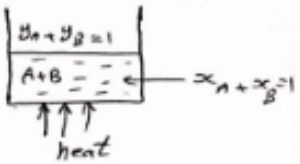
Distillation

The seperation of liquid mixture into their several components. It is the key operation of oil refinery.

where x = Mol fraction of component in the liquid face

y = Mol fraction of a component in the gas phase.



$P_A \neq P_B$ = partial pressure of A, B

P_A°, P_B° = vapor pressure of A, B

Dalton law: Total pressure is equal to the summation of the partial pressure.

$$P_T = \sum P_A$$

$$P_A = y_A P_T$$

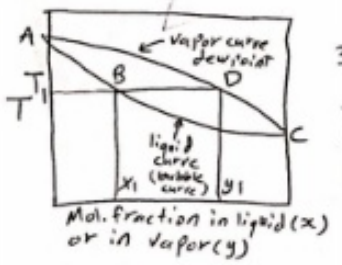
Raults law:

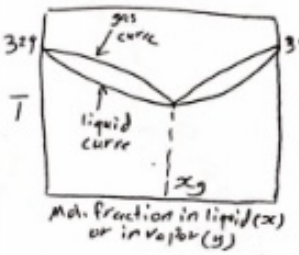
$$P_A = P_A^\circ x_A$$

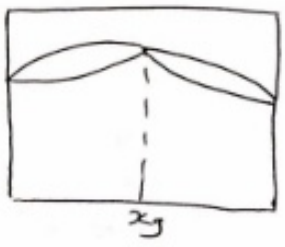
$$P_B = P_B^\circ x_B$$

Henrys law: $P_A = H x_A$

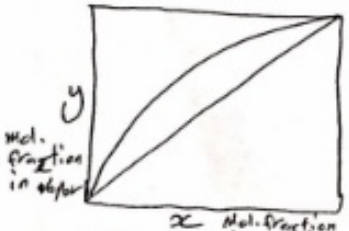
Vapor liquid Equilibrium



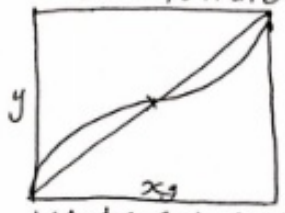




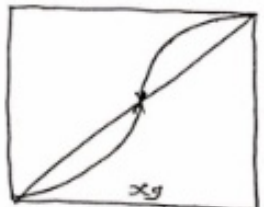
Temp. composition diagram



a) Benzen - Toluen



b) Aceton - Carbondisulphide



c) Aceton - chloroform

Vapor composition as a function of liquid composition at constant pressure and temp.

- The curve ABC shows the composition of the liquid which boils at any given temperature.
- The curve ADC the corrosponding composition of the vapor at that temp.

Thus, a liquid of composition x_1 will boil at temperature T_1 and the vapor in equilibrium is indicated by point D of composition y_1. And for any liquid composition x the vapour formed will be richer in More volatile Component (M.V.C) (where x = mol. fraction of volatile component in the liquid. y = mol. fraction of more volatile component in the vapor.)

Fig. b&c

There is a critical composition x_g where the vapor has the same composition as the liquid, so that no change occurs on boiling. Such critical mixtures are called azeotropes.

equilibrium curve:-

$$P_A = y_A P_T \Longrightarrow y_A = \frac{P_A}{P_T} \quad \text{---(1)}$$

for ideal mixture $P_A = P_A^\circ x_A$ ---(2) Raoult's law

$P_A = H x_A$ Henry's law

$$y_A = \frac{P_A^\circ x_A}{P_T} \quad \& \quad y_B = \frac{P_B^\circ x_B}{P_T}$$

$$y_A + y_B = 1$$

$$\frac{P_A^\circ x_A}{P_T} + \frac{P_B^\circ (1 - x_A)}{P_T} = 1 \Longrightarrow \boxed{x_A = \frac{P_T - P_B^\circ}{P_A^\circ - P_B^\circ}} \quad \text{---(3)}$$

Volatility & Relative Volatility (α)

$$\text{Volatility of } A = \frac{P_A}{x_A}, \quad \text{Volatility of } B = \frac{P_B}{x_B}$$

$$\text{Relative Volatility } \alpha = \frac{\text{Volatility of A}}{\text{Volatility of B}} = \frac{P_A / x_A}{P_B / x_B}$$

$$= \frac{P_A x_B}{P_B x_A} \quad \text{---(4)}$$

$$P_A = y_A P_T \quad \& \quad P_B = y_B P_T$$

$$\alpha = \frac{P_A^\circ x_A / x_A}{P_B^\circ x_B / x_B} = \frac{P_A^\circ}{P_B^\circ}$$

$$\therefore \alpha = \frac{y_A x_B}{y_B x_A} \Rightarrow \boxed{\frac{y_A}{y_B} = \alpha \frac{x_A}{x_B}} \quad \text{---- (5)}$$

