Mass Transfer Operations

Lec 11: Analysis of binary distillation
McCabe Thiele Graphical Method

Content

McCabe Thiele Method, Reflux Considerations
Special Cases for Rectification, No. of Real Plates

Prof. Zayed Al-Hamamre

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Content

- McCabe Thiele Method
- Reflux Considerations
- Special Cases for Rectification
- No. of Real Plates

Learning Outcomes

- After this lecture you should be able to…..
  - Explain the importance of Reflux
  - Apply the McCabe Thiele graphical design method to determine the number of equilibrium stages in a column
  - Describe column internals (trays and packing)
  - Define tray efficiency
  - Explain the height of packing equivalent to a theoretical plate
  - Thanks to Ming Tham's website for pictures and information. An excellent presentation on distillation:
  - http://lorien.ncl.ac.uk/ming/distil/copydist.htm

McCabe Thiele Method

- Developed in 1925
- Suitable for binary mixtures in continuous columns
- Can be used as a preliminary study for multi component
- Combination of VLE data and operating lines
- Assumes:
  - Constant molal overflow
  - No heat loss from column
- Not suitable if:
  - Relative volatility is < 1.3 or > 5
  - Reflux ratio < 1.1 Rmin
  - More than 25 theoretical stages needed

Specifications required

- The following information is needed

F	Total feed rate
z_F	Mol fraction of MVC in the feed
P	Column operating pressure
x_D	Mol fraction of MVC in the distillate
x_B	Mol fraction of MVC in the bottoms
R/R_{min}	Ratio of reflux to minimum reflux

Results provided

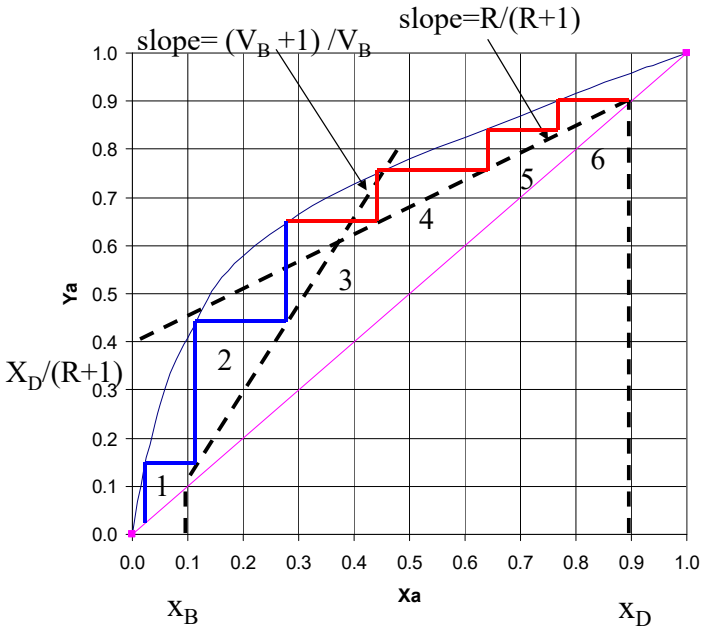
- McCabe Thiele provides the following results

D	Overall distillate flowrate
B	Overall bottoms flowrate
N_{min}	Minimum number of equilibrium stages
R_{min}	Minimum reflux ratio (L_{min}/D)
R	Reflux ratio (L/D)
V_B	Boilup ratio (V/B)
N	Number of equilibrium stages
x_n, y_n	Vapour and liquid compositions at each of the stages

Design Procedure

- Start with VLE data and draw an x-y diagram
- Determine the three operating lines and draw them on the x-y diagram
- Using the rectifying operating line and the equilibrium curve draw steps from the distillate composition to the feed point
- Count the number of steps. Each one equals an equilbrium or theoretical stage
- Using the stripping operating line and the equilibrium curve draw steps from the bottoms composition to the feed point
- Add the number of steps to the previous to give the total number of equilibrium stages
- Convert this to a number of plates by dividing by the plate efficiency.

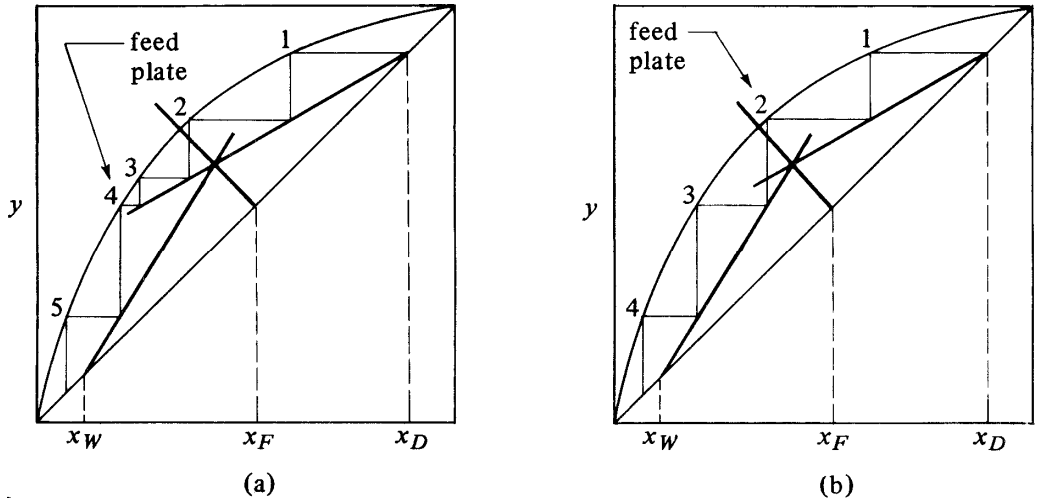
Design Illustrated



1. Obtain VLE curve
2. Pick distillate composition, X_D, and Reflux ratio, R, and draw operating line for rectification
3. Pick bottoms composition, X_B, and Boilup ratio, V_B, and draw operating line for stripping
4. Start stepping off from the distillate end until the intersection of the two operating lines is passed.
5. Continue stepping but use the stripping operating line
6. Count the number of stages
7. Subtract one for the reboiler to give the number of theoretical trays (or leave as safety factor)

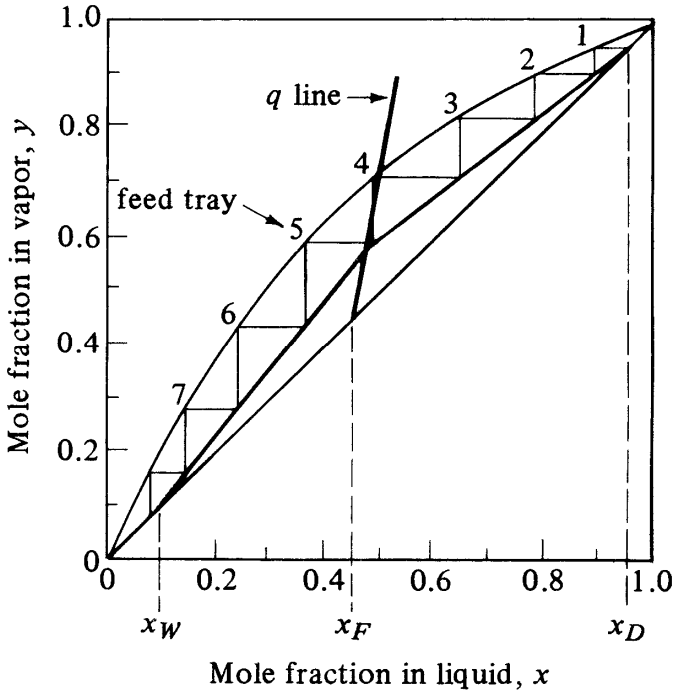
Design Illustrated

- The feed stage/tray is the stage that crosses from the rectifying or enriching section into the stripping section



Method of stripping off number of theoretical trays and location of feed plate: (a) **improper** location of feed on tray 4; (b) **proper** location of feed on tray 2 to give minimum number of steps.

Design Illustrated



Example 11.4-1 (Geankoplis)

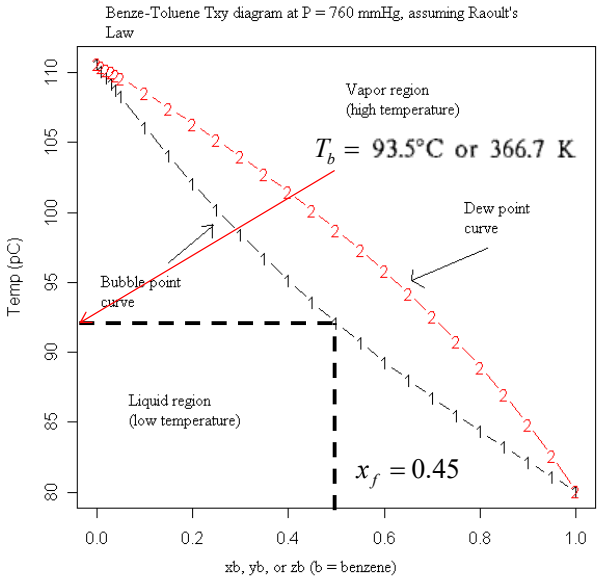
Rectification of a Benzene - Toluene Mixture

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

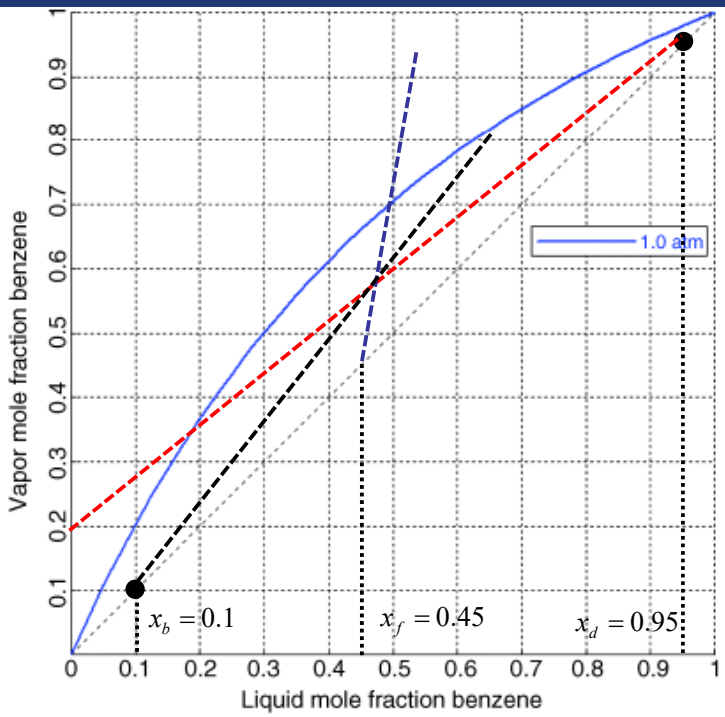
*b.p of the feed = 366.7 K

Example Contd.

