الجامعة التكنولوجية

قسم الهندسة الكيمياوية

المرحلة الثالثة

انتقال كتلة

أ.م. د. عامر عزيز

Eqn of q-line	point of q line
(a) $y = \frac{q}{q-1}x - \frac{Z_f}{q-1}$	$(Z_f, Z_f), \left(\frac{Z_f}{q}, 0\right)$
(b) $x = Z_f$	vertical through (Z_f).
(c) $y = -\left(\frac{1-f_v}{f_v}\right)x + \frac{Z_f}{f_v}$	$(Z_f, Z_f), \left(0, \frac{Z_f}{f_v}\right)$.
(d) $y = Z_f$	horiz. Through (Z_f).
(e) $y = \frac{q}{q-1}x - \frac{Z_f}{q-1}$	$(Z_f, Z_f), \left(0, \frac{Z_f}{1-q}\right)$

H.W Try to obtain the above relations.

(a) If the feed is saturated liquid, $q=1$, the slope of the feed line is (∞), and feed line will be vertical line through (Z_f, Z_f).

(b) If the feed is a saturated vapour, $q=0$, the slope of the feed line is zero, and feed line will be horizental line through (Z_f, Z_f).

(c) If the feed is a mixture of liquid and vapor or superheated vap. or subcooled liquid, then the slope of the feed line can be calculated using equation (24), where :-

$$\boxed{H_L = C_{p_s} \cdot M_{av.} \cdot (T - T_o) + \Delta H_s} \quad \text{---- (26)}$$

Enthalpy of liquid.

H_L = molar enthalpy of solution at temp. T, kJ / Kmol.
C_{p_s} = Specific heat of solution, kJ/kg. K.
$M_{av.}$ = average molecular weight of solution.
T_o = ref. Temp.
ΔH_s = heat of solution at ref. Temp. T_o, kJ/Kmol.

$$\boxed{H_V = \sum y_i^* * M_i \left[C_{p_i} (T - T_o) + \lambda_i \right]} \quad \text{---- (27)}$$

Enthalpy of vapour.

H_V = Enthalpy of the vapour, kJ / kmol
λ = heat of vaporization, kJ/kg.

$H_F = \sum$ Latent heat + Sensible heat

$$\boxed{H_F = \sum (x_i * M_i * \lambda_i) + (x_i * C_{p_i} * (T - T_o))} \quad \text{---- (28)}$$

H_F = Enthalpy of the feed.

Ex. (2) :- (How to draw feed line for different feed conditions)

A mixture of benzene and toluene containing 58% mole benzene is to be seperated in a continuous column operating at 1 atm total press. Draw the feed line for the following feed condition :-

ⓐ Saturated vapour. (at dew point).
ⓑ Saturated liquid. (at bubble point).
ⓒ 65 mass % vapour. (partial vaporiza)
ⓓ vapour at 120 °C. (super-heated)
ⓔ liquid at 50 °C. (subcooled)

Given that :-

$Cp_{(B)L} = 146.5$ kJ/kmol.K

$Cp_{(B)V} = 97.6$ kJ/kmol.K

$M_B \lambda_{(B)} = 30,770$ kJ/kmol.

$Cp_{(T)L} = 170$ kJ/kmol.K

$Cp_{(T)V} = 124.3$ kJ/kmol.K

$M_T \lambda_{(T)} = 32,120$ kJ/kmol.

ref. Temp. = 90 °C

Sol. :-

Using eq. (25) to draw feed line.

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

(a) For saturated vapour. $\therefore q = 0$

Slope $= \frac{0}{0-1} \Rightarrow$ horizontal line through point F (Z_f, Z_f).

(b) For saturated liquid. $\therefore q = 1$

slope $= \frac{q}{q-1} = \infty \Rightarrow$ vertical line through point F.

(c) For 65% vap. $\Rightarrow$ $\therefore$ feed contain 35% liquid.

$\therefore q = 0.35 \Rightarrow$ slope $= \frac{0.35}{0.35 - 1} = -0.538$

intercept $= \frac{-Z_f}{q-1} \Rightarrow$ inter. $= \frac{-0.58}{0.35 - 1} = 0.892$

We have (F) point and intercept.