for binary mixture $y_B = 1 - y_A$

$x_B = 1 - x_A$

$$\alpha = \frac{y_A (1 - x_A)}{(1 - y_A) x_A}$$

$$y_A = \frac{\alpha x_A}{1 + (\alpha - 1) x_A} \quad \text{---- (6)}$$

$$x_A = \frac{y_A}{\alpha - (\alpha - 1) y_A} \quad \text{---- (7)}$$

if $\alpha = 1 \Rightarrow$ no seperation occur يحدث فصل لا

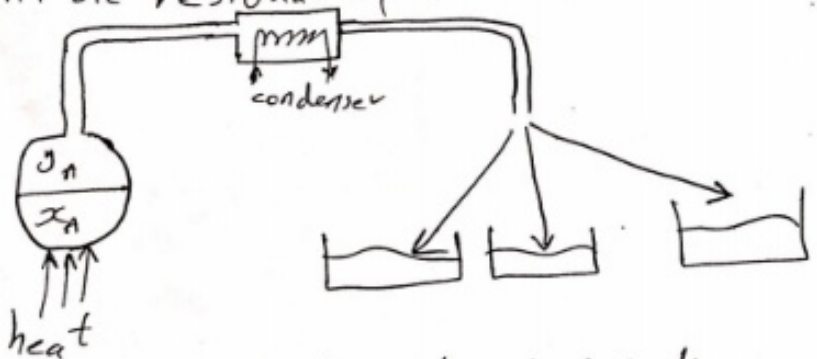
$(x_A = y_A)$

$\alpha > 1 \Rightarrow$ seperation occur يحدث فصل

Type of Distillation

1- Differential Distillation

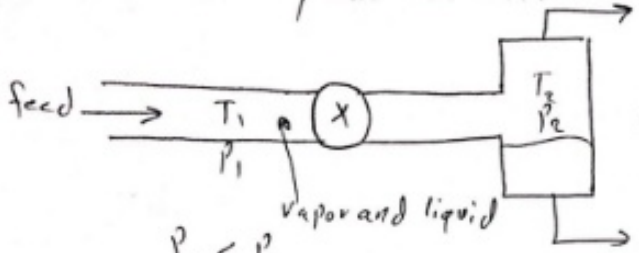
Starting with a still pot, initially full, heated at a cons
rate. In this process the vapor formed con boiling the liqu
is removed at once from the system. Since the vapor is
richer in the more volatile component than the liquid.
The vapor formed over a short period is in equilibr
with the liquid, the total vapor formed is not in equili.
with the residual liquid



differential distillation

Flash Distillation

is a continuous process. The feed is passed into the still, where part is vaporised, the vapor remaining in contact with the liquid. The mixture of vapor and liquid leaves the still and is seperated so that the vapor is in equilibrium with the liquid.

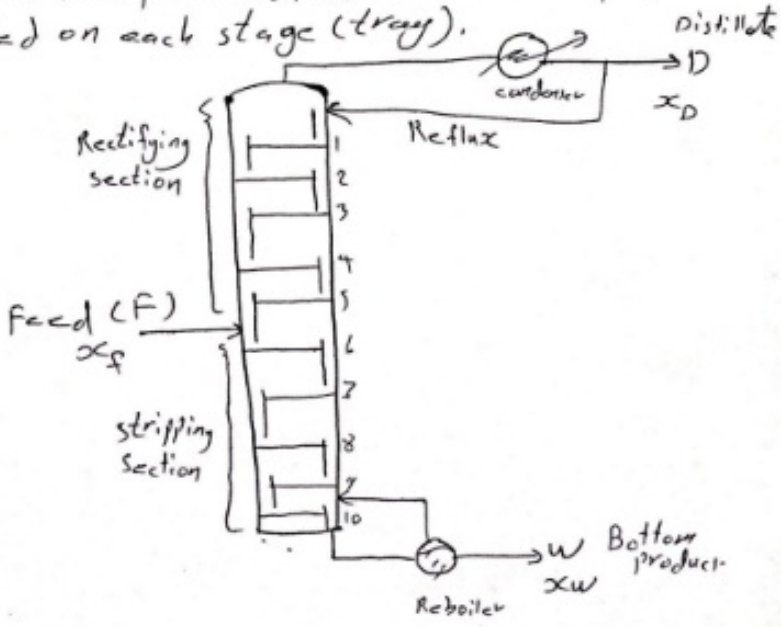


$P_2 < P_1$

$T_2 < T_1$

3- Multi stage continuous distillation "Rectification" (Fractionating column):

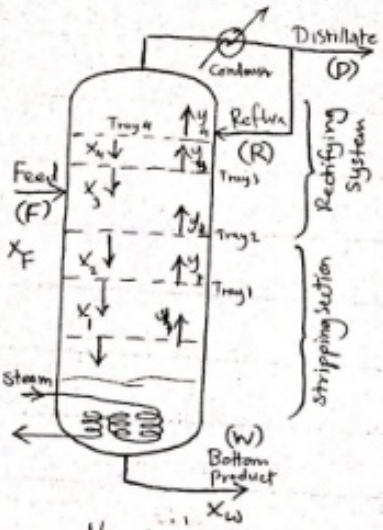
There is a partial condensation of rising vapor and partial vaporisation of the reflux liquid is repeated on each stage (trays).

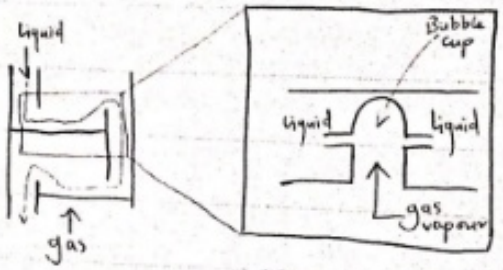


[3] Multi Stage continous distillation (Rectification)

[fractionating column]; because

There is a partial condensation of the rising vapour & Partial vaporization of the liquid Reflux.