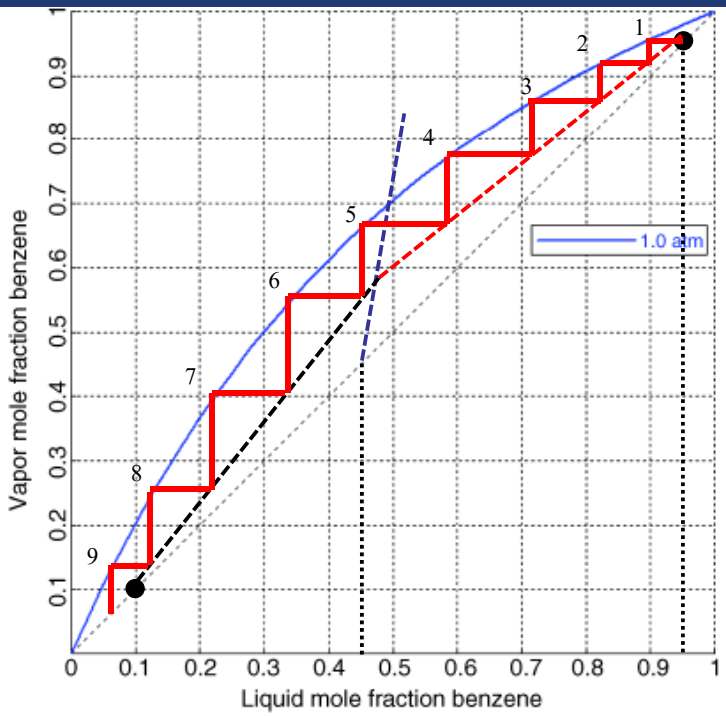
Example Contd.



Example Contd.



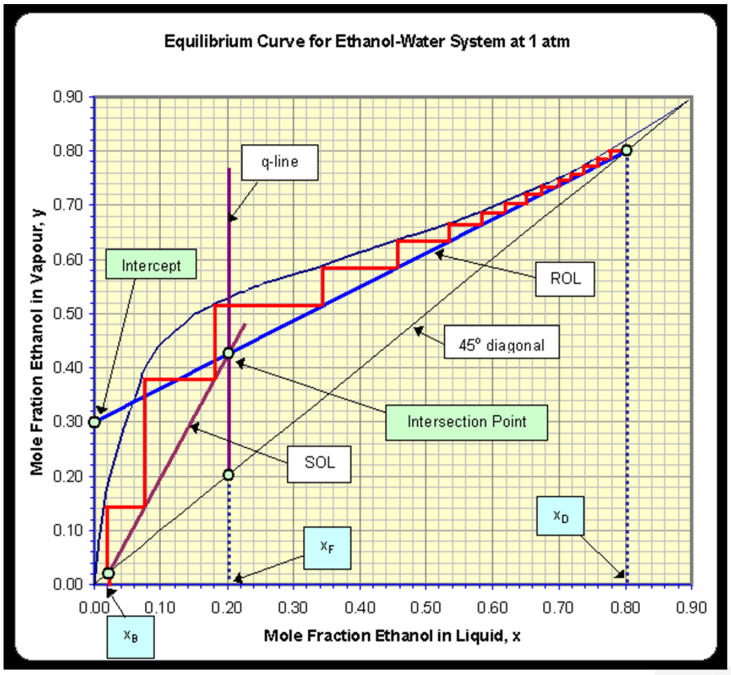
Example Contd.



Example

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

Example Contd.



Reflux Considerations

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

- The rectifying operating line slope depends on R.
- Therefore, the number of stages required for a given separation depends on R.
- Effective reflux ratio can be greater than R if the column is poorly insulated and may be subject to change due to weather conditions. A well lagged column is more consistent.
- The amount of reflux lies somewhere between the two extremes of total reflux and minimum reflux

Reflux Considerations

- In distillation of a binary mixture *A* and *B* the feed conditions, distillate composition, and bottoms composition are usually specified and the number of theoretical trays are to be calculated.
- However, the number of theoretical trays needed depends- upon the operating lines. To fix the operating lines, the reflux ratio (R= *L*/D) at the top of the column must be set.
- The two extremes are Total Reflux and Minimum Reflux
- Total Reflux - all condensate is returned, no product is taken off and no feed added. This gives the minimum number of stages to achieve the separation.
- Minimum Reflux - reducing R requires more stages to achieve the separation. Further reduction creates a pinch point where an infinite number of stages is required.
- Minimum reflux happens when the intersection of the operating lines lies on the equilibrium curve

Total Reflux

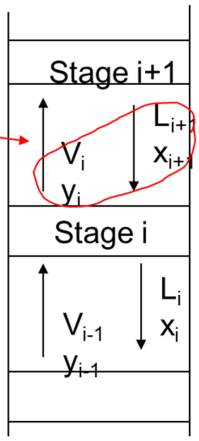
- One of the limiting values of reflux ratio is that of total reflux, or $R = \infty$.
- All vapour is condensed and returned as liquid

→ The slop of the enriching operating line

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

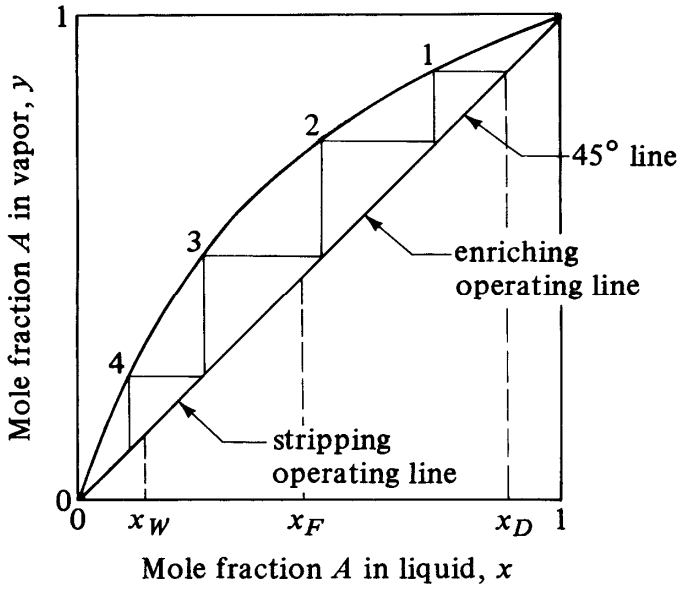
approach 1.0

Total reflux:
$V_i = L_{i+1}$
$y_i = x_{i+1}$



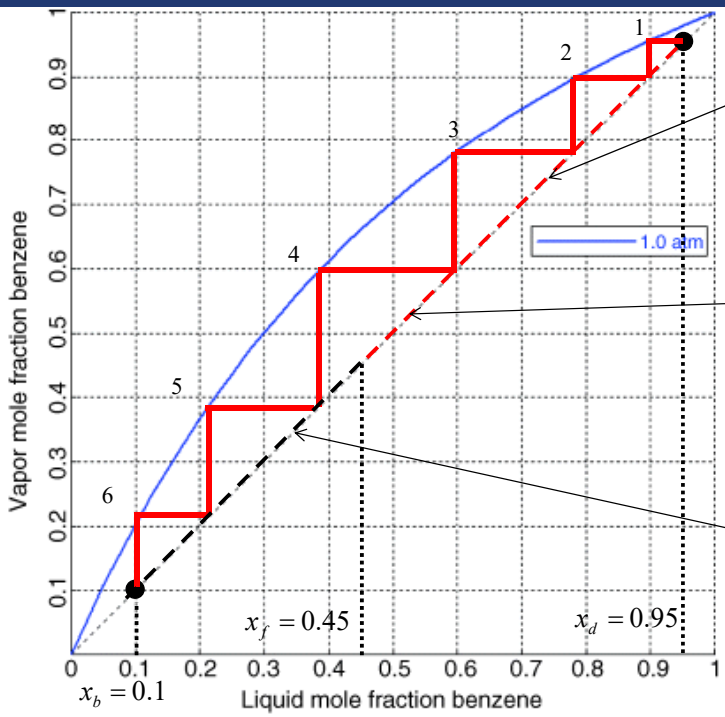
- Hence, the operating lines of both sections of the column coincide with the *45°* diagonal line,
- This gives the minimum number of trays that can possibly be used to obtain the given separation.

Total Reflux



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Total Reflux



The rectifying (enriching) operating line

x= y line

Stripping operating line

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Total Reflux

- Can use Fenske equation to calculate N_{min}

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

$$\alpha_{ab} = \frac{\frac{P_a}{x_a}}{\frac{P_b}{x_b}}$$

- Sometimes a column is operated in total reflux at startup

- This condition of total reflux can also be interpreted as requiring infinite sizes of condenser, reboiler, and tower diameter for a given feed rate.

Feed Location

iv. Determination of Feed Location

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

Kirkbride Equation

Ratio of the no. of stages in rectifying to stripping section.

Z_H and Z_L – mole fraction of heavy and light key respectively in feed.

$X_{B,L}$ and $X_{D,H}$ – mole fraction of light key in bottom product and heavy key in top product respectively.

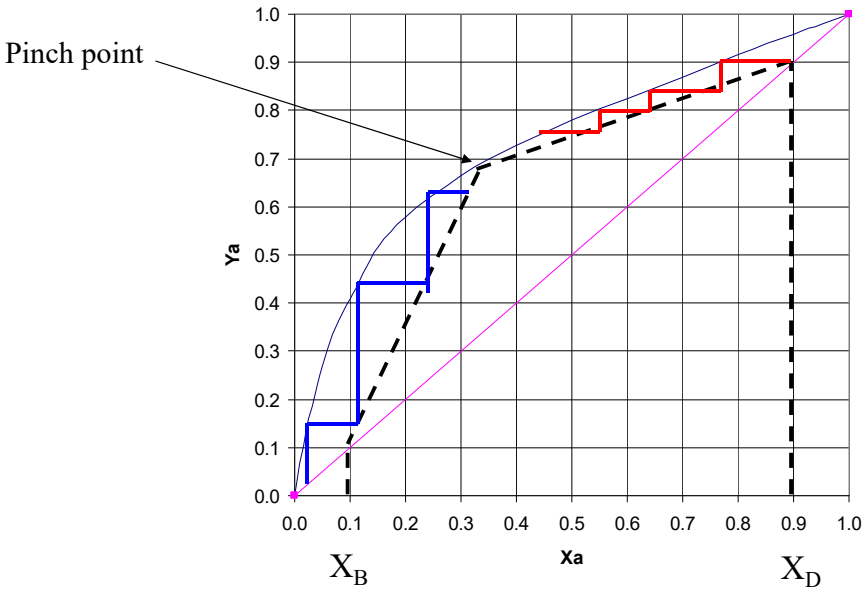
B and D – molar flow of bottom product and distillate.