(d) For vapour at 120°C $\Rightarrow$ super heated vapour.

$\therefore$ Eq. (24) is to be used.

$$q = (H_v - H_f) / (H_v - H_L)$$

$\Delta H_s = 0$ (pure component)

$H_L = Cp_{sol.} \cdot M_{av} (T - T_o) + \cancel{\Delta H_s}^{\,0}$

For the solution at its bubble pt ($T = 90\,^\circ C$), take $T_o = 90$: then :-

$\boxed{H_L = 0}$ for $\Delta T = 0$

(*) Enthalpy of vapour :- for $x = 0.58$, $y^* = 0.78$ (من الرسم)

$$H_V = y^*_B \cdot M_B [Cp_B (\cancel{T - T_o})^{0} + \lambda_B] + y^*_T M_T [\cancel{Cp_T (T - T_o)}^{0} + \lambda_T]$$

$H_V = 0.78 * 30770 + (1 - 0.78) * 32120$

$H_V = 31{,}067$ KJ/kmol.

(*) Feed enthalpy :-

We assume that 0.58 kmol benzene and 0.42 kmol of toluene are vaporized seperately at $90\,^\circ C$, the vapours are heated to $120\,^\circ C$ and then mixed

$$H_F = 0.58 * 30770 + 0.58 * 97.6 (120 - 90)] + [0.42 * 32120 + 0.42 (124.3)(120 - 90)]$$

$H_F = 34601$

Now :- $q = \dfrac{H_V - H_F}{H_V - H_L}$

$$q = \frac{31067 - 34601}{31067 - 0} = -0.114$$

$$\text{intercept} = \frac{-Z_f}{q-1} = \frac{-0.58}{-0.114-1} = 0.521$$

② The Feed is subcooled at 50°C

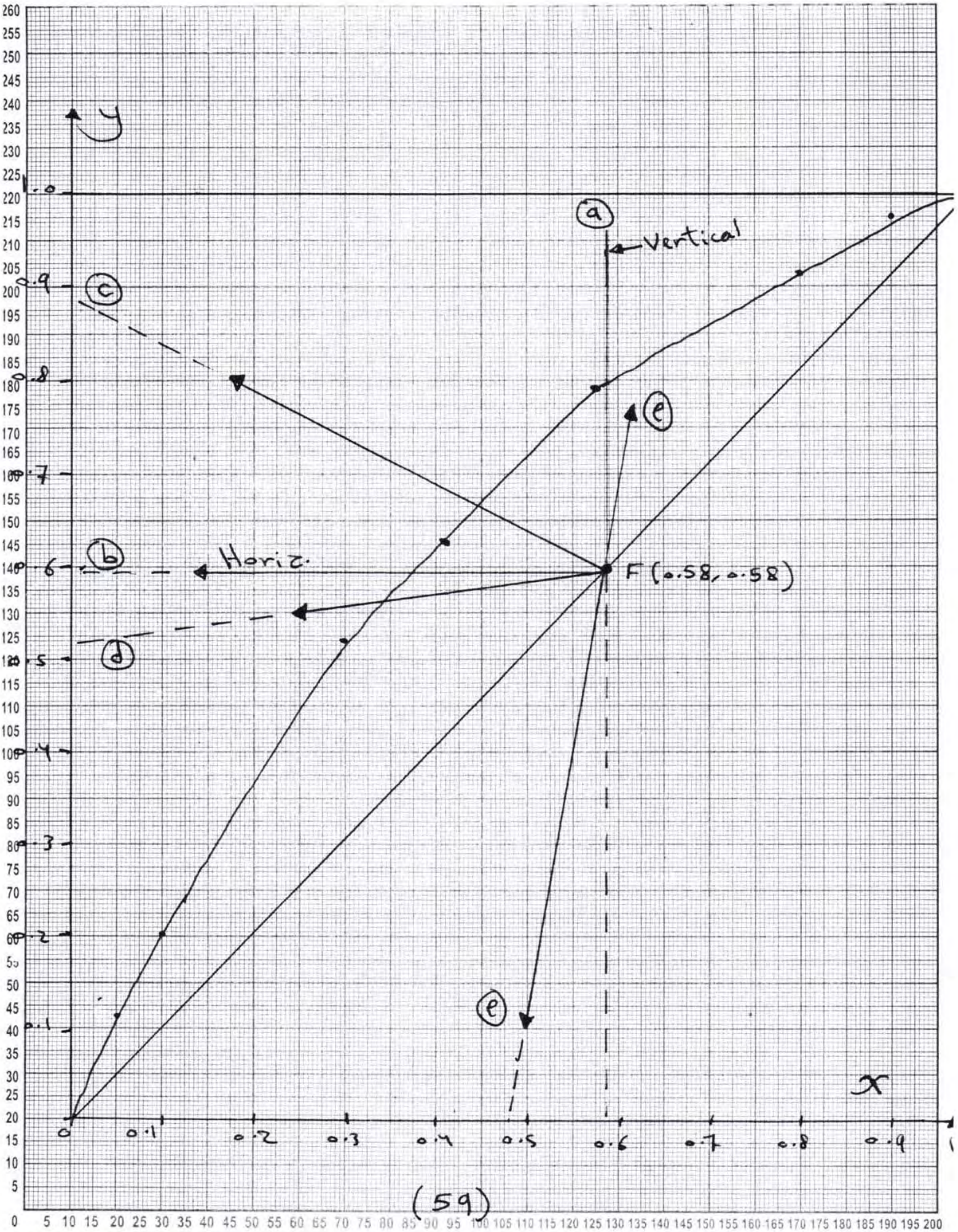
We assume that 0.58 kmol of (B) and (0.42) kmol of (T) are cooled from ref. Temp. ($T_o = 90°C$) to ($T = 50°C$) and then mixed to get 1 kmol of feed.