∴ The disturbance will occur if it is change

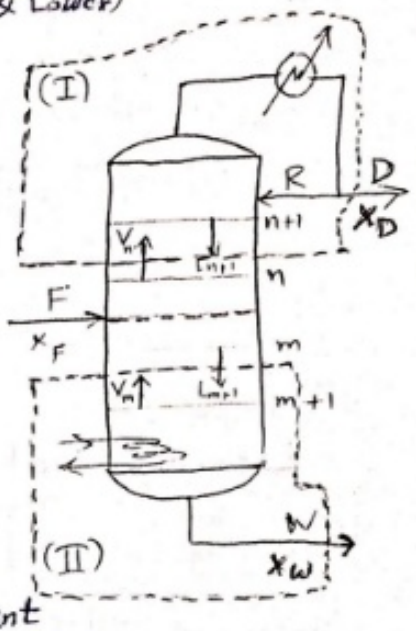
- The feed stream must introduce on same intermediate tray where the liquid has approximate the same composition as the feed [on each tray the system tends to reach equilibrium because]

[A] from the rising vapour some of the less volatile component (L.V.C) condensed into the liquid. thus increasing the concentration of more volatile component (M.V.C) in the vapour.

[B] From the liquid in the tray some of the (M.V.C) is vaporized thus increasing the concentration of (L.V.C) in the liquid.

Calculation of number of plates:

we have two operating line (upper & Lower)



- Top of the column (Box I) upper:

$$V_n = L_{n+1} + D \quad \text{—(1)}$$

(M.V.C) Materail Balance:

$$y_n V_n = L_{n+1} \cdot x_{n+1} + D x_D \quad \text{—(2)}$$

$$y_n = \frac{L_{n+1}}{V_n} x_{n+1} + \frac{D}{V_n} x_D \quad \text{—(3)}$$

The moles of liquid over flow are constant

$$L_n = L_{n+1}$$

$$\boxed{y_n = \frac{L_n}{V_n} x_{n+1} + \frac{D}{V_n} x_D} \quad \text{—(4)}$$

upper operating line or Top ope
line equ

The relation between composition of the vapour rising to the plate & the composition of the liquid on any plate above the feed

Material Balance in the Bottom (Box II)

$$L_{m+1} = V_m + W$$

$$L_m = L_{m+1}$$

(M.V.C) Material Balance:

$$y_m V_m = L_{m+1} x_{m+1} - W x_w \qquad (5)$$

$$y_m = \frac{L_m}{V_m} x_{m+1} - \frac{W}{V_m} x_w$$

Lower operating line equation

The relation between the composition of vapour rising to the plate & the liquid to the plate (for any plate below the feed).

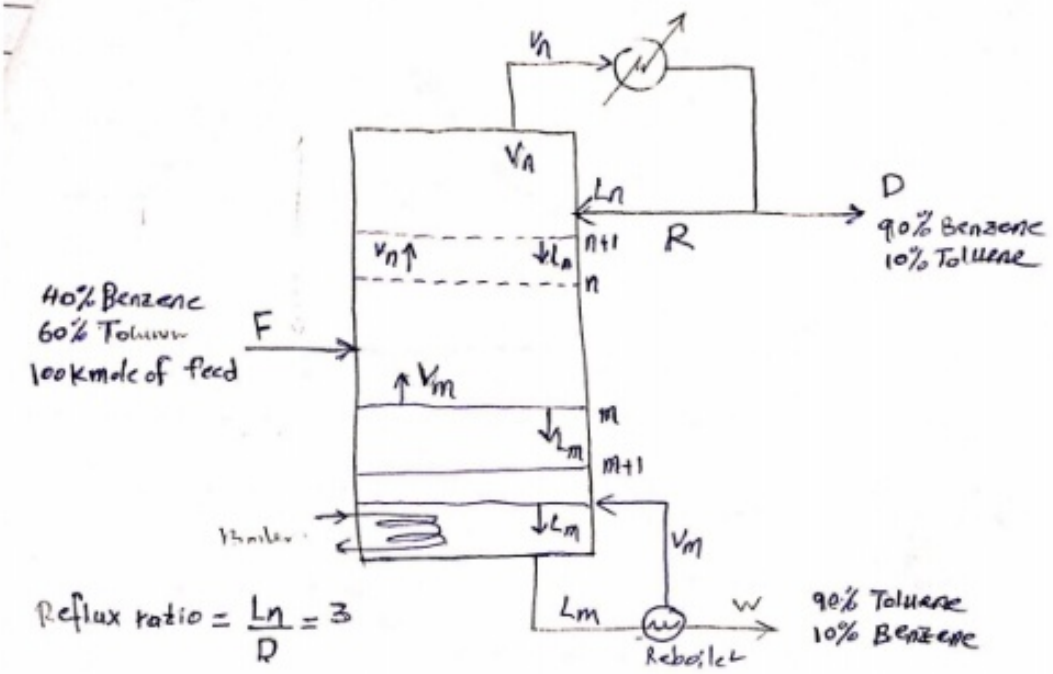
Example (11-7): (stage to stage calculation)
Lewis-sorel method

mole fraction of Benzen in liquid (x)	0	0.05	0.1	0.2	0.3	0.4	0.5	0.6	0.7	0.8	0.9
mole fraction of benzene in vapour (y)	0	0.118	0.205	0.37	0.51	0.62	0.71	0.79	0.85	0.91	0.95

from question:

from the question:

