Mass Transfer Operations I

Prof. BishnupadaMandal
Department of Chemical Engineering
Indian Institute of Technology, Guwahati

Module -5
Distillation
Lecture - 8
Fractional Distillation: Subcooled Reflux, Tray Efficiency and Use of Open Steam

Welcome to the eighth lecture of module 5. We are discussing distillation; in the last lecture we have covered total reflux, another is minimum reflux. At total reflux we use to get the minimum number of trays or stages required to obtain a given separation. The last class, we have discussed how to obtain the minimum reflux using McCabe–Thiele method.

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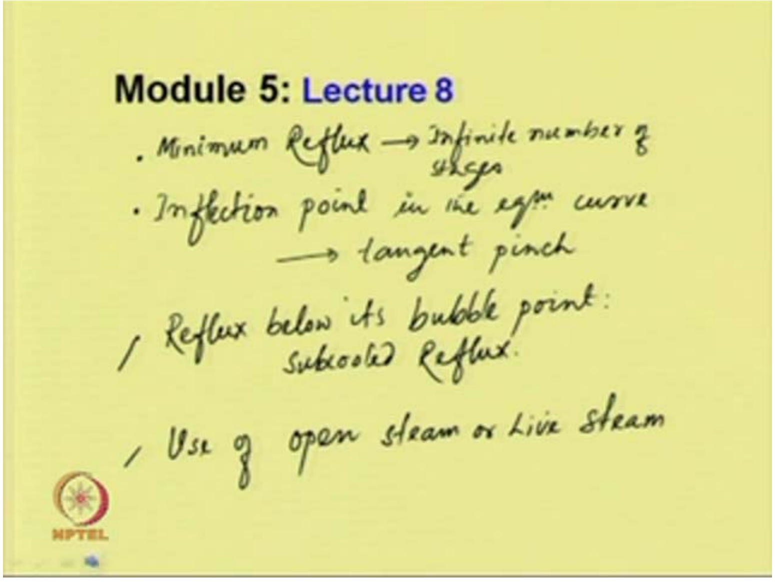

Module 5: Lecture 8

- Minimum Reflux → Infinite number of stages
- Inflection point in the eqm curve → tangent pinch

/ Reflux below its bubble point: Subcooled Reflux.

/ Use of open steam or Live steam

So, we have discussed in the last class minimum reflux, infinite number of stages. We have seen how to calculate using Fenske equation. So, we have also seen that if inflection point in the equilibrium curve, and we generally use tangent pinch point or tangent pinch.

So today, we will discuss another important thing, when we say reflux Whether it is saturated reflux added bubble points or reflux below its bubble points, reflux below its bubble point or we can say subcooled reflux. This is one thing we will discuss today, another topic which we will discuss is that, the use of live steam, use of open steam or live steam. Let us start with reflux below its bubble point or subcooled reflux. Sometimes the vapor which is coming out at the top of the tower is condensed and then cooled below its bubble point and when subcooled reflux returned to the tower, at the top of the section the subcooled reflux immediately reaches its bubble point and for that it takes the energy from the upcoming vapor.

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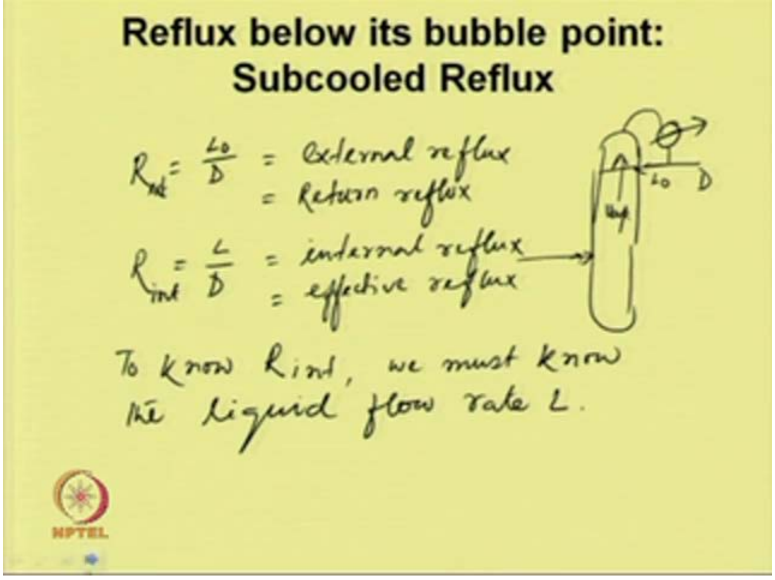

Reflux below its bubble point:
Subcooled Reflux

$R_{ext} = \frac{L_0}{D}$ = External reflux
= Return reflux

$R_{int} = \frac{L}{D}$ = internal reflux
= effective reflux

To know R_{int}, we must know the liquid flow rate L.

If we see this is the feed and this is the top section which is cooled and then returned the reflux, this is L naught and this is D. So the reflux ratio R is equal to L naught by D, hence as the liquid which is coming out will not over here, it will take the heat from the vapor which is upcoming and then the vapor gets condensed and mixed with these L naught and flows down. So the liquid flows in these sections become more and vapor flow become reduced. It is very important to find out the reflux Which is coming out over here is known as external reflux, this is known as external reflux or we an say is returned reflux. So it is called external e x g and we can say also the returned reflux, reflux and the liquid which is coming we should be R inside the tower it should be L by D. So this is known as internal reflux ratio, internal reflux or we can say effective reflux.

So we can name it as internal. In order to know R internal, we must know the liquid flow rate L, we will discuss now how to calculate this L.

