Sapartion1

2st
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Flash Distillation

D1. One of the prerequisites for study of separations is the ability to convert from weight to mole fractions and vice versa. As a refresher in this conversion, solve the following problem: We have a flow rate of 1500 kmol/h of a feed that is 40 mol% ethanol and 60 mol% water. What is the weight fraction of ethanol, and what is the total flow rate in pounds per hour?

D1. *New problem for 3rd edition.* Basis 1kmol feed.

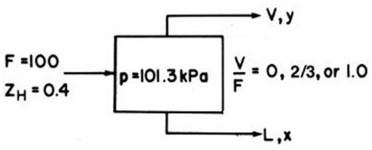
$$.4 \text{ kmole E } = (.4)(\text{MW} = 46) = 18.4 \text{ kg}$$

$$.6 \text{ kmol Water } = .6\ (\text{MW} = 18) = \frac{10.8 \text{ kg}}{\text{total} = 29.2 \text{ kg}}$$

Weight fraction ethanol = 18.4/29.2 = 0.630

Flow rate = (1500 kmol/hr)[(29.2kg)/(1 kmol)] = 43,800 kg/hr.

A flash distillation chamber operating at 101.3 kPa is separating an ethanol-water mixture. The feed mixture is 40 mol% ethanol and F = 100 kmol/h. (a) What is the maximum vapor composition and (b) what is the minimum liquid composition that can be obtained if V/F is allowed to vary? (c) If V/F = 2/3, what are the liquid and vapor compositions? (d) Repeat step c, given that F is specified as 1000 kmol/h.



Mass Balances:

$$F = V + L$$

$$Fz = Vy + Lx$$

Solve for y:

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

From the overall balance, L = F – V. Thus

when V/F = 0.0, V = 0, L = F, and L/V = F/0 = ∞

when V/F = 2/3, V = (2/3)F, L = (1/3)F, and L/V = (1/3)F/[(2/3)F] = 1/2

when V/F = 1.0, V = F, L = 0, and L/V = 0/F = 0

Thus the slopes (–L/V) are –∞, –1/2, and –0.

If we solve for the y = x interception, we find it at y = x = z = 0.4 for all cases. Thus we can plot three operating lines through y = x = z = 0.4, with slopes of –∞, –1/2 and –0. These operating lines were shown in Figure 2-8.

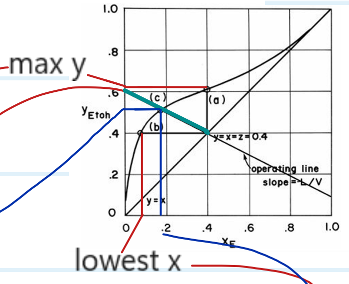
$y = -\frac{1}{2}x + \frac{F}{V}\cdot z$

intercept

$\frac{3}{2} \times 0.4$

$= 0.6$

Figure 2-8.



a. Highest y is for V/F = 0: y = 0.61 [x = 0.4]

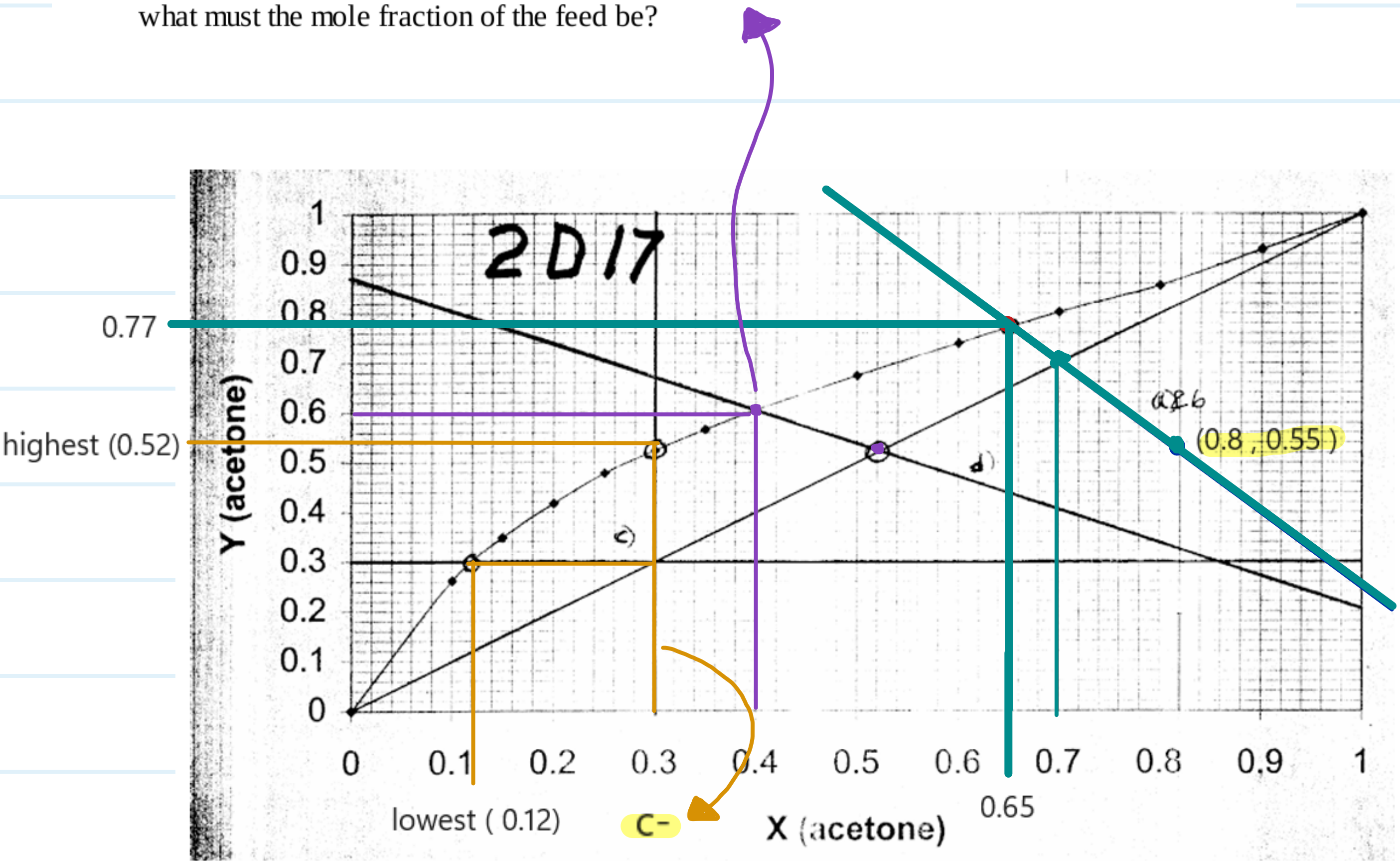
b. Lowest x is for V/F = 1.0: x = 0.075 [y = 0.4]

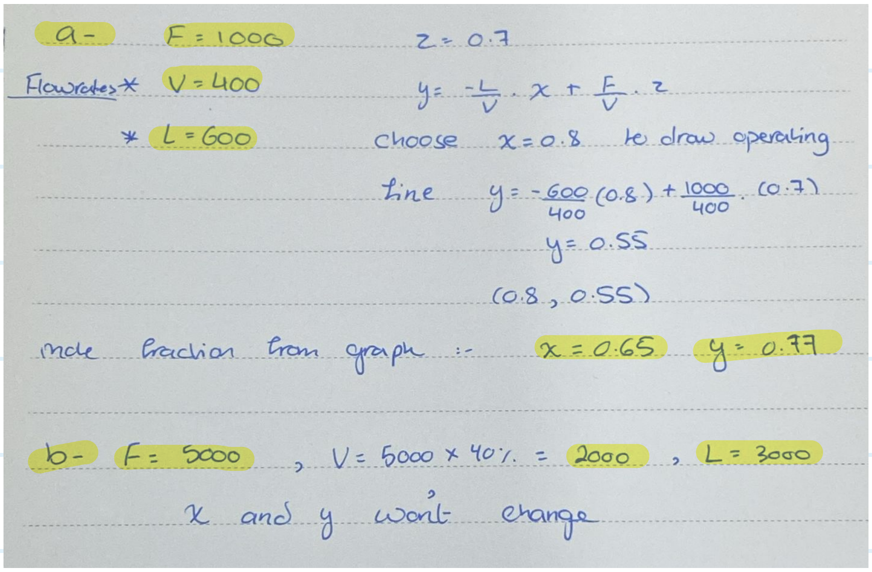
c. When V/F is 2/3, y = 0.52 and x = 0.17

d. When F = 1,000 with V/F = 2/3, the answer is exactly the same as in part c

D17. We are separating a mixture of acetone (MVC) from ethanol by flash distillation at p = 1 atm. Equilibrium data are listed in Problem 4.D7. Solve graphically.

- **a.** 1000 kmol/day of a feed that is 70 mol% acetone is flash distilled. If 40% of the feed is vaporized, find the flow rates and mole fractions of the vapor and liquid products.
- **b.** Repeat part a for a feed rate of 5000 kmol/day.
- **c.** If feed is 30 mol% acetone, what is the lowest possible liquid mole fraction and the highest possible vapor mole fraction?
- **d.** If we want to obtain a liquid product that is 40 mol% acetone while flashing 60% of the feed, what must the mole fraction of the feed be?




a- F = 1000 z = 0.7

Flowrates* V = 400 $y = -\frac{L}{V} \cdot x + \frac{F}{V} \cdot z$

* L = 600 choose x = 0.8 to draw operating

line $y = -\frac{600}{400}(0.8) + \frac{1000}{400} \cdot (0.7)$

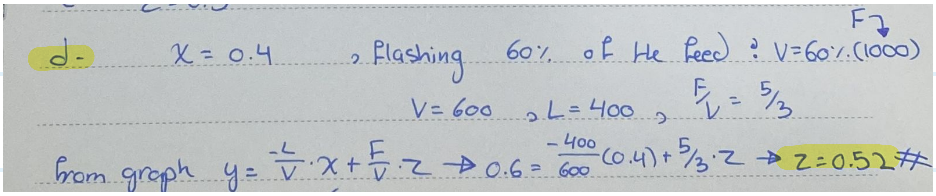
y = 0.55

(0.8 , 0.55)

mole fraction from graph :- x = 0.65 y = 0.77

b- F = 5000 , V = 5000 × 40% = 2000 , L = 3000

x and y won't change


d- x = 0.4 , Flashing 60% of the feed : $V = 60\%.\overset{F}{(1000)}$

V = 600 , L = 400 , $\frac{F}{V} = \frac{5}{3}$

from graph $y = -\frac{L}{V} \cdot x + \frac{F}{V} \cdot z \rightarrow 0.6 = \frac{-400}{600}(0.4) + \frac{5}{3} \cdot z \rightarrow z = 0.52$#

Energy separating agent

mass separating agent

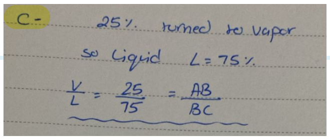
1.4. Use of an ESA or an MSA.

Compare the advantages and disadvantages of making separations using an ESA versus using an MSA.

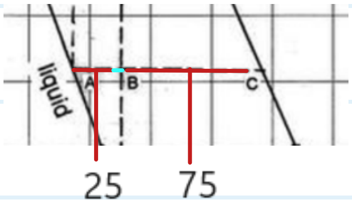
Analysis: With an MSA, an additional separator is needed to recover the MSA. Also, some MSA will be lost, necessitating the need for MSA makeup. If the MSA is incompletely recovered, a small amount of contamination may result. The use of an MSA can make possible a separation that can not be carried out with an ESA. An ESA separation is easier to design.

4.6. Partial vaporization of a nonideal binary mixture.

A liquid mixture containing 25 mol% benzene and 75 mol% ethyl alcohol, in which components are miscible in all proportions, is heated at a constant pressure of 1 atm from 60°C to 90°C. Using the following *T–x–y* experimental data, determine (a) the temperature where vaporization begins; (b) the composition of the first bubble of vapor; (c) the composition of the residual liquid when 25 mol % has evaporated, assuming that all vapor formed is retained in the apparatus and is in equilibrium with the residual liquid. (d) Repeat part (c) for 90 mol% vaporized. (e) Repeat part (d) if, after 25 mol% is vaporized as in part (c), the vapor formed is removed and an additional 35 mol% is vaporized by the same technique used in part (c).

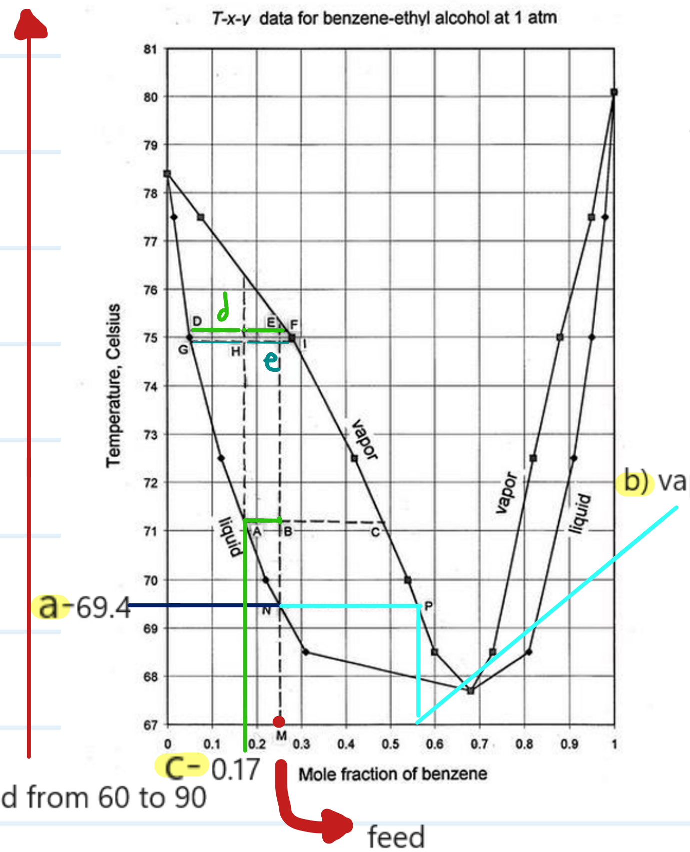

c- 25% turned to vapor
so Liquid L = 75%

$\frac{V}{L} = \frac{25}{75} = \frac{AB}{BC}$



بقرأ النقطة الي خط ال vapor ل خط ال liquid بيعطيني 25/75

The benzene mole fraction in the equilibrium liquid 0.175 at 71.2C



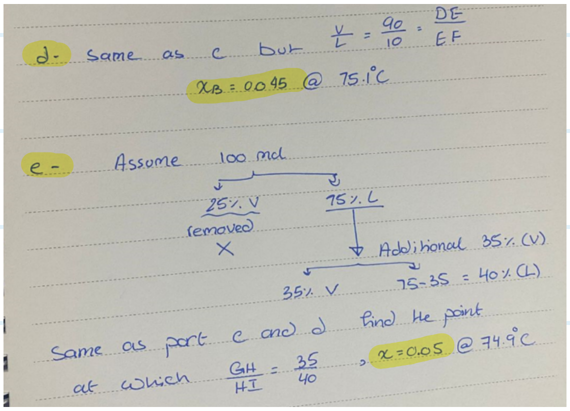
a-69.4

C- 0.17

heated from 60 to 90

feed

b) vapor composition?
0.56


d- Same as c but $\frac{V}{L} = \frac{90}{10} = \frac{DE}{EF}$

$x_B = 0.045$ @ 75.1°C

e- Assume 100 mol

25% V removed ×

75% L

Additional 35% (V)

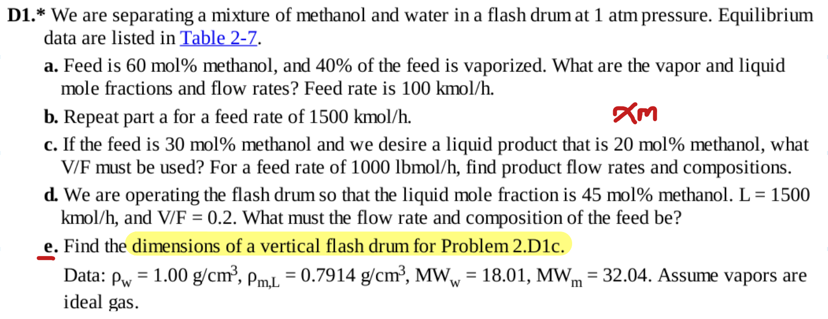
35% V

75-35 = 40% (L)

Same as part c and d find the point at which $\frac{GH}{HI} = \frac{35}{40}$, $x = 0.05$ @ 74.9°C

d- 0.045 e- 0.05

محلول بالسلايدات (a-d) ⭐



D1.* We are separating a mixture of methanol and water in a flash drum at 1 atm pressure. Equilibrium data are listed in Table 2-7.

a. Feed is 60 mol% methanol, and 40% of the feed is vaporized. What are the vapor and liquid mole fractions and flow rates? Feed rate is 100 kmol/h.

b. Repeat part a for a feed rate of 1500 kmol/h.

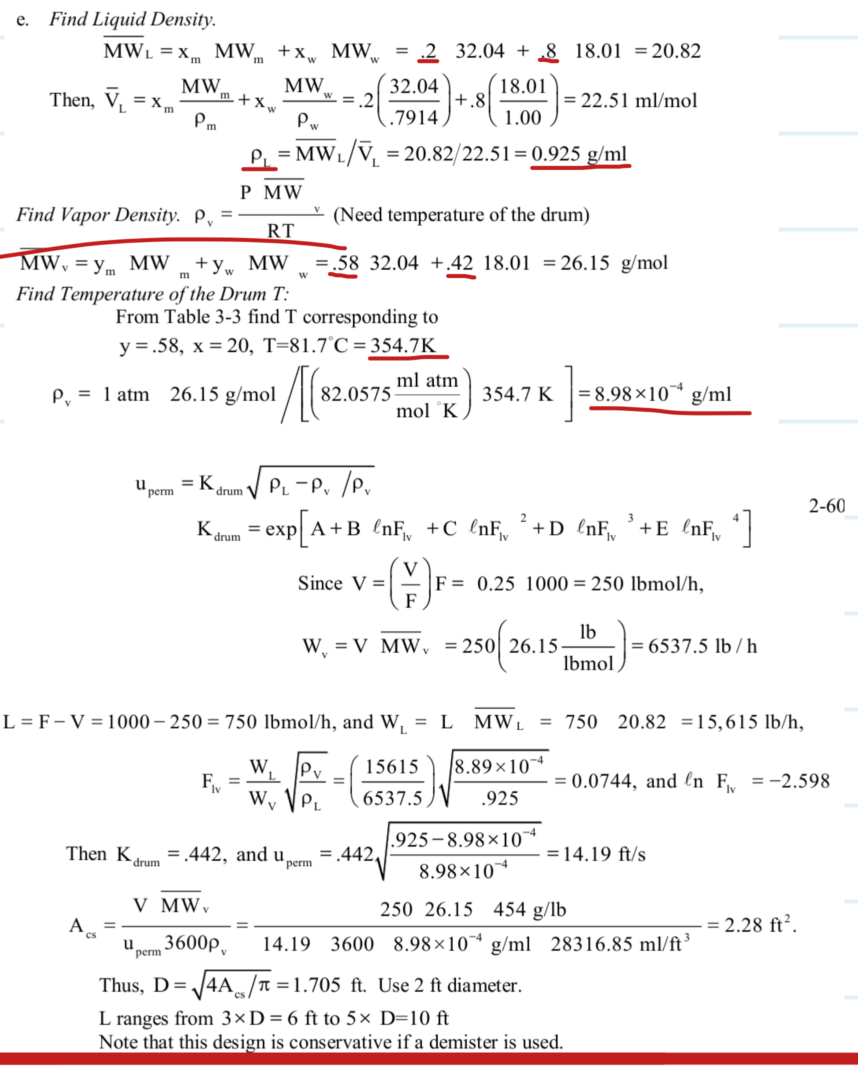
xm

c. If the feed is 30 mol% methanol and we desire a liquid product that is 20 mol% methanol, what V/F must be used? For a feed rate of 1000 lbmol/h, find product flow rates and compositions.

d. We are operating the flash drum so that the liquid mole fraction is 45 mol% methanol. L = 1500 kmol/h, 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 Problem 2.D1c.

Data: $\rho_w = 1.00$ g/cm^3, $\rho_{m,L} = 0.7914$ g/cm^3, $MW_w = 18.01$, $MW_m = 32.04$. Assume vapors are ideal gas.



e. *Find Liquid Density.*

$$\overline{MW}_L = x_m\ MW_m\ + x_w\ MW_w\ = .2\ 32.04\ + .8\ 18.01\ = 20.82$$

$$\text{Then, } \overline{V}_L = x_m \frac{MW_m}{\rho_m} + x_w \frac{MW_w}{\rho_w} = .2\left(\frac{32.04}{.7914}\right) + .8\left(\frac{18.01}{1.00}\right) = 22.51 \text{ ml/mol}$$

$$\rho_L = \overline{MW}_L / \overline{V}_L = 20.82/22.51 = 0.925 \text{ g/ml}$$

Find Vapor Density. $\rho_v = \dfrac{P\ \overline{MW}_v}{RT}$ (Need temperature of the drum)

$$\overline{MW}_v = y_m\ MW_m + y_w\ MW_w = .58\ 32.04\ + .42\ 18.01\ = 26.15 \text{ g/mol}$$

from part c

Find Temperature of the Drum T:

From Table 3-3 find T corresponding to

$y = .58,\ x = 20,\ T = 81.7^\circ C = 354.7K$

$$\rho_v = 1 \text{ atm}\quad 26.15 \text{ g/mol} \Big/ \left[\left(82.0575 \frac{\text{ml atm}}{\text{mol }^\circ K}\right) 354.7 \text{ K}\right] = 8.98\times10^{-4} \text{ g/ml}$$

$$u_{perm} = K_{drum}\sqrt{\rho_L - \rho_v\ /\rho_v}$$

2-60

$$K_{drum} = \exp\left[A + B\ \ell n F_{lv}\ + C\ \ell n F_{lv}\ ^2 + D\ \ell n F_{lv}\ ^3 + E\ \ell n F_{lv}\ ^4\right]$$

$$\text{Since } V = \left(\frac{V}{F}\right) F = 0.25\ 1000 = 250 \text{ lbmol/h},$$

$$W_v = V\ \overline{MW}_v\ = 250\left(26.15 \frac{\text{lb}}{\text{lbmol}}\right) = 6537.5 \text{ lb / h}$$

$$L = F - V = 1000 - 250 = 750 \text{ lbmol/h, and } W_L = L\ \overline{MW}_L\ = 750\ 20.82\ = 15{,}615 \text{ lb/h},$$

$$F_{lv} = \frac{W_L}{W_V}\sqrt{\frac{\rho_V}{\rho_L}} = \left(\frac{15615}{6537.5}\right)\sqrt{\frac{8.89\times10^{-4}}{.925}} = 0.0744, \text{ and } \ell n\ F_{lv}\ = -2.598$$

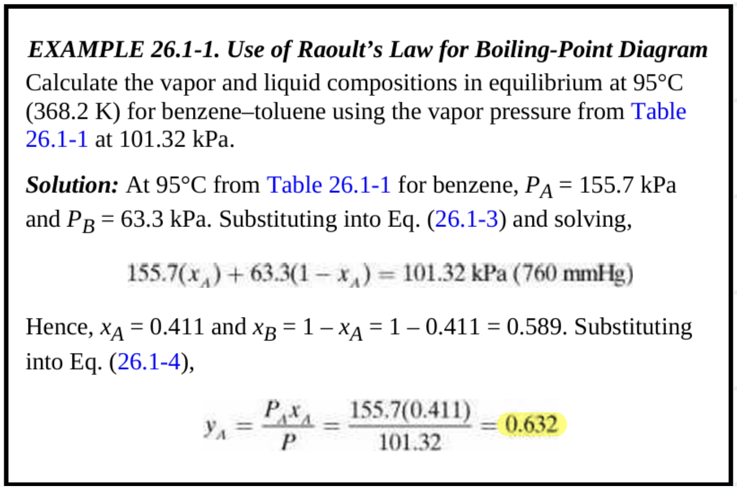
$$\text{Then } K_{drum} = .442, \text{ and } u_{perm} = .442\sqrt{\frac{.925 - 8.98\times10^{-4}}{8.98\times10^{-4}}} = 14.19 \text{ ft/s}$$

$$A_{cs} = \frac{V\ \overline{MW}_v}{u_{perm} 3600 \rho_v} = \frac{250\ 26.15\ \ 454 \text{ g/lb}}{14.19\ \ 3600\ \ 8.98\times10^{-4} \text{ g/ml}\ \ 28316.85 \text{ ml/ft}^3} = 2.28 \text{ ft}^2.$$

$\text{Thus, } D = \sqrt{4A_{cs}/\pi} = 1.705$ ft. Use 2 ft diameter.

L ranges from $3\times D = 6$ ft to $5\times\ D = 10$ ft

Note that this design is conservative if a demister is used.



EXAMPLE 26.1-1. Use of Raoult's Law for Boiling-Point Diagram

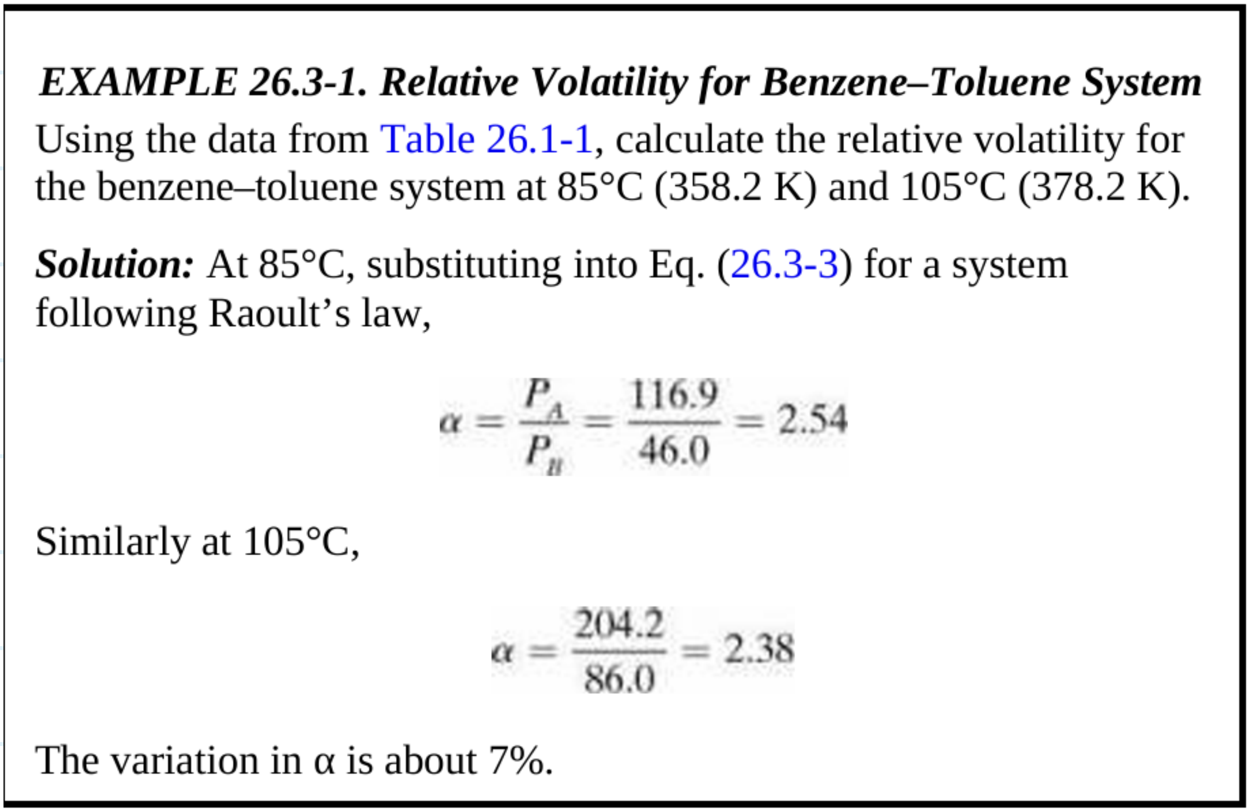
Calculate the vapor and liquid compositions in equilibrium at 95°C (368.2 K) for benzene–toluene using the vapor pressure from Table 26.1-1 at 101.32 kPa.

Solution: At 95°C from Table 26.1-1 for benzene, $P_A = 155.7$ kPa and $P_B = 63.3$ kPa. Substituting into Eq. (26.1-3) and solving,

$$155.7(x_A) + 63.3(1 - x_A) = 101.32 \text{ kPa } (760 \text{ mmHg})$$

Hence, $x_A = 0.411$ and $x_B = 1 - x_A = 1 - 0.411 = 0.589$. Substituting into Eq. (26.1-4),

$$y_A = \frac{P_A x_A}{P} = \frac{155.7(0.411)}{101.32} = 0.632$$



EXAMPLE 26.3-1. *Relative Volatility for Benzene–Toluene System*

Using the data from Table 26.1-1, calculate the relative volatility for the benzene–toluene system at 85°C (358.2 K) and 105°C (378.2 K).

Solution: At 85°C, substituting into Eq. (26.3-3) for a system following Raoult's law,

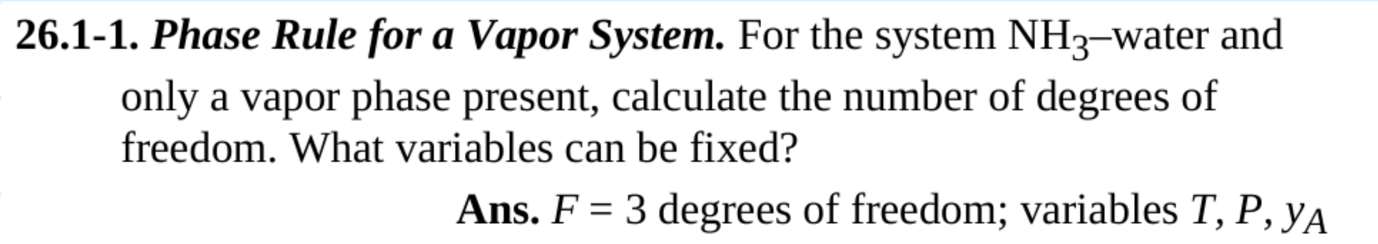
$$\alpha = \frac{P_A}{P_B} = \frac{116.9}{46.0} = 2.54$$

Similarly at 105°C,

$$\alpha = \frac{204.2}{86.0} = 2.38$$

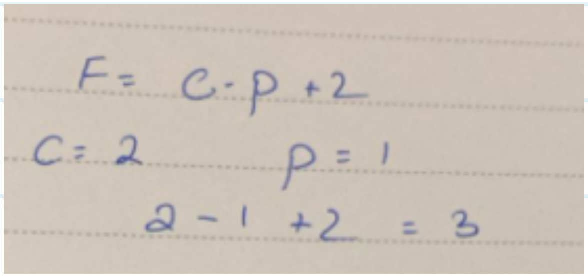
The variation in α is about 7%.

Just reminder that the relative volatility is the vapor pressure of A/ vapor pressure of B + its (ya/xa)/(yb/xb)



26.1-1. *Phase Rule for a Vapor System.* For the system NH_3–water and only a vapor phase present, calculate the number of degrees of freedom. What variables can be fixed?

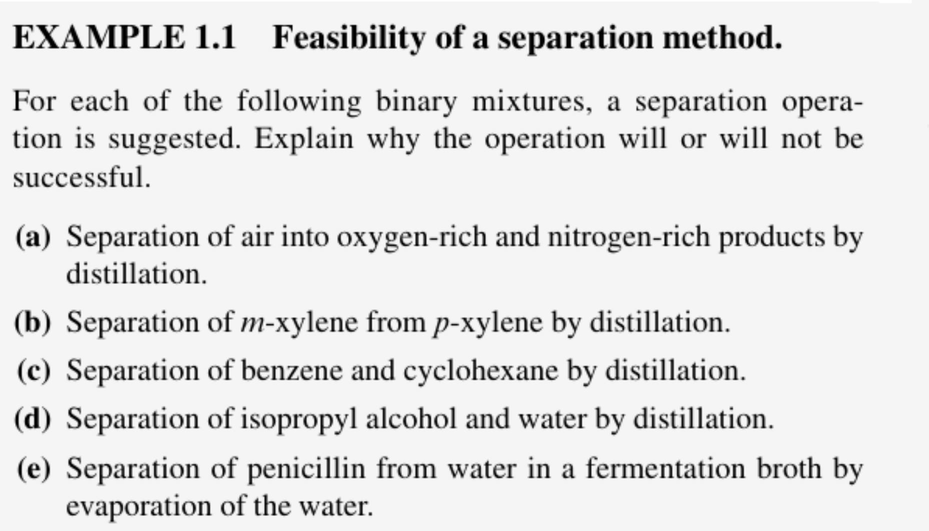
Ans. $F = 3$ degrees of freedom; variables T, P, y_A



$F = C - P + 2$

$C = 2$ $P = 1$

$2 - 1 + 2 = 3$



EXAMPLE 1.1 Feasibility of a separation method.

For each of the following binary mixtures, a separation operation is suggested. Explain why the operation will or will not be successful.

(a) Separation of air into oxygen-rich and nitrogen-rich products by distillation.

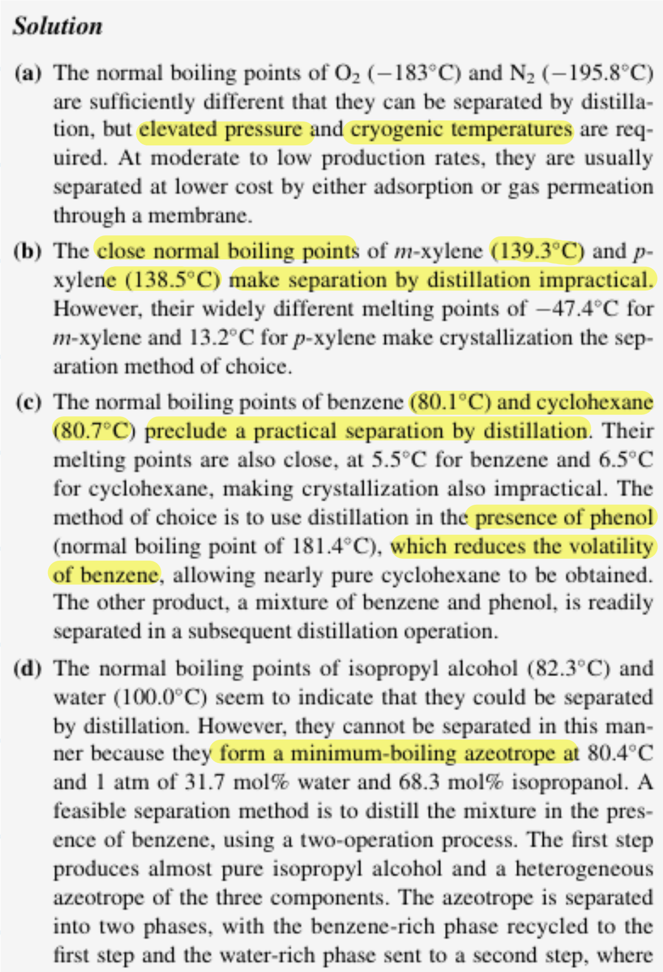
(b) Separation of *m*-xylene from *p*-xylene by distillation.

(c) Separation of benzene and cyclohexane by distillation.

(d) Separation of isopropyl alcohol and water by distillation.

(e) Separation of penicillin from water in a fermentation broth by evaporation of the water.

★ general senarios why we can't separate sometimes



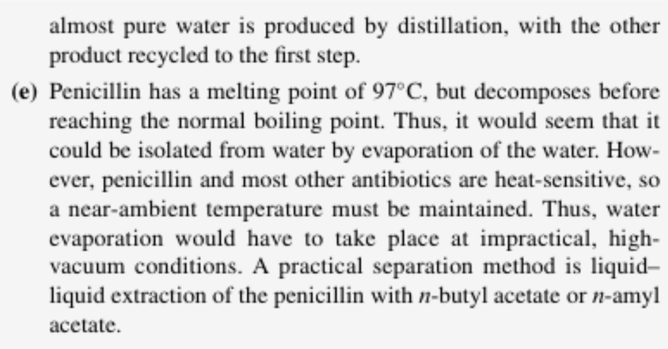
Solution

(a) The normal boiling points of O_2 (−183°C) and N_2 (−195.8°C) are sufficiently different that they can be separated by distillation, but elevated pressure and cryogenic temperatures are required. At moderate to low production rates, they are usually separated at lower cost by either adsorption or gas permeation through a membrane.

(b) The close normal boiling points of *m*-xylene (139.3°C) and *p*-xylene (138.5°C) make separation by distillation impractical. However, their widely different melting points of −47.4°C for *m*-xylene and 13.2°C for *p*-xylene make crystallization the separation method of choice.

(c) The normal boiling points of benzene (80.1°C) and cyclohexane (80.7°C) preclude a practical separation by distillation. Their melting points are also close, at 5.5°C for benzene and 6.5°C for cyclohexane, making crystallization also impractical. The method of choice is to use distillation in the presence of phenol (normal boiling point of 181.4°C), which reduces the volatility of benzene, allowing nearly pure cyclohexane to be obtained. The other product, a mixture of benzene and phenol, is readily separated in a subsequent distillation operation.

(d) The normal boiling points of isopropyl alcohol (82.3°C) and water (100.0°C) seem to indicate that they could be separated by distillation. However, they cannot be separated in this manner because they form a minimum-boiling azeotrope at 80.4°C and 1 atm of 31.7 mol% water and 68.3 mol% isopropanol. A feasible separation method is to distill the mixture in the presence of benzene, using a two-operation process. The first step produces almost pure isopropyl alcohol and a heterogeneous azeotrope of the three components. The azeotrope is separated into two phases, with the benzene-rich phase recycled to the first step and the water-rich phase sent to a second step, where

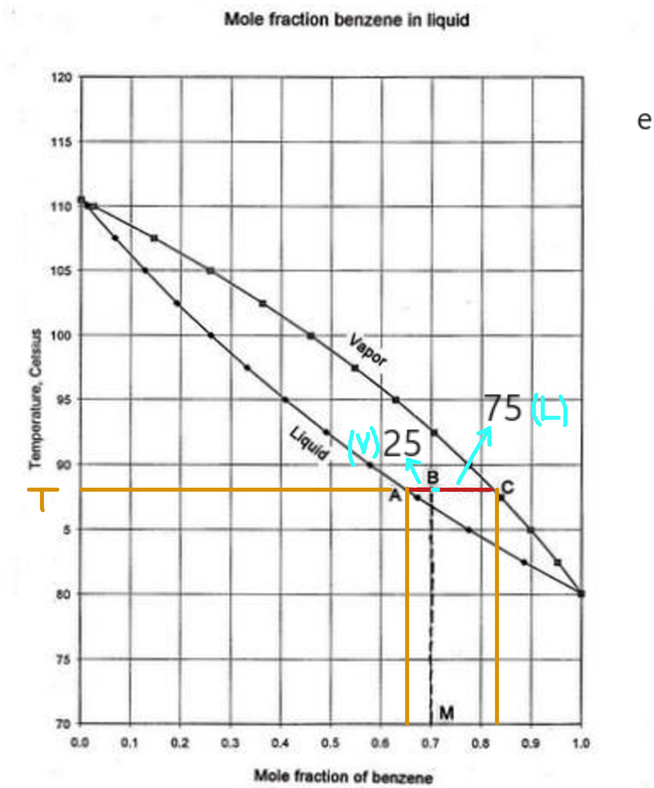


almost pure water is produced by distillation, with the other product recycled to the first step.

(e) Penicillin has a melting point of 97°C, but decomposes before reaching the normal boiling point. Thus, it would seem that it could be isolated from water by evaporation of the water. However, penicillin and most other antibiotics are heat-sensitive, so a near-ambient temperature must be maintained. Thus, water evaporation would have to take place at impractical, high-vacuum conditions. A practical separation method is liquid–liquid extraction of the penicillin with *n*-butyl acetate or *n*-amyl acetate.

4.8. Equilibrium plots for benzene–toluene.

The relative volatility, α, of benzene to toluene at 1 atm is 2.5. ~~Construct x–y and T–x–y diagrams for this system at 1 atm. Repeat the construction of the x–y diagram using vapor pressure data for benzene from Exercise 4.6 and for toluene from the table below, with Raoult's and Dalton's laws.~~ Use the diagrams for the following: (a) A liquid containing 70 mol% benzene and 30 mol% toluene is heated in a container at 1 atm until 25 mol% of the original liquid is evaporated. Determine the temperature. The phases are then separated mechanically, and the vapors condensed. Determine the composition of the condensed vapor and the liquid residue. ~~(b) Calculate and plot the K-values as a function of temperature at 1 atm.~~



temperature is 88C

equilibrium vapor (0.85), equilibrium liquid (0.65) composition

4.10. Continuous, single-stage distillation.

Saturated-liquid feed of $F = 40$ mol/h, containing 50 mol% A and B, is supplied to the apparatus in Figure 4.35. The condensate is split so that reflux/condensate = 1.

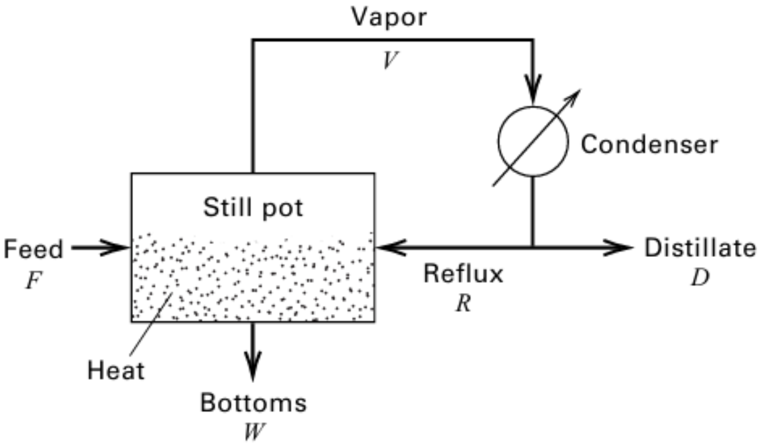
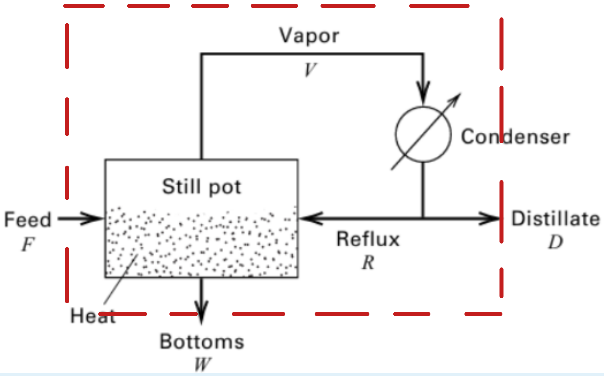


Figure 4.35 Conditions for Exercise 4.10.

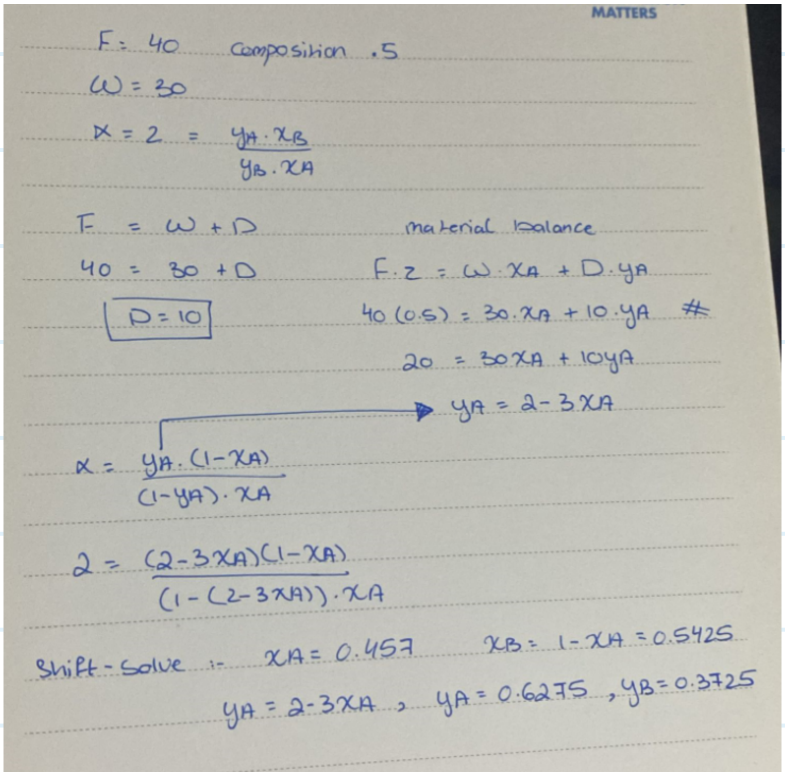
(a) If heat is supplied such that $W = 30$ mol/h and $\alpha = 2$, as defined below, what will be the composition of the overhead and the bottoms product?

$$\alpha = \frac{P_A^s}{P_B^s} = \frac{y_A x_B}{y_B x_A}$$



overall balance

F = W + D



$F = 40$ composition $.5$

$W = 30$

$\alpha = 2 = \frac{y_A \cdot x_B}{y_B \cdot x_A}$

$F = W + D$ material balance

$40 = 30 + D$ $F \cdot z = W \cdot x_A + D \cdot y_A$

$\boxed{D = 10}$ $40(0.5) = 30 \cdot x_A + 10 \cdot y_A$ #

$20 = 30x_A + 10y_A$

$\rightarrow y_A = 2 - 3x_A$

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

$2 = \frac{(2 - 3x_A)(1 - x_A)}{(1 - (2 - 3x_A)) \cdot x_A}$

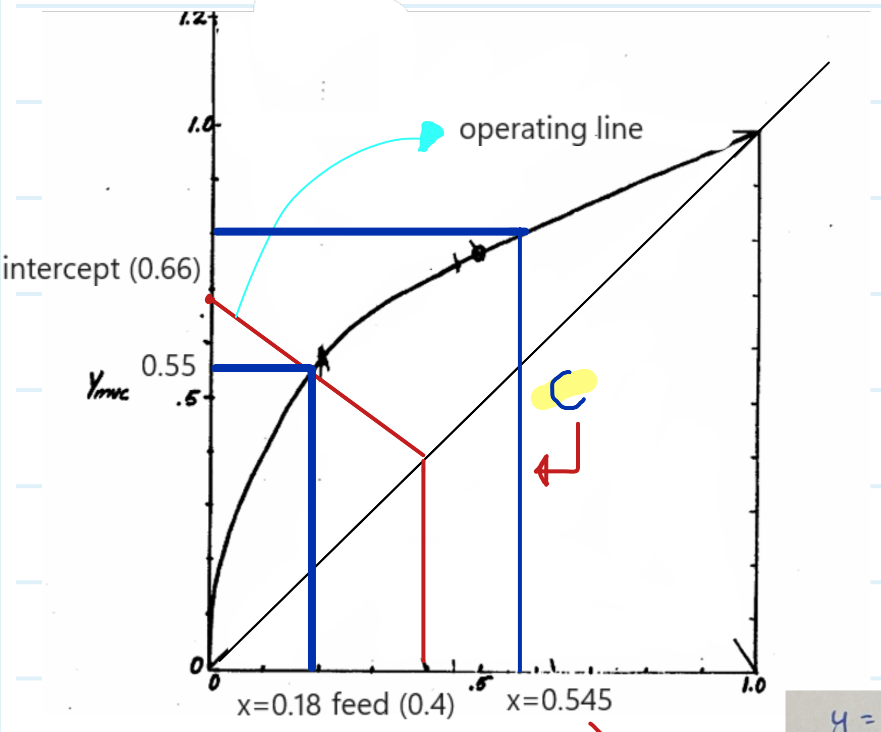
Shift - Solve :- $x_A = 0.457$ $x_B = 1 - x_A = 0.5425$

$y_A = 2 - 3x_A$, $y_A = 0.6275$, $y_B = 0.3725$

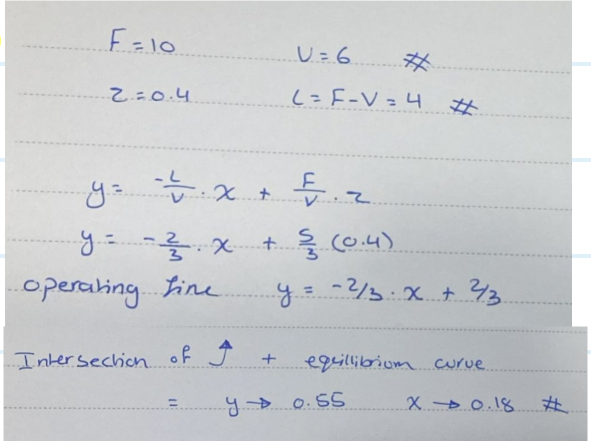
D3. We are separating a mixture of methanol and water in a flash drum at 1.0 atm pressure. Use the equilibrium data listed in Table 2-7. Feed rate is 10.0 kmol/h.

a. The feed is 40 mol% methanol and 60% of the feed is vaporized. Find mole fractions and flow rates of vapor and liquid products.

c. We desire a vapor product that is 80 mol% methanol and want to operate at V/F = 0.3. What must the feed composition be?



a-



$F = 10$ $V = 6$ #

$z = 0.4$ $L = F - V = 4$ #

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

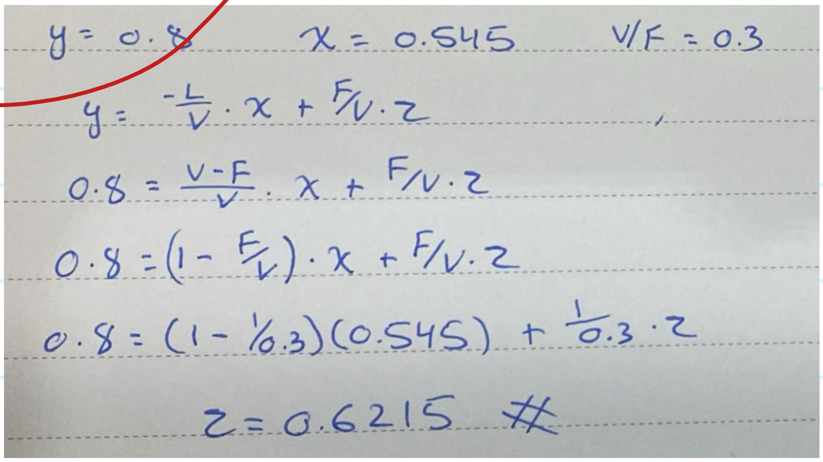
$y = -\frac{2}{3} \cdot x + \frac{5}{3} (0.4)$

operating line $y = -2/3 \cdot x + 2/3$

Intersection of ↑ + equillibrium curve

= $y \to 0.55$ $x \to 0.18$ #

c- we need x from the graph



$y = 0.8$ $x = 0.545$ $V/F = 0.3$

$y = \frac{-L}{V} \cdot x + F/V \cdot z$

$0.8 = \frac{V-F}{V} \cdot x + F/V \cdot z$

$0.8 = (1 - \frac{F}{V}) \cdot x + F/V \cdot z$

$0.8 = (1 - 1/0.3)(0.545) + \frac{1}{0.3} \cdot z$

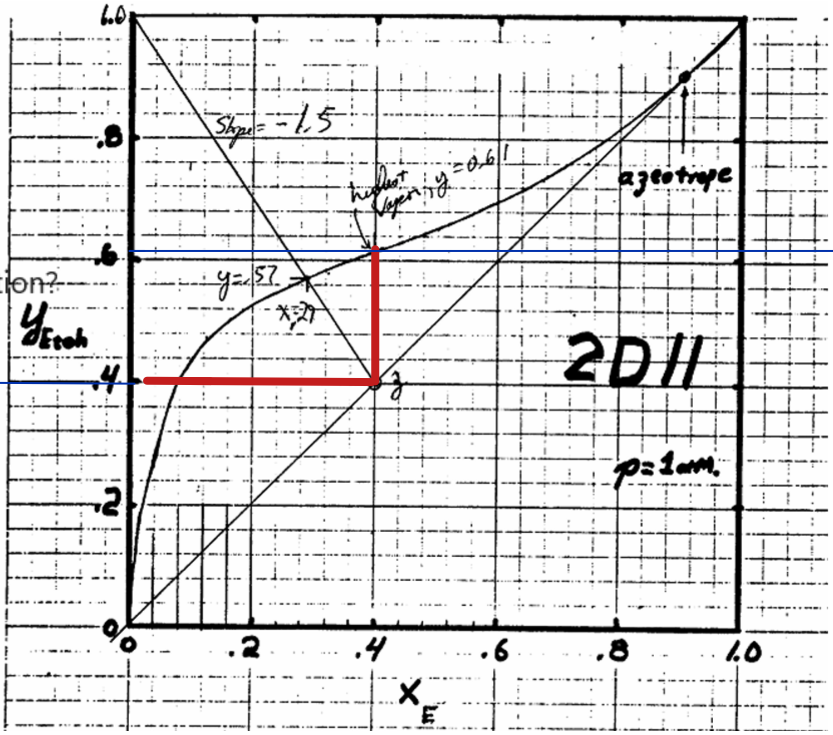
$z = 0.6215$ #

D11. A vapor stream which is 40.0 mol% ethanol and 60.0 mol% water is partially condensed and sent to a flash drum operating at 1.0 atm.

a. What is the highest vapor mole fraction which can be produced?

$$L/F = 0.6 \Rightarrow V/F = 0.4 \ \& \ L/V = 1.5$$

$$\text{Operating line: Slope } = -1.5, \text{ through } y = x = z = 0.4$$



lowest vapor mole fraction?

y min (0.4)

a-

y= (0.61) maximum

D12. Find the dimensions (h_{total} and D) for a horizontal flash drum for Problem 2.D1c. Use $h_{total}/D = 4$.

-Problem in page 4

$$K_{drum,horizontal} = 1.25 \times K_{drum,vertical}$$

نفس الخطوات لحد K drum بضرب ال K drum ال vertical ب 1.25

و بكمل نفس الخطوات

$$K_{drum,vertical} = 0.442, \text{ and } K_{drum,horiz} = 0.5525$$

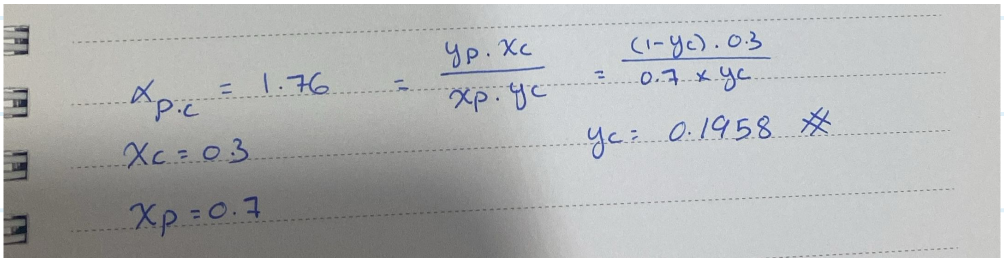
$$u_{perm} = 0.5525\sqrt{\frac{0.925 - 8.98\times10^{-4}}{8.98\times10^{-4}}} = 17.74 \text{ ft/s}$$

$$A_{cs} = \frac{V\ \overline{MW}_v}{u_{perm}\ 3600\rho_v} = \frac{250\ \ 26.15\ \ \ 454 \text{ g/lbm}}{17.74\ \ \ 3600\ \ \ 8.98\times10^{-4} \text{ g/ml}\ \ \ 28316.85 \text{ ml/ft}^3}$$

$$A_{Cs} = 1.824 \text{ ft}^2, \quad A_T = A_{Cs}/0.2 = 9.12 \text{ ft}^2$$

With L/D = 4, $D = \sqrt{4A_T/\pi} = 3.41$ ft and $L = 13.6$ ft

D13. The phenol-cresol system has an approximately constant relative volatility at $\alpha_{p-c} = 1.76$. Phenol is more volatile. At equilibrium, if cresol mole fraction in the liquid is 0.3, what is the mole fraction of cresol in the vapor?



$\alpha_{p.c} = 1.76 = \frac{y_p \cdot x_c}{x_p \cdot y_c} = \frac{(1-y_c) \cdot 0.3}{0.7 \times y_c}$

$x_c = 0.3$

$x_p = 0.7$

$y_c = 0.1958$ #

D14. We are feeding 100 kmol/h of a mixture that is 30 mol% n-butane and 70 mol% n-hexane to a flash drum. We operate with V/F = 0.4 and $T_{drum} = 100°C$. Use Raoult's law to estimate K values from vapor pressures. Use Antoine's equation to calculate vapor pressure,

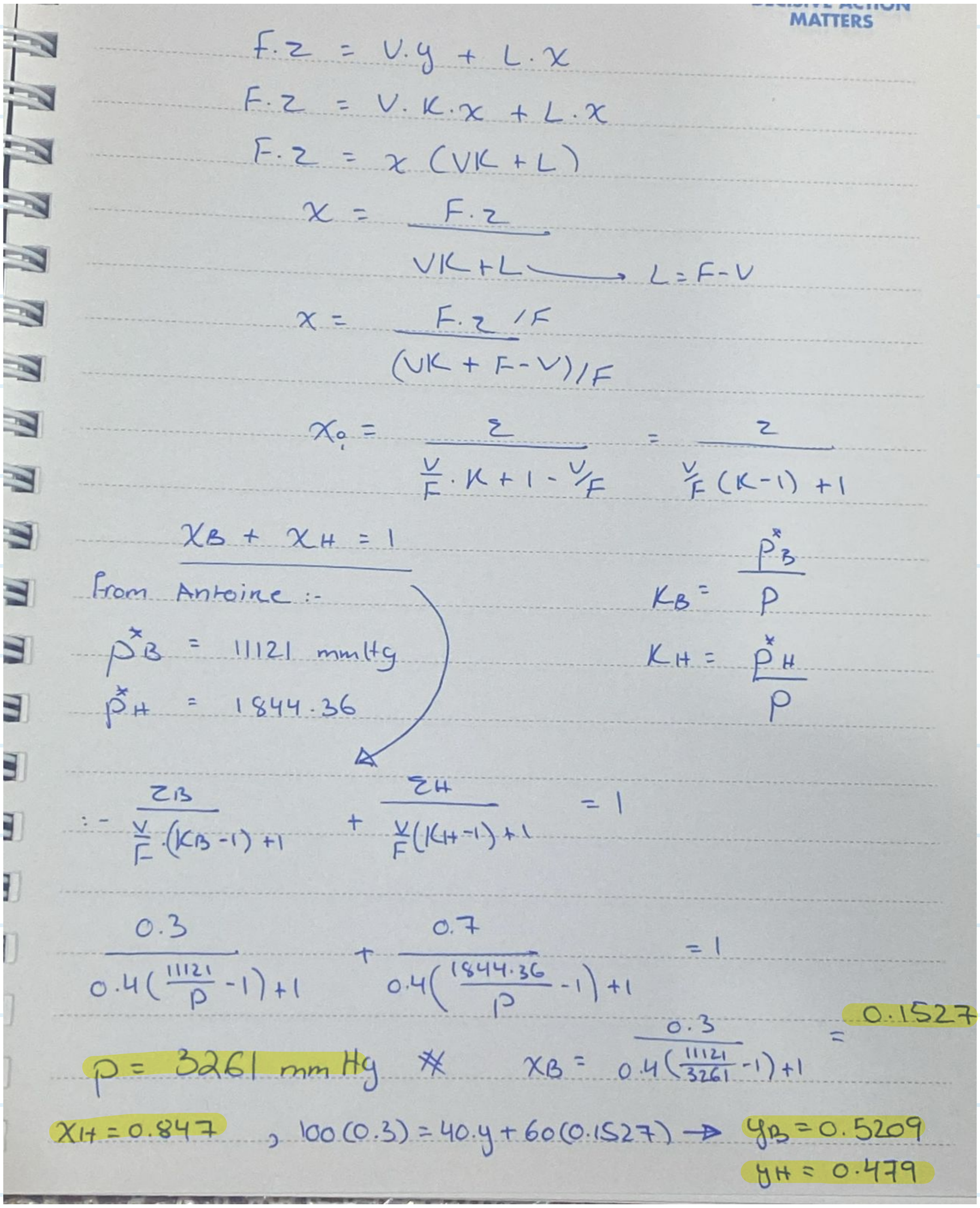
$$\log_{10}(VP) = A - \frac{B}{(T+C)}$$

where VP is in mm Hg and T is in °C.

n-butane: A = 6.809, B = 935.86, C = 238.73

n-hexane: A = 6.876, B = 1171.17, C = 224.41

Find p_{drum}, x_i and y_i



$F.z = V.y + L.x$

$F.z = V.K.x + L.x$

$F.z = x(VK + L)$

$$x = \frac{F.z}{VK + L} \longrightarrow L = F - V$$

$$x = \frac{F.z/F}{(VK + F - V)/F}$$

$$x_i = \frac{z}{\frac{V}{F}.K + 1 - V/F} = \frac{z}{V/F(K-1)+1}$$

$\underline{x_B + x_H = 1}$

From Antoine :-

$p^*_B = 11121$ mmHg

$p^*_H = 1844.36$

$$K_B = \frac{p^*_B}{p}$$

$$K_H = \frac{p^*_H}{p}$$

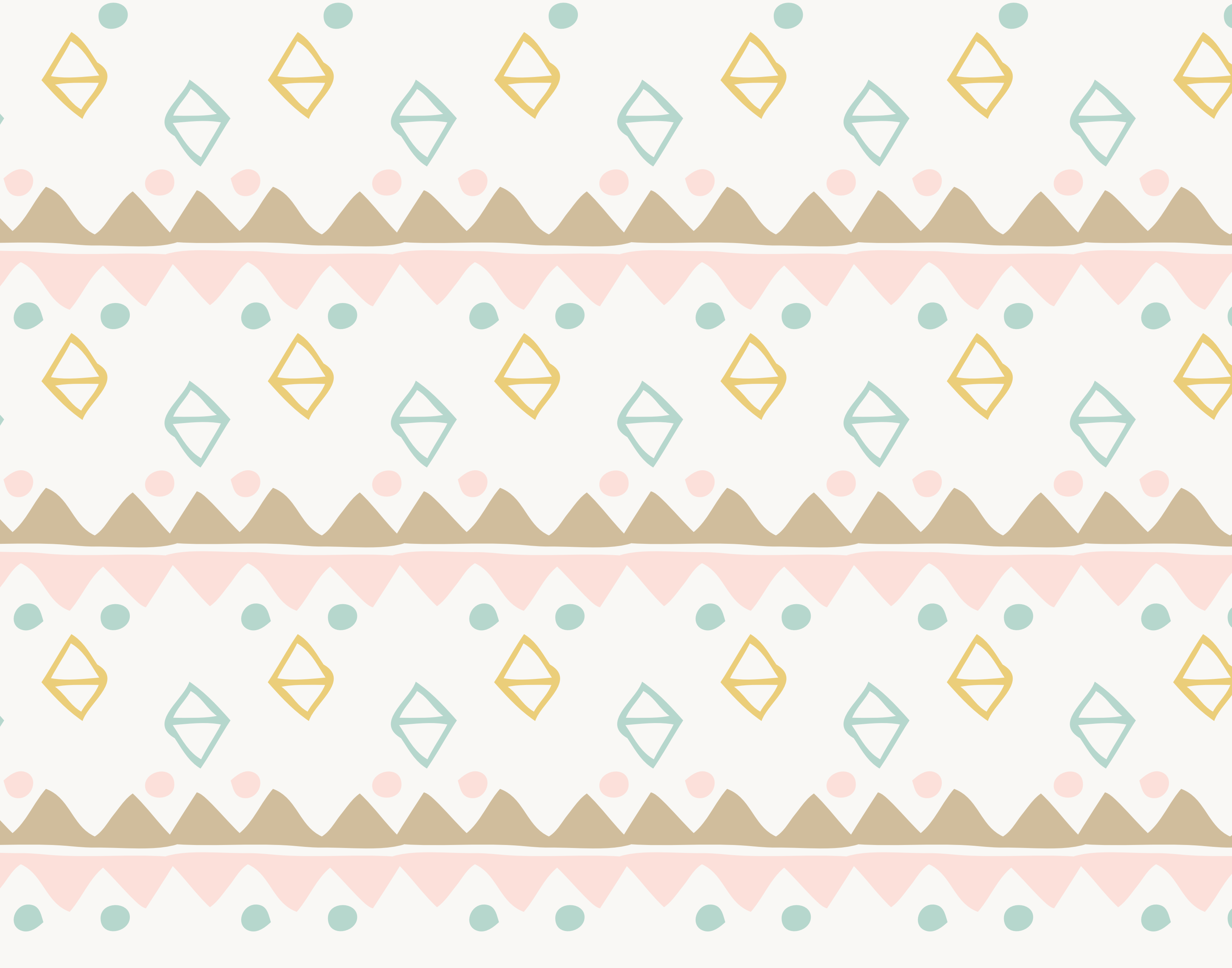
$$:- \frac{z_B}{\frac{V}{F}(K_B - 1) + 1} + \frac{z_H}{\frac{V}{F}(K_H - 1) + 1} = 1$$

$$\frac{0.3}{0.4\left(\frac{11121}{p} - 1\right) + 1} + \frac{0.7}{0.4\left(\frac{1844.36}{p} - 1\right) + 1} = 1$$

$p = 3261$ mmHg ※ $x_B = \frac{0.3}{0.4\left(\frac{11121}{3261} - 1\right) + 1} = 0.1527$

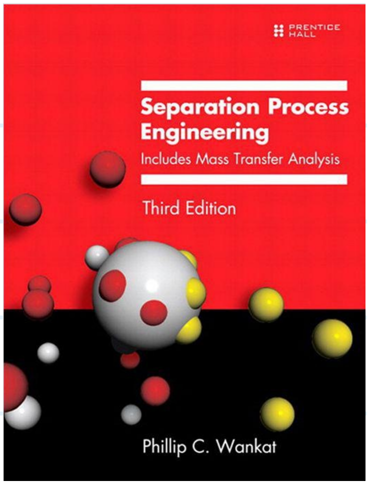
$x_H = 0.847$, $100(0.3) = 40.y + 60(0.1527) \rightarrow y_B = 0.5209$

$y_H = 0.479$



CH4

Continuous Distillation



Key problems

- D2
- D4
- D5
- D7
- D10
- D14
- D15
- D18
- D19
- D21
- D23
- D25
- D26
- D27
- D30

Practice Problems

- D1
- D6
- D8
- D9
- D11
- D13
- D17
- D22
- D24
- D28
- D29
- D32
- D33

D1. A continuous, steady-state distillation column with a total condenser and a partial reboiler is separating methanol from water at one atmosphere (see Table 2-7 for data). The feed rate is 100 kmol/h. The feed is 55 mol% methanol and 45 mol% water. We desire a distillate product that is 90 mol% methanol and a bottoms product that is 5 mol% methanol. Assume CMO.

a. If the external reflux ratio L/D = 1.25 plot the top operating line.

b. If the boilup ratio $\overline{V}/B$ = 2.0 plot the bottom operating line.

c. Step off stages starting at the bottom with the partial reboiler. Use the optimum feed stage. Report the optimum feed stage and the total number of stages.

