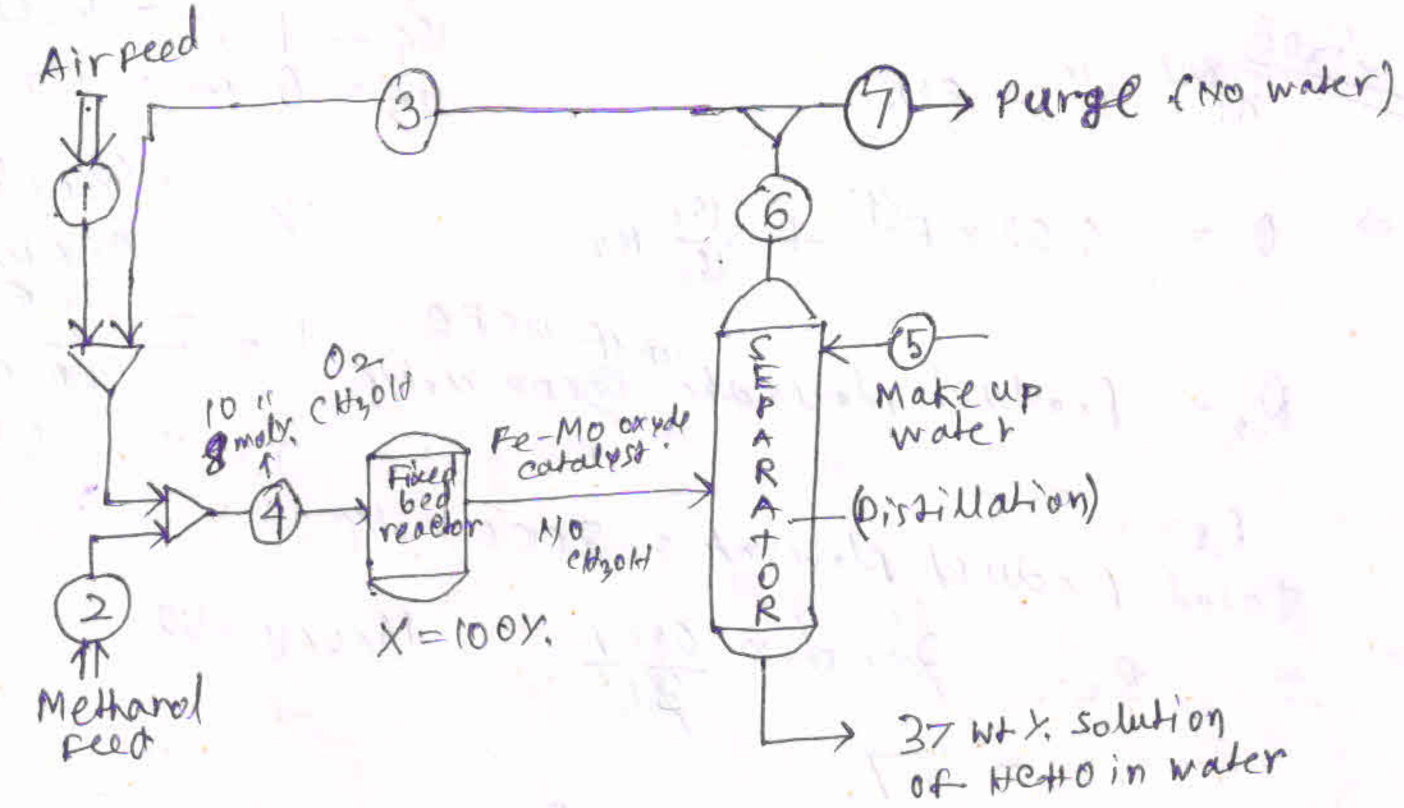
$x_A = \frac{-\upsilon_{LR}\xi}{F^{i}_{LR\,in}}$

$\xi = \frac{x_A F_{LR}}{-\upsilon_{LR}}$

$N^{i}_{out} = N^{i}_{in} + \frac{\upsilon_i}{(-\upsilon_{LR})} x_A F_{LR}$

Production of formaldehyde

$$CH_3OH + \frac{1}{2}O_2 \xrightarrow{\text{Iron-molybdenum oxyde catalyst.}} HCHO + H_2O$$



calculate flowrate, composition & R.

