



Separation 1

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Separation processes → operations which transform a mixture of substances into two or more products which differ from each other in composition.

* Heat and mass transfer are examples on separation processes.

- mixing → spontaneous process
- separation → non-spontaneous process

Reasons of separation?

- ① Main goal is purification
- ② preparation of raw materials into their component
- ③ remove toxins or inactive components from solution "drugs"
- ④ Needs for ultra-pure samples for testing or analysis

separations includes: Enrichment, purification, isolation, concentration, Refining

Note! A typical chemical plant is (reactor) surrounded by separators. → heart of the process!
[50 - 90 %] are separators!

* All natural processes take place to increase the entropy *

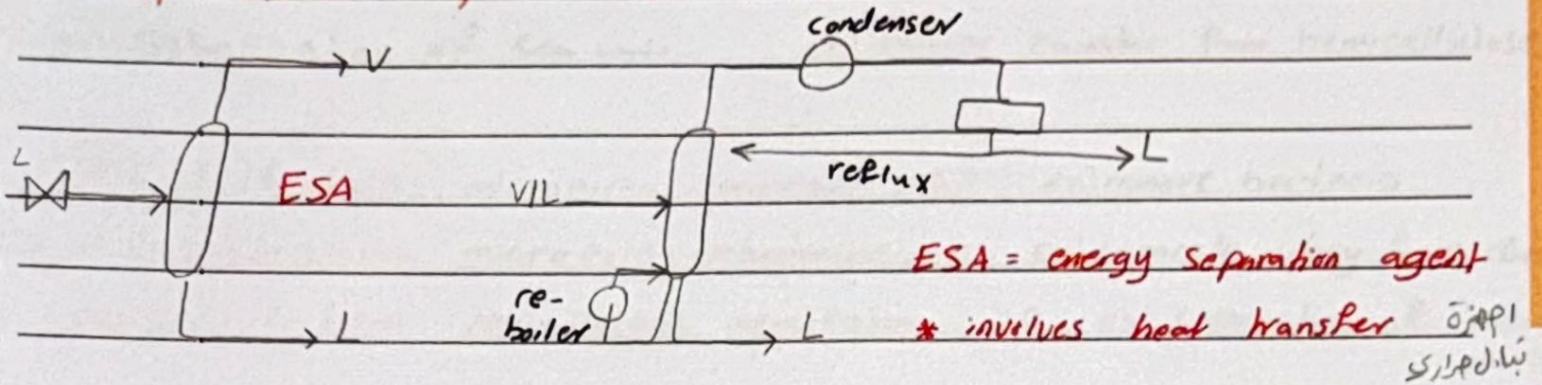
* To separate a mixture we must supply the equivalent energy (heat or work) *

Note! When the volatility difference small ∴ more trays are required!

Separation Techniques 8

1 Separation by phase creation

Evaporation, Distillation



Drying ex) remove H_2O from PVC, ESA

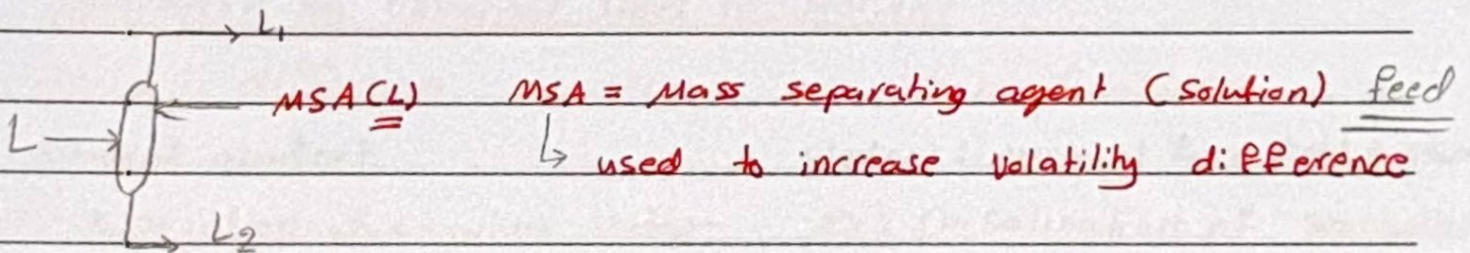
ESA Crystallization ex) crystallization of high pure silicon for semiconductor

ESA Desublimation ex) recovery of phthalic anhydride from noncondensable gas

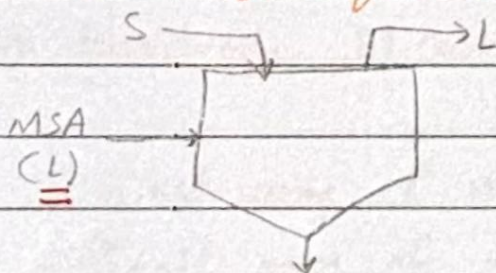
Humidification "for cooling" = cooling tower

2 Separation by phase addition

liquid-liquid extraction



leaching (liquid-solid extraction), Foam fractionation $MSA(C)$



Absorption $MSA(L)$, stripping $MSA(C)$



Extractive distillation :

[MSA + ESA]



example on absorption:

Separation CO_2 from
combustion product using
ethanolamine

example on stripping:

removal of light ends from
naphtha, kerosene, gas oil

3 Separation by barrier

Reverse osmosis , Dialysis

nonporous membrane , ΔP

porous membrane , ΔP

ex) Desalination of sea water

ex) recover caustic from hemicellulose

Microfiltration microporous membrane , ΔP , ex) remove bacteria

Ultrafiltration microporous membrane , ΔP , ex) separate whey from cheese

pervaporation Non-porous membrane , ΔP , ex) separation of azeotropic mixture

4 separation by solid agent

Adsorption , chromatography , ion exchange

ex) purification of

ex) separation of xylene

ex) removing calcium ion

p-xylene

and ethylbenzene

from water

solid adsorbent

solid / liquid adsorbent

resin with ion active sites

on solid support

5 separation by external field or gradient

Thermal diffusion ,

Electrodialysis

thermal gradient

gradient: electrical force field + membrane

ex) separation of uranium isotopes

ex) Desalination of seawater

Factors of choice separation processes:

a) Feasibility b) cost c) product value and quality d) selectivity

Easiest way of separation $\rightarrow$ physical separation, ex) filtration

difficult way of separation $\rightarrow$ membranes



Distillation depends on:

- 1) difference in boiling points
- 2) distributions of the components between liquid and gas phase in the mixture
- 3) vapor pressure characteristics

How vapor pressure created? by supplying heat as separating agent

* Distillation is the most widely used method for separating liquid mixtures *

Notes on distillation:

- it is a very energy intensive technique
- mostly carried out in multi tray columns
- packed column used in distillation

Equilibrium stage concept → Streams leaving a stage are in equilibrium.

[each stage has equilibrium]

phase equilibrium → the rate of vaporizing is equal to the rate of condensing. * the composition of L and V are not equal $y_A \neq x_A$ *

Note! if the compositions were equal, no separation could be achieved in any equilibrium process.

Types of equilibrium :-

Thermal $T^{\alpha} = T^{\beta}$

Mechanical $P^{\alpha} = P^{\beta}$

Phase $\mu_i^v = \mu_i^l$
↳ chemical potential

$$P_T = P_A + P_B$$

Note! Henry's law for gas-liquid equilibrium



* if temperature and pressure are constant $\therefore$ equal rates of vaporization and condensation require a minimum in free energy of the system *

Two limitations of separation:

- ① equilibrium (needs trays) ② Second law of thermodynamics

VLE: Raoult's law

when we can use Raoult's law? low pressure + same intermolecular forces

$\hookrightarrow \therefore$ we can treat vapor $\rightarrow$ ideal gas

liquid $\rightarrow$ ideal solution

$$y_i P = x_i P_i^{\text{sat}}$$

Raoult's law

$$* P_i = x_i P_i^{\text{sat}}$$

$$* P_i = y_i P$$

$[y_i P = \text{partial pressure}]$
 $[P_i]$

Note: Same intermolecular forces means species are chemically similar

$$y_A P = x_A P_A^{\text{sat}}$$

$$y_B P = x_B P_B^{\text{sat}}$$

$$((x_A + x_B = 1))$$

$$P = x_A P_A^{\text{sat}} + (1 - x_A) P_B^{\text{sat}}$$

$$((y_A + y_B = 1))$$

$\therefore$ at equilibrium

$$y_A = \frac{x_A P_A^{\text{sat}}}{x_A P_A^{\text{sat}} + (1 - x_A) P_B^{\text{sat}}}$$

$$x_A P_A^{\text{sat}} + (1 - x_A) P_B^{\text{sat}}$$

Limitation of Raoult's law?

applied only to species of known vapor pressure means a subcritical species

Vapor pressure → pressure exerted by the vapor, when dynamic equilibrium between V and L.

[Vapor pressure $\propto$ temperature] * each substance has its own vapor pressure *

$$\ln \frac{P_2^{sat}}{P_1^{sat}} = \ominus \frac{\Delta H_{vap}}{R} \left[\frac{1}{T_2} - \frac{1}{T_1} \right] \quad \text{Clausius - clapeyron}$$

$$\ln P^{sat} = A - \frac{B}{T+C} \quad \text{Antoine equation}$$

Note! Antoine equation not good near critical point.

Bubble point T_{bp} → Temp at which liquid molecules have just enough kinetic energy to escape to gas.

Dew point T_{dp} → Temp at which vapor molecule has just enough kinetic energy to condense.

* Saturated liquid : liquid at bubble point *

* Saturated vapor : vapor at it's dew point *

Azeotrope → a constant boiling mixture, mixture behaves like a single component.

Volatility → degree to which it tend to transfer from liquid to vapor state. = liquids with

$$\text{Volatility} = \frac{P_a}{x_a} \quad \begin{array}{l} \rightarrow \text{partial pressure} \\ \rightarrow \text{mole fraction} \end{array}$$



Liquid : $T = T^{\text{sat}}$ $\therefore$ saturated liquid
 $T < T^{\text{sat}}$ $\therefore$ subcooled liquid

vapor : $T = T^{\text{sat}}$ $\therefore$ saturated vapor
 $T > T^{\text{sat}}$ $\therefore$ superheated vapor

Example. 26.1.18- Vapor + liquid composition?

$T = 95^\circ\text{C} = 368.2\text{ K}$ benzene and toluene

$P = 101.32\text{ kPa}$

$P_{\text{benzene}} = 155.7\text{ kPa}$

$P_{\text{toluene}} = 63.3\text{ kPa} = 9.4$

$$[P = P_{\text{benzene}} X_{\text{benzene}} + P_{\text{toluene}} X_{\text{toluene}}]$$

$$101.32\text{ kPa} = 155.7 X_b\text{ kPa} + 63.3 (1 - X_b)\text{ kPa}$$

$$101.32 = 155.7 X_b + 63.3 - 63.3 X_b$$

$$X_b = 0.4115$$

$$y_b = \frac{P_b X_b}{P} = 0.632$$



Relative Volatility :-

$$\alpha_{AB} = \frac{y_A/x_A}{y_B/x_B} = \frac{P_A^{\text{sat}}}{P_B^{\text{sat}}} = \frac{P_A/x_A}{P_B/x_B}$$

$$\left(y_A = \frac{P_A^{\text{sat}}}{P} \cdot X_A \right) \quad \left(P_A = P_A^{\text{sat}} \cdot X_A \right)$$

* High α_{AB} : the separation easier to occur by distillation

$$y_a = \frac{\alpha_{ab} x_a}{1 + (\alpha_{ab} - 1) x_a}$$

- ① if $\alpha = 1$: Separation is not possible
- ② if α close to 1 : separation is difficult
- ③ if α is large : Separation is easy

b.p $\downarrow$: Volatility $\uparrow$

• To get lever rule equations : $F = L + V$ and $Fz = Lx + Vy$

Note 1, pure liquids have boiling point
mixtures have boiling range = difference between T_{bp} and T_{dp}

* Distillation column encourages repeated partial vaporisation and condensation.

* going from bottom to top in the distillation : T becomes lower

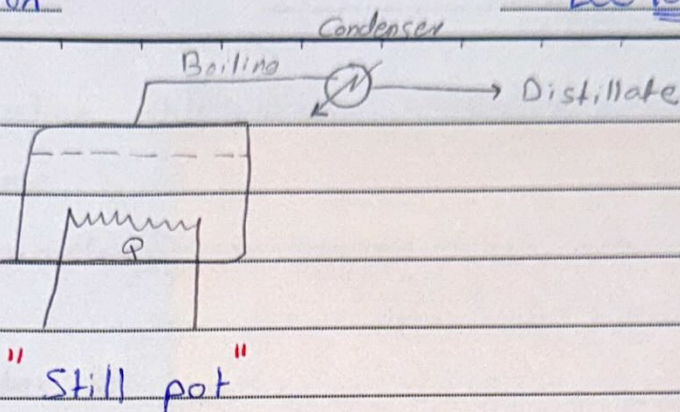
$\uparrow$ T , lower

Continuous Distillation

Lec 10

pros [advantage]:

Simple to make and
operate + Cheap Component
+ Flexible

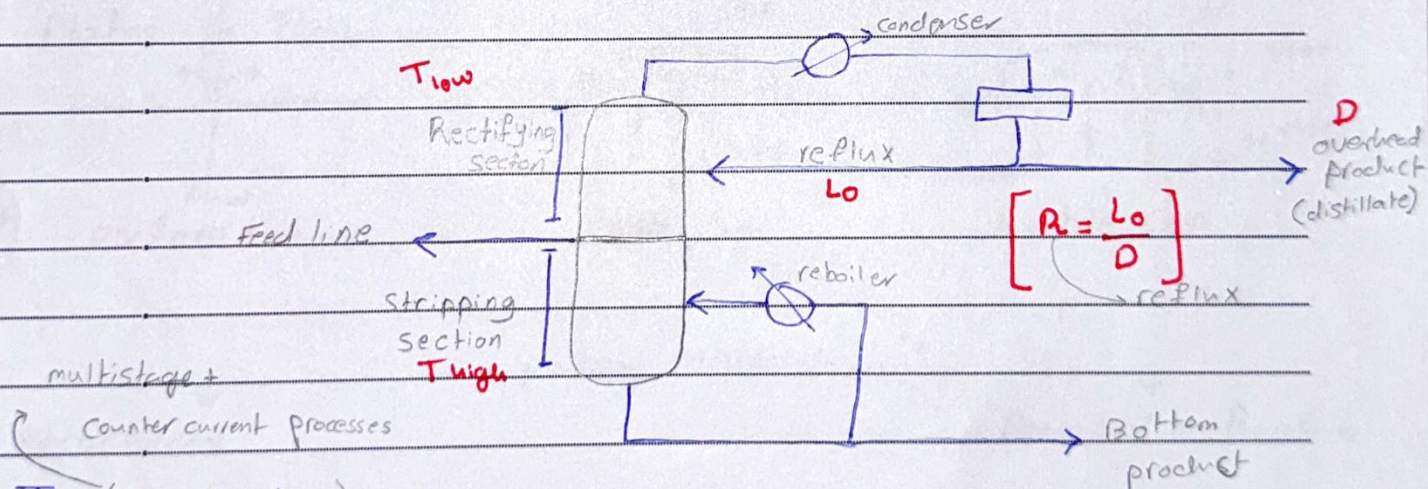


cons [disadvantage]:

Small amount of high purity + large amount of low purity
+ Energy intensive + Slow

Note! boiling and condensing results in change of composition.

Note! boiling and condensing don't just occur in reboiler and condenser, they happen inside Distillation column.



The (Distillation) column is -

- Fractionation / stage distillation with reflux = series of flash vaporization stages and vapor-liquid products from each stage flow counter current to each other.
- The more volatile component increased in the vapor and decreased in the liquid.

