Introduction:

→ Seperation processes are defined as those operations which transform a mixture of substances into two or more products which differ from each other in composition.

→ The main goal of seperation processes is to purify solution

↳ Cause different transport of species or convection of species

→ Most separation processes involve differential transport

→ Separation includes:

- Enrichment
- Purification
- Isolation
- Concentration
- Refining

→ Substances are tend to mix together naturally & Spontaneously

→ All natural processes take place to increase the entropy, or randomness of the universe

→ To separate the mixtures into products ⟶ Supply the equivalent of Energy (heat or work)

Second Law of Thermodynamics

* Rate of separation is governed by mass transfer (Rate - Controlled separation)

* Extent of separation is Limited by thermodynamics (Equilibrium - staged separation)

→ Separation Techniques:

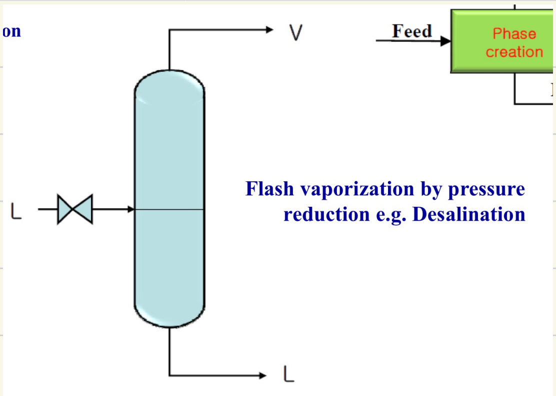
- Separation by phase creation
- Separation by phase addition
- Separation by barrier
- separation by solid agent
- Separation by external field or gradient

1- Separation by phase creation

↳ Ex: Evaporation (mostly solid & liquid separation)

Entering stream (liquid) & by adding Energy (heat)
→ phase creation of vapor

Downstream: Concentrated liquid solution



Flash vaporization by pressure reduction e.g. Desalination

↳ Ex: Distillation

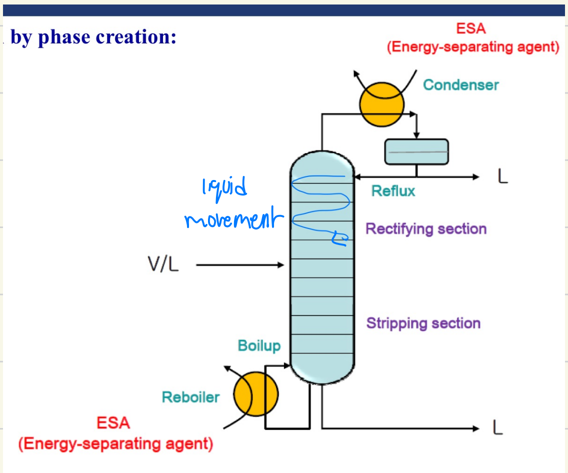
- Downstream: liquid fraction
- Boiler : provides heat liquid → vapor
- Condensor: remove heat from upstream vapor
- Trays with holes to allow vapor to move to upstream,

every tray allows contact between liq & vapor phases

↳ Flash unit composition at every tray is different to achive separation

- Reflux till desired purity
- Heat is the separating agent

↳ Heat equivalent to boiling Temp of heavy component

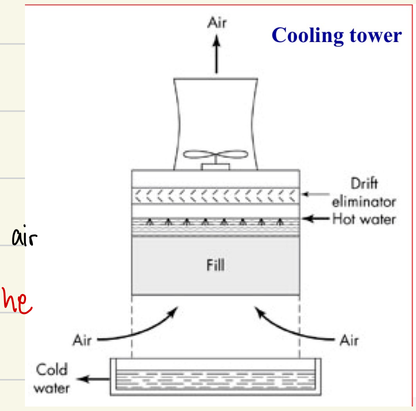

by phase creation:

↑ light component ↓ heavy component

↳ Ex: Drying, Evaporation, Crystallization, Desublimation

↳ Ex: Humidification

- Single step separations are rare ⇒ multiple separations
- Cooling Towers: Direct contact heat exchanger
  - Spraying hot water, natural draft or forced draft to contact water & air


Cooling tower

↳ water droplets (Evaporate) [takes heat from inner part of the droplet] to cool the Temp.

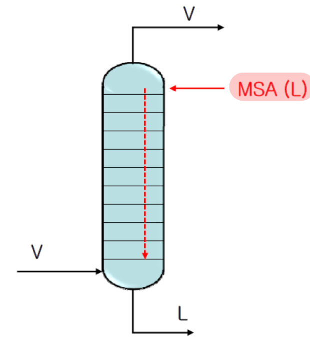
* Performance depends on surrounding conditions.

2- Separation by phase addition

↳ Ex: liquid - Liquid Extraction

↳ Ex: leaching (liq - solid extraction) , Foam fractionation

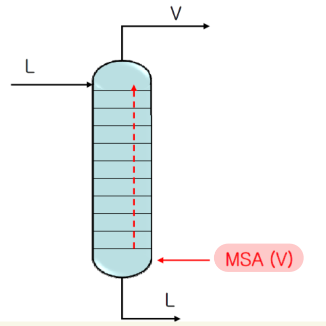
↳ Ex: Absorption & Stripping



Absorption: solvent selective to gas A & solvent → downstream

Inerts & undesired → upstream

(Use of significant amount of solvent)



Stripping : Desired product is taken from liquid into upstream

Undesired into downstream

↳ Ex: Extractive distillation [more complicated]

3 - Separation by barrier

↳ Ex: Reverse osmosis , Dialysis

↳ Ex: Microfiltration, Ultrafiltration, Pervaporation

4- Separation by solid agent

↳ Ex: Adsorption , Chromatography , Ion exchange

5- Separation by external field or gradient

↳ Ex: Thermal diffusion [Thermal gradient]

↳ Ex: Electrodialysis [Electrical force field & membrane]

Lecture 1: Distillation - Basics & Theory

Introduction:

→ Distillation is a method of separation of components from a liquid mixture which depends on the differences in boiling points of the indivisual components & distributions between liquid & gas phase

→ Different boiling point characteristics depending on the concentrations of the components

→ Distillation depends on the vapor pressure characteristics of liquid mixtures

→ Vapor pressure is created by supplying heat as the separating agent [new phases differ from the original by their heat content]

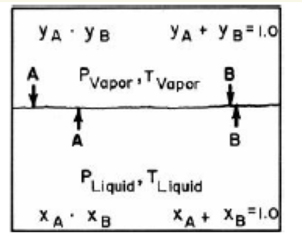
* use steam as a heat source when water is already a part of the mixture

* Very energy intensive technique, especially when the relative volatility of the component is low

→ Carried out in multi tray columns, Packed columns with efficient structured packing

↳ more efficient due to more surface area for interactions

Concept of Equilibrium



- Equilibrium stage concept: the streams leaving a stage are in equilibrium
  - → liquid molecules are continually vaporizing while vapor molecules are continually condensing
- when not in equilibrium, the liquid & the vapor can be at different Pressures & Temperatures & be present in different mole fractions.
- At equilibrium the Temperature, Pressure, & fraction of the two phases cease to change

net mass transfer = 0

$y_A P = x_A P^{sat}$

$y_B P = x_B P^{sat}$

↳ P^{sat}: antione eq.

$x_A \neq y_A$

$x_B \neq y_B$

$\Sigma x_i = 1$

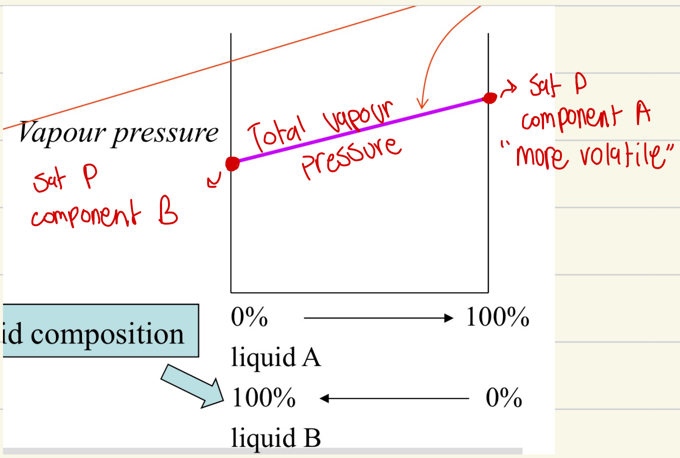
$\Sigma y_i = 1$

- In phase equilibrium, the rate at which each species is vaporizing is just equal to the rate at which its condensing → There is no change in composition [the compositions of liq & vapor are not equal]

  → if the compositions were equal, no separation could be achived in any equilibrium process

$$\mu^{v} = \mu^{L}$$ Chemical potenial

Two-component system, Mixture of Two Miscible Liquids



* more volatile component:

- higher saturation Pressure
- lower Boiling Point

$$P_1 + P_2 = P_{tot}$$ at given composition

VLE: Raoult's Law

$$y_i P = x_i P^{sat}$$

vapor composition ↲ ↳ liquid composition

$y_i P$ = Partial pressure of species i

→ Applied to VLE, & directly results from the criteria for equilibrium under the special circumstances.

- Vapor Phase is an ideal gas
- liquid phase is an ideal solution

- The first assumption means that Raoult's Law can apply only for low to moderate Pressures
- It can have approximate validity only when the species that comprimise the sys are chemically similar
- Ideal solution behaviour is often approximated by liquid phases where in the molecular species are not too different in size & are of the same chemical nature.

To calculate mole fraction

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

* Raoult's Law states that the vapor pressure of a component in a mixture at a given temperature is directly proportional to its mole fraction in the liquid mixture, & is equal to the product of its mole fraction in the liquid mixture & the vapor pressure of the pure component at that Temperature

* If we have a binary mixture of A & B, then we can write Raoult's Law for both components, Adding the equations → Convenient relationship between the liquid mole fractions & the total P.

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

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

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

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

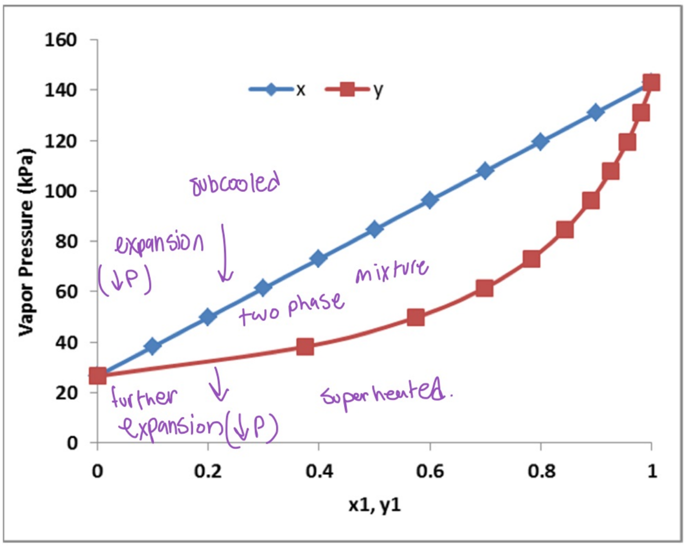
$$y_i = K x_i$$

$$\hookrightarrow \frac{P_i^{sat}}{P_{tot}}$$

Pxy

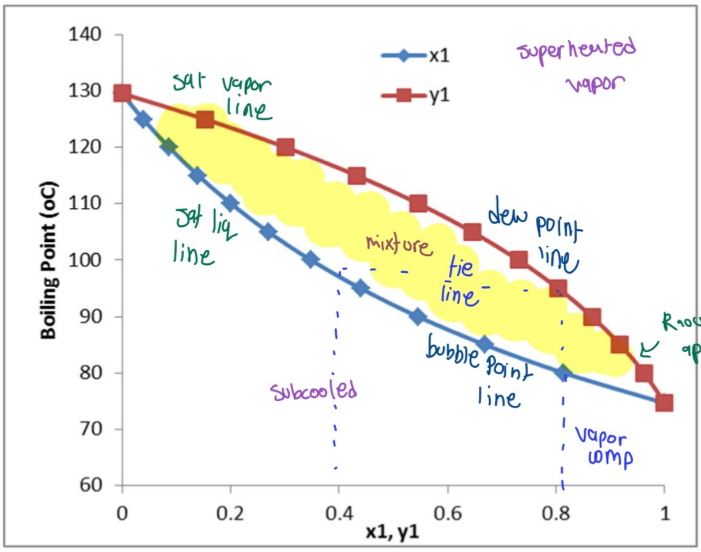
At constant temperature

↓P → goes to liq phase



Txy

At constant pressure



↙ Raoult's Law applicable inside only

tie line = isothermal line

Ideal Binary Solution

✳ A limitation of Raoult's Law is that it can be applied only to species of known vapor pressure

→ subcritical : to be at a temperature below its critical Temperature.

Vapor Pressure

→ Vapor Pressure is the pressure exerted by the vapor, when dynamic equilibrium between the vap & liq exists

→ Vapor Pressure changes with temperature (increase)

→ Different substances have different vapor pressures

→ Vapor pressure can be calculated by Antoine Equation or Clausius - Clapeyron Equation

$$\ln \frac{P_2^{sat}}{P_1^{sat}} = -\frac{\Delta h_{vap,i}}{R}\left[\frac{1}{T_2} - \frac{1}{T_1}\right]$$

Clausius - Clapeyron Eq.

*Assuming Δh_{vap} is independent of temperature

$$\ln P_i^{sat} = A - \frac{B}{T+C}$$

Antione Eq

* reasonably good over a moderate range of temperatures
* Doesn't do very well near the critical point.

Definitions:

→ Bioling / Bubble point T_{bp}: temperature at which the avg liq molecule has enough kinetic energy to escape from the surface of the liquid into gas phase

* Kinetic Energy follows Boltzmann distribution, molecules with higher than avg KE can still escape from the surface at $T < T_{bp}$ by evaporation

→ Saturated liquid: liquid at it's boiling / bubble point

→ Dew point T_{dp}: temperature at which the avg vap molecule has enough kinetic energy to condensate

→ Saturated vapor: vapor at its dew point

→ Vapor pressure: pressure at which the vapor & liquid phases are in equilibrium at a given temperature

→ Azeotrope: a constant boiling mixture ; mixture that behaves as a single component

Volatility

- liquids with high vapor pressures are said to be volatile
- Vapor pressure is a measure of volatility
- Volatile liquids have low boiling points

$$\text{volatility} = \frac{P_a \rightarrow \text{Partial Pressure}}{X_a \rightarrow \text{mole fraction}}$$

Relative Volatility

- Relative Volatility α_{AB} , is relative volatility of A with respect to B in the binary system
  - High α_{AB} → the separation is easier to occur

$$\alpha_{AB} = \frac{y_A / x_A}{y_B / x_B} = \frac{y_A / x_A}{(1-y_A)/(1-x_A)}$$

* Ratio between light & heavy component

$$y_A P = x_A P^{sat} \rightarrow y_A = \frac{x_A P_A^{sat}}{P_A}$$

$$y_B = \frac{x_B P_B^{sat}}{P_B}$$

$$\boxed{\alpha_{AB} = \frac{P_A^{sat}}{P_B^{sat}} = \frac{P_A / x_A}{P_B / x_B}}$$

$$\boxed{y_A = \frac{\alpha_{AB}\, x_A}{1 + (\alpha_{AB} - 1)\, x_A}}$$

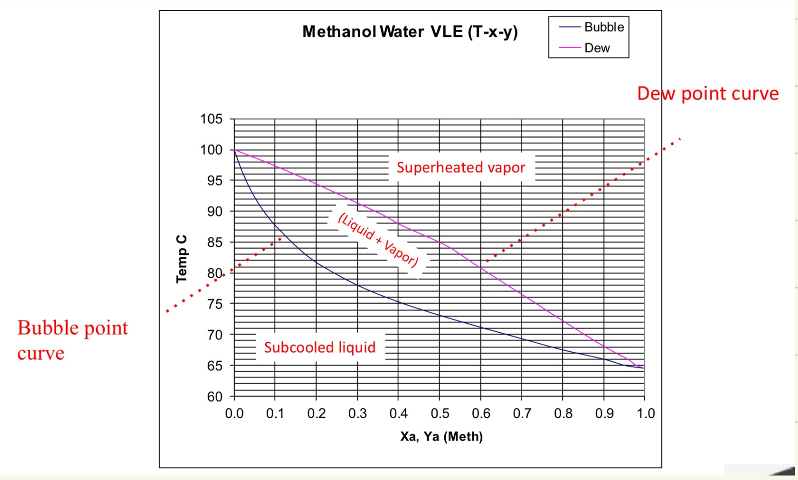
* P_A = Partial P of more volatile component

- if $\alpha = 1$ → no separation
- if α is close to 1 → separation is difficult
- if α is large → separation is easy

* α is a function of Temperature ⇒ it can change throughout the column

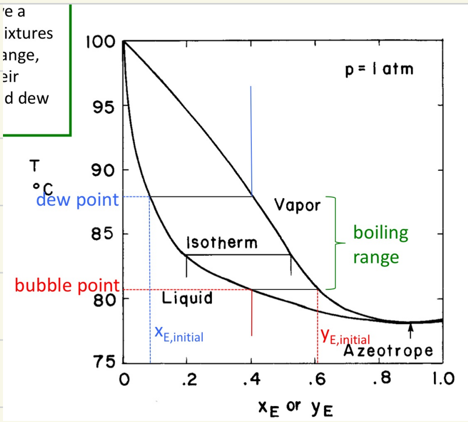
Phase Diagram

* Boiling point of the mixture depends on the relative amounts of the two components present

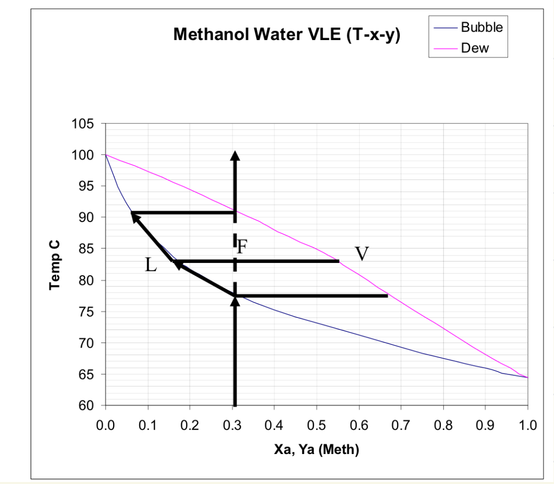


The more volatile component is labeled as x in the liquid phase concentration & y is the vapor

Bubble point = Dew point for pure components only



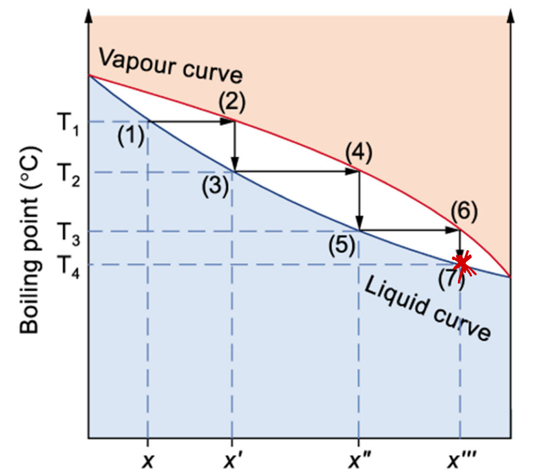
Pure liquids have a boiling point; mixtures have a boiling range



→ The relative amounts of liquid, vapor, & feed are:

$$\frac{\text{Liquid}}{\text{Feed}} = \frac{FV}{LV} = \frac{55-30}{55-15} = 62.5\%$$

$$\frac{\text{Vapor}}{\text{Feed}} = \frac{FL}{LV} = \frac{30-15}{55-15} = 37.5\%$$



to increase the concentration of light component A

* highest purity of A

Starting with a liquid mixture of composition x, after a series of consecutive distillations, Pure A is obtained in the final distillate, Pure B is obtained in the final residue.

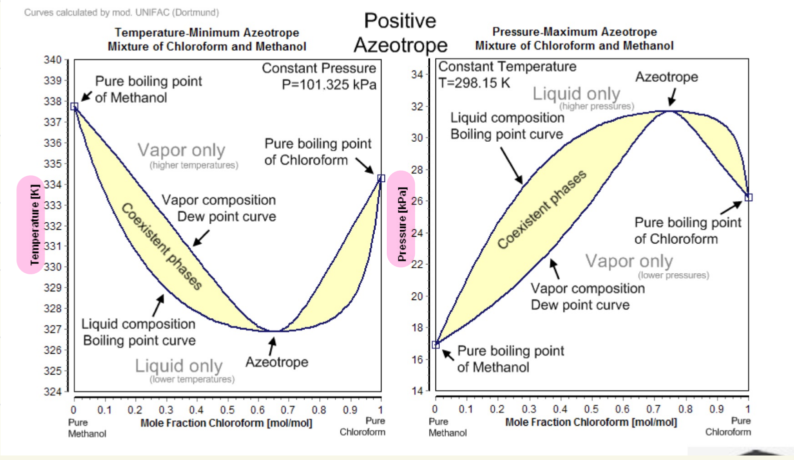
Azeotropes

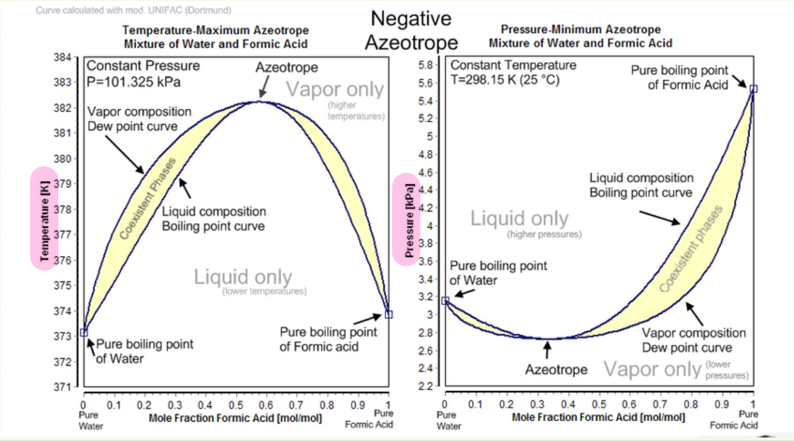
- An azeotrope is formed when the liquid & vapor compositions are the same
- separation by conventional distillation is not possible
- Dew point & bubble Point are the same at the azeotrope
- Usually occurs at a particular mole fraction, outside this point separation is possible
- Can have a minimum or maximum boiling point azeotrope.
- Changing the pressure can be the solution - Extractive Distillation

* Azeotrope (constant boiling mixtures):

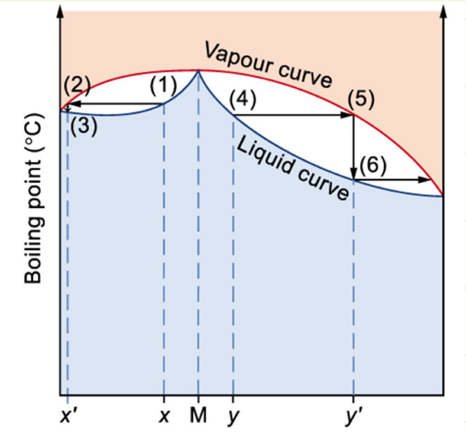
→ A boiling liquid at this composition produces a vapor of exactly the same composition

→ No separation is possible by distillation at azeotropic composition Diagram





Negative Deviation from Raoult's Law

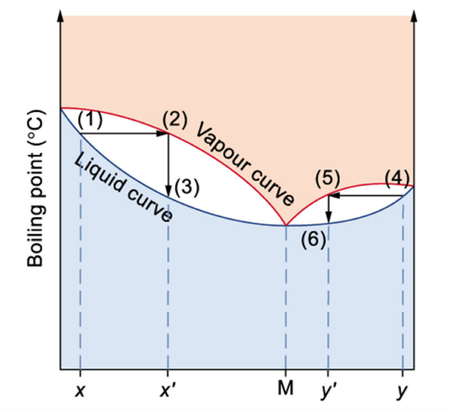


at M , vapor has the same conc as liquid

at x , fractional distillation leads to pure B in vapor phase & an azeotropic mixture residue

at y' , fractional distillation leads to pure A in vapor phase & an azeotropic mixture residue

Positive Deviation from Raoult's law



Lecture 2: Flash Distillation

Introduction:

→ The basic requirement for the separation of the components by distillation is that the composition of the vapor be different from the composition of the liquid with which it is at equilibrium at the boiling point of the liquid

→ In Evaporation (Ex: solution of salt & water), the water is vaporized but the salt is not

→ The process of absorption differs from distillation, one of the components in absorption is essentially insoluble in the liquid phase (Ex: Absorption of Ammonia from air by water), where air is insoluble in the water-Ammonia solution.

* Two methods of Distillation:

1- The first method of distillation involves the production of a vapor by boiling the liquid mixture to be separated in a single stage & recovering & condensing the vapors

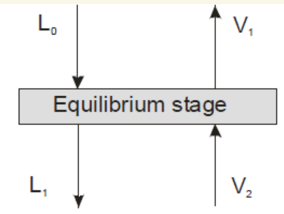
- No liquid is allowed to return to the single-stage still to contact the rising vapors.
- Do not involve refraction, Ex: Equilibrium or flash distillation, simple batch or differential distillation, & Simple steam distillation

2- The second method of distillation involves the returning of a portion of the condensate to the still

- The vapors rise through a series of stages or trays, & part of the condensate flows downwards through the series of stages or trays counter currently to the vapors
- Fractional distillation, distillation with a reflux, or rectification

Single-Stage Equilibrium Contact

- A single stage process can be defined as one in which two different phases are bought into intimate contact with eachother & then separated
- During contact → intimate mixing occurs & the various component diffuse & redistribute themselves between the two phases
- If mixing time is long enough, the components are essentially at equilibrium in the two phases after separation & the process is considered a single equilibrium stage



$L_1 + V_1 = L_0 + V_2$

out = in

* Product streams → valid for Raoult's law

→ VLE : Raoult's law

→ GLE : Henery's law

→ LLE : Distribution factor

* V_1 & L_1 (Exit streams) → Equilibrium after mixing & sufficient time

- For a binary mixture of A & B, if sensible heat effects are small & the latent heats of both compounds are the same [constant molar overflow]
  - when 1 mole of A condenses, 1 mole of B must vaporize

    ∴ The total molar flow V_1 will equal the total molar flow V_2

    ∴ The total molar flow $L_1 = L_0$

→ when CMO is valid, compositions in streams V_1 & L_1 can be solved from material balances & equilibrium rxns.

↳ Energy balance is not required

Overall Material Balance:

$$V_2 + L_0 = V_1 + L_1$$

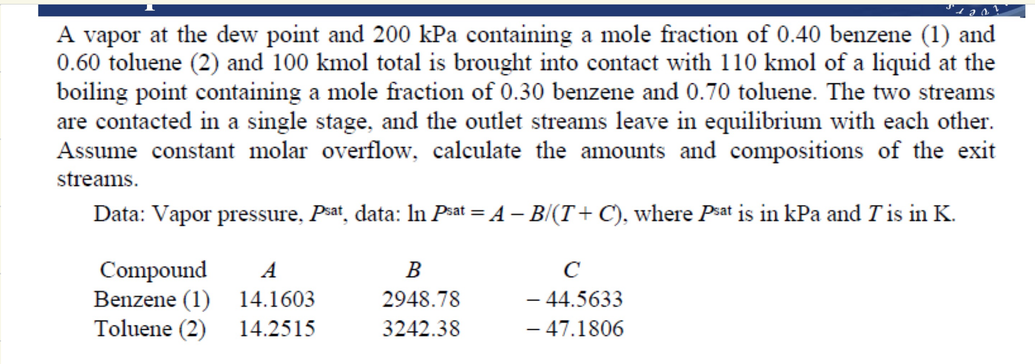
in = out

Species Balance:

$$y_2 V_2 + x_0 L_0 = y_1 V_1 + x_1 L_1$$

* The compositions of the streams leaving the process (y_1 & x_1) are related by the equilibrium distribution relation

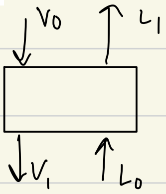
$$y_i = \Psi(x_i)$$



A vapor at the dew point and 200 kPa containing a mole fraction of 0.40 benzene (1) and 0.60 toluene (2) and 100 kmol total is brought into contact with 110 kmol of a liquid at the boiling point containing a mole fraction of 0.30 benzene and 0.70 toluene. The two streams are contacted in a single stage, and the outlet streams leave in equilibrium with each other. Assume constant molar overflow, calculate the amounts and compositions of the exit streams.

Data: Vapor pressure, P^{sat}, data: $\ln P^{sat} = A - B/(T + C)$, where P^{sat} is in kPa and T is in K.

Compound	*A*	*B*	*C*
Benzene (1)	14.1603	2948.78	− 44.5633
Toluene (2)	14.2515	3242.38	− 47.1806



V = 100 kmol

L = 110 kmol $\quad x_B = 0.3 \quad y_B = 0.4$

$x_T = 0.7 \quad y_T = 0.6$

* for CMO $\quad L_1 = L_0 = 110$ & $V_1 = V_0 = 100$ kmol

$$x_0 L_0 + y_0 V_0 = x_1 L_1 + y_1 V_1$$

$$0.3(110) + 0.4(100) = x_1(110) + y_1(100)$$

$$73 = 110 x_1 + 100 y_1$$

Rearrange to find T $\quad$ $y_1 = 0.73 - 1.1 x_1$ $\quad \leftarrow$ operating line equation

Raoult's law on B $\quad$ $y_1 P = x_1 P^{sat}$

$$y_1 (200) = x_1 \left[\exp \frac{14.1603 - 2948.78}{T + -44.5633} \right]$$

Raoult's law on T $\quad$ $$(1 - y_1)(200) = (1 - x_1) \left[\exp \frac{14.2515 - 3242.38}{T - 47.1806} \right]$$

$y_1 = 0.4425 \quad x_1 = 0.2164$

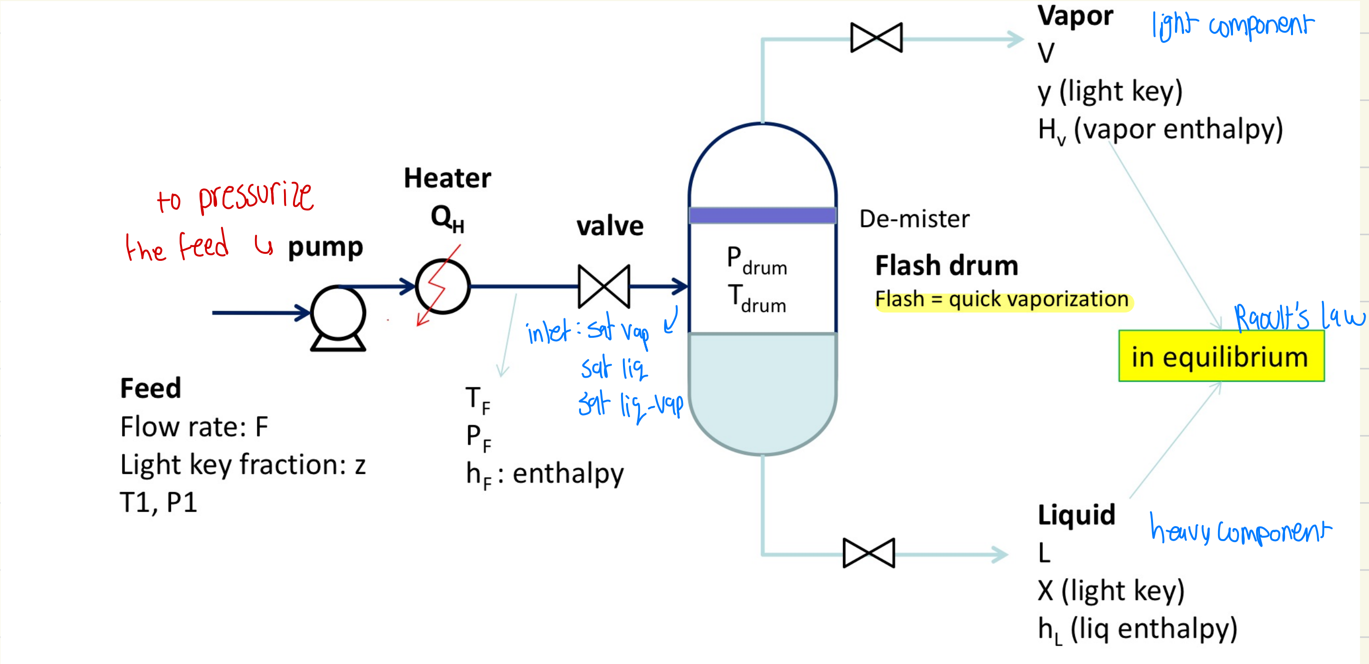
Equilibrium or flash Distillation

* flash distillation → one of the simplest Separation processes

→ A pressurized feed stream [liq phase], is passed through a throttling valve/nozzle (it may be passed through a heater sometimes to pre-heat the feed) connected to a tank or drum "flash" drum.

→ After being passed through the valve/nozzle, the feed enters the tank, whose pressure is low; thus there is a substantial pressure drop in the feed stream, causing the feed to partially vaporize

→ Vapor is allowed to come to equilibrium with the liquid → separation



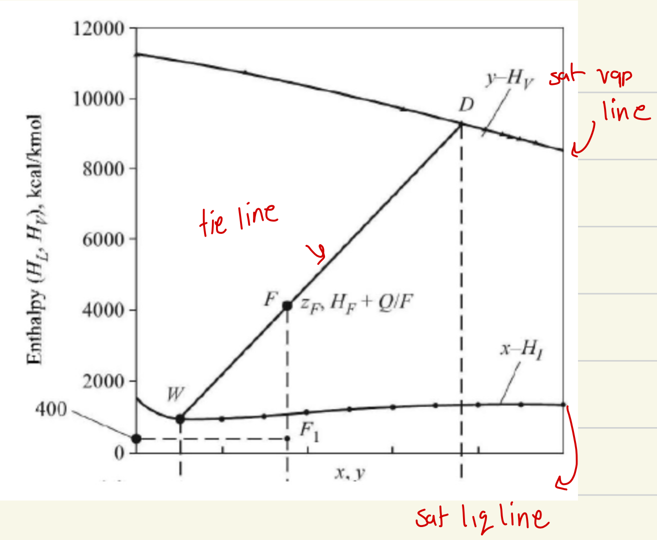
$P_{drum} < P_{feed}$ So the more volatile component evaporates quickly

* Heater is placed after the pump, rather than before it to avoid cavitation & damage to the pump
(formation of vapor bubbles)

The overall material Balance $F = L + V$

Species Balance $z_i F = x_i L + y_i V$

Energy Balance $h_F F + Q_{Flash} = h_L L + h_V V$



$$-\frac{L}{V} = \frac{y_i - z_i}{x_i - z_i} = \frac{H_V - \left(H_F + \frac{Q_{Flash}}{F}\right)}{h_L - \left(H_F + \frac{Q_{Flash}}{F}\right)}$$

$$\Rightarrow \quad F = L + V$$

$$zF = xL + yV$$

$$zL + zV = xL + yV$$

$$zL - xL = yV - zV$$

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

$$-\frac{L}{V}(x - z) = (y - z)$$

$$-\frac{L}{V} = \frac{y - z}{x - z}$$

$$y_i = -\frac{L}{V} x_i + \frac{F}{V} z$$

* linear eq on x-y diagram

* To determine the amount of Q_H (or to determine the size of the heater), an energy Balance around the heater is performed.

$$h_1 F + Q_H = h_F F$$

Degrees of Freedom

$$F = C - P + 2$$

Binary Flash Distillation

Operating line Equation

$$\boxed{y_i = \underbrace{-\frac{L}{V}}_{\text{Slope}} x_i + \underbrace{\frac{F}{V}}_{\text{Intercept}} z_i}$$

$$F = L + V$$

$$L = F - V \quad \rightarrow \quad y_i = \frac{V - F}{F} x_i + \frac{F}{V} z_i$$

* $\frac{V}{F} = f$ fraction of the feed that vaporises

$f + q = 1$

* $\frac{L}{F} = q$ fraction of the feed that remain liquid

→ f depends on the enthalpy of the liquid feed, the enthalpies of vap & liq leaving the separator $0 < f < 1$

$$\therefore \frac{L}{V} = \frac{F-V}{V} = \frac{1-\frac{V}{F}}{\frac{V}{F}} = \frac{1-f}{f}$$

$$\boxed{y_1 = \underbrace{\frac{f-1}{f}}_{\text{Slope}} x_1 + \underbrace{\frac{1}{f}}_{\text{Intercept}} z_i}$$

Alternatively: $V = F - L$

$$\frac{L}{V} = \frac{L}{F-L} = \frac{L/F}{L/F - 1} = \frac{q}{q-1}$$

$$\frac{F}{V} = \frac{F}{F-L} = \frac{1}{1-L/F} = \frac{1}{1-q}$$

$$\boxed{y_1 = \underbrace{\frac{q}{1-q}}_{\text{Slope}} x_i + \underbrace{\frac{1}{1-q}}_{\text{intercept}} z_i}$$

☐ Material Balance Operating Line Equations

→ The intersection of the equilibrium line & the operating line is the point where the system reaches equilibrium.

when x = y

$$y_1 = -\frac{L}{V} y_1 + \frac{F}{V} z_1$$

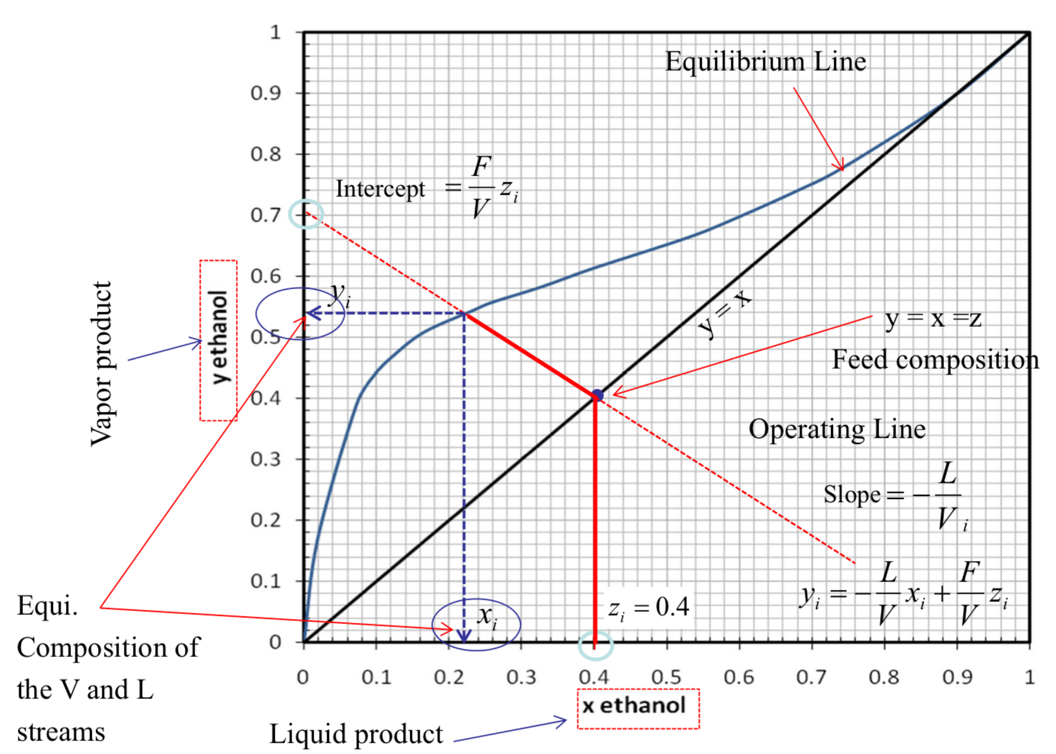
$$\left(1+\frac{L}{V}\right) y_1 = \frac{F}{V} z_i \quad \text{or} \quad \frac{V+L}{V} y_1 = \frac{F}{V} z_1$$

$V + L = F$ then: $\frac{F}{V} y_1 = \frac{F}{V} z_1$

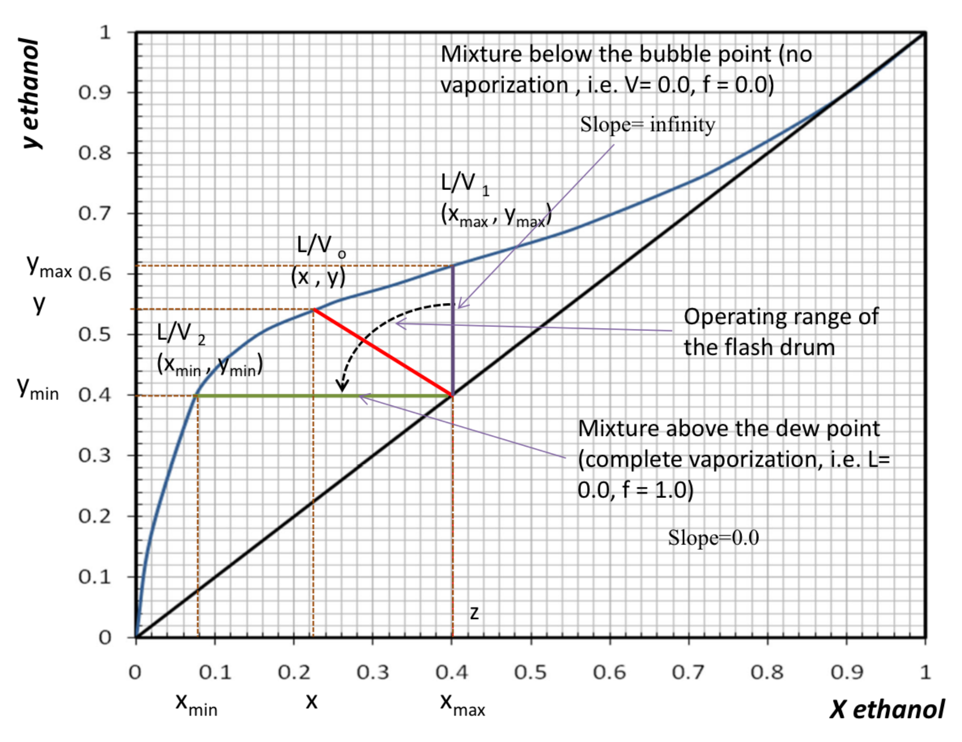
$$x_i = y_i = z_i$$

↳ The intersection of the operating line & the x = y line is the feed composition

Graphical Solution for a Flash Distillation



* To draw xy diagram → Data from T xy

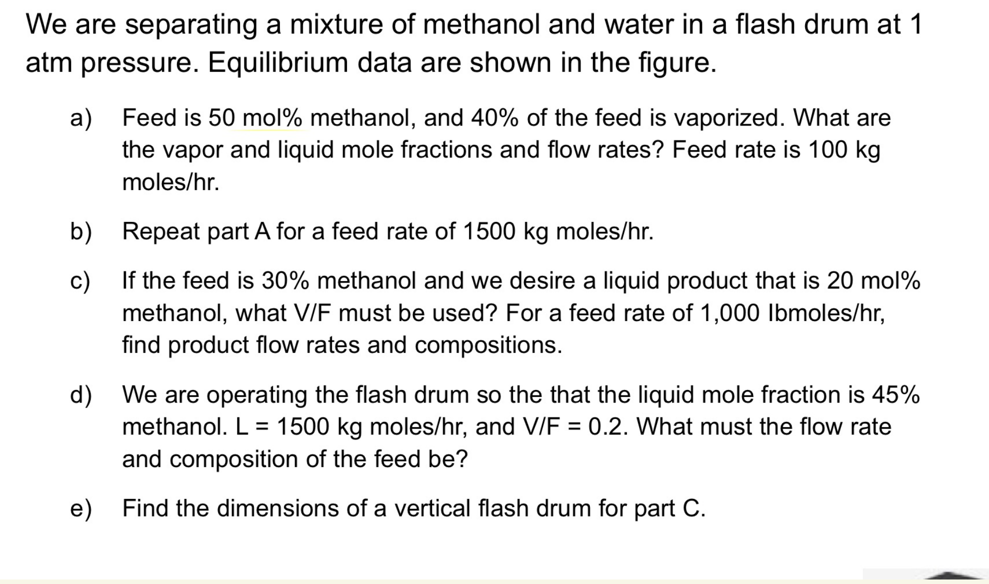


* Slope = ∞ → No vaporization , Mixture below bubble point

V = 0 (x_{max} , y_{max})

* Slope = 0 → Complete vaporization , Mixture above dew point

L = 0 (x_{min} , y_{min})



We are separating a mixture of methanol and water in a flash drum at 1 atm pressure. Equilibrium data are shown in the figure.

a) Feed is 50 mol% methanol, and 40% of the feed is vaporized. What are the vapor and liquid mole fractions and flow rates? Feed rate is 100 kg moles/hr.

b) Repeat part A for a feed rate of 1500 kg moles/hr.

c) If the feed is 30% methanol and we desire a liquid product that is 20 mol% methanol, what V/F must be used? For a feed rate of 1,000 lbmoles/hr, find product flow rates and compositions.

d) We are operating the flash drum so the that the liquid mole fraction is 45% methanol. L = 1500 kg moles/hr, and V/F = 0.2. What must the flow rate and composition of the feed be?

