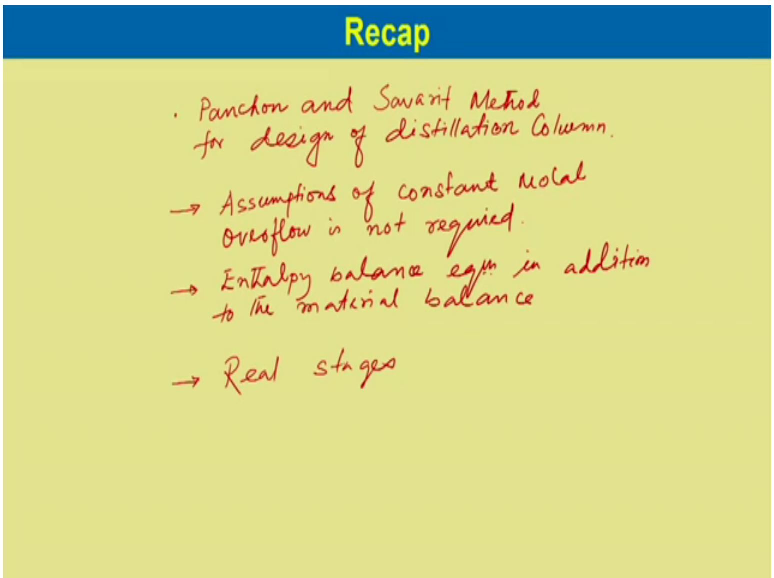
Mass Transfer Operations-I
Prof. Bishnupada Mandal
Department of Chemical Engineering
Indian Institute of Technology, Guwahati

Distillation-V
Lecture - 38
Packed Distillation

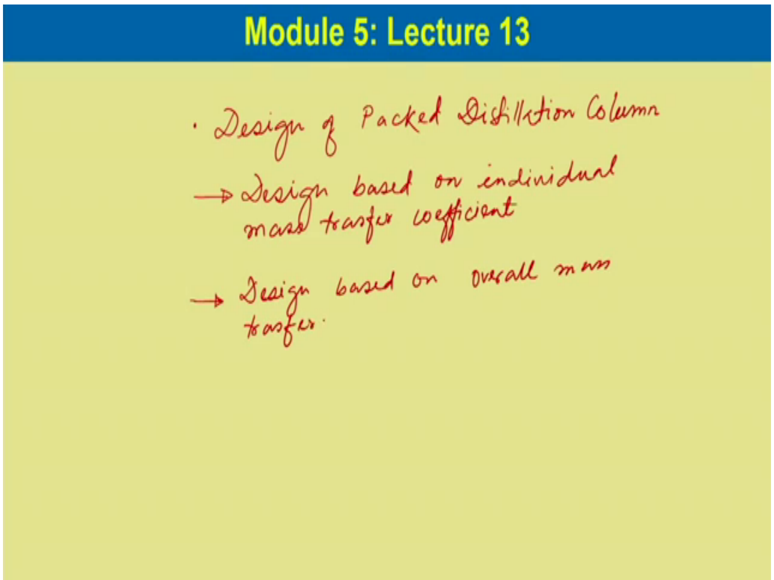
Welcome to the 13th lecture of module 5 of Distillation Operation. In this module, we are discussing distillation operation before going to this lecture let us have brief recap on our previous lecture.

(Refer Slide Time: 00:58)



In our previous lecture, we have mainly consider Panchon and Savarit method for design of distillation column. In this method, as you know we need not to have the assumptions as we had with the McCabe Thiele method assumptions of constant Molal overflow is not required. So, this consider enthalpy balance equation in addition to the material balance. So, the stages obtained using this method is basically the actual number of stages or the real stages. So, in this method we can obtain real stages.

(Refer Slide Time: 02:33)

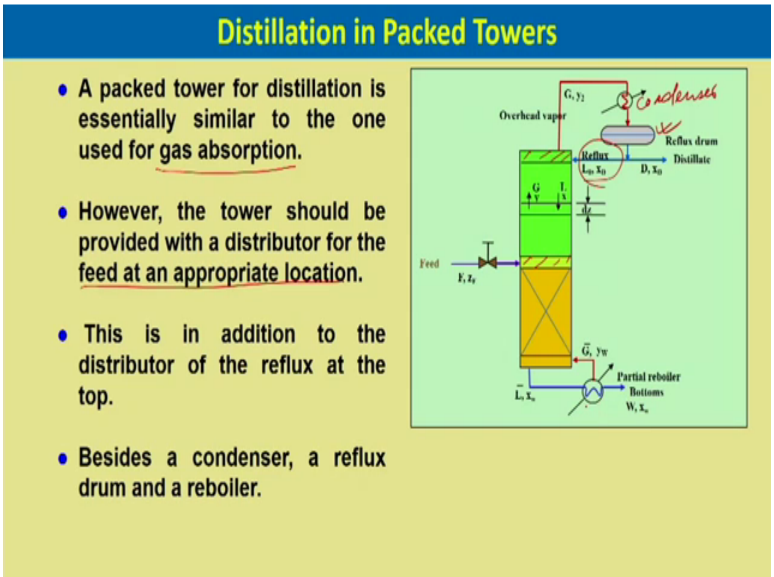


Now, in this lecture we will mainly focus on design of packed distillation column. So, under which we will first consider design based on individual mass transfer coefficient and second we will consider design based on overall mass transfer coefficient.

(Refer Slide Time: 03:22)

