Process Design Decisions and Project Economics
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Module - 4
Design of Separation Processes
Lecture - 20
Design of Distillation Columns - Part II
(Plate and Packed Towers, Number of Plates, Diameter and Height of the Column)

Separation Processes, in the previous 2 lectures, we got introduce to the basic aspects of design of the separation process module or separation process systems of a chemical process. And there after we had some discussion on the design aspects of distillation column, which is the most popular or most frequently used separation process in most of the process industries.

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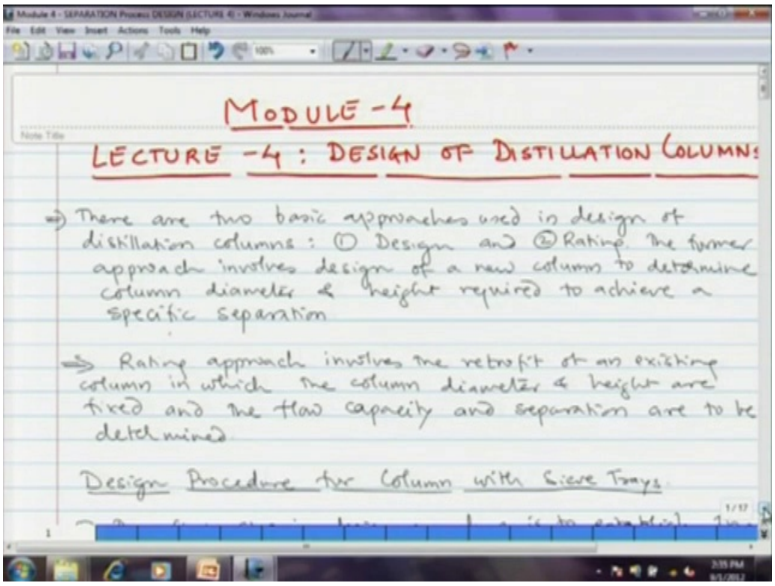

MODULE - 4

LECTURE - 4 : DESIGN OF DISTILLATION COLUMNS

⇒ There are two basic approaches used in design of distillation columns : ① Design and ② Rating. The former approach involves design of a new column to determine column diameter & height required to achieve a specific separation

⇒ Rating approach involves the retrofit of an existing column in which the column diameter & height are fixed and the flow capacity and separation are to be determined

Design Procedure for Column with Sieve Trays

In this lecture we shall go ahead and see the design of distillation column, the aspects of exact design of distillation column. There are 2 basic approaches used in design of distillation column; first is the design, and second is the rating. The former approach involves design of a new column to determine the column diameter and height required to achieve a specific separation, while the rating approach involves retrofit of an existing column in which the column diameter and height are already fixed. And the flow capacity and separation are to be determined, means fitting of an existing column for another purpose other than what it was designed for.

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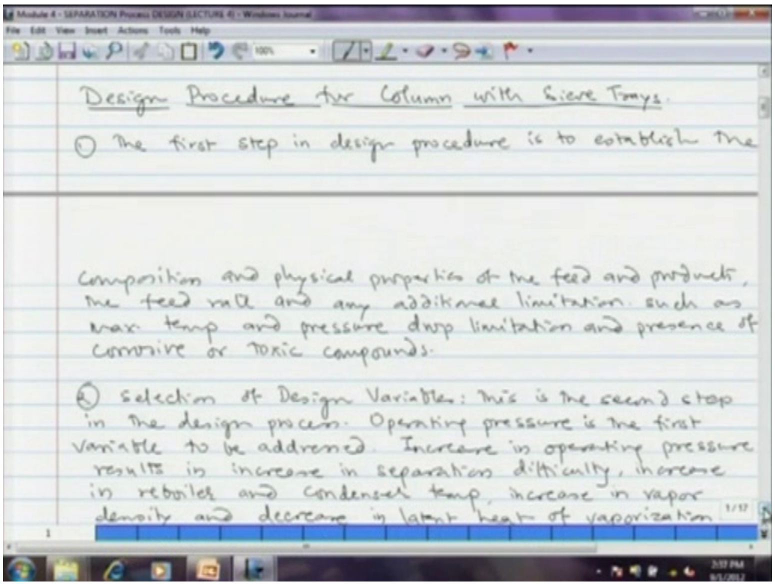

Design Procedure for Column with Sieve Trays.

① The first step in design procedure is to establish the

composition and physical properties of the feed and products, the feed rate and any additional limitation such as max. temp and pressure drop limitation and presence of corrosive or toxic compounds.

② Selection of Design Variables: This is the second step in the design process. Operating pressure is the first variable to be addressed. Increase in operating pressure results in increase in separation difficulty, increase in reboiler and condenser temp, increase in vapor density and decrease in latent heat of vaporization

There are 2 types of distillation columns - first is the pack distillation column, and second is plate distillation column. You have learnt the basic aspects of these columns in the NPTEL course of mass transfer. Today we shall see some practical aspects of these columns and also some practical design procedures then what you have learnt in mass transfer operation course of NPTEL.

Let us start with the design procedure of a column with sieve trays, sieve trays are the most popular columns, column internals that are used. The first step in the design procedure is to establish the composition and physical properties of the feed and the products that is the overall mass balance across a distillation column. The feed rate and any additional limitation also has to be determined limitation could be such as the maximum temperature and pressure drop limitation or presence of some corrosive or toxic component in the feed which has to be removed in priority to others mass transfer operations course of NPTEL.

You have mostly dealt with binary mixtures, but in industry you will rarely find a binary mixture to be distilled. The mixtures that are encountered in industry are multi component mixture, mixtures with more than 5 6 or may be even 20 30 components. So, we need a different approach although the basic principles remains the same, we need a different approach for design of columns for this multi component separation and that is what is going to be today’s discussion.

The second aspect of or the second step in the design of column with sieve trays is that of the selection of design variable. Operating pressure is the first variable to be addressed; in the previous lecture I showed you some aspects of design of distillation columns in which operation operating pressure variation can give flexibility in terms of energy integration. That thermally couple columns and also those things we have seen in previous lecture. Increase in operating pressure results in increase in the separation difficulty, because alpha the relative volatility decreases.

Increase in the re-boiler and condenser temperature, increase in the vapour density and decrease in the latent heat of vaporisation. The lower limit is set by the desire to avoid vacuum operation and the use of external refrigeration for condenser to avoid raise in the operating cost. As I told you in previous lecture if you increase the pressure of a column then re-boiler temperature will increases, condenser temperature increases. And if you reduce the pressure of a column, then both of these temperatures decrease.

Now we want to set an optimum pressure in such a way that we can use cooling water which is a cheapest utility for operating the condenser of the distillation column. And we can use either a medium or high pressure steam for operating the re-boiler, if the process constraints permit and the pressure is, so set that the cooling water and air cooling can be used in condenser.

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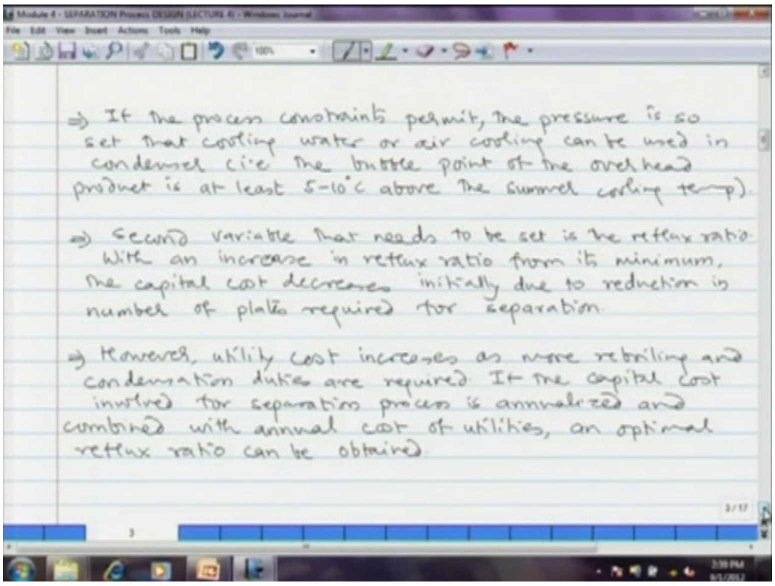

⇒ If the process constraints permit, the pressure is so set that cooling water or air cooling can be used in condenser (i.e the bubble point of the overhead product is at least 5-10°C above the summer cooling temp).