* each stage has equilibrium *

assumptions of distillation column :

- 1 ideal stage Streams have good time , good contact between L & V
- 2 constant molel [equimolar]

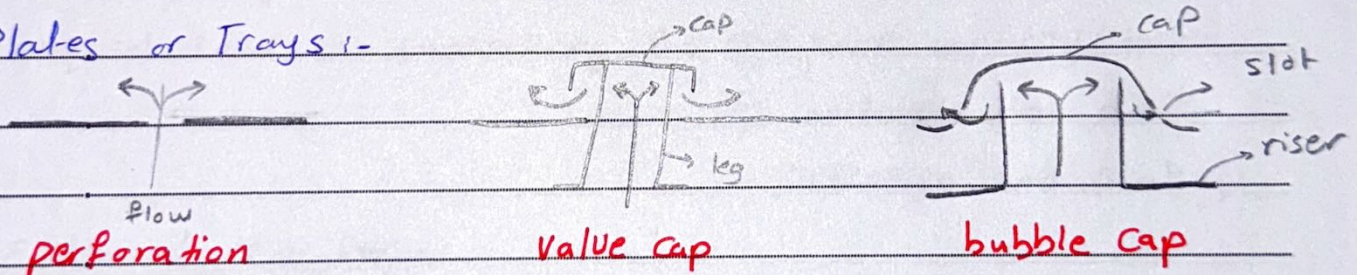
Stages = sieve plates = trays

Note! all reboilers are partial !!

Note! downcomer = liquid level ~ 15 + ~ 15 cm on each stage

Note! The mass transfer will happen only on the surface of the tray when there is no packed material.

Plates or Trays :-

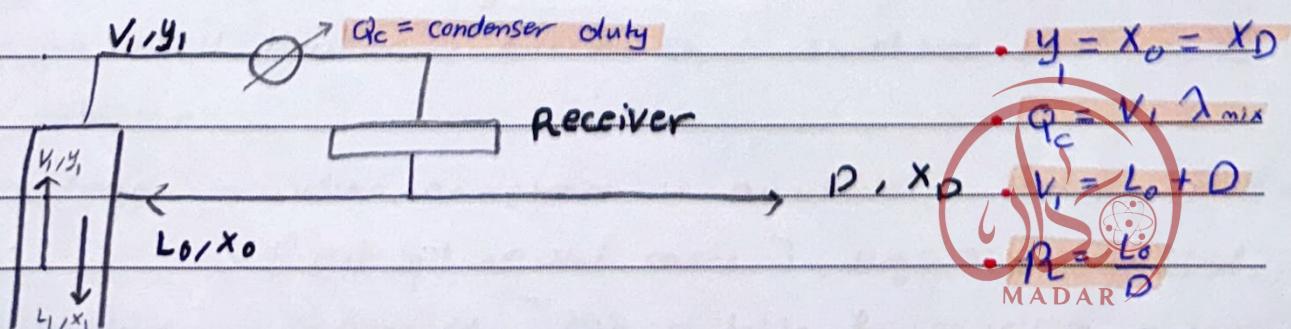


packing materials

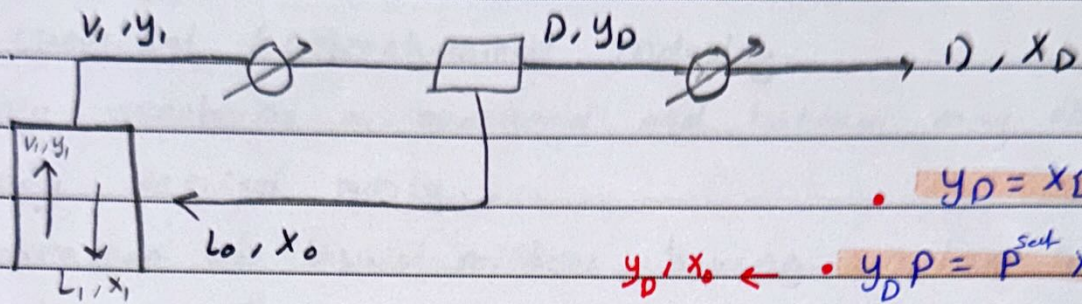
Structured Dumped or Random

ex) Nutter ring , wire mesh packing

Total condenser :- tray 1



partial condenser :- tray يعتبر acts as one plate



• $y_D = x_D$

• $y_D p = p^{sat} x_D$
equilibrium

• $V, y_1 = D, x_D + L, x_0$

Reboilers :-

1] Internal reboiler.

- * Tubular heat exchanger built into the bottom.
- * Cleaning requires a shut down of distillation operation.
- * Liquid level is controlling for ① Small capacity tower
② materials that don't
corrosive and scaling

2] External reboiler.

A- Thermosiphon reboiler.

- * Vertical
- * Heating medium outside the tubes

B- Kettle reboiler.

- * heating medium inside the tubes
- * large size
- * better control
- * long residence time
- * Not used for thermally sensitive materials

☛ pipe still heater / Furnace : sometimes used inside the reboiler.

Advantage → when no steam is available, Heating media is [hot oil or hot gases], High T_v is needed.

Disadvantage → Expensive, Not suitable for sensitive materials.

Applications of Distillation :-

- 1) most widely used large scale for separation fluid mixtures in chemical / petrochemical industry.
- 2) if No azeotropes = overhead and bottom may obtained in any desired purity.
- 3) Separation of liquid mixtures having similar boiling points at (low relative volatility but $\alpha > 1.05$)

Equipment of Distillation :-

Tray columns [stagewise contact]

Packed columns [continuous contact on the surface of packing material]

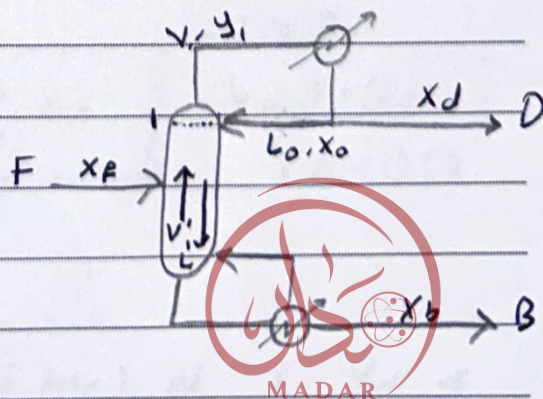
McCabe - Thiele Method : mathematical - graphical method uses material balance for determining the number of theoretical stages or trays needed for a given separation of binary mixture

assumptions :-

- * equal and constant latent heats λ
- * well insulated
- * pressure is constant = no pressure drop
- * constant molar flow rate
- * K_E, P_E are negligible

Column mass balance :-

- $L' = L_0 + qF$
- $V = V' + (1-q)F$
- $F = D + B$
- $FX_F = DX_d + BX_b$
- $V = L + D$



Activity 8 - $x_F = 0.4$ Acetic Acid $R = 3$

$x_d = 0.9$ " "

$x_b = 0.1$ " "

The feed is heating to boiling point $\Rightarrow$ Sat. liquid, $q=1$

$D=?$ $B=?$ $L=?$ $V=?$ $L'=?$ $V'=?$

$$F = 100 \text{ mol/time} \quad F = D + B \rightarrow 100 = D + B \dots (1)$$

$$F x_F = D x_d + B x_b \rightarrow 40 = 0.9D + 0.1B \dots (2)$$

$$[D = 37.5 \text{ mol/time}] \quad [B = 62.5 \text{ mol/time}]$$

$$R = \frac{L_0}{D} \rightarrow [L_0 = 112.5] \quad L' = L_0 + qF$$

$$[L' = 212.5]$$

$$V' = V - (1-q)F \quad [V = L_0 + D = 150]$$

$$[V' = 150]$$

* بالواقع فاعني constant molar flow في كل ال stages

The Rectifying section 8 - "enriching section"

mass balance: vapor = liquid + distillate

$$V_{n+1} y_{n+1} = L_n x_n + D x_d$$

$\Rightarrow$ because I assume constant flow $\therefore V_{n+1} = V$ and $L_n = L$

$$y_{n+1} = \frac{L}{V} x_n + \frac{D}{V} x_d$$

$$\bullet L = RD$$

$$\bullet V = L + D =$$

$$(R+1)D$$

$$\frac{L}{V} = \frac{RD}{(R+1)D} = \frac{R}{R+1}$$

* it intersects the $y=x$ line (diagonal line) at $x = x_d$ *

Continue the Activity 8 - operating lines!
rectifying operating line:

$$D = 37.5$$

$$V = 150$$

$$y_{n+1} = \frac{L}{V} x_n + \frac{D}{V} x_d$$

$$\frac{L}{V} = \frac{R}{R+1} = \frac{3}{4} \quad \text{and} \quad x_d = 0.9$$

$$y_{n+1} = \frac{3}{4} x_n + 0.225 \quad \times$$

stripping operating line:

$$y_m = \frac{L'}{V'} x_{m+1} - \frac{B}{V'} x_b$$

$$x_b = 0.1$$

$$\frac{L'}{V'} = \frac{212.5}{150} = 1.42$$

$$B = F - D = 100 - 37.5 = 62.5$$

$$y_m = 1.42 x_{m+1} - 0.042 \quad \times$$

* Between stages : operating line

$$y_1 = \frac{L}{V} x_0 + \frac{D}{V} x_d$$

$$y_2 = \frac{L}{V} x_1 + \frac{D}{V} x_d \quad \text{and so on...}$$

* On stages : equilibrium

$$y_1, x_1 = \text{equilibrium}$$

$$x_2, y_2 = \text{equilibrium} \quad \text{and so on...}$$

For total condenser $\Rightarrow x_0 = x_d$ so $y_1 = x_d$

Note. If we draw the stages between operating line and equilibrium line.

The Stripping Section :-

$$L'_{m+1} x_{m+1} = V'_m y_m + B x_b$$

because we assume constant flow

$$L'_{m+1} = L' \quad \text{and} \quad V'_m \rightarrow V'$$

$$y_m = \frac{L'}{V'} x_{m+1} - \frac{B}{V'} x_b$$

Boilup ratio $\rightarrow$ ratio of vapor returned to the column to the bottoms product.

$$\bullet \quad \underline{V_B = \frac{V'}{B}}$$

$$\bullet \quad \underline{L' = V' + B}$$

* it intersects the $y=x$ line at $x = x_b$ *

Feed stage consideration 8-

to find the intersection of rectifying and stripping lines:-

① ... $V'_m y = L'_m x - Bx_b$ bottom part

② ... $V_n y = L_n x + Dx_d$ upper part

1 - 2 $\Rightarrow V'_m y - V_n y = L'_m x - L_n x - Bx_b - Dx_d$

so $\rightarrow (V'_m - V_n) y = (L'_m - L_n) x - \underbrace{(Bx_b + Dx_d)}_{Fx_F}$

$\rightarrow \left[y \frac{(V'_m - V_n)}{F} = \frac{(L'_m - L_n)}{F} x - x_F \right] *$

* y and x = points of intersection *

to find q-line equation:-

material balance $\rightarrow F + L_n + V'_m = V_n + L'_m$

$F + (V'_m - V_n) = (L'_m - L_n)$

$\div F$

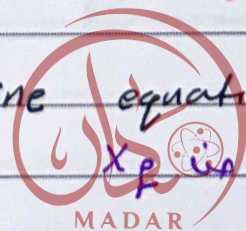
$1 + \frac{(V'_m - V_n)}{F} = \frac{(L'_m - L_n)}{F} = q$

$\therefore \left[q = \frac{(L'_m - L_n)}{F} \text{ and } q = \frac{(V'_m - V_n)}{F} \right] *$

substitution these two equations * $\rightarrow$

$y_q = \frac{q}{q-1} x_q - \frac{x_F}{q-1}$

the q-line equation



* The Slope of the q -line depends on the feed condition *

The feed could be:

- 1 Subcooled liquid
- 2 Bubble point liquid = Boiling point $\leftarrow$ = sat. liquid
- 3 partially Vaporised feed
- 4 Dew point vapor = sat. vapor
- 5 Superheated vapor

$$q = \frac{H_v - H_F}{H_v - H_L}$$

H_v = enthalpy of feed at dew point

H_F = enthalpy of feed at it's condition

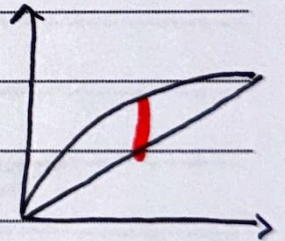
H_L = enthalpy of feed at B.P

1 Saturated liquid feed.

$$(h_F = h_L) \quad \therefore q = \frac{H_v - h_L}{H_v - h_L} = \frac{\lambda}{\lambda} = 1$$

$$\therefore \text{slope} = \frac{q}{q-1} = \frac{1}{1-1} = \infty$$

then q -line is a vertical line

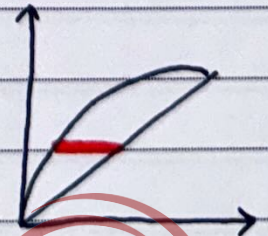


2 Saturated vapor feed.

$$(h_F = h_v) \quad \therefore q = \frac{H_v - h_v}{H_v - h_L} = \frac{0}{\lambda} = 0$$

$$\therefore \text{slope} = \frac{q}{q-1} = \frac{0}{0-1} = 0$$

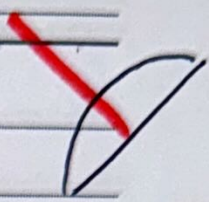
then q -line is a horizontal line



[3] Partial vaporized Feed.

$$(0 < q < 1)$$

q-line has -ve slope



[4] Subcooled feed.

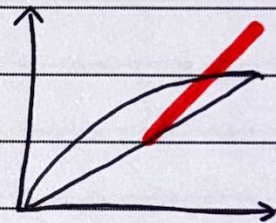
$$(h_f < h_v)$$

$$(q > 1)$$

$$\therefore q = 1 + \frac{C_{PL}(T_b - T_F)}{\lambda}$$

C_P : heat capacity of liquid
 T_F : Feed temp
 T_b : bubble point

q-line in the first quarter



[5] Superheated vapor feed.

$$(h_f > h_v)$$

$$(q < 1)$$

$$\therefore q = - \frac{C_{PV}(T_F - T_d)}{\lambda}$$

q-line in the third quarter



Assumption of McCabe Thiele Method:

- * Constant molar overflow
- * No heat loss from column

McCabe Method Not suitable if:

- * Relative volatility < 1.3 or Relative volatility > 5
- * Reflux ratio $< 1.1 R_{min}$
- * More than (25) theoretical stages needed

Notes!