Minimum Reflux

- The minimum reflux ratio can be defined as the reflux ratio R_m that will require an infinite number of trays for the given separation desired of x_d and x_b.
- This corresponds to the minimum vapor flow in the tower, and hence the minimum reboiler and condenser sizes.
- If R is decreased, the slope of the enriching operating line $R/(R + 1)$ is decreased, and the intersection of this line and the stripping line with the q line moves farther from the 45° line and closer to the equilibrium line.
- As a result, the number of steps required to give a fixed x_d and x_b increases

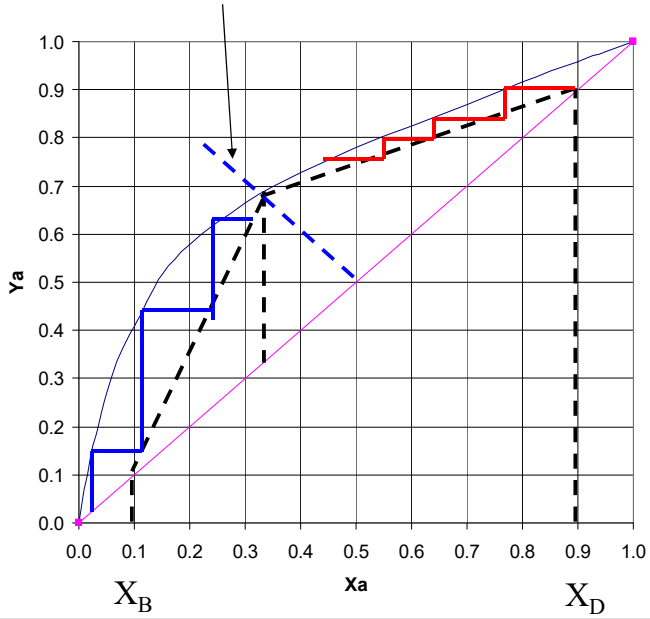
Minimum Reflux

- Min Reflux happens when the two operating lines intersect on equilibrium curve (intersection point is called a pinch point)



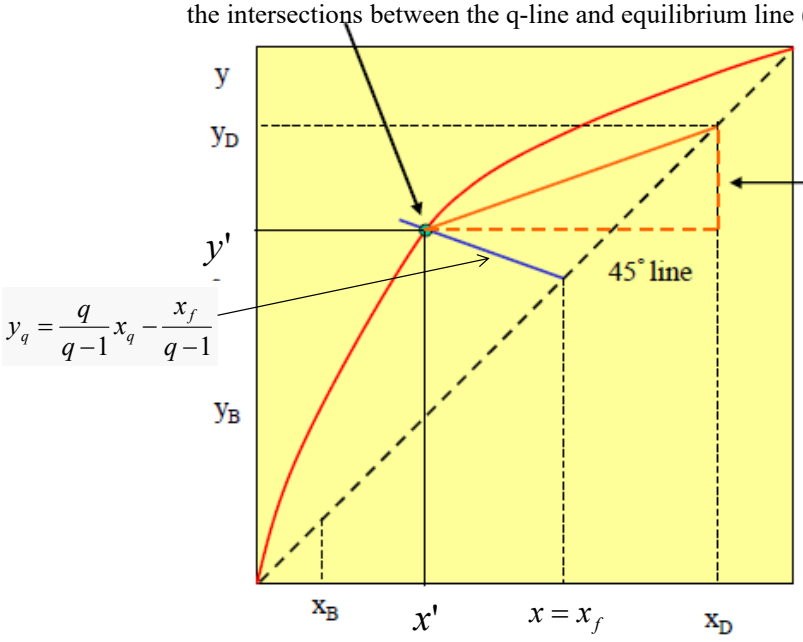
Minimum Reflux

- Don't forget the q line. Min reflux occurs at intersection with equilibrium curve because all three lines should intersect



Minimum Reflux

the intersections between the q-line and equilibrium line (**pinch point**)



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

$$slop = \frac{R_{min}}{R_{min}+1} = \frac{x_D - y'}{x_D - x'} = \left(\frac{L}{V}\right)_{min}$$

- From the minimum reflux, and R/R_{min} we can determine the reflux R.
- We determine the slope of the rectifying section operating line from:

$$Slop = \frac{R}{R+1}$$

Minimum Reflux

- Can also be calculated (if feed is liquid at b.p.)

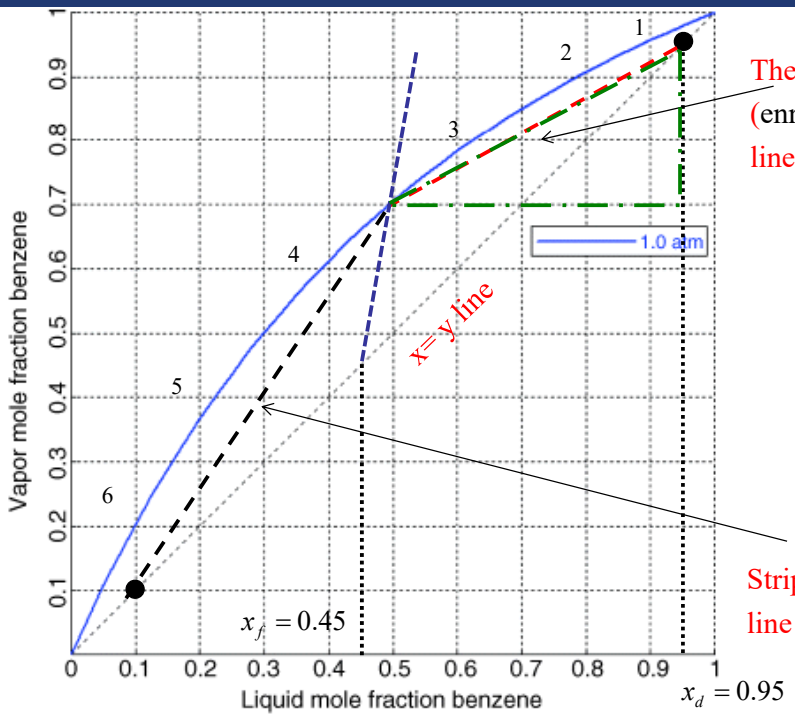
$$R_{\min} = \frac{1}{(\alpha - 1)}\left[\frac{x_D}{x_F} - \frac{\alpha(1 - x_D)}{(1 - x_F)}\right]$$

- Or, using definition of slope of line = R/(R+1)