⇒ Second variable that needs to be set is the reflux ratio. With an increase in reflux ratio from its minimum, the capital cost decreases initially due to reduction in number of plates required for separation.

⇒ However, utility cost increases as more reboiling and condensation duties are required. If the capital cost involved for separation process is annualized and combined with annual cost of utilities, an optimal reflux ratio can be obtained.

That is the bubble point of the overhead product is at least 5 to 10 degrees above the summer cooling temperature. The second variable that needs to set is the reflex ratio, you have seen the variation of the column plates with reflex ratio in the course of mass transfer. You know that with increase in the reflex ratio permits minimum the capital cost will reduce because the number of plates will reduce.

However, the utility cost will go up, because more re-boiling will be required more condensation will be required. And if the capital cost involved in the separation processes annualised and combined with the annual cost of utilities, you will always find that the total cost passes through a minimum against a reflex ratio.

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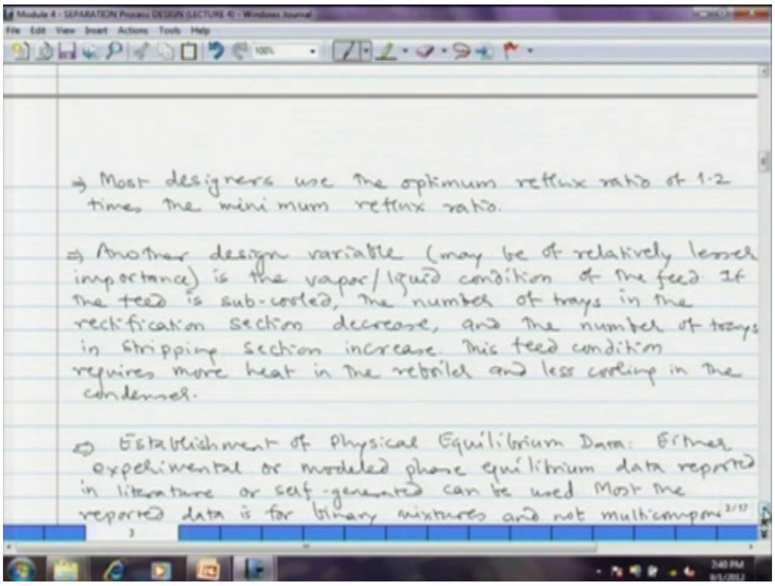

⇒ Most designers use the optimum reflux ratio of 1.2 times the minimum reflux ratio.

⇒ Another design variable (may be of relatively lesser importance) is the vapor/liquid condition of the feed. If the feed is sub-cooled, the number of trays in the rectification section decrease, and the number of trays in stripping section increase. This feed condition requires more heat in the reboiler and less cooling in the condenser.

⇒ Establishment of Physical Equilibrium Data: Either experimental or modeled phase equilibrium data reported in literature or self-generated can be used. Most the reported data is for binary mixtures and not multicompon

Most designers use the optimum reflex ratio as 1.2 times the minimum reflex ratio, this is a thumb rule of distillation column design.

Another design variable which may be of relatively lesser importance is the vapour liquid condition of the feed, again in mass transfer operations course you have learnt this particular aspect of distillation. In that course you have defined a Q line which is quality line and Q represents the quality of the feed. For Q equal to 1 the feed is a saturated liquid and for Q equal to 0 the feed is saturated vapour, but for other than 0 and 1 values if the feed is sub cooled, then the number of trays in the rectification section decrease.

And the number of trays in the stripping section increase and this feed condition requires more heat in the re-boiler and less cooling in the condenser. And if the feed is separated vapour, then reverse happens establishment of the physical equilibrium data is the next step in the distillation column design. Equally physical equilibrium data means basically the really data vapour liquid equilibrium data, either experimental or moderate phase equilibrium data reported in the literature or self-generated data can be used.

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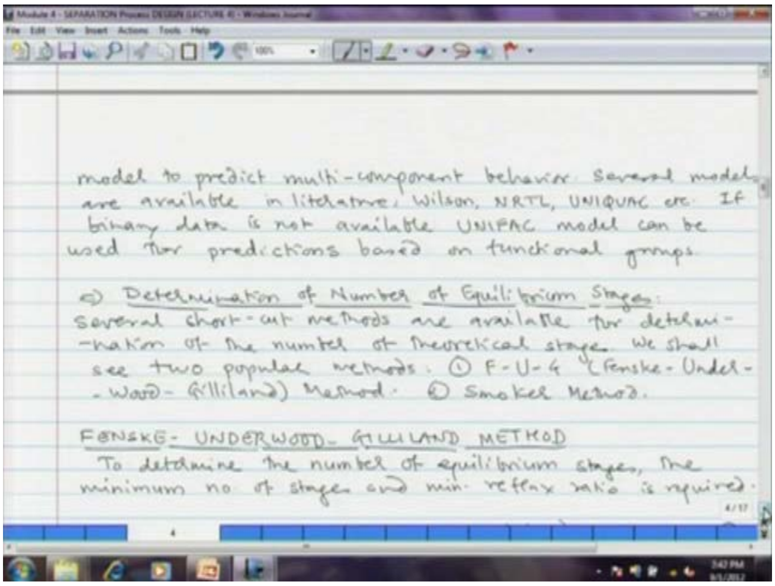

model to predict multi-component behavior. Several models are available in literature: Wilson, NRTL, UNIQUAC etc. If binary data is not available UNIFAC model can be used for predictions based on functional groups.

c) Determination of Number of Equilibrium Stages:
Several short-cut methods are available for determination of the number of theoretical stage. We shall see two popular methods. ① F-U-G (Fenske-Underwood-Gilliland) Method. ② Smoker Method.

FENSKE- UNDERWOOD- GILLILAND METHOD
To determine the number of equilibrium stages, the minimum no of stages and min. reflux ratio is required.

Now a days there is tremendous data base available in the literature, so for practically any mixture you will have the data already reported in literature. You have to look through proper sources, but most of the data that is reported in the journals and handbooks is for binary mixture and not for a multi component mixture. And therefore, it is necessary to combine this data with a model to predict the multi component behaviour. Several models are available in the lecture like, Wilson model, N R T L model, uniquack model, universal quasi chemical and unifac model which is the most vigorous approach.

If the binary data is not available, you can use this model for prediction based on the functional groups that are present in the compounds to be separated. Then comes the aspect of determination of number of equilibrium stages, in mass transfer course of NPTEL. You have learnt the mack up kelly method or punch on separate method for determination of the theoretical number of equilibrium stages. But, this is mostly for binary mixtures how do we use the same method for a multi component mixture. Of course, the approach has to be changed the it has to be more rigorous.

