Mass Transfer Operations I
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Module - 5
Distillation
Lecture - 12
Ponchon and Savarit Method and Packed Tower Distillation

Welcome to the 12th lecture of module 5, in this module we are discussing Distillation.

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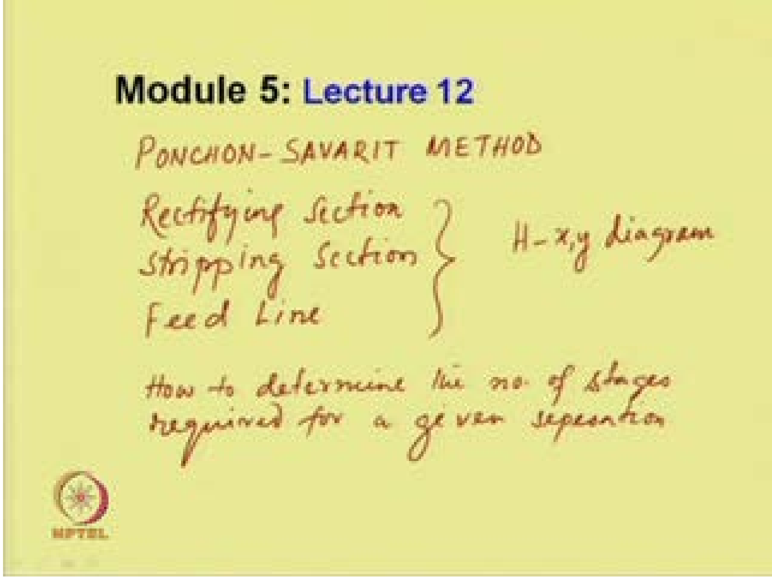

Module 5: Lecture 12

PONCHON-SAVARIT METHOD

Rectifying Section
Stripping Section } H-x,y diagram
Feed Line

How to determine the no. of stages required for a given separation

Let us have a recap of our previous lecture on this topic; in our previous lecture we have primarily considered the design of a distillation column, by Ponchon and Savarit method, Ponchon-Savarit method. Primarily we have discussed for a fractionators, what is the balance equations for the rectifying section section, stripping section and feed line. All the balance equations we have derived. And we know how to construct the H x, y diagram; that is enthalpy concentration diagram.

So, we will continue our previous lecture with how to determine the number of stages, how to determine the number of stages required for our given separation, using enthalpy concentration diagram. And in this case, the main thing is that as we said earlier we assume the constant molar overflow, in case of McCabe-Thiele method, but in case of

Ponchon and Savarit we need the enthalpy data for different compositions, and we need not to assume the constant molar overflow rate.

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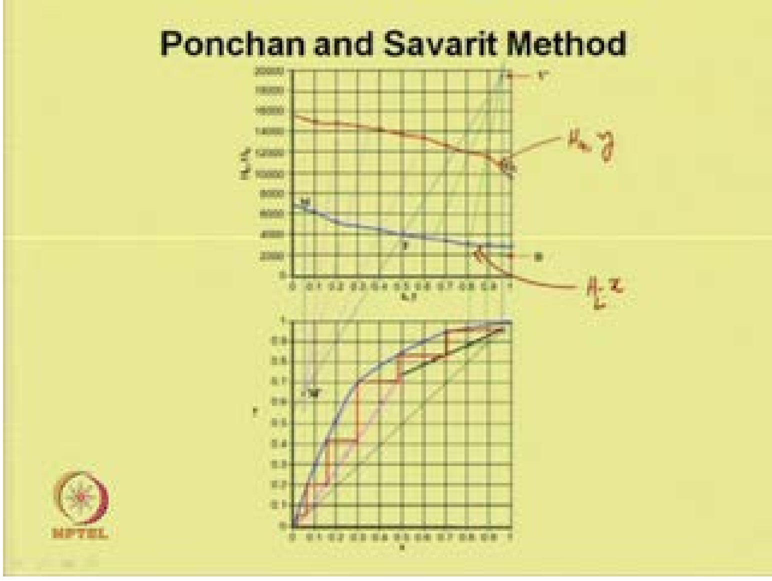

Ponchan and Savarit Method

In this case we know the equilibrium data, x, y data and we know the H enthalpy of the liquid, and enthalpy of the vapor. And so we can plot the lower curve represents, H L x and this is H G y.

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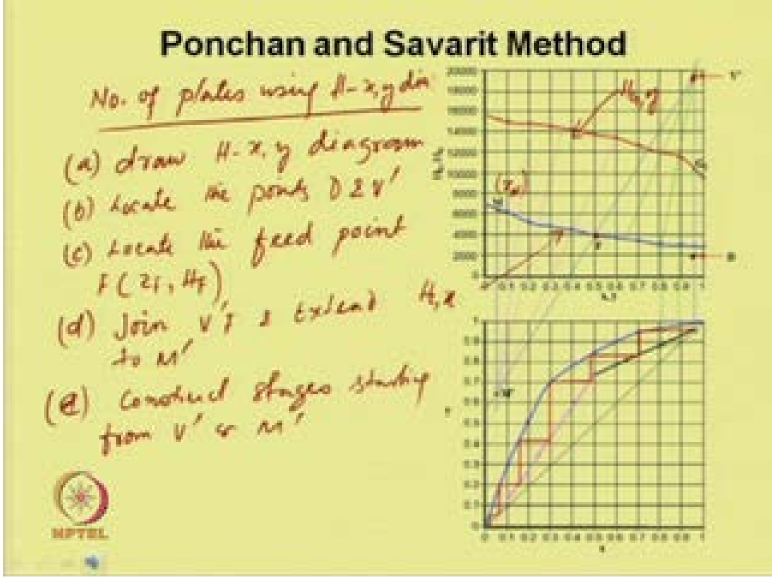

Ponchan and Savarit Method

No. of plates using H-x,y dia

(a) draw H-x, y diagram

(b) Locate the points D 2 V'

(c) Locate the feed point F (zF , HF)

(d) Join V' F & extend H, x to M'

(e) Construct stages starting from V' & M'

First we will consider number of plates, number of plates using H x, y diagram here, first we have to draw H x, y diagram with the help of given data which is plotted over here, this is H L x and this is H G y. Then we have to locate locate the points D and V dash which is over here, this is point d and this is point v dash; and we know how to calculate the enthalpy at point V dash, whose concentration at this point is Z, D. Then locate the feed point which is F is Z F and H F which is located over here, now join V dash F and extend to M dash.

The point M dash which is based on the bottom compositions, which is x w at point M and the concentration is x w, we can draw a perpendicular line at point M downwards, and then we can extend the line V dash F to M dash. Then construct the stages construct a stage starting from V dash or M dash, a change over has to be made at the feed point at this locations F and we have to proceed the construction till the end point is reached. For this case, if we use only the enthalpy concentration diagram to find out the number of plates, required for a given separations we need to have the equations for the tie line; if tie lines are not available it will be easier to handle with the use of the x, y diagram, in conjunction with the H x, y diagram.