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

- Where x' , y' is the intersection of line with curve

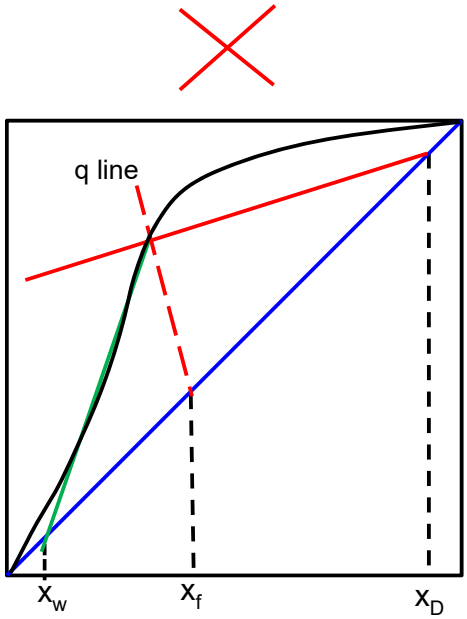
Minimum Reflux

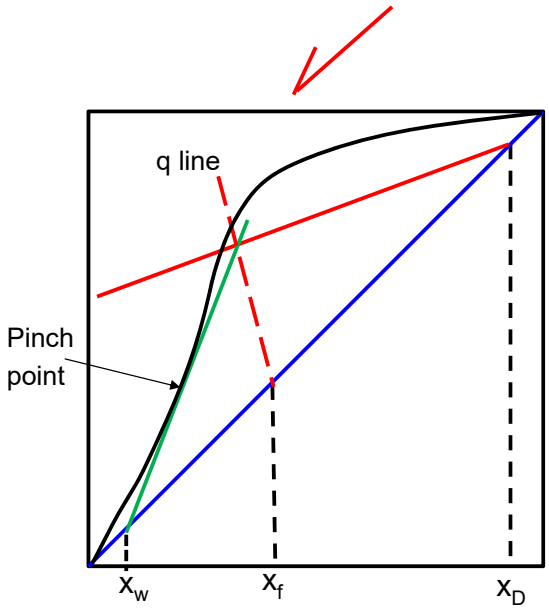


The rectifying (enriching) operating line

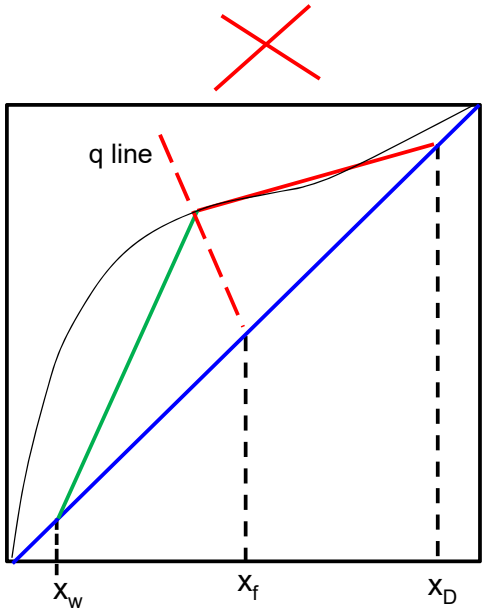
Stripping operating line

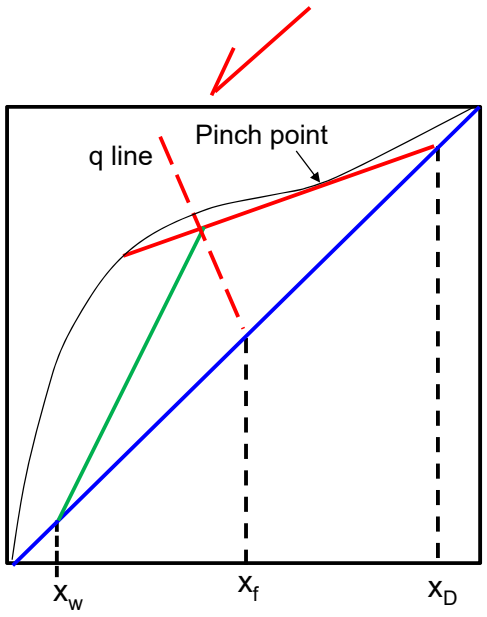
Minimum Reflux





Minimum Reflux

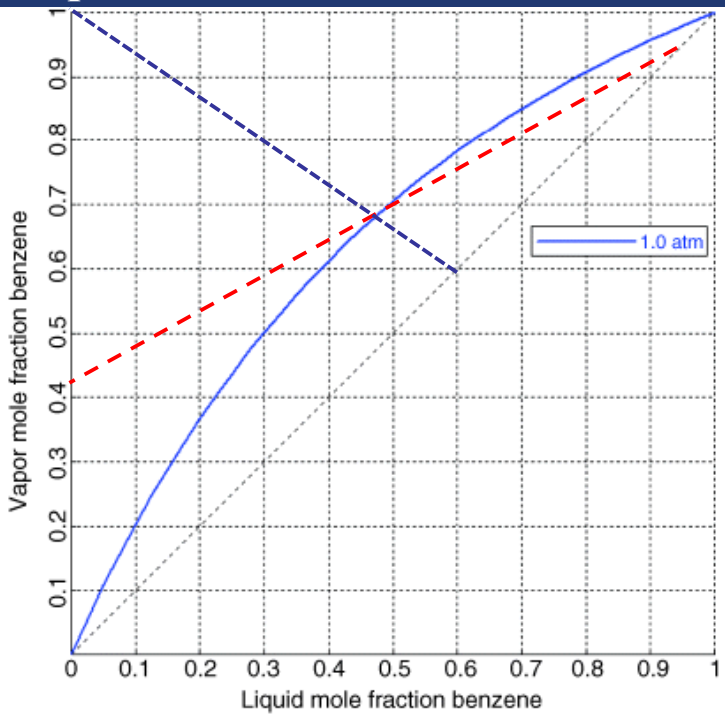




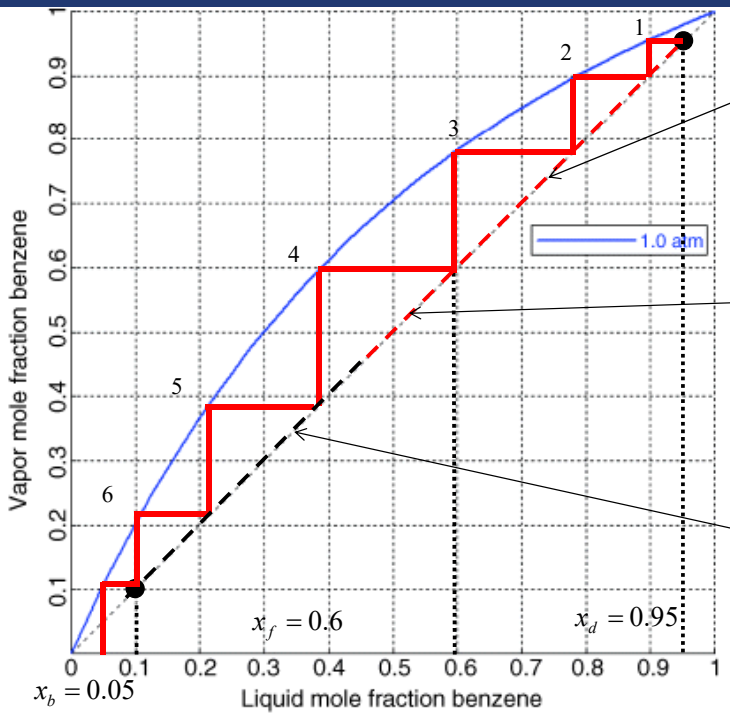
Activity

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

Activity-Equi. Data



Total Reflux

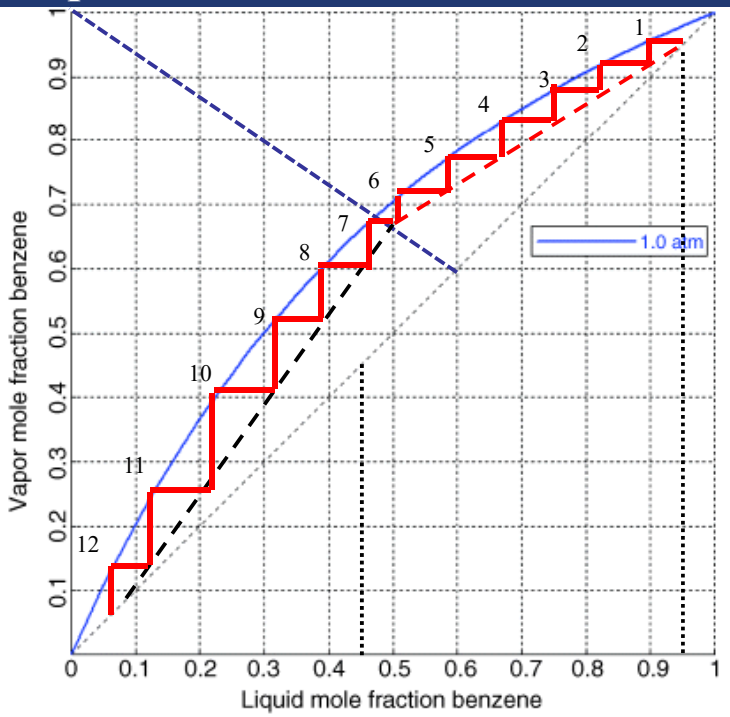


The rectifying (enriching) operating line

x= y line

Stripping operating line

Activity-Equi. Data

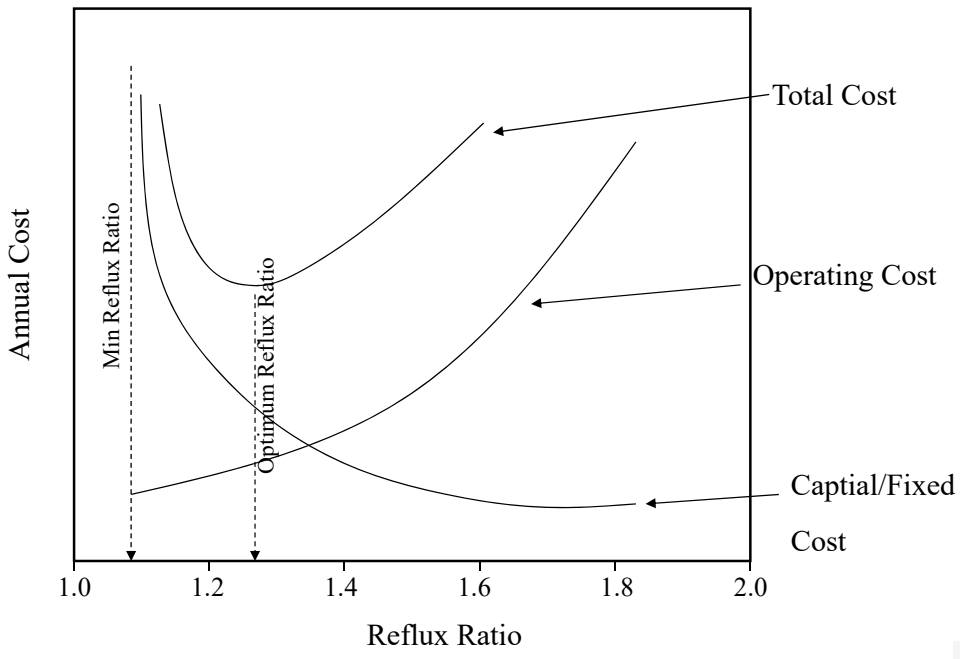


Optimum Reflux Ratio

- Increase R
- Diameter (D) gets bigger
  - Since D = fn(V, L)
  - R ↑ ⇒ V & L ↑ ⇒ D ↑ ∴ Capital Cost ↑
- No. of Plates gets smaller
  - Height ↓ ∴ Capital Cost ↓
- Heat exchangers get bigger
  - Boiling/Condensing ↑ ∴ Capital & Running Cost ↑

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Optimum Reflux Ratio



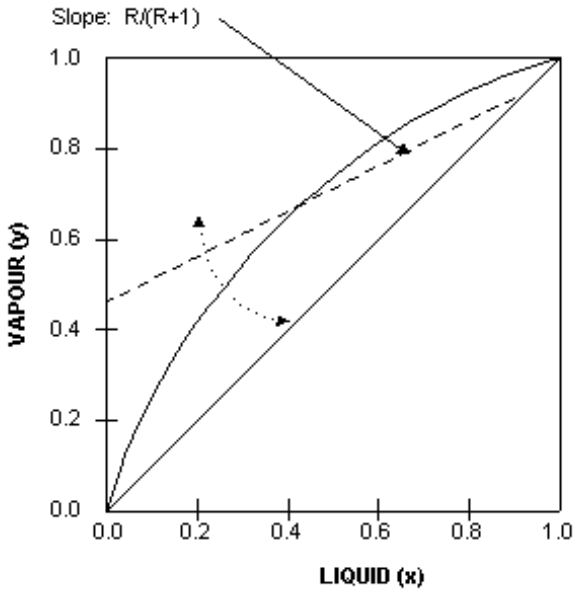
Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Reflux and Cost

- A trade off between operating cost and capital cost is needed.
- Increase R ⇒ Less stages, less capital cost BUT also ⇒ More boiling and condensing
- Decrease R ⇒ More stages, More capital cost BUT also ⇒ Less boiling and condensing
- Capital and operating costs combine to give a total cost. This is minimised with the following
- Rule of thumb: R = 1.2 to 1.5 times R_{min}

R Determines the Slope

- Change in slope means less or more stages. As R increases the slope approaches 1. More MVC returned to the column.



- Less material is removed as distillate. Separation improves, fewer trays needed
- As R is decreased, the slope decreases towards the equilibrium line.
  A 'pinch' between operating and equilibrium lines becomes more pronounced and more and more trays are required

Activity

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

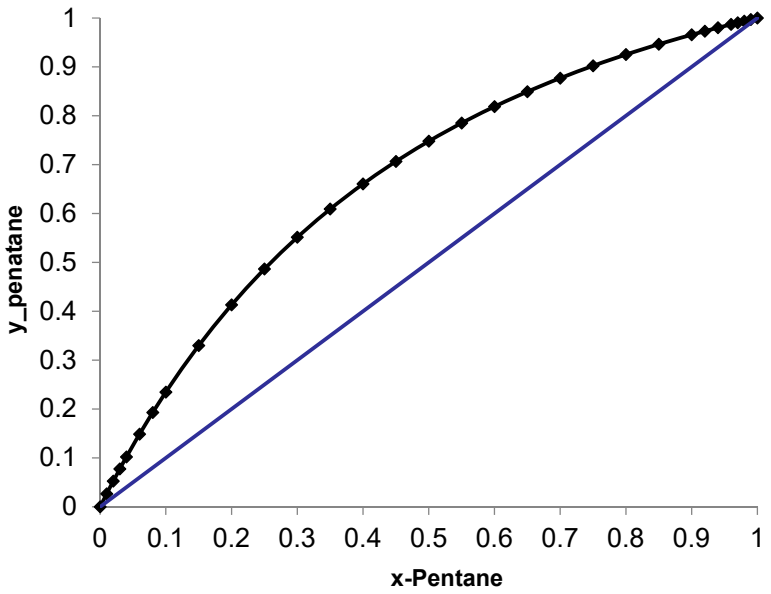
Determine the minimum number of equilibrium trays and the minimum reflux ratio.

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

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

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

Activity Cont.d



http://vle-calc.com/phase_diagram.html

Solution

Distillate and bottoms flow rates Overall material balance over the entire tower gives

$$D + B = 2{,}500$$

Material balance for n-pentane over the entire tower gives

$$0.97D + 0.02B = (0.4)(2{,}500) = 1{,}000 \Rightarrow 0.97(2{,}500 - B) + 0.02B = 1{,}000$$

Solving for B and D from the above equations we have

$$B = 1{,}500 \text{ lbmoles/hr and } D = 1000 \text{ lbmoles/hr.}$$

Heating and cooling loads:

$$Q_C = V_1(H_1 - h_D) \approx V_1 \Delta H_{evap}$$
$$= V_1(0.97\lambda_{C5} + 0.03\lambda_{C6})$$
$$V_1 = L_o + D = (R + 1)D = 4D = (4)(1000) = 4000 \text{ lbmoles/hr}$$

$$Q_C = (4{,}000)(0.97\times 11{,}369 + 0.03\times 13{,}572) = 4.574\times 10^7 \text{ Btu/h}$$

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Solution

From the Txy diagram, $T_D \approx 309$ K and $T_B \approx 342$ K

Since the boiling point of a 40 mole % n-pentane is 324.79 K, the feed enters the column at 30°C (or 303.15 K) is a subcooled liquid.

Making an energy balance over the column, we obtain

$$Fh_F + Q_R = Dh_D + Q_C + Bh_B$$

Let the reference state be liquid at 30°C, then $h_F = 0$.