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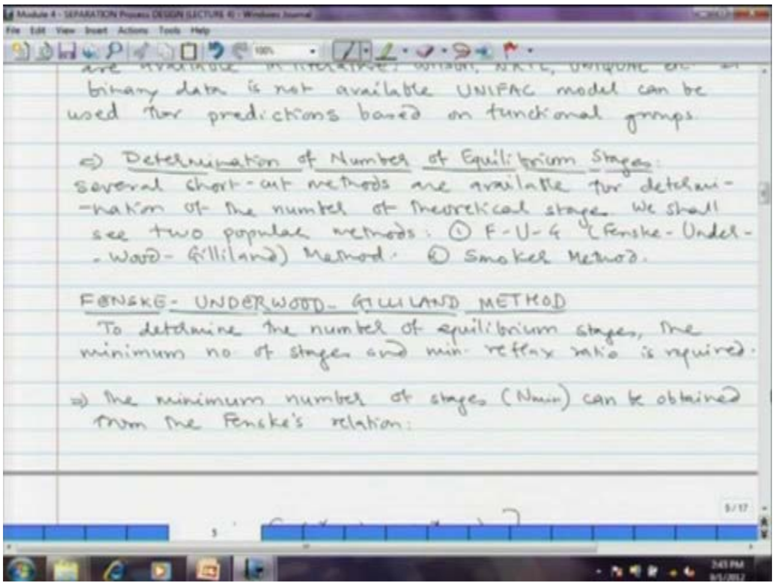
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⇒ Determination of Number of Equilibrium Stages:
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FENSKE- UNDERWOOD- GILLILAND METHOD
To determine the number of equilibrium stages, the minimum no of stages and min. reflux ratio is required.

⇒ The minimum number of stages (N_{min}) can be obtained from the Fenske's relation:

And there are several short cut methods that are reported in literature, for determination of number of theoretical stages for a separation of multi component mixture. We shall see in today's lecture 2 popular methods that are used for this purpose. The first method is F-U-G method Fenske Underwood Gilliland method and second method is the smoker method which is essentially the analytical approach to the mack up kelly method which is a graphical solution. Let us first see the Fenske Underwood Gilliland method, to determine the number of equilibrium stages the minimum number of stages and minimum reflex ratio is required.

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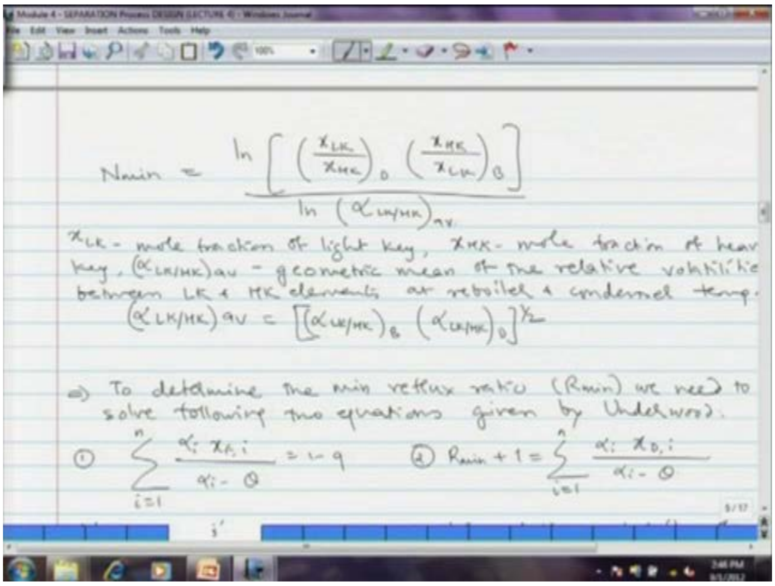
$$N_{min} = \frac{\ln\left[\left(\frac{x_{LK}}{x_{HK}}\right)_D \left(\frac{x_{HK}}{x_{LK}}\right)_B\right]}{\ln\left(\alpha_{LK/HK}\right)_{av}}$$

x_{LK} - mole fraction of light key, x_{HK} - mole fraction of heav key, $(\alpha_{LK/HK})_{av}$ - geometric mean of the relative volatilitie between LK & HK elements at reboiler & condenser temp.

$$(\alpha_{LK/HK})_{av} = \left[(\alpha_{LK/HK})_B \, (\alpha_{LK/HK})_D\right]^{1/2}$$

⇒ To determine the min reflux ratio (R_{min}) we need to solve following two equations given by Underwood.

① $\sum_{i=1}^{n} \frac{\alpha_i x_{F,i}}{\alpha_i - \theta} = 1 - q$ ② $R_{min} + 1 = \sum_{i=1}^{n} \frac{\alpha_i x_{D,i}}{\alpha_i - \theta}$

The minimum number of stages and minimum can be obtained from Fenske’s equation. N minimum is equal to l n into x L K divided by x H K distillate into x H K by x L K bottoms divided by l n the log of alpha relative volatility between light key that is L K and heavy key H K component. Exactly, same expression we have derived in mass transfer operations course for binary mixture, in that case you had define x D and x B where, you had only 1, when you had 2 components.

One more volatile and other less volatile, then x D by x W the mole fraction of more volatile component in distillate divided by x W, the composition of more volatile component in bottoms. Exactly same thing is here, but we are now defining L K and H K components which is light key component and heavy key components. This is for multi component mixture, suppose we have a mixture of 5 6 or 10 components.

Then we mark one component as light key component which is among the lower boiling range and 1 component as heavy key component which is in the in the higher boiling range. And this expression that appears on the screen is exact analogy of the binary expression that we have learnt in mass transfer operation for multi component mixture. Now, here in the denominator we have used the average value of alpha relative volatility, now why because in case of multi component mixture the relative volatility between components.

May sometimes change with temperature, it may also change sometimes with the composition as the separation proceeds. And therefore, we have to use an average value average of that minima and maxima of the relative volatility. So, usually the volatility is not mathematically or numerically algebraically averaged, but it is the geometric mean that is taken that is alpha L K by L K H K average is alpha L K H K at bottoms.

That is relative volatility between light key and heavy key component in the bottom, at the highest temperature and alpha L K H K at distillate that is the volatility at distillate which is lowest temperature in the column, there is to half this is the geometric mean of the 2 volatilities. After we have determined the minimum number of pages then we have to determine the minimum reflex ratio, for that purpose we have to solve 2 equations which are given by under wood.

The first equation is summation i running from 1 to n where, n is the number of components in the mixture. alpha i into x F i, alpha i is the relative volatility of any component i with respect to heavy key component. Except i is the composition or mole fraction of that component i in the feed alpha i into x f i divided by alpha i minus theta is equal to 1minus Q. This Q is again the quality factor that we have seen earlier, Q equal to 1 implies feed being at saturated liquid condition, Q equal to 0 implies feed at saturated vapour condition.

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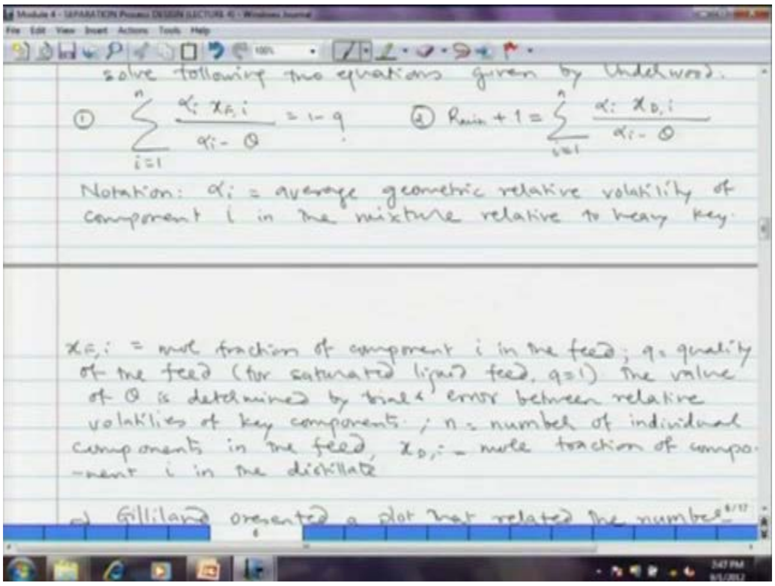

solve following two equations given by Underwood.