$CH_3OH - 1$; $O_2 - 2$; $HCHO - 3$; $H_2O - 4$; $N_2 - 5$

$x_2^{(1)} = 0.21$; $x_5^{(1)} = 0.79$; $x_1^{(4)} = 0.08$; $x_2^{(4)} = 0.1$; $x_5^{(4)} = 0.82$

Mixers before reactor

CH3OH:
$$F^{(2)} = x_1^{(4)} F^{(4)}$$
$$F^{(2)} = 0.08\, F^{(4)} \quad - ①$$

O_2:
$$x_2^{(1)} F^{(1)} + x_2^{(3)} F^{(3)} = x_2^{(4)} F^{(4)}.$$
$$0.21\, F^{(1)} + x_2^{(3)\,①} F^{(3)} = 0.1\, F^{(4)} - ②$$

N_2:
$$x_5^{(1)} F^{(1)} + x_5^{(3)} F^{(3)} = x_5^{(4)} F^{(4)}$$
$$0.79\, F^{(1)} + x_5^{(3)\,②} F^{(3)} = 0.82\, F^{4} - ③$$

definition $x_5^{(3)} + x_2^{(3)} = 1 \quad - ④$

Reactor + separator.
+

$\nu_1 = -1 \rightarrow CH_3OH$
$\nu_2 = -1/2 \rightarrow O_2$
$\nu_3 = 1 \longrightarrow HCHO$
$\nu_4 = 1 \longrightarrow H_2O$
$\nu_5 = 0 \longrightarrow N_2$

CH_3OH

$n_i^{out} = n_i^{in} + \xi \nu_i$

$\Rightarrow \quad 0 = 0.08 \times F^{(4)} + \frac{\nu_1}{\nu_3} R_3$

$X = \frac{-(\nu_{LR})\xi}{n_{LR\,feed\,input}}$

$1 = \frac{-(-1)\xi}{~~\nu_{LR}~~ \cdot n_{LRFi}}$

$\xi = n_{LFR} \cdot 0.9$

$R_3 =$ Product flow rate of HCHO mol/s.

Total product flowrate $= 3000$ mol/s.

$\therefore \quad R_3 = \cancel{3000} \times \frac{0.37}{\cancel{30}}$

$= 37$

$M_{HCHO} = 30.$

$0.08\, F^{(4)} - F^{(8)} \frac{0.37}{30} = 0$ — ⑤

H_2O:

$\frac{0.63 \times F_8}{18} = F^{(5)} + \frac{\nu_4}{\nu_3} F^{(8)} \times \frac{0.37}{30}$

$\frac{0.63}{18} F_8 = F^5 + F^{(8)} \frac{0.37}{30}$ — ⑥

O_2:

$X_2^{(6)} F^{(6)} = 0.1\, F^{(4)} + \frac{\nu_2}{\nu_3} F^{(8)} \frac{0.37}{30}$

$\therefore X_2^{(6)} F^{(6)} = 0.1\, F^{(4)} - \frac{1}{2} F^{(8)} \frac{0.37}{30}$ — ⑦

N_2

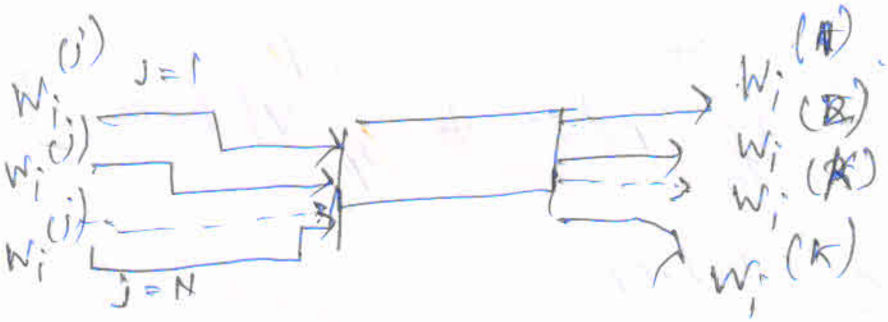
$X_5^{(6)} F^{(6)} = 0.82\, F^{(4)}$. — ⑧

splitter $F^{(3)} + F^{(7)} = F^{(6)}$. — ⑨

$F_0^1, F_0^2, F^3, F_5^5, \cancel{F^8}, F^6, F^7, F^4,$ ⑧ variables $x_2^3\, x_5^3,\, x_2^6,\, x_5^6$