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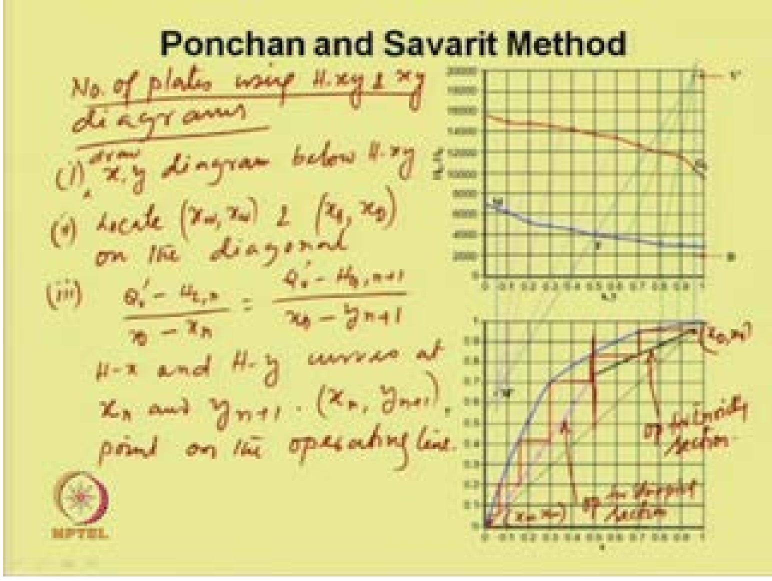

Ponchan and Savarit Method

No. of plates using H.xy & xy diagrams

(i) draw x.y diagram below H.xy

(ii) locate (x_w, x_w) & (x_D, x_D) on the diagonal

(iii) $\frac{Q_v' - H_{L,n}}{x_D - x_n} = \frac{Q_v' - H_{G,n+1}}{x_D - y_{n+1}}$

H-x and H-y curves at x_n and y_{n+1}. (x_n, y_{n+1}), point on the operating line.

Now, let us see how to obtain the number of plates using H x, y and x, y diagram. So, first we have to draw the x, y diagram below H x, y diagram with the similar scale, and then locate x w, x w and x D, x D on the diagonal, which is located over here, and another point is located over here. So, on the diagram the point x w, x w and x D, x D, so this points are located and then as we know from the earlier derivations Q v dash minus

H L, n divided by x D minus x n is equal to Q v dash minus H G n plus 1 divided by x D minus y n plus 1.

So, that any line which is immediate from point V dash or point M dash, intersects the H x and H y curves at x n and y n plus 1, this is the points x n and y n plus 1 is the point on the operating line. If we draw several lines starting from at point V, and if we draw the line starting from X D and X D, and we obtain several points on the operating lines; and we put join all these points, and we will obtain the operating line for the enriching system, this the operating line for enriching section.

Similarly, if we start drawing different lines from point M dash in a similar procedure, we could obtain operating line for the striping section; this is the operating line for the striping section. From the two different curve H x and H y, we can obtain the equilibrium curve which is given below, we know the equilibrium curve, we know the operating line for the different section, and we know the feed point; if feed condition is known to us we can draw the feed line. And then similar procedure as we followed like McCabe-Thiele, we could follow the same procedure that is (( )) constructions of the stages, and we could obtain the number of stages required for the given separation.

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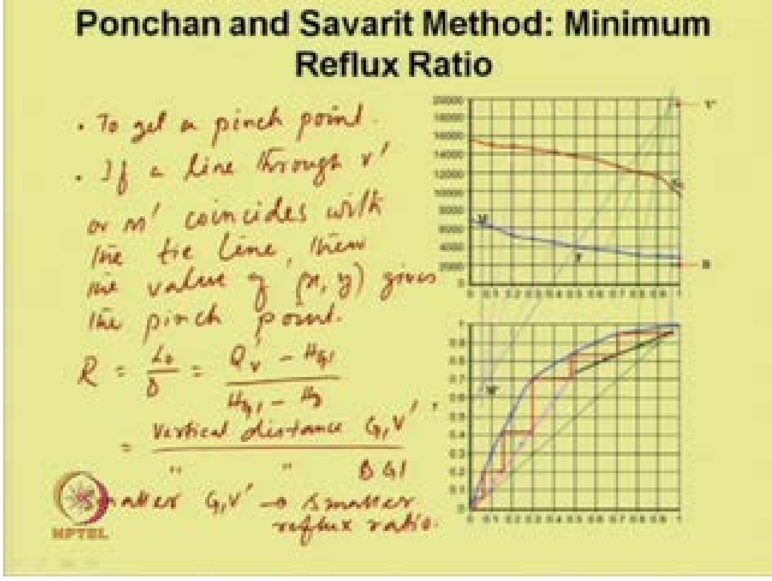

Ponchan and Savarit Method: Minimum Reflux Ratio

- To get a pinch point
- If a line through v' or m' coincides with the tie line, then the value of (x, y) gives the pinch point.

$$R = \frac{L_0}{D} = \frac{Q_{v'} - H_{G1}}{H_{G1} - H_D}$$

$$= \frac{\text{Vertical distance } G_1V'}{\text{" " } DG_1}$$

Smaller G_1V' → smaller reflux ratio.

Now, using Ponchon and Savarit method, we could obtain the minimum reflux ratio to obtain the minimum reflux ratio which is R mean, we need to obtain the pinch point, first to get a pinch point to obtain the pinch point if a line through V dash or M dash coincides with the tie line. Then the value of (x, y) gives the pinch point, we know the reflux ratio equations R is equal to L naught by D is equal to Q dash v minus H G 1 divided by H G 1 minus H D which is equal to vertical distance vertical distance G 1 V dash divided by vertical distance D G 1. If V dash is nearer to G 1 or the distance G 1 V dash is as small, it increases the number of plates required or the reflux ratio is smaller smaller G 1 V dash smaller reflux ratio.

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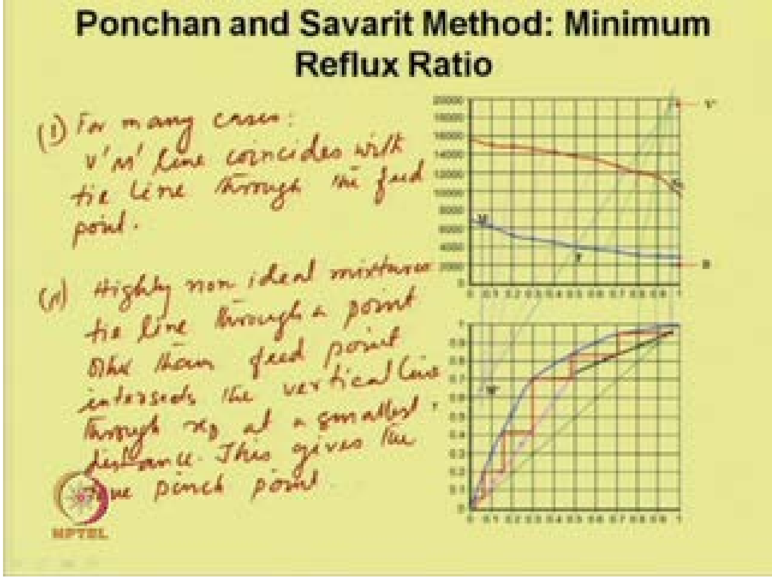

Ponchan and Savarit Method: Minimum Reflux Ratio

(i) For many cases:
V'M' line coincides with tie line through the feed point.