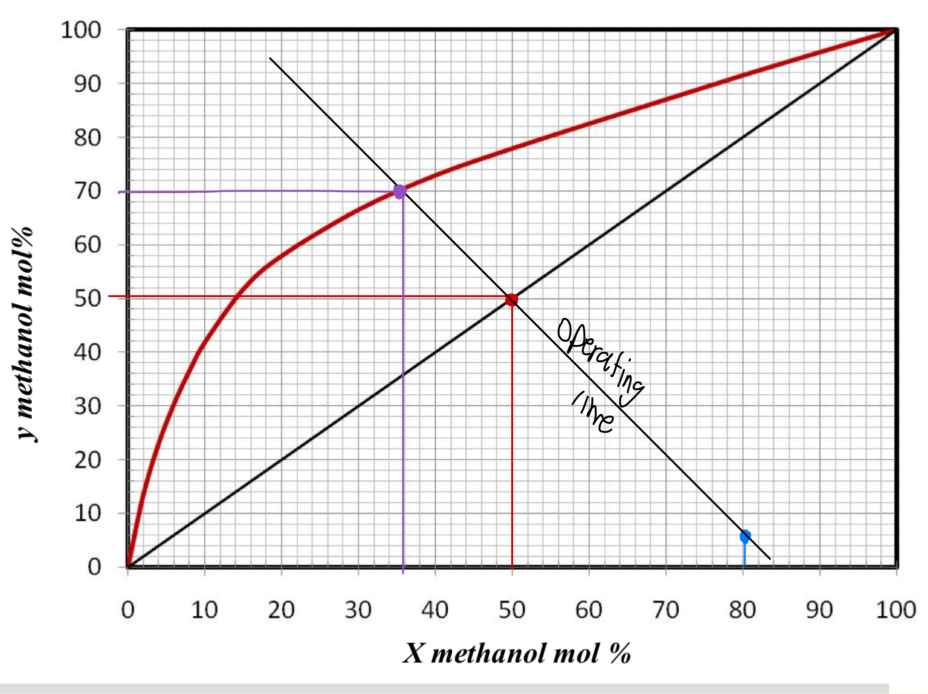
e) Find the dimensions of a vertical flash drum for part C.

a) $z = 0.5$ methanol

$f = 0.4$

$F = 100$ kg moles/hr

$x = z = y$



$$y = \frac{f-1}{f} x_i + \frac{1}{f} z_i$$

$$y = -1.5 x_i + 1.25$$

$$y = \frac{0.4-1}{0.4} x_i + \frac{1}{0.4}(0.5)$$

→ Assume $x = 0.8$ $y = 0.05$

Slope = -1.5

→ $x_{methanol} = 0.35$ $y_{methanol} = 0.7$

Intercept = 1.25

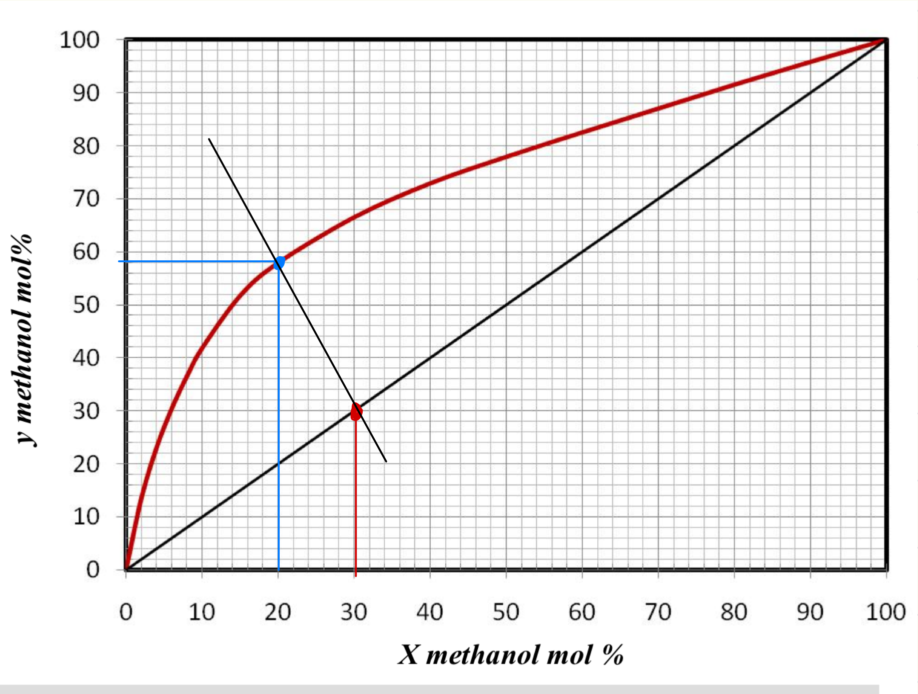
c) $z = 0.3$

→ $x = 0.2$

$\frac{V}{F} = ?$

$F = 1000$ lbmoles/hr

→ $x = y = z$



$* \; f = \frac{V}{F}$

$x = 0.2 \quad y = 0.58$

$x = 0.3 \quad y = 0.3$

$$y = \frac{f-1}{f} x_i + \frac{1}{f} z$$

$$\text{Slope} = \frac{f-1}{f} = \frac{0.58 - 0.3}{0.2 - 0.3} = -2.8$$

$$\frac{f-1}{f} = -2.8 \qquad f = 0.26$$

$$\frac{V}{F} = 0.26$$

d) $x = 0.45$ ←

$L = 1500$ kg moles / hr

$\frac{V}{F} = 0.2$

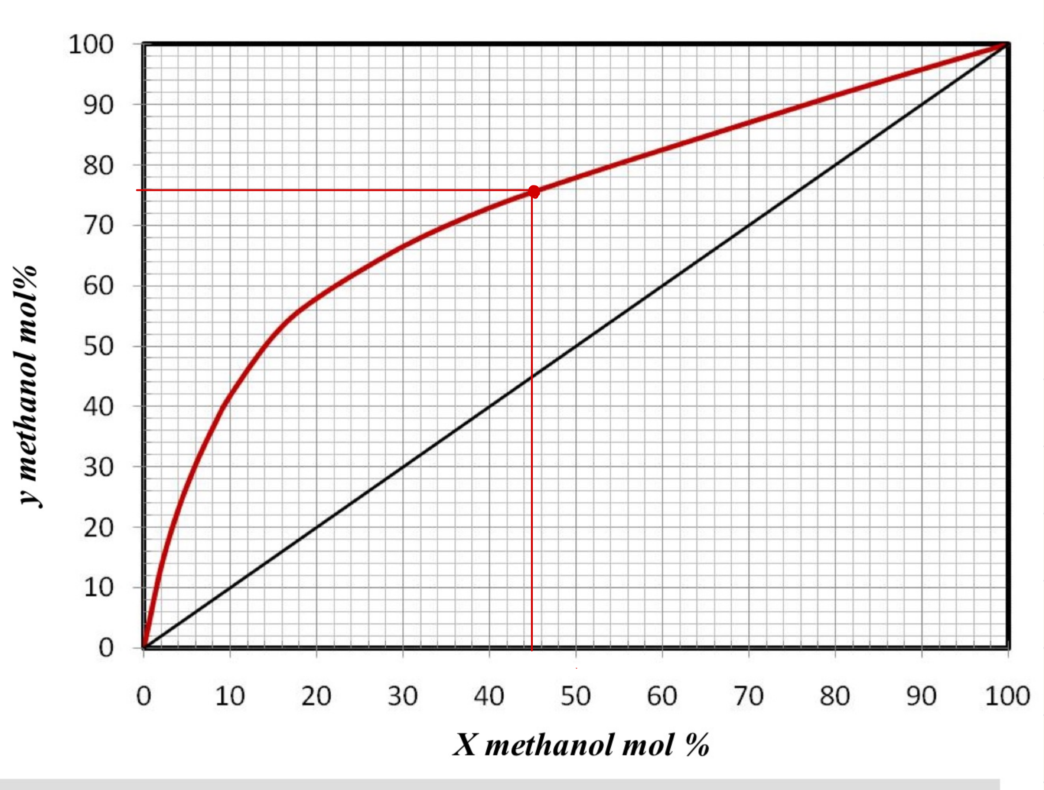
$f = 0.2$

$q = 1 - 0.2 = 0.8$

$\frac{L}{F} = q$ $\qquad \frac{1500}{F} = 0.8$ $\qquad F = 1875$ kmol/h

$\frac{V}{F} = f$ $\qquad \frac{V}{1875} = 0.2$ $\qquad V = 375$ Kmol/h

e)



$x = 0.45$ $\qquad y : 0.76$

$f = 0.2$ $\qquad q : 0.8$

$$y = \frac{f-1}{f} x_i + \frac{1}{f} z$$

$$0.76 = \frac{0.2-1}{0.2}(0.45) + \frac{1}{0.2} z$$

$z = 0.512$

* Flow rates effect the size of the drum ↓ flow rate ↓ size

* Temp & feed effect the separation process subcooled "vertical" max x & y
Super heated "horizontal" min x & y

Design of Flash Drum

→ Vertical is preferred when:

- small liquid load
- limited plot space
- ease of level control is desired

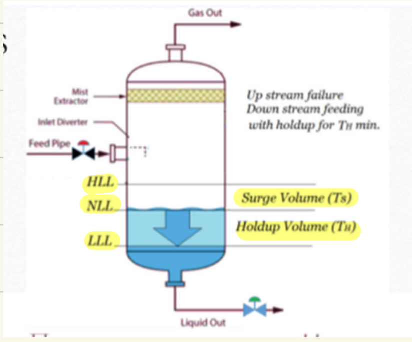
→ Horizontal is preferred when:

- Large liquid loads are involved, consequently hold-up will set the size
- Three phases are present.

Hold-up & surge volumes

Hold up: time it takes to reduce the liquid level from normal (NLL) to empty (LLL), while maintaining a normal outlet without feed makeup.

Surge: time it takes for the liquid level to rise from normal (NLL) to maximum (HLL), while maintaining a normal feed without any outlet flow



Size of vertical drum

Step 1: calculate the vapor liquid separation factor

$$F_{LV} = \frac{W_L}{W_V}\sqrt{\frac{\rho_V}{\rho_L}}$$

* W_L & W_V : liq & vap flow rates (mass /time)

Step 2: Calculate the vapor velocity factor

$$K_{drum} = \exp\left[A + B\ln F_{LV} + C(\ln F_{LV})^2 + D(\ln F_{LV})^3 + E(\ln F_{LV})^3\right]$$

A, B, C, D, & E → constants.

$0.1 < K_{drum} < 0.35$

Step 3: Calculate the maximum vapor velocity u_{perm}

$$u_{perm} = K_{drum}\sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

* u_{perm}: max permitted vapor velocity (ft/s)

→ Usually for safety issues: $U_V = 0.75\, u_{perm}$

Step 4: Find Drum cross-sectional Area from vapor velocity & flow rates

$$D_{VD} = \sqrt{\frac{4\,Q_V}{\pi\, U_V}} \quad \text{(ft)}$$

← vapor volumetric flow rate

$\dot{m} = \rho v A_c$

$A_c = 4\pi D^2$

$W_L = \rho Q_L$

$$Q_V = \frac{W_V}{3600\,\rho_V} \quad (\mathrm{ft^3/s})$$

conversion h→s

$$D_{vert} = \sqrt{\frac{4A_c}{\pi}}$$

* 3 to 6 inches need to be added to the Diameter if the mist eliminator is present in the vessel

Step 5: Approximate the vapor - liquid inlet nozzle based on the following criteria

$$\max V = \frac{100}{\sqrt{\rho}} \quad ft/sec$$

$$\min V = \frac{60}{\sqrt{\rho}} \quad ft/sec$$

or feed line nozzle diameter

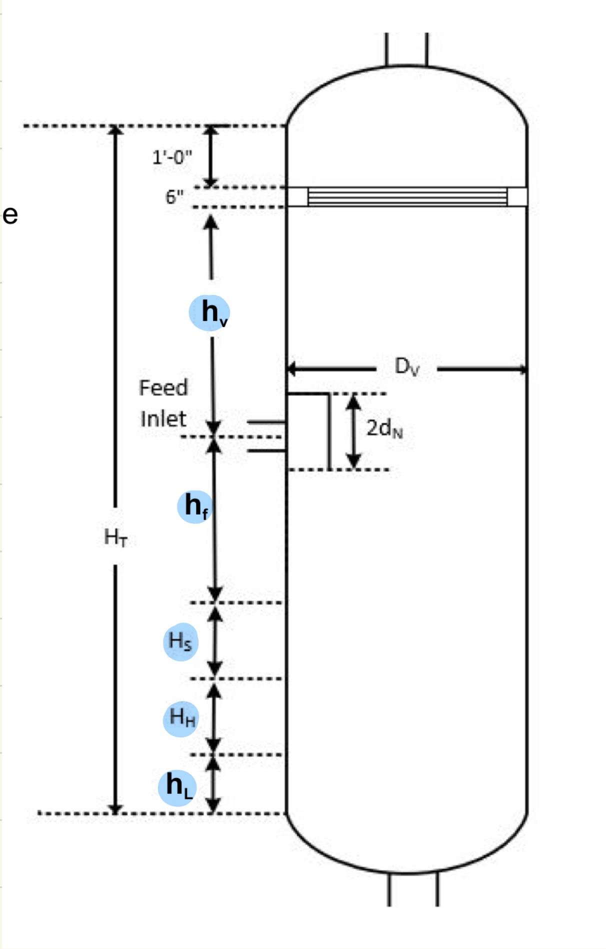
$$d_N \geq \left(\frac{4\,Q_m}{\frac{60\pi}{\sqrt{\rho_m}}} \right)^{0.5} \quad (ft)$$

$$\rho_m = \lambda \rho_L + \rho_V (1-\lambda) \quad (lb/ft^3)$$

$$\lambda = \frac{Q_L}{Q_L + Q_V}$$

$$Q_m = Q_L + Q_V \quad (ft^3/s)$$

Step 6: Make preliminary sizing as in the following figure:



h_v [if no mist eliminator is present]:

$h_v = 0.5 D_v$

$= 36" + \frac{1}{2}$ Diameter of feed line

h_v [if mist eliminator is present]:

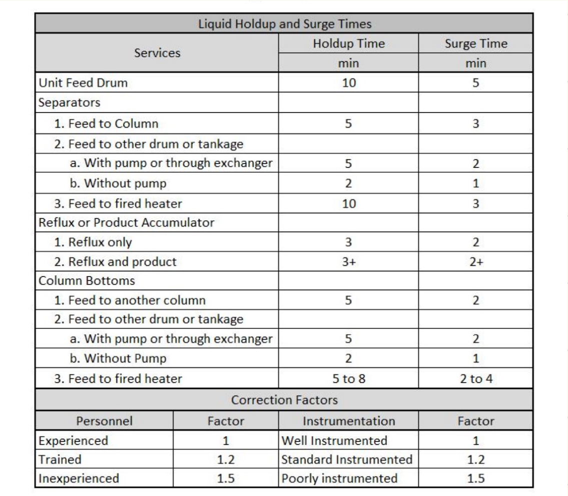
$h_v = 0.5 D_v$

$= 24" + \frac{1}{2} d_N$

h_f (with inlet Diverter) $= 12 + d_N$

h_f (without inlet Diverter) $= 12 + \frac{1}{2} d_N$

Step 7: Select hold up time from table to calculate holdup volume



Liquid Holdup and Surge Times

Services	Holdup Time	Surge Time
	min	min
Unit Feed Drum	10	5
Separators		
1. Feed to Column	5	3
2. Feed to other drum or tankage		
a. With pump or through exchanger	5	2
b. Without pump	2	1
3. Feed to fired heater	10	3
Reflux or Product Accumulator		
1. Reflux only	3	2
2. Reflux and product	3+	2+
Column Bottoms		
1. Feed to another column	5	2
2. Feed to other drum or tankage		
a. With pump or through exchanger	5	2
b. Without Pump	2	1
3. Feed to fired heater	5 to 8	2 to 4

Correction Factors

Personnel	Factor	Instrumentation	Factor
Experienced	1	Well Instrumented	1
Trained	1.2	Standard Instrumented	1.2
Inexperienced	1.5	Poorly instrumented	1.5

$$V_H = T_H \, Q_L$$

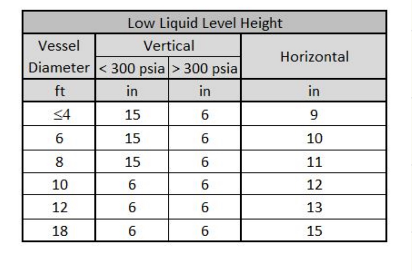
$$H_H = \frac{V_H}{\frac{\pi}{4} D_V^2}$$

Step 8: Select surge time from the table to calculate surge volume

$$V_S = T_S \, Q_L$$

$$H_S = \frac{V_S}{\frac{\pi}{4} D_V^2}$$

Step 9: select the low liquid level time from the table



Low Liquid Level Height

Vessel Diameter	Vertical		Horizontal
	< 300 psia	> 300 psia	
ft	in	in	in
≤4	15	6	9
6	15	6	10
8	15	6	11
10	6	6	12
12	6	6	13
18	6	6	15

h_L

Step 10: Calculate H_T

$$H_T = h_V + h_f + H_S + H_H + h_L + \underbrace{H_{ME}}_{\text{will be zero for no mist eliminator}}$$

Step II: Check geometry

$$3 < \frac{L}{D} < 5$$

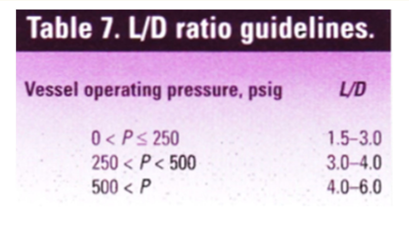
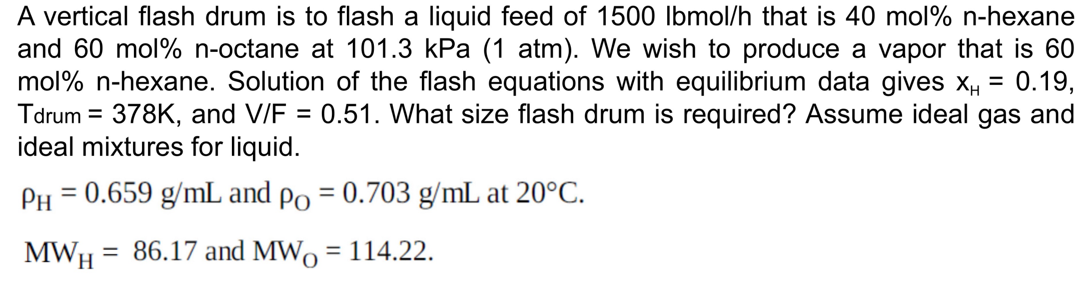


Table 7. L/D ratio guidelines.

Vessel operating pressure, psig	L/D
0 < P ≤ 250	1.5–3.0
250 < P < 500	3.0–4.0
500 < P	4.0–6.0

if $\frac{L}{D} < 3$ → increase V_{pool}

$\frac{L}{D} > 5$ → Horizontal Drum



A vertical flash drum is to flash a liquid feed of 1500 lbmol/h that is 40 mol% n-hexane and 60 mol% n-octane at 101.3 kPa (1 atm). We wish to produce a vapor that is 60 mol% n-hexane. Solution of the flash equations with equilibrium data gives $x_H = 0.19$, $T_{drum} = 378K$, and $V/F = 0.51$. What size flash drum is required? Assume ideal gas and ideal mixtures for liquid.

$\rho_H = 0.659$ g/mL and $\rho_O = 0.703$ g/mL at 20°C.

$MW_H = 86.17$ and $MW_O = 114.22$.

$y_h = 0.6$ $\quad y_o = 0.4$

$x_h = 0.19$ $\quad x_o = 0.81$

$\frac{V}{F} = 0.51$

Step 1: $F_{lv} = \frac{W_L}{W_V}\sqrt{\frac{\rho_V}{\rho_L}}$

$W_L = L(\overline{MW_L}) = (735)\left[0.19(86.17) + 0.81(114.22)\right] = 80034.5175$ lb/h

$\frac{V}{F} = 0.51$ $\quad F = 1500$ lbmol/h $\quad V = 765$ lbmol/h

$V + L = F$ $\quad L = 735$ lbmol/h

$W_V = V(\overline{MW_V}) = 765\left[0.6(86.17) + 0.4(114.22)\right] = 74503$ lb/h

ρ_L (ideal mix Assumtion) $\frac{1}{\rho_L} = \frac{x_H}{\rho_H} + \frac{x_o}{\rho_o} = \frac{0.19}{0.659} + \frac{0.81}{0.703} = 0.69$ g/mL

ρ_V (ideal gas Assumtion) $\rho_V = P\overline{MW_V}/RT = 1(97.39)/82.0575(378) = 3.14\times10^{-3}$ g/ml

$$F_{LV} = \frac{80034}{74503}\sqrt{\frac{3.14\times10^{-3}}{0.69}} = 0.0722$$

Step 2: K_{Drum}

$$K_{Drum} = \exp\left[-1.877478097 - 0.8145804597\ln(0.0722) - 0.1870744085(\ln 0.0722)^2 - 0.0145228667(\ln 0.0722)^3 - 0.0010148518(\ln 0.0722)^4\right]$$

$$K_{Drum} = 0.4433 \qquad 0.1 < K_{Drum} < 0.35$$

$$\Rightarrow K_{Drum} = 0.35$$

Step 3: U_{perm}

$$U_{perm} = K_{Drum}\sqrt{\frac{\rho_L - \rho_V}{\rho_V}} = 0.35\sqrt{\frac{0.69 - 3.14\times10^{-3}}{3.14\times10^{-3}}} = 5.2\ ft/s$$

$$U_V = 0.75\,U_{perm} = 0.75(5.2) = 3.9\ ft/s$$

Step 4: Cross-sectional Area

$$A_c = \frac{V(\overline{MW}_V)}{U_V(3600)\rho_V} = \frac{765\ lb\cancel{mol}}{\cancel{h}}\left|\frac{97.39\ \cancel{lb}}{lb\cancel{mol}}\right|\frac{\cancel{s}}{3.9\ \cancel{ft}}\left|\frac{\cancel{h}}{3600\ \cancel{s}}\right|\frac{\cancel{mL}}{3.14\times10^{-3}\ \cancel{g}}\left|\frac{454\ \cancel{g}}{\cancel{lb}}\right|\frac{ft^{\cancel{3}\ 2}}{28316.85\ \cancel{mL}}$$

$$A_c = 27.1\ ft^2$$

$$D = \sqrt{\frac{4A_c}{\pi}} = \sqrt{\frac{4(27.1)}{\pi}} = 5.87\ ft$$

$\rightarrow$ choose 6 ft

$*$ General rule of thumb $\frac{h_{tot}}{D} = 4\ ft$ $\qquad \frac{h_{tot}}{6} = 4 \qquad h_{tot} = 24\ ft$

Size a vertical separator with a mist eliminator pad to separate the following mixture .

W_v =145600 lb_m/h (ρ_v = 4.01 ~~ft^3/lb_m~~) lbm / ft3

W_L =46100 lb_m/h (ρ_L = 38.83 ~~ft^3/lb_m~~) lbm / ft3

+14.7 → psia

The operating pressure is 975 psig and the holdup and surge are to be 10 min and 5 min respectively. Use a design temperature of 650°F .

↳ to K

$P = 975 + 14.7$

$P_{sig} = 989.7$

1. $F_{lv} = \frac{W_L}{W_V}\sqrt{\frac{\rho_v}{\rho_L}}$

$$= \frac{46100}{145600}\sqrt{\frac{4.01}{38.83}} = 0.102$$

2. - $k_{drum} = 0.43$

$0.1 < K_{drum} < 0.35$

$40 \leq P \leq 5500$

- $K = 0.43 - 0.023 \ln P$

$$k = 0.43 - 0.023 \ln(989.7) = 0.2713$$

- $K_{GPSA} = 0.35 - 0.0001\,(P - 100)$

$$0.35 - 0.0001\,(989.7 - 100) \simeq 0.2625$$

$k = k_{min} = K_{GPSA} = 0.2625$

3. $u_{perm} = k\sqrt{\frac{\rho_L - \rho_v}{\rho_v}} = 0.2625\sqrt{\frac{38.83 - 4.01}{4.01}} = 0.77\ ft/s$

$U_v = 0.75\, u_{perm} = 0.75\,(0.77) = 0.58\ ft/s$

4. $D_{VD} = \sqrt{\frac{4\,Q_v}{\pi U_v}} = \sqrt{\frac{4\,(10.09)}{\pi\,(0.58)}} = 4.7\ ft \cong 5\ ft$

$Q_v = \frac{W_v}{\rho_v} = \frac{145600}{4.01 \times 3600} = 10.09\ ft^3/s$

5. $d_N \geq \left(\frac{\frac{4 Q_m}{60\pi}}{\sqrt{\rho_m}} \right)^{0.5}$

$Q_m = Q_L + Q_V = 10.09 + 0.33 = 10.42$

$Q_L = \frac{46100}{38.83 \times 3600} = 0.33$

$\rho_m = \lambda \rho_L + (1-\lambda)\rho_v \rightarrow 0.0316\ (38.83) + (1 - 0.0316)(4.01) = 5.11$

$\lambda = \frac{Q_L}{Q_L + Q_V} = \frac{0.33}{0.33 + 10.09} = 0.0316$

$$d_N \geq \left(\frac{\frac{4(10.42)}{60\pi}}{\sqrt{5.11}} \right)^{0.5}$$

$d_N \geq 0.71$ ft $= 8.52$ in ⇒ choose 10"

6. with mist eliminator

$h_v = 24'' + \frac{1}{2} d_N$

$24 + \frac{1}{2}(10) = 29''$

$h_f = 12 + \frac{1}{2}(10) = 17''$

7. $V_H = \tau_H Q_L$

$\tau_H = 10 \text{ min}$

$Q_L = \frac{46100}{38.83 \times 60} = 19.8 \text{ ft}^3/\text{min}$

$V_H = 10 \times 19.8 = 198 \text{ ft}^3$

$h_H = \frac{V_H}{\frac{\pi}{4} D_V^2} = \frac{198}{\frac{\pi}{4}(5)^2} = 10.1 \text{ ft} = 121.2''$

$V_S = \tau_s \, Q_L$

$5 \times 19.8 = 99 \text{ ft}^3$

$$H_S = \frac{99}{\frac{\pi}{4}(5)^2} = 5.04 \text{ ft} \longrightarrow 60.5''$$

$6'' + 1 \text{ ft} = 18''$

8. $H_T = h_v + h_f + H_H + H_S + h_L + H_{ME}$

$\hookrightarrow 6''$ (Table)

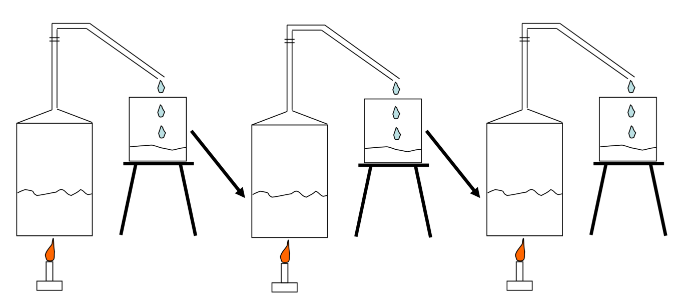
$H_T = 29 + 17 + 121.2 + 60.5 + 6 + 18 = 251.7 \text{ in} \rightarrow 20.9 \text{ ft}$

9. check geometry $\frac{L}{D} = \frac{20.9}{5} = 4.18$

$3 < L/D < 5$

lecture 3: Continous Distillation

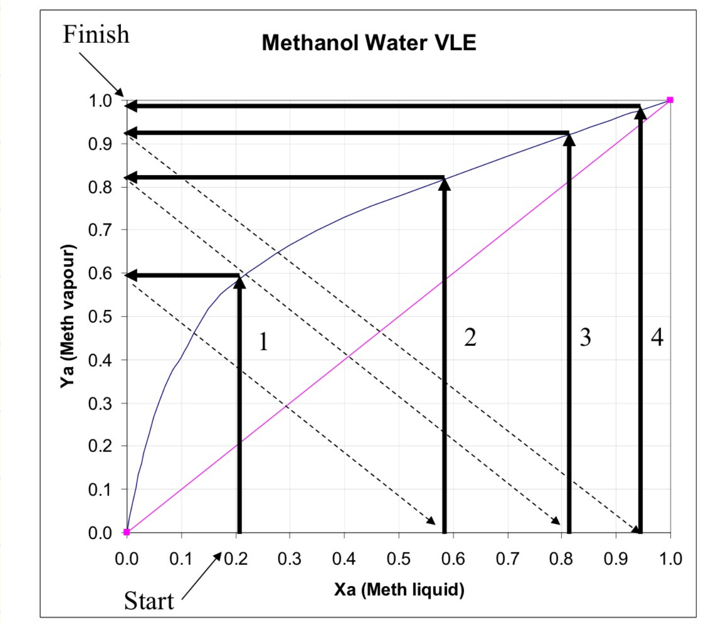
Binary mixture separation - Pot Still



- Biol the mixture, condense the vapor & collect the distillate. Repeat the procedure until the desired purity is obtained

Pros: Simple to make & operate, cheap component, Flexible; can collect over time before using

Cons: Small amount for high purity, large amount for low purity, Energy Intensive, slow



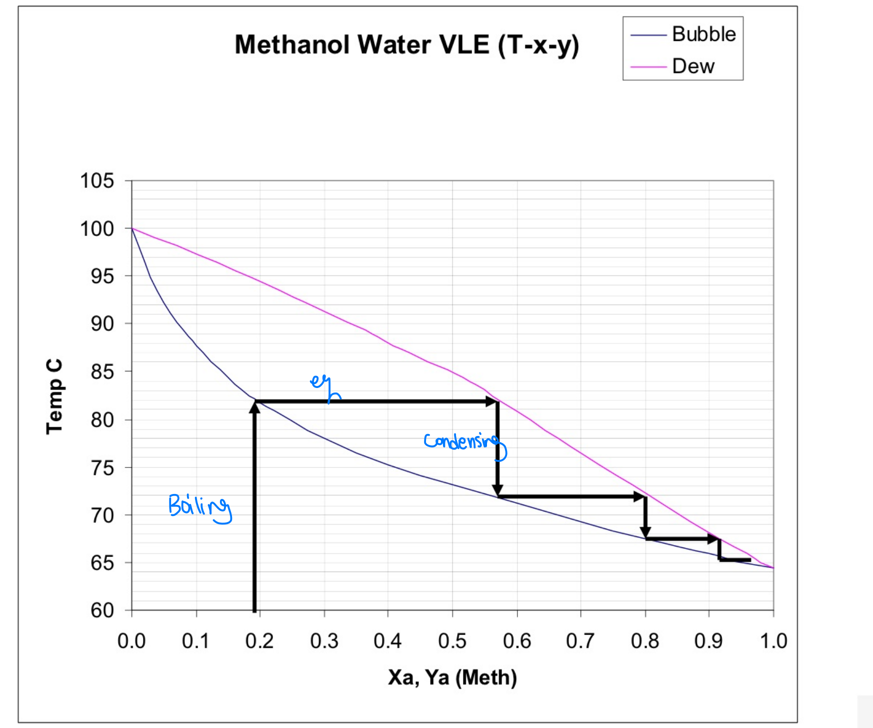
1- Boiling liq with $x_a = 0.2$ produces $y_a = 0.6$

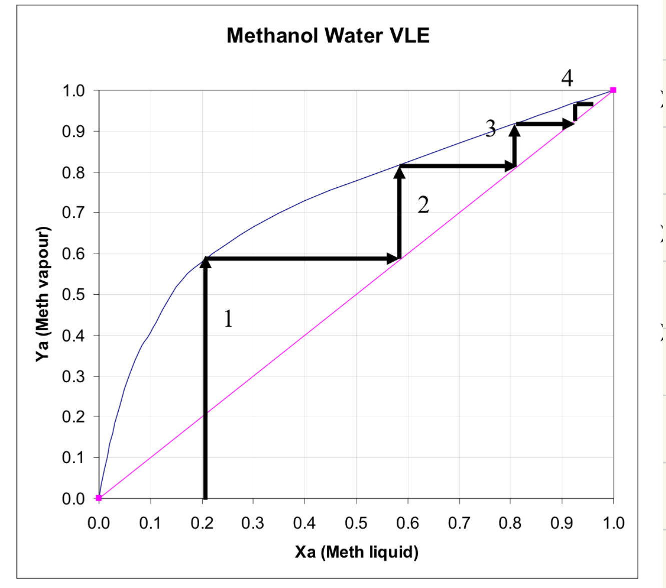
2- Boiling $x_a = 0.6$ produces $y_a = 0.8$

3- Boiling $x_a = 0.8$ produces $y_a = 0.9$

4- Boiling $x_a = 0.9$ produces $y_a = 0.98$

Alternative use of Txy Diagram



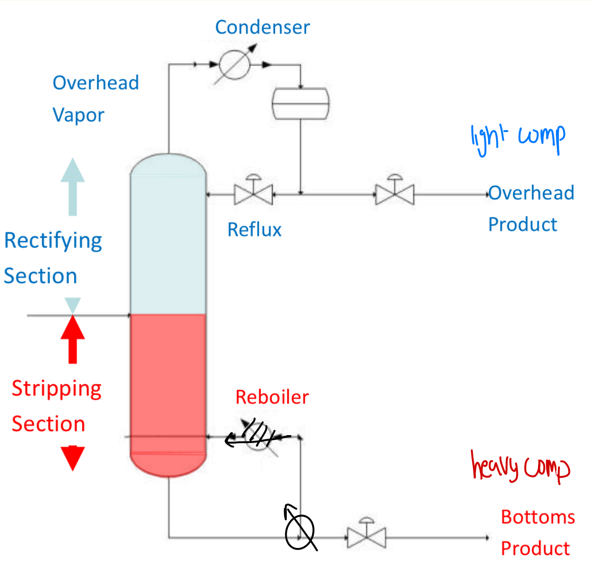


- Each still is a step on the x-y Diagram
  → Each step is an ideal stage in distillation

The Distillation Column

→ How to improve the pot still?

- Boiling results in a change of composition, & so does condensing
- Therfore, combine the two processes inside the column to improve the distillation processes
- Falling liquid meets rising vapor, boiling & condensing don't just occur in the reboiler & condensor, they happen inside the column also.



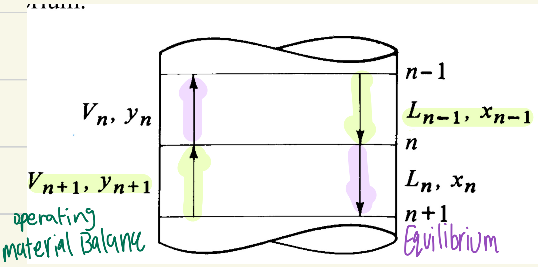
Assumptions of Analysis:

- Adiabatic [condensor, Reboiler, tower]
- Constant molar overflow [1mol condenses, 1 mol vaporize]
- Ideal Trays [outlets in equilibrium]

$$\frac{\text{ideal \# of Trays}}{\text{efficiency}} = \text{Actual \# of Trays}$$

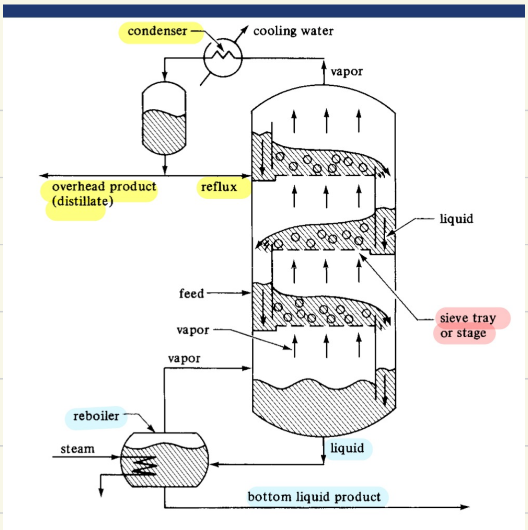
Rectification (fractionation), or stage distillation with reflux → processes in which a series of flash-vaporization stages are arranged in a series in such manner that the vapor & liquid products from each stage flow counter-currently to each other [multi stage, countercurrent separation processes]

Assuming the feed is a liquid at its boiling point, the feed flows down the stripping section to the bottom of the column, in which a definite level of liquid is maintained



- In each stage V & L enter, are mixed & equilibrated; exit streams are in equilibrium

✱ The concentration of the more volatile liquid (lower boiling point) is being increased in the vapor from each stage going upward & decreased in the liquid from each stage going downward

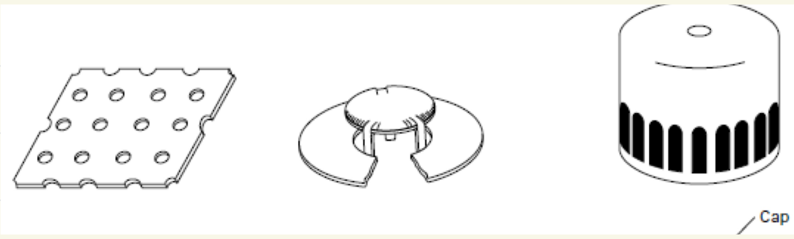


- The liquid bottom product is richer in the less volatile component (LVC), withdrawn from the reboiler
- The overhead vapor richer in the more volatile component (MVC), is completely condensed to a bubble point liquid distillate & a liquid reflux is returned to the top stage.

* Inside the column, the liqs & vapors are always at their bubble & dew points respectively, so that the highest temp is at the bottom, & the lowest at the top.

