**Mass Transfer Operations-I**
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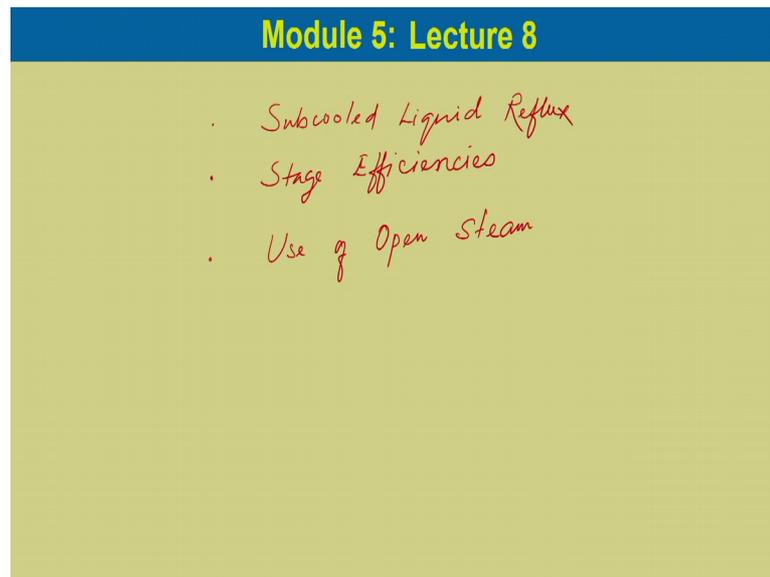
Lecture - 33
Distillation

Welcome to the 8th lecture of module 5 of Mass Transfer Operation. In this module we are discussing Distillation operation, before going to this lecture let us have brief recap on our previous lecture. On our previous lecture, we have mostly discussed few important things.
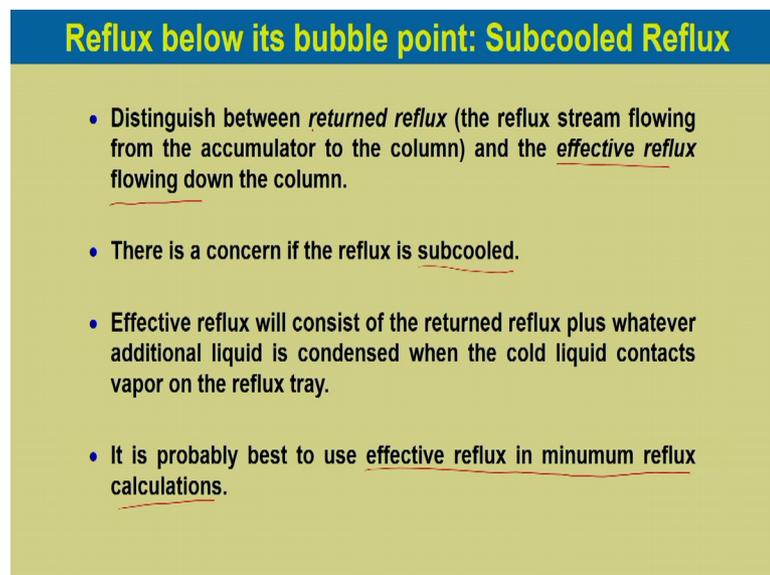
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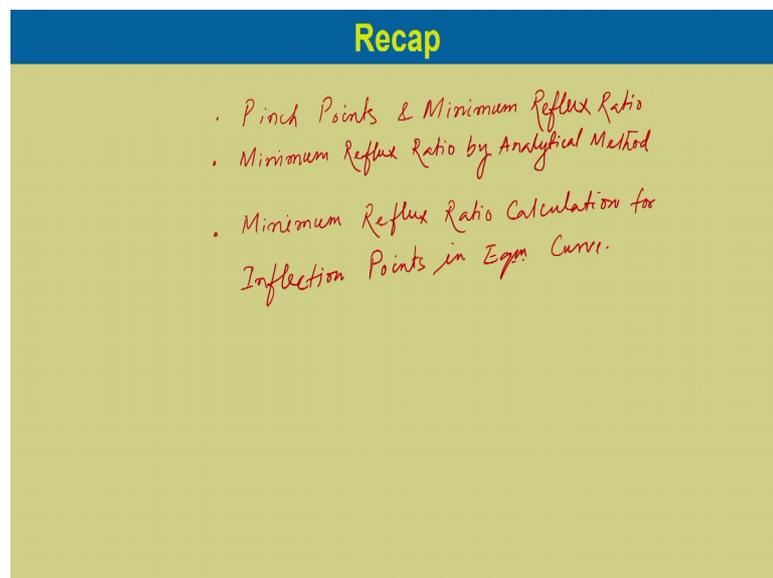
One is pinch points and concept of minimum reflux ratio. The second thing we have discussed here is how to calculate minimum reflux ratio by analytical method and the third thing which we have discussed is the minimum reflux ratio calculation for the inflection points in equilibrium curve. So, these are the three important topics which we have discussed in our last lecture.