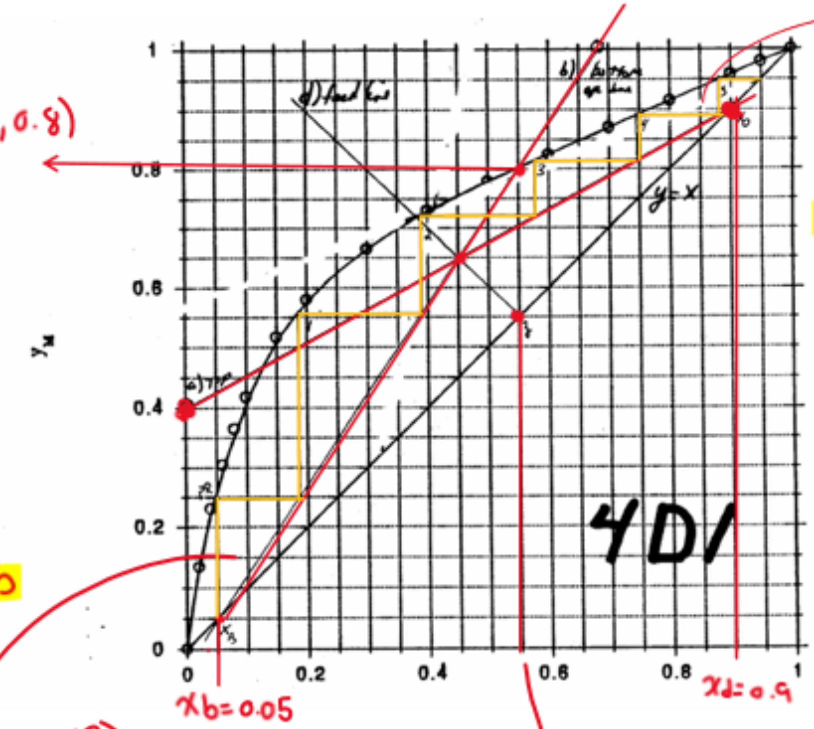
d. Plot the feed line. Calculate its slope. Calculate q. What type of feed is this?

(1)

$y=0.55x+0.4$

-a

(0.55, 0.8)



C- 6 stages

1 Reboiler

5 internal

-b

(2)

$y=1.5x-0.025$

$x=0.55 \rightarrow y=1.5(0.55)-0.025$

$y=0.8$

2nd point → (0.55, 0.8)

$x_{feed}=0.55$

another point for q line

(1) and (2) line intercept

(0.45, 0.65)

$\text{slope}=\frac{q}{q-1}=\frac{y_2-y_1}{x_2-x_1}$

$\frac{q}{q-1}=\frac{0.55-0.65}{0.55-0.45} \rightsquigarrow q=0.5$

mixture

$0<q<1$

-d

D2. We are separating ethanol and water. All percentages are mol%. Find the q values and plot the feed lines.

a. Feed is 60% ethanol and flashes in the column with V/F = 0.37.

b. Feed is 40% ethanol and is a two-phase mixture with liquid and vapor in equilibrium at a temperature of 84.1°C.

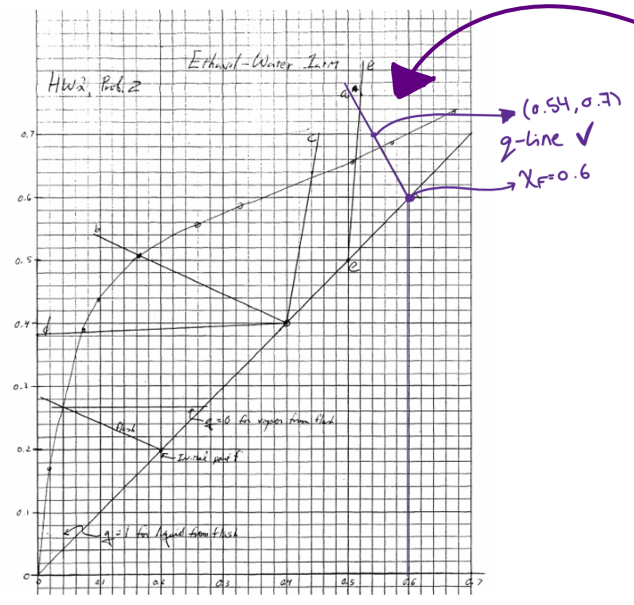
c. Feed is 40% ethanol and is a liquid at 20°C.

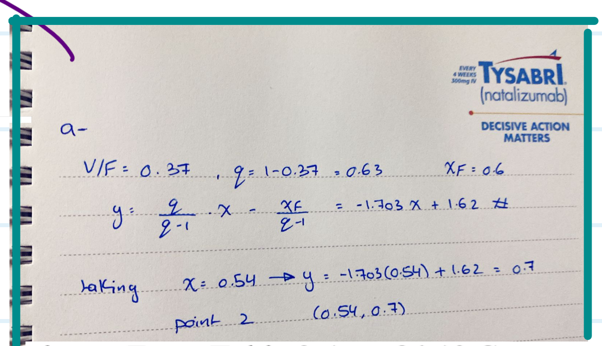
d. Feed is 40% ethanol and is a vapor at 120°C.

نطلع q من قيم Hv , Hl , hf

e. Feed is 50% ethanol and is a subcooled liquid. One mole of vapor must be condensed to heat 12 moles of feed to their boiling point.

f. Feed is 20% ethanol and 70% is vaporized in a flash distillation system. The products of the flash system are fed to distillation column. Calculate the two q values and plot the two feed lines.





a-

$V/F = 0.37$, $q = 1 - 0.37 = 0.63$ $x_F = 0.6$

$y = \frac{q}{q-1} \cdot x - \frac{x_F}{q-1} = -1.703\, x + 1.62$ #

taking $x = 0.54 \rightarrow y = -1.703(0.54) + 1.62 = 0.7$

point 2   (0.54, 0.7)

b. From Table 2-1, at 84.1° C   y = .5089

c. liquid at 20°C $q = \frac{H - h_F}{H - h}$ and 40 mole % ethanol.

The pressure in Figure 2-4 is very close to 1.0 atm, thus it can be used, but must convert to wt frac.

Basis 1kmol feed.

$$.4 \text{ kmole E} = .4 \quad MW = 46 \quad = 18.4 \text{ kg} \qquad 0.63 \text{ wt frac.}$$

$$.6 \text{ kmol Water} = .6 \quad MW = 18 \quad = \frac{10.8 \text{ kg}}{\text{total} = 29.2 \text{ kg}}$$

From Figure 2-4 $H \sim 398 \text{ kcal/kg}, h \sim 75, h_F \;\; 20^\circ C \;\; = 10$

$$q \sim \frac{398 - 10}{398 - 75} = 1.20$$

$$\text{Slope} = \frac{q}{q-1} = \frac{1.2}{+.2} = 6$$

Alternate Solution: 40 mole % ethanol boils at 84.1°C (Table 2-1).
Then if pick reference as saturated liquid at 40 mole %

$$h_F = C_{p,40\%liq} \;\; 20 - 84.1$$

$$h = 0, \quad H = \lambda\big|_{40\%E}$$

d. $40 \text{ mole } \%E = 63 \text{ wt}\%, H \sim 398 \text{ kcal/kg}, \; h \sim 65, \; h_F \sim 398 \frac{\text{kcal}}{\text{kg}} + C_{P_{vapor}} \;\; 120 - 84.1$

$$C_{P_{vapor}} = y_E C_{P_{Evapor}} + y_w C_{P_{w,vapor}}$$

Assume only 1st and 2nd terms in C_p equations are significant.

From Problem 2.D9

$$C_{P_{vapor}} = .4 \;\; 14.66 + 0.03758T \;\; + .6 \;\; 7.88 + .0032T$$

$\uparrow$ kcal/kmol T is °C

which simplifies to $C_{P_{vapor}} = 10.592 + 0.16952T$

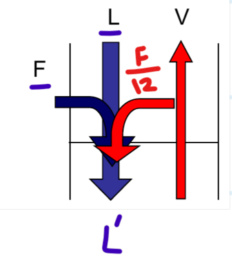
For linear $\int_{84.1}^{120} C_p dT$ is equal to $C_{P_{vapor}}$ @ T_{avg}

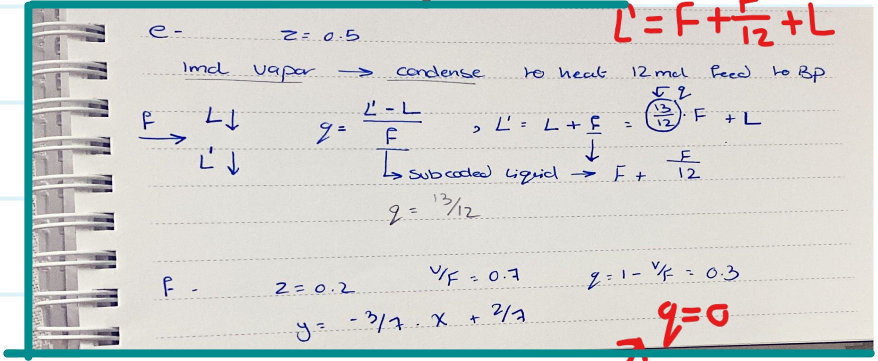
$$T_{avg} = \;\; 84.1 + 120 \;\; /2 = 102.05 \text{. Then } C_{P_{v,avg}} = 10.592 + \;\; 0.16952 \quad 102.05 \;\; = 12.32 \frac{\text{kcal}}{\text{kmol}}$$

$$h_F = 398 \frac{\text{kcal}}{\text{kg}} + 12.32 \;\; 120 - 84.1 \;\; \frac{\text{kcal}}{\text{kmol}} \left| \frac{1 \text{ kmol}}{27.2 \text{ kg}} \right|$$

$$h_F = 398 + 15.149 = 413.15 \text{ kcal/kg}$$

$$q = \frac{398 - 413.15}{398 - 65} = \frac{-15.147}{333} = -0.045.$$





$L' = F + \frac{F}{12} + L$

e- $z = 0.5$

1mol vapor → condense to heat 12 mol feed to BP

F → L↓ L'↓

$q = \frac{L' - L}{F}$, $L' = L + F = \underset{q}{\boxed{\frac{13}{12}}} \cdot F + L$

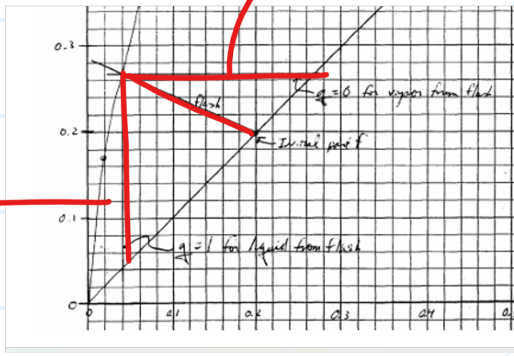
$\hookrightarrow$ subcooled liquid → $F + \frac{F}{12}$

$q = 13/12$

f- $z = 0.2$ $V/F = 0.7$ $q = 1 - V/F = 0.3$

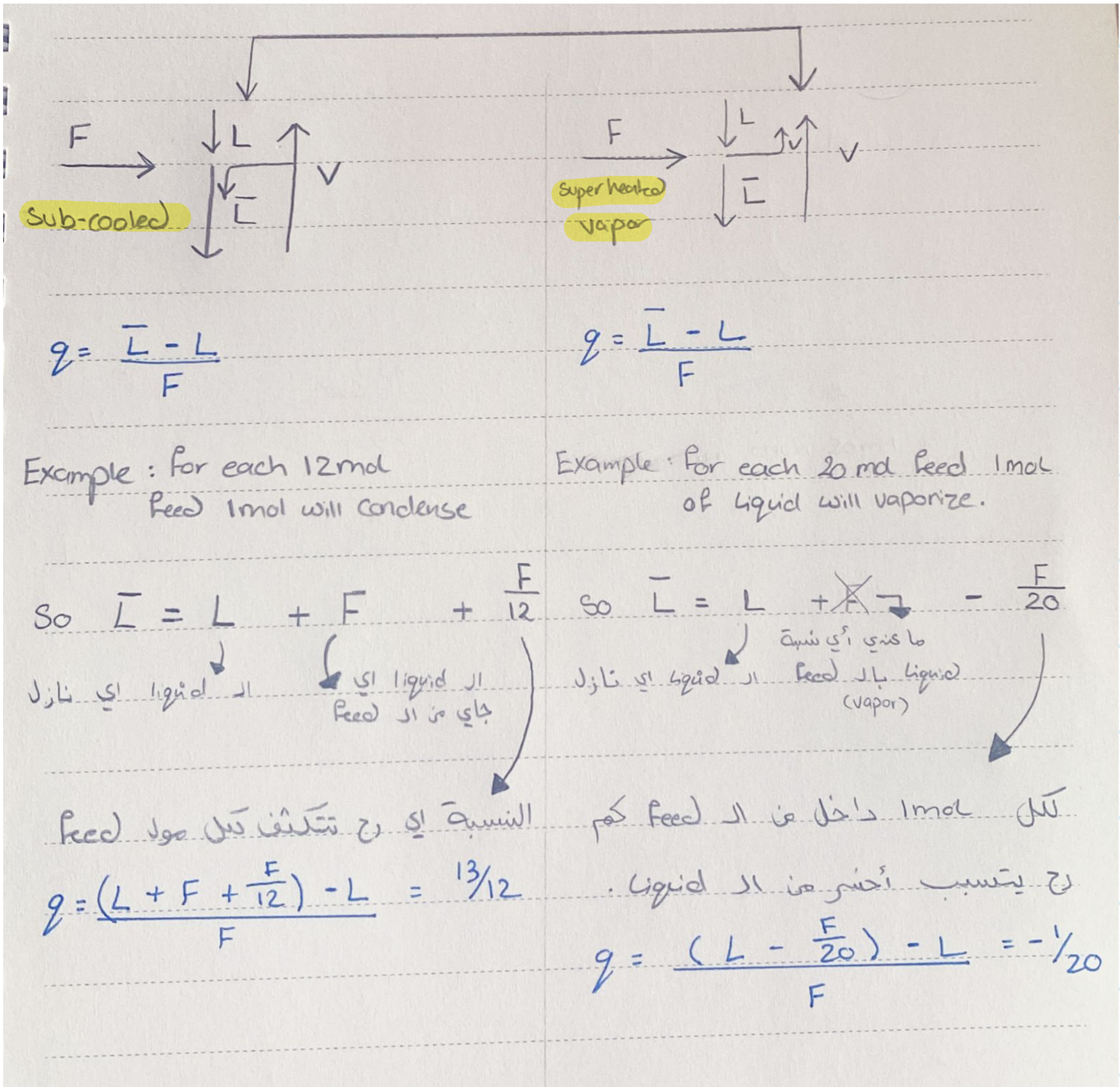
$y = -3/7 \cdot x + 2/7$

$q = 0$



$q = 1$

How to get (q) value when the feed is a subcooled liquid (D2) or Superheated vapor (D26)



$$q = \frac{\bar{L} - L}{F}$$

Example : for each 12mol feed 1mol will condense

$$\text{So} \quad \bar{L} = L + F + \frac{F}{12}$$

الـ liquid اي نازل

ال liquid اي جاي من ال feed

النسبة اي رح تتكثف لكل مول feed

$$q = \frac{\left(L + F + \frac{F}{12}\right) - L}{F} = {}^{13}/_{12}$$

$$q = \frac{\bar{L} - L}{F}$$

Example : for each 20 mol feed 1mol of liquid will vaporize.

$$\text{So} \quad \bar{L} = L + \cancel{F} - \frac{F}{20}$$

ال liquid اي نازل

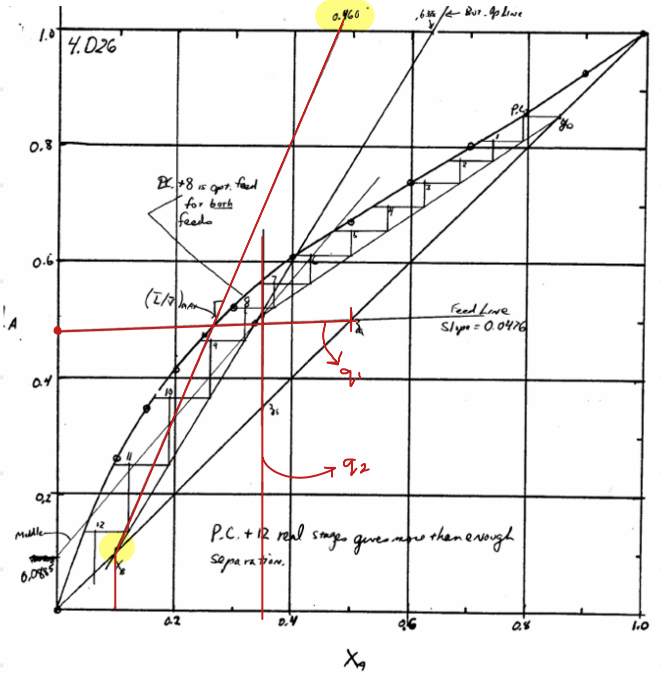
ما عندي أي نسبة liquid بالـ feed (vapor)

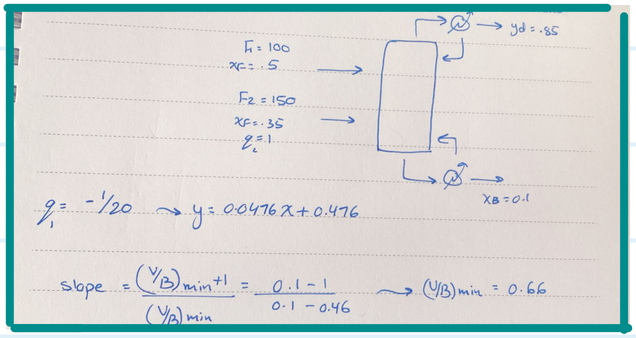
لكل 1mol داخل من ال feed كم رح يتسبب أكثر من ال liquid.

$$q = \frac{\left(L - \frac{F}{20}\right) - L}{F} = -{}^{1}/_{20}$$

D26. A distillation column with a partial condenser and a total reboiler is separating acetone and ethanol. There are two feeds. One feed is 50.0 mol% acetone, flows at 100.0 mol/min, and is a superheated vapor where approximately 1 mole of liquid will vaporize on the feed stage for each 20 moles of feed. The other feed is a saturated liquid, flows at 150.0 mol/min and is 35.0 mol% acetone. We desire a distillate product that is $y_D = 0.85$ mole fraction acetone and a bottoms product that is $x_B = 0.10$ mole fraction acetone. The column has a partial condenser and a total reboiler. Boilup is returned as a saturated vapor. Column operates at a pressure of 1.0 atm. Assume CMO and use a McCabe-Thiele diagram. VLE data are given in Problem 4.D7.

a. Find $(\overline{V}/B)_{min}$. (Plot both feed lines to decide which one to use.)





$F_1 = 100$
$x_F = .5$

$F_2 = 150$
$x_F = .35$
$q_2 = 1$

$y_d = .85$

$x_B = 0.1$

$q_1 = -1/20 \rightarrow y = 0.0476x + 0.476$

$$\text{slope} = \frac{(V/B)_{min}+1}{(V/B)_{min}} = \frac{0.1-1}{0.1-0.46} \rightarrow (V/B)_{min} = 0.66$$

D4.*

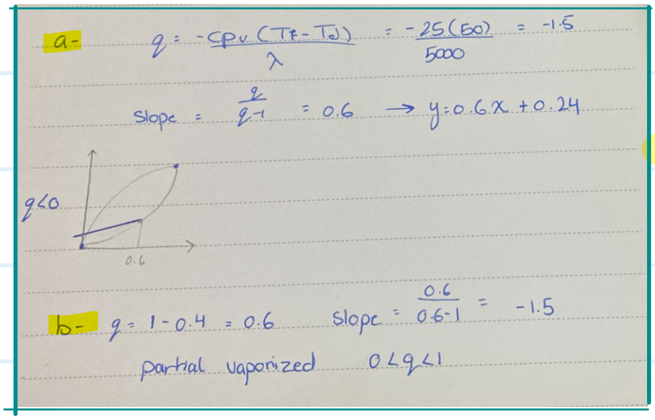
a. We have a superheated vapor feed of 60 mol% more volatile component at 350°C. Feed flow rate is 1000 kmol/h. On the feed plate the temperature is 50°C. For this mixture the average heat capacities are

$C_{PL} = 50$ cal/(mol- °C), $C_{PV} = 25$ cal/(mol- °C)

while the latent heat of vaporization is $\lambda = 5000$ cal/mol. Plot the feed line for this feed.

b. If a feed to a column is a two-phase feed that is 40 mol% vapor, find the value of q and the slope of the feed line.

c. If the feed to a column is a superheated vapor and 1 mole of liquid is vaporized on the feed



a- $q = \frac{-C_{pv}(T_F - T_d)}{\lambda} = \frac{-25(50)}{5000} = -1.5$

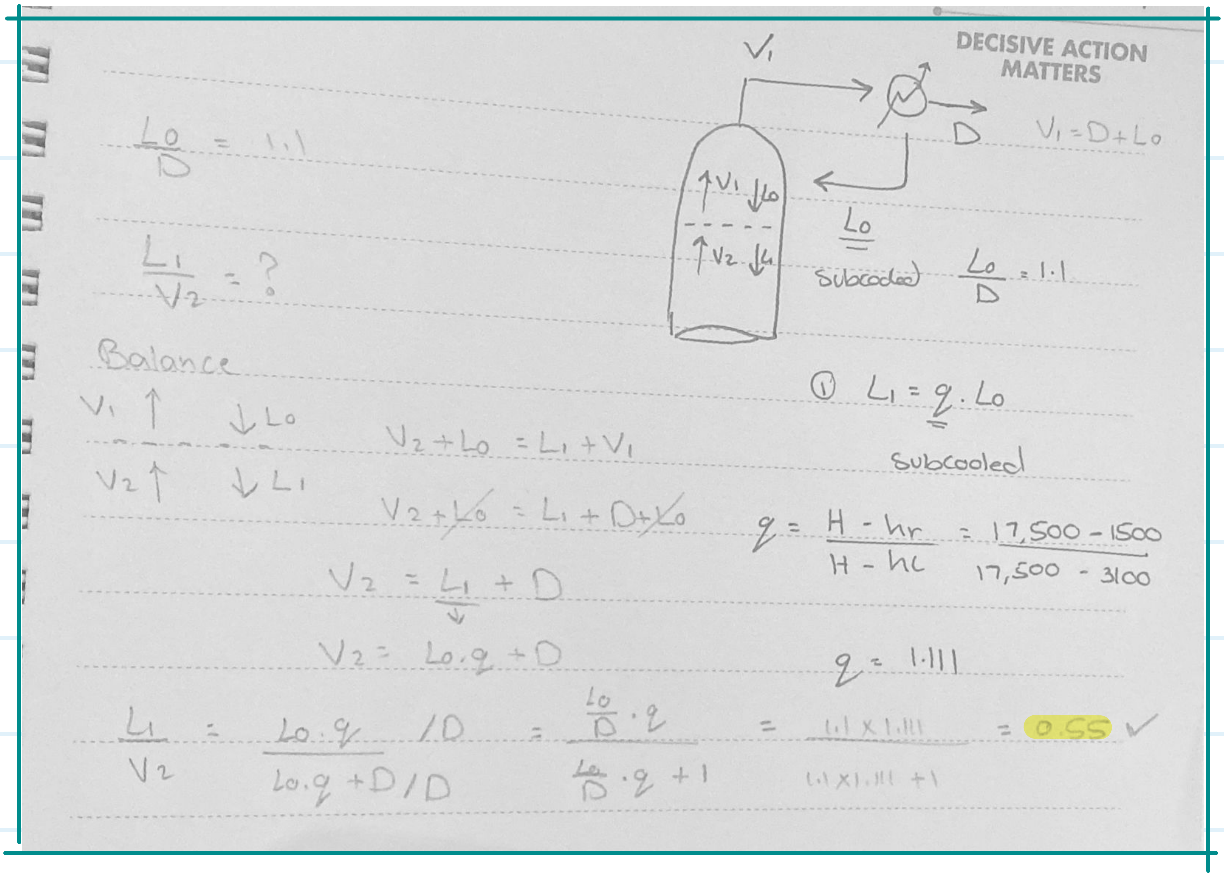
Slope $= \frac{q}{q-1} = 0.6 \rightarrow y = 0.6x + 0.24$

b- $q = 1 - 0.4 = 0.6$ Slope $= \frac{0.6}{0.6-1} = -1.5$

partial vaporized $0 < q < 1$

c. $q = \bar{L} - L \,/F$ where $\bar{L} = L - F/5$. $q = L - F/5 - L \,/F = -1/5$, slope $= q/\ q - 1\ = 1/6$

D5.* A distillation column is operating with a subcooled reflux. The vapor streams have an enthalpy of $H_1 = H_2 = 17{,}500$ Btu/lbmol, while the saturated liquid $h_1 = 3100$ Btu/lbmol. Enthalpy of the reflux stream is $h_0 = 1500$ Btu/lbmol. The external reflux ratio is set at $L_0/D = 1.1$. Calculate the internal reflux ratio inside the column, L_1/V_2.



$\frac{L_0}{D} = 1.1$

$\frac{L_1}{V_2} = ?$

$V_1 = D + L_0$

$\underline{\underline{L_0}}$ subcooled $\frac{L_0}{D} = 1.1$

Balance

$V_1 \uparrow \quad \downarrow L_0$

$V_2 \uparrow \quad \downarrow L_1$

$V_2 + L_0 = L_1 + V_1$

$V_2 + \cancel{L_0} = L_1 + D + \cancel{L_0}$

$V_2 = \underline{L_1} + D$

$V_2 = L_0 . q + D$

① $L_1 = \underline{\underline{q}} . L_0$

subcooled

$q = \frac{H - h_r}{H - h_L} = \frac{17{,}500 - 1500}{17{,}500 - 3100}$

$q = 1.111$

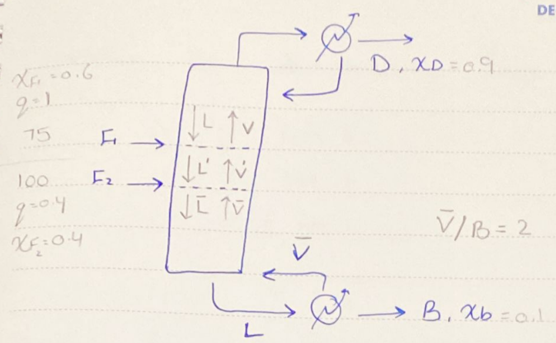
$\frac{L_1}{V_2} = \frac{L_0 . q \ /D}{L_0 . q + D / D} = \frac{\frac{L_0}{D} . q}{\frac{L_0}{D} . q + 1} = \frac{1.1 \times 1.111}{1.1 \times 1.111 + 1} = 0.55$ ✓

D6. We are separating a mixture of acetone and ethanol in a distillation column operating at one atmosphere pressure. The column has a total condenser and a partial reboiler. The distillate is 90 mol% acetone, and the bottoms is 10 mol% acetone. The reflux is returned as a saturated liquid. Use a boilup ratio of $\overline{V}/B = 2.0$. Two feeds are fed to the column. The first feed has a flow rate of 75 kmol/h, it is a saturated liquid, and it is 60 mol% acetone. The second feed has a flow rate of 100 kmol/h, it is a two-phase mixture that is 60% vapor, and it is 40 mol% acetone. Use the optimum feed location for each feed. Assume CMO. Equilibrium data are in Problem 4.D7.

a. Find the distillate and bottoms flow rates, D and B.

b. Plot the two feed lines and the three operating lines.

c. Step off stages and find the optimum feed location for each feed stream and the total number of equilibrium stages required.



① D? B?

$F_1 + F_2 = D + B$

$F_1 x_{F_1} + F_2 x_{F_2} = D x_D + B x_B$

$D = 84.375$, $B = 90.625$ # -a

Plotting 2 feed line : -b

For F_1 $q = 1$ vertical line ✓

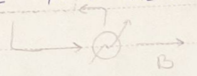
For F_2 $q = 0.4$ slope $= \frac{q}{q-1} = -\frac{2}{3}x + \frac{2}{3}$ ✓

plotting the 3 operating lines :-

$F_1 \rightarrow$ ↓L ↑V Line 3

$F_2 \rightarrow$ ↓L' ↑V' Line 2

↓$\bar{L}$ ↑$\bar{V}$ $\bar{V}$ Line 1



Line 1 : $B = 90.625$ $\bar{V}/B = 2 \rightarrow \bar{V} = 181.25$

Balance around Reboiler $\bar{L} = \bar{V} + B = 271.875$

F_2 L'↓ $\bar{L}$↓ $\rightsquigarrow \overset{\checkmark}{\bar{L}} = \overset{?}{L'} + \overset{\checkmark}{q_2} \cdot \overset{\checkmark}{F_2} = L' = \bar{L} - q_2 \cdot F_2 = 231.875$

F_2 ↑V' ↑$\bar{V}$ $\rightsquigarrow \overset{\checkmark}{\bar{V}} + (1-q) \cdot F_2 = \overset{?}{V'} = 181.25 + 0.6(100) = 241.25$

* Stripping operating line (Line 1) $= y = \frac{V_B+1}{V_B} x - \frac{1}{V_B} \cdot x_B$

① $y = \frac{3}{2}x - 0.05$

Line 2 : $L' + F_2 + \bar{V} = V' + \overset{(\bar{V}+B)}{\bar{L}} \rightsquigarrow L' + F_2 + \cancel{\bar{V}} = V' + \cancel{\bar{V}} + B$

$L' + F_2 = V' + B \rightarrow L' \cdot x + F_2 \cdot x_{F_2} = V' \cdot y + B x_B$

So $y = \frac{L'}{V'} \cdot x + \frac{F_2 \cdot x_{F_2} - B x_b}{V'} \rightarrow 0.128$

$\frac{231.875}{241.25} = 0.961$

② $y = 0.961x + 0.128$

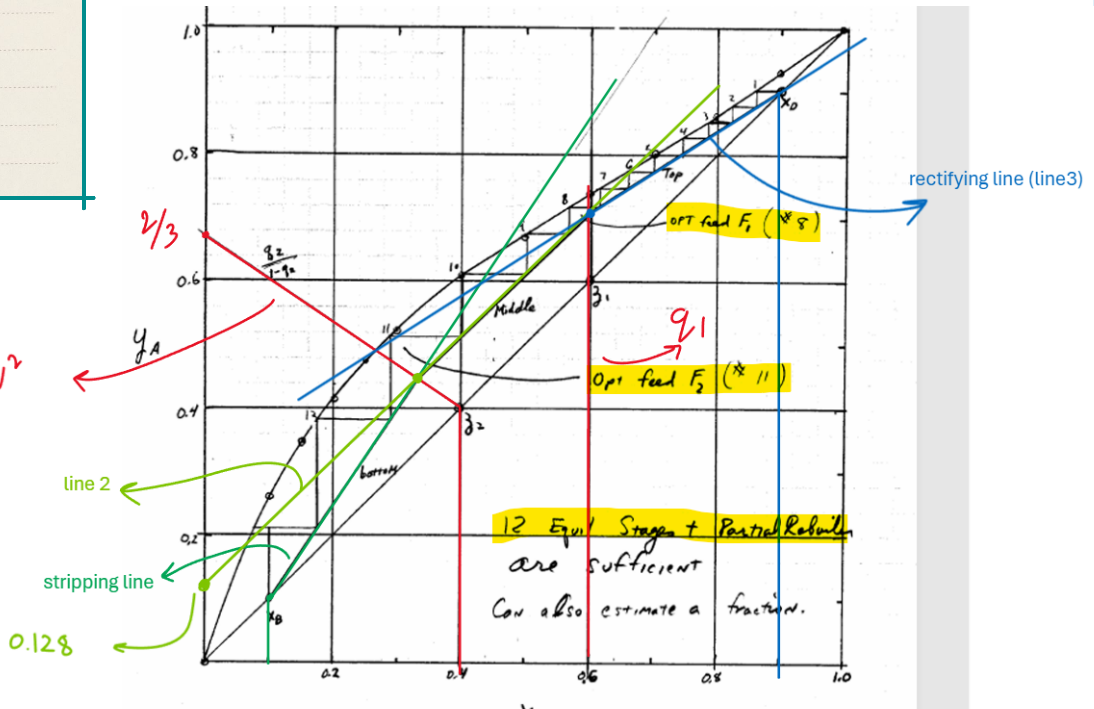
Line 3 : [you can get it by graph] + ↓

F_1 ↓L ↓L' $L' = \overset{(1)}{q} F_1 + L \rightarrow L = 231.875 - 75 = 156.875$

F_1 ↑V ↑V' $q = 1$ so $V' = V = 241.25$

$y = \frac{L}{V} \cdot x_o + \frac{D}{V} \cdot x_d$

③ $y = 0.65 x_o + 0.314$



stage 1 to stage 8	Line 1
stage 8 to stage 11	Line 2
stage 11 to stage 13	Line 3

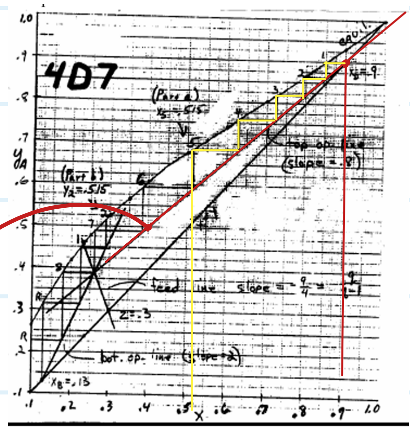
D7.*

a. A distillation column with a total condenser is separating acetone from ethanol. A distillate concentration of $x_D = 0.90$ mole fraction acetone is desired. Since CMO is valid, L/V = constant. If L/V is equal to 0.8, find the composition of the liquid leaving the fifth stage below the total condenser.

b. A distillation column separating acetone and ethanol has a partial reboiler that acts as an equilibrium contact. If the bottoms composition is $x_B = 0.13$ mole fraction acetone and the boilup ratio $\overline{V}/B = 1.0$, find the vapor composition leaving the second stage above the partial reboiler.

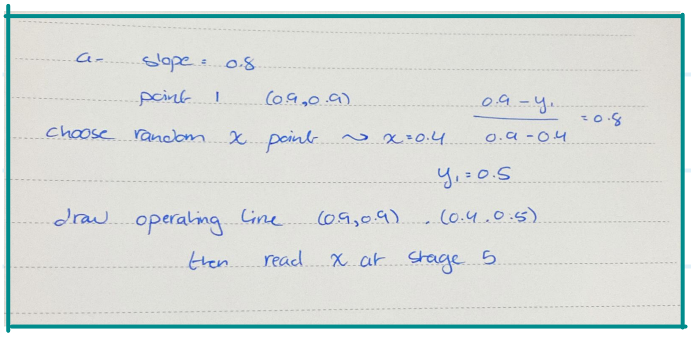
c. The distillation column in parts a and b is separating acetone from ethanol and has $x_D = 0.9$, $x_B = 0.13$, L/V = 0.8, and $\overline{V}/B = 1.0$. If the feed composition is z = 0.3 (all concentrations are mole fraction of more volatile component), find the optimum feed plate location, total number of stages, and required q value of the feed. Equilibrium data for acetone and ethanol at 1 atm (Perry *et al.*, 1963, pp. 13–4) are

x_A	.10	.15	.20	.25	.30	.35	.40	.50	.60	.70	.80	.90
y_A	.262	.348	.417	.478	.524	.566	.605	.674	.739	.802	.865	.929



(0.4,0.5)

a- 0.52



a- slope = 0.8

point 1 (0.9,0.9)

$$\frac{0.9 - y_1}{0.9 - 0.4} = 0.8$$

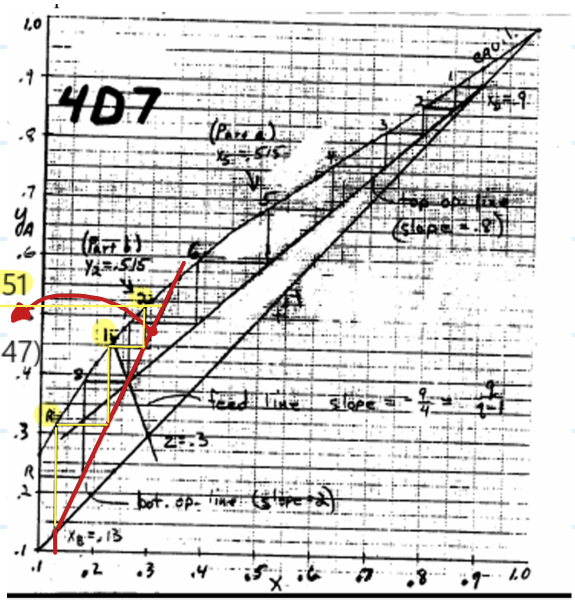
choose random x point ⇝ x = 0.4

$y_1 = 0.5$

draw operating line (0.9,0.9) . (0.4 . 0.5)

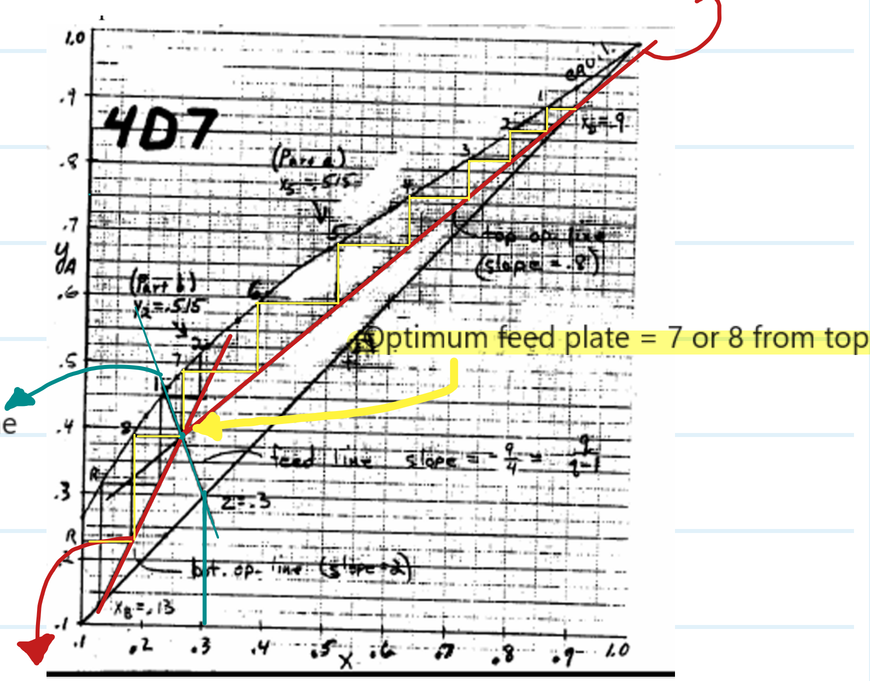
then read x at stage 5

rectifying line from a



b)~ 0.51

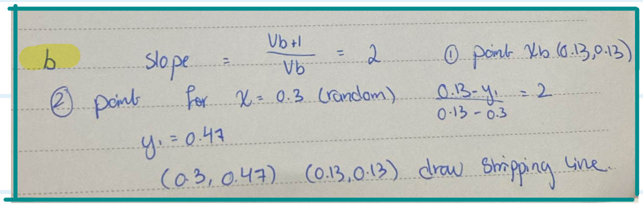
(0.3,0.47)



Optimum feed plate = 7 or 8 from top

q line

stripping line from b

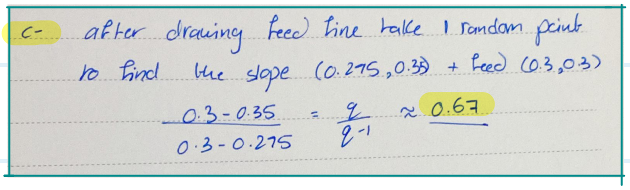


b slope = $\frac{Vb+1}{Vb}$ = 2 ① point xb (0.13,0.13)

② point for x = 0.3 (random) $\frac{0.13 - y_1}{0.13 - 0.3} = 2$

$y_1 = 0.47$

(0.3, 0.47) (0.13,0.13) draw Stripping line.



c- after drawing feed line take 1 random point to find the slope (0.275,0.35) + feed (0.3,0.3)

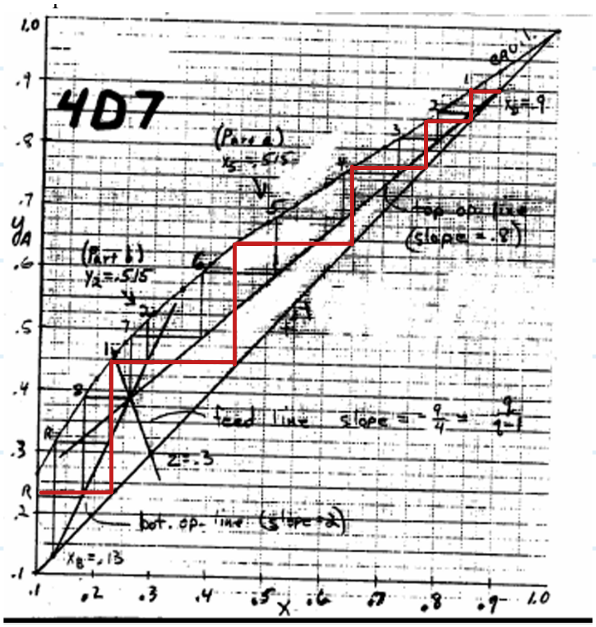
$$\frac{0.3 - 0.35}{0.3 - 0.275} = \frac{q}{q-1} \approx \underline{0.67}$$

D8.* For Problem 4.D7c for separation of acetone from ethanol, determine:

a. How many stages are required at total reflux?

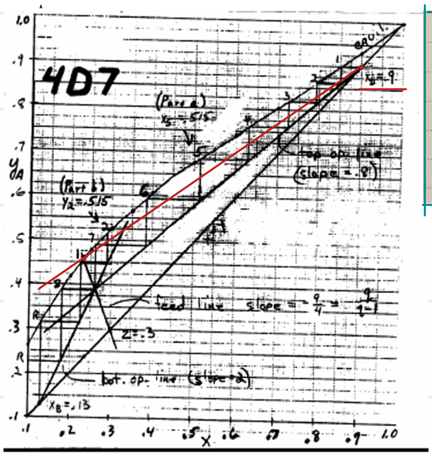
b. What is $(L/V)_{min}$? What is $(L/D)_{min}$?

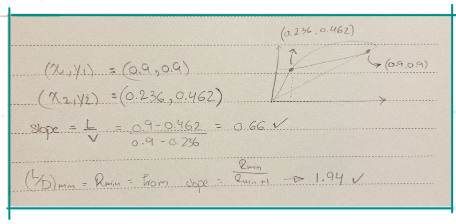
c. The L/D used is how much larger than $(L/D)_{min}$?



a) 6 stages

b)



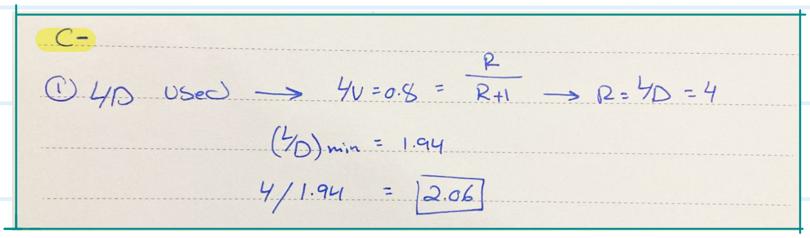


$(x_1, y_1) = (0.9, 0.9)$

$(x_2, y_2) = (0.236, 0.462)$

slope $= \frac{L}{V} = \frac{0.9-0.462}{0.9-0.236} = 0.66$ ✓

$(L/D)_{min} = R_{min}$ = from slope $= \frac{R_{min}}{R_{min}+1} \rightarrow 1.94$ ✓



C-

① L/D used $\longrightarrow$ $L/V = 0.8 = \frac{R}{R+1}$ $\longrightarrow$ $R = L/D = 4$

$(L/D)_{min} = 1.94$

$4/1.94 = \boxed{2.06}$

D9. A distillation column with two feeds is separating ethanol (E) and water at a pressure of 1.0 atm. The column has a total condenser with saturated liquid reflux and a partial reboiler. Feed 1 is a saturated liquid and is 42 mol% ethanol. Feed 2 flow rate is F_2 = 100 kmol/h. Feed 2 is 18 mol% ethanol and is a two-phase mixture that is 30% vapor. The external reflux ratio is L/D = 1/2, and the distillate flow rate is D = 80 kmol/h. We desire a distillate mole fraction of x_D = 0.66 mole fraction ethanol and a bottoms that is x_B = 0.04 mole fraction ethanol. You can assume that CMO is valid. Equilibrium data are in Table 2-1.

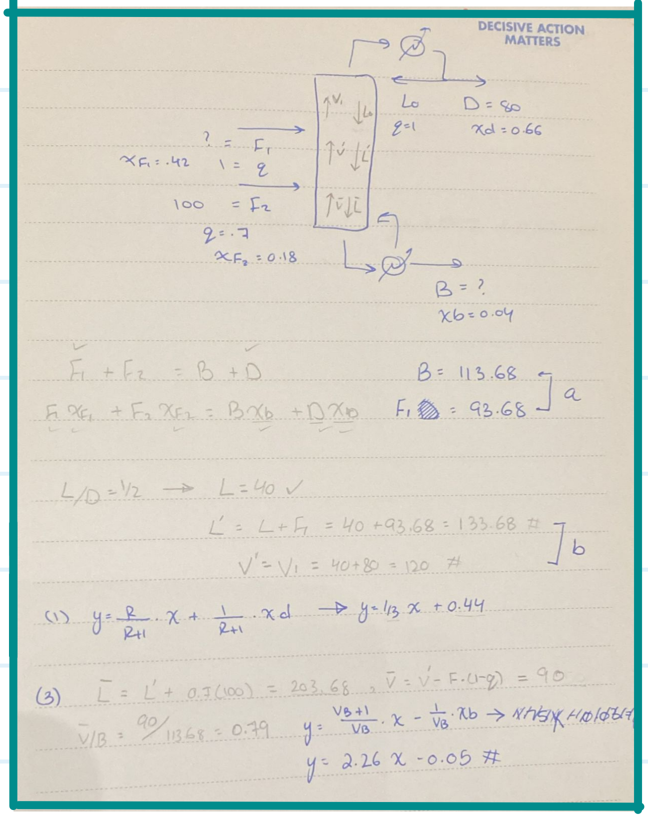
Same as D6

a. Find the flow rates F_1 and B.

b. Find the liquid and vapor flow rates in the middle section, L′ and V′.

c. Determine and plot the operating lines. Be neat.

d. Find both optimum feed locations (above partial reboiler) and the total number of equilibrium stages needed. Step off stages from the bottom up. Be neat.



L_0
$q=1$

$D = 80$
$x_d = 0.66$

$? = F_1$
$x_{F_1} = .42$ $1 = q$

$100 = F_2$
$q = .7$
$x_{F_2} = 0.18$

$B = ?$
$x_b = 0.04$

$F_1 + F_2 = B + D$

$F_1 x_{F_1} + F_2 x_{F_2} = B x_b + D x_D$

$B = 113.68$
$F_1 = 93.68$] a

$L/D = 1/2 \rightarrow L = 40$ ✓

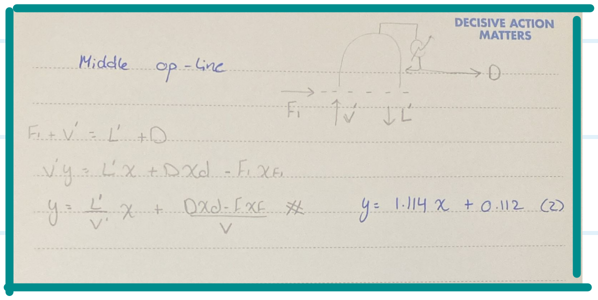
$L' = L + F_1 = 40 + 93.68 = 133.68$ #
$V' = V_1 = 40 + 80 = 120$ #] b

(1) $y = \frac{R}{R+1} \cdot x + \frac{1}{R+1} \cdot x_d \rightarrow y = 1/3\, x + 0.44$

(3) $\bar{L} = L' + 0.7(100) = 203.68$, $\bar{V} = V' - F \cdot (1-q) = 90$

$\bar{V}/B = 90/113.68 = 0.79$ $y = \frac{V_B + 1}{V_B} \cdot x - \frac{1}{V_B} \cdot x_b \rightarrow$ ~~[illegible]~~

$y = 2.26\, x - 0.05$ #



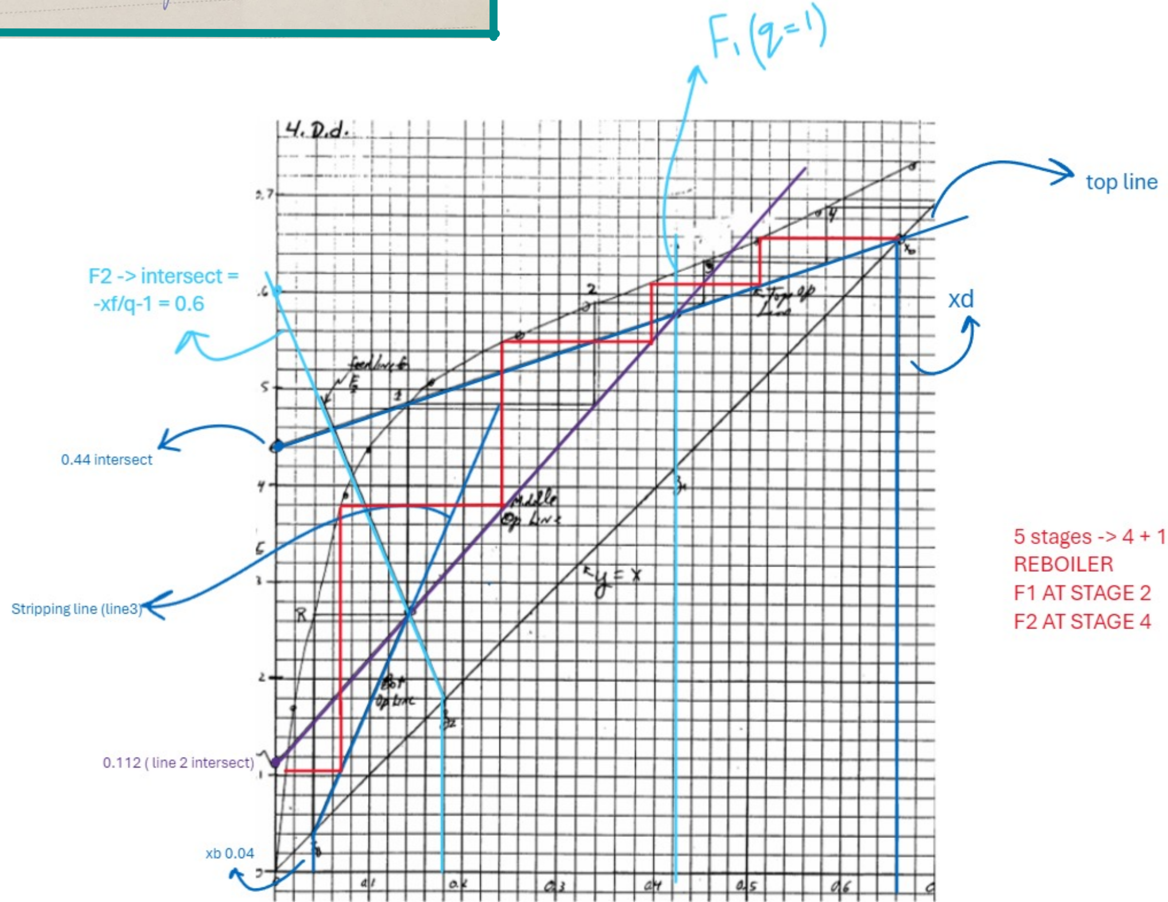
Middle op-line

$F_1 + V' = L' + D$

$V' y = L' x + D x_d - F_1 x_{F_1}$

$y = \frac{L'}{V'} x + \frac{D x_d - F x_F}{V}$ #

$y = 1.114\, x + 0.112$ (2)

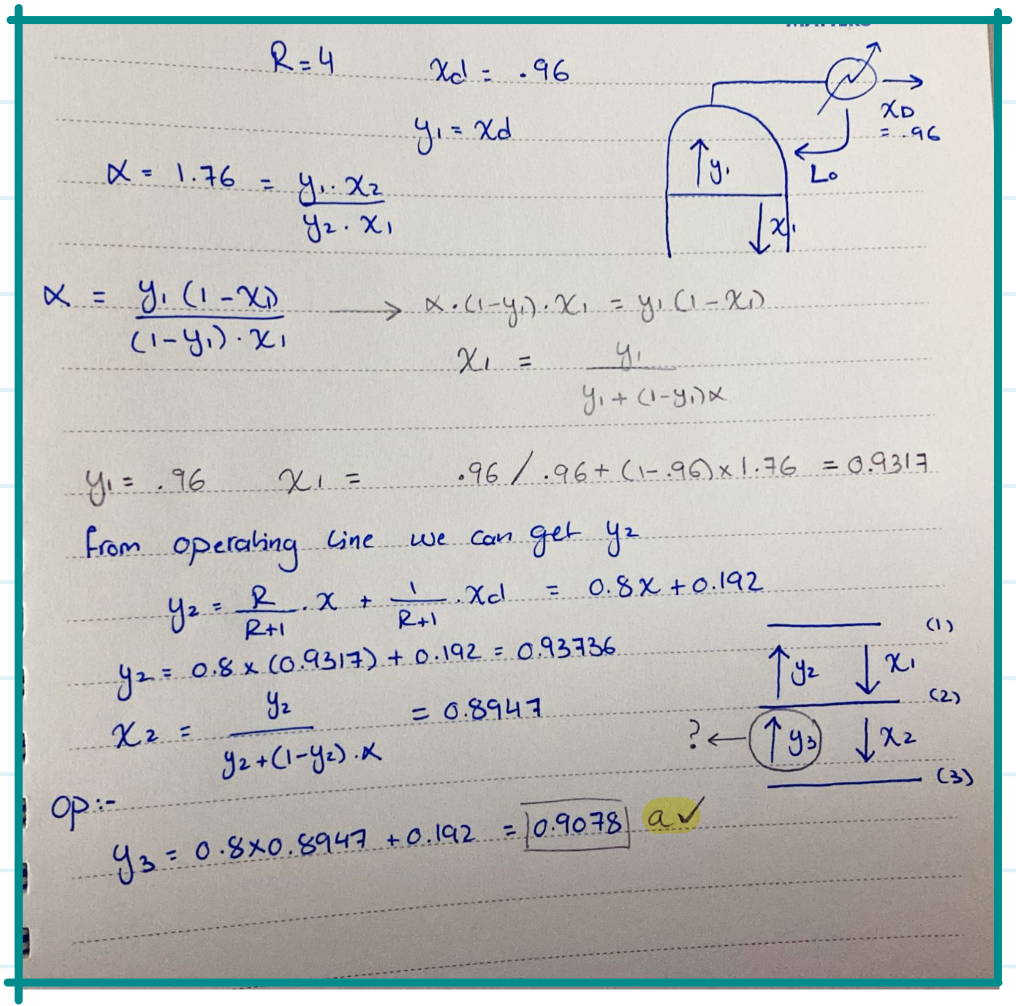


5 stages -> 4 + 1 for REBOILER
F1 AT STAGE 2
F2 AT STAGE 4

D10.* A distillation column is separating phenol from p-cresol at 1 atm pressure. The distillate composition desired is 0.96 mole fraction phenol. An external reflux ratio of L/D = 4 is used, and the reflux is returned to the column as a saturated liquid. The equilibrium data can be represented by a constant relative volatility, $\alpha_{phenol-cresol} = 1.76$ (Perry et al., 1963, pp. 13–3). CMO can be assumed.

a. What is the vapor composition leaving the third equilibrium stage below the total condenser? Solve this by an analytical stage-by-stage calculation alternating between the operating equation and the equilibrium equation.

b. What is the liquid composition leaving the sixth equilibrium stage below the total condenser? Solve this problem graphically using a McCabe-Thiele diagram plotted for $\alpha_{p-c} = 1.76$.



$R = 4$ $\quad x_d = .96$

$y_1 = x_d$

$\alpha = 1.76 = \frac{y_1 \cdot x_2}{y_2 \cdot x_1}$

$\alpha = \frac{y_1 (1 - x_1)}{(1 - y_1) \cdot x_1} \longrightarrow \alpha \cdot (1 - y_1) \cdot x_1 = y_1 (1 - x_1)$

$x_1 = \frac{y_1}{y_1 + (1 - y_1)\alpha}$

$y_1 = .96 \quad x_1 = .96 / .96 + (1 - .96) \times 1.76 = 0.9317$

From operating line we can get y_2

$y_2 = \frac{R}{R+1} \cdot x + \frac{1}{R+1} \cdot x_d = 0.8x + 0.192$

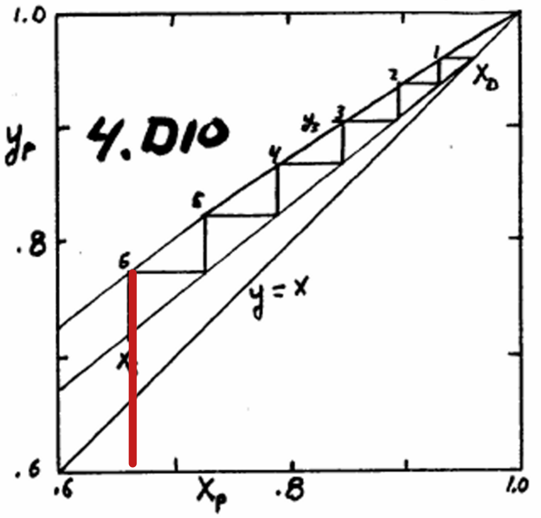
$y_2 = 0.8 \times (0.9317) + 0.192 = 0.93736$

$x_2 = \frac{y_2}{y_2 + (1 - y_2) \cdot \alpha} = 0.8947$

op:-

$y_3 = 0.8 \times 0.8947 + 0.192 = \boxed{0.9078}$ a ✓

b) Plot equilibrium curve and operating line. (See Figure). Slope = L/V = .8, y intercept (x = 0) = 0.192, y = x = x_D = 0.96. Find x_6 = 0.660.



D11. A mixture of methanol and water is being separated in a distillation column with open steam. The feed is 100.0 kmol/h. Feed is 60.0 mol% methanol and is at 40°C. The column is at 1.0 atm. The steam is pure steam ($y_M = 0$) and is a saturated vapor. The distillate product is 99.0 mol% methanol and leaves as a saturated liquid. The bottoms is 2.0 mol% methanol and since it leaves an equilibrium stage must be a saturated liquid. The column is adiabatic. The column has a total condenser. External reflux ratio is L/D = 2.3. Assume CMO is valid. Equilibrium data are in Table 2-7. Data for water and methanol are available in Problem 3.E1.

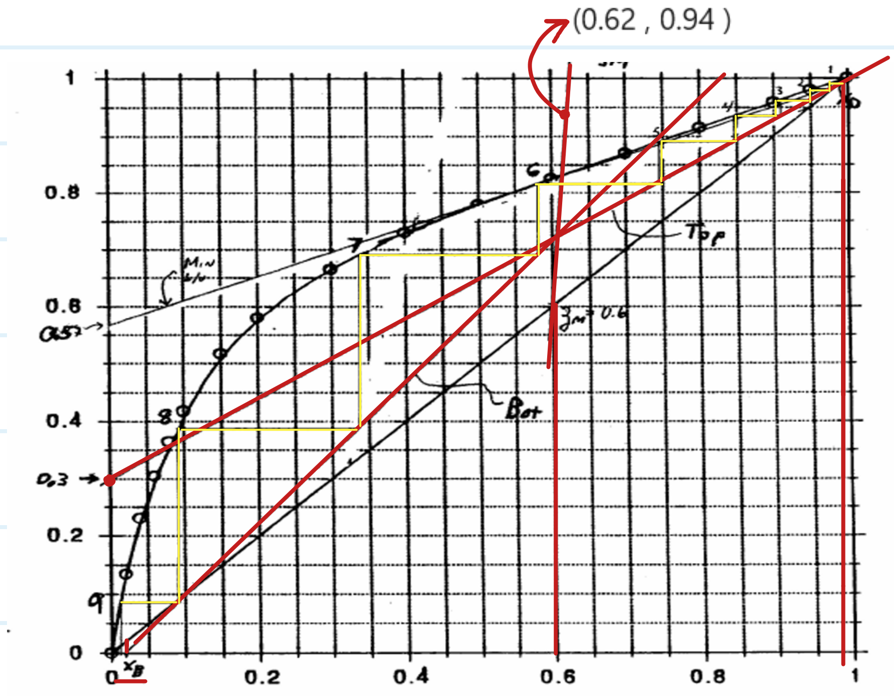
a. Estimate q.

b. Find optimum feed stage and total number of equilibrium stages (step off stages from top down). Use a McCabe-Thiele diagram.

c. Find $(L/D)_{min}$. Use a McCabe-Thiele diagram.

from $q = 1 + \frac{c_{pL}(T_b - T_F)}{\lambda}$

q = 1.0668



b)

top line

y= R/R+1 + xd/R+1

y= 0.7x+0.3

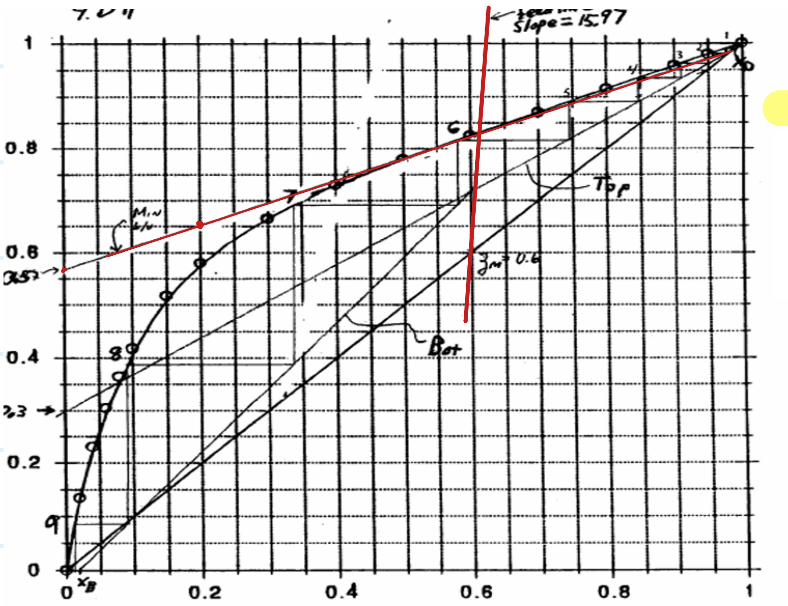
q line

y=16x-9

random point (0.62,0.94)

Total number of stages = 9

feed at stage 6



c)

draw a line from xd to q-line and eq curve intersect and find the slope

slope = L/V = 0.4242 = R/R+1

Rmin= L/D(min) = 0.73684

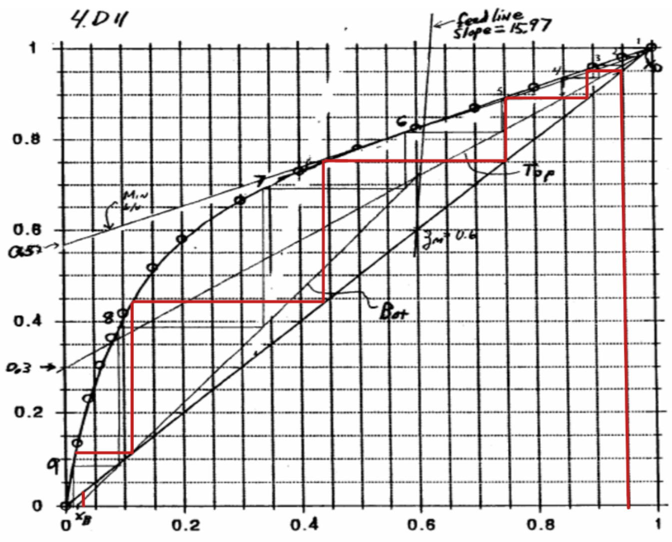
D13. A distillation column is separating a 30% methanol–70% water feed. The feed rate is 237 kmol/h and is a saturated liquid. The column has a partial reboiler and a partial condenser. We desire a distillate mole fraction of $y_{D,M} = 0.95$ and a bottoms mole fraction of $x_{B,M} = 0.025$. Assume CMO is valid. Data are in Table 2-7 and Problem 3.E1.

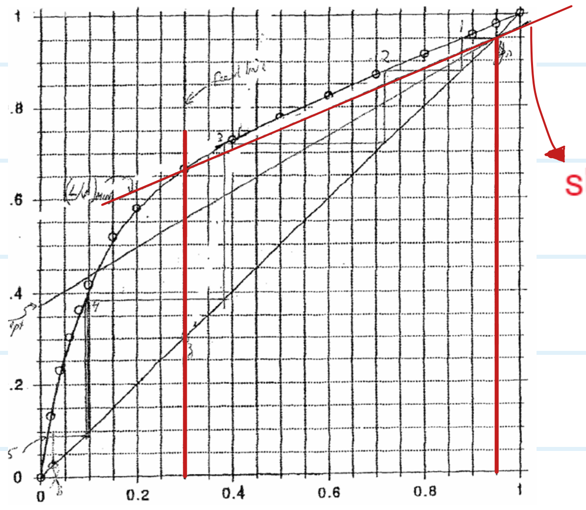
a. Find N_{min}.

b. Find $(L/V)_{min}$ and $(L/D)_{min}$.

c. If $L/D = 2.0\ (L/D)_{min}$, find the optimum feed plate location and the total number of equilibrium stages required.

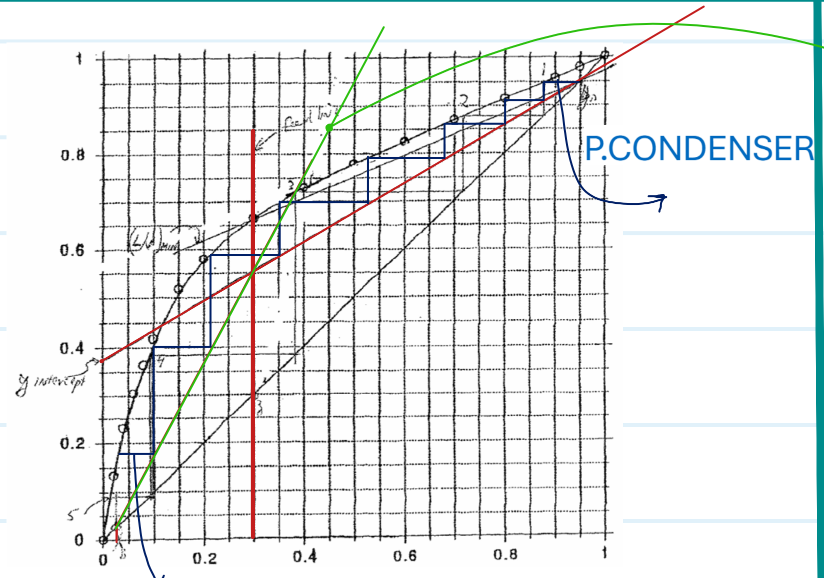
d. Determine the boilup ratio used.





a) 5 Total stages
(including 1 for condenser
and 1 for reboiler)

b)
slope = L/V = R/(R+1)
2 points
(0.95,0.95) , (0.3,0.665)
slope = 0.438
R(min) = L/D(min) = 0.78



for Reboiler

(0.45,0.85)

d)
slope = (Vb+1)/Vb
(0.025,0.025) , (0.45,0.85)
slope = 1.94 -> Vb = 1.0625

c)
R=2*0.78= 1.56
y=(R/R+1) *x + xd/(R+1)
y= 0.6 x + 0.37
feed line ; q=1

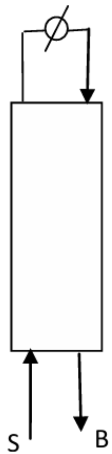
8 total stages
1 for reboiler
1 for P.C
feed at stage 6

D14. A distillation column with open steam heating is separating a feed that is 80.0 mol% methanol and 20.0 mol% water in a steady state operation. The column has 10 stages, a total condenser, and the feed is on stage 5. Operation is at 1.0 atm. The steam is pure water and is a saturated vapor. CMO can be assumed to be valid. At 2:16 a.m. 25 days ago the feed and distillate flows were shut off (D = F = 0), but the steam rate was unchanged and the total condenser is still condensing the vapor to a saturated liquid. The column has now reached a new steady state operation.

a. What is the current methanol mole fraction in the bottoms?

b. At the new steady state estimate the methanol mole fraction in the liquid leaving the total condenser.

a)



New S.S.