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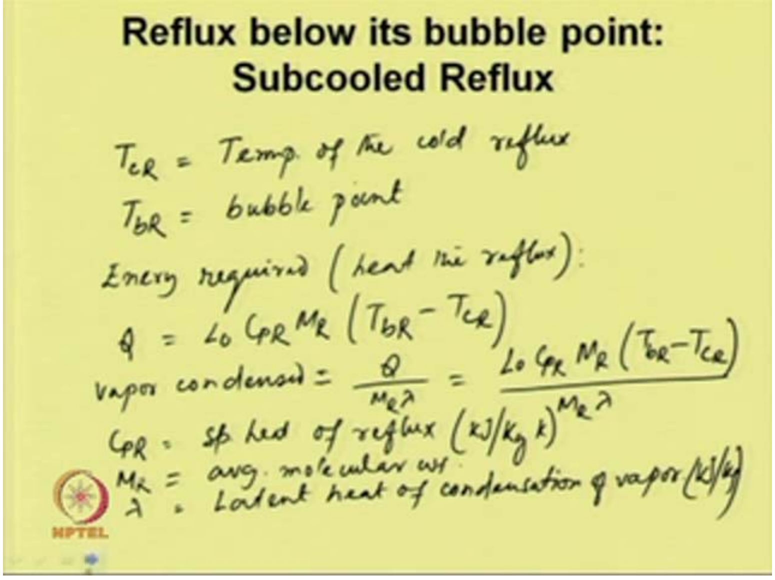

Reflux below its bubble point: Subcooled Reflux

T_{cR} = Temp. of the cold reflux

T_{bR} = bubble point

Enery required (heat the reflux):

$$Q = L_0 C_{pR} M_R (T_{bR} - T_{cR})$$

$$\text{vapor condensed} = \frac{Q}{M_R \lambda} = \frac{L_0 C_{pR} M_R (T_{bR} - T_{cR})}{M_R \lambda}$$

C_{pR} = sp hed of reflux (kJ/kg K)

M_R = avg. molecular wt.

λ = Latent heat of condensation of vapor (kJ/kg)

Let us take TCR is the temperature of the cold reflux and TBR is the bubble point. Then the energy required to heat there flux, energy required heat the reflux is equal to we can write Q is equal to L naught is the flow rate CPR, M R, TBR minus TCR and vapor condensed, condensed which is equal to Q by MR lambda, which is equal to L naught CPR, M R, TBR minus TCR divided by M R lambda. CPR this is specific heat of reflux in unit of Kilo Joule per K G Kelvin. M R is the average molecular weight, lambda, Latin heat of condensation of vapor, heat of condensation of vapor. So which is Kilo Joule per Kg, then we can calculate the internal liquid flow rate.

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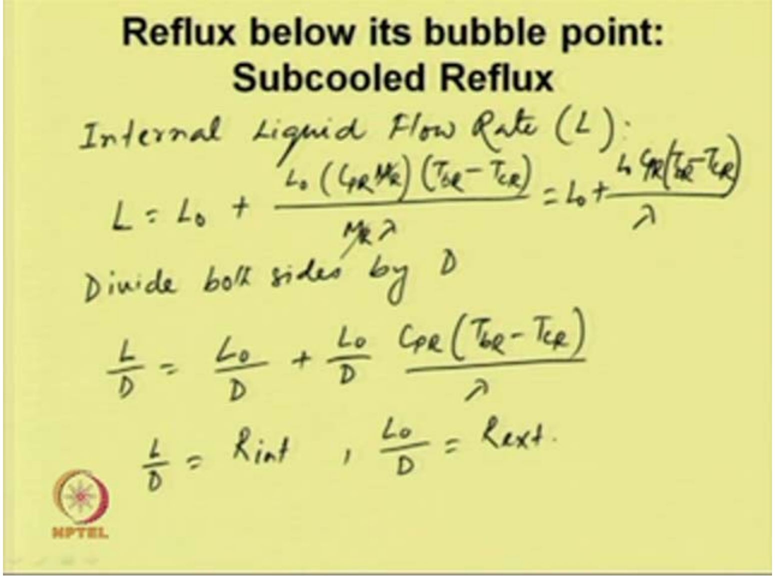

Reflux below its bubble point: Subcooled Reflux

Internal Liquid Flow Rate (L):

$$L = L_0 + \frac{L_0 (C_{pR} M_R)(T_{bR} - T_{cR})}{M_R \lambda} = L_0 + \frac{L_0 C_{pR}(T_{bR} - T_{cR})}{\lambda}$$

Divide both sides by D

$$\frac{L}{D} = \frac{L_0}{D} + \frac{L_0}{D} \frac{C_{pR}(T_{bR} - T_{cR})}{\lambda}$$

$$\frac{L}{D} = R_{int} \quad , \quad \frac{L_0}{D} = R_{ext}$$

So, the internal liquid flow rate L which is L is equal to L naught plus L naught CPR, M R into TBR minus TCR divided by M R lambda. So this M R, M R will cancel out. So we can write L naught plus L naught CPR, TBR minus TCR divided lambda. So if we divide both sides by D, so we will have L by D would be equal to L naught by D plus L naught by D, CPR, TBR minus TCR divided by lambda. As we know L by D is equal to R internal and L naught by D is equal to R external. So we can write this equations R internal would be R external in plus R external in to this term.

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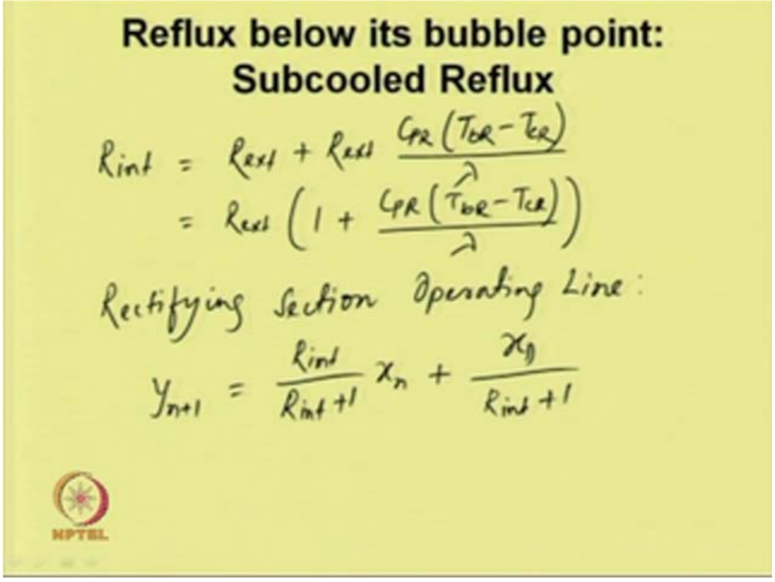

Reflux below its bubble point:
Subcooled Reflux

$$R_{int} = R_{ext} + R_{ext} \frac{C_{PR}(T_{bR} - T_{cR})}{\lambda}$$

$$= R_{ext} \left(1 + \frac{C_{PR}(T_{bR} - T_{cR})}{\lambda}\right)$$

Rectifying Section Operating Line:

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

So we can write R internal would be equal to R external plus R external into CPR into TBR minus TCR divided by lambda. So, this we can write R external into 1 plus CPR, TBR minus TCR divided by lambda this term. So the rectifying sections operating line Y n plus 1 would be equal to R internal divided by R internal plus 1 x n plus x D divided by R internal plus 1.