And in this lecture we will mainly focus on three important things, one is subcooled liquid reflux and second thing different stage efficiencies and third thing we will considered use of open steam in distillation. So, let us start with the subcooled liquid reflux.

That is reflux below its bubble point and we call it subcooled reflux. So, distinguish between the returned reflux, the reflux stream flowing from the accumulator to the column and the effective reflux flowing down the column. So, if you just think about the

distillation column vapour which is coming from the top tray named as tray 1 and it goes to the condenser. So, it condense the vapour into the liquid form.

So, if it is a total condenser it will cool all the or condense all the vapour to liquid and then if it is cooled further below its bubble point so that will be called as subcooled liquid. So, if that is returned from the reflux drum to the first tray, at top tray as a returned reflux which is liquid flow say L naught and d is the product which is taken out. So, L naught by d is the liquid reflux which is given from to the top tray. Now, that is the returned reflux and the another one is the effective reflux flowing down. What is that? So, there is a concern if the reflex is subcooled. So, if the reflux which is coming after the condenser to the reflux drum and then coming to the top tray if that is subcooled then there is a major concern for the design.

So, effective reflux will consist of the returned reflux plus whatever additional liquid is condensed when the cold liquid contacts vapour on the reflux tray. So, on the reflux tray what will happen that is the top tray, the vapour is going out and reflux is coming to that tray so it is subcooled. So, because of the low temperature it will condense some of the vapour which is going out from tray 1, top tray. So, the liquid flow inside the column below the reflux tray will increase. So, it is probably best to use effective reflux in minimum reflux calculation in state of the returned reflux.

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Reflux below its bubble point: Subcooled Reflux

- Vapour from the top tray may sometimes be condensed and then cooled below its bubble point (Subcooled) in the condenser. This is 'cold reflux'.

- On entering the top tray, the 'cold reflux' reaches its bubble point very quickly.

- The required energy is supplied by condensation of some of the vapour on the tray.

Vapour from the top tray may sometimes be condensed and then cooled below its bubble point that is called subcooled in the condenser this is called no cold reflux on entering the top tray the cold reflux reaches its bubble point very quickly.

So, because of this reaching the bubble point very quickly what happens? The energy required is supplied by the condensation of some of the vapour on that tray. So, on that top tray the condensation of the vapour which is going out is supplied because of the energy for this condensation is supplied by the vapour of that tray.

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And as a result there is no vapour flow which is coming from the top tray that is G 1, vapour from the top tray will be less that tray lower than that trays. So, G 1 which is going out that will be less than the G 2 that is the below that tray or the subsequent trays. The liquid flow rate L in the rectifying section will be larger than the reflux L naught that is feed to the column from the condenser. As a result the internal reflux ratio that is R internal is L by D will be greater than the external reflux ratio which is defined R external is equal to L naught by D.

So, this is external reflux ratio and this is called internal reflux ratio. So, since L naught is increasing to L. So, the internal reflux ratio will increase in order to calculate the internal reflux ratio that is R internal we need to know the liquid flow rate L. So, we have to calculate the liquid flow rate.

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Reflux below its bubble point: Subcooled Reflux

T_{cR} = temperature of the cold reflux ; T_{bR} = its bubble point

Energy required for heating the reflux:

$$Q = L_o \, C_{pR} \, M_R \left(T_{bR} - T_{cR} \right)$$

Energy required for vapour condensed:

$$\frac{Q}{M_R \lambda} = \frac{L_o \, C_{pR} \, M_R \left(T_{bR} - T_{cR} \right)}{M_R \lambda}$$

C_{pr} = Specific heat of the reflux (kJ/kg . °C)

M_R = its average molecular weight (this is the same as that of the vapour from the top tray)

λ = latent heat of condensation of the vapour (kJ/kg)

Now, let us consider T cR is the temperature of the cold reflux and T bR is the temperature its bubble point of the cold reflux fluid.

So, energy required for heating the reflux is Q would be equal to L naught C pR M R into the temperature gradient that is no it reaches the bubble point on that tray. So, cold reflux temperature bubble point temperature minus the cold reflux temperature that is the temperature gradient. And so, the energy required for vapour condense we can just divide by M R lambda to the left hand side.

So, it will be L Q by M R lambda would be equal to L naught C pR M R T bR minus T cR divided by M R lambda. Here C pR is the specific heat of the reflux in kilo Joule per kg degree centigrade and M R is its average molecular weight, this is the same as that of the vapour from the top tray and lambda is the latent heat of condensation of the vapour that is kilo joule per kg.

So, this is basically the liquid which will increase, this is the liquid due to the condensation of the vapour this much liquid will increase.