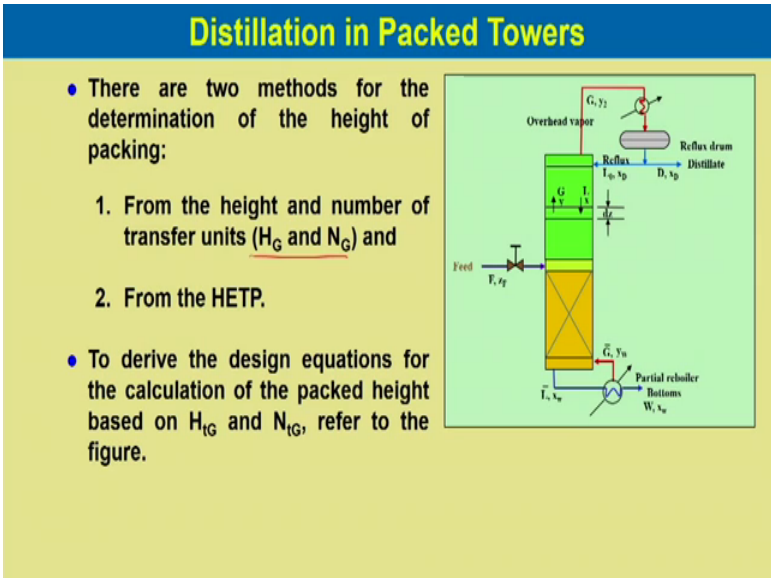
So, let us start with distillation in packed towers. Distillation of a liquid mixture in packed towers is preferred now let us start with distillation in packed towers. Distillation of a liquid mixture in a packed tower is preferred in the following cases; one is vacuum or low pressure operation. So, if we have vacuum or low pressure operation then packed tower may be used. The second case is the low capacity of the feed and third case small allowable pressure drop if we need small allowable pressure drop, then we can use packed towers and third one if there is corrosive services like the system corrode the equipment and other parts of the column then packed towers may be preferred.

(Refer Slide Time: 04:31)



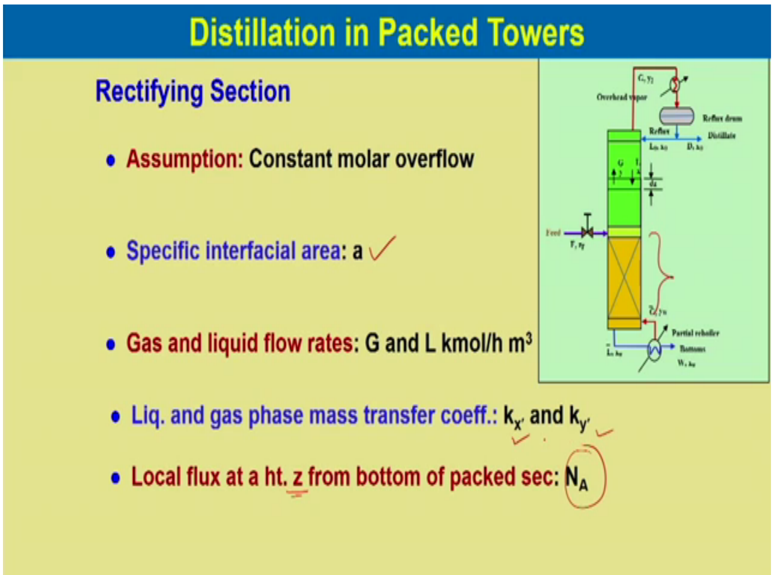
This is a typical packed distillation columns a packed tower for distillation is essentially similar to the one used for gas absorption. However, the tower should be provided with a distributor for the feed at appropriate location so we should have distributor at the feed point. This is in addition to the distributor of the reflux at the top. So, you can see we have a reflux and also we need to provide the distributor here. Besides a condenser over here and a reflux drum which is over here and a reboiler. So, this is the reflux drum and this is the reboiler.

(Refer Slide Time: 05:30)



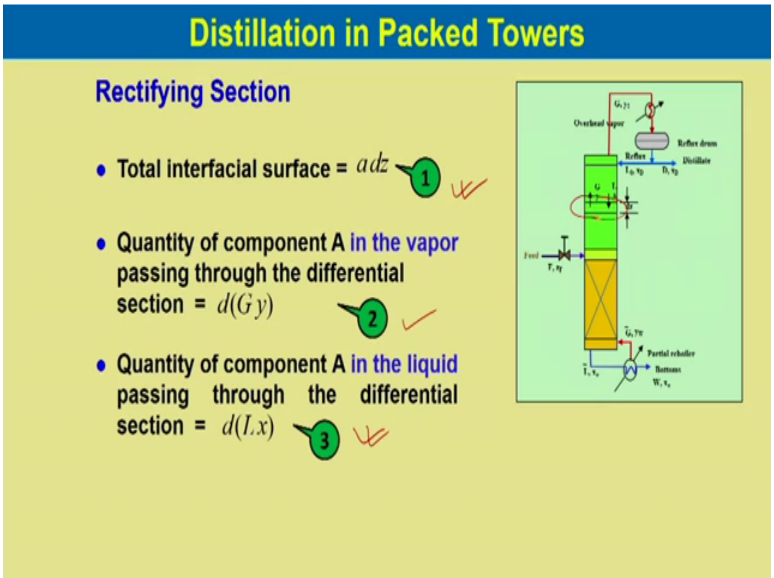
There are two methods for the determination of the height of the packing one of them is from the height and number of transfer unit as we have discussed in case of packed towers in absorption and second method is based on the HETP Height Equivalent to a Theoretical Plate. To derive the design equations for the calculation of the packed height based on H tG and N tG that is Height of transfer unit Gas phase transfer unit and N tG is Number of gas phase transfer unit we need to refer to this figure.

(Refer Slide Time: 06:14)



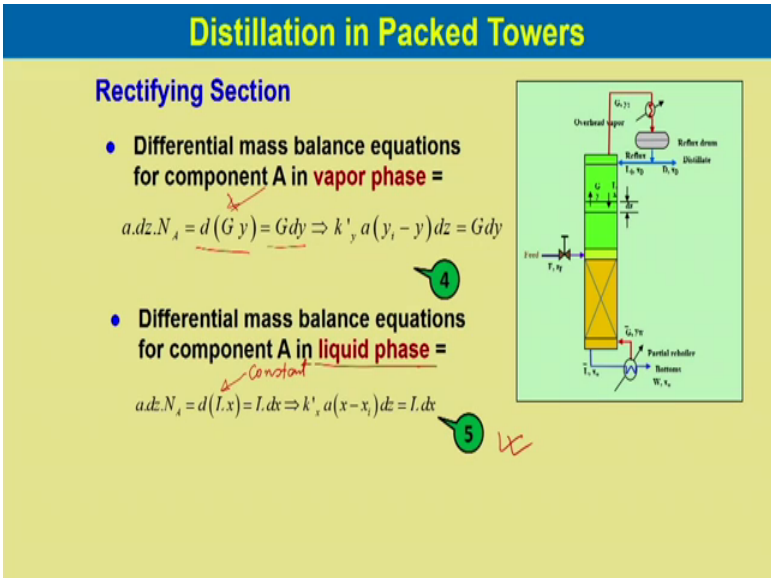
Now, we have two sections above the feed locations is the rectifying sections and below the feed location is the stripping section. Now assumptions over here is constant molal overflow and specific interfacial area is a. So, unit of a is meter square per meter cube of packing, gas and liquid flow rate G and L in kilo mole per hour meters cube, liquid and gas phase mass transfer coefficient k x dash and k y dash. Now local flux at height z from bottom of the packed section we can define by N A.