- N_{min} = minimum number of equilibrium stages
- N = Number of theoretical stages
- $V_B = V/B$ Boilup ratio

Example 11.4-1 8- (Benzene - Toluene) $p = 101.3 \text{ kPa}$

$$F = 100 \text{ kgmol/h} \quad X_F = 0.45 \quad T_F = 327.6 \text{ K}$$

$$X_D = 0.95 \quad X_B = 0.1 \quad R = 4 \quad \bar{C}_P = 159 \text{ kJ/kg} \cdot \text{mol} \cdot \text{K}$$

$$\bar{\lambda} = 32099 \text{ kJ/kg} \cdot \text{mol} \cdot \text{K} \quad \text{Boiling point of feed} = 366.7 \text{ K}$$

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

$T_F < T_{bp}$ $\therefore$ Subcooled liquid

$$* y_{n+1} = 0.8 x_n + 0.19 \rightarrow \text{rectifying line equation}$$

$$y_q = \frac{q}{q-1} x_q - \frac{x_F}{q-1}$$

$$q = 1 + \frac{C_P (T_b - T_F)}{\bar{\lambda}} = 1.19 \approx 1.2$$

$$* y_q = 6 x_q - 2.25$$

Feed line equation

$$\text{if } x_q = 0.5 \quad \therefore y_q = 0.75$$

$$[\bar{C}_p = x_1 C_{p1} + x_2 C_{p2}]$$

$$[\bar{\lambda} = x_1 \lambda_1 + x_2 \lambda_2]$$

* $R \uparrow \therefore \text{slope} \uparrow \therefore N \downarrow$ *

* being more closer to equilibrium
line $\therefore N \uparrow$ *



Example 8- $p = 1 \text{ atm}$ $F = 1000 \text{ kgmol/h}$ $x_f = 0.2$

Feed is (saturated liquid) $x_D = 0.8$ $x_B = 0.02$

$$R = 5/3 = 1.67 \quad \rightarrow q = 1$$

$$F = D + B \rightarrow (1000 = D + B)$$

$$F x_f = D x_D + B x_B \rightarrow (200 = 0.8D + 0.02B)$$

$$D = 230.8 \text{ kgmol/h} \quad B = 769.2 \text{ kgmol/h}$$

$$y_{n+1} = \frac{R}{R+1} x_n + \frac{x_D}{R+1} \Rightarrow y_{n+1} = 0.63 x_n + 0.3$$

rectifying line equation

* The number of stages required for a given separation depends on R *

$$\left[R \propto \frac{1}{\text{stages}} \right] \quad (R \downarrow \therefore \text{slope } (R/(R+1)) \downarrow \therefore \text{stages } \uparrow)$$

Total reflux $\rightarrow$ all condensate is returned, no product and no feed added $\therefore$ gives minimum number of stages

Minimum reflux $\rightarrow$ gives more stages to achieve separation $\therefore$ infinite number of stages

* total reflux happens when operating lines lie on a 45° line

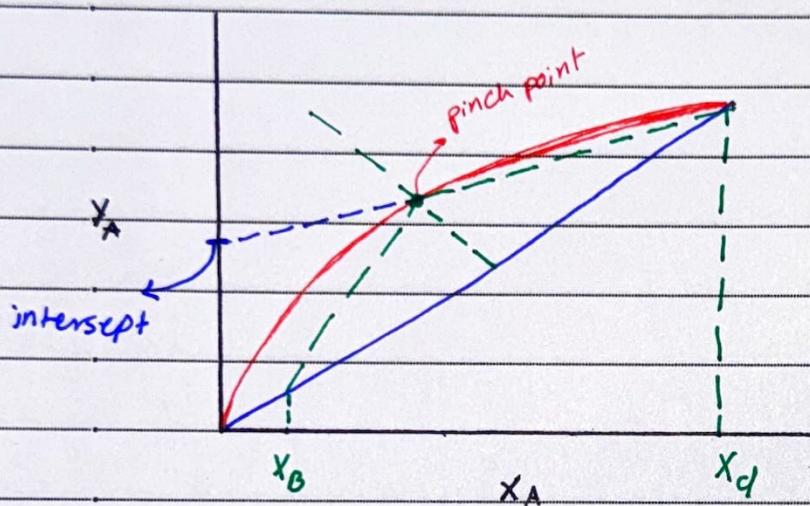
* minimum reflux happens when operating lines are on equilibrium curve

Pinch point $\rightarrow$ in minimum reflux, when two operating lines intersect on equilibrium curve, the intersection point is called pinch point.



يُعرف R_{min} من الرسم 8. يُرسم q-line ثم نجد خط من تقاطع x_d مع 45° line إلى q-line مع equilibrium curve. pinch point $\Leftarrow$ نقطة التقاطع مع y-axis

$$\text{Intercept} = \frac{x_d}{R_{min} + 1} \Rightarrow \text{I can get } R_{min}$$



$$R = (1.2 \text{ to } 1.5) * R_{min}$$

Note! increasing R needs bigger diameter (D)

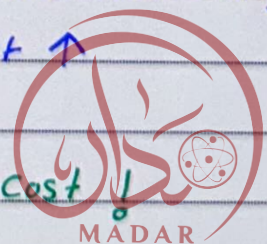
$$R \uparrow \therefore D \uparrow \therefore \text{cost} \uparrow \text{ capital cost}$$

Note! Number of plates $\uparrow \therefore$ height $\uparrow \therefore$ cost $\uparrow$
 Number of plates $\downarrow \therefore$ height $\downarrow \therefore$ cost $\downarrow$ capital cost

Note! $R \uparrow \therefore$ less stages but more boiling and condensing
 $\therefore$ operating cost $\uparrow$, capital cost $\downarrow$

$R \downarrow \therefore$ more stages but less boiling and condensing
 $\therefore$ operating cost $\downarrow$, capital cost $\uparrow$

* capital and operating costs give total cost!



Activity 8 -

$F = 204 \text{ kmol/h}$, $x_f = 0.6$, $x_D = 0.95$, $x_B = 0.05$
 $\frac{R}{R_{\min}} = 1.3$, use McCabe method to find $N_{\min}$, $R_{\min}$
 and number of stages.

Note 1. The feed enters the column with a molar percent vaporization equal to the distillate to feed ratio.

$$\rightarrow \left(q = 1 - \frac{D}{F} \right) \quad \left(R = \frac{L_0}{D} \right)$$

rectifying section equation :- $y_{n+1} = \frac{L}{V} x_n + \left(x_D \cdot \frac{D}{V} \right)$

$$F = D + B \Rightarrow 204 = D + B \dots (1)$$

$$F x_F = D x_D + B x_B \Rightarrow 122.4 = 0.95 D + 0.05 B \dots (2)$$

$$\left[\begin{array}{l} D = 124.7 \text{ kmol/h} \\ B = 79.3 \text{ kmol/h} \end{array} \right] \therefore [q = 0.4]$$

q-line equation :- $y_q = \frac{q}{q-1} x - \frac{x_F}{q-1}$

$$\Rightarrow (y_q = -0.67 x_q + 1) \quad \text{if } x_q = 0.5 \quad y_q = 0.67$$

intersection of q-line with line $\uparrow$ on the graph to find the q-line

from x_D at 45° line = 0.42 $\rightarrow$ من تقاطع الخط 45° مع خط q

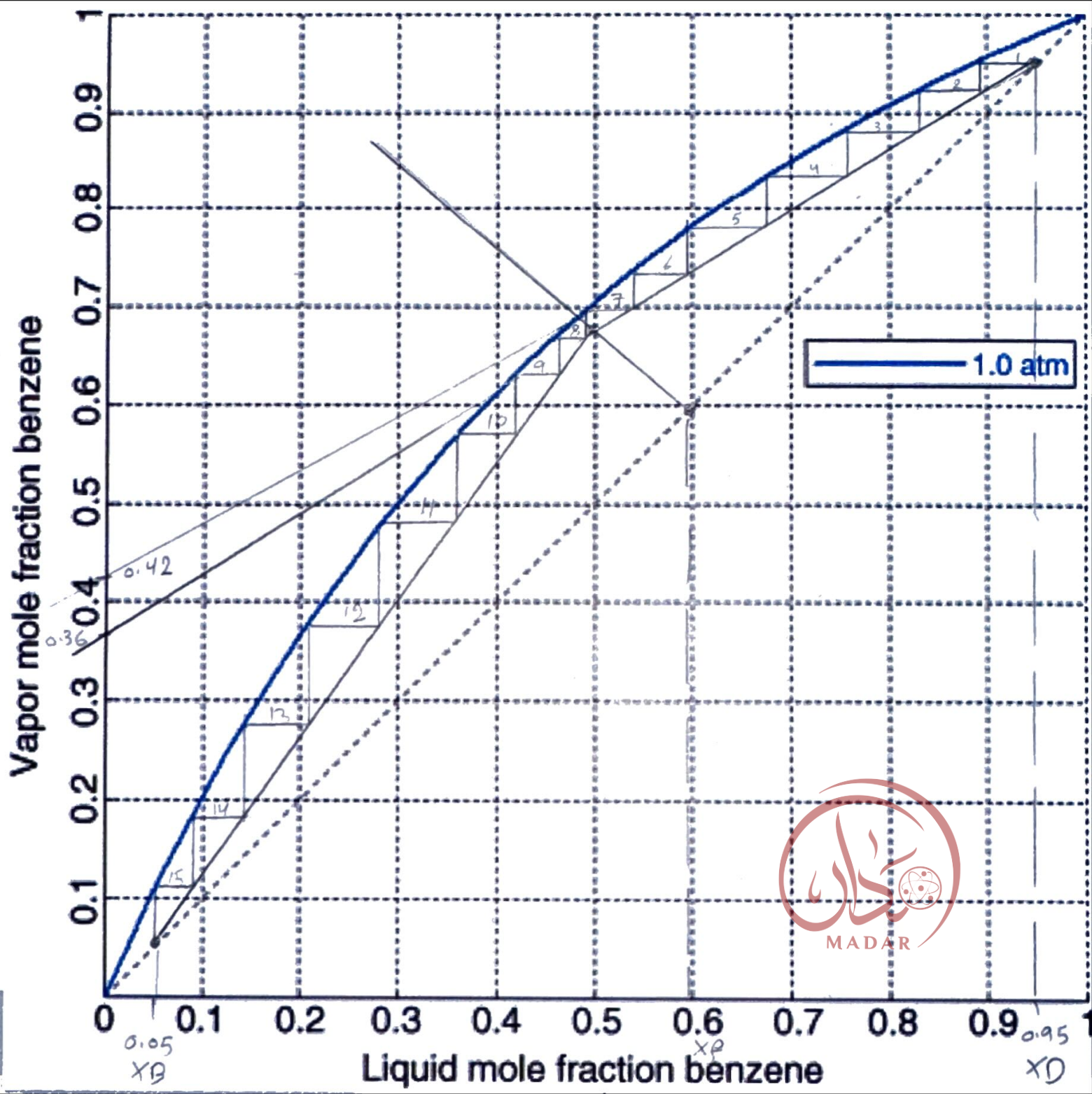
$$0.42 = \frac{x_D}{R_{\min} + 1} \Rightarrow R_{\min} = 1.3 \quad \times$$

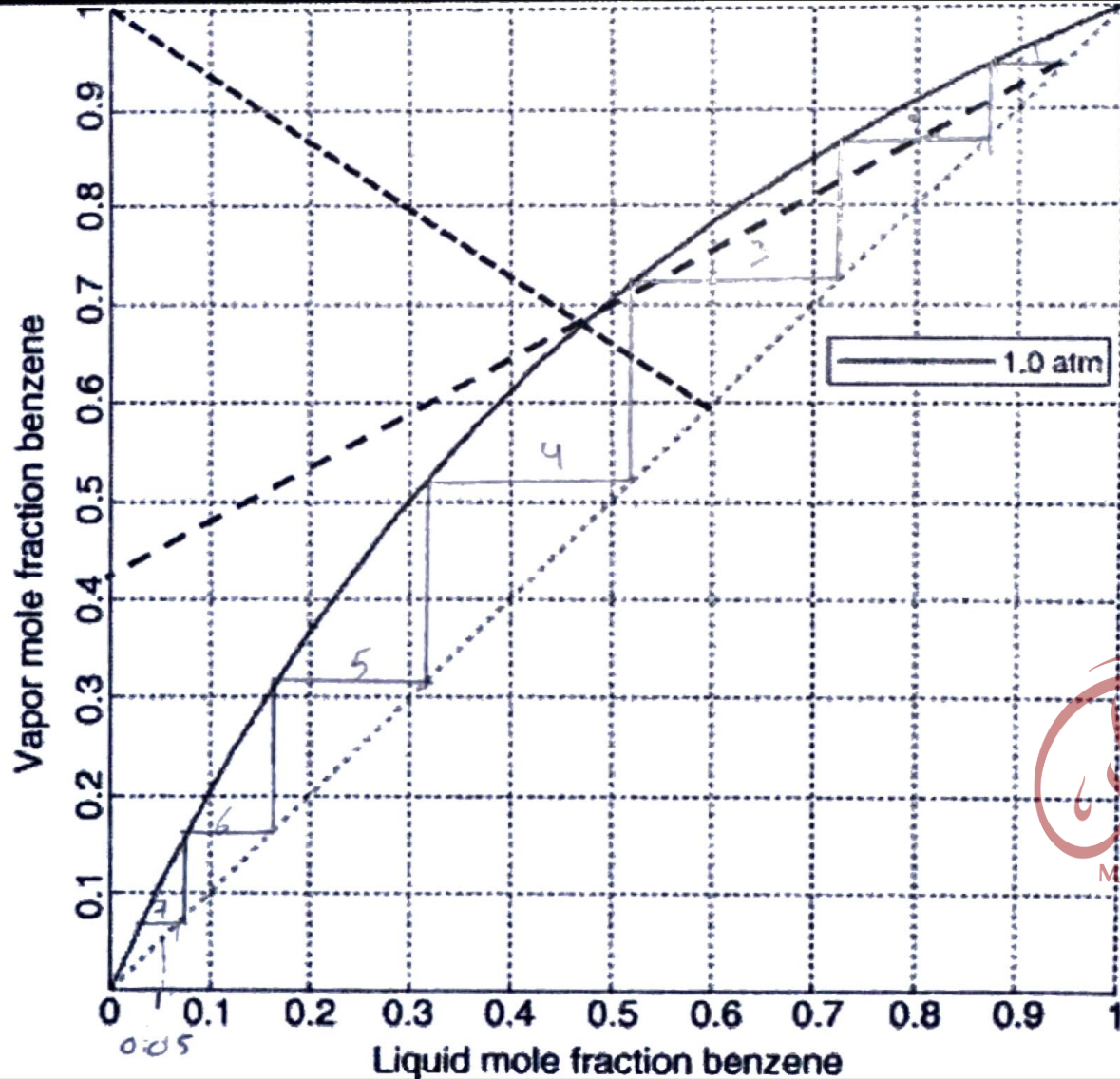
$$R = 1.64 \quad \times$$

rectifying equation $\Rightarrow y_{n+1} = 0.62 x_n + 0.36 \quad \times$

Number of stages = 15

$N_{\min} = 7 \rightarrow$ عدد المراحل الدنيا
 45° line & operating lines





$$N_{\min} = 7$$



$$Q_c = V_1 (\Delta H_{\text{evap}})$$

$$= V_1 (x_A \lambda_A + x_B \lambda_B)$$

$$\text{energy balance} \Rightarrow Fh_F + Q_R = Dh_D + Q_c + Bh_B$$

* at reference state $h=0$ *

$$h_D = c_p (T_D - T_F)$$

$$h_B = c_p (T_B - T_F)$$

Special cases :-

[1] Stripping column distillation.

* Feed enters as liquid

* feed to be distilled is not supplied

* no reflux and liquid returned back

$$V = D$$

if the feed is saturated liquid $\therefore L = F$

if feed is cold liquid [below boiling point] $\therefore q > 1 \therefore (L = qF)$

[2] Enriching column distillation.

* feed enters as vapor

[3] Rectification with Direct steam injection.

Note!

Steam stripping $\rightarrow$ آلية تستخدم لتفريق المواد وإزالة المواد العضوية
والتي تكون بنسب بسيطة في الماء

How? by entering a steam to the water at stripping section
and the steam remove the organic, volatile component
because of the higher temperature in stripping section.

Example :- $x_F = 0.1$ Fed at boiling point in Sat. liquid

Stripping tower, $D = \text{sat. vapor} = V$

$$x_B = 0.02$$

$$R = \frac{L_0}{D} = 0$$

$$3F \rightarrow 2V$$

$$y_{m+1} = \frac{L}{V} x_m - \frac{B}{V} x_B$$

$$F = L$$

$$100 \text{ mol/h} = 100 \text{ mol/h}$$

$$\frac{L}{V} = \frac{100}{150} = 1.5$$

$$* V = \frac{2}{3} F = \frac{2}{3} * 100 = 66.7$$

$$\frac{B}{V} * x_B = \frac{33.3 * 0.02}{66.7}$$

$$L = V + B$$

$$* B = 33.3 \text{ mol/h}$$

$$= 0.0099 \approx 0.01$$

$$\Rightarrow y_{m+1} = 1.5 x_m - 0.01$$

$$\text{if } x = 0.06 \quad y = 0.08$$

on the graph

$$N = 4$$

to find stripping
operating line



Generally, the heat is applied to one side = reboiler
 So, the steam doesn't contact with boiling solution

$\Rightarrow$ the heat provided by open steam injected directly at the bottom.