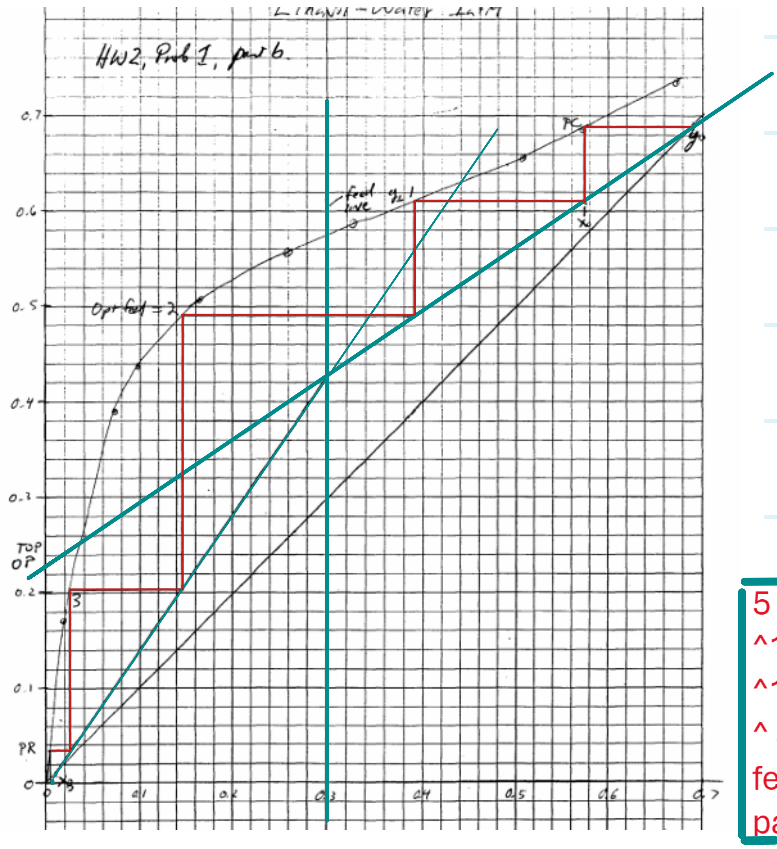
External M.B. S = B

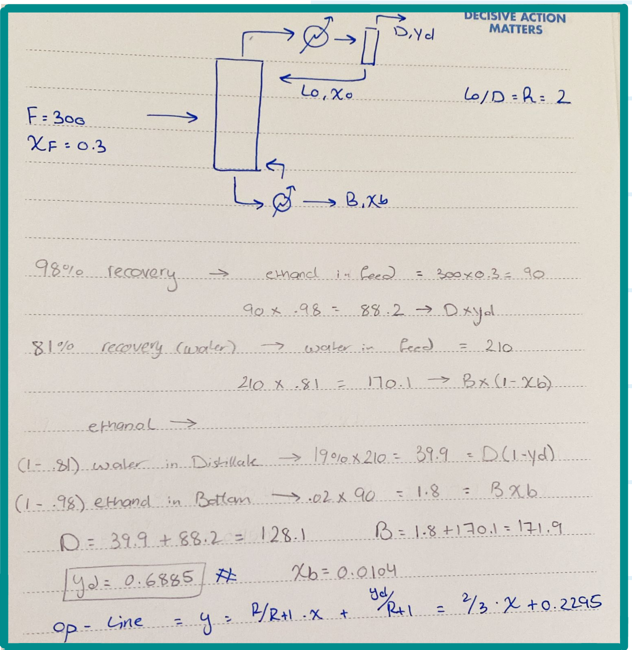
$Sy_s = Bx_B$. Since $y_S = 0$ (pure water)

$xb=0$

b) Common sense is all methanol leaks out and $x_{MA} = 0$.

D15. A partial condenser takes vapor leaving the top of a distillation column and condenses a portion of it. The vapor portion of mole fraction y_D is removed as the distillate product. The liquid portion of mole fraction x_0 is returned to the column as reflux. The liquid and vapor leaving partial condensers and partial reboilers can be assumed to be in equilibrium. A distillation column with a partial condenser and a partial reboiler is separating 300 kmol/h of a mixture that is 30 mol% ethanol and 70 mol% water and is a saturated liquid. We desire a 98% recovery of the ethanol in the vapor distillate and an 81% recovery of water in the bottoms. If $L_0/D = 2.0$ find the optimum feed location and the total number of equilibrium stages.





$F = 300$

$x_F = 0.3$

$L_0/D = R = 2$

98% recovery → ethanol in feed = 300×0.3 = 90

90 × .98 = 88.2 → D×yd

81% recovery (water) → water in feed = 210

210 × .81 = 170.1 → B×(1−xb)

ethanol →

(1−.81) water in Distillate → 19%×210 = 39.9 = D(1−yd)

(1−.98) ethand in Bottom → .02 × 90 = 1.8 = B xb

D = 39.9 + 88.2 = 128.1

B = 1.8 + 170.1 = 171.9

$\boxed{yd = 0.6885}$ #

xb = 0.0104

op-line = $y = \frac{R}{R+1} \cdot x + \frac{yd}{R+1} = \frac{2}{3} \cdot x + 0.2295$

5 total stages

^1 for P.C

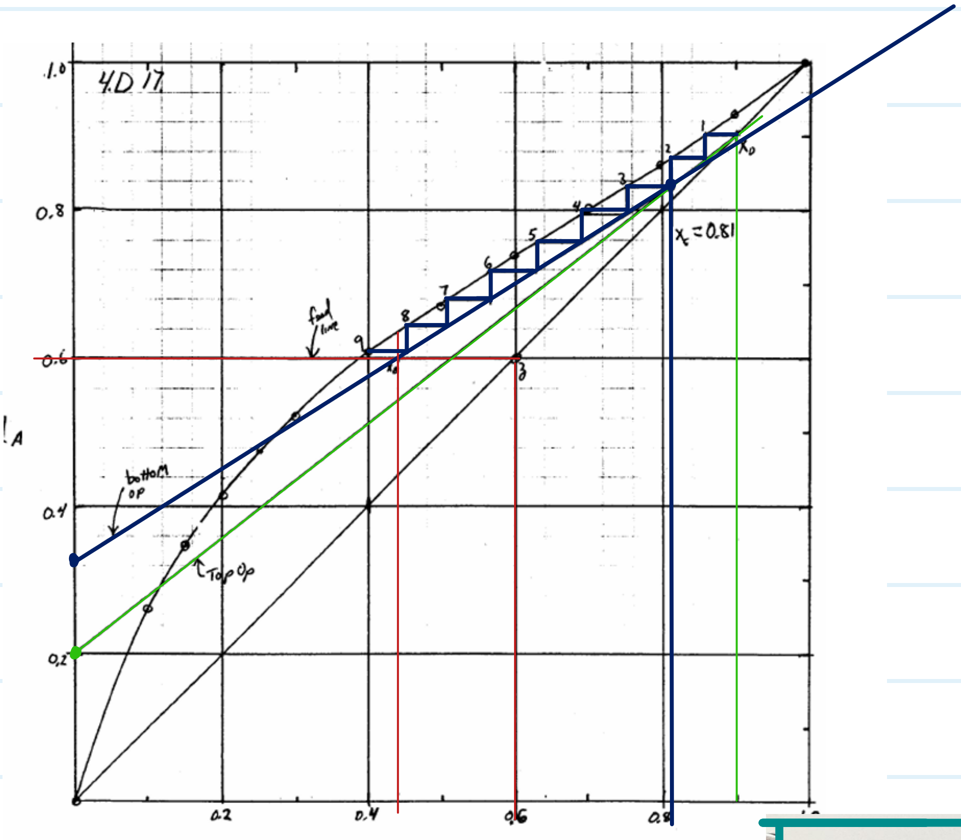
^1 for P.R

^ 3 equillibrium stages

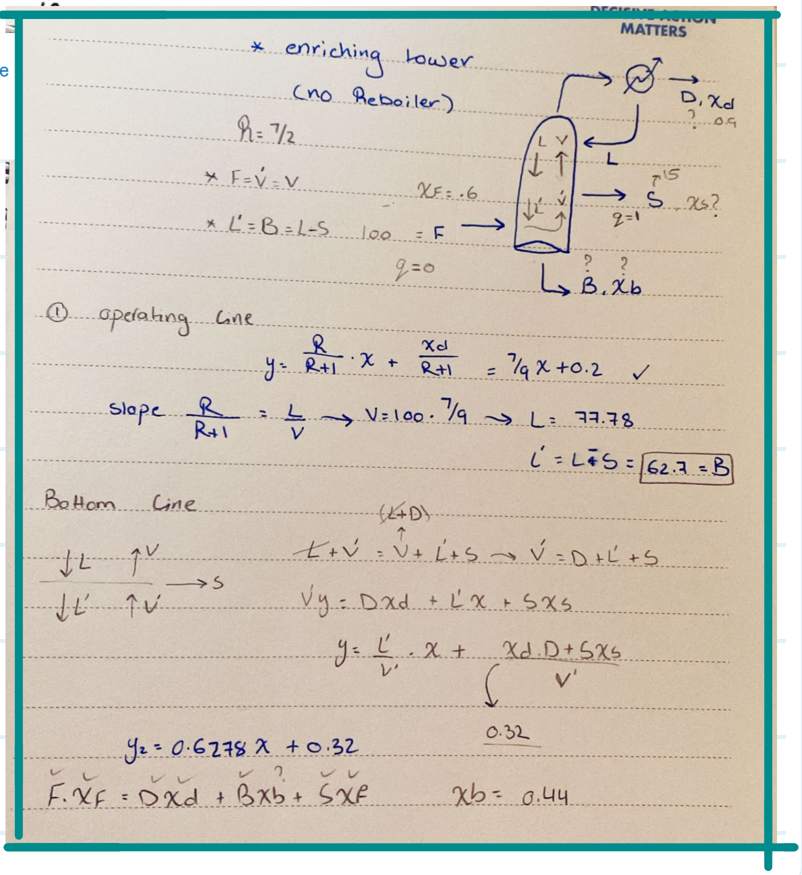
feed at stage 2 below partial condenser

D17. A mixture of acetone and ethanol (acetone is more volatile) is fed to an enriching column that has a liquid side stream withdrawn. The feed flow rate is 100.0 mol/min. Feed is 60.0 mol% acetone and is a saturated vapor. The liquid side product is withdrawn from the second stage below the total condenser at a flow rate of S = 15.0 mol/min. The reflux is returned as a saturated liquid. The distillate should be 90.0 mol% acetone. The external reflux ratio is L/D = 7/2. Column pressure is 1.0 atm. Column is adiabatic and CMO is valid. Equilibrium data are in Problem 4.D7. Note: Trial and error is **not** required.

Find the mole fraction of acetone in the sidestream x_S, the mole fraction of acetone in the bottoms x_B, and the number of equilibrium stages required.



given Inf; the side stream is withdrawn second stage below condenser so we can get xs from graph ~ xs=0.81



* enriching tower

(no Reboiler)

$R = 7/2$

* $F = \acute{V} = V$ $\quad x_F = .6$

* $L' = B = L - S$ $\quad 100 = F$

$q = 0$

① operating Line

$$y = \frac{R}{R+1} \cdot x + \frac{x_d}{R+1} = 7/9\, x + 0.2 \;\checkmark$$

$$\text{slope } \frac{R}{R+1} = \frac{L}{V} \rightarrow V = 100 \cdot 7/9 \rightarrow L = 77.78$$

$$L' = L - S = \boxed{62.7 = B}$$

Bottom Line

$$\overset{\overset{(\cancel{L}+D)}{\uparrow}}{\cancel{L}} + \acute{V} = V + L' + S \rightarrow \acute{V} = D + L' + S$$

$$\acute{V} y = D x_d + L' x + S x_s$$

$$y = \frac{L'}{V'} \cdot x + \frac{x_d \cdot D + S x_s}{V'}$$

0.32

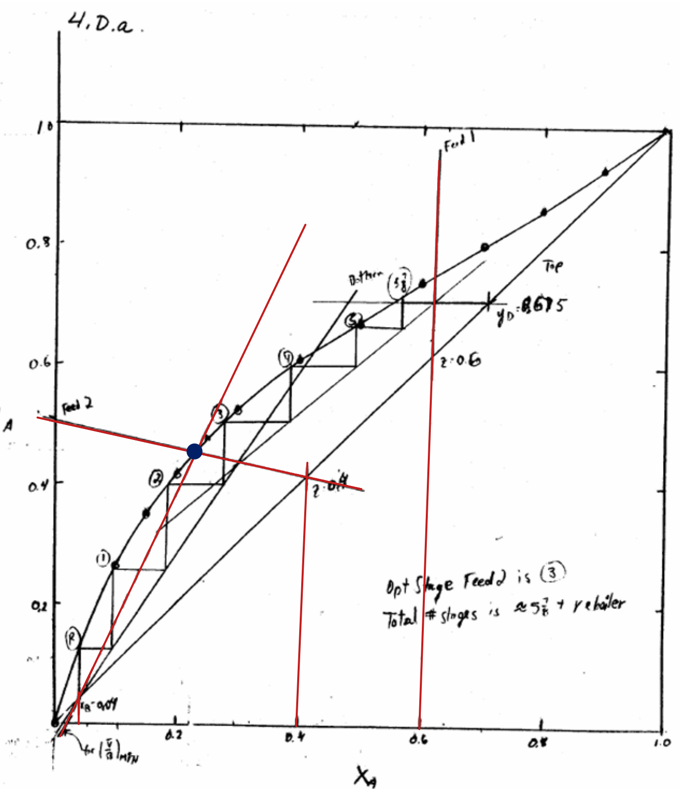
$$y_2 = 0.6278\, x + 0.32$$

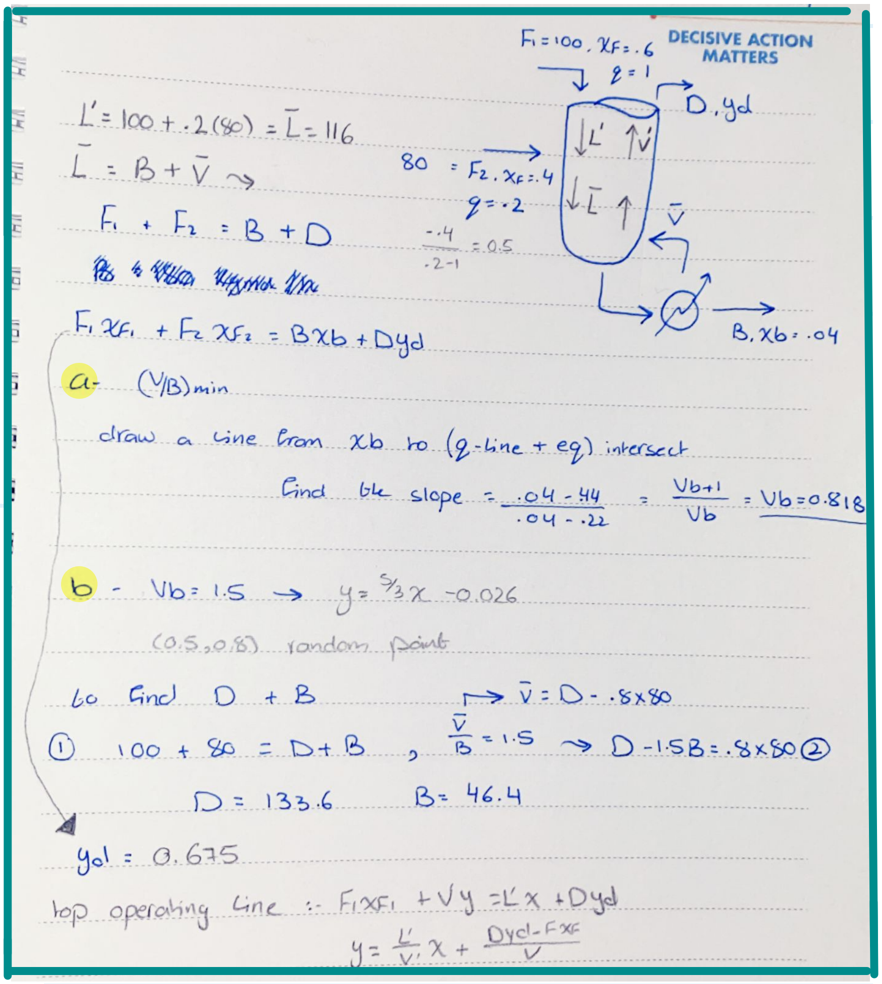
$$F \cdot x_F = D x_d + B x_b + S x_F \qquad x_b = 0.44$$

D18. We have a stripping column with two feeds separating acetone and ethanol at 1 atm. Feed F_1 is a saturated liquid and is fed into the top of column (no condenser). Flow rate of F_1 is 100 kgmol/h, and this feed is 60 mol% acetone. Feed F_2 is 40 mol% acetone, it is a two-phase feed that is 80% vapor, and flow rate is 80 kmol/h. We desire a bottoms mole fraction that is 0.04 mole fraction acetone. The column has a partial reboiler. Equilibrium data are in Problem 4.D7.

a. Calculate $(\overline{V}/B)_{min}$.

b. If $(\overline{V}/B) = 1.5$, find D and y_D, the optimum feed stage for feed F_2, and the total number of stages. Please step off stages from the bottom upwards.





$F_1 = 100$, $x_F = .6$

$q = 1$

D, yd

DECISIVE ACTION MATTERS

$L' = 100 + .2(80) = \bar{L} = 116$

$\bar{L} = B + \bar{V}$ ↝

$80 = F_2$, $x_F = .4$

$q = .2$

$\frac{-.4}{.2-1} = 0.5$

$F_1 + F_2 = B + D$

B, xb = .04

$F_1 x_{F_1} + F_2 x_{F_2} = B x_b + D y_d$

a- (V/B)min

draw a line from xb to (q-line + eq) intersect

find the slope $= \frac{.04 - .44}{.04 - .22} = \frac{V_b+1}{V_b} = \underline{V_b = 0.818}$

b - $V_b = 1.5 \rightarrow y = \frac{5}{3}x - 0.026$

(0.5, 0.8) random point

to find D + B → $\bar{V} = D - .8 \times 80$

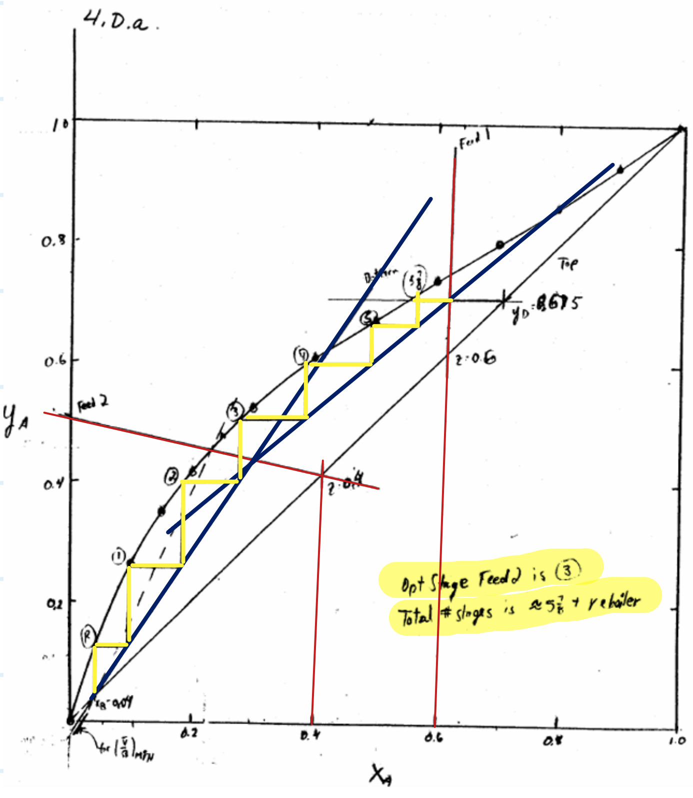
① $100 + 80 = D + B$, $\frac{\bar{V}}{B} = 1.5 \rightarrow D - 1.5B = .8 \times 80$ ②

$D = 133.6$ $B = 46.4$

$y_d = 0.675$

top operating line :- $F_1 x_{F_1} + V y = L' x + D y_d$

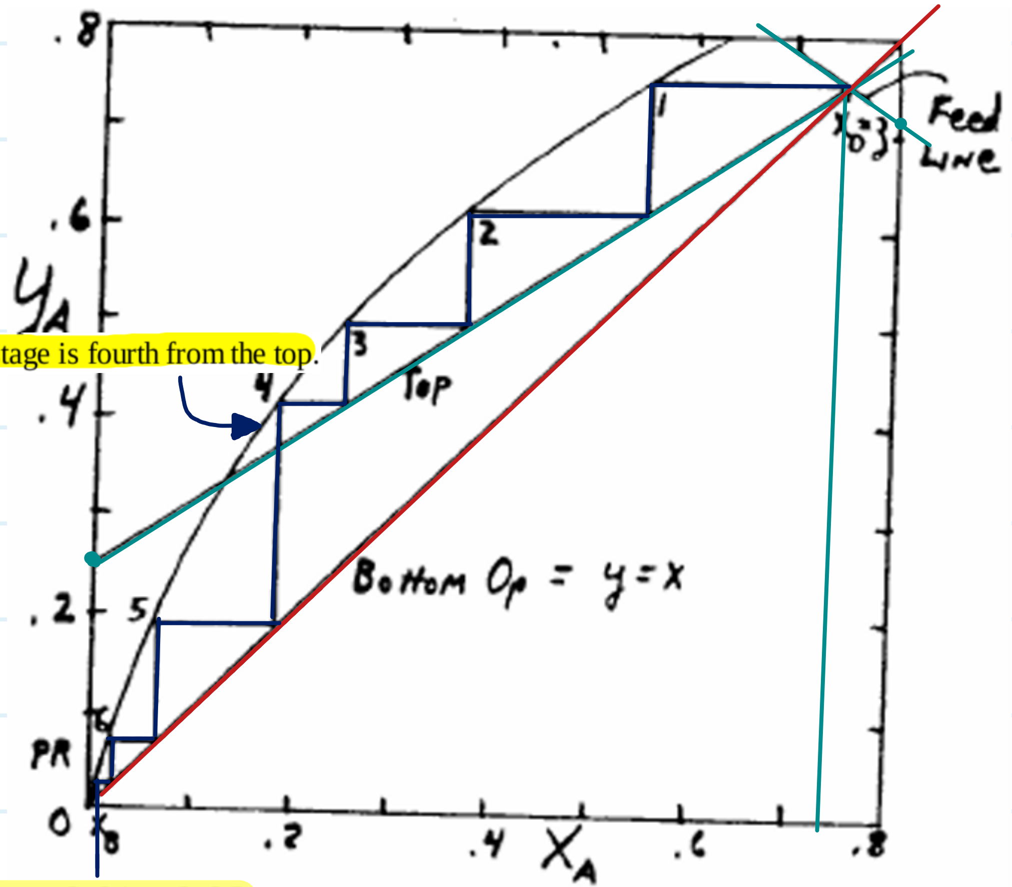
$y = \frac{L'}{V'} x + \frac{D y_d - F x_F}{V}$



Opt Stage Feed 2 is ③

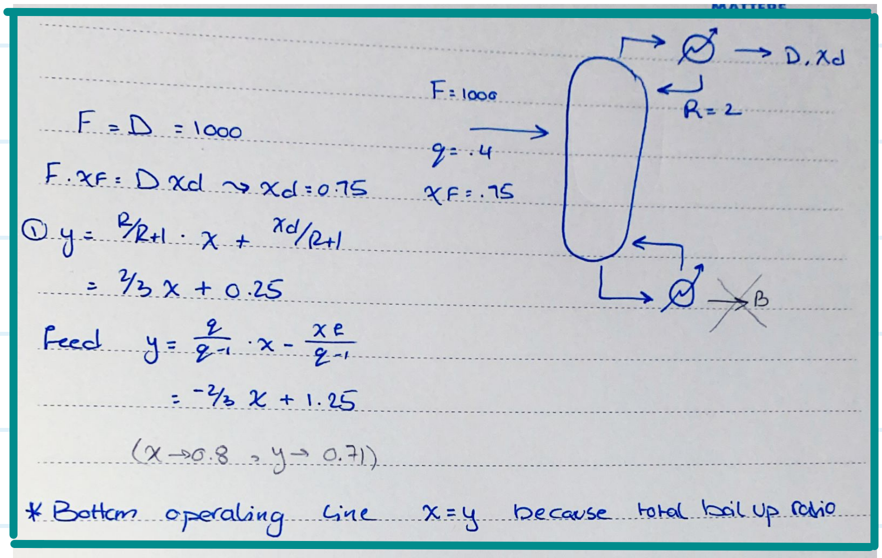
Total # stages is ≈5 3/8 + reboiler

D19.* A distillation column is separating acetone and ethanol. The column effectively has six equilibrium stages plus a partial reboiler. Feed is a two-phase feed that is 40% liquid and 75 mol% acetone. Feed rate is 1000 kmol/h, and the feed stage is fourth from the top. The column is now operating at a steady state with the bottoms flow valve shut off. However, a distillate product is drawn off, and the vapor is boiled up in the reboiler. $L_0/D = 2$. Reflux is a saturated liquid. CMO can be assumed. p = 1 atm. Equilibrium data are in Problem 4.D7. Find the distillate composition. If one drop of liquid in the reboiler is withdrawn and analyzed, predict x_B.



feed stage is fourth from the top.

xb = 0.01 - 0.02


F = D = 1000

F . xF = D xd ⇝ xd = 0.75

① $y = \frac{R}{R+1} \cdot x + \frac{xd}{R+1}$

$= \frac{2}{3} x + 0.25$

Feed $y = \frac{q}{q-1} \cdot x - \frac{xF}{q-1}$

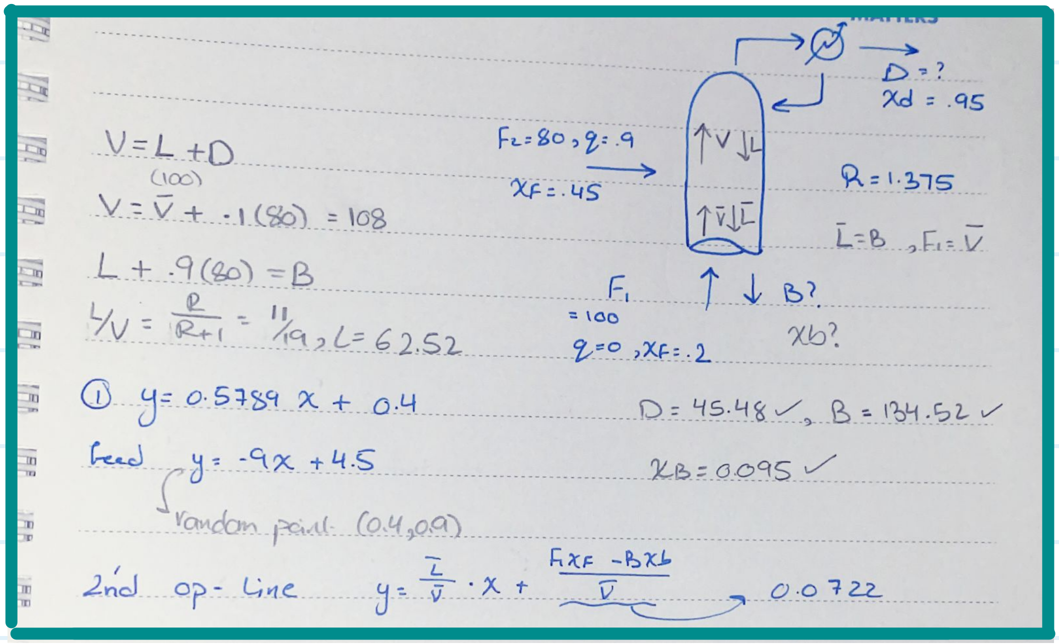
$= -\frac{2}{3} x + 1.25$

(x → 0.8 , y → 0.71)

* Bottom operating line x = y because total boil up ratio

D21. An enricher column has two feeds. Feed F_1 (input at the bottom) is a saturated vapor. Flow rate F_1 = 100.0 kmol/h. This feed is 20.0 mol% methanol and 80.0 mol% water. Feed F_2 (input part way up the column) is a two-phase mixture that is 90 % liquid. Flow rate F_2 = 80.0 kmol/h. Feed F_2 is 45.0 moles % methanol and 55.0 mol% water. We desire a distillate that is 95.0 mol% methanol. Reflux is returned as a saturated liquid. Pressure is one atmosphere. L/D = 1.375. Assume CMO. Data are available in Table 2-7.

Find: D, B, x_B, optimum feed location and number of equilibrium stages required.



$V = L + D$ (100)

$V = \bar{V} + .1(80) = 108$

$L + .9(80) = B$

$L/V = \frac{R}{R+1} = 11/19$, $L = 62.52$

① $y = 0.5789\,x + 0.4$

feed $y = -9x + 4.5$

random point (0.4,0.9)

2nd op-Line $y = \frac{\bar{L}}{\bar{V}} \cdot x + \frac{F_1 x_F - B x_b}{\bar{V}}$ → 0.0722

$F_2 = 80$, $q = .9$

$x_F = .45$

$D = ?$

$x_d = .95$

$R = 1.375$

$\bar{L} = B$, $F_1 = \bar{V}$

$F_1 = 100$

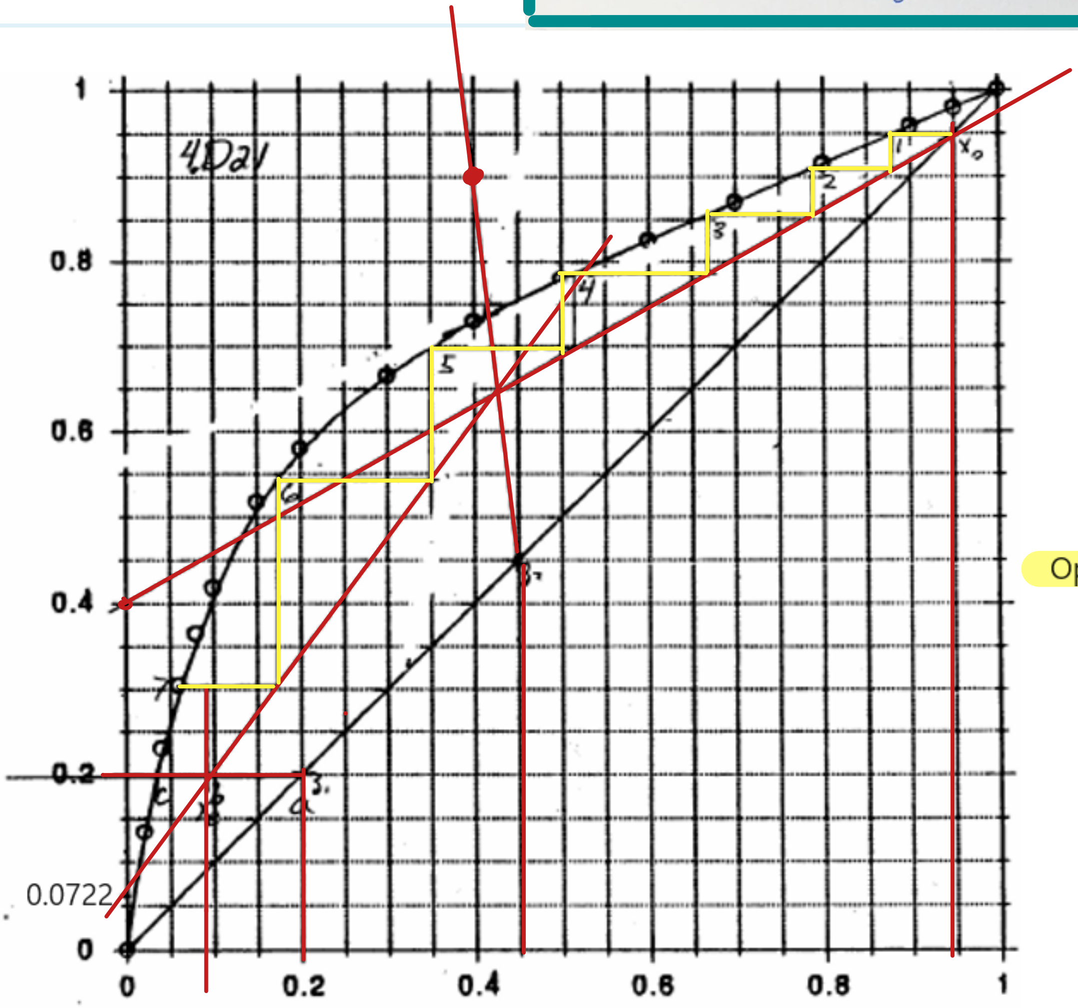
$q = 0$, $x_F = .2$

B?

x_b?

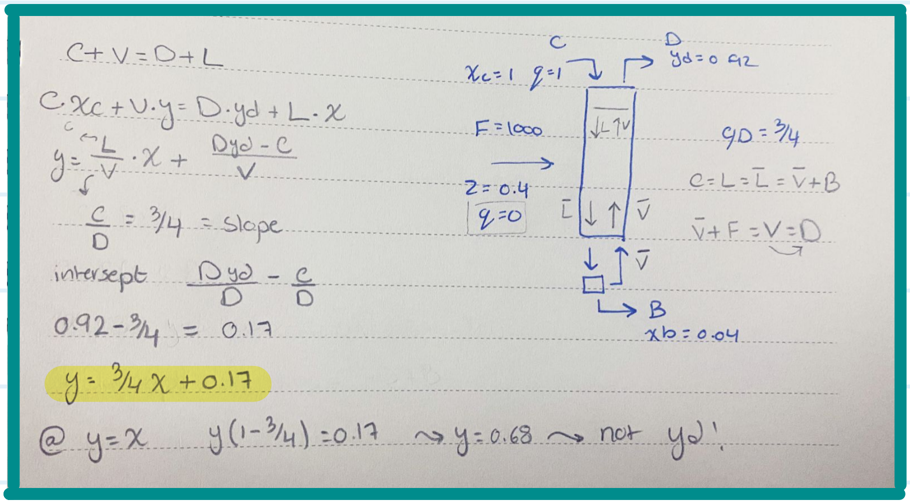
$D = 45.48$ ✓, $B = 134.52$ ✓

$x_B = 0.095$ ✓



Optimum feed = 5, 7 equilibrium stages

D22.* When water is the more volatile component we do not need a condenser but can use direct cooling with boiling water. This was shown in Problem 3.D3. We set $y_D = 0.92$, $x_B = 0.04$, $z = 0.4$ (all mole fractions water), feed is a saturated vapor, feed rate is 1000 kmol/h, p = 1 atm, CMO is valid, the entering cooling water (C) is a saturated liquid and is pure water, and C/D = 3/4. Derive and plot the top operating line. Note that external balances (that is, balances around the entire column) are not required.



$C + V = D + L$

$C \cdot x_c + V \cdot y = D \cdot y_d + L \cdot x$

$y = \frac{L}{V} \cdot x + \frac{D y_d - C}{V}$

$\frac{C}{D} = 3/4 = \text{slope}$

intersept $\frac{D y_d}{D} - \frac{C}{D}$

$0.92 - 3/4 = 0.17$

$y = 3/4\, x + 0.17$

@ $y = x$ $y(1 - 3/4) = 0.17 \rightsquigarrow y = 0.68 \rightsquigarrow$ not y_d!

D23.* When water is the more volatile component we do not need a condenser but can use direct cooling. This was illustrated in Problem 3.D3. We set $y_D = 0.999$, $x_B = 0.04$, $z = 0.4$ (all mole fractions water), feed is a saturated liquid, feed rate is 1000 kmol/h, p = 1 atm, CMO is valid, the entering cooling water (C) is pure water and C/D = 3/4. The entering cooling water is at 100 °F while its boiling temperature is 212 °F.

$$C_{PL,W} = 18.0 \frac{\text{Btu}}{\text{lb mol °F}}, \qquad MW_w = 18, \qquad \lambda_W = 17{,}465.4 \frac{\text{Btu}}{\text{lb mol}}$$

Find the slope of the top operating line, L/V.

Note: Equilibrium data are not needed.

Note L/V ≠ C/D since C is subcooled. Let c = amount condensed. The energy required to heat stream W to the boiling point must come from this condensation. That is,

$$(H - h)\, c = (h - h_W)\, C$$

$$c = \frac{h - h_W}{H - h} C = \frac{C_p\, \Delta T}{\lambda} C = \frac{18\,(212 - 100)}{17465.4} C = 0.1154C$$

$$L = C + c = 1.1154C$$

$$V = D + c = D + 0.1154C$$

In addition, C/D = ¾ or D/C = 4/3.

$$\frac{L}{V} = \frac{1.1154C}{D + .1154C} = \frac{1.1154}{4/3 + .1154} = \frac{1.1154}{4/3 + .1154} = 0.77$$

This compares to L/V = 0.75 if entering water is a saturated liquid. Very little effect since λ is very large.

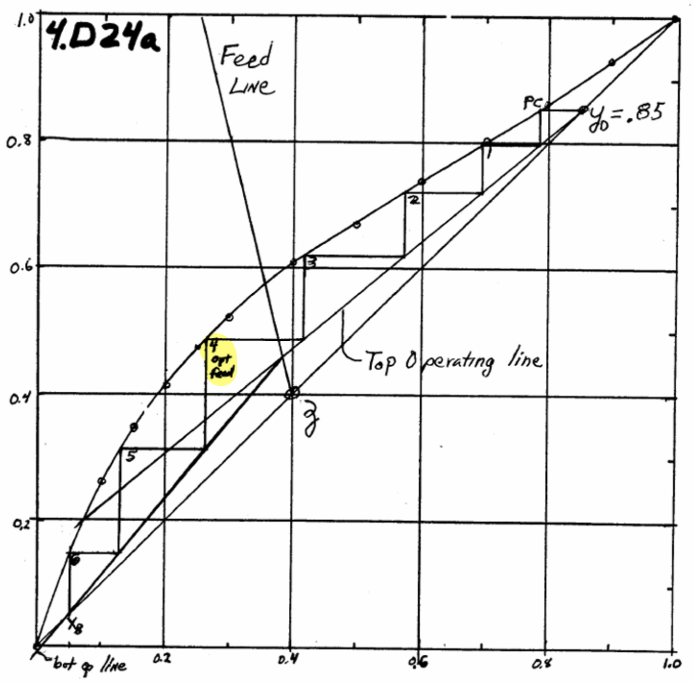
Same As (D22) but instead of C entering as a sat-liquid, it enters as a subcooled liquid

D24. We have a distillation column with a partial condenser and a total reboiler separating a feed of 200.0 kmol/h. The feed is 40.0 mol% acetone and 60.0 mol% ethanol. The feed is a two-phase mixture that is 80% liquid. We desire a distillate vapor that is 85.0 mol% acetone and a bottoms that is 5.0 mol% acetone. Column pressure is one atmosphere. Reflux is returned as a saturated liquid and L/D = 3.25. Assume CMO. Equilibrium data are in Problem 4.D7.

a. Find the optimum feed location and the total number of equilibrium stages.

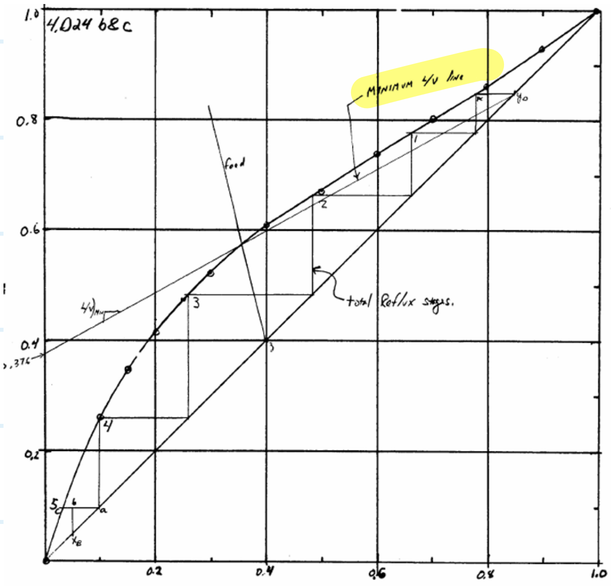
b. What is the minimum external reflux ratio?

c. What is the minimum number of stages (total reflux)?

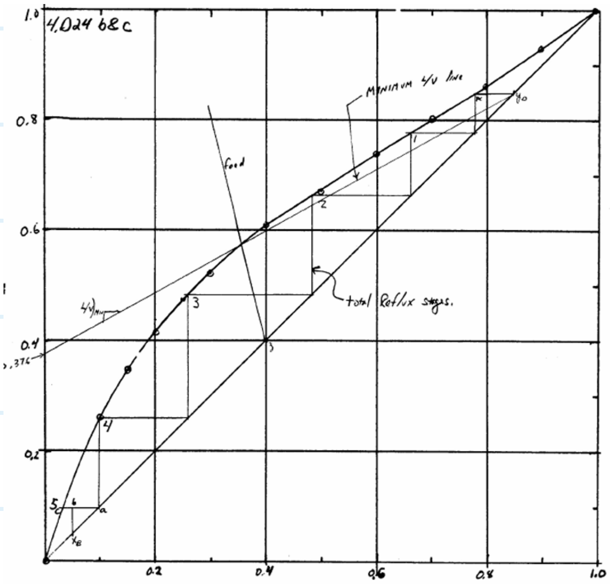


a) PC + 6 STAGES
feed at stage 4

b) Slope = Rmin/(Rmin+) =0.557
Rmin = 1.26



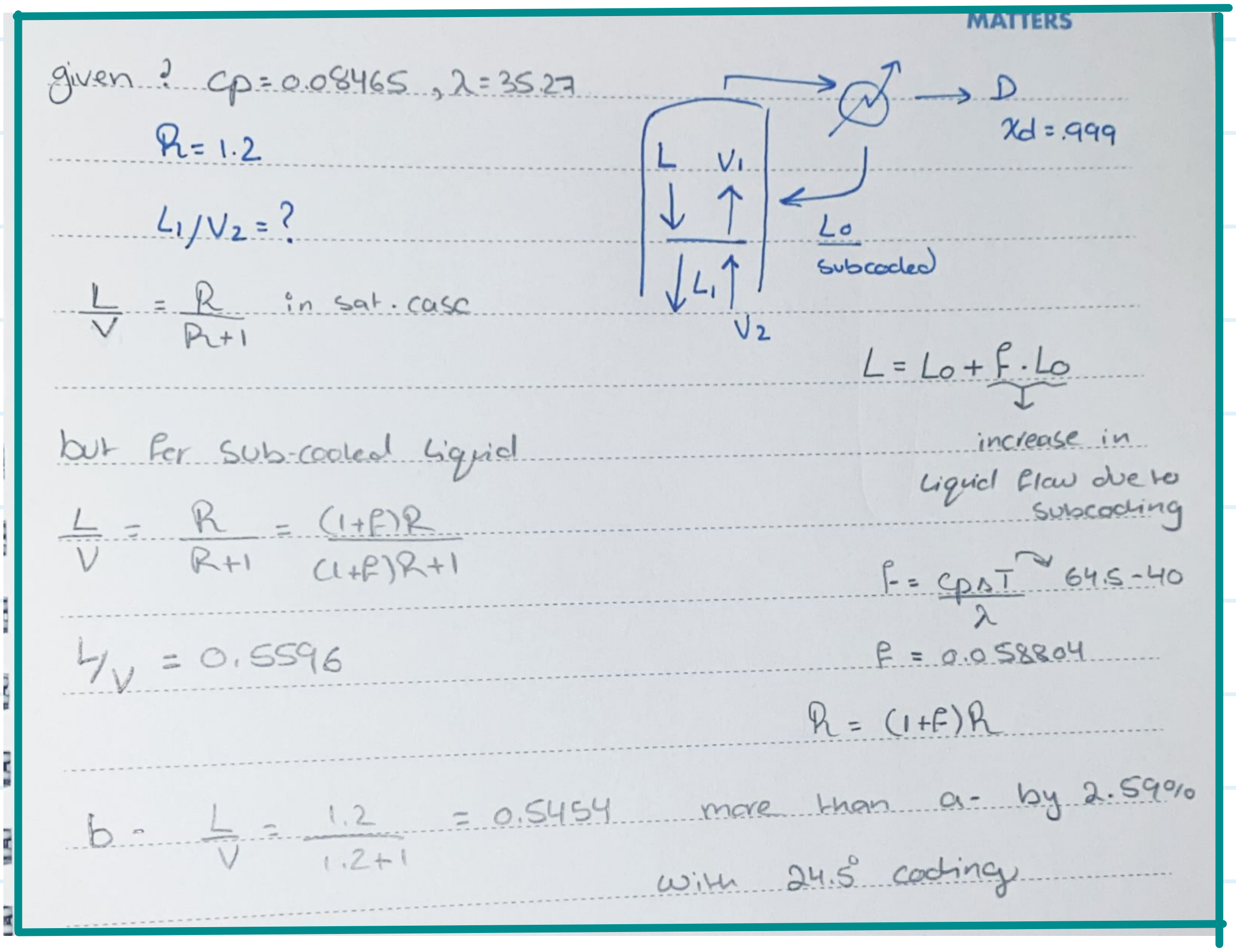
c) PC + 5 STAGES



D25. We are separating methanol and water. Calculate the internal reflux ratio inside the column, L_1/V_2, for the following cases. The column is at 101.3 kPa. Data are available in Table 2-7 and in Problem 3.E1.

a. Distillate product is 99.9 mol% methanol. External reflux ratio is $L_0/D = 1.2$. Reflux is cooled to 40.0°C. (i.e., it is subcooled).

b. Repeat part a except for a saturated liquid reflux. Compare with Part a.



given : $c_p = 0.08465$, $\lambda = 35.27$

$R = 1.2$

$L_1/V_2 = ?$

$\frac{L}{V} = \frac{R}{R+1}$ in sat. case

$L = L_o + \underbrace{f \cdot L_o}$

increase in liquid flow due to subcooling

but for sub-cooled liquid

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

$f = \frac{c_p \Delta T}{\lambda}$ 64.5 − 40

$f = 0.058804$

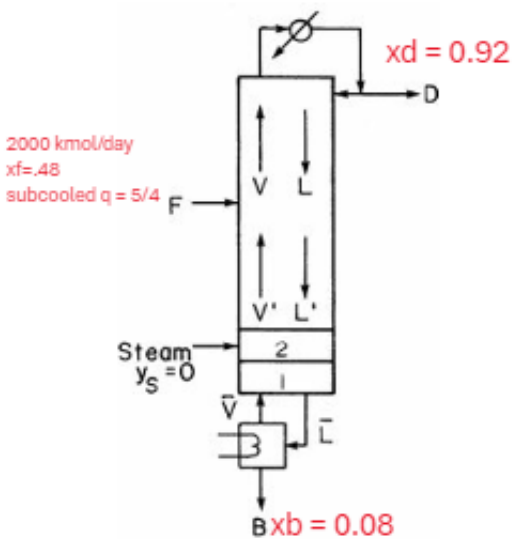
$R = (1+f)R$

$L/V = 0.5596$

b - $\frac{L}{V} = \frac{1.2}{1.2+1} = 0.5454$ more than a- by 2.59%

with 24.5° cooling

D27.* A distillation column is separating methanol from water at 1 atm pressure. The column has a total condenser and a partial reboiler. In addition, a saturated vapor stream of pure steam is input on the second stage above the partial reboiler (see figure).

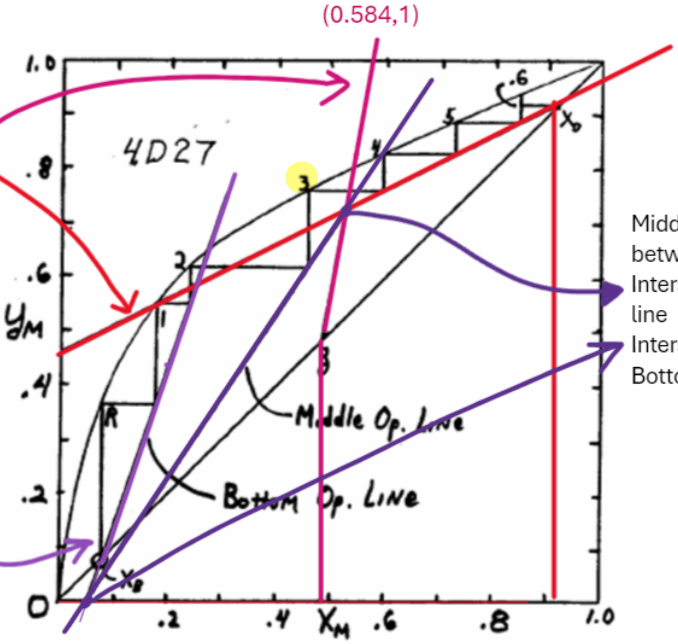


q= (L'-L) / F
L' = L + F + F/4
q = 5/4

The feed flow rate is 2000 kmol/day. Feed is 48 mol% methanol and 52 mol% water and is a subcooled liquid. For every 4 moles of feed, 1 mole of vapor must condense inside the column. Distillate composition is 92 mol% methanol. Reflux is a saturated liquid, and $L_0/D = 1.0$. Bottoms composition is 8 mol% methanol. Boilup ratio is $\bar{V}/B = 0.5$. Equilibrium data are given in Table 2-7. Assume that CMO is valid. Find the optimum feed plate location and the total number of equilibrium stages required.

R= 1 , Vb = 0.5
Top line = y = 0.5x + 0.46
Bottom line = y = 3x - 0.16
q line = y = 5x - 1.92

6 stages + 1 for Reboiler

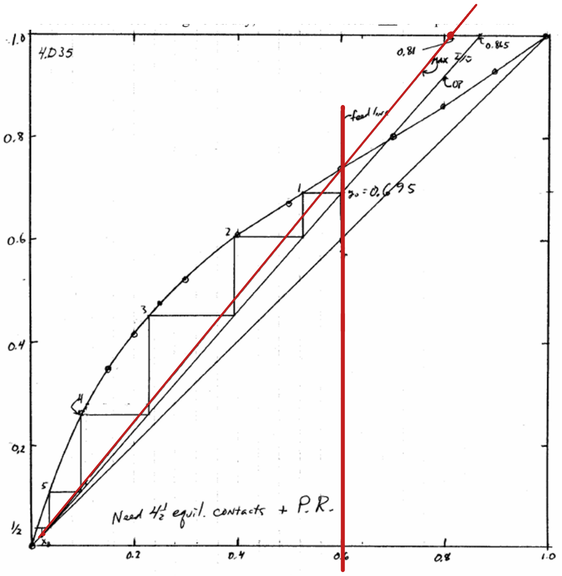


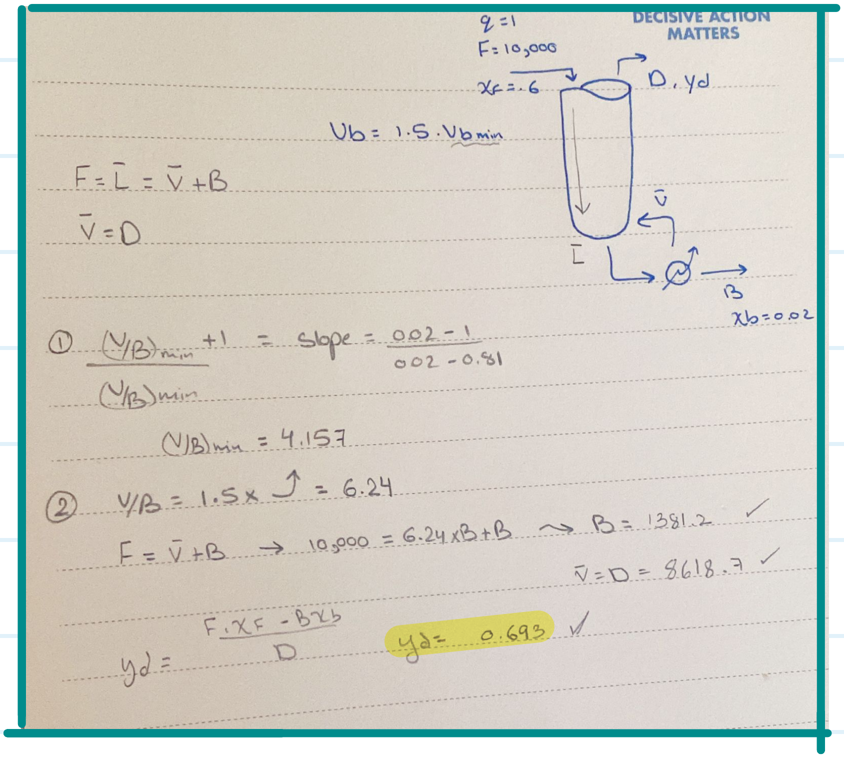
Middle Op.Line between 2 point
Intersection of q1 and op line
Intersection of q2 and Bottom line

Second feed line
q = 0 (Horizontal)
at ys= 0

D28. A stripping column is separating 10,000.0 kmol/day of a saturated liquid feed. The column has a partial reboiler. The feed is 60.0 mol% acetone and 40.0 mol% ethanol. We desire a bottoms that is 2.0 mol% acetone. Operate at a boilup rate $\overline{V}/B = 1.5(\overline{V}/B)_{min}$. Find y_D, N, D, and B. Equilibrium data are in Problem 4.D7.

1) Finding V/B (min)





q = 1
F = 10,000
x_F = .6
x_b = 0.02

$V_b = 1.5 \cdot V_{b\,min}$

$F = \bar{L} = \bar{V} + B$

$\bar{V} = D$

① $\dfrac{(V/B)_{min} + 1}{(V/B)_{min}}$ = slope = $\dfrac{0.02 - 1}{0.02 - 0.81}$

$(V/B)_{min} = 4.157$

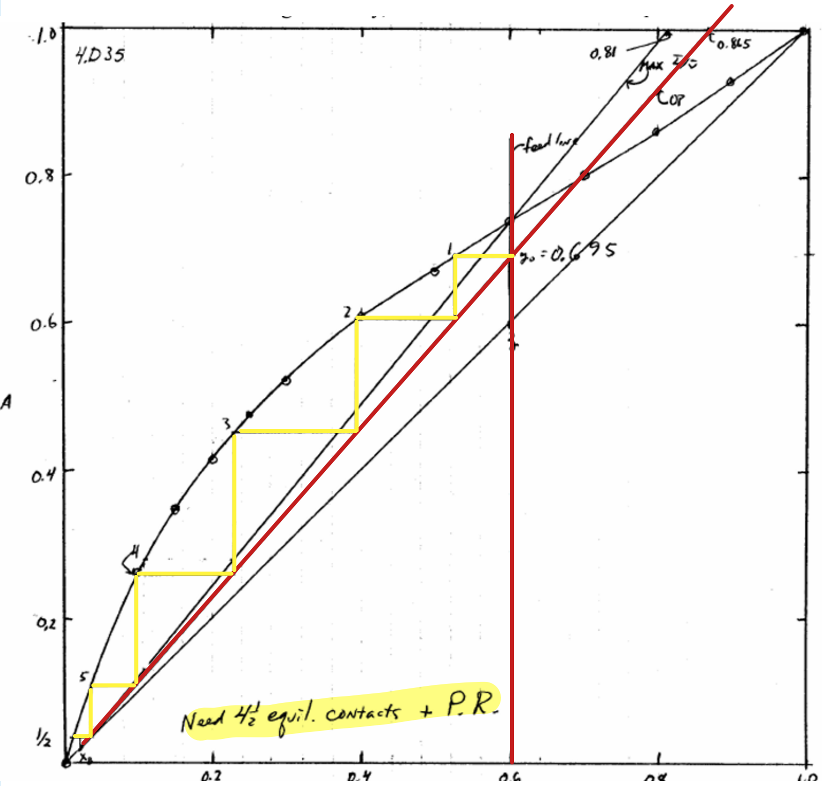
② $V/B = 1.5 \times \curvearrowright = 6.24$

$F = \bar{V} + B \rightarrow 10{,}000 = 6.24 \times B + B \rightsquigarrow B = 1381.2$ ✓

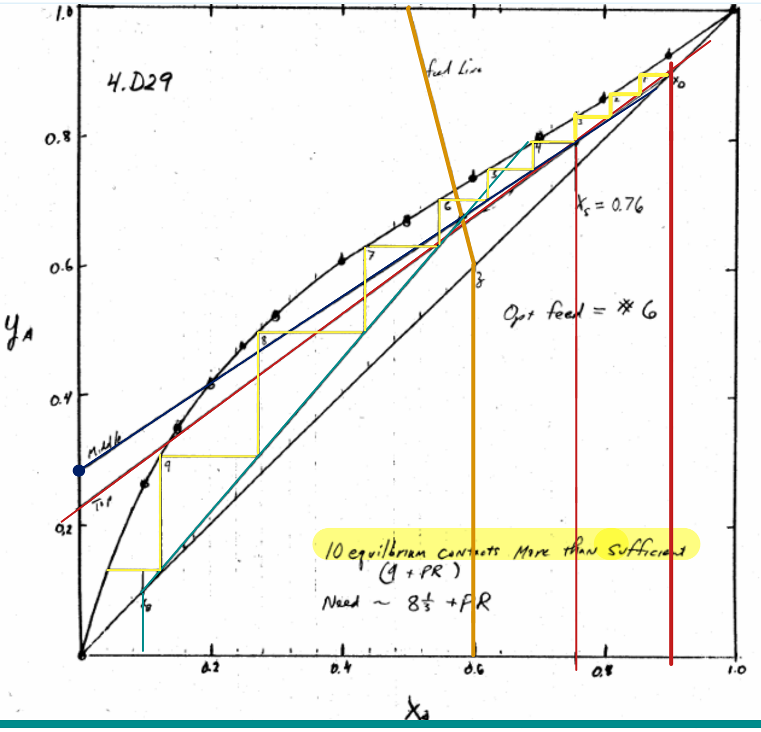
$\bar{V} = D = 8618.7$ ✓

$y_d = \dfrac{F \cdot x_F - B x_b}{D}$ $y_d = 0.693$ ✓

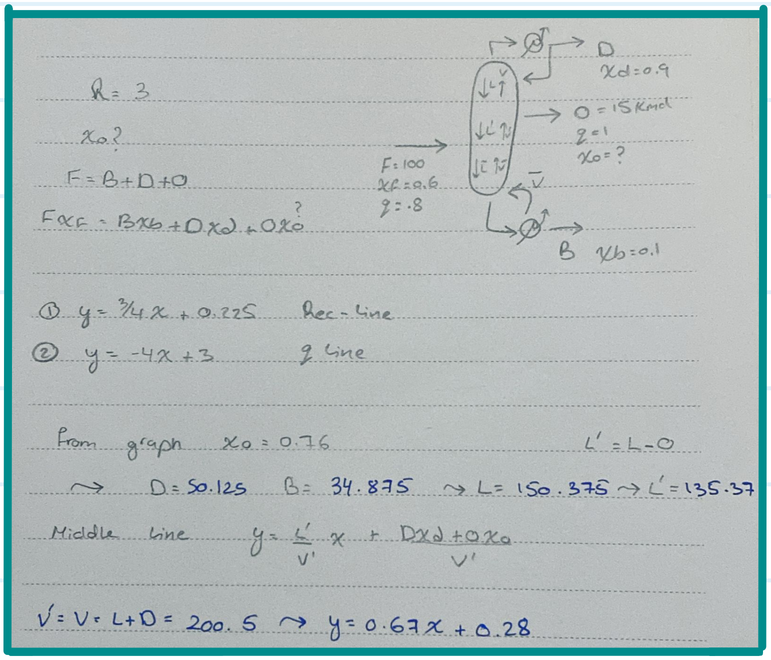
2) finding N



D29. A distillation column will use the optimum feed stage. A liquid side stream is withdrawn on the third stage below the total condenser at a rate of 15.0 kmol/h. The feed is a two phase mixture that is 20% vapor. Feed to the column is 100.0 kmol/h. The feed is 60.0 mol% acetone and 40.0 mol% ethanol. We desire a distillate composition that is 90.0 mol% ethanol. We operate with an external reflux ratio of L/D = 3. The bottoms product is 10.0 mol% acetone. A partial reboiler is used. Find the mole fraction ethanol in the side stream x_s, the optimum feed location, and the total number of equilibrium contacts needed. Equilibrium data are available in Problem 4.D7.



10 equilibrium contacts More than sufficient
(9 + PR)
Need ~ $8\frac{1}{3}$ + PR



R = 3

x_o?

F = B + D + O

$Fx_F = Bx_b + Dx_d + O\overset{?}{x_o}$

D
$x_d = 0.9$

O = 15 Kmol
q = 1
x_o = ?

F = 100
$x_f = 0.6$
q = .8

$\bar{V}$

B $x_b = 0.1$

① $y = \frac{3}{4}x + 0.225$ Rec-Line

② $y = -4x + 3$ q Line

From graph $x_o = 0.76$ $L' = L - O$

→ D = 50.125 B = 34.875 → L = 150.375 → L' = 135.37

Middle Line $y = \frac{L'}{V'}x + \frac{Dx_d + Ox_o}{V'}$

$V' = V = L + D = 200.5$ → $y = 0.67x + 0.28$

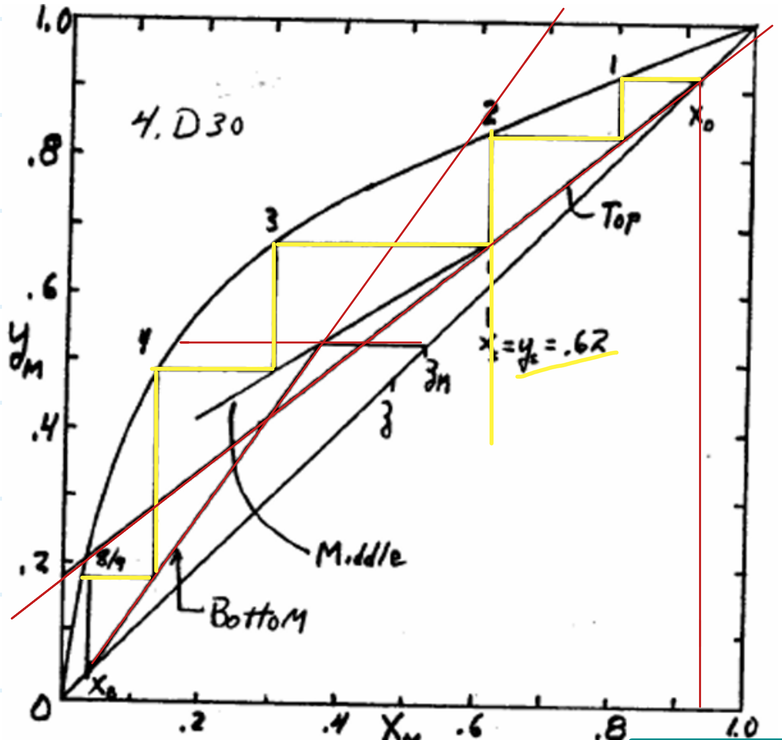
D32.* A distillation column is separating acetone from ethanol. Feed is a saturated liquid that is 40 mol% acetone. Feed rate is 50 kmol/h. Operation is at 1 atm and CMO can be assumed. The column has a total condenser and a partial reboiler. There are eight equilibrium stages in the column, and the feed is on the third stage above the reboiler. Three months ago the distillate flow was shut off (D = 0), but the column kept running. The boilup ratio was set at the value of $\overline{V}/B$ = 1.0. Equilibrium data are given in Problem 4.D7. What is x_B?

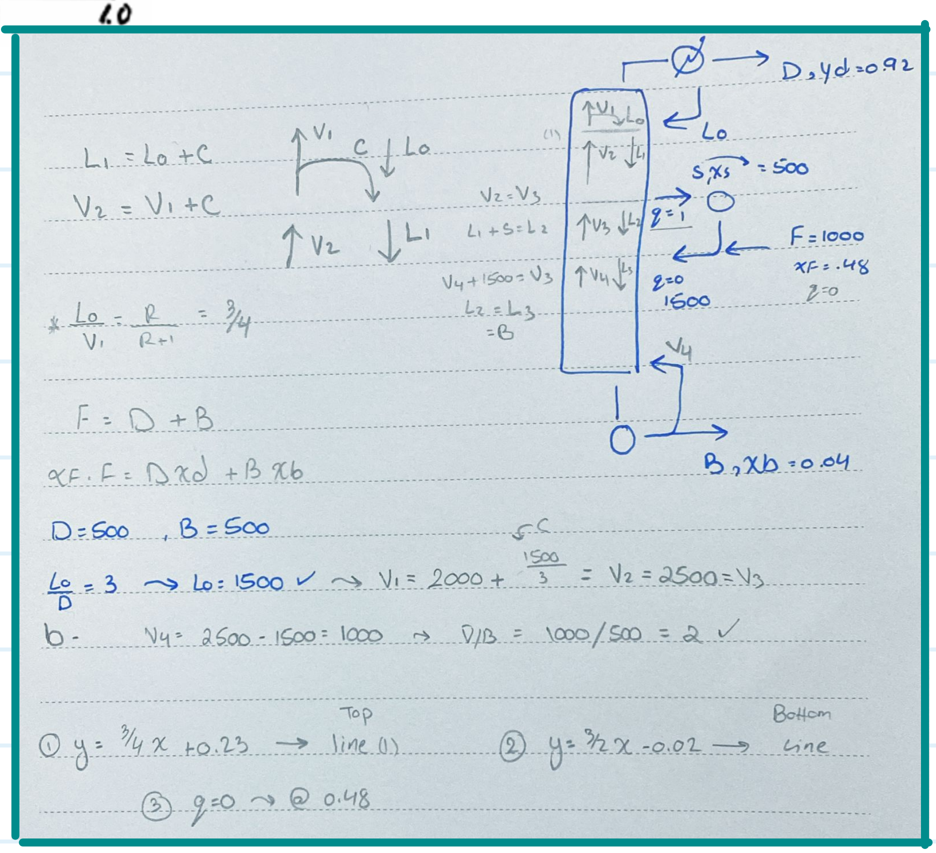
External Balance: $F = B = 50$, and $Fz = Bx_B$. Thus $x_B = z = 0.4$.

D30.* A distillation column is separating methanol from water. The column has a total condenser that subcools the reflux so that 1 mole of vapor is condensed in the column for each 3 moles of reflux. $L_0/D = 3$. A liquid side stream is withdrawn from the second stage below the condenser. This side stream is vaporized to a saturated vapor and then mixed with the feed and input on stage 4. The side withdrawal rate is S = 500 kmol/h. The feed is a saturated vapor that is 48 mol% methanol. Feed rate is F = 1000 kmol/h. A total reboiler is used, which produces a saturated vapor boilup. We desire a distillate 92 mol% methanol and a bottoms 4 mol% methanol. Assume CMO. Equilibrium data are given in Table 2-7. Find:

a. The total number of equilibrium stages required.

b. The value of $\overline{V}/B$.