$$\therefore H_F = 0.58 * 146.5 * (50-90) + 0.42 * 170 (50-90)$$
$$H_F = -6240 \text{ kJ/kmol.}$$

$$\therefore q = \frac{31067 - (-6240)}{31067 - 0} = 1.2$$

$$\text{intercept with x-axis} = \frac{Z_f}{q} = \frac{0.58}{1.2} = 0.484$$

Note:- ① For a liq. feed, it is introduced just above the feed tray. But a vapour feed is introduced just below it.
② If the feed is a mixture of liq. and vap., it is desirable that it is seperated into the vap., and the liq. phases first. The liq. part should enter the column just above feed tray, and the vap. part just below it. However, this is not always done in practice, and a mixed feed is often introduced as a whole over the feed tray.


y
x
Vertical
Horiz.
F (0.58, 0.58)
a
b
c
d
e
e
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0
0.1
0.2
0.3
0.4
0.5
0.6
0.7
0.8
0.9
1

Ex. (3) :- Determination of No. of ideal trays.

A stream of aqueous methanol having 45 mol % CH_3OH is to be seperated into a top product having 96 mol % methanol and a bottom liquid with 4 % methanol. The feed is at its bubble point and the operating pressure is 101.3 kPa. A reflux ratio of 1.5 is used.

(a) Determine the No. of ideal trays.

(b) Find the No. of real trays if overall tray eff. is 40 %. On which real tray should the feed introduced.

The equilib^m. Data for the system.

x:	0	0.02	0.04	0.06	0.08	0.1	0.2	0.3	-----	0.7	0.8	0.9	1
y:	0	0.134	0.23	0.304	0.365	0.41	0.58	0.66	---	0.87	0.92	0.96	1
T:	100	96.4	93.5	91.2	89.3	87.7	84.4	78.0	----	69.3	67.6	66	64

Solution :- $x_D = 0.96$, $x_w = 0.04$, $Z_f = 0.45$

Plot equil^m. data, the points $D(0.96, 0.96)$, $W(0.04, 0.04)$ and $F(0.45, 0.45)$ are located on the $x-y$ plane.

(a) The upper operating line is drawn through the point (D) with the intercept with the y-axis

$$\text{Intercept} = \frac{x_D}{R+1} = \frac{0.96}{1.5+1} = 0.384$$

∵ Feed is a saturated liquid, then feed line is vertical line through point (F).