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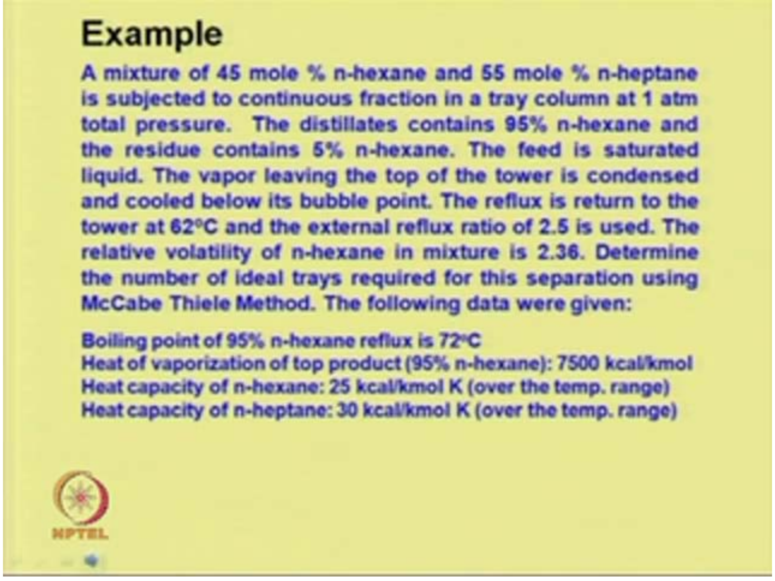

Example

A mixture of 45 mole % n-hexane and 55 mole % n-heptane is subjected to continuous fraction in a tray column at 1 atm total pressure. The distillates contains 95% n-hexane and the residue contains 5% n-hexane. The feed is saturated liquid. The vapor leaving the top of the tower is condensed and cooled below its bubble point. The reflux is return to the tower at 62°C and the external reflux ratio of 2.5 is used. The relative volatility of n-hexane in mixture is 2.36. Determine the number of ideal trays required for this separation using McCabe Thiele Method. The following data were given:

Boiling point of 95% n-hexane reflux is 72°C
Heat of vaporization of top product (95% n-hexane): 7500 kcal/kmol
Heat capacity of n-hexane: 25 kcal/kmol K (over the temp. range)
Heat capacity of n-heptane: 30 kcal/kmol K (over the temp. range)

Now, let us take an example, how to calculate the internal reflux ratio and then the number of plates required for a given separation. A mixture of 45 mole percent n-hexane and 55 mole percent n-heptane is subjected to continuous fractionation in a tray column at 1 atmosphere total pressure. The distillates contains 95 percent n-hexane and the residue contains 5 percent n-hexane. The feed is saturated liquid, the vapor leaving the top of the tower each is condensed and cooled bellow its bubble point.

The reflux is return to the tower at 62 degree centigrade and the reflux ratio of 2.5 is used. The relative volatility of n-hexane in the mixture is 2.36. Determine the number of ideal trays required for this separation using McCabe–Thiele method. The following data were given boiling point of 95 percent n-hexane reflux is 72 degree centigrade; heat of vaporization of the top product of 95 percent n-hexane, which is 7,500 kilo calorie per kilo mole. Heat capacity of n-hexane is 25 kilo calorie per kilo mole Kelvin and heat capacity of n-heptane is 30 kilo calorie per kilo mole Kelvin

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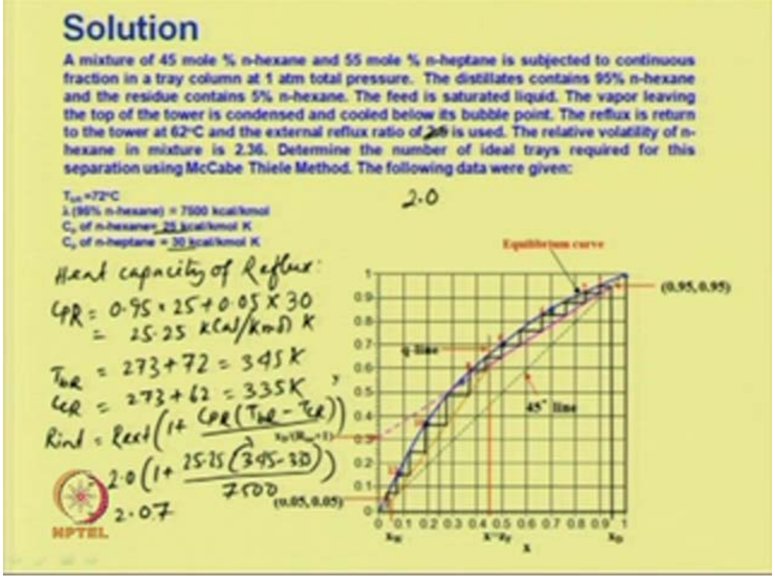

Solution

A mixture of 45 mole % n-hexane and 55 mole % n-heptane is subjected to continuous fraction in a tray column at 1 atm total pressure. The distillates contains 95% n-hexane and the residue contains 5% n-hexane. The feed is saturated liquid. The vapor leaving the top of the tower is condensed and cooled below its bubble point. The reflux is return to the tower at 62°C and the external reflux ratio of ~~2.5~~ is used. The relative volatility of n-hexane in mixture is 2.36. Determine the number of ideal trays required for this separation using McCabe Thiele Method. The following data were given:

2.0

T_{bR} = 72°C

λ (95% n-hexane) = 7500 kcal/kmol

C_p of n-hexane = 25 kcal/kmol K

C_p of n-heptane = 30 kcal/kmol K

Heat capacity of Reflux:

$$C_{PR} = 0.95 \times 25 + 0.05 \times 30$$
$$= 25.25 \text{ kCal/kmol K}$$
$$T_{bR} = 273 + 72 = 345 \text{ K},$$
$$T_{cR} = 273 + 62 = 335 \text{ K}$$
$$R_{int} = R_{ext}\left(1 + \frac{C_{PR}(T_{bR} - T_{cR})}{\lambda}\right)$$
$$2.0\left(1 + \frac{25.25(345 - 335)}{7500}\right)$$
$$2.07$$

So we can calculate the heat capacity of reflux, heat capacity of reflux from the data given. CPR is equal to 95 percent of n-hexane into 25 which is given over hear, plus 0.05 into 30, which is over hear, so which is equal to 25.25 kilo calorie per K mole Kelvin. Now TBR is given TBR which is equal to 273 plus 72, which equal to 345 K and TCR is 273 plus 62 is 335 K and we know that R internal would be equal to R external, into 1 plus CPR into TBR minus TCR divided by lambda. So, if we substitute external reflux ratio which is given, so let us consider instead of 2.5 is 2.2 if we use then it would be 2 into 1 plus 25.25 into 345 minus 335 divided by 7,500, so which would be equal to 2.07. So, internal reflux ration is 2.07.

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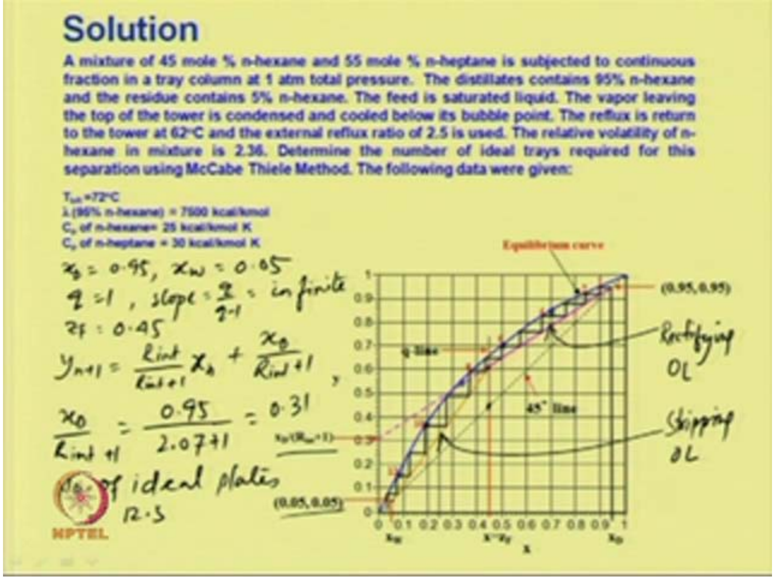

Solution

A mixture of 45 mole % n-hexane and 55 mole % n-heptane is subjected to continuous fraction in a tray column at 1 atm total pressure. The distillates contains 95% n-hexane and the residue contains 5% n-hexane. The feed is saturated liquid. The vapor leaving the top of the tower is condensed and cooled below its bubble point. The reflux is return to the tower at 62°C and the external reflux ratio of 2.5 is used. The relative volatility of n-hexane in mixture is 2.36. Determine the number of ideal trays required for this separation using McCabe Thiele Method. The following data were given:

T_{bt} =72°C
λ (95% n-hexane) = 7500 kcal/kmol
C_p of n-hexane= 25 kcal/kmol K
C_p of n-heptane = 30 kcal/kmol K

$x_D = 0.95$, $x_W = 0.05$

$q = 1$, slope $= \frac{q}{q-1}$ = infinite

$z_F = 0.45$

$$y_{n+1} = \frac{R_{int}}{R_{int}+1} x_n + \frac{x_D}{R_{int}+1},$$

$$\frac{x_D}{R_{int}+1} = \frac{0.95}{2.07+1} = 0.31$$

No. of ideal plates
12.5

Now, the data which are given x D is 0.95 x W is 0.05, q is equal to 1 saturated liquid. So slope is equal to q by q minus 1 is equal to infinite. So this is the point, feed point and this is the q line which is vertical line at the feed point of 0.45 because z F is 0.45. So, we can calculate y n plus 1 would be R internal divided by R internal plus 1 x n plus x D by R internal plus 1. So, intersect x D by R internal plus 1 would be equal to 0.95 by 2.07 plus 1, which would be equal to 0.31.

So at 0.31 which is x D by R internal plus 1, so this is the value. So we know the intersect values and we know the x D point over hear. So, we can plot the rectifying section operating line, so this is the rectifying section operating line, rectifying operating line which is extended to the intersect and then we know x W, x W point and we know the feed line. So intersection point we know, so we plot between these two point, stripping section operating line, this is the stripping section operating line and then if we follow constructions of the stages ,we see that the number of ideal plates, number of ideal plates are 12.5.

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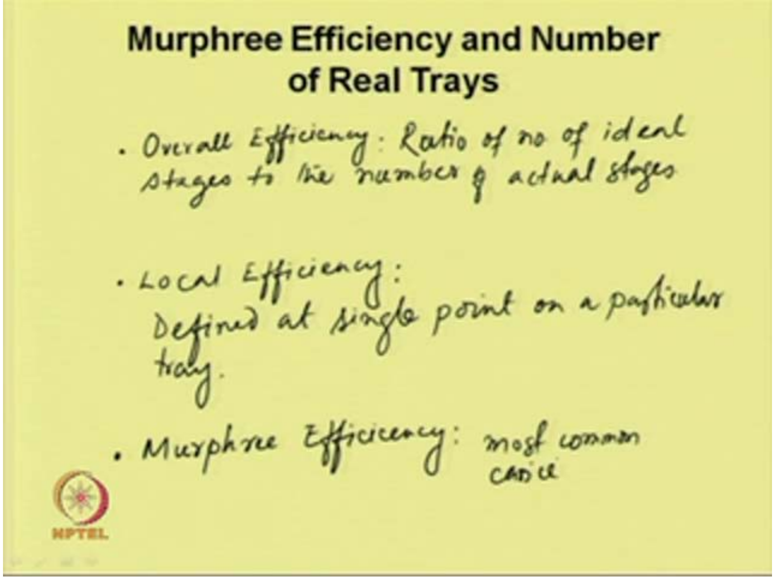