(ii) Highly non ideal mixtures tie line through a point other than feed point intersects the vertical line through x_D at a smallest distance. This gives the true pinch point

So, in many cases V dash M dash line coincides with tie line through the feed point. In some situation where for highly non ideal mixtures, the tie line through a point other than feed point intersects the vertical line, through X D at a smallest distance, this give the true pinch point. As soon as the pinch point is obtain, then the same can be calculated using the equations described earlier, using this same equations we can use to obtain the minimum reflux ratio.

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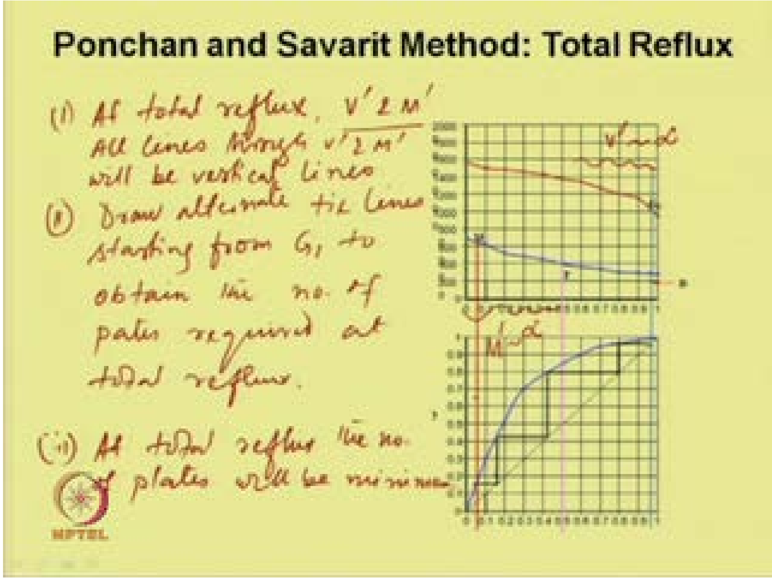

Ponchan and Savarit Method: Total Reflux

(i) At total reflux, V' & M'
All lines through V' & M'
will be vertical lines

(ii) Draw alternate tie lines
starting from G_1 to
obtain the no. of
plates required at
total reflux.

(iii) At total reflux the no.
of plates will be minimum

Now, we will discuss the total reflux, at total reflux total reflux the point V dash and M dash this two points will be add infinity (No audio from 16:54 to 17:10), this two points will be M dash and V dash will be at infinity, the line which will joint between M dash and V dash will be a vertical line. So, all the lines between M dash and V dash will be the vertical line, so V dash and M dash will be vertical lines.

So, draw alternate tie lines starting from G 1 to obtain the obtain the number of plates required at total reflux, so at total reflux reflux the number of plates will be minimum. Now, it will be easy if we use the x, y diagram instead of the enthalpy concentration diagram, like we have done before we can draw the x, y curve using the H x and H y data which is plotted over here.

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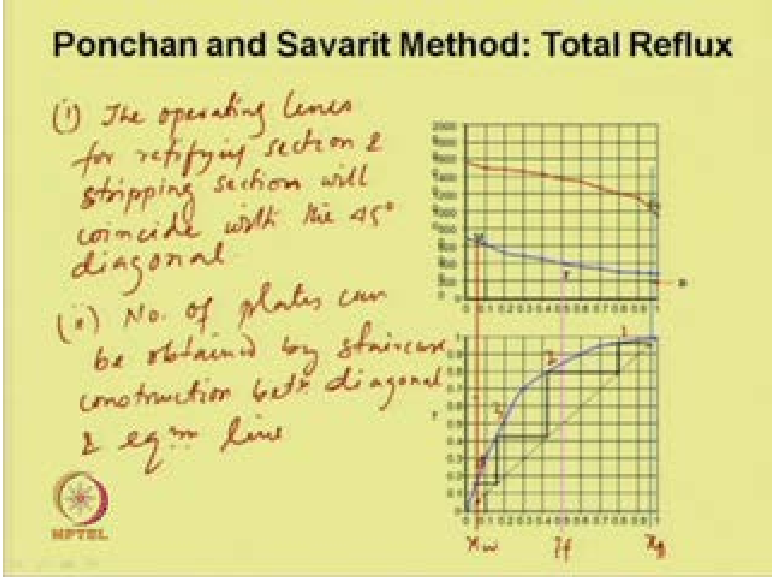

Ponchan and Savarit Method: Total Reflux

(i) The operating lines for rectifying section & stripping section will coincide with the 45° diagonal

(ii) No. of plates can be obtained by staircase construction betn diagonal & eqm line

The operating lines for rectifying section section, and stripping section will coincide with the 45 degree diagonal, and the number of plates number of plates can be obtained by staircase contraction, between diagonal and equilibrium line; which is shown over here, start at the point of X D, this is the point of X D. So, locate this point, this is the feed point Z F and this is X W, so this points are located and then we know the equilibrium curve, so join them this is 1, 2, 3 and 4, so this is the number of plates at total reflux.

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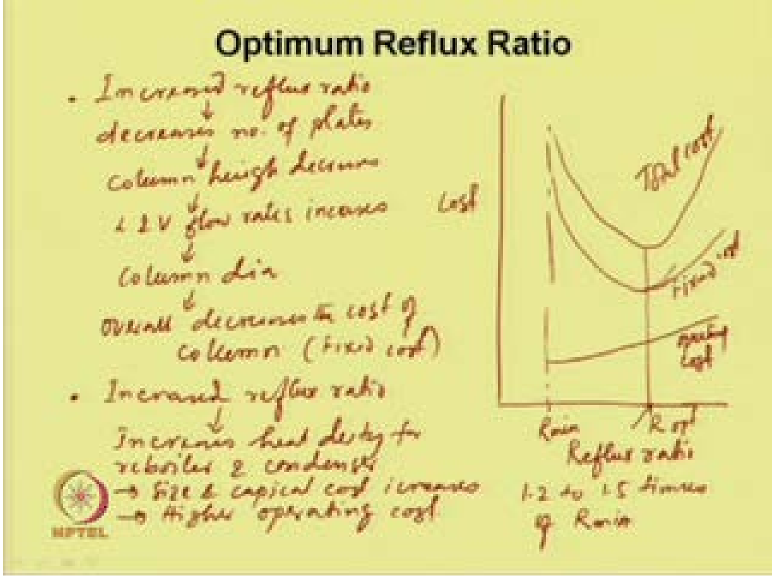