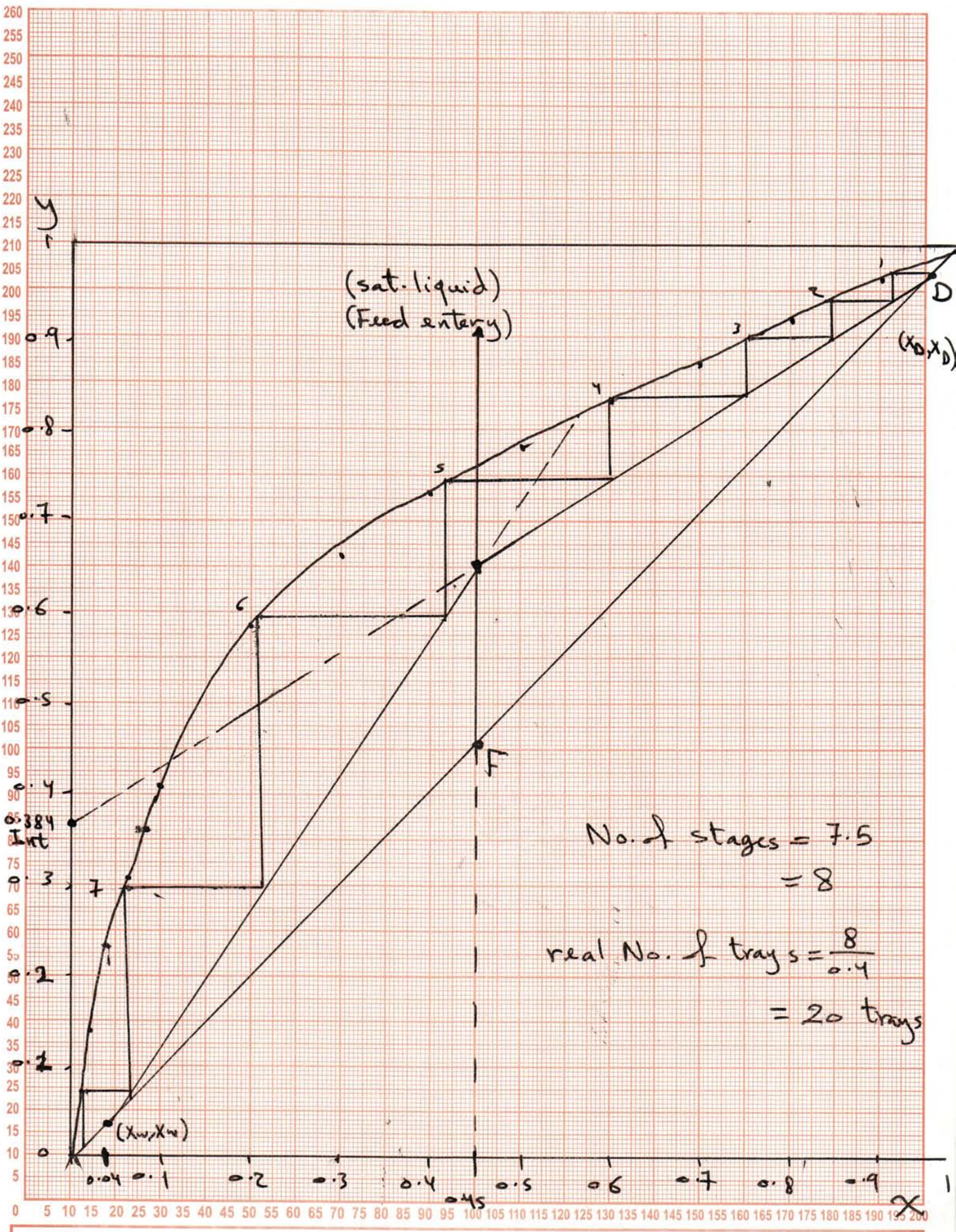
* The upper operating line meets the feed line at point (M).

* The point (M) and (W) are joined to get the lower or stripping section operating line.

(b) No. of ideal plates $= 7.5$ From the figure
$\simeq 8$

No. of real plates $= 8 / 0.4$
$= 20$ trays plus the reboiler which is assumed to act like an ideal stage.