Murphree Efficiency and Number of Real Trays

- Overall Efficiency: Ratio of no of ideal stages to the number of actual stages
- Local Efficiency: Defined at single point on a particular tray.
- Murphree Efficiency: most common choice

Now, we will discuss stage efficiency, particularly would be more interested to Murphree efficiency and number of real trays. What we do in mass transfer operations for the calculations of the number of ideal stages required for a given separation, we assume that the each stage is in equilibrium, the vapor and liquid living the trays are in equilibrium, but in reality they are not true. So the ideal stage assumptions or equilibrium assumptions is an approximation. So there is a deviation about the efficiency of the ideal trays we assumed in reality. To calculate these efficiency of the trays there are three types of efficiency which we generally use, one is known as overall efficiency, another term which we use is the local efficiency and third which we use is the Murphree efficiency.

So, overall efficiency is use the simplistic approach and it is the ratio ratio of number of ideal stages to the number of actual stages. So this is the single efficiency which generally not valid because the efficiency of the stage varies from trays to trays and it depends on the deferent factors like the geometry and design of trays and then flow rates and paths of all streams compositions, this varies from trays to trays. This is the simplistic approach to use the overall efficiency. Another type of efficiency which is the most accurate, but very difficult to apply is the local efficiency. Defined at single point on a particular tray, use of this efficiency is very difficult the Murphree efficiency which is the common choice, most common choice.

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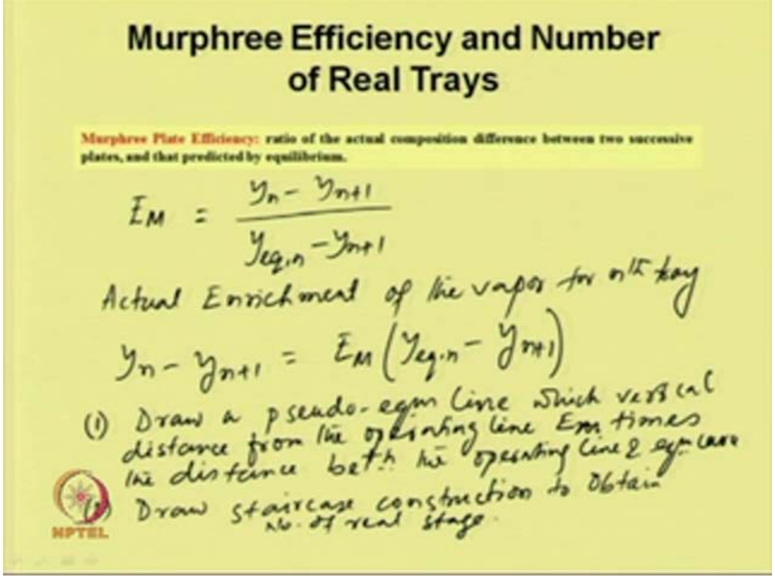

Murphree Efficiency and Number of Real Trays

Murphree Plate Efficiency: ratio of the actual composition difference between two successive plates, and that predicted by equilibrium.

$$E_M = \frac{y_n - y_{n+1}}{y_{eq,n} - y_{n+1}}$$

Actual Enrichment of the vapor for nth tray

$$y_n - y_{n+1} = E_M \left(y_{eq,n} - y_{n+1} \right)$$

(1) Draw a pseudo-eqm line which vertical distance from the operating line Em times the distance betn the operating line & eqm curve

(2) Draw staircase construction to obtain no. of real stage.

Now, we will discuss Murphree plate efficiency. It is the ratio of the actual composition difference between two successive plates and that predicted by equilibrium, which we can write EM is equal to y n minus y n plus 1, these are the two successive stages composition divided by y equilibrium of n minus y n plus one. So this is most common practice to use and the actual enrichment, enrichment of the vapor for nth tray. We can write is y n minus y n plus 1 would be equal to EM times y equilibrium minus y n plus 1. To find out the number of real trays required for a particular operations, what we have to do using this efficiency we have to draw a pseudo equilibrium line, draw a pseudo equilibrium line which is vertical, which is vertical distance vertical distance from the operating line, EM times the distance between, the distance between the operating line line and equilibrium curve, equilibrium curve.

We have to draw another pseudo equilibrium line and the vertical distance between the operating line and the equilibrium line should be EM times and then we have to do the staircase construction, as we did before. Draw staircase construction to obtain number of real stages. So this staircase construction should be between the pseudo equilibrium line and the operating line.

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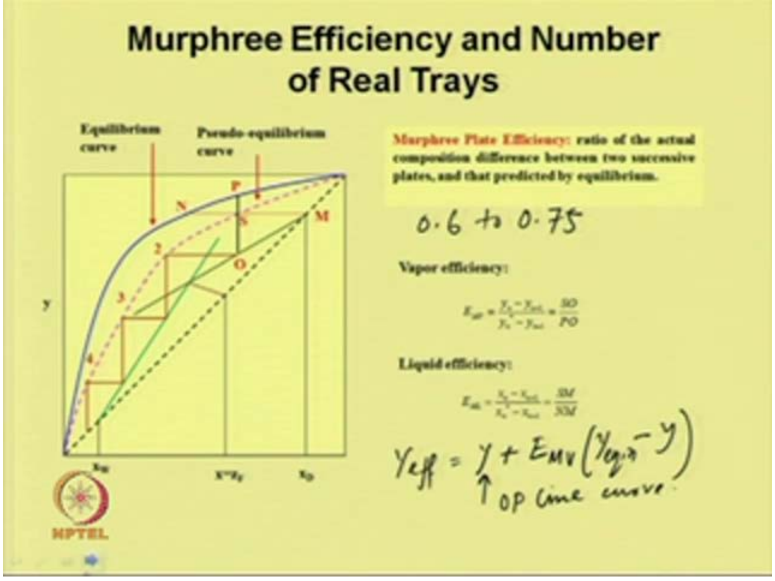

Murphree Efficiency and Number of Real Trays

Murphree Plate Efficiency: ratio of the actual composition difference between two successive plates, and that predicted by equilibrium.