① $$\sum_{i=1}^{n} \frac{\alpha_i x_{F,i}}{\alpha_i - \Theta} = 1 - q$$

② $$R_{min} + 1 = \sum_{i=1}^{n} \frac{\alpha_i x_{D,i}}{\alpha_i - \Theta}$$

Notation: α_i = average geometric relative volatility of component i in the mixture relative to heavy key.

$x_{F,i}$ = mole fraction of component i in the feed; q = quality of the feed (for saturated liquid feed, q=1) The value of Θ is determined by trial & error between relative volatilies of key components; n = number of individual components in the feed, $x_{D,i}$ = mole fraction of component i in the distillate

c) Gilliland presented a plot that related the number

Theta is a intermediate is the basically mathematical parameter, which has to be determined. Now if you have a multi component mixture and if you write down this summation and simplify, you will get a polynomial. And roots of those polynomials are the values of theta, you will get n minus 1 values of theta for n components, but you have to take only that value of theta. For the second equation which is r minimum plus 1 is equal to summation i running from 1 to n alpha i into x D i.

Now, x D i is the mole fraction of component i in the distillate divided by alpha i minus theta. So, choose solve the first equation get all the values of theta and take only that value of theta which is between the relative volatility of light key and heavy key component. And then use that value of theta in the next expression and solve for r minimum. x B i is the mole fraction of component i in the distillate and that can be determined from the minimum number of plates.

Distillate in bottom distribution of the non key components in the feed can be evaluated after calculation of minimum number of stages, but minimum number of stages again are determined by Fenske’s equation. So, the simple relation is x D i divided by x B i is equal to alpha i raise to n minimum alpha i average raise to n minimum into x H K in the distillate and x H K in the bottom. That is the simplest very simple relation between the composition of component in distillate and composition of the same component in the bottoms.

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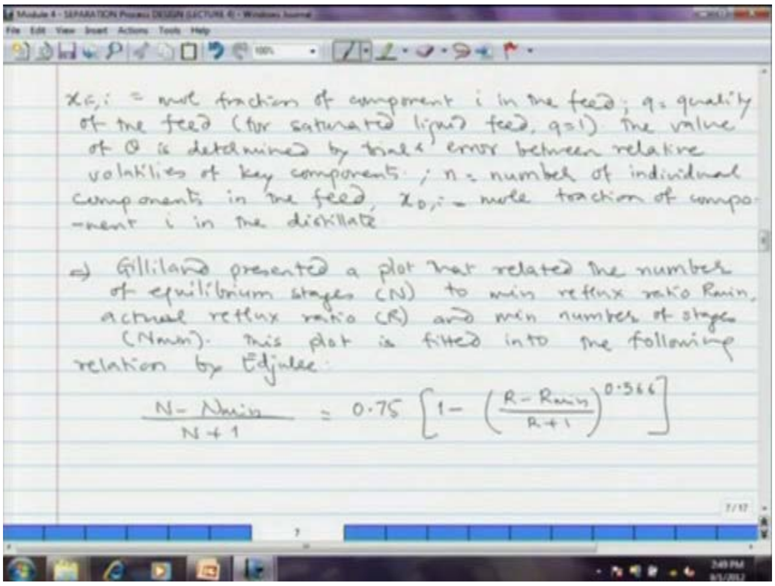

$x_{F,i}$ = mole fraction of component i in the feed; q = quality of the feed (for saturated liquid feed, q=1) the value of θ is determined by trial & error between relative volatilies of key components; n = number of individual components in the feed, $x_{D,i}$ = mole fraction of component i in the distillate

⇒ Gilliland presented a plot that related the number of equilibrium stages (N) to min reflux ratio Rmin, actual reflux ratio (R) and min number of stages (Nmin). This plot is fitted into the following relation by Edjulee:

$$\frac{N - N_{min}}{N + 1} = 0.75 \left[1 - \left(\frac{R - R_{min}}{R + 1} \right)^{0.566} \right]$$

Then after determination of r minimum and n minimum, we have to calculate the actual number of plates, but now remember actual number of theoretical plates not the actual number of actual plates or practical plates. Gilliland has presented a plot that related the quantity n minus n minimum divided by n plus 1 to the quantity r minus r minimum divided by r plus 1.

And a relation is later on transformed into an expression as a correlation, essentially by edjulee and the correlation that is n minus n minimum divided by n plus 1 is equal to 0.75 into 1 minus r minus r minimum divided by r plus 1 raise to 0.566. So, after determination of n minimum from Fenske’s equation after determination of r minimum from Underwood’s equations, which you can set as r equal to 1.2 times r minimum and then determine the number of plates n. So, this is a very simple procedure, although it is implicit because you have to solve 2 polynomial equations to get the roots which are essentially values of theta, but you can easily program this and then get the number of plates that are required for a particular separation.

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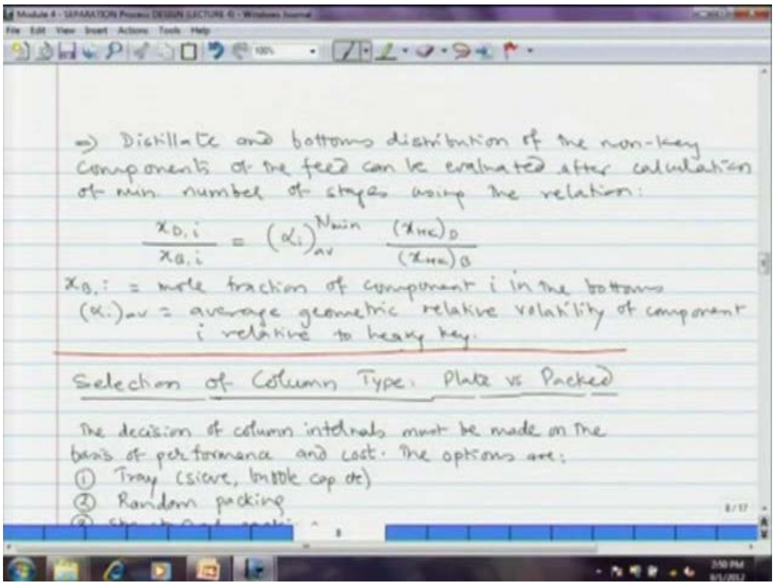

⇒ Distillate and bottoms distribution of the non-key components of the feed can be evaluated after calculation of min number of stages using the relation:

$$\frac{x_{D,i}}{x_{B,i}} = (\alpha_i)_{av}^{N_{min}} \frac{(x_{HK})_D}{(x_{HK})_B}$$

$x_{B,i}$ = mole fraction of component i in the bottoms

$(\alpha_i)_{av}$ = average geometric relative volatility of component i relative to heavy key.

Selection of Column Type: Plate vs Packed

The decision of column internals must be made on the basis of performance and cost. The options are:

① Tray (sieve, bubble cap etc)

② Random packing

We shall see another approach to determination of number of the equilibrium plates.