* Feed enters in a place between (4 - 5)



No. of stages = 7.5
= 8

real No. of trays = $\frac{8}{0.4}$
= 20 trays

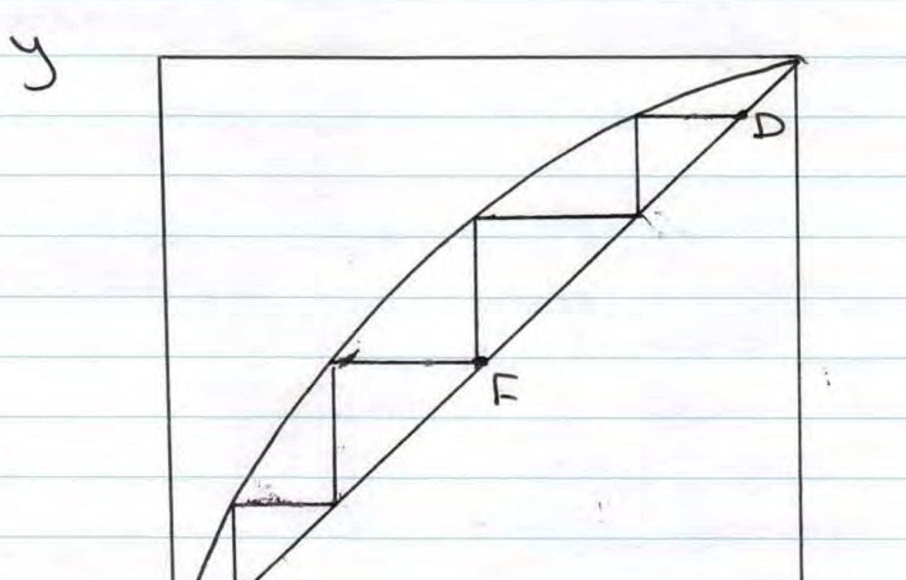
Analytical Determination of The No. of Ideal Stages.

ⓐ Total Reflux ⟹ Give (N_{min})

∴ $R = \frac{L_o}{D}$ and at total (R) where no distillate is accumulated then $D = 0$

∴ $R = \infty$ and No. product is drawn from reboiler

At total reflux, the slope of the rectifying section operating line is $\left(\frac{R}{R+1}\right)$ for $R \longrightarrow \infty$. The slope of the line is unity and pass through the point (x_D, x_D) on the diagonal. Therefore, the operating line coincides with the diagonal. So does the stripping section line. The No. of ideal stages is obtained by stair case construction between equil^m line and the diagonal. Total reflux is very often used during the startup of the column till steady state condition reached.



No. of plates at total R.
(min^m. No. of plates)

Fenske's equation :-

This equation can be used to calculate theoretically the "Minimum. No. of Trays" if the relative volatility remains constant ($\alpha_{avg.}$)

$$y_n = \frac{R}{R+1} x_{n+1} + \frac{x_D}{R+1} \qquad \text{at } R = \infty$$

$\therefore y_n = x_{n+1}$ for rectifying section

Same $y_m = x_{m+1}$, stripping section

$$\alpha = \frac{y_A / x_A}{y_B / x_B} \Longrightarrow \boxed{\frac{y_{AW}}{y_{BW}} = \alpha_o \left(\frac{x_{AW}}{x_{BW}} \right)}$$

vap. rise from the reboiler (y_{AW}) have a relation with liquid fall in stage one, joining in operating line.

$$x_{A_1} = y_{AW} \quad , \quad x_{B_1} = y_{BW}$$

$$\frac{x_{A1}}{x_{B_1}} = \frac{y_{AW}}{y_{BW}}$$

$$\boxed{\frac{x_{A_1}}{x_{B_1}} = \alpha_o \left(\frac{x_{AW}}{x_{BW}} \right)}$$

