

MT processes are characterised by transfer of components from one to another.

Molecular diffusion is important.

Interface always at equilibrium.

$L$  so thin that there is no mass transfer resistance.

In all practical MT processes, atleast two phases should be there.

Crystallisation is used to separate p-xylene from a mixture of xylenes.

#### • Fractional Crystallisation

p-xylene used for making synthetic fibres.

$G - L$ ;  $L - L$ ;  $G - G$ ;  $G - S$ ;  $S - L$ ;  $G - L - S$

Gas - Liquid : Absorption / Desorption, Distillation, Humidification / Dehumidification

Liquid - Liquid : Extraction, Membrane separation

Gas - Gas : Membrane separation

Gas - Solid : Adsorption

Solid - Liquid : Adsorption, Leaching, Crystallisation

Gas - Liquid - Solid : Drying

In drying and humidification, liquid phase should be pure so no resistance from the other component.

## Equipment for Liquid + Gas Mass Transfer

### Operations :

For effective MT, we need to disperse one phase into another.

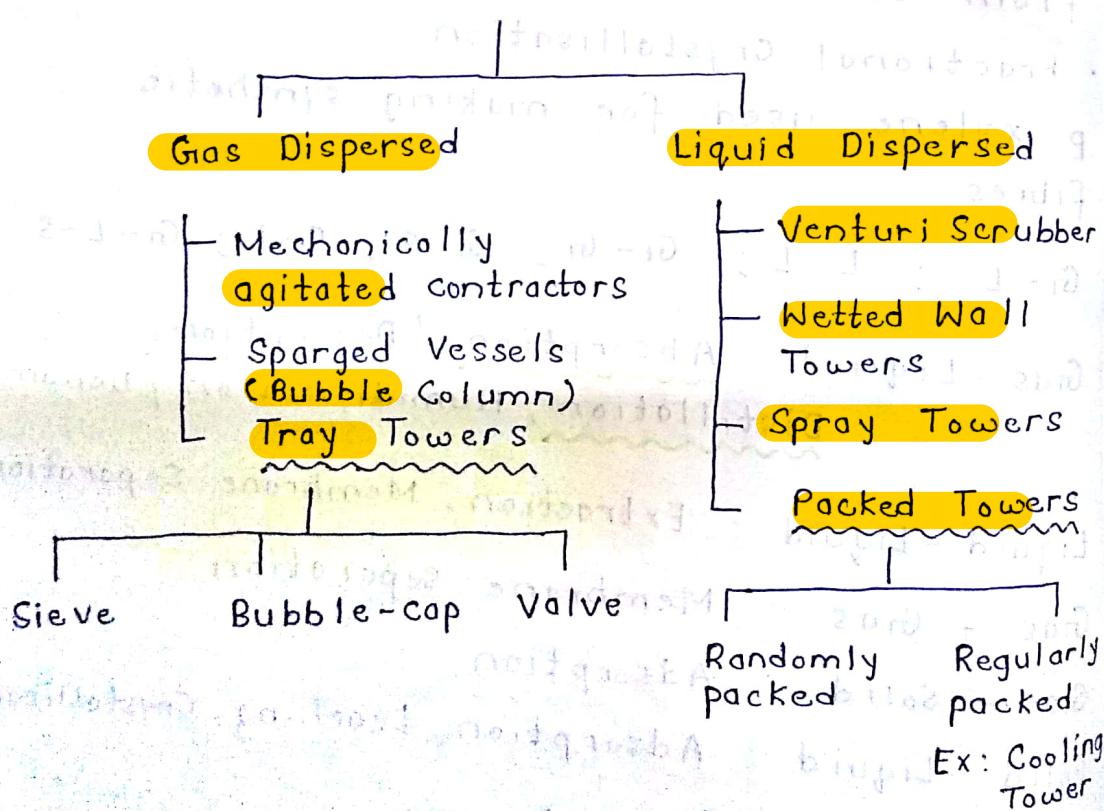
Intimate contact b/w the two phases is required.

Large interfacial area  $\Rightarrow$  Large Mass Transfer.

Disperse gas in the form of bubbles into liquid.

Disperse liquid in the form of droplets into gas.

### Gas - Liquid Contactors



Agitated Vessels are used whenever multi-stage counter current effects are not required. This is the case when a chemical rxn b/w the dissolved gas and a constituent

of the liquid

Ex : Carbonation of vegetable

In most of the liquids.

tendency to solids can be easily in ag

Cascade Bubble

To get better perforated tr

### Tray Towers

Flow is co-current.

Multi stage

Each tray stage

Conc. of sol as they move tower.

### Stage : A de

intimate abtion b/w the phases

Ex: Cooling Tower

Feed

of the liquid is required.

**Ex:** Carbonation of lime slurry; hydrogenation of vegetable oils.

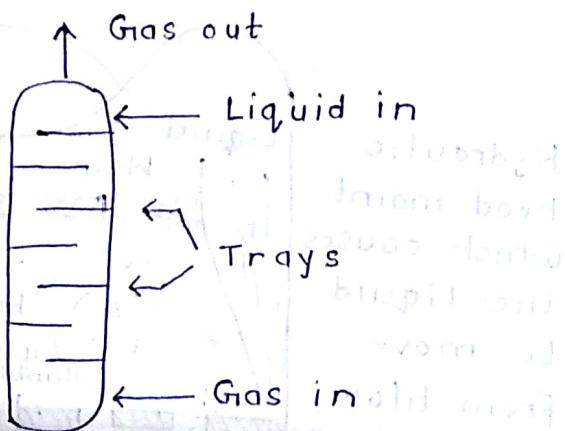
In most of them solids are suspended in the liquids. Counter current towers have a tendency to clog with such solids and solids can be suspended in the liquids easily in agitated vessels.

### Cascade Bubble Column-

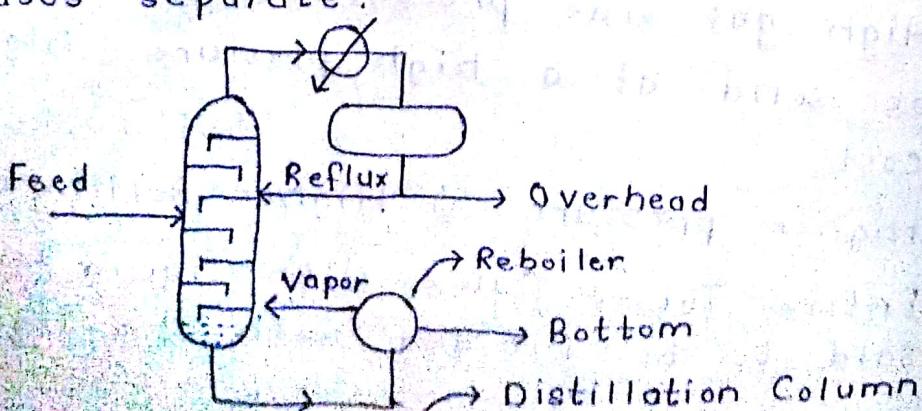
To get better dispersion of the gas perforated trays are placed.

### Tray Towers:-

- Flow is counter-current.
- Multi stage.
- Each tray is a stage.
- Conc<sup>n</sup> of phases change ↓ Liquid out as they move up the tower.



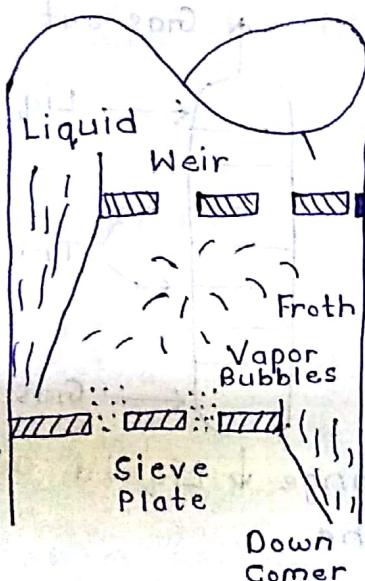
**Stage:** A device where two phases contact intimately, MT takes place and then the phases separate.



## Selection of Trays

	Bubble Cap	Valves	Sieve
Relative Cost	2.0	1.2	1.0
$\Delta P$	Highest	Intermediate	Lowest
$\eta$	Highest	Highest	Lowest
Vapor Capacity	Lowest	Highest	Highest
Turndown ratio	4	2	2

Hydraulic head maintained which causes the liquid to move from low P to high P.



- Liquid pool to be high
- Gas to be passed at **high velocity**, so that it may disperse in the form of small bubbles.

**Liquid Entrainment:** Film of liquid on the bubble. Less mass transfer.

**High gas side pressure drop.** Gas needs to be sent at a high pressure; higher cost.

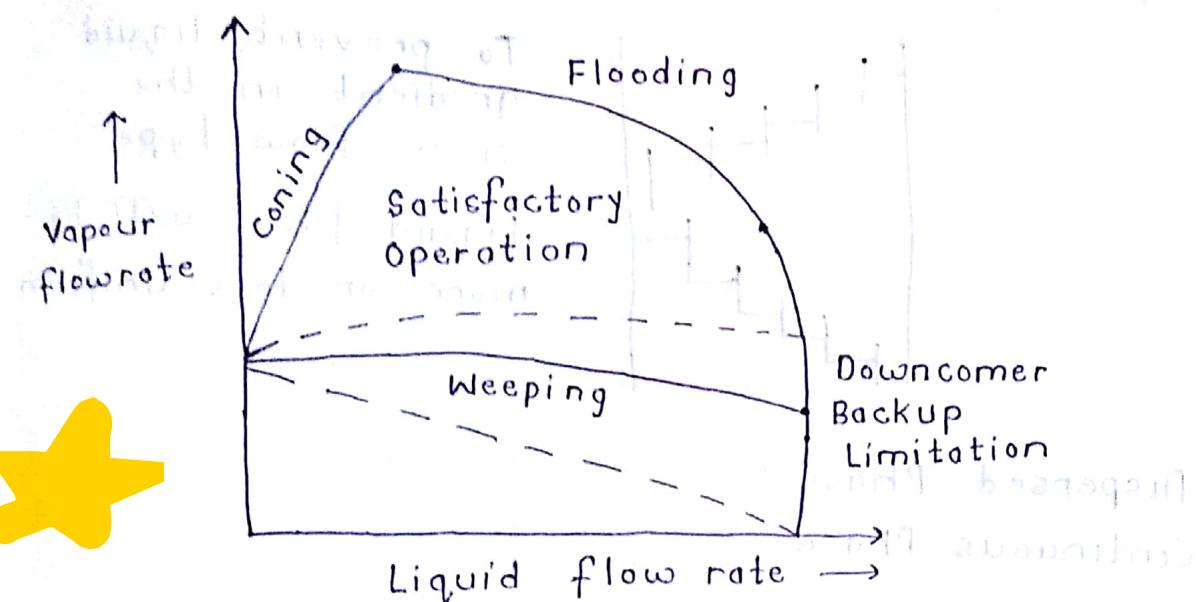
Higher pressure  $\rightarrow$  Higher Boiling point.

Entire Tower filled with liquid  $\rightarrow$  Tower is said to be **flooded** with the liquid.

**Priming**: too much liquid entrainment.

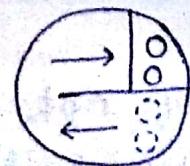
**Dumping**: liquid flows through the holes at very low gas rates. Extreme case of weeping in which liquid will not reach the downcomer.

**Coning**: At high gas rates but low liquid rates.



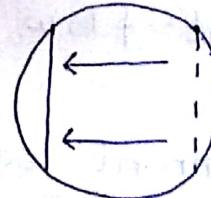
**Type of Trays** depending on liquid distribution of tray

a) Reverse flow

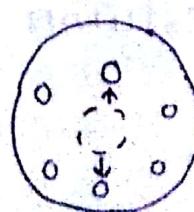
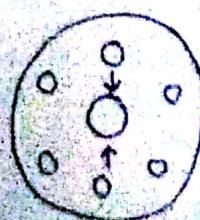


Small dia. towers

b) Cross flow



c) Radial flow



Large Diameter Towers

d) split flow

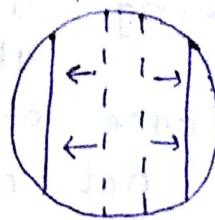
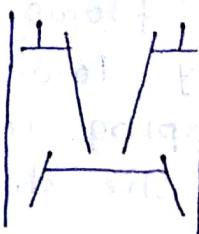
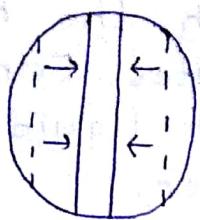
arrest

process

process

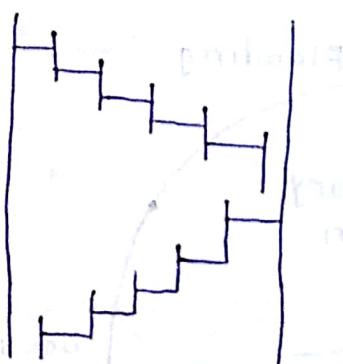
process

process



b) upflow and downflow separation stage of A primary

e) Cascade Trays



To prevent liquid gradient in the cross-flow type.

Liquid flow will be more or less uniform.

Dispersed Phase :

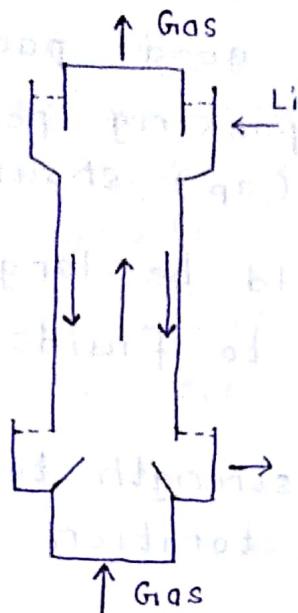
Continuous Phase :

Liquid Dispersed :

Venturi Scrubber :-

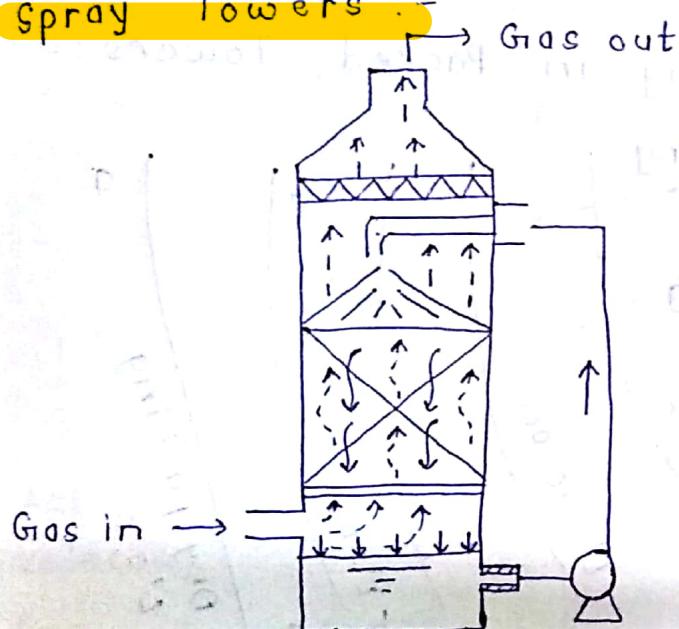
- Gas - Liquid Separation
- Effective in removing acidic gases like  $\text{SO}_x$ .
- Co-current flow as there is chemical reaction.
- In co-current cascade, we will not get the benefit of several equilibrium stages; or ideal stages.