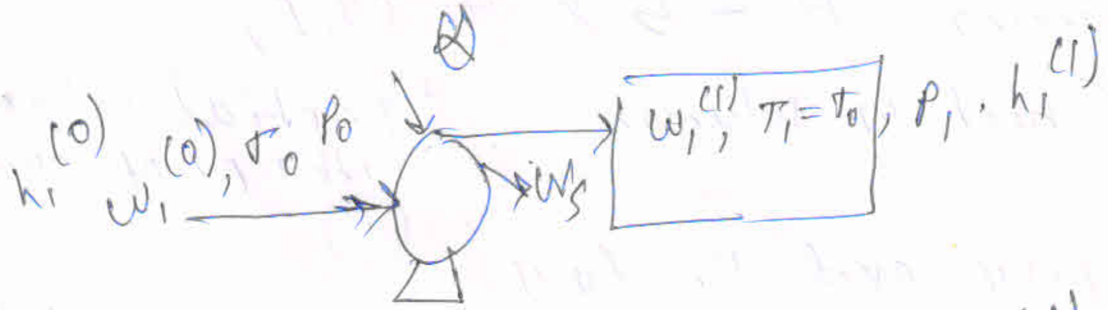
in macroscopic view

$$Q + \sum w_i^{(j)} \left[\bar{h}_i + \frac{1}{2} v^2 + g h \right]_{\text{stream } j} = \sum w_i^{(k)} \left[\bar{h}_i + \frac{1}{2} v^2 + g h \right]_{\text{stream } k} + W_s$$



shaft work

W_s = shaft work



$$Q + \sum w_i^{(0)} \bar{h}_i^{(0)} = \sum w_i^{(1)} \bar{h}_i^{(1)} + W_s$$

Enthalpy of liquid & Solid

$$H = H(T_0, P_0) + \int_{T_0}^{T}\left(\frac{\partial H}{\partial T}\right)_P dT + \int_{P_0}^{P}\left(\frac{\partial H}{\partial P}\right)_T dP$$

$$H = \int_{T_0}^{T} C_p\, dT + \int_{P_0}^{P}\left[V - T\left(\frac{\partial V}{\partial T}\right)_P\right] dP + H(T_0, P_0)$$

$$\alpha_T = \frac{1}{V_T}\left(\frac{\partial V}{\partial T}\right)_P$$ = coefficient of thermal expansion.

$$H = H(T_0, P_0) + \int_{T_0}^{T} C_p\, dT + \int_{P_0}^{P} V[1 - \alpha_T T]\, dP$$

$\frac{PV}{m}$

Non ideal liquid solution.

$\Delta H_{mixing} = H - \sum x_i \tilde{H}_i^{+}(T, P)$.

↳ heat of solution

↳ partial molar enthalpy of component ⓘ

Energy flow and 1st law in flow system

total energy = kinetic + potential + internal + work done on the body by pressure force

$\frac{m^2}{s^2}$

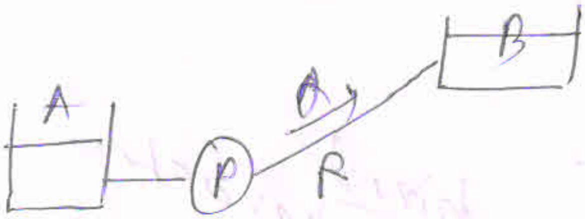
$= \frac{1}{2}v^2 + gh + u + \frac{P}{\rho}$ Energy/unit mass

$\frac{J}{kg}$ $\frac{kg}{m^3}$

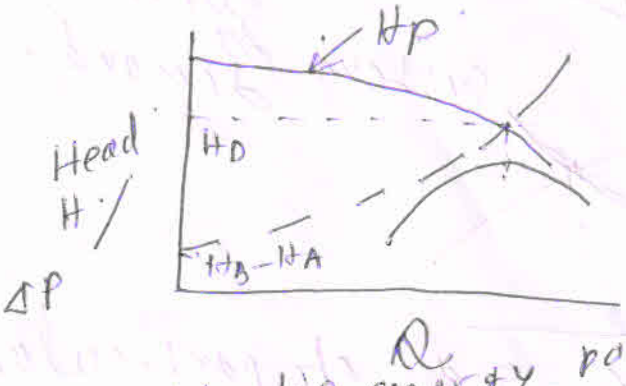
$u + \frac{P}{\rho} = \left(u + \frac{PV}{m}\right) = h_i$