Same for stage one :-

$$\alpha_1 = \frac{y_{A_1} / x_{A_1}}{y_{B_1} / x_{B_1}}$$

$$\boxed{\frac{x_{A2}}{x_{B2}} = \alpha_0 \alpha_1 \left(\frac{x_{Aw}}{x_{Bw}} \right)}$$

continue to stage (n)

$$\left(\frac{x_A}{x_B} \right)_D = \underbrace{\alpha_0 \alpha_1 \alpha_2 \text{----} \alpha_n}_{} \left(\frac{x_A}{x_B} \right)_w$$

$\hookrightarrow \alpha_{avg.}$ = avg. volatility of The M.V.C

$$\boxed{N_{min.} = \frac{\mathrm{Log}\left[\frac{x_D (1 - x_w)}{x_w (1 - x_D)} \right]}{\mathrm{Log}\, \alpha_{avg.}} - 1}$$

Fenske's equation (at total Reflux)

(b) Minimum Reflux Ratio (R_m) for (N_∞)

(1) Graphically :-

Can be defined as that ratio at which an infinte No. of stages are needed to obtain the desired over head and bottom products.

The determination of (R_m) is based on identifying the "pinch point".

For a particular reflux ratio (R_1), DE_1 is the enriching section operating line with slope $= (R/R+1)$. It intersects the feed line at point (M_1), WM_1 is the stripping section operating line.

As the reflux ratio (R_1) decreases to (R_2) the slope of U.O.L decreases, but intercept increases, the point (E_1) moves to (E_2). DM_2 is the U.O.L and WM_2 is the S.O.L, and they intersect the feed line at M_2. Then the driving force is less and the No. of theo. trays will be more.

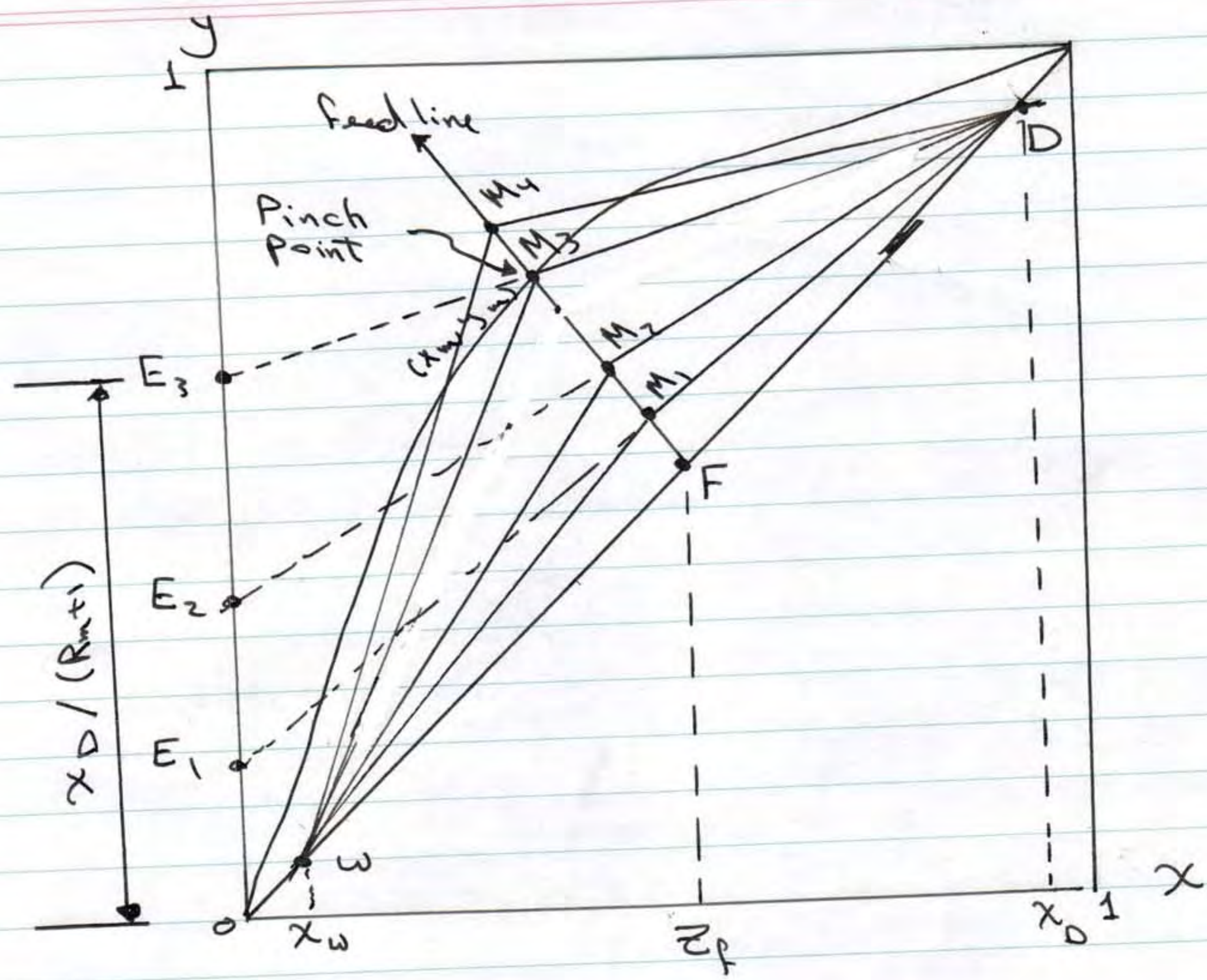
If the reflux ratio is gradually reduced, to (E_3), then line DE_3 intersect the feed line at (M_3) which lies on the equilibrium curve.