Wetted Wall Column :



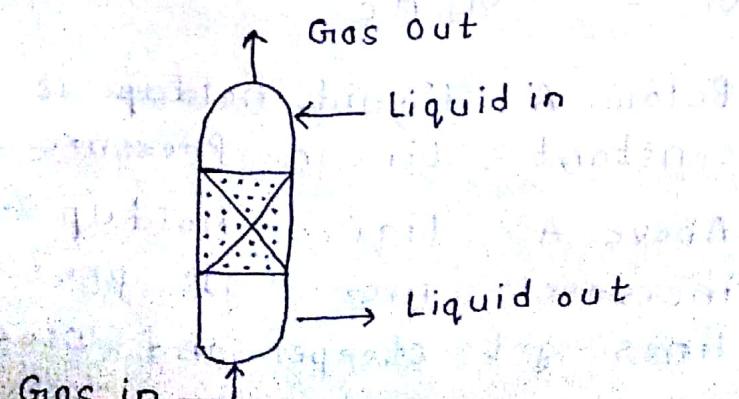
- Interfacial area can be more or less accurately calculated.
- Formation of hydrochloric acid by HCl gas into  $H_2O + HCl$

### Spray Towers:



Horizontal Spray Tower is Spray Chamber. Cocurrent contact is followed in spray chamber. No loss in efficiency. Used in air conditioners.

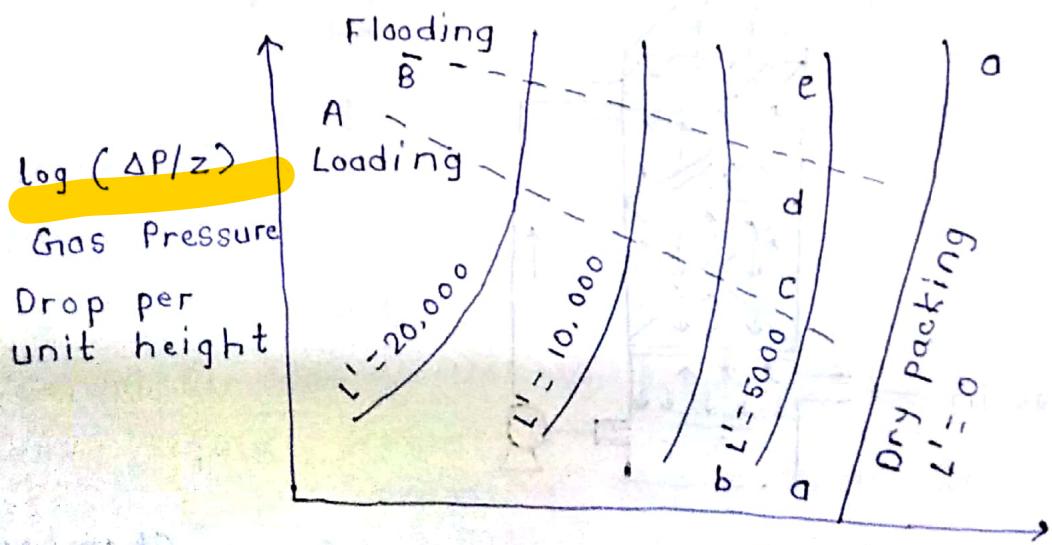
### Packed Towers:



Characteristics of good packing material.

- The surface of packing per unit volume of packed space ( $\Delta P$ ) should be large.
- Void space should be large.
- Chemically inert to fluids being processed.
- Have structural strength to permit easy handling and restoration.
- Low Cost.
- Large interfacial area b/w liquid and gas.

Loading and Flooding in Packed Towers:-



Log  $(\Delta P/z)$ , Gas Pressure Drop per unit height

Log  $(G_1')$ , superficial gas mass velocity

$$L' = L/A_c$$

Gas rate high  $\rightarrow$  Gas pressure drop high

$$G_1' = G_1/A_c$$

Below A, liquid holdup is more or less constant. Linear Pressure Variation.

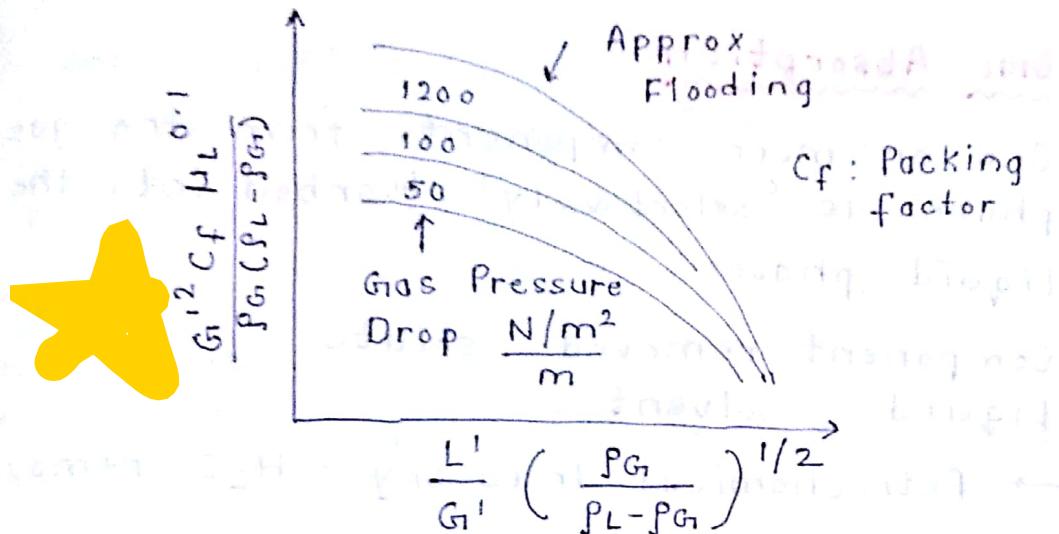
Above A, liquid holdup in the tower increase, more gas pressure drop. The lines get steeper and steeper

Above B., tower is entirely filled with liquid. Liquid is the continuous phase and gas comes out in the form of bubbles. This condition is called **flooding**.

**phase inversion** : liquid changes from dispersed to continuous phase.

y

gas.



$$\frac{L'}{G'} \left( \frac{P_G}{P_L - P_G} \right)^{1/2}$$

$$\frac{L'}{G'} = \frac{L}{G}$$

Actual velocity is 75-80% of the flooding velocity.

### Tray Towers Vs. Packed Towers

1. **Gas Pressure Drop** : Packed Towers require smaller  $\Delta P$ .
2. **Liquid Holdup** : P.T. provide substantially smaller liquid hold up.
3. **Liquid / Gas ratio** : Low  $L/G$  : T.T.s  
High  $L/G$  : P.T.s
4. **Liquid Cooling** : Tray towers are suitable.
5. **Side streams** : More readily removed from T.T.s.

6. Foaming Systems : P.T.s are more suitable.
7. Corrosion : P.T.s are more suitable
8. Cleaning : Frequent cleaning is easier with tray towers.

11/10/2018

### Gas Absorption

One or more components from the gas phase is 'selectively' absorbed into the liquid phase.

Component removed : solute

Liquid : solvent

→ Petrochemical Industry : H<sub>2</sub>S removed

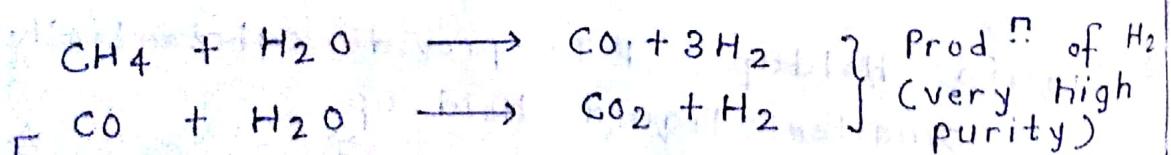
Desorption / Stripping :-

MT from liquid to gas phase.

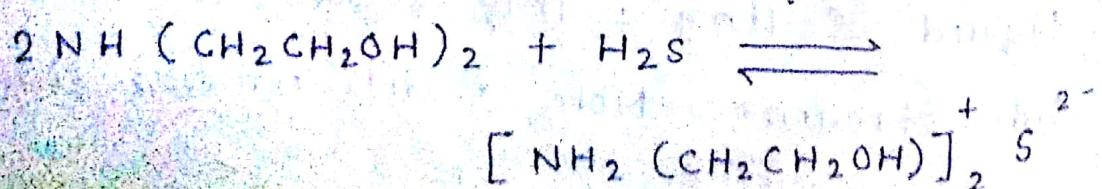
Aq. alkanol amine is used for absorption of H<sub>2</sub>S.

DEA → NH(CH<sub>2</sub>CH<sub>2</sub>OH)<sub>2</sub> (25-30%)

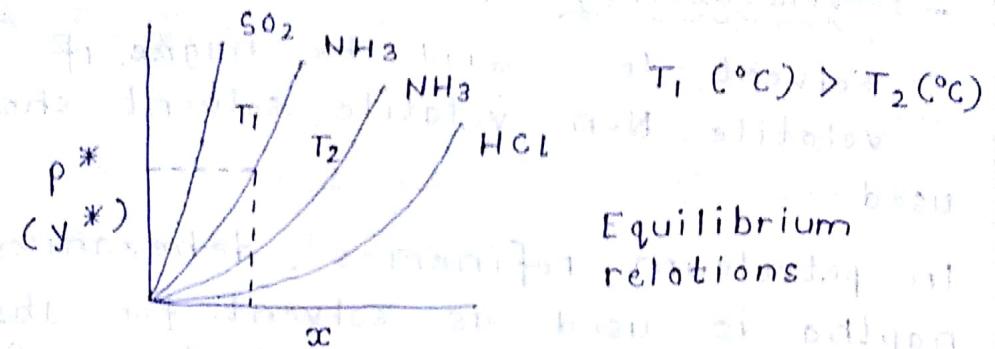
Steam Reforming of CH<sub>4</sub>/C<sub>2</sub>H<sub>6</sub>



↳ CO<sub>2</sub> formed is removed by absorption into DEA.



For different solutes



Equilibrium relationship

For ideal system:  $\rightarrow$  Raoult's Law

The curves are linear:  $p^* = P \cdot x$

In reality, only a small portion of the curve can be approximated as a line

Henry's Law:

$$y^* = \frac{P^*}{P_t} = mx \quad \{ \text{For real solutions} \}$$

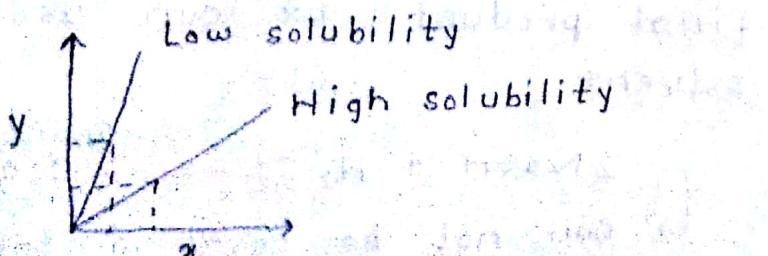
At low concentrations.

### Choice / Selection of Solvent:

- Very high absorption rates required.
- Should selectively remove 1 or 2 component.

Desirable Properties

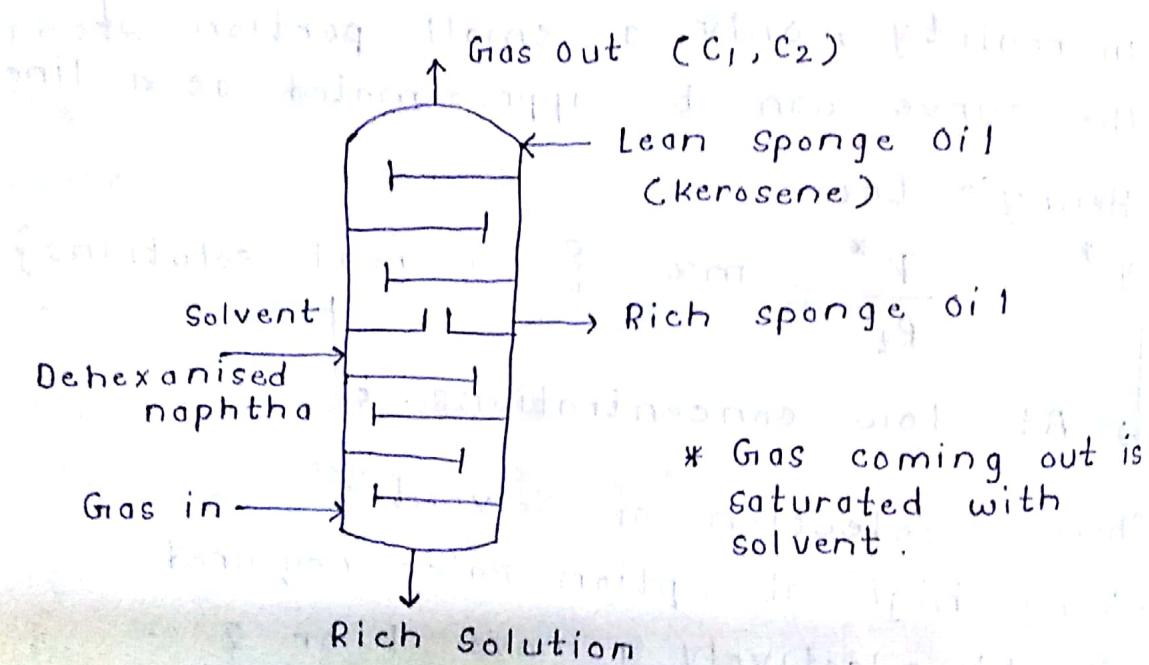
- $\rightarrow$  Solubility of solute in that solvent should be high
- If molecular weight is less, solubility will be higher; required in less quantity.



## → Volatility of Solvent :

Solvent loss will be huge if it is volatile. Non-volatile solvent should be used.

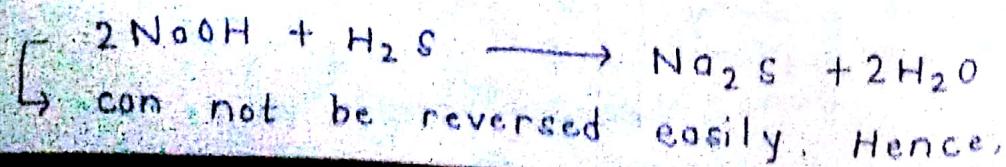
In petroleum refineries, dehexanized naphtha is used as solvent for absorbing lighter hydrocarbons. But it has sufficient volatility. To overcome this, sponge absorbers (like kerosene) are used.



• If solute and solvent react, solubility will be even higher, but the reaction should be reversible.

• To get hydrochloric acid, we have to absorb HCl gas in water.

→ However, if the solution is not the final product, we can choose the solvent.



$\text{NaOH}$  cannot be used as solvents to absorb  $\text{H}_2\text{S}$ .

### Corrosiveness:

Solvent should not be corrosive, else the fixed cost will go up.

### Viscosity:

Solvent should be less viscous.

- . Higher MT rates.
- . Pumping / operating cost less.

### Freezing Point:

Solvent should have a lower freezing point as closer to freezing point, the viscosity is very high.

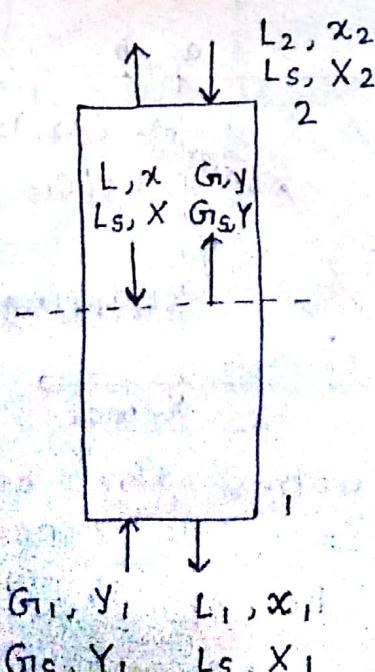
### Cost:

Material should be of low cost and should be available readily.

### Others:

Should not be toxic, non-flammable.