* reboiler is not needed

* Steam injected as small bubble into the liquid in the bottom

overall balance

$$F + S = D + B$$

$$F x_F + S y_s = D x_D + B x_B$$

$$(y_{s,A} = 0) / A = \underline{\underline{\text{more volatile}}}$$

• Enriching operating line is the same: $y_{n+1} = \frac{L}{V} x_n + \frac{D}{V} x_D$

• Stripping operating line: $L_m + S = V_{m+1} + B$

$$L_m x_m + S y_s = V_{m+1} y_{m+1} + B x_B$$

$$\Rightarrow y_{m+1} = \frac{L_m}{V_{m+1}} x_m - \frac{B x_B}{V_{m+1}}$$

for saturated steam entering: $S = V_{m+1}$ and $L_m = B$

$$y_{m+1} = \frac{B}{S} x_m - \frac{B x_B}{S}$$

liquid ↓

B زبط ز

كانت داخل S

less vapor is a, less

here the stripping line starts from $y=0$

intersection of stripping line with 45° line $\Rightarrow$

$$x = \frac{B x_B}{B - S}$$

* The advantage of open steam is simpler construction of the heater.

Example 2- feed is two phase mixture, $x_F = 0.5$ methanol
 with 50% vaporized $q = 1 - 0.5 \Rightarrow q = 0.5$
 $x_D = 0.95$ $x_B = 0.05$ $R = 2$

open stream used insted of reboiler, $N = ?$

feed line q-

$$y_q = \frac{q}{q-1} x_q - \frac{x_F}{q-1} \Rightarrow y_q = -x_q + 1 \quad \text{at } x = 0.4$$

$$y = 0.6$$

rectifying line q-

$$y_{n+1} = \frac{R}{R+1} x_n + \frac{x_D}{R+1} \Rightarrow y_{n+1} = 0.67 x_n + 0.32$$

$$N = 6$$

4 Rectification tower with side stream.

here I have 4 operating lines.

$$\text{Total Balance} \Rightarrow F = D + B + 0$$

(in = out)

$$V_{s+1} = L_s + D + 0$$

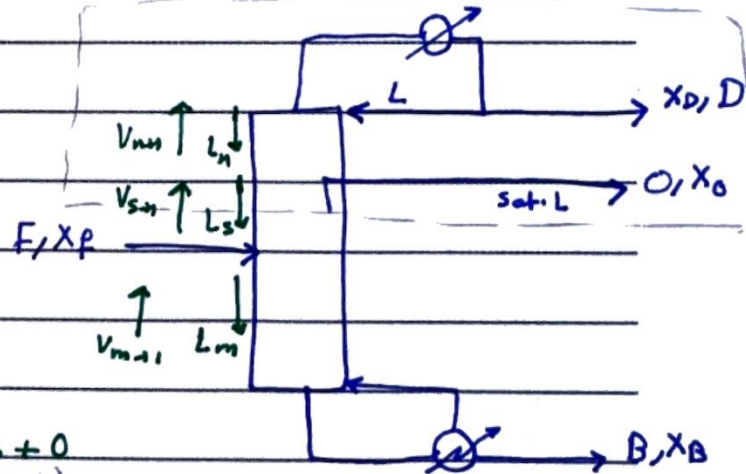
because side stream is sat.L

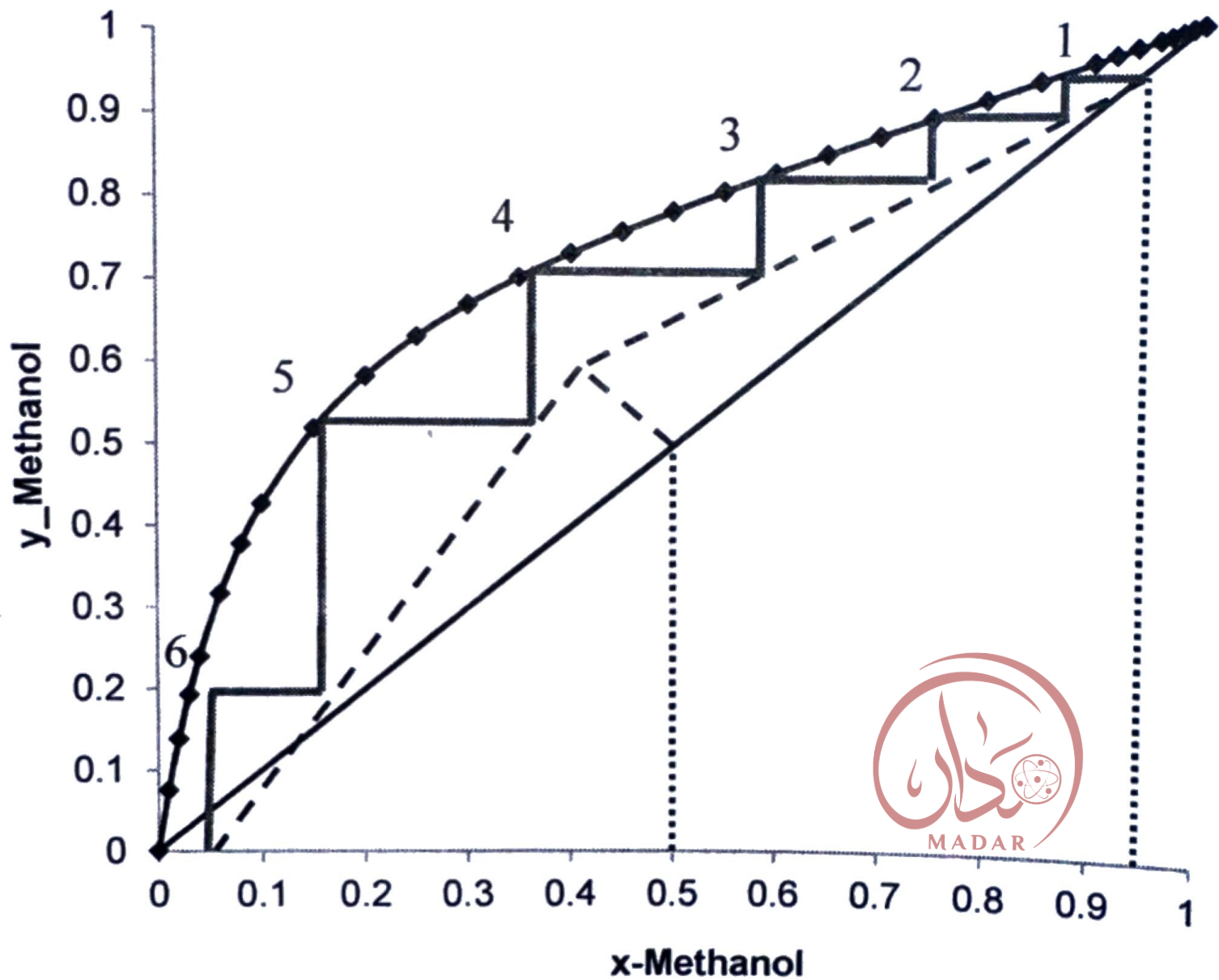
$$V_{s+1} = V_{n+1} \quad \text{and} \quad L_n = L_s + 0$$

(in = out)

$$V_{s+1} y_{s+1} = L_s x_s + D x_D + 0 x_0$$

$$\Rightarrow y_{s+1} = \frac{L_s}{V_{s+1}} x_s + \frac{D x_D + 0 x_0}{V_{s+1}} \rightarrow \text{operating line for intermediate section}$$





When we use this method ?

كيفية الحل ..

simple distillation

- * Non-ideal system
- * molar latent heat is no longer constant
- * Substantial heat of mixing

here we use → enthalpy - composition chart

Data we need to solve equations :-

- Heat capacity
- heat of mixing, small ∴ usually ignored
- latent heat of vaporization
- Bubble-point temperature

$$H_x = x_A C_{PLA} (T_b - T_o) + (1 - x_A) C_{PLB} (T_b - T_o) + \Delta H_{sol}$$

$T_b \rightarrow$ boiling T

$T_o \rightarrow$ reference T

$H_x \rightarrow$ sat. L enthalpy

$$H_y = y_A [\Delta H_{VA} + C_{PVA} (T_d - T_o)] + (1 - y_A) [\Delta H_{VB} + C_{PVB} (T_d - T_o)]$$

$T_d \rightarrow$ dew T

$H_y \rightarrow$ sat. V enthalpy

$$\begin{aligned} \Delta H_{VA} &= C_{PLA} (T_{bA} - T_o) + \underbrace{\Delta H_{VAb}}_{\text{From tables for pure A and B}} - C_{PVA} (T_{bA} - T_o) \\ \Delta H_{VB} &= C_{PLB} (T_{bB} - T_o) + \underbrace{\Delta H_{VBb}}_{\text{From tables for pure A and B}} - C_{PVB} (T_{bB} - T_o) \end{aligned}$$

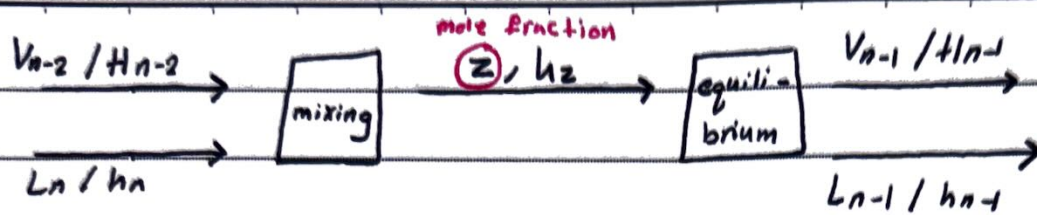
From tables
for pure A and B

at normal boiling point

$$\left[H_A + \frac{Q}{m_A} = H_C \right]$$

* معادلة الطاقة
الانثالبي الكلية (Hc) إذا
كمية الحرارة (Q) مع
النقل من نقطة A إلى C





1. energy balance:

$$\text{mixing} \rightarrow V_{n-2} H_{n-2} + L_n h_n = (V_{n-2} + L_n) h_z$$

$$\text{equilibrium} \rightarrow (V_{n-2} + L_n) h_z = V_{n-1} H_{n-1} + L_{n-1} h_{n-1}$$

2. material balance:

$$\text{mixing} \rightarrow y_{n-2} V_{n-2} + x_n L_n = z (V_{n-2} + L_n)$$

$$\text{equilibrium} \rightarrow z (V_{n-2} + L_n) = y_{n-1} V_{n-1} + x_{n-1} L_{n-1}$$

$$\therefore \frac{L}{V} = \frac{y_{n-1} - z}{z - x_{n-1}} \quad \text{or} \quad \frac{L}{V} = \frac{h_{n-1} - h_z}{h_z - h_{n-1}}$$

y_{n-1} and x_{n-1} = equilibrium !

Using lever rule :- $F = V + L$ [1] , $F x_F = V y + L x$ [2]

$$V x_F + L x_F = V y + L x \rightarrow V x_F - V y = L x - L x_F$$

$$V (x_F - y) = L (x - x_F)$$

$$\rightarrow \left[\frac{L}{V} = \frac{x_F - y}{x - x_F} \right]$$

OR $F = v + L$ [1] , $F h_F = V H + L h$ [2]

$$V h_F + L h_F = V H + L h \rightarrow V h_F - V H = L h - L h_F$$

$$V (h_F - H) = L (h - h_F)$$

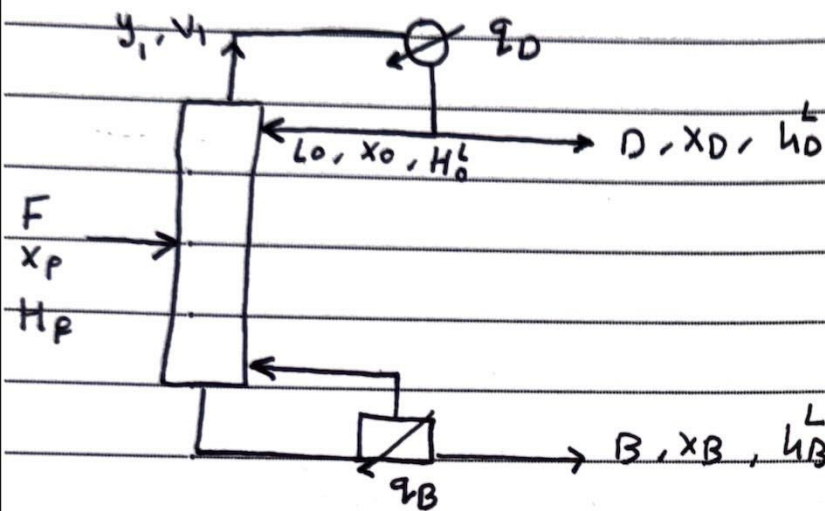
$$\rightarrow \left[\frac{L}{V} = \frac{(h_F - H)}{(h - h_F)} \right]$$

$$\therefore \left[\frac{x_F - y}{x - x_F} = \frac{h_F - H}{h - h_F} \right]$$

$$\left[\frac{h_F - H}{x_F - y} = \frac{h - h_F}{x - x_F} \right]$$

slope of $\overline{VF}$ slope of $\overline{FL}$

Balance on Continuous Distillation column :-



overall balance :-

$$F = D + B$$

Component balance :-

$$F x_F = D x_D + B x_B$$

Enthalpy balance :-

$$F h_F + q_B = D h_D + q_D + B h_B$$

in total condenser $(y_1 = x_0 = x_D)$

material balance around the condenser:

$$[1] \quad V_1 = L_0 + D$$

$$[2] \quad V_1 y_1 = L_0 x_0 + D x_D \rightarrow y_1 = x_0 = x_D \text{ جو } x \text{ کا } y \text{ میں برابر ہونا}$$

$$[3] \quad V_1 H_1 = L_0 h_0 + D h_D + q_D \quad \text{enthalpy balance}$$

$$Q_D [\text{specific}] = \frac{q_D}{D}$$

enthalpy balance becomes $\Rightarrow V_1 H_1 = L_0 h_0 + D (h_D + Q_D)$ [4]

* External reflux $\rightarrow R = \frac{L_0}{D} = \frac{h_D - H_1}{H_1 - h_D}$ from [1] and [4]

* internal reflux $\rightarrow \frac{L_0}{V_1} = \frac{h_D - H_1}{h_D - h_0}$ stage in reboiler (تحت)



$$L_0 = RD \rightarrow V = (R+1)D$$

$$[V = D + L_0]$$

$h_D \rightarrow$ موجوده بکدال
operating lines !