(8)



overall material Balance: (O.M.B)

$$F = D + W$$

$$100 = D + W \quad \text{——(1)}$$

(M.V.E) M.B:

$$0.4F = 0.9D + 0.1W \quad \text{——(2)}$$

$$\therefore W = 62.5 \text{ Kmole}$$

$$D = 37.5 \text{ Kmole}$$

$$\therefore R = \frac{L_n}{D} \Rightarrow L_n = R \times D = 3 \times 37.5$$

$$\therefore L_n = 112.5 \text{ Kmole}$$

$$\because V_n = L_n + D = 112.5 + 375 = 150 \text{ Kmole}$$

Top operating line:

$$y_n = \frac{L_n}{V_n} x_{n+1} + \frac{D}{V_n} x_D$$

$$y_n = \frac{112.5}{150} x_{n+1} + \frac{37.5}{150} * 0.9$$

$$\boxed{y_n = 0.75\, x_{n+1} + 0.225}$$ Top operating line

$\because L_m = L_n + F = 112.5 + 100 = 212.5$ kmole

$L_m = V_m + W \Rightarrow V_m = L_m - W = 212.5 - 62.5$

$V_m = 150 = V_n$

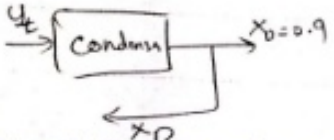
Lower operating line:

$$y_m = \frac{L_m}{V_m} x_{m+1} - \frac{W}{V_m} x_W$$

$$y_m = \frac{212.5}{150} x_{m+1} - \frac{62.5}{150} * 0.1$$

$$\therefore \boxed{y_m = 1.415\, x_{m+1} - 0.042}$$ Lower operating line

$y_t = 0.9$



the composition of liquid at the top, which is at equilibrium with (y_t)

Since all the vapor from the column is condensed the composition of the vapor $y_t = x_D = x_R$ (composition of the liquid returned as reflux.

Now we will use the lower operating line equation:

$$y_{t-4} = (1.415 \times 0.38) - 0.042$$

$$y_{t-4} = 0.49$$

$$x_{t-4} = 0.29 \quad \text{(from graph)}$$

using lower operating line equ.

$$y_{t-5} = (1.1415 \times 0.29) - 0.042$$

$$y_{t-5} = 0.379$$

$$x_{t-5} = 0.21$$

Using lower operating line equ.

$$y_{t-6} = (1.145 \times 0.21) - 0.042$$

$$y_{t-6} = 0.25$$

$$x_{t-6} = 0.13$$

Using lower operating line equ.

$$y_{t-7} = (1.145 \times 1.3) - 0.042$$

$$y_{t-7} = 0.127$$

$$x_{t-7} = 0.055 \quad < x_w$$

كفاءة التقطير تعتمد على :
1- bubble cup
2- Reboiler
3- Condenser
4- high of weed

Thus mean we have (3) stages in the bottom

3 Stages in bottom

3 Stages in Top & 1 stage for boilers

No. of theortical stage = 3+3+1 = 7 stages.

Actual No. of stages = $\dfrac{\text{theortical No. of stages}}{\text{efficiency of D...}}$

McCabe Thiele Diagram

Now by using the Mccabe Thiele diagram to calculate the number of stage.

$$\therefore\ y_n = \frac{L_n}{V_n} x_{n+1} + \frac{D}{V_n} x_D \quad \text{---(1)}$$ T.o.L Top operating line

$V_n = V_{n+1}$
$L_n = L_{n+1}$
} for constant molar heat of vaporization & no heat loss

If: $x_{n+1} = x_D$

$\therefore V_n = L_n + D$

$\therefore L_n = V_n - D$

Substitute in (1):

$$y_n = \frac{V_n - D}{V_n} x_D + \frac{D}{V_n} x_D$$

$$y_n = \left(1 - \frac{D}{V_n}\right) x_D + \frac{D}{V_n} x_D$$

$$y_n = x_D - \frac{D}{V_n} x_D + \frac{D}{V_n} x_D$$

$$\therefore y_n = x_D$$

$\therefore$ Upper operating line passes through two point

(x, x) & $\left(0, \frac{Dx_D}{V_D}\right)$

- Now draw $(x_D, x_B)(0.9, 0.9)$ in graph

- we know from solution:

$\frac{L_n}{D} = 3$, $V_n = V_m = 150$, $L_m = 112.5$, $L_m = 212.5$, $D = 37.5$

$\therefore$ Intersept $= \frac{D}{V_n} x_D = \frac{37.5}{150} * 0.9$

$\therefore$ Intesept $= 0.225$

$\because x_F = 0.4$

Lower operating line

$$y_m = \frac{L_m}{V_m} x_{m+1} + \frac{W}{V_m} x_w$$

if: $x_{m+1} = x_w$

$\because L_m = V_m + W$

$$\therefore y_m = \left(\frac{V_m + W}{V_m}\right) x_w - \frac{W}{V_m} x_w$$

$$= x_w + \frac{W}{V_m} x_w + \frac{W}{V_m} x_w$$

$\therefore y_m = x_w$ $\quad (x_w, x_w)$ $(0.1, 0.1)$

$\therefore$ the lower operating line passes through two points (x_w, x_w) & slope $\left(\frac{L_m}{V_m}\right)$