Should be chemically stable.



$G_1$ : molar flow rate of total gas

$G_s$ : solute free flow rate

$$Y = \frac{y}{1-y}$$

$$G_1 = G_s (1+Y)$$

$$G_s = G_1 (1-y)$$

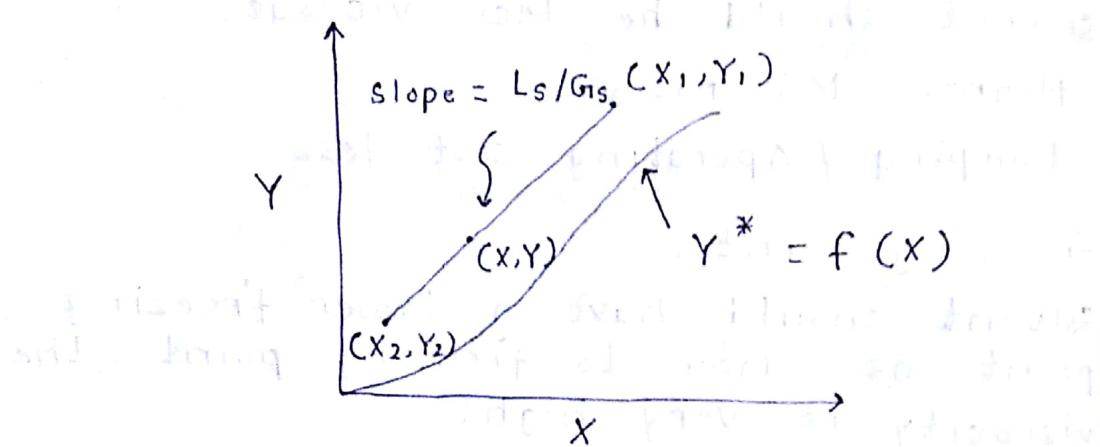
At steady state, component balance gives,

$$G_s Y_1 + L_s X = G_s Y + L_s X_1$$

$$\Rightarrow G_s (Y_1 - Y) = L_s (X_1 - X)$$

for the whole tower or infinite height

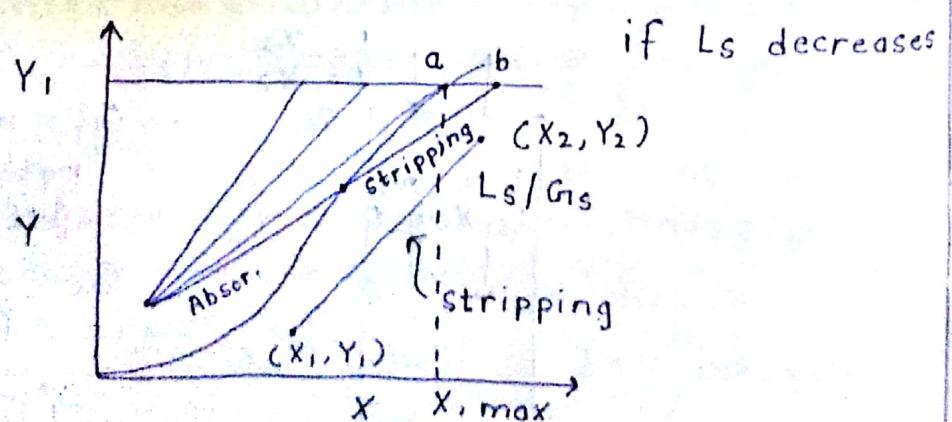
$$G_s (Y_1 - Y_2) = L_s (X_1 - X_2)$$



Any point on the operating line, the coordinates will represent the composition of the streams at any section.

To find the flowrate of liquid :-

$X_2$ ,  $Y_2$ ,  $Y_1$  and  $G_s$  are known to us.



If the OP touches the equilibrium line, driving force is zero.

$\Rightarrow$  Tower should be of infinite height

for condition b, both absorption and stripping will take place, which we don't want.  
Intersection of O.L and E.L : pinching.

'a' corresponds to minimum solvent flow rate.

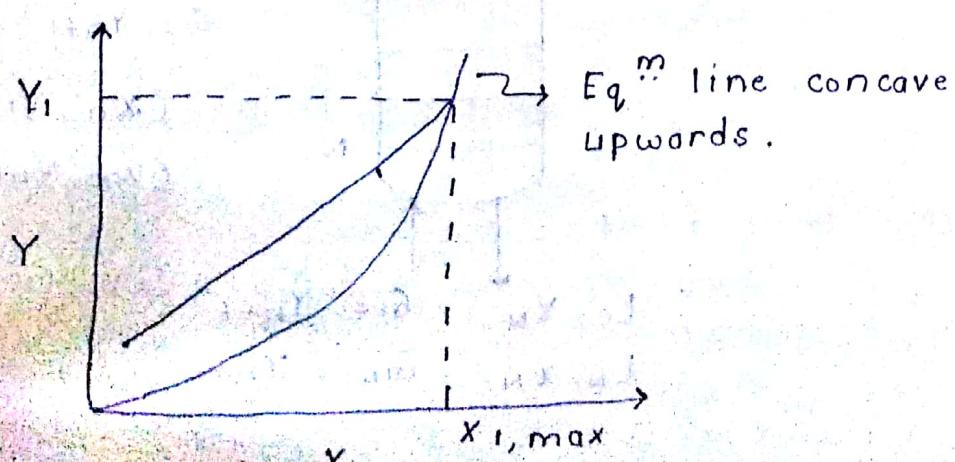
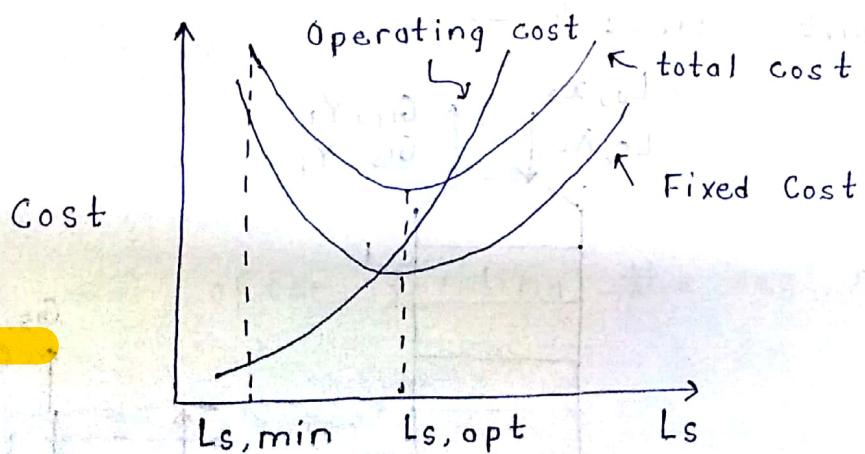
Desired value of  $L_s$  :-

$$L_s = (1.2 \text{ to } 2.0) \times L_{s,\min}$$

Minimum Liquid flow rate, solution is concentrated. For dilute solutions, the separation is difficult.

Higher liquid flow rate  $\Rightarrow$  size of tower decreases  $\Rightarrow$  less capital cost.

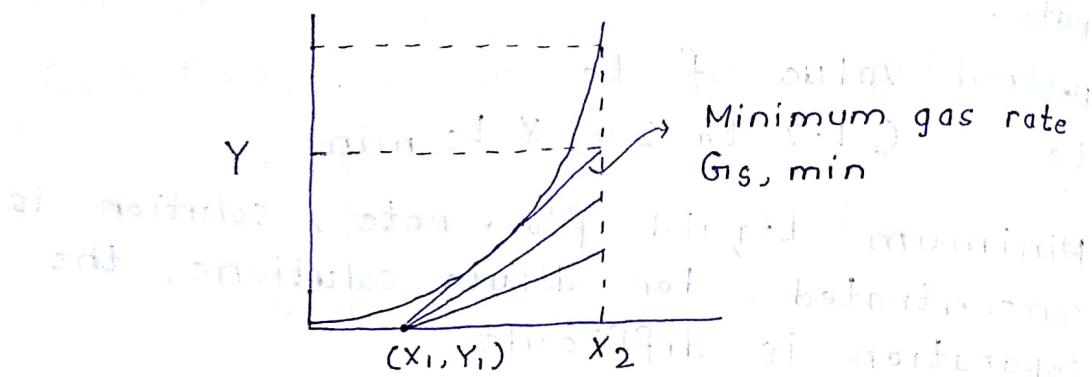
But diameter has to be increased to avoid flooding of the tower.



$$G_{ls}(Y_1 - Y_2) = L_{s,\min}(X_{1,\max} - X_2)$$

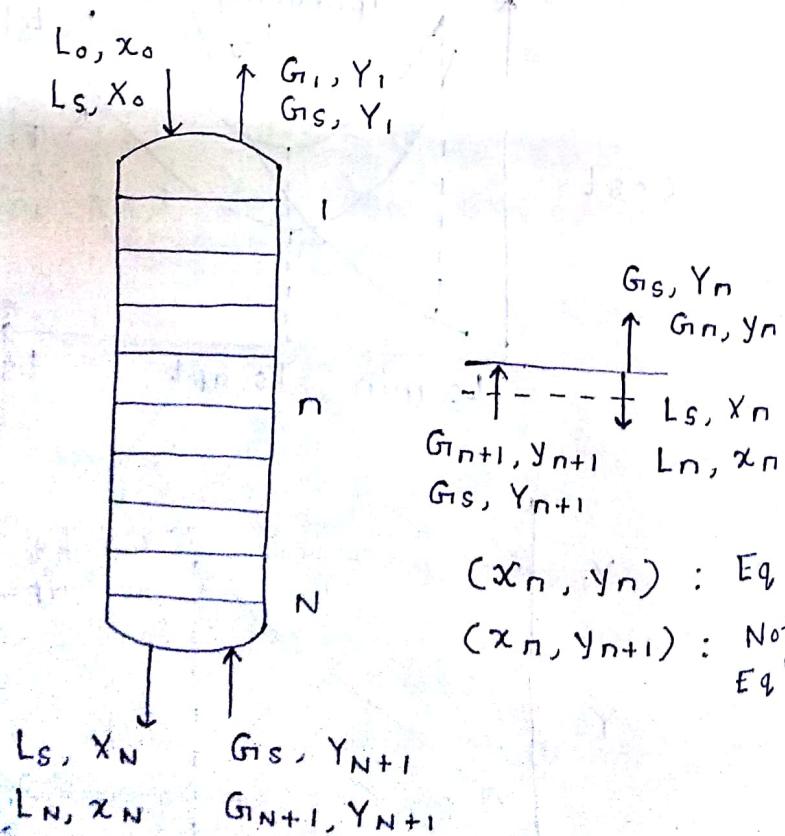
For stripping :  $L_s, X_2, X_1 \leq Y_1$  ( $\approx 0$  as mostly pure gas)

are known minimum of stripping



We don't get benefit of a multistage in case of cocurrent contactor, whatever be the size of the tower.

### Number of Ideal Stages for counter-current contact

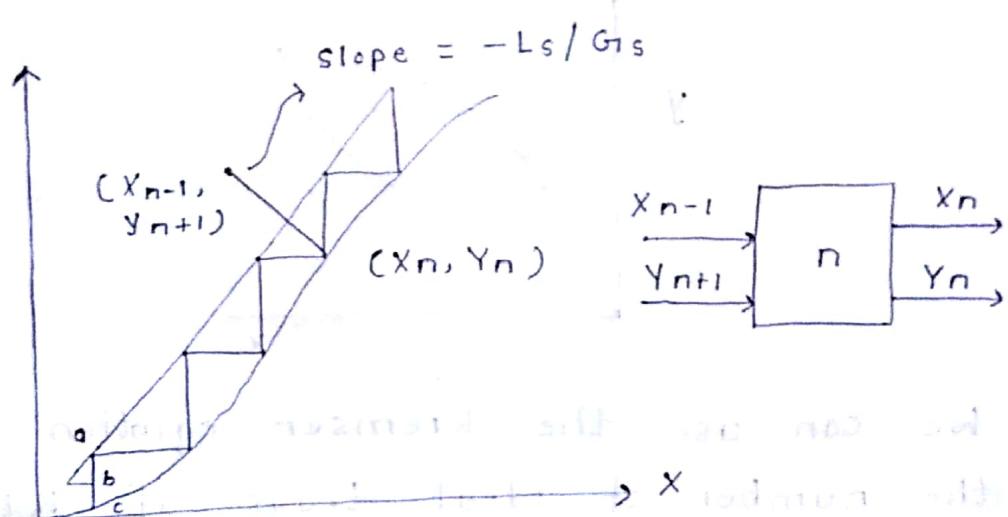
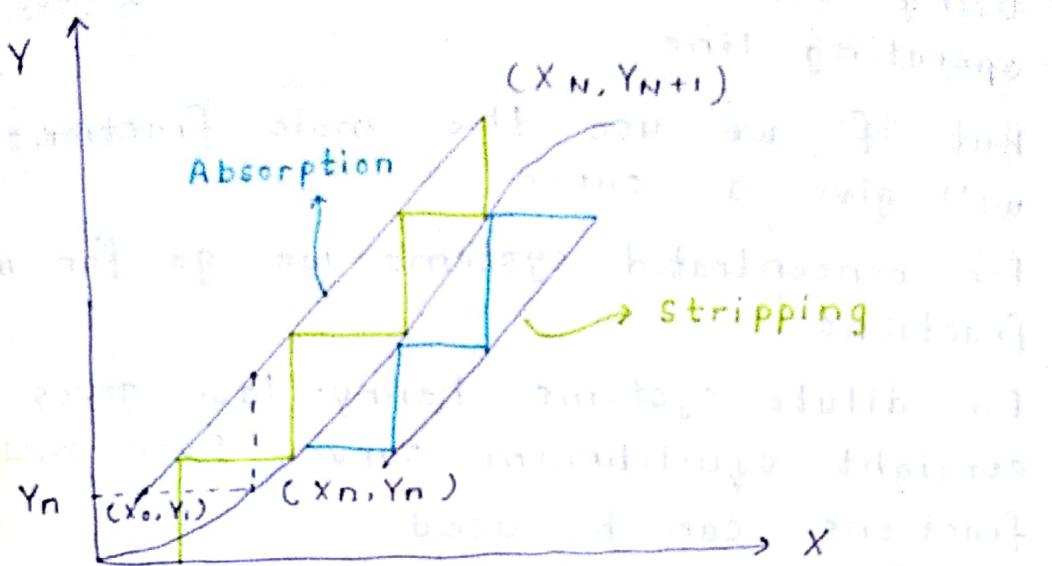


$$(x_n, y_n) : Eq^m$$

$$(x_n, y_{n+1}) : Not \in Eq^m$$

Fraction  
Round off  
of trays.  
 $N_{real}$

For stripping

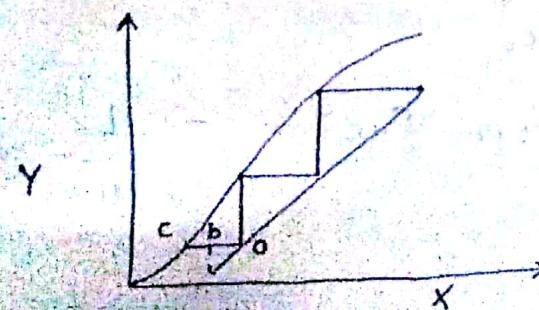


$$\text{Fraction of tray} = \frac{ab}{ac}$$

Round off after finding the real number of trays.

$$N_{\text{real}} = \frac{N_{\text{ideal}}}{\eta_{\text{tray}}}$$

For stripping:



$$\text{Fraction of tray} = \frac{ab}{ac}$$

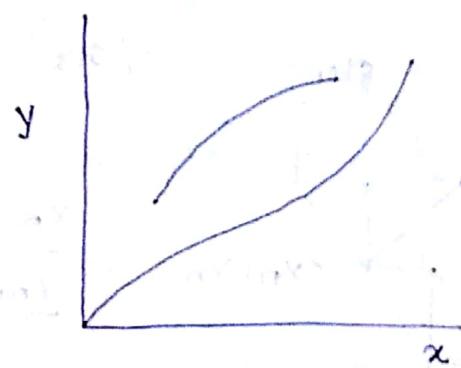
Eq m  
Not in  
Eq m

Using the solute-free basis, gives straight operating line.