Plates or Trays, & Packing Materials

* liq level has a pressure head that is sufficient enough that it won't carry over by the vapor



Sieve tray or Perforation

valve cap

bubble cap

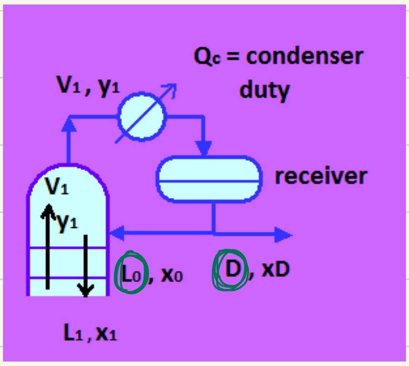
* Vapor velocity is high enough that liquid doesn't go through the caps

* Packing materials depend on allowable pressure drop & materials

↳ Sometimes used in exothermic reactions for cooling

Total & Partial Condensors

1- **Total condensor**: All saturated vapors at the top of the distillation column are condensed into sat. liq.



$$y_1 = x_1 = x_D$$

condensor duty ↰

$$Q_c = V_1 \lambda_{mix}$$ ↱ latent heat

$$\lambda_{mix} = y_1 \lambda_1 + y_2 \lambda_2$$

$$Q_c = (L_0 + D)\lambda_{mix}$$

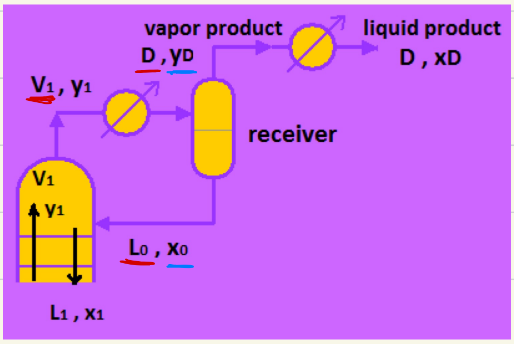
$$\frac{Q_c}{D} = \left(\frac{L_0}{D} + 1\right)\lambda_{mix}$$

$$\frac{Q_c}{D} = (R + 1)\lambda_{mix}$$

↳ Reflux Ratio

$$R = \frac{L_0}{D} = \frac{\text{liquid flow rate returning to the top plate}}{\text{Distillate flow rate}}$$

2- **Partial Condensor**: Acts as one plate with y_D in equilibrium with top plate condensate x_D



$$y_D P = x_0 P^{sat}$$

& $y_D = x_D$

- ∗ Partial condensor & Reboiler → Equilibrium stage
- ∗ Total condensor → not Equilibrium stage.

$V_1 = D + L_0$ → Material Balance

y_D & x_0 → Equilibrium

$y_D = x_D$

Reboilers.

1- Internal Reboiler.

Tubular heat exchanger built into the bottom of the tower provides large surface area

Cleaning requires a shut down of the distillation operation

2- External Reboilers:

A- Thermosiphon Reboiler: vertical thermosiphon with heating medium outside the tubes, can be operated so as to vaporize all the liquid entering it to produce a vapor of the same composition as the residue product, in which no enrichment is provided.

B- Kettle Reboiler: heating medium is inside the tubes, provides vapor to the tower in equilibrium with the residue product → behaves like a theoretical stage (large, better to control, long residence time)

C- Pipe still heaters: sometimes furnaces or pipe still heaters are used instead of reboilers

→ Used when no steam is available (hot oil or hot gas), high temp is needed

→ Expensive & not suitable for sensitive materials

Applications & Distillation Equipment

The most widely used large-scale method for separating homogeneous fluid mixtures in the chemical & petrochemical industry.

If no azeotropes are encountered, overhead & bottom products may be obtained at any desired purity

Suitable for the separation of liquid mixtures of components having similar boiling points (at low relative volatility $\alpha > 1.05$)

McCabe Thiele Method

- A mathematical - graphical method for determining the number of theoretical trays or stages needed for a given separation of a binary mixture

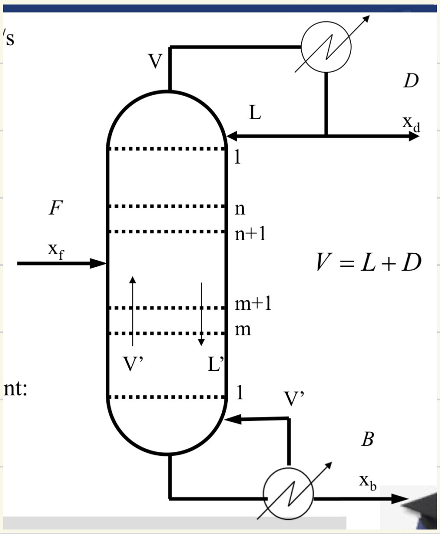
Assumptions & Simplifications:

1) constant molar enthalpies of vaporization (latent heat)
2) Heat capacity changes & heat of mixing are negligable compared to the heat of vaporization (ideal behaviour of binary mixture)
3) The condensor & Reboiler are well insulated so that heat losses to the environment are negligable.
4) Constant Pressure
   → Constant Molar overflow
5) kinetic & Potenial Energies are negligable
6) continous steady state operation
7) The streams leaving each stage are in vapor - liquid equilibrium

Column Mass Balance

Overall Material Balance

$$F = D + B$$



Component Material Balance

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

- A mixture of Acetic Acid and Acetic Anhydride containing 40 mol % Acetic Acid is to be separated by distillation. The top product is to be 90 mol % Acetic Acid and the bottom product 10 mol % Acetic Acid.
- The feed is heated to its boiling point. The vapour is condensed but not cooled and some is returned at a reflux ratio of 3 kmol/kmol product.
- Carry out a mass balance on this column

$x_F = 0.4$ $R = 3$

$x_D = 0.9$

$x_B = 0.1$

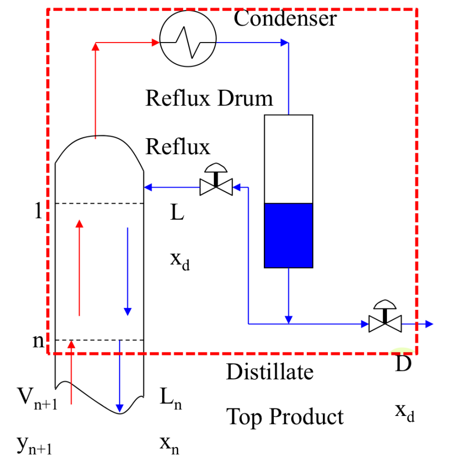
$F = D + B$ $x_F F = x_B B + x_D D$

$100 = D + B$ $0.4(100) = 0.1\ B + 0.9\ D$

$D = 37.5$

$B = 62.5$

The Rectifying Section



Reflux Ratio = Reflux / distillate

overall → $V_{n+1} = D + L_n$

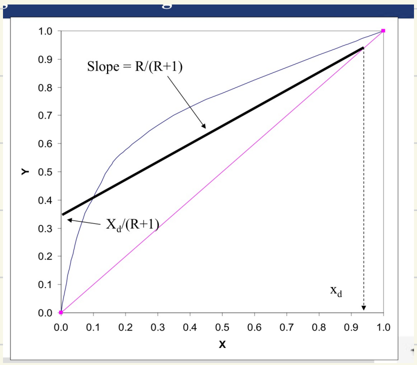
component → $V_{n+1}\, y_{n+1} = x_d D + x_n L_n$

operating line equation [enriching section] : $y_{n+1} = \frac{L_n}{V_{n+1}} x_n + \frac{D}{V_{n+1}} x_d$

↳ Valid from reflux to feed

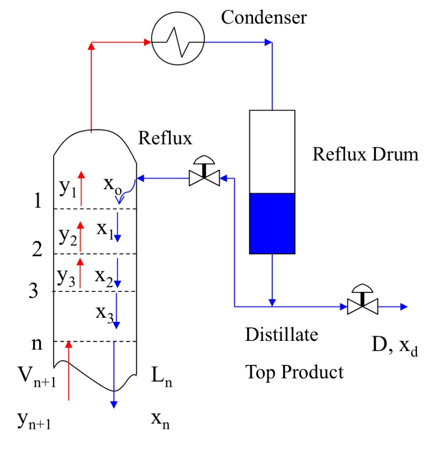
- Some condensed liquid is removed from the column as distillate, some is returned. The Reflux Ratio is the ratio of liquid returned to the coloumn over the amount removed

$R = \frac{L}{D}$ $L = DR$ & $V = L + D$



The rectifying operating line $y_{n+1} = \frac{R}{R+1} x + \frac{x_D}{R+1}$

↳ it intersects the $y = x$ line at $x = x_d$



Between stages → operating line from material Balance

$$y_1 = \frac{R}{R+1} x_0 + \frac{x_D}{R+1}$$

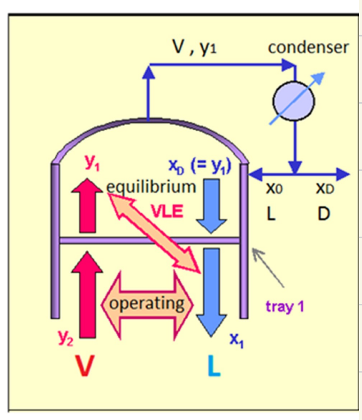
$$y_2 = \frac{R}{R+1} x_1 + \frac{x_D}{R+1}$$

On stages → Equilibrium

y_1 & x_1

y_2 & x_2

y_3 & x_3



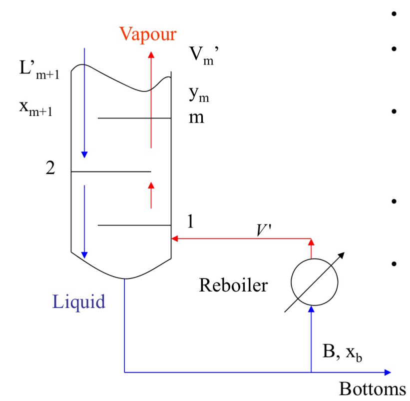
→ At the top of the column $x_0 = x_D = y_1$

& $y_1 = \frac{L}{V} x + \frac{D}{V} x_D$

y_1 & x_1 ⇒ VLE

→ Between stages ⇒ Operating

The Stripping Section



- Reboiler heats liquid to its BP & vapor rises
- Liquid that does not vaporise is removed as bottoms product
- Vapor leaving the partial reboiler is assumed to be in equilibrium with the Liquid bottoms product ⇒ partial Reboiler equilibrium stage

$$L' = V' + B$$

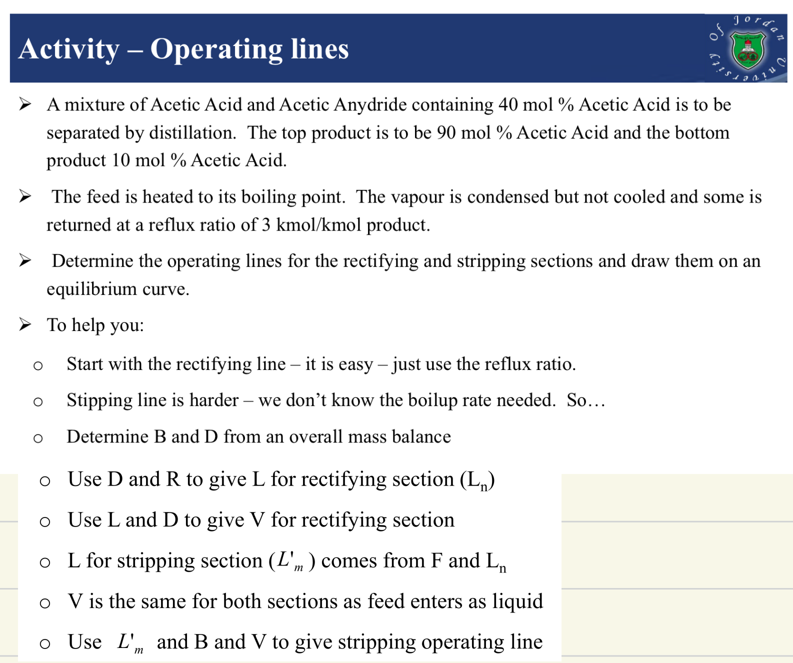
$$x L' = y V' + x_B B \rightarrow y = \frac{L'}{V'} x - \frac{B}{V'} x_B$$

* The Boilup Ratio → the ratio of vapor returning to the column to the bottoms flow rate.

$$V_B = \frac{V'}{B}$$

$$L = V' + B \Rightarrow L = (V_B + 1)B$$

$$\therefore\ y = \frac{V_B+1}{V_B}x - \frac{1}{V_B}x_B$$



Activity – Operating lines

- ➢ A mixture of Acetic Acid and Acetic Anhydride containing 40 mol % Acetic Acid is to be separated by distillation. The top product is to be 90 mol % Acetic Acid and the bottom product 10 mol % Acetic Acid.
- ➢ The feed is heated to its boiling point. The vapour is condensed but not cooled and some is returned at a reflux ratio of 3 kmol/kmol product.
- ➢ Determine the operating lines for the rectifying and stripping sections and draw them on an equilibrium curve.
- ➢ To help you:
  - ○ Start with the rectifying line – it is easy – just use the reflux ratio.
  - ○ Stipping line is harder – we don't know the boilup rate needed. So…
  - ○ Determine B and D from an overall mass balance
  - ○ Use D and R to give L for rectifying section (L_n)
  - ○ Use L and D to give V for rectifying section
  - ○ L for stripping section (L'_m) comes from F and L_n
  - ○ V is the same for both sections as feed enters as liquid
  - ○ Use L'_m and B and V to give stripping operating line

$x_F = 0.4$ Acetic Acid

$x_D = 0.9$

$x_B = 0.1$

$R = 3$

Enriching line $\quad y = \frac{R}{R+1}x + \frac{x_D}{R+1}$

$$y = \frac{3}{4}x + \frac{0.9}{4} \qquad y = 0.75x + 0.225$$

Stripping line $\quad y = \frac{L'}{V'}x - \frac{Bx_B}{V'}$

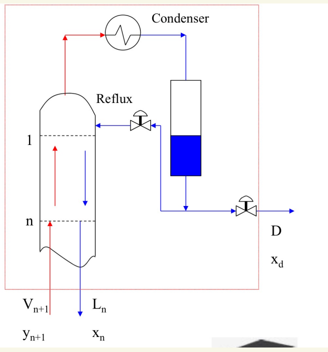
overall Balance $\quad F = D + B \qquad 100 = D + B$

component $\quad x_F F = x_D D + x_B B \qquad 0.4(100) = 0.9D + 0.1B$

$D = 37.5$

$B = 62.5$

$R = L/D$ $\quad 3 = L/37.5$ $\quad L = 112.5$



$V = L_n + D$

$V = 112.5 + 37.5$ $\quad V = 150$ mol/h

Saturated liq ⇒ $V' = V$

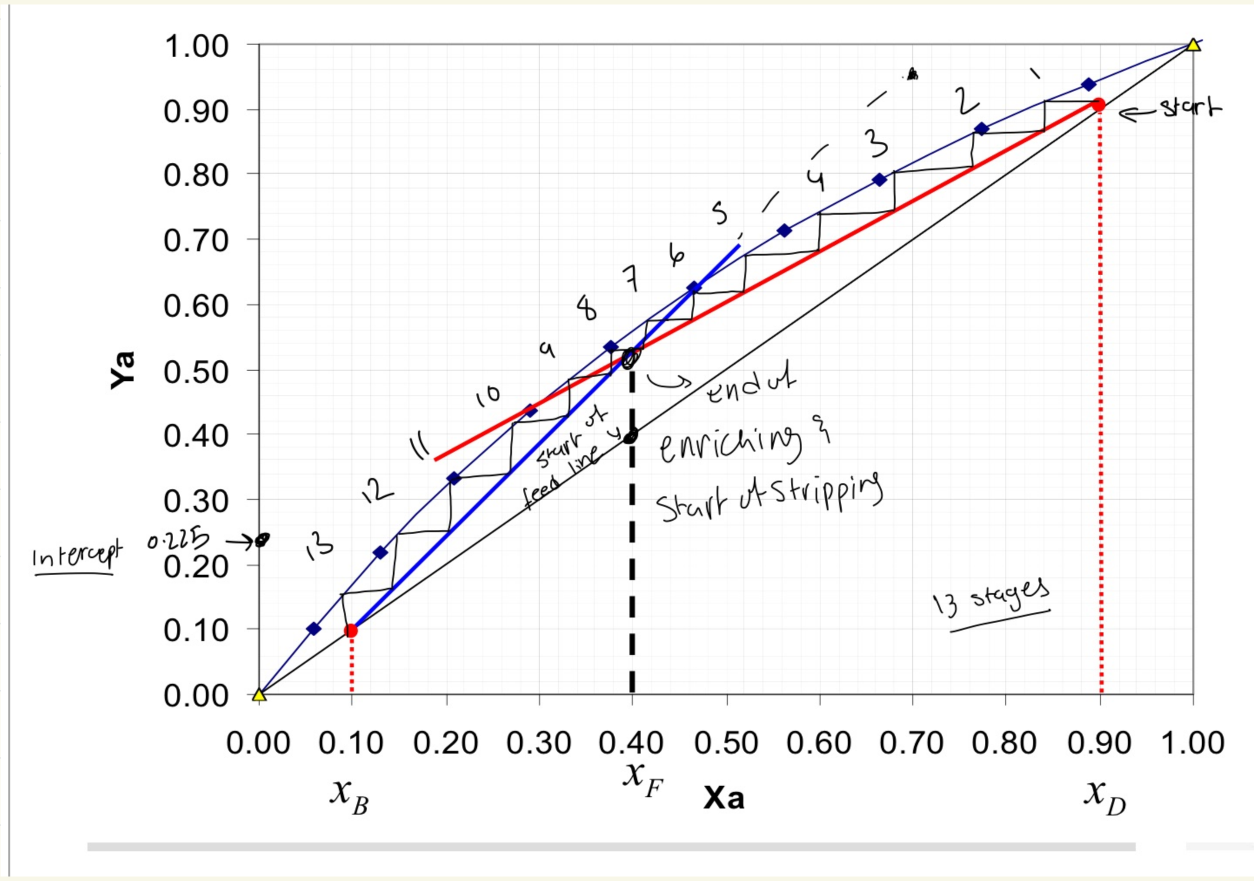
no vapor fraction in the feed

$L' = L_0 + qF$ $\quad q = 1$ (sat liquid)

$L' = 112.5 + 100 = 212.5$ mol/h

stripping operating line $\quad y = \frac{212.5}{150}x - \frac{62.5(0.1)}{150}$

$y = 1.42x - 0.042$



Feed Stage Consideration

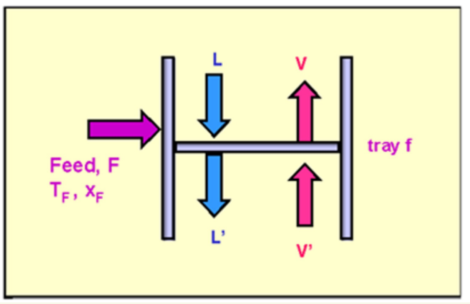
The point of intersection of the enriching & stripping operating line equations on the xy

$$V'y = L'x - Bx_b$$
$$Vy = Lx + Dx_D$$

subtracting the equations: $(V'-V)y = (L'-L)x - (Dx_D + Bx_B)$

$$(V'-V)y = (L'-L)x - Fx_F$$

$$y\frac{(V'-V)}{F} = \frac{(L'-L)}{F}x - x_f$$



∗ Material Balance on feed entrance

$$F + L + V' = V + L'$$

$$F + (V'-V) = (L'-L)$$

$L' = L + qF$

$L' - L = qF$

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

$$\therefore \quad q = \frac{(L'-L)}{F} \qquad q-1 = \frac{(V'-V)}{F}$$

$$y = \frac{q}{q-1}x - \frac{x_f}{q-1}$$ The q - line Equation

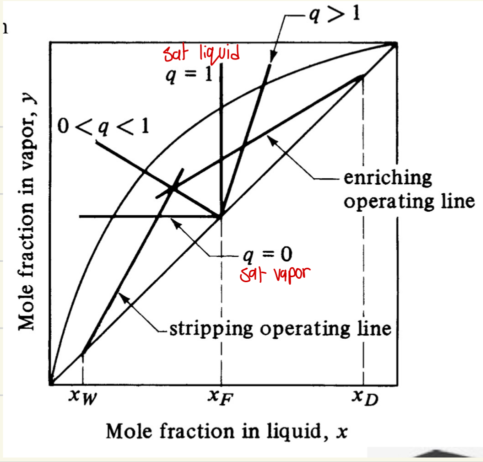
∗ q line slope for sat liq $= \infty$ (vertical)

∗ q line slope for sat vap $= 0$ (horizontal)

∗ q line slope for sat vap, liq mixture $0 < q < 1$

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

- The condition of the feed stream F entering the tower determines the relation between the Vapor in the stripping & V in enriching also L' & L.



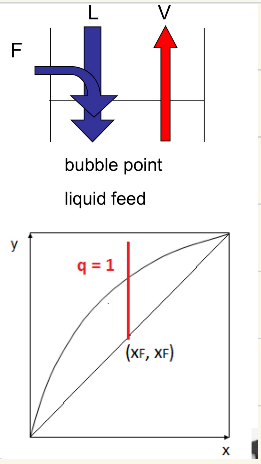
q = the enthalpy change needed to bring the feed to a dew point vapor divided by the enthalpy of vaporization of the feed

$$q = \frac{\text{Heat to vaporise 1mol of Feed}}{\text{Molar latent heat of vaporization of Feed}}$$

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

- H_V : Enthalpy of Feed at Dew point
- H_F = Enthalpy of Feed at Entrance Conditions
- H_L : Enthalpy of Feed at Bubble point

1- Saturated liquid Feed



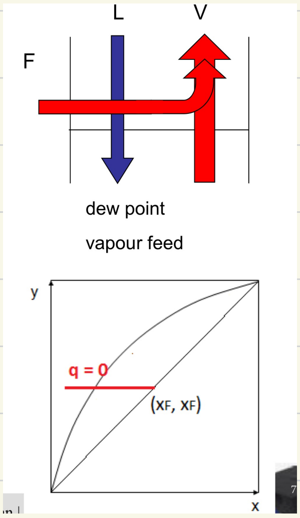
$$q = \frac{H_V - H_F}{H_V - H_L} \qquad * \ H_f = H_L$$

$$q = \frac{H_V - H_L}{H_V - H_L} = \frac{\lambda}{\lambda} = 1$$

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

vertical line

2- Saturated Vapor Feed



dew point vapour feed

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

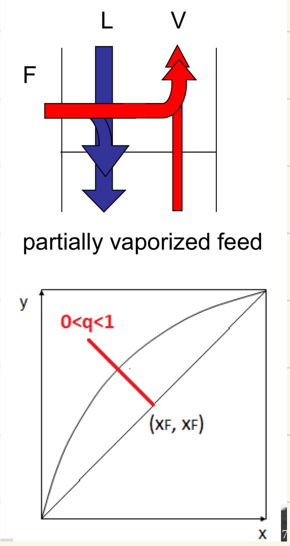
$* \; H_V = H_F$

$$q = \frac{H_V - H_V}{H_V - H_L} = 0$$

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

Horizontal line

3- Partial Vaporized Feed



partially vaporized feed

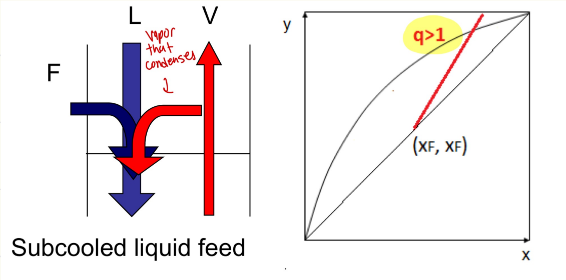
$$0 < q < 1$$

ex: $q = 0.7$

$$\text{q-line slope} \quad \frac{q}{q-1} = \frac{0.7}{-0.3}$$

Negative slope line

4- Subcooled Feed



Subcooled liquid feed

$$H_F < H_V$$

$$q = \frac{H_V - H_F}{H_V - H_L} = \frac{m C_p \Delta T - \lambda}{\lambda} > 1$$

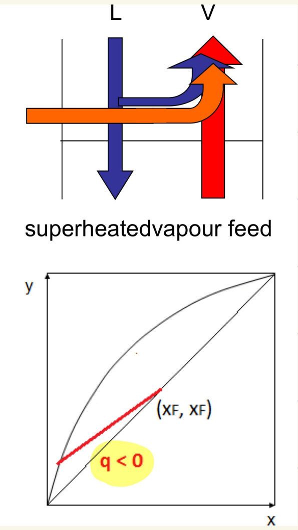
ex: $q = 1.5$

$$\text{q-line slope} = \frac{q}{q-1} = \frac{1.5}{0.5}$$

Positive slope (first quarter)

$$q = 1 + \frac{C_{P,L}(\overset{\text{BP}\downarrow}{T_b} - \overset{\downarrow\text{feed}}{T_F})}{\lambda}$$

5 - Super heated Feed



$H_F > H_V$

$$q = \frac{H_V - H_F}{H_V - H_L} = \frac{-}{+}$$

ex: $q = -1$

$$q\text{-line slope } \frac{q}{q-1} = \frac{-1}{-2} = \frac{1}{2}$$

Third quarter line

$$q = \frac{-C_{pV}\,(T_F - T_d)}{\lambda}$$

Lecture 4: Analysis of Binary Distillation McCabe Thiele Graphical Method

→ The McCabe Thiele Method Assumes:

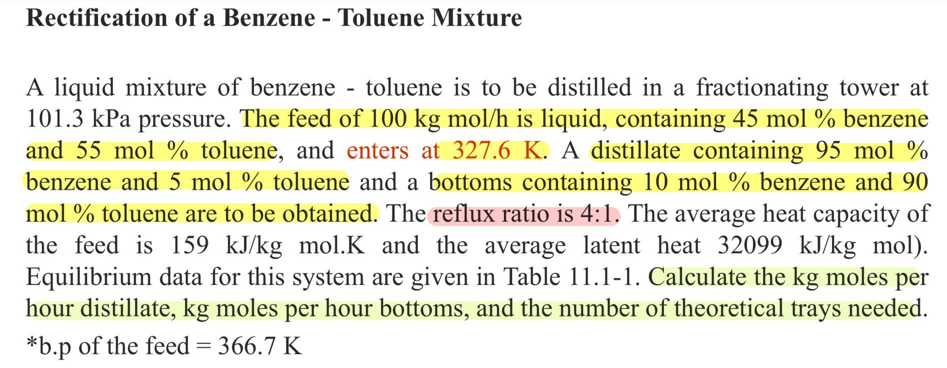
- Constant molal overflow
- No heat losses from column.

→ Not suitable if

- Relative volatility < 1.3 or > 5
- Reflux Ratio $< 1.1\ R_{min}$
- More than 25 theoretical stages needed

* To convert from theoretical stages to number of plates ⇒ Divide by the plate efficiency

Ex 11.4-1



Rectification of a Benzene - Toluene Mixture

A liquid mixture of benzene - toluene is to be distilled in a fractionating tower at 101.3 kPa pressure. The feed of 100 kg mol/h is liquid, containing 45 mol % benzene and 55 mol % toluene, and enters at 327.6 K. A distillate containing 95 mol % benzene and 5 mol % toluene and a bottoms containing 10 mol % benzene and 90 mol % toluene are to be obtained. The reflux ratio is 4:1. The average heat capacity of the feed is 159 kJ/kg mol.K and the average latent heat 32099 kJ/kg mol). Equilibrium data for this system are given in Table 11.1-1. Calculate the kg moles per hour distillate, kg moles per hour bottoms, and the number of theoretical trays needed.

*b.p of the feed = 366.7 K

Bp of feed = 366.7 K > Entering T = 327.6 ⇒ subcooled

overall Balance $F = B + D$ $100 = B + D$

$x_F F = x_B B + x_D D$ $0.45(100) = 0.1\,B + 0.95\,D$

$B = 58.8$ kgmol/h

$D = 41.2$ kgmol/h

Theoretical Stages needed:

Enriching operating line

$$y = \frac{R}{R+1} x + \frac{x_D}{R+1}$$

$$y = \frac{4}{4+1} x + \frac{0.95}{4+1}$$

$$y = 0.8x + 0.19$$

q-line

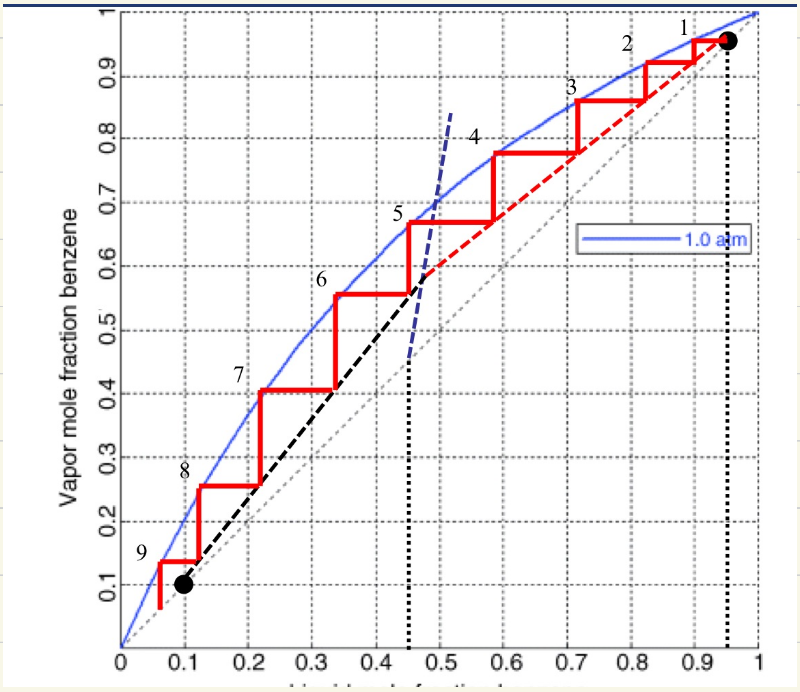
$$q = 1 + \frac{C_{p,L}(T_b - T_f)}{\lambda}$$

$$q = 1 + \frac{159(366.7 - 327.6)}{32099}$$

$$q = 1.195$$

$$y = \frac{q}{q-1} x - \frac{x_f}{q-1} \Rightarrow y = \frac{1.195}{1.195-1} x - \frac{0.45}{1.195-1}$$

$$y = 6.12x - 2.307$$



9 Stages

A distillation column operating at 1 atm is to be designed for separating an ethanol-water mixture. The feed is 20 mole% ethanol and the feed flow rate is 1000 kg-mole/hr of saturated liquid. A distillate composition of 80 mole% ethanol and a bottoms composition of not more than 2 mole% ethanol are desired. The reflux ratio is 5/3.

$x_F = 0.2$
$x_D = 0.8$
$x_B = 0.02$

$F = D + B$ $\qquad$ $1000 = D + B$

$x_F F = x_D D + x_B B$ $\qquad$ $0.2(1000) = 0.8\ D + 0.02\ B$

$D = 230.77$

$B = 769.23$

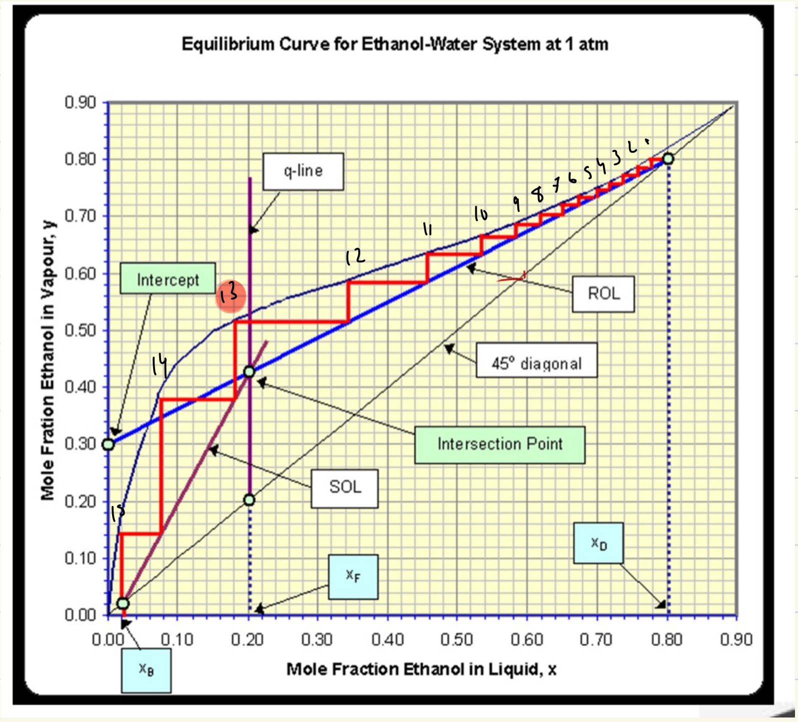
Enriching operating line

$$y = \frac{R}{R+1} x + \frac{x_D}{R+1}$$

$$y = \frac{5/3}{5/3+1} x + \frac{0.8}{5/3+1}$$

$$y = 0.625 + 0.3$$

q-line $\qquad$ $q = 1$ (sat liquid) ⇒ vertical line



15 stages

14 Trays $\qquad$ 1 Reboiler

stage 13 ⇒ optimum feed

Reflux Considerations

$$* \quad y = \frac{R}{R+1} x + \frac{x_D}{R+1} \qquad \uparrow R \ \uparrow \text{slope}$$

$$\therefore y = \frac{R}{R} x + \frac{x_D}{R} \qquad \Rightarrow \quad y = x$$

- The closer to the $y = x$ line (Tray n ↓)
- when ↑R the closer it is to the $y = x$ line ⇒ Tray n ↓
- At $y = x$ → Minimum number of stages.

$$R = \frac{\text{Flow returned as reflux}}{\text{Flow of top product taken off}}$$

* The rectifying operating line depends on R, therfore the required number of stages for given separation depends on R

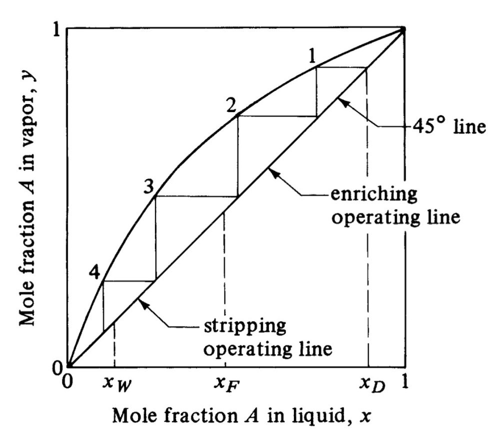
* Effective reflux ratio can be greater than R if the column is poorly insulated & may be subjected to change due to wether conditions

* The amount of reflux lies somewhere between the two extremes of total reflux & minimum reflux

Total Reflux: All condensate is returned, no product is taken off & no feed added [minimum number of stages]

Minimum Reflux: Reducing R requires more stages to achieve the separation, further reduction creates a pinch point where an infinite number of stages is required.

Minimum Reflux → when the intersection of the operating lines lie on the equilibrium curve.
100% total reflux at the begining of condensation



Total Reflux → when $R = \infty$, the operating lines of both sections coincide with the 45° diagonal line [$y = x$ line]

$$N_{min} = \frac{\ln\left[\frac{x_D(1-x_B)}{x_B(1-x_D)}\right]}{\ln \alpha_{AB}} - 1$$

Fenske Equation

Minimum number of stages ↲

$$\alpha_{AB} = \frac{P_A / x_A}{P_B / x_B}$$

mole fraction of Heavy comp in feed

mole fraction of light in bottom product

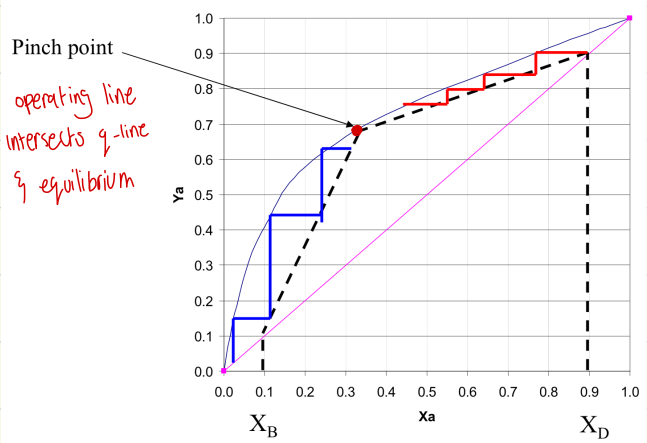
$$\log \frac{N_R}{N_S} = 0.206 \log \left[\frac{z_H}{z_L} \cdot \frac{B}{D} \cdot \left(\frac{x_{B.L}}{x_{D.H}} \right)^2 \right]$$

Kirkbride Equation

mole fraction of light comp in feed

mole fraction of heavy in top product

Ratio of no. of stages in rectifying to stripping ↲



operating line intersects q-line & equilibrium

Minimum Reflux → Enriching & stripping line intersection with the q-line moves further from the 45° line & closer to the equilibrium line

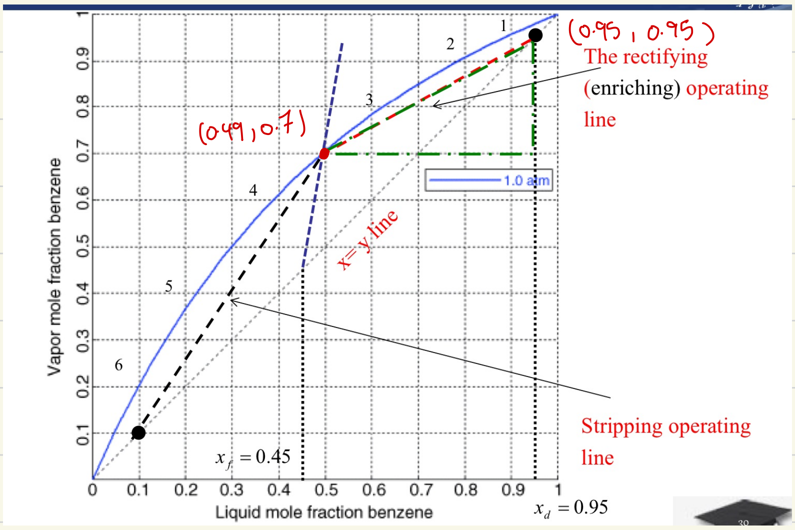
$$R_{optimum} = 1.2 \text{ to } 1.5 \, (R_{min})$$

$$\text{Slope} = \frac{R_{min}}{R_{min}+1} \qquad \text{Intercept} = \frac{x_D}{R_{min}+1}$$

* From the minimum reflux, & R/R_{min} we can determine the reflux R

→ when feed is at its boiling point (sat feed $q=1$)

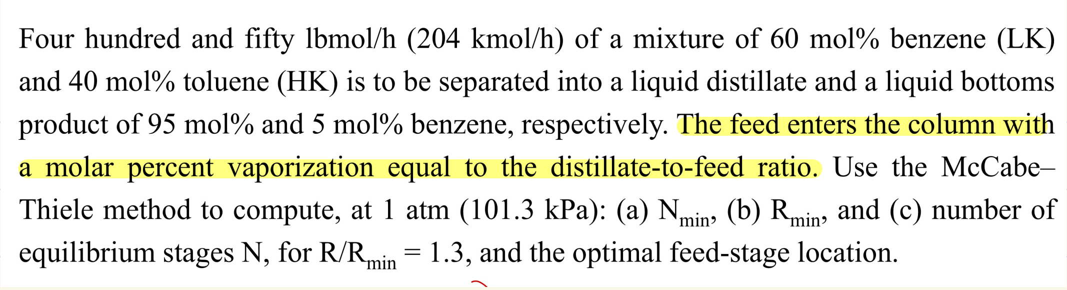
$$R_{min} = \frac{1}{(\alpha-1)}\left[\frac{x_D}{x_f} - \frac{\alpha(1-x_D)}{(1-x_F)}\right]$$



$$\frac{x_D - y'}{x_D - x'} = \frac{R_{min}}{R_{min}+1}$$

$$\frac{0.95 - 0.7}{0.95 - 0.49} = \frac{R_{min}}{R_{min}+1}$$

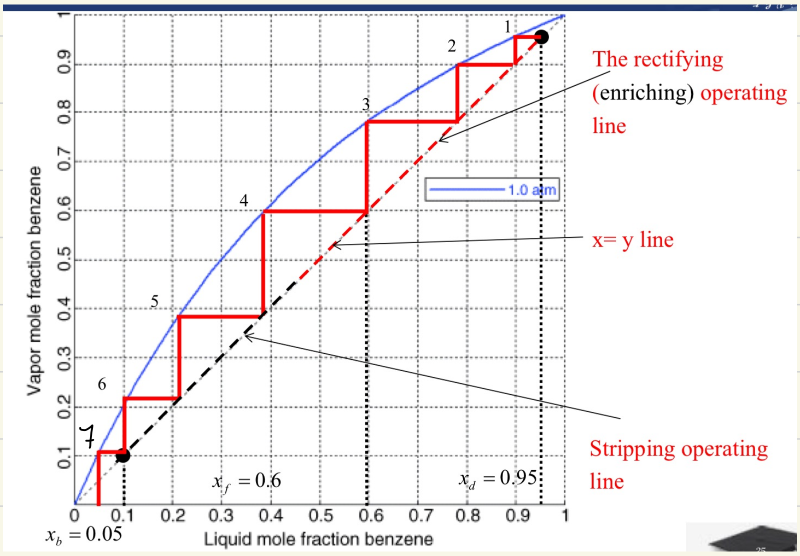
$$R_{min} = 1.19$$



Four hundred and fifty lbmol/h (204 kmol/h) of a mixture of 60 mol% benzene (LK) and 40 mol% toluene (HK) is to be separated into a liquid distillate and a liquid bottoms product of 95 mol% and 5 mol% benzene, respectively. The feed enters the column with a molar percent vaporization equal to the distillate-to-feed ratio. Use the McCabe–Thiele method to compute, at 1 atm (101.3 kPa): (a) N_{min}, (b) R_{min}, and (c) number of equilibrium stages N, for R/R_{min} = 1.3, and the optimal feed-stage location.