Optimum Reflux Ratio

- Increased reflux ratio
  ↓
  decreases no. of plates
  ↓
  column height decreases
  ↓
  L & V flow rates increase
  ↓
  column dia
  ↓
  overall decreases the cost of column (fixed cost)
- Increased reflux ratio
  ↓
  Increases heat duty for reboilers & condensers
  → size & capital cost increases
  → Higher operating cost

1.2 to 1.5 times of Rmin

Now, what is the optimum reflux ratio, when we design a new column we need to use a optimum reflux ratio, as we increase the reflux ratio increased reflux ratio what happens, it decreases the number of plates decreases number of plates required, so the column height decreases. On the other hand what happens, liquid and vapor flow rates increases, so it increases the column diameter. And overall decreases the cost of column, this is the fixed cost; another thing is that as we increase the reflux ratio, increases the heat duty increases heat duty for reboiler and condenser.

So, the size size and capital cost increases also higher operating cost, so overall if we plot cost verses reflux ratio, there will be a minimum reflux ratio which is R min and the operating cost would gradually increase, as we increase the reflux ratio this is operating cost. And the fixed cost initially will be higher, and then it decrease and then it increases, so this is a fixed cost. So, the total cost will increase, and so there will be a minimum of total cost curve, so this will give the optimum are optimum that is optimum reflux ratio for the operations, this in general 1.2 to 1.5 times of R m R min.

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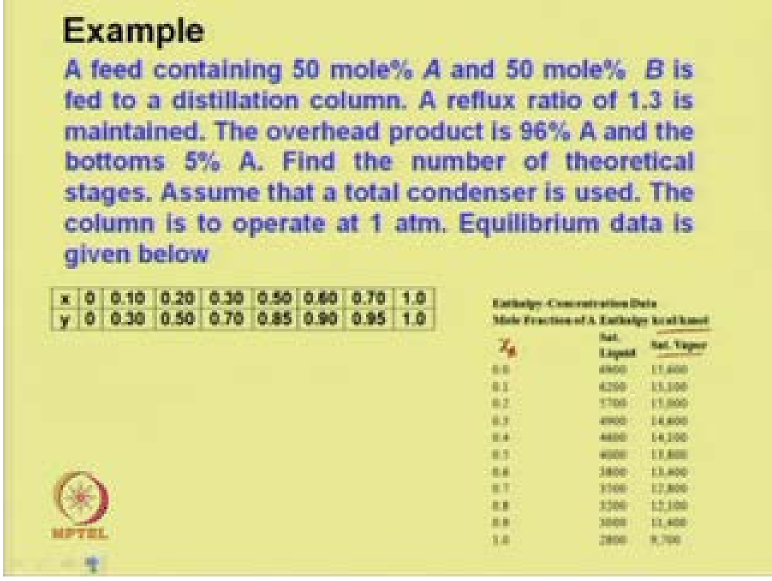

Example

A feed containing 50 mole% *A* and 50 mole% *B* is fed to a distillation column. A reflux ratio of 1.3 is maintained. The overhead product is 96% A and the bottoms 5% A. Find the number of theoretical stages. Assume that a total condenser is used. The column is to operate at 1 atm. Equilibrium data is given below

x	0	0.10	0.20	0.30	0.50	0.60	0.70	1.0
y	0	0.30	0.50	0.70	0.85	0.90	0.95	1.0

Enthalpy-Concentration Data

Mole Fraction of A x_A	Enthalpy kcal/kmol Sat. Liquid	Sat. Vapor
0.0	4900	17,600
0.1	6200	15,100
0.2	5700	15,000
0.3	4900	14,600
0.4	4600	14,200
0.5	4000	13,800
0.6	3800	13,400
0.7	3500	12,800
0.8	3200	12,100
0.9	3000	11,400
1.0	2800	9,700

Let us take an example for calculations of the number of plates, using Ponchon-Savarit method a feet containing 50 mole percent A, and 50 mole percent B a feat to a distillation column. A reflux ratio of 1.3 each maintains the over head product is 96 percent A, and the bottom is 5 percent A, find the number of theoretical stages. Assume that a total condenser is use, the column is to operate at one atmosphere equilibrium data is given below, this represent mole fraction of component A that is X A. And this is for saturated liquid enthalpy in terms of kilo calorie per kilo mole, and this is saturated liquid, and this is saturated vapor.

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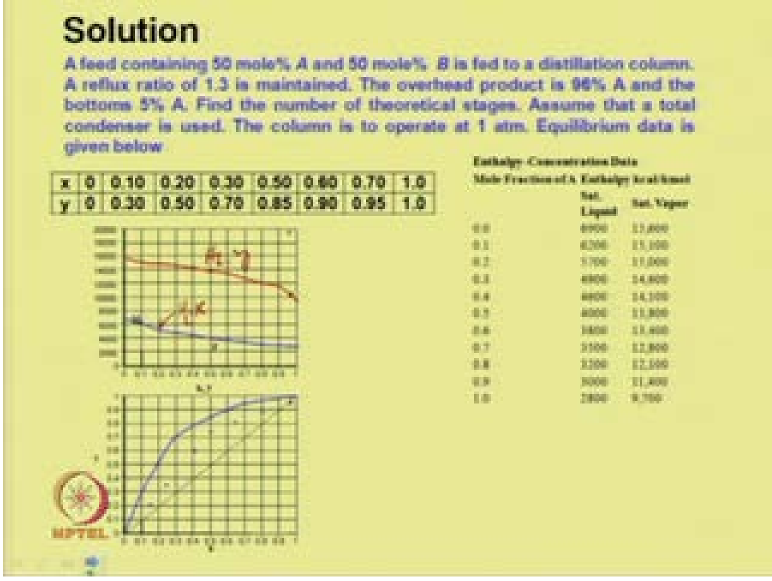

Solution

A feed containing 50 mole% *A* and 50 mole% *B* is fed to a distillation column. A reflux ratio of 1.3 is maintained. The overhead product is 96% A and the bottoms 5% A. Find the number of theoretical stages. Assume that a total condenser is used. The column is to operate at 1 atm. Equilibrium data is given below

x	0	0.10	0.20	0.30	0.50	0.60	0.70	1.0
y	0	0.30	0.50	0.70	0.85	0.90	0.95	1.0

Enthalpy-Concentration Data

Mole Fraction of A	Enthalpy kcal/kmol Sat. Liquid	Sat. Vapor
0.0	6900	15,600
0.1	6200	15,100
0.2	5700	15,000
0.3	4900	14,600
0.4	4600	14,100
0.5	4000	13,800
0.6	3800	13,400
0.7	3500	12,800
0.8	3200	12,100
0.9	3000	11,400
1.0	2800	9,700

With the help of this data we can plot the equilibrium curve, and we can plot the H x, y diagram H x and H L x and H L y these are the two plots. And the points which is giving feed points, and this is the bottom composition and this is the X D, which is located over here is given, so this points are located.