خطوات اکل عم total condenser :

1. ایبار x_D

2. ایبار h_D

3. ایبار H_1

4. ایبار h_D [اذا من جافزة عنی]

من R بطلعها

Note: $H \rightarrow \text{sat. Vapor}$

$h \rightarrow \text{sat. liquid}$

MADAR

in partial condenser:

$$[1] V_1 = L_0 + D$$

$$[2] y_1 V_1 = x_0 L_0 + D y_D \rightarrow \text{هون ستمه لاین}$$

$$[3] V_1 H_1 = L_0 h_0 + Q_D + D H_D \quad \text{upper \& partial condenser}$$

از بالا Vaper با Lاین ستمه H_D و h_0

* external reflux $\rightarrow R = \frac{L_0}{D} = \frac{H_D + Q_D - H_1}{H_1 - h_0}$

* internal reflux $\rightarrow \frac{L_0}{V_1} = \frac{H_D + Q_D - H_1}{H_D + Q_D - h_0}$

* نفس خطوات اکل دکن نبه؟ ستمه y_D

$L_0, V_1, \Delta = \text{at same line}$

$L_1, V_2, \Delta = \text{" " "}$

$L_2, V_3, \Delta = \text{" " "}$ نظرو

$[X_D = X_D] \quad h_D = h_D + Q_D \rightarrow \text{total condenser}$

$[X_B = X_B] \quad h_B = h_B - Q_B \rightarrow \text{partial condenser}$

Rectifying section :- $\Delta (X_D, h_D) \Leftarrow \text{اصطیاقه } \Delta$

stripping section :- $\bar{\Delta} (X_B, h_B) \Leftarrow \text{اصطیاقه } \bar{\Delta}$

$$h_D = h_D + Q_D$$

$$h_B = h_B - Q_B$$

$$\left[\frac{h}{F} = C_p \Delta T \right] \quad \left[\Delta T = (T_F - T) \right]$$



Example :- St. liquid

$$x_F = 0.42$$

$$R = 2.5$$

[total condenser]

$$x_D = 0.97$$

$$x_B = 0.01$$

$$F = 100 \text{ mol/hr} \quad [1] \quad F = D + B \rightarrow 100 = D + B$$

$$[2] \quad F x_F = D x_D + B x_B \rightarrow 42 = 0.97D + 0.01B$$

$$* D = 42.71 \text{ mol/hr}$$

$$* B = 57.29 \text{ mol/hr}$$

$$h_F = 5500 \text{ cal/mol}$$

$$h_D = \bar{h}_d + Q_D$$

$$h_d = 5000 \text{ cal/mol}$$

$$h_D = \bar{h}_d + \frac{Q_D}{D}$$

$$h_b = 5900 \text{ cal/mol}$$

$$H_1 = 13000 \text{ cal/mol}$$

to find $Q_D \Rightarrow$

$$R = \frac{h_D - H_1}{H_1 - h_d} \rightarrow 2.5 = \frac{h_D - 13000}{13000 - 5000}$$

$$* h_D = 33000 \text{ cal/mol}$$

$$33000 = 5000 + Q_D$$

$$Q_D = 28000 \text{ cal/hr}$$

$$Q_D = 1.2 \times 10^6 \text{ cal/mol}$$

$$F h_F + Q_B = D h_d + Q_D + B h_b$$

$$Q_B = 1.2 \times 10^6 \text{ cal/mol}$$

$$Q_B = 20973 \text{ cal/hr}$$



Absorption → gas mixture in contact with liquid [absorbent, solvent] to selectively dissolve component of the gas by mass transfer.



inert gas & carrier gas ← الغاز الخامل وغاز الناقل

Stripping → opposite to absorption $L \rightarrow G$

Solute → the component transferred to the liquid.

Note! in Absorption and stripping separation based on

- adding mass separating agent -

in Distillation separation based on

- adding energy separation agent -

* The main factor is the Solubility *

Uses of Absorption :-

- 1) Purification of gases. Ex) remove H_2S from hydrocarbons by Amines
 - 2) separation of gases.
 - 3) production of useful liquid product. Ex) (HCl , HNO_3 , ammonia)
- من أجل ذلك يتم استخدام الماء كعامل فصل

Applications :-

* remove ethanol from carbon dioxide by H_2O

* remove acetone from air by H_2O

* remove carbon dioxide from air by $NaOH$

Note! absorption : purification of gas stripping : purification of liquid

Factors of choice a Solvent :-

- High absorption power. $\rightarrow$ gas solubility should be high in solvent
- Highly selective.
- Easy to recover.
- Small viscosity. $\rightarrow$ Power $\downarrow$, $\Delta P \downarrow$
- Low volatility.
- Cost
- Corrosiveness [Non, less]
- Toxicity and hazard.

Absorption equipments :-

(A) Plate towers.

" multistage, $D \uparrow$, capacity $\uparrow$, separation $\uparrow$, (highly pressure drop) ^{سحب}
easy to be clean, cooling stage is provided

(B) Packed Columns.

" used for high corrosive material, $D \downarrow$, (Not easy to clean), ^{سحب}
 $\Rightarrow$ cooling done by dividing the column to many sections **out side**
the column.

[more than one section $\therefore$ surface area increased] $\rightarrow$ performance $\uparrow$

* packing materials are made from: ceramics, stones, steel, wood,

(C) Spray column. Simplest Column

" continuous contact, low ΔP , low efficiency, low cost (empty) "
gas phase controlling, considered as one stage.

(D) Wetted wall column. used in labs for mass transfer coefficient

(E) Tubular Reactor. used for high exothermic rxn , low mass transfer

For absorption and stripping we assume that:

1. Carrier gas is insoluble
2. Solvent is non-volatile $L \rightarrow G$
3. The system is isothermal and isobaric

- $V = A + C$ C : carrier gas
- $V' = V(1 - y_A)$ = flow of the carrier gas only!
- moles of $A = y_A V$
- $L' =$ moles of the solvent

Note! moles of L' and V' stay the same

- * if the equilibrium pressure of gas is high $\therefore$ insoluble in liquid.
 - * if the equilibrium pressure of gas is low $\therefore$ solubility in liquid high
 - * if the equilibrium temperature $\uparrow$ $\therefore$ solubility of gas $\downarrow$
- $\hookrightarrow$ so, it's necessary to fit coolers to the absorber to keep its temp low then eff $\uparrow$

Henry's law $\rightarrow$ for diluted gas at equilibrium

$$P_A = H X_A$$

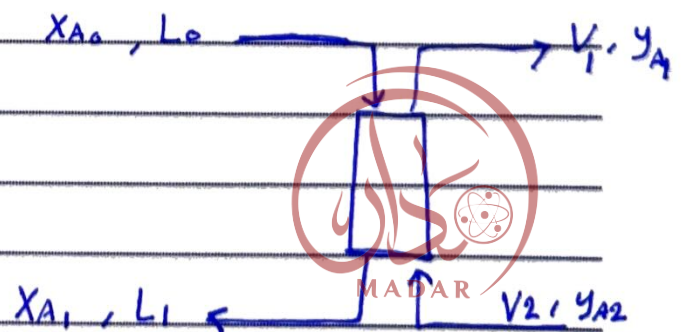
$$y_{Ai} = m X_{Ai}$$

$$m = H_A / P_{\text{total}}$$

$$H = \frac{K P_A}{\text{mole fraction}}$$

in absorption or stripping operation carried out in counter current

- * gas introduced in bottom
- * liquid introduced in top
- * $y_{A1} \leftarrow X_{A2} \rightarrow$ equilibrium



Example 8: single stage gas [air + CO₂] with pure water

inlet gas $V = 100 \text{ kg mol/h}$

$$L' = ?$$

$$y_{A2} = 0.20$$

$$V' = ?$$

liquid inlet $L = 300 \text{ kg mol/h}$

$$x_{A1}, y_{A1} = ?$$

$$H = 0.142 \times 10^4 \text{ atm / mol fra}$$

$$L' = L(1 - x_{A0}) \quad \text{pure water} \therefore x_{A0} = 0$$

$$[L' = 300 \text{ kg mol/h}]$$

$$V' = V(1 - y_{A2}) = 100(1 - 0.2) = [80 \text{ kg mol/h} = V']$$

$$y_{A1} = 0.142 \times 10^4 x_{A1} \quad \text{--- (1)}$$

$$\textcircled{2} \dots 300 \left(\frac{0}{1-0} \right) + 80 \left(\frac{0.2}{1-0.2} \right) = 300 \left(\frac{x_{A1}}{1-x_{A1}} \right) + 80 \left(\frac{y_{A1}}{1-y_{A1}} \right)$$

$$20 = 300 \left(\frac{x_{A1}}{1-x_{A1}} \right) + 80 \left(\frac{0.142 \times 10^4 x_{A1}}{1 - 0.142 \times 10^4 x_{A1}} \right)$$

$$\boxed{\begin{array}{l} x_{A1} = 1.41 \times 10^{-4} \\ y_{A1} = 0.2 \end{array}}$$



$$V_2 y_{A2} + L_0 x_{A0} = V_1 y_{A1} + L_1 x_{A1}$$

$$L' = L(1 - x_A)$$

$$V' = V(1 - y_A)$$

$$= L / (1 + x_A)$$

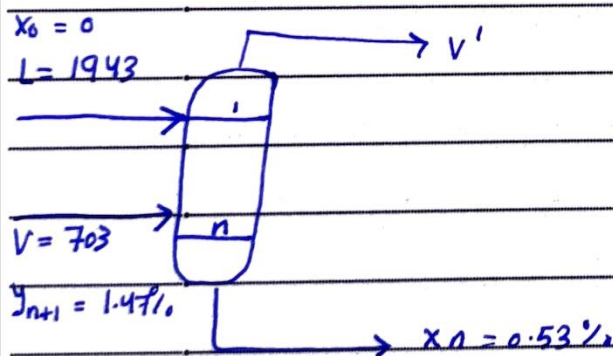
$$= V / (1 + y_A)$$

A → only component that redistributes between two phases!

$$\underbrace{L' \left(\frac{x_{A0}}{1 - x_{A0}} \right)}_{x_{A0}} + \underbrace{V' \left(\frac{y_{A2}}{1 - y_{A2}} \right)}_{y_{A2}} = \underbrace{L' \left(\frac{x_{A1}}{1 - x_{A1}} \right)}_{x_{A1}} + \underbrace{V' \left(\frac{y_{A1}}{1 - y_{A1}} \right)}_{y_{A1}}$$

$$y_{n+1} = \frac{L'}{V'} x_n + \left(y_1 - \frac{L'}{V'} x_0 \right) \Rightarrow \text{operating line equation}$$

Example :-



$$L' = L(1 - x_{A0}) = 1943(1 - 0)$$

$$L' = 1943 \text{ kmol/h}$$

$$V' = V(1 - y_{n+1}) = 703(1 - 0.0147)$$

$$V' = 692.7 \text{ kmol/h}$$

$$X_{A_n} = \frac{x_{An}}{1 - x_{An}} = \frac{0.0053}{1 - 0.0053} = 0.0053$$

$$y_{n+1} = \frac{0.0147}{1 - 0.0147} = 0.0149$$



Graphical Design

- 1) Draw an equilibrium line
 - 2) Draw an operating line
 - 3) locate the point (X_0, Y_1) on the operating line
- $\downarrow$ in $\downarrow$ out-final (TOP)

* in Absorption the operating line is above equilibrium line *

Note! Solute free basis $\rightarrow$ use the ratio X, Y

- 4) Find X_N and Y_{N+1} and locate them
- $\downarrow$ out-final $\downarrow$ in (Bottom)

$$y = H_A x \Rightarrow Y = \frac{H_A X}{1 + X - H_A X}$$

Henry's law in term of Ratio

* in Stripping the operating line is below equilibrium line *

* لكن اذا ما عدي H واعطاني equilibrium relation $y_A = 2.53x_A$
باخذ حدود، كتر من نقطة x_A و y_A $\rightarrow$ حسب التوزيع التي بيها $0 < 1$
ومن هاهنا العلم بعد حدود x_A و y_A و X_A و Y_A *

Absorption in packed towers

Lec "16"

* continuous contacting equipment (continuous differential contact) *

Steps for the design of packed tower : select a solvent $\rightarrow$ selection of packing $\rightarrow$ find L_{min} $\rightarrow$ D $\rightarrow$ height of column

Note! $L' \approx L$ and $V' \approx V$

Packing materials characteristics : Cost, Surface area (large),
Void Volume (high), Fouling resistance, Mechanical strength,
Uniform flow of streams

Notes! * efficiency $\uparrow$ = low pressure drop "Lids" *

- High void volume = low pressure
- Bigger packing materials = low fouling resistance
- Non uniform flow leads to stagnant liquid pool = low mass transfer

Packing materials

- Random, ex: Rasching rings, metal ball rings
- Structured, ex: Mellapak, Flexipak, wire mesh

(Random = dumped)

* باء کانی استعمال اکثرین نوع packing باء tower, لوازم *

Packing Height (Z)

Equilibrium stage analysis HETP

$$Z = HETP \cdot N$$

$N =$ Number of stages

HETP = Height equivalent
to a theoretical plate

Mass transfer analysis IITU

$$Z = HTU * NTU$$

HTU = height of transfer unit = H_o

MTU = number of trans per unit
= N_0

* in stage wise column = we calculate the number of stages!

* in continuous differential column (Packed bed) = calculate the height of packing

$$\boxed{Z = \frac{V}{K_y A_c} \int_{y_{in}}^{y_{out}} \frac{dy}{(y^* - y)}} \Rightarrow \boxed{Z = H_{OG} \times N_{OG}}$$

$$\boxed{Z = \frac{L}{K_x A_c} \int_{x_{in}}^{x_{out}} \frac{dx}{(x - x^*)}} \Rightarrow \boxed{Z = H_{OL} \times N_{OL}}$$

y^* → bulk of the gas

x^* → bulk of the liquid

- * if the operating line parallel to eq. line $\therefore NTU = N$ *
- * if the operating line has slope goes to y-axis $\therefore NTU > N$ *
- * if the operating line has slope goes to x-axis $\therefore NTU < N$ *

$$V y_A + L x_A = V (y_A - dy_A) + L (x_A + dx_A)$$

$$V dy_A = L dx_A \quad \text{or} \quad d(V y_A) = d(L x_A)$$

$$V y_A - V_1 y_{A1} = L x_A - L_1 x_{A1} \quad \text{or} \quad x_{in} L_{in} + y V = x L + y_{out} V_{out}$$

$$y_{n+1} = \frac{L'}{V'} x_n + (y_1 - \frac{L'}{V'} x_0) \Rightarrow \text{operating line equation}$$

For absorption $\therefore y = x \left(\frac{L}{V} \right) + (y_{out} - \left(\frac{L}{V} \right) x_{in})$

For stripping $\therefore y = x \left(\frac{L}{V} \right) + (y_{in} - \left(\frac{L}{V} \right) x_{out})$

Two Film theory :-

- driving force in gas = ΔP • driving force in liquid = ΔC
- ! phase of species is just in equilibrium

$$\text{Flux} = N_A = \overset{\text{Small } k}{k_y} (y_{Ab} - y_{Ai}) = \overset{\text{Small } k}{k_x} (x_{Ai} - x_{Ab})$$

$y_{Ai} \Rightarrow$ at the equilibrium curve.

* in general thickness of gas boundary layer ($\delta \sim 1 \text{ mm}$)
and thickness of liquid boundary layer ($\delta = 0.1 \text{ mm}$)

* concentrations on interface are in equilibrium *

$$\boxed{-\frac{k_x}{k_y} = \frac{y_{Ab} - y_{Ai}}{x_{Ab} - x_{Ai}}} = \text{slope of the line between } (x_b, y_b) \text{ and } (x_i, y_i)$$

$\xrightarrow{\text{capital } K} \quad \xrightarrow{\text{capital } K}$
 $N_A = K_x a (X_A^* - X_A) = K_y a (y_A - y_A^*)$

a : surface area per

unit volume packing

$$\frac{1}{K_y a} = \frac{1}{k_y a} + \frac{m}{k_x a}$$

$$\frac{1}{K_x a} = \frac{1}{k_x a} + \frac{1}{m k_y a}$$

$y = y_i - \frac{k_x a}{k_y a} (x - x_i) \rightarrow$ bulk composition y , x_i at interface, y_i at equilibrium composition x_i

y, x = bulk "at operating line"

y_i, x_i = interface "at equilibrium curve"

For constant mass flux :-

$$d\bar{N}_A = k_y (y - y_A^*) a A dz = k_x (X_A^* - X_A) a A dz$$

For diluted system $\Rightarrow [k'_y \approx k_y, k'_x \approx k_x]$

$$Z = \frac{V}{k'_y a A} \int_{y_{out}}^{y_{in}} \frac{dy}{y - y_A^*}$$

$\downarrow$ HoG $\downarrow$ NoG

$$Z = \frac{L}{k'_x a A} \int_{x_{in}}^{x_{out}} \frac{dx}{X_A^* - X_A}$$

$\downarrow$ HoL $\downarrow$ NoL

Note 1. $L' = (1.2 - 2) L'_{min}$

overall mass balance equation -

$$L' \left(\frac{X_{A0}}{1 - X_{A0}} \right) + V' \left(\frac{y_{A2}}{1 - y_{A2}} \right) = L' \left(\frac{X_{A1}}{1 - X_{A1}} \right) + V' \left(\frac{y_{A1}}{1 - y_{A1}} \right)$$

Numerical Evaluations B - • (2-Points): $\frac{h}{2} [f(x_0) + f(x_1)]$ $h = x_1 - x_0$

• (3-Points): $\frac{h}{3} [f(x_0) + 4f(x_1) + f(x_2)]$ ($x_1 = x_0 + h$) $h = \frac{x_2 - x_0}{2}$

• (4-Points): $\frac{3h}{8} [f(x_0) + 3f(x_1) + 3f(x_2) + f(x_3)]$ ($x_1 = x_0 + h, x_2 = x_0 + 2h$) $h = \frac{x_3 - x_0}{3}$

MADAR

Separation by phase addition $\Rightarrow$ Extraction

liquid-liquid extraction (solvent extraction) $\rightarrow$ mass transfer operation in which a feed carrier + solute is contacted with immiscible or nearly immiscible liquid solvent or organic phase.