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Reflux below its bubble point: Subcooled Reflux

Then the internal liquid flow rate:

$$I_r = I_o + \frac{L_o\left(C_{pR}M_R\right)\left(T_{bR} - T_{cR}\right)}{M_R\lambda} = I_o + \frac{I_o\,C_{pR}\left(T_{bR} - T_{cR}\right)}{\lambda} \checkmark$$

$$\frac{L}{D} = \frac{I_o}{D} + \frac{I_o}{D}\frac{C_{pR}\left(T_{bR} - T_{cR}\right)}{\lambda} \qquad\qquad R_{int} = \frac{L}{D} \checkmark$$

$$R_{int} = R_{ext} + R_{ext}\frac{C_{pR}\left(T_{bR} - T_{cR}\right)}{\lambda} = R_{ext}\left(1 + \frac{C_{pR}\left(T_{bR} - T_{cR}\right)}{\lambda}\right) \qquad R_{ext} = \frac{I_0}{D} \checkmark$$

Equation of the rectifying operating line

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

So, then the internal liquid flow rate we can calculate L naught, L is equal to L naught plus L naught C pR into M R into T bR minus T cR divided by M R lambda. So, M R M R will cancelled out so it will be L naught plus L naught into C pR into T bR minus T cR by lambda. And again, if we further simplify and divide both the sides with D so it will be L by D would be equal to L naught by D plus L naught by D into C pR into T bR minus T cR by lambda. So, as we know L by D is the internal reflux ratio L by D.

So, and we can write these R internal would be R external because R external is L naught by D. So, R internal would be equal to R external plus R external into C pR into the temperature you know difference divided by lambda. So, we can take R external outside. So, it would be R external into one plus C pR into T bR minus T cR by lambda. So, if this specific heat and the temperature bubble point temperature and the cold reflux temperature and the lambda is known, heat capacity and specific heat these are known then we can calculate the internal reflux ratio.

So, the equation for the rectifying operating line will then change to R y m plus 1 would be equal to R internal by R internal plus 1 into x n plus x D by R internal plus 1.

Now, let us take an example to calculate the number of trays are required by McCabe-Thiele method using the internal reflux ratio. A mixture of 45 mole percent of n-hexane and 55 mole percent of n-heptane is subjected to continuous fractionation in a tray column at 1 atmosphere total pressure. The distillate contents 95 percent n-hexane and the residue contents 5 percent n-hexane. The feed is saturated liquid. The vapour leaving the top of the tray tower is condensed and cooled below its bubble point. The reflux is return to the tower at 62 degree centigrade temperature and the external reflux ratio is 2 is used. The relative volatility of n hexane in mixture is given 2.36.

Now determine the number of ideal trays required for the separation using McCabe-Thiele method and following data are given. So, boiling point of 95 percent n hexane of reflux is 72 degree centigrade. Heat of vaporization of top product that is 95 percent n hexane is 7500 kilo calorie per kilo mole, heat capacity of n hexane is 25 kilo calorie per kilo mole over the temperature range and heat capacity of n-heptane is 30 kilo calorie per kilo mol Kelvin that is over the temperature range.

So, heat capacity of the reflux we can calculate C pR would be equal to 0.95, that is 95 percent into heat capacity of n hexane is given and 0.05 would be the n-heptane that is 0.05 into 30. So, we can get C pR would be 25.25 kilo calorie per kilo mol Kelvin. So, C pR we can calculate this way. Now T bR which is 72 degree centigrade temperature over here and that should be T bR would be equal to 273 plus 72. So, 345 Kelvin temperature

the return reflux is at the top of the tower is 62 degree centigrade the reflex which is coming. So, T cR it would be 273 plus 62. So, it is 335 Kelvin.

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Example 1

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 distillation 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 2.0 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 temperature range)
Heat capacity of n-heptane: 30 kcal/kmol K (over the temperature range)

Solution:

We know that $\quad R_{int} = R_{ext}\left(1 + \dfrac{C_{pR}\left(T_{bR} - T_{cR}\right)}{\lambda}\right)$

$\Longrightarrow \quad R_{int} = 2.0\left(1 + \dfrac{25.25(345 - 335)}{7500}\right) = 2.0(1 + 0.0337) = 2.07$

So, we know that the internal reflux ratio is equal to external reflux ratio into 1 plus C pR into T bR minus T cR by lambda. Now, if we substitute the values R internal would be equal to R external which is given to the external reflux ratio 2 into 1 plus 25.25 into 345 minus 335 divided by lambda is given 7500. So, it would be 2.07. So, that is the internal reflux ratio.

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Example 1

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 distillation 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 2.0 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:

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Solution:

Given:

$x_D = 0.95$; $x_W = 0.05$; saturate liquid feed: q=1, slope of q-line = infinite

$y_{m+1} = \dfrac{R_{int}}{R_{int} + 1}x_n + \dfrac{x_D}{R_{int} + 1}$ $\quad\Longrightarrow\quad$ $\dfrac{x_D}{R_{int} + 1} = \dfrac{0.95}{2.07 + 1} = 0.31$

Now, the other data which are given that is the top product is 95 percent n-hexane the distillate which is point normal hexane which is 0.95 x D and x W the bottom products having the hexane is 0.05 and the feed is saturated liquid. So, when it is saturated liquid so q is 1 and slope of the q-line will be infinite so, it would be a vertical line.

So, from here we can calculate R internal by the operating line equation for the rectifying section is this one, y m plus 1 would be equal to R internal divided by R internal plus 1 into x n plus x D by R internal plus 1.