$= h_i$ = enthalpy/unit mass

$= \frac{1}{2}v^2 + gh + h_i$



$$\dot{W}_p = \frac{\gamma Q H_p}{\eta}$$



H_D = demand head.

$$f = \frac{\tau_0}{\frac{1}{8}\rho V^2}$$

assume kinetic energy parts are negligibl

$$H_B - H_A = \Sigma h_f = \left(f\frac{L}{D}\frac{1}{2gA^2} + \frac{\Sigma K}{2gA^2}\right)Q^2$$

$$h_L = \frac{fLV^2}{D\,2g}$$

Darcy-weisbach eq%.

$$\frac{P_B}{\rho g} + Z_B - \frac{P_A}{\rho g} - Z_A + h_L = H_p$$

~~$H_B - H_A$~~ $H_B - H_A + h_L = H_p$

$H_B - H_A +$ $RQ^2 = H_p$

put H_p = at maximum efficiency

colebrook-white

$$\frac{1}{\sqrt{f_D}} = -2\log\left(\frac{2.51}{Re\sqrt{f_D}}\left(1+\frac{R}{3.3}\right)\right)$$

$$\frac{1}{\sqrt{f}} = -2\log\left(\frac{\varepsilon}{3.7D_h} + \frac{2.51}{Re\sqrt{f}}\right).$$

Moody

$$f = 0.0055\left[1 + \left(2\times10^4\cdot\frac{\varepsilon}{D} + \frac{10^6}{Re}\right)^{1/3}\right]$$

$$\varepsilon/D = 0 - 0.01$$

Head loss in a pipeline h_L is

$$h_L = f \frac{L}{D} \frac{V^2}{2g}$$ [Darcy-weisbach equation]

valid both for laminar & turbulent flow.

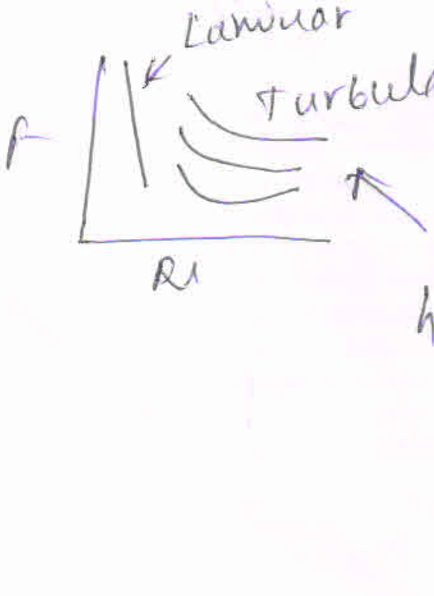
for laminar flow $f = \frac{64}{Re}$.

f = friction factor.

$$\frac{-\dot{W}_s}{\dot{m} g} = \frac{V_2^2}{2g} + \frac{P_2}{\rho g} + z_2 \cancel{P_1} - \frac{P_1}{\rho g} - \frac{V_1^2}{2g} - z_1 + h_L$$

$$h_L = k \frac{V^2}{2g}, \quad k = \text{loss co-eff.}$$

$$f = \text{function}\left(\frac{e}{D}, \frac{VD\rho}{\mu}\right).$$ $\frac{e}{D}$ relative roughness.



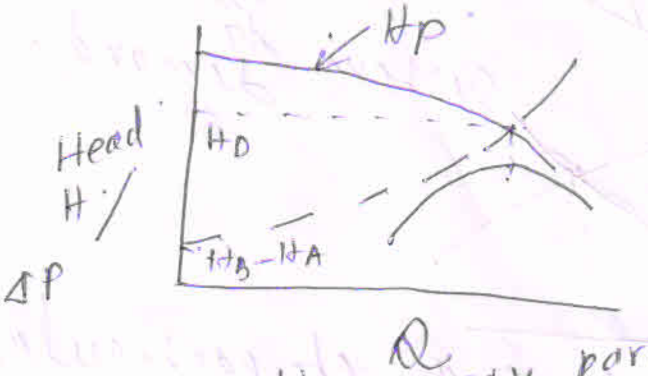
It is presented in Moody diagram,

$$h_L = 1.07 \frac{Q^2 L}{g D^5} \left\{ \ln\left[\frac{e}{3.7D} + 4.62\left(\frac{\nu D}{Q}\right)^{0.9}\right]\right\}^{-2}$$

$$10^{-6} < \frac{e}{D} < 10^{-2}, \quad \underline{3000 < Re < 3 \times 10^8}$$

$$h_L = \frac{V^2}{2g}\Big[\underbrace{f \frac{L}{D}}_{\text{straight pipe}} + \Sigma K\Big]$$ → bending, expansion, contraction

$$W_p = \frac{\gamma Q H_p}{\eta}$$



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Darcy-Weisbach eq%.