✓ Energy saving ✓ easier separation ✓ lower temperature

LL E steps 8-

* mixing / contacting (turbulent contact)

* phase separation drops of same phase comes together due to (density difference)

* collection the phases leaving the unit

Raffinate (R) $\rightarrow$ feed phase

Extract (E) $\rightarrow$ solvent phase

carrier [C] and solute [A] are $\Rightarrow$ miscible	[C] and solvent [S] are $\Rightarrow$ immiscible	[A] is soluble in [C] and [S]
---	--	-------------------------------

Note ! normally one of the two phases is an organic phase and the other is aqueous phase.

Distribution law $\Rightarrow$ $\left[K_D = \frac{(Y_A)_E}{(X_A)_R} \right]$ under equilibrium!
or partition

The separator could be :- ① column with stages or packing
② column with moving internals ③ Single stage (mixer/settler)

Why we use LLE? when I have:

- 1) Relative volatilities are close to unity $\alpha < 1.1$, or azeotropic
- 2) Thermally sensitive components
- 3) Solute concentration is low

Applications:

- purification
- Wash polar components or acid/base

properties of extraction solvent:-

- selectivity
- solvent should be immiscible or slightly miscible with feed
- low boiling point [high vapor pressure]
- stable / non-reactive
- recoverability of solute from solvent
- safe, cheap, available

Note: Solubility of organic compounds is a function of the polarities of solvent and solute. "Like dissolve Like"

* The higher value of K_D indicates higher extraction efficiency
ideal condition $K_D \gg 1$ or $K_D \ll 1$

$$\frac{X}{W_0} = \left[\frac{V_{\text{feed}}}{V_{\text{feed}} + K_D V_{\text{solvent}}} \right]^n$$

X : solute remain in aqueous.s

W_0 : initial mass of solute in feed

n : number of extractions

V_i : volumes

MADAR

Example 8 - $K_D = 10$ 150 mL ether 100 mL H_2O
 ether will extract 5g of organic molecule from water
 Find the extract efficiency? (concentration = $\frac{m}{V}$)

[1] Assuming

$$\text{efficiency} = \frac{\text{mass out}}{\text{mass total}}$$

$$K_D = \frac{(Y_A)_E}{(X_A)_R}$$

$m_1 \rightarrow$ stay at carrier phase (water)

$5 - m_1 \rightarrow$ in the ether (solvent) phase (E)

$$K_D = 10 = \frac{\frac{5 - m_1}{150}}{\frac{m_1}{100}} \Rightarrow m_1 = 0.3125 \text{ g}$$

$$5 - 0.3125 = 4.6875 \text{ g}$$

الكتلة المتبقية في
 ether في المرحلة

$$\text{efficiency} = \frac{4.6875}{5} = 93.75\%$$

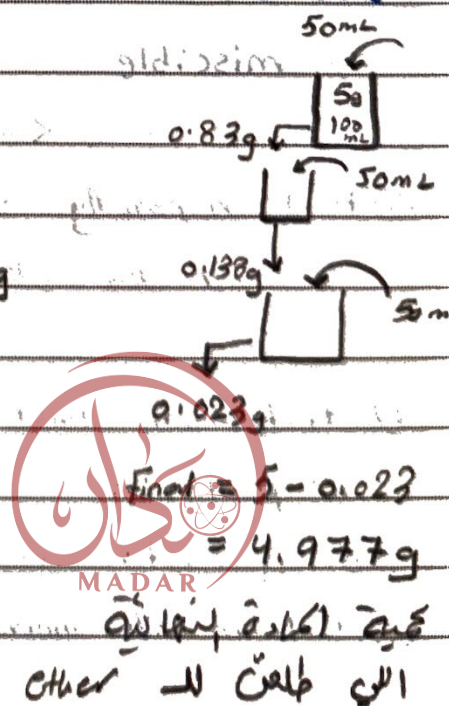
[2] in three portions with 50 mL of ether each time.

$$10 = \frac{(5 - m_1)/50}{m_1/100} \Rightarrow m_1 = 0.83 \text{ g}$$

$$10 = \frac{(0.83 - m_2)/50}{m_2/100} \Rightarrow m_2 = 0.138 \text{ g}$$

$$10 = \frac{(0.138 - m_3)/50}{m_3/100} \Rightarrow m_3 = 0.023 \text{ g}$$

$$\text{efficiency} = \frac{5 - 0.023}{5} = 99.54\%$$



in Liquid-liquid equilibrium :-

$$f_a = x_a \gamma_a f_a^o \rightarrow \text{fugacity for liquids !}$$

if I have two phases (L^α and L^β) at same reference state

$$f_a^\alpha = f_a^\beta \Rightarrow \cancel{f_a^o} x_a^\alpha \gamma_a^\alpha = \cancel{f_a^o} x_a^\beta \gamma_a^\beta$$

$$\Rightarrow [x_a^\alpha \gamma_a^\alpha = x_a^\beta \gamma_a^\beta]$$

(γ_a : activity coefficient) γ_a : activity coefficient

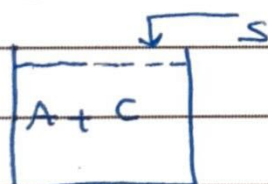
Gibbs law : $F = C - P + 2$

in LLE [$F=3$] : The system defined by 3 parameters

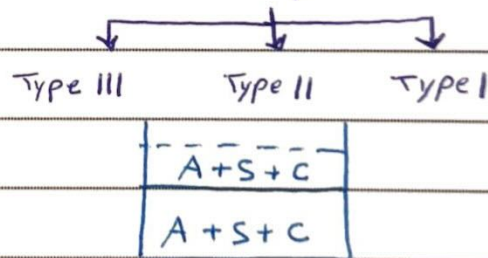
Temperature, $(y_A)_E$, $(x_A)_R$

when LLE reaches equilibrium

homogeneous phase



two phases



Type III : when A is soluble in C and S $\therefore$ The extract and raffinate phases will be binary mixture.

Type I : A dissolves completely in C and S [single phase] and C/S are partially miscible [two phases].

Type II : (A-S) and (S-C) are partially miscible but A and C are totally soluble.



* miscibility curve داخل ال

* two phases عى

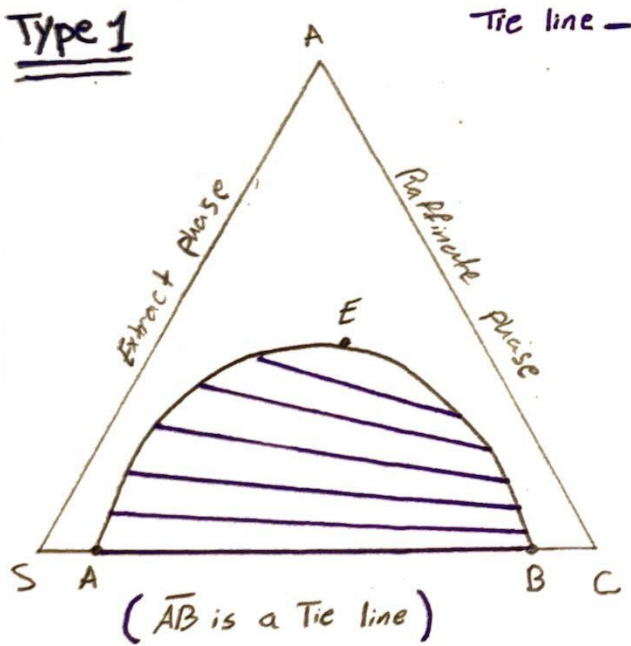
• Raffinate between A and C

• Extract between A and S

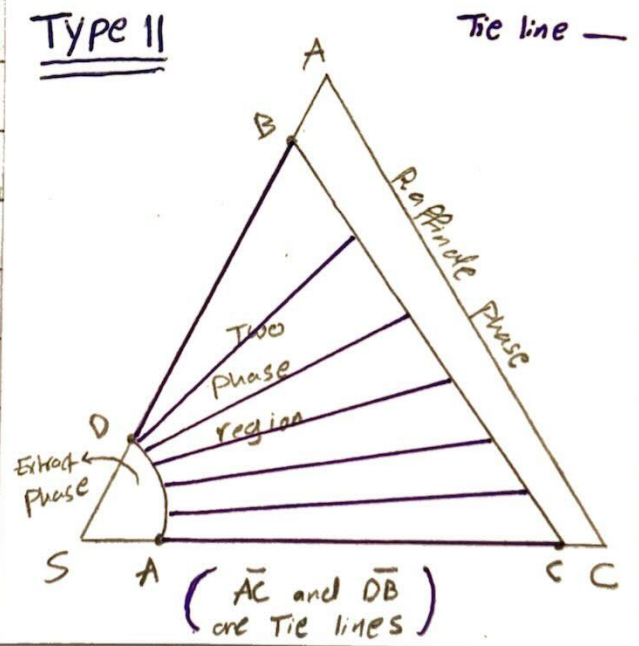
* I have plait point (E)

in Type I = azeotrope

Type I



Type II



TYPE I $\Rightarrow$ Extraction using moderate amount of solvent

TYPE II $\Rightarrow$ Extraction using larger amount of solvent

TYPE I $\Rightarrow$ closed ternary system

TYPE II $\Rightarrow$ open ternary system

Note! $T \uparrow \therefore$ Solubility $\uparrow \therefore$ two phase region $\downarrow \therefore$
separation is hard

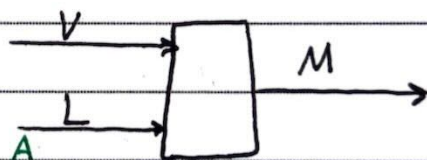
$\Rightarrow$ For this reason most LL extraction operates at low temperature and sometimes require refrigeration

Note! pressure has little effect on solubility.

Single stage equilibrium ϕ -

$$V + L = M \quad \dots (1) \quad \text{overall balance}$$

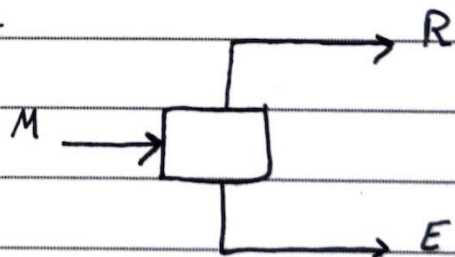
$$V y_A + L x_A = M x_{AM} \quad \dots (2) \quad \text{balance on A}$$



Single stage equilibrium extraction ϕ -

$$M = R + E \quad \dots (1)$$

$$x_{AM} M = x_{AR} R + y_{AE} E \quad \dots (2)$$



$$M = S + F \quad \dots (1)$$

$$x_{AM} M = y_{AS} S + x_{AF} F \quad \dots (2)$$



(1) أخذ دالة النقطة F

(2) فصل بين F و S

(3) نجد x_{AM} ونقاط مع الخط كإبي، نقطة M

Extraction

Batch

Continuous

single stage

multiple stages

counter-current

Co-current

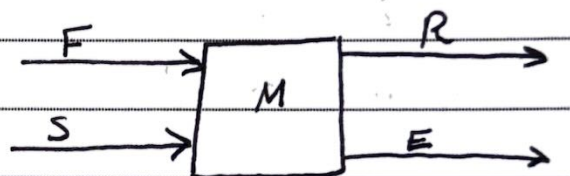
(Mixer-settler)

Note: cross flow extraction → if the feed into each stage mixed with more fresh solvent.

المرحلة "Raffinate" هو الذي يرجع إلى المرحلة التالية *
 في كل مرحلة

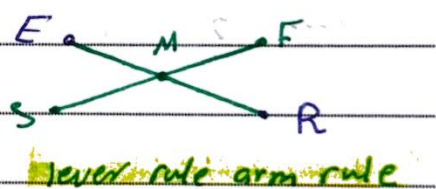
1 Mixer-settler [single stage extraction] :-

$$F + S = R + E = M \dots ①$$



Fraction of solute recovery:

$$\left(F = 1 - \frac{X_{AR} R}{X_{AF} F} \right)$$



Minimum Solvent "D" :-

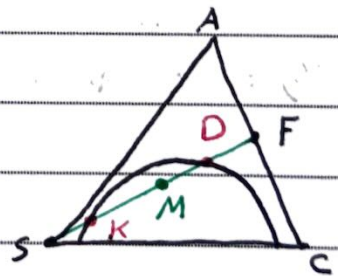
$$D = F + S_{min} \dots 1$$

$$X_{AD} = X_{AF} F + y_{AS} S_{min} \dots 2$$

Maximum solvent "K" :-

$$K = F + S_{max} \dots 1$$

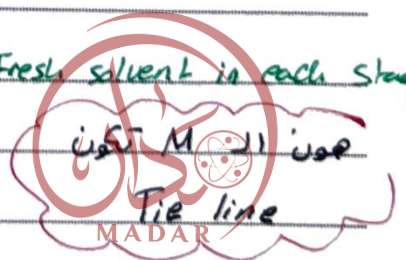
$$X_{AK} K = X_{AF} F + y_{AS} S_{max} \dots 2$$



2 Multistage cross-current Extraction :- Fresh solvent in each stage

$$F + S_1 = R_1 + E_1 = M_1 \text{ stage } ①$$

$$R_1 + S_2 = R_2 + E_2 = M_2 \text{ stage } ②$$



Example :- single stage extraction

$$F = 250 \text{ kg} \quad X_{AF} = 0.24 \quad X_{CF} = 0.76$$

$$S = 100 \text{ kg} \quad \underline{\text{Pure}}$$

$$F + S = M \Rightarrow [M = 350 \text{ kg}]$$

$$X_F F + y_A S^0 = X_{AM} M$$

$$60 = X_{AM} \times 350 \Rightarrow [X_{AM} = 0.17]$$

$$M = E + R \Rightarrow 350 = E + R \quad \dots (1)$$

$$59.5 = 0.3 E + 0.09 R \quad \dots (2)$$

$$\begin{bmatrix} E = 133.3 \text{ kg} \\ R = 216.7 \text{ kg} \end{bmatrix}$$

$$\text{recovery} = 1 - \frac{0.09 \times 216.7}{0.24 \times 250} = 0.675$$

Min solvent :-

$$D = 250 + S_{\min} \Rightarrow D - S_{\min} = 250 \quad \dots (1)$$

$$0.23 D = 250 \times 0.24 + 0 \quad \dots (2)$$

$$D = 260.9 \text{ kg} \quad [S_{\min} = 10.87 \text{ kg}]$$



* Solute recovery = $\left(1 - \frac{R_N X_{RNA}}{F X_{FA}} \right)$

* overall solute concentration in Extract : $\left(\bar{y}_{EA} = \frac{\sum E_i y_{EiA}}{\sum E_i} \right)$

* cross-flow JL، Tie line م تقع على الأضلاع، تقسم على الأضلاع
! extract , raffinate !

[3] Continuous multistage Countercurrent Extraction :-

* Re-use of solvent *

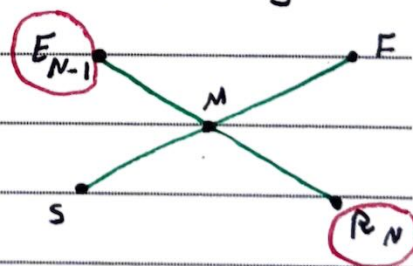
* The overall concentration : $\bar{y}_{E1A} = y_{E1A}$

Note ! • between final stage and initial stage I have mass

Balance !