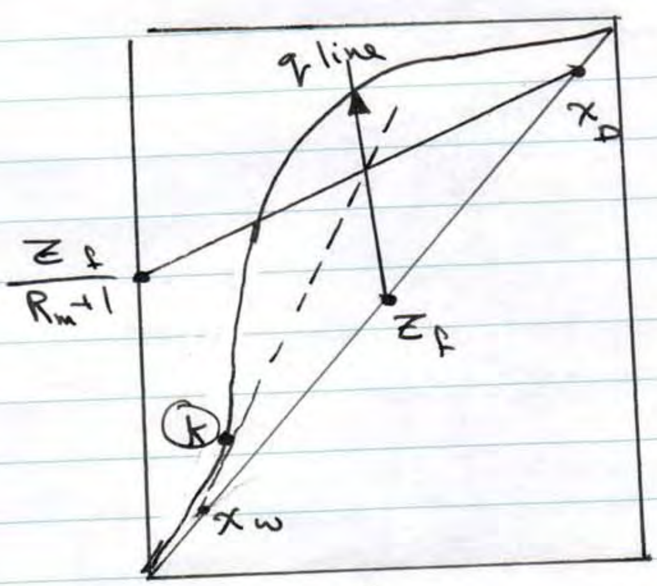
Then driving force will be zero at M_3, it is the "pinch point".

The No. of Theo. plates will be infinite.

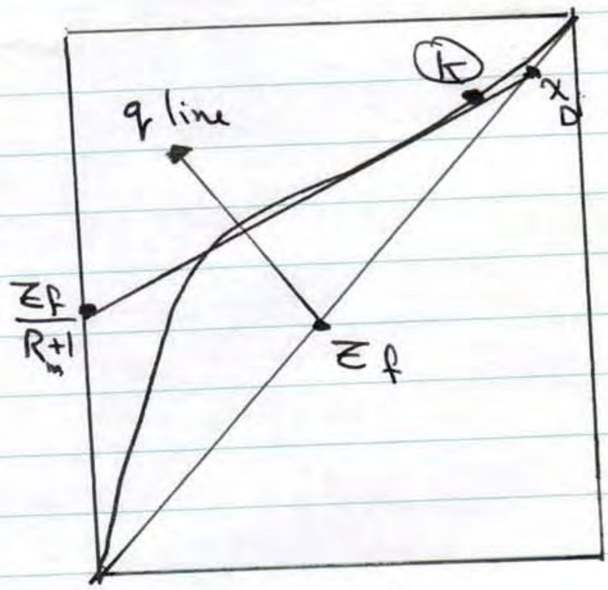
The operating line (DM_3) corresponds to the min. reflux ratio.

For further reduced in (R) the operating line will intersect the feed line at a point above the equil^m. Curve, this is impossible.





Pinch Point for stripping section



Pinch Point for Rectifying section