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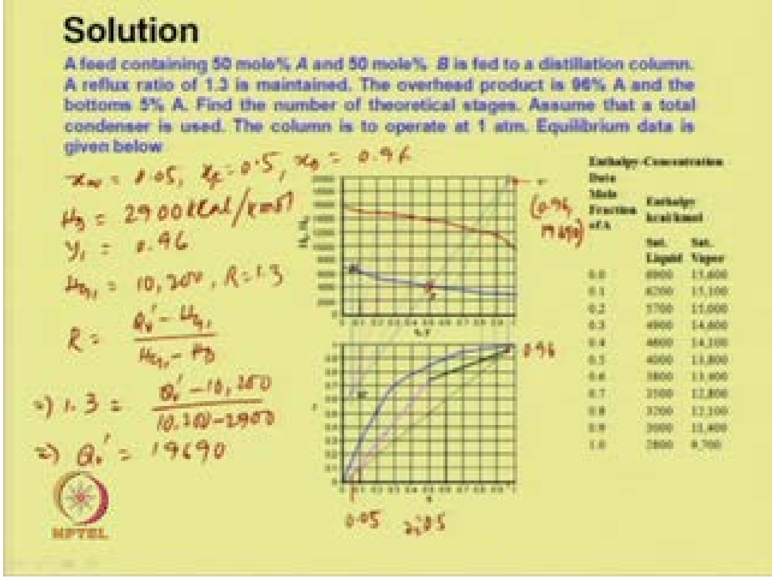

Solution

A feed containing 50 mole% *A* and 50 mole% *B* is fed to a distillation column. A reflux ratio of 1.3 is maintained. The overhead product is 96% A and the bottoms 5% A. Find the number of theoretical stages. Assume that a total condenser is used. The column is to operate at 1 atm. Equilibrium data is given below

$x_W = 0.05,\ x_F = 0.5,\ x_D = 0.96$

$H_D = 2900\ \text{kcal/kmol}$

$y_1 = 0.96$

$H_{G_1} = 10,300,\ R = 1.3$

$$R = \frac{Q_D' - H_{G_1}}{H_{G_1} - H_D}$$

$$\Rightarrow 1.3 = \frac{Q_D' - 10,300}{10,300 - 2900}$$

$$\Rightarrow Q_D' = 19690$$

Enthalpy-Concentration Data

Mole Fraction of A	Enthalpy kcal/kmol Sat. Liquid	Sat. Vapor
0.0	6900	15,600
0.1	6200	15,100
0.2	5700	15,000
0.3	4900	14,600
0.4	4600	14,100
0.5	4000	13,800
0.6	3800	13,400
0.7	3500	12,800
0.8	3200	12,100
0.9	3000	11,400
1.0	2800	9,700

And now, we can plot the operating line between the n reaching section, so we have located the point V dash given that X W is equal 0.05 and X F is 0.5 and X D is 0.96, So, this points are located, this is a 0.96, this is X W which is 0.05 and this is the feed is 0.5. Now, from the table we can see the enthalpy of distillate H D is given 2900 kilo calorie per k mole, and y 1 is 0.96 H G 1 is 10200 corresponding to this, and then the reflux ratio is given R is equal to 1.3.

Now, if we use R is equal to Q v dash minus H G 1 divided by H G 1 minus H D, so it will be 1.3 would be equal to Q v dash minus 10200 divided by 10200 minus 2900. So, from this you can obtain Q v dash is equal to 19690 which are located over here, V dash which is value at 0.96 and 19690, so this is the point which is located V dash. Now, the overall enthalpy lines, we can draw the lines from point V dash to M dash and which will intersect the point at F, feed point F which is Z F is equal to 0.5.

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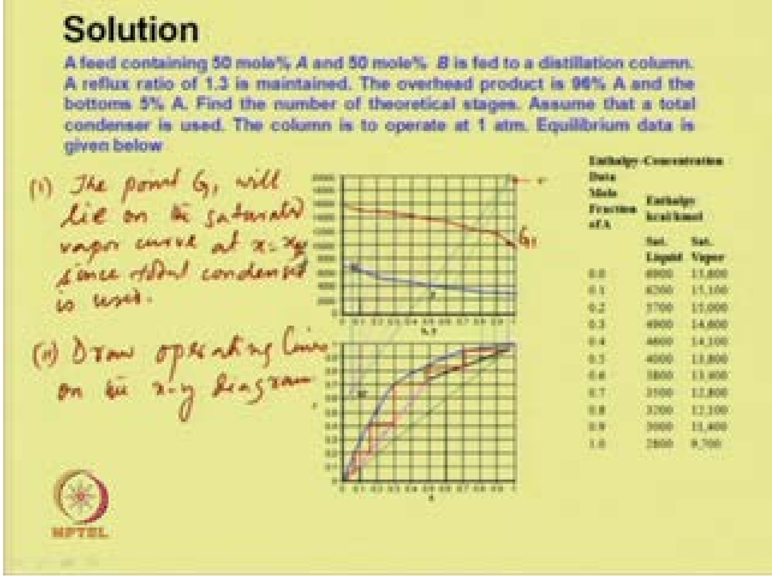

Solution

A feed containing 50 mole% *A* and 50 mole% *B* is fed to a distillation column. A reflux ratio of 1.3 is maintained. The overhead product is 96% A and the bottoms 5% A. Find the number of theoretical stages. Assume that a total condenser is used. The column is to operate at 1 atm. Equilibrium data is given below

(i) The point G_1 will lie on the saturated vapor curve at $x = x_D$ since total condenser is used.

(ii) Draw operating line on the x-y diagram

Enthalpy-Concentration Data

Mole Fraction of A	Enthalpy kcal/kmol Sat. Liquid	 Sat. Vapor
0.0	6900	15,800
0.1	6300	15,100
0.2	5700	15,000
0.3	4900	14,600
0.4	4600	14,100
0.5	4000	13,800
0.6	3800	13,400
0.7	3500	12,800
0.8	3200	12,100
0.9	3000	11,400
1.0	2800	9,700

And then the point G 1 which is over here, G 1 will lie on the saturated vapor curve at x is equal to X D, since total condenser is used. Now, we can draw operating lines on the x, y diagram operating lines are drawn between these two sections, and then the contraction was followed from this points over here.