Now draw in graph

$\therefore$ No of stage = 6 stage

Example (11-7) in page- 564 Volume (2)

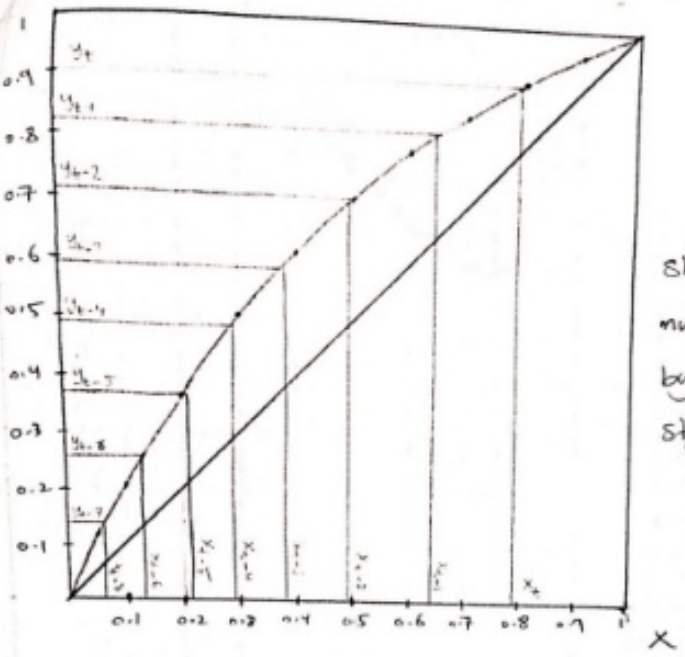


Fig (1): Show how to find number of stage by using Stage-to Stage calculation

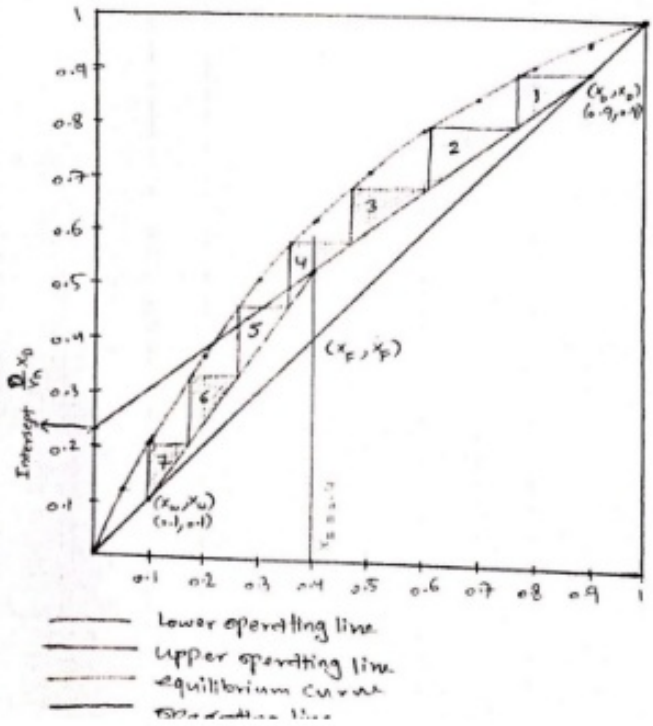
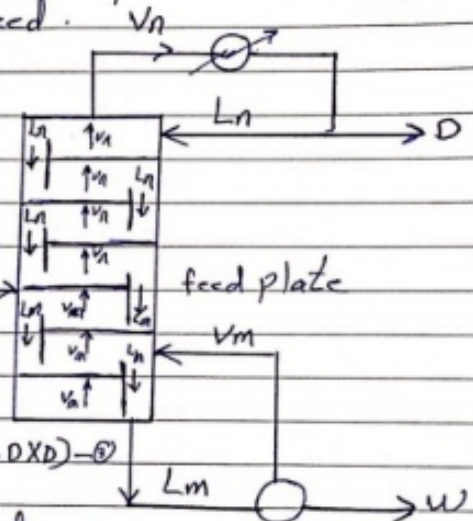


Fig (2) show how to find number of stage by using Mecabe Thiele diagram

q-line equation

The intersection of the two operating lines.

- The feed enters as liquid at its boiling point
- The two operating lines intersect at a point having an x-coordinate of x_f
- The intersection of the two operating lines (x_q, y_q) depend on temperature and physical condition of the feed.



$V_n = D + L_n$

$L_m = V_m + W$

$V_m = L_m - W$

$V_n y_q = L_n x_q + D x_D$ — ①

$V_m y_q = L_m x_q - W x_W$ — ②

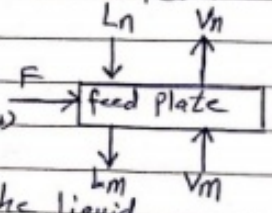
② − ① $y_q (V_m - V_n) = (L_m - L_n) x_q - (W x_W + D x_D)$ — ③

Material Balance over the feed

$F + L_n + V_m = L_m + V_n$

$V_m - V_n = L_m - L_n - F$ — ④

or $L_m = L_n + F + (V_m - V_n)$ — ④



IF the feed in the form of the liquid at its boiling point.

and H_{fs} = the enthalpy of one mole of feed at its boiling point.