(Refer Slide Time: 07:09)



So, now the total interfacial surface would be a into dz. So, if we consider a differential area dz differential sections dz the total interfacial surface would be a into dz which is shown in equation 1. Now quantity of component a in the vapour passing through the differential sections this sections dz would be d G y which is shown in equation 2. Now quantity of component A in the liquid passing through the differential section should be d L into x so which is shown in equation 3.

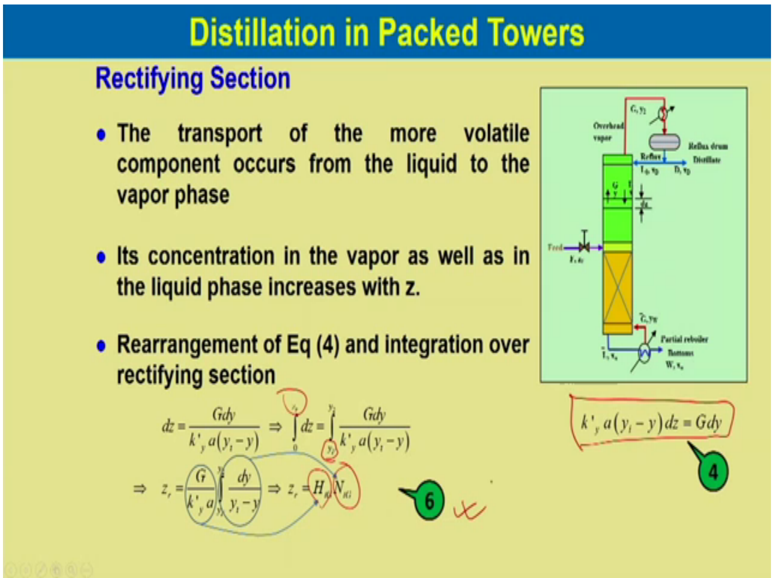
(Refer Slide Time: 07:57)



Now if we do the differential mass balance equation for component A in the vapour phase we can write a into dz into N A would be equal to d G y which is equal to G into dy which is equal to k y dash a into y i minus y into dz which is equal to G dy. So, since G is constant as we have considered constant molar overflow. So, this we can write G into d y, k y dash a is the individual mass transfer coefficient in the gas phase.

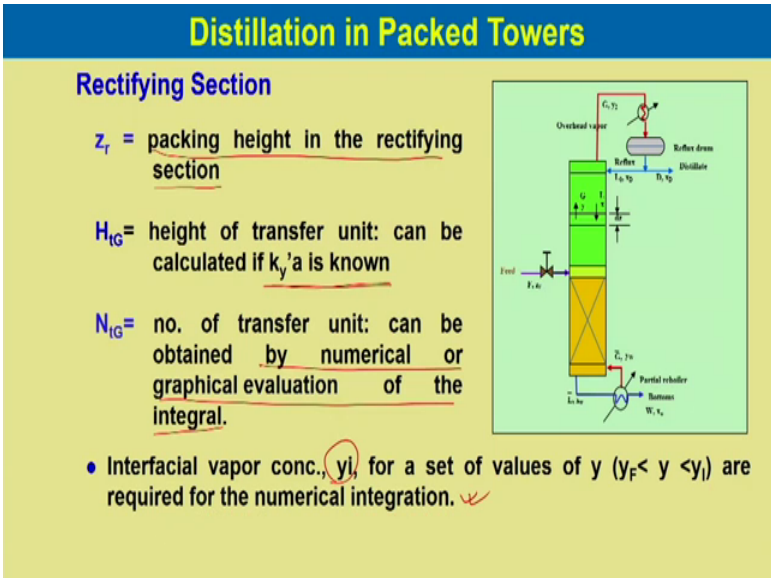
Now differential mass balance equation for component A in the liquid phase, we can write a dz into N A is equal to d Lx which is equal to L into dx it is similar assumptions we have taken the equimolal overflow or constant molar overflow for gas and liquid so L is constant. So, in that case we will have L into d x so from here we can write k x dash into a x minus x i into dz would be equal to L d x so which is shown in equation 5.

(Refer Slide Time: 09:34)



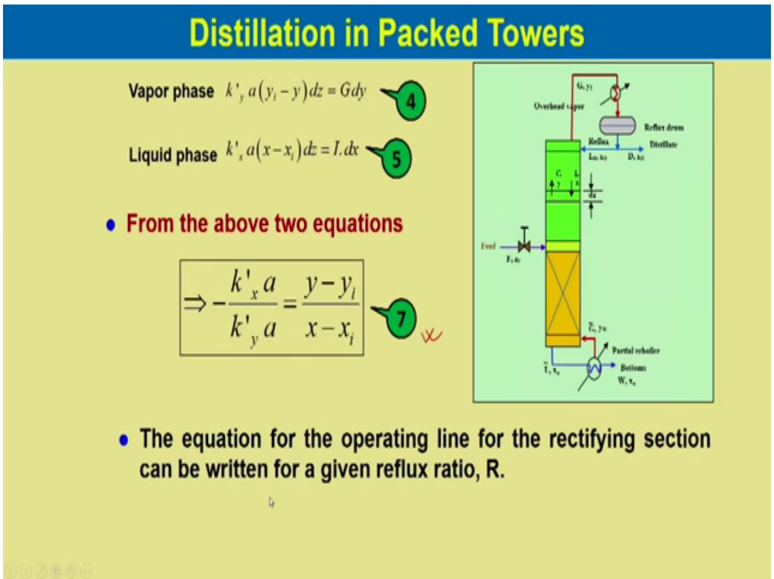
Now, the transport of more volatile component occurs from the liquid to the vapour phase. Its concentration in the vapour phase as well as in the liquid phase increases with z. Now if we rearrange equation 4 and integration over the rectifying section if we do this is equation 4 and if we do the rearrangement and integration over the height that is from 0 to z r dz. This z r is the height of the rectifying section packing height which is equal to integral y f to y 2 y f is the concentration of the feed to y 2 Gdy divided by k y dash a into y i minus y. So, after integrations it would be z r would be equal to G by k y dash a integral y f to y 2 dy by y i minus y. So, from here we can write this part is basically H tG and the other integral part is the N tG. So, which is shown in equation 9 equation 6.