• between (E2 and R2) or (E3 and R3) or.... I have equilibrium

here : M point is not a Tie line , there is no splitting
will happen on a Tie line



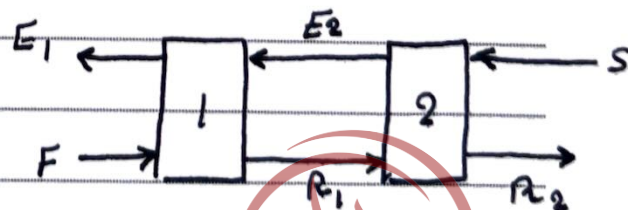
ex: if I have R_N , I draw a line
between (R_N and M) to find E_{N-1} , and
to find R_{N-1} : Tie line from E_{N-1}

mass balance of stage ①

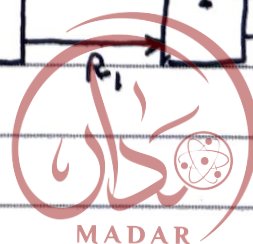
$$F + E_2 = E_1 + R_1$$

$$F - E_1 = R_1 - E_2$$

$$F - E_1 = \Delta \quad R_1 - E_2 = \Delta$$



$$F - E_1 = \Delta \quad R_N - R_{N+1} = \Delta \quad R_N - S = \Delta$$



Example :- Cross-current

$$F = 250 \quad X_{FA} = 0.24 \quad S = 100 \text{ kg "pure"}$$

Find min-stages to achieve 85% recovery?

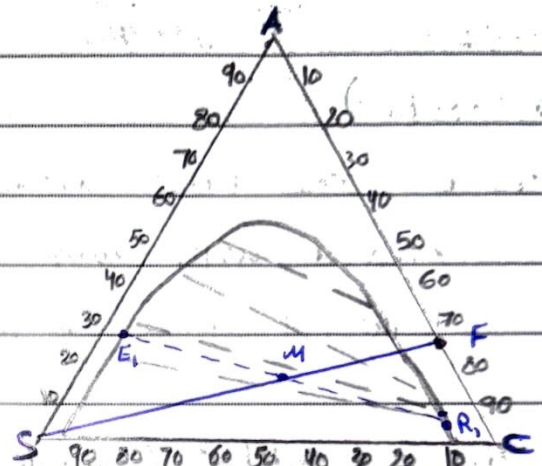
Stage 1

$$F + S = M = 350 \text{ kg}$$

$$\textcircled{1} 350 = R_1 + E_1$$

$$X_{AM} = \frac{250 \times 0.24}{350} = 0.17$$

$$\textcircled{2} 60 = 0.1 R_1 + 0.35 E_1$$



$$R_1 = 250 \text{ kg}$$

$$E_1 = 100 \text{ kg}$$

Stage 2

$$S + R_1 = M_2 = 350$$

$$\text{from balance of A} \Rightarrow X_{AM_2} = 0.071$$

$$M_2 = R_2 + E_2 \Rightarrow \textcircled{1} 350 = R_2 + E_2$$

$$\textcircled{2} 24.85 = 0.04 R_2 + 0.11 E_2$$

$$\left[\begin{array}{l} R_2 = 195 \text{ kg} \\ E_2 = 155 \text{ kg} \end{array} \right]$$

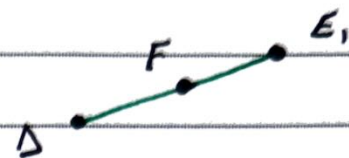
$$\text{recovery} = 1 - \frac{195 \times 0.04}{250 \times 0.24}$$

$$= 87\% > 85\%$$

∴ the stages = two!



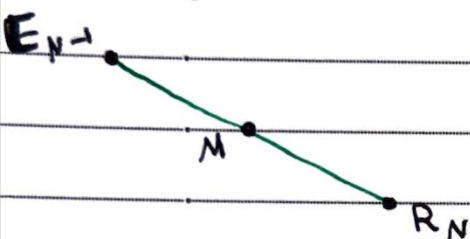
* Δ determined by two lines !



$\Delta = \text{pseudo flow rate}$

* Δ هو نقطة توازن الكتلة : Δ هو نقطة توازن الكتلة

Note ! Δ is a fictitious (خيالي) operating point.



* دالة X_{AM} و M في نقطة توازن

Total mass balance $\Rightarrow F + S = E_1 + R_N = M$

Example 8-

$F = 112 \text{ kg/h}$, $X_{AF} = 0.25$, $X_{CF} = 0.75$, $S = 28 \text{ kg/h}$

A) Find number of stages to achieve $X_{AR_N} = 0.025$

B) Find overall recovery . $F + S = M = 140 \text{ kg/h}$

$X_{AM} = 0.12$

From the graph $N \approx 5$

① $140 = E_1 + R_N$ $y_{AE_1} \approx 0.45$

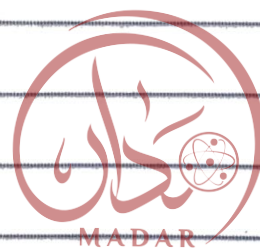
② $(0.25 * 140) = (0.45 * E_1) + (0.025 * R_N)$

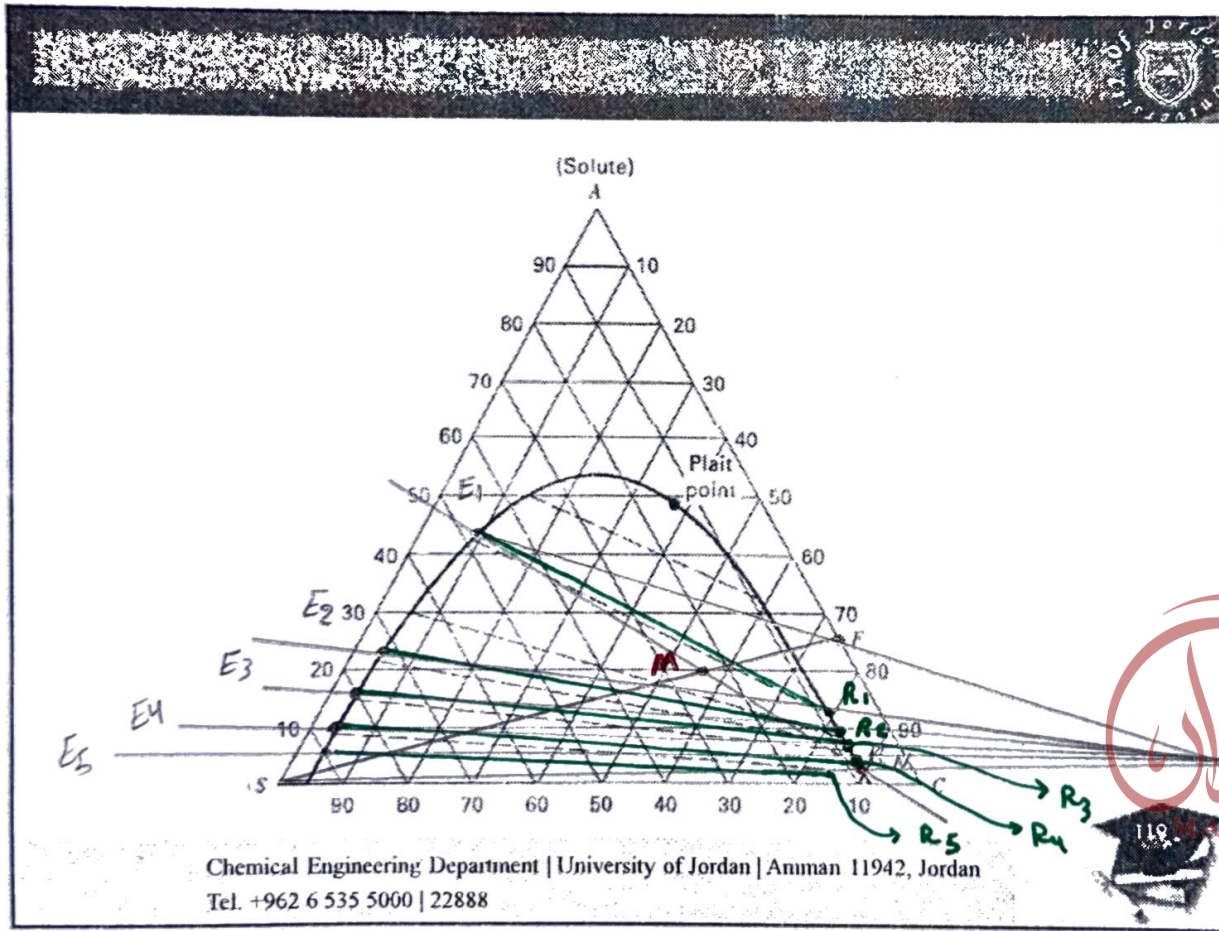
$\Rightarrow E_1 = 74.12 \text{ kg/h}$

$R_N = 65.9 \text{ kg/h}$

Recovery = $1 - \frac{R_N X_{AN}}{F X_{AF}} = 94\%$

* القيم x و y تغير حسب قراءته في ternary diagram





Determining Minimum solvent 8.

1. ترسیم tie line بر بار F و tie line بر بار F و یوای افقی tie line موجود

2. گذر E_1 و $E_{1,min}$

3. نقطه بین F و S

4. نقطه بین R_N و S

5. نقطه بین $E_{1,min}$ و R_N لنج M_{min}

$$[F + S_{min} = M_{min}]$$

$$M_{min} = E_1 + R_N$$

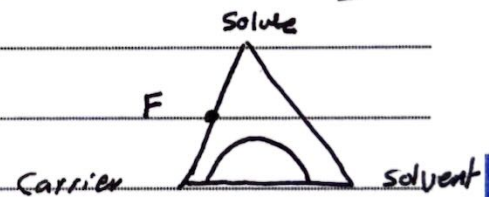
$$\frac{S_{min}}{F} = \frac{X_{AF} - X_{Am,min}}{X_{Am,min} - y_{AS}}$$

$$\left[\frac{S}{F} = 1.5 \frac{S_{min}}{F} \right]$$

* فی حال از عکس در ternary diagram

بهای بالای افقی tie line از زیر به F

Tangent line بین $E_{1,min}$ و F



Determining Maximum Solvent 8.

* if amount of solvent (S) increases $\therefore$ point M moves toward point S *

! as I have single stage !

$$\frac{S_{max}}{F} = \frac{X_{AF} - X_{Am}}{X_{Am} - y_{AS}}$$

[4] countercurrent extraction with immiscible liquids 8-

- only the solute is distributed -
- ternary diagram is not helpful -

E' : Flow rate of solvent in the extract

R' : Flow rate of carrier in the raffinate



$$R_0' = F(1 - X_{AF}) = R_1' = R_2' = R'$$

$$E' = S(1 - y_{N+1}) = E_1' = E_2' = E'$$

ratio of solute to carrier:

$$X = \frac{x}{1-x} = \frac{\text{kg of A}}{\text{kg of C}}$$

ratio of solute to solvent:

$$Y = \frac{y}{1-y} = \frac{\text{kg of A}}{\text{kg of S}}$$



operating line equation $\rightarrow \left[y_{n+1} = \underbrace{\frac{R'}{E'}}_{\text{Slope}} x_n + \underbrace{y_1 - \frac{R'}{E'} x_0}_{\text{intercept}} \right]$

$$E'_{\min} = \frac{R'}{\text{slope}_{\max}}$$

$$S_{\min} = \frac{E'_{\min}}{1 - y_{n+1}}$$

Example : $S = 200 \text{ kg/h}$ $y_{A5} = 0.0005 = y_{N+1}$

$F = 100 \text{ kg/h}$ $x_{A0} = 0.01 = x_0$

$x_N = 0.001$

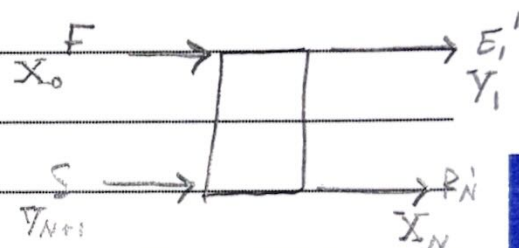
$R' = F(1 - x_{A0}) = 100(1 - 0.01) = 99 \text{ kg/h}$

$E' = S(1 - y_{A5}) = 200(1 - 0.0005) = 199.9 \text{ kg/h}$

$\bar{x}_0 = \frac{x_0}{1 - x_0} = 0.01$

$\bar{x}_N = \frac{x_N}{1 - x_N} = 0.001$

$y_{N+1} = 0.0005$



using operating line equation $\rightarrow$

$0.0005 = (0.5 \times 0.001) + y_1 - (0.5 \times 0.01)$

$[y_1 = 0.005]$

مقدار سولنت

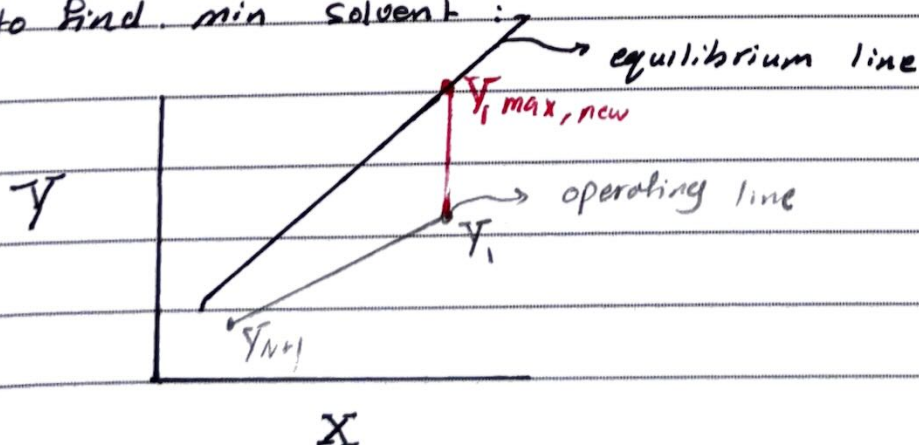
تساوی اکل

* Stripping From bottom to top *

$(x_0, y_1) \rightarrow$ top

$(x_N, y_{N+1}) \rightarrow$ bottom

to find min solvent :

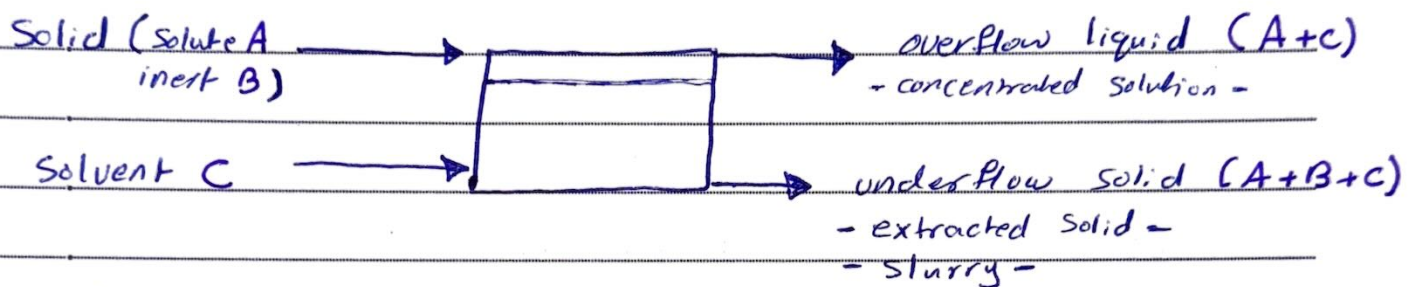


$\text{slope}_{\max} = \frac{y_{\text{new}} - y_{N+1}}{x_0 - x_N}$



- * Removal of constituent from a mixture of solids by contacting between solid material and liquid solvent.
- * The solute diffuses from solid to liquid phase.