$h_D = C_{pL,C5}(T_D - T_F) = (39.7)(309 - 303)(1.8) = 428.8$ Btu/lbmol

$h_B = C_{pL,C6}(T_B - T_F) = (51.7)(342 - 303)(1.8) = 3{,}629.3$ Btu/lbmol

$Q_R = Dh_D + Q_C + Bh_B = (1{,}500)(\ 3{,}629.3) + 4.574\times 107 + (1{,}000)(428.8)$

$= 5.1613\times 107$ Btu/hr

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Solution

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

Since $R = 3$ and $x_D = (0.97)$ we have

$$y_{n+1} = \frac{3}{3+1} x_n + \frac{1}{3+1}(0.97) = 0.75\, x_n + 0.2425$$

The *q*-line is determined

$$q = \frac{H_V - H_F}{H_V - h_L} = \frac{H_V - h_L + h_L - H_F}{H_V - h_L} = 1 + \frac{C_{pL}(T_{boil} - T_F)}{H_V - h_L}$$

Solution

$$H_V - h_L = \text{latent heat} = 0.4\lambda_{C5} + 0.6\lambda_{C6} = (0.4)(11{,}369) + (0.6)(13{,}572)$$

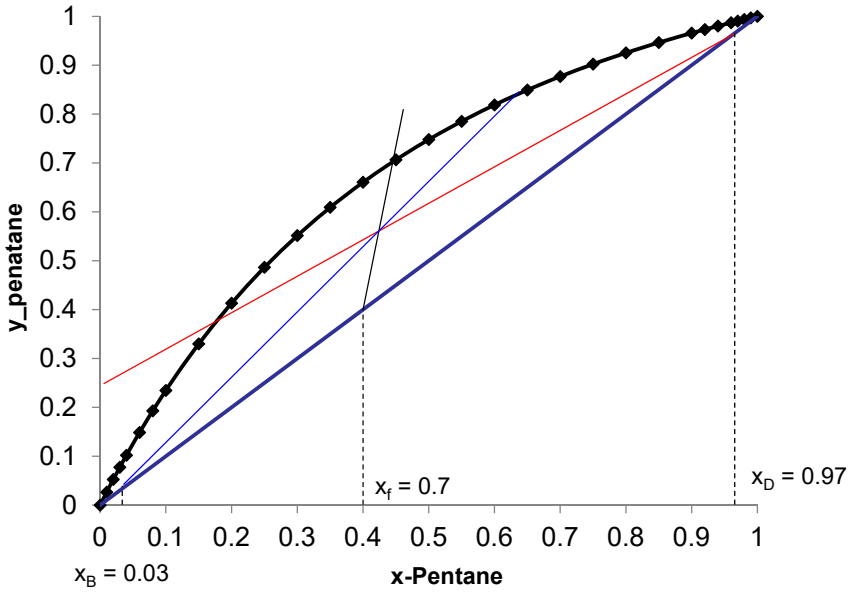
$$H_V - h_L = 12{,}691 \text{ Btu/lbmol}$$

$$C_{pL} = 0.4C_{pL,C5} + 0.6C_{pL,C6} = (0.4)(39.7) + (0.6)(51.7) = 46.9 \text{ Btu/lbmol}\cdot{}^{\circ}\text{F}$$

$$q = 1 + \frac{(46.9)(324.79 - 303.15)}{12{,}691} = 1.08$$

$$y = \frac{q}{q-1} x - \frac{x_F}{q-1} = \frac{1.08}{1.08-1} x - \frac{0.4}{1.08-1} = 13.5x - 5$$

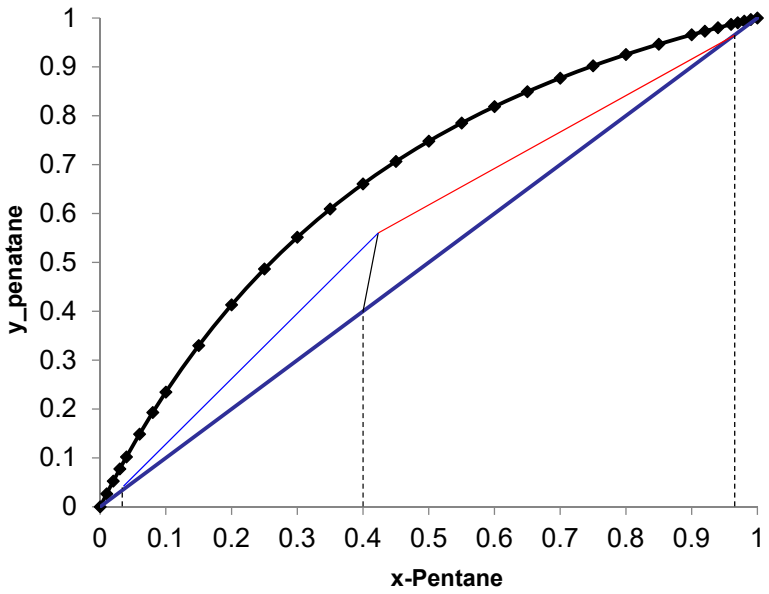
Activity Cont.d



http://vle-calc.com/phase_diagram.html

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

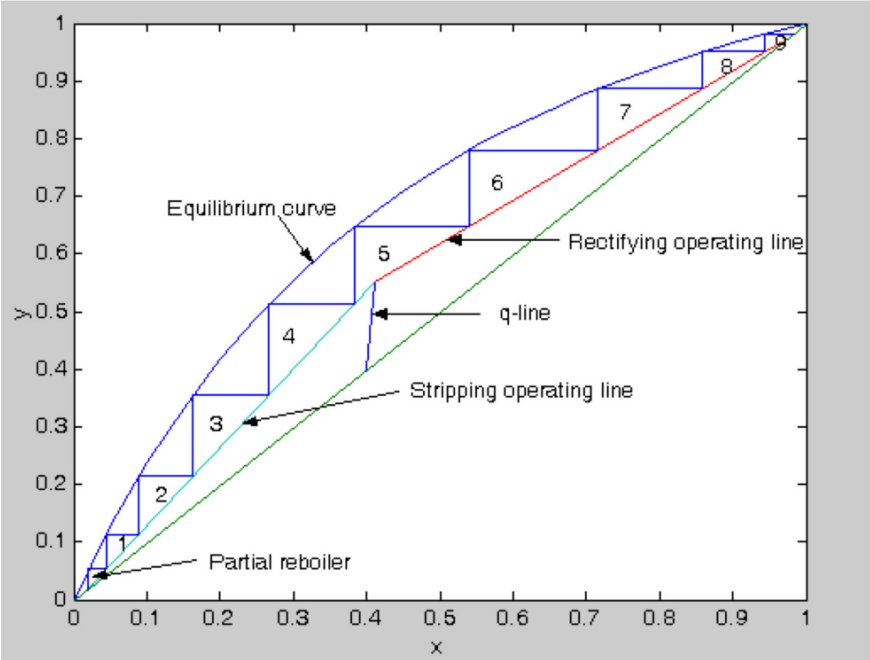
Activity Cont.d



http://vle-calc.com/phase_diagram.html

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

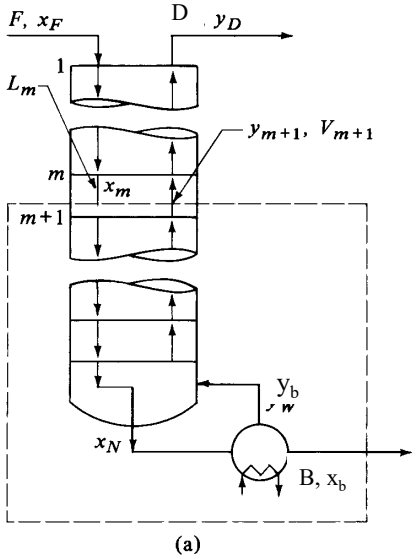
Solution



Special Cases for Rectification Using McCabe-Thiele Method

Stripping-Column Distillation

- In some cases the feed to be distilled is not supplied to an intermediate point in a column but is added to the top of the stripping column.
- The feed is usually a saturated liquid at the boiling point and the overhead product *D* is the vapor rising from the top plate, which goes to a condenser with no reflux or liquid returned back to the tower.
- The bottoms product *B* usually has a high concentration of the less volatile component *B*.
- Hence, the column operates as a stripping tower with the vapor removing the more volatile *A* from the liquid as it flows downward

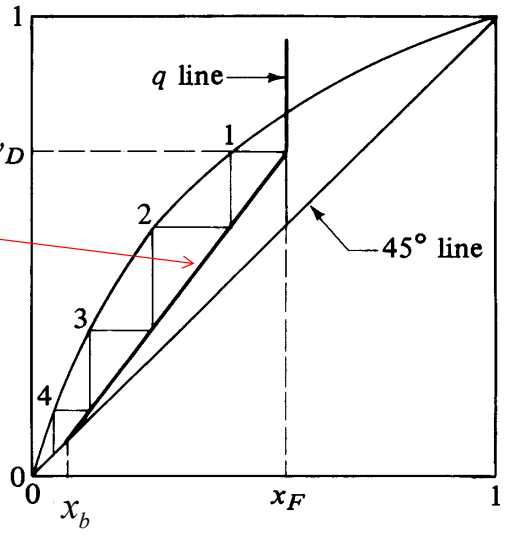


(a)

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

Stripping-Column Distillation

$$slop = \frac{L_m}{V_{m+1}}$$



- If the feed is saturated liquid:

$$L_m = F$$

- If the feed is cold liquid below the boiling point (q >1):

$$L_m = qF$$

- It intersects the *y* = x line (45° diagonal line) at $x = x_b$

Stripping-Column Distillation

- Enriching towers 'are also sometimes used, where the feed enters the bottom of the tower as a vapor.
- The overhead distillate is produced in the same manner as in a complete fractionating tower and is usually quite rich in the more volatile component *A*.
- The liquid bottoms is usually comparable to the feed in composition, being slightly leaner in component *A*.

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

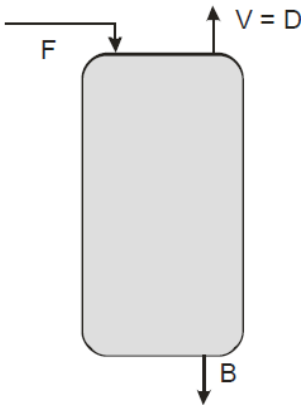
- The q line equation can also be represented as:

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

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Example

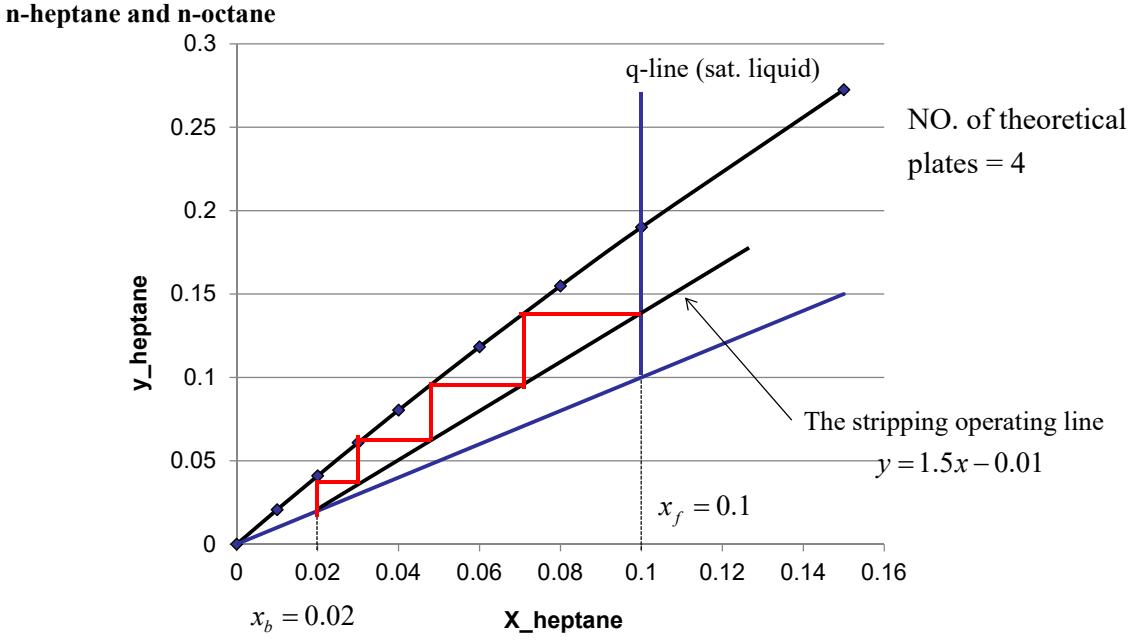
[1]A liquid mixture containing 10 mol % n-heptane and 90 mol % n-octane is fed at its boiling point to the top of a stripping tower at 101.32 kPa. Figure E4.4-5 depicts a stripping tower where the feed stream is the saturated liquid and the distillate stream is the saturated vapor. There is no reboiler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



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Example

Example Contd.



http://vle-calc.com/phase_diagram.html

Direct Steam Injection

- Stream stripping for water clean-up is essentially a distillation process where the heavy product is water and the light product is a mixture of volatile organics.
- These organics are present in the feed water, in relatively small concentrations.
- Since the volatility of the organics is a very strong function of temperature, the high stripping temperature inherent in stream stripping allow for the removal of heavier more soluble organics that are not strippable with air.
- No off-gas treatment is needed ad the only waste stream generated is a small amount of very concentrated organics.