So, if we substitute the values it would be x d by R internal plus 1 would be equal to 0.95 by 2.07 plus 1 so it is 0.31. So, and for the intercept of this rectifying section operating line the interceptive is 0.31 and we know the starting point x D x D for the stripping section operating line and the equilibrium data which we have discussed in earlier problems. So, with the reflux know with the relative volatility of 2.36, we can draw the equilibrium curve.

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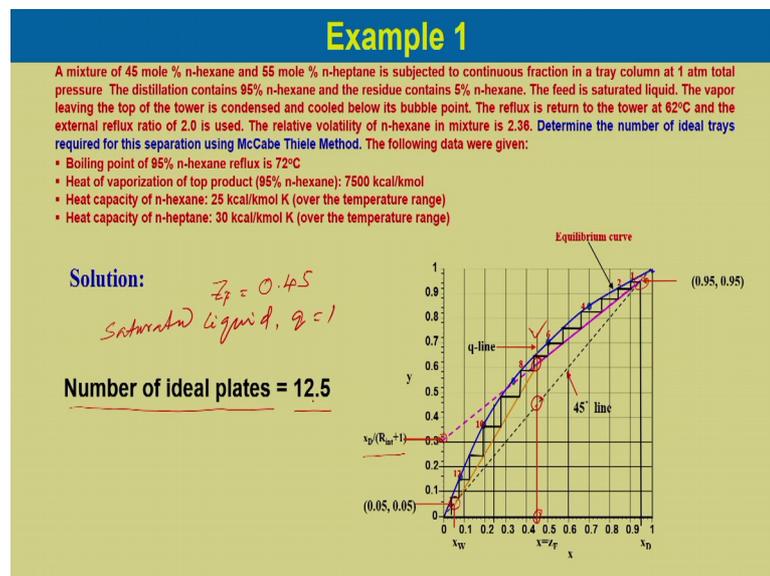
So, equilibrium curve is blue line shown over here in this figure equilibrium curve we can draw, we can locate this point x D x D and we can no locate the point that is x D by R internal plus 1 which is 0.31 over here.

So, we can just draw this operating line in the rectifying section then, we know the feed condition that is z F is equal to 0.45. So, which is over here and if we go vertically it will

no cut over here to the 45 degree diagonal and from that the feed condition is saturated liquid. So, q is equal to 1 and then q line slope of the q line is infinite.

So, this is the vertical line and q line is plotted over here. So, you could locate the intersection between the q line and the rectifying section operating line and knowing the rectifying section starting point because that this now bottoms contains the residue contains 5 percent no n-hexane. So, it is 0.05 0.5 at this location on the 45 degree diagonal and you can join this two intersection points and to get the stripping section operating line.

Now, it is starting from tray 1 over here from x d x d is equal to 1 if you draw the staircase arrangement between the operating line and the equilibrium line, you can calculate the total number of trays ideal trays required by McCabe-Thiele method and also you can locate the feed tray location as well. So, from this we can see that the total number of ideal plates is about 12.5.

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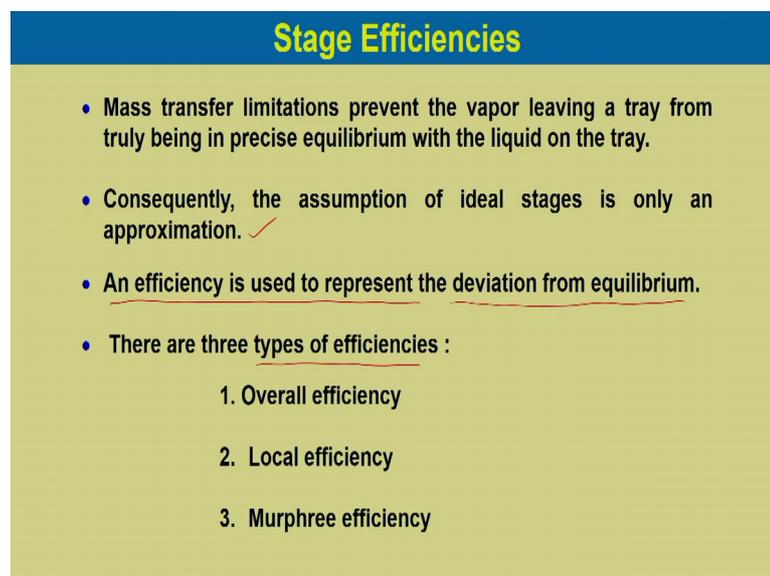
Now, we will discuss another important topic is stage efficiencies. Mass transfer limitation prevent the vapour leaving a tray from truly being in precise equilibrium with the liquid on the tray. So, although ideal trays as we have discussed earlier, an ideal tray should reach the equilibrium between the living stream from that tray. Basically, the vapour going off from the tray and the liquid flowing down from that tray the living streams they would be in equilibrium. But the mass transfer limitations prevent the

vapour which is living on that tray being precisely in equilibrium with the liquid of that tray which is no living down.