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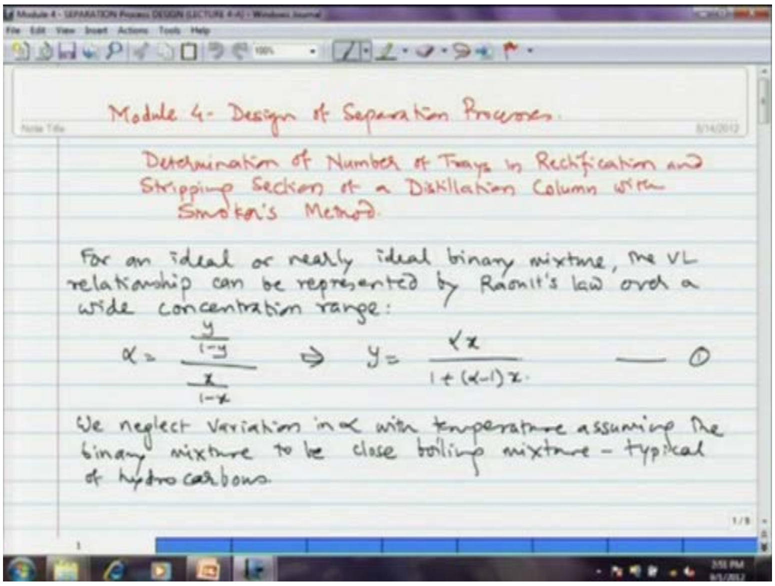

Module 4- Design of Separation Processes.

Determination of Number of Trays in Rectification and Stripping Section of a Distillation Column with Smoker's Method.

For an ideal or nearly ideal binary mixture, the VL relationship can be represented by Raoult's law over a wide concentration range:

$$\alpha = \frac{\frac{y}{1-y}}{\frac{x}{1-x}} \Rightarrow y = \frac{\alpha x}{1+(\alpha-1)x} \quad \text{—— ①}$$

We neglect variation in α with temperature assuming the binary mixture to be close boiling mixture – typical of hydrocarbons.

This is essentially analytical solution of method that you have seen in mass transfer operations course of NPTEL. This solution was given by smoker a scientist in US. And we shall see now, the derivation of this method that is the number of trays in the rectification stripping section of a distillation column. For an ideal or nearly ideal binary mixture the vapour liquid relationship can be represented by Raoult's law over a wide range of concentration range.

This you have seen in a mass transfer operation course alpha is equal to y divided by 1 minus y divided by x Divided by 1 minus x. And this can be rearranged as y is equal to alpha x Divided by 1 plus alpha minus 1 into x. Now we had talked of variation in alpha sometimes I because we have we had taken in earlier expressions average value of alpha, but in many situations when the molecules are not very chemically different or when the temperature variation across the column is not very high. In that case we can neglect the variation in alpha with temperature assuming that the binary mixture to be a close boiling mixture of typically of hydro carbons or you can use value of average alpha also that is also possible.

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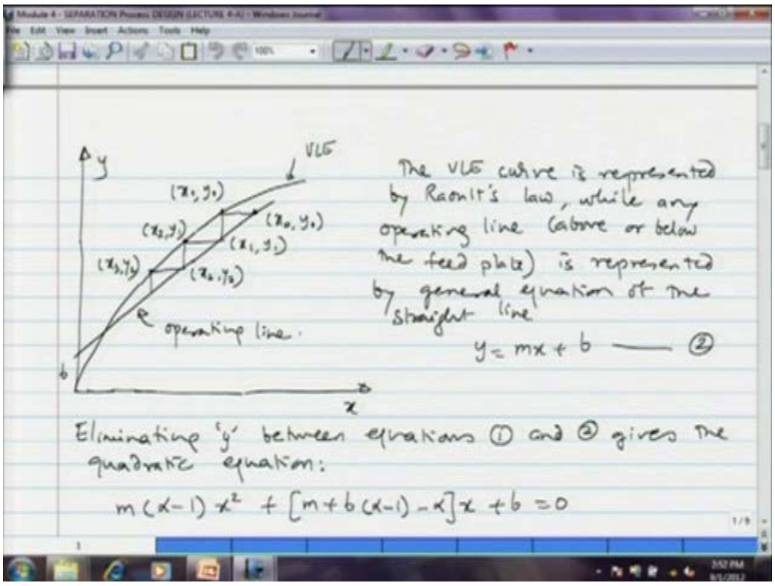

The VLE curve is represented by Raoult's law, while any operating line (above or below the feed plate) is represented by general equation of the straight line

$$y = mx + b \quad \text{——} \quad ②$$

Eliminating 'y' between equations ① and ② gives the quadratic equation:

$$m(\alpha-1)x^2 + [m + b(\alpha-1) - \alpha]x + b = 0$$

Now, the V L E curve given by the Raoult’s law is represented as the diagram now appears on your screen. The V L E curve is essentially a parabolic curve and then we have to plot the 2 lines operating, 2 operating lines. One for the rectification section and another for the stripping section. Rectification operating line is above the feed plate and stripping line is below the feed plate. Now let us say that we give a general expression for this operating line as y is equal to m x plus b, and then we plot specific points on operating line and the corresponding points on the vapour liquid equilibrium curve x 0, y 0.

As you see on the screen is the point on operating line and x 1, y 0 is the corresponding point on the V L E curve, then x 1 y 1 is the next point on the operating line. x 2 y 1 is the corresponding point on V L E curve and so on, and so forth. Now if we eliminate y between equation 1 and 2 equation 1 being the Raoult's law and equation 2 being the general equation of a line y equal to m x plus b. Then you get a quadratic expression in x, that is m into alpha minus 1 x square plus m plus b into alpha minus 1 minus alpha into x plus b equal to 0.

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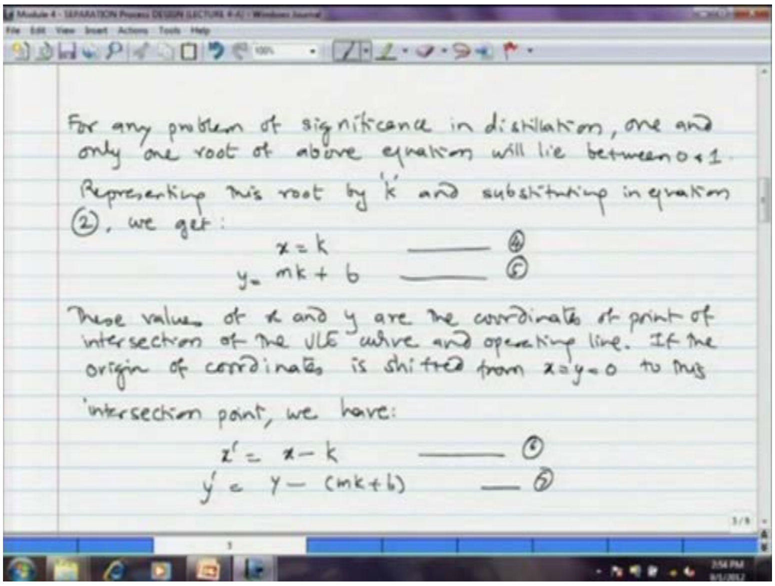

For any problem of significance in distillation, one and only one root of above equation will lie between 0 + 1.

Representing this root by 'k' and substituting in equation ②, we get:

$$x = k \quad \text{———} \quad ④$$

$$y = mk + b \quad \text{———} \quad ⑤$$

These values of x and y are the coordinates of point of intersection of the VLE curve and operating line. If the origin of coordinates is shifted from x = y = 0 to this intersection point, we have:

$$x' = x - k \quad \text{———} \quad ⑥$$

$$y' = y - (mk + b) \quad \text{———} \quad ⑦$$

Now, you can very easily work out this, so I leave this as exercise, this is equation 3. Quadratic equation will yield in 2 values of x for any problem of significance in distillation 1 and only 1 root of the above equation will lie between 0 and 1. Since, x is the mole fraction it is value has to bounce 0 and 1. If the number is negative below 0 then that is absolute, if the number is above 1 then also it is absolute. We represent the root between 0 and 1 as k for the for that quadratic equation x equal to k and then y corresponding y becomes m k plus b. And these values of x and y are the coordinates of the point of intersection of the V L E curve.