$L_1 = L_0 + C$

$V_2 = V_1 + C$

$V_2 = V_3$

$L_1 + S = L_2$

$V_4 + 1500 = V_3$

$L_2 = L_3 = B$

$* \frac{L_0}{V_1} = \frac{R}{R+1} = 3/4$

$F = D + B$

$x_F \cdot F = D x_d + B x_b$

$D = 500$, $B = 500$

$\frac{L_0}{D} = 3 \rightarrow L_0 = 1500$ ✓ $\rightarrow V_1 = 2000 + \overset{\curvearrowleft C}{\frac{1500}{3}} = V_2 = 2500 = V_3$

b- $V_4 = 2500 - 1500 = 1000 \rightarrow \bar{V}/B = 1000/500 = 2$ ✓

① $y = 3/4 x + 0.23 \rightarrow$ Top line (1)

② $y = 3/2 x - 0.02 \rightarrow$ Bottom line

③ $q = 0 \rightarrow$ @ 0.48

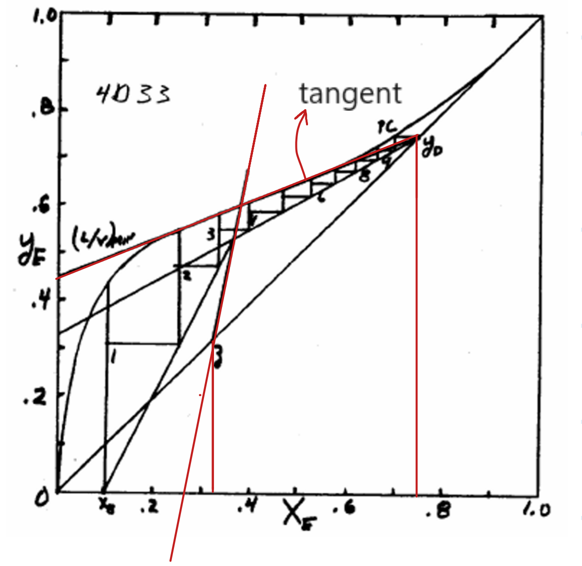
D33.* A distillation column is separating 1000 mol/h of a 32 mol% ethanol, 68 mol% water mixture. The feed enters as a subcooled liquid that will condense 1 mole of vapor on the feed plate for every 4 moles of feed. The column has a partial condenser and uses open steam heating. We desire a distillate product $y_D = 0.75$ and a bottoms product $x_B = 0.10$. CMO is valid. The steam used is pure saturated water vapor. Data are in Table 2-1 and Figure 2-2.

a. Find the minimum external reflux ratio.

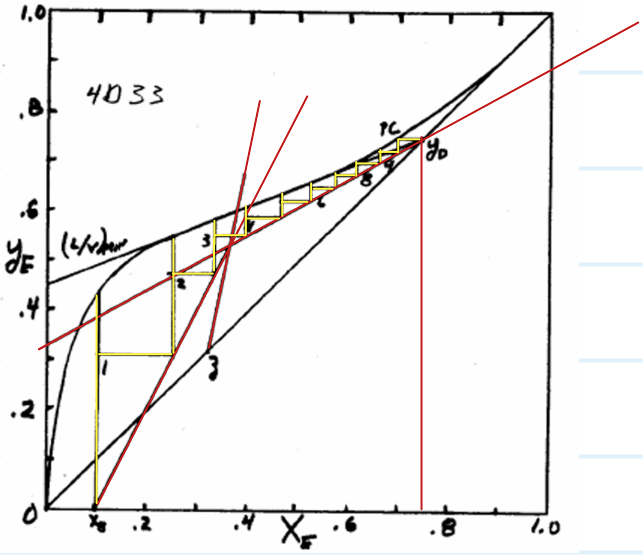
b. Use $L/D = 2.0(L/D)_{min}$, and find the number of real stages and the real optimum feed location if the Murphree vapor efficiency is 2/3 for all stages.

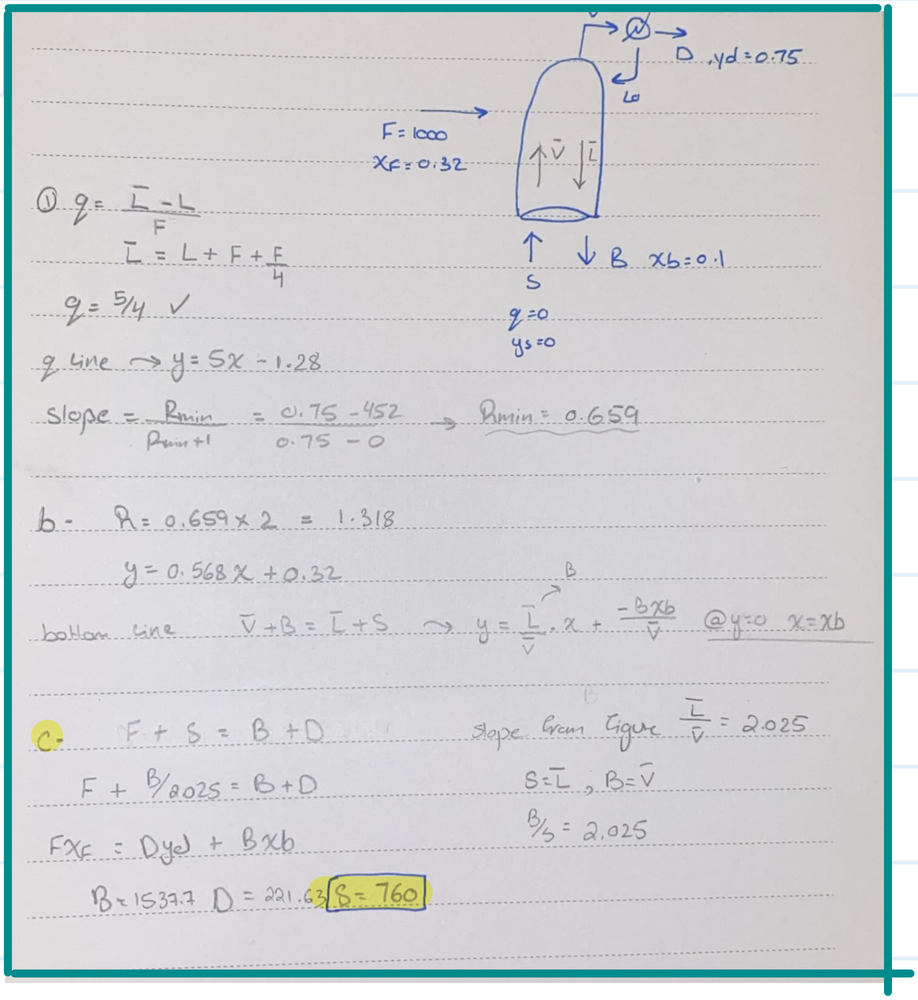
c. Find the steam flow rate used.

a)



b)





① $q = \frac{\bar{L} - L}{F}$

$\bar{L} = L + F + \frac{F}{4}$

$q = 5/4$ ✓

q line → $y = 5x - 1.28$

$\text{slope} = \frac{R_{min}}{R_{min}+1} = \frac{0.75 - .452}{0.75 - 0} \rightarrow R_{min} = 0.659$

b- $R = 0.659 \times 2 = 1.318$

$y = 0.568x + 0.32$

bottom line $\bar{V} + B = \bar{L} + S \rightarrow y = \frac{\bar{L}}{\bar{V}} \cdot x + \frac{-Bx_b}{\bar{V}}$ @ $y = 0$ $x = x_b$

c- $F + S = B + D$

$F + B/2.025 = B + D$

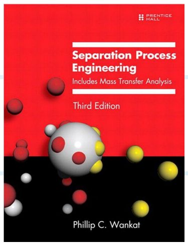
$Fx_F = Dy_d + Bx_b$

$B = 1537.7$ $D = 221.63$ $\boxed{S = 760}$

slope from figure $\frac{\bar{L}}{\bar{V}} = 2.025$

$S = \bar{L}$, $B = \bar{V}$

$B/S = 2.025$



Separation Process Engineering

Includes Mass Transfer Analysis

Third Edition

Phillip C. Wankat

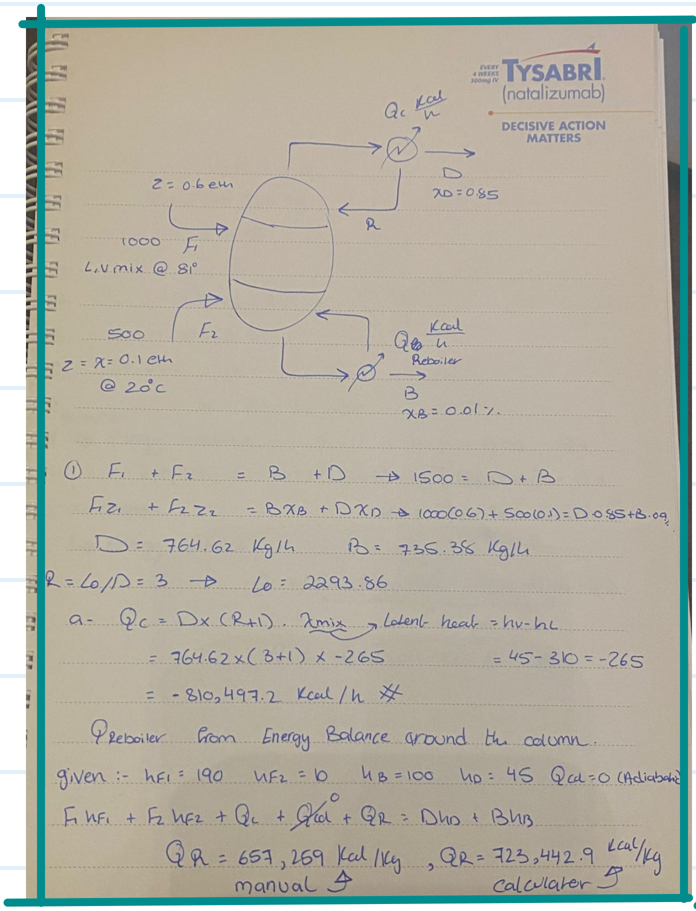
CH3 Different column distillation Cases

1) #Partial Reboiler with total condenser and 2 Feeds

D1. A distillation column with two feeds is separating ethanol from water. The first feed is 60 wt % ethanol, has a total flow rate of 1000 kg/h, and is a mix of liquid and vapor at 81°C. The second feed is 10 wt % ethanol, has a total flow rate of 500 kg/h, and is liquid at 20°C. We desire a bottoms product that is 0.01 wt % ethanol and a distillate product that is 85 wt % ethanol. The column operates at 1 kg/cm^2 and is adiabatic. The column has a partial reboiler, which acts as an equilibrium contact, and a total condenser. The distillate and reflux are saturated liquids. Find B and D in kg/h, and find Q_c and Q_R in kcal/h. Use data in Figure 2-4. Do both parts a and b.

a. External reflux ratio, $L_0/D = 3.0$

b. Boilup ratio, $\overline{V}/B = 2.5$.



① $F_1 + F_2 = B + D \rightarrow 1500 = D + B$

$F_1 z_1 + F_2 z_2 = B x_B + D x_D \rightarrow 1000(0.6) + 500(0.1) = D\,0.85 + B \cdot 0.01\%$

$D = 764.62$ Kg/h $B = 735.38$ Kg/h

$R = L_0/D = 3 \rightarrow L_0 = 2293.86$

a- $Q_c = D \times (R+1) \cdot \lambda_{mix}$ → Latent heat $= h_v - h_L$

$= 764.62 \times (3+1) \times -265$ $= 45 - 310 = -265$

$= -810{,}497.2$ Kcal/h #

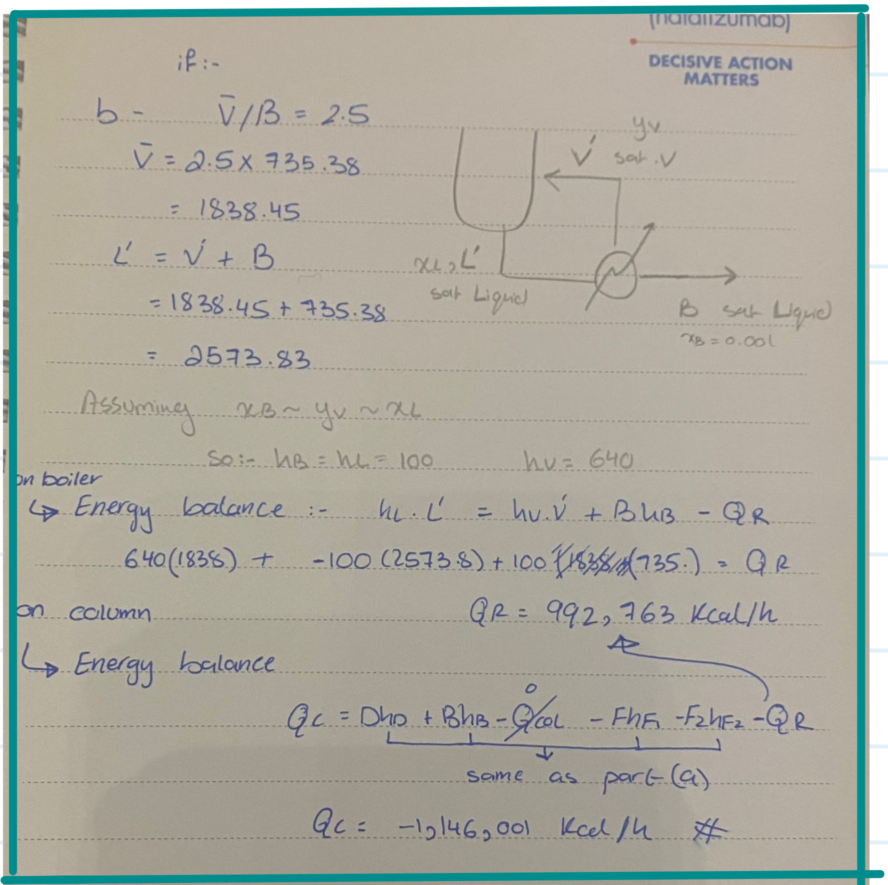
Q Reboiler from Energy Balance around the column.

given :- $h_{F1} = 190$ $h_{F2} = 10$ $h_B = 100$ $h_D = 45$ $Q_{col} = 0$ (Adiabatic

$F_1 h_{F1} + F_2 h_{F2} + Q_c + \overset{0}{\cancel{Q_{col}}} + Q_R = D h_D + B h_B$

$Q_R = 657{,}269$ Kcal/Kg , $Q_R = 723{,}442.9$ Kcal/Kg

manual ↑ calculater ↑



if :-

b- $\bar{V}/B = 2.5$

$\bar{V} = 2.5 \times 735.38$

$= 1838.45$

$L' = V' + B$

$= 1838.45 + 735.38$

$= 2573.83$

Assuming $x_B \sim y_v \sim x_L$

So:- $h_B = h_L = 100$ $h_v = 640$

on boiler

↳ Energy balance :- $h_L \cdot L' = h_v \cdot V' + B h_B - Q_R$

$640(1838) + -100(2573.8) + 100\,\cancel{(1838)}(735.) = Q_R$

on column $Q_R = 992{,}763$ Kcal/h

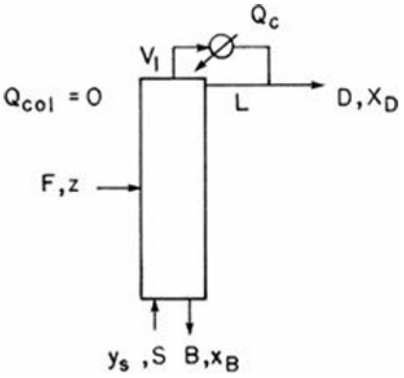
↳ Energy balance

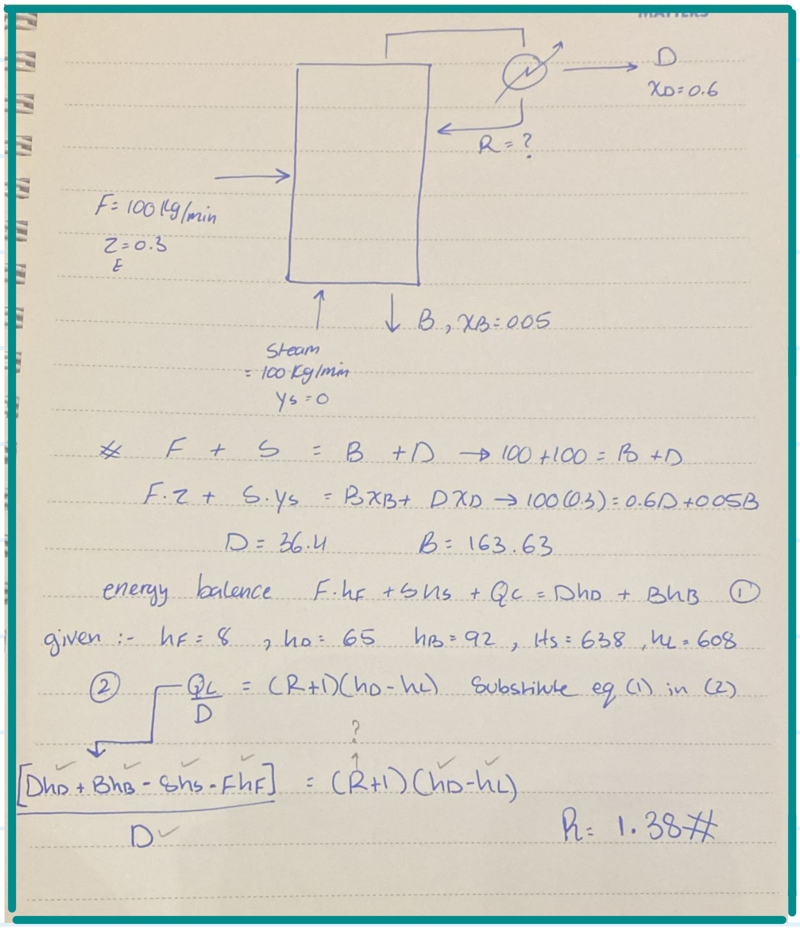
$Q_c = \underbrace{D h_D + B h_B - \overset{0}{\cancel{Q_{col}}} - F h_{F1} - F_2 h_{F2}}_{\text{same as part (a)}} - Q_R$

$Q_c = -1{,}146{,}001$ Kcal/h #

2) #Total condenser and a steam of water vapor instead of reboiler

D2.* A distillation column separating ethanol from water is shown. Pressure is 1 kg/cm^2. Instead of having a reboiler, steam (pure water vapor) is injected directly into the bottom of the column to provide heat. The injected steam is a saturated vapor. The feed is 30 wt % ethanol and is at 20 °C. Feed flow rate is 100 kg/min. Reflux is a saturated liquid. We desire a distillate concentration of 60 wt % ethanol and a bottoms product that is 5 wt % ethanol. The steam is input at 100 kg/min. What is the external reflux ratio, L/D?





D

$x_D = 0.6$

R = ?

F = 100 kg/min

z = 0.3

E

B, x_B = 005

Steam = 100 kg/min

$y_s = 0$

※ $F + S = B + D \rightarrow 100 + 100 = B + D$

$F.z + S.y_s = Bx_B + Dx_D \rightarrow 100(0.3) = 0.6D + 005B$

$D = 36.4 \qquad B = 163.63$

energy balence $F.h_F + Sh_s + Q_c = Dh_D + Bh_B$ ①

given :- $h_F = 8$, $h_D = 65$ $h_B = 92$, $H_s = 638$, $h_L = 608$

② $\frac{Q_L}{D} = (R+1)(h_D - h_L)$ Substitute eq (1) in (2)

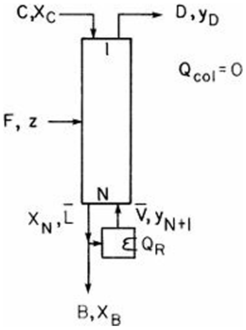
?

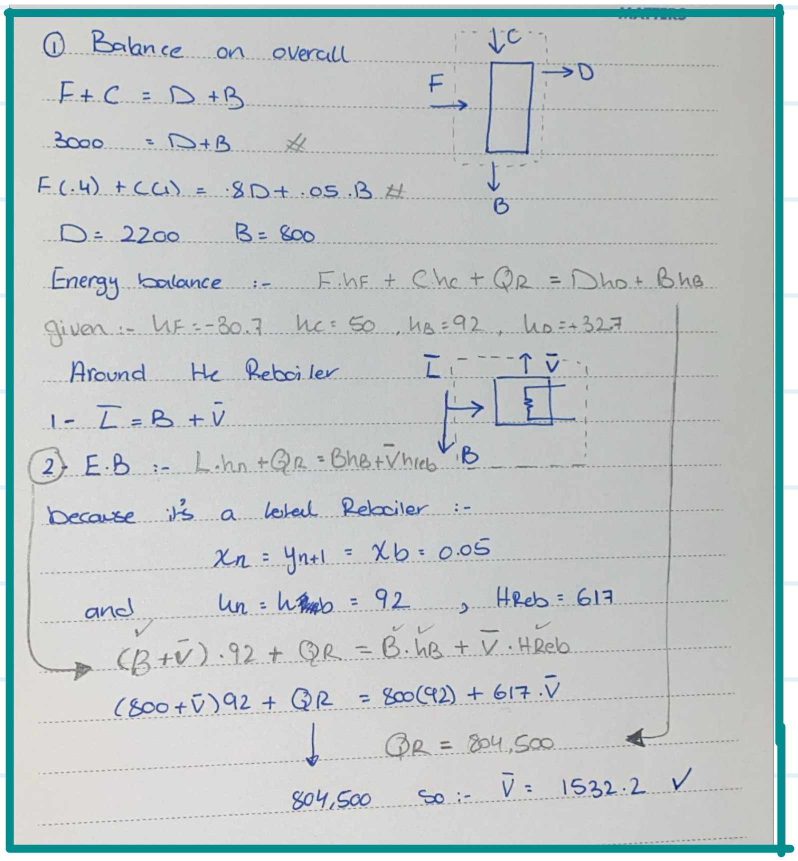
$\frac{[Dh_D + Bh_B - Sh_s - Fh_F]}{D} = (R+1)(h_D - h_L)$

R = 1.38#

3) Total Reboiler and instead of condenser stream of pure ethanol

D3.* A distillation column separating ethanol from water is shown. Pressure is 1 kg/cm^2. Instead of having a condenser, a stream of pure liquid ethanol is added directly to the column to serve as the reflux. This stream is a saturated liquid. The feed is 40 wt % ethanol and is at −20 °C. Feed flow rate is 2000 kg/h. We desire a distillate concentration of 80 wt % ethanol and a bottoms composition of 5 wt % ethanol. A total reboiler is used, and the boilup is a saturated vapor. The cooling stream is input at C = 1000 kg/h. Find the external boilup rate, $\overline{V}$. Note: Set up the equations, solve in equation form for $\overline{V}$ including explicit equations for all required terms, read off all required enthalpies from the enthalpy composition diagram (Figure 2-4), and then calculate a numerical answer.





① Balance on overall

$F + C = D + B$

$3000 = D + B$ #

$F(.4) + C(1) = .8D + .05.B$ #

$D = 2200 \quad B = 800$

Energy balance :- $F.h_F + C h_C + Q_R = D h_D + B h_B$

given :- $h_F = -30.7 \quad h_C = 50, \; h_B = 92, \; h_D = +327$

Around the Reboiler

1- $\overline{L} = B + \overline{V}$

② E.B :- $L.h_n + Q_R = B h_B + \overline{V} h_{reb}$

because it's a total Reboiler :-

$x_n = y_{n+1} = x_b = 0.05$

and $h_n = h_{reb} = 92 \quad , \quad H_{Reb} = 617$

$(B + \overline{V}).92 + Q_R = B.h_B + \overline{V}.H_{Reb}$

$(800 + \overline{V})92 + Q_R = 800(92) + 617.\overline{V}$

$Q_R = 804,500$

804,500 so :- $\overline{V} = 1532.2$ ✓

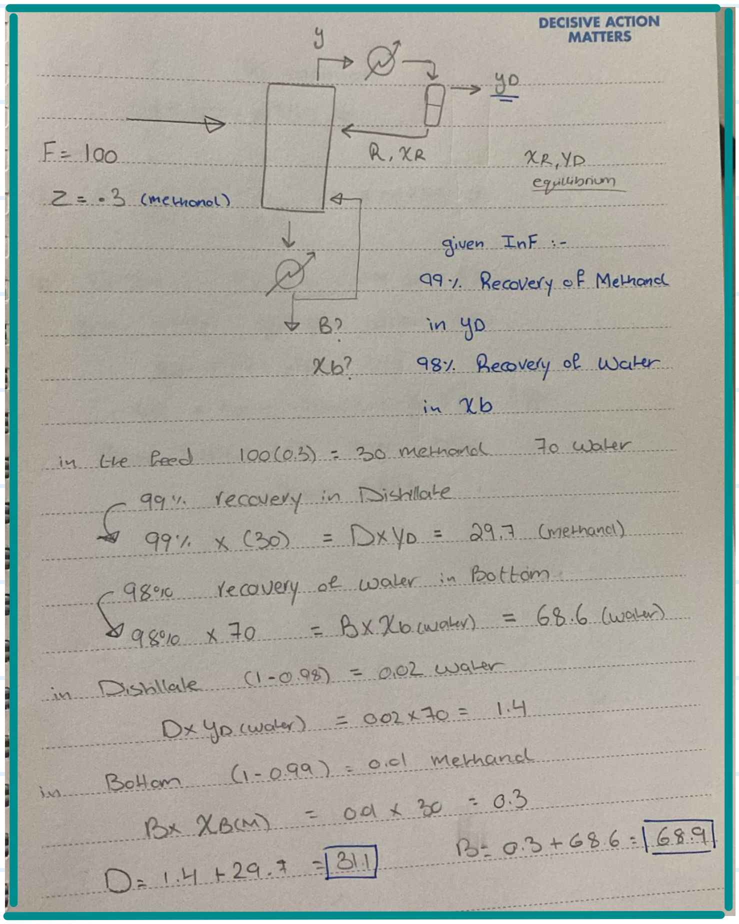
4) Partial condenser , Partial Reboiler

D4. A partial condenser takes vapor leaving the top of a distillation column and condenses a portion of it. The vapor portion of mole fraction y_D is removed as the distillate product. The liquid portion of mole fraction x_0 is returned to the column as reflux. The liquid and vapor leaving the partial condenser can be assumed to be in equilibrium.

A distillation column with a partial condenser and a partial reboiler is separating 100 kmol/h of a mixture that is 30 mol% methanol and 70 mol% water and is a saturated liquid. Column pressure is 1.0 atm. We desire a 99% recovery of the methanol in the vapor distillate and a 98% recovery of water in the bottoms. Equilibrium data are in Table 2-7 (in Problem 2.D1), and other data are in Problem 3.E1.

a. Find D, B, $y_{D,M}$, and $x_{B,M}$.

b. If L/D = 2.0, find $x_{0,M}$ and L_0 where subscript 0 refers to the reflux stream.



F= 100

Z = .3 (methanol)

XR, YD equilibrium

given InF :-

99%. Recovery of Methanol in yD

98%. Recovery of Water in Xb

in the feed 100(0.3) = 30 methanol 70 water

99% recovery in Distillate

99% x (30) = D x YD = 29.7 (methanol)

98% recovery of water in Bottom

98% x 70 = B x Xb (water) = 68.6 (water)

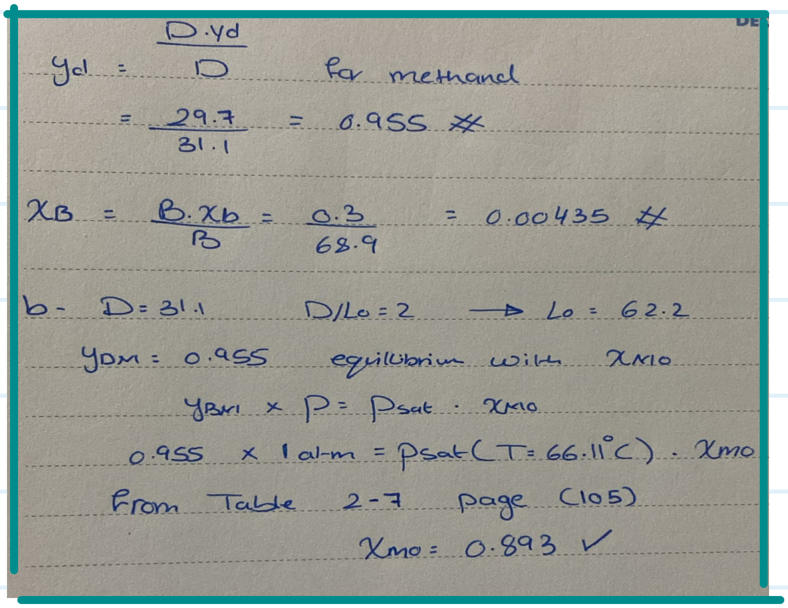
in Distillate (1-0.98) = 0.02 water

D x YD (water) = 0.02 x 70 = 1.4

in Bottom (1-0.99) = 0.01 methanol

B x XB(M) = 0.01 x 30 = 0.3

$D = 1.4 + 29.7 = \boxed{31.1}$ $B = 0.3 + 68.6 = \boxed{68.9}$



$y_d = \frac{D \cdot y_d}{D}$ for methanol

$= \frac{29.7}{31.1} = 0.955$ #

$x_B = \frac{B \cdot x_b}{B} = \frac{0.3}{68.9} = 0.00435$ #

b- D = 31.1 D/Lo = 2 → Lo = 62.2

$y_{DM} = 0.955$ equilibrium with x_{M0}

$y_{BM} \times P = P_{sat} \cdot x_{M0}$

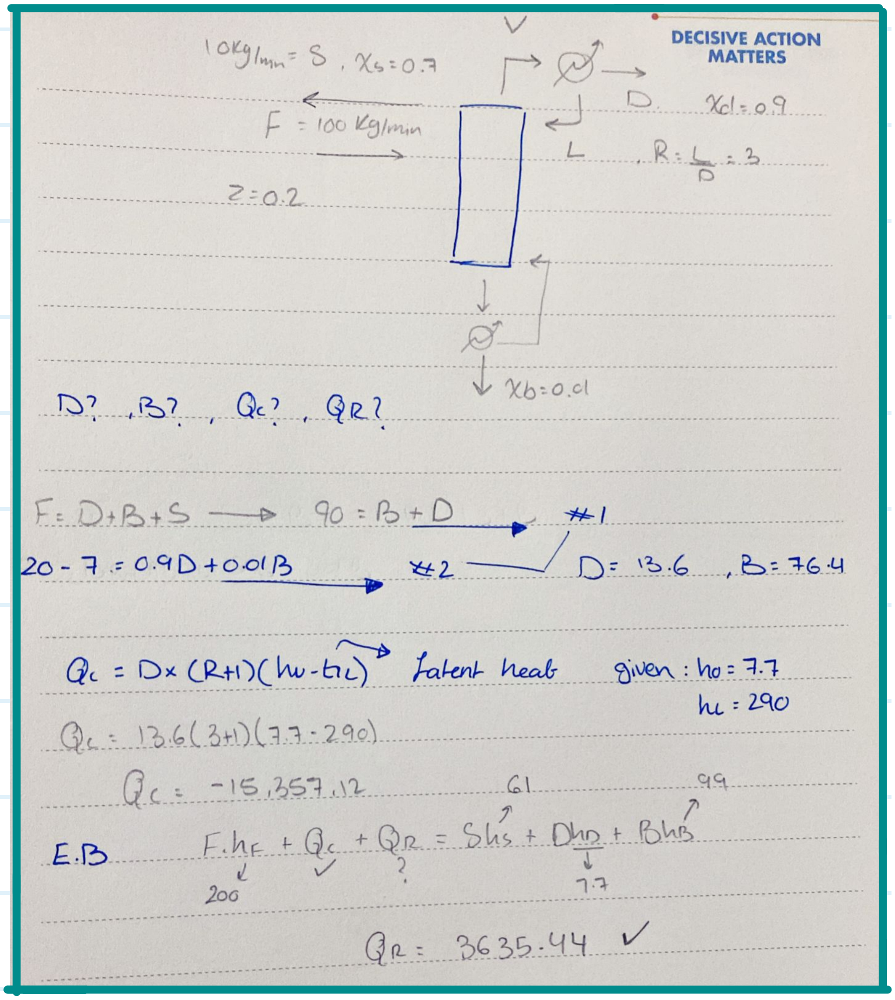
$0.955 \times 1 \text{ atm} = P_{sat}(T = 66.11°C) \cdot x_{m0}$

From Table 2-7 page (105)

$x_{m0} = 0.893$ ✓

5) Total condenser , Partial Reboiler with a liquid side stream

D5.* A distillation column is separating ethanol from water at a pressure of 1 kg/cm^2. A two-phase feed of 20 wt% ethanol at 93°C is input at 100 kg/min. The column has a total condenser and a partial reboiler. The distillate composition is 90 wt % ethanol. Distillate and reflux are at 20°C. Bottoms composition is 1 wt % ethanol. Reflux ratio is $L_0/D = 3$. A liquid side stream is withdrawn above the feed stage. Side stream is 70 wt % ethanol, and side stream flow rate is 10 kg/min. Find D, B, Q_c, and Q_R. Data are in Figure 2-4.



D? , B? , Q_c? , Q_R?

$F = D + B + S \longrightarrow 90 = B + D$ #1

$20 - 7 = 0.9D + 0.01B$ #2

D = 13.6 , B = 76.4

$Q_c = D \times (R+1)(h_v - h_L)$ → latent heat

given : $h_D = 7.7$

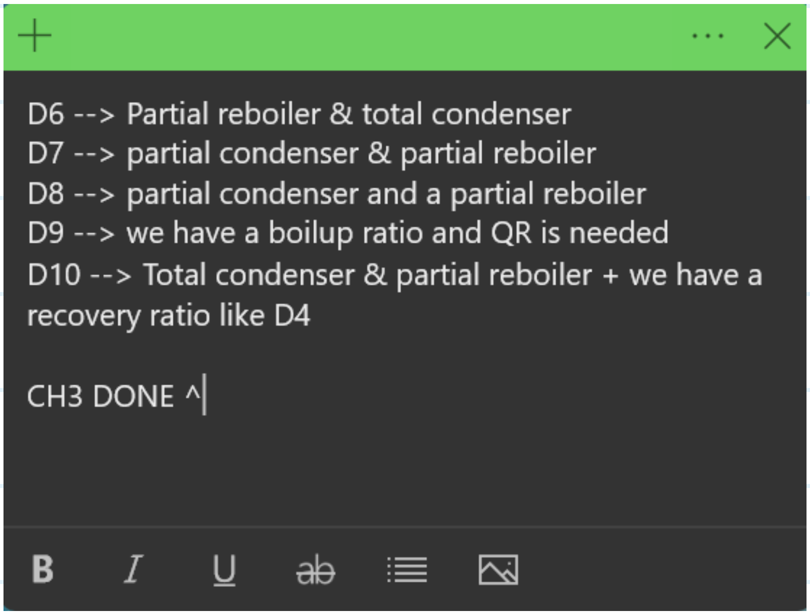
$h_L = 290$

$Q_c = 13.6(3+1)(7.7 - 290)$

$Q_c = -15,357.12$

E.B $\quad \underset{200}{F.h_F} + \underset{✓}{Q_c} + \underset{?}{Q_R} = \overset{61}{Sh_s} + \underset{7.7}{Dh_D} + \overset{99}{Bh_B}$

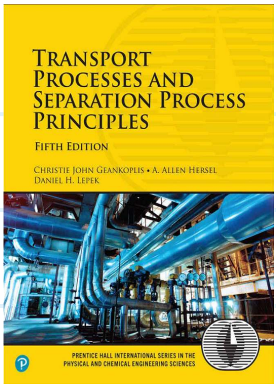
$Q_R = 3635.44$ ✓



D6 --> Partial reboiler & total condenser
D7 --> partial condenser & partial reboiler
D8 --> partial condenser and a partial reboiler
D9 --> we have a boilup ratio and QR is needed
D10 --> Total condenser & partial reboiler + we have a recovery ratio like D4

CH3 DONE ^

26.4 Binary Distillation with Reflux Using the McCabe–Thiele and Lewis Methods


TRANSPORT PROCESSES AND SEPARATION PROCESS PRINCIPLES

FIFTH EDITION

CHRISTIE JOHN GEANKOPLIS • A. ALLEN HERSEL
DANIEL H. LEPEK

PRENTICE HALL INTERNATIONAL SERIES IN THE PHYSICAL AND CHEMICAL ENGINEERING SCIENCES

EXAMPLE 26.4-3. Number of Trays in a Stripping Tower

A liquid feed at the boiling point of 400 kg mol/h containing 70 mol % benzene (*A*) and 30 mol % toluene (*B*) is fed to a stripping tower at 101.3 kPa pressure. The bottoms product flow is to be 60 kg mol/h containing only 10 mol % *A* and the rest *B*. Calculate the kg mol/h overhead vapor, its composition, and the number of theoretical steps required.

Solution: Referring to Fig. 26.4-14a, the known values are $F = 400$ kg mol/h, $x_F = 0.70$, $W = 60$ kg mol/h, and $x_W = 0.10$. The equilibrium data from Table 26.1-1 are plotted in Fig. 26.4-15. Making an overall material balance,

$$F = W + V_D$$

$$400 = 60 + V_D$$

Solving, $V_D = 340$ kg mol/h. Making a component *A* balance and solving,

$$Fx_F = Wx_W + V_D y_D$$

$$400(0.70) = 60(0.10) + 340(y_D)$$

$$y_D = 0.806$$

For a saturated liquid, the *q* line is vertical and is plotted in Fig. 26.4-15. The operating line is plotted through the point $y = x_W = 0.10$ and the intersection of $y_D = 0.806$ with the *q* line. Alternatively, Eq. (26.4-25) can be used, with a slope of $L_m/V_{m+1} = 400/340$. Stepping off the trays from the top, 5.3 theoretical steps or 4.3 theoretical trays plus a reboiler are needed.

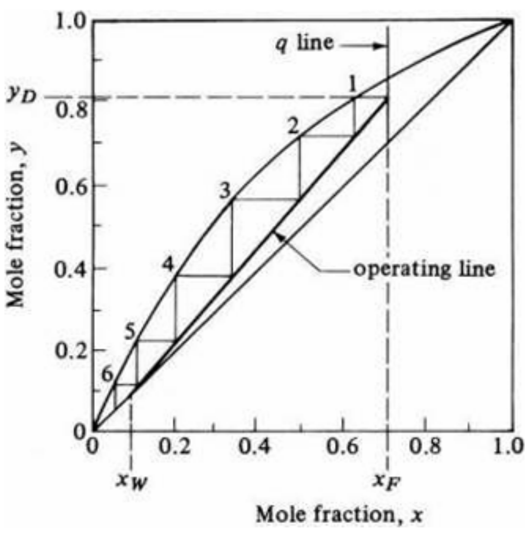
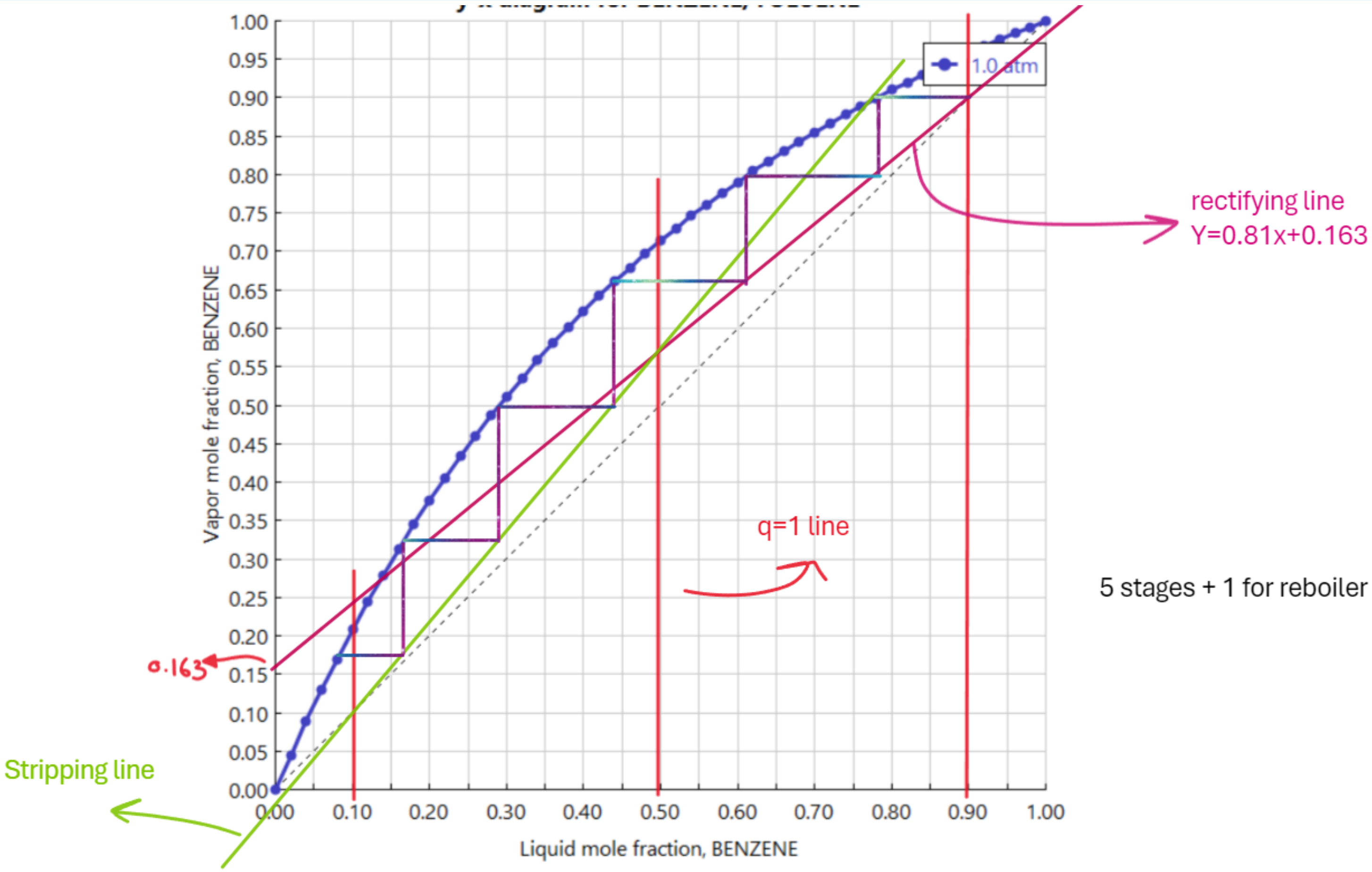


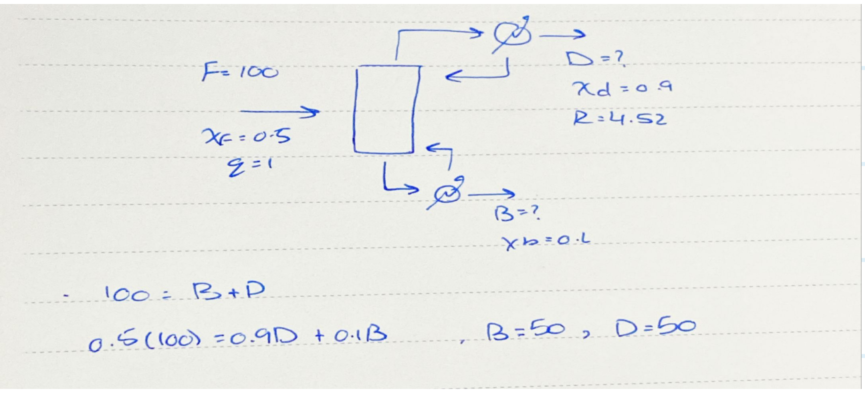
Figure 26.4-15. *Stripping tower for Example 26.4-3.*

26.4-1. *Distillation Using the McCabe–Thiele Method.* A rectification column is fed 100 kg mol/h of a mixture of 50 mol % benzene and 50 mol % toluene at 101.32 kPa abs pressure. The feed is liquid at the boiling point. The distillate is to contain 90 mol % benzene and the bottoms 10 mol % benzene. The reflux ratio is 4.52:1. Calculate the kg mol/h distillate, kg mol/h bottoms, and the number of theoretical trays needed using the McCabe–Thiele method.

Ans. D = 50 kg mol/h, W = 50 kg mol/h, 4.9 theoretical trays plus reboiler

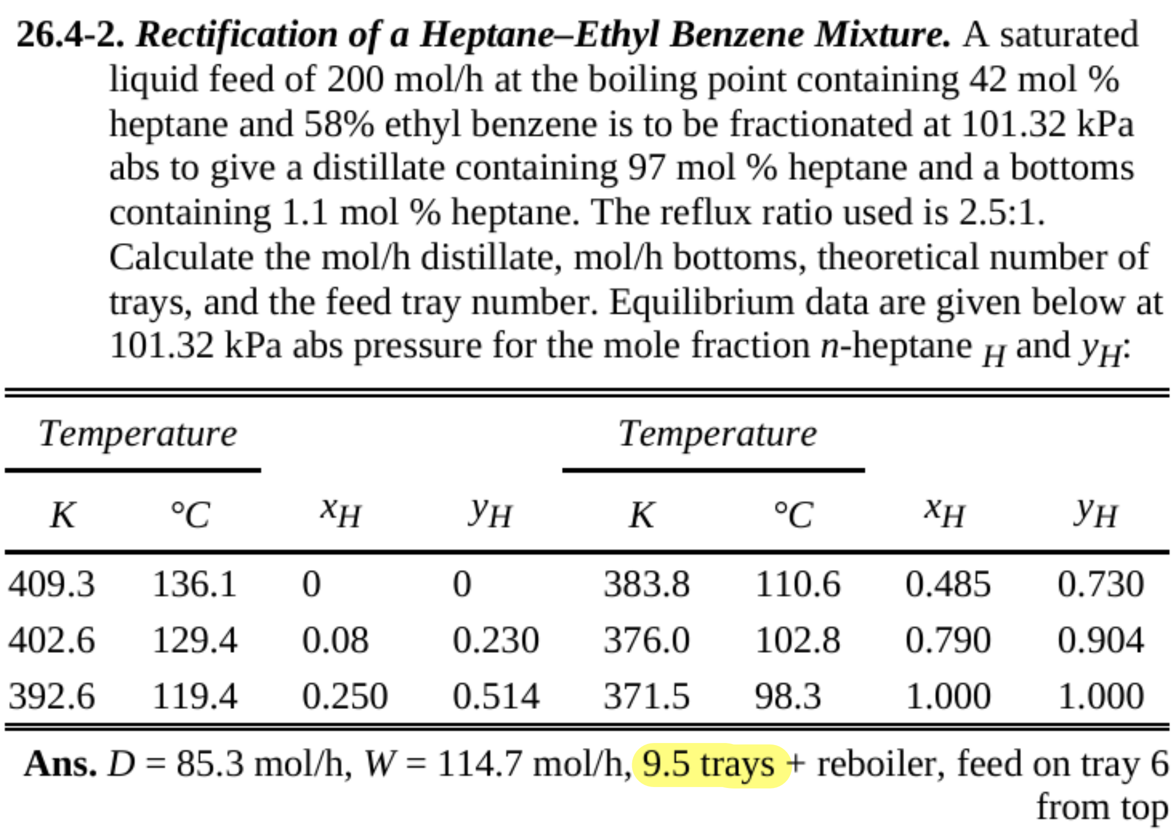


5 stages + 1 for reboiler



100 = B+D

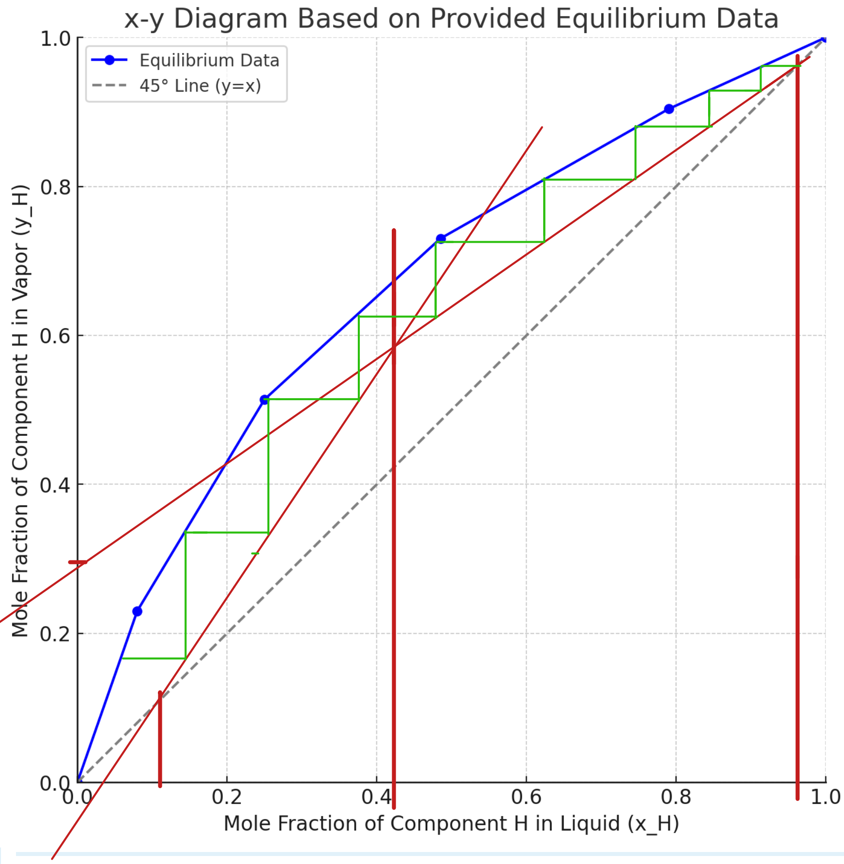
0.5(100) =0.9D + 0.1B , B=50 , D=50

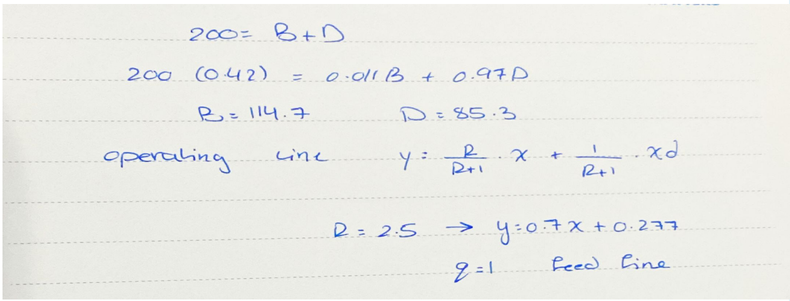


26.4-2. *Rectification of a Heptane–Ethyl Benzene Mixture.* A saturated liquid feed of 200 mol/h at the boiling point containing 42 mol % heptane and 58% ethyl benzene is to be fractionated at 101.32 kPa abs to give a distillate containing 97 mol % heptane and a bottoms containing 1.1 mol % heptane. The reflux ratio used is 2.5:1. Calculate the mol/h distillate, mol/h bottoms, theoretical number of trays, and the feed tray number. Equilibrium data are given below at 101.32 kPa abs pressure for the mole fraction *n*-heptane $_H$ and y_H:

Temperature				*Temperature*			
K	°C	x_H	y_H	K	°C	x_H	y_H
409.3	136.1	0	0	383.8	110.6	0.485	0.730
402.6	129.4	0.08	0.230	376.0	102.8	0.790	0.904
392.6	119.4	0.250	0.514	371.5	98.3	1.000	1.000

Ans. $D = 85.3$ mol/h, $W = 114.7$ mol/h, 9.5 trays + reboiler, feed on tray 6 from top





200= B+D

200 (0.42) = 0.011 B + 0.97 D

B = 114.7 D = 85.3

operating line $y = \frac{R}{R+1} \cdot x + \frac{1}{R+1} \cdot xd$

$R = 2.5 \rightarrow y = 0.7x + 0.277$

$q = 1$ feed line

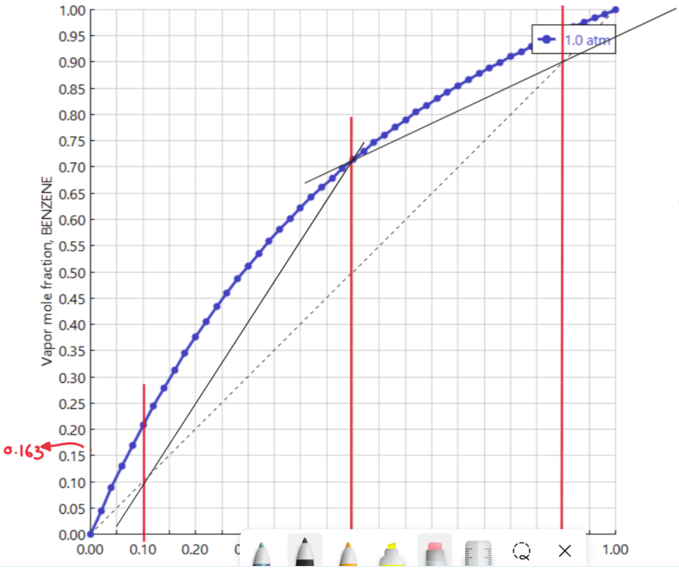
8 trays + 1 for reboiler + feed at stage 6

26.4-3. ***Graphical Solution for a Minimum Reflux Ratio and Total Reflux.*** For the rectification given in Problem 26.4-1, where an equimolar liquid feed of benzene and toluene is being distilled to give a distillate of composition $x_D = 0.90$ and a bottoms of composition $x_W = 0.10$, calculate the following using graphical methods:

a. Minimum reflux ratio R_m

b. Minimum number of theoretical plates at total reflux

Ans. (a) $R_m = 0.91$; (b) 4.0 theoretical trays plus a reboiler

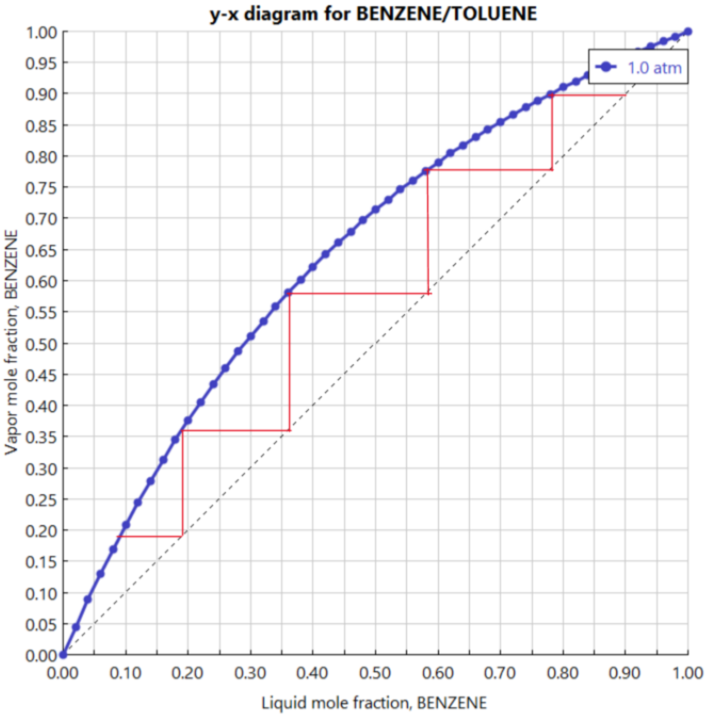


$x_1 = 0.5$
$y_1 = 0.71$

$x_2 = 0.9$
$y_2 = 0.9$

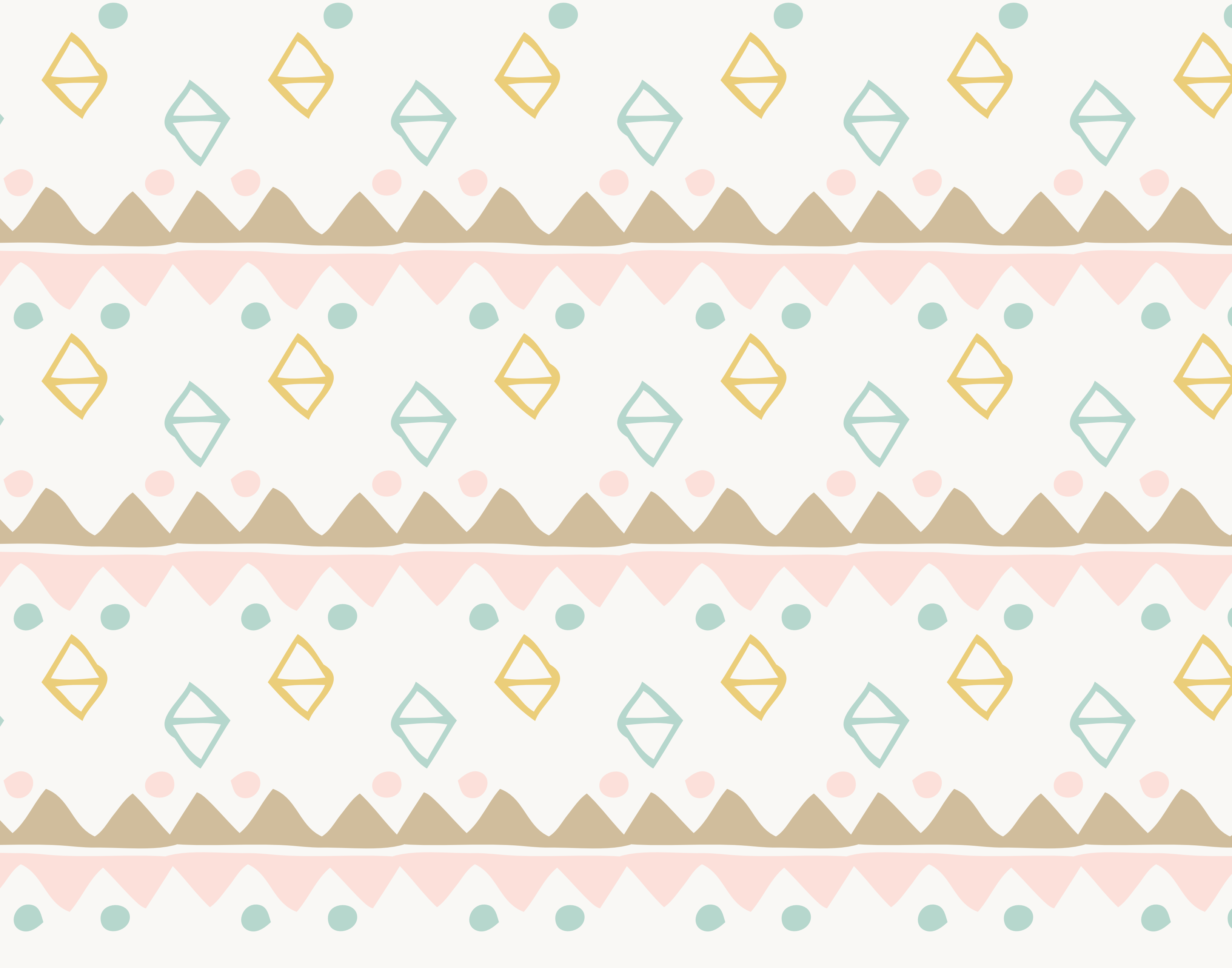
$$\frac{R_{min}}{R_{min}+1} = \frac{y_2 - y_1}{x_2 - x_1}$$

$R_{min} = 0.905$

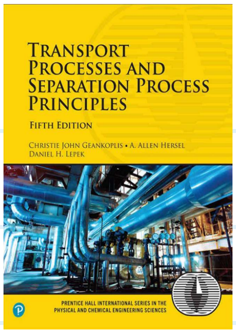


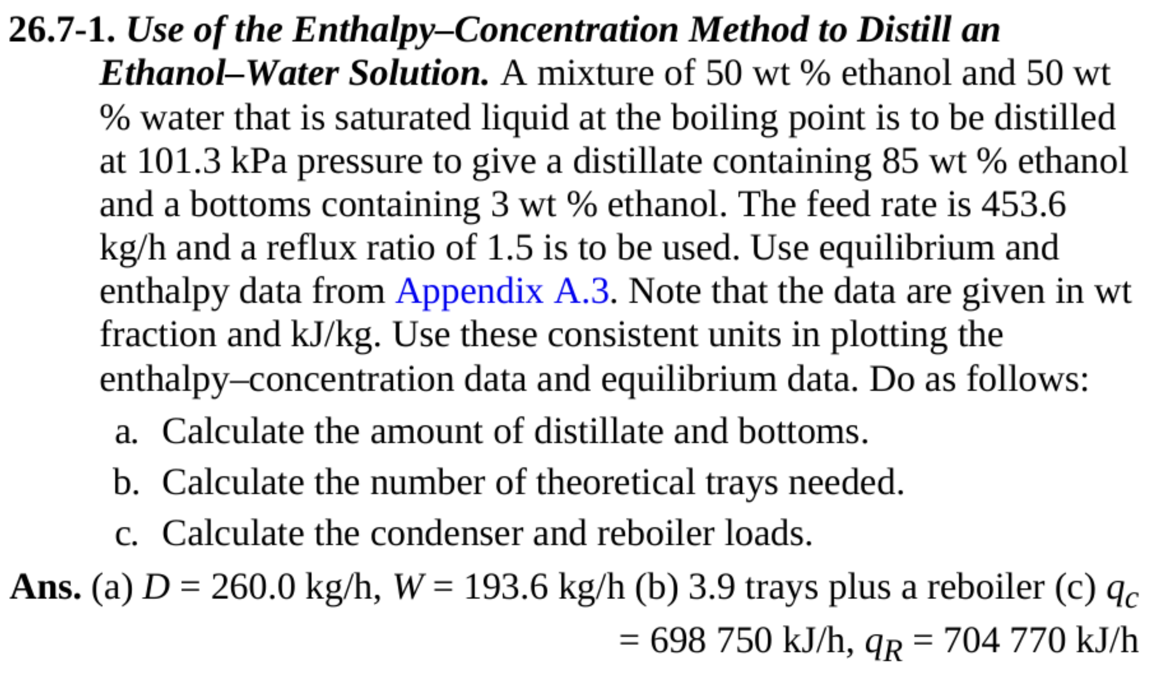
b

5 stages ; 4 trays and 1 for Reboiler



Ponchon savarit

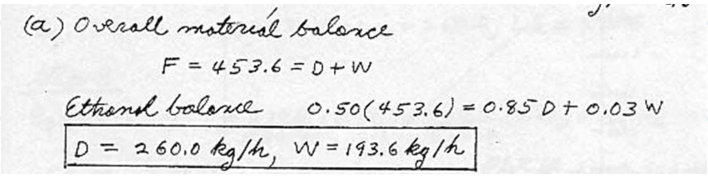




26.7-1. ***Use of the Enthalpy–Concentration Method to Distill an Ethanol–Water Solution.*** A mixture of 50 wt % ethanol and 50 wt % water that is saturated liquid at the boiling point is to be distilled at 101.3 kPa pressure to give a distillate containing 85 wt % ethanol and a bottoms containing 3 wt % ethanol. The feed rate is 453.6 kg/h and a reflux ratio of 1.5 is to be used. Use equilibrium and enthalpy data from Appendix A.3. Note that the data are given in wt fraction and kJ/kg. Use these consistent units in plotting the enthalpy–concentration data and equilibrium data. Do as follows:

a. Calculate the amount of distillate and bottoms.
b. Calculate the number of theoretical trays needed.
c. Calculate the condenser and reboiler loads.

Ans. (a) D = 260.0 kg/h, W = 193.6 kg/h (b) 3.9 trays plus a reboiler (c) q_C = 698 750 kJ/h, q_R = 704 770 kJ/h

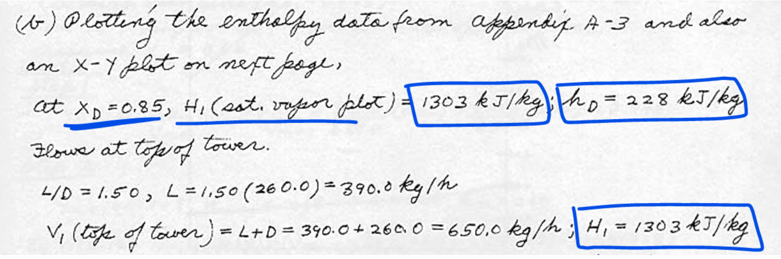


(a) Overall material balance

$F = 453.6 = D + W$

Ethanol balance $0.50(453.6) = 0.85D + 0.03W$

$\boxed{D = 260.0 \text{ kg/h}, \quad W = 193.6 \text{ kg/h}}$



(b) Plotting the enthalpy data from appendix A-3 and also an X-Y plot on next page,

at $x_D = 0.85$, H_1 (sat. vapor plot) = $\boxed{1303 \text{ kJ/kg}}$; $\boxed{h_D = 228 \text{ kJ/kg}}$

Flows at top of tower.