(Refer Slide Time: 11:07)



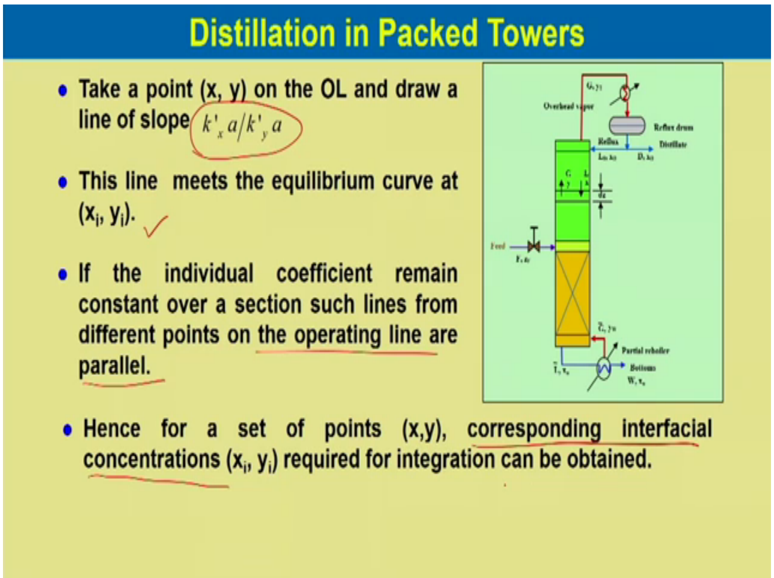
Now, z r is the packing height in the rectifying section and H tG is the height of the transfer unit and can be calculated if k y dash a is known. N tG is the number of transfer unit and can be obtained by numerical or graphical evaluation of the integral. Interfacial vapour composition that is yi as we have done before in case of absorption, for a set of values of y y F less than y less than y 1 are required for the numerical integration.

(Refer Slide Time: 11:59)



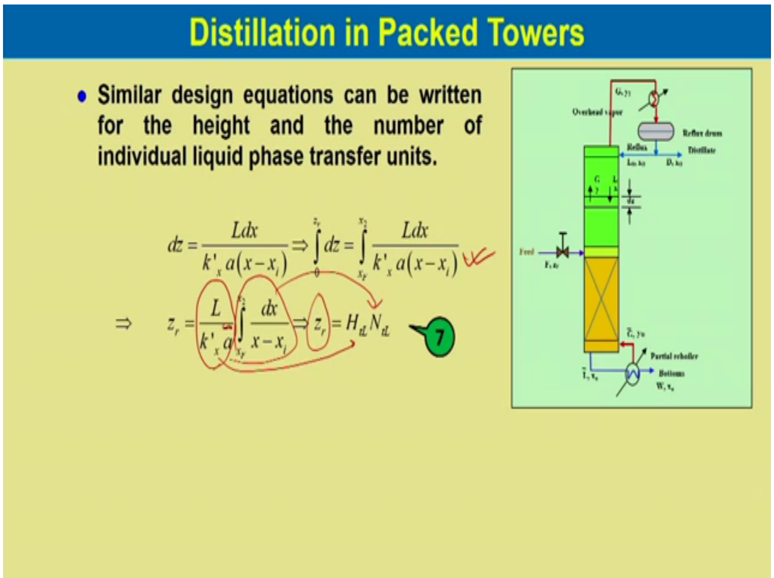
Now, vapour phase equations 4 as we have obtained k y dash a y i minus y into dz is equal to G into dy, liquid phase equations we have obtained k x dash a x minus x i dz is equal to L into dx. From this two equations, we can obtain minus k x dash a divided by k y dash a is equal to y minus y i divided by x minus x i. The equation for the operating line for the rectifying section can be written for a given reflux R.

(Refer Slide Time: 12:48)



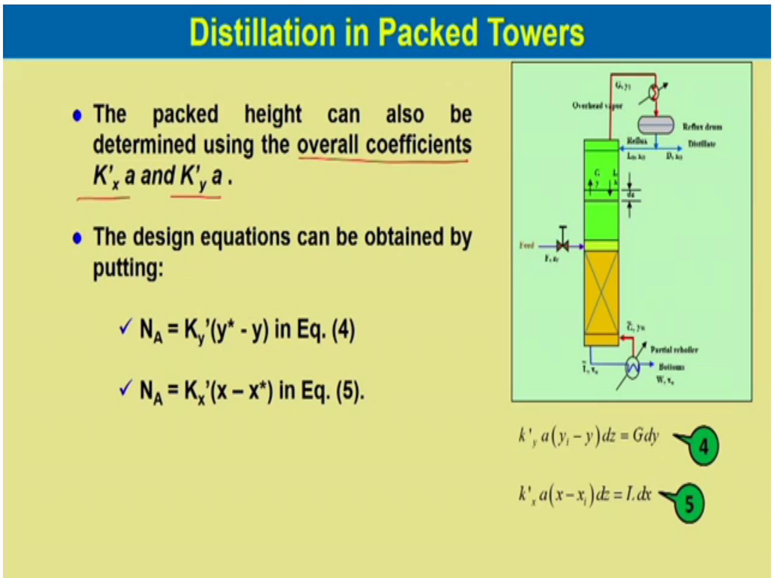
Then take a point x, y on the operating line and draw a line of slope k x dash a by k y dash a. This line meets the equilibrium curve at x i, y i. If the individual coefficient remain constant over a section such line from different points on the operating line are parallel. So, if this ratio remains constant throughout then, we can draw the parallel line from the operating line on the equilibrium curve. Hence for a set of points x, y corresponding interfacial concentrations that is x i, y i required for the integration can be obtained.