But if we use the mole fractions, it will give a curve.

For concentrated systems we go for mole fractions.

For dilute systems, Henry law gives the straight equilibrium curve. Even mole fractions can be used.



We can use the Kremser equation to find the number of ideal trays, if both the operating line and eq<sup>m</sup> curve are straight lines.

$$N = \log \left[ \frac{y_{N+1} - mx_0}{y_1 - mx_0} \left( 1 - \frac{1}{A} \right) + \frac{1}{A} \right]$$

for absorption  $\log A$  for  $A \neq 1$

$$N = \frac{y_{N+1} - y_1}{y_1 - mx_0} \quad \text{for } A = 1$$

$$A = \text{Absorption factor} = \frac{L_s}{mG_s} \approx \frac{L}{mG}$$

straight for stripping, we can start from the top

it  $N = \log \left[ \frac{\frac{x_0 - y_{N+1}/m}{x_N - y_{N+1}/m} \left( 1 - \frac{1}{s} \right) + \frac{1}{s}}{\log s} \right]$

else  $\log s$  for  $s \neq 1$

the  $N = \frac{x_0 - x_N}{x_N - y_{N+1}/m}$  for  $s = 1$

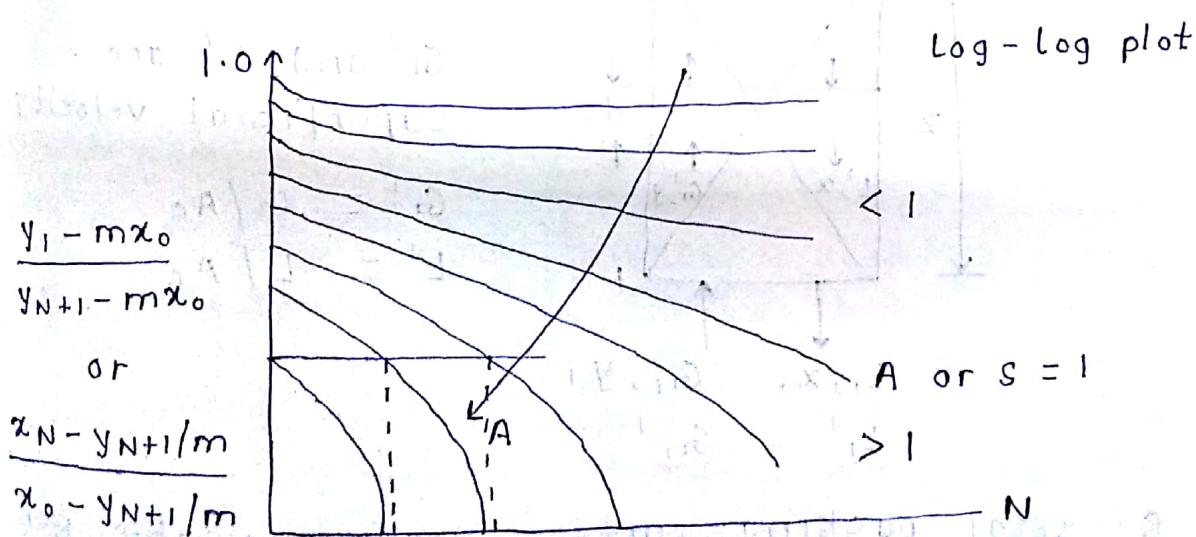
$s$  = Stripping factor  $= 1/A = mG/L$

4/10/2018

Optimum Absorption factor is to be applied

$$A = \frac{L}{mG} \rightarrow \text{fixed}$$

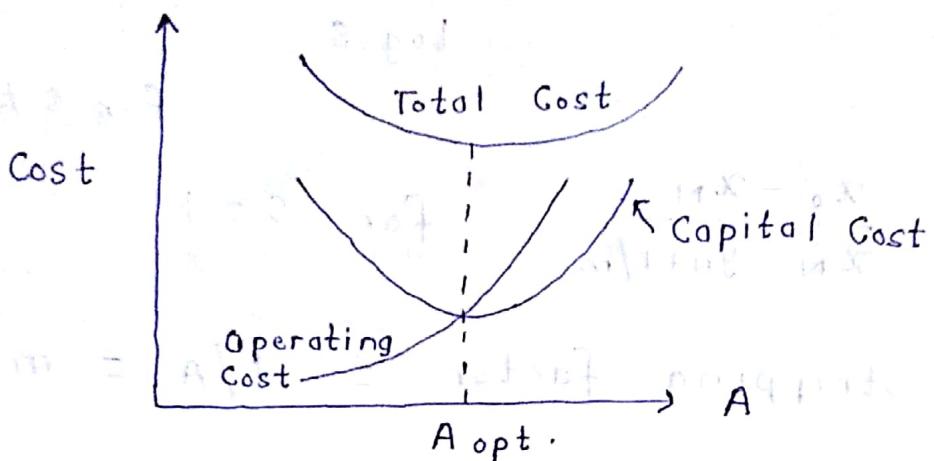
$\hookrightarrow$  fixed for system



if  $A$  (or  $s$ )  $< 1$ , degree of removal is limited and vice-versa.

For fixed degree of removal, higher the value of  $A$ , lesser will be the number of trays required.

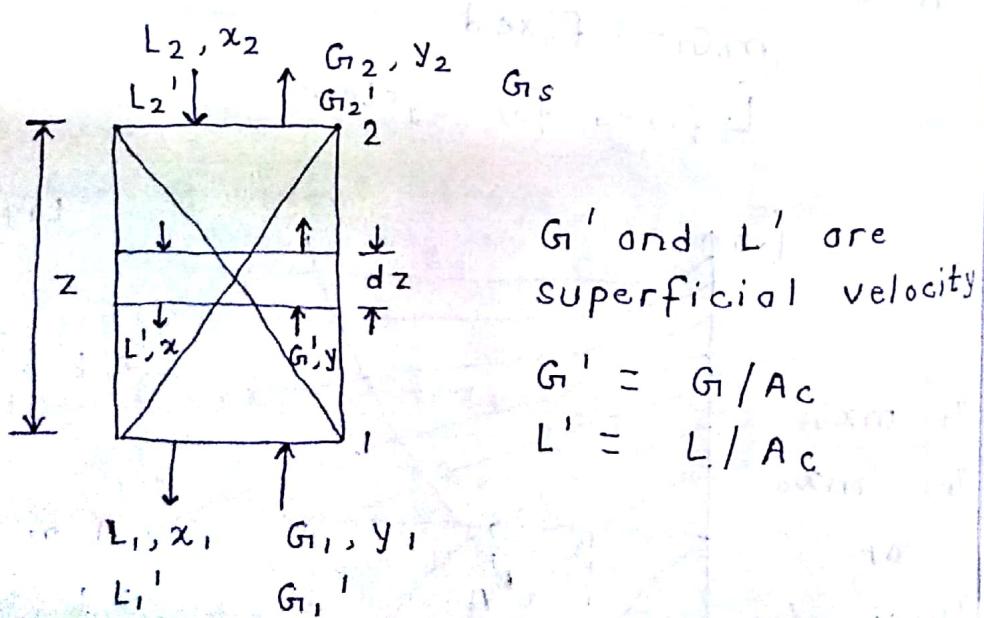
But the soln gets more and more diluted which leads to an increase in the operating cost.



$$A_{opt} = 1.25 \text{ to } 2.00$$

## Design of Continuous Contact Equipment:

### - Packed Towers



$S$ : total packing surface area available per unit cross section

$$S = \sigma X z ; \sigma = \text{specific interfacial area}$$

$$ds = \sigma dz = \text{surface area/volume}$$

$$-d(G'y) = N_A \alpha dz$$

$-dG'$ : decrease in total gas flow rate  
↓ due to transfer of A only

$$\Rightarrow -dG' \cdot y - dy \cdot G' = N_A \alpha dz$$

$$-G' dy + y N_A \alpha dz = N_A \alpha dz$$

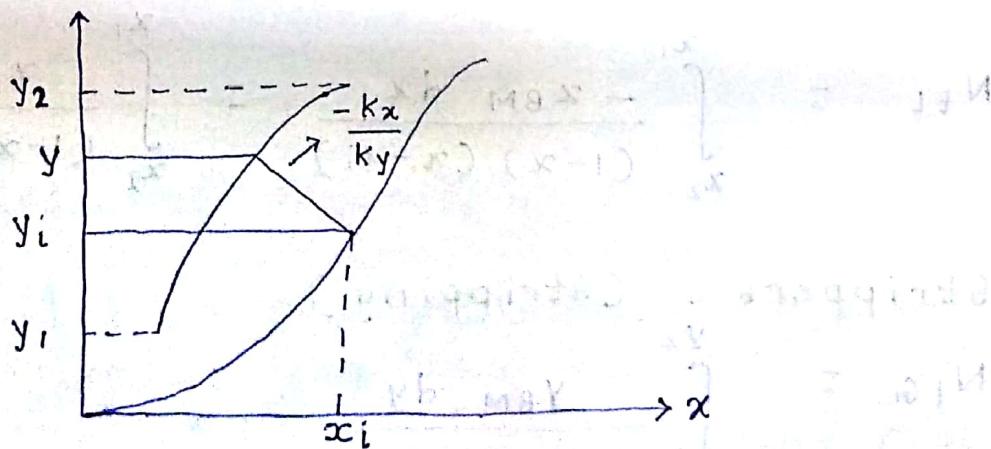
$$-G' dy = (1-y) N_A \alpha dz$$

$$N_A = k_y (y - y_i)$$

$$-G' dy = (1-y) k_y (y - y_i) dz$$

$$\Rightarrow \int_0^z dz = \int_{y_1}^{y_2} \frac{-G' dy}{k_y \alpha (1-y)(y-y_i)}$$

$$\Rightarrow z = - \int_{y_1}^{y_2} \frac{G' dy}{k_y \alpha (1-y)(y-y_i)}$$



$$y_{BM} = \frac{(1-y_i) - (1-y)}{\ln \left( \frac{1-y_i}{1-y} \right)}$$

$\frac{G'}{k_y \alpha y_{BM}}$  remains constant,

$$\frac{G'}{k_y \alpha y_{BM}}$$

$$\Rightarrow Z = \frac{G_1'}{ky_a Y_{BM}} \int_{y_2}^{y_1} \frac{Y_{BM} dy}{(1-y)(y-y_i)}$$

$$= H_{tG} \times N_{tG}$$

$H_{tG}$  : height of gas-phase transfer unit

$N_{tG}$  : no. of gas-phase transfer unit

$H_{tG}$  indicates inversely the relative ease with which a tower can accomplish the MT job.

$N_{tG}$  indicates the difficulty involved in the desired degree of separation.

In terms of liquid phase parameters,

$$Z = H_{tL} \times N_{tL}$$

$$H_{tL} = \frac{L'}{K_x a X_{BM}}$$

$$N_{tL} = \int_{x_2}^{x_1} \frac{-X_{BM} dx}{(1-x)(x-x_i)} = \int_{x_2}^{x_1} \frac{X_{BM} dx}{(1-x)(x_i-x)}$$

Strippers : (stripping)

$$N_{tG} = \int_{y_1}^{y_2} \frac{Y_{BM} dy}{(1-y)(y_i-y)}$$

$$N_{tL} = \int_{x_1}^{x_2} \frac{X_{BM} dx}{(1-x)(x-x_i)}$$

For Abs  
For stri

$ky_a =$

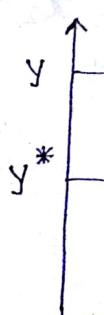
For non  
 $G_1' =$

$y_{BM} =$

$Z = H$

$H_{tOG} =$

$y_{BM}^* =$



For moderate  
(5 - 11 %) :

$y_{BM}^* \approx$

For Absorption : gas phase transfer unit  
For stripping : liquid phase transfer unit

$k_y \cdot a$  = volumetric mass transfer coefficient

For non-dilute solutions :

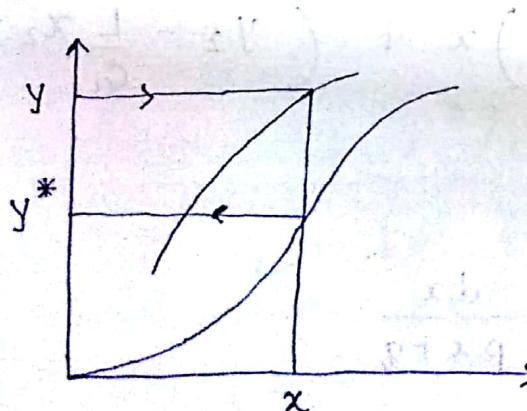
$$G_1' = \frac{G_{1,1}' + G_{1,2}'}{2}$$

$$y_{BM} = \frac{(y_{BM})_1 + (y_{BM})_2}{2}$$

$$Z = H_{tOG} \times N_{tOG}$$

$$H_{tOG} = \frac{G_1'}{k_y \cdot a \cdot y_{BM}^*} \quad N_{tOG} = \int_{y_2}^{y_1} \frac{y_{BM}^* dy}{(1-y)(y-y^*)}$$

$$y_{BM}^* = \frac{(1-y^*) - (1-y)}{\ln \left( \frac{1-y^*}{1-y} \right)}$$



For moderately concentrated solutions (gases)  
(5-11%) :

$$y_{BM}^* \approx \frac{(1-y^*) + (1-y)}{2} = \frac{2(1-y) + (y-y^*)}{2}$$

$$N_{\text{tot}G_i} = \int_{y_2}^{y_1} \frac{dy}{y - y^*} + \frac{1}{2} \int_{y_2}^{y_1} \frac{dy}{1-y} \\ = \frac{1}{2} \ln \left( \frac{1-y_2}{1-y_1} \right) + \int_{y_2}^{y_1} \frac{dy}{y - y^*}$$

For dilute gases :-

$$1-y \approx 1$$

$$y_{\text{eqm}}^* \approx 1$$

$$N_{\text{tot}G_i} = \int_{y_2}^{y_1} \frac{dy}{y - y^*}$$



$$y^* = mx + r$$

↳ Eq<sup>m</sup> line

$$y - y_2 = \frac{L}{G} (x - x_2)$$

$$y = \frac{L}{G} (x - x_2) + y_2$$

$$y - y^* = \frac{L}{G} (x - x_2) + y_2 - mx - r$$

$$= \left( \frac{L}{G} - m \right) x + \left( y_2 - \frac{L}{G} x_2 - r \right)$$

$$= px + q$$

$$\therefore N_{\text{tot}G_i} = \frac{L}{G} \int_{x_2}^{x_1} \frac{dx}{px + q}$$

$$= \frac{L}{Gp} \ln \left( \frac{px_1 + q}{px_2 + q} \right)$$

$$\frac{L}{G} = \frac{y_1 - y_2}{x_1 - x_2}$$

$$p = \frac{L}{G} - m$$

$$= (y_1 - y_2)$$

$$\frac{L}{Gp} = \frac{y_1}{(y_1 - y_2)}$$

$$\Rightarrow N_{\text{tot}G_i} =$$

$$= \frac{L}{(y_1 - y_2)}$$

$$= \frac{R}{C}$$

where

$$(y - y^*)_M$$

y

$$= \frac{(y_1 - y_2) - m(x_1 - x_2)}{x_1 - x_2}$$

$$\frac{L}{G_P} = \frac{y_1 - y_2}{(y_1 - mx_1 - r) - (y_2 - mx_2 - r)}$$

$$\Rightarrow N_{t \rightarrow G} = \frac{y_1 - y_2}{(y_1 - y_1^*) - (y_2 - y_2^*)} \ln \frac{y_1 - y_1^*}{y_2 - y_2^*}$$