0.6 to 0.75

Vapor efficiency:

$$E_{MV} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}} = \frac{SO}{PO}$$

Liquid efficiency:

$$E_{ML} = \frac{x_n - x_{n+1}}{x_n^* - x_{n+1}} = \frac{SM}{NM}$$

$$y_{eff} = \underset{\uparrow}{y} + E_{MV}(y_{eq,n} - y)$$

op line curve

As we said the Murphree efficiency we can define as the ratio of the actual composition difference between the two successive plates and that predicted by equilibrium. So there are two types of Murphree efficiency we can define, one is in terms of the vapor, another in terms of the liquid.

So, EM g which we can define y n minus y n plus 1 divided by y n star which is equilibrium minus y n plus 1 and this we can see the ratio between the S O and P O. Ratio between these vertical so this is the operating line and this is the actual equilibrium line, this one the equilibrium line and this the ratio of S O by P O this is the vapor efficiency. Similarly, in terms of the liquid we can write EM L is x n minus x n plus 1 divided by x n star minus x n plus 1 which is the ratio between the S M divided by N M. So, this is the ratio, so this is the liquid efficiency and this is the vapor efficiency. In shift trays generally we use this efficiency between 0.6 to 0.75 and it is very easy to calculate the effective equilibrium curve. We can write y effective would be equal to y plus E MV multiplied by y equilibrium of n minus y.

So, this y is nothing but the operating line curve. So then we can calculate the y effective and we can draw the pseudo equilibrium curve which shown over here in dotted lines. So this is the pseudo equilibrium line.

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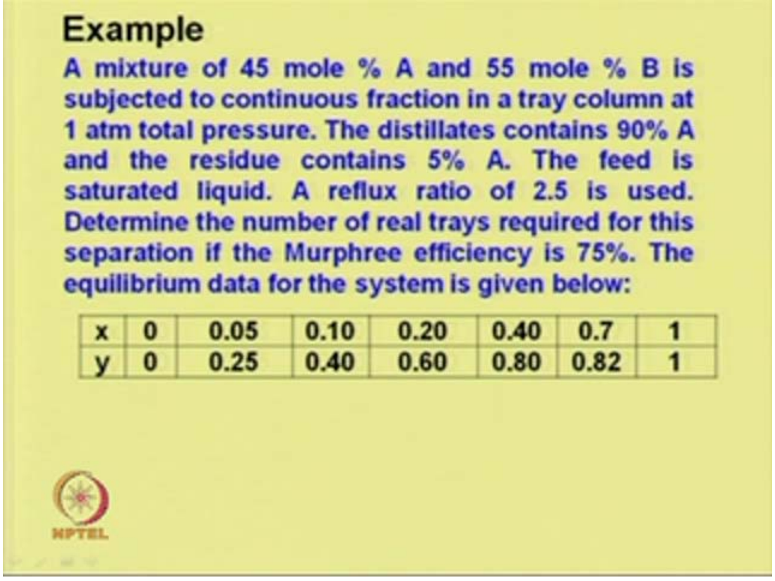

Example

A mixture of 45 mole % A and 55 mole % B is subjected to continuous fraction in a tray column at 1 atm total pressure. The distillates contains 90% A and the residue contains 5% A. The feed is saturated liquid. A reflux ratio of 2.5 is used. Determine the number of real trays required for this separation if the Murphree efficiency is 75%. The equilibrium data for the system is given below:

x	0	0.05	0.10	0.20	0.40	0.7	1
y	0	0.25	0.40	0.60	0.80	0.82	1

Now let us take an example, to see how to calculate the number of stages using Murphree efficiency. A mixture of 45 mole percent A and 55 mole percent B is subject to continuous fractionation in a tray column at 1 atmosphere total pressure. The distillates contains 90 percent A and the residue contains 5 percent A. The feed is saturated liquid, a reflux ratio of 2.5 is used determine the number of real trays required for this separation, if the Murphree efficiency 75 percent. The equilibrium data for the system is given bellow.

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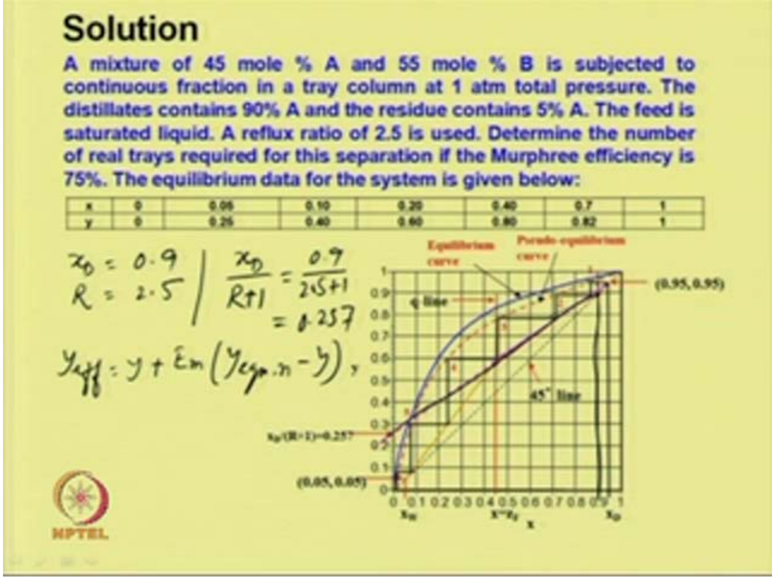

Solution

A mixture of 45 mole % A and 55 mole % B is subjected to continuous fraction in a tray column at 1 atm total pressure. The distillates contains 90% A and the residue contains 5% A. The feed is saturated liquid. A reflux ratio of 2.5 is used. Determine the number of real trays required for this separation if the Murphree efficiency is 75%. The equilibrium data for the system is given below:

x	0	0.05	0.10	0.20	0.40	0.7	1
y	0	0.25	0.40	0.60	0.80	0.82	1

$x_D = 0.9$

$R = 2.5$

$\frac{x_D}{R+1} = \frac{0.9}{2.5+1} = 0.257$

$y_{eff} = y + E_m (y_{eqm,n} - y),$

So these are the equilibrium data which is the given. Using this equilibrium data which are given we can plot the equilibrium curve and then the data which are given x D is equal to 0.9 and R is given 2.5. So we can calculate x D by R plus 1 which is equal to 0.9 divided by 2.5 plus 1 which is equal to 0.257. We know the point x D, x D, we know the point at 0.9 and 0.9 and then we can plot the operating line through this. So this is the point we can calculate y effective using the equation given earlier y plus EM into y equilibrium n minus y. So using this we can obtained y effective and with the y effective this is the plot which is drawn over here, this the pseudo equilibrium curve.