So, consequently the assumption of ideal tray is only an approximation. So, the ideal tray is only approximation, but actually they are not reached in equilibrium. So, an efficiency is used to represent the deviation from the equilibrium. So, efficiency has to be used which are departure or deviation from the equilibrium. There are three types of efficiencies which can be defined, one of them is overall efficiency that is the efficiency throughout the column considered as same for all the trays. So, that is overall efficiency and the second one is the local efficiency and third efficiency is the Murphree efficiency.

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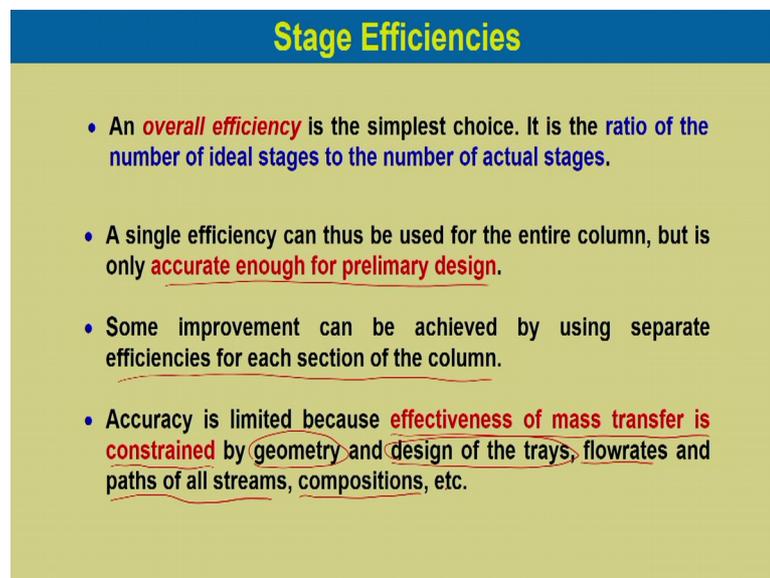
So, we will just discuss one by one, an overall efficiency is the simplest choice; it is the ratio of the number of ideal trays to the number of actual trays. For a given separation no if we just assume that each stage are in equilibrium and we calculate the or each stage we considered as ideal trays and we calculate the number of trays required by McCabe-Thiele or whatever method the graphical method of solution or whatever method we can calculate the ideal trays required for a particular separation.

Now, if we perform the experiments and if we see the total number of actual trays required is would be much higher or it would be higher than the ideal trays because there is no usually there is deviation from the ideal tray behaviour or equilibrium tray behaviour. So, if we divide the ideal trays divided by the actual number of trays required

for that separation which will give the lumped overall efficiency of each tray. A single efficiency can thus be used for the entire column, but is only accurate enough for preliminary design. So, this is basically for the preliminary design.

Some improvement of this design can be achieved by using separate efficiencies for each section of the column, like accuracy is limited because the effectiveness of mass transfer is constrained by the geometry and design of the tray, flow rates and paths of all streams, compositions and so on. So, because of several parameters which influence the effective mass transfer in a particular tray or in a particular column. So, there is deviation from tray to tray or sections to section in a distillation column. So, if we can use different efficiencies for different sections or for different tray that would be helpful for finding out the actual tray required.

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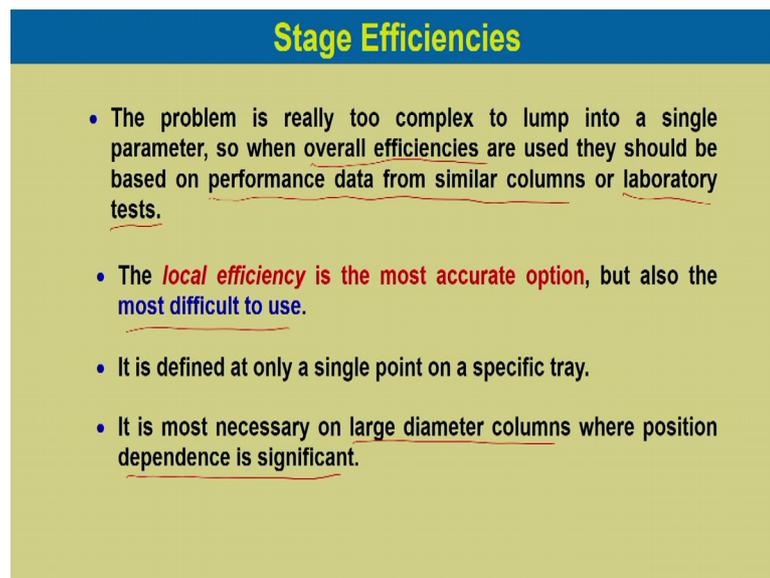
Stage Efficiencies

- The problem is really too complex to lump into a single parameter, so when overall efficiencies are used they should be based on performance data from similar columns or laboratory tests.

- The *local efficiency* is the most accurate option, but also the most difficult to use.

- It is defined at only a single point on a specific tray.

- It is most necessary on large diameter columns where position dependence is significant.