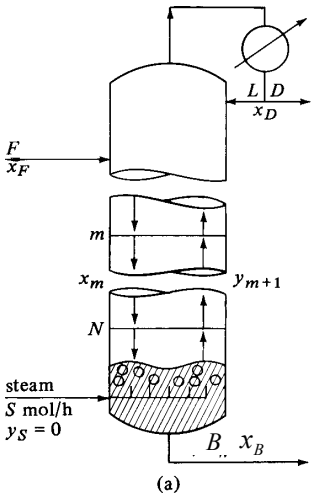
Typical Steam Stripping Applications

Benzene removal from wastewaters
Sour water (h2O and NH3) stripping
Acetone removal/recovery from wastewaters
Oxygenate (MTBE. MEK) removal/recovery
Removal of chloroform, bromoform and other halogenated organics from water
Removal of organics from quench waters
Organics recovery from leachates
Alcohol (ethanol, propanol, IPA, butanol) removal from water
Solvents recovery or removal (tetrahydrofuran, hexane, heptane)

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Rectification with Direct Steam Injection

- Generally, the heat to a distillation tower is applied to one side of a heat exchanger (reboiler) and the steam does not directly contact the boiling solution,
- However, when an aqueous solution of more volatile *A* and water *B* is being distilled, the heat required may be provided by the use of open steam injected directly at the bottom of the tower.
- The reboiler exchanger is then not needed.
- The steam is injected as small bubbles into the liquid in the tower bottom.
- The vapor leaving the liquid is then in equilibrium with the liquid if sufficient contact is obtained



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Rectification with Direct Steam Injection

- Overall balance on the tower

$$F + S = D + B$$

- Balance on component A,

$$Fx_{f,a} + Sy_{s,a} = Dx_{d,a} + Bx_{b,a}$$

where

S mol/h of steam and

$y_{s,a}$ mole fraction of A in steam $(y_{s,a} = 0)$

- The enriching operating-line equation is the same as for indirect steam.

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

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Rectification with Direct Steam Injection

- For the stripping-line equation, an overall balance and a balance on component A are as follows:

$$L_m + S = V_{m+1} + B$$
$$L_m x_m + S(0) = V_{m+1} y_{m+1} + B x_B$$

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

- For saturated steam entering, $S = V_{m+1}$, thus the total balance revels that $L_m = B$

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

The stripping operating line with direct steam injection

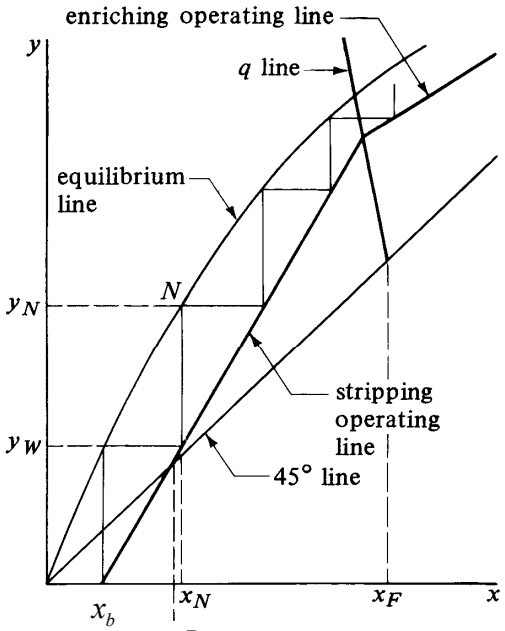
Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Rectification with Direct Steam Injection

- It intersects the $y = x$ line (45° diagonal line) at

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

- The stripping line passes through the point $y = 0$ at $x = x_b$
- For a given reflux ratio and overhead distillate composition, the use of open steam rather than closed requires an extra fraction of a stage, since the bottom step starts below the $y = x$ line.
- The advantage of open steam lies in simpler construction of the heater.



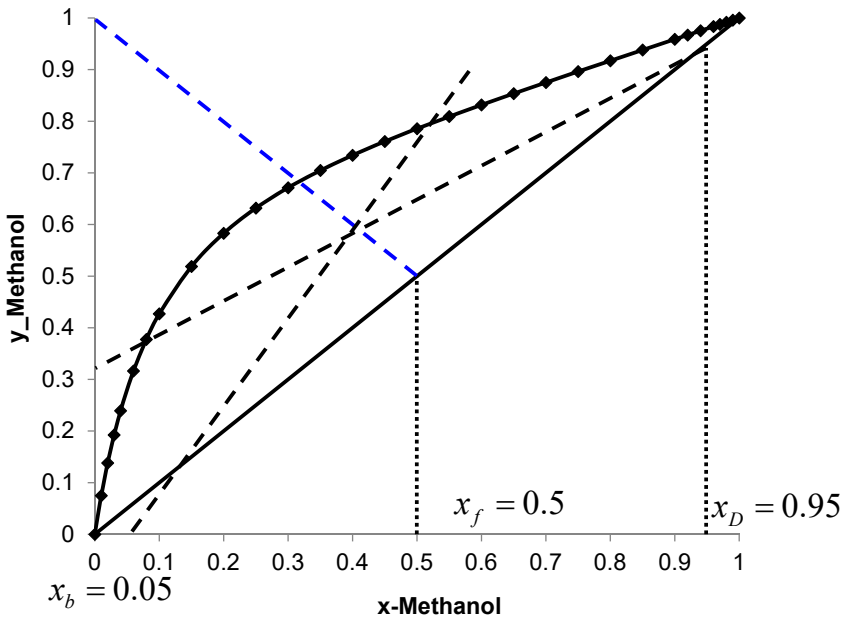
$$x = \frac{Bx_B}{B - S}$$

Example

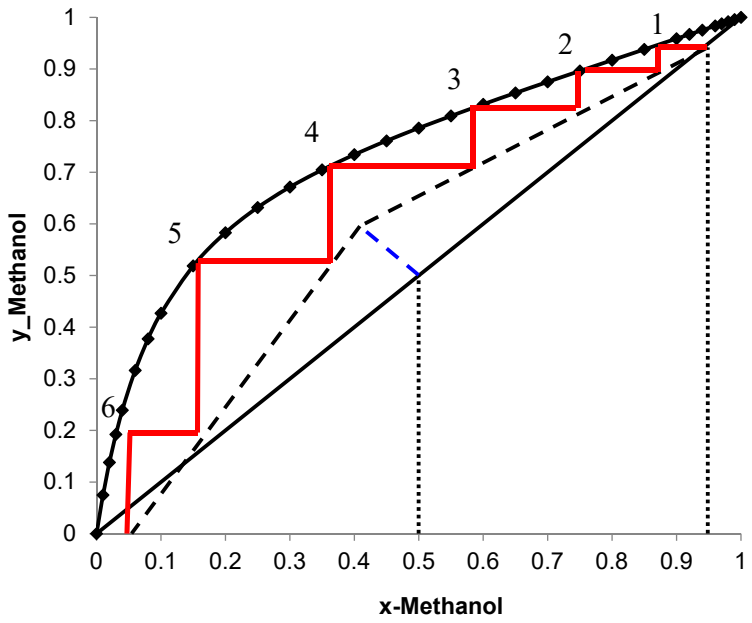
A distillation column operating at 1 atm is to be designed for separating an methanol-water mixture. The feed is a two phase mixture with 50 mole% methanol with 50 % vaporized. A distillate composition of 95 mole% ethanol and a bottoms composition of not more than 5 mole% ethanol are desired. The reflux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

Example Contd.

Example Contd.



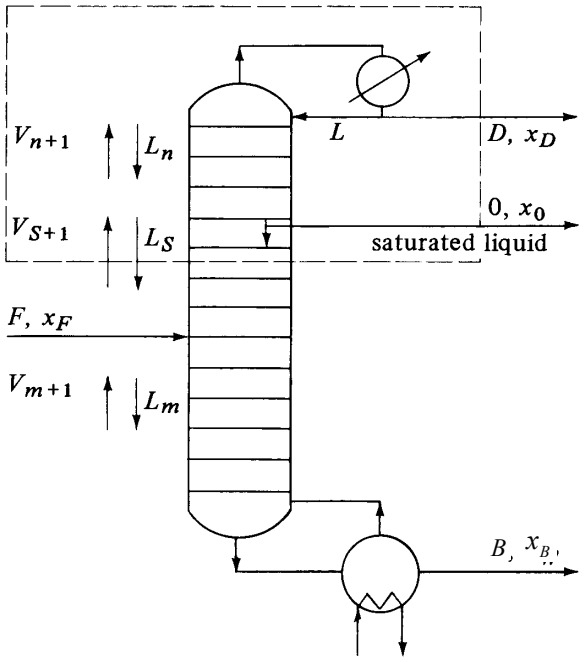
Example Contd.



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Rectification Tower With Side Stream.

- In certain situations, intermediate product or side streams are removed from sections of the tower between the distillate and the bottoms.
- The side stream may be vapor or liquid and may be removed at a point above the feed entrance or below depending on the composition desired.
- The top enriching operating line above the liquid side stream and the stripping operating line below the feed are found in the usual way.
- The equation of the *q* line is also unaffected by the side stream and is found as before.



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Rectification Tower With Side Stream

- Total balance on the tower $F = D + B + O$
- Component balance $Fx_F = x_B B + x_o O + x_D D$
- The liquid side stream alters the liquid rate below it, and hence the material balance or operating line in the middle portion between the feed and liquid side stream plates.

Total material balance on the top portion of the tower

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

where O is mol/h saturated liquid removed as a side stream.