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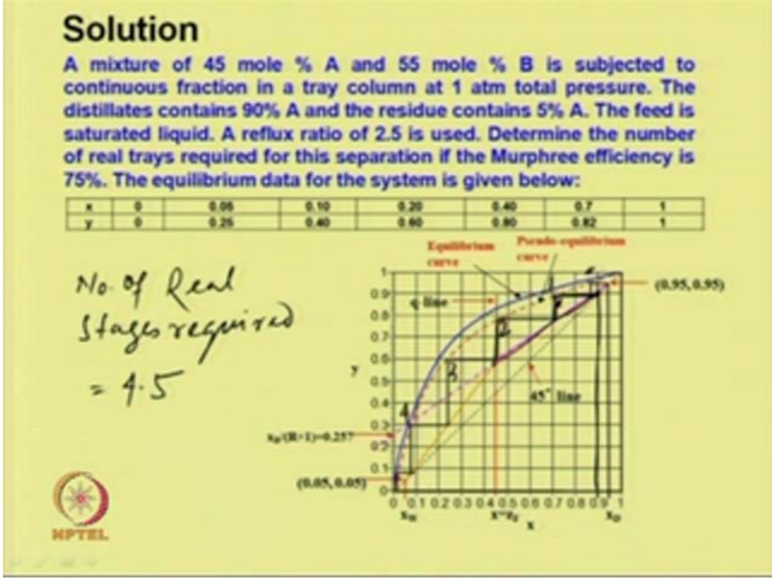

Solution

A mixture of 45 mole % A and 55 mole % B is subjected to continuous fraction in a tray column at 1 atm total pressure. The distillates contains 90% A and the residue contains 5% A. The feed is saturated liquid. A reflux ratio of 2.5 is used. Determine the number of real trays required for this separation if the Murphree efficiency is 75%. The equilibrium data for the system is given below:

x	0	0.05	0.10	0.20	0.40	0.7	1
y	0	0.25	0.40	0.60	0.80	0.82	1

No. of Real Stages required = 4.5

We plot from this line and if we plot the staircase construction from this point, so the number of stages would be one, two, three and this is four. Total number of real stages required is equal to 4.5. The staircase construction is drawn between the operating line and the pseudo equilibrium line which is drawn using the Murphree efficiency.

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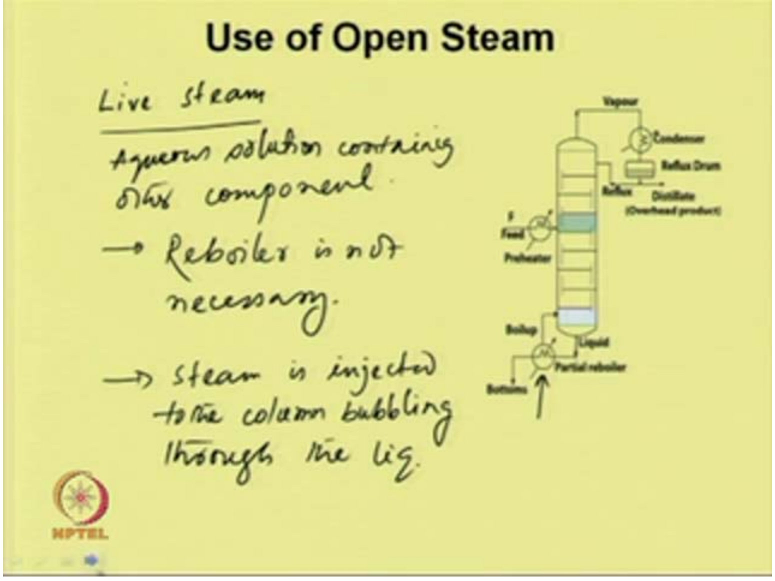

Use of Open Steam

Live steam

Aqueous solution containing other component.

→ Reboiler is not necessary.

→ steam is injected to the column bubbling through the liq.

So, now let us considered another important topic, which is use of open steam. In general we supply the heat using the partial reboiler, heat is supply to this and by taking the liquid which is coming out from the tower, that is heated with the reboiler, where steam is not directly injected to the tower. Here particularly for a solution, which contains water or some other components then this aqua solutions may be convenient to separate using live steam. Live steam is used to separate aqueous solutions aqueous solutions containing other components. So, in this case the reboiler which is used in general is not necessary. Reboiler is not necessary; the steam, which is injected at the bottom directly to the column bubbling through the liquid, bubbling through the liquid.

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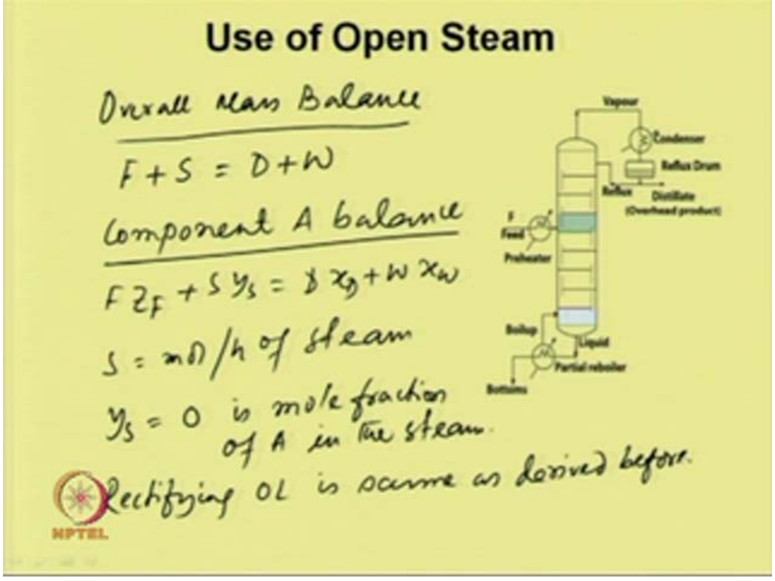

Use of Open Steam

Overall Mass Balance

$F + S = D + W$

Component A balance

$F z_F + S y_S = D x_D + W x_W$

S = mol/h of steam

$y_S = 0$ is mole fraction of A in the steam.