$x_F = 0.6$ $F = 204$ $x_D = 0.95$ $x_B = 0.05$ $(1-q) = D/F$

① N_{min} (operating lines = 45°)



7 stages

② Material Balances $F = D + B$ $204 = D + B$

$x_F F = x_D D + x_B B$ $0.6(204) = 095D + 0.05B$

$D = 124.66$

$B = 79.33$

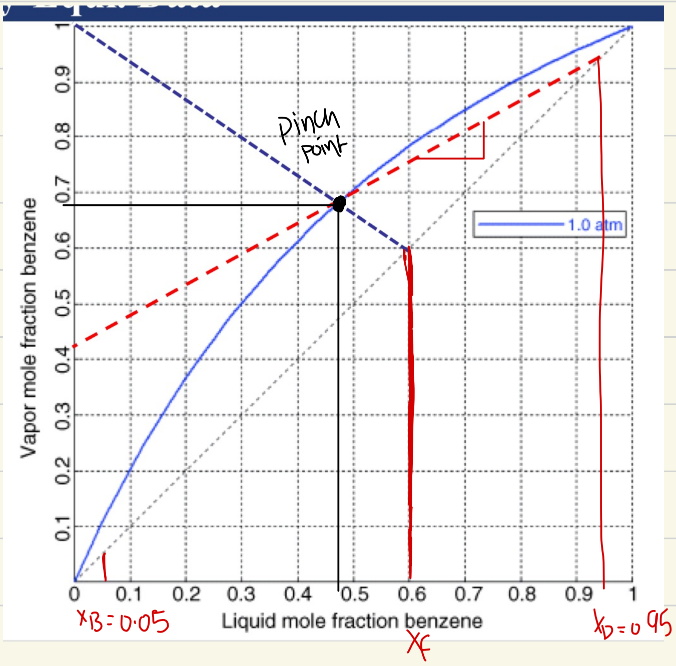
③ $(1-q) = D/F = \frac{124.66}{204}$ $q = 0.38$

q-line eq: $y = \frac{q}{q-1} x - \frac{x_F}{q-1}$

$y = \frac{0.38}{0.38-1} x - \frac{0.6}{0.38-1}$

$y = -0.61 x + 0.98$

q-line:



④ $$\frac{R_{min}}{R_{min}+1} = \frac{x_D - y'}{x_D - x'}$$

$$\frac{R_{min}}{R_{min}+1} = \frac{0.95 - 0.68}{0.95 - 0.48}$$

$R_{min} = 1.35$

⑤ when $R/R_{min} = 1.3$

$R/1.35 = 1.3$

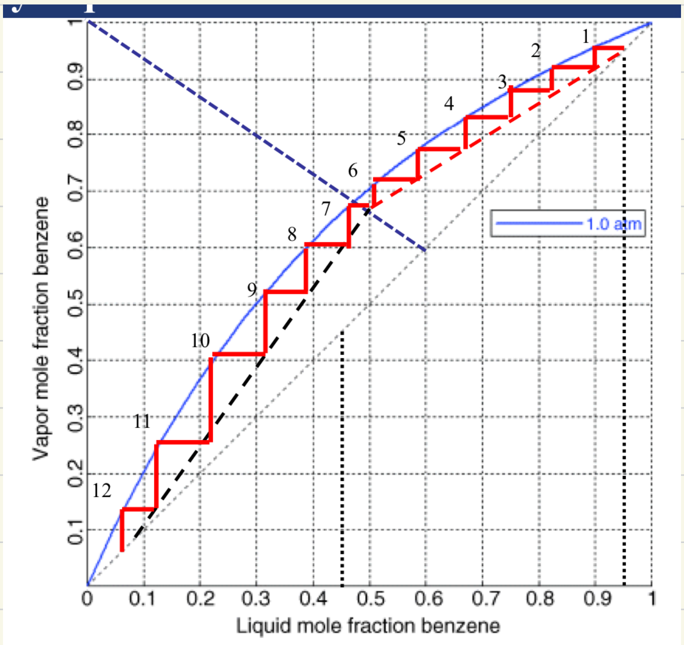
$R = 1.755$

⑥ Enriching operating line

$$y = \frac{R}{R+1}x + \frac{x_D}{R+1}$$

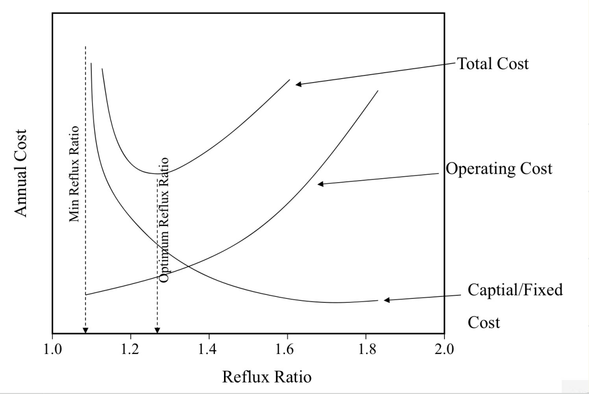
$$y = \frac{1.755}{2.755}x + \frac{0.95}{2.755}$$

$y = 0.637x + 0.3448$



* The further from y=x line, more stages.

Optimum Reflux Ratio



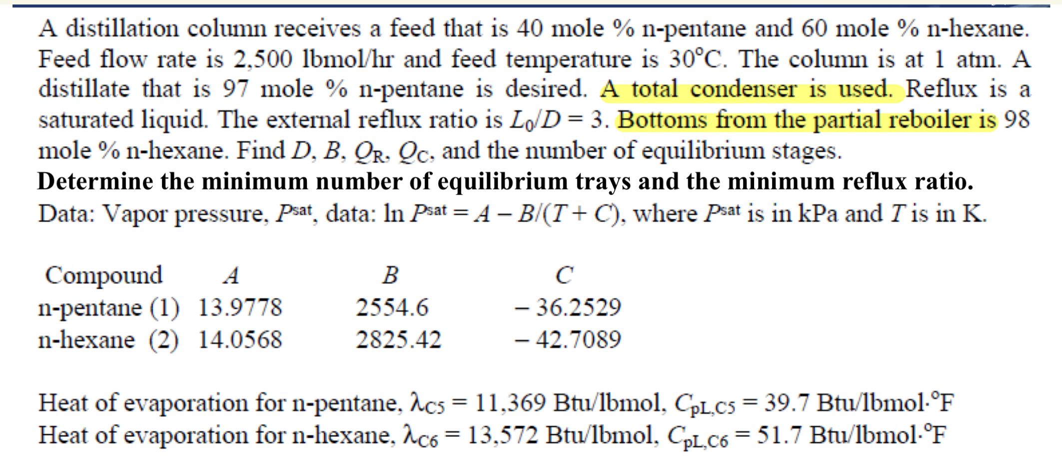
- ↑R ↓ stages ⇒ ↓ Height ↓ Capital cost
- ↑R ↑ D ⇒ ↑ volume ↑ length ↑ Capital Cost
- ↑R ↑ Heat exchanger → ↑ Boiling / Condensation ↑ Capital Cost

✳ Capital & operating costs combine to give a total cost, that is minimized by

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

✳ As R increases the slope approaches 1, More MVC is returned to the column

→ less material is removed as distillate, improving separation



A distillation column receives a feed that is 40 mole % n-pentane and 60 mole % n-hexane. Feed flow rate is 2,500 lbmol/hr and feed temperature is 30°C. The column is at 1 atm. A distillate that is 97 mole % n-pentane is desired. A total condenser is used. Reflux is a saturated liquid. The external reflux ratio is $L_0/D = 3$. Bottoms from the partial reboiler is 98 mole % n-hexane. Find D, B, Q_R, Q_C, and the number of equilibrium stages.

Determine the minimum number of equilibrium trays and the minimum reflux ratio.

Data: Vapor pressure, P^{sat}, data: $\ln P^{sat} = A - B/(T + C)$, where P^{sat} is in kPa and T is in K.

Compound	*A*	*B*	*C*
n-pentane (1)	13.9778	2554.6	– 36.2529
n-hexane (2)	14.0568	2825.42	– 42.7089

Heat of evaporation for n-pentane, $\lambda_{C5} = 11{,}369$ Btu/lbmol, $C_{pL,C5} = 39.7$ Btu/lbmol·°F
Heat of evaporation for n-hexane, $\lambda_{C6} = 13{,}572$ Btu/lbmol, $C_{pL,C6} = 51.7$ Btu/lbmol·°F

$X_F : 0.4$

$F :- 2500$

$X_D = 0.97$ pentane

R = sat liq

$R = L_0/D = 3$

$X_B : 0.98$ Hexane

$X_B = 0.02$ pentane

① Material Balance

$F = D + B$ $\qquad 2500 = D + B$

$x_F F = x_D D + x_B B$ $\qquad 0.4(2500) = 0.97\,D + 0.02\,B$

$D = 1000$ lbmol/hr

$B = 1500$ lbmol/hr

Enriching operating line

$$y = \frac{R}{R+1}x + \frac{X_D}{R+1}$$

$$y = \frac{3}{3+1}x + \frac{0.97}{3+1}$$

$$y = 0.75x + 0.2425$$

q - line (sat liquid)

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

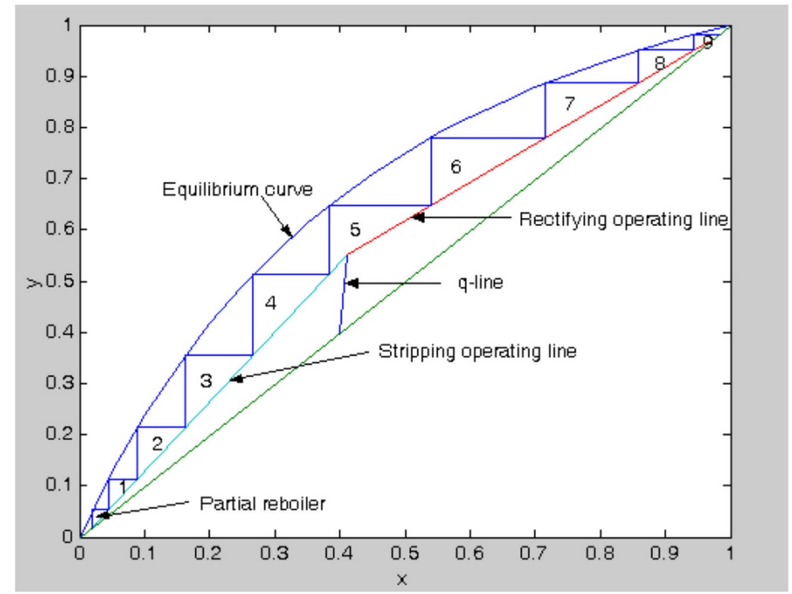
$C_{PL} = 0.4(39.7) + 0.6(51.7) = 49.6$

$H_V - h_L = 0.4(11369) + 0.6(13572) = 12691$

$$q = 1 + \frac{49.6\,(324.79 - 303.15)}{12691} = 1.08$$

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

$$y = \frac{1.08}{0.08}x - \frac{0.4}{0.08} \qquad y = 13.5x - 5$$



10 stages

9 Trays

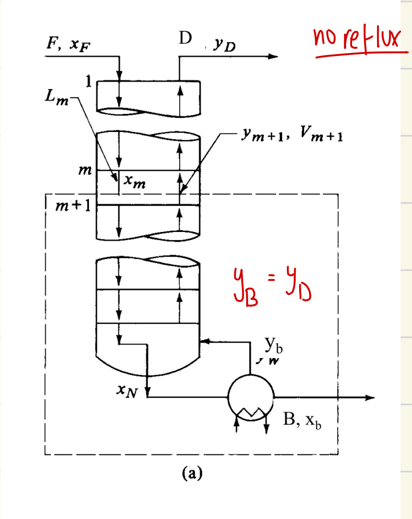
1 partial Reboiler.

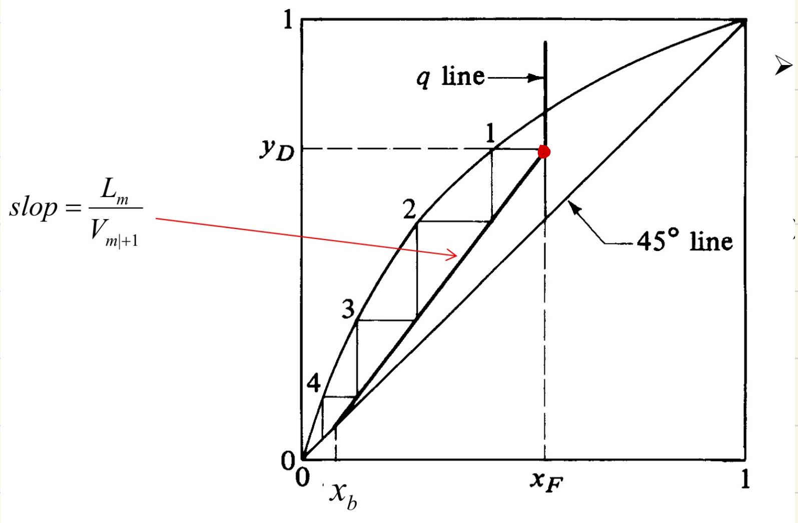
Special cases for Rectification Using McCabe Thiele Method

→ Stripping Column Distillation

- The feed is usually a saturated liq at the boiling point & the overhead product D is the vapor rising from the top plate, which goes to a condensor with no reflux or liq returned back to the tower
- The Bottom product usually has high conc. of the less volatile component
- The column operates as a stripping tower with the vapor removing the more volatile A from the liq as it flows downward

$$y = \frac{L}{V}x - \frac{Bx_B}{V}$$





if the feed is saturated liquid

$L_m = F$

if the feed is a subcooled liquid $(q>1)$

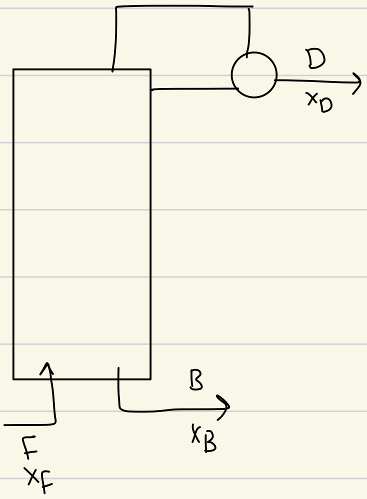
$L_m = qF$

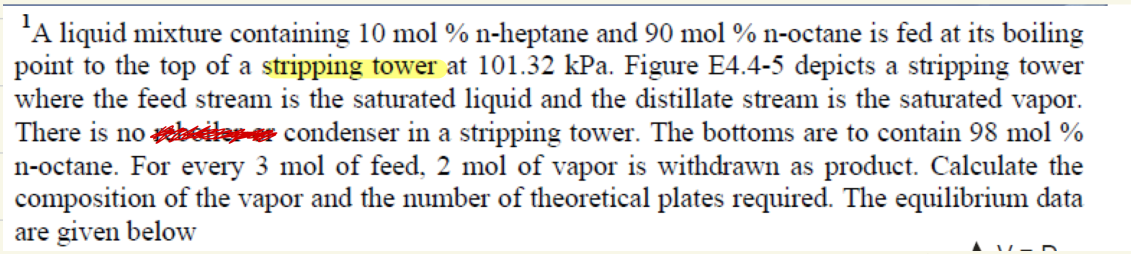
→ Enriching Column Distillation

- Enriching towers are sometimes used, where the feed enters the bottom of the tower as a vapor.
- The overhead product → produced in same manner as in complete fractionation tower [rich in MVC A]

$$y = \frac{L}{V}x + \frac{Dx_D}{V}$$

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





[1] A liquid mixture containing 10 mol % n-heptane and 90 mol % n-octane is fed at its boiling point to the top of a stripping tower at 101.32 kPa. Figure E4.4-5 depicts a stripping tower where the feed stream is the saturated liquid and the distillate stream is the saturated vapor. There is no ~~reboiler or~~ condenser in a stripping tower. The bottoms are to contain 98 mol % n-octane. For every 3 mol of feed, 2 mol of vapor is withdrawn as product. Calculate the composition of the vapor and the number of theoretical plates required. The equilibrium data are given below

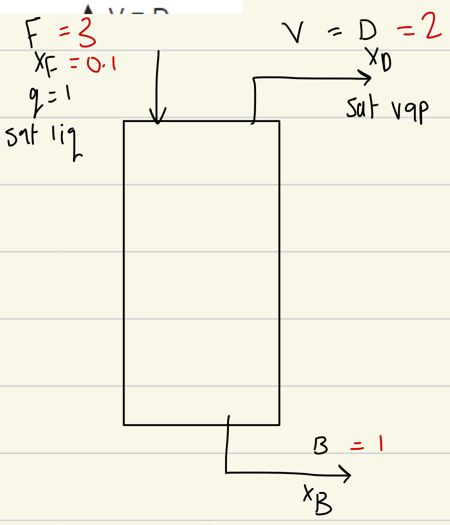
Material Balance $F = D + B$

$3 = 2 + B$ $\quad B = 1$

$x_F F = y_D D + x_B B$

$0.1(3) = y_D(2) + 0.02(1)$

$y_D = 0.14$



Stripping operating line $\quad y = \frac{L}{V}x - \frac{Bx_B}{V}$

$L = F$ (sat liq)

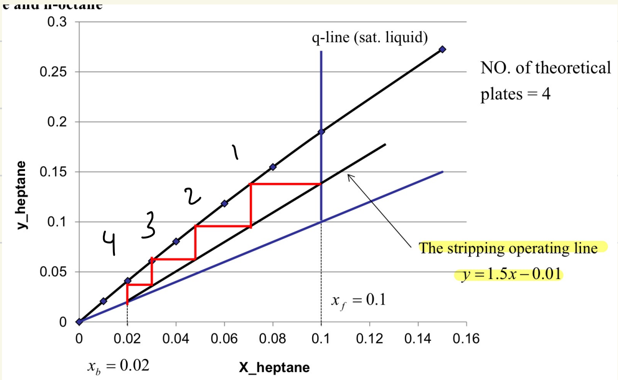
$V = D$ (sat vap)

$$y = \frac{3}{2}x - \frac{1(0.02)}{2}$$

$$y = 1.5x - 0.01$$

x_B octane = 0.98

x_B heptane = 0.02



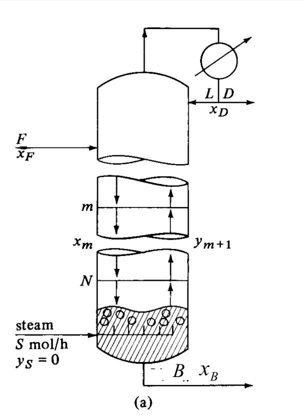
NO. of theoretical plates = 4

The stripping operating line $y = 1.5x - 0.01$

→ Direct Steam Injection

- Water is the heavy product & the light product is a mixture of volatile organics (small concentrations)
- Since volatility is a very strong function in temp, the high stripping temp inherent in steam stripping allow for the removal of heavier, more soluble organics that are not strippable with air

Rectification with Direct Steam Injection



(a)

- Generally, the heat to a distillation tower is applied to one side of a heat exchanger (Reboiler) & the steam does not directly contact the boiling solution
- When an ageous sol of more volatile A & water B is being distilled, the heat required may be provided by the use of open steam injected directly at the bottom of the tower

✳ The Reboiler is not needed

Overall Balance $F + S = D + B$

component Balance $x_F F + y_S S = x_D D + x_B B$

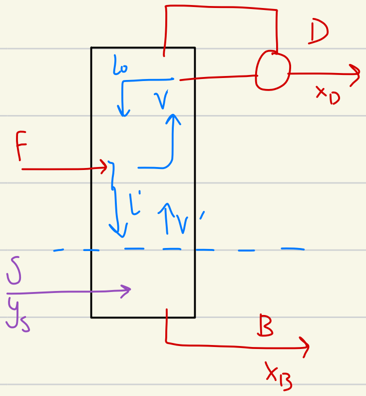
✳ No change in enriching operating line $y = \frac{L}{V} x + \frac{x_D D}{V}$

✳ Stripping line equation

overall $L' + S = V' + B$

component $x L' + y_S S = y V' + x_B B$

$$y = \frac{L'}{V'} x + \frac{y_S S - x_B B}{V'}$$



if stream is pure $y_S = 0$ $y = \frac{L'}{V'} x - \frac{x_B B}{V'}$

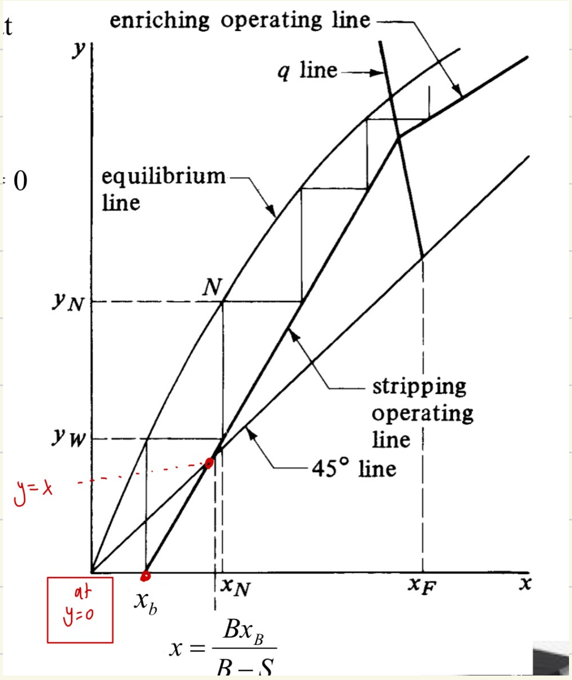
$L' = L_o + qF$

$V = V' + (1-qF)$

For sat steam entering $V' = S$ & $L' = B$

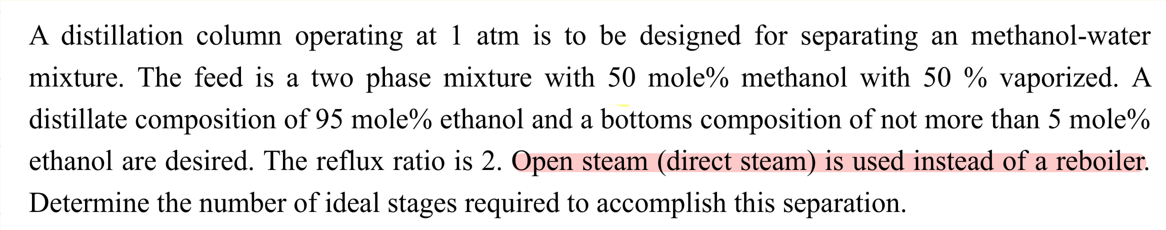
$$y = \frac{B}{S}x - \frac{x_B B}{S}$$

* line starts at x_B when $y = 0$



- Intersects the $y=x$ line at $x = \frac{Bx_B}{B-S}$

* For a given Reflux ratio (R) & overhead distillation comp (D), the use of open steam rather than closed requires an extra fraction of a stage, since the bottom step starts below the $y=x$ line



A distillation column operating at 1 atm is to be designed for separating an methanol-water mixture. The feed is a two phase mixture with 50 mole% methanol with 50 % vaporized. A distillate composition of 95 mole% ethanol and a bottoms composition of not more than 5 mole% ethanol are desired. The reflux ratio is 2. Open steam (direct steam) is used instead of a reboiler. Determine the number of ideal stages required to accomplish this separation.

$1 - q = 0.5$ $q = 0.5$

$x_D = 0.95$ $x_B = 0.05$ $R = 2$ $x_F = 0.5$

① Enriching operating line $y = \frac{R}{R+1}x + \frac{x_D}{R+1}$

$$y = 0.667x + 0.32$$

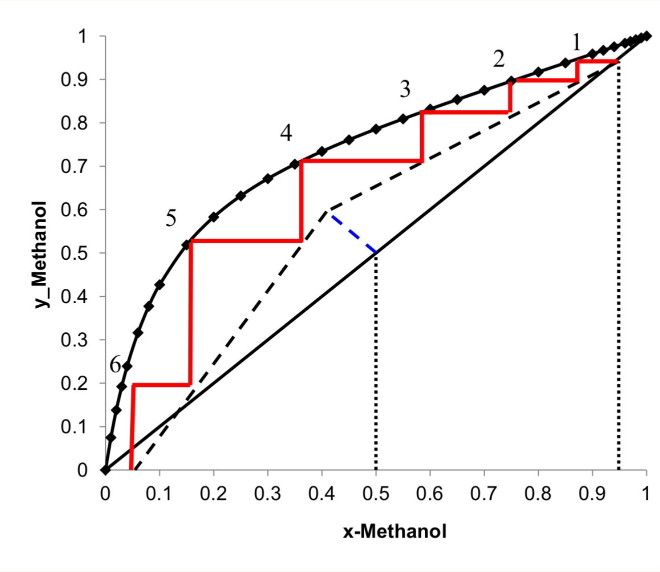
② q-line $y = \frac{q}{q-1}x - \frac{x_F}{q-1}$

$$y = -x + 1$$

③ Stripping line → from Intersection of Enriching & q-line

Start from $y = 0$ $x = x_B$

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



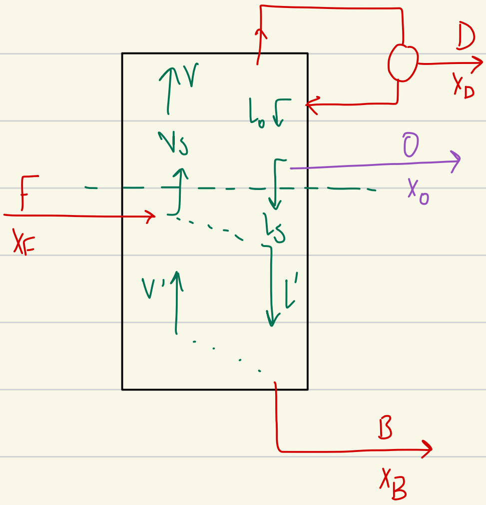
Direct Steam Injection ⇒ No reboiler

6 Trays

→ Rectification Column with side streams

In certain situations, intermedaite product or side streams are removed from sections of the tower between D & B

q-line equation is unaffected by the side stream



$V_s = V' + (1-q)F$ (Reboiler + feed)

$V = V_s$ (side stream doesn't affect it)

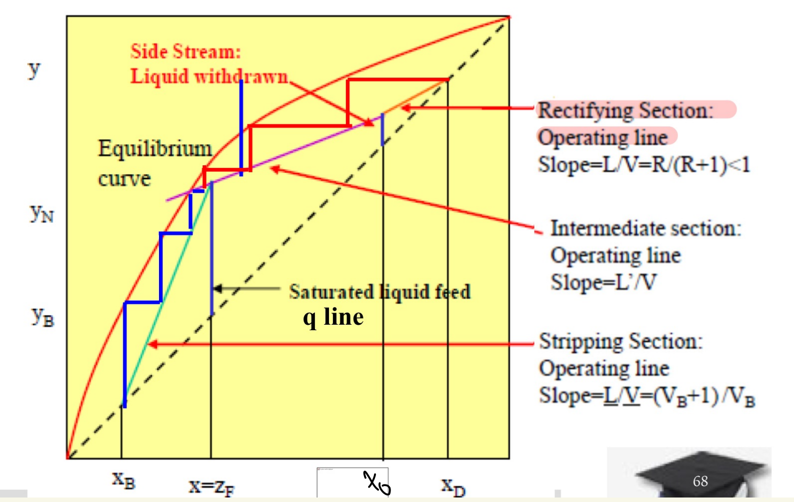
$L_s = L_o - O$ (from condensor - Removed by side stream)

$L' = L_s + qF$ (after side + Feed)

overall Balance on top part $V_s = L_s + O + D$

Component Balance $yV_s = xL_s + x_o O + x_D D$

$$y = \frac{L_S}{V_S} x + \frac{O x_O + D x_D}{V_S}$$



① Enriching [x_D to x_O]

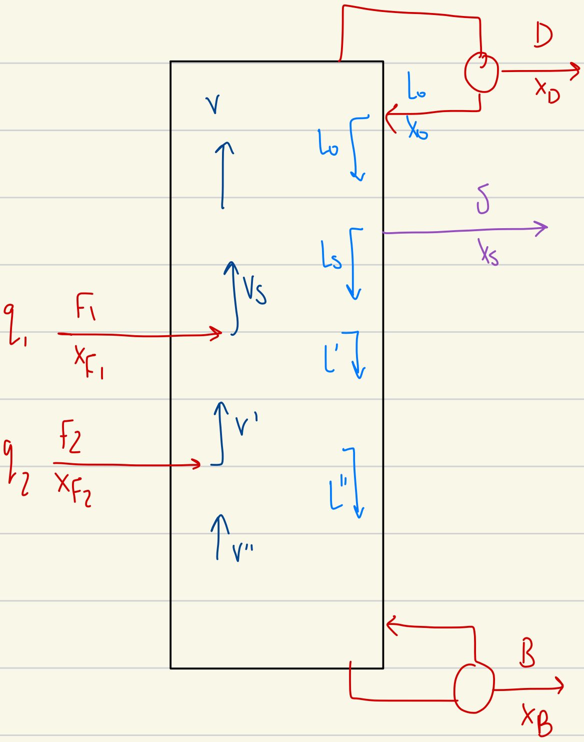
② side stream [x_O to q-line]

③ Stripping section [q-line to x_B]

→ Partial Condensers

- In normal distillation columns → total condensor
- Partial condensor → Equilibrium stage $y_D \neq x_R$

→ Multiple Feeds



$L_o = R/D$

$L_s = L_o - S$

$L' = L_s + q_1 F_1$

$L'' = L' + q_2 F_2$

$V' = V'' + (1 - q_2) F_2$

$V_s = V' + (1 - q_1) F_1$

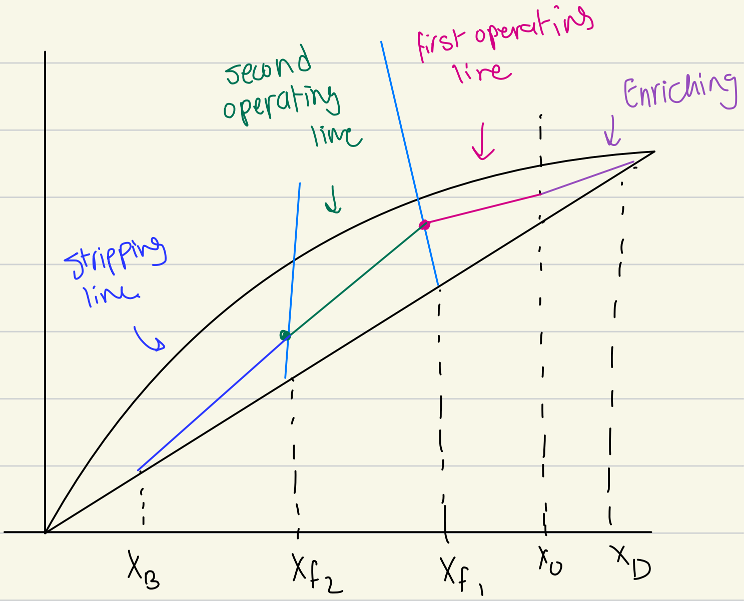
$V = V_1 = (R+1) D$

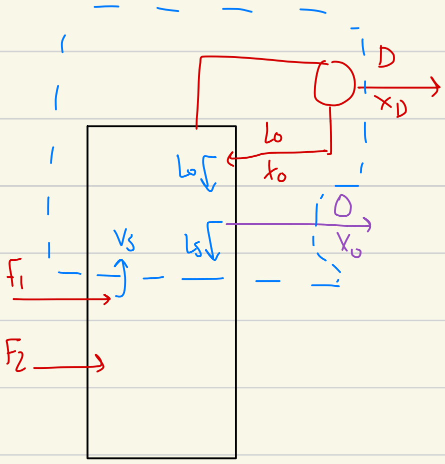
q_1 : $$y = \frac{q_1}{q_1 - 1} x - \frac{x_{F1}}{q_1 - 1}$$

q_2 : $$y = \frac{q_2}{q_2 - 1} x - \frac{x_{F2}}{q_2 - 1}$$

Enriching Operating line $$y = \frac{R}{R+1} x + \frac{x_D}{R+1}$$

Stripping operating line $$y = \frac{L''}{V''} x - \frac{B x_B}{V''}$$



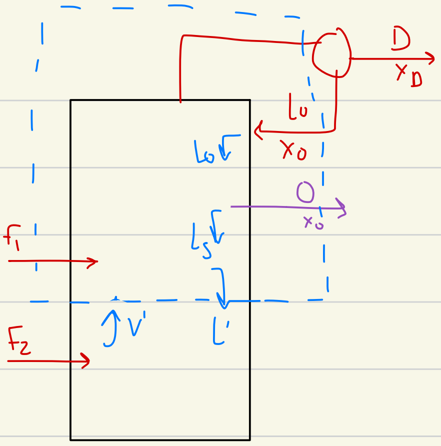


To find first operating line

Overall Balance $$V_S = L_S + D + O$$

Component Balance $$y_S V_S = x_S L_S + x_D D + x_O O$$

Operating line $$y = \frac{L_S}{V_S} x + \frac{D x_D + O x_O}{V_S}$$

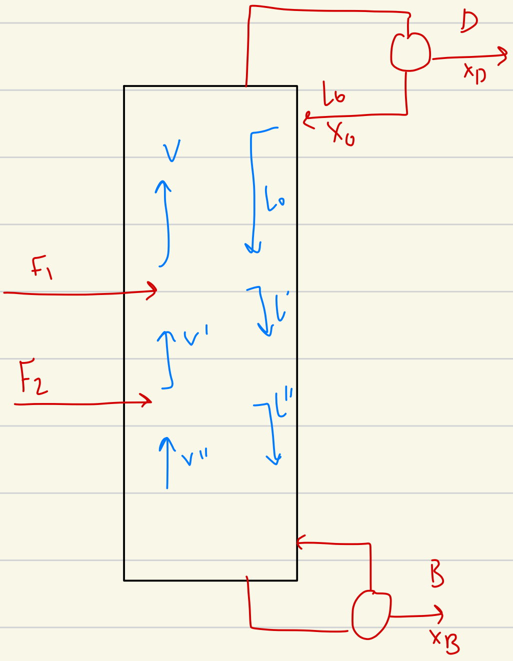


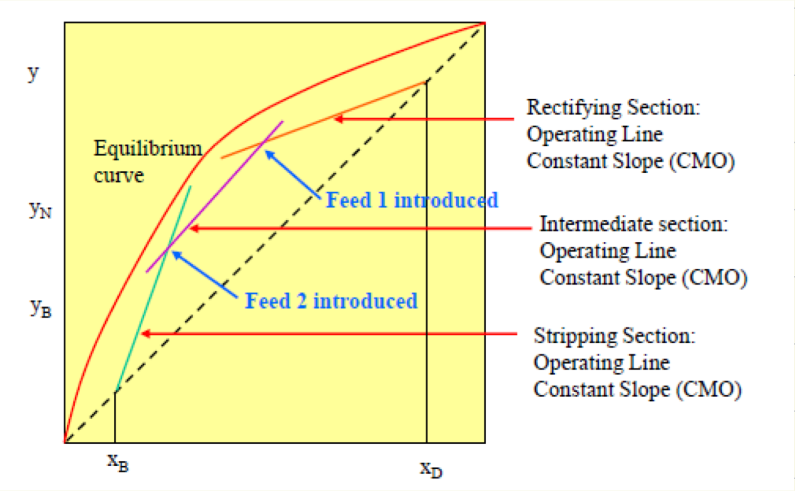
To find Second operating line

Overall Balance $V' + F_1 = L' + O + D$

Component Balance $yV' + x_{F_1}F_1 = xL' + x_O O + x_D D$

Operating line $y = \frac{L'}{V'}x + \frac{Ox_O + Dx_D - x_{F_1}F_1}{V'}$

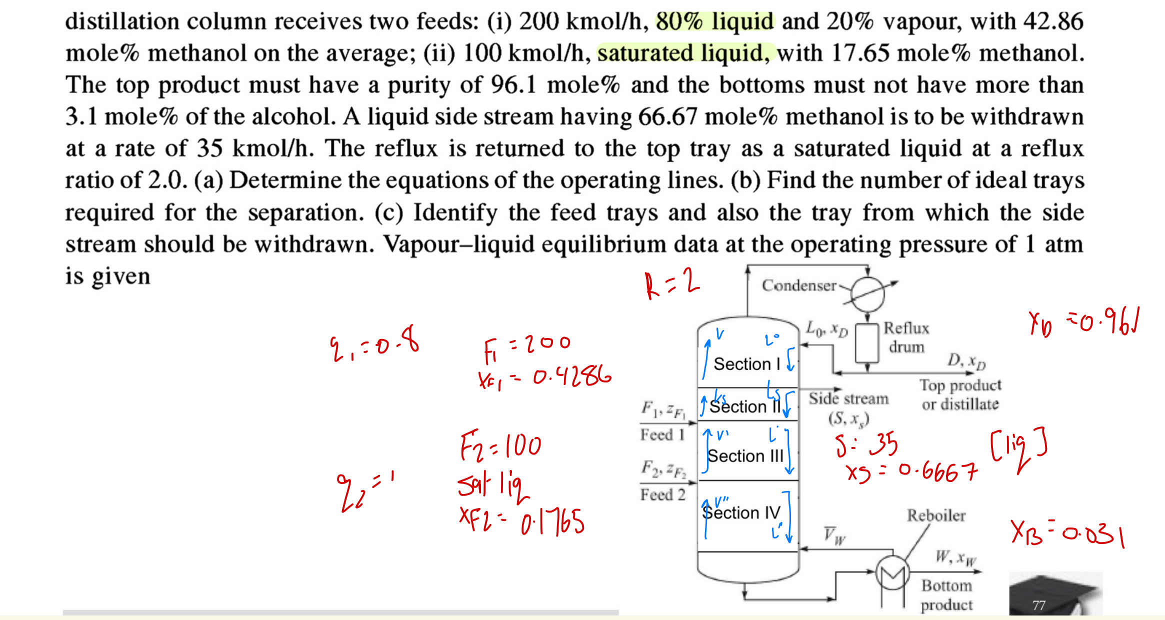




① Enriching operating line $y = \frac{L}{V}x + \frac{Dx_D}{V}$

② first operating line $y = \frac{L'}{V'}x + \frac{Dx_D - F_1x_{F_1}}{V'}$

③ stripping operation line $y = \frac{L''}{V''}x - \frac{Bx_B}{V''}$



distillation column receives two feeds: (i) 200 kmol/h, 80% liquid and 20% vapour, with 42.86 mole% methanol on the average; (ii) 100 kmol/h, saturated liquid, with 17.65 mole% methanol. The top product must have a purity of 96.1 mole% and the bottoms must not have more than 3.1 mole% of the alcohol. A liquid side stream having 66.67 mole% methanol is to be withdrawn at a rate of 35 kmol/h. The reflux is returned to the top tray as a saturated liquid at a reflux ratio of 2.0. (a) Determine the equations of the operating lines. (b) Find the number of ideal trays required for the separation. (c) Identify the feed trays and also the tray from which the side stream should be withdrawn. Vapour–liquid equilibrium data at the operating pressure of 1 atm is given

$q_1 = 0.8$

$F_1 = 200$
$x_{F_1} = 0.4286$

$F_2 = 100$
sat liq
$x_{F_2} = 0.1765$

$q_2 = 1$

$R = 2$

$x_D = 0.961$

$S = 35$
$x_S = 0.6667$ [liq]

$x_B = 0.031$

Overall Balance $F_1 + F_2 = D + B + S$ $200 + 100 = D + B$

component Balance $x_{F_1} F_1 + x_{F_2} F_2 = x_D D + x_B B + x_S S$

$0.4286(200) + 0.1765(100) = 0.961 D + 0.031 B + 0.6667(35)$

$D = 77.5$

$B = 187.75$

① Enriching operating line $y = \frac{R}{R+1} x + \frac{x_D}{R+1}$

$y = 0.667x + 0.3203$

② Operating line 1 $y V_S = x L_S + D x_D + S x_S$

$y = \frac{L_S}{V_S} x + \frac{D x_D + S x_S}{V_S}$

$L_S = L_0 - S$
$L_0 = RD$

$y = \frac{119.5}{231.75} x + \frac{77.5(0.961) + 35(0.6667)}{231.75}$

$L_0 = 2(77.5) = 154.5$
$L_S = 154.5 - 35 = 119.5$

$y = 0.5156x + 0.421$

$V_1 = V = (R+1) D$
$(2+1)(77.5) = 231.75$

③ Operating line 2

$$yV' + x_{F1}F_1 = xL' + Dx_D + Sx_S$$

$$y = \frac{L'}{V'}x + \frac{Dx_D + Sx_S - x_{F1}F_1}{V'}$$

$L' = L_S + qF_1$

$L' = 119.5 + 0.8(200) = 279.5$

$$y = \frac{279.5}{191.75}x + \frac{77.5(0.961) + 35(0.6667) - 200(0.4286)}{191.75}$$

$V' = V_S - (1-q)F_1$

$V' = 231.75 - (1-0.8)200 = 191.75$

$y = 1.4576x + 0.0618$

④ stripping line

$$y = \frac{L''}{V''}x - \frac{Bx_B}{V''}$$

$L'' = L' + qF_2$

$L'' = 279.5 + 100 = 379.5$

(Draw from Intersections)

$V'' = V' - (1-q)F_2$

$V'' = 191.75 - (1-1)100 = 191.75$

q line D

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

$$y = \frac{0.8}{0.8-1}x - \frac{0.4286}{0.8-1}$$

$y = -4x + 2.143$

Overall Efficiency

- Every tray is assumed to have the same efficiency
- Overall efficiency depends on:
  - → Geometry & design of the contacting trays
  - → Flow rates & flow paths of vapor & liquid streams
  - → Compositions & properties

Drickamer - Bradford empirical correlation

$$E_o = 13.3 - 66.8 \log(\mu)$$

↳ valid for hydrocarbon mixtures $342\,K < T < 488.5\,K$

$1\,atm < P < 25\,atm$

$0.066 < \mu < 0.355\ cP$

O'Connell correlation

$$E_o = 50.3\,(\alpha\mu)^{-0.226}$$

$$E_o = 0.52782 - 0.27511 \log_{10}(\alpha\mu) + 0.044923\,[\log_{10}(\alpha\mu)]^2$$

α = relative volatility

μ = viscosity

Murphee Efficiency

$$E_{MV} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} \qquad E_{MV} = \frac{x_n - x_{n+1}}{x_n^* - x_{n+1}}$$

y_n : Avg actual concentration leaving tray n

y_{n+1} : Avg actual concentration of time mixed vapor entering tray n

y_n^* = Conc. of vapor that would be in equilibrium with the liq conc. leaving the tray

Van Winkle's Correlation

$$E_{MV} = 0\,07\ \underset{\hookrightarrow \text{surface tension}}{Dg^{0.14}}\ \overset{\text{schmidt number}}{Sc^{0.25}}\ \underset{\hookrightarrow \text{Reynolds number}}{Re^{0.08}}$$

Lecture 5: Analysis of Binary Distillation by Ponchon - Savarit Method

- For a non-ideal system, where molar latent heat is not constant & there is substantial heat of mixing ⇒ Ponchon - Savarit method
- An Enthalpy concentration diagram for a vapor - liquid mixture of A & B takes into account latent heats, heats of solution or mixing, & sensible heats of the components

Data needed:

1. Heat capacity as a function of temperature, composition & Pressure.
2. Heat of mixing & dilution as a function of temp & composition.
3. Latent heats of vaporization as a function of temp, pressure, & composition
4. Bubble point temperature as a function of composition & pressure.