$L/D = 1.50$, $L = 1.50(260.0) = 390.0$ kg/h

V_1 (top of tower) $= L + D = 390.0 + 260.0 = 650.0$ kg/h; $\boxed{H_1 = 1303 \text{ kJ/kg}}$

$$\overset{(1.5)}{R} = \frac{h_D - \overset{1303}{H_V}}{H_V - \underset{228}{h_L}}$$

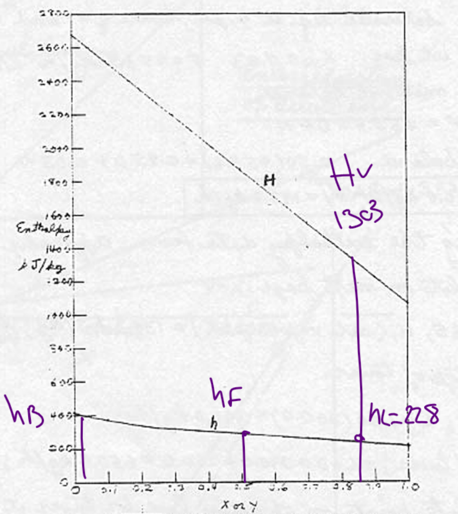
$$h_D = 2915.5 = Q_D + \overset{228}{h_D}$$

$$Q_D = 2687.5 \rightsquigarrow = q_D/D \rightarrow q_D = \underline{698750}$$

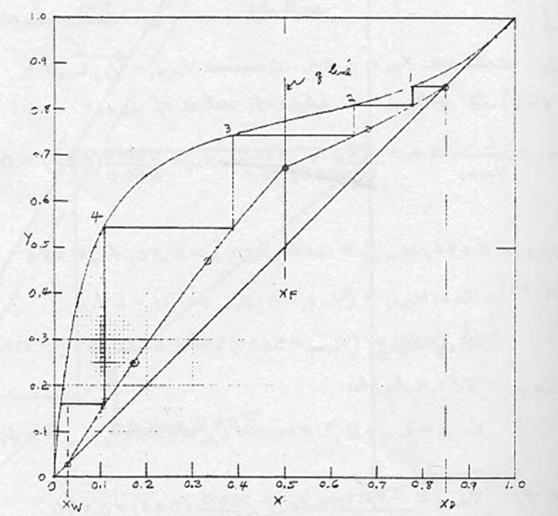
✓ ? ✓ ✓ ✓

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

$$q_B = \underline{704,770}$$



Plot of Enthalpy vs x or y for Prob. 11.6-1



Plot of Operating Lines for Prob. 11.6-1

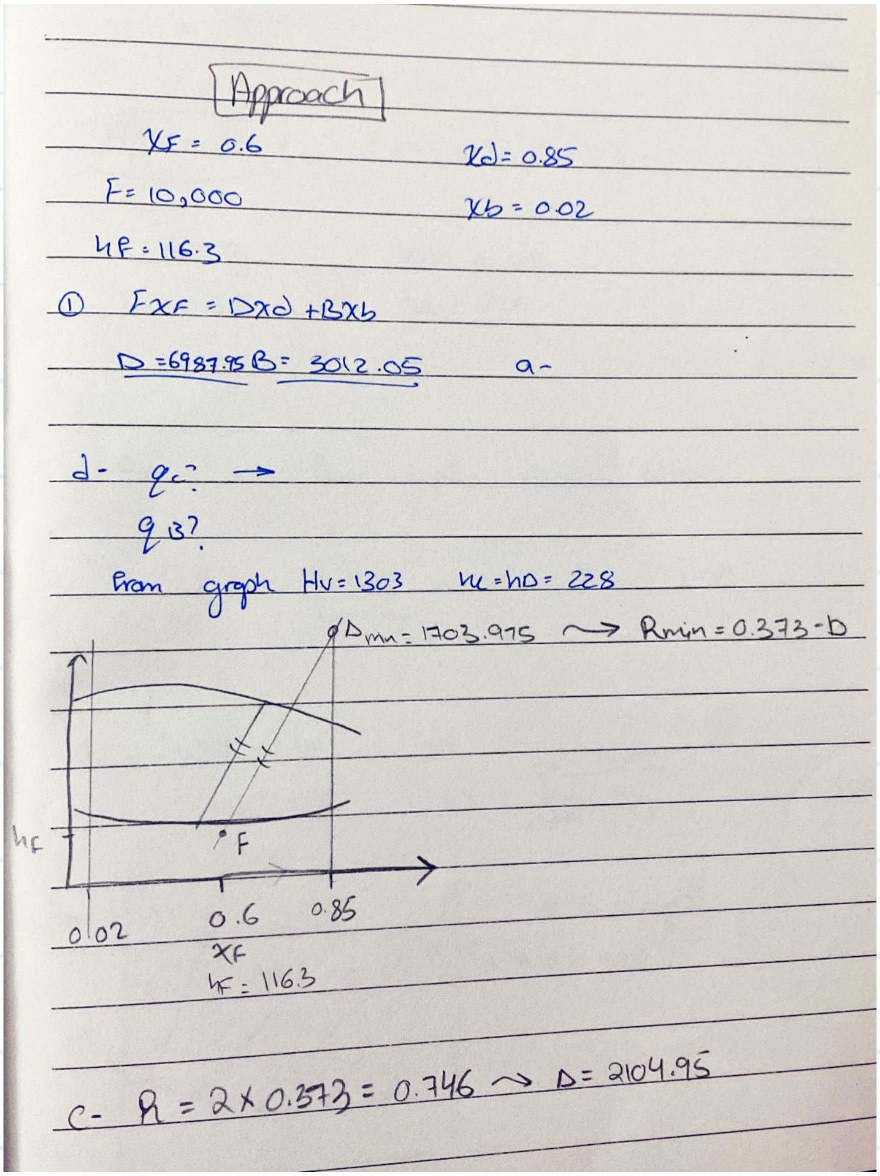
from x-y diagram

\# of trays

26.7-3. *Minimum Reflux and the Theoretical Number of Trays.* A feed of ethanol–water containing 60 wt % ethanol is to be distilled at 101.32 kPa pressure to give a distillate containing 85 wt % ethanol and a bottoms containing 2 wt % ethanol. The feed rate is 10000 kg/h and its enthalpy is 116.3 kJ/kg (50 btu/lb$_m$). Use consistent units of kg/h, weight fraction, and kJ/kg.

a. Calculate the amount of distillate and bottoms.
b. Determine the minimum reflux ratio using enthalpy–concentration data from Appendix A.3.
c. Using 2.0 times the minimum reflux ratio, determine the theoretical number of trays needed.
d. Calculate the condenser and reboiler heat loads.
e. Determine the minimum number of theoretical plates at total reflux.

Ans. (b) R_m = 0.373 (c) 4.4 theoretical trays plus reboiler (d) q_c = 3634 kW, q_R = 4096 kW (e) 2.8 theoretical trays plus reboiler



Approach

$x_F = 0.6$ $\quad$ $x_d = 0.85$

$F = 10{,}000$ $\quad$ $x_b = 0.02$

$h_F = 116.3$

① $F x_F = D x_d + B x_b$

$\underline{D = 6987.95}$ $\underline{B = 3012.05}$ $\quad$ a-

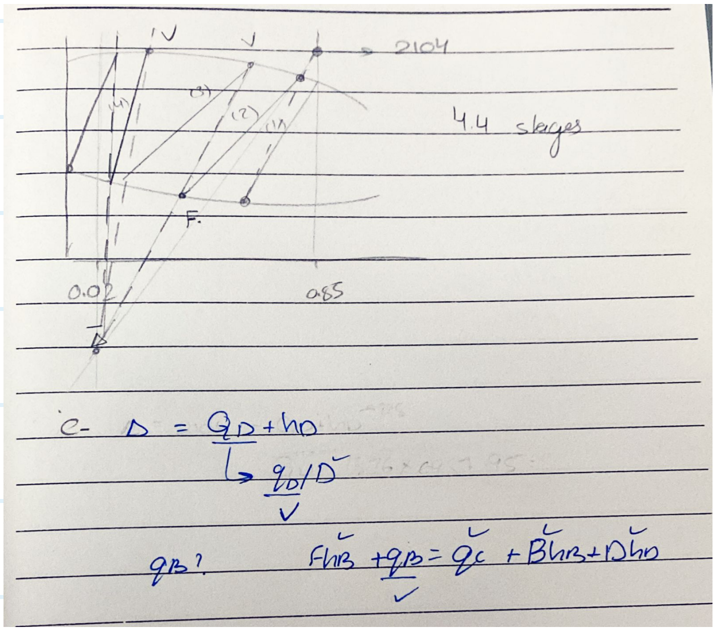
d- q_c? →

q_B?

From graph $H_V = 1303$ $\quad$ $h_L = h_D = 228$

$\Delta_{min} = 1703.975 \rightsquigarrow R_{min} = 0.373$ - b

c- $R = 2 \times 0.373 = 0.746 \rightsquigarrow \Delta = 2104.95$



4.4 stages

c- $\Delta = \underline{Q_D} + h_D$

$\hookrightarrow \underline{q_D / D}$ ✓

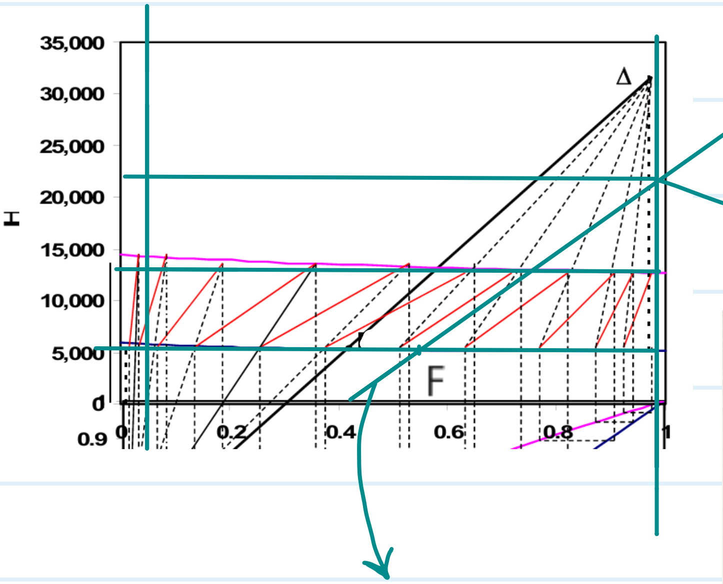
q_B? $\quad$ $F h_B + \underline{q_B} = q_c + B h_B + D h_D$

26.7-4. *Distillation of Benzene–Toluene Feed Using the Enthalpy–Concentration Method.* A liquid feed of 100 kg mol/h of benzene–toluene at the boiling point contains 55 mol % benzene and 45 mol % toluene. It is being distilled at 101.32 kPa pressure to give a distillate with $x_D = 0.98$ and bottoms of $x_W = 0.04$. Using a reflux ratio of 1.3 times the minimum and the enthalpy-concentration method, do as follows:

a. Determine the theoretical number of trays needed.

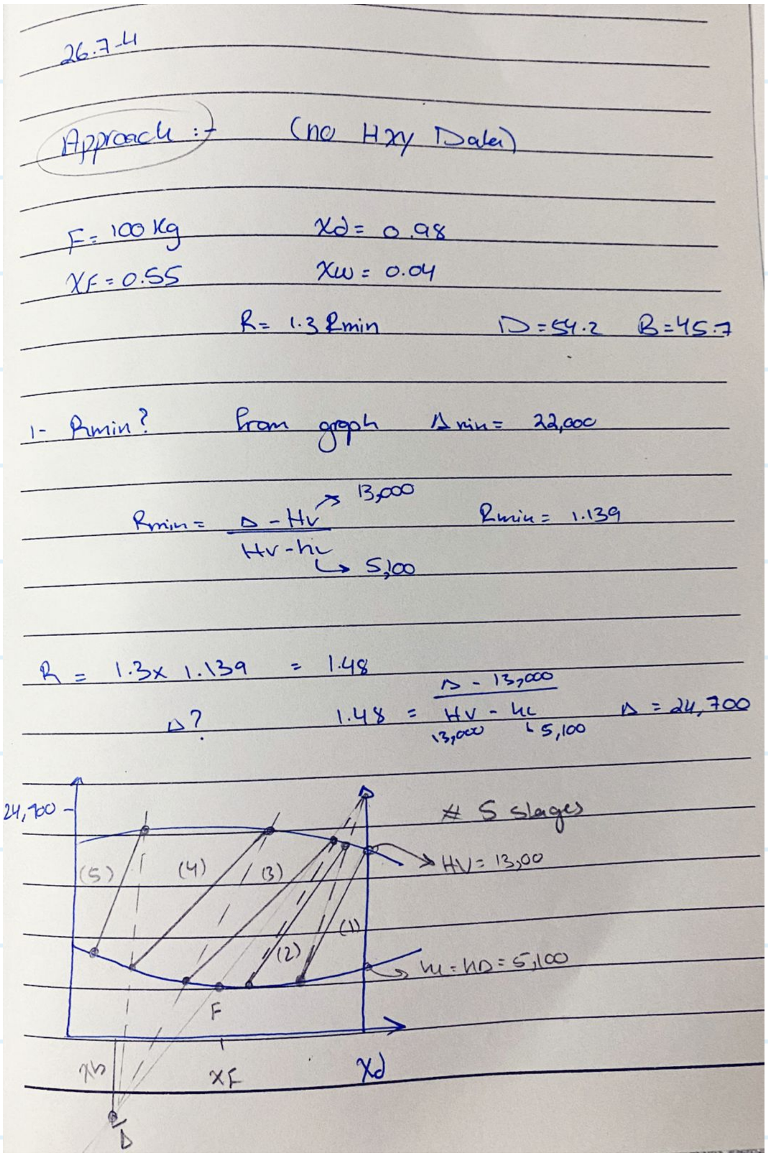
b. Calculate the condenser and reboiler heat loads.

c. Determine the minimum number of theoretical trays at total reflux.



^ min = 22,000

parallel to nearest eq-line



26.7-4

Approach :- (no Hxy Data)

$F = 100$ Kg $x_d = 0.98$

$x_F = 0.55$ $x_w = 0.04$

$R = 1.3 R_{min}$ $D = 54.2$ $B = 45.7$

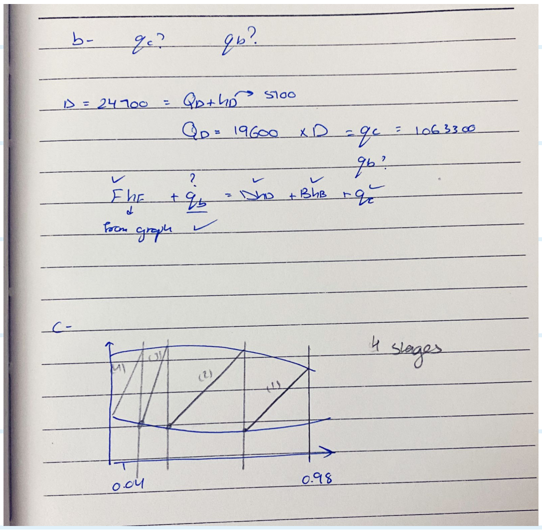
1- R_{min}?   From graph $\Delta_{min} = 22,000$

$$R_{min} = \frac{\Delta - H_V}{H_V - h_L} \quad H_V \rightarrow 13,000 \quad h_L \rightarrow 5,100 \qquad R_{min} = 1.139$$

$R = 1.3 \times 1.139 = 1.48$

Δ? $$1.48 = \frac{\Delta - 13,000}{H_V - h_L} \quad H_V = 13,000 \quad h_L = 5,100 \qquad \Delta = 24,700$$

5 stages



b- q_c? q_b?

$D = 24700 = Q_D + h_D \rightarrow 5100$

$Q_D = 19600 \times D = q_c = 1063300$

q_b?

$$\overset{✓}{Fh_F} + \overset{?}{q_b} = \overset{✓}{Dh_D} + \overset{✓}{Bh_B} + \overset{✓}{q_c}$$

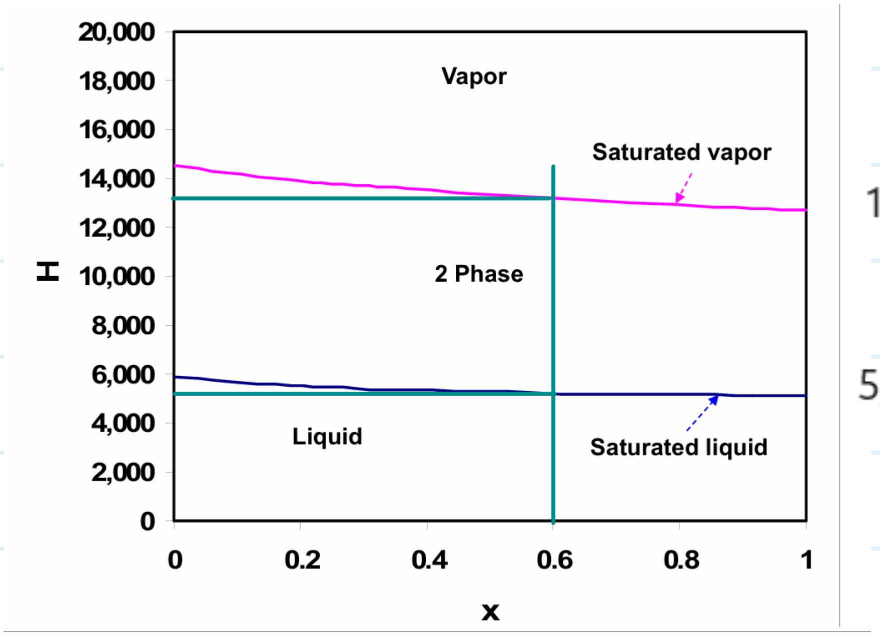
from graph ✓

c-

4 stages

26.7-5. ***Use of an Enthalpy–Concentration Plot.*** For the benzene–toluene system, do as follows:

a. Plot the enthalpy–concentration data using values from Table 26.7-2. For a value of $x = 0.60 = y$, calculate the saturated liquid enthalpy h and the saturated vapor enthalpy H, and plot these data on the graph.



13,200

5,100

EXAMPLE 26.7-2. Distillation Using the Enthalpy–Concentration Method

A liquid mixture of benzene–toluene is being distilled under the same conditions as in Example 26.4-1, except that a reflux ratio of 1.5 times the minimum reflux ratio is to be used. The value $R_m = 1.17$ from Example 26.4-2 will be used. Use enthalpy balances to calculate the flow rates of the liquid and vapor at various points in the tower, and plot the curved operating lines. Determine the number of theoretical stages needed.

Solution: The given data are as follows: $F = 100$ kg mol/h, $x_F = 0.45$, $x_D = 0.95$, $x_W = 0.10$, $R = 1.5R_m = 1.5(1.17) = 1.755$, $D = 41.2$ kg mol/h, $W = 58.8$ kg mol/h. The feed enters at 54.4°C and $q = 1.195$. The flows at the top of the tower are calculated as follows:

$$\frac{L}{D} = 1.755; \qquad L = 1.755(41.2) = 72.3;$$

$$V_1 = L + D = 72.3 + 41.2 = 113.5$$

From Fig. 26.1-1, the saturation temperature at the top of the tower for $y_1 = x_D = 0.95$ is 82.3°C. Using Eq. (26.7-2),

$$H_1 = 0.95[30820 + 96.3(82.3 - 80.1)]$$
$$+ (1 - 0.95)[34224 + 138.2(82.3 - 80.1)] = 31206$$

This value of 31 206 could also have been obtained from the enthalpy–concentration plot, Fig. 26.7-1. The boiling point of the distillate D is obtained from Fig. 26.1-1 and is 81.1°C. Using Eq. (26.7-5),

$$h_D = 0.95(138.2)(81.1 - 80.1) + (1 - 0.95)(167.5)(81.1 - 80.1) = 139$$

Again, this value could have been obtained from Fig. 26.7-1.

Following the procedure outlined for the enriching section for step 1, a value of $x_n = 0.55$ is selected. Assuming a straight operating line for Eq. (26.7-8), an approximate value of y_{n+1} is obtained:

$$y_{n+1} = \frac{72.3}{113.5}x_n + \frac{41.2}{113.5}(0.95) = 0.637(x_n) + 0.345$$
$$= 0.637(0.55) + 0.345 = 0.695$$

Starting with step 2 and using Fig. 26.7-1, for $x_n = 0.55$, $h_n = 1590$, and for $y_{n+1} = 0.695$, $H_{n+1} = 33\,240$. Substituting into Eq. (26.7-12) and solving,

$$V_{n+1}(33240) = (V_{n+1} - 41.2)1590 + 113.5(31206) - 72.3(139)$$
$$V_{n+1} = 109.5$$

Using Eq. (26.7-6),

$$109.5 = L_n + 41.2 \quad \text{or} \quad L_n = 68.3$$

For step 3, substituting into Eq. (26.7-8),

$$y_{n+1} = \frac{68.3}{109.5}(0.55) + \frac{41.2}{109.5}(0.95) = 0.700$$

This calculated value of $y_{n+1} = 0.700$ is sufficiently close to the approximate value of 0.695 that no further trials are needed.

Selecting another value for $x_n = 0.70$ and, using Eq. (26.7-8), an approximate value of y_{n+1} is calculated:

$$y_{n+1} = \frac{72.3}{113.5}(0.70) + \frac{41.2}{113.5}(0.95) = 0.791$$

Using Fig. 26.7-1 for $x_n = 0.70$, $h_n = 1000$, and for $y_{n+1} = 0.791$, $H_{n+1} = 32\,500$. Substituting into Eq. (26.7-12) and solving,

$$V_{n+1}(32500) = (V_{n+1} - 41.2)1000 + 113.5(31206) - 72.3(139)$$
$$V_{n+1} = 110.8$$

Using Eq. (26.7-6),

$$L_n = 110.8 - 41.2 = 69.6$$

Substituting into Eq. (26.7-8),

$$y_{n+1} = \frac{69.6}{110.8}(0.70) + \frac{41.2}{110.8}(0.95) = 0.793$$

In Fig. 26.7-3, the points for the curved operating line in the enriching section are plotted. This line is approximately straight and is very slightly above that for constant molal overflow.

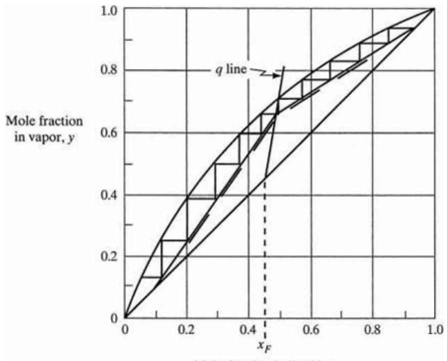


Figure 26.7-3. *Plot of curved operating lines using enthalpy–concentration method for Example 26.7-2. Solid lines are for enthalpy–concentration method and dashed lines are for constant molal overflow.*

Using Eq. (26.7-10), the condenser duty is calculated:

$$q_c = 113.5(31\,206) - 72.3(139) - 41.2(139)$$
$$= 3526\,100 \text{ kJ/h}$$

To obtain the reboiler duty q_R, values for h_W and h_F are needed. Using Fig. 26.7-1 for $x_W = 0.10$, $h_W = 4350$. The feed is at 54.5°C. Using Eq. (26.7-5),

$$h_F = 0.45(138.2)(54.5 - 80.1) + (1 - 0.45)(167.5)(54.5 - 80.1)$$
$$= -3929$$

Using Eq. (26.7-17),

$$q_R = 41.2(139) + 58.8(4350) + 3526100 - 100(-3929) = 4180500 \text{ kJ/h}$$

Using Fig. 26.4-5 and making a material balance below the bottom tray and around the reboiler,

$$L_N = W + V_W \qquad \textbf{(26.7-18)}$$

Rewriting Eq. (26.7-16) for this bottom section,

$$V_W H_W = (V_W + W)h_N + q_R - Wh_W \qquad \textbf{(26.7-19)}$$

From the equilibrium diagram, Fig. 26.1-2, for $x_W = 0.10$, $y_W = 0.207$, which is the vapor composition leaving the reboiler.

For equimolal overflow in the stripping section, using Eqs. (26.4-14) and (26.4-15),

$$L_m = L_n + qF = 72.3 + 1.195(100) = 191.8 \qquad \textbf{(26.4-14)}$$

$$V_{m+1} = V_{n+1} - (1 - q)F = 113.5 - (1 - 1.195)100 = 133.0 \qquad \textbf{(26.4-15)}$$

Selecting $y_{m+1} = y_W = 0.207$, and using Eq. (26.7-15), an approximate value of $x_m = x_N$ is obtained:

$$y_{m+1} = \frac{L_m}{V_{m+1}}x_m - \frac{Wx_W}{V_{m+1}} \qquad \textbf{(26.7-15)}$$
$$0.207 = \frac{191.8}{133.0}(x_N) - \frac{58.8}{133.0}(0.10)$$

Solving, $x_N = 0.174$. From Fig. (26.7-1), for $x_N = 0.174$, $h_N = 3800$, and for $y_W = 0.207$, $H_W = 37\,000$. Substituting into Eq. (26.7-19),

$$V_W(37000) = (V_W + 58.8)(3800) + 4180500 - 58.8(4350)$$

Solving, $V_W = 125.0$. Using Eq. (26.7-18), $L_N = 183.8$. Substituting into Eq. (26.7-15) and solving for x_N,

$$0.207 = \frac{183.8}{125.0}(x_N) - \frac{58.8}{125.0}(0.10)$$
$$x_N = 0.173$$

This value of 0.173 is quite close to the approximate value of 0.174.

Assuming a value of $y_{m+1} = 0.55$ and using Eq. (26.7-15), an approximate value of x_m is obtained:

$$y_{m+1} = 0.55 = \frac{191.8}{133.0}(x_m) - \frac{58.8}{133.0}(0.10)$$
$$x_m = 0.412$$

From Fig. (26.7-1), for $x_m = 0.412$, $h_m = 2300$, and for $y_{m+1} = 0.55$, $H_m = 34400$. Substituting into Eq. (26.7-16),

$$V_{m+1}(34400) = (V_{m+1} + 58.8)(2300) + 4180500 - 58.8(4350)$$

Solving, $V_{m+1} = 126.5$. Using Eq. (26.7-13),

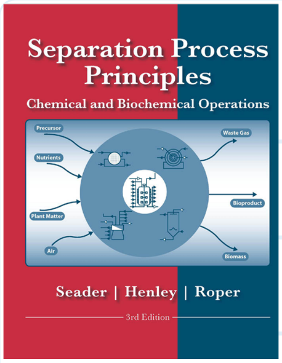
$$L_m = W + V_{m+1} = 58.8 + 126.5 = 185.3$$

Substituting into Eq. (26.7-15) and solving for x_m,

$$y_{m+1} = 0.55 = \frac{185.3}{126.5}x_m - \frac{58.8}{126.5}(0.1)$$
$$x_m = 0.407$$

This value of 0.407 is sufficiently close to the approximate value of 0.412 that no further trials are needed. The two points calculated for the stripping section are plotted in Fig. 26.7-3. This stripping line is also approximately straight and is very slightly above the operating line for constant molal overflow.

Using the operating line for the enthalpy balance method, the number of theoretical steps is 10.4. Using the equimolal method, 9.9 steps are obtained. This difference would be larger if the reflux ratio of 1.5 times R_m were decreased to, say, 1.2 or 1.3. At larger reflux ratios, this difference in number of steps would be less.



7.54. Use of an enthalpy-concentration diagram.

Figure 7.37 is an enthalpy-concentration diagram for *n*-hexane (H), and *n*-octane (O) at 101 kPa. Use this diagram to determine the: (a) mole-fraction composition of the vapor when a liquid containing 30 mol% H is heated from Point A to the bubble-point temperature at Point B; (b) energy required to vaporize 60 mol% of a mixture initially at 100°F and containing 20 mol% H (Point G); and (c) compositions of the equilibrium vapor and liquid resulting from part (b).

Exercise 7.54

Subject: Determining compositions and energy requirements for a mixture of nC_6 and nC_8 with an enthalpy-concentration diagram for 101 kPa.

Given: Enthalpy-concentration diagram in Fig. 7.37.

Find: (a) Composition of vapor at bubble-point for 30 mol% nC_6.
(b) Energy to vaporize 60 mol% of a mixture initially at 100°F with 20 mol% nC_6 .
(c) Vapor and liquid compositions resulting from part (b).

Analysis: (a) In Fig. 7.37, below, the bubble point for 30 mol% nC_6 in nC_8 is shown as point B, which corresponds to a temperature of 211°F. An equilibrium tie line is drawn from point B to point M at the intersection with the saturated vapor line. The vapor composition at point M is 69 mol% nC_6 and 31 mol% nC_8.

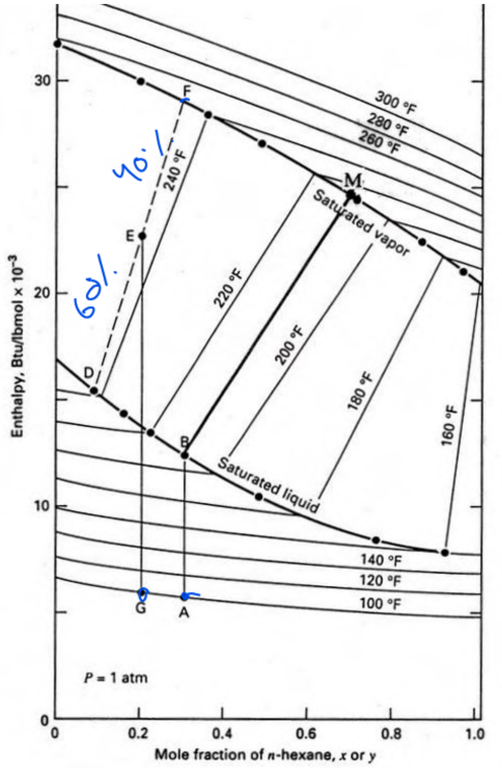
(b) From Fig. 7.37, reproduced below, the enthalpy of the initial mixture of 20 mol% nC_6 at 100°F, shown at point G, is 6,500 Btu/lbmol = H_G . A vertical line is drawn upward from point G until it reaches the point E, located on the dashed tie line between the saturated liquid and saturated vapor curves, at a point that is 60% of the way from the saturated liquid curve to the saturated vapor curve. The enthalpy of the two-phase mixture at point ~~E is 23,000~~ Btu/lbmol = H_E. Therefore, the energy required = $Q = H_E - H_G$ = 23,000 - 6,500 = 16,500 Btu/lbmol.

(c) From the reproduction of Fig. 7.37 below, the equilibrium liquid phase composition is at point D, which is 7.5 mol% nC_6 , while the equilibrium vapor composition is at point F, which is 28.5 mol% nC_6 . Checking this by material balance on nC_6 , for a basis of 100 lbmoles of mixture:

$$0.20(100) = 20 \text{ lbmoles}$$

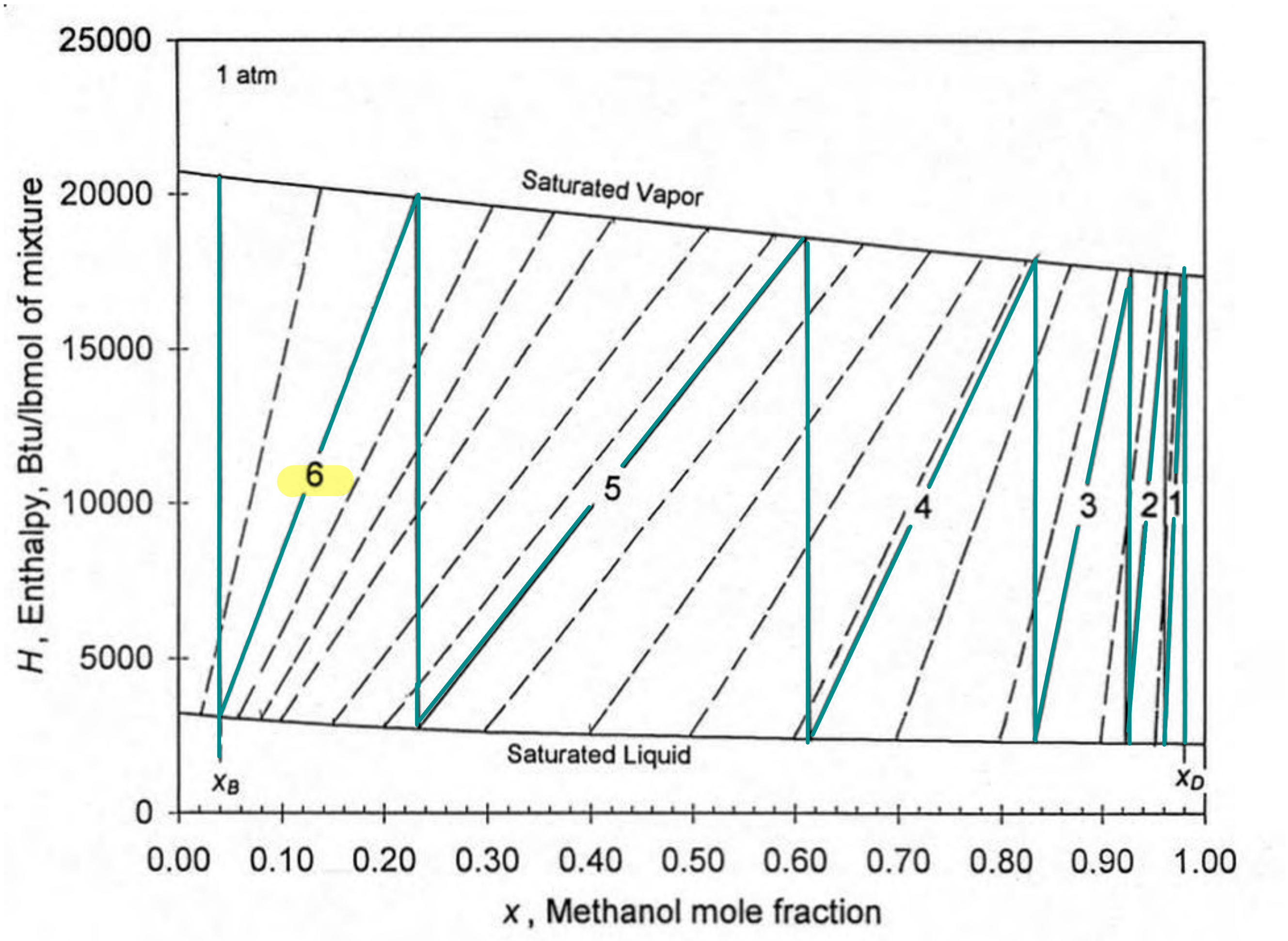
$$0.075(40) + 0.285(60) = 20.1 \text{ lbmoles}$$

This is a reasonably good check.

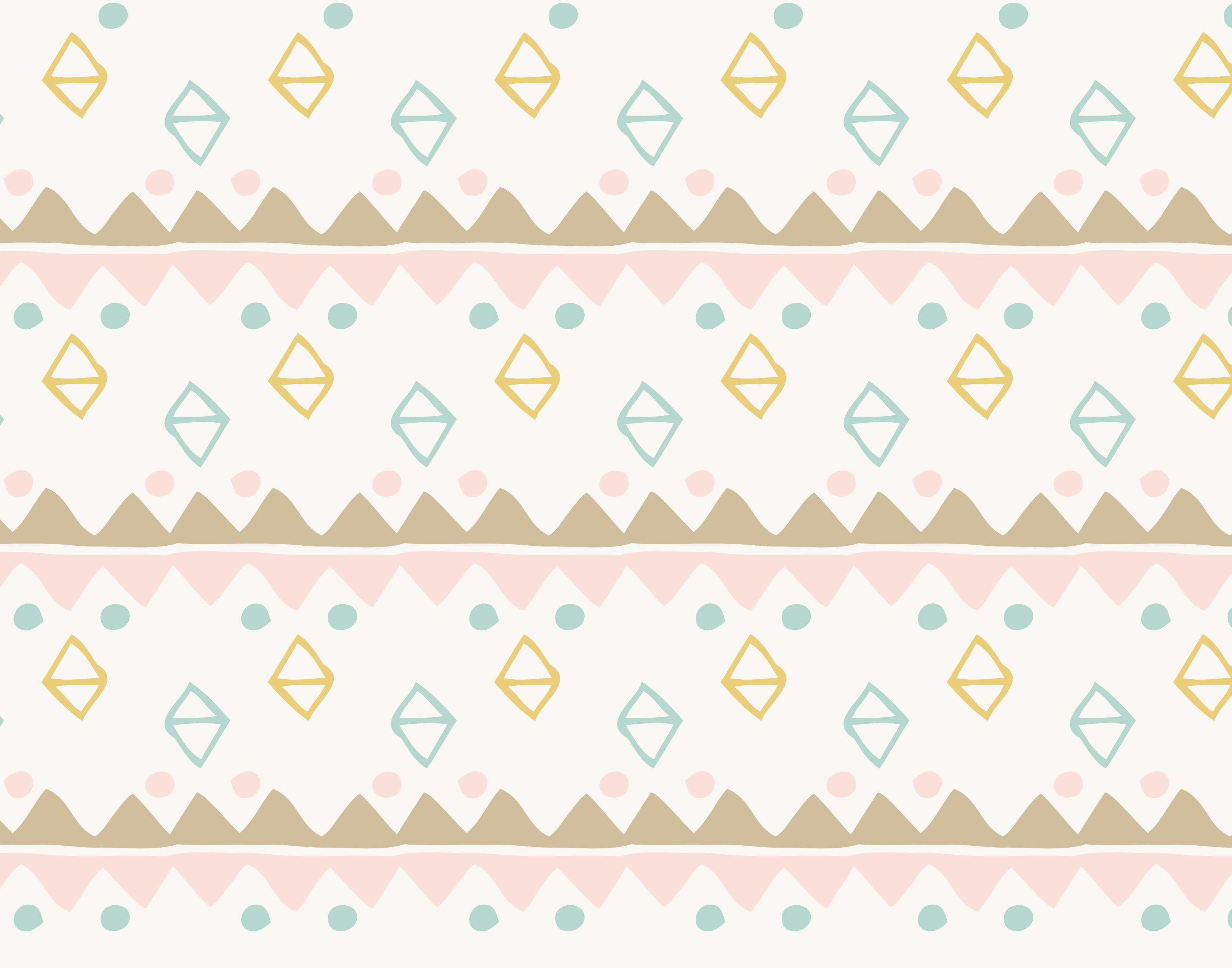


7.56. Plotting an enthalpy-concentration diagram for distillation calculations.

One hundred lbmol/h of 60 mol% methanol in water at 30°C and 1 atm is to be separated by distillation into a liquid distillate containing 98 mol% methanol and a bottoms containing 96 mol% water. Enthalpy and equilibrium data for the mixture at 1 atm are given in Table 7.8. The enthalpy of the feed mixture is 765 Btu/lbmol. (a) Using the given data, plot an enthalpy-concentration diagram. (b) Devise a procedure to determine, from the diagram of part (a), the minimum number of equilibrium stages for the condition of total reflux and the required separation. (c) From the procedure developed in part (b), determine N_{min}. Why is the value independent of the feed condition? (d) What are the temperatures of the distillate and bottoms?



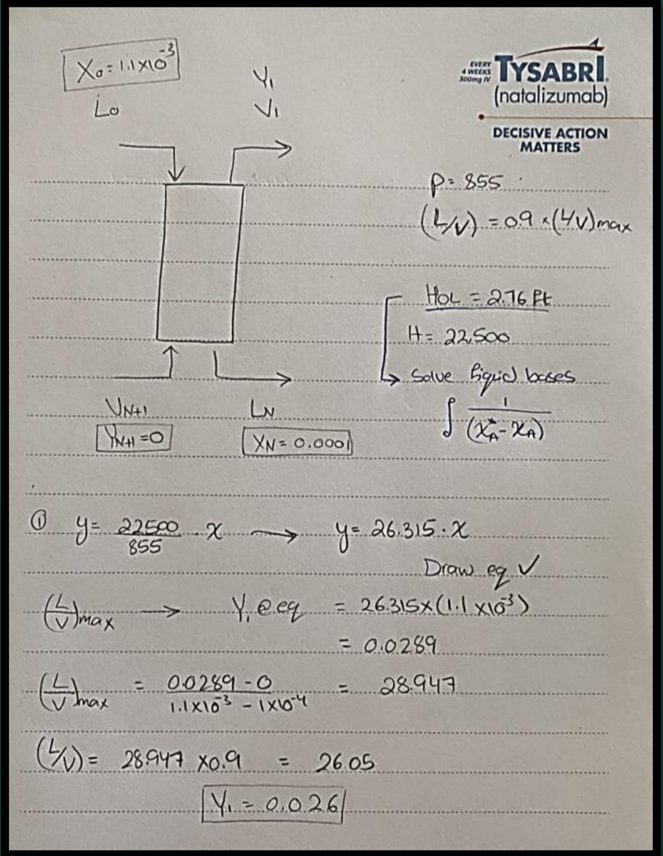
The minimum number of stages is seen to be 6. The value is independent of the feed condition because at total reflux, there is no feed. The temperature of the bottoms is 93.5°C. The temperature of the distillate is 64.7°C



ABSORPTION IN PACKED TOWERS

LEC-8

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.



$\boxed{X_0 = 1.1\times10^{-3}}$

L_0

Y_1

V_1

V_{N+1}

$\boxed{Y_{N+1} = 0}$

L_N

$\boxed{X_N = 0.0001}$

$P = 855$

$(L/V) = 0.9\times(L/V)_{max}$

$H_{OL} = 2.76$ ft

$H = 22{,}500$

→ Solve liquid bases

$\int \frac{1}{(x_A^* - x_A)}$

① $y = \frac{22500}{855} \cdot x \longrightarrow y = 26.315 \cdot x$

Draw eq ✓

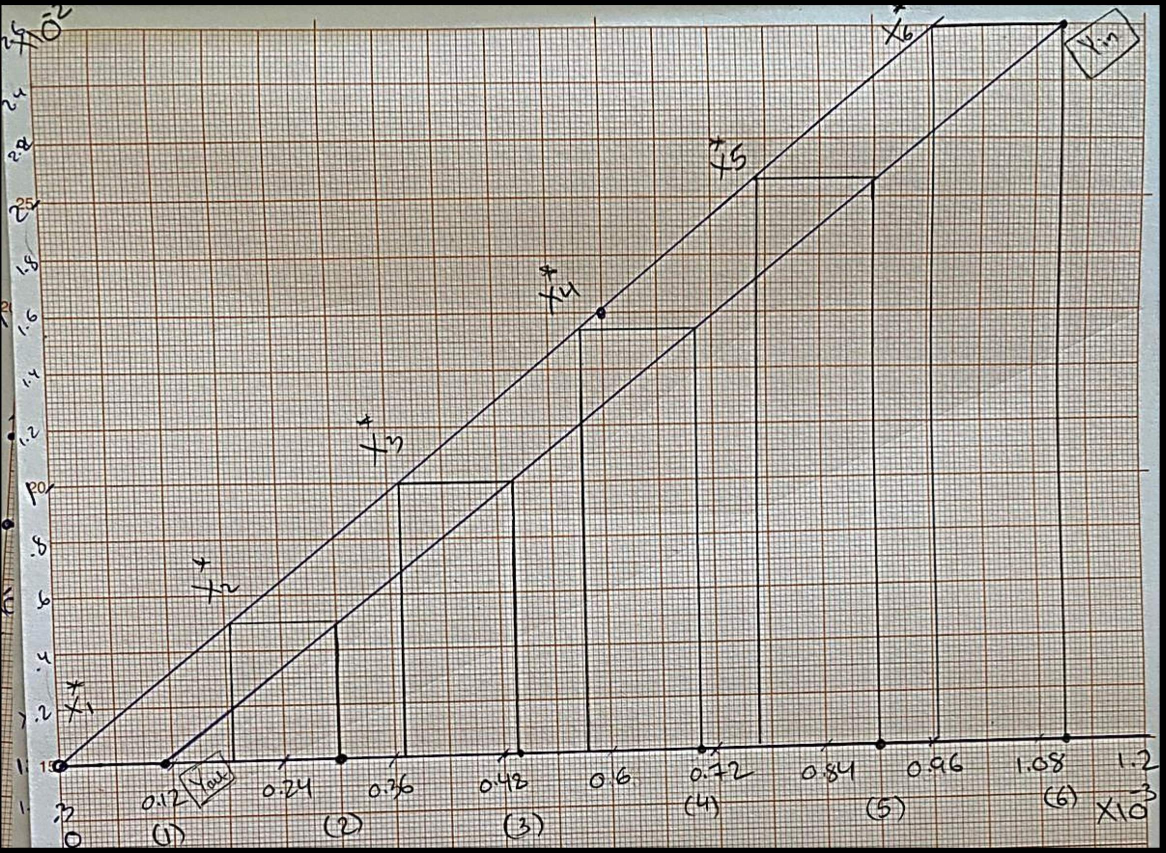
$(\frac{L}{V})_{max} \longrightarrow Y_1 @eq = 26.315\times(1.1\times10^{-3})$

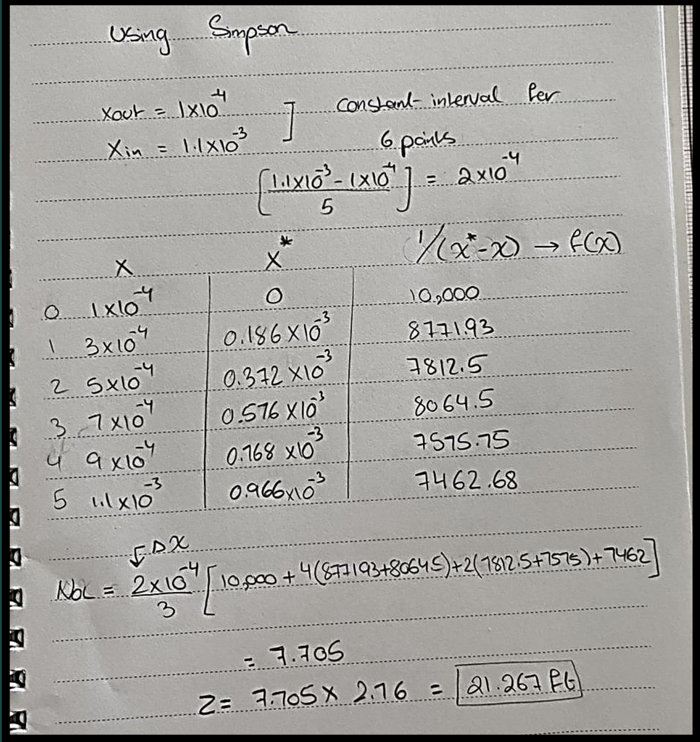
$= 0.0289$

$(\frac{L}{V})_{max} = \frac{0.0289 - 0}{1.1\times10^{-3} - 1\times10^{-4}} = 28.947$

$(L/V) = 28.947 \times 0.9 = 26.05$

$\boxed{Y_1 = 0.026}$





Using Simpson

$x_{out} = 1\times10^{-4}$

$x_{in} = 1.1\times10^{-3}$] constant interval for 6 points

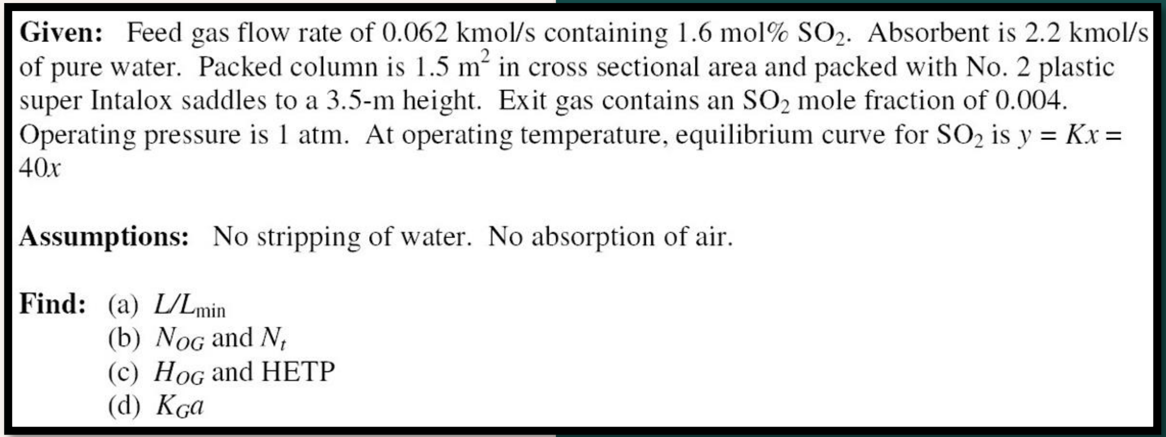
$\left[\frac{1.1\times10^{-3} - 1\times10^{-4}}{5}\right] = 2\times10^{-4}$

	x	x^*	$1/(x^*-x) \rightarrow f(x)$
0	1×10^{-4}	0	10,000
1	3×10^{-4}	0.186×10^{-3}	8771.93
2	5×10^{-4}	0.372×10^{-3}	7812.5
3	7×10^{-4}	0.576×10^{-3}	8064.5
4	9×10^{-4}	0.768×10^{-3}	7575.75
5	1.1×10^{-3}	0.966×10^{-3}	7462.68

$N_{OL} = \frac{\overset{\Delta x}{2\times10^{-4}}}{3}\left[10{,}000 + 4(8771.93 + 8064.5) + 2(7812.5 + 7575) + 7462\right]$

$= 7.705$

$Z = 7.705 \times 2.76 = \boxed{21.267 \text{ ft}}$

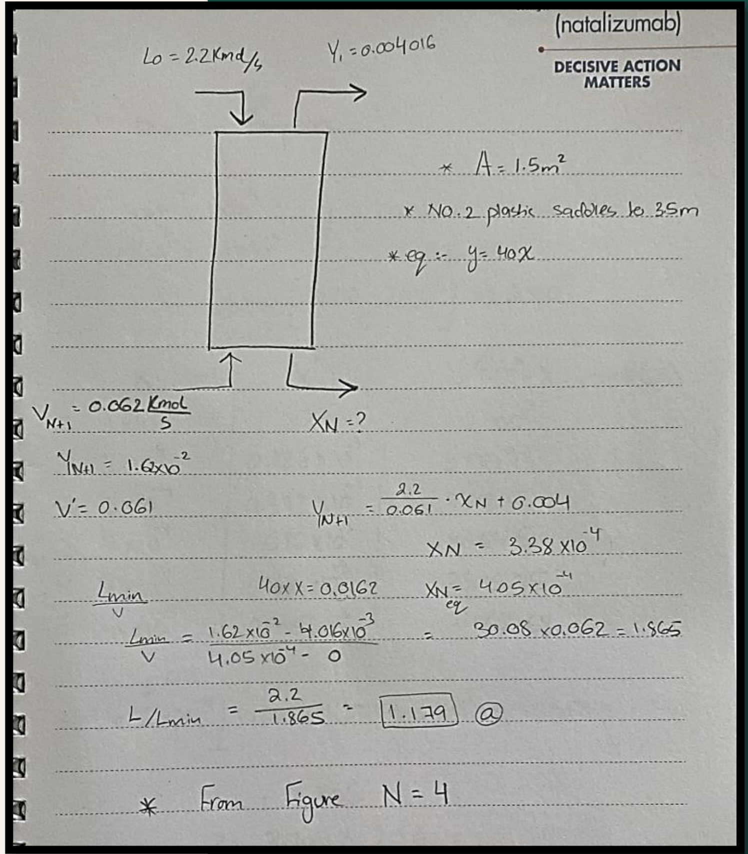


Given: Feed gas flow rate of 0.062 kmol/s containing 1.6 mol% SO_2. Absorbent is 2.2 kmol/s of pure water. Packed column is 1.5 m^2 in cross sectional area and packed with No. 2 plastic super Intalox saddles to a 3.5-m height. Exit gas contains an SO_2 mole fraction of 0.004. Operating pressure is 1 atm. At operating temperature, equilibrium curve for SO_2 is $y = Kx = 40x$

Assumptions: No stripping of water. No absorption of air.

Find: (a) $L/L_{\min}$
(b) N_{OG} and N_t
(c) H_{OG} and HETP
(d) $K_G a$

(A) + Nt



(natalizumab)

DECISIVE ACTION MATTERS

$L_0 = 2.2\,\text{Kmol/s}$ $Y_1 = 0.004016$

* $A = 1.5\,m^2$

* NO.2 plastic saddles to 3.5m

* eq :- $y = 40x$

$V_{N+1} = 0.062\,\frac{\text{Kmol}}{\text{s}}$ $X_N = ?$

$Y_{N+1} = 1.62\times10^{-2}$

$V' = 0.061$ $Y_{N+1} = \frac{2.2}{0.061} \cdot x_N + 0.004$

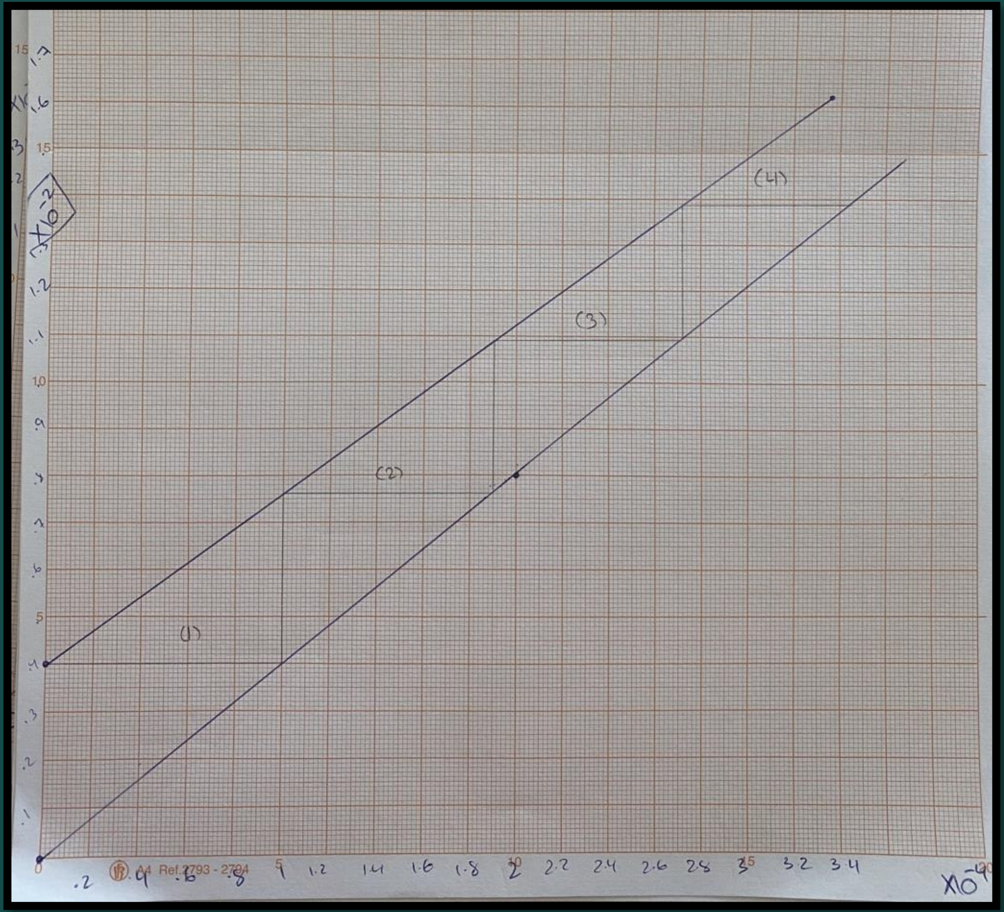
$X_N = 3.38\times10^{-4}$

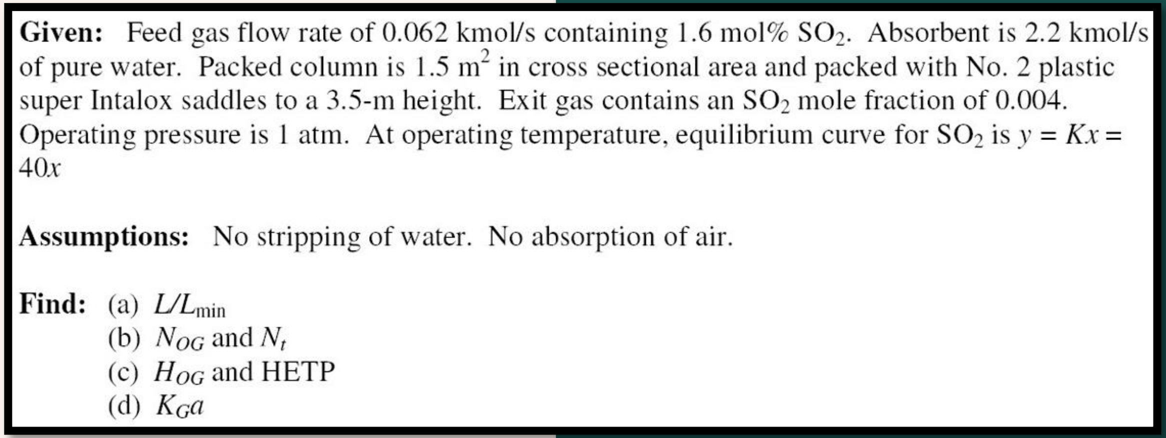
$\frac{L_{min}}{V}$ $40\times X = 0.0162$ $X_{N_{eq}} = 4.05\times10^{-4}$

$\frac{L_{min}}{V} = \frac{1.62\times10^{-2} - 4.016\times10^{-3}}{4.05\times10^{-4} - 0} = 30.08 \times 0.062 = 1.865$

$L/L_{min} = \frac{2.2}{1.865} = \boxed{1.179}$ ⓐ

* From Figure N = 4



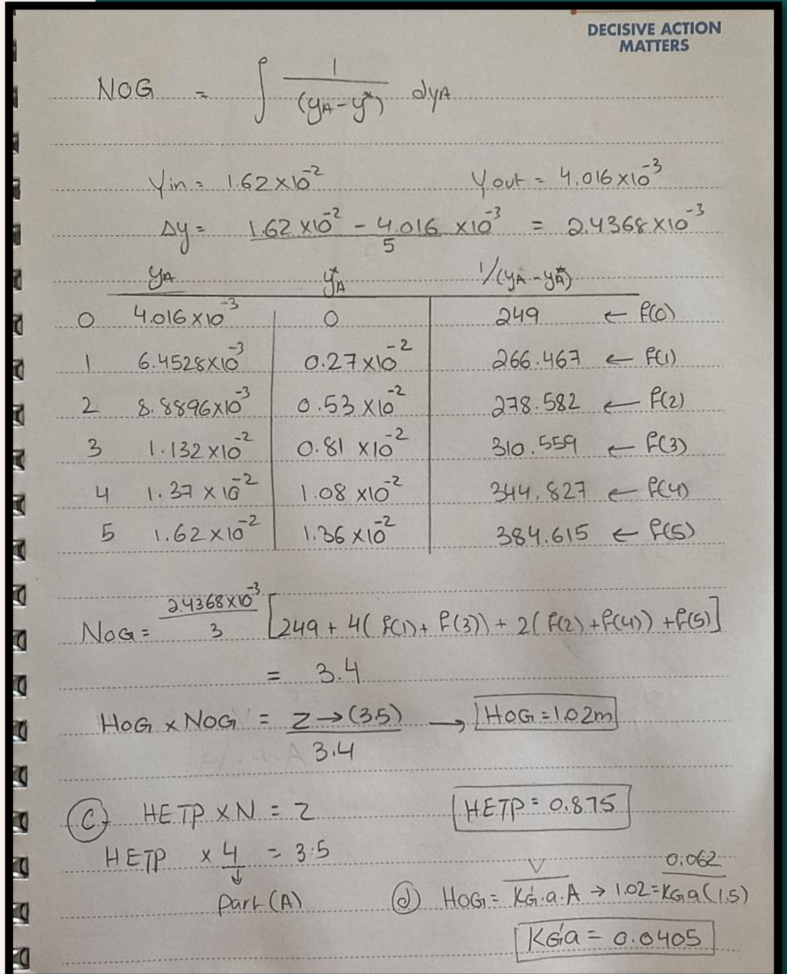


Given: Feed gas flow rate of 0.062 kmol/s containing 1.6 mol% SO_2. Absorbent is 2.2 kmol/s of pure water. Packed column is 1.5 m^2 in cross sectional area and packed with No. 2 plastic super Intalox saddles to a 3.5-m height. Exit gas contains an SO_2 mole fraction of 0.004. Operating pressure is 1 atm. At operating temperature, equilibrium curve for SO_2 is $y = Kx = 40x$

Assumptions: No stripping of water. No absorption of air.

Find:
- (a) L/L_{min}
- (b) N_{OG} and N_t
- (c) H_{OG} and HETP
- (d) $K_G a$

(B) + (C) +(D)



DECISIVE ACTION MATTERS

$$N_{OG} = \int \frac{1}{(y_A - y^*)} dy_A$$

$y_{in} = 1.62 \times 10^{-2}$ $y_{out} = 4.016 \times 10^{-3}$

$$\Delta y = \frac{1.62 \times 10^{-2} - 4.016 \times 10^{-3}}{5} = 2.4368 \times 10^{-3}$$

	y_A	y_A^*	$1/(y_A - y_A^*)$
0	4.016×10^{-3}	0	249 ← f(0)
1	6.4528×10^{-3}	0.27×10^{-2}	266.467 ← f(1)
2	8.8896×10^{-3}	0.53×10^{-2}	278.582 ← f(2)
3	1.132×10^{-2}	0.81×10^{-2}	310.559 ← f(3)
4	1.37×10^{-2}	1.08×10^{-2}	344.827 ← f(4)
5	1.62×10^{-2}	1.36×10^{-2}	384.615 ← f(5)

$$N_{OG} = \frac{2.4368 \times 10^{-3}}{3}\left[249 + 4(f(1) + f(3)) + 2(f(2) + f(4)) + f(5)\right]$$

$$= 3.4$$

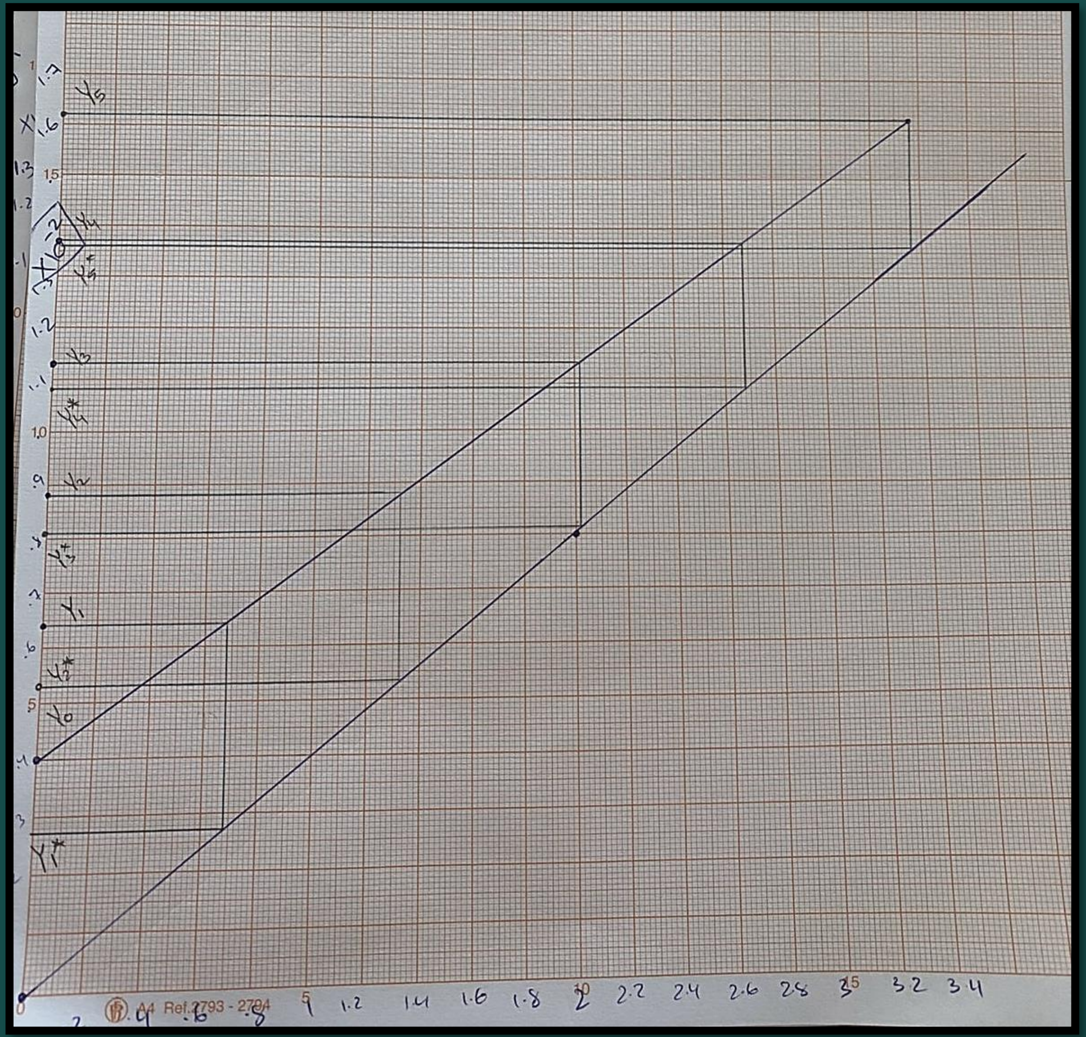
$$H_{OG} \times N_{OG} = \frac{Z \rightarrow (3.5)}{3.4} \longrightarrow \boxed{H_{OG} = 1.02m}$$

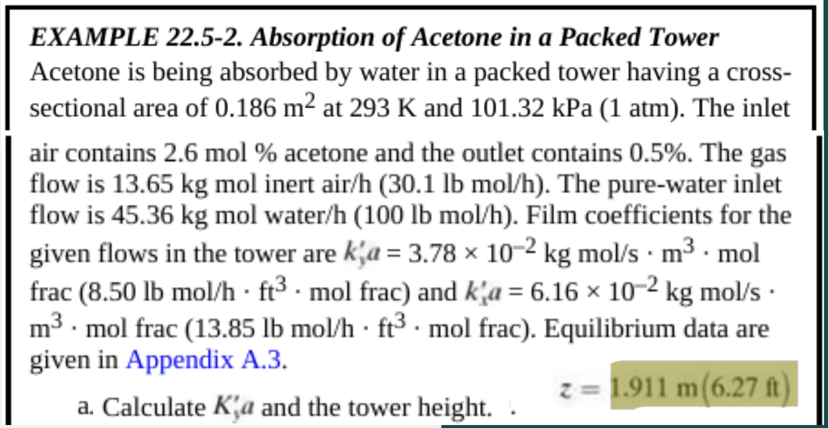
Ⓒ $HETP \times N = Z$ $\boxed{HETP = 0.875}$

$HETP \times \underset{\text{part (A)}}{\underline{4}} = 3.5$

ⓓ $H_{OG} = \frac{V}{K_G' \cdot a \cdot A} \rightarrow 1.02 = \frac{0.062}{K_G a (1.5)}$

$$\boxed{K_G'a = 0.0405}$$



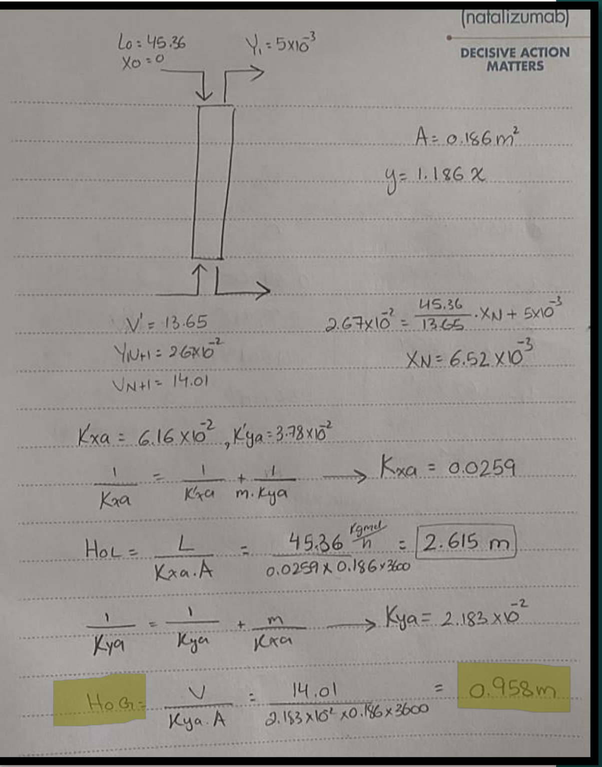


EXAMPLE 22.5-2. Absorption of Acetone in a Packed Tower

Acetone is being absorbed by water in a packed tower having a cross-sectional area of 0.186 m^2 at 293 K and 101.32 kPa (1 atm). The inlet air contains 2.6 mol % acetone and the outlet contains 0.5%. The gas flow is 13.65 kg mol inert air/h (30.1 lb mol/h). The pure-water inlet flow is 45.36 kg mol water/h (100 lb mol/h). Film coefficients for the given flows in the tower are $k'_y a = 3.78 \times 10^{-2}$ kg mol/s · m^3 · mol frac (8.50 lb mol/h · ft^3 · mol frac) and $k'_x a = 6.16 \times 10^{-2}$ kg mol/s · m^3 · mol frac (13.85 lb mol/h · ft^3 · mol frac). Equilibrium data are given in Appendix A.3.

a. Calculate $K'_y a$ and the tower height.