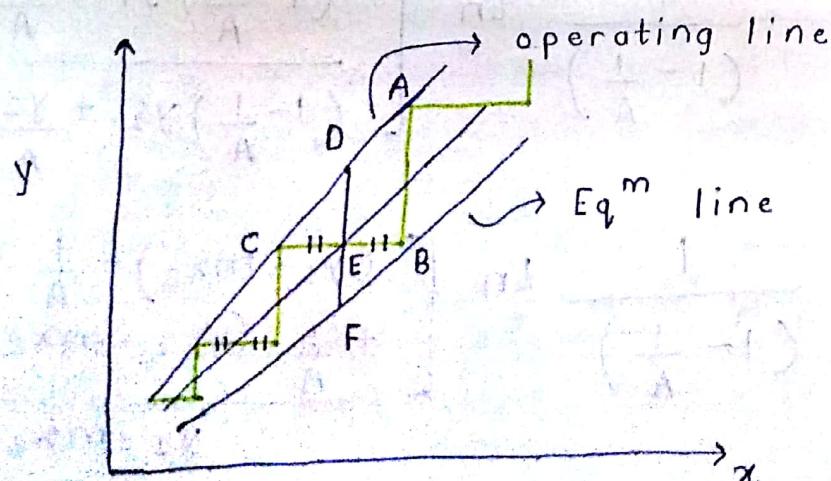
$$= \frac{y_1 - y_2}{(y - y^*)_1 - (y - y^*)_2}$$

$$\ln \frac{(y - y^*)_1}{(y - y^*)_2}$$

$$= \frac{y_1 - y_2}{(y - y^*)_M}$$

where

$$(y - y^*)_M = \frac{(y - y^*)_1 - (y - y^*)_2}{\ln \frac{(y - y^*)_1}{(y - y^*)_2}}$$



$$\frac{AB}{BC} = \frac{DE}{CE}$$

$$AB = \frac{BC}{CE} \cdot DE$$

$$= 2DE = DF$$

If Henry's Law is followed:-

$$y - y_2 = \frac{L}{G_1} (x - x_2)$$

$$= \frac{L}{G_1} \left( \frac{y^*}{m} - x_2 \right)$$

$$= \frac{L}{m G_1} (y^* - mx_2)$$

$$y^* = \frac{1}{A} (y - y_2) + mx_2$$

$$y - y^* = \left( 1 - \frac{1}{A} \right) y + \frac{y_2}{A} - mx_2$$

$$N_{tOG_1} = \int_{y_2}^{y_1} \frac{dy}{y - y^*}$$

$$= \int_{y_2}^{y_1} \frac{dy}{\left( 1 - \frac{1}{A} \right) y + \frac{y_2}{A} - mx_2}$$

$$? = \frac{1}{\left( 1 - \frac{1}{A} \right)} \ln \left[ \frac{\left( 1 - \frac{1}{A} \right) y_1 + \frac{y_2}{A} - mx_2}{\left( 1 - \frac{1}{A} \right) y_2 + \frac{y_2}{A} - mx_2} \right]$$

$$= \frac{1}{\left( 1 - \frac{1}{A} \right)} \ln \left[ \frac{(y_1 - mx_2) - \frac{1}{A} (y - mx_2)}{y_2 - mx_2} + \frac{\frac{1}{A} \cdot (y_2 - mx_2)}{y_2 - mx_2} \right]$$

$$N_{tOG_1} =$$

For strip

$$N_{tOL} =$$

$$\frac{y_2 - mx_2}{y_1 - mx_2}$$

or

$$\frac{x_1 - y_1 / m}{x_2 - y_1 / m}$$

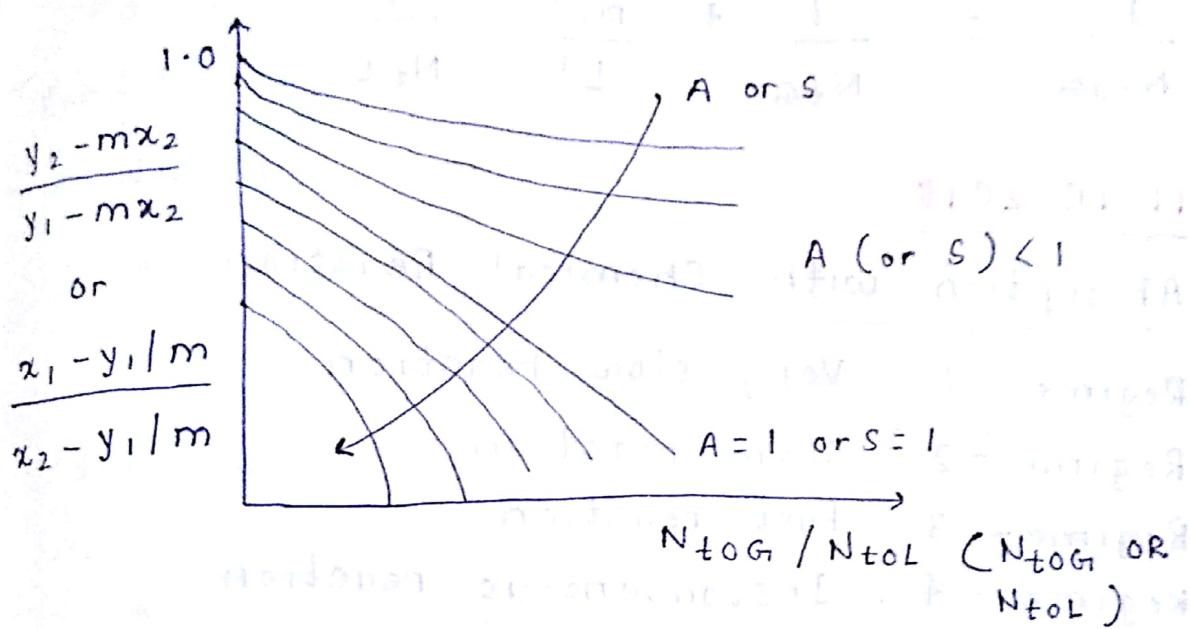
$$\frac{1}{Ky_A}$$

$$H_{tOG_1}$$

$$N_{tOG} = \frac{1}{1 - \frac{1}{A}} \ln \left[ \frac{y_1 - mx_2}{y_2 - mx_2} \left( 1 - \frac{1}{A} \right) + \frac{1}{A} \right]$$

For stripping -

$$N_{tOL} = \frac{1}{1-A} \ln \left[ \frac{x_2 - y_1/m}{x_1 - y_1/m} (1-A) + A \right]$$



$$\frac{1}{K_y a} = \frac{1}{K_y a} + \frac{m}{K_x a}$$

$$H_{tOG} = \frac{G_1^1}{K_y a y_{BM}^*} = \frac{G_1^1}{y_{BM}^*} \left( \frac{1}{K_y a} + \frac{m}{K_x a} \right)$$

$$\left[ \frac{x_2}{x_2} \right] = \frac{G_1^1}{K_y a y_{BM}} \cdot \frac{y_{BM}}{y_{BM}^*} + \frac{m G_1^1}{L^1} \cdot \frac{L^1}{K_x a x_{BM}}$$

$$\left[ \frac{x_{BM}}{y_{BM}^*} \right] = H_{tG} \times \frac{y_{BM}}{y_{BM}^*} + \frac{m G_1^1}{L^1} H_{tL} \frac{x_{BM}}{y_{BM}^*}$$

$$\frac{1}{N_{tOG}} = \frac{1}{N_{tG_1}} \frac{y_{BM}}{y_{BM}^*} + \frac{m G_1'}{L'} \frac{1}{N_{tL}} \frac{x}{y_{BM}}$$

For dilute systems,

$$y_i \approx y^*$$

$$H_{tOG} = H_{tG_1} + \frac{m G_1'}{L'} H_{tL}$$

$$\frac{1}{N_{tOG}} = \frac{1}{N_{tG_1}} + \frac{m G_1'}{L'} \frac{1}{N_{tL}}$$

11-10-2018

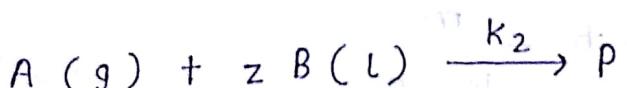
### Absorption with Chemical Reaction

Regime - 1 : Very slow reaction

Regime - 2 : Slow reaction

Regime - 3 : Fast reaction

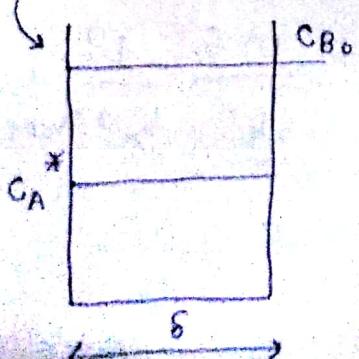
Regime - 4 : Instantaneous reaction



A is limiting reactant. B is in excess.

$$k_L a (C_A^*) >>$$

Interface



$$(k_2 C_B_0 C_A^*)$$

↳ fractional liquid holdup

$$(R_A)^d = L k_2 C_B_0 C_A^*$$

↳ rate of absorption  
= rate of reaction

For slow  
 $k_L a \ll$

$$C_A^*$$

$R_A^d$   
condition

$$\sqrt{M}$$

$$\sqrt{M}$$

$$C_A^*$$

Condition

$$R_A^d =$$

$$\phi =$$

• Instantaneous reaction

Reactor

For slow reaction

$$K_L \alpha \ll L k_2 C_{B_0}$$



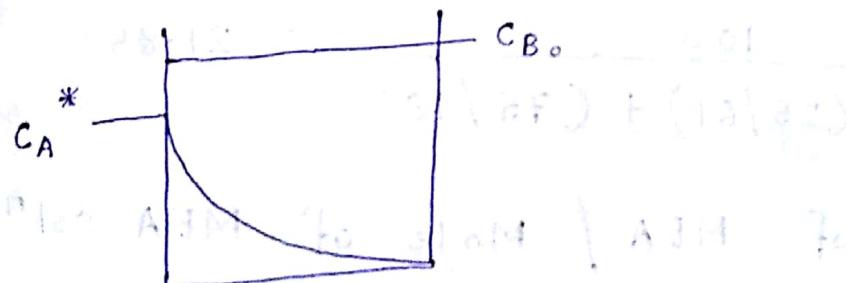
$$R_A \alpha = K_L \alpha C_A^*$$

Condition:

$$\sqrt{M} \ll 1$$

$$\sqrt{M} = \frac{\sqrt{D_A k_2} C_{B_0}}{K_L} = H_a \quad (\text{Hatta no.})$$

For fast reaction (Regime 3)



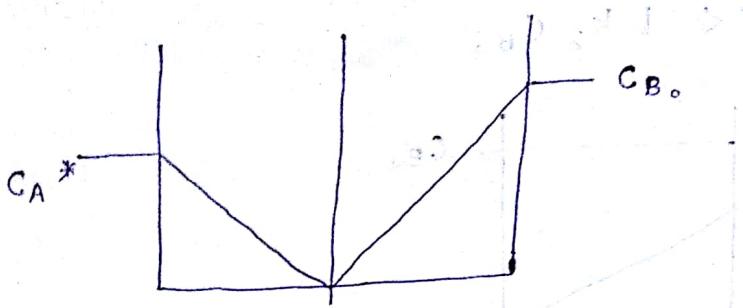
Condition:  $\sqrt{M} > 3$  (fast) (1st)

$$R_A \alpha = K_L \alpha C_A^* \frac{\sqrt{M}}{\tanh \sqrt{M}}$$

$$\phi = \frac{\sqrt{M}}{\tanh \sqrt{M}} \quad \left\{ \begin{array}{l} \phi = \text{enhancement factor} \\ \text{reaction} \end{array} \right.$$

Instantaneous reaction

Reactants will not co-exist



$$R_A \alpha = k_L \alpha C_A^* \left[ 1 + \frac{D_B C_{B_0}}{Z D_A C_A^*} \right]$$

$q \gg 1$

$$\phi = 1/(1+q)$$

### Assignment

2. Mol. wt. of 25% MEA soln {conc. soln}

$$= \frac{100}{(25/61) + (75/18)} = 21.85$$

Moles of MEA / Mole of MEA soln

$$= (25/61) (100/21.85) = 0.0895$$

$$X = \frac{\text{Moles H}_2\text{S}}{\text{mole MEA}} \times 0.0895$$

$$X \quad 0.024 \quad 0.033 \quad 0.052 \quad 0.061 \quad 0.067$$

$$0.072 \quad 0.076$$

$$Y \quad 0.0066 \quad 0.0133 \quad 0.0411 \quad 0.0704 \quad 0.1014$$

$$0.1515 \quad 0.25$$

From the

$y_{N+1} =$

$y_1 =$

$x_0 = 0$

The eq.

$p(x_0, y$

line wil

liq/gas

at ordin

Material

Gs/GYN

For min

Ls,m

Gs

From t

$x_N$ , md

$\therefore L_s, mi$

Gs

$\therefore L, m$

G

$$Y = \frac{P_{H_2S} \cdot 10^3}{P_T - P_{H_2S}} = \left\{ e^{\frac{y}{1-y}} \right\}$$

$$\frac{5}{760 - 5} = 6.62 \times 10^{-3}$$

From the given cond<sup>n</sup>  $y_{N+1}$  is plotted  $\downarrow$

$$y_{N+1} = 0.20 \text{ or } Y_{N+1} = 0.25$$

$$y_1 = 0.01 \text{ or } Y_1 = 0.0101$$

$$x_0 = 0.0152$$

The eq<sup>m</sup> data were plotted and point  $P(x_0, Y_1)$  was located. The operating line will start from  $P$  and for min. liq/gas ratio will cut the eq<sup>m</sup> line at ordinate  $Y_{N+1}$ .

Material Balance,

$$G_s(Y_{N+1} - Y_1) = L_s(X_N - X_0)$$

For min. liq/gas ratio —

$$\underline{L_s, \min} = \frac{Y_{N+1} - Y_1}{X_N, \max - X_0}$$

$$G_s = \frac{X_N, \max - X_0}{Y_{N+1} - Y_1}$$

From the plot,

$$X_N, \max = 0.076$$

$$\therefore \frac{L_s, \min}{G_s} = \frac{0.25 - 0.0101}{0.076 - 0.0152} = 3.94$$

$$\therefore \frac{L, \min}{G_l} = \frac{L_s, \min (1 + X_0)}{G_s (1 + Y_{N+1})}$$

$$= \frac{3.94 \times 1.0152}{1.25} = 3.20$$

Therefore, minimum liquid to gas ratio is 3.2 mol/mol.

$$b) L_s/G_{ls} = 1.15 \times \frac{L_s, \text{min}}{G_{ls}} = 1.15 \times 3.94$$

$$= 4.53, Y = 10.0$$

From the material balance, eq?

$$X_N = X_0 + \frac{Y_{N+1} - Y_1}{L_s/G_{ls}}$$

$$= 0.0152 + \frac{0.25 - 0.0101}{4.53}$$

$$= 0.068$$

Now, operating line drawn b/w points  $(X_0, Y_1)$  and  $(X_N, Y_{N+1})$  and the number of theoretical stages was found to be 2.7.

3. c) For single stage absorption,

$$G_{ls}(Y_1 - y_2) = L_s(x_2 - x_1)$$

$$\Rightarrow 237.6(0.01 - y_2) = 502.8(x_2 - 0.001)$$

24.10.2018

## DISTILLATION

Process of separating the components of mixture/solution into two phases.  
All the components present in all the phases.