- Small heats of solutions (hydrocarbon mixtures) ⇒ neglected

Enthalpy - Composition Diagram

Saturated liquid enthalpy:

$$H_x = x_A C_{PLA} (T_b - T_0) + (1 - x_A) C_{PLB} (T_b - T_0) + \Delta H_{sol}$$

T_b : boiling Temp

T_0 : reference Temp

C_{PLA}: Heat capacity of A

C_{PLB}: Heat capacity of B

ΔH_{sol}: Heat of mixing (usually neglected)

Saturated vapor Enthalpy

$$H_y = y_A \left[\Delta H_{vA} + C_{pvA} (T_d - T_0) \right] + (1 - y_A) \left[\Delta H_{vB} + C_{pvB} (T_d - T_0) \right]$$

T_d : Dew Temperature
T_0 : Reference Temperature
C_{pvA} : Vapor Heat capacity of A
C_{pvB} : Vapor Heat capacity of B
ΔH_{vA} : latent heat of A at the reference T_0 & calculated from that at the normal boiling Temp (T_{bA})

- $\Delta H_{vA} = C_{pLA} (T_{bA} - T_0) + \Delta H_{vAb} - C_{pvA} (T_{bA} - T_0)$

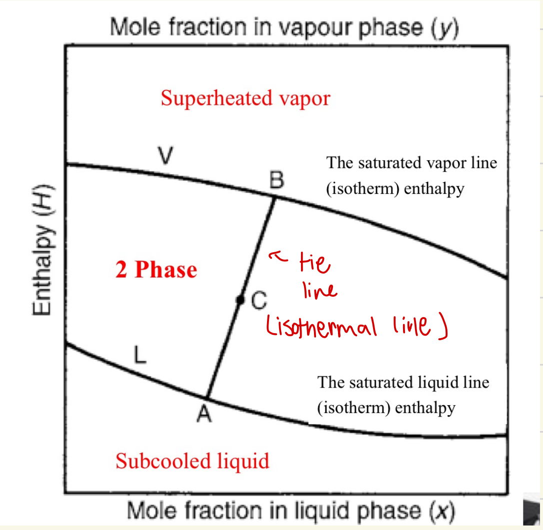
- $\Delta H_{vB} = C_{pLB} (T_{bB} - T_0) + \Delta H_{vBb} - C_{pvB} (T_{bB} - T_0)$

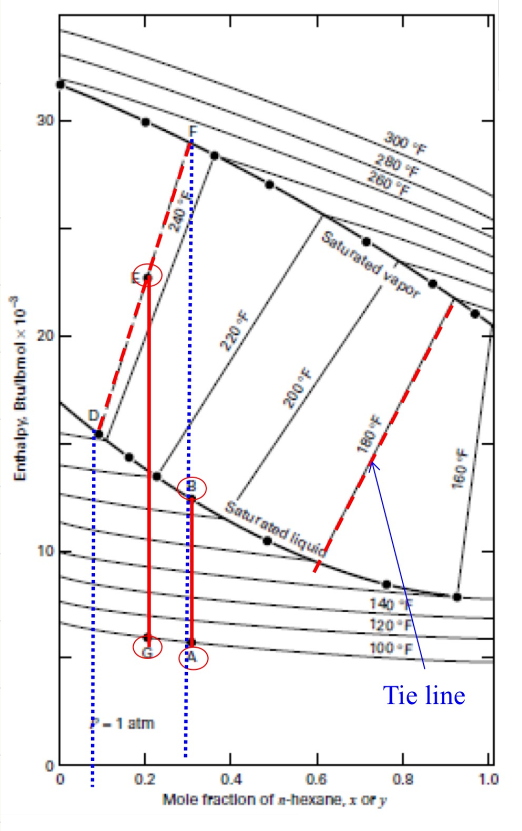
* The liquid is first changed from temperature T_0 to T_b & then evaporated at T_b, the vapor temp is then changed from T_b to T_0

Enthalpy Composition Diagram

if an amount Q of heat is added to a mass m_A of a phase, the increase in enthalpy from H_A to H_C will be given by :

$$H_A + \frac{Q}{m_A} = H_C$$

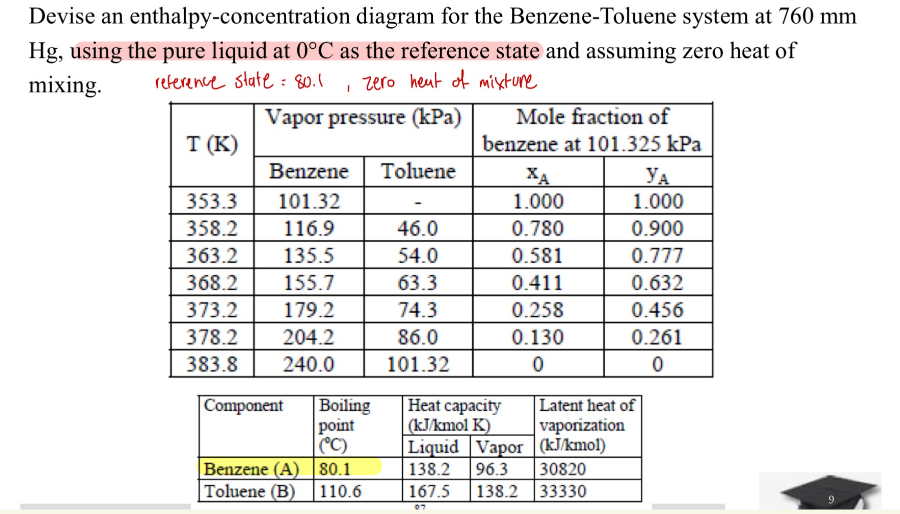




Point A → mixture 30% mol hexane , 70% mol octane [subcooled liq]

Point B → Heating A to bubble point Temp

Point E → Heating point G [at equilibrium vapor point F , liq point D]



Devise an enthalpy-concentration diagram for the Benzene-Toluene system at 760 mm Hg, using the pure liquid at 0°C as the reference state and assuming zero heat of mixing.

reference state : 80.1 , zero heat of mixture

T (K)	Vapor pressure (kPa)		Mole fraction of benzene at 101.325 kPa	
	Benzene	Toluene	x_A	y_A
353.3	101.32	-	1.000	1.000
358.2	116.9	46.0	0.780	0.900
363.2	135.5	54.0	0.581	0.777
368.2	155.7	63.3	0.411	0.632
373.2	179.2	74.3	0.258	0.456
378.2	204.2	86.0	0.130	0.261
383.8	240.0	101.32	0	0

Component	Boiling point (°C)	Heat capacity (kJ/kmol K)		Latent heat of vaporization (kJ/kmol)
		Liquid	Vapor	
Benzene (A)	80.1	138.2	96.3	30820
Toluene (B)	110.6	167.5	138.2	33330

At reference Temp = 80.1

liq enthalpy of pure Benzene ($x_A = 1$) at T=0°C

- first point → pure toluene ($x_A = 0$)

→ liq toluene at T_b = 110.6

$$H_x = x_A C_{PLA} (T_b - T_0) + (1 - x_A) C_{PLB} (T_b - T_0)$$

$$H_x = 0 + (1 - 0)\ 167.5\ (110.6 - 80.1) = 5109 \text{ KJ/mol}$$

→ for sat vapor enthalpy of pure toluene

1. $\Delta H_{VB}(80.1) = C_{PLB}(T_{bB} - T_0) + \Delta H_{VBb} - C_{PVB}(T_{bB} - T_0)$

not at reference BP ✓

$\Delta H_{VB} = 167.5(110.6 - 80.1) + 33330 - 138.2(110.6 - 80.1) = 34224$ KJ/Kmol

$y_A = 0$ ← 2. $H_y = y_A[\Delta H_{VA} + C_{PVA}(T_d - T_0)] + (1-y_A)[\Delta H_{VB} + C_{PVB}(T_d - T_0)]$

$H_y = 0 + (1-0)[34224 + 138.2(110.6 - 80.1)] = 38439$ KJ/Kmol

- Second point → Pure Benzene ($x_A = y_A = 1$)

$H_x = x_A C_{PLA}(T_b - T_0) + (1-x_A) C_{PLB}(T_b - T_0)$

$1(138.2)(80.1 - 80.1) + (1-1)\,167.5(80.1 - 80.1) = \underline{0}$

at point $x_A = 0.5$ $T_b = 92$ / $y_A = 0.5$ $T_d = 98.8$ [T-xy Diagram]

$H_x = x_A C_{PLA}(T_b - T_0) + (1-x_A) C_{PLB}(T_b - T_0)$

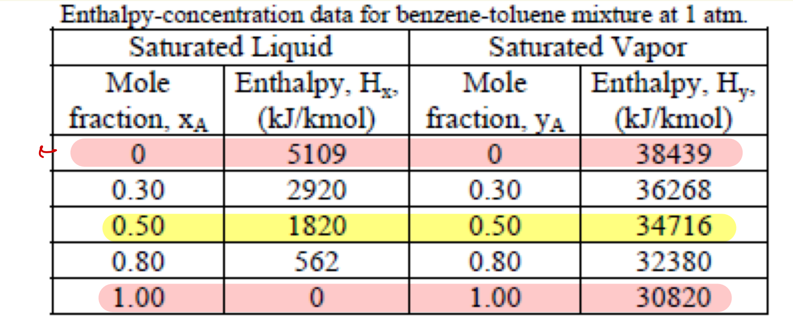
$0.5(138.2)(92 - 80.1) + (1-0.5)(167.5)(92 - 80.1) = 1820$ KJ/Kmol

↑ ΔH_{VA} already at reference BP

$H_y = y_A[\Delta H_{VA} + C_{PVA}(T_d - T_0)] + (1-y_A)[\Delta H_{VB} + C_{PVB}(T_d - T_0)]$

$H_y = 0.5[30820 + 96.3(98.8 - 80.1)] + (1-0.5)[34224 + 138.2(98.8 - 80.1)]$

$H_y = 34716$ KJ/Kmol



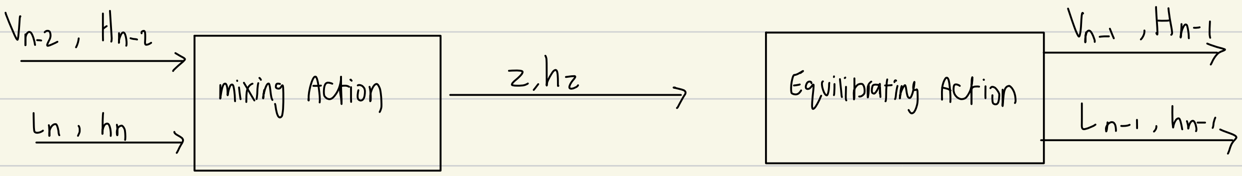
Enthalpy-concentration data for benzene-toluene mixture at 1 atm.

Saturated Liquid		Saturated Vapor	
Mole fraction, x_A	Enthalpy, H_x, (kJ/kmol)	Mole fraction, y_A	Enthalpy, H_y, (kJ/kmol)
0	5109	0	38439
0.30	2920	0.30	36268
0.50	1820	0.50	34716
0.80	562	0.80	32380
1.00	0	1.00	30820

at x_A = zero pure toluene ←

↳ At $y_A = 1$ = Latent heat of vaporization

Application to Equilibrium Stage Calculation



→ Energy Balance Equations for stage (n-1)

Mixing: $V_{n-2} H_{n-2} + L_n h_n = (V_{n-2} + L_n) h_z$ ①

Equilibration: $(V_{n-2} + L_n) h_z = V_{n-1} H_{n-1} + L_{n-1} h_{n-1}$ ②

→ The material balance eq. for the light components:

Mixing $y_{n-2} V_{n-2} + x_n L_n = z(V_{n-2} + L_n)$ ③

Equilibration $z(V_{n-2} + L_n) = y_{n-1} V_{n-1} + x_{n-1} L_{n-1}$ ④

① & ③ $$\frac{H_{n-2} - h_z}{y_{n-2} - z} = \frac{h_z - h_n}{z - x_n}$$

② & ④ $$\frac{H_{n-1} - h_z}{y_{n-1} - z} = \frac{h_z - h_{n-1}}{z - x_{n-1}}$$

y_{n-1} & x_{n-1} are in equilibrium

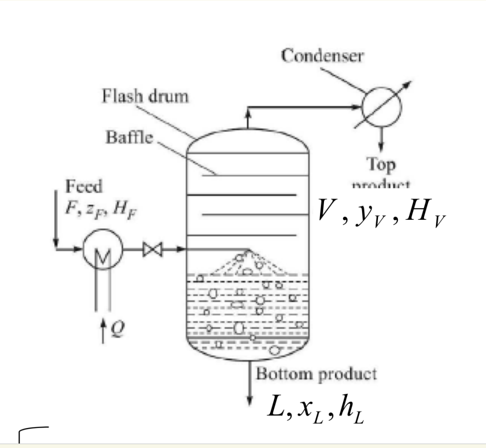
→ Amounts of liq & vapor phases → Lever Rule

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

$$\therefore \frac{H_{n-1} - h_z}{h_z - h_{n-1}} = \frac{y_{n-1} - z}{z - x_{n-1}}$$

Rearranging $$\frac{H_{n-1} - h_z}{y_{n-1} - z} = \frac{h_z - h_{n-1}}{z - x_{n-1}}$$

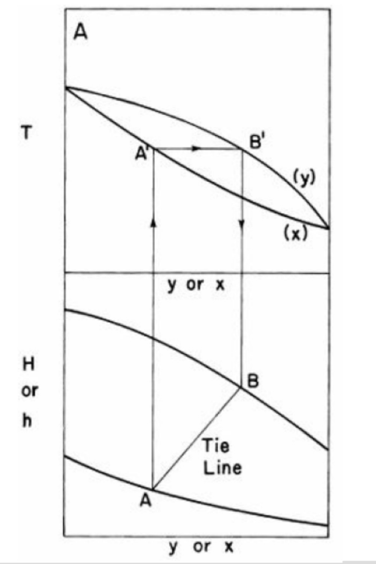
Flash vaporization of a Binary Mixture



Total Material Balance: $F = V + L$

Component Balance: $F z_F = y_V V + x_L L$

$$(V+L) z_F = y_V V + x_L L$$



Enthalpy Balance: $F H_F + Q = V H_V + L h_L$

$$-\frac{L}{V} = \frac{H_V - (Q/F + H_F)}{h_L - (Q/F + H_F)}$$

for $Q = 0$

$$-\frac{L}{V} = \frac{H_V - H_F}{h_L - H_F}$$

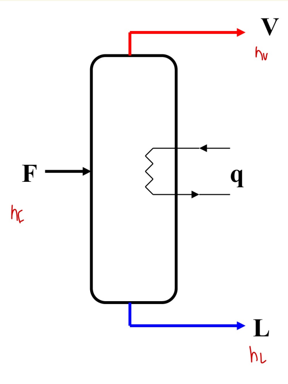
$$-\frac{L}{V} = \frac{y_V - z_F}{x_L - z_F}$$

$$y = \frac{L}{V} x_i + \frac{F}{V} z_i$$

$$y = \frac{f-1}{f} x_i + \frac{1}{f} z_i$$

$$f = \frac{V}{F}$$

Balance on Continous Distillation Column



Overall Material Balance : $F = L + V$

Component Balance : $Fz_F = x_L L + y_V V$

Enthalpy Balance : $F H_F = L h_L + V H_V$

For adiabatic process $q = 0$

$$(V+L) H_F = L h_L + V H_V \qquad (V+L) z_F = L x + y V$$

$$V H_F + L H_F = L h_L + V H_V \qquad V z_F + L z_F = L x - V y$$

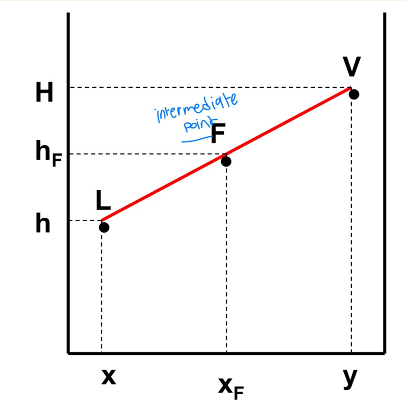
$$V (H_V - H_F) = L (H_F - h_L) \qquad V (y - z_F) = L (z_F - x)$$

$$\frac{L}{V} = \frac{H_V - H_F}{H_F - h_L} \qquad \frac{L}{V} = \frac{y - z_F}{z_F - x}$$

↓ Energy Balance ↓ Mass Balance

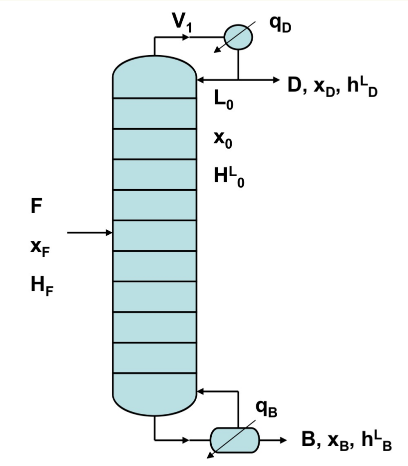
$$\frac{H_V - H_F}{y - z_F} = \frac{H_F - h_L}{z_F - x}$$

↳ equal slopes ⇒ parallel or extension



* since both lines go through the same point F, the lines lie on the same straight line.

Enthalpy conc. lines [Adiabatic $q = 0$]



Overall Material Balance : $F = D + B$

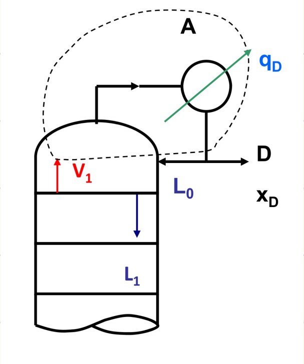
Component Balance : $F x_F = x_D D + x_B B$

$$F x_F = D x_D + x_B (F - D)$$

$$D = \frac{F (x_F - x_B)}{x_D - x_B}$$

Enthalpy Balance : $F h_F + q_B = D h_D + B h_B + q_D$

Total Condensor



Material Balance around condensor : $V_1 = L_0 + D$

Component material Balance : $yV = L_0 x_0 + D x_D$

Enthalpy Balance : $V_1 H_1 = L_0 h_L + q_D + D h_D$

$$Q_D = \frac{q_D}{D}$$

$$V_1 H_1 = L_0 h_L + D (h_D + Q_D)$$

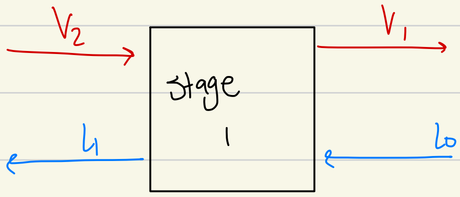
↳ heat on condensor [liq]

Reflux Ratio $R = \frac{L_0}{D} = \frac{h_D + Q_D - H_1}{H_1 - h_D}$

Internal Reflux $\frac{L_0}{V_1} = \frac{\overset{h_\Delta}{\boxed{h_D + Q_D}} - H_1}{\underset{h_\Delta}{\boxed{h_D + Q_D}} - h_0}$

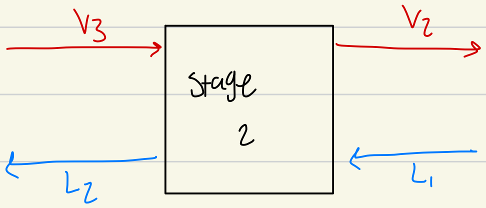
→ Internal Reflux between each plate, until a point in the column is reached where a stream is added or removed, shown as:

$$\frac{L_n}{V_{n+1}} = \frac{h_D + Q_D - H_{n+1}}{h_D + Q_D - h_n}$$



$$L_0 + V_2 = V_1 + L_1$$

$$L_0 - V_1 = L_1 - V_2 = \Delta$$



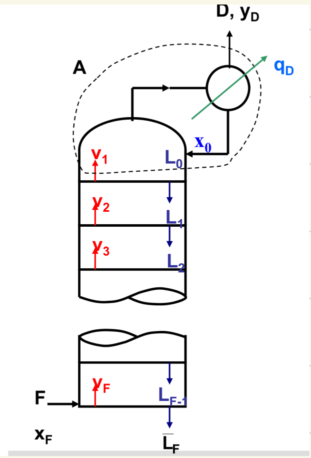
$$L_1 + V_3 = L_2 + V_2$$

$$L_1 - V_2 = L_2 - V_3 = \Delta$$

$$L_n - V_{n+1} = -D = \Delta$$

Δ → shared between all stages ⇒ straight line from Δ to L, passing V

Partial Condensor



Overall material Balance : $V_1 = L_0 + D$

Component Balance : $y_1 V_1 = x_0 L_0 + y_D D$

Enthalpy Balance : $H_1 V_1 = h_0 L_0 + q_D + D H_D$

$$Q_D = \frac{q_D}{D}$$

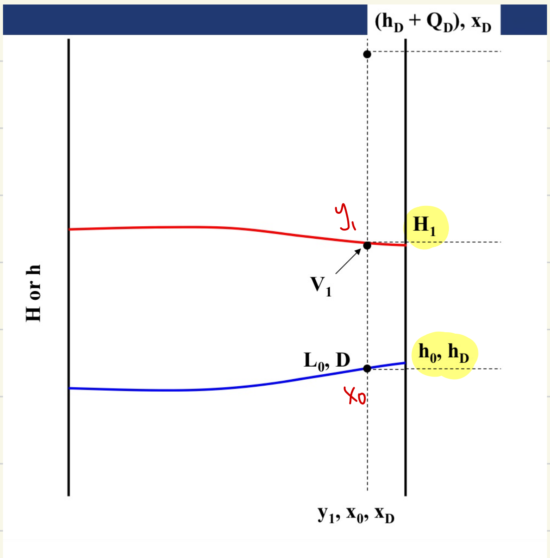
$$H_1 V_1 = h_0 L_0 + D(H_D + Q_D)$$

↳ heat on condensor [vap]

Internal Reflux Ratio $\frac{L_n}{V_{n+1}} = \frac{\overset{\Delta}{\boxed{H_D + Q_D}} - H_{n+1}}{\underset{\Delta}{\boxed{H_D + Q_D}} - h_n}$

Difference between total & partial condensor

Total condensor



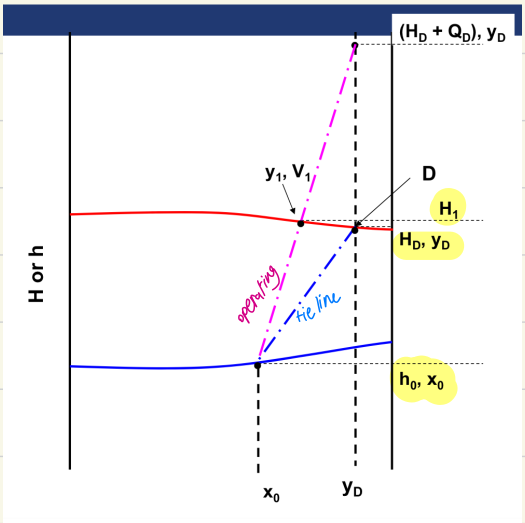
$$y_1 = x_0 = x_D$$

$$h_0 = h_D$$

$$\Delta = h_D + Q_D$$

$$R = \frac{L_0}{D} = \frac{h_D + Q_D - H_1}{H_1 - h_0}$$

Partial condensor



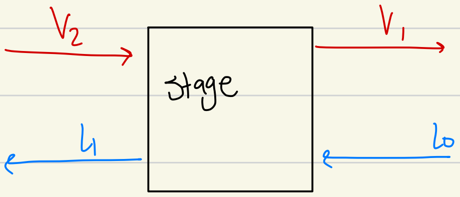
$$y_D \neq x_0$$

at y_D find x_D → xy diagram & draw the tie line

connect x_0 to Δ to find y_1 (H_1)

$$R = \frac{L}{D} = \frac{H_D + Q_D - H_1}{H_1 - h_0}$$

Rectifying section



$$L_0 x_0 + y_2 V_2 = x_1 L_1 + y_1 V_1$$

$$L_0 x_0 - y_1 V_1 = x_1 L_1 - y_2 V_2 = -D x_D = \Delta x_\Delta$$

$$x_D = x_\Delta$$

Enthalpy Balance: $L_0 h_0 - V_1 H_1 = L_1 h_1 - V_2 H_2 = -D(h_D + Q_D) = \Delta h_\Delta$

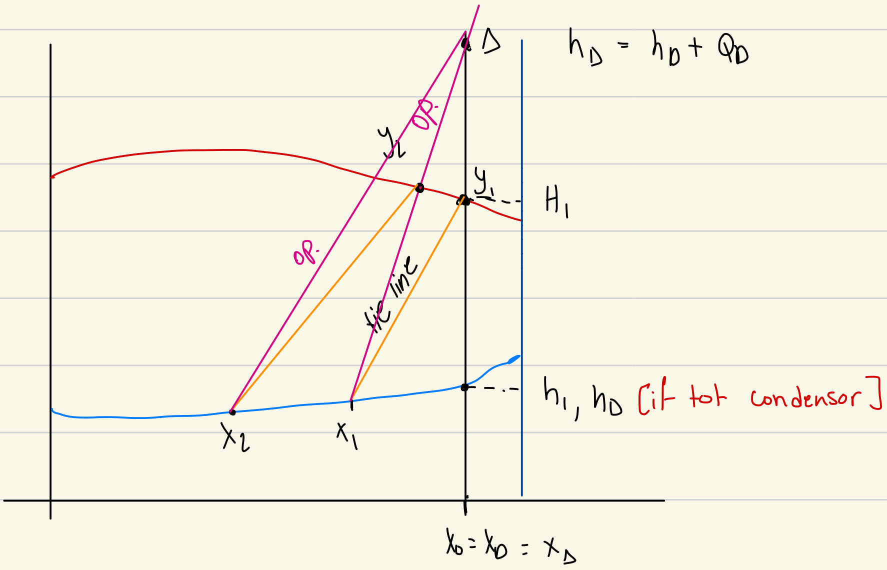
$$h_\Delta = h_D + Q_D$$

* Δ → The difference point, & all lines corrosponding to the combined material & enthalpy Balance equations (operating line equations) for the rectifying section pass through this intersection

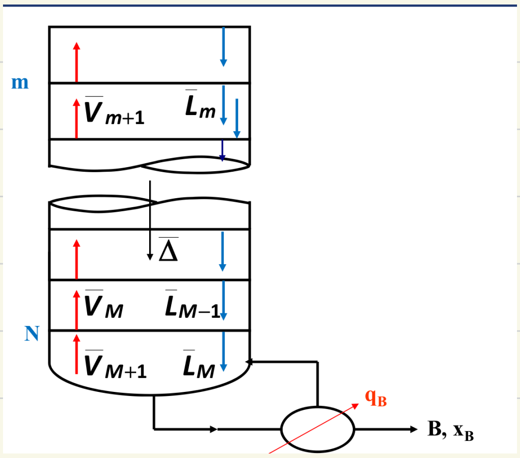
Plate to plate graphical procedure for Determining the number of equilibrium stages:

1. Establish the location of point Δ with $x_D = x_\Delta$ & $h_\Delta = h_D + Q_D$ (Total condensor) & R
   or $h_\Delta = H_D + Q_D$ (Partial condensor)

2. Use equilibrium data to establish L_1 at (x_1, h_1) [sat liq line]
3. Draw the operating line between L_1 & Δ → Intersecting V_2 (y_2, H_2)



Stripping Section



overall:

$$\bar{L}_m - \bar{V}_{m+1} = B = \bar{\Delta}$$

Component:

$$x_m \bar{L}_m - \bar{V}_{m+1} y_{m+1} = \bar{\Delta} x_{\bar{\Delta}}$$

$$x_{\bar{\Delta}} = x_B$$

Enthalpy

$$h_m \bar{L}_m - H_{m+1} \bar{V}_{m+1} = \bar{\Delta} h_{\bar{\Delta}}$$

$$h_{\bar{\Delta}} = h_B - Q_B$$

$$\bar{\Delta} = \Delta + F$$

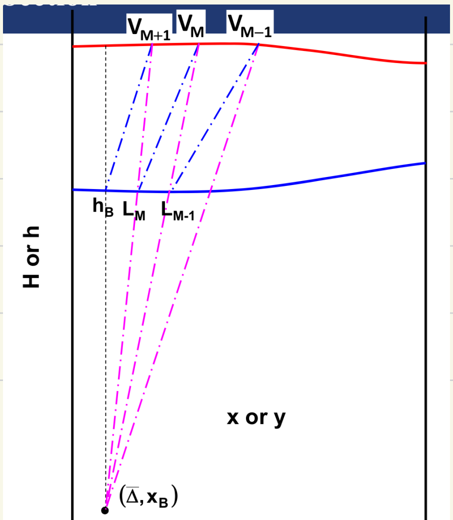
↳ $\bar{\Delta}$ lies on the extension of the straight line passing through F & Δ

Plate to Plate Graphical procedure for determining the number of equilibrium stages

Draw a straight line passing through Δ & F

Draw a vertical line starting from x_B till it intersects the Δ & F line ⇒ $\bar{\Delta}$

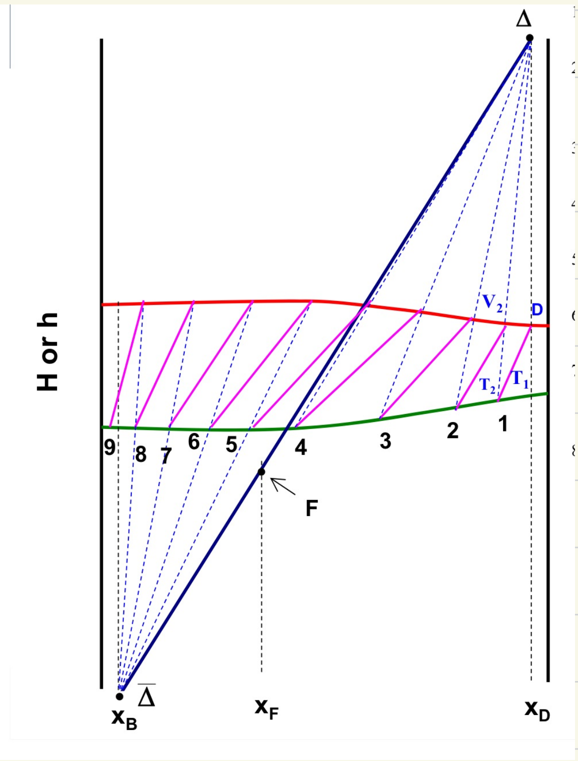
Assuming the reboiler to be an equilibrium stage, the vapor V_{m+1} is in eq with the bottom stream.

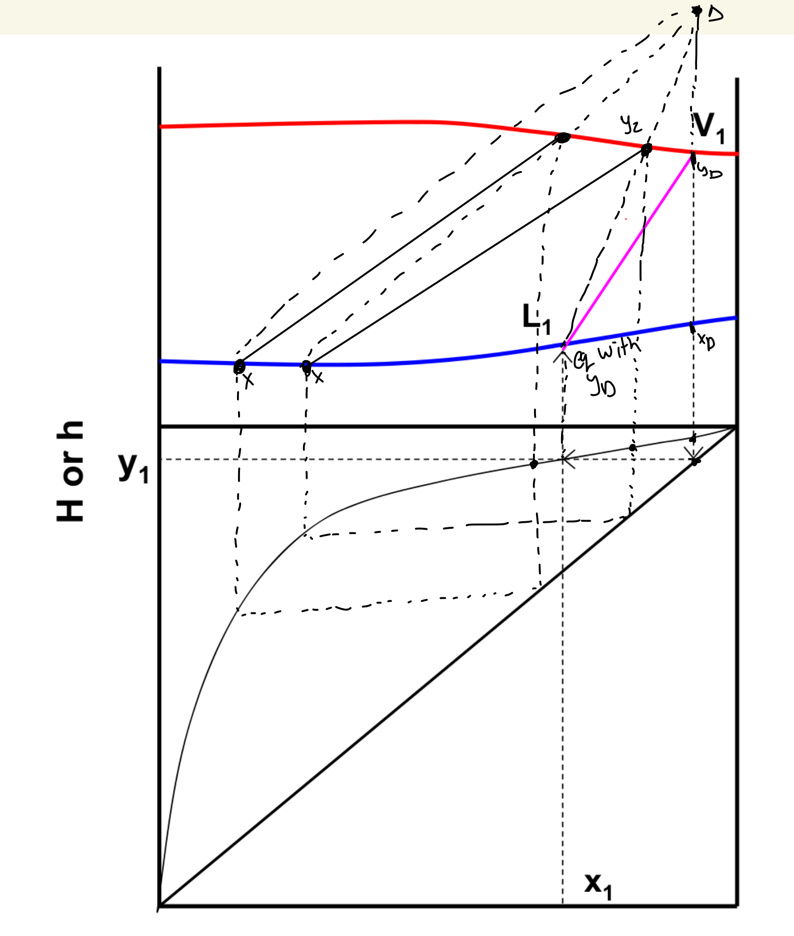


Total Column

→ The construction may start from either side [rectifying or stripping]

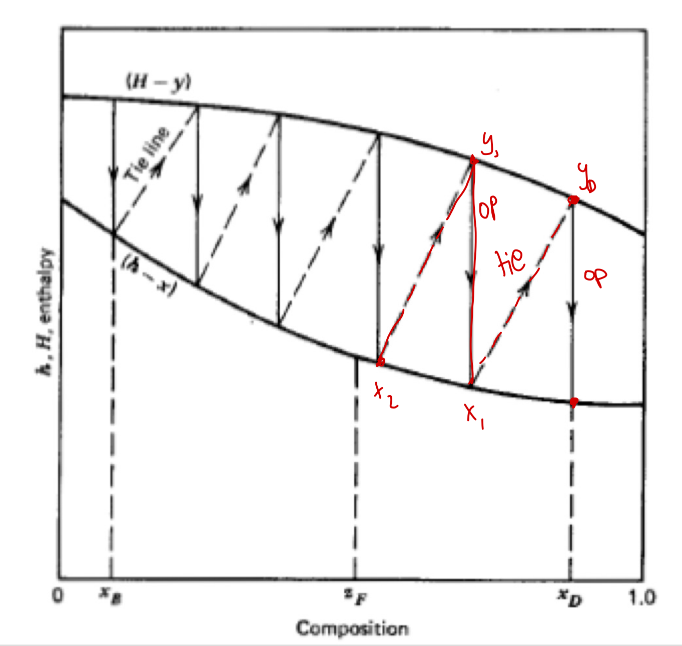
→ In either case, when an equilibrium tie line crosses the line connecting the different points through the feed condition, the other difference point is used to complete the construction





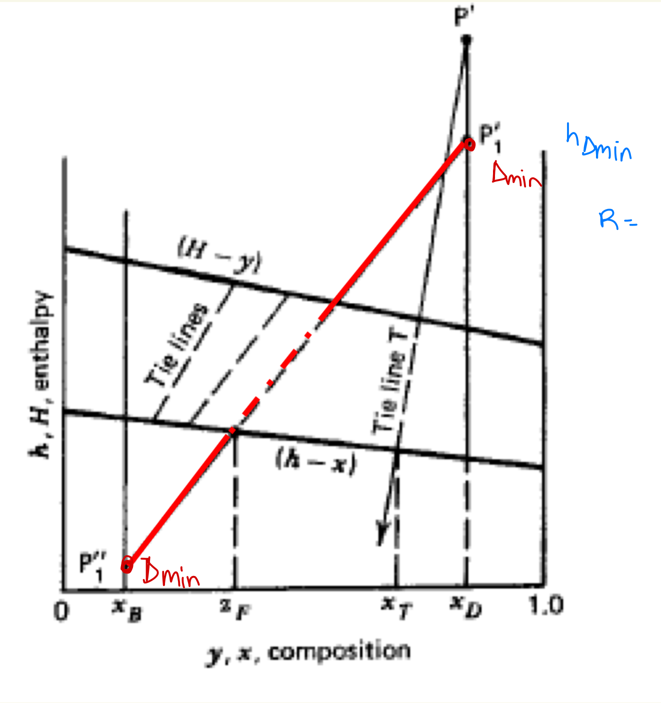
Total Reflux

* minimum number of stages can be obtained by drawing vertical operating lines.

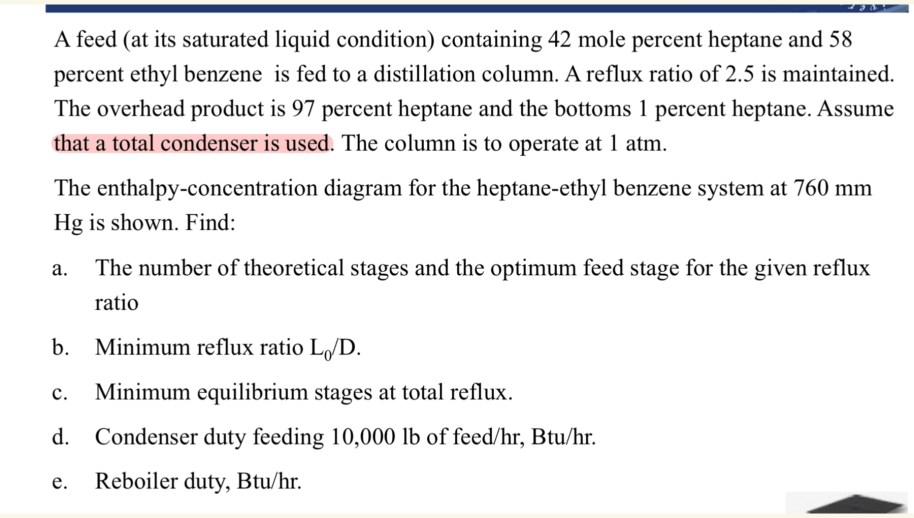


Minimum Reflux

Can be obtained by extending the tie line through the feed composition to intersect a vertical line drawn through x_D, & x_B



$$R = \frac{h_{\Delta min} - H_1}{H_1 - h_0}$$



A feed (at its saturated liquid condition) containing 42 mole percent heptane and 58 percent ethyl benzene is fed to a distillation column. A reflux ratio of 2.5 is maintained. The overhead product is 97 percent heptane and the bottoms 1 percent heptane. Assume that a total condenser is used. The column is to operate at 1 atm.

The enthalpy-concentration diagram for the heptane-ethyl benzene system at 760 mm Hg is shown. Find:

a. The number of theoretical stages and the optimum feed stage for the given reflux ratio

b. Minimum reflux ratio L_0/D.

c. Minimum equilibrium stages at total reflux.

d. Condenser duty feeding 10,000 lb of feed/hr, Btu/hr.

e. Reboiler duty, Btu/hr.