$z = 1.911$ m (6.27 ft)



(natalizumab)

DECISIVE ACTION MATTERS

$L_0 = 45.36$

$X_0 = 0$

$Y_1 = 5 \times 10^{-3}$

$A = 0.186 \, m^2$

$y = 1.186 \, x$

$V' = 13.65$

$Y_{N+1} = 2.6 \times 10^{-2}$

$V_{N+1} = 14.01$

$$2.67 \times 10^{-2} = \frac{45.36}{13.65} \cdot X_N + 5 \times 10^{-3}$$

$X_N = 6.52 \times 10^{-3}$

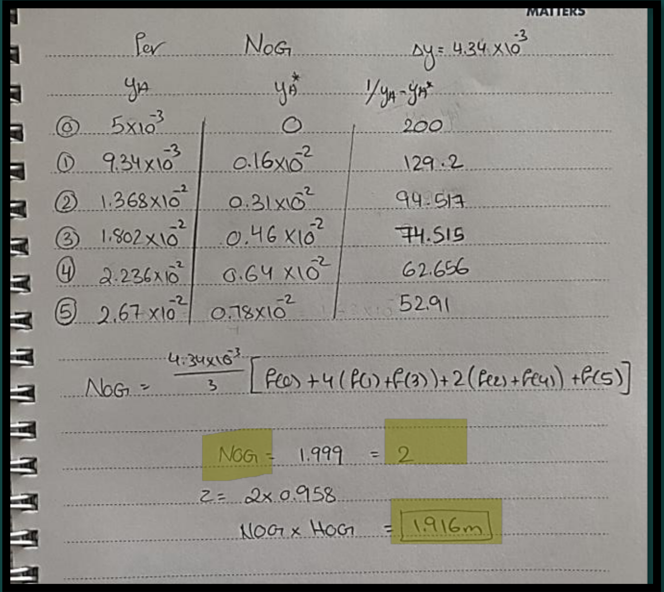
$K'xa = 6.16 \times 10^{-2}$, $K'ya = 3.78 \times 10^{-2}$

$$\frac{1}{Kxa} = \frac{1}{K'xa} + \frac{1}{m \cdot Kya} \longrightarrow Kxa = 0.0259$$

$$H_{OL} = \frac{L}{Kxa \cdot A} = \frac{45.36 \, \frac{kgmol}{h}}{0.0259 \times 0.186 \times 3600} = \boxed{2.615 \text{ m}}$$

$$\frac{1}{Kya} = \frac{1}{Kya} + \frac{m}{Kxa} \longrightarrow Kya = 2.183 \times 10^{-2}$$

$$H_{OG} = \frac{V}{Kya \cdot A} = \frac{14.01}{2.183 \times 10^{-2} \times 0.186 \times 3600} = 0.958 \text{m}$$



MATTERS

Per N_{OG}

$\Delta y = 4.34 \times 10^{-3}$

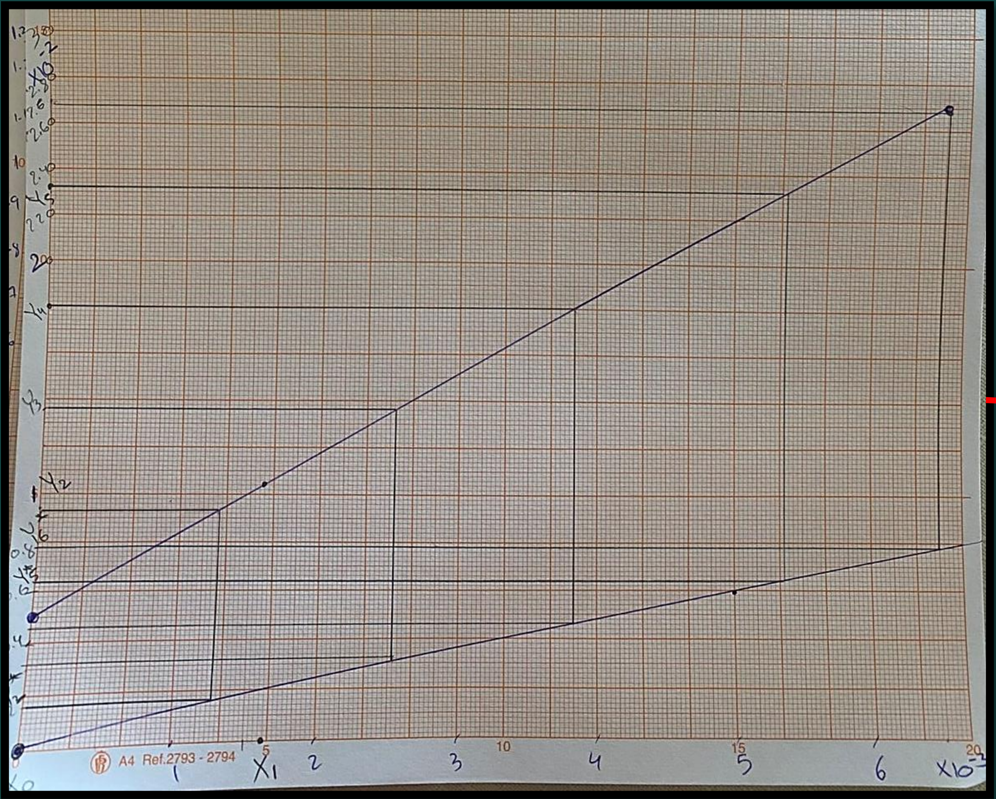
	y_A	y_A^*	$1/y_A - y_A^*$
⓪	5×10^{-3}	0	200
①	9.34×10^{-3}	0.16×10^{-2}	129.2
②	1.368×10^{-2}	0.31×10^{-2}	94.517
③	1.802×10^{-2}	0.46×10^{-2}	74.515
④	2.236×10^{-2}	0.64×10^{-2}	62.656
⑤	2.67×10^{-2}	0.78×10^{-2}	52.91

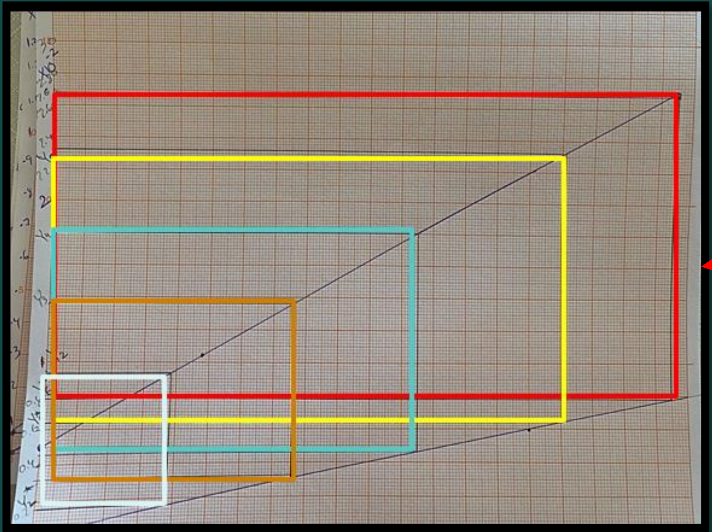
$$N_{OG} = \frac{4.34 \times 10^{-3}}{3} \left[f(0) + 4(f(1) + f(3)) + 2(f(2) + f(4)) + f(5) \right]$$

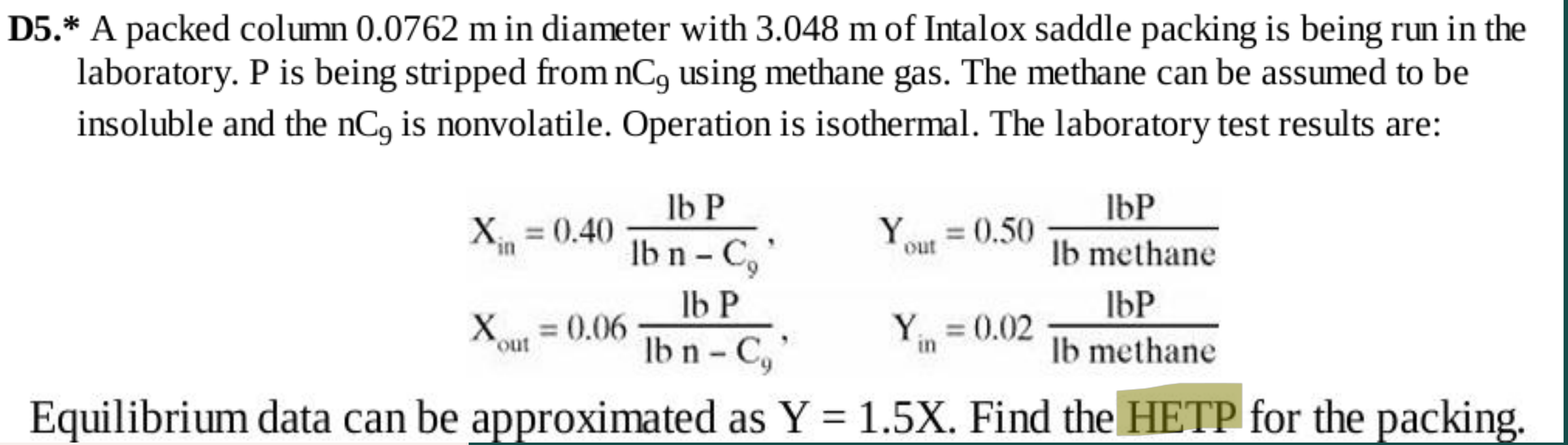
$N_{OG} = 1.999 = 2$

$z = 2 \times 0.958$

$N_{OG} \times H_{OG} = \boxed{1.916 \text{m}}$





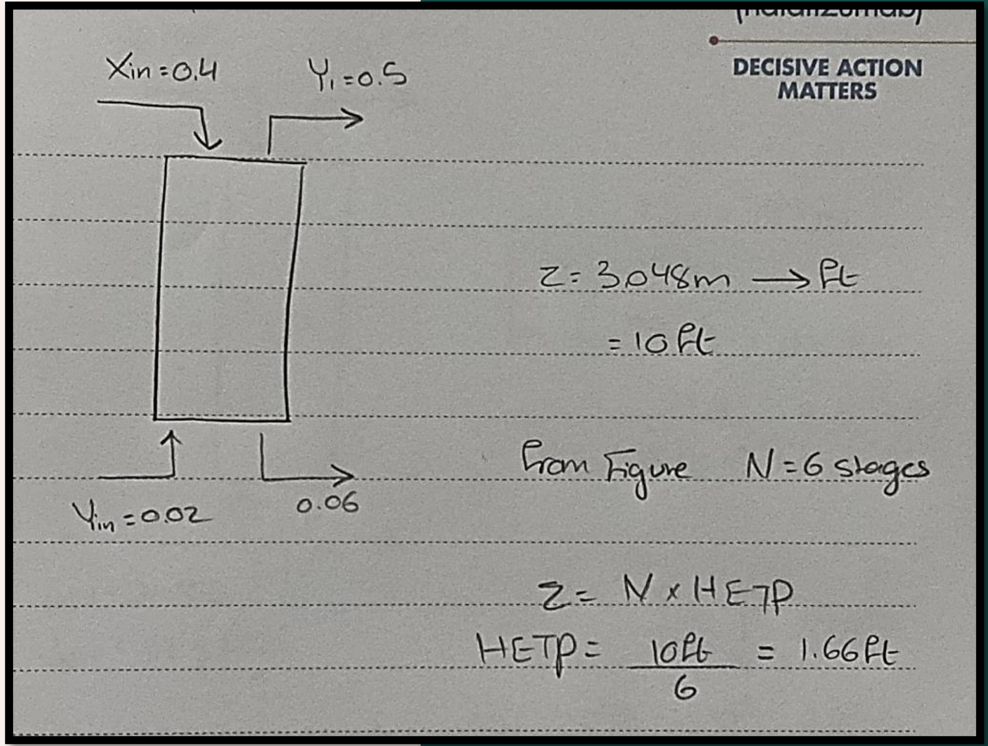


D5.* A packed column 0.0762 m in diameter with 3.048 m of Intalox saddle packing is being run in the laboratory. P is being stripped from nC_9 using methane gas. The methane can be assumed to be insoluble and the nC_9 is nonvolatile. Operation is isothermal. The laboratory test results are:

$$X_{in} = 0.40 \ \frac{\text{lb P}}{\text{lb n} - C_9}, \qquad Y_{out} = 0.50 \ \frac{\text{lbP}}{\text{lb methane}}$$

$$X_{out} = 0.06 \ \frac{\text{lb P}}{\text{lb n} - C_9}, \qquad Y_{in} = 0.02 \ \frac{\text{lbP}}{\text{lb methane}}$$

Equilibrium data can be approximated as Y = 1.5X. Find the HETP for the packing.



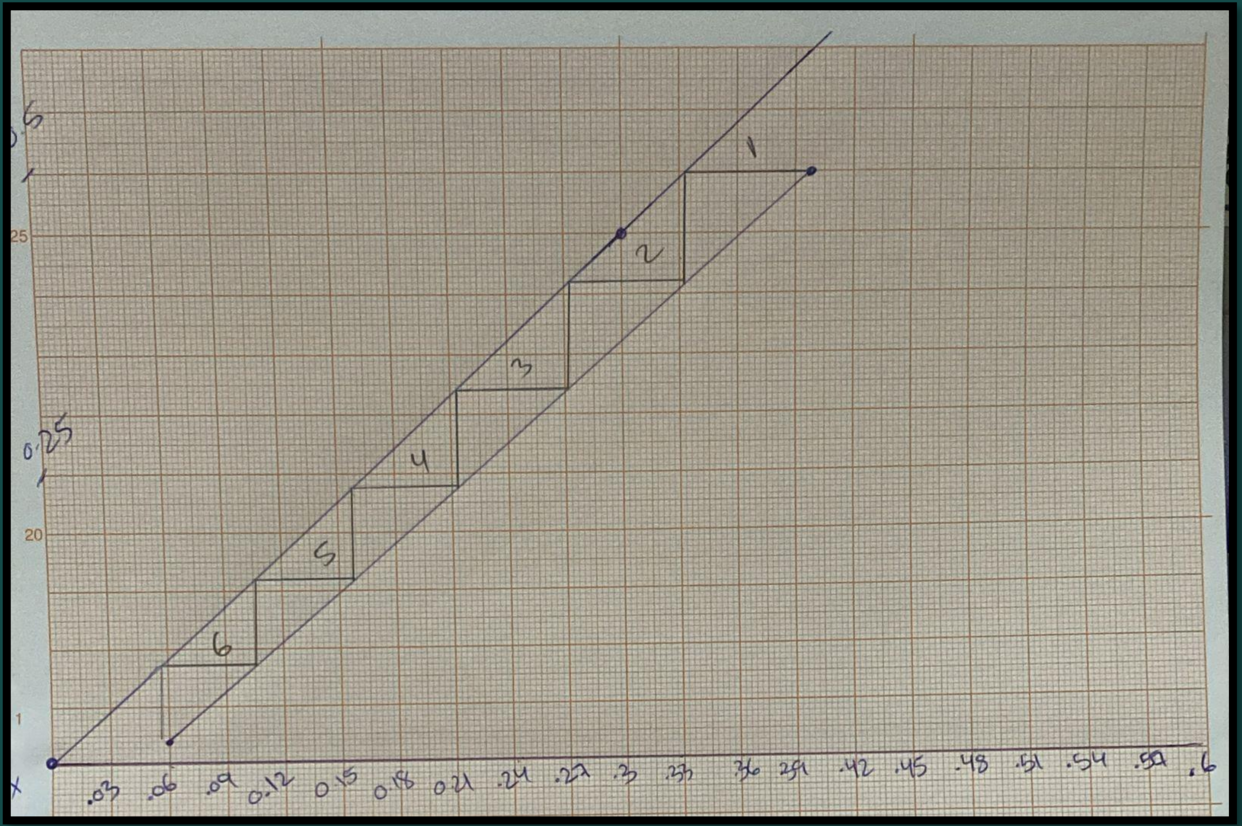
Z = 3.048m → ft

= 10 ft

From Figure N = 6 stages

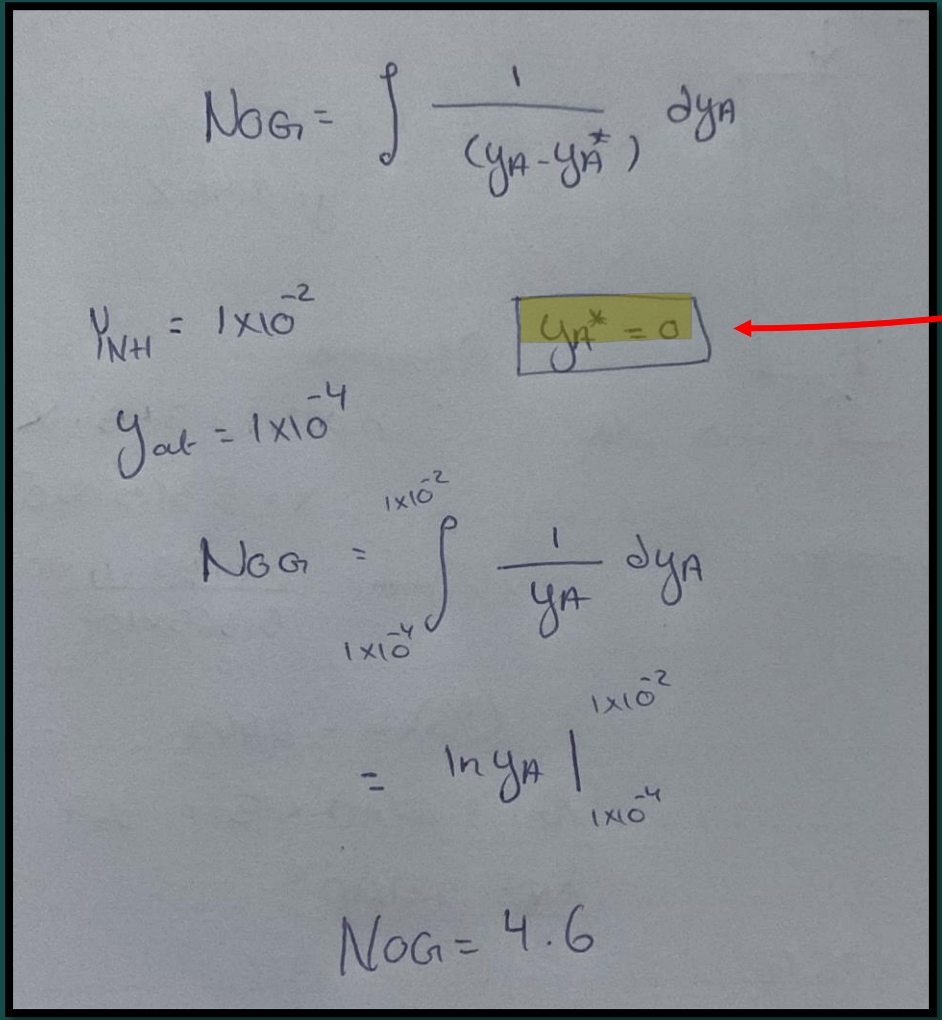
$Z = N \times HETP$

$HETP = \frac{10 \text{ ft}}{6} = 1.66 \text{ ft}$



D10.* A packed tower is used to absorb ammonia from air using aqueous sulfuric acid. The gas enters the tower at 31 lbmol/(h-ft^2) and is 1 mol% ammonia. Aqueous 10 mol% sulfuric acid is fed at a rate of 24 lbmol/(h-ft^2). The equilibrium partial pressure of ammonia above a solution of sulfuric acid is zero. We desire an outlet ammonia composition of 0.01 mol% in the gas stream.

a. Calculate n_{OG} for a countercurrent column.



$$N_{OG} = \int \frac{1}{(y_A - y_A^*)} dy_A$$

$Y_{NH} = 1\times10^{-2}$

$\boxed{y_A^* = 0}$

$y_{at} = 1\times10^{-4}$

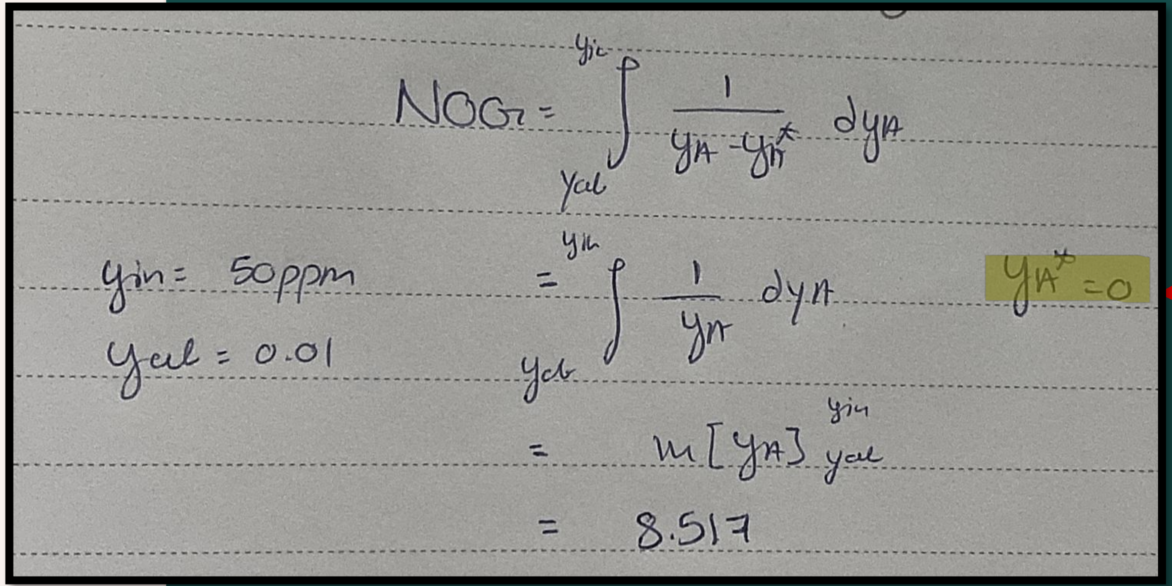
$$N_{OG} = \int_{1\times10^{-4}}^{1\times10^{-2}} \frac{1}{y_A} dy_A$$

$$= \ln y_A \Big|_{1\times10^{-4}}^{1\times10^{-2}}$$

$$N_{OG} = 4.6$$

D11. An air stream containing 50 ppm (mole) of H_2S is to be absorbed with a dilute NaOH solution. The base reacts irreversibly with the acid gas H_2S so that at equilibrium there is no H_2S in the air. An outlet gas that contains 0.01 ppm (mole) of H_2S is desired. L/V = 0.32.

Calculate n_{OG} for a countercurrent system.



$$N_{OG} = \int_{y_{al}}^{y_{in}} \frac{1}{y_A - y_A^*} \, dy_A$$

$y_{in} = 50 ppm$

$y_{al} = 0.01$

$$= \int_{y_{ab}}^{y_{in}} \frac{1}{y_A} \, dy_A$$

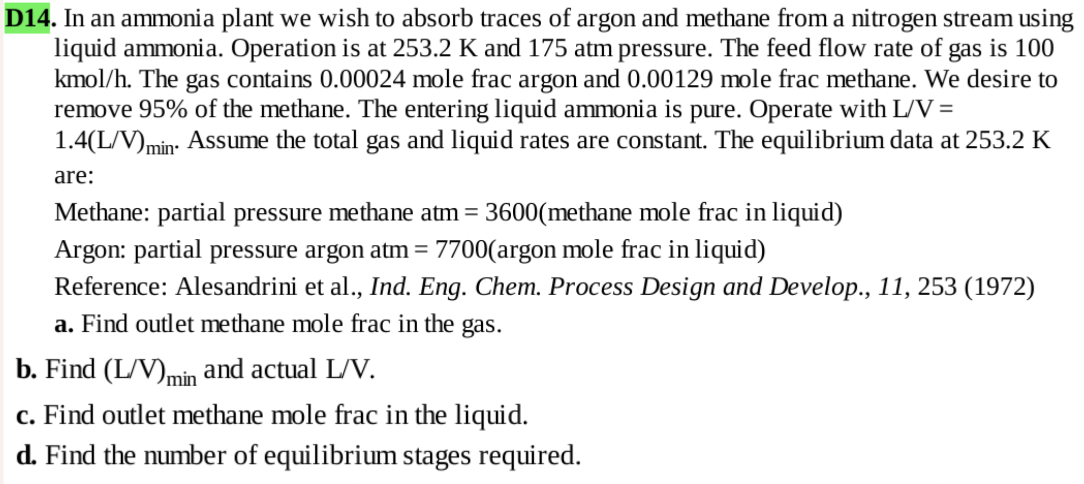
$y_A^* = 0$

$$= \ln [y_A]_{y_{al}}^{y_{in}}$$

$$= 8.517$$

ABSORPTION & STRIPPING

LEC6



D14. In an ammonia plant we wish to absorb traces of argon and methane from a nitrogen stream using liquid ammonia. Operation is at 253.2 K and 175 atm pressure. The feed flow rate of gas is 100 kmol/h. The gas contains 0.00024 mole frac argon and 0.00129 mole frac methane. We desire to remove 95% of the methane. The entering liquid ammonia is pure. Operate with L/V = 1.4$(L/V)_{min}$. Assume the total gas and liquid rates are constant. The equilibrium data at 253.2 K are:

Methane: partial pressure methane atm = 3600(methane mole frac in liquid)

Argon: partial pressure argon atm = 7700(argon mole frac in liquid)

Reference: Alesandrini et al., *Ind. Eng. Chem. Process Design and Develop.*, *11*, 253 (1972)

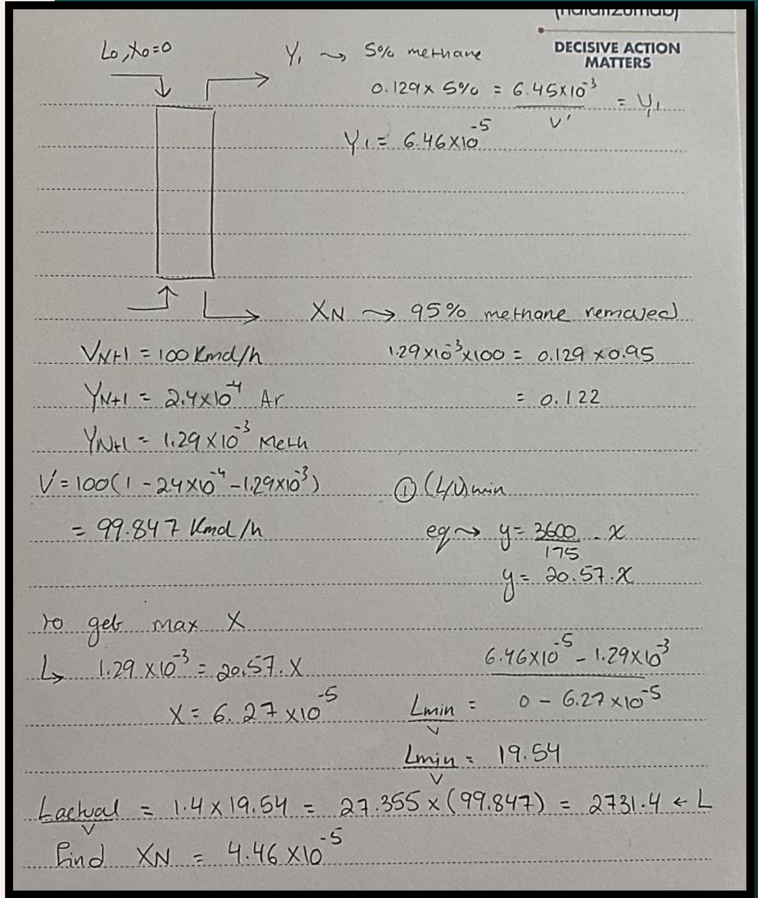
a. Find outlet methane mole frac in the gas.

b. Find $(L/V)_{min}$ and actual L/V.

c. Find outlet methane mole frac in the liquid.

d. Find the number of equilibrium stages required.

6 stages



DECISIVE ACTION MATTERS

$L_0, x_0 = 0$

$Y_1 \rightsquigarrow$ 5% methane

$0.129 \times 5\% = \frac{6.45 \times 10^{-3}}{V'} = Y_1$

$Y_1 = 6.46 \times 10^{-5}$

$X_N \rightsquigarrow$ 95% methane removed

$V_{N+1} = 100$ Kmol/h

$1.29 \times 10^{-3} \times 100 = 0.129 \times 0.95$

$= 0.122$

$Y_{N+1} = 2.4 \times 10^{-4}$ Ar

$Y_{N+1} = 1.29 \times 10^{-3}$ Meth

$V' = 100(1 - 2.4 \times 10^{-4} - 1.29 \times 10^{-3})$

$= 99.847$ Kmol/h

① $(L/V)_{min}$

eq $\rightsquigarrow$ $y = \frac{3600}{175} \cdot x$

$y = 20.57 \cdot x$

to get max X

$\hookrightarrow 1.29 \times 10^{-3} = 20.57 \cdot X$

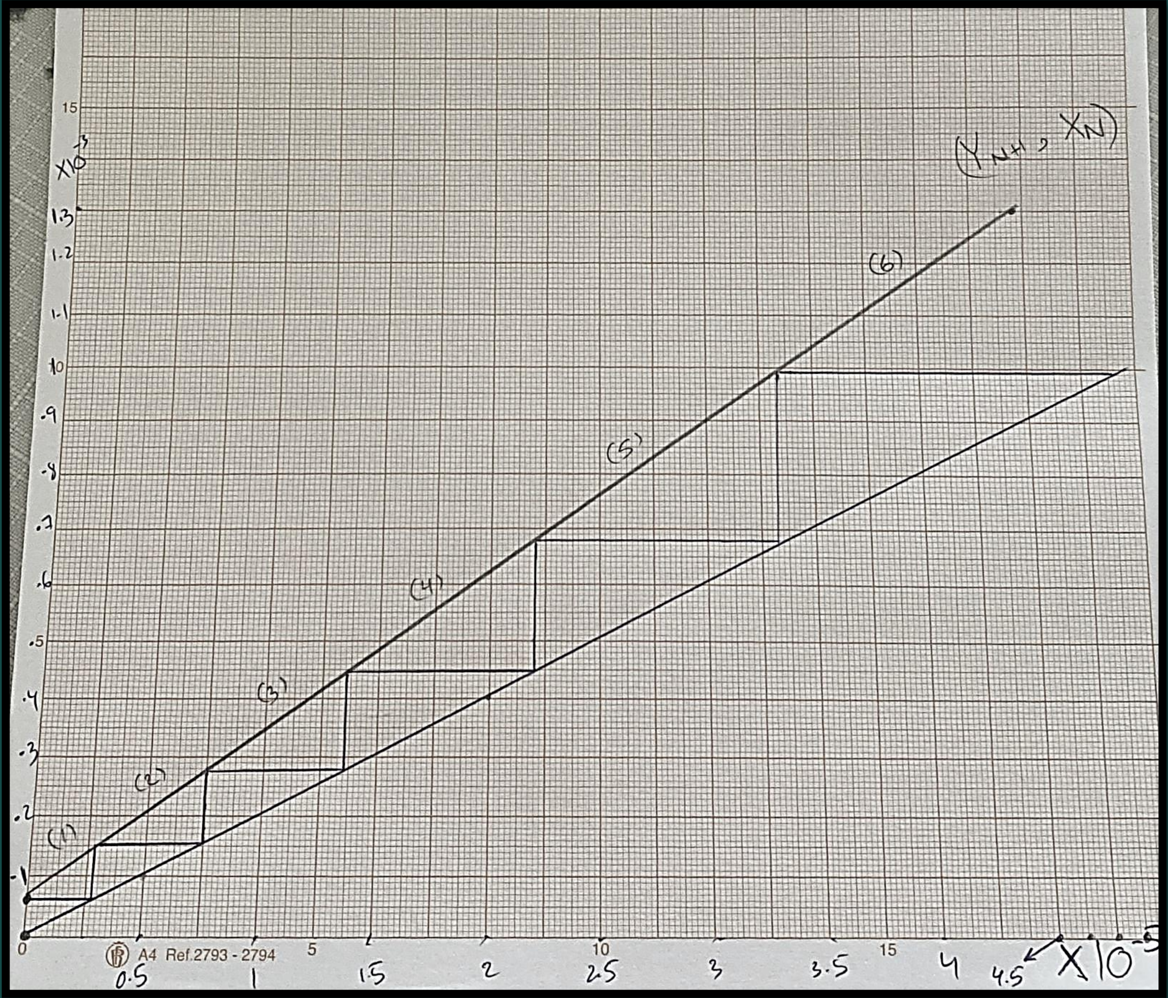
$X = 6.27 \times 10^{-5}$

$\frac{L_{min}}{V} = \frac{6.46 \times 10^{-5} - 1.29 \times 10^{-3}}{0 - 6.27 \times 10^{-5}}$

$\frac{L_{min}}{V} = 19.54$

$\frac{L_{actual}}{V} = 1.4 \times 19.54 = 27.355 \times (99.847) = 2731.4 \leftarrow L$

Find $X_N = 4.46 \times 10^{-5}$

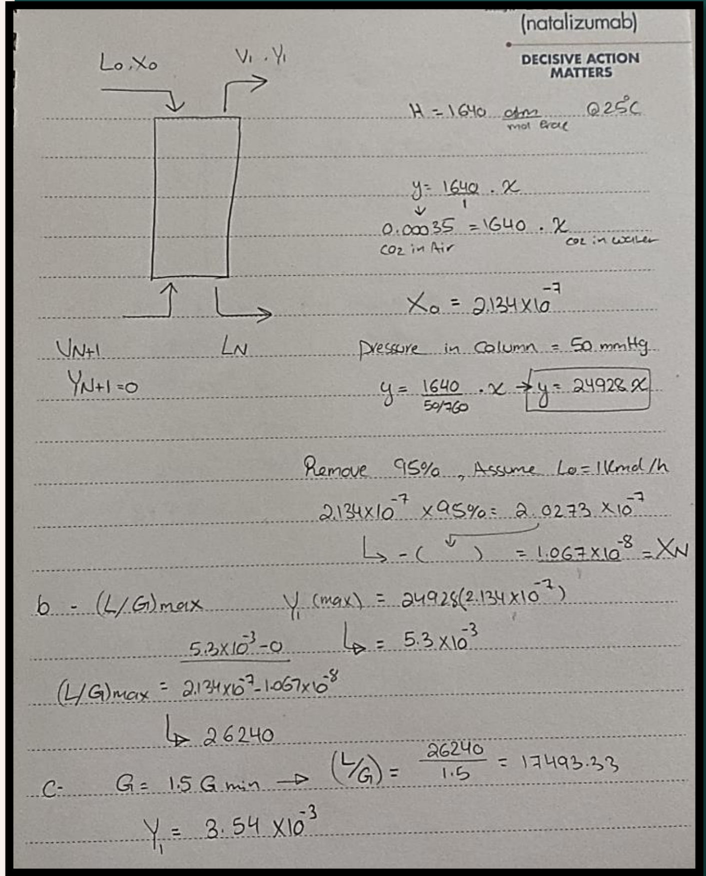


D16. You have a water feed at 25°C and 1 atm. The water has been in contact with air and can be assumed to be in equilibrium with the normal CO_2 content of air (about 0.035 volume %) (Note: This dissolved CO_2 will decrease the pH of the water.) We wish to strip out the CO_2. A countercurrent stripping column will be used operating at 25°C and 50 mm Hg pressure. The stripping gas used will be nitrogen gas saturated with pure water at 25°C. Assume the nitrogen is insoluble in water. Assume ideal gas (vol % = mol%). Data: See Table 12-1. We want to remove 95% of the initial CO_2 in the water.

a. Calculate the inlet and outlet mole fracs of CO_2 in water.

b. Determine $(L/G)_{max}$. If L = 1 kgmol/h, calculate G_{min} (corresponds to $(L/G)_{max}$).

c. If we operate at G = 1.5 × G_{min} find the CO_2 mole frac in the outlet nitrogen and the number of equilibrium stages needed.



(natalizumab)

DECISIVE ACTION MATTERS

L_0, X_0 V_1, Y_1

V_{N+1} L_N

$Y_{N+1} = 0$

$H = 1640 \frac{atm}{mol\ frac}$ @25°C

$y = \frac{1640}{1} \cdot x$

$\underset{CO_2\ in\ Air}{0.00035} = 1640 \cdot \underset{CO_2\ in\ water}{x}$

$X_0 = 2.134\times10^{-7}$

Pressure in Column = 50 mmHg

$y = \frac{1640}{50/760} \cdot x \rightarrow \boxed{y = 24928x}$

Remove 95%, Assume L_0 = 1 kmol/h

$2.134\times10^{-7} \times 95\% = 2.0273\times10^{-7}$

$\hookrightarrow -(\ \ ) = 1.067\times10^{-8} = X_N$

b - $(L/G)_{max}$ $Y_1(max) = 24928(2.134\times10^{-7})$

$\hookrightarrow = 5.3\times10^{-3}$

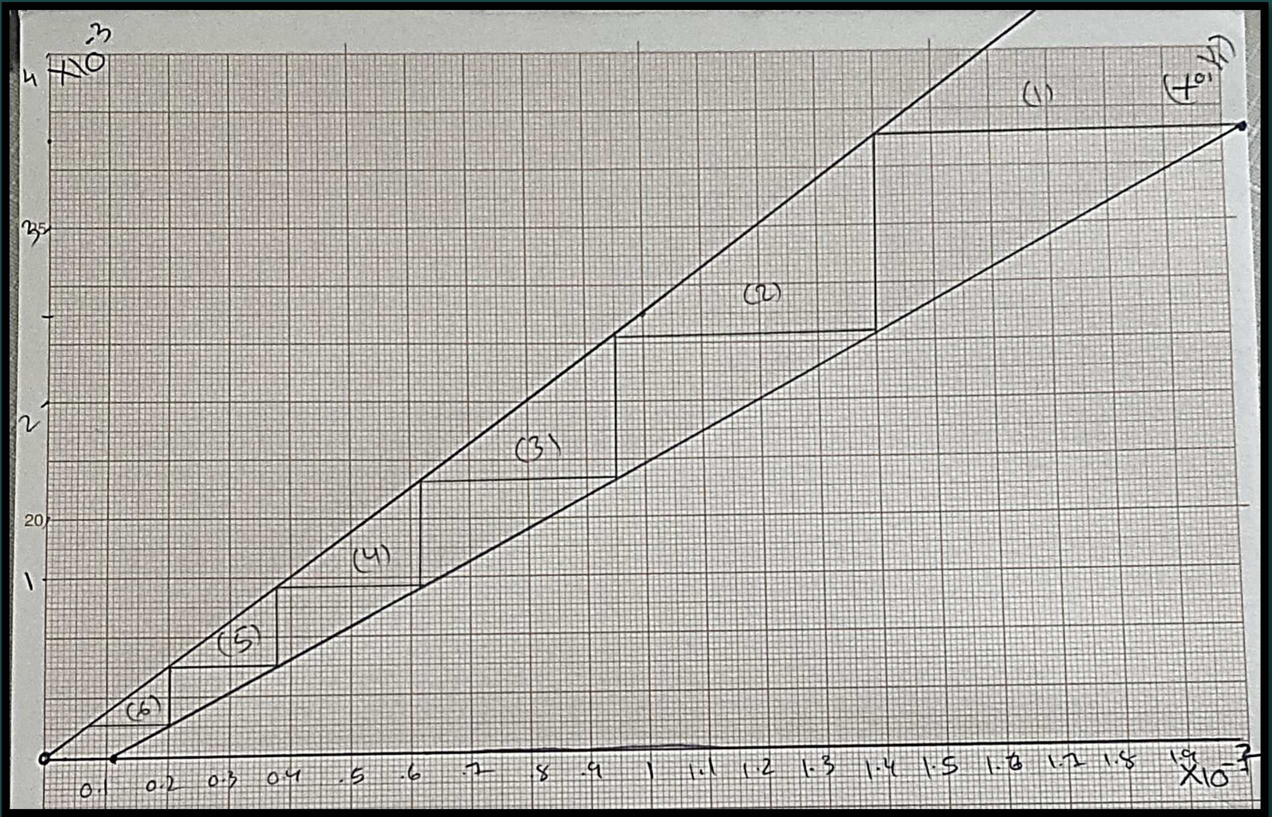
$(L/G)_{max} = \frac{5.3\times10^{-3} - 0}{2.134\times10^{-7} - 1.067\times10^{-8}}$

$\hookrightarrow 26240$

c- $G = 1.5\ G_{min} \rightarrow (L/G) = \frac{26240}{1.5} = 17493.33$

$Y_1 = 3.54\times10^{-3}$

6 stages

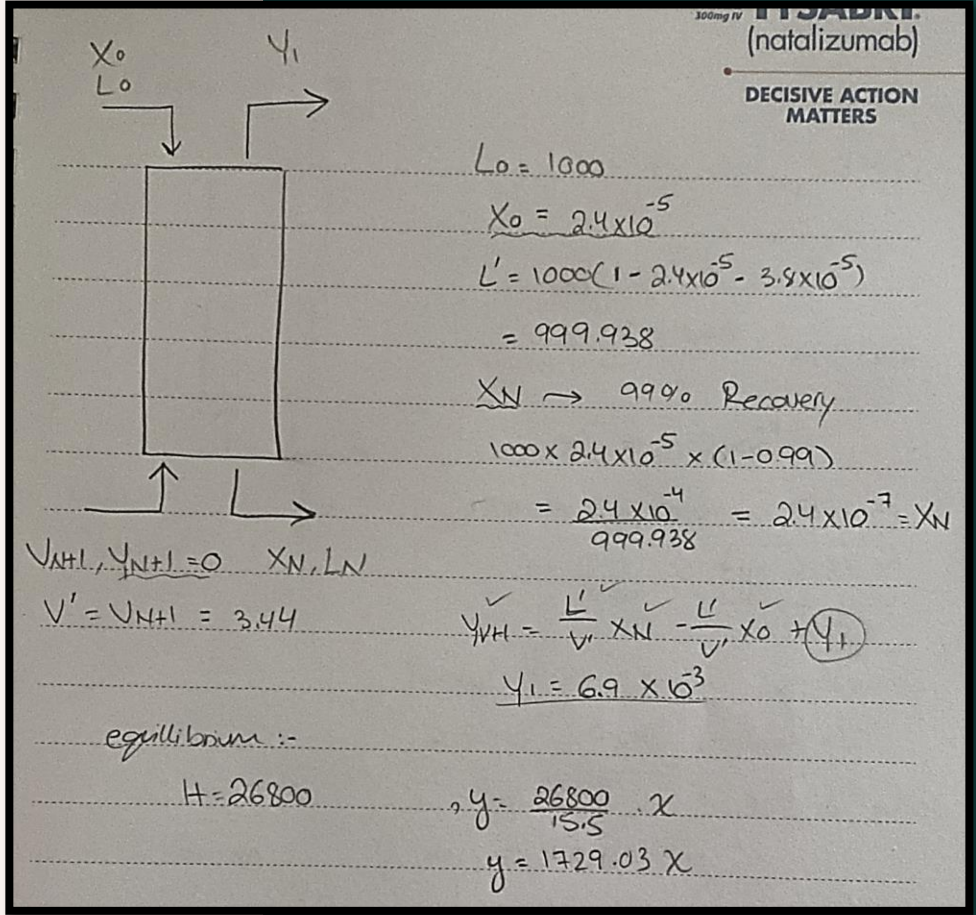


D20. We need to remove H_2S and CO_2 from 1000 kmol/h of a water stream at 0°C and 15.5 atm. The inlet liquid contains 0.000024 mole frac H_2S and 0.000038 mole frac CO_2. We desire a 99% recovery of the H_2S in the gas stream. The gas used is pure nitrogen at 0°C and 15.5 atm. The nitrogen flow rate is 3.44 kmol/h. A staged countercurrent stripper will be used. Assume water flow rate and air flow rate are constant. Data are in Table 12-1. Note: Watch your decimals.

a. Determine the H_2S mole fractions in the outlet gas and liquid streams.

b. Determine the number of equilibrium stages required.

3 stages



$L_0 = 1000$

$X_0 = 2.4\times10^{-5}$

$L' = 1000(1 - 2.4\times10^{-5} - 3.8\times10^{-5})$

$= 999.938$

$X_N \rightarrow$ 99% Recovery

$1000 \times 2.4\times10^{-5} \times (1-0.99)$

$= \frac{2.4\times10^{-4}}{999.938} = 2.4\times10^{-7} = X_N$

$V_{N+1}, Y_{N+1} = 0$ X_N, L_N

$V' = V_{N+1} = 3.44$

$Y_{N+1} = \frac{L'}{V'} X_N - \frac{L'}{V'} X_0 + \boxed{Y_1}$

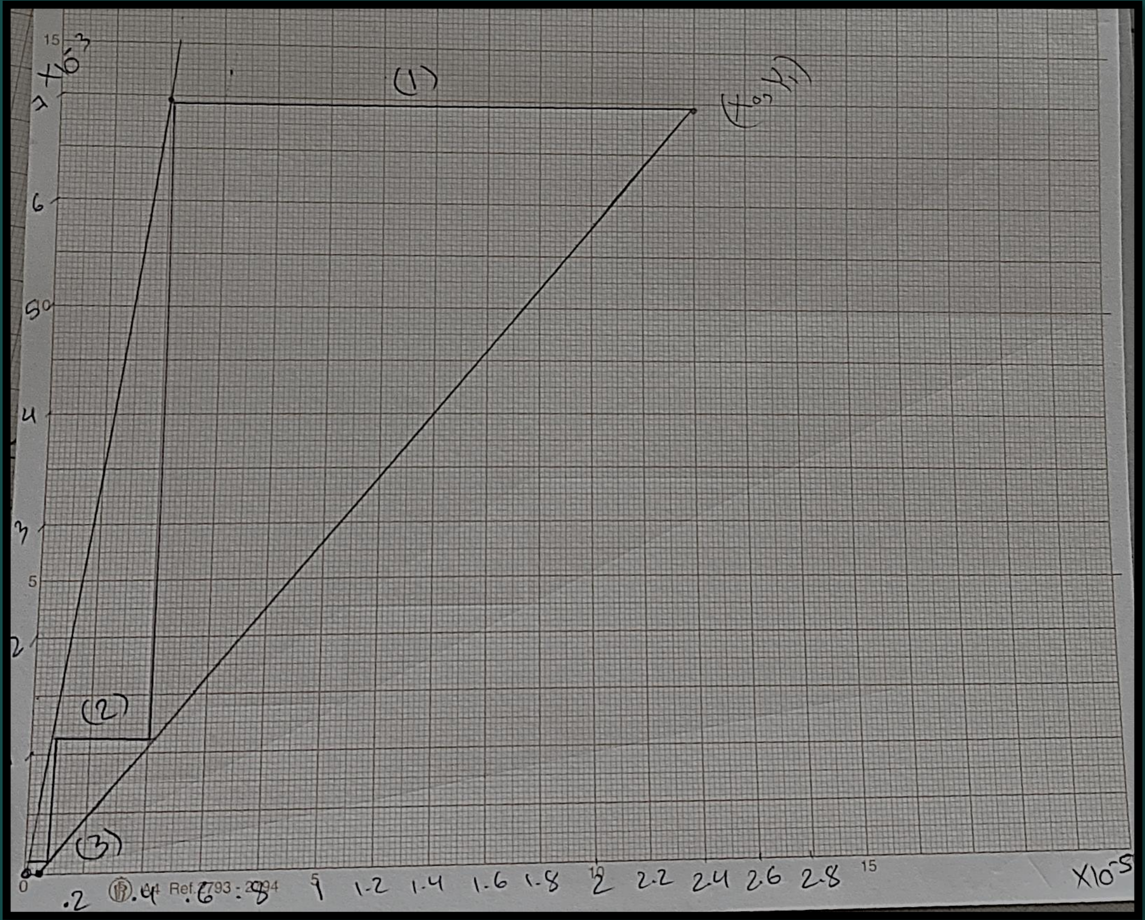
$Y_1 = 6.9\times10^{-3}$

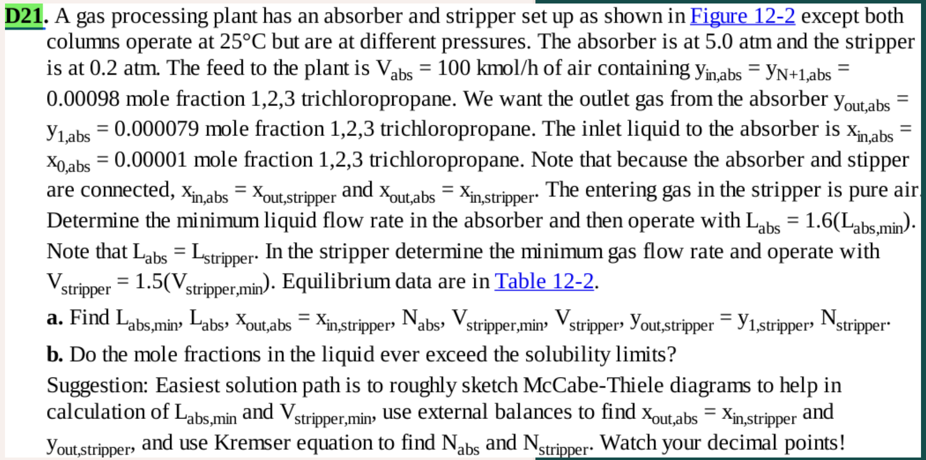
equillibrium :-

H = 26800

$y = \frac{26800}{15.5} x$

$y = 1729.03\, x$



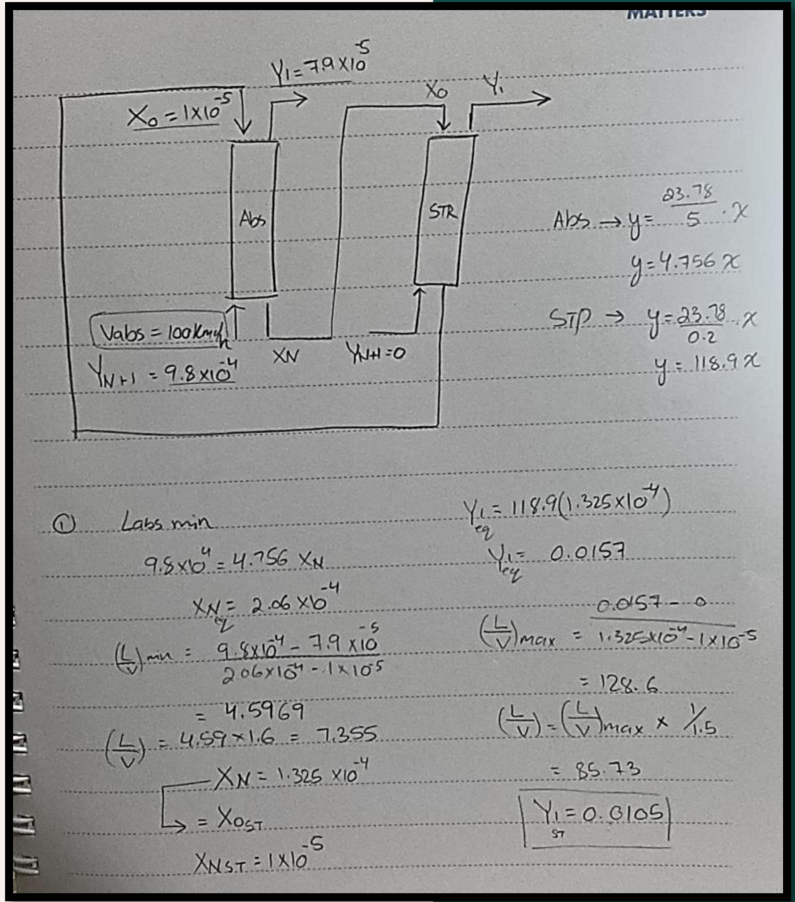


D21. A gas processing plant has an absorber and stripper set up as shown in Figure 12-2 except both columns operate at 25°C but are at different pressures. The absorber is at 5.0 atm and the stripper is at 0.2 atm. The feed to the plant is V_{abs} = 100 kmol/h of air containing $y_{in,abs} = y_{N+1,abs}$ = 0.00098 mole fraction 1,2,3 trichloropropane. We want the outlet gas from the absorber $y_{out,abs} = y_{1,abs}$ = 0.000079 mole fraction 1,2,3 trichloropropane. The inlet liquid to the absorber is $x_{in,abs} = x_{0,abs}$ = 0.00001 mole fraction 1,2,3 trichloropropane. Note that because the absorber and stripper are connected, $x_{in,abs} = x_{out,stripper}$ and $x_{out,abs} = x_{in,stripper}$. The entering gas in the stripper is pure air. Determine the minimum liquid flow rate in the absorber and then operate with $L_{abs} = 1.6(L_{abs,min})$. Note that $L_{abs} = L_{stripper}$. In the stripper determine the minimum gas flow rate and operate with $V_{stripper} = 1.5(V_{stripper,min})$. Equilibrium data are in Table 12-2.

a. Find $L_{abs,min}$, L_{abs}, $x_{out,abs} = x_{in,stripper}$, N_{abs}, $V_{stripper,min}$, $V_{stripper}$, $y_{out,stripper} = y_{1,stripper}$, $N_{stripper}$.

b. Do the mole fractions in the liquid ever exceed the solubility limits?

Suggestion: Easiest solution path is to roughly sketch McCabe-Thiele diagrams to help in calculation of $L_{abs,min}$ and $V_{stripper,min}$, use external balances to find $x_{out,abs} = x_{in,stripper}$ and $y_{out,stripper}$, and use Kremser equation to find N_{abs} and $N_{stripper}$. Watch your decimal points!



Abs → $y = \frac{23.78}{5} \cdot x$

$y = 4.756\,x$

STP → $y = \frac{23.78}{0.2} x$

$y = 118.9\,x$

① Labs min

$9.8 \times 10^{-4} = 4.756\, X_N$

$X_{N_{eq}} = 2.06 \times 10^{-4}$

$\left(\frac{L}{V}\right)_{min} = \frac{9.8 \times 10^{-4} - 7.9 \times 10^{-5}}{2.06 \times 10^{-4} - 1 \times 10^{-5}}$

$= 4.5969$

$\left(\frac{L}{V}\right) = 4.59 \times 1.6 = 7.355$

$X_N = 1.325 \times 10^{-4}$

$= X_{0_{ST}}$

$X_{N_{ST}} = 1 \times 10^{-5}$

$Y_{1_{eq}} = 118.9(1.325 \times 10^{-4})$

$Y_{1_{eq}} = 0.0157$

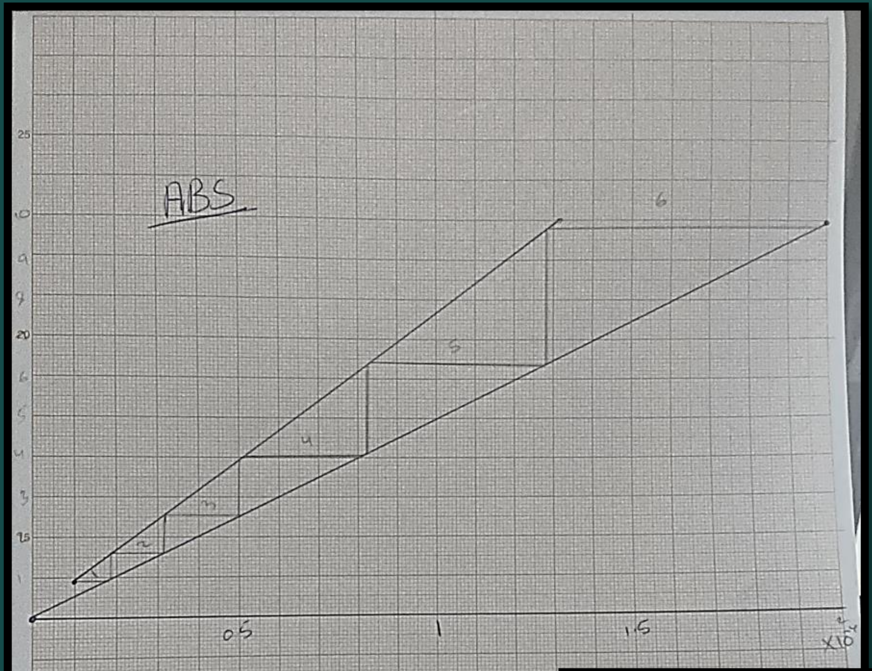
$\left(\frac{L}{V}\right)_{max} = \frac{0.0157 - 0}{1.325 \times 10^{-4} - 1 \times 10^{-5}}$

$= 128.6$

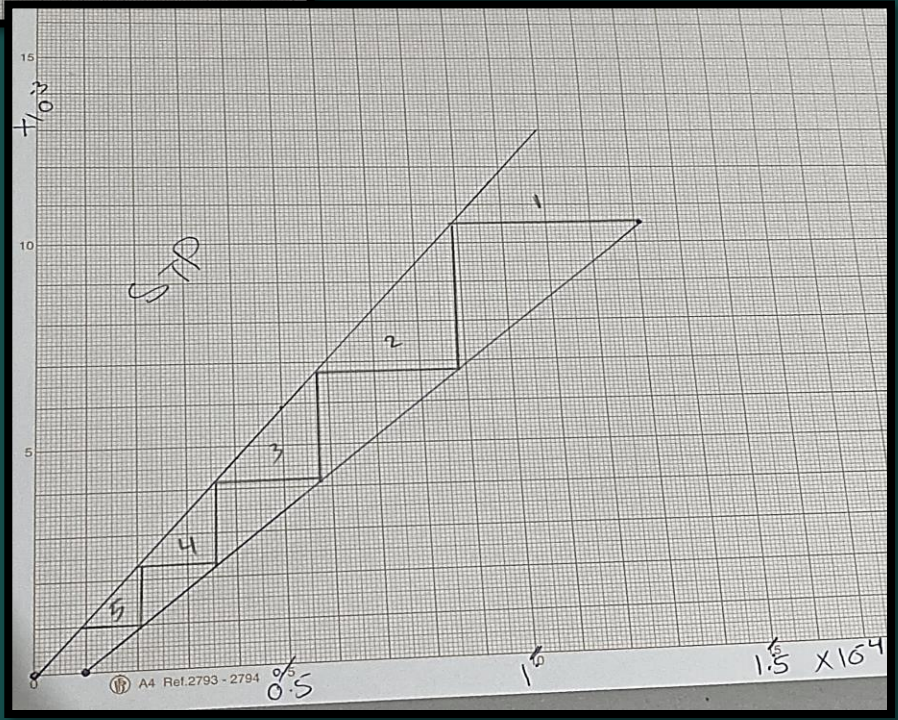
$\left(\frac{L}{V}\right) = \left(\frac{L}{V}\right)_{max} \times 1/1.5$

$= 85.73$

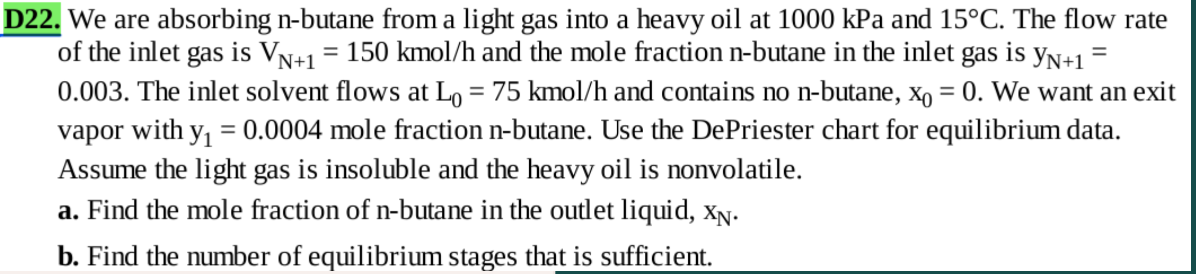
$\boxed{Y_{1_{ST}} = 0.0105}$



6 stages



5 stages

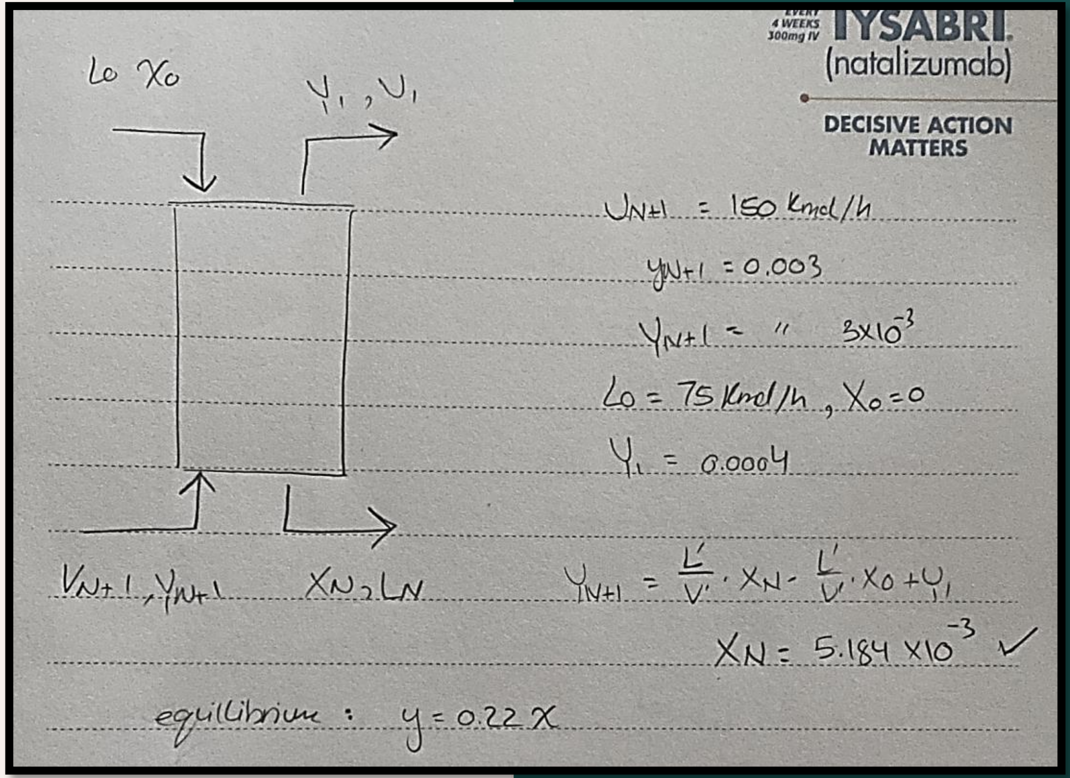


D22. We are absorbing n-butane from a light gas into a heavy oil at 1000 kPa and 15°C. The flow rate of the inlet gas is V_{N+1} = 150 kmol/h and the mole fraction n-butane in the inlet gas is y_{N+1} = 0.003. The inlet solvent flows at L_0 = 75 kmol/h and contains no n-butane, x_0 = 0. We want an exit vapor with y_1 = 0.0004 mole fraction n-butane. Use the DePriester chart for equilibrium data. Assume the light gas is insoluble and the heavy oil is nonvolatile.

a. Find the mole fraction of n-butane in the outlet liquid, x_N.

b. Find the number of equilibrium stages that is sufficient.