The problem is really too complex to lump into a single parameter. So, when overall efficiencies are used they should be based on performance data from similar column or laboratory test. So, if we use the overall efficiencies for a particular separation, then there should be a performance data of the similar column or laboratory test, so that we can scale of into that particular application that what would be the actual number of trays required and what would be the efficiency overall efficiency. The local efficiency is the most accurate option, but also it is most difficult to use or it is definitely no defined at only a single point on a specific tray.

So, local efficiency is the efficiency at a particular location in a particular tray. So, this point efficiency is very difficult to use, because you have so many locations in a tray and the you have so many trays. So, local efficiency although it is most accurate, but it is very very difficult to use, and it only a local efficiency calculate the efficiency at a single point; this is important when we have a large diameter column we have the tray area is much bigger where the efficiency of the tray depends on the geometry of the column or geometry of the tray. So, it is most necessary on large diameter columns where position dependency is significant.

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A Murphree efficiency is probably the most common choice, since it represent a workable compromise between accuracy and ease of use. So, it has the same form as a local efficiency, but is based on the tray average compositions. So, it does not take the composition at a particular location to calculate the tray efficiency.

In this case the Murphree stage efficiency takes the average composition of on a particular tray, but it varies from tray to tray or sections to section. So, it is very close to the local efficiency. So, it is defined by E M Murphree stage efficiency is equal to y n minus y n plus 1 divided by y equilibrium in n stage n minus y n plus 1.

So, the actual enrichment of the vapour over the n th tray we can just write y n this is the actual enrichment, y n minus y n plus 1 would be equal to e m into y equilibrium n minus y n plus 1. So, this is actual enrichment in that column in that tray.

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So, in this case what we used to do? A pseudo equilibrium line is drawn such that its vertical distance from the operating line is E M times the distance between the operating line and the equilibrium line. So, like if you consider point O and point P over here; O S is the E M times of the total vertical distance between the operating line and the actual equilibrium line which is blue line O P.

So, staircase construction between the pseudo equilibrium line which is know pink colour over here shown over here between these will give you the actual or real number of trays or plates required for the particular separation. So, typical values for E M is generally 0.6 to 0.75 for the sieve trays, it is most common 0.6 to 0.75.

We know that the liquid leaving a tray is not really the same as the tray average. So, a Murphree efficiency effectively assumes perfect mixing on the tray, because it considered the average concentration of that tray, but actually the composition which is varying throughout the tray it is not the steam which is leaving is not exactly same as the average temperature.

So, in practice we normally measure the liquid composition and get the vapour composition from an equilibrium calculation or diagram. So, in the case of multi component system the efficiencies are different for each component.

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The overall efficiency and the Murphree efficiency are not directly related and you cannot use an average Murphree efficiency in place of an overall value. To use local or Murphree efficiency with a graphical method the true equilibrium line as we said the true equilibrium line is blue lines shown over here is replaced with the effective equilibrium curve located between the which is pink line located between the operating line and the equilibrium line. The effective curve is used to count the stages.

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Now, in this case note that that the efficiency does not apply to the reboiler. So, the true equilibrium curve should be used for the last stage of the stripping section. So, it does not include the no reboiler as an equilibrium stage in this calculation for the stage efficiency. So, to construct an effective equilibrium curve is not difficult, the effective curve is given by y effective would be equal to y plus E M into y equilibrium at that stage at n minus y; y equilibrium n is the equilibrium concentration of that tray vapour composition and y is the values of the operating lines or operating equations.

So, y represents the operating curve data; now a plot of y effective will produce the an interior line on the equilibrium diagram construction. So, the pink line drawn over here is data calculated based on y effective. So, we will have one example to draw the effective equilibrium curve and how to calculate the number of aerial trays required for a particular separation.

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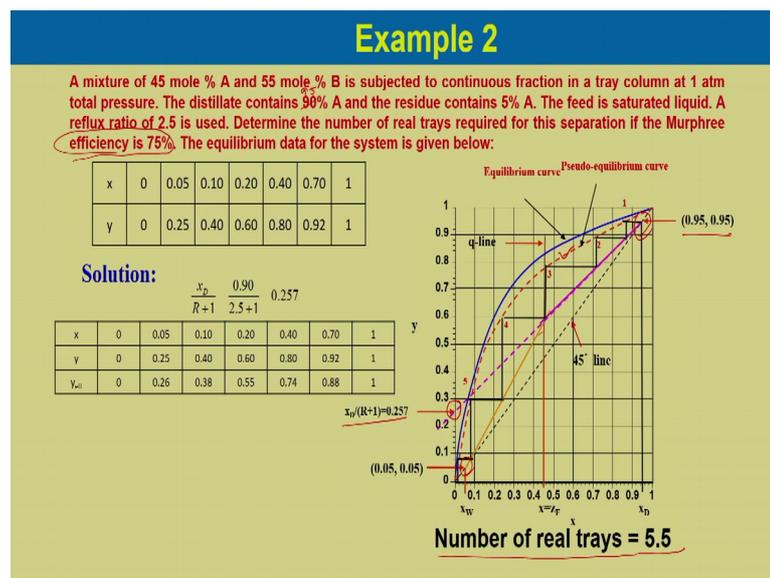
A mixture of 45 mole percent A and 55 mole percent B is subjected to continuous fractionation in a tray column at 1 atmosphere total pressure. The distillate contains 90 percent A and the residue contents 5 percent A. The feed is saturated liquid. A reflux ratio of 2.5 is used. Determine the number of real trays required for the separation if the Murphree efficiency is 75 percent. The equilibrium data for the system is given over here this is the x y data.