Boiling point difference b/w the two components is the driving force.

Unless the difference in b.p.s is very high, we don't get a very high degree of separation.

We do not use any third component.

Evaporation: The vapor phase is pure solvent. Hence, there is no mass transfer as such.

The vapor and liquid are saturated. Therefore, only latent heats get exchanged.

If the latent heats of vaporisation of the components are equal, we have equimolar counter diffusion.

Vapor - Liquid Equilibrium:-

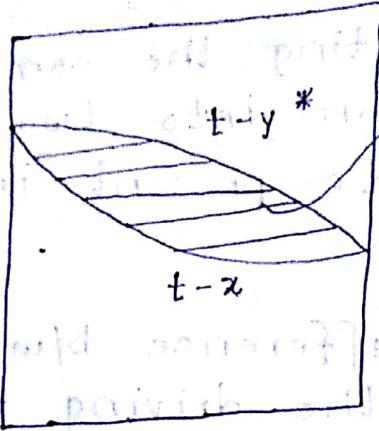
t = x, y at constant pressure

p = x, y at constant temperature

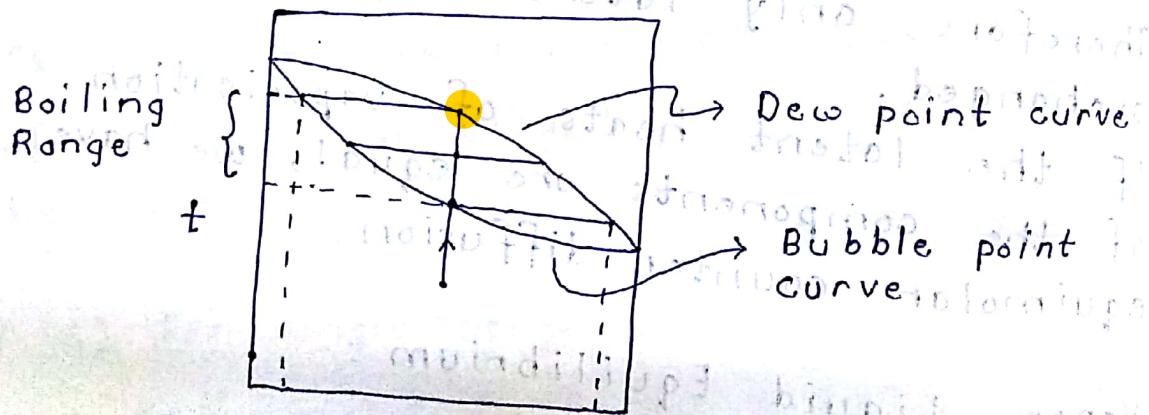
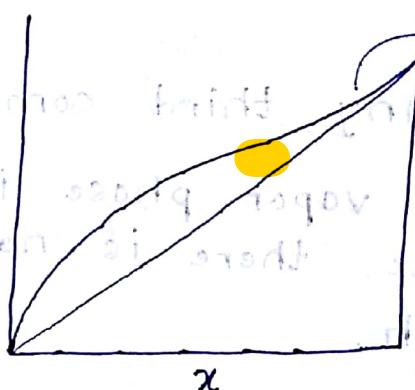
$$f = C - P + 2 = 2 - 2 + 2$$

$$f = 2$$

Variables: P, T, x, y

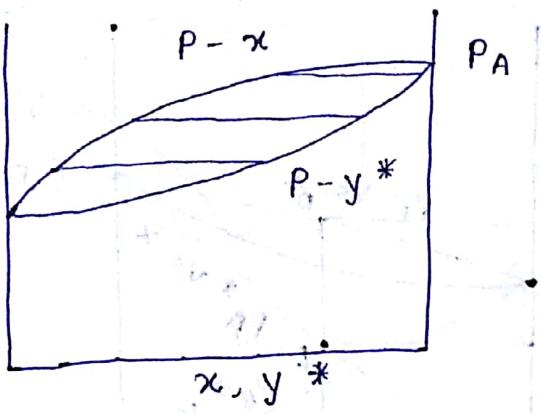


Equilibrium line: tie line



For a mixture, there is no fixed boiling temperature, there is a temperature range.

P - x, y VLE :-



Relative Volatility :  $\alpha$  or  $\alpha_{AB}$

$$\alpha = \frac{y^* / (1-y^*)}{x / (1-x)}$$

$\alpha_{AB}$

If the value of  $\alpha_{AB}$  is higher, the separation will be better.

If  $\alpha = 1$ , liq. and vap. have same composition and separation is not possible.

Azeotropic mixture.

$y^* = \frac{\alpha_{AB} x}{1 + (\alpha_{AB}-1)x}$

Ideal system (Raoult's Law applicable)

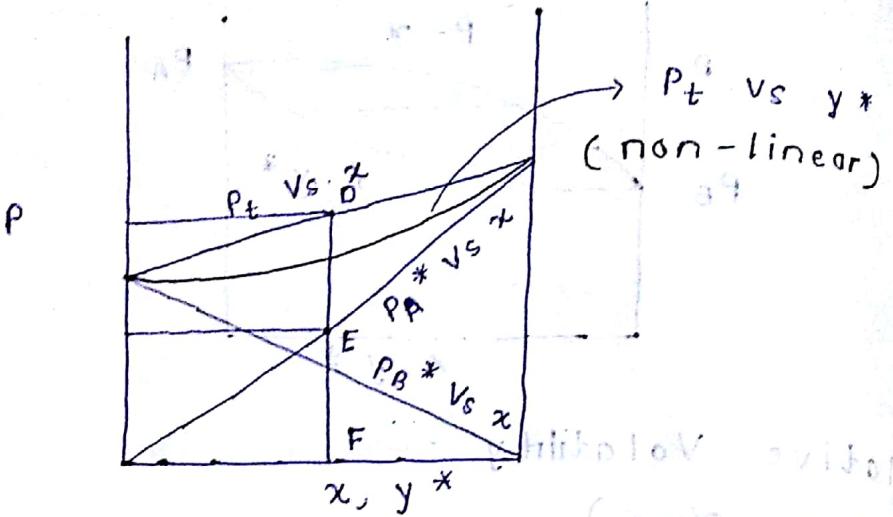
Raoult's Law :-

$$p_A^* = P_A x$$

$$p_B^* = P_B (1-x)$$

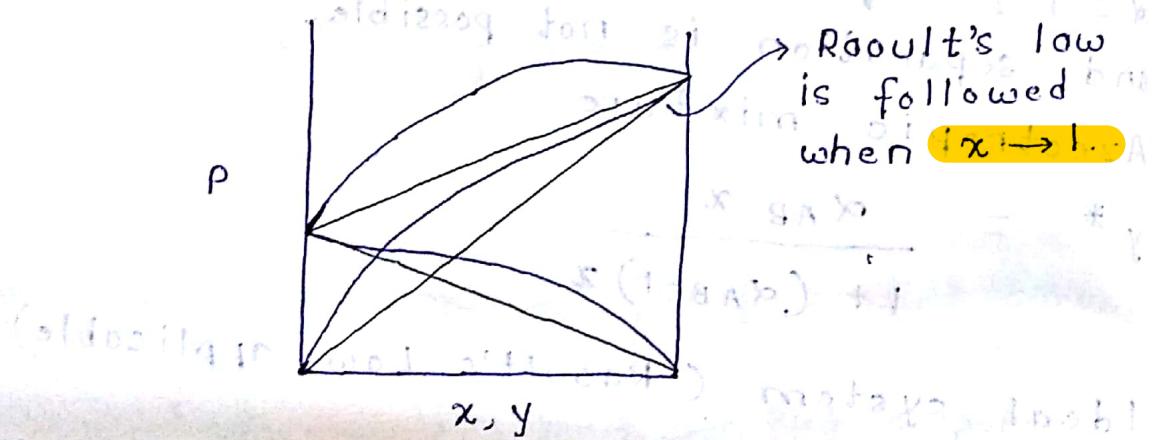
$$P_t = p_A^* + p_B^* = P_A x + P_B (1-x)$$

$$= P_B + (P_A - P_B)x$$



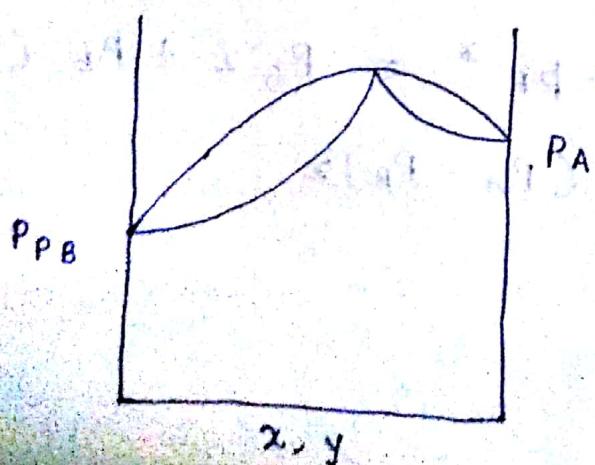
$$y^* = \left( \frac{P_t}{P_A^*} \right)^{-1} \quad C^* = \frac{P_A^* x}{P_t} = \frac{EF}{DF}$$

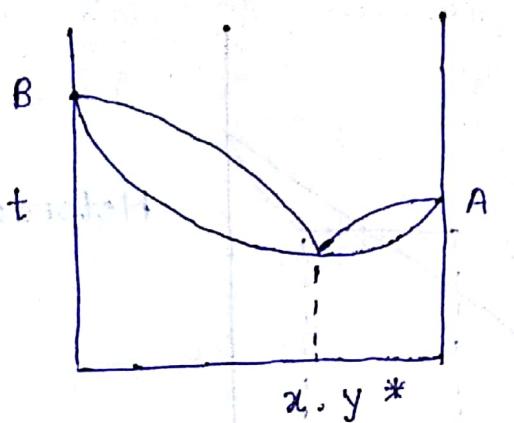
**Positive deviation:** Total and partial pressure more than that determined by Raoult's Law.



**Azeotropic mixture:**

Large positive deviation from ideality.





Minimum Boiling,  
Azeotrope

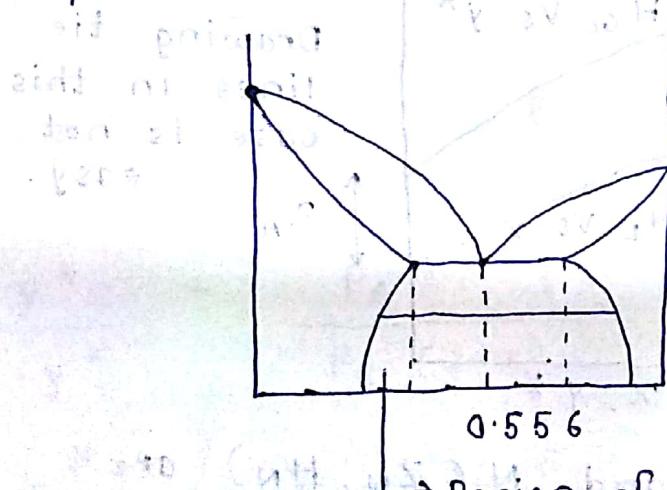
Minimum Boiling Azeotropes are very common.

Example : Ethanol in water.

Large Negative deviation  
Max Maximum Boiling Azeotrope.

Not very common. HCl in water.

At severe deviation from ideality, the liquid phase is not miscible at all compositions.



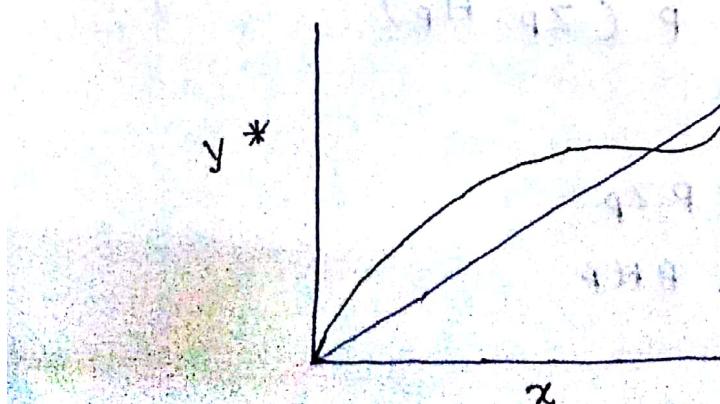
Heteroazeotrope

### Water-Toluene

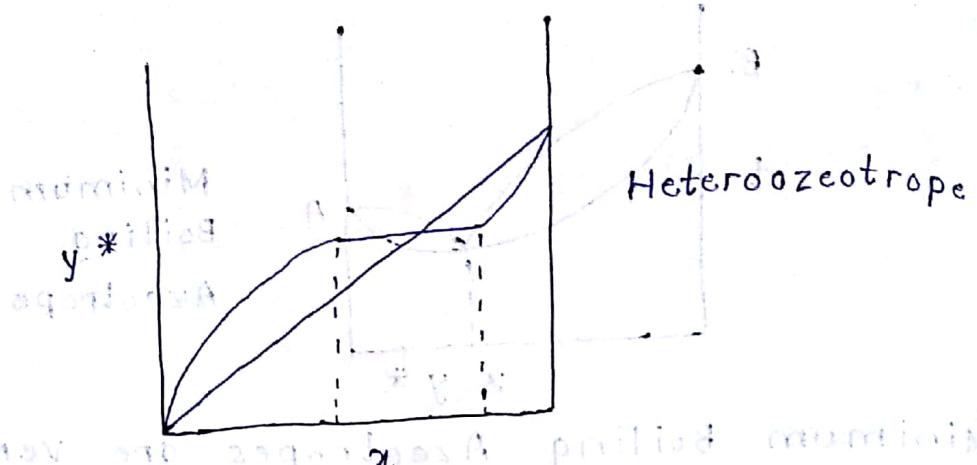
100 °C      110.8 °C

84.1 °C      44.4  
mole % toluene

Region of immiscibility



Minimum Boiling Azeotrope



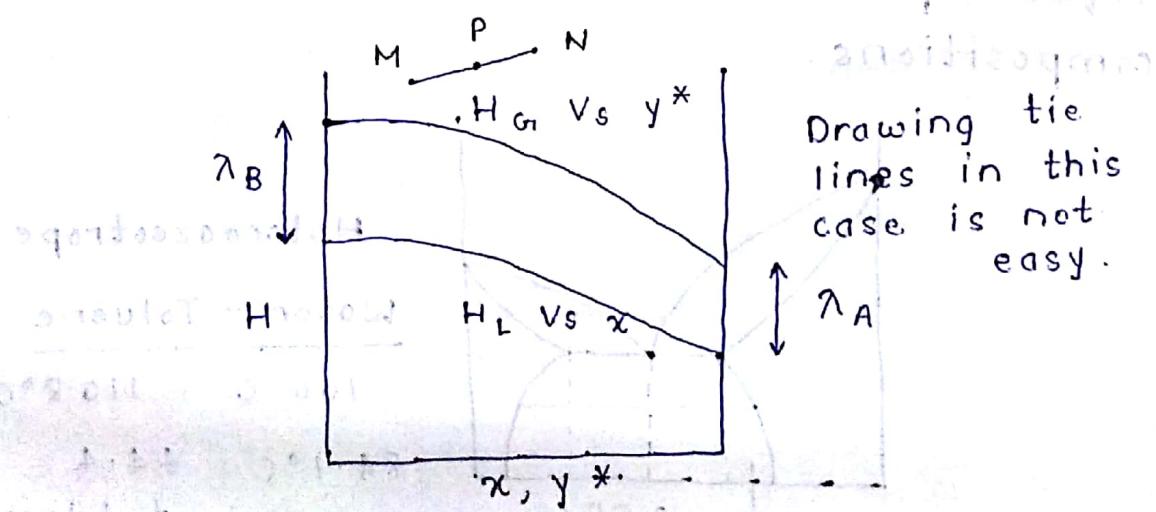
$$H_L = C_L (t_{Lsat} - t_0) M_{ov} + \Delta H_s$$