Rectifying OL is same as derived before.

Now, if we do the overall mass balance, what the tower using live steam. Overall mass balance, here F is the feed plus the steam S would be equal to D plus W. So this is the overall mass balance. Now component balance, component A balance, F zF plus S yS is equal to D xD plus W xW, where s is the mole of steam, mole per hour of steam injected and y S would be 0 is the mole fraction, fraction of A in the steam. So, the operating line in the rectifying section above the feed tray would be same as we derived earlier. Rectifying OL, rectifying operating line is same as derived before.

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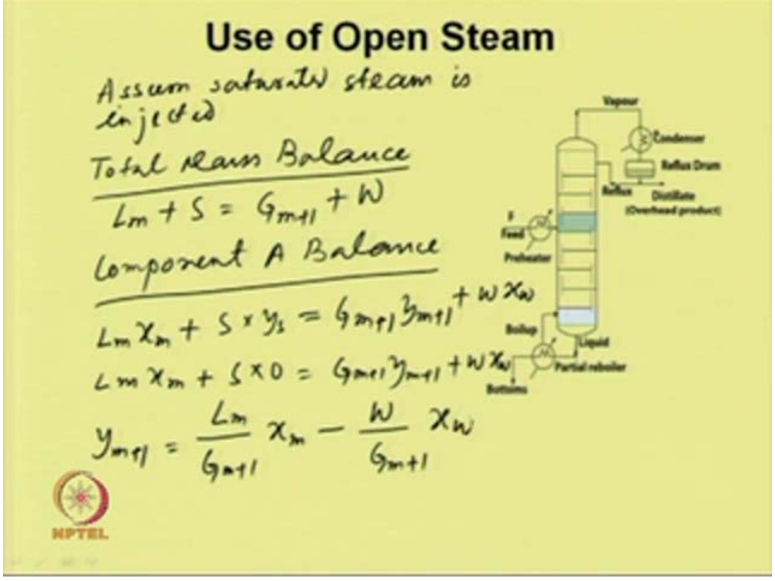

Use of Open Steam

Assume saturated steam is injected

Total Mass Balance

$L_m + S = G_{m+1} + W$

Component A Balance

$L_m x_m + S \times y_S = G_{m+1} y_{m+1} + W x_W$

$L_m x_m + S \times 0 = G_{m+1} y_{m+1} + W x_W$

$$y_{m+1} = \frac{L_m}{G_{m+1}} x_m - \frac{W}{G_{m+1}} x_W$$

Now, we will do the total mass balance and material balance over the stripping section. We assume saturated steam is injected, then the total mass balance L m plus S is equal to G m plus 1 plus W and component balance, A balance if we do that we can write, L m X m plus S into Y S is equal to G m plus 1, y m plus 1 plus W XW. Now, if we solve it Lm X m plus S into 0 would be equal to Gm plus 1, y m plus 1 plus W XW. It will give Y m plus 1 would be L m by G m plus 1 X m minus W by G m plus 1 X W.

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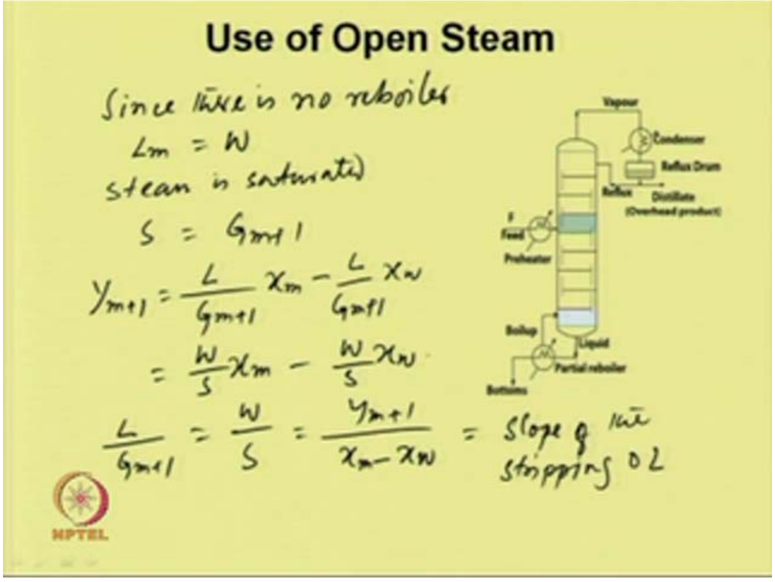

Use of Open Steam

Since there is no reboiler

$L_m = W$

steam is saturated

$S = G_{m+1}$

$$y_{m+1} = \frac{L}{G_{m+1}} x_m - \frac{L}{G_{m+1}} x_w$$

$$= \frac{W}{S} x_m - \frac{W}{S} x_w$$

$$\frac{L}{G_{m+1}} = \frac{W}{S} = \frac{y_{m+1}}{x_m - x_w}$$ = slope of the stripping OL

So, we can write since there is no reboiler Lm would be W and steam is saturated, steam is saturated. So we can write S equal to G m plus 1. So therefore, the operating line for stripping section would be Y m plus 1 would be equal to L by G m plus 1, X m minus L by G m plus 1, XW or we can write W by S, X m minus W by S, XW. And then slope of the operating line is L by G m plus 1 or W by S is equal to Y m plus 1 divided by X m plus XW. So this is the slope of the operating line, stripping O L. In the next lecture we will discuss how to calculate the steam requirement and number of stages required for a given separation.

Thank you.