2 stages



V_{N+1} = 150 Kmol/h

y_{N+1} = 0.003

Y_{N+1} = " 3×10^{-3}

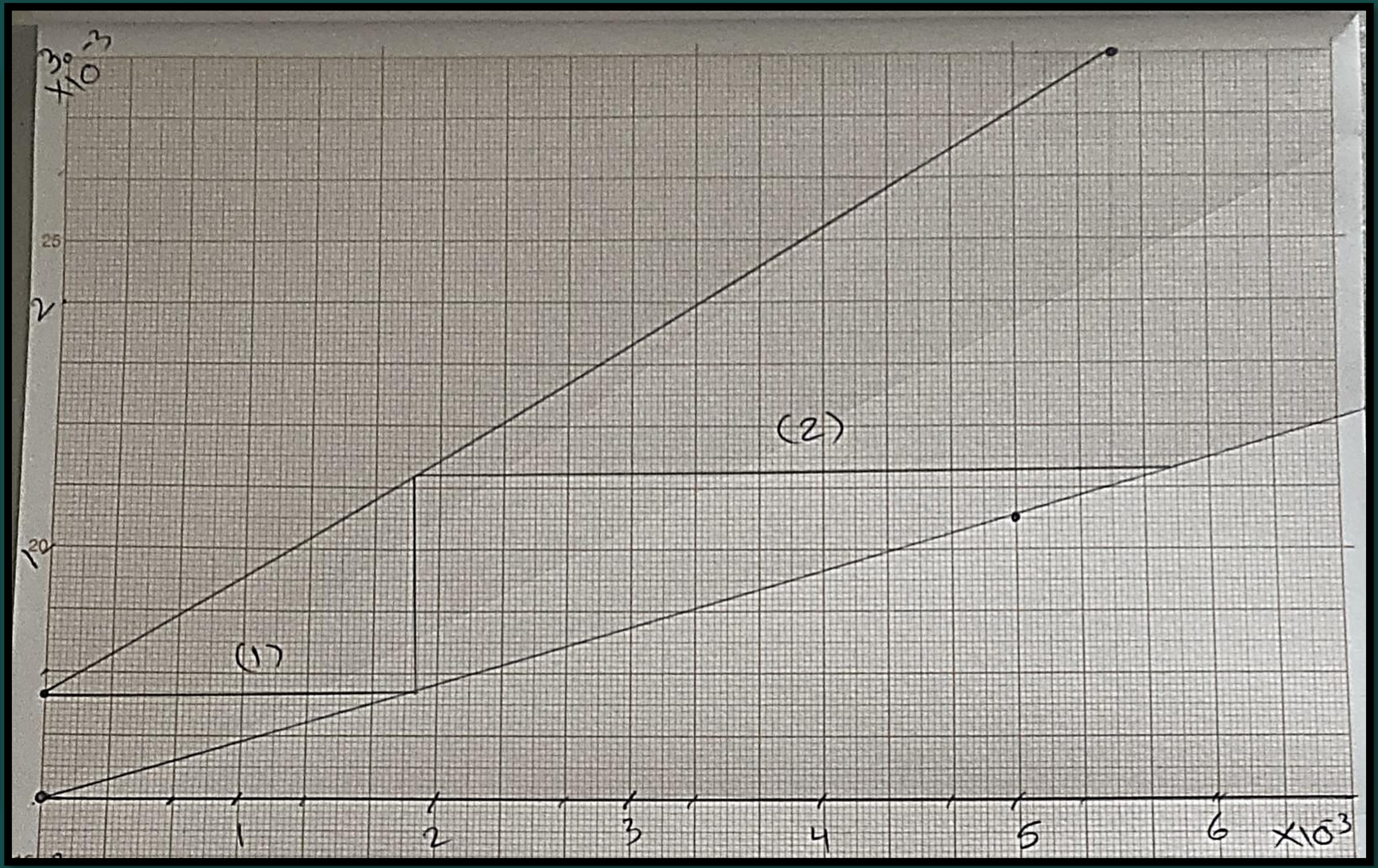
L_0 = 75 Kmol/h, $X_0 = 0$

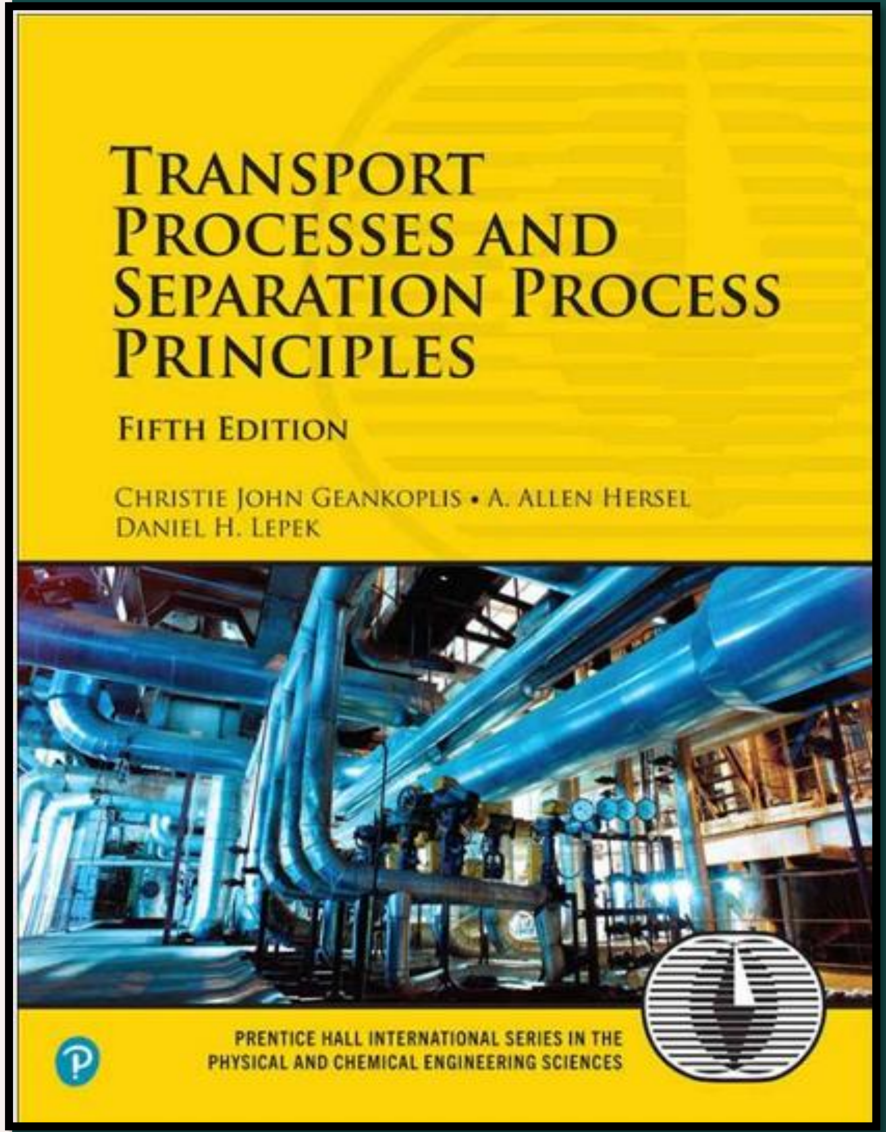
Y_1 = 0.0004

$$Y_{N+1} = \frac{L'}{V'}\cdot X_N - \frac{L'}{V'}\cdot X_0 + Y_1$$

$X_N = 5.184\times10^{-3}$ ✓

equilibrium : $y = 0.22x$





TRANSPORT PROCESSES AND SEPARATION PROCESS PRINCIPLES

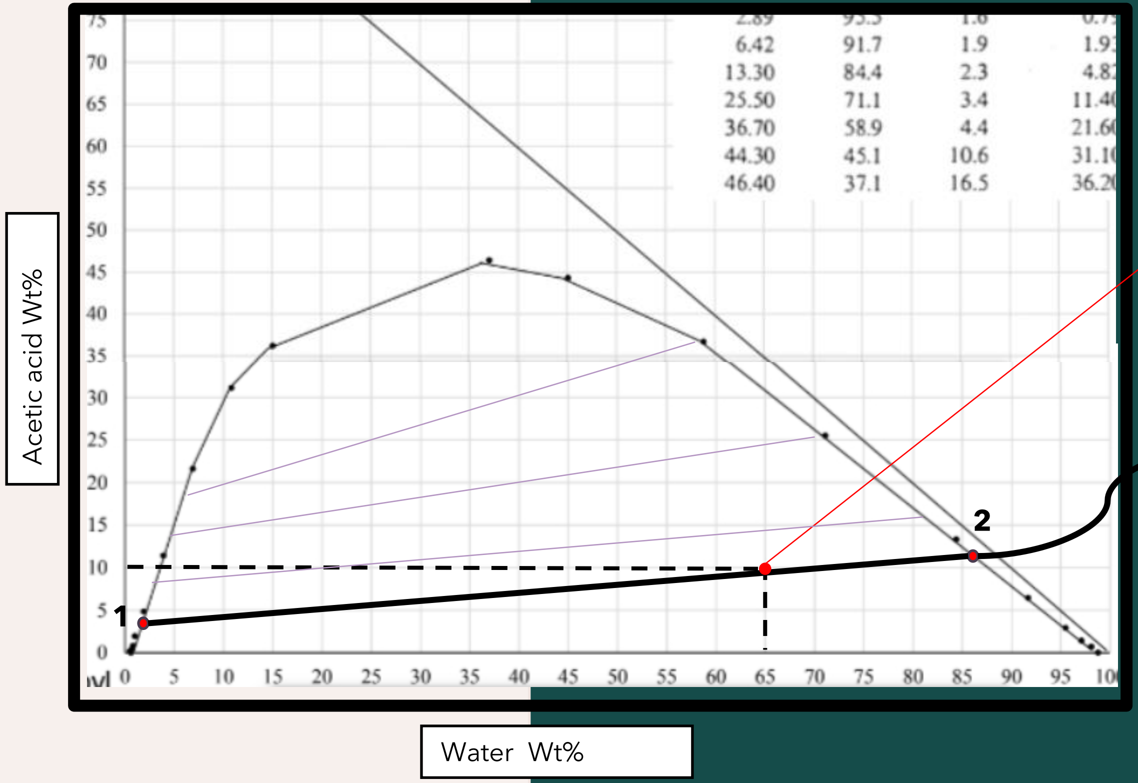
FIFTH EDITION

CHRISTIE JOHN GEANKOPLIS • A. ALLEN HERSEL
DANIEL H. LEPEK

PRENTICE HALL INTERNATIONAL SERIES IN THE PHYSICAL AND CHEMICAL ENGINEERING SCIENCES

Chapter 27 : Liquid–Liquid Extraction

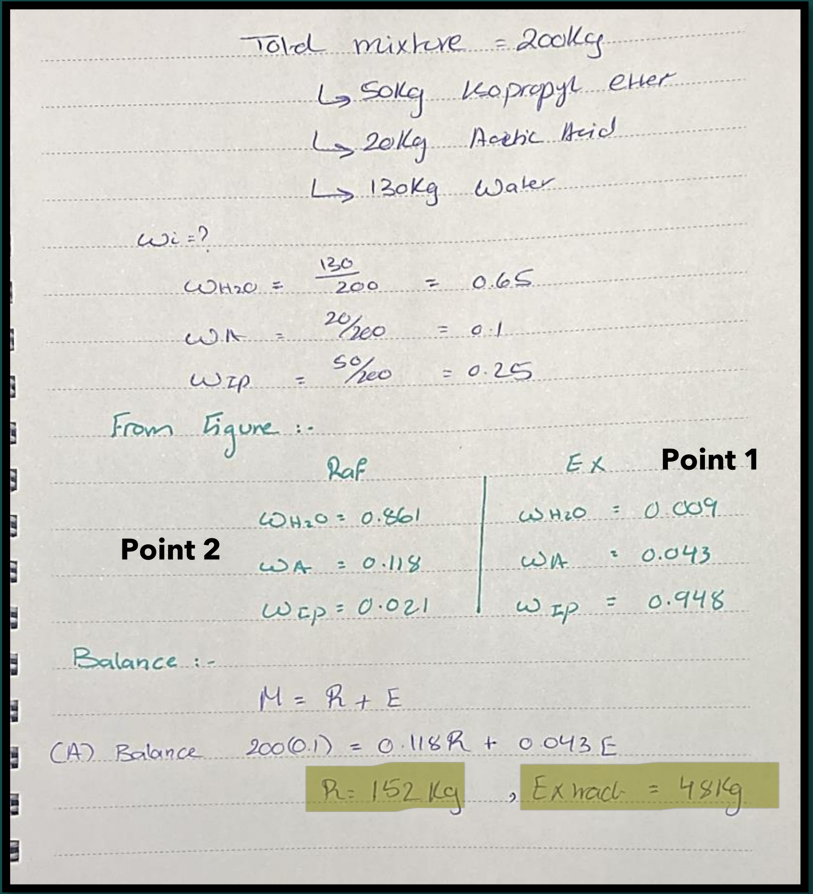
27.1-1. *Composition of Two Liquid Phases in Equilibrium.* An original mixture weighing 200 kg and containing 50 kg of isopropyl ether, 20 kg of acetic acid, and 130 kg of water is equilibrated in a mixer–settler and the phases are separated. Determine the amounts and compositions of the raffinate and extract layers. Use equilibrium data from Appendix A.3.



2.89	95.5	1.6	0.7
6.42	91.7	1.9	1.9
13.30	84.4	2.3	4.8
25.50	71.1	3.4	11.4
36.70	58.9	4.4	21.6
44.30	45.1	10.6	31.1
46.40	37.1	16.5	36.2

x(AC)= 0.1
x(H2O)= 0.65
x(IPE)= 0.25

Corresponding tie-line for this point.



Total mixture = 200kg
↳ 50kg Isopropyl ether
↳ 20kg Acetic Acid
↳ 130kg Water

$\omega_i = ?$

$\omega_{H_2O} = \frac{130}{200} = 0.65$

$\omega_A = 20/200 = 0.1$

$\omega_{IP} = 50/200 = 0.25$

From Figure :-

Raf	Ex
$\omega_{H_2O} = 0.861$	$\omega_{H_2O} = 0.009$
$\omega_A = 0.118$	$\omega_A = 0.043$
$\omega_{IP} = 0.021$	$\omega_{IP} = 0.948$

Point 1

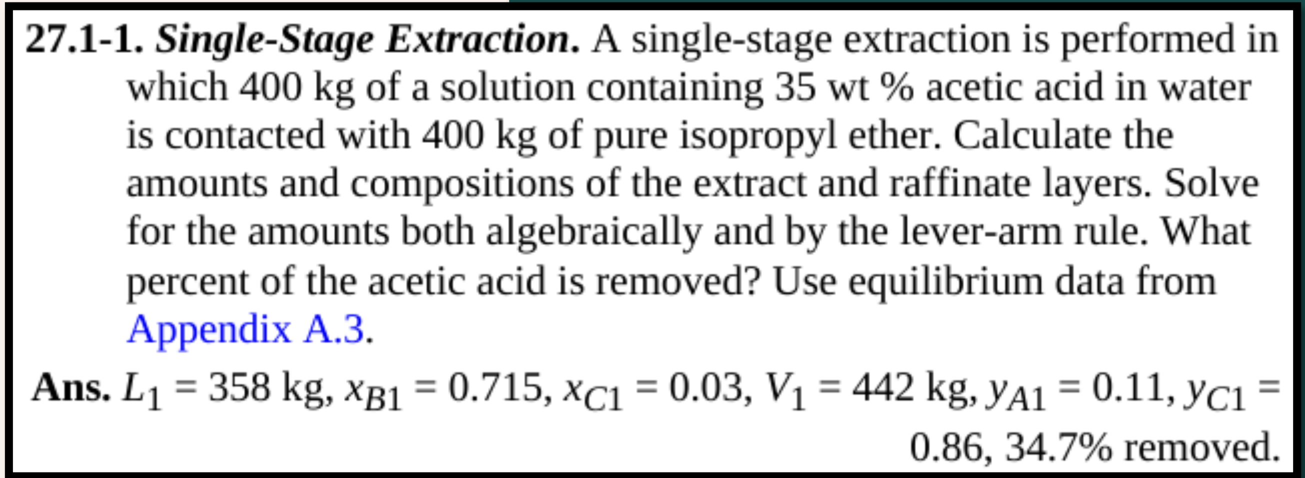
Point 2

Balance :-

$M = R + E$

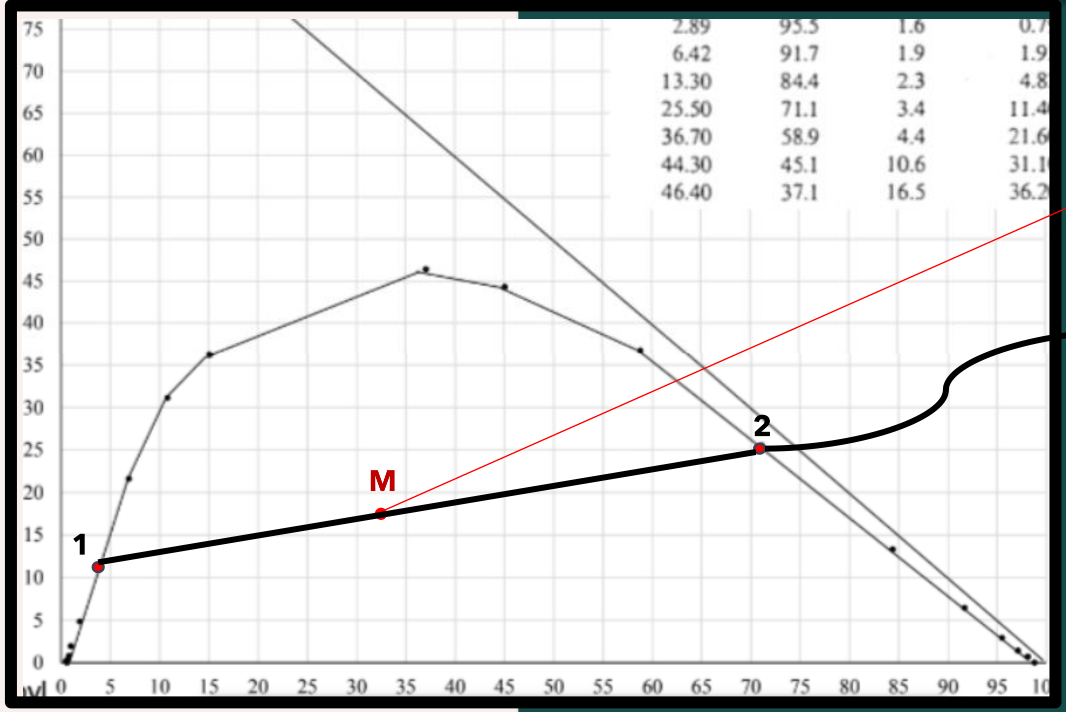
(A) Balance $200(0.1) = 0.118R + 0.043E$

R = 152 Kg , Extract = 48Kg



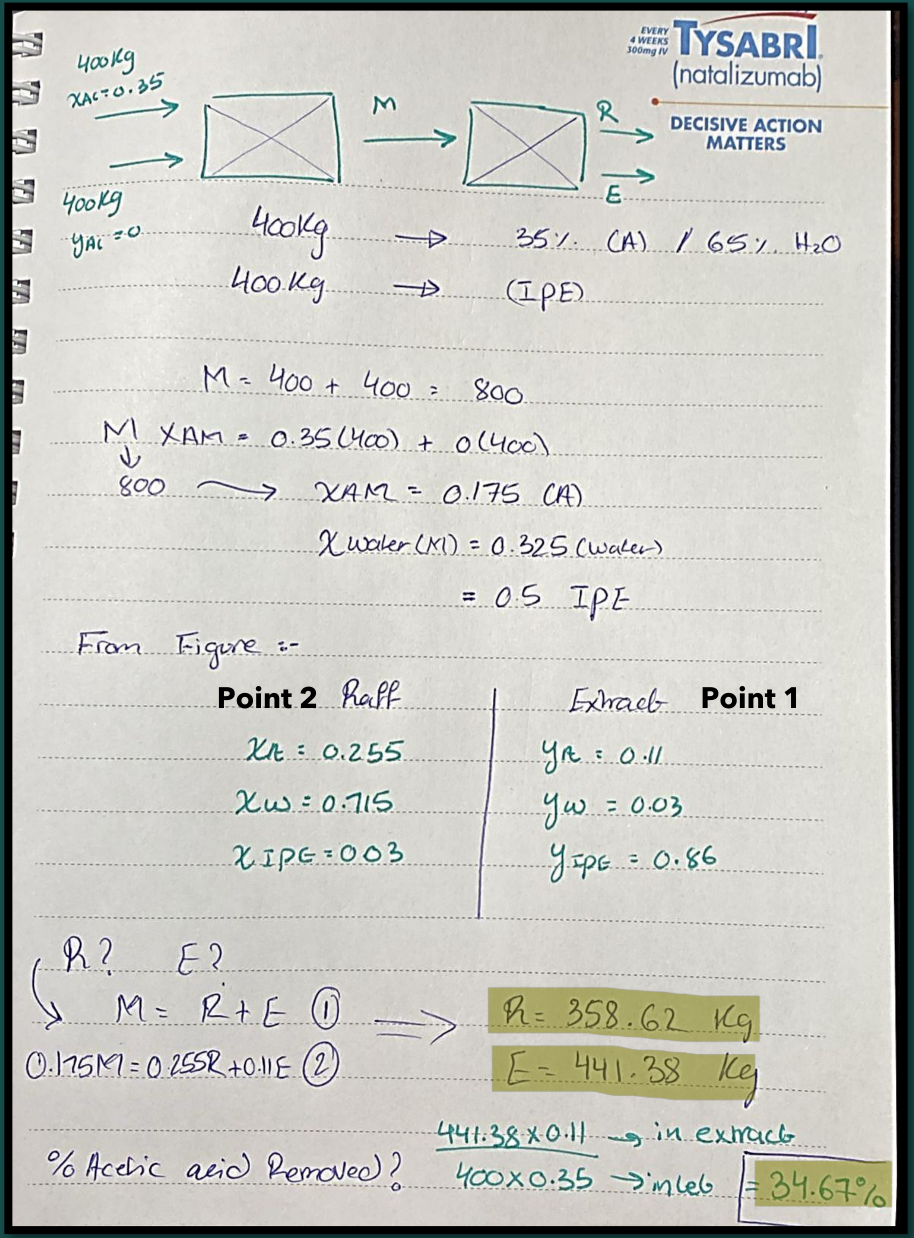
27.1-1. *Single-Stage Extraction.* A single-stage extraction is performed in which 400 kg of a solution containing 35 wt % acetic acid in water is contacted with 400 kg of pure isopropyl ether. Calculate the amounts and compositions of the extract and raffinate layers. Solve for the amounts both algebraically and by the lever-arm rule. What percent of the acetic acid is removed? Use equilibrium data from Appendix A.3.

Ans. $L_1 = 358$ kg, $x_{B1} = 0.715$, $x_{C1} = 0.03$, $V_1 = 442$ kg, $y_{A1} = 0.11$, $y_{C1} = 0.86$, 34.7% removed.



x(AC)= 0.11
x(H2O)= 0.03
x(IPE)= 0.86

Corresponding tie-line for this point.



400Kg → 35% (A) / 65% H_2O

400 Kg → (IPE)

$M = 400 + 400 = 800$

$\underset{\substack{\downarrow \\ 800}}{M}\, X_{AM} = 0.35(400) + 0(400)$

$\longrightarrow X_{AM} = 0.175$ (A)

$X_{water}(M) = 0.325$ (water)

$= 0.5$ IPE

From Figure :-

Point 2 Raff	Extract **Point 1**
$x_A = 0.255$	$y_A = 0.11$
$x_w = 0.715$	$y_w = 0.03$
$x_{IPE} = 0.03$	$y_{IPE} = 0.86$

R? E?

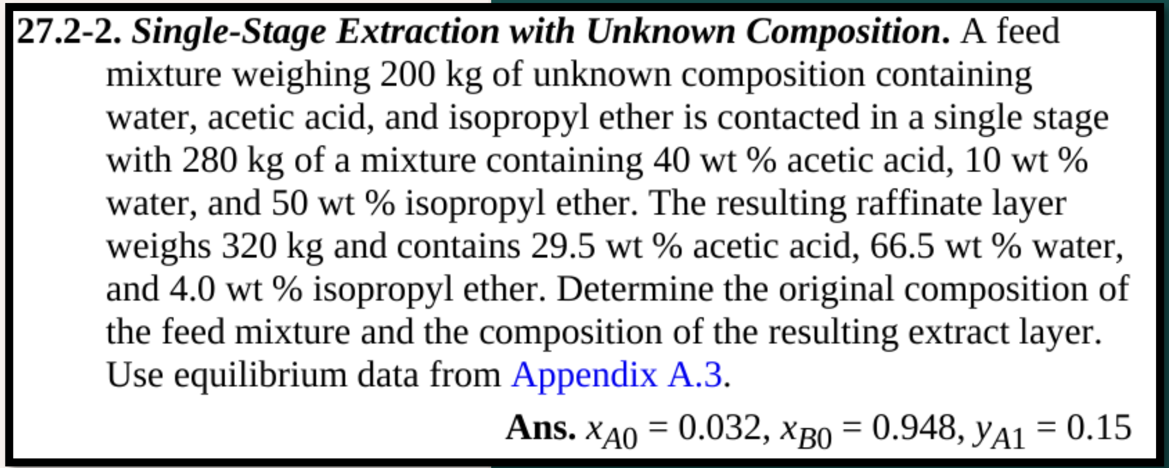
$M = R + E$ ①

$0.175M = 0.255R + 0.11E$ ②

$\Longrightarrow$ R = 358.62 Kg

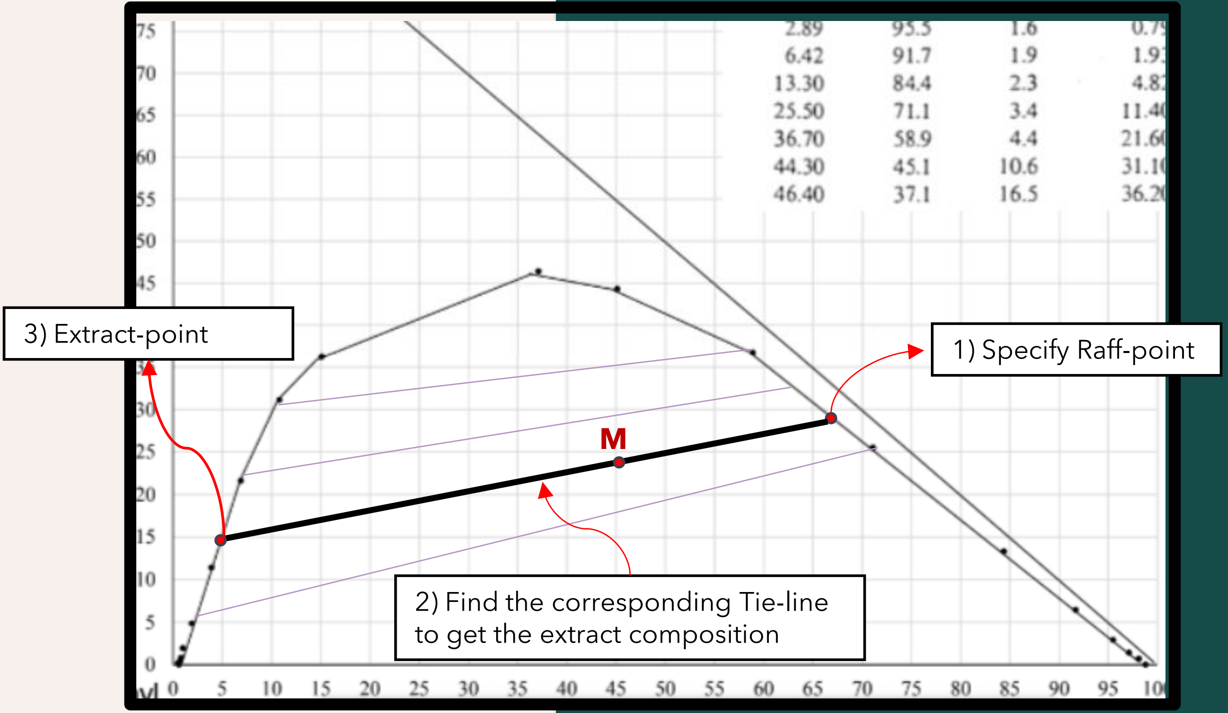
E = 441.38 Kg

% Acetic acid Removed? $\dfrac{441.38 \times 0.11 \rightarrow \text{in extract}}{400 \times 0.35 \rightarrow \text{inlet}}$ $\boxed{= 34.67\%}$



27.2-2. *Single-Stage Extraction with Unknown Composition.* A feed mixture weighing 200 kg of unknown composition containing water, acetic acid, and isopropyl ether is contacted in a single stage with 280 kg of a mixture containing 40 wt % acetic acid, 10 wt % water, and 50 wt % isopropyl ether. The resulting raffinate layer weighs 320 kg and contains 29.5 wt % acetic acid, 66.5 wt % water, and 4.0 wt % isopropyl ether. Determine the original composition of the feed mixture and the composition of the resulting extract layer. Use equilibrium data from Appendix A.3.

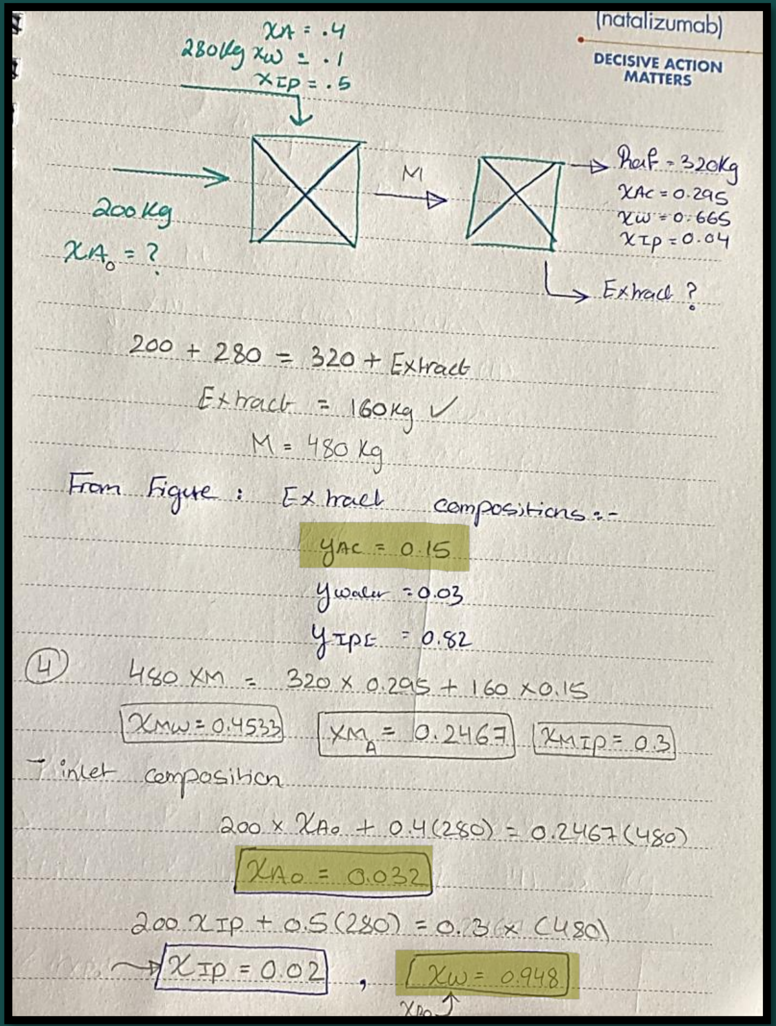
Ans. $x_{A0} = 0.032$, $x_{B0} = 0.948$, $y_{A1} = 0.15$



3) Extract-point

1) Specify Raff-point

2) Find the corresponding Tie-line to get the extract composition



$x_A = .4$
280 kg $x_w = .1$
$x_{IP} = .5$

200 kg
$x_{A_0} = ?$

M

Raf = 320 kg
$x_{Ac} = 0.295$
$x_w = 0.665$
$x_{IP} = 0.04$

Extract ?

200 + 280 = 320 + Extract

Extract = 160 kg ✓

M = 480 kg

From Figure : Extract compositions :-

$y_{AC} = 0.15$

$y_{water} = 0.03$

$y_{IPE} = 0.82$

④ $480 \, x_M = 320 \times 0.295 + 160 \times 0.15$

$\boxed{x_{Mw} = 0.4533}$ $\boxed{x_{M_A} = 0.2467}$ $\boxed{x_{MIP} = 0.3}$

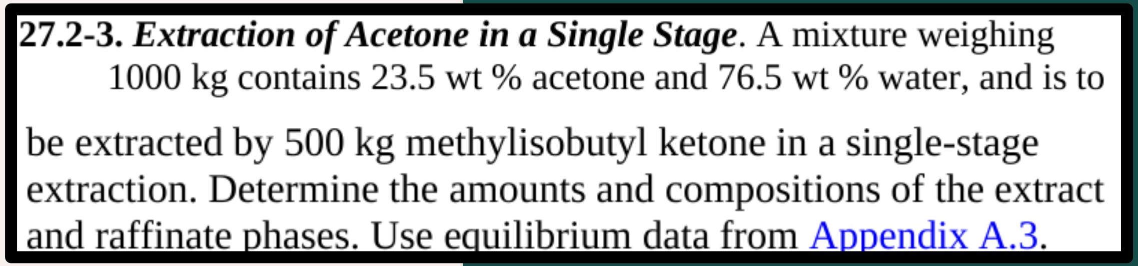
→ inlet composition

$200 \times x_{A_0} + 0.4(280) = 0.2467(480)$

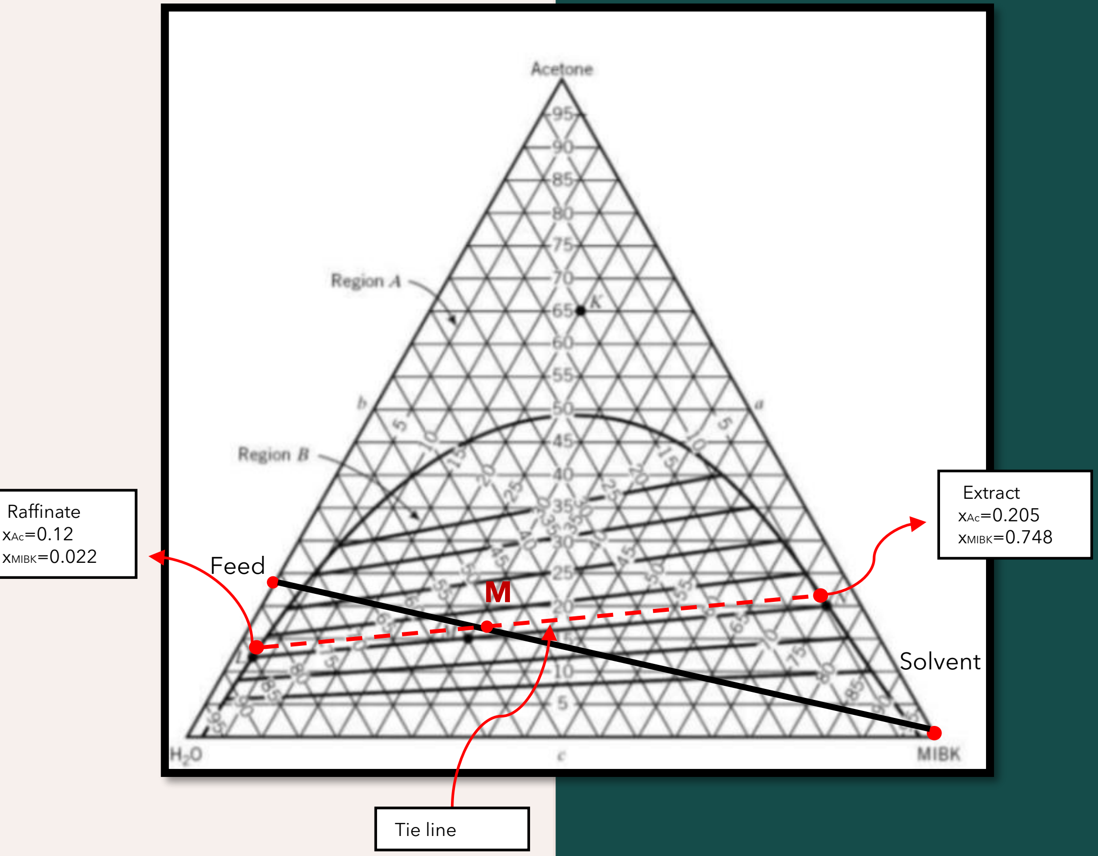
$\boxed{x_{A_0} = 0.032}$

$200 \, x_{IP} + 0.5(280) = 0.3 \times (480)$

→ $\boxed{x_{IP} = 0.02}$, $\boxed{x_w = 0.948}$
x_{B_0}



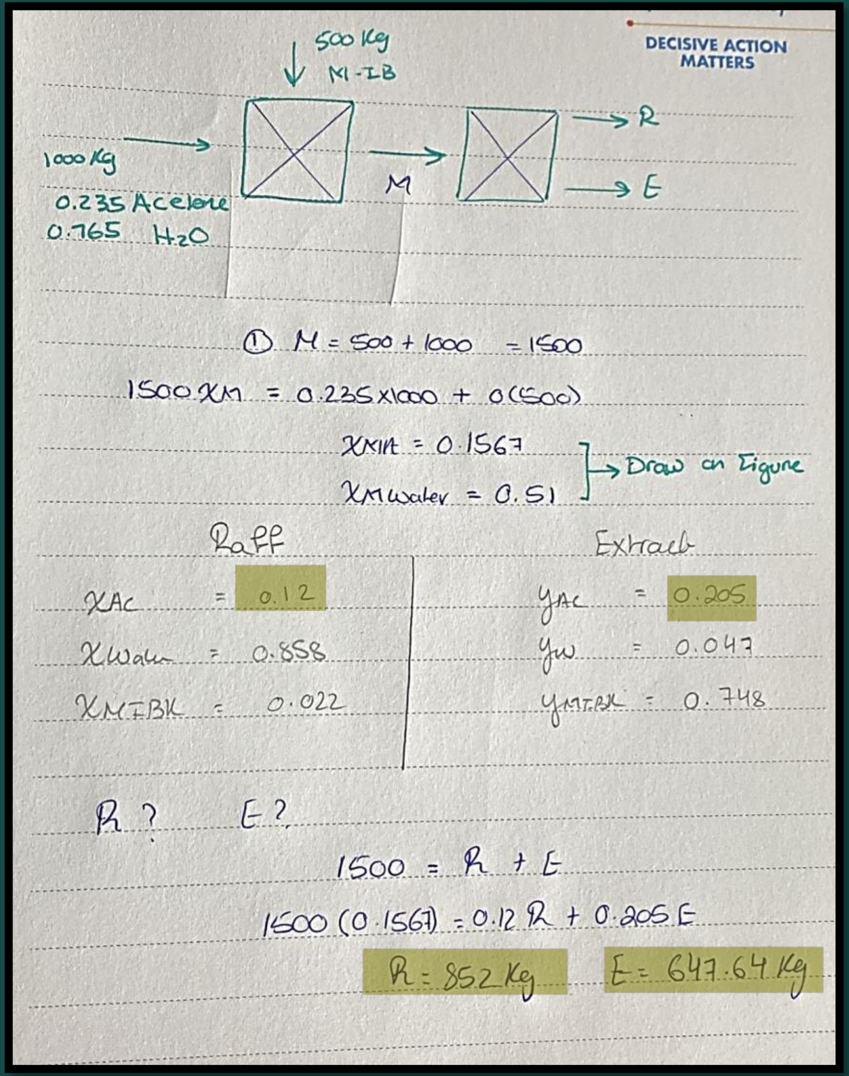
27.2-3. *Extraction of Acetone in a Single Stage*. A mixture weighing 1000 kg contains 23.5 wt % acetone and 76.5 wt % water, and is to be extracted by 500 kg methylisobutyl ketone in a single-stage extraction. Determine the amounts and compositions of the extract and raffinate phases. Use equilibrium data from Appendix A.3.



Raffinate
x_{Ac}=0.12
x_{MIBK}=0.022

Extract
x_{Ac}=0.205
x_{MIBK}=0.748

Tie line



① $M = 500 + 1000 = 1500$

$1500\, x_M = 0.235 \times 1000 + 0(500)$

$x_{M\,Ac} = 0.1567$

$x_{M\,water} = 0.51$

→ Draw on Figure

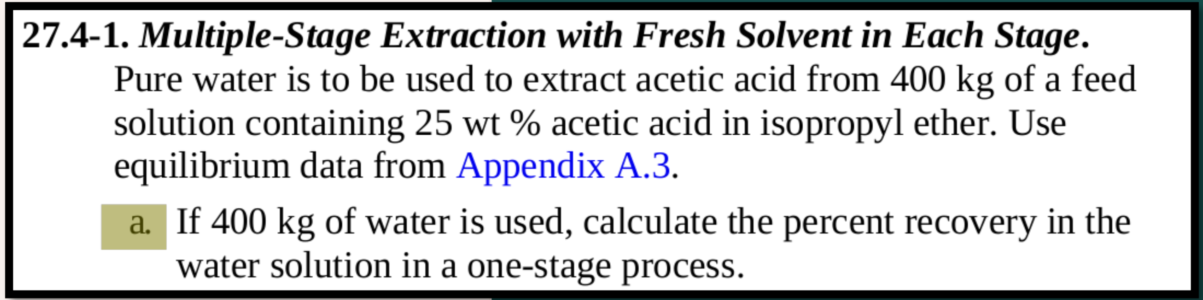
Raff		Extract	
x_{Ac}	= 0.12	y_{Ac}	= 0.205
x_{Water}	= 0.858	y_w	= 0.047
x_{MIBK}	= 0.022	y_{MIBK}	= 0.748

R ? E ?

$1500 = R + E$

$1500\,(0.1567) = 0.12\,R + 0.205\,E$

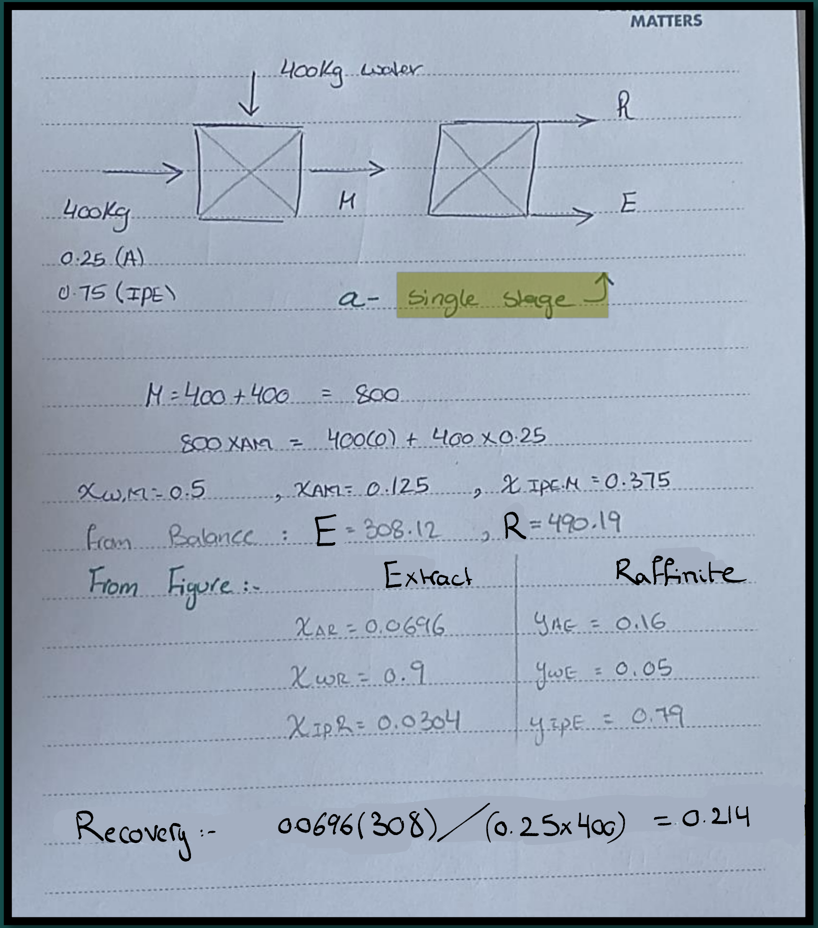
R = 852 Kg E = 647.64 Kg



27.4-1. ***Multiple-Stage Extraction with Fresh Solvent in Each Stage.*** Pure water is to be used to extract acetic acid from 400 kg of a feed solution containing 25 wt % acetic acid in isopropyl ether. Use equilibrium data from Appendix A.3.

a. If 400 kg of water is used, calculate the percent recovery in the water solution in a one-stage process.

6.42	91.7	1.9	1.9
13.30	84.4	2.3	4.8
25.50	71.1	3.4	11.4
36.70	58.9	4.4	21.6
44.30	45.1	10.6	31.1
46.40	37.1	16.5	36.2



0.25 (A)
0.75 (IPE)

a- single stage

$M = 400 + 400 = 800$

$800\, x_{AM} = 400(0) + 400 \times 0.25$

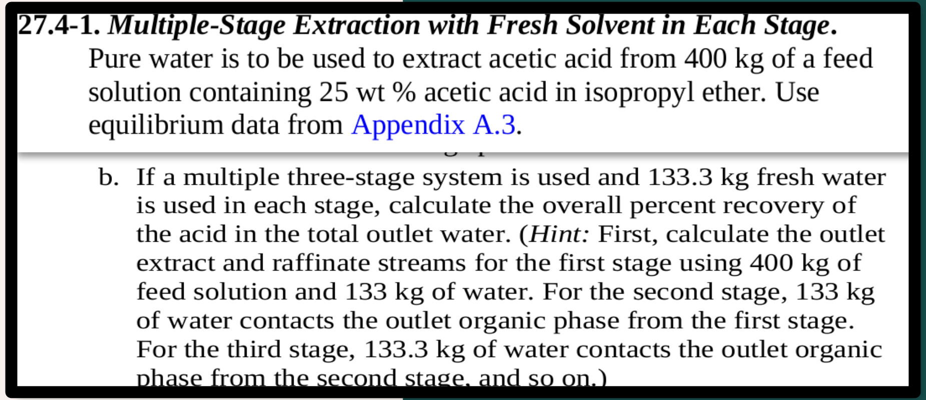
$x_{w,M} = 0.5$, $x_{AM} = 0.125$, $x_{IPE.M} = 0.375$

From Balance : $E = 308.12$, $R = 490.19$

From Figure :-

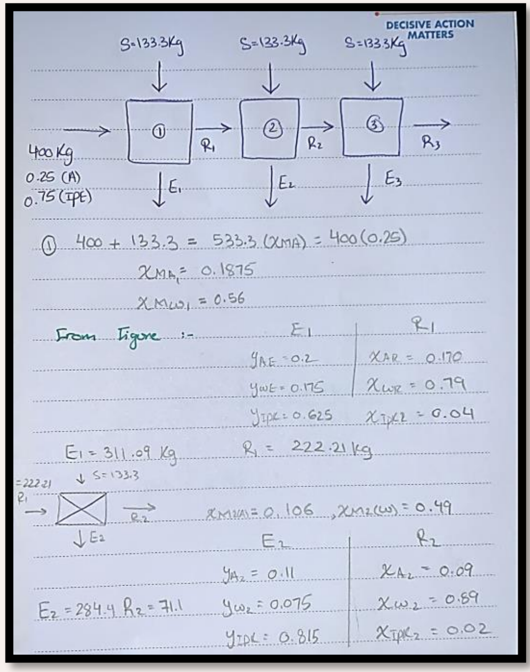
	Extract	Raffinite
	$x_{AR} = 0.0696$	$y_{AE} = 0.16$
	$x_{wR} = 0.9$	$y_{wE} = 0.05$
	$x_{IPR} = 0.0304$	$y_{IPE} = 0.79$

Recovery :- $0.0696(308) / (0.25 \times 400) = 0.214$



27.4-1. ***Multiple-Stage Extraction with Fresh Solvent in Each Stage.*** Pure water is to be used to extract acetic acid from 400 kg of a feed solution containing 25 wt % acetic acid in isopropyl ether. Use equilibrium data from Appendix A.3.

b. If a multiple three-stage system is used and 133.3 kg fresh water is used in each stage, calculate the overall percent recovery of the acid in the total outlet water. (*Hint:* First, calculate the outlet extract and raffinate streams for the first stage using 400 kg of feed solution and 133 kg of water. For the second stage, 133 kg of water contacts the outlet organic phase from the first stage. For the third stage, 133.3 kg of water contacts the outlet organic phase from the second stage, and so on.)



① $400 + 133.3 = 533.3\,(x_{MA}) = 400(0.25)$

$x_{MA_1} = 0.1875$

$x_{MW_1} = 0.56$

From Figure :-

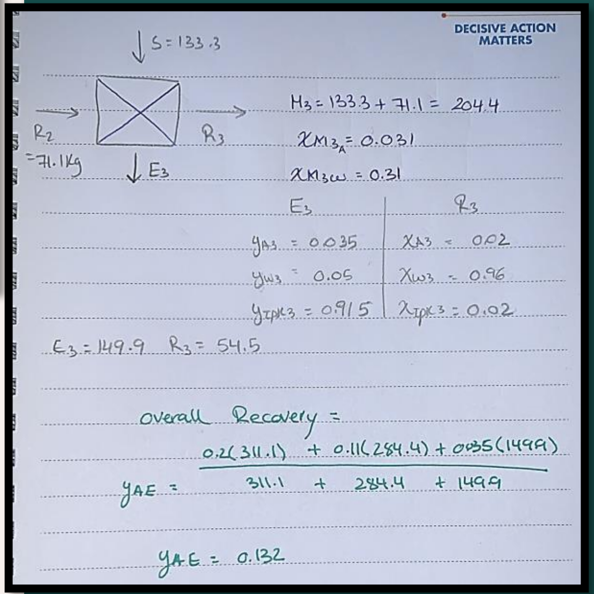
E_1	R_1
$y_{AE} = 0.2$	$x_{AR} = 0.170$
$y_{WE} = 0.175$	$x_{WR} = 0.79$
$y_{IPE} = 0.625$	$x_{IPE2} = 0.04$

$E_1 = 311.09$ Kg $\quad R_1 = 222.21$ kg

$x_{M2(A)} = 0.106$, $x_{M2(W)} = 0.49$

E_2	R_2
$y_{A_2} = 0.11$	$x_{A_2} = 0.09$
$y_{W_2} = 0.075$	$x_{W_2} = 0.89$
$y_{IPE} = 0.815$	$x_{IPE_2} = 0.02$

$E_2 = 284.4 \quad R_2 = 71.1$



$M_3 = 133.3 + 71.1 = 204.4$

$x_{M3_A} = 0.031$

$x_{M3W} = 0.31$

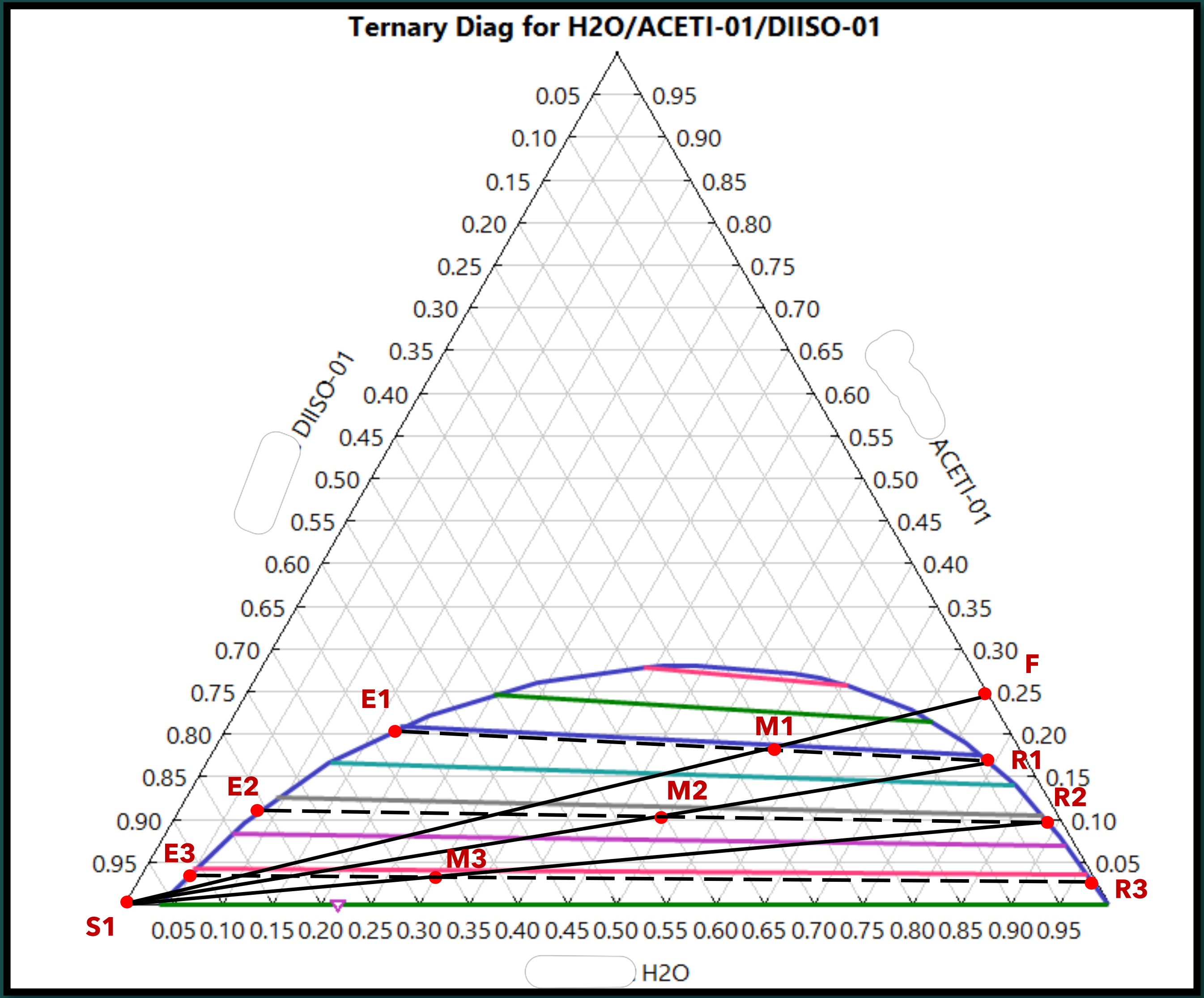
E_3	R_3
$y_{A3} = 0.035$	$x_{A3} = 0.02$
$y_{W3} = 0.05$	$x_{W3} = 0.96$
$y_{IPE3} = 0.915$	$x_{IPE3} = 0.02$

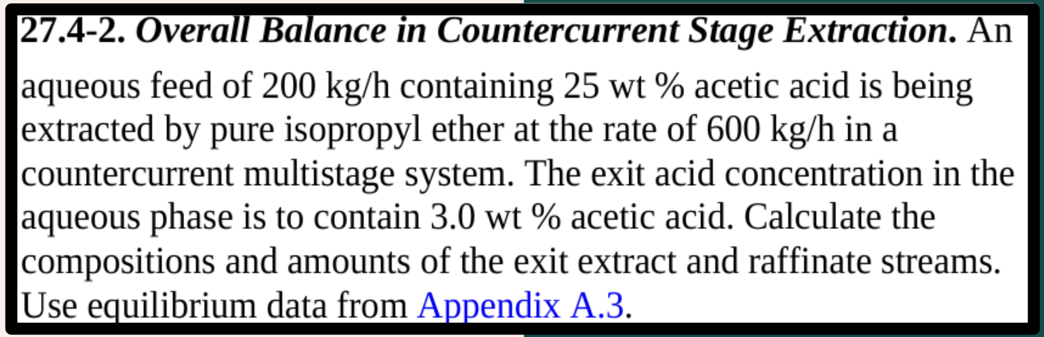
$E_3 = 149.9 \quad R_3 = 54.5$

overall Recovery =

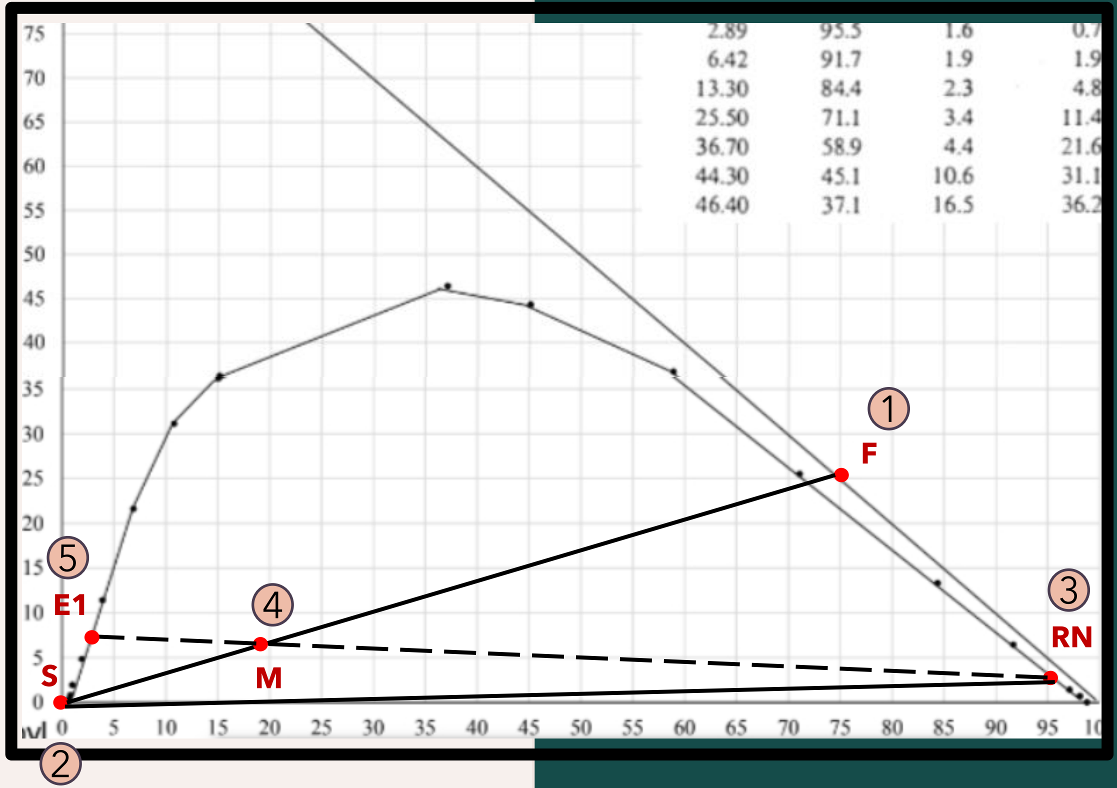
$$y_{AE} = \frac{0.2(311.1) + 0.11(284.4) + 0.035(149.9)}{311.1 + 284.4 + 149.9}$$

$y_{AE} = 0.132$

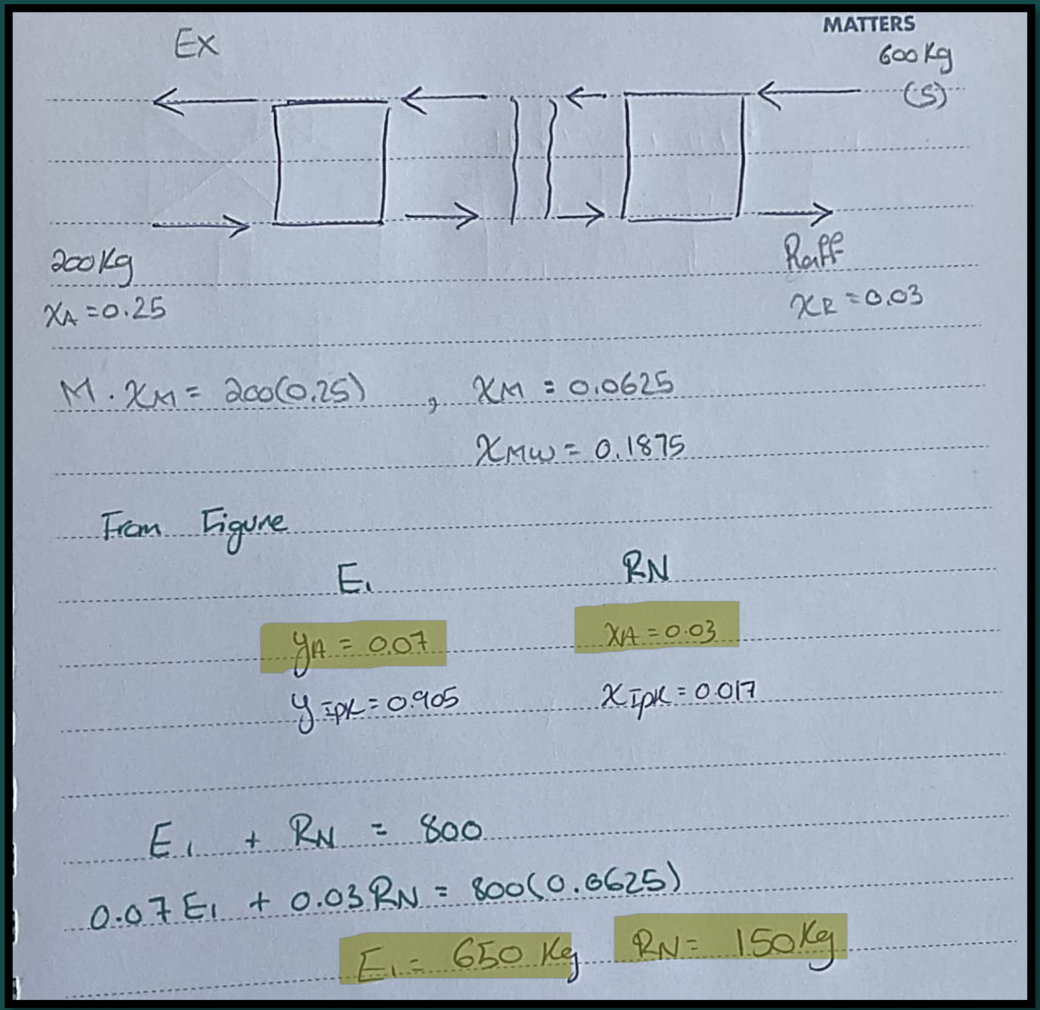




27.4-2. *Overall Balance in Countercurrent Stage Extraction.* An aqueous feed of 200 kg/h containing 25 wt % acetic acid is being extracted by pure isopropyl ether at the rate of 600 kg/h in a countercurrent multistage system. The exit acid concentration in the aqueous phase is to contain 3.0 wt % acetic acid. Calculate the compositions and amounts of the exit extract and raffinate streams. Use equilibrium data from Appendix A.3.



2.89	95.5	1.6	0.7
6.42	91.7	1.9	1.9
13.30	84.4	2.3	4.8
25.50	71.1	3.4	11.4
36.70	58.9	4.4	21.6
44.30	45.1	10.6	31.1
46.40	37.1	16.5	36.2



MATTERS

Ex

600 Kg
(S)

200 Kg
$x_A = 0.25$

Raff
$x_R = 0.03$

$M \cdot x_M = 200(0.25)$, $x_M = 0.0625$

$x_{MW} = 0.1875$

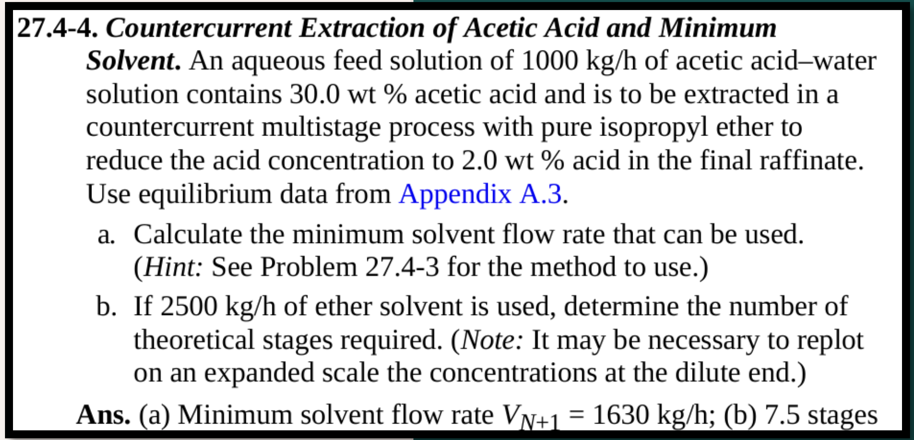
From Figure

E_1	RN
$y_A = 0.07$	$x_A = 0.03$
$y_{IPK} = 0.905$	$x_{IPK} = 0.017$

$E_1 + R_N = 800$

$0.07 E_1 + 0.03 R_N = 800(0.0625)$

$E_1 = 650$ Kg $R_N = 150$ Kg

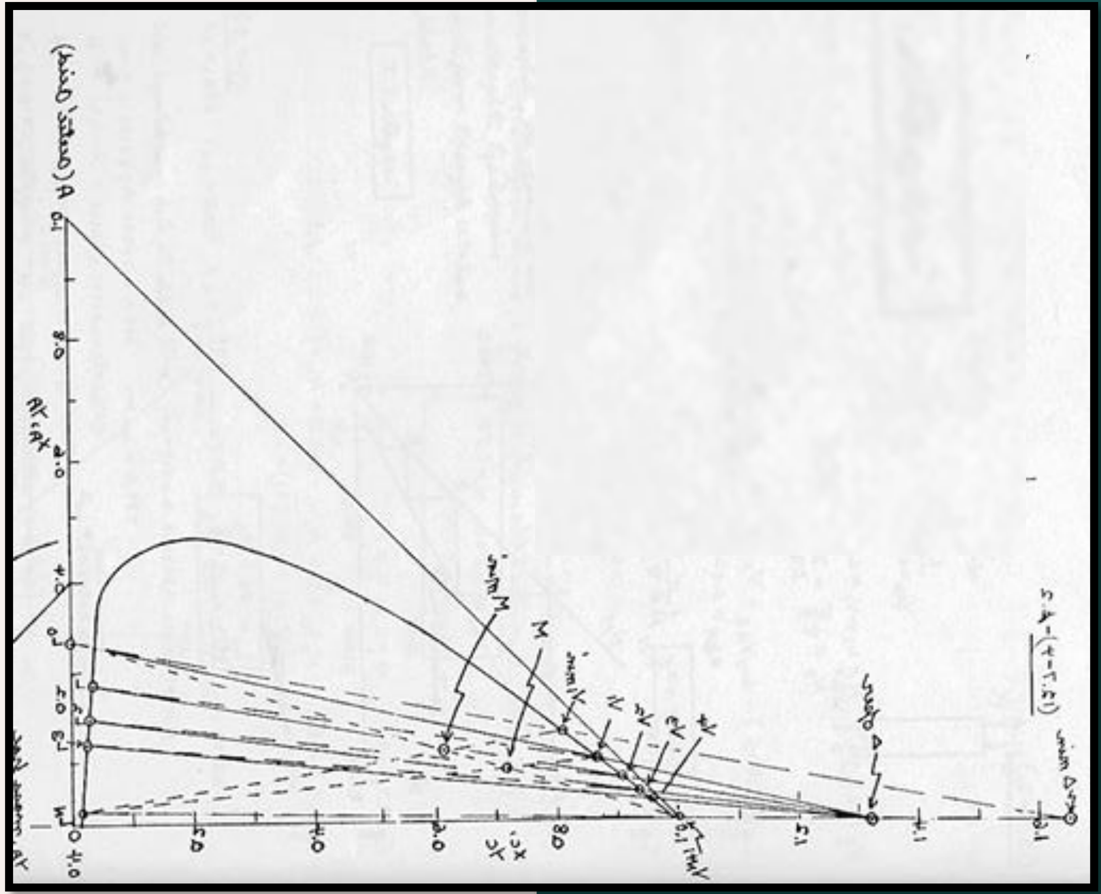


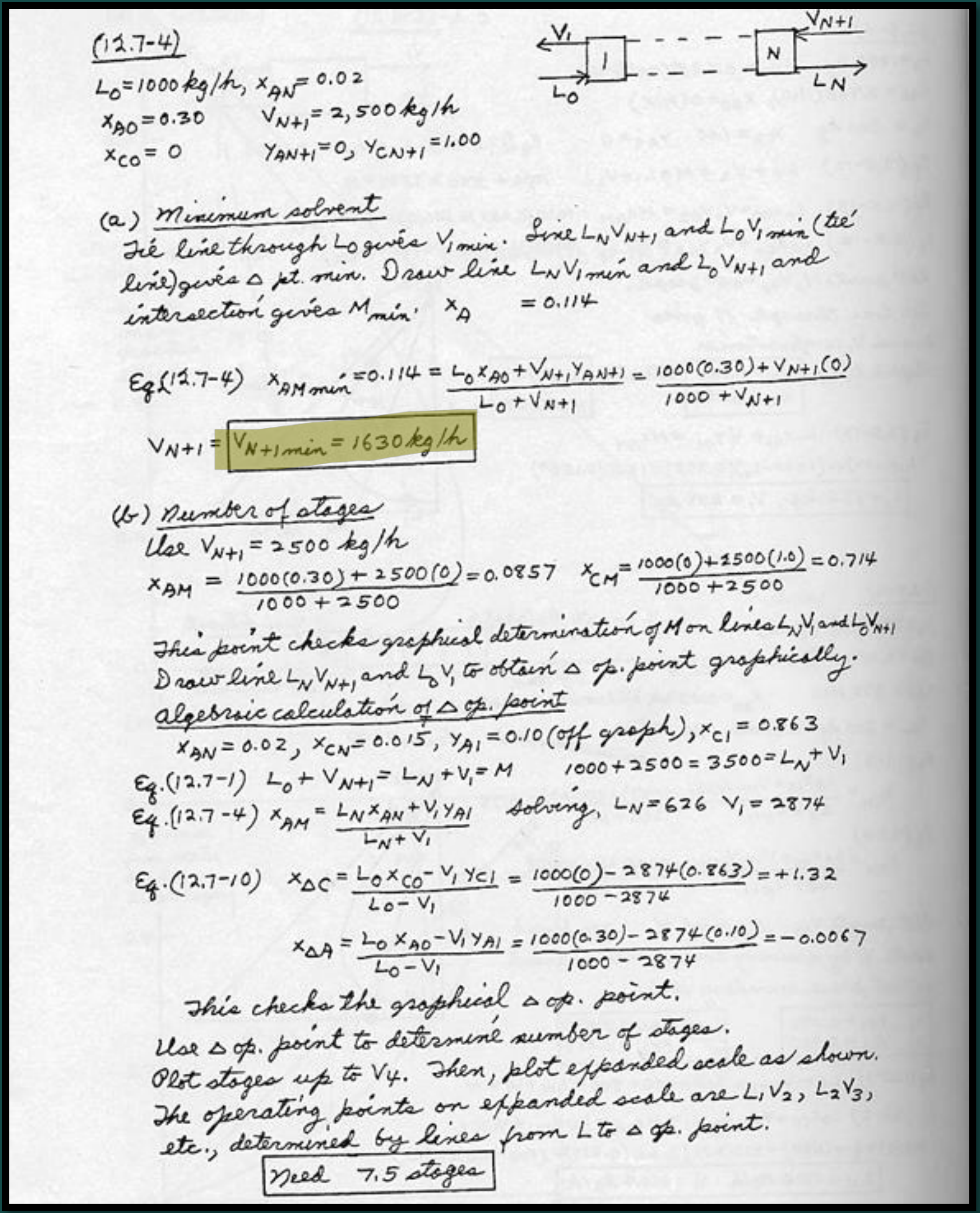
27.4-4. *Countercurrent Extraction of Acetic Acid and Minimum Solvent.* An aqueous feed solution of 1000 kg/h of acetic acid–water solution contains 30.0 wt % acetic acid and is to be extracted in a countercurrent multistage process with pure isopropyl ether to reduce the acid concentration to 2.0 wt % acid in the final raffinate. Use equilibrium data from Appendix A.3.

a. Calculate the minimum solvent flow rate that can be used. (*Hint:* See Problem 27.4-3 for the method to use.)

b. If 2500 kg/h of ether solvent is used, determine the number of theoretical stages required. (*Note:* It may be necessary to replot on an expanded scale the concentrations at the dilute end.)