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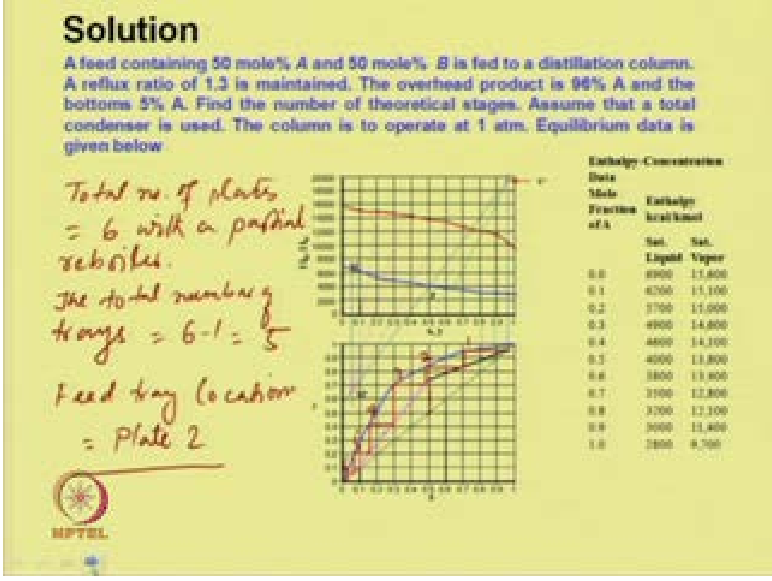

Solution

A feed containing 50 mole% *A* and 50 mole% *B* is fed to a distillation column. A reflux ratio of 1.3 is maintained. The overhead product is 96% A and the bottoms 5% A. Find the number of theoretical stages. Assume that a total condenser is used. The column is to operate at 1 atm. Equilibrium data is given below

Total no. of plates = 6 with a partial reboiler.

The total number of trays = 6-1 = 5

Feed tray location = Plate 2

Enthalpy-Concentration Data

Mole Fraction of A	Enthalpy kcal/kmol	
	Sat. Liquid	Sat. Vapor
0.0	6900	15,600
0.1	6200	15,100
0.2	5700	15,000
0.3	4900	14,600
0.4	4600	14,100
0.5	4000	13,800
0.6	3800	13,400
0.7	3500	12,800
0.8	3200	12,100
0.9	3000	11,400
1.0	2800	9,700

So, 1, 2, 3, 4, 5, 6 total number of plates is equal to 6 with a partial reboiler, the total number of total number of trays is equal to 6 minus 1 is equal to 5. And the optimum feed tray locations feed tray location is equal to we can see which is the saturated liquid which is over here, plate number two plate two this way we can obtain the number of ideal stages required, and locate the feed tray.

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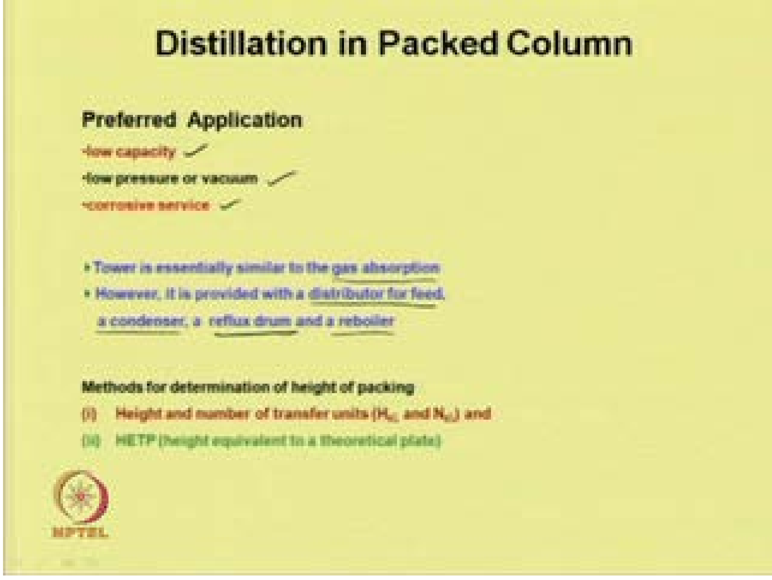

Distillation in Packed Column

Preferred Application

- low capacity
- low pressure or vacuum
- corrosive service

- Tower is essentially similar to the gas absorption
- However, it is provided with a distributor for feed, a condenser, a reflux drum and a reboiler

Methods for determination of height of packing

(i) Height and number of transfer units (H_{OG} and N_{OG}) and

(ii) HETP (height equivalent to a theoretical plate)

Now, we will discuss distillation in packed column, in some cases where packed column distillation is vapor, particularly in case of low capacity applications or low pressure or vacuum applications, and corrosive services. And the tower used for this packed column distillation is essentially similar to, the one which you have used for gas absorption which we have discussed before, but with a little modification.

And this tower is particularly modified with a distributor for the feed, a condenser, a reflux drum, and a reboiler, in this case we have to determine then height of packing required for a given separation. How to determine the two general methods, generally (( )) one is the height and number of transfer unit method, which is HTG or NTG method, and then HTP method, HTP is equivalent way theoretical plate, height equivalent way theoretical plate method which essentially similar, which we have discuss for packed tower absorption.

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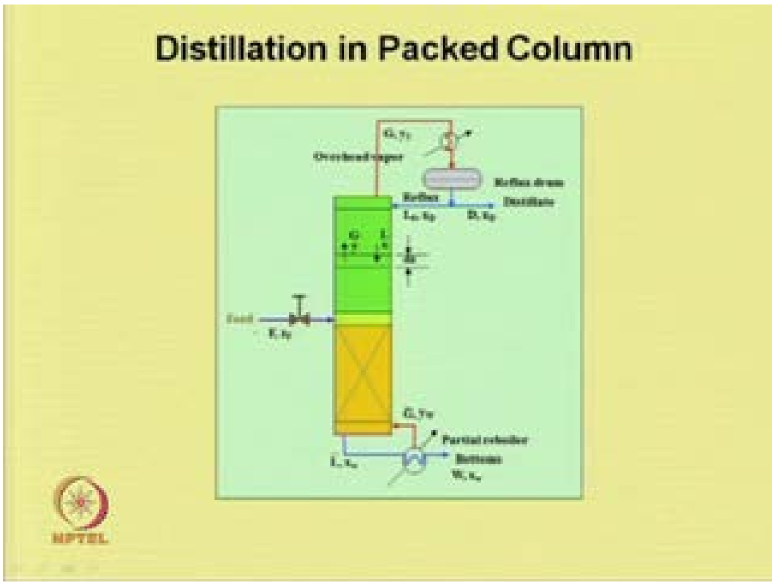

Distillation in Packed Column

Now, we will discuss height and number of transfer unit method, this is typical pack distillation column, where feed enters at particular locations with a composition Z F. And we have reflux dram, and the distillate there is a condenser and similarly, at the bottom we have partial reboiler as in distillation.