Since the liquid side stream is saturated,

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

Balance on the most volatile component (A)

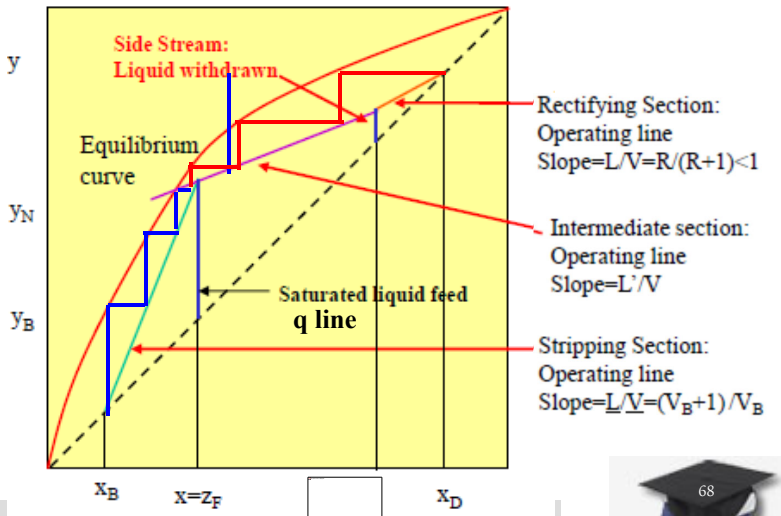
$$V_{s+1} y_{s+1} = L_s x_s + O x_o + D x_D$$

Rectification Tower With Side Stream

Solving for y_{s+1}, the operating line for the region between the side stream and the feed is

$$\Rightarrow y_{s+1} = \frac{L_s}{V_{s+1}} x_s + \frac{O x_O + D x_D}{V_{s+1}}$$

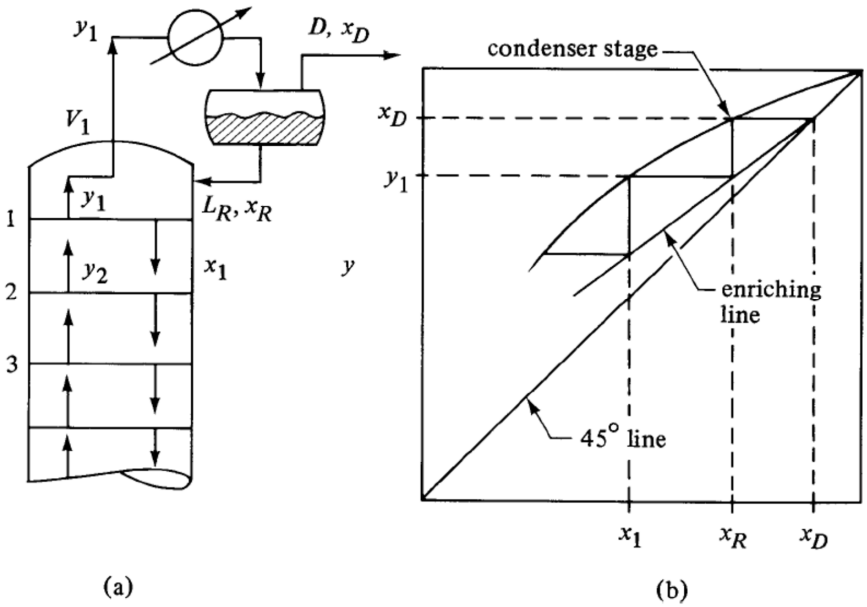
- The line can be located by the *q* line, which determines the intersection of the stripping line and operating line of the side stream, or it may be fixed by the intersection of the side stream line and the enriching section operating line at x_o.



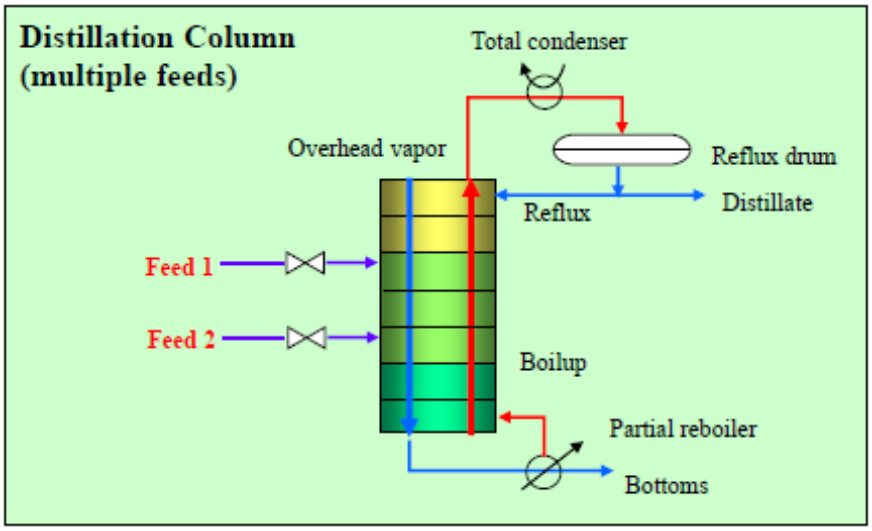
Partial Condensers

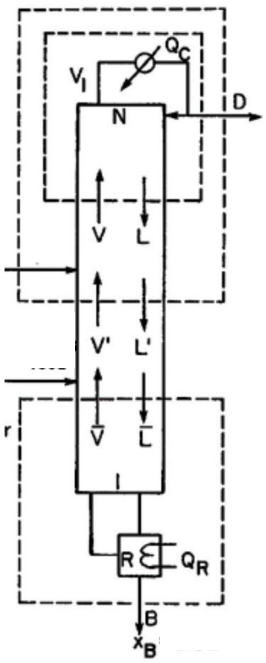
- In the *normal* distillation, the condenser used in the operation is a *total* condenser, at which *all vapour* from the top of the column is *condensed* to *liquid.*
- On the contrary, a ***partial*** condenser ***condenses*** only a ***fraction* of vapour** from the top of the column.
- There is an *equilibrium* between *vapour* (*distillate*: D, with the concentration of x_D) and *liquid* (*reflux*: L_R , with the concentration of x_R) at the condenser. Thus
  - The concentration of the distillate (y_D) is no longer equal to the concentration of the reflux (x_R) ,
  - But x_R can be read from the equilibrium curve when y_D is known
  - The *partial condenser* is considered an *additional* equilibrium *stage*

Partial Condensers



Multiple Feeds

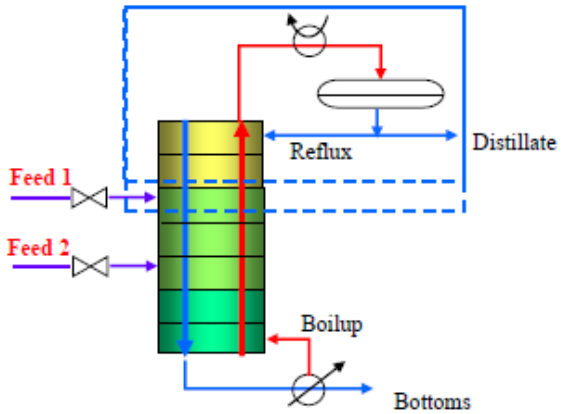


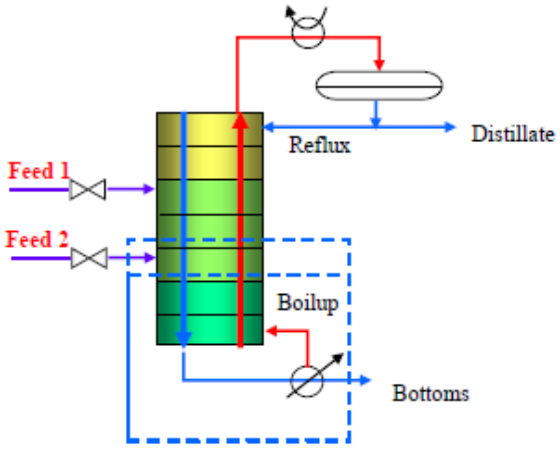


$$F_1 + F_2 = B + D$$

$$F_1 x_{F1} + F_2 x_{F2} = x_B B + x_D D$$

Multiple Feeds



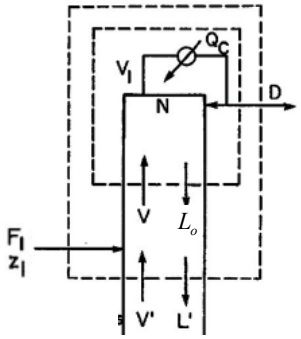


- Each feed stream changes the slope of the operating line from section to section.
- The feed stream changes the flow rates in the stages above and below it. Consequently, it changes the mass balances and the slopes of the operating lines.

Multiple Feeds

- The flow rates above Feed 1 are constant due to constant molar overflow (CMO).
- The feed changes the slope depending on the feed condition.
- Flow rates in the intermediate section are constant, but change when Feed 2 is introduced.
- The *top* operating line or the operating line for the top section – the *section* between the *top* of the column and *Feed 1:*

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



- **<u>The operating line for the middle section</u>** (or the middle operating line) – note that the middle section is the section between the two feed streams,

$$F_1 + V' = D + L'$$

$$x_{F1}F + y'\ V' = Dx_d + x'\ L'$$

$$L' = L_o + q_1 F_1$$

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

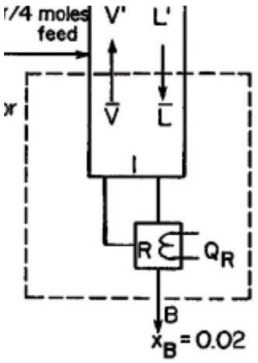
Multiple Feeds

- The material balance lead to

$$\Rightarrow y = \frac{L'}{V'} x + \frac{(x_d D - x_{F1} F_1)}{V'}$$

- The q line for feed 1: $y_q = \frac{q_1}{q_1 - 1} x_q - \frac{x_{f1}}{q_1 - 1}$
- The *middle* operating line must pass through the intersection of the top operating line and the feed line of Feed 1
- The q line for feed 2: $y_q = \frac{q_2}{q_2 - 1} x_q - \frac{x_{f2}}{q_2 - 1}$
- The bottom (stripping) operating line

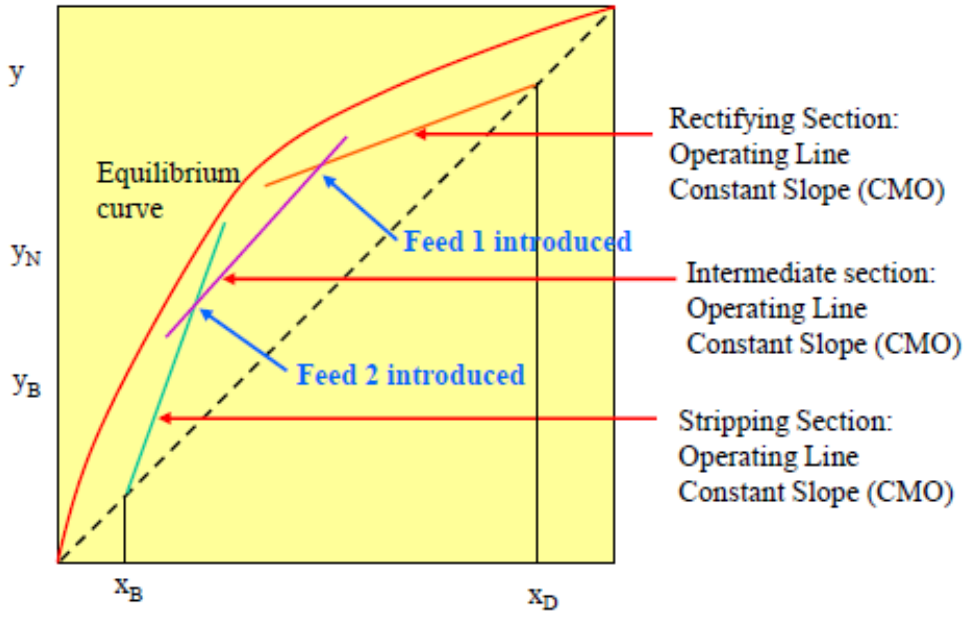
$$y = \frac{\bar{L}}{\bar{V}} x - \frac{B x_B}{\bar{V}}$$