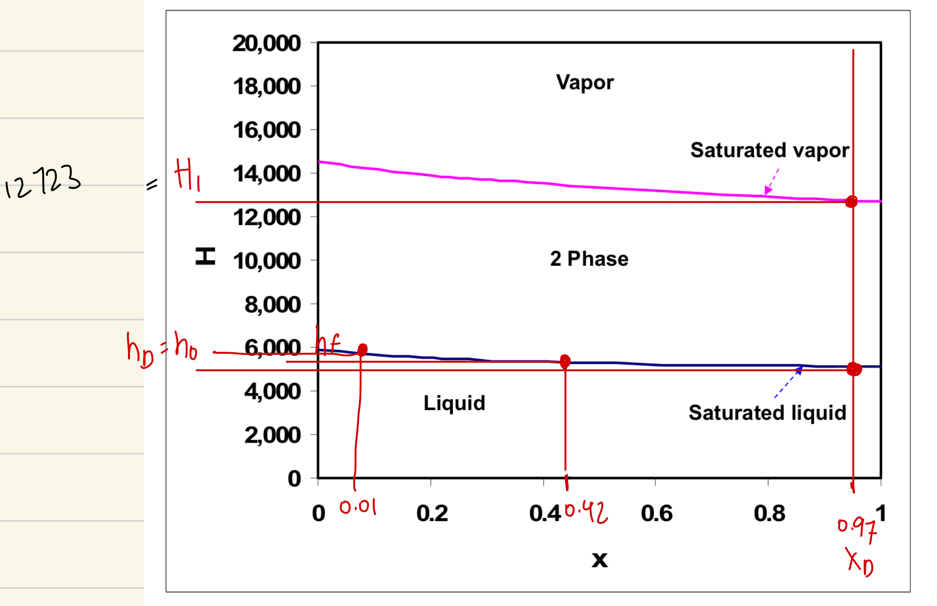
$x_F = 0.42 \rightarrow h_F = 5329.9$

$R = 2.5$

$x_D = 0.97 \rightarrow 5117$

$x_B = 0.01 \rightarrow 5886$

$F = 9.9937$ kmol/hr



$F = D + B$ $\qquad$ $9.993 = D + B$

$z_F F = x_D D + x_B B$ $\qquad$ $0.42(9.993) = 0.97(D) + 0.01 B$

$D = 4.261$ kmol/hr

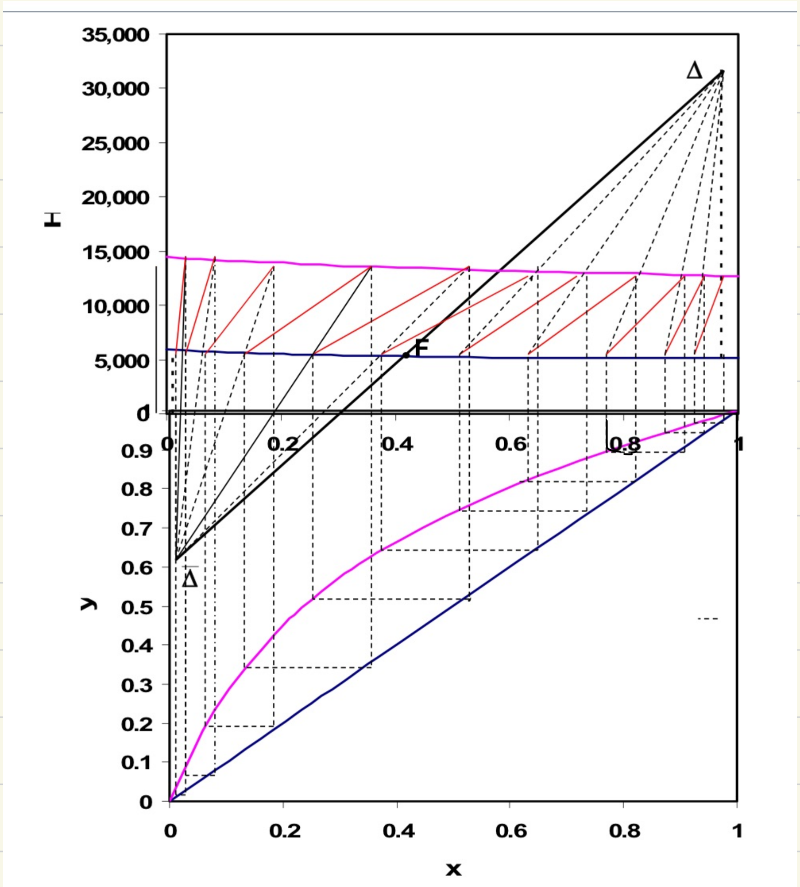
$B = 5.7256$ kmol/hr

$$R = \frac{L_0}{D} = \frac{h_D + Q_D - H_1}{H_1 - h_0} \rightarrow 2.5 = \frac{5117 + Q_D - 12723}{12723 - 5117}$$

$Q_D = 26621$ cal/mol

$h_\Delta = h_D + Q_D$

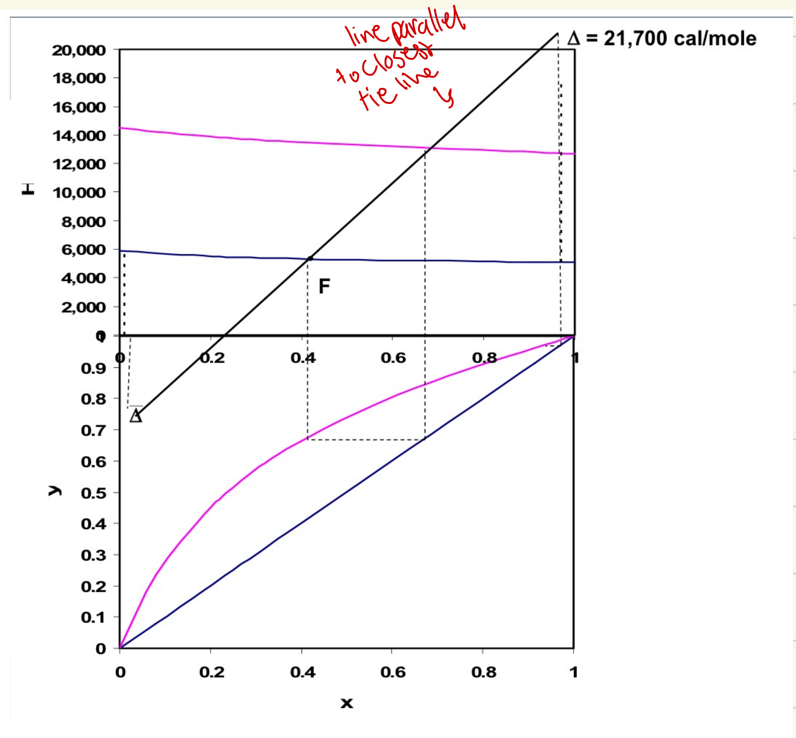
$h_\Delta = 5117 + 26621 = 31738$



11 stages

Stage 7 → feed stage

b. Minimum reflux ratio



$h_{\Delta min} = 21700$ cal/mol

$$R = \frac{L}{D} = \frac{h_{\Delta min} - H_1}{H_1 - h_0}$$

$$R = \frac{21700 - 12723}{12723 - 5117}$$

$R = 1.18$

c. total reflux (vertical operating lines) → 7 stages

D. Condensor duty

$Q_D = \frac{q_D}{D}$ $\qquad q_D = Q_D \times D$

$26621 \times 4261 = 113\,621$ cal/hr

E. Reboiler duty

Enthalpy balance

$$Fh_F + q_B = h_D D + q_D + B\,h_B$$

$$9.993(5329.9) + q_B = 4.2681(5117) + 113621 + 5886(5.7256)$$

$$q_B = 116770 \text{ cal/hr}$$

lecture 6 : Absorption & stripping

Absorption

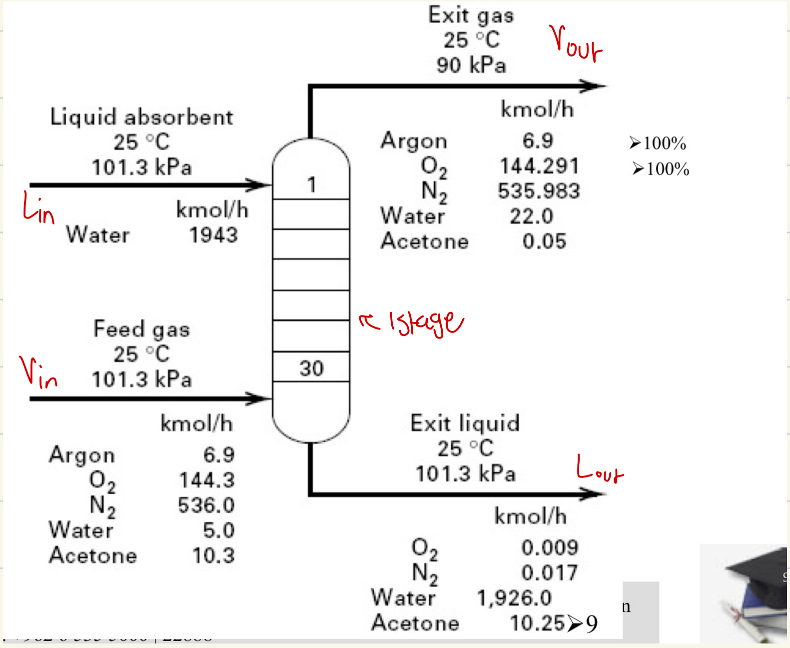
- Gas absorption is the removal of one or more pollutants from a contaminated gas stream → by allowing the gas to come into intimate contact with the liquid that enables the pollutants to become dissolved by the liq.
  → Solubility of pollutants in the absorbing liquid

- Absorption : a gas mixture in contact with liq (absorbant or solvent) to selectively dissolve one or more components of the gas by mass transfer from gas → liq
- Solute (absorbate) : the component transferred to the liquid
- Stripping : opposite to absorption (liquid → gas)
- Stripping & Absorption ⇒ mass separating agent
- Distillation ⇒ Energy separating agent

Uses of Absorption:

1. Purification of gasses
2. Separation of gasses
3. Production of useful liquid product

Typical Absorption Process:



Choice of Solvent for Gas Absorption

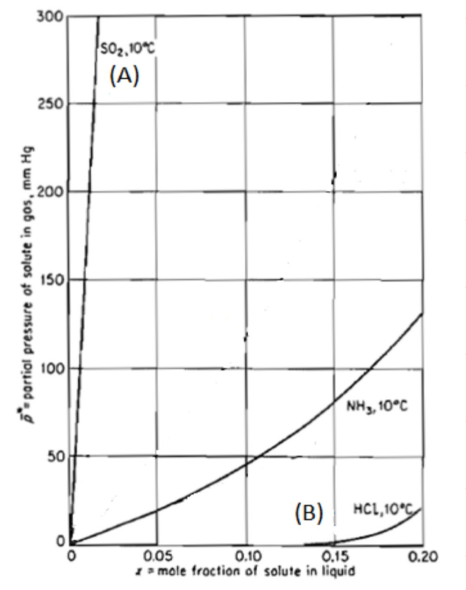
1. High absorption power   gas solubility should be high in the solvent (increase the rate of absorption & decrease the rate of solvent required)
2. Highly Selective
3. Easy to recover
4. low volatility : the solvent should have low vapor pressure to reduce the loss of solvent in the gas leaving
5. Small viscosity · rapid absorption rates, improving flooding characteristics, low pressure drops & good heat transfer characteristics.
6. Cost
7. Corrosiveness
8. Toxicity & Hazard

Absorption & Stripping Equilibria

Assumptions:

1. Carrier gas is insoluble (into a solvent)
2. Solvent is non-volatile (loss of solvent due to vaporization → negligable)
3. System is isothermal & isobaric

↳ The equilibrium data is represented by Henry's law constant.



→ if the equilibrium pressure of a gas at a given liquid concentration is high, as in case A, the gas is relatively insoluble in the solvent

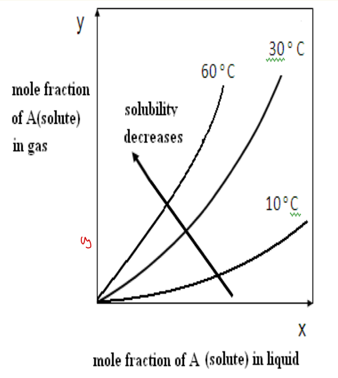
→ while if it's low, as for B, the solubility is high.

* The solubility of any gas is influenced by the temperature.

↑ Temp ↓ Solubility

Absorption processes is usually accompanied by evolution of heat.

↳ fit coolers to the absorber to keep its temperature sufficiently low.



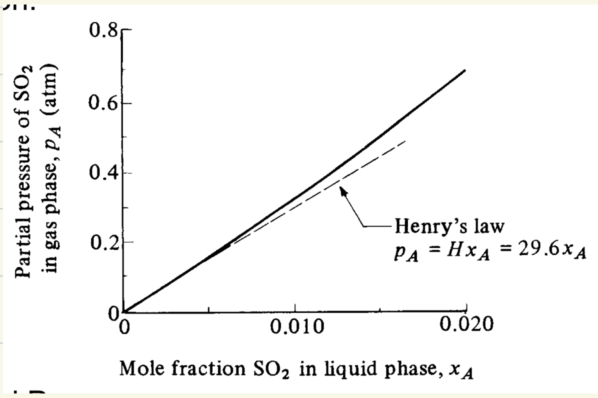
Henry's Law

* For dilute concentrations of many gasses the equilibrium relationship ⇒ Henry's Law

→ It relates the partial pressure developed by a dissolved solute (A) in a liquid solvent (S)

$$P_A = H x_A \quad \text{or} \quad y_A = m x_A$$

$$m = \frac{H_A}{P_{tot}}$$



Single Component Absorption

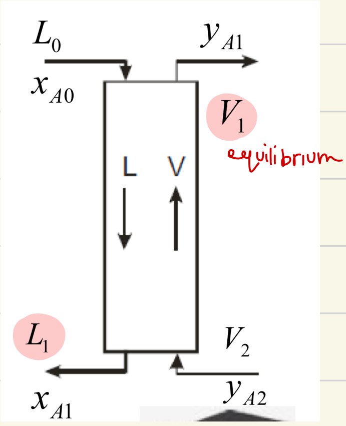
* Carried out in counter current flow processes, gas flow is introduced in the bottom of the column & the liquid solvent is introduced at the top of the column

The overall material Balance:

$$L_0 + V_2 = L_1 + V_1$$

Component Balance:

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



- In the usual gas- liquid system, the solute A is in the gas phase with inert B, & liquid L with inert water C
  - → Binary gas phase A-B
  - → Binary liquid phase A-C

* Carrier gas & solvent rates remain constant throughout the absorber

$L' = L(1 - x_A)$
↓
flow rate of nonvolatile solvent
(inert water C)

$V' = V(1 - y_A)$
↓
carrier gas flow rate
(inert Air B)

Balance on A:

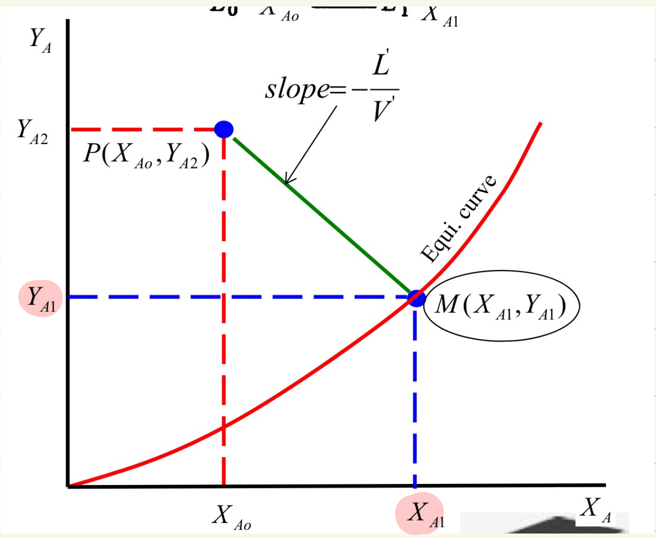
$$L'\left(\frac{x_{A0}}{1-x_{A0}}\right) + V'\left(\frac{y_{A2}}{1-y_{A2}}\right) = L'\left(\frac{x_{A1}}{1-x_{A1}}\right) + V'\left(\frac{y_{A1}}{1-y_{A1}}\right)$$

or:

$$L' X_{A0} + V' Y_{A2} = L' X_{A1} + V' Y_{A1}$$

$$Y_1 = -\frac{L'}{V'} X_1 + \left(\frac{L'}{V'} X_0 + Y_2\right)$$

$$\text{slope} = -\frac{L'}{V'} : \frac{Y_{A2} - Y_{A1}}{X_{A0} - X_{A1}}$$

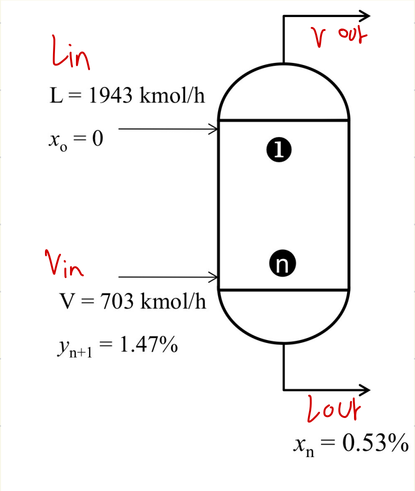


exit compositions
from graph

* For Dilute systems $x = X$, $y = Y$

* The composition of the streams leaving the process (y_{A1} & x_{A1}) are related by the equilibrium distribution relation

$$y_{A1} = H' x_{A1}$$



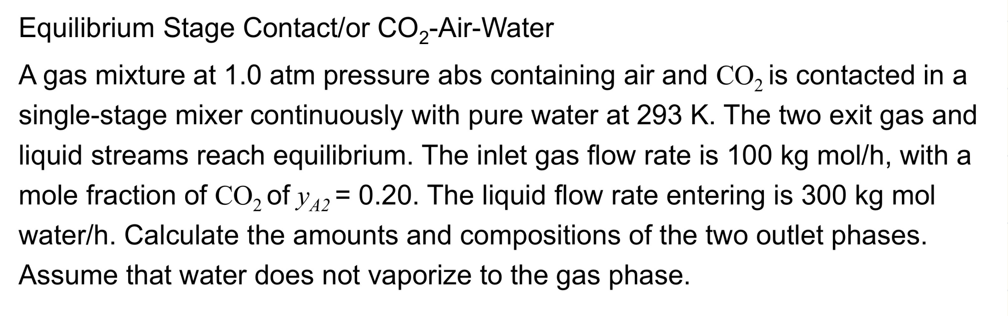
$L' = L(1 - x_{A0}) = 1943(1-0) = 1943$ Kmol/h

$V' = V(1-y_A) = 703(1-0.0147) = 692\ 66$ Kmol/h

$$Y = \frac{y_A}{1-y_A} = \frac{0.0147}{1-0.0147} = 0.0149$$

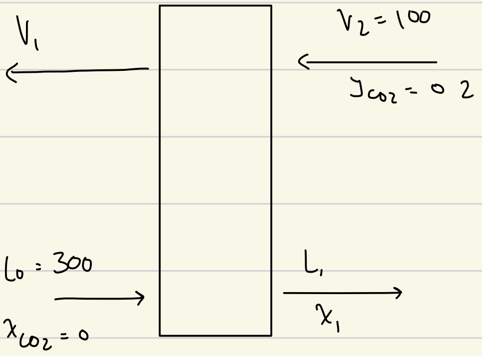
$$X_0 = \frac{x_{A0}}{1-x_{A0}} = \frac{0}{1-0} = 0$$

$$X_n = \frac{x_n}{1-x_n} = \frac{0.0053}{1-0.0053} = 0.0053$$



Equilibrium Stage Contact/or CO_2-Air-Water

A gas mixture at 1.0 atm pressure abs containing air and CO_2 is contacted in a single-stage mixer continuously with pure water at 293 K. The two exit gas and liquid streams reach equilibrium. The inlet gas flow rate is 100 kg mol/h, with a mole fraction of CO_2 of $y_{A2} = 0.20$. The liquid flow rate entering is 300 kg mol water/h. Calculate the amounts and compositions of the two outlet phases. Assume that water does not vaporize to the gas phase.



Overall Balance

$$L_0 + V_2 = V_1 + L_1$$

Component Balance

$$x_0 L_0 + y_2 V_2 = y_1 V_1 + x_1 L_1$$

$$X_0 L_0' + Y_2 V_2' = Y_1 V_1' + X_1 L_1'$$

Operating line $Y = -\frac{L'}{V'} X + \left(\frac{L'}{V'} X_0 + Y_2 \right)$

$L' = L(1-x) = 300(1-0) = 300$

$V' = V(1-y) = 100(1-0.2) = 80$

$Y = \frac{y}{1-y} = \frac{0.2}{1-0.2} = 0.25$

$X_0 = 0$

$$Y = -\frac{300}{80} X + \left(\frac{300}{80}(0) + 0.25 \right)$$

$Y = -3.75X + 0.25$

→ X vs Y plot:

$H = 0.142 \times 10^4$ atm / mole fraction

$m = H/P = 1420 \frac{\cancel{atm}}{\text{mole fraction}} \bigg| \frac{}{1 \cancel{atm}} = 1420$ mole fraction^{-1}

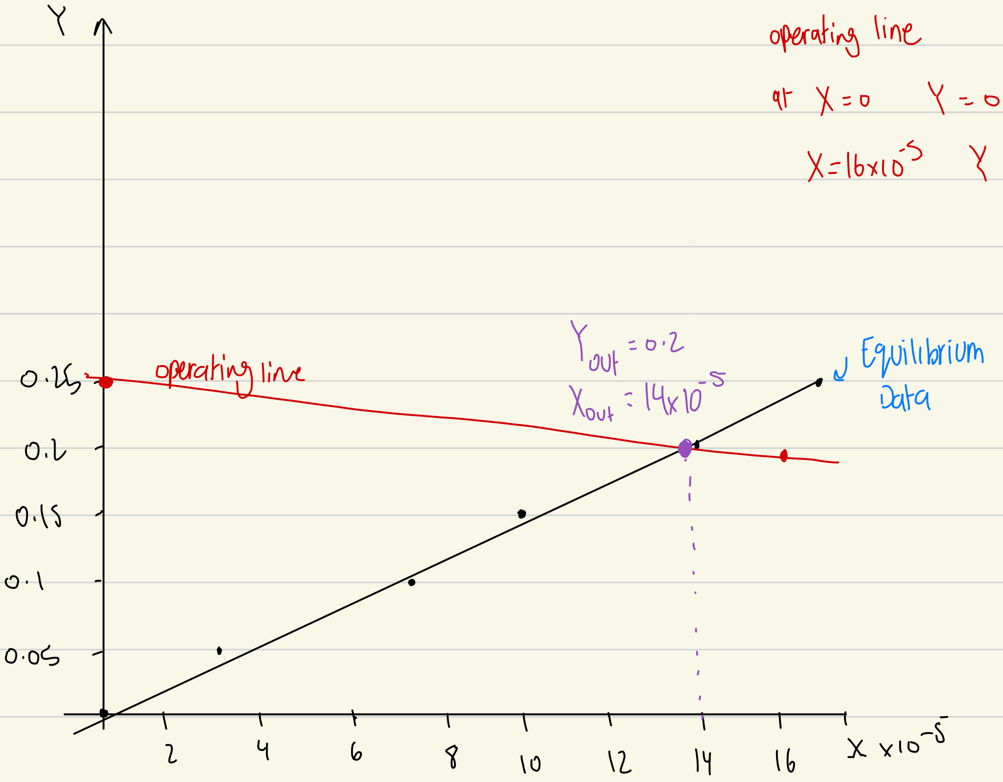
$y = mx$

y	x	Y	X
0	0	0	0
0.05	3.52×10^{-5}	0.0526	3.52×10^{-5}
0.1	7.04×10^{-5}	0.111	7.04×10^{-5}
0.15	10.5×10^{-5}	0.176	10.5×10^{-5}
0.2 by $_{max}$	14.08×10^{-5}	0.25	14.08×10^{-5} (Dilute)

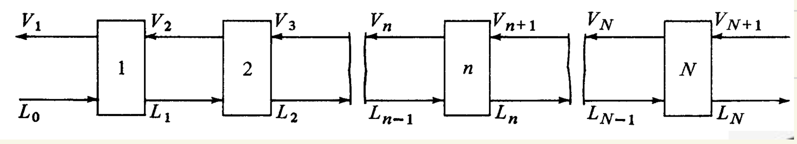
operating line

at $X = 0$ $Y = 0.25$

$X = 16 \times 10^{-5}$ $Y = 0.2494$



Counter Current Multiple Contact Stages



- To conserve the use of the L_0 stream & to get a more concentrated product, countercurrent multiple stage is generally used

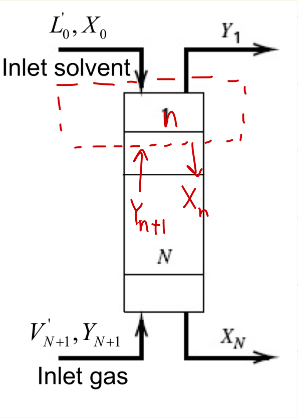
Component A is being exchanged between L & V streams

$V'_{N+1} = V'_N = V'_{N-1} = V'$ flow rate of insoluble carrier gas

$L'_N = L'_{N-1} = L'$ flow rate of the non-volatile solvent

* x,y → mole fraction

X, Y → mole Ratio

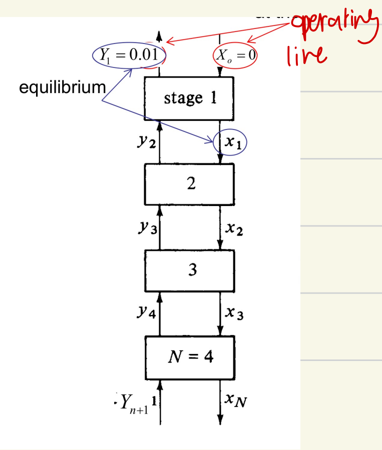


$$Y = \frac{L'}{V'} X_n + \left(Y_1 - \frac{L'}{V'} X_0 \right)$$

X_0 & Y_1 ⇒ known & Regulated (Points on operating line)

V' & L' ⇒ Constants

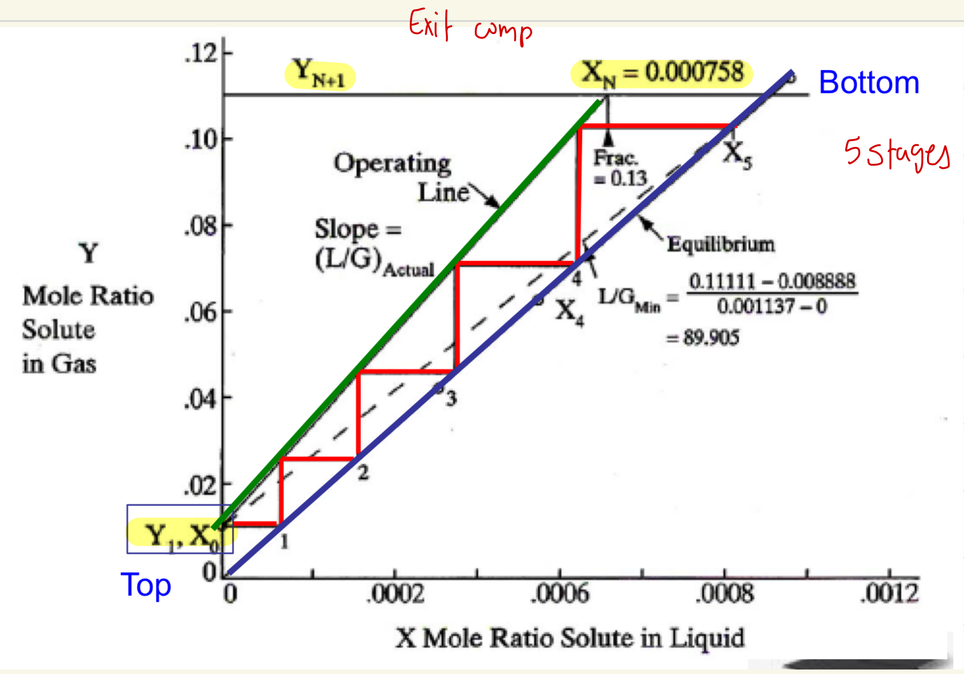
Graphical Design for Absorption Staged Tower

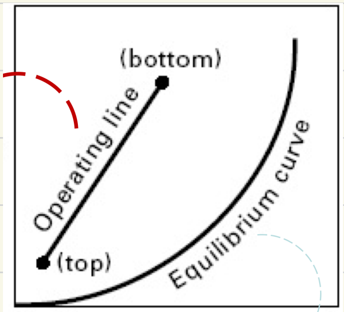


* In Absorption the operating line is above the equilibrium line.

$$Y = \frac{L'}{V'} X + \left(Y_1 - \frac{L'}{V'} X_0 \right)$$

* solvent inlet $x_0 = 0$ for pure water





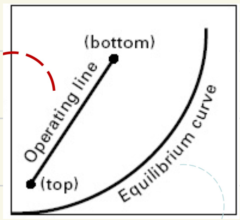
Equilibrium curve: $y = H_A x$

$x \quad y \quad | \quad X \quad Y$

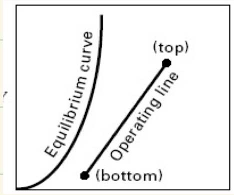
Operating line: $Y = \frac{L'}{V'} X + \left(Y_1 - \frac{L'}{V'} X_0 \right)$

Absorption vs. Stripping

Absorption



Stripping



Absorption & Separation factors

Absorption: $A = \frac{L}{K \times V}$ $K_{value} = \frac{y}{x}$

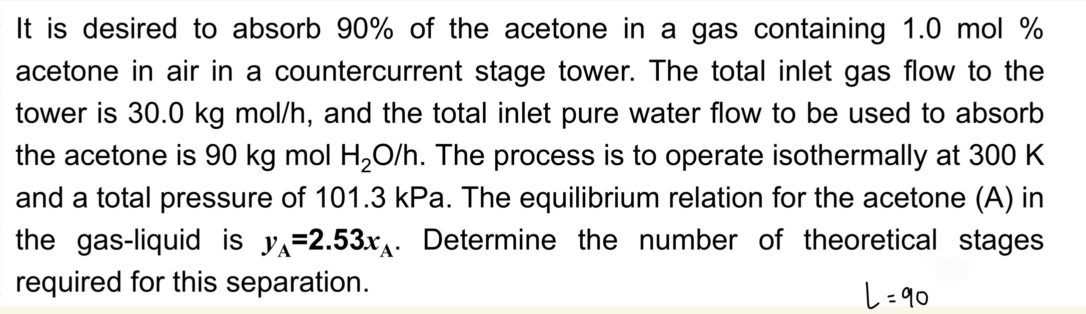
* Absorption takes place if $A > 1$

$1.25 < A < 2$ → A optimum $= 1.4$

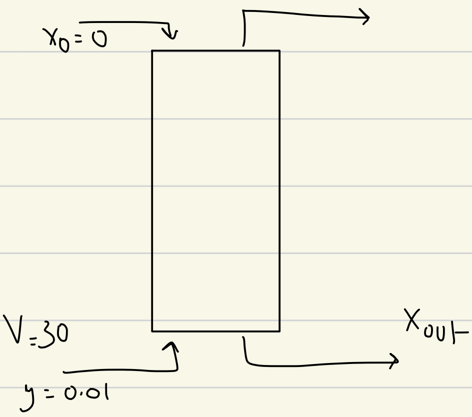
* As A↑, number of stages ↓ (↑L)
* To increase A: ↑L ↓K ↓V
* To ↓K (function (exp T)/P): ↓T ↑P

Stripping $S = \frac{1}{A} = \frac{KV}{L}$

* To ↑S: ↑K ↑V ↓L
* ↑K by ↑T ↓P



It is desired to absorb 90% of the acetone in a gas containing 1.0 mol % acetone in air in a countercurrent stage tower. The total inlet gas flow to the tower is 30.0 kg mol/h, and the total inlet pure water flow to be used to absorb the acetone is 90 kg mol H_2O/h. The process is to operate isothermally at 300 K and a total pressure of 101.3 kPa. The equilibrium relation for the acetone (A) in the gas-liquid is $y_A = 2.53 x_A$. Determine the number of theoretical stages required for this separation.



$L' = L(1-x) = 90$

$V' = V(1-y) = 30(1-0.01) = 29.7$

$X_0 = 0$

$Y_0 = \frac{0.01}{1-0.01} = 0.0101$

$X_{out} = \frac{0.9(30)(0.01)}{90} = 0.003$

$Y_{out} = \frac{0.1(30)(0.01)}{29.7} = 0.001$

→ Equilibrium Data $y = 2.53x$

$x_{max} = \frac{0.01}{2.53} = 3.95\times10^{-3}$ $y_{max} = 0.01$ [Diluted system]

$X \simeq x$	$Y \simeq y$
0	0
3.95×10^{-3}	0.01

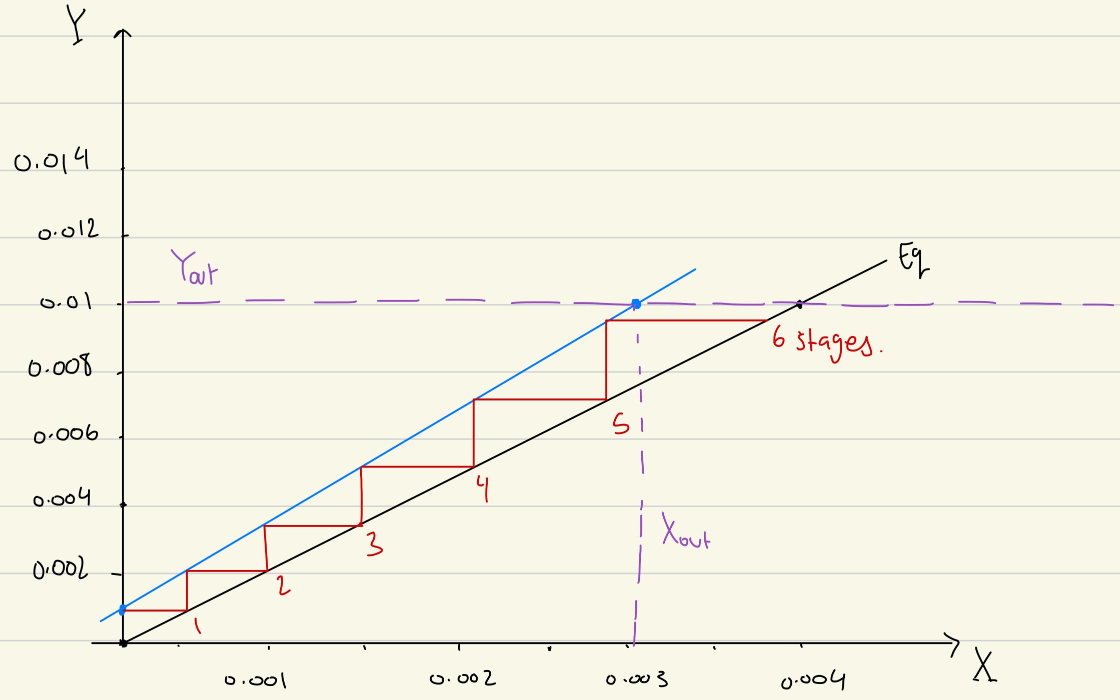
→ Operating line $Y = \frac{L'}{V'}X + \left(Y_1 - \frac{L'}{V'}X_0\right)$

$$Y = \frac{90}{29.7}X + \left(0.001 - \frac{90}{29.7}(0)\right)$$

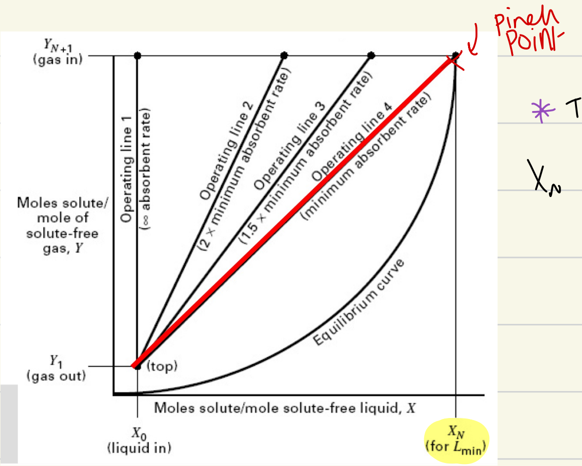
$$Y = 3.03X + 0.001$$

Points of operating line: X_0 & Y_{out} (0, 0.001)

X_{out} & Y_0 (0.003, 0.0101)

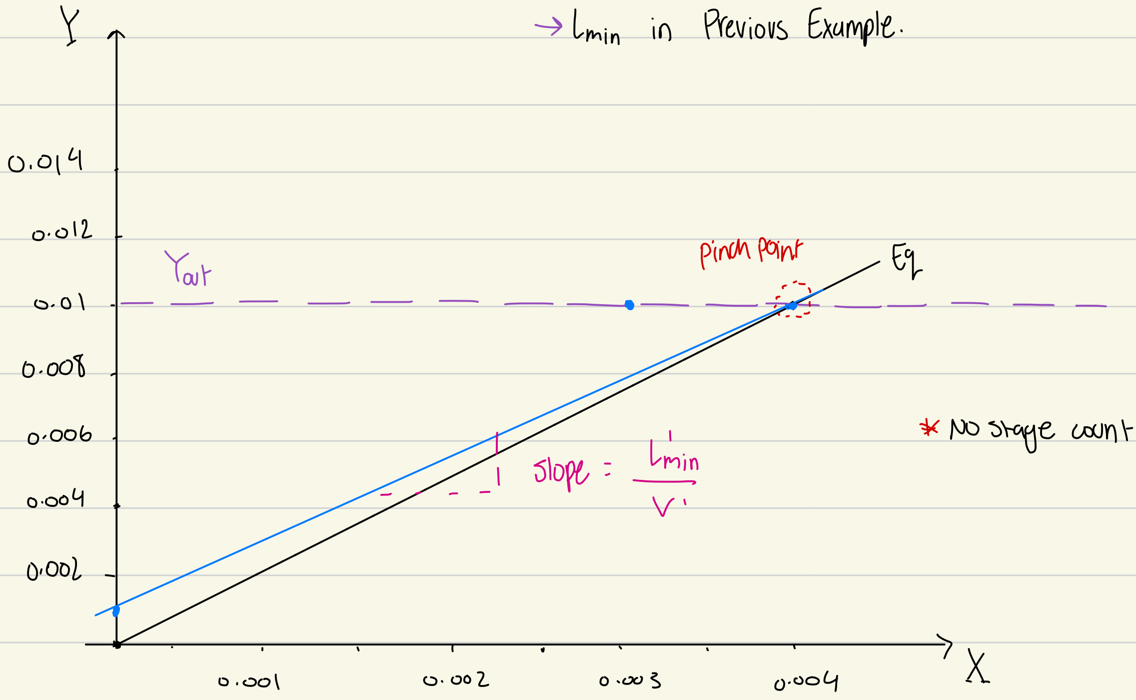


Minimum Solvent flow Rate (Absorption)



* The minimum absorbent flow rate L_{min}, corresponds to a value of X_N, in eq with Y_{N+1} (the solute conc. in the feed gas).

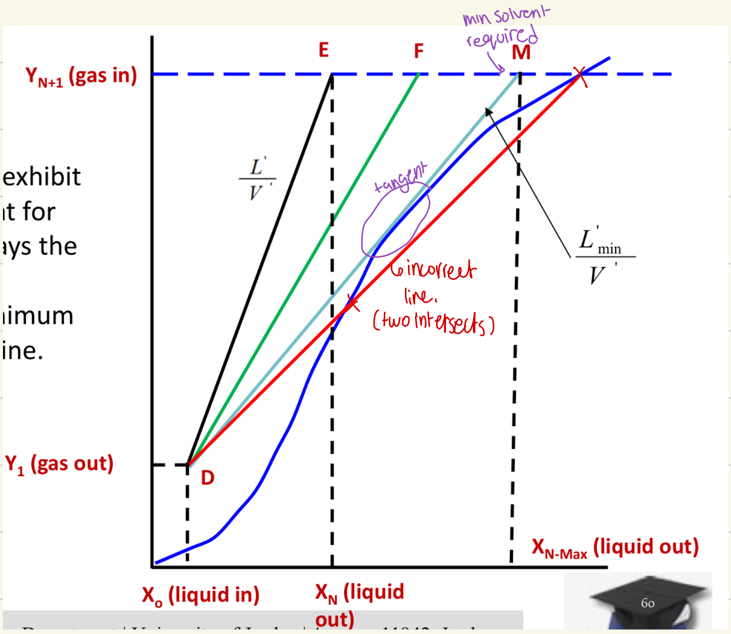
→ L_{min} in Previous Example.



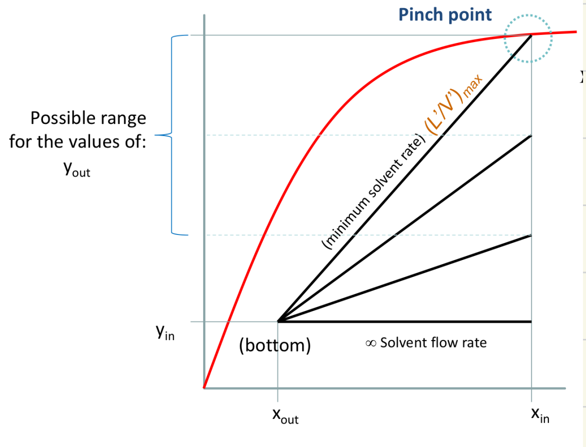
* No stage counting.

The equilibrium data may exhibit other forms.

So a pinch point for minimum may not always be the case, for a concave down eq. curve, the minimum is taken as the tangential line

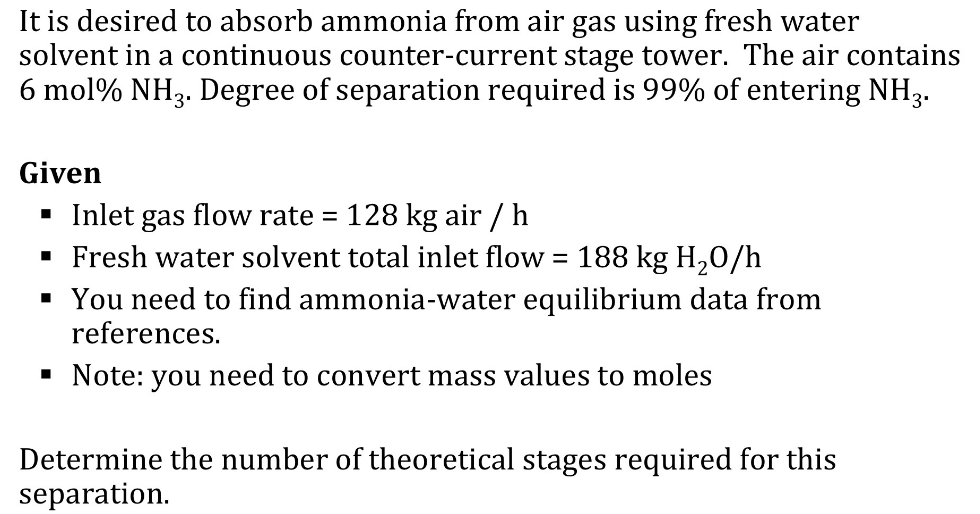


Minimum Solvent Flow Rate (Stripping)



$$Y = \frac{L'}{V'} X + \left(Y_1 - \frac{L'}{V'} X_0 \right)$$

$$\frac{L'}{V'_{min}} = \text{pinch point} \quad \left(\frac{L'}{V'} \right)_{max}$$

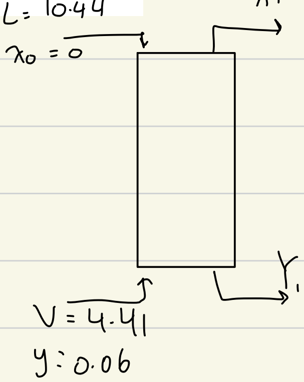


It is desired to absorb ammonia from air gas using fresh water solvent in a continuous counter-current stage tower. The air contains 6 mol% NH_3. Degree of separation required is 99% of entering NH_3.

Given

- Inlet gas flow rate = 128 kg air / h
- Fresh water solvent total inlet flow = 188 kg H_2O/h
- You need to find ammonia-water equilibrium data from references.
- Note: you need to convert mass values to moles

Determine the number of theoretical stages required for this separation.