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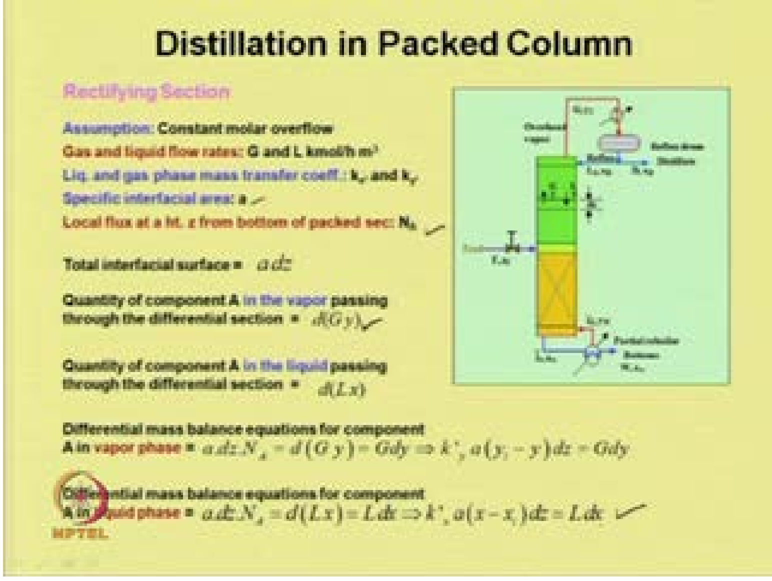

Distillation in Packed Column

Rectifying Section

Assumption: Constant molar overflow

Gas and liquid flow rates: G and L kmol/h m²

Liq. and gas phase mass transfer coeff.: k_x and k_y

Specific interfacial area: a

Local flux at a ht. z from bottom of packed sec: N_A

Total interfacial surface = $a\,dz$

Quantity of component A in the vapor passing through the differential section = $d(G\,y)$

Quantity of component A in the liquid passing through the differential section = $d(L\,x)$

Differential mass balance equations for component A in vapor phase = $a\,dz\,N_A = d(G\,y) = G dy \Rightarrow k'_y\, a(y_i - y)dz = G dy$

Differential mass balance equations for component A in liquid phase = $a\,dz\,N_A = d(L\,x) = L dx \Rightarrow k'_x\, a(x - x_i)dz = L dx$

Let us considered the rectifying section how to do the balance, for the rectifying section we have following assumptions constant molar over flow, liquid and gas flow rate G and L respectively, liquid and gas phase mass transfer co-efficient small k, x dash and k, y dash, specific interfacial area A. And local flax at a particular height Z from the bottom of the packed section is N A, now total interfacial surface would be a d z, so we have taken a essential section of thickness d z.

So, the total interfacial surface would be a, interfacial are d z and the vapor component which is passing through this differential section would be d (G, y). Similarly, the liquid which is passing through this different section would be d (L x), so if we do the differential mass balance for the vapor phase, it would be a d z and N a, would be d of D into y, so we will get k y dash a (y i minus y) d z is equal to G d y, where y i is the interfacial concentration. Similarly, for the liquid phase we can obtain k x dash a into x minus x i d z is equal to L ds.

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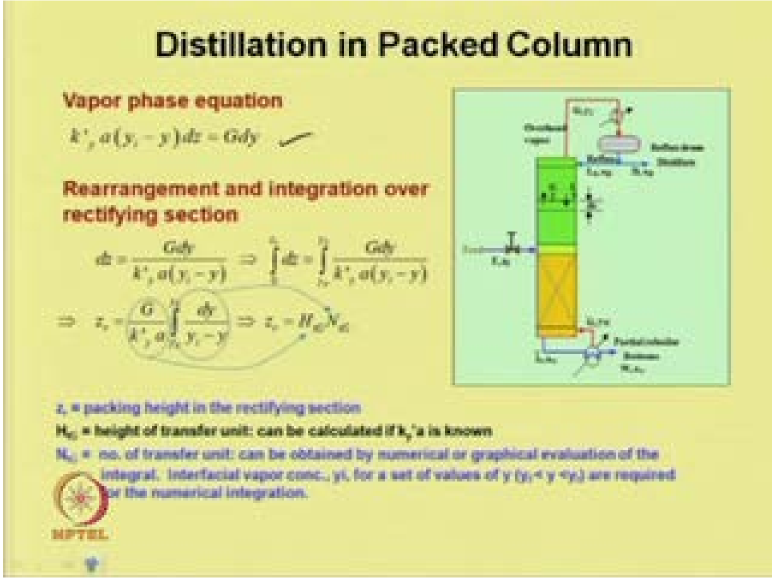

Distillation in Packed Column

Vapor phase equation

$$k'_y a(y_i - y)dz = Gdy$$

Rearrangement and integration over rectifying section

$$dz = \frac{Gdy}{k'_y a(y_i - y)} \Rightarrow \int_0^{z_r} dz = \int_{y_f}^{y_2} \frac{Gdy}{k'_y a(y_i - y)}$$

$$\Rightarrow z_r = \frac{G}{k'_y a}\int_{y_f}^{y_2} \frac{dy}{y_i - y} \Rightarrow z_r = H_{tG} N_{tG}$$

z_r = packing height in the rectifying section

H_{tG} = height of transfer unit: can be calculated if $k_y'a$ is known

N_{tG} = no. of transfer unit: can be obtained by numerical or graphical evaluation of the integral. Interfacial vapor conc., yi, for a set of values of y ($y_f < y < y_2$) are required for the numerical integration.

If we take the vapor phase equation to rearrange and integral over the rectifying section from 0 to z r, we considered the height of the rectifying section is z r, then the z r if we integrate would be G by k dash y a integral y f to y 2 d y divided by y i minus y. So, this is known as NTG integral term, and this term HTG the first term is height of transfer unit, and this can be calculated if k y dash a is known. And NTG can be calculated by graphical integration, if we know the interfacial concentration y i, for a set of values of y, we can get interfacial concentration y I, and then we can do the numerical integration.

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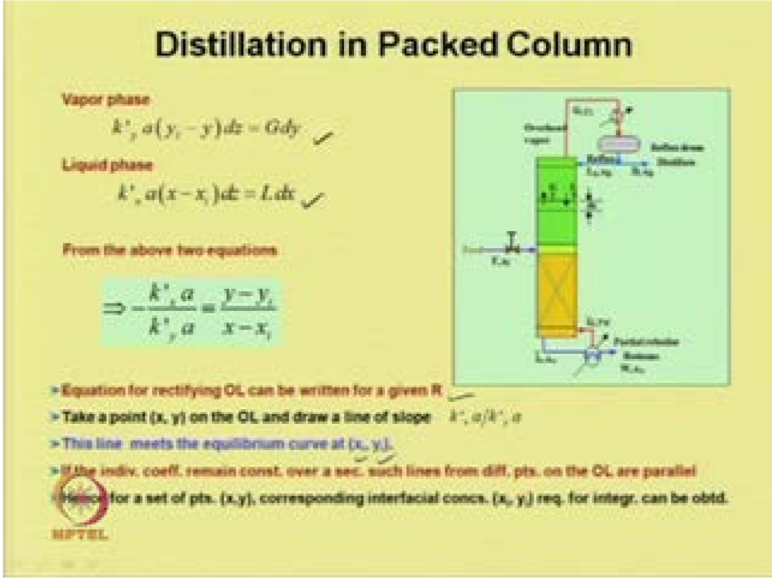

Distillation in Packed Column

Vapor phase

$$k'_y a(y_i - y)dz = Gdy$$

Liquid phase

$$k'_x a(x - x_i)dz = Ldx$$

From the above two equations

$$\Rightarrow -\frac{k'_x a}{k'_y a} = \frac{y - y_i}{x - x_i}$$

- Equation for rectifying OL can be written for a given R
- Take a point (x, y) on the OL and draw a line of slope $-k'_x a/k'_y a$
- This line meets the equilibrium curve at (x_i, y_i).
- If the indiv. coeff. remain const. over a sec. such lines from diff. pts. on the OL are parallel
- Hence for a set of pts. (x,y), corresponding interfacial concs. (x_i, y_i) req. for integr. can be obtd.