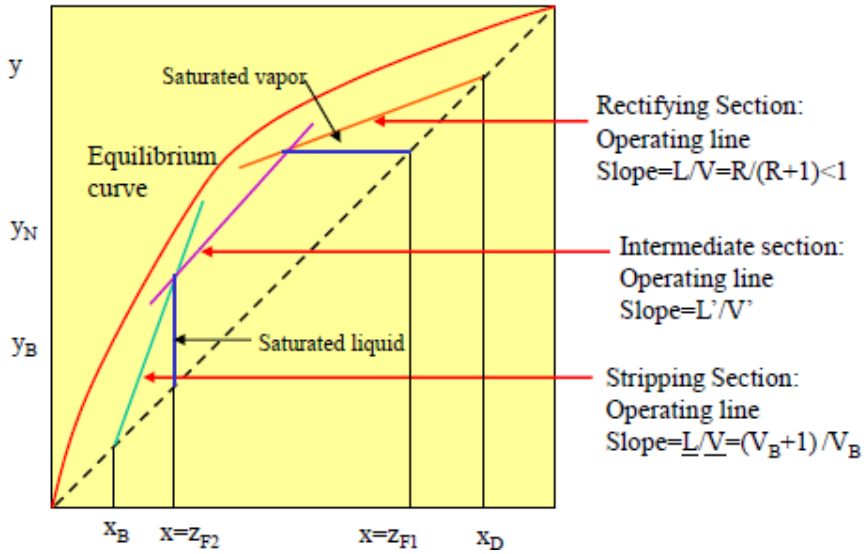
$$\bar{L} = L' + q_2 F_2$$

$$\bar{V} = \bar{V} + (1 - q_2)F_2$$

Multiple Feeds



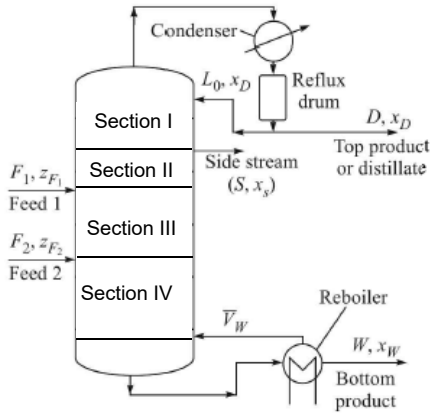
Multiple Feeds



Feed 1 a saturated vapor of composition z_{F1}, and
Feed 2 a saturated liquid of composition z_{F2}

Example

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



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Example Contd

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Example Contd

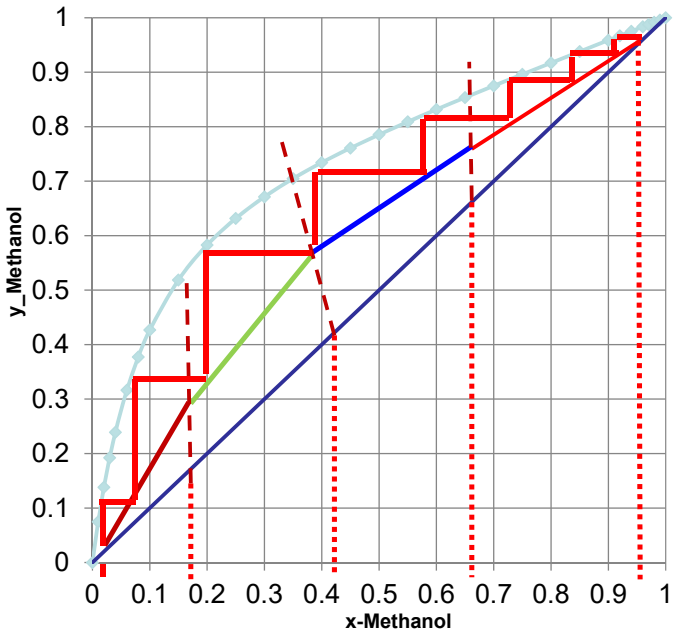
Example Contd

Example Contd

The flow rates of the streams in different sections and the equations of the operating lines are given below.

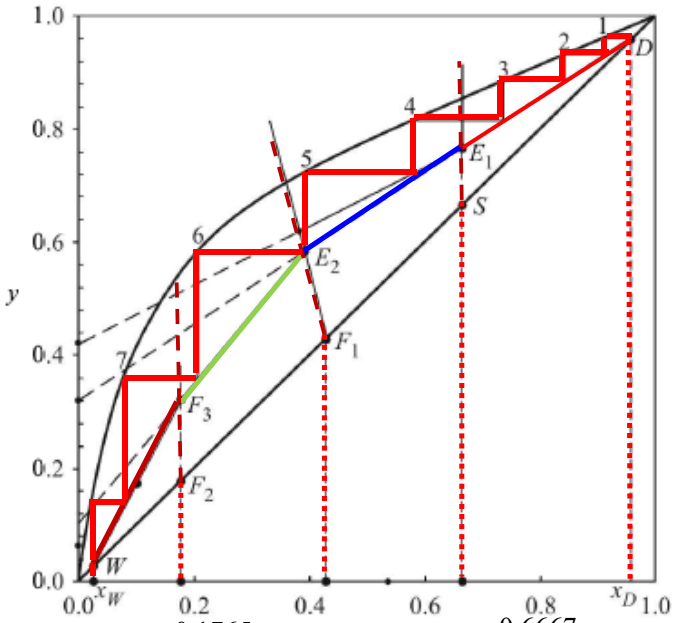
Section	*Liquid rate*	*Vapour rate*	*Slope of the operating line*	*Equation of the operating line*
I	$L(\text{I}) = L_0 = 154.5$	$V = 231.75$	$L_0/V = 0.6667$	$y = 0.6667x + 0.3203$
II	$L(\text{II}) = L(\text{I}) - S = 119.5$	$V = 231.75$	$L(\text{II})/V = 0.5156$	$y = 0.5156x + 0.421$
III	$L(\text{III}) = L(\text{II}) + 0.8F_1$	$V = 231.75 - 0.2F_1$	$L(\text{III})/V = 1.4576$	$y = 1.4576x + 0.06176$
	$= 119.5 + (0.8)(200) = 279.5$	$= 231.75 - 40$	$L(\text{IV})/V = 1.9791$	$y = 1.9791x - 0.03028$
		$= 191.75$		
IV	$L(\text{IV}) = L(\text{III}) + F_2$	$V = 191.75$		
	$= 279.5 + 100 = 379.5$			

Example Contd



- The No. of ideal trays is 7.8.
- Feed 1 enters at the 5th tray,
- Feed 2 enters at the 7th tray
- The side stream is drawn from tray No. 4

Example Contd



- The No. of ideal trays is 7.8.
- Feed 1 enters at the 5th tray,
- Feed 2 enters at the 7th tray
- The side stream is drawn from tray No. 4

$x_W = 0.031$ $x_{f2} = 0.1765$ $x_{f1} = 0.4286$ $x_S = 0.6667$ $x_{D1} = 0.961$

Actual Real Plates

- For an equilibrium tray, the vapor and liquid leaving the tray are in thermodynamic equilibrium.
- In a real tray, equilibrium will rarely be attained.
- The concept of tray efficiency is used to link the performance of a real tray to an equilibrium tray.
- The no. of equilibrium stages must be converted to a number of actual real plates.
- Tray efficiencies can be defined in several ways, the overall tray efficiency, E_o is defined as:

$$E_o = \frac{No.\ of\ theoratical\ trays}{No.\ of\ actual\ trays}$$

Overall Efficiency

- It is applied for the whole column.
- Every tray is assumed to have the same efficiency.
- The overall efficiency depends on the
  i. Geometry and design of the contacting trays,
  ii. Flow rates and flow paths of vapor and liquid streams,
  iii. Compositions and properties of vapor and liquid streams.
- Typical values for tray efficiency are 0.5 to 0.75, I.e. they are 0.5 to 0.75 times as effective as an ideal stage.
- Divide the number of ideal stages by the efficiency, add 10% extra trays.
- The overall efficiency is extremely easy to measure and use; thus, it is the most commonly used efficiency value in the plant.
- However, the overall efficiency is not representative of column operation because the different compositions on the various trays result in different tray efficiencies.

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Overall Efficiency

- The overall efficiency can be also calculated from the following correlations:

i. The Drickamer-Bradford empirical correlation:

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

The corrrelation is valid for hydrocarbon mixtures in the range of 342 K < T < 488.5 K, 1 atm < P < 25 atm and 0.066 < μ < 0.355 cP

ii. The O'Connell correlation:

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

- Add 10% extra trays.

iii. Another equation can be also used

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

where α = relative volatility of key components, and μ = viscosity of feed in cP.

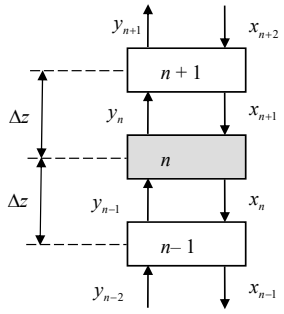
Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Murphree Efficiency

- The efficiency of the tray can also be calculated based on semi-theoretical models which can be interpreted by the Murphree Tray Efficiency E_M.
- In this case it is assumed that the vapor and liquid between trays are well-mixed and have uniform composition.
- It is defined for each tray according to the separation achieved on each tray based on either the liquid phase or the vapor phase.
- For a given component, it can be expressed as:

$$E_{MV} = \frac{y_n - y_{n+1}}{y^*_n - y_{n+1}}$$ Based on vapor phase

$$E_{ML} = \frac{x_n - x_{n-1}}{x^*_n - x_{n-1}}$$ Based on liquid phase



Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Murphree Efficiency

where

y_n is the average actual concentration of the mixed vapor leaving the tray n

y_{n+1} the average actual concentration of time mixed vapor entering tray n,

y^*_n the concentration of the vapor that would be in equilibrium with the liquid of concentration x_n leaving the tray to the downcomer.

- The liquid entering the tray has a concentration of x_{n-1} and as it travels across the tray, its concentration drops to x_n at the outlet.
- Hence, there is a concentration gradient in the liquid as it flows across the tray.
- The vapor entering the tray contacts liquid of different concentrations, and the outlet vapor will not be uniform in concentration.

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888

Murphree Efficiency

- Van Winkle's correlation can be used to predict plate efficiencies for binary systems.
- The data used to derive the correlation covered both bubble-cap and sieve plates.

$$E_{MV} = 0.07\, Dg^{0.14}\, Sc^{0.25}\, Re^{0.08}$$

where

Dg = surface tension number = $(\sigma_L/\mu_L u_v)$,
u_v = superficial vapor velocity,
σ_L = liquid surface tension,
μ_L = liquid viscosity,
Sc = liquid Schmidt number = $(\mu_L/\rho_L D_{LK})$,
ρ_L = liquid density,
D_{LK} = liquid diffusivity, light key component,
Re = Reynolds number = $(h_w u_v \rho_v/\mu_L(FA))$,
h_w = weir height,
ρ_v = vapor density,
(FA) = fractional area = (area of holes or risers)/(total column cross-sectional area)

Chemical Engineering Department | University of Jordan | Amman 11942, Jordan
Tel. +962 6 535 5000 | 22888