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$$\cancel{H_B - H_A}\ H_B - H_A + h_L = H_p$$

$$H_B - H_A + RQ^2 = H_p$$

put H_p = at maximum efficiency

colebrook – white

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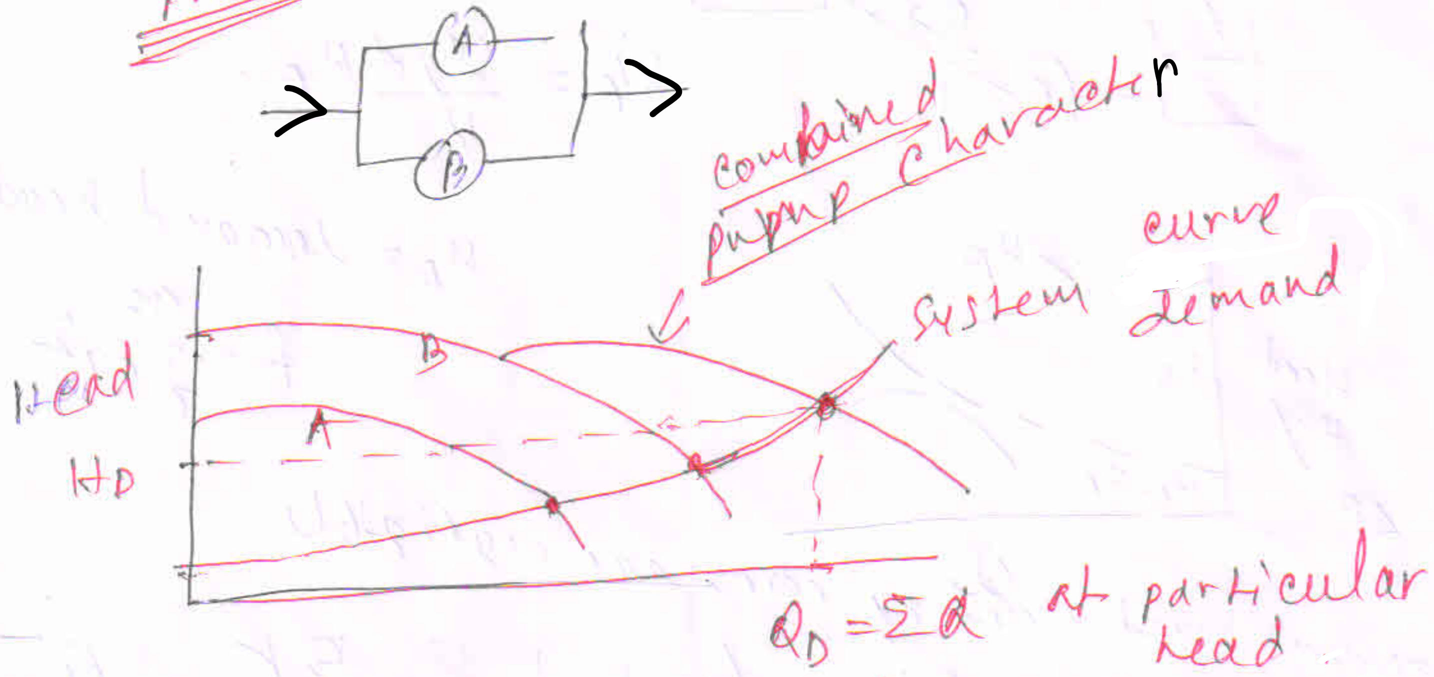
$$\frac{1}{\sqrt{f}} = -2\log\left(\frac{\varepsilon}{3.7D_h} + \frac{2.51}{Re\sqrt{f}}\right).$$

Moody

$$f = 0.0055\left[1 + \left(2\times10^4\cdot\frac{\varepsilon}{D} + \frac{10^6}{Re}\right)^{1/3}\right]$$

$$\varepsilon/D = 0 - 0.01$$

parallel



Series