Now, how to get this interfacial concentration for a set of values of y, if we considered the vapor phase and the liquid phase this two relation, and if we divide this two equitation the ratio of mass transfer co efficient. We know the equation of operating line for a rectifying section for a given R, and we can take a point x, y on the operating and line draw a line of the slope of a x dash a divided by k y dash a.

So, if we draw this line will meet in the equilibrium line at point x i and y i, and if the individual co-efficient remain constant over that section this lines from different points, on the operating line there will be parallel line. So, for a set of points x, y correspondence interfacial concentration x i and y i can be obtain, which can be used for the integration.

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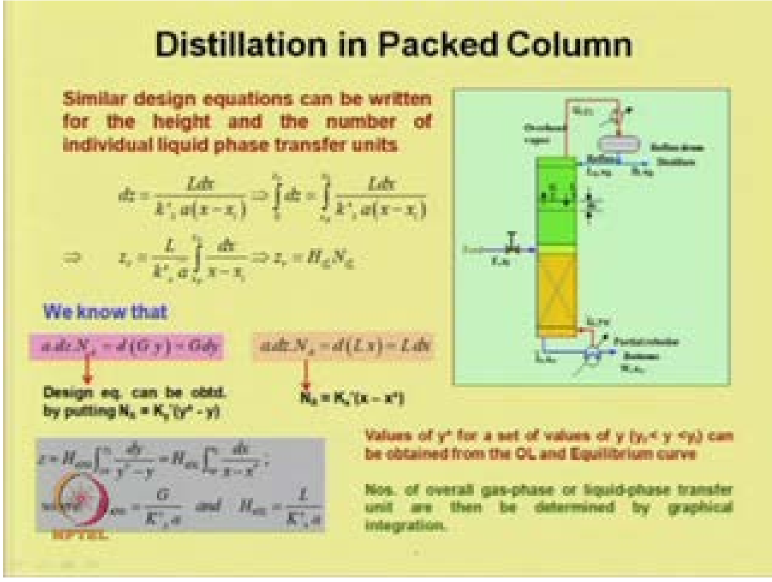

Distillation in Packed Column

Similar design equations can be written for the height and the number of individual liquid phase transfer units

$$dz = \frac{L\,dx}{k'_x a(x-x_i)} \Rightarrow \int_0^{z_r} dz = \int_{x_b}^{x_t} \frac{L\,dx}{k'_x a(x-x_i)}$$

$$\Rightarrow \quad z_r = \frac{L}{k'_x a}\int_{x_b}^{x_t}\frac{dx}{x-x_i} \Rightarrow z_r = H_L N_L$$

We know that

$a\,dz\,N_A = d(G\,y) = G\,dy$ → Design eq. can be obtd. by putting $N_A = K_y'(y^* - y)$

$a\,dz\,N_A = d(L\,x) = L\,dx$ → $N_A = K_x'(x - x^*)$

$$z = H_{tOG}\int_{y_b}^{y_t}\frac{dy}{y^*-y} = H_{tOL}\int_{x_b}^{x_t}\frac{dx}{x-x^*};$$

$$H_{tOG} = \frac{G}{K'_y a} \quad \text{and} \quad H_{tOL} = \frac{L}{K'_x a}$$

Values of y^* for a set of values of y ($y_b < y < y_t$) can be obtained from the OL and Equilibrium curve

Nos. of overall gas-phase or liquid-phase transfer unit are then be determined by graphical integration.

Similar design equations can be written for height and number of transfer units in the liquid phase, where we can calculate z r would be H T L into N T L height of transfer unit in the liquid phase, and number of transfer unit in the liquid phase. Now, we know that from the differential mass balance a d z N a is equal to G into d y, so if we write in terms of the overall mass transfer co-efficient, over all gas phase mass transfer co-efficient k y dash into y star minus y, y star is the equilibrium concentration.

Then we can write Z would be equal to HTOG, HTOG is nothing but G y capital K Y dash a gas flow rate divided by overall gas phase mass transfer co efficient. Similarly, Z can be obtaining based on the overall mass transfer co-efficient in the liquid phase, which is given over here. The values of y star for a set of y can be obtained from operating line and equilibrium line, as we discussed before; now numbers of overall gas phase or liquid phase transfer unit are then be determine by graphical integration.

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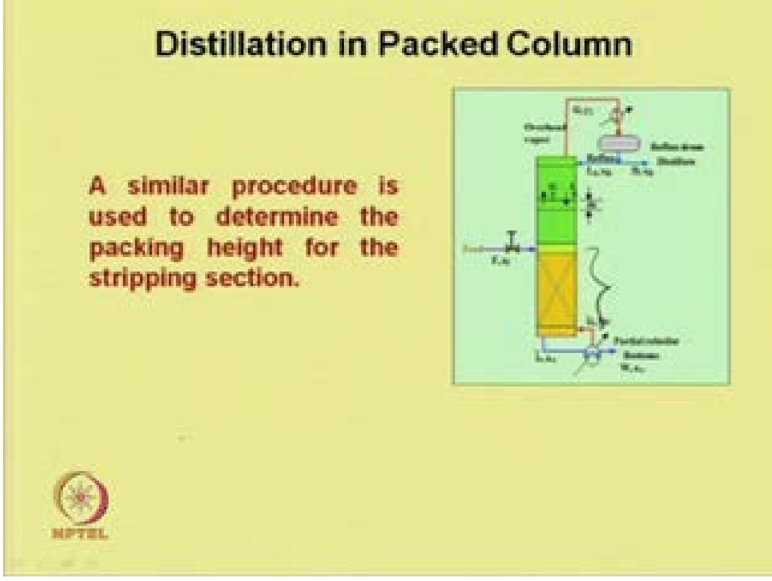

Distillation in Packed Column

A similar procedure is used to determine the packing height for the stripping section.

A similar procedure is used to determine the packing height for the stripping section. So, far we have discussed the design procedure, how to obtain the packing height in the rectifying section, similar procedure should be adopted to obtain the height of packing in the stripping section. So, that the total height required for a given separation can be obtained.

Thank you.