Now x D by R plus 1, we can calculate 0.9 x D is 0.9 divided by 2.5 plus 1. So, that is 0.257 and equilibrium data is given so we can calculate y effective from the equations we have shown.

So, now we can plot the data over here is known which is 0.95, 0.95. So, this should be no not 90 percent this is 95 percent and you can calculate the intercept so which is over here. So, intercept and you can plot the operating line for the no rectifying section, the feed condition is given like earlier. So, you can plot the q line for saturated liquid. So, q line would be vertical, you know the bottom no location which is 0.05, 0.05. So, you can have the stripping section operating line, now we have to draw the pseudo equilibrium curve with this data y effective and the x data.

You can plot the pseudo equilibrium curve shown over here with the red line. So, now, this the actual equilibrium data is not important to calculate the real number of trays. So, what we need to do? We need to calculate so to calculate y effective we have used the stage efficiency no Murphree efficiency of 0.75 and that is how we calculated the pseudo equilibrium curve.

Now, if you just draw the staircase arrangement like McCabe-Thiele method between the pseudo equilibrium curve and the operating line between stripping section and the rectifying section you will obtain the total number of real trays required for the separation, which is about 5.5.

So, if you just remove the pseudo equilibrium curve and draw the staircase arrangement with the actual equilibrium curve, the number of trays required will obviously reduced because, the vertical distance between the equilibrium curve pseudo equilibrium curve and the operating line is less than the actual equilibrium curve and the operating lines.

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Now, another important thing which we will discuss is the use of open steam. So, in the distillation column in general heat to a distillation operation is supplied by reboiler. So, which is over here at the bottom and the steam does not come in direct contact with the boiling solutions. So, the solution which is coming to be separated into the different fractions, they does not come in contact with the steam over here. The boiler and mixture or solution is heated and part of that solvent is in the vapour which is returned as a boilup to the tower at the bottom.

So, in this case the boiling solution does not come in contact with the steam. However, when separating an aqueous solution if the residue or bottom products of water and traces of the solute, sometimes it may be convenient to feed live steam at the bottom of the tower. In this case no you need not to have reboiler at the bottom. So, live steam you can just inject into the tower bottom.

The steam is injected as small bubbles into the liquid in the tower bottom. So, that will heat up the solution. The vapour leaving the liquid is then in equilibrium with the liquid if sufficient contact is obtained.

Now, if we you know use the balance equation, the overall balance on the tower and a component balance on A. This is overall balance you had feed F and you have a steam, live steam injected is S would be equal to your distillate and the bottoms. Now, if we do the species mol balance F into z F plus S into y S would be equal to D x D plus W x W,

here S is the mole per hour of steam and y S generally no solute present in the steam. So, we considered y S is equal to 0, the y S is the mol fractions of a in the steam.

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The operating line equation for the rectifying section remains the same as derived earlier operating line equation, but that for the stripping section is; obviously, different the total mass balance and material balance of component A over a stripping envelope is we can do that that is L m is the liquid flow out plus the steam which is given S would be equal to G m plus 1 plus W. Then if we do the species mol balance it would be L m x m plus S into 0 because there is no solute in the steam would be equal to G m plus 1 into y m plus 1 plus W x W.

So, each terms have their usual significance as we discussed before and we assume that the inlet steam is saturated. So, if we solve these it would give us y m plus 1 would be equal to L m by G m plus 1 into x m minus W by G m plus 1 into x W. So, this is the stripping section operating line.

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Since, there is no reboiler L m would be equal to W and for saturated steam entering. So, S would be equal to G m plus 1 and then this operating line equation in the stripping section would be y m plus 1 would be equal to W by S x m minus W by S x W. So, if you just take out W by S from the both term. So, it would be we can write L by G m plus 1 would be equal to W by S would be equal to y m plus 1 by x m minus x W. So, this is L by G plus G m plus 1 or W by S or y m plus 1 by x m minus x W is the slope of the operating line in the stripping section.

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Now, let us take an example: An aqueous solution of ethanol which contains 60 mol percent of methanol is subjected to a fractionation at 1 atmosphere total pressure. Open steam is used as the source of heat energy at a rate of 100 kilo mol per hour. The top product contains 90 percent methanol and the bottom products contains 5 percent methanol. Although it is given 5 percent, but actually it would be 10 percent methanol, but considering 5 percent we will know how to do the operating line my equations. The feed is saturated liquid and reflux ratio of 2.5 is used.