H_f = the enthalpy of one mole of feed at temperature T_f.

then the heat to be supplied to bring feed to the boiling point = $F(H_{fs} - H_f)$

and the number of moles of vapor to be condensed if $(H_f < H_{fs})$ to provide this heat = $\frac{F(H_{fs} - H_f)}{\lambda}$

$= V_m - V_n$

Where λ = molar latent heat of the vapor

From equation (4)

$$L_m = L_n + F \rightarrow \frac{F(H_{fs} - H_f)}{\lambda}$$

$$L_m = L_n + F\left[1 + \frac{H_{fs} - H_f}{\lambda}\right]$$

$$L_m = L_n + F\left[\frac{\lambda + (H_{fs} - H_f)}{\lambda}\right]$$

$$L_m = L_n + qF \quad \text{---(5)} \Rightarrow L_m - L_n = qF$$

Where $q = \frac{\lambda + (H_{fs} - H_f)}{\lambda}$

$$= \frac{\text{heat required to vaporised 1 mole of feed}}{\text{molar latent heat of vaporisation of the feed}}$$

From equation (4) & (5)

$$V_m - V_n = qF - F \quad \text{---(6)}$$

Material Balance of the M.V.C. over the whole column.

$$FX_f = DX_D + WX_W$$

From equations (3), (5), (6)

$$y_L(\underbrace{qF - F}_{V_m - V_n}) = \underbrace{qFx_q}_{L_m - L_n} - \underbrace{Fx_f}_{DX_D + WX_W}$$

$$y_L F(q-1) = qFx_q - Fx_f$$

$$\boxed{y_L = \frac{q}{q-1}x_q - \frac{x_f}{q-1}}$$

q-line equation of slope $\frac{q}{q-1}$ and intersept $= \frac{x_f}{q-1}$

when $y_L = 0$

$$0 = \frac{q}{q-1}x_q - \frac{x_f}{q-1}$$

$$\frac{x_f}{q-1} = \frac{q}{q-1}x_q$$

$$x_f = qx_q \Rightarrow \boxed{x_q = \frac{x_f}{q}}$$

intersept of q line with X-axis

If $x_q = x_f$

$$y_q = \frac{q}{q-1} x_f - \frac{x_f}{q-1}$$

$$y_q = \frac{2x_f - x_f}{q-1} = \frac{x_f (q-1)}{(q-1)}$$

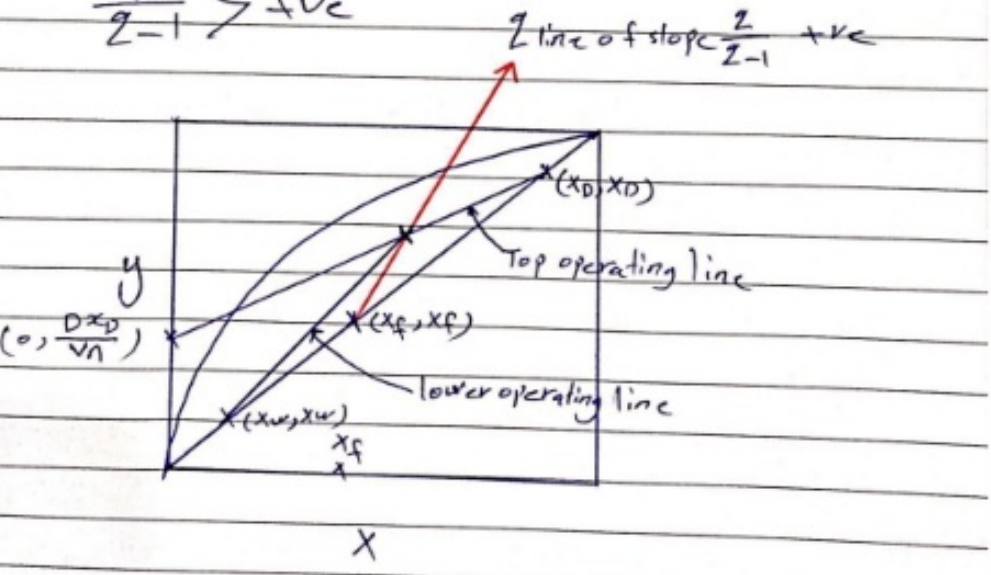
$\therefore y_q = x_f \Rightarrow$ the q line passing through the point (x_f, x_f)

$$q = \frac{\lambda + (Hfs - Hf)}{\lambda}$$

Cases of the feed:

① IF feed is cold i.e $Hf < Hfs \Rightarrow q > 1$ and slope of q line positive +ve