The difference between Leaching and LLE is difficulty to transport the solid from stage to stage.



Applications :

- Leaching of sugar from sugar beets with hot water
- production of oils with organic solvents → (hexane acetone)
(peanuts, soybeans, ... oils)
- Leaching of mucins from homogenized animal material

biological substances

- Leaching Gold from its ore using sodium cyanide solution
- Extraction of copper oxide from low grade ores with dilute H_2SO_4
- Leaching Nickel salts using (sulfuric acid + ammonia + O_2) mixture

Metal industries [metal salts]

Leaching rate steps : Solvent trans from bulk to solid surface →

- solvent diffuses into solid → solute dissolves from solid to solvent
- the mixture diffuses to surface → solute trans to bulk

* diffusion into or out the solid surface = intraparticle diffusion *

① : rapid

② : low (slow) or rapid

* rate of diffusion of solute
 $\Rightarrow$ controlling resistance

* if solid made of porous

$\therefore$ diffusion through porous

described by effective diffusivity

* resistance to mass. trans of

solute from solid surface to

bulk small than diffusion

within the solid itself *



$$N_A = K_L A (C_{As} - C_A)$$

rate of A dissolved $\left[\frac{\text{kg}}{\text{s}} \right]$ $\rightarrow$ N_A $\rightarrow$ m/s
 K_L $\rightarrow$ Surface area of particles m^2
 A $\rightarrow$ Saturated concentration kg/m^3
 "Batch system"

$$\Rightarrow V \frac{dC_A}{dt} = N_A \Rightarrow \frac{C_{As} - C_A}{C_{As} - C_{A0}} = \exp \left[- \frac{K_L A}{V} t \right]$$

* Dissolving is the controlling process : solution approach saturated condition exponentially

Leaching Steps :-

[1] Contacting step. (mass trans step) by diffusion and dissolution

* this process may result in formation of pores in solid material

$\Rightarrow$ new surfaces

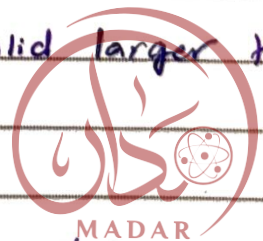
[2] Separation step. (momentum trans step) by settling or filtration

Note! these two steps may be conducted in separate equipment or in one same equipment.

Ideal contacting (mixing) yields to Thermodynamic equilibrium
 ((No mass trans + No heat trans + No momentum trans))

Why is difficult to reach ideal stage?

- 1) rate of mass transfer is slow due to pore resistance
- 2) Some part of solute is not exposed to solvent [inside closed pores]
- 3) Sometimes, adsorption of solute on surface of solid larger than solubility in solvent :
 $\Rightarrow$ Long time operation batch reactor
 or $\Rightarrow$ excessively large apparatus continuous reactor



- speed $\uparrow \therefore$ thickness of B.L $\downarrow$
- $T \uparrow \therefore$ Solubility $\uparrow \therefore$ diffusion $\uparrow$
 $\therefore N \downarrow \therefore$ Film thickness $\downarrow$
 $\therefore$ pore resistance $\downarrow$

- Size of solid $\downarrow \therefore$ mass. trans $\uparrow$
 $\therefore$ area $\uparrow$, $\boxed{\text{settling velocity} \downarrow}$
 $\therefore$ Separation of Sol / liq
 more difficult

Note! Solvent Quantity and Moisture content of the seeds are [Factors] affecting leaching.

- Vegetable and animal materials are cellular in structure and the wall cell provide a resistance to diffusion
- Drying original material rupture (تفتت) the cell wall

Preparation of leaching & - Size reduction

- * Crushing + grinding of solid $\rightarrow$ increasing rate of leaching
- * Simple washing if the soluble substance distributed on solid surface
- * in Biological material, they cut into wedge-shape slices
- * in Soybeans and Vegetable seeds they rupture by rolling + flaking
- * For Pharmaceutical product, drying of material

3-phase system: $A \rightarrow$ Solute $C \rightarrow$ Solvent $B \rightarrow$ inert solid or leached solid (insoluble solid)

Concentration of B $\Rightarrow$
$$N = \frac{kg B}{kg A + kg C}$$
 in the mixture $\rightarrow$ Solution

mass composition of solute A $\rightarrow$
$$X_A = \frac{kg A}{kg A + kg C}$$
 overflow phase
$$y_A = \frac{kg A}{kg A + kg C}$$
 underflow phase (slurry)

Note! At equilibrium $X_A = y_A$

The shape of equilibrium curve [settling characteristics] depends on: ① M_v + relative density of the liquid.

- ② concentration of solute in slurry. ③ solubility of solute in solvent.
④ Adsorptivity of solute on inert solid B.

Case [1] when P and M are constant and No adsorption.

- $y = x \Rightarrow 45^\circ$ line
- amount of solution is constant (mass of solution)
- Composition stay the same
- KL line is horizontal
- Vertical tie lines
- amount of underflow is constant

Constant Underflow: Conditions are occur to regulate the operation of Continuous thickeners.

Case [2] when P and M are not constant and adsorption occurs.

- $y \neq x$, $y > x$
- more solution will be drawn off with settled solid
- $N \downarrow$ with $x \uparrow$
- rate of solid settling $\downarrow$ with solut concentration $\uparrow$
- Tie lines are not vertical
- * unequally distribution between solid and liquid at equilibrium *

Case [3] when overflow curve is above $N=0$.

why it's above $N=0$? B is partially soluble in the solvent
OR incompletely settled liquid has been withdrawn.

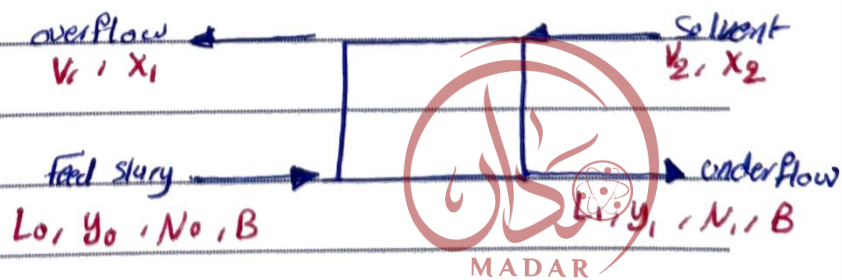
$\Rightarrow$ Tie lines are not vertical

Single stage leaching 8 -

• $V = A + C$ mass of overflow

• $L = A + C = \frac{B}{N}$

• $N = \frac{B}{L}$

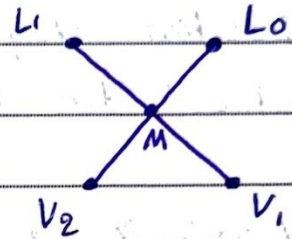


Material Balance :-

$$\textcircled{1} L_0 + V_2 = L_1 + V_1 = M$$

$$\textcircled{2} L_0 y_{A_0} + V_2 x_{A_2} = L_1 y_{A_1} + V_1 x_{A_1} = x_{Am} M$$

$$[B = N_0 L_0 = N_1 L_1 = N_m M]$$



خطوط عملی - 2

(1) $L_0 \leftarrow N_0$ و y_0 کدر

(2) نقطه خط بین V_2 و L_1

(3) x_{Am} و x_{Am} خط

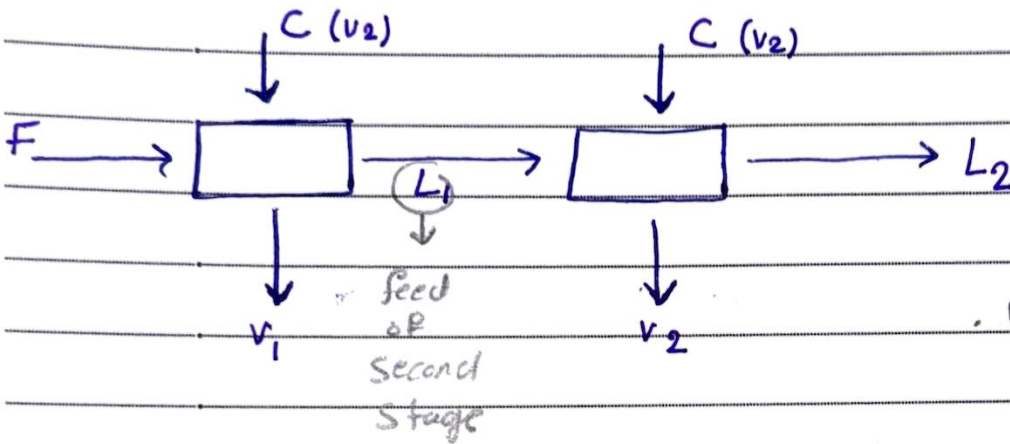
(4) $[ideal stage]$ x_1 و y_1 خط x_{Am} و Vertical line

↓ overFlow ↓ underFlow

Multistage cross-current leaching 8- "always Fresh solvent"

* more efficient *

in each step



* گند ی و نه

کای د ل

خارجی

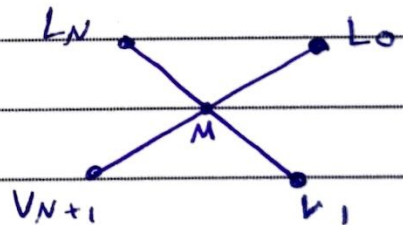
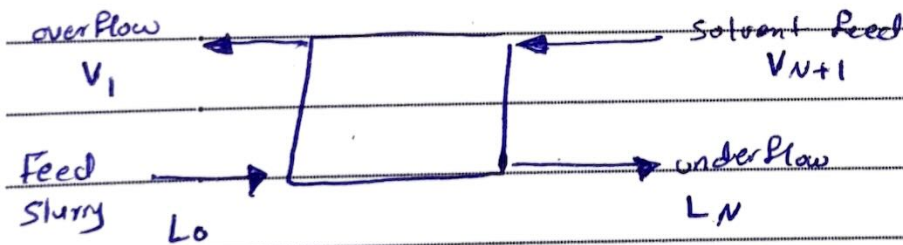
* فصل خط بین v_2 و L_2

* خط X_{AM}

* N_1, X_1, y_1 بخار Vertical tie line *

Multistage Counter Current Leaching 8-

here we use Δ -point ((between L_{in} and V_{out}))



$$V_1 - L_0 = V_2 - L_1 = V_N - L_N = \Delta \quad \left(\begin{array}{l} \text{between } L_{in} \text{ and } V_{in} \\ \text{or } L_{out} \text{ and } V_{out} \end{array} \right)$$

M

Balance on Solid :-

$$B = N_0 L_0 = N_N L_N = N_M M = N_\Delta \Delta$$

* على الرسم فصل بین L_N و V_1 يمر بـ M

* فصل بین L_0 و V_1

* L_N و V_{N+1}



Example

Continuous Countercurrent multistage $F = 2000 \text{ kg/h} + (800 \text{ kg/hr A}) + C = 50 \text{ kg/h}$, here the Feed slurry contain $(A+B+C)$,

Fresh solvent inFlow $V_{N+1} = 1310 \text{ kg/h (C)}$ and 20 kg/h (A)

Underflow $L_N = 120 \text{ kg/h (A)}$ and $B = 2000 \text{ kg/h}$ and $C = ?$

$A = ?$ V_1
 $C = ?$

V_{N+1} 4) $V_{N+1} = 1330$ $N=0$
 $C = 1310 \text{ kg/h}$
 $A = 20 \text{ kg/h}$ $X_{N+1} = 0.015$

5)

L_0 $A = 800 \text{ kg/h}$
 $B = 2000 \text{ kg/h}$
 $C = 50 \text{ kg/h}$

L_N $A = 120 \text{ kg/h}$
 $B = 2000 \text{ kg/h}$
 $C = ?$

$N_N = ?$

1) $L_0 = A + C = 850$

2) $N_0 = \frac{B}{L_0} = 2.35$

$L_N = A + C$

$C = 1010.1 - 120 = 890.1$

3) $Y_0 = \frac{A}{L_0} = 0.94$

6) $M = V_{N+1} + L_0 = 1330 + 850 = 2180 \text{ kg}$

7) $X_{AM} = \frac{(0.015 * 1330) + (0.94 * 850)}{2180} = 0.376$

8) From the graph $N_m \approx 0.9$

9) to find Y_N and N_N :-

$M = V_1 + L_N$

$N_N = \frac{B}{L_N}$ $Y_N = \frac{A}{L_N}$

10)

From the graph

Slope = $\frac{N_N}{Y_N} = \frac{B}{A} = \frac{2000}{120} = 16.67$

• $N_N = 1.98$

• $Y_N = 0.12$

• $L_N = 1010.1$

• $V_1 = 11.70 \therefore X_{A1} = 0.6$

$\therefore N_N = 16.67 \frac{Y_N}{Y_N}$

if $Y_N = 0, N_N = 0$

and $Y_N = 0.1, N_N = 1.66$

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Lec 3:-

Single stage leaching, Feed = 100 kg

$$\frac{B}{\frac{A+C}{L}} = 1.5, \text{ Find } V, L, X_A, Y_A, ?$$

$$V_2 + L_0 = M = 120 \text{ kg}$$

$$X_2 V_2 + L_0 Y_0 = X_{Am} M \quad (X_{2A} = 0)$$

$$\frac{20}{120} = X_{Am}$$

$$X_{Am} = 0.167$$

→ For ideal stage

$$X_{Am} = Y_1 = X_1 = 0.167$$

$$L_1 = \frac{80}{1.5} = 53.3 \text{ kg}$$

$$V_1 + L_1 = M$$

$$V_1 = 120 - 53.3$$

$$V_1 = 66.7 \text{ kg}$$

$$X_0 = 0.2 F, C = 100 \text{ kg Fresh}$$

$$X_{A2} = 0$$



$$(A+B)$$

$$A = 20 \text{ kg}$$

$$B = 80 \text{ kg}$$

$$L_0 = A = 20 \text{ kg}$$

$$N_0 = \frac{B}{L_0}$$

$$= 4$$

$$Y_0 = \frac{A}{A+C} = 1$$

$$N = 1.5 = \frac{80}{L_1}$$

$$B = 80 \text{ kg}$$

$F = 1000 \text{ kg}$ of Crushed oil (19.5% oil, 80.5% meal)
 $C = 1500 \text{ kg}$ pure [Batch extraction = single stage]
 Find Fraction of oil extracted.

$$(X_{A2} = 0)$$

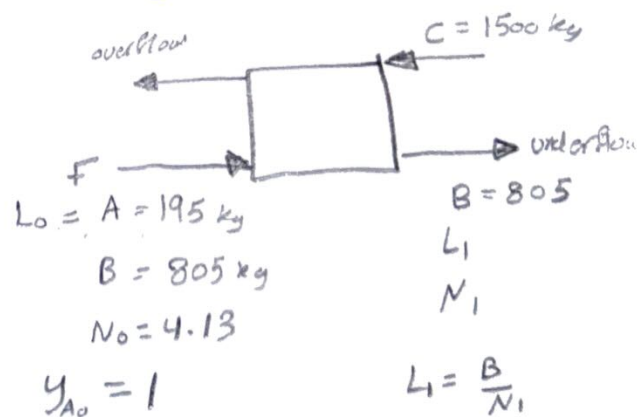
$$L_0 + V_2 = M = 1695 \text{ kg}$$

$$L_1 + V_1 = 1695$$

$$X_{Am} = \frac{195}{1695} = 0.12 = Y_1 = X_1 \quad \text{ideal stage}$$

From the graph we found $N_1 \Rightarrow \approx 2.1$

$$\text{Fraction of oil extracted} = \frac{A_1 - A_2}{A_1} = \frac{195 - 46}{195} = 0.764$$



$$L_0 = A = 195 \text{ kg}$$

$$B = 805 \text{ kg}$$

$$N_0 = 4.13$$

$$Y_{A0} = 1$$

$$L = \frac{B}{N_1}$$

$$L \approx \frac{805}{2.1}$$

$$L \approx 383.3$$

$$A_2 = 0.12 \times 383.3$$

$$A_2 = 46$$

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