(Refer Slide Time: 13:51)



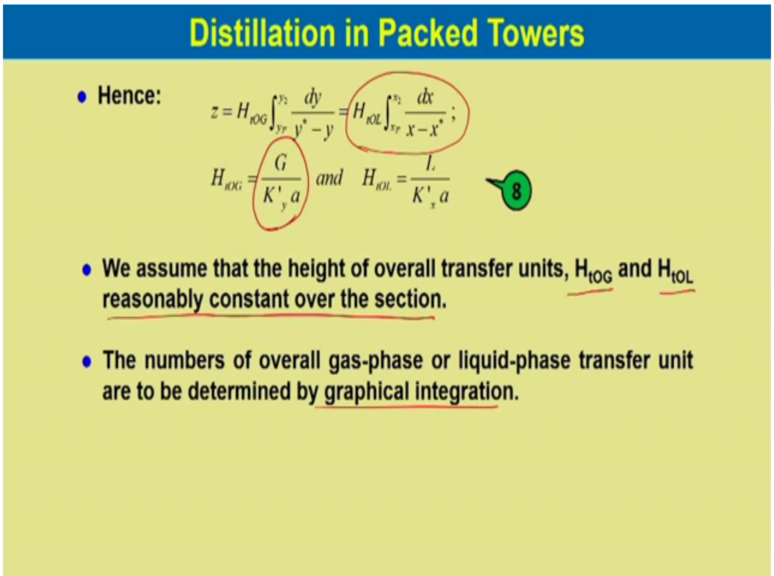
Similar design equations can be written for the height and the number of individual liquid phase transfer unit. So, for liquid phase we can write dz would be equal to Ldx divided by k x dash a into x minus x i and then if we integrate 0 to z r dz would be equal to integral x f to x 2 Ldx divided by k x dash a into x minus x i. Now after integration the left hand side would be z r which is equal to L by k x dash a integral x f to x 2 dx by x minus x i. So, from here we can write z r is equal to this part is nothing but H tL the height of liquid phase transfer unit and this part is nothing but number of liquid phase transfer unit. So from here we can calculate the height of the packing in the rectifying section based on the liquid phase.

(Refer Slide Time: 15:28)



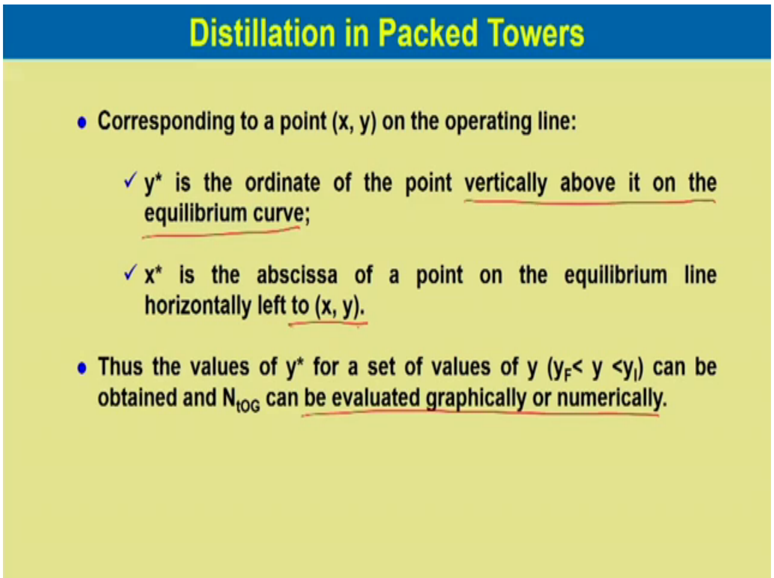
The packed height can also be determined using the overall coefficient that is capital K x a dash and capital K y dash a. The design equations can be obtained by putting N A is equal to K y dash into y star minus y in equation 4 and N A is equal to K x dash x minus x star in equation 5.

(Refer Slide Time: 16:14)



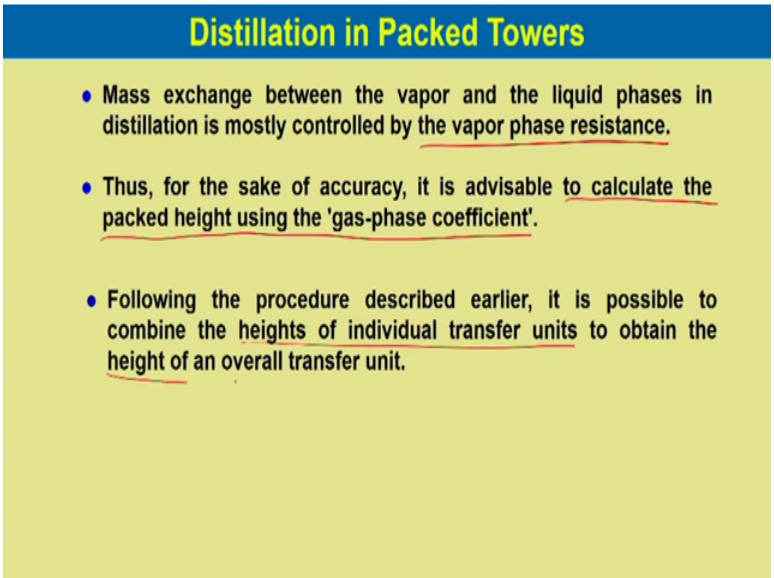
So, if we do that we will obtain z is equal to H tOG that is g by capital K dash y a and integral y f to y 2 dy by y star minus y this part is nothing but the N tOG and we can also write in terms of H tOL that is H tOL is L by capital K x dash a. We assume that height of overall transfer unit that is H tOG and H tOL reasonably constant over the section. The number of overall gas phase or liquid phase transfer unit are to be determined by graphical integration as we have done for the individual mass transfer coefficient.

(Refer Slide Time: 17:28)



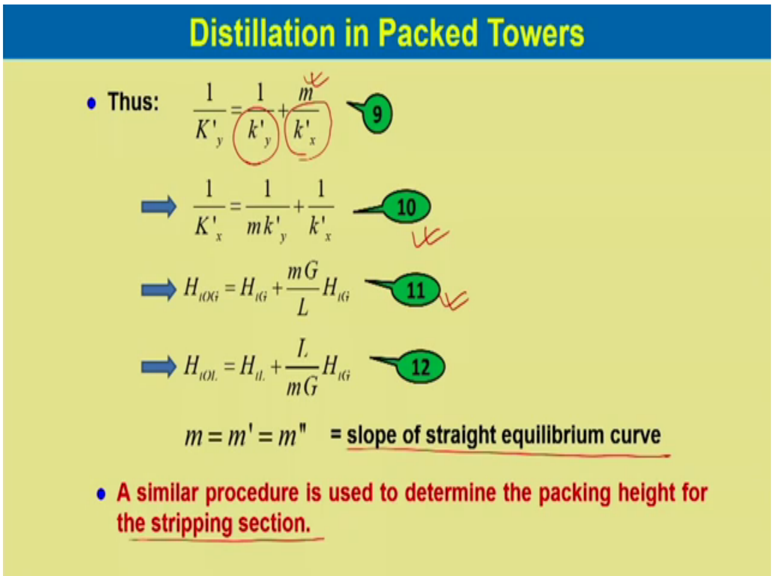
Corresponding to a point x, y on the operating line we can obtain y star is the ordinate of the point vertically above it on the equilibrium curve and x star is the abscissa of a point on the equilibrium line horizontally left to x, y. Thus the values of y star for a set of values of y which is y F less than y less than y i can be obtained and N tOG can be evaluated graphically or numerically.