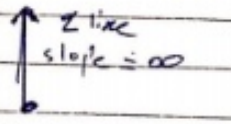
$$\frac{q}{q-1} > +ve$$

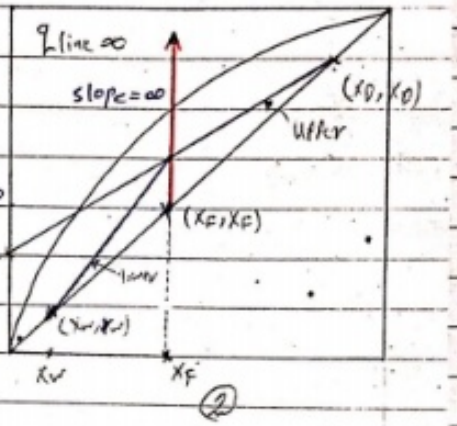


② If feed is saturated liquid (at boiling point)

$H_f = H_{fs} \Rightarrow q = 1$

and slope of q line $= \frac{q}{q-1} = \infty$





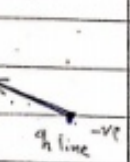
③ If the feed is partly vapor

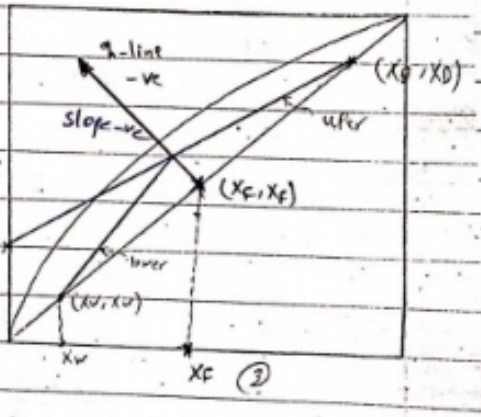
$H_{fs} < H_f < H_{fs} + \lambda$

$0 < q < 1$

(q line $< -ve$)

and slope $\left(\frac{q}{q-1}\right) < -ve$

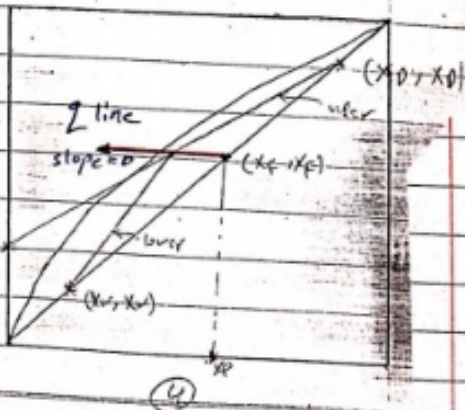




④ If Feed is saturated vapor

$H_f = H_{fs} + \lambda \Rightarrow q = 0$

$\text{slop} = \left(\frac{q}{q-1}\right) = 0$

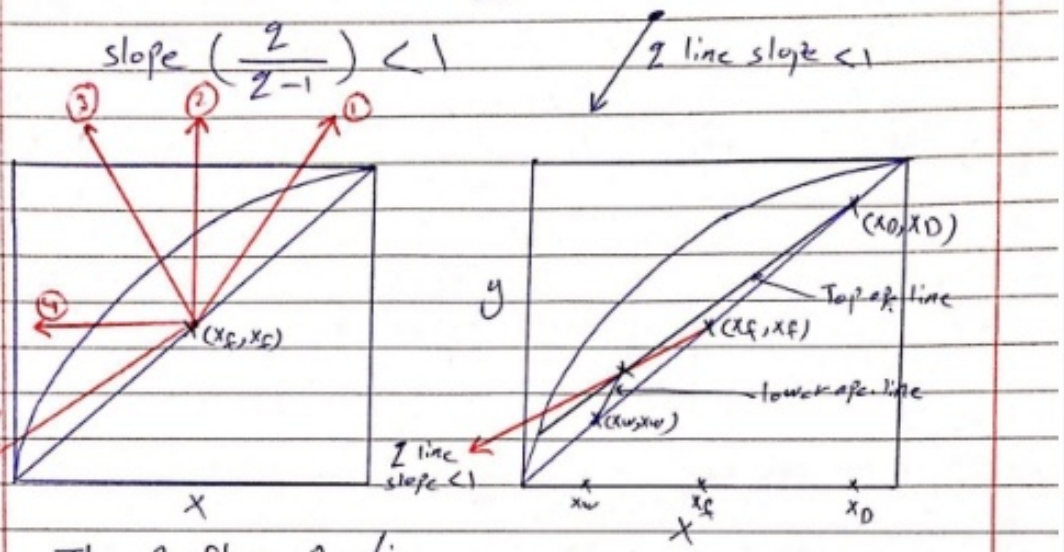


Rama .Co

(5) IF feed is superheated vapor

$H_f > H_{fs} + \lambda \Longrightarrow q < 0$

slope $\left(\frac{q}{q-1}\right) < 1$



The Reflux Ratio

only in the top operating line

$$y_n = \frac{L_n}{V_n} x_{n+1} + \frac{D}{V_n} x_D \quad \text{(Top operating line)}$$

$$R = \frac{L_n}{D} \quad \& \quad V_n = L_n + D$$

$$y_n = \frac{L_n}{L_n + D} x_{n+1} + \frac{D}{L_n + D} x_D$$

$$y_n = \frac{\frac{L_n}{D}}{\frac{L_n}{D} + \frac{D}{D}} x_{n+1} + \frac{\frac{D}{D}}{\frac{L_n}{D} + \frac{D}{D}} x_D$$

$$\boxed{y_n = \frac{R}{R+1} x_{n+1} + \frac{x_D}{R+1}}$$

Top operating line صورة أخرى من معادلة التشغيل

$$y_n = \frac{R}{R+1} x_{n+1} + \frac{x_D}{R+1}$$ Top operating line

Top operating line.