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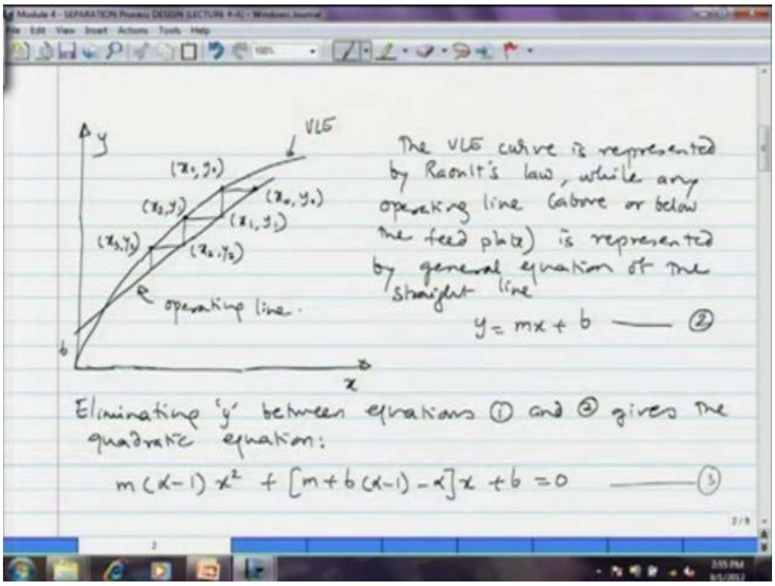

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Eliminating 'y' between equations ① and ② gives the quadratic equation:

$$m(\alpha - 1)x^2 + [m + b(\alpha - 1) - \alpha]x + b = 0 \quad \text{——} \quad ③$$

And the operating line which is essentially this point which now appears on the screen this point. If we shift the origin of coordinates from x equal to y equal to 0 to this point. This point of intersection between the operating line and the V L E curve, then we are exactly addressing the rectification section. We are no more taken into consideration this stripping section. And then the coordinates of that point that shifted origin are x Dash equal to x minus k and y dash equal to y minus m k plus b.

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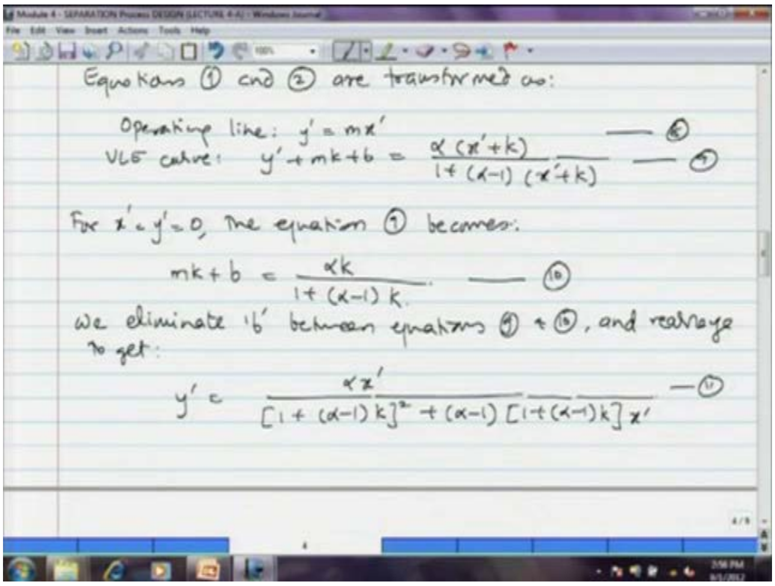

Equations ① and ② are transformed as:

Operating line: $y' = mx'$ —— ⑧

VLE curve: $y' + mk + b = \dfrac{\alpha(x' + k)}{1 + (\alpha - 1)(x' + k)}$ —— ⑨

For $x' = y' = 0$, the equation ⑨ becomes:

$$mk + b = \frac{\alpha k}{1 + (\alpha - 1)k} \quad \text{——} \quad ⑩$$

We eliminate 'b' between equations ⑨ + ⑩, and rearrange to get:

$$y' = \frac{\alpha x'}{[1 + (\alpha - 1)k]^2 + (\alpha - 1)[1 + (\alpha - 1)k]x'} \quad \text{——} \quad ⑪$$

And if you substitute these x x Dash and y dash, the new coordinates of origin into the equation of operating line and Raoult’s law equation. Then operating line equation becomes y dash is equal to m x Dash V L E curve becomes, y dash is equal y dash plus m k plus b is equal to alpha into x Dash plus k divided by 1 plus alpha minus 1 into x Dash plus k. And for x Dash y dash equal to 0, when we take when we assume that particular a point of intersection as the 0 0 point origin point.

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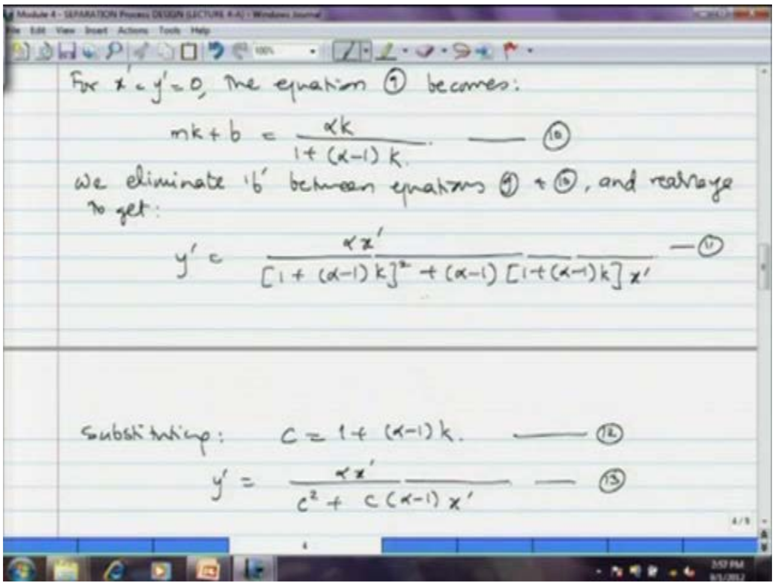

For $x' = y' = 0$, the equation ⑨ becomes:

$$mk + b = \frac{\alpha k}{1 + (\alpha - 1)k} \quad — ⑩$$

We eliminate 'b' between equations ⑨ + ⑩, and rearrange to get:

$$y' = \frac{\alpha x'}{[1 + (\alpha - 1)k]^2 + (\alpha - 1)[1 + (\alpha - 1)k]x'} \quad — ⑪$$

Substituting: $c = 1 + (\alpha - 1)k$ — ⑫

$$y' = \frac{\alpha x'}{c^2 + c(\alpha - 1)x'} \quad — ⑬$$

Then equation 9 which is the V L E curve equation becomes m k plus b is equal to alpha k divided by 1 plus alpha minus 1 into k. And if we eliminate b from equation 9 and 10 V L E curve. And the V L E curve when x Dash y dash equal to 0, then we get y dash is equal to alpha x Dash divided by 1 plus alpha minus 1 k square plus alpha minus 1 into 1plus alpha minus 1 k into x Dash.

Eliminating b that means we are not considering the intercept any more of the. So, that we are considering just the V L E curve and the operating line which ends at origin. So, b is 0 is it, we substitute a new variable c for 1 plus alpha minus 1 into k, which is appearing 2 times in the denominator. So, that the expression becomes handy easy to write and then y dash is equal to alpha into x Dash divided by c square plus c into alpha minus 1 into x Dash.

This is our equation 13, equation 13 represents the Raoult’s law expressed in coordinates, determined by the intersection of the original Raoult’s law equation and any operating line. We may this expression equation 13 explicit in terms of x Dash, so just simple rearrangement x dash is equal to c square into y dash divided by alpha minus c into alpha minus 1 into y dash, x 0 x 0 dash y 0 dash in the figure that I had shown initially, represented any point on the operating line y dash is equal to m x Dash.