(Refer Slide Time: 18:12)



Now mass exchange between the vapour and the liquid phases in distillation is mostly controlled by the vapour phase resistance. Thus, for the sake of accuracy it is advisable to calculate the packed height using gas phase coefficient. Following the procedure described earlier it is possible to combine the height of individual transfer unit to obtain the height of an overall transfer unit.

(Refer Slide Time: 18:49)

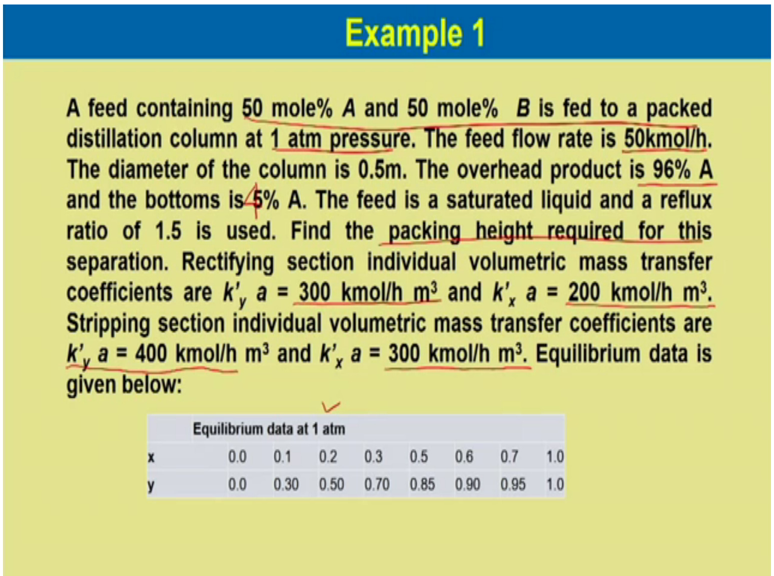


Thus 1 by K y capital K y dash would be equal to 1 by k y dash plus equal to 1 by small k y dash plus m by small k x dash here capital K y dash is the overall gas phase mass transfer coefficient and small k y dash is the individual gas phase mass transfer coefficient and small k x dash is the individual mass transfer coefficient in the liquid phase and m is the Henry's constant.

Similarly, we can write in terms of the overall mass transfer coefficient based on the liquid phase. So, we can write 1 by capital K x dash would be equal to 1 by m small k y dash plus 1 by small k x dash so this is equation 10.

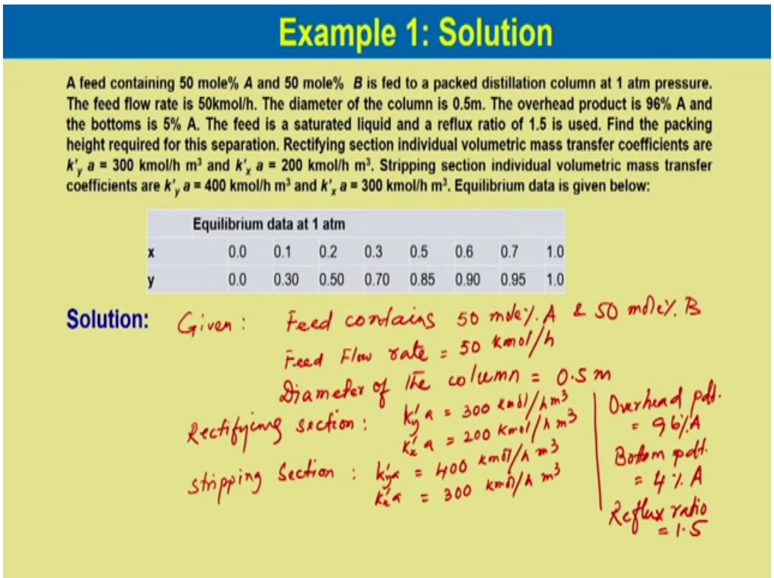
Now we can relate with the equation derived earlier H tOG would be equal to H tG plus m G by L H tG and similarly for the liquid phase we can write H tOL would be equal to H tL plus L by m G H tG. So, m equal to m dash equal to m double dash which is the slope of the straight equilibrium curve. A similar procedure is to be used to determine the packing height for the stripping section. So, once we can get the packing height in the stripping section considering the procedure we followed in the rectifying section we can find out the total height of the packing required for a distillation operation.

(Refer Slide Time: 21:35)