$$q_h = \frac{\lambda + (H_{fs} - H_f)}{\lambda}$$

Ex:- 11.5 A continous fractionating column, operate at atmospheric pressure is to be designed to separate amixture containing 15.67% CS_2 and 84.33% CCl_4 into an over head product containing 91% CS_2 and a waste of 97.3% CCl_4 (all weight present) Assume a plate efficiency of 70% and reflux of 3.16 Kmol per Kmol of product. Using the data below to determine the number of plates required. feed enters at 290 K with a specific heat of 1.7 KJ/Kg K and boiling point of 33.6 K. λ of CS_2 and CCl_4 = 25400 $\frac{KJ}{Kmol}$

M.Wt CCl_4 = 154

CS_2 = 76 ($C = 12$, $Cl = 35.45$, $S = 32$)

y mol% CS_2 in vapor	0	8.23	15.55	26.6	33.2	49.5	63.4	79.7	84.4	87.8
x mol% of CS_2 in liquid	0	2.96	6.15	11.06	14.35	25.85	39.	51.18	66.7	75.75

(cold) تغذية الجزء

$$q_h = \frac{25900 + (14200 - 3832)}{25900} = 1.4$$

q line equation $y_q = \frac{q_h}{q_h - 1} x_q - \frac{x_f}{q_h - 1}$

Slop of q line $= \frac{1.4}{1.4 - 1} = 3.5$

The intercept of q line on the X-axis (i.e $y = o$) $= \frac{XF}{q_h}$

$\frac{XF}{q_h} = \frac{0.274}{1.4} = 0.196$

- The q line is drawn through (X_f, X_f) (0.274, 0.274) and (0.196, 0)

$R = 3.16$

Top operating line $y_n = \frac{R}{R+1} x_{n+1} + \frac{X_D}{R+1}$

intercept $= \frac{X_D}{R+1} = \frac{0.953}{1 + 3.16} = 0.229$

- Top operating line may be drawn through (X_D, X_D) (0.953, 0.953) and (0, 0.229)

The lower op. line is drawn by joining the intersection of the top op. line and the q line with the point (X_w, X_w) (0.053, 0.053)

From the drawing the number of theratical plate = 9

Plate efficiency = 70 %

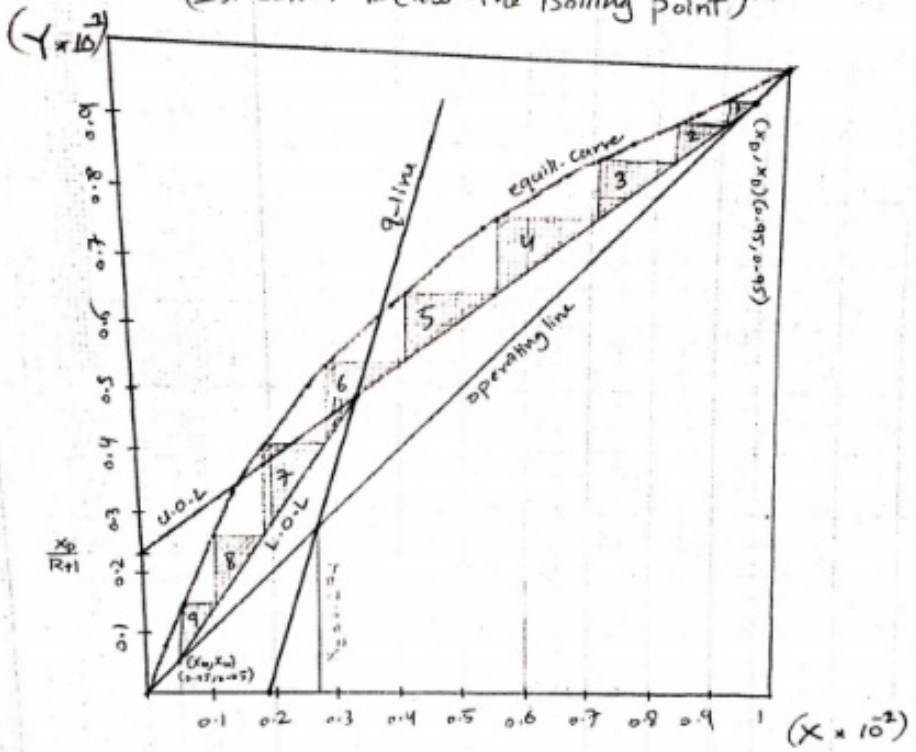
the number of actual plate $= \frac{9}{0.7} = 12.85 = 13$ plate

Rama .Co

Example (11-5):

Show how to draw & calculate the number of stages by q-line equation

(1st case: Below the Boiling point)



- The intersep of q-line with x-axis when (y=0)

$$\frac{x_F}{q} = \frac{0.274}{1.4} = 0.196$$

- we can find the slope of q-line by knowing the slope

Slope $= \frac{q}{q-1} = \frac{1.4}{1.4-1} = 3.5$

——— equilibrium Curve — Upper operating line

— Operating line — Lower operating line

— q-line