↓ reference temp ↓

Average saturation temperature Enthalpy of mixing.

$$H_G = y [C_{L,A} (t_{G_i} - t_0) M_A + \gamma_A] +$$

$$(1-y) [C_{L,B} (t_{G_i} - t_0) M_B + \gamma_B]$$



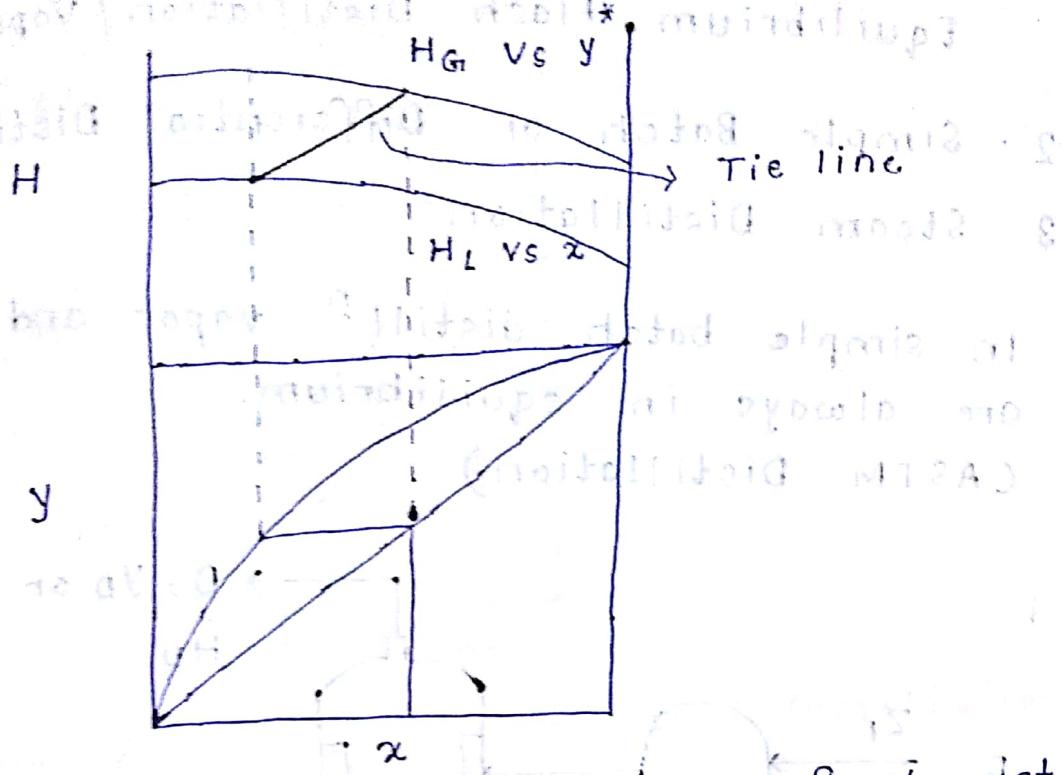
$M(z_M, H_M)$  and  $N(z_N, H_N)$  are mixed to get,  $P(z_P, H_P)$ .

$$M + N = P$$

$$M z_M + N z_N = P z_P$$

$$M H_M + N H_N = P H_P$$

$$\frac{M}{N} = \frac{z_N - z_P}{z_P - z_M} = \frac{H_N - H_P}{H_P - H_M}$$



For ideal systems, the values of  $\alpha_{AB}$  at different temp. are so close

$$y^* = \frac{P_A^*}{P_t} = \frac{P_A x}{P_t}$$

$$1-y^* = \frac{P_B^*}{P_t} = \frac{P_B (1-x)}{P_t}$$

$$\frac{y^*}{1-y^*} = \frac{\frac{P_A x}{P_t}}{\frac{P_B (1-x)}{P_t}} = \frac{P_A x}{P_B (1-x)}$$

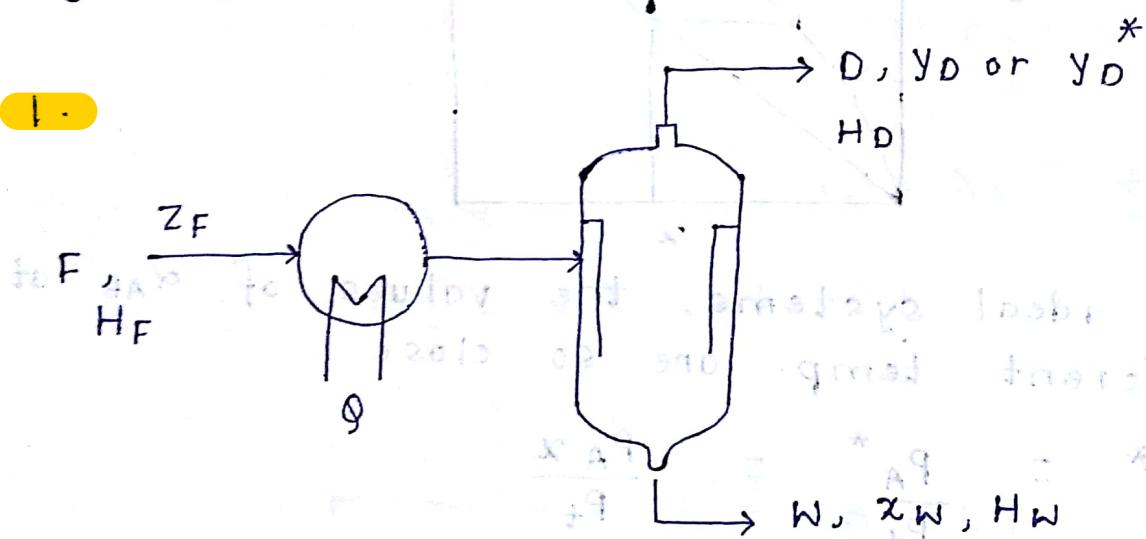
$$\frac{y^*}{1-y^*} / \frac{x}{1-x} = \alpha_{AB} = \frac{P_A}{P_B}$$

ratio of vapor pressures

## Simple Distillation Methods

1. Flash Distillation / Vaporisation
2. Simple Batch or Differential Distillation
3. Steam Distillation

In simple batch distill. vapor and liquid are always in equilibrium.  
(ASTM Distillation) :



2. Simp

Initially  
mol of

$$D + W = F$$

$$D y_D + W x_W = F z_F$$

$$D H_D + W H_W = F H_F + Q = F \left( H_F + \frac{Q}{F} \right)$$

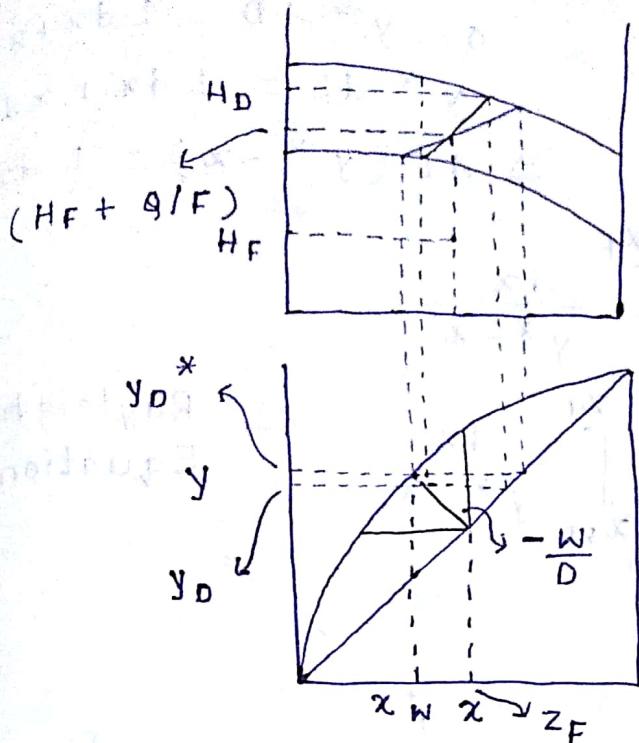
$$-\frac{W}{D} = \frac{y_D - z_F}{x_W - z_F} = \frac{H_D - (H_F + Q/F)}{H_W - (H_F + Q/F)}$$

• Mostly practised in petroleum industry.

Total

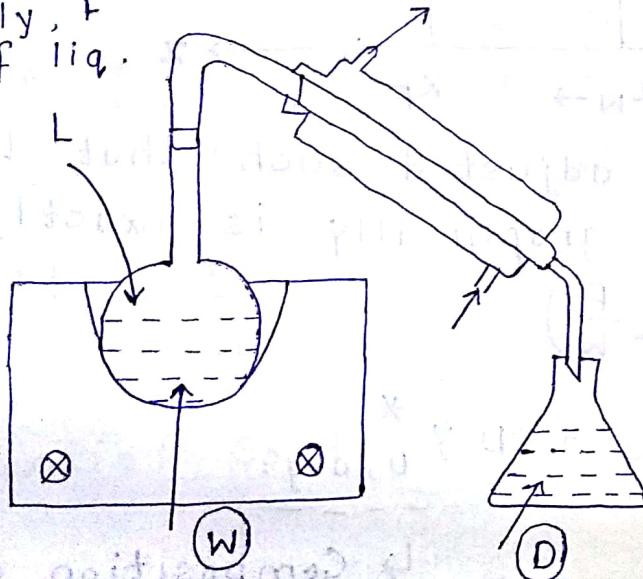
In

isolation  
ation  
iquid



## Q. Simple Batch or Differential Distillation

Initially,  $F$   
mol of liq.



Infinite number  
of successive  
equilibrium flash  
vaporisation.

Vapor and Liquid  
are always in  
equilibrium.

At any instant,  $L$  moles of liquid.

Total Material

Component A

In	0
Out	$dD$
Accum.	$dL$

$$y^* dD$$

$$d(Lx) = Ldx + x dL$$

$$\text{In-out} = \text{Acc}$$

$$0 - dD = dL$$

$$\Rightarrow dL = -dD$$

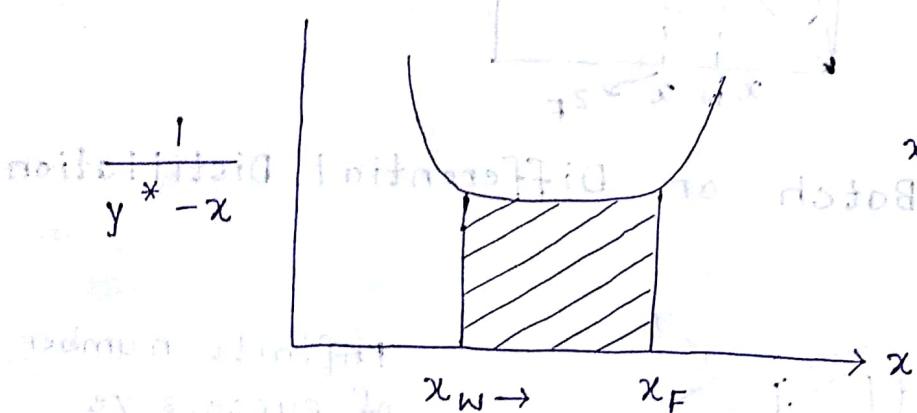
$$0 - y^* dD = L dx + x_d$$

$$y^* dL = L dx + x_d$$

$$\Rightarrow dL (y^* - x) = L dx \quad \ln \frac{F}{W}$$

$$\Rightarrow \int_w^F \frac{dL}{L} = \int_{x_w}^{x_F} \frac{dx}{y^* - x}$$

$$\Rightarrow \ln \left( \frac{F}{w} \right) = \int_{x_w}^{x_F} \frac{dx}{y^* - x} \quad \text{Rayleigh Equation}$$



$x_F$  : fixed  
amount

$$\ln \frac{F}{w} = (\alpha - 1)$$

$x_w$  has to be adjusted such that the area calculated graphically is exactly equal to  $\ln \left( \frac{F}{w} \right)$ .

$$F x_F = w x_w + D y^*_{\text{avg.}}$$

↳ Composition of the  
composited distillate

For ideal system,

$$y^* = \frac{\alpha x}{1 + (\alpha - 1)x}$$

$$y^* - x = \frac{\alpha x - x - (\alpha - 1)x^2}{1 + (\alpha - 1)x}$$

$\Rightarrow \log$

Different

$$\ln \frac{F}{D}$$

$$\begin{aligned}
 &= \frac{(\alpha-1)(1-x)x}{1+(\alpha-1)x} \\
 \ln \frac{F}{W} &= \int_{x_w}^{x_F} \frac{dx}{y^* - x} \\
 &= \int_{x_w}^{x_F} \frac{1+(\alpha-1)x}{(\alpha-1)x(1-x)} dx \\
 &= \frac{1}{\alpha-1} \int_{x_w}^{x_F} \frac{dx}{x(1-x)} + \int_{x_w}^{x_F} \frac{dx}{1-x}
 \end{aligned}$$

$$\begin{aligned}
 \ln \frac{F}{W} &= \ln \frac{x_F}{x_w} + \ln \frac{1-x_w}{1-x_F} + (\alpha-1) \ln \frac{1-x_w}{1-x_F} \\
 &\quad (\alpha-1) \ln \frac{1-x_w}{1-x_F}
 \end{aligned}$$

$$\Rightarrow \ln \frac{F x_F}{W x_w} = \alpha \ln \frac{F (1-x_F)}{W (1-x_w)}$$

$$\Rightarrow \log \frac{F x_F}{W x_w} = \alpha \log \frac{F (1-x_F)}{W (1-x_w)}$$

Differential Condensation:-

$$\ln \frac{F}{D} = \int_{y_F}^y \frac{dy}{y^* - x}$$

## Steam Distillation :-

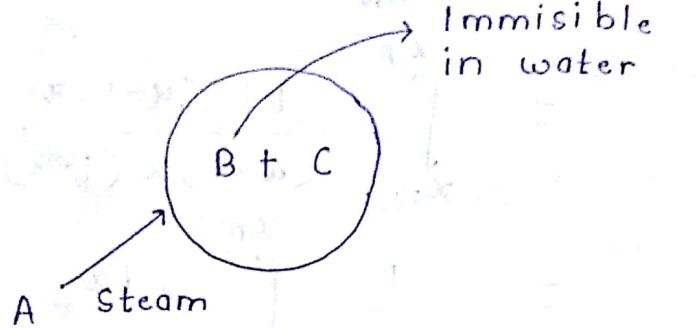
$B + C$

↓

very high  
boiling point

volatile but  
decomposes at  
its b.p.