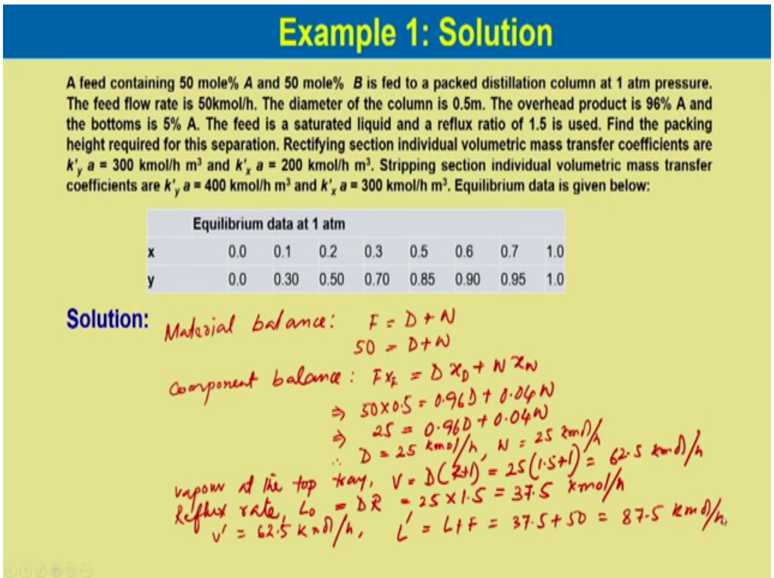
Let us consider an example to solve the packed tower problems. A feed containing 50 mole percent A and 50 mole percent B is fed into a packed distillation column at atmospheric pressure. The feed flow rate is 50 kilo mole per hour, the diameter of the column is 0.5 meter, the overhead product is 96 percent A and the bottom is 4 percent A, the feed is a saturated liquid and a reflux ratio of 1.5 is used. Find the packing height required for this separation. Rectifying section individual volumetric mass transfer coefficient are k y dash a is 300 kilo mole per hour meter cube and k x dash a is 200 kilo mole per hour meter cube. Stripping section individual volumetric mass transfer coefficients are k y dash a is 400 kilo mole per hour meter cube and k x dash a is 300 kilo mole per hour meter cube. Equilibrium data is given in this table, now we need to find out the height of the packing required for this separation.

(Refer Slide Time: 23:00)



So, the data which are given ; the feed condition 50 mole percent A and 50 mole percent B , feed flow rate 50 kilo mole per hour, diameter of the column which is 0.5 meter, rectifying section volumetric mass transfer coefficient k y dash a is 300 kilo mole per hour meter cube and k x dash a is 200 kilo mole per hour meter cube, stripping section k y dash a is 400 kilo mole per hour meter cube and k x dash a is 300 kilo mole per hour meter cube. The overhead product is equal to 96 percent A and bottom products 4 percent A, reflux ratio given which is equal to 1.5. So, these are the parameters given, we have listed them.

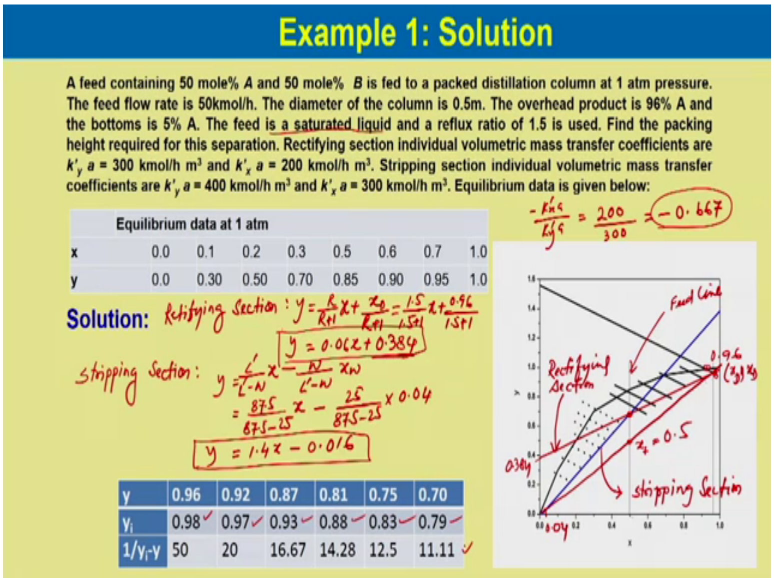
(Refer Slide Time: 25:20)



Now let us do the material balance. Overall material balance is F is equal to D plus W F is given 50 which is equal to D plus W. Now component balance on A it would be F x F would be equal to D x D plus W x W. Now if we put F this 50 into 0.5 would be equal to 0.96 D plus 0.04 W. So, from here we can write 25 would be equal to 0.96 D plus 0.04 W.

So, therefore, we can calculate by substituting the total material balance on this equations we can get D is equal to 25 kilo mole per hour and W is equal to 25 k mole per hour. Now vapour at the top tray V would be equal to D into R plus 1. So, we can substitute D 25 into 1.5 plus 1 which is equal to 62.5 k mole per hour. Now reflux rate which is L naught would be equal to D into R D is 25 into L naught into R is 1.5 which is equal to 37.5 kilo mole per hour. So, V dash is; V dash is 62.5 kilo mole per hour and L dash would be equal to L plus F which is 50, 37.5 plus 50 which is equal to 87.5 kilo mole per hour.

(Refer Slide Time: 28:57)



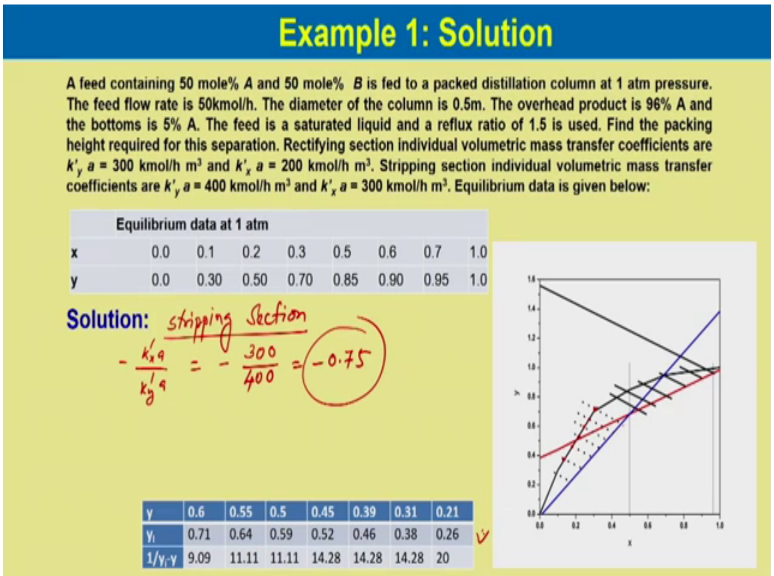
Now, for the rectifying section, we can obtain the operating line equation rectifying section if we know the operating line equation which is y is equal to R by R plus 1 into x plus x D by R plus 1. So, if we substitute the values, it is 1.5 by 1.5 plus 1 into x plus 0.96 divided by 1.5 plus 1. So, this will give y is equal to 0.06 x plus 0.384. So, 0.384 is the intercept on the y axis. Similarly for the stripping section we can write y is equal to L dash by L dash minus W into x minus W by L dash minus W into x W. So, if we substitute then it would be 87.5 divided by 87.5 minus 25 x minus 25 by 87.5 minus 25 into 0.04. So, this would give y is equal to 1.4 x minus 0.016. This is the stripping section operating line equation and this is the rectifying section operating line equation.