Now, we have to obtain the stripping section operating line, determine the number of ideal plates required for the given separation, the steam rate required for the separation is we have to calculate. The equilibrium data for methanol water system at 1 atmosphere pressure is given. So, this is the equilibrium data.

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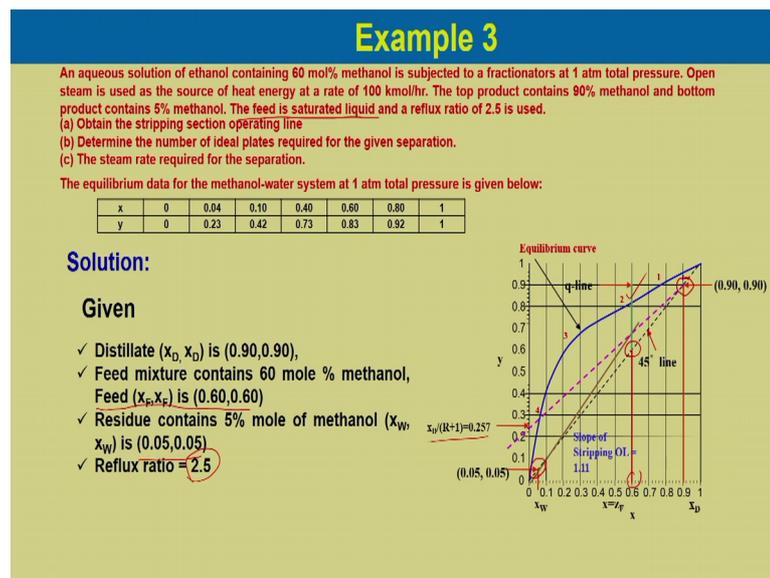
Now, distillate x D, x D is 0.9 and 0.9 and feed mixture contains 60 mol percent of methanol. So, feed x F, x F is 0.6, 0.6 residue 5 mol percent of methanol so x W, x W would be 0.05, 0.05. So, these data are given and reflux ratio is 2.5. So, if you plot them so this is the starting point in the top sections of the column and this is the starting for end points in the bottom section of the columns.

The feed condition is given is point 60 mol percent. So 0.6, so we can locate the point on the 45 degree diagonal and feed is saturated liquid which is given over here. So, we can plot the q line which is vertical from this 45 degree diagonal we calculate x D by R plus

1 is 0.257. So, we can just draw the operating line in the rectifying section and we can also draw the, connect the operating line in the stripping section.

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So, intercept of the rectifying section is 0.257, Q line vertical z F is equal to 0.6 saturated feed. So, and this is the equation for the stripping section operating line. So, slope of the operating line we can calculate 1.1. So, we have to calculate the obtain the stripping section operating line. So, intercept is no 0.056. So, equation for the operating line for the stripping section would be y is equal to 1.11 x minus 0.056. So, this is the equation for the stripping section operating line.

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Example 3

An aqueous solution of ethanol containing 60 mol% methanol is subjected to a fractionators at 1 atm total pressure. Open steam is used as the source of heat energy at a rate of 100 kmol/hr. The top product contains 90% methanol and bottom product contains 5% methanol. The feed is saturated liquid and a reflux ratio of 2.5 is used.
(a) Obtain the stripping section operating line
(b) Determine the number of ideal plates required for the given separation.
(c) The steam rate required for the separation.

The equilibrium data for the methanol-water system at 1 atm total pressure is given below:

x	0	0.04	0.10	0.40	0.60	0.80	1
y	0	0.23	0.42	0.73	0.83	0.92	1

Solution: (b)

(b) By McCabe Thiele method the number of ideal stages = 4.0

Now, the second part we have to do the number of ideal plates required for the separation by McCabe-Thiele method or no starting from this point, we can calculate the number of you know ideal trays required.

So, if we just plot the horizontal line and the vertical line. So, from starting from this point go to the equilibrium curve and go vertically to the operating line and then horizontal line and then go to vertically and then go horizontally; so vertical line and then the horizontal line. So, if you look into this so about know 4 number of the ideal stages are required for this particular separation and then the third problem which is given to calculate the steam rate required for the separation.

So, in order to calculate the steam rate we have to use the material balance equation. So, that is F plus S is equal to D plus W. So, feed and steam rate and then the distillate rate and the bottoms. So, steam is given 100 plus S would be equal to D plus W. Now for saturated steam entering S would be equal to G m plus 1. So, we can write 100 plus G would be equal to D plus W and if you just consider the species mol balance equation, it is 100 into no z F that is 0.6 plus S into 0 that is the steam does not contain any solute. D into 0.9 plus W into 0.05. So, from here we can calculate, no this is 60 would be equal to D into 0.9 plus W into 0.05.

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So, from here W would be, W by G would be 1.11. So, G would be W by 1.11. So, we can calculate D is given 25 kilo mol and W is 750 kilo mol. So, we can calculate G would be equal to 750 kilo mol per hour, which is 675 kilo mol per hour. So, that is 12150 kg steam per hour.

So, thank you for you know attending this lecture. So, we will continue our discussion on distillation in the next class.