$V = \frac{128}{29} = 4.41$ $\qquad$ $L = \frac{188}{18} = 10.44$

$V' = V(1-y) = 4.41(1-0.06) = 4.143$

$L' = L(1-x_0) = 10.44$

$Y_0 = \frac{0.06}{1-0.06} = 0.064$

$X_0 = 0$

$X_1 = \frac{0.99(0.064)(4.41)}{10.44} = 0.0268$

$Y_1 = \frac{(1-0.99)(0.064)(4.41)}{4.143} = 0.00068$

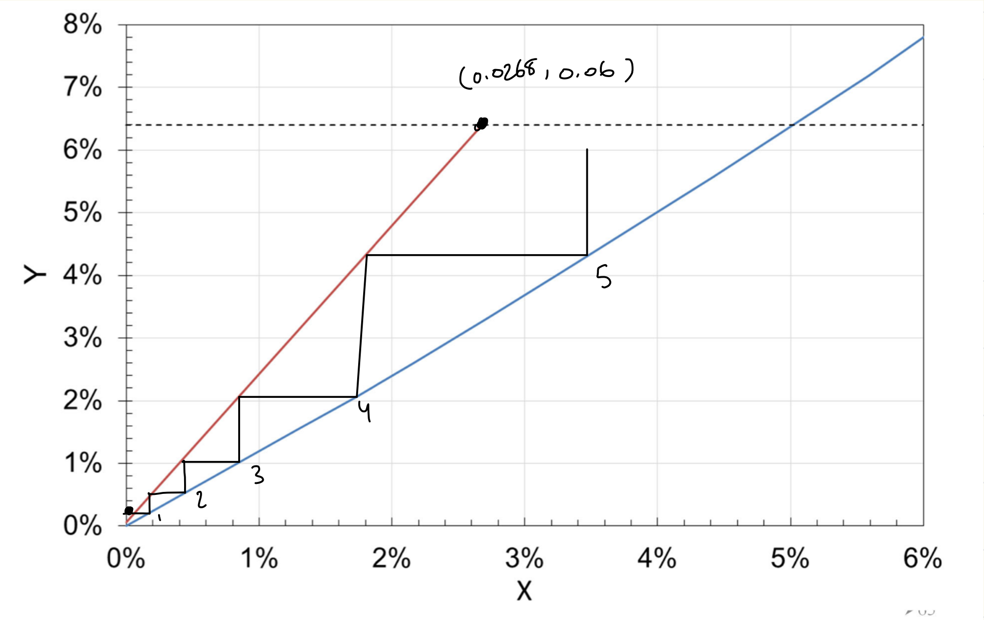
Operating line $$Y = \frac{L'}{V'} X + \left(Y_1 - \frac{L'}{V'} X_0\right)$$

$$Y = \frac{10.44}{4.143} X + 0.00068$$

$$Y = 2\ 52X + 0.00068$$

→ Points for op line X_0, Y_1 $(0, 0.00068)$

X_1, Y_0 $(0.0268, 0.06)$



6 stages

Analytical Equations for countercurrent stage contact.

* For transfer of solute A from L to V (stripping)

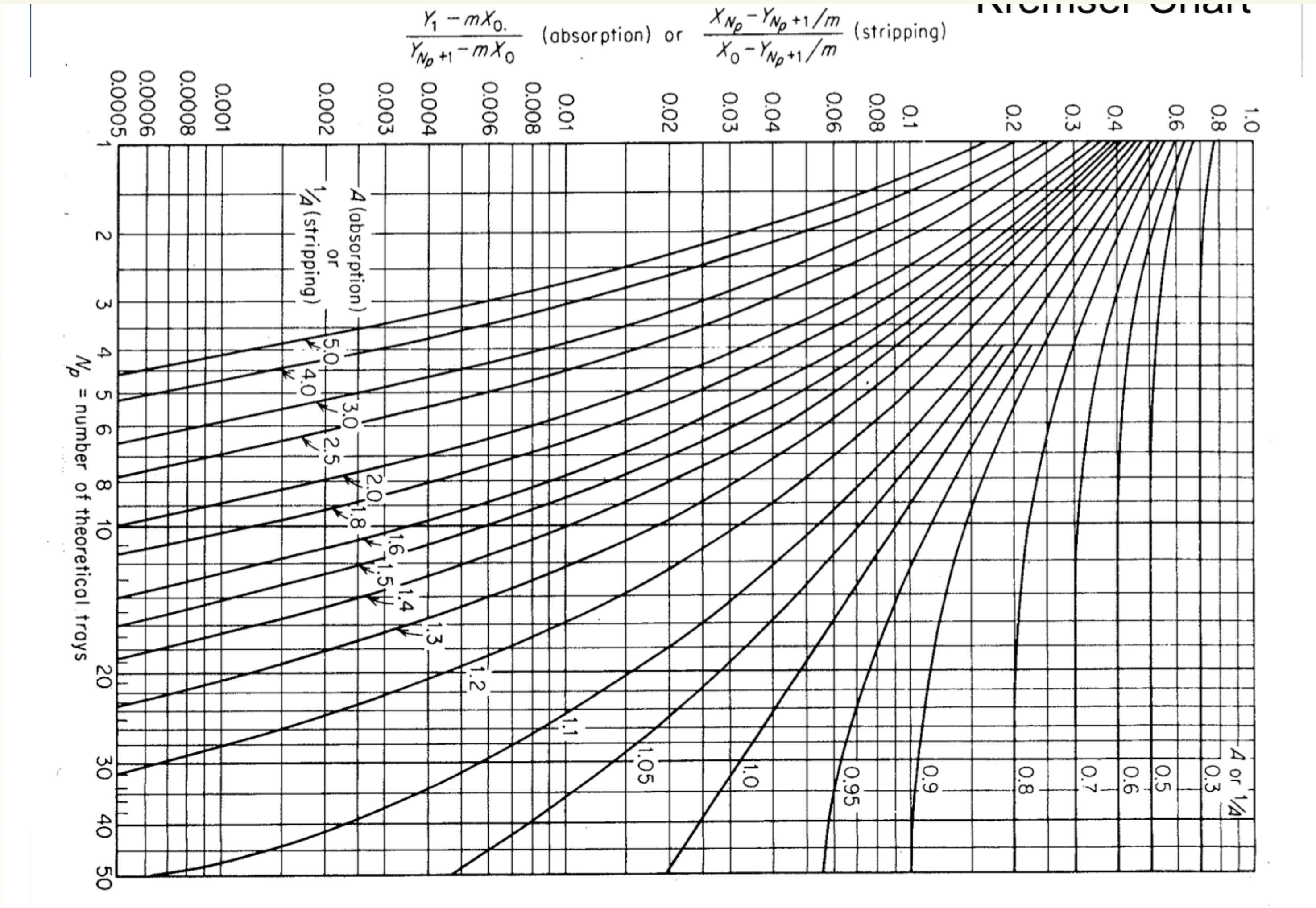
$$N = \frac{\log\left(\frac{x_0 - (y_{N+1}/m)}{x_N - (y_{N+1}/m)}(1 - A) + A\right)}{\log(1/A)}$$

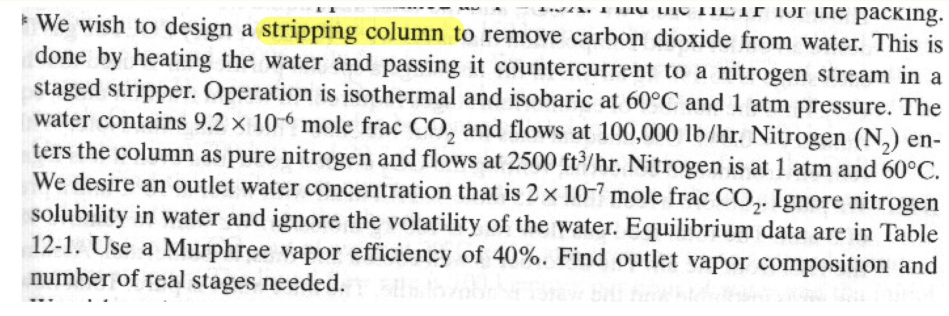
$$A = \frac{L}{mV}$$

* for transfer of solute A from V to L (Absorption)

$$N = \frac{\log\left(\frac{y_{N+1} - m x_0}{y_1 - m x_0}\left(1 - \frac{1}{A}\right) + \frac{1}{A}\right)}{\log A}$$

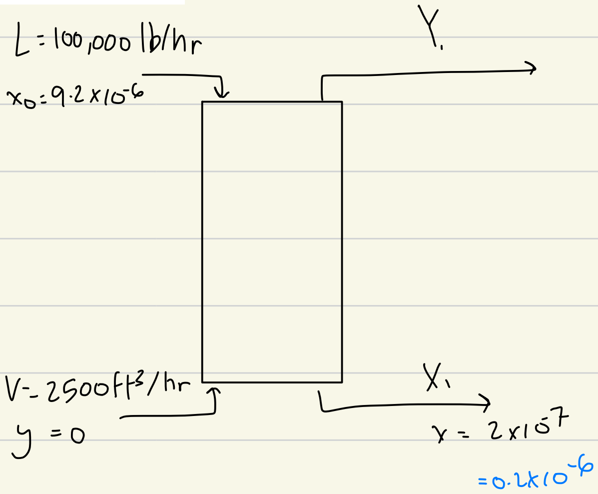
⇒ Kremser Chart





* We wish to design a stripping column to remove carbon dioxide from water. This is done by heating the water and passing it countercurrent to a nitrogen stream in a staged stripper. Operation is isothermal and isobaric at 60°C and 1 atm pressure. The water contains 9.2×10^{-6} mole frac CO_2 and flows at 100,000 lb/hr. Nitrogen (N_2) enters the column as pure nitrogen and flows at 2500 ft³/hr. Nitrogen is at 1 atm and 60°C. We desire an outlet water concentration that is 2×10^{-7} mole frac CO_2. Ignore nitrogen solubility in water and ignore the volatility of the water. Equilibrium data are in Table 12-1. Use a Murphree vapor efficiency of 40%. Find outlet vapor composition and number of real stages needed.

Stripping
liq → Gas



$$V = \frac{2500}{437} = 5.72 \text{ lbmol/h}$$

$$L = \frac{100\,000}{18} = 5560 \text{ lbmol/h}$$

$$Y = \frac{L'}{V'} X_{in} + \left(Y_0 - \frac{L'}{V'} X_1 \right)$$

$$Y_{out} = \frac{5560}{5.72} \left(9.2x10^{-6} \right) + \left(0 - \frac{5560}{5.72} \left(2x10^{-7} \right) \right)$$

$$Y_{out} = 8.75 \times 10^{-3} = 0.00875$$

⇒ Equilibrium Data

$y = 3410\, x$

$x \cong X$	$y \cong Y$
0	0
$2.65x10^{-6}$	0.00875

⇒ operating line points

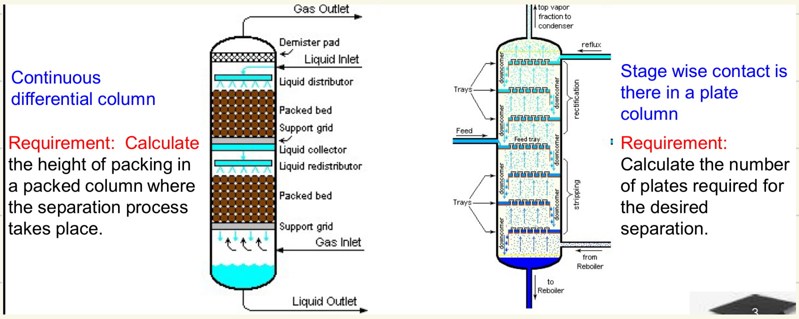
$X_0, Y_1 \quad (9.2x10^{-6}, 8.75x10^{-3})$

$X_1, Y_0 \quad (0.2x10^{-6}, 0)$

Lecture 5: Absorption in packed Tower

* In continous contacting equipment, the up-flowing gas remains in contact with down-flowing liquid throughout the packing, at every point of the tower.

* Therfore packed tower is known as "continous differential contact equipment".



Continuous differential column

Requirement: Calculate the height of packing in a packed column where the separation process takes place.

Stage wise contact is there in a plate column

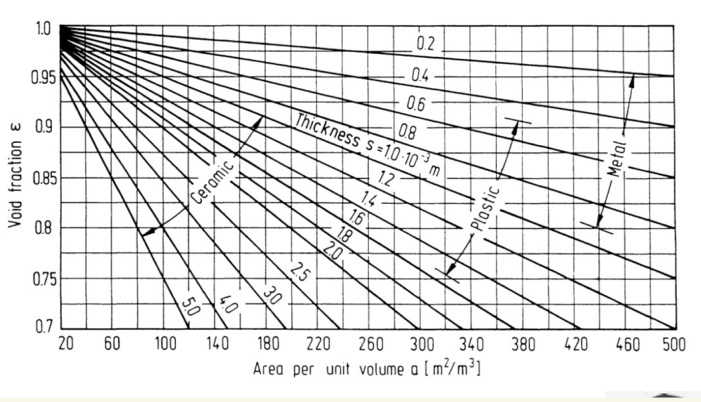
Requirement: Calculate the number of plates required for the desired separation.

* Packing Material are utilized to provide large interfacial area of contact between two phases
    - → low cost
    - → large interfacial area of contact [higher pressure drop] $\downarrow P = \frac{F}{\uparrow A}$
    - → High void volume to maintain low pressure drop
    - → Packing material should not trap suspended solids present in liq [Bigger packing materials ⇒ low fouling resistance]
    - → Mechanical strength
    - → Uniform flow of streams → uniform void spaces [non-uniform flow of streams leads to stagnant liquid pool → low mass transfer].

Types of Packing:

Random packing (Dumped Packing)

Structured Packing → Due to low gas Pressure drop & improved efficiency



Packing Height (Z)

1- Equilibrium Stage Analysis (HETP) Method

$$Z = HETP \times N$$

Hieght equivalent to a Theoretical Plate & ↳ number of theoretical stages obtained from McCabe Thiele

Represents the height of packing that gives similar separation to as a theoretical stage.

HETP values are provided for each type of packing (used for preliminary estimates)

2- Mass Transfer Analysis (HTU) Method

$$Z = HTU \times NTU$$

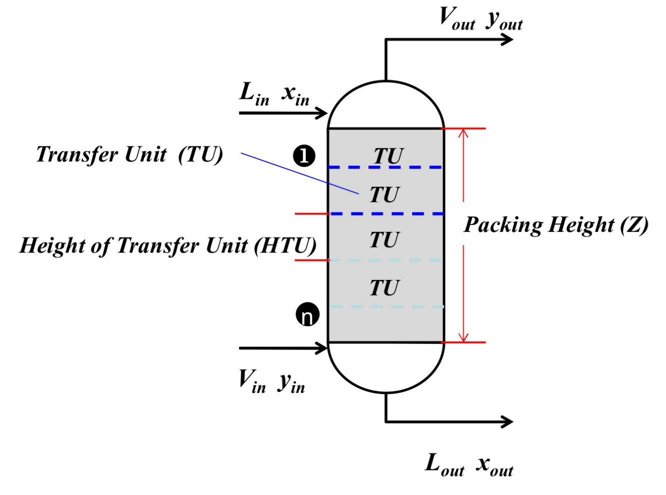
Height of Transfer unit & ↳ Number of Transfer units (obtained by numerical integration)

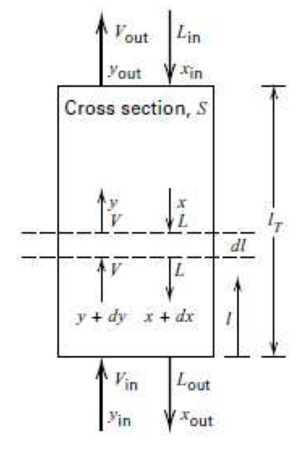
$$Z = \frac{V}{K_y \, a \, A_c} \int_{y_{A\,in}}^{y_{A\,out}} \frac{dy}{(y^*_{A} - y_A)}$$

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

$$Z = \frac{L}{k_x \, a \, A_L} \int_{x_{A\,in}}^{x_{A\,out}} \frac{dx}{(x_A - x_A^*)}$$

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





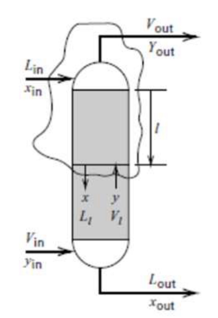
Operating Line for Packed Absorption Tower

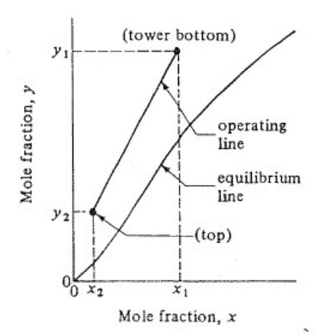
$$Y_{n+1} = \frac{L'}{V'} X_n + \left(Y_1 - \frac{L'}{V'} X_0 \right)$$

For dilute solutions: $V' \cong V$ $L' \cong L$

Absorption vs. Stripping

Absorption

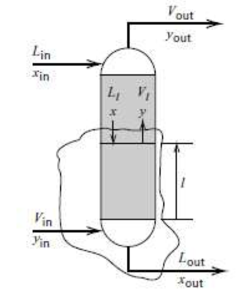


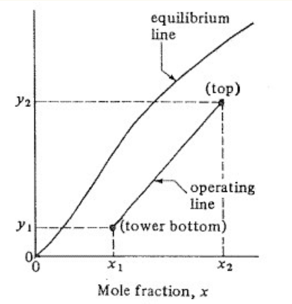


(a)

$$Y = X \frac{L'}{V'} + \left(Y_{out} - X_{in} \frac{L'}{V'} \right)$$

Stripping

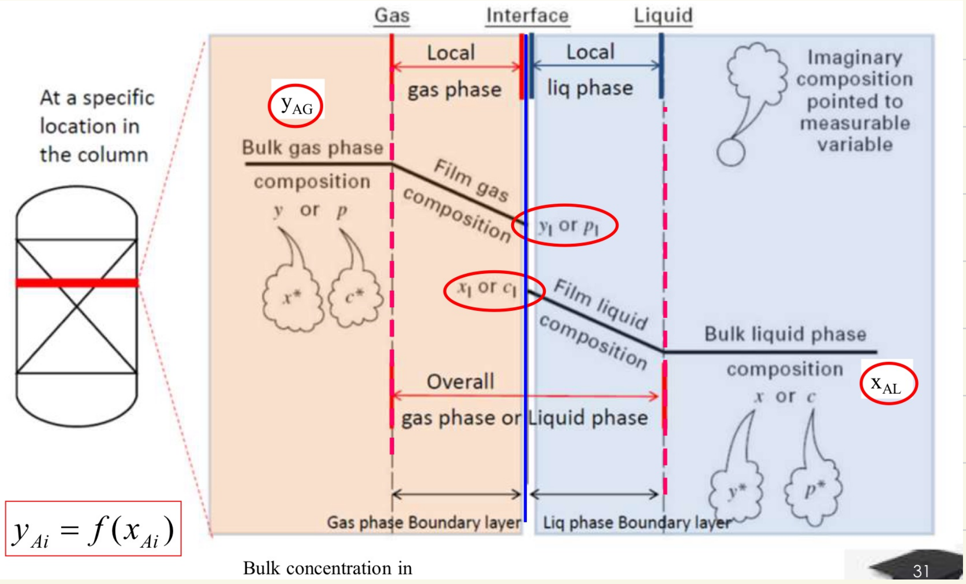




(b)

$$Y = X \frac{L'}{V'} + \left(Y_{in} - X_{out} \frac{L'}{V'} \right)$$

Two Film Theory:



* Two Film Theory (Thin Film): Mass Transfer resistance in binary systems before the interface.

- Mass Transfer → concentration boundary layer

→ y_{AG} & x_{AL} → Bulk compositions

→ y_I & x_I → compositions on Interface

→ y^* → Equilibrium mole fraction of solute in vapor corresponding to x_A in liq

→ x^* → Equilibrium mole fraction of solute in liq corresponding to y_A in vapor.

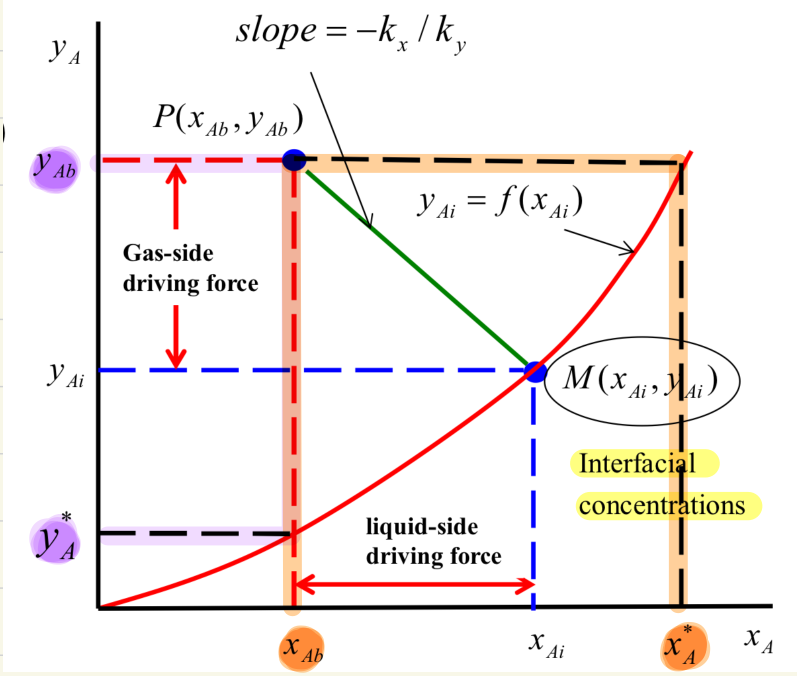
* Consider a steady state Mass Transfer

$$N_A = k_y(y_{Ab} - y_{Ai}) = k_x(x_{Ai} - x_{Ab})$$

$$-\frac{k_x}{k_y} = \frac{y_{Ab} - y_i}{x_{Ab} - x_i}$$

x_I & y_i → interfacial concentrations

x_{Ab} & y_{Ab} → Bulk concentrations.



✳ Overall Mass Transfer Approach

$$N_A = k_y a (y - y_I) = k_x a (x_I - x)$$

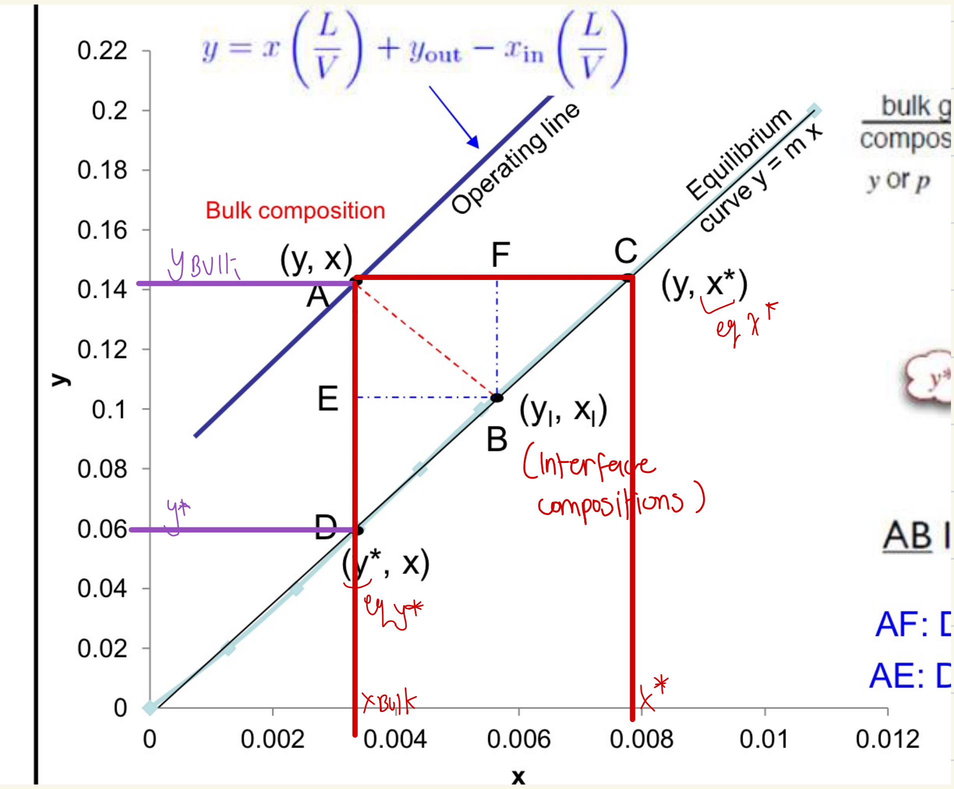
a = surface Area per unit volume of packing

$= \frac{\text{interfacial area}}{\text{unit volume}}$.

$$y = y_I - \frac{k_x a}{k_y a}(x - x_I)$$

↳ Relative resistance of mass transfer between the two phases

$$N_A = k_x a (x^* - x) = k_y a (y - y^*)$$



$$x^* = \frac{y}{m}$$

$$y^* = m x$$

Determining Height of Packing in Tower [HTU method]

For the gas phase, the differential rate of mass transfer of component A is equal to the differential rate of change of the mass of A in the incoming gas stream in a height dz.

$$d\bar{N}_A = \partial(Vy_A) = \partial(Lx_A)$$

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

$$\therefore Z = \int_{y_{out}}^{y_{in}} \underbrace{\left[\frac{V}{K_y a A (1-y_A)_{LM}} \right]}_{HTU_{OG}} \underbrace{\left[\frac{(1-y_A)_{LM} \; dy_A}{(1-y_A)\;(y_A - y_A^*)} \right]}_{NTU_{OG}}$$

* The smaller the HTU, the more efficient is the contacting

* The larger the NTU, the greater is the extent of contacting required

→ For Dilute $(1-y_A) \cong (1-y_A)_{LM} \cong 1$

$$K_y' \cong K_y$$

Based on vapor phase

$$Z = \underbrace{\frac{V}{K_y' a A}}_{} \int_{y_{out}}^{y_{in}} \frac{1}{(y_A - y_A^*)} dy_A$$

↳ H_{OG} : constant through the absorption column.

Based on liquid phase

$$Z = \frac{L}{K_x' a A} \int_{x_{in}}^{x_{out}} \frac{1}{(x_A^* - x_A)} dx_A$$

Step by Step Procedure.

1) Draw the equilibrium curve $[y = mx]$

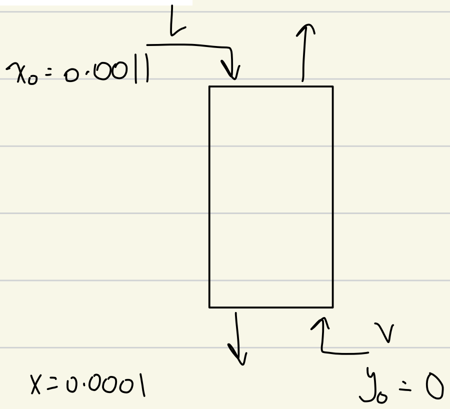
2) Draw the operating line $Y = X\frac{L'}{V'} + (Y_{out} - X_{in}\frac{L'}{V'})$

3) A point (A) (x, y) is taken on the operating line (k_x & k_y) → a line with slope $\frac{k_x}{k_y}$ is drawn to equilibrium line → (x_1, y_1)

We wish to strip SO_2 from water using air at 20C. The inlet air is pure. The outlet water contains 0.0001 mole fraction SO_2, while the inlet water contains 0.0011 mole fraction SO_2. Operation is at 855 mmHg and $L/V = 0.9\times(L/V)_{max}$. Assume H_{OL} = 2.76 feet and that the Henry's law constant is 22,500 mmHg/mole frac SO_2.

$H_{OL} = 2.76$ ft

$H = 22500$ mmHg / mole frac SO_2



$$m = \frac{22500}{855} = 26.315$$

→ Equilibrium $y = mx$

$y = 26.315x$

Dilute system

$Y \cong y \quad X \cong x$

$y_{max} = 26.3\ (0.0011) = 0.0289$

$\frac{L}{V} = 0.9\left(\frac{L}{V}\right)_{max}$ $\left(\frac{L}{V}\right)_{max} = \frac{28.9\times10^{-3} - 0}{11\times10^{-4} - 1\times10^{-4}} = 28.9$

$\frac{L}{V} = 0.9(28.9) = 26.3$

Operating line $Y = \frac{L'}{V'}X + \left(Y_{in} - \frac{L'}{V'}X_{out}\right)$

$Y = 26.3(11\times10^{-4}) + (0 - 26.3(1\times10^{-4}))$

$Y_{out} = 26.3\times10^{-3}$

* → Draw operating line points X_{in}, Y_{out} $(11\times10^{-4}, 26\,3\times10^{-3})$

X_{out}, Y_{in} $(1\times10^{-4}, 0)$

→ Packing Height

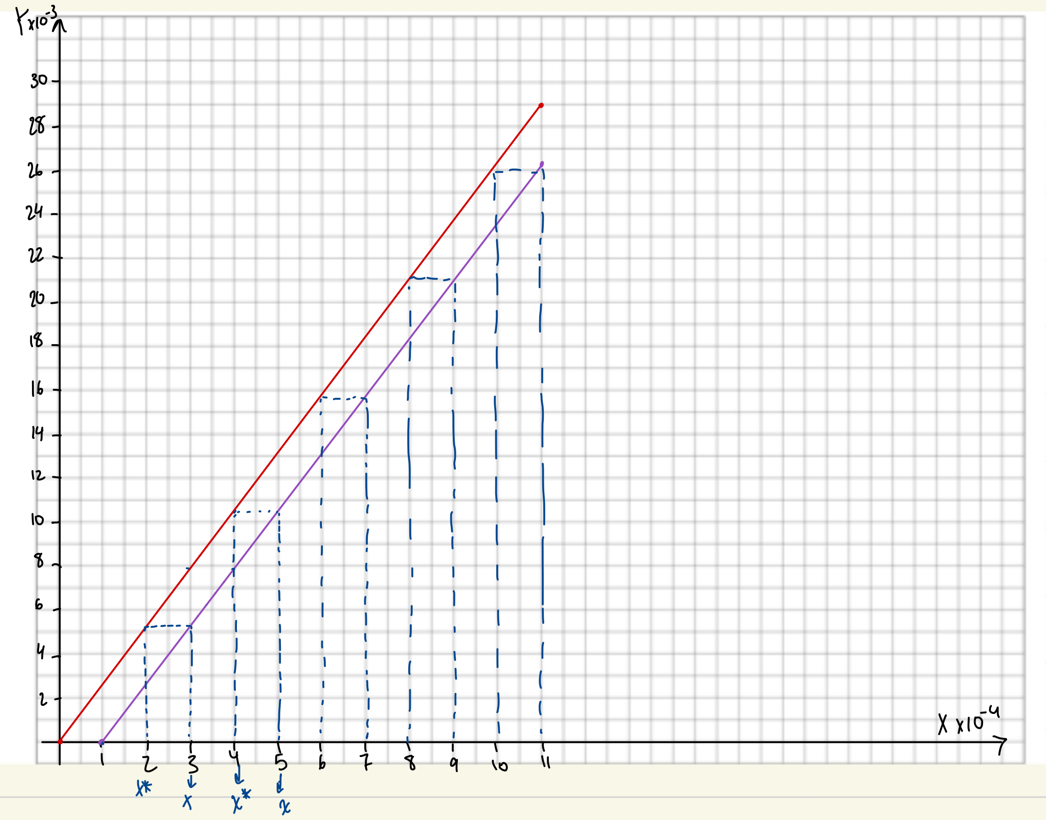
$$Z = \frac{L}{k_x a A}\int_{x_{in}}^{x_{out}} \frac{1}{(x_A^* - x_A)}\,dx_A$$

$$Z = H_{OL}\int_{x_{in}}^{x_{out}} \frac{1}{(x - x^*)}\,dx$$

	X	x^*	$\frac{1}{x-x^*}$
1	1×10^{-4}	0	10 000
2	3×10^{-4}	2×10^{-4}	10 000
3	5×10^{-4}	4×10^{-4}	10000
4	7×10^{-4}	6×10^{-4}	10 000
5	9×10^{-4}	8×10^{-4}	10000
6	11×10^{-4}	10×10^{-4}	10 000

Numerical Integration: Simpson's Rule $\frac{(3-1)\times10^{-4}}{3}\left[10\,000 + 4(2(10000)) + 2(2(10000)) + 10\,000\right]$

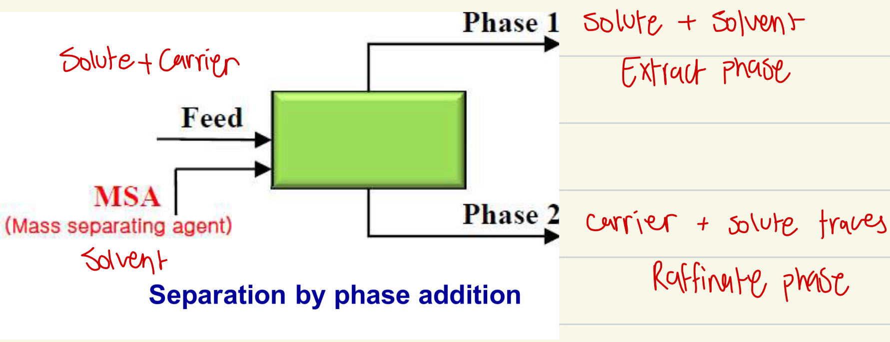
$N_{OL} = 9.3$



$Z = 2.76\ (9.3) = 25.76$ ft

lecture 8: liquid-liquid Extraction

- liquid-liquid extraction is a mass transfer operation, in which a liquid solution (Feed = Solute + carrier) is contacted with an immiscible or nearly immiscible liquid (solvent or organic phase), that exhibits preferential affinity or selectivity towards [solute].



Separation by phase addition

→ Alternative to absorption / Stripping

→ Easier Separation [between solute & solvent]

→ Energy Savings [lower temperatures]

* The simplest liq - liq extraction involves only a ternary system

→ The Feed consists of two miscible components → Carrier (C) & Solute (A)

→ Addition of a second phase → Solvent (S)

- C & S → immiscible or partially soluble

Immiscible liquids form two distinct phases when mixed.

- A & C → miscible
- A & S → miscible or partially soluble

Under Equilibrium Conditions the distribution of solute A over the two phases is determined by the distribution law.

$$K_D = \frac{(y_A)_E}{(x_A)_R}$$

← composition of solute in Extract phase

← composition of solute in raffinate phase

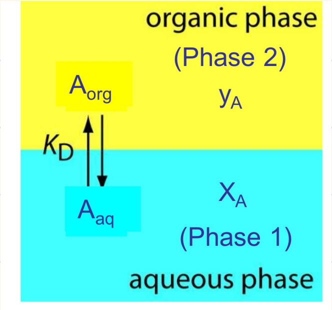
Why Solvent Extraction?

- Relative volatilities are close to unity ($\alpha < 1.1$), or azeotropic → Costly distillation
- Thermally sensative components → will not permit high temperatures.
- When Solute Concentration is low

Properties of Extraction Solvents

- The solvent should be immiscible or only be slighty miscible with Feed aqeous phase
- Relatively low boiling point, for easier removal
- Stable & inert

Distribution Coefficients:



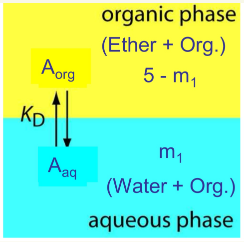
- The partitioning of solute between two chemical phases, is described by the equilibrium constant K_D

$$K_D = \frac{\text{Solute conc. in Extract phase}}{\text{Solute conc. in Raffinate phase}}$$

- Higher value of K, indicates higher extraction efficiency.

Example

An organic molecule with distribution coefficient (or partition coefficient) of 10 between ether and water, and 150 mL of ether will be used to extract 5.0 g of such organic molecule from 100 mL of water.



$$K_D = 10 = \frac{\frac{5-m}{150}}{\frac{m}{100}} \qquad m = 0.31\ g$$

Amount of organic in Ether: $5 - 0.31 = 4.69g$

Extraction efficiency $= \frac{4.69}{5} \times 100\% = 93.8\%$

Example 2

Same amount of ether (150 mL) is used to extract the same organic from 100 mL water but in three portions, each with 50 mL of ether.

cycle 1 .

$$10 = \frac{\frac{5-m}{50}}{\frac{m}{100}} \qquad 10 = \frac{5-m}{50} \times \frac{100}{m}$$

$$500 - 100\,m = 500\,m$$

$$m = 0.833$$

Amount in Extract phase $= 5 - 0.833 = 4.16$

efficiency $= 4.16 / 5 = 83.3\%$

cycle 2 :

$$10 = \frac{\frac{0.83-m}{50}}{\frac{m}{100}} \qquad 10 = \frac{0.83-m}{50} \times \frac{100}{m}$$

$$500\,m = 83 - 100\,m$$

$$m = 0.14$$

Amount in Extract phase : $5 - 0.14 = 4.86$

efficiency : $\frac{4.86}{5} = 97.2\%$

cycle 3 :

$$10 = \frac{\frac{0.14-m}{50}}{\frac{m}{100}} \qquad 500\,m = 14 - 100\,m$$

$$m = 0.023$$

Amount in Extract · $5 - 0.02 = 4.98$

efficiency $\frac{498}{5} = 99.6\%$

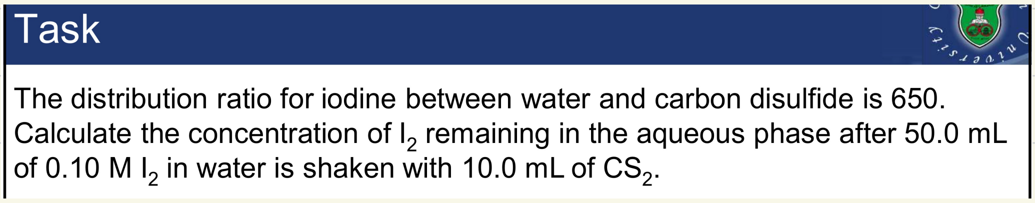
solute remains in aq. phase

initial mass of solute in feed

number of cycles

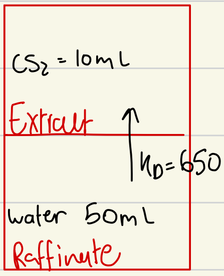
$$\frac{x}{W_0} = \left[\frac{V_{feed}}{V_{feed} + K_D V_{solvent}}\right]^n$$

$$\frac{x}{5} = \left[\frac{100}{100 + 10(50)}\right]^3 \qquad x = 0.02$$



Task

The distribution ratio for iodine between water and carbon disulfide is 650. Calculate the concentration of I_2 remaining in the aqueous phase after 50.0 mL of 0.10 M I_2 in water is shaken with 10.0 mL of CS_2.