So, we know the feed condition and we know this point over here. Once we draw the operating line for the rectifying section we locate x D x D on the 45 degree diagonal which is not shown over here. So, there will be a 45 degree diagonal over here and once we locate this point with the intercept on the y axis that is 0.384, we can draw the operating line for the rectifying section and we know the feed condition which is 50 percent A and 50 percent B.

So, 0.5; 0.5 o is over here if we end the feed condition is given so we locate this point on the 45 degree diagonal that is x F is equal to 0.5 and the feed condition is given is saturated liquid. So, the feed line slope it will go vertically. So, from this point we draw a vertical line. So, this is the q line or feed line and we locate the intersection point and we know the starting point where the bottom conditions 0.04 and we join this two point to get the operating line for the rectifying section for the stripping section.

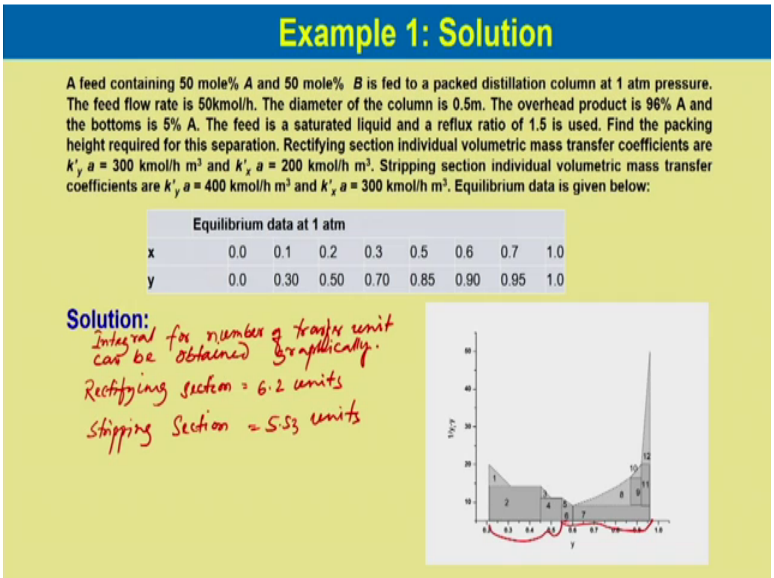
Now, what we need to do, we know the slope of the line that is in the rectifying section we know minus k x dash a divided by k y dash a which is equal to 200 by 300 which is equal to minus 0.667. So, with this, this is the slope of the tie line and with this we can draw several point lines in the rectifying section starting from the 0.96 over here. So, with 0.96 we can draw with the slope which will connect to the equilibrium line and we will obtain the value of y i which is shown over here. So, accordingly with different points, we can draw the parallel line if slope does not change then we can obtain the values of y i that is the equilibrium concentration or interfacial concentration then once we know this points we can calculate 1 by y i minus y from this table. So, we can obtain this value.

(Refer Slide Time: 35:29)



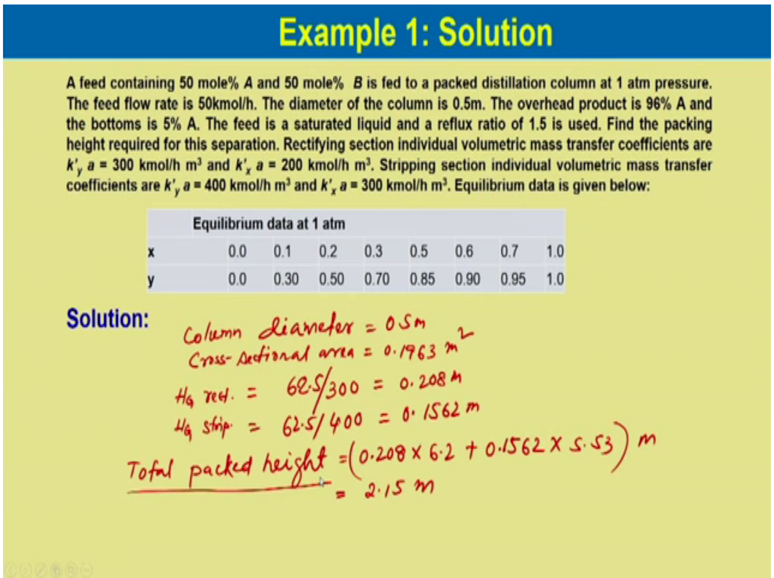
Now, for the stripping section, we can obtain the slope of the line by minus k x dash a divided by minus k y dash a which is equal to minus 300 divided by 400 which is equal to minus 0.75. So, similar to the rectifying section for the stripping section we can also from the operating line we can draw several lines with this slope minus 0.75 and we can obtain this table. So, this is y i we can get from the equilibrium intersection between this line and the equilibrium line. So, at different points from the stripping section we can obtain the values of y i. So, then we can also similar way we can calculate y by y i minus y. So, this is for the stripping section.

(Refer Slide Time: 36:56)



If we plot both the sections whatever both tables which we have obtained earlier like this table for the stripping section and this table is for the rectifying section, if we take this data and plot y by y i minus y versus y so we will obtain this curve. Now the integral for number of transfer unit can be obtained by graphically. So, the area under the curve in the rectifying section, we can obtain from here this is for the rectifying section about 6.2 unit and similarly from this point we have about stripping section is about 5.53 unit. So, once we obtain this then we can calculate the height of the tower as we know the column diameter is 0.5 meter, so we can calculate the cross sectional area of the tower is about 0.1963 meter square.

(Refer Slide Time: 38:33)



Now H G in the rectifying section is equal to 62.5 divided by 300 which is about 0.208 meter and stripping section height of transfer unit in the stripping section is about 62.5 divided by 400 so which is about 0.1562 meter.

Now, the total packed height would be equal to the sum of the height of the two sections. So, that would be 0.208 H tG into N N tG that is 6.2 plus 0.1562 into 5.53 unit. So, this much meter is the height of the tower and it would give about 2.15 meter. So, this way we can solve a packed distillation column problems for a particular separation we can obtain the height of the tower. So, thank you for your patience hearing and we will continue our discussion on distillation in the next class.