Ans. (a) Minimum solvent flow rate V_{N+1} = 1630 kg/h; (b) 7.5 stages





(12.7-4)

$L_0 = 1000$ kg/h, $x_{AN} = 0.02$

$x_{A0} = 0.30$ $\quad V_{N+1} = 2{,}500$ kg/h

$x_{C0} = 0$ $\quad y_{AN+1} = 0,\ y_{CN+1} = 1.00$

(a) Minimum solvent

Tie line through L_0 gives V_1 min. Line L_NV_{N+1} and L_0V_1 min (tie line) gives Δ pt. min. Draw line L_NV_1 min and L_0V_{N+1} and intersection gives M_{min}: $x_A = 0.114$

$$\text{Eq.(12.7-4)}\quad x_{AM\,min} = 0.114 = \frac{L_0 x_{A0} + V_{N+1} y_{AN+1}}{L_0 + V_{N+1}} = \frac{1000(0.30) + V_{N+1}(0)}{1000 + V_{N+1}}$$

$$V_{N+1} = \boxed{V_{N+1\,min} = 1630\ \text{kg/h}}$$

(b) Number of stages

Use $V_{N+1} = 2500$ kg/h

$$x_{AM} = \frac{1000(0.30) + 2500(0)}{1000 + 2500} = 0.0857 \qquad x_{CM} = \frac{1000(0) + 2500(1.0)}{1000 + 2500} = 0.714$$

This point checks graphical determination of M on lines L_NV_1 and L_0V_{N+1}

Draw line L_NV_{N+1} and L_0V_1 to obtain Δ op. point graphically.

algebraic calculation of Δ op. point

$x_{AN} = 0.02$, $x_{CN} = 0.015$, $y_{A1} = 0.10$ (off graph), $x_{C1} = 0.863$

$$\text{Eq.(12.7-1)}\quad L_0 + V_{N+1} = L_N + V_1 = M \qquad 1000 + 2500 = 3500 = L_N + V_1$$

$$\text{Eq.(12.7-4)}\quad x_{AM} = \frac{L_N x_{AN} + V_1 y_{A1}}{L_N + V_1} \qquad \text{Solving, } L_N = 626 \quad V_1 = 2874$$

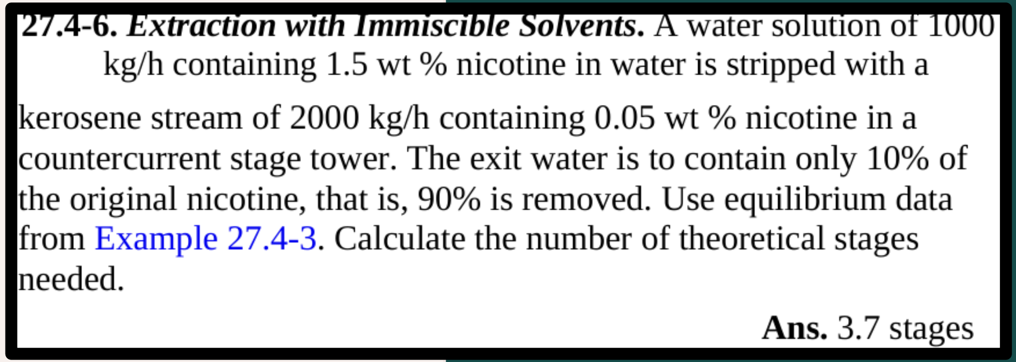
$$\text{Eq.(12.7-10)}\quad x_{\Delta C} = \frac{L_0 x_{C0} - V_1 y_{C1}}{L_0 - V_1} = \frac{1000(0) - 2874(0.863)}{1000 - 2874} = +1.32$$

$$x_{\Delta A} = \frac{L_0 x_{A0} - V_1 y_{A1}}{L_0 - V_1} = \frac{1000(0.30) - 2874(0.10)}{1000 - 2874} = -0.0067$$

This checks the graphical Δ op. point.

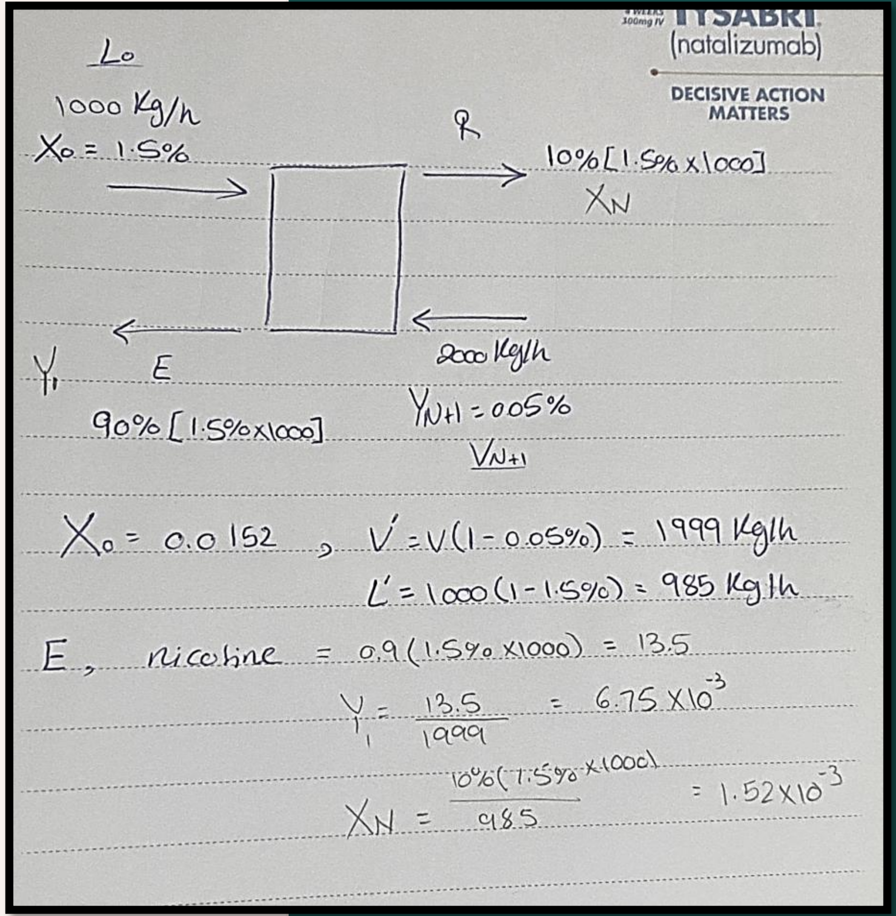
Use Δ op. point to determine number of stages. Plot stages up to V_4. Then, plot expanded scale as shown. The operating points on expanded scale are L_1V_2, L_2V_3, etc., determined by lines from L to Δ op. point.

$\boxed{\text{Need 7.5 stages}}$



27.4-6. *Extraction with Immiscible Solvents.* A water solution of 1000 kg/h containing 1.5 wt % nicotine in water is stripped with a kerosene stream of 2000 kg/h containing 0.05 wt % nicotine in a countercurrent stage tower. The exit water is to contain only 10% of the original nicotine, that is, 90% is removed. Use equilibrium data from Example 27.4-3. Calculate the number of theoretical stages needed.

Ans. 3.7 stages



L_o
1000 Kg/h
$X_o = 1.5\%$

R
$10\%[1.5\% \times 1000]$
X_N

Y_1 E
$90\%[1.5\% \times 1000]$

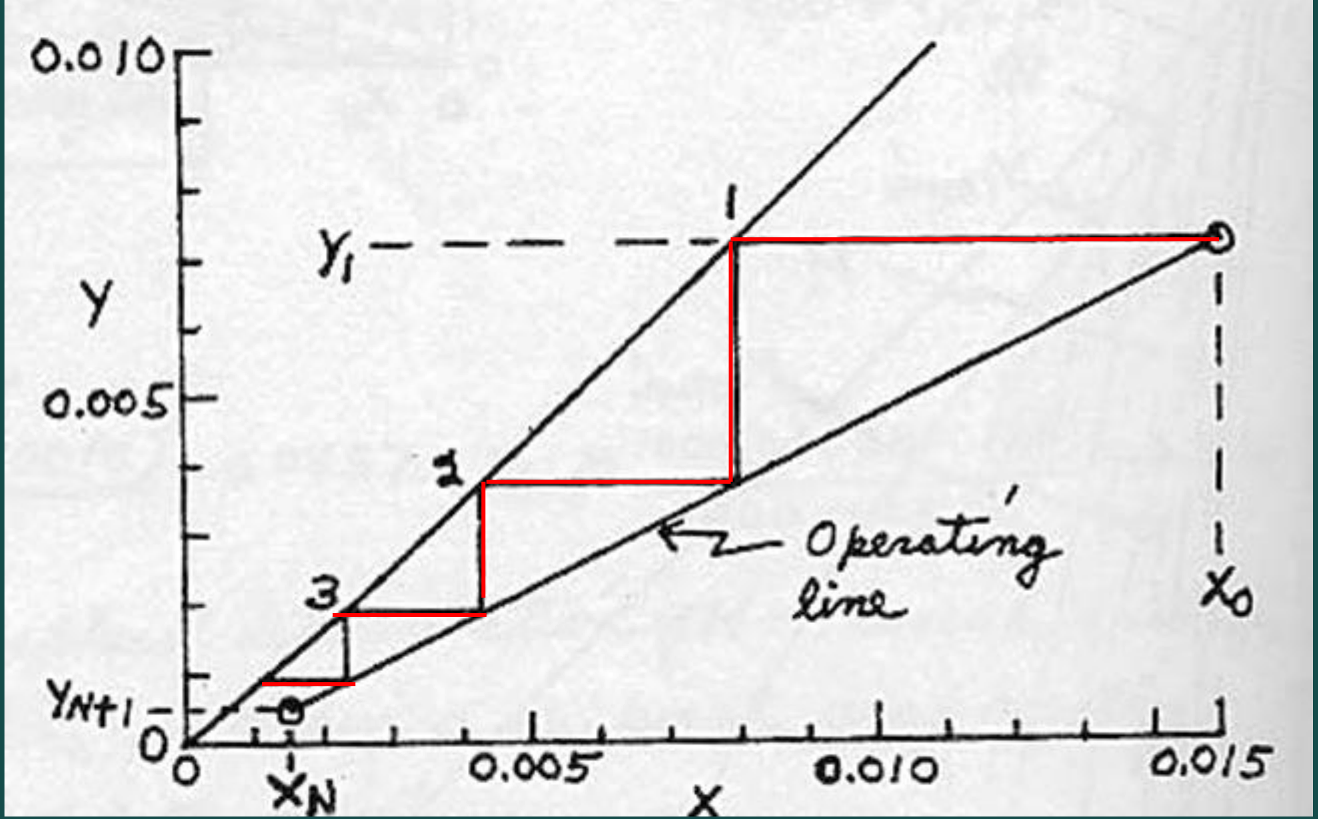
2000 Kg/h
$Y_{N+1} = 0.05\%$
V_{N+1}

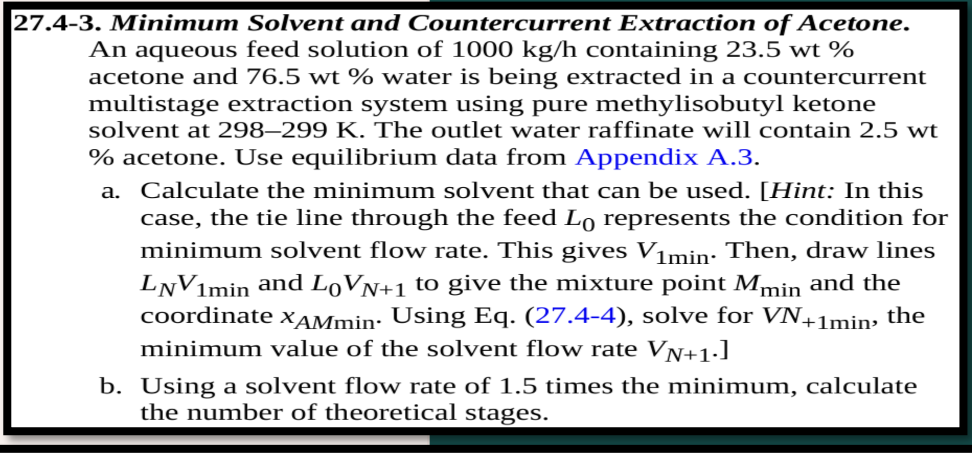
$X_o = 0.0152$, $V' = V(1 - 0.05\%) = 1999$ Kg/h
$L' = 1000(1 - 1.5\%) = 985$ Kg/h

E, nicotine $= 0.9(1.5\% \times 1000) = 13.5$

$Y_1 = \frac{13.5}{1999} = 6.75 \times 10^{-3}$

$X_N = \frac{10\%(1.5\% \times 1000)}{985} = 1.52 \times 10^{-3}$

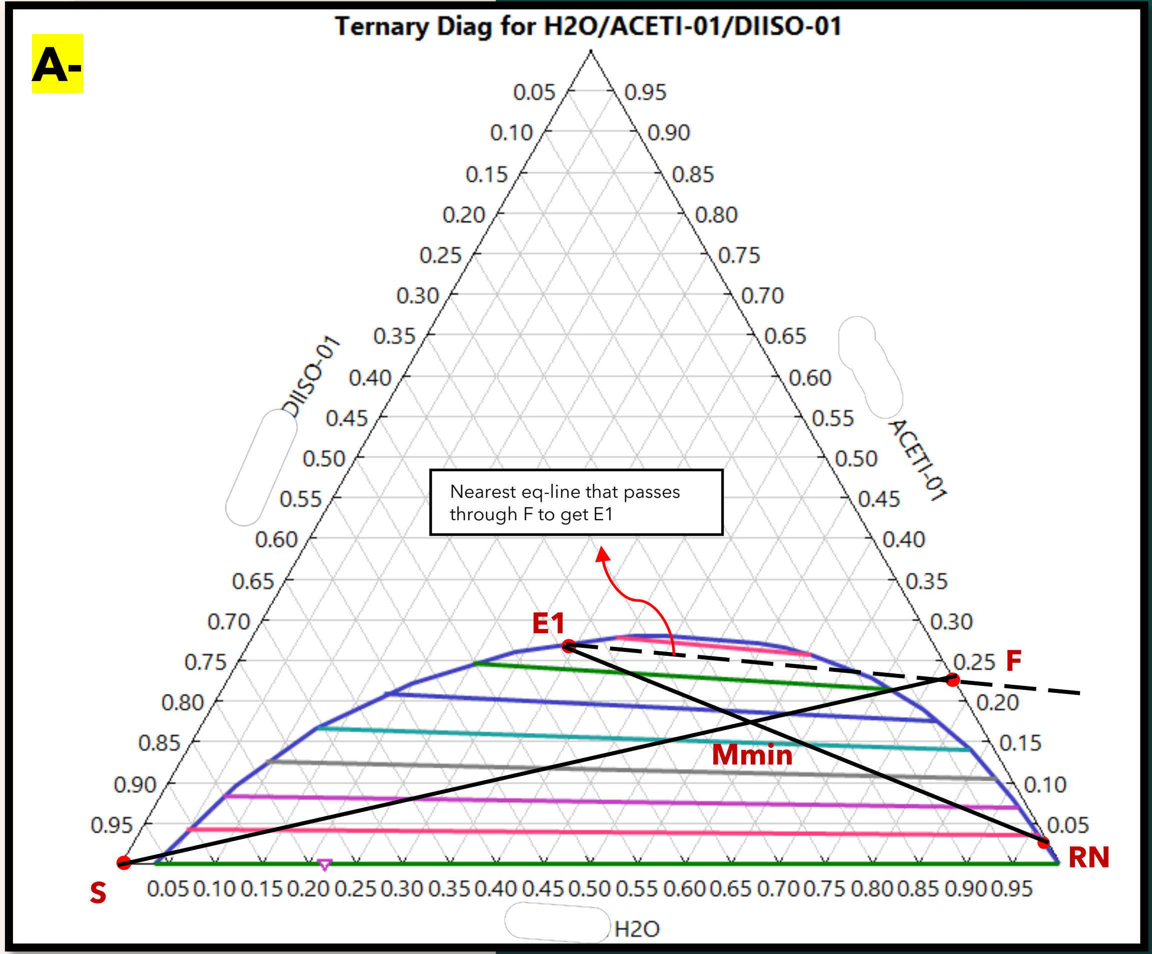




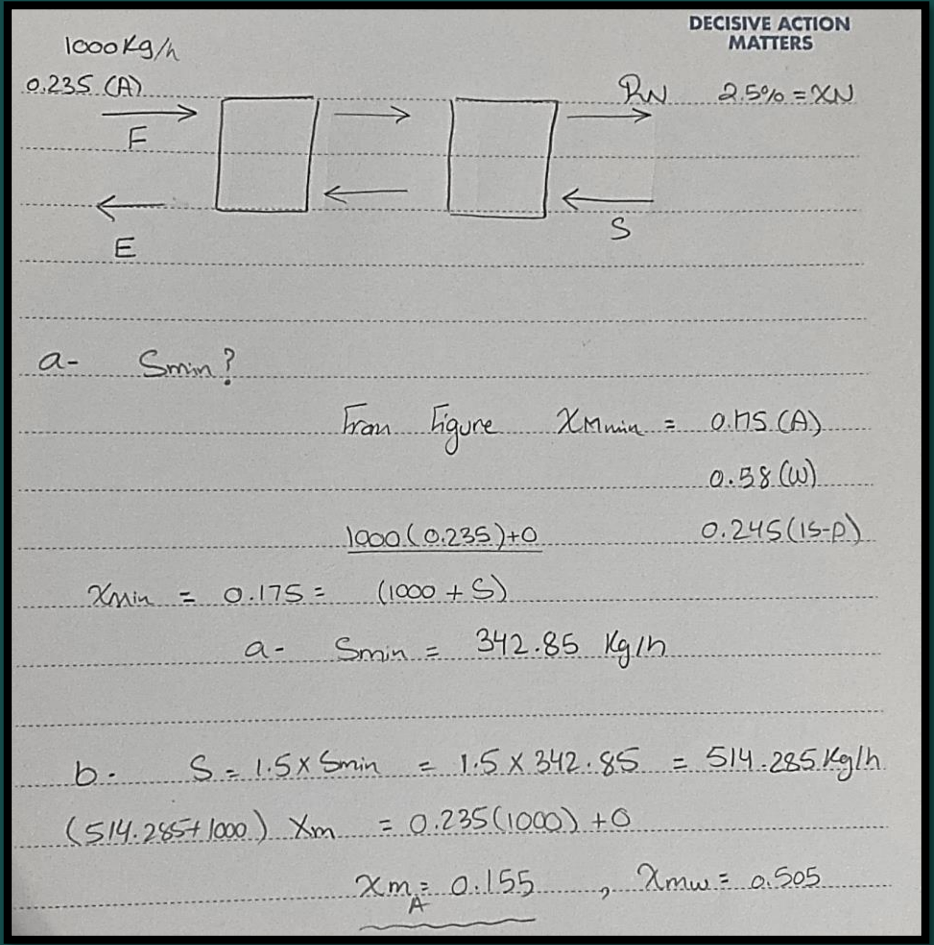
27.4-3. *Minimum Solvent and Countercurrent Extraction of Acetone.* An aqueous feed solution of 1000 kg/h containing 23.5 wt % acetone and 76.5 wt % water is being extracted in a countercurrent multistage extraction system using pure methylisobutyl ketone solvent at 298–299 K. The outlet water raffinate will contain 2.5 wt % acetone. Use equilibrium data from Appendix A.3.

a. Calculate the minimum solvent that can be used. [*Hint:* In this case, the tie line through the feed L_0 represents the condition for minimum solvent flow rate. This gives $V_{1\min}$. Then, draw lines $L_N V_{1\min}$ and $L_0 V_{N+1}$ to give the mixture point $M_{\min}$ and the coordinate $x_{AM\min}$. Using Eq. (27.4-4), solve for $VN_{+1\min}$, the minimum value of the solvent flow rate V_{N+1}.]

b. Using a solvent flow rate of 1.5 times the minimum, calculate the number of theoretical stages.



Nearest eq-line that passes through F to get E1



DECISIVE ACTION MATTERS

1000 Kg/h

0.235 (A)

F

E

R_N 2.5% = X_N

S

a- S_{min}?

From Figure X_{Mmin} = 0.175 (A)

0.58 (W)

0.245 (IS-P)

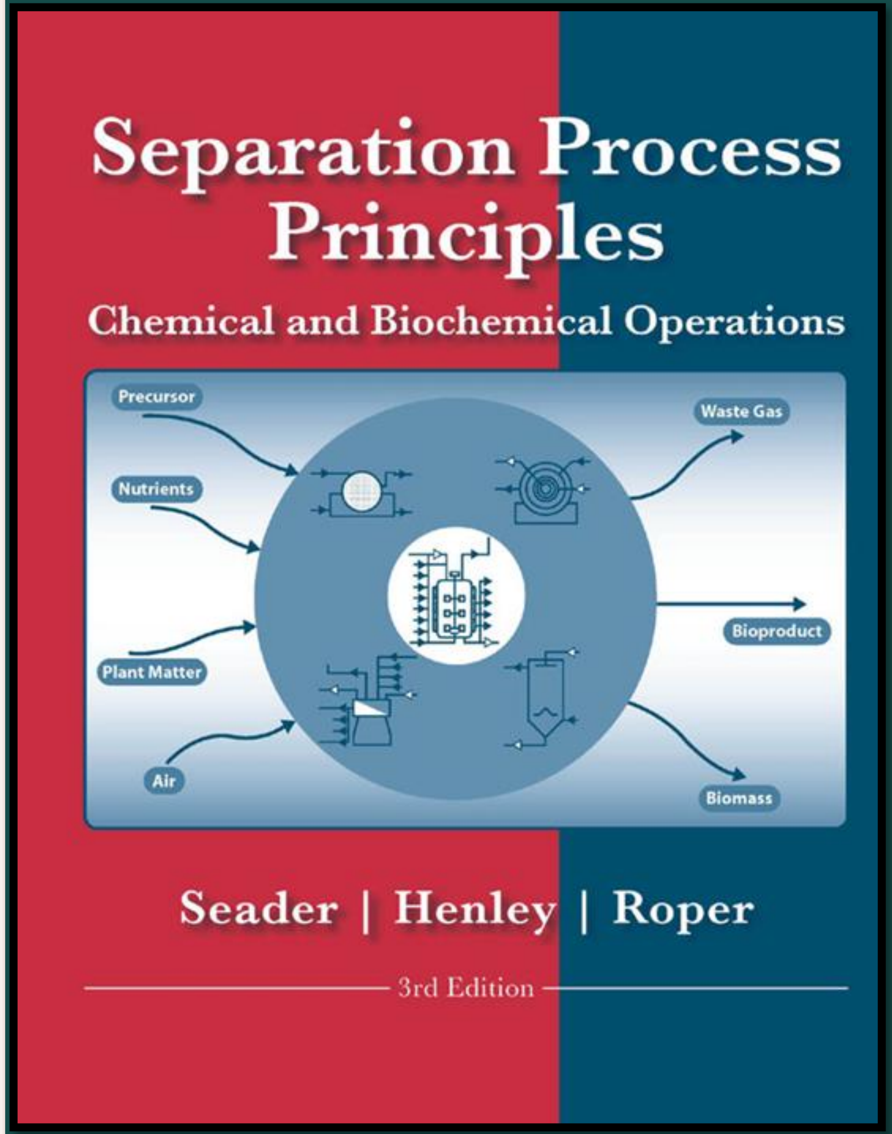
$X_{min} = 0.175 = \dfrac{1000(0.235)+0}{(1000+S)}$

a- S_{min} = 342.85 Kg/h

b. $S = 1.5 \times S_{min} = 1.5 \times 342.85 = 514.285$ Kg/h

$(514.285 + 1000)\ X_m = 0.235(1000) + 0$

$\underline{x_{m_A} = 0.155}$, $x_{mw} = 0.505$



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Ch8. Liquid–Liquid Extraction with Ternary Systems

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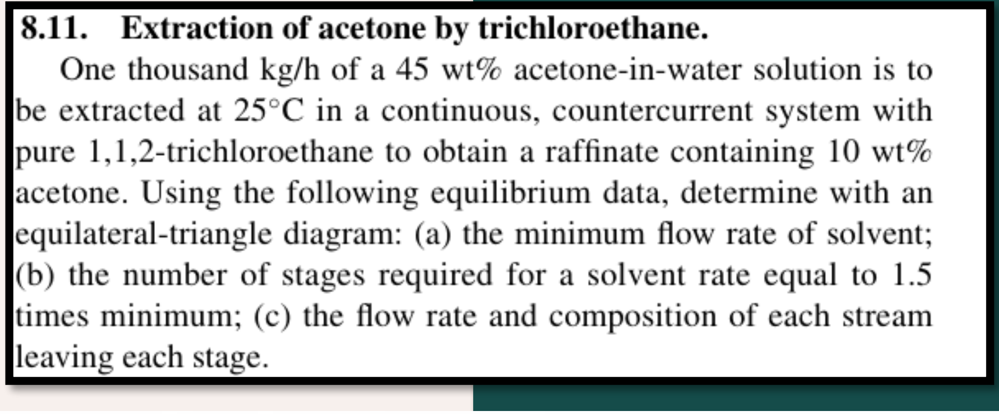
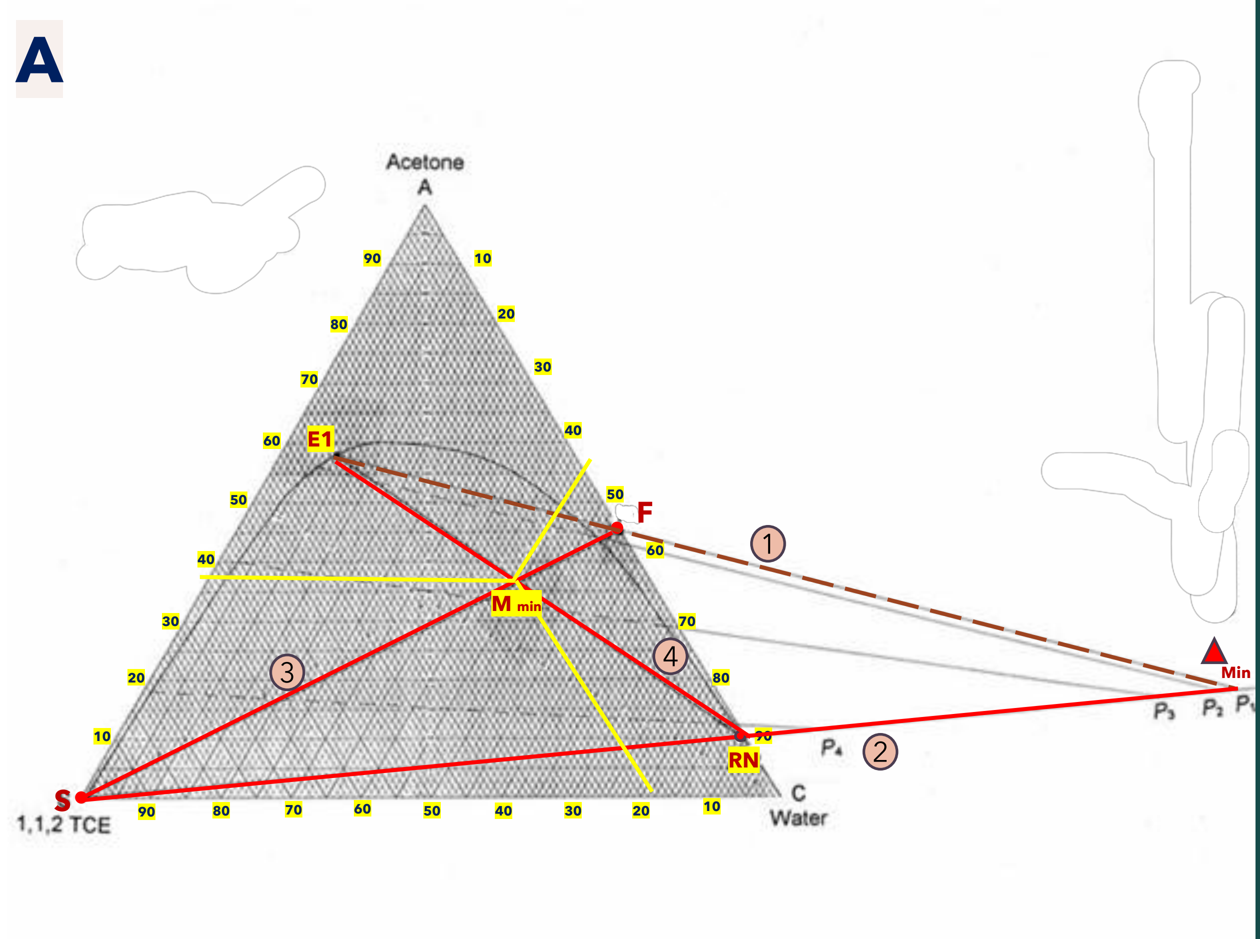
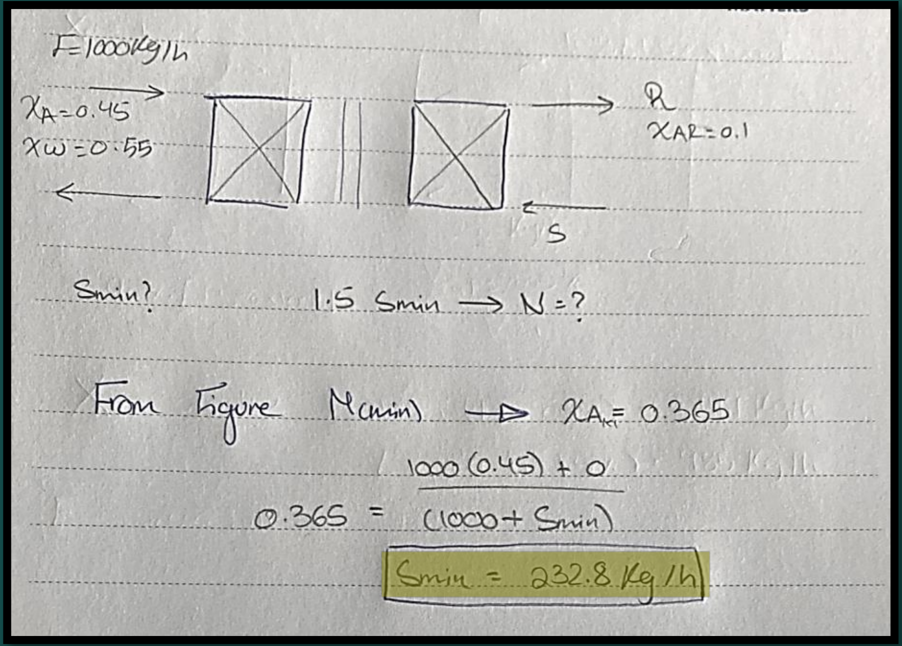


8.11. Extraction of acetone by trichloroethane.

One thousand kg/h of a 45 wt% acetone-in-water solution is to be extracted at 25°C in a continuous, countercurrent system with pure 1,1,2-trichloroethane to obtain a raffinate containing 10 wt% acetone. Using the following equilibrium data, determine with an equilateral-triangle diagram: (a) the minimum flow rate of solvent; (b) the number of stages required for a solvent rate equal to 1.5 times minimum; (c) the flow rate and composition of each stream leaving each stage.





$F = 1000\ kg/h$

$x_A = 0.45$

$x_w = 0.55$

R

$x_{AR} = 0.1$

S

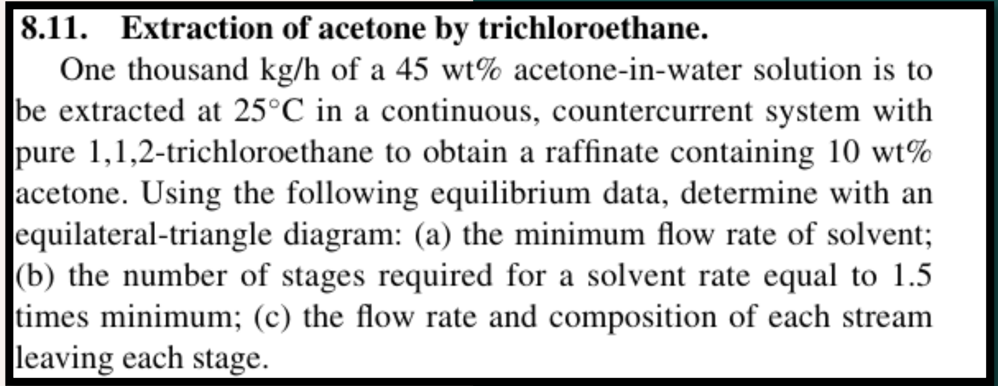
S_{min}?

$1.5\ S_{min} \rightarrow N = ?$

From Figure $M_{(min)} \longrightarrow x_{A_{M}} = 0.365$

$$0.365 = \frac{1000(0.45) + 0}{(1000 + S_{min})}$$

$$\boxed{S_{min} = 232.8\ Kg/h}$$



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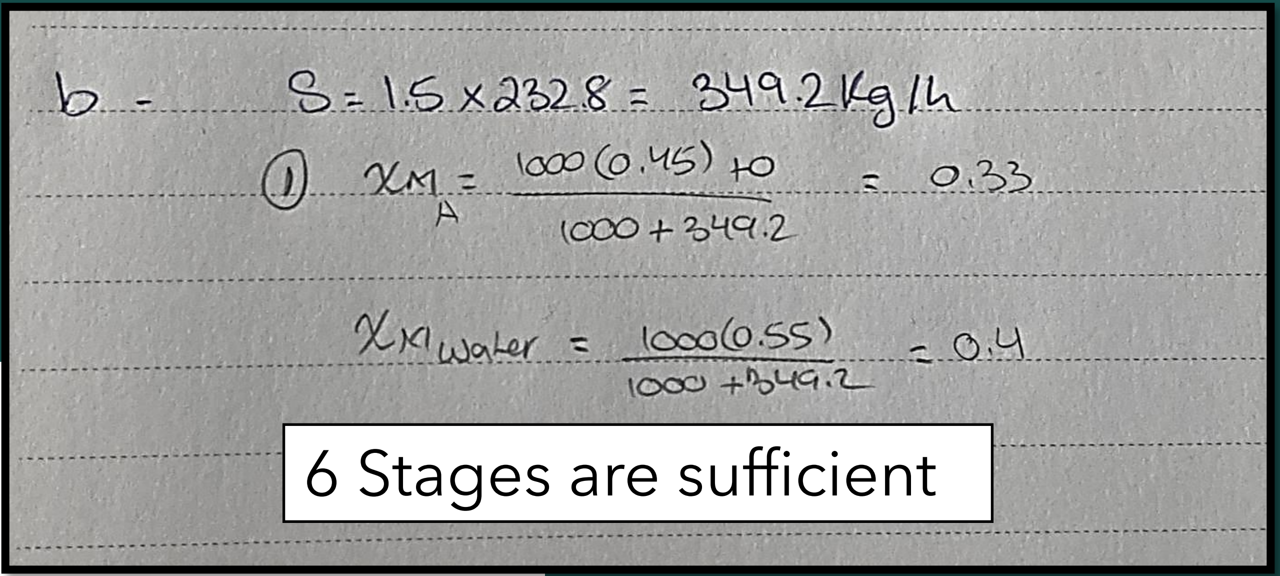
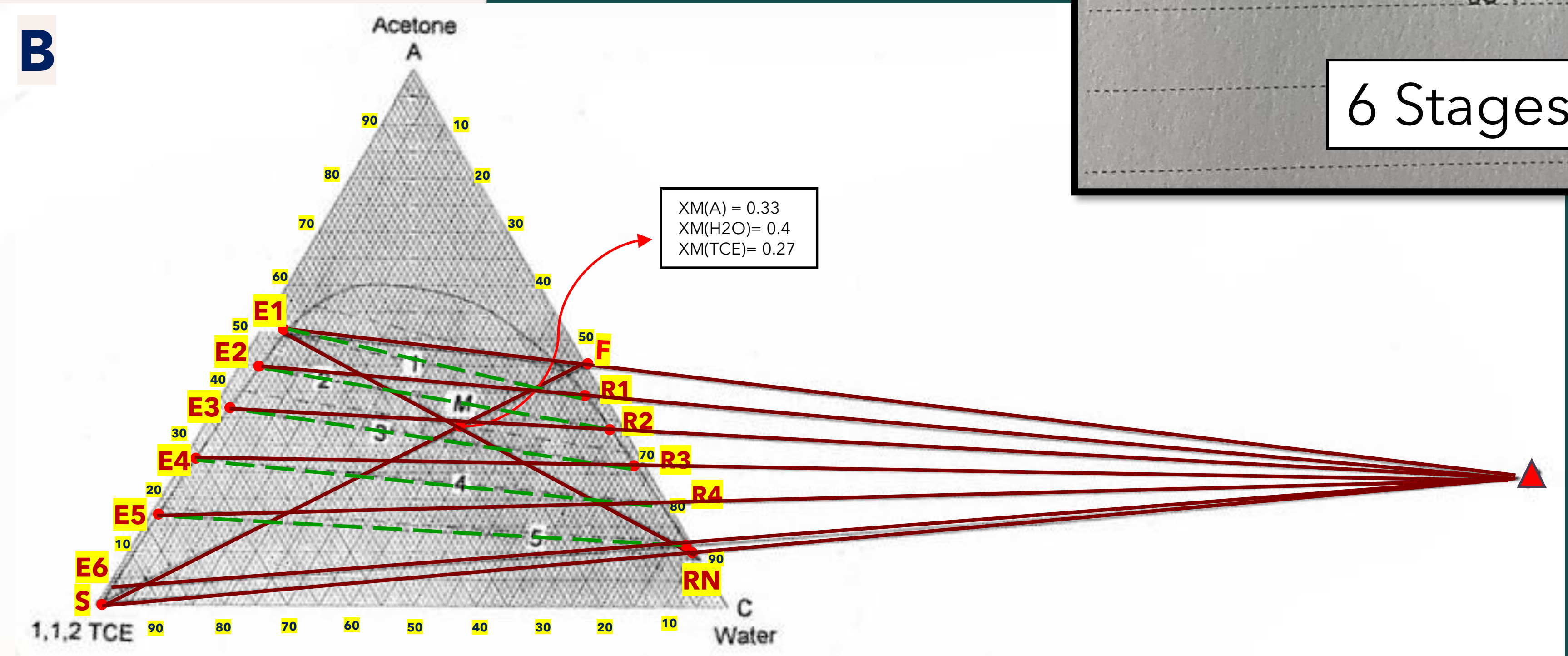


b - $S = 1.5 \times 232.8 = 349.2$ Kg/h

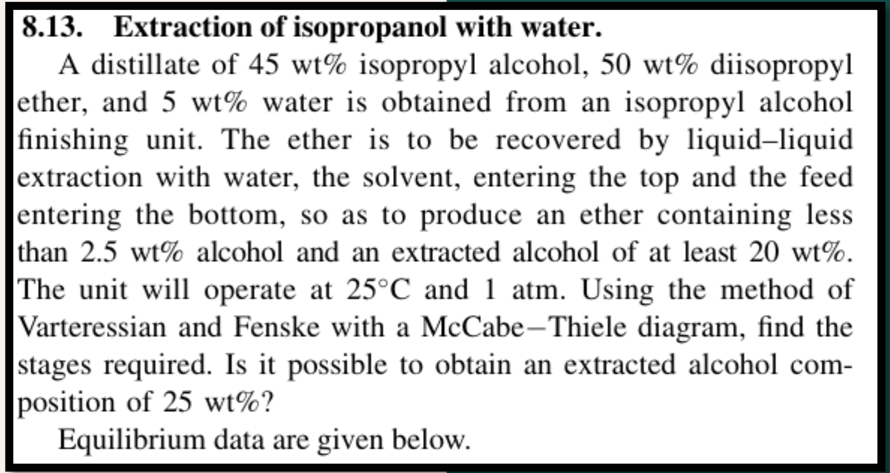
① $x_{M_A} = \frac{1000(0.45)+0}{1000+349.2} = 0.33$

$x_{M_{water}} = \frac{1000(0.55)}{1000+349.2} = 0.4$

6 Stages are sufficient



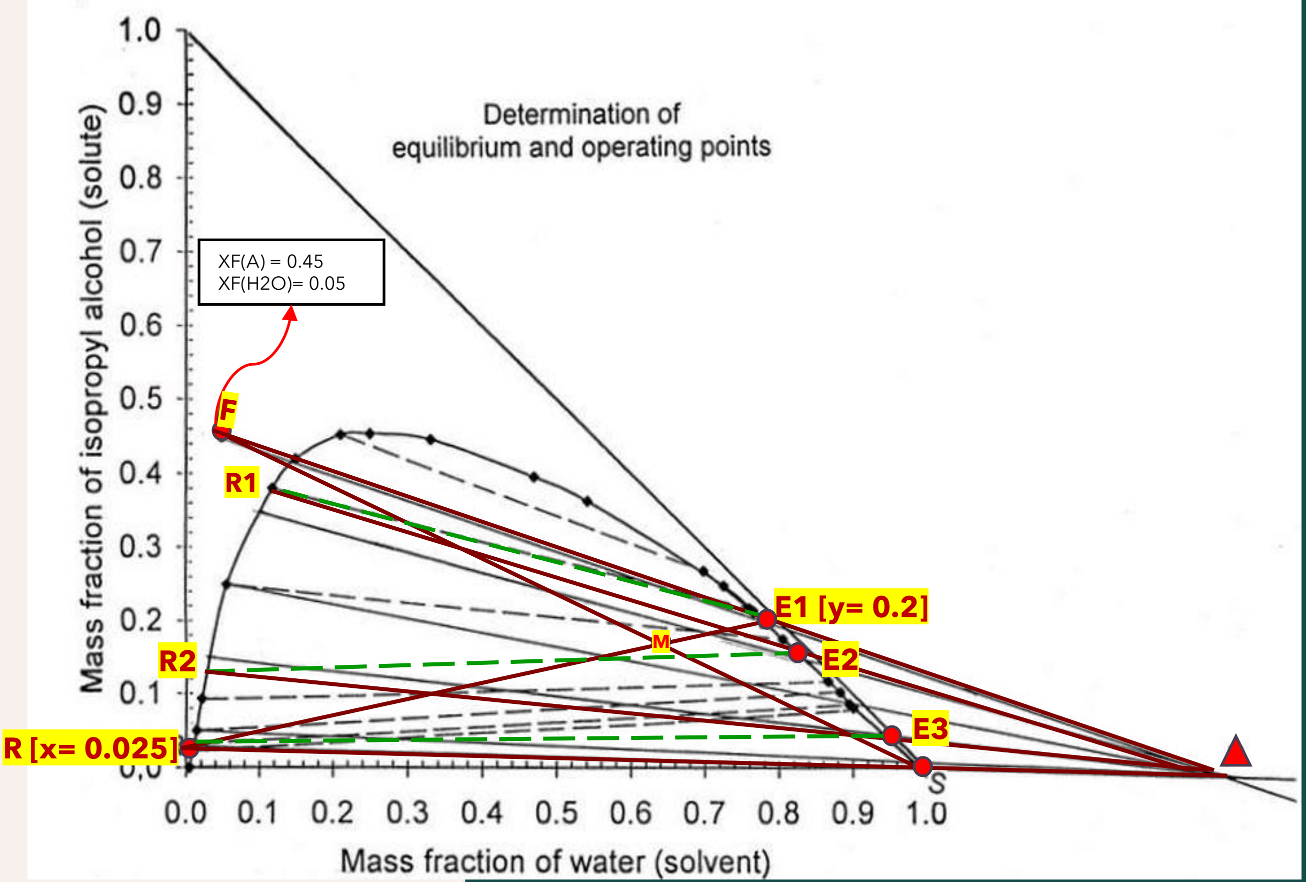
XM(A) = 0.33
XM(H2O)= 0.4
XM(TCE)= 0.27

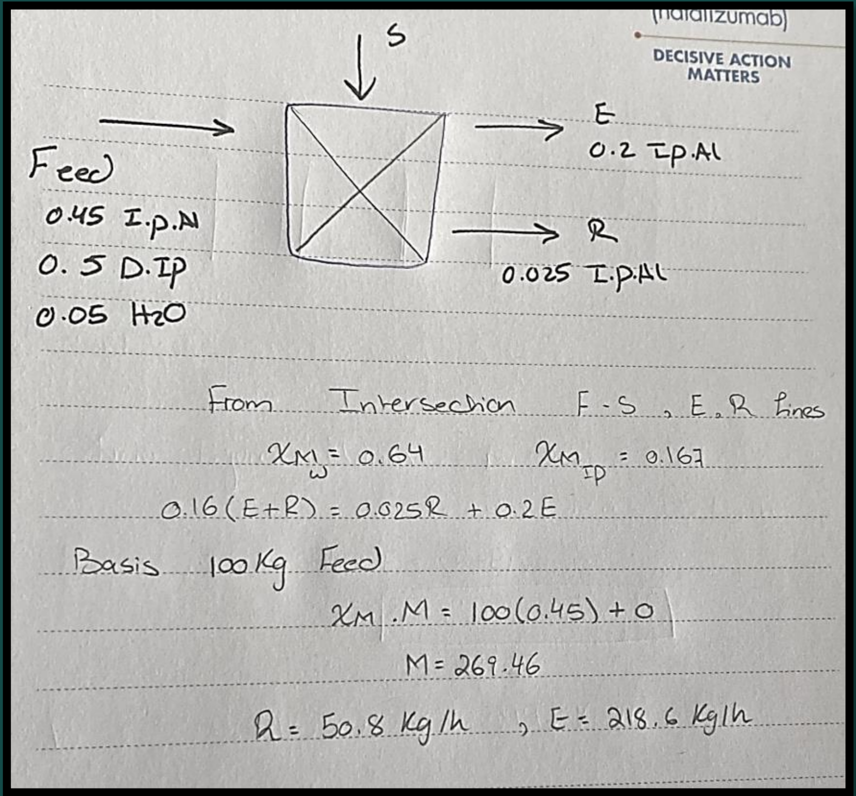


8.13. Extraction of isopropanol with water.

A distillate of 45 wt% isopropyl alcohol, 50 wt% diisopropyl ether, and 5 wt% water is obtained from an isopropyl alcohol finishing unit. The ether is to be recovered by liquid–liquid extraction with water, the solvent, entering the top and the feed entering the bottom, so as to produce an ether containing less than 2.5 wt% alcohol and an extracted alcohol of at least 20 wt%. The unit will operate at 25°C and 1 atm. Using the method of Varteressian and Fenske with a McCabe–Thiele diagram, find the stages required. Is it possible to obtain an extracted alcohol composition of 25 wt%?

Equilibrium data are given below.





(nataliizumab)

DECISIVE ACTION MATTERS

S

Feed

0.45 I.p.Al

0.5 D.Ip

0.05 H_2O

E

0.2 Ip.Al

R

0.025 I.p.Al

From Intersection F-S , E.R lines

$x_{M_w} = 0.64$ $x_{M_{IP}} = 0.167$

$0.16(E+R) = 0.025R + 0.2E$

Basis 100 Kg Feed

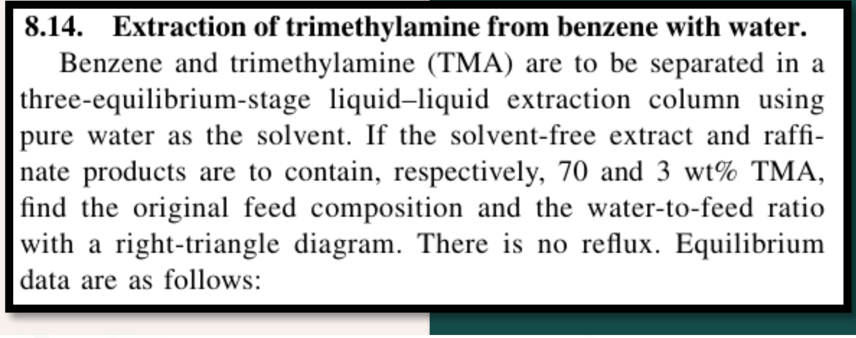
$x_M . M = 100(0.45) + 0$

$M = 269.46$

$R = 50.8$ Kg/h , $E = 218.6$ Kg/h

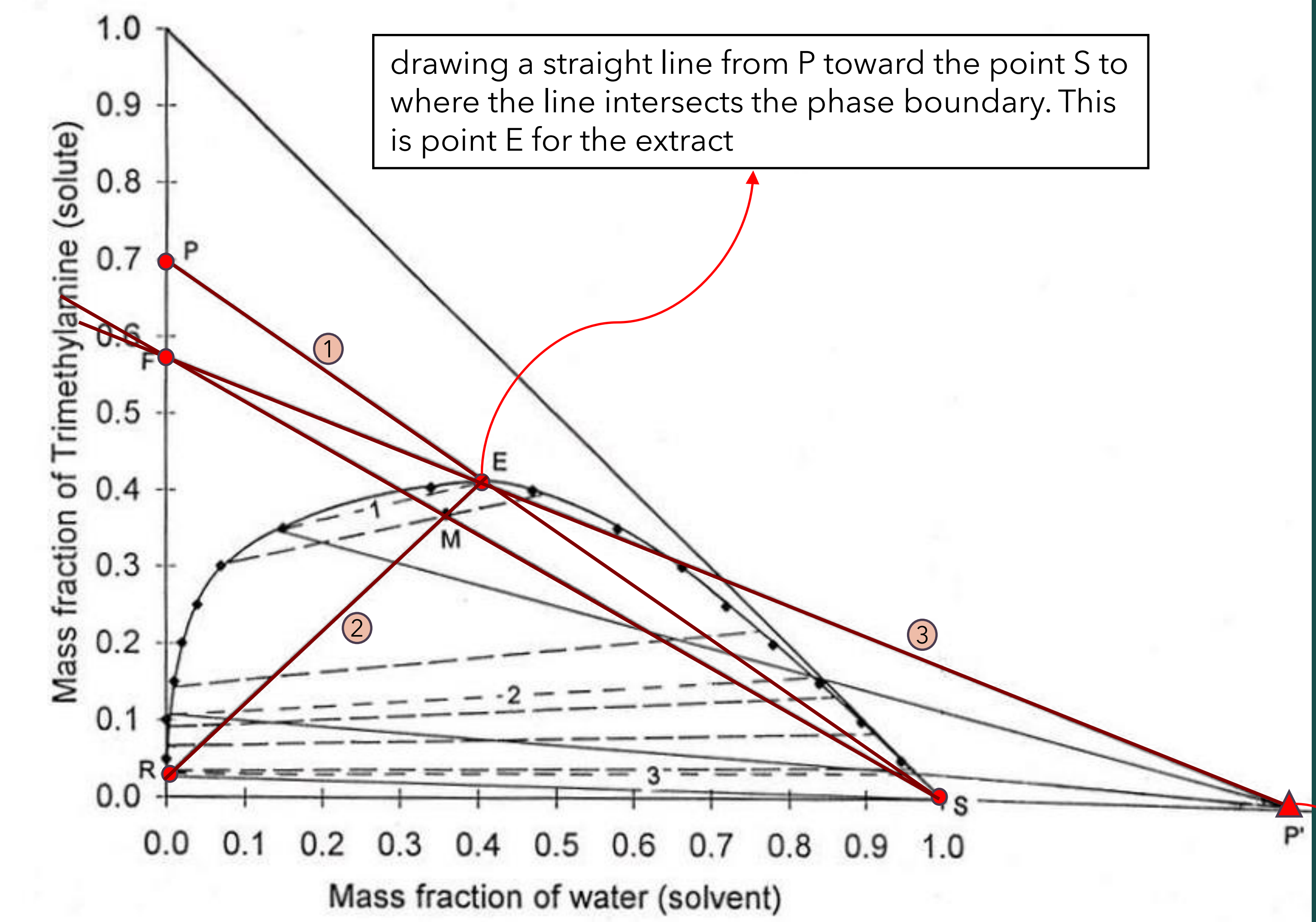
3 Stages are sufficient

*An extract with an alcohol content of greater than 25 wt% A is impossible because the operating line would have a slope less than an overlapping tie line, making it impossible to step off stages in the right direction.



8.14. Extraction of trimethylamine from benzene with water.

Benzene and trimethylamine (TMA) are to be separated in a three-equilibrium-stage liquid–liquid extraction column using pure water as the solvent. If the solvent-free extract and raffinate products are to contain, respectively, 70 and 3 wt% TMA, find the original feed composition and the water-to-feed ratio with a right-triangle diagram. There is no reflux. Equilibrium data are as follows:



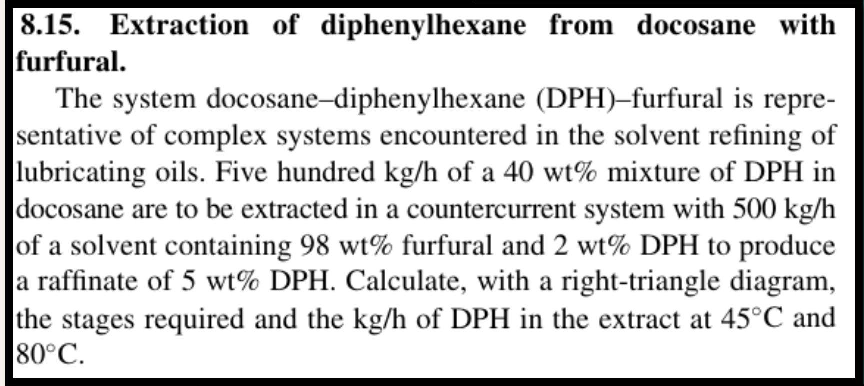
drawing a straight line from P toward the point S to where the line intersects the phase boundary. This is point E for the extract

1- Assuming given M → Point F will be the intersection between (S-M) line + (delta-E) line

OR ;

2- Assuming water-free feed→ Point F will be the intersection between y-axis + (delta-E) line then you can get point M

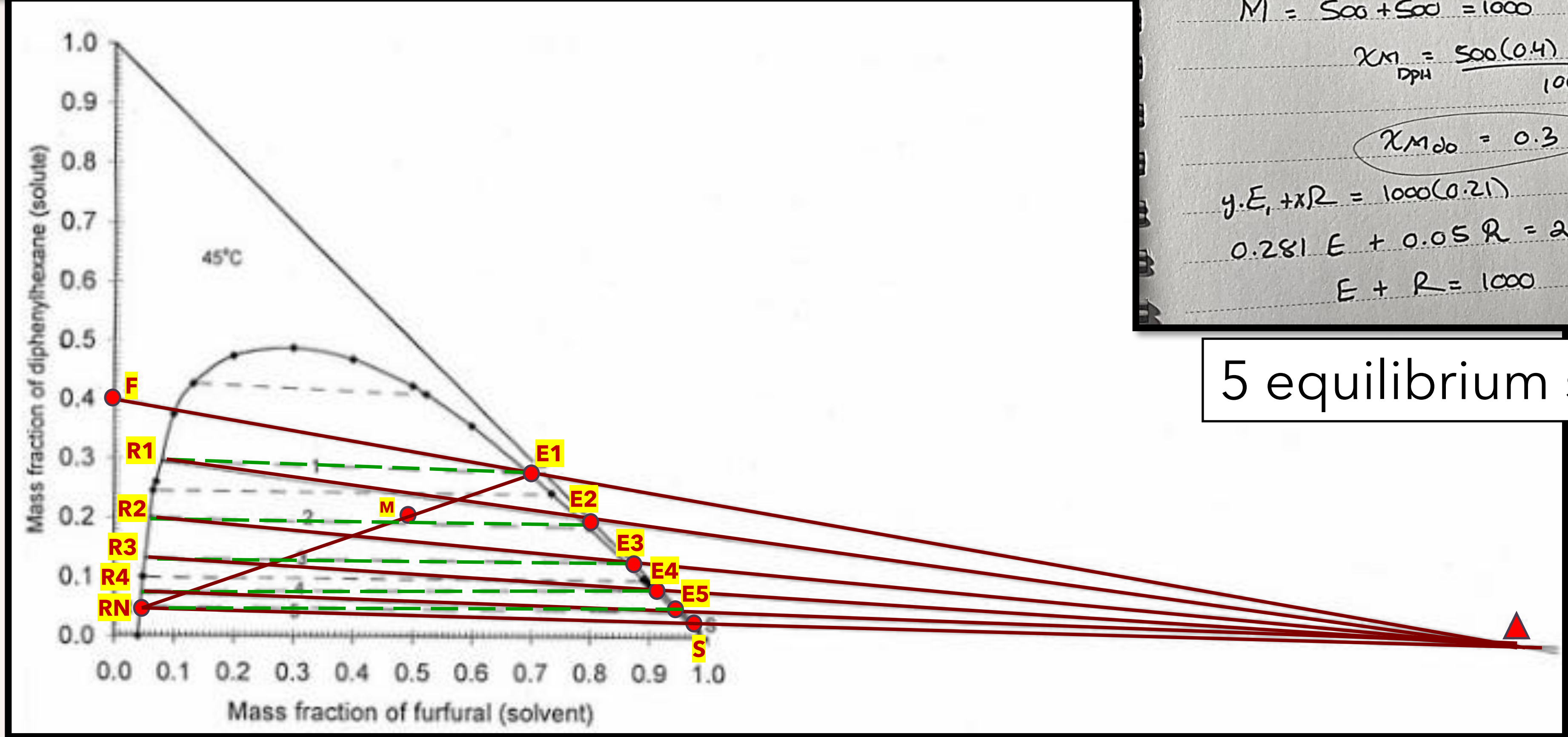
Assumed given [found by trial and error]

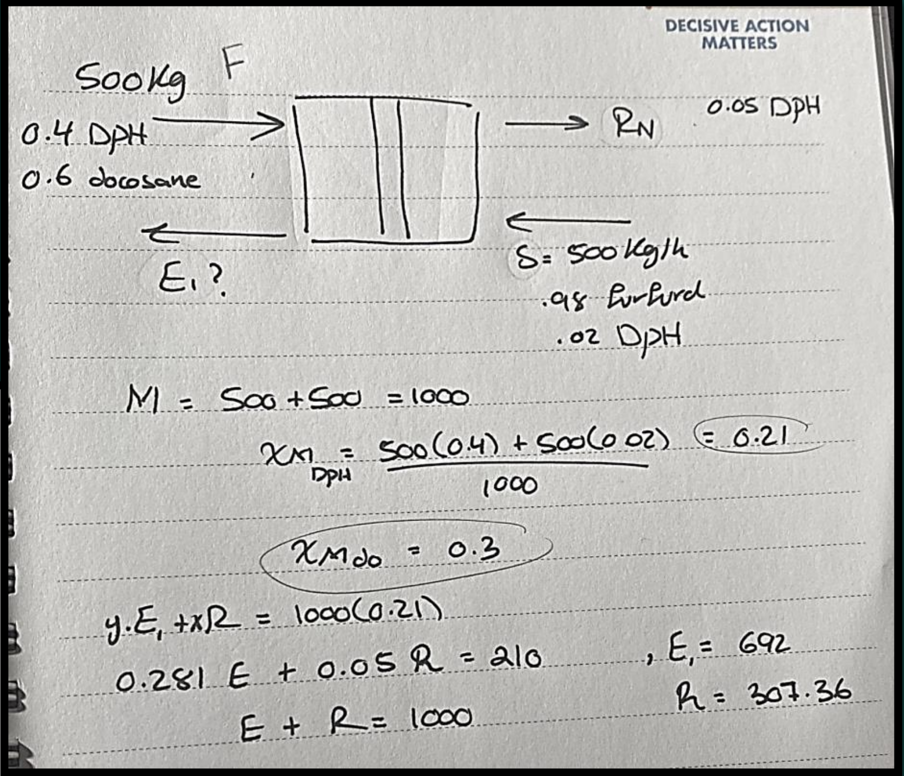


8.15. Extraction of diphenylhexane from docosane with furfural.

The system docosane–diphenylhexane (DPH)–furfural is representative of complex systems encountered in the solvent refining of lubricating oils. Five hundred kg/h of a 40 wt% mixture of DPH in docosane are to be extracted in a countercurrent system with 500 kg/h of a solvent containing 98 wt% furfural and 2 wt% DPH to produce a raffinate of 5 wt% DPH. Calculate, with a right-triangle diagram, the stages required and the kg/h of DPH in the extract at 45°C and 80°C.

@45°





500kg F

0.4 DPH

0.6 docosane

R_N 0.05 DPH

E_1?

S= 500 kg/h

.98 furfurd

.02 DPH

$$M = 500 + 500 = 1000$$

$$x_{M_{DPH}} = \frac{500(0.4) + 500(0.02)}{1000} \boxed{= 0.21}$$

$$\boxed{x_{M_{do}} = 0.3}$$

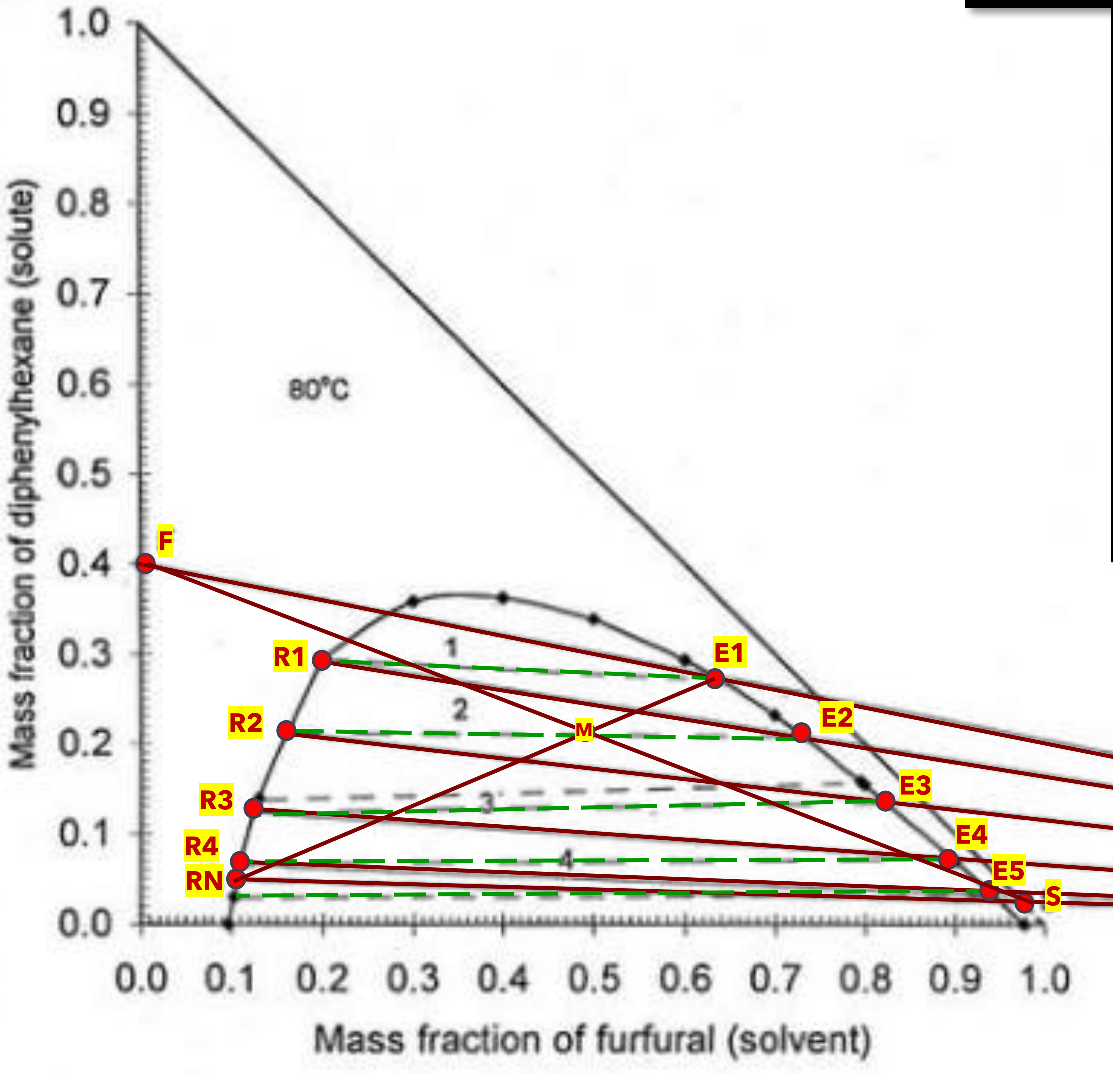
$$y \cdot E_1 + xR = 1000(0.21)$$

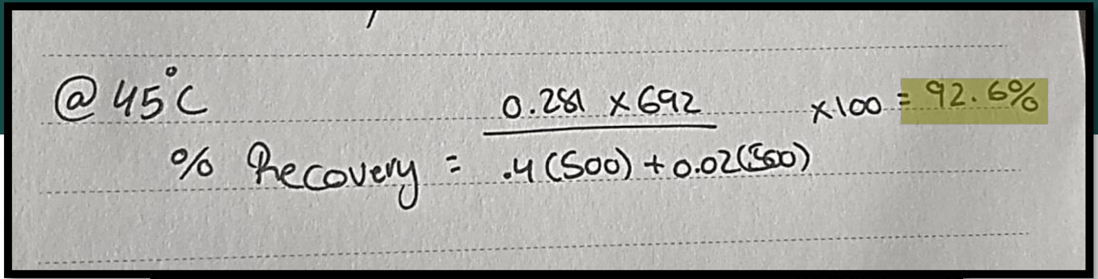
$$0.281\,E + 0.05\,R = 210 \qquad , E_1 = 692$$

$$E + R = 1000 \qquad R = 307.36$$

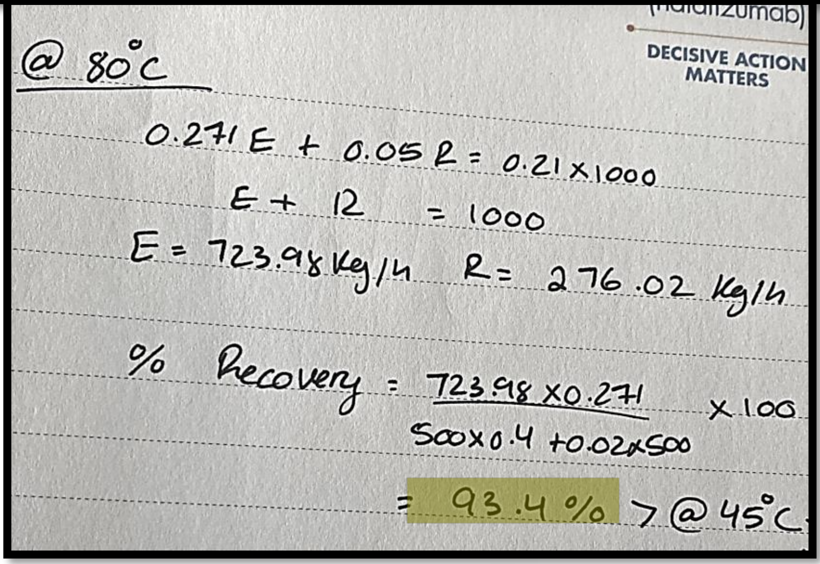
5 equilibrium stages

@80°




@ 45°C

$$\% \text{ Recovery} = \frac{0.281 \times 692}{.4(500) + 0.02(500)} \times 100 = 92.6\%$$


DECISIVE ACTION MATTERS

@ 80°C

$$0.271E + 0.05R = 0.21 \times 1000$$

$$E + R = 1000$$

$E = 723.98$ Kg/h $R = 276.02$ Kg/h

$$\% \text{ Recovery} = \frac{723.98 \times 0.271}{500 \times 0.4 + 0.02 \times 500} \times 100$$

$$= 93.4\% > @ 45°C$$

4+ equilibrium stages