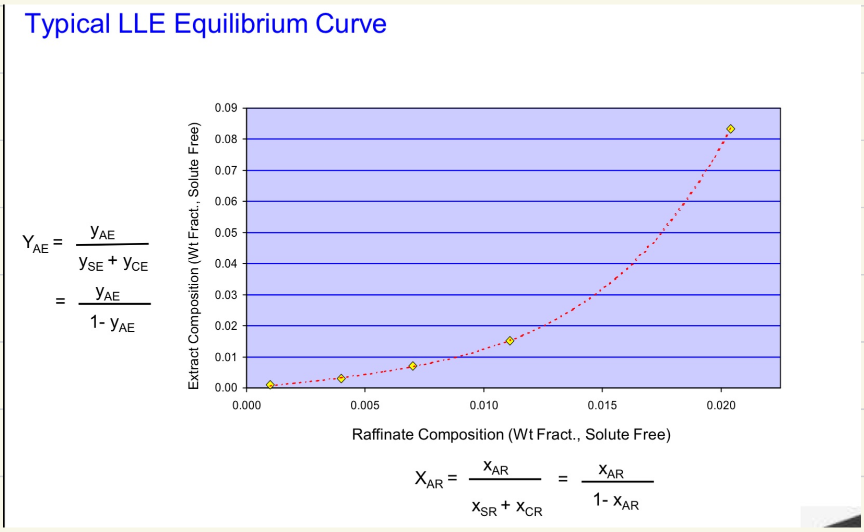
$m = 50(0.1) = 5$ mmol of Iodine

$$K_D = 650 = \frac{\frac{5-m}{10}}{\frac{m}{50}} \qquad m = 0.038 \text{ mmol}$$

conc of Iodine in water $\frac{0.038}{50} = 0.00076$ M

In CS_2 = $5 - 0.038 = 4.962$

Liquid - Liquid System

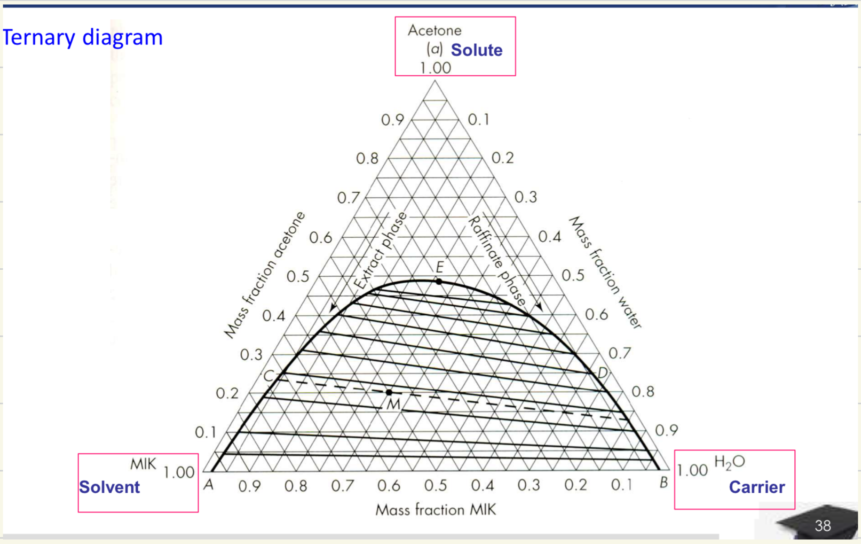


Typical LLE Equilibrium Curve

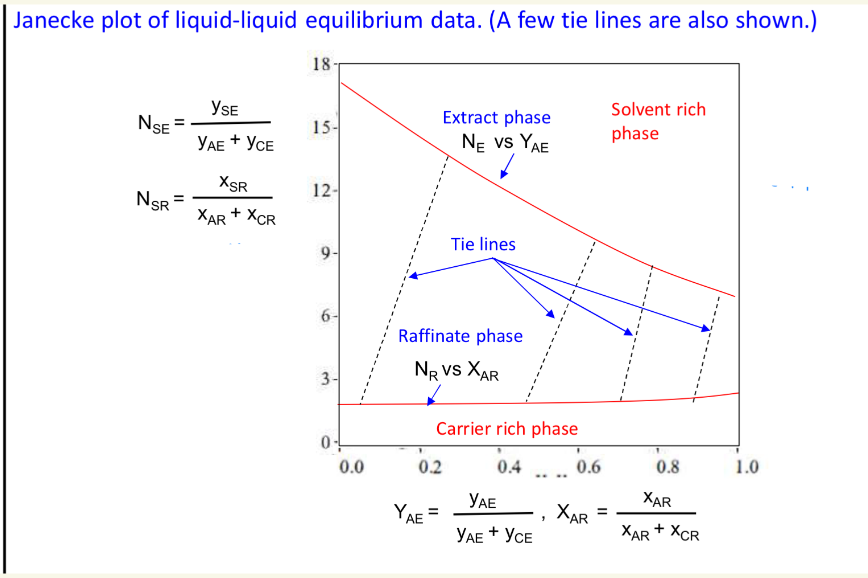
$$Y_{AE} = \frac{y_{AE}}{y_{SE} + y_{CE}}$$

$$= \frac{y_{AE}}{1 - y_{AE}}$$

$$X_{AR} = \frac{x_{AR}}{x_{SR} + x_{CR}} = \frac{x_{AR}}{1 - x_{AR}}$$



Ternary diagram

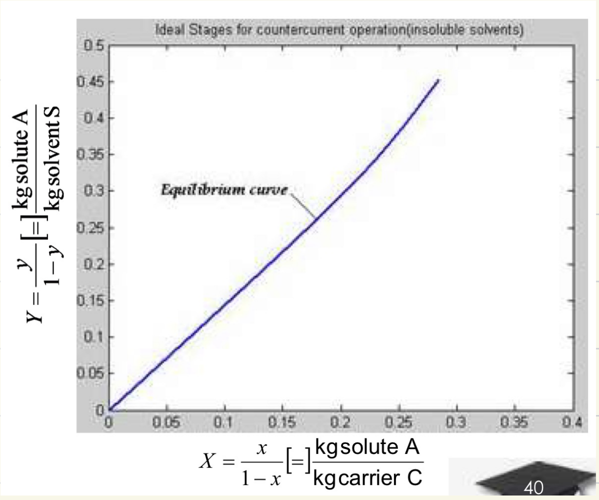


Janecke plot of liquid-liquid equilibrium data. (A few tie lines are also shown.)

$$N_{SE} = \frac{y_{SE}}{y_{AE} + y_{CE}}$$

$$N_{SR} = \frac{x_{SR}}{x_{AR} + x_{CR}}$$

$$Y_{AE} = \frac{y_{AE}}{y_{AE} + y_{CE}}, \quad X_{AR} = \frac{x_{AR}}{x_{AR} + x_{CR}}$$



Type III : A → soluble in C & S
C → completely insoluble in S

Extract & Raffinate → Binary mixtures

Type I : A → soluble in C & S completely
C → partially miscible in S

Type II : A → completely soluble in C , partially in S
C → partially miscible in S

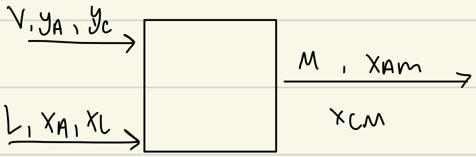
* Assumptions: Isothermal
Isobaric
Negligable heat of mixing
No chemical reactions

Effect of Temperature on Solubility

Usually, the solubility increases as the temp increases, for this reason most liq - liq extraction systems operate at low T

Pressure has little effect on solubility

Single Stage Equilibrium



Overall mass Balance $M = V + L$

Component Balance on A $x_{Am} M = x_A L + y_A V$

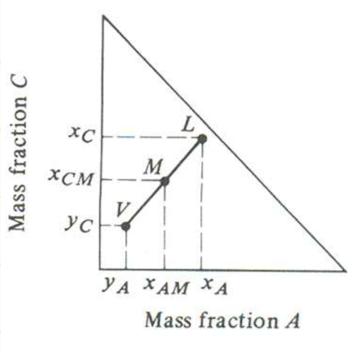
Component Balance on C $x_{cm} M = x_c L + y_c V$

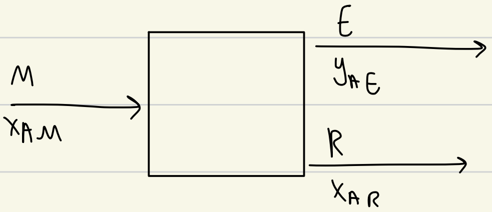
* Lever Rule Derivation

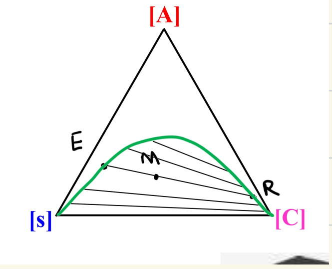
$$\frac{L}{V} = \frac{y_A - x_{Am}}{x_{Am} - x_A} = \frac{y_c - x_{cm}}{x_{cm} - x_c}$$

Lever arm Rule $\frac{L}{V} = \frac{\overline{VM}}{\overline{ML}}$

$$\frac{L}{M} = \frac{\overline{VM}}{\overline{LV}}$$



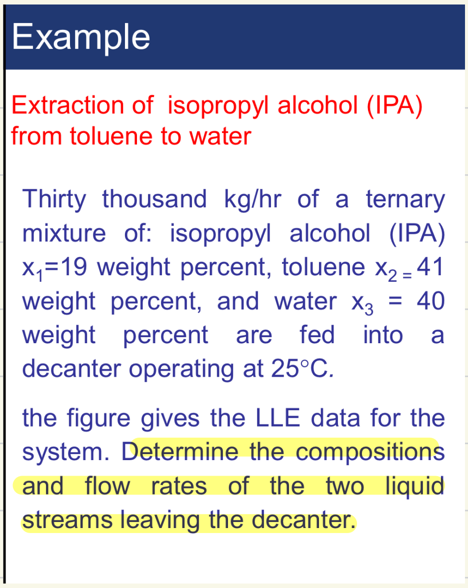




Overall Balance $M = E + R$

Component Balance $x_{AM} M = y_{AE} E + x_{AR} R$

$$\frac{R}{E} = \frac{y_{AE} - x_{AM}}{x_{AM} - x_{AR}} = \frac{\overline{EM}}{\overline{MR}}$$

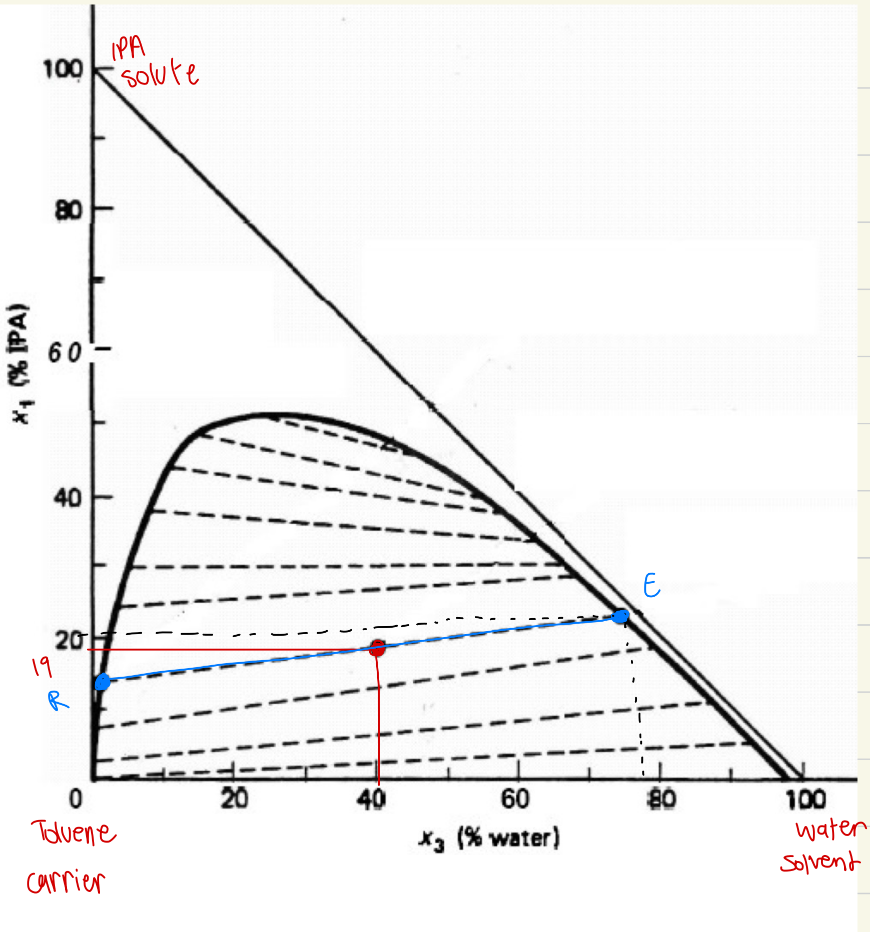


Example

Extraction of isopropyl alcohol (IPA) from toluene to water

Thirty thousand kg/hr of a ternary mixture of: isopropyl alcohol (IPA) x_1=19 weight percent, toluene x_{2} = 41 weight percent, and water x_3 = 40 weight percent are fed into a decanter operating at 25°C.

the figure gives the LLE data for the system. Determine the compositions and flow rates of the two liquid streams leaving the decanter.



Compositions:

E: 74 % water

23% IPA

3% Toluene

R: 2% H_2O

14% IPA

84% toluene

Flow Rates:

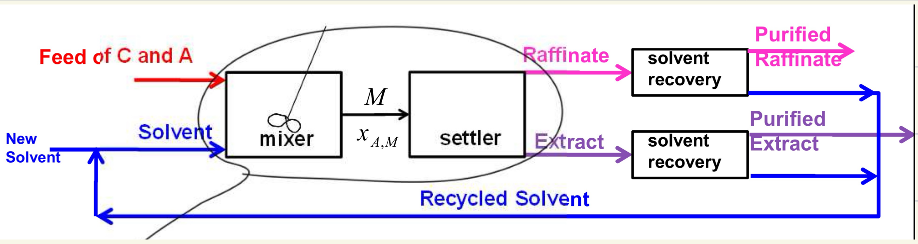
Overall Balance $M = E + R$ $30000 = E + R$

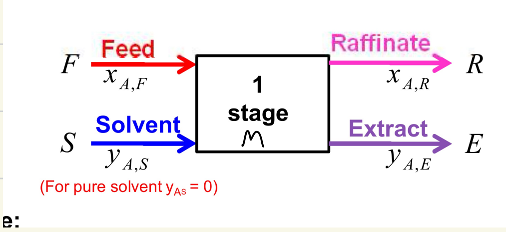
Component Balance (water) $x_{AM} M = y_A E + x_A R$ $30000(0.4) = E(0.74) + R(0.02)$

E: 15833 3 Kg/h

R: 14166.6 Kg/h

Mixer - Settler (single stage extraction)





Overall Balance : $F + S = R + E = M$

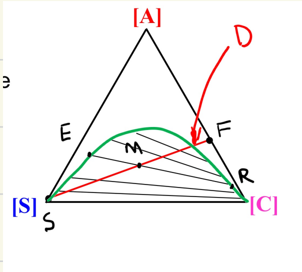
$$x_{AF} F + y_{AS} S = x_{AR} R + y_{AE} E = x_{AM} M$$

Given

$$x_{AM} = \frac{x_{AF} F + y_{AS} S}{F + S}$$

Fraction of solute Recovered $f = 1 - \frac{X_{AR} R}{X_{AF} F}$

→ The minimum Solvent required

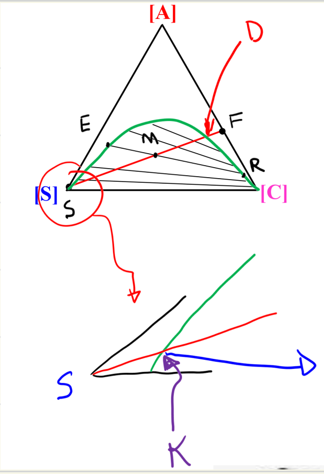


$$F + S_{min} = D$$

$$x_{AF} F + y_{AS} S_{min} = x_{AD} D$$

$$\frac{S_{min}}{F} = \frac{\overline{FD}}{\overline{DS}} = \frac{x_{AF} - x_{AD}}{x_{AD} - y_{AS}}$$

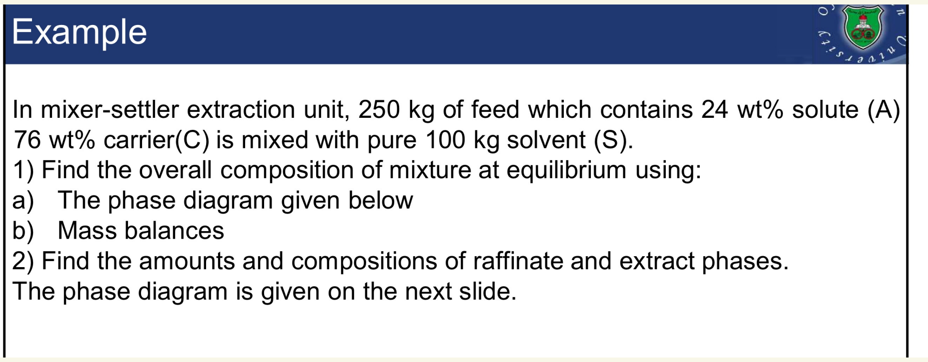
→ Maximum Solvent



$$F + S_{max} = k$$

$$x_{AF} F + y_{AS} S_{max} = x_{AK} K$$

$$\frac{S_{max}}{F} = \frac{\overline{Fk}}{\overline{KS}} = \frac{x_{AF} - x_{AK}}{x_{AK} - y_{AS}}$$



Example

In mixer-settler extraction unit, 250 kg of feed which contains 24 wt% solute (A) 76 wt% carrier(C) is mixed with pure 100 kg solvent (S).

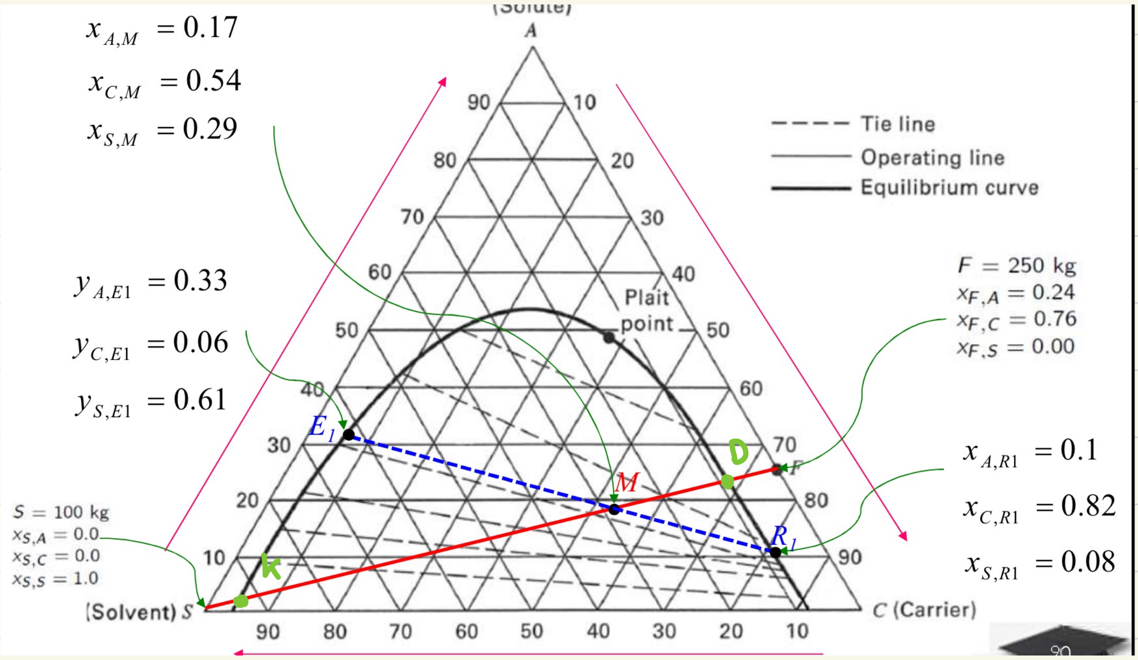
1) Find the overall composition of mixture at equilibrium using:

a) The phase diagram given below

b) Mass balances

2) Find the amounts and compositions of raffinate and extract phases.

The phase diagram is given on the next slide.



$x_{A,M} = 0.17$

$x_{C,M} = 0.54$

$x_{S,M} = 0.29$

$y_{A,E1} = 0.33$

$y_{C,E1} = 0.06$

$y_{S,E1} = 0.61$

$S = 100$ kg
$x_{S,A} = 0.0$
$x_{S,C} = 0.0$
$x_{S,S} = 1.0$

$F = 250$ kg
$x_{F,A} = 0.24$
$x_{F,C} = 0.76$
$x_{F,S} = 0.00$

$x_{A,R1} = 0.1$

$x_{C,R1} = 0.82$

$x_{S,R1} = 0.08$

$$F + S = M$$

$$250 + 100 = M$$

$$M = 350$$

$$x_{AF} F + y_{AS} S = x_{AM} M$$

$$0.25(250) + 0 = x_{AM}(350)$$

$$x_{Am} = 0.17$$

$R + E = M$ $\qquad$ $x_{RA} R + y_{AE} E = x_{AM} M$

$R + E = 350$ $\qquad$ $0.1 R + 0.33 E = 0.17 (350)$

$R = 243.47$

$E = 106.52$

Solute Recovery : $f = 1 - \frac{0.1(243.47)}{0.24(250)}$ $= 59.4\%$

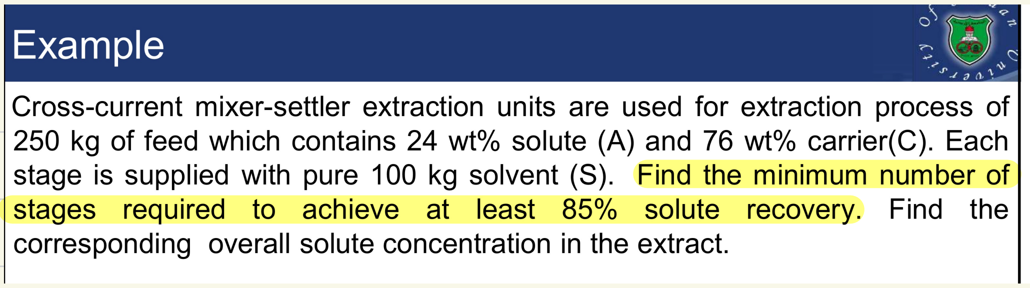
Minimum Solvent:

$x_{CD} = 0.68$ $\qquad$ $\frac{S_{min}}{F} = \frac{FD}{DS} = \frac{x_{AF} - x_{AD}}{x_{AD} - y_{AS}}$

$x_{AD} = 0.23$

$x_{SD} = 0.09$ $\qquad$ $\frac{S_{min}}{250} = \frac{0.24 - 0.23}{0.23 - 0}$

$S_{min} = 10.87$ kg



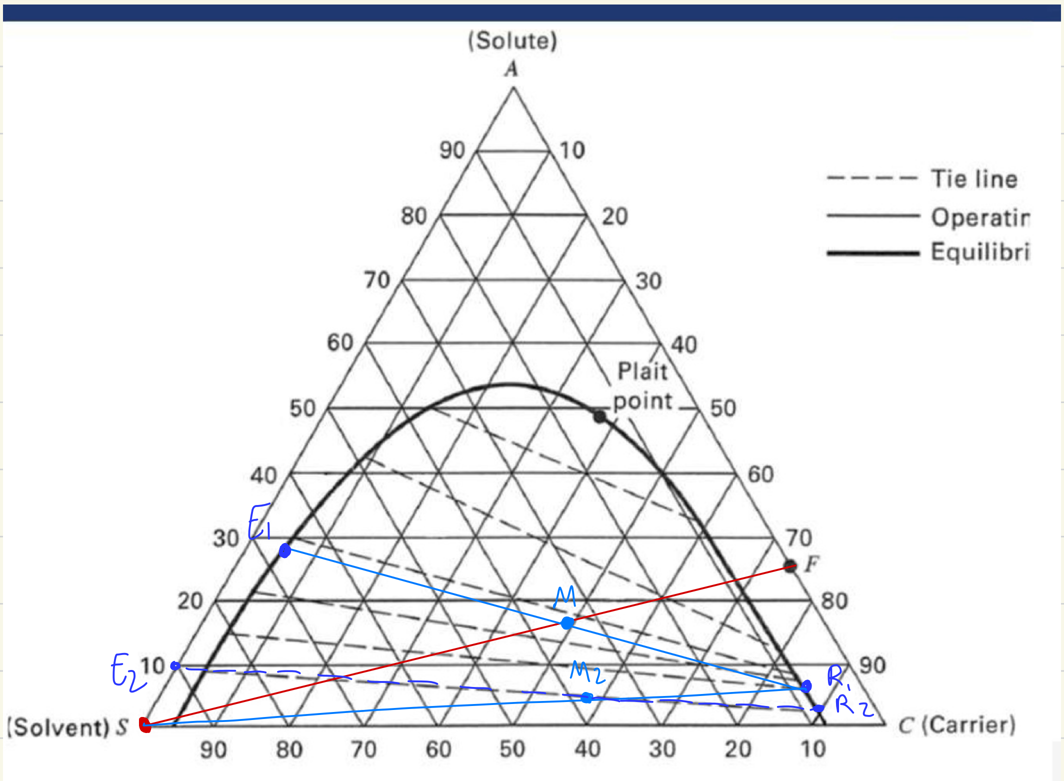
Example

Cross-current mixer-settler extraction units are used for extraction process of 250 kg of feed which contains 24 wt% solute (A) and 76 wt% carrier(C). Each stage is supplied with pure 100 kg solvent (S). Find the minimum number of stages required to achieve at least 85% solute recovery. Find the corresponding overall solute concentration in the extract.

$F = 250$ $S = 100$

$x_{AF} = 0.24$ $y_{AS} = 0$

$x_{AC} = 0.76$



Overall Balance $F + S = M$ $250 + 100 = M$ $M = 350\,kg$

Component Balance $x_{AF}F + y_{AS}S = x_{Am}M$ $0.24(250) + 0 = x_{Am}(350)$ $x_{AM} = 0.17$

$E + R = M$ $E + R = 350$

$y_{Ae}E + x_{AR}R = x_{Am}M$ $0.33E + 0.1R = 0.17(350)$

$R_1 = 243.37\ kg$

$E_1 = 106.52\ kg$

Stage 2. overall Balance $R_1 + S_1 = M_2$ $243.37 + 100 = 343.37$

component $x_{AR} R + y_{AS} S = x_{AM_2} M_2$ $0.1(243.37) + 0 = x_{AM_2}(343\ 37)$

$x_{AM_2} = 0.071$

overall $E + R = M$ $E_2 + R_2 = 343.37$

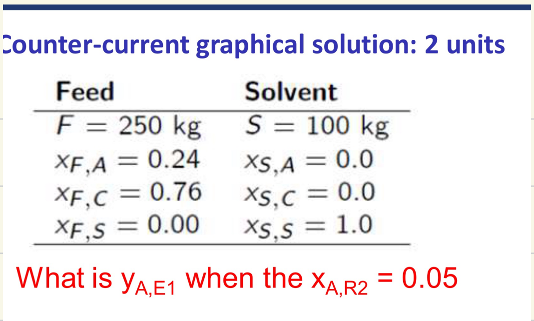
component $y_{AE} E + x_{AR} R = x_{AM} M$ $0.11 E_2 + 0.04 R_2 = 0.071(343.37)$

$E_2 = 152\ 06$

$R_2 = 191.3$

Solute Recovery : $f = 1 - \dfrac{0.04(191.3)}{0.24(250)}$ $= 87.24$ > Required

Overall solute concentration $= \dfrac{\Sigma y_{AE} E}{\Sigma E} = \dfrac{0.33(106.52) + 0.11(152.06)}{106.52 + 152.06} = 0.2$



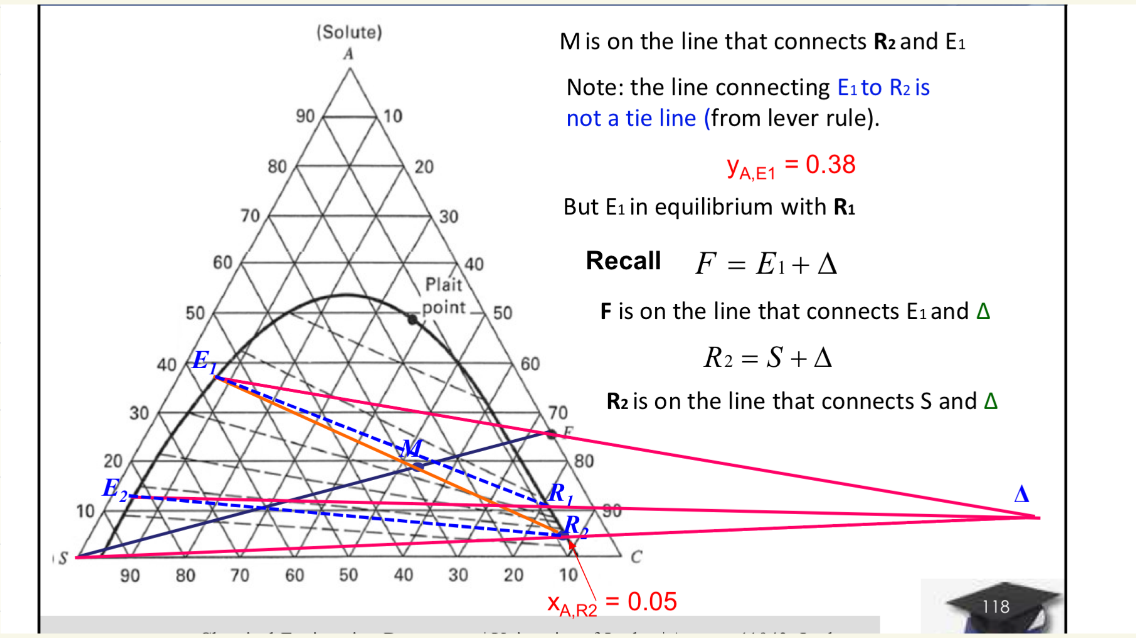
ounter-current graphical solution: 2 units

Feed	Solvent
$F = 250$ kg	$S = 100$ kg
$x_{F,A} = 0.24$	$x_{S,A} = 0.0$
$x_{F,C} = 0.76$	$x_{S,C} = 0.0$
$x_{F,S} = 0.00$	$x_{S,S} = 1.0$

What is $y_{A,E1}$ when the $x_{A,R2} = 0.05$

$F + S = M \qquad 250 + 100 = 350$

$0.24(250) + 0 = x_{Am}(350) \qquad x_{Am} = 0.17$



$x_{A,R2} = 0.05$

M is on the line that connects **R**$_2$ and E$_1$

Note: the line connecting E$_1$ to R$_2$ is not a tie line (from lever rule).

$y_{A,E1} = 0.38$

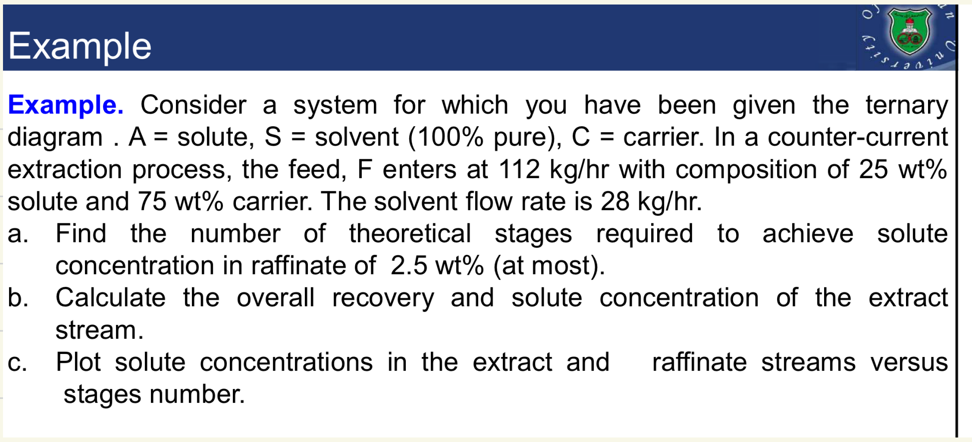
But E$_1$ in equilibrium with **R**$_1$

Recall $F = E_1 + \Delta$

F is on the line that connects E$_1$ and Δ

$$R_2 = S + \Delta$$

R$_2$ is on the line that connects S and Δ



Example

Example. Consider a system for which you have been given the ternary diagram . A = solute, S = solvent (100% pure), C = carrier. In a counter-current extraction process, the feed, F enters at 112 kg/hr with composition of 25 wt% solute and 75 wt% carrier. The solvent flow rate is 28 kg/hr.

a. Find the number of theoretical stages required to achieve solute concentration in raffinate of 2.5 wt% (at most).

b. Calculate the overall recovery and solute concentration of the extract stream.

c. Plot solute concentrations in the extract and raffinate streams versus stages number.

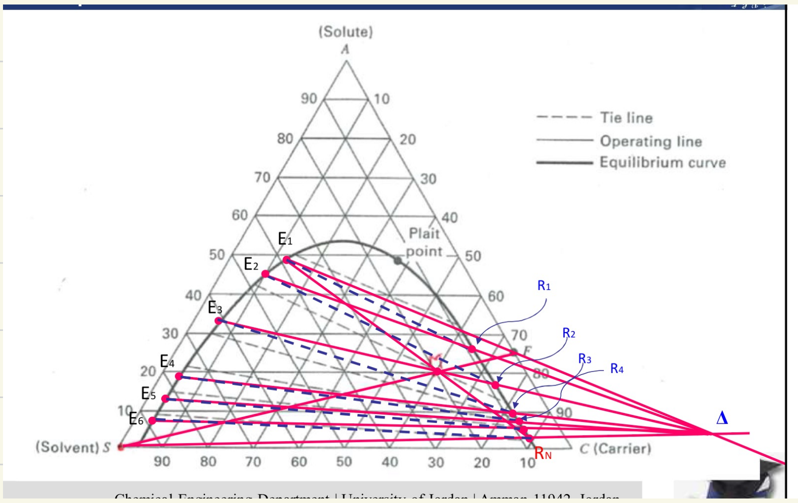
$F = 112\ kg/hr$ $S = 28\ kg/hr$

$x_{AF} = 0.25$ $y_{AS} = 0$

$x_{CF} = 0.75$

$F + S = M$ $x_{AF}F + y_{AS}S = x_{Am}M$

$112 + 28 = 140$ $0.25(112) + 0 = x_{AM}(140)$ $x_{AM} = 0.2$



6 stages

Solute Recovery $f = 1 - \frac{x_{AR}R}{x_{AF}F}$

$E + R = 140$ $0.48E_1 + 0.025R_N = 0.2(140)$

$E_1 = 53.84\ kg/hr$

$R_N = 86.15\ kg/hr$

$f = 1 - \frac{0.025(86.15)}{0.25(112)} = 92.3\%$

$F - E_1 = \Delta$

$112 - 53.84 = 58.16$

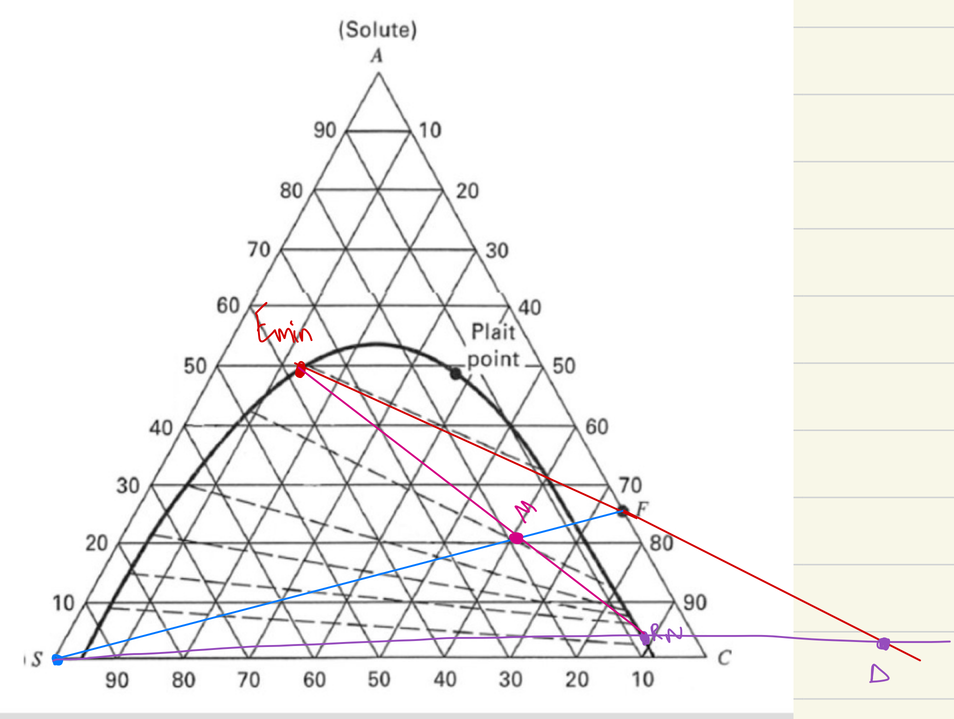
Determining Minimum Solvent

① Draw tie line passing through feed to determine E_{min}

② Draw FS line

③ Draw S & R_N line to determine Δ

④ Draw R_N & E_{min} line to find M_{min}



$$F + S_{min} = M_{min} = R + E$$

$$x_{AF} F + y_{AS} S_{min} = x_{AM} M_{min}$$

$$\frac{S_{min}}{F} = \frac{FM}{MS_{min}} = \frac{x_{AF} - x_{AM}}{x_{AM} - y_{AS}}$$

with $\frac{S_{min}}{F}$, Number of theoretical stages → ∞

Determining Maximum Solvent

$\frac{S_{max}}{F}$ occurs when point M reaches the extract phase curve

$$\frac{S_{max}}{F} = \frac{x_{AF} - x_{AM}}{x_{AM} - y_{AS}}$$

Counter current Stage Extraction with immiscible liquids

- The carrier C is immiscible in solvent S
  - ↳ Raffinate → Binary mixture of A & C
  - ↳ Extract → Binary mixture of A & S
- xy equilibrium diagram is used.

$$K_D = \frac{(y_A)_E}{(x_A)_R}$$

$$F + S = R + E$$

$$x_{AF} F + y_{AS} S = x_{AR} R + y_{AE} E$$

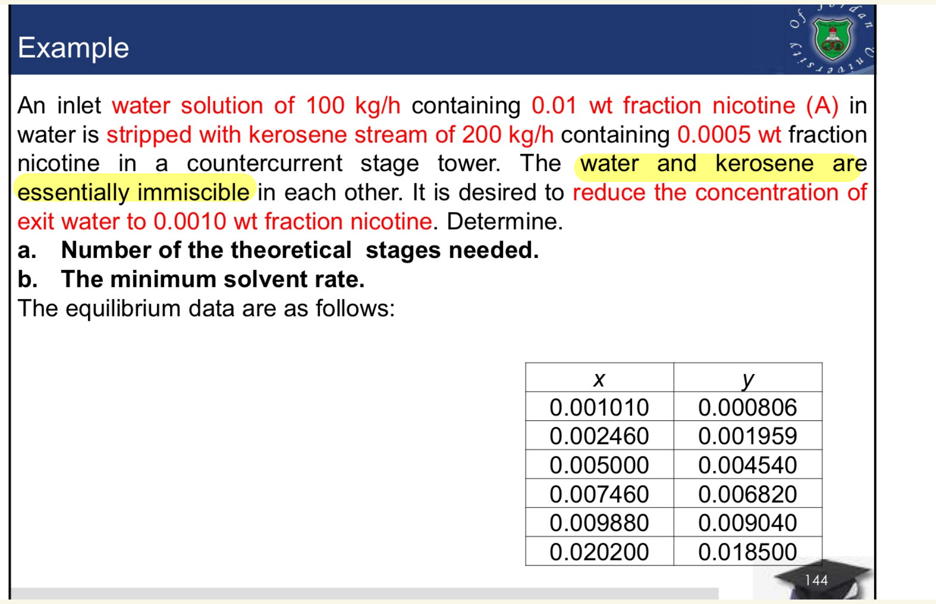
X → Solute to carrier ratio in Raffinate

Y → Solute to solvent ratio in extract

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

$$R' = F(1 - x_{AF}) = R(1-x)$$

Operating line $\underset{(in)}{Y_{N+1}} = \frac{R'}{E'} \underset{(in)}{X_N} + \underset{(out)}{Y_1} - \frac{R'}{E'} \underset{(out)}{X_0}$



Example

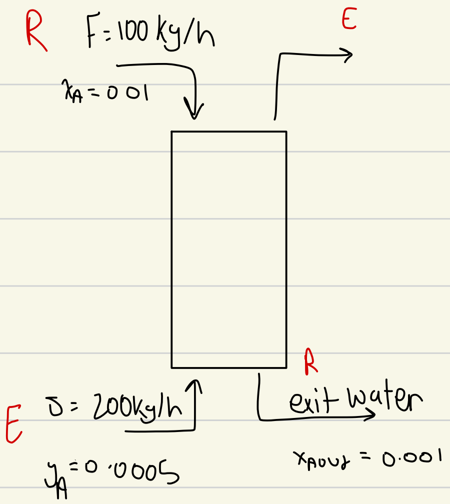
An inlet water solution of 100 kg/h containing 0.01 wt fraction nicotine (A) in water is stripped with kerosene stream of 200 kg/h containing 0.0005 wt fraction nicotine in a countercurrent stage tower. The water and kerosene are essentially immiscible in each other. It is desired to reduce the concentration of exit water to 0.0010 wt fraction nicotine. Determine.

a. Number of the theoretical stages needed.

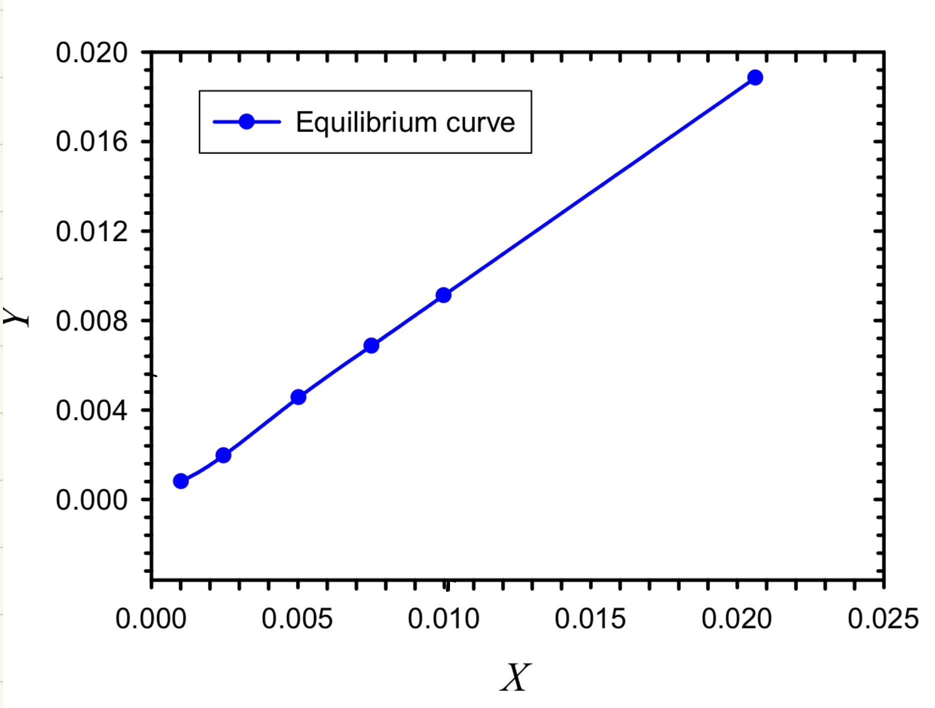
b. The minimum solvent rate.

The equilibrium data are as follows:

x	*y*
0.001010	0.000806
0.002460	0.001959
0.005000	0.004540
0.007460	0.006820
0.009880	0.009040
0.020200	0.018500



$X \simeq x \qquad Y \simeq y \Rightarrow$ Dilute system



$R' = R(1-x) = 100(1-0.01) = 99$

$E' = E(1-y) = 200(1-0.0005) = 199.9$

Operating line $\quad Y_{in} = \frac{R'}{E'} X_{out} + Y_{out} - \frac{R'}{E'} X_{in}$

$Y_{in} = \frac{0.0005}{1-0.0005} = 0.0005$

$X_{in} = \frac{0.01}{1-0.01} = 0.01$

$X_{out} = \frac{0.001}{1-0.001} = 0.001$

$$0.0005 = \frac{99}{199.9}(0.001) + Y_{out} - \frac{99}{199.9}(0.01)$$

$$Y_{out} = 0.005$$

Operating line points: X_{in} Y_{out} $(0.01, 0.005)$

X_{out} Y_{in} $(0.001, 0.0005)$