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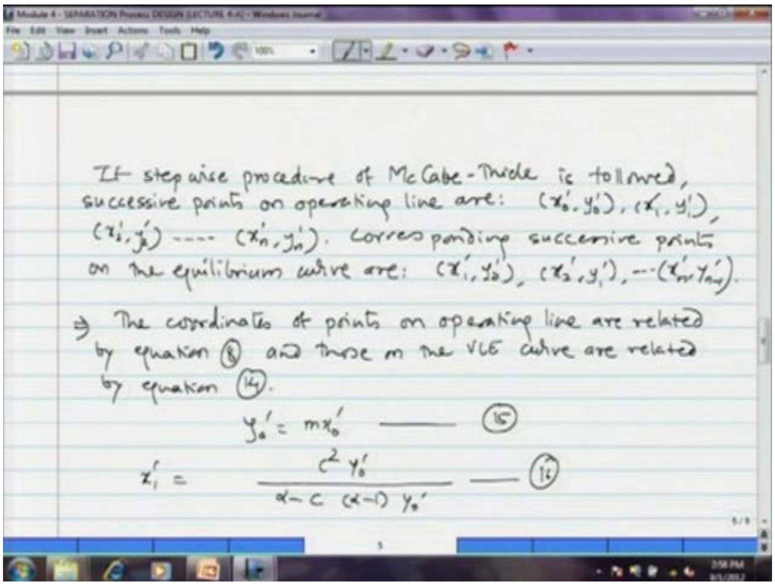

If step wise procedure of McCabe-Thiele is followed, successive points on operating line are: (x_0', y_0'), (x_1', y_1'), (x_2', y_2') ---- (x_n', y_n'). Corresponding successive points on the equilibrium curve are: (x_1', y_0'), (x_2', y_1'), --- (x_n', y_{n-1}').

⇒ The coordinates of points on operating line are related by equation ⑧ and those on the VLE curve are related by equation ⑭.

$$y_0' = m x_0' \quad \text{———} \quad (15)$$

$$x_1' = \frac{c^2 y_0'}{\alpha - c(\alpha - 1) y_0'} \quad \text{———} \quad (16)$$

If we step wise start the procedure of McCabe Thiele method; that means, we assume x 0 dash y 0 dash acts the point representing the distillate of the distillation column. Then we get corresponding points as I earlier mentioned that x 0 x 0 dash y 0 dash extended horizontally to a equilibrium curve will give x 1 dash y 0 dash. Then that point projected downwards perpendicular downward to operating line will give x 1 dash y 1 dash.

And then x 1 dash y 1 dash projected horizontally to equilibrium curve will give x 2 dash y 1 dash and so on, and so forth. The coordinates of the point on operating line are related by equation 8 and those on the V L E curve are related by equation 14. Equation 14 is the V L E curve in shifted coordinates and equation 8 was the operating line in terms of shifted coordinates y dash equal to m x Dash, but after setting b equal to 0; that means, after shift of origin. And then follows a stepwise method, first y 0 dash equal to m x 0 dash x 1 dash is equal to c square into y 0 dash divided by alpha minus c into alpha minus 1 y 0 dash.

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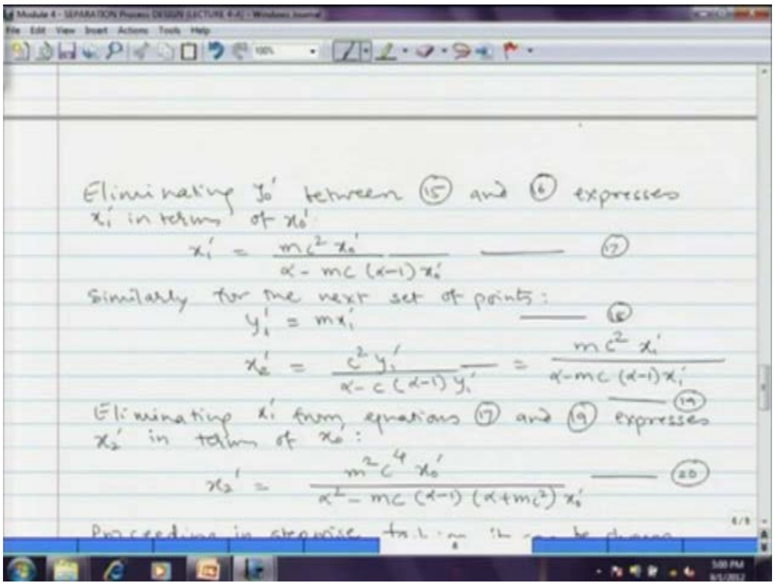

Eliminating y_0' between ⑮ and ⑥ expresses x_1' in terms of x_0'

$$x_1' = \frac{mc^2 x_0'}{\alpha - mc(\alpha-1)x_0'} \quad \text{———} \quad ⑰$$

Similarly for the next set of points:

$$y_1' = mx_1' \quad \text{———} \quad ⑱$$

$$x_2' = \frac{c^2 y_1'}{\alpha - c(\alpha-1)y_1'} = \frac{mc^2 x_1'}{\alpha - mc(\alpha-1)x_1'} \quad \text{———} \quad ⑲$$

Eliminating x_1' from equations ⑰ and ⑲ expresses x_2' in terms of x_0':

$$x_2' = \frac{m^2 c^4 x_0'}{\alpha^2 - mc(\alpha-1)(\alpha+mc^2)x_0'} \quad \text{———} \quad ⑳$$

Proceeding in stepwise

Now eliminating y 0 dash between the above 2 equation 15 and 16 expresses x 1 dash in terms of x 0 dash, x 1 dash is the composition of liquid on the first plate. And x 0 dash is the composition of the liquid which is big obtained as distillate, it is the same as y 1 dash because the same vapour is condensed, but x 1 dash is different. Now, why have we made the calculation in terms of x and not y, because it is very easy to measure the composition of liquid.

For example, if you are doing a simple experiment of benzene toluene distillation, the refractive index of the mixture changes with composition. And with simple measurement of just refracting index you can measure, you can determine the composition of the liquid. And that is why we have made all our calculation explicit in terms of x and not y. Determination of composition of vapour requires special techniques such as gas chromatograph or liquid chromatography.

Basically, gas chromatograph and then that is more tedious, but measurement of the composition of liquid is much easier, so we proceed. Next point y 1 dash is equal to m x 1 dash and then corresponding point on the V L E curve, x 2 dash is equal to c square y 1 dash divided by alpha minus c into alpha minus 1 into y 1 dash. This you can rearrange by writing y 1 dash in terms of x 1 dash. So, m c square x 1 dash divided by alpha minus m c into alpha minus 1 into x 1 dash.