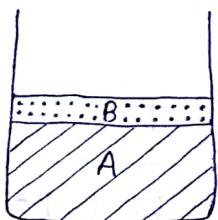
non-volatile



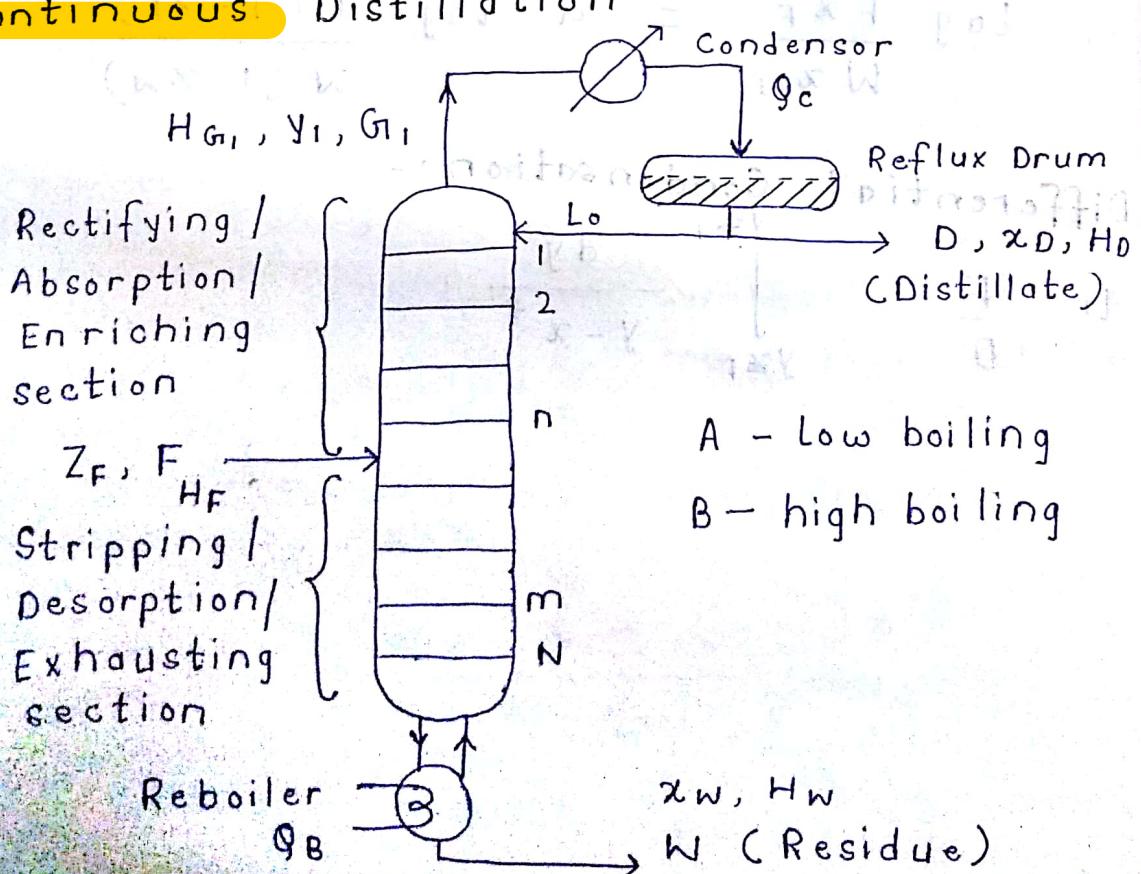
Two layers in the condensate.

$$\text{At temp } t \quad P_A + P_B = P_T$$

if A is more volatile than B;



## Continuous Distillation



A gets  
,  
B  
Condenser  
on equili  
partial C  
required

Reflux

Reflux

$G_1$  =

$G_1$  =

Enthal

$G_1, H_G$

$Q_C$  =

$Q_C$  :

Reboil

$F_H F$

$Q_B$  =

1. Ponc

2. McC

A gets concentrated in vapour phase.  
 B , , , , , , , , liquid , , , ,

Condenser is fully condensing mostly (not on equilibrium stage).

partial condenser condenses vapour that is required for reflux (vap - liq. equilibrium)

$$\text{Reflux Ratio } (R) = \frac{L_0}{D}$$

Reflux Drum :-

$$G_{T1} = L_0 + D$$

$$G_1 = RD + D = D(R+1)$$

Enthalpy balance -

$$G_1 H_{G_1} = L_0 H_{L_0} + D H_D + Q_c$$

$$Q_c = G_1 H_{G_1} - L_0 H_{L_0} - D H_D$$

$$Q_c = D [ (R+1) H_{G_1} - R H_{L_0} - H_D ]$$

Heat lost

Reboiler :-

$$F H_F + Q_B = D H_D + W H_W + Q_c + Q_L$$

$$Q_B = D H_D + W H_W + Q_c + Q_L - F H_F$$

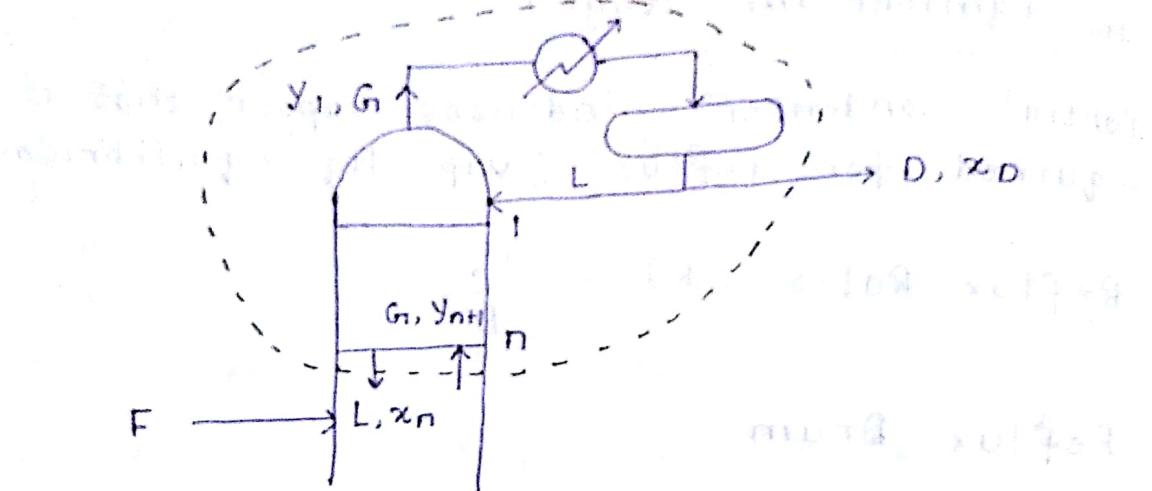
1. Ponchon and Savarit
2. McCabe & Thiele

To determine the number of ideal stages

## McCabe & Thiele Method

Principle: Equimolar overflow and vaporisation

Latents: one almost same, the other



Reflux liquid is at its bubble point.

$$R = L/D$$

$$G_1 = L + D$$

$$G_1 = RD + D = D(R+1)$$

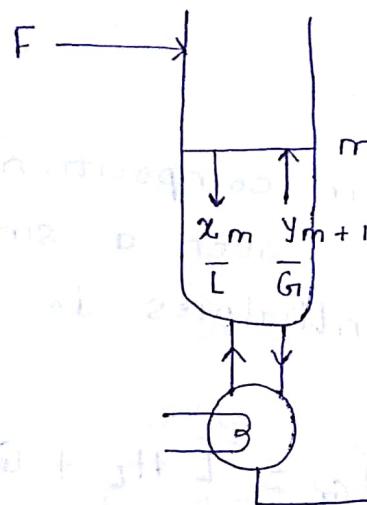
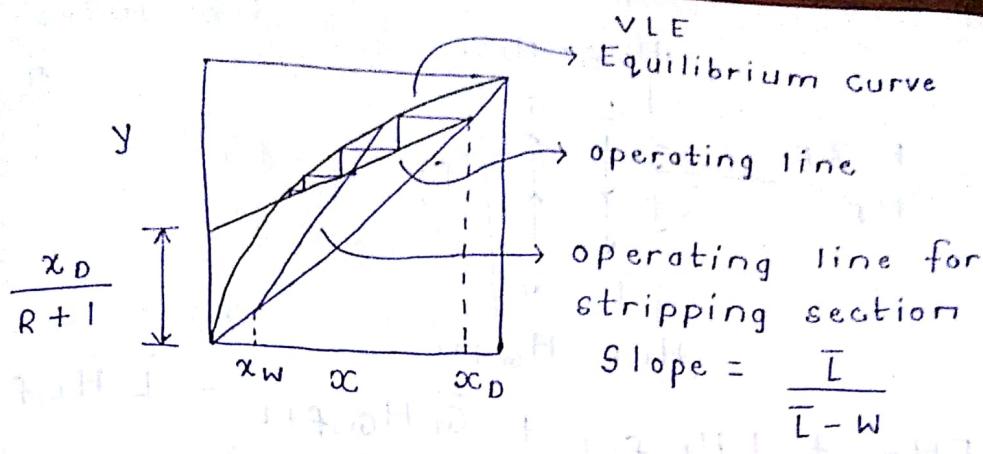
$$G_1 y_{n+1} = L x_n + D x_D$$

$$\Rightarrow y_{n+1} = \frac{L}{G_1} x_n + \frac{D x_D}{G_1}$$

$$= \frac{R\phi}{\phi(R+1)} x_n + \frac{\phi}{\phi(R+1)} x_D$$

$$\Rightarrow y_{n+1} = \frac{R}{R+1} x_n + \frac{x_D}{R+1}$$

Eq<sup>n</sup> of operating line!



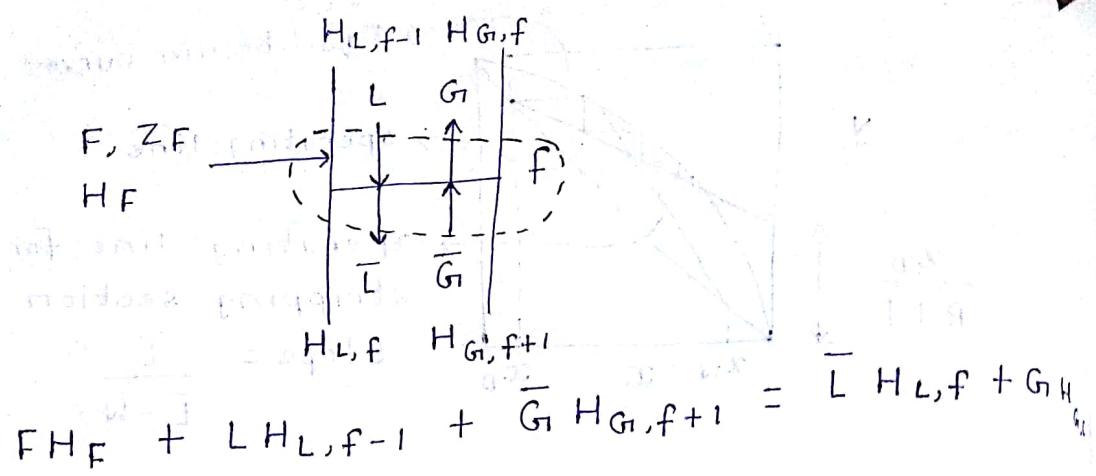
$$\bar{L} = \frac{\bar{G}_1 + W}{\bar{G}_1} = \frac{\bar{H}_1 - \bar{H}(1 - f)}{\bar{G}_1} = \frac{\bar{H}_1 - \bar{H}(1 - f)}{\bar{G}_1}$$

$$\Rightarrow \frac{\bar{G}_1}{\bar{G}_1} = \frac{\bar{H}_1 - \bar{H}(1 - f)}{\bar{L} - W} = \frac{\bar{H}_1 - \bar{H}(1 - f)}{\bar{L}} + \frac{\bar{H}(1 - f)}{\bar{L}} = \frac{\bar{H}_1 - \bar{H}}{\bar{L}} + \frac{\bar{H}f}{\bar{L}}$$

$$\Rightarrow \bar{L}x_m = \frac{\bar{H}_1 - \bar{H}}{\bar{G}_1} y_{m+1} + \frac{\bar{H}f}{\bar{G}_1} x_w + \frac{\bar{H}}{\bar{G}_1}$$

$$\Rightarrow y_{m+1} = \frac{\bar{L}}{\bar{G}_1} x_m - \frac{W}{\bar{G}_1} x_w + 1$$

$$\Rightarrow y_{m+1} = \frac{\bar{L}}{\bar{L} - W} x_m - \frac{W}{\bar{L} - W} x_w$$



$$F H_F + L H_{L,f-1} + \bar{G} H_{G,f+1} = \bar{L} H_{L,f} + G H_G$$

$$F + L + \bar{G} = \bar{L} + G$$

$$\Rightarrow \bar{G} - G = (\bar{L} - L) - F$$

In practice change in composition of the streams is very less over a single tray that molar enthalpies do not change.

$$\therefore F H_F + L H_L + \bar{G} H_G = \bar{L} H_L + G H_G$$

$$(\bar{G} - G) H_G = (\bar{L} - L) H_L - F H_F$$

$$(\bar{L} - L) H_G - F H_G = (\bar{L} - L) H_L - F H_F$$

$$\Rightarrow (\bar{L} - L) (H_G - H_L) = F (H_G - H_F)$$

$$\Rightarrow \frac{\bar{L} - L}{F} = \frac{H_G - H_F}{H_G - H_L} = q$$

= Heat required to convert 1 mole of heat to saturated vapor

Molar Latent heat of vap.

For saturated liquid,

$$H_F = H_L ; q = 1$$

q-line for  
saturated

For saturated vapor,  
 $q = 0$

$$\Rightarrow \bar{G} - G_1 = (\bar{L} - L) - F = Fq - F = F(q-1)$$

$$\bar{G} = G_1 + F(q-1)$$

$$\bar{G} = D(R+1) + F(q-1)$$

$H_{L,f} + G_{H_{G_f}}$

$$G_1 y = Lx + Dx_D$$

$$\bar{G} y = \bar{L}x - w x_w$$

$$(\bar{G} - G_1)y = (\bar{L} - L)x - (w x_w + D x_D)$$

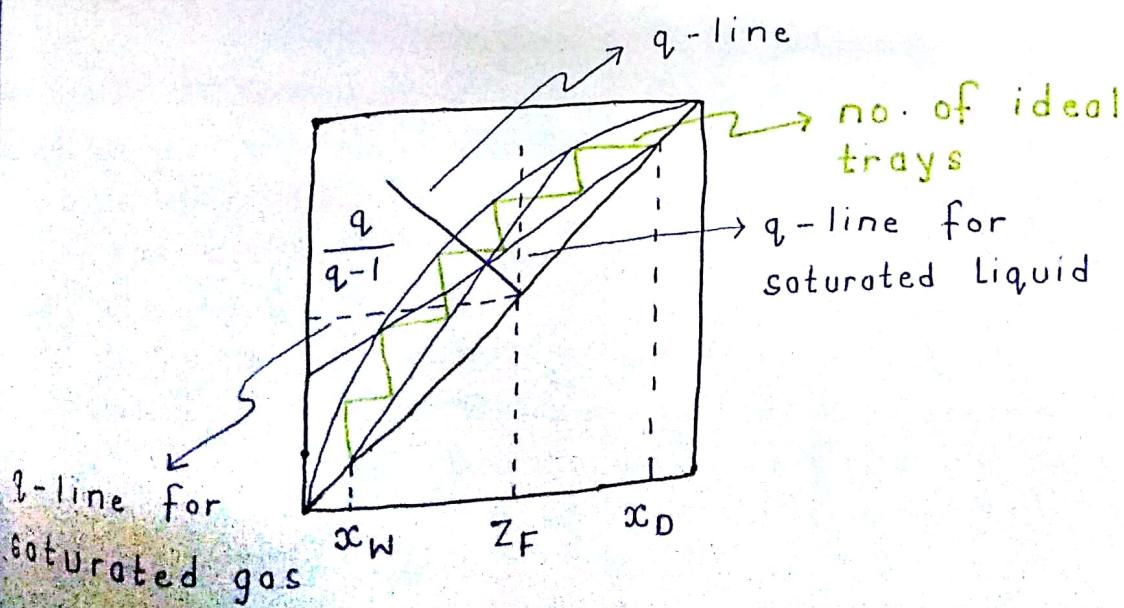
$$y = \frac{(\bar{L} - L)x}{\bar{G} - G_1} - \frac{F z_F}{\bar{G} - G_1}$$

Total material balance

$$y = \frac{F q}{F(q-1)} x - \frac{F z_F}{F(q-1)}$$

$$y = \frac{q}{q-1} x - \frac{z_F}{q-1}$$

$$x = z_F, \quad y = z_F$$



1 mole  
r