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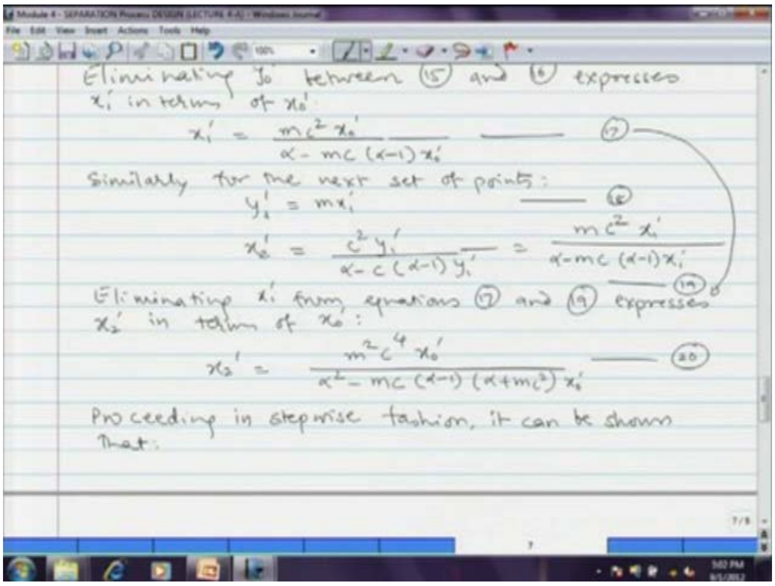

Eliminating y_0 between ⑮ and ⑯ expresses x_1' in terms of x_0'

$$x_1' = \frac{mc^2 x_0'}{\alpha - mc(\alpha-1)x_0'} \quad \text{———} \quad ⑰$$

Similarly for the next set of points:

$$y_1' = mx_1' \quad \text{———} \quad ⑱$$

$$x_2' = \frac{c^2 y_1'}{\alpha - c(\alpha-1)y_1'} = \frac{mc^2 x_1'}{\alpha - mc(\alpha-1)x_1'} \quad \text{———} \quad ⑲$$

Eliminating x_1' from equations ⑰ and ⑲ expresses x_2' in terms of x_0':

$$x_2' = \frac{m^2c^4 x_0'}{\alpha^2 - mc(\alpha-1)(\alpha+mc^2)x_0'} \quad \text{———} \quad ⑳$$

Proceeding in stepwise fashion, it can be shown that:

And now x 1 dash you can write in terms of x 0 dash here, eliminating x 1 dash from equation 17 and 19, expresses x 2 dash in terms of x 0 dash. So, x 2 dash related to m square c raise to 4 x 0 dash divided by alpha square minus m c into alpha minus 1 into alpha plus m c square into x 0 dash.

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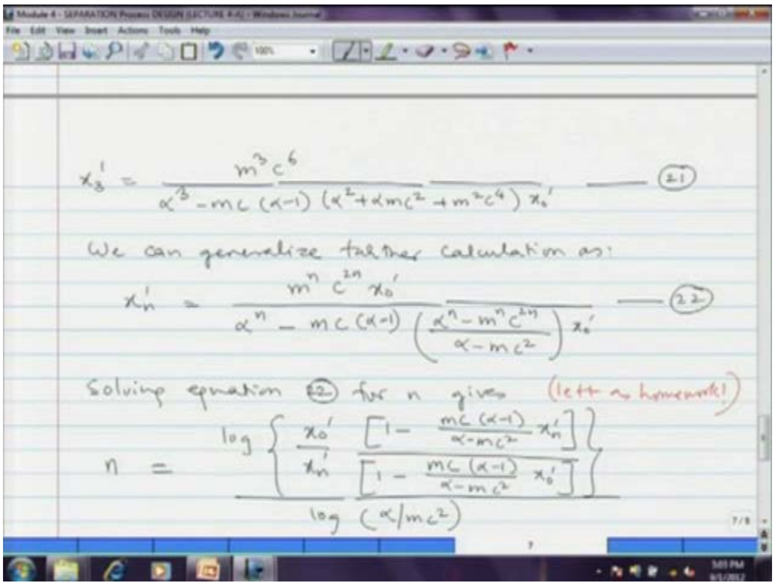

$$x_3' = \frac{m^3c^6}{\alpha^3 - mc(\alpha-1)(\alpha^2+\alpha mc^2+m^2c^4)x_0'} \quad \text{———} \quad ㉑$$

We can generalize further calculation as:

$$x_n' = \frac{m^n c^{2n} x_0'}{\alpha^n - mc(\alpha-1)\left(\frac{\alpha^n - m^n c^{2n}}{\alpha - mc^2}\right)x_0'} \quad \text{———} \quad ㉒$$

Solving equation ㉒ for n gives (left as homework!)

$$n = \frac{\log\left\{\frac{x_0'}{x_n'}\frac{\left[1-\frac{mc(\alpha-1)}{\alpha-mc^2}x_n'\right]}{\left[1-\frac{mc(\alpha-1)}{\alpha-mc^2}x_0'\right]}\right\}}{\log(\alpha/mc^2)}$$

And if you proceed stepwise like this, then you can write expression for x 3 dash. Now this is a big expression, but obtaining it is rather straight forward, we have to first get x 3 dash in terms of y 3 dash and then substitute for y 3 dash. And then substitute for x 3 dash in terms of x 2 dash x 2 dash in terms of x 1 dash like that. So, generalising these stepwise procedure we can obtain a general equation as x n dash is equal to m raise to n c raise to 2 n into x 0 dash divided by alpha raise to n minus m into c into alpha minus 1 into the bracket alpha raise to n minus m raise to n c raise to 2 n divided by alpha minus m c square into x 0 dash.

You can rearrange expression 22 I am leaving this as home work as I have written here, left as homework. So, you can take log on both sides and then rearrange this equation and express it in terms of n. So, n is equal to log x 0 dash divided by x n dash into bracket 1 minus m c alpha minus 1 into x n dash divided by alpha minus 1 alpha minus m c square divided by 1 minus m c alpha minus 1 divided by alpha minus m c square into x 0 dash and the whole numerator divided by log alpha divided by m c square.

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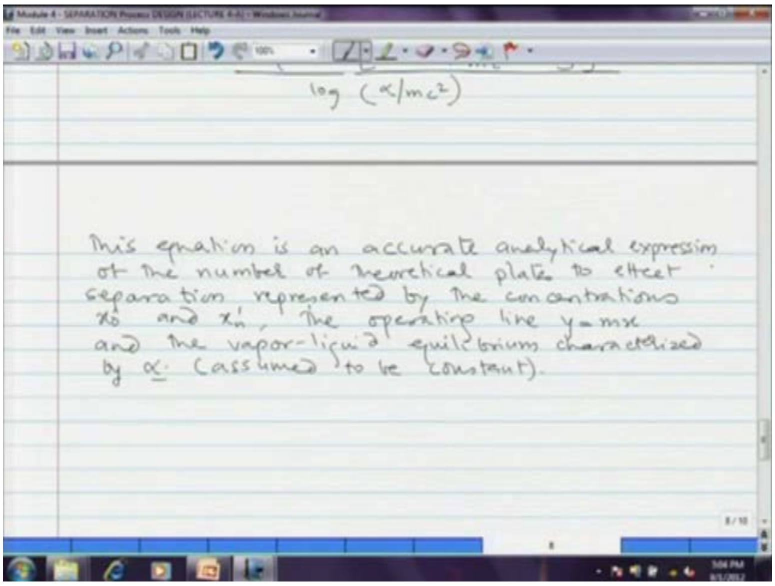

$\log(\alpha/mc^2)$

This equation is an accurate analytical expression
of the number of theoretical plate to effect
separation represented by the concentrations
x_0' and x_n', the operating line $y = mx$
and the vapor-liquid equilibrium characterized
by $\underline{\alpha}$. (assumed to be constant).

This equation is an accurate analytical expression for the number of plates, number of theoretical plates represented by concentrations x 0 dash and x 1 dash, the operating line y equal to m x and a Vapour Liquid Equilibrium curve characterized by alpha. Now what is the beauty of this expression? we have already learnt calculation of or determination of the equation of the operating line in the course of mass transfer operations of an NPTEL.

For example, the equation of the operating line of the rectification section is equal to y is equal to r divided by r plus 1 into x plus x D which is the composition of more volatile component in distillate divided by r plus 1. So, if we want to fit the y is equal to m x plus b kind of expression to this, then m which is the slope of operating line is equal to r divided by r plus 1. And b which is the intercept of the operating line is x D divided by r plus 1 simply by substituting these values here for a given separation. That means, what should be value of x 0 which is essentially x D and x N here remember is the feed plate. This expression was solved for only half of the total Mc Café Thiele diagram. How did we do this by taking the intercept to be 0 because we shifted our origin to the point of intersection of the operating line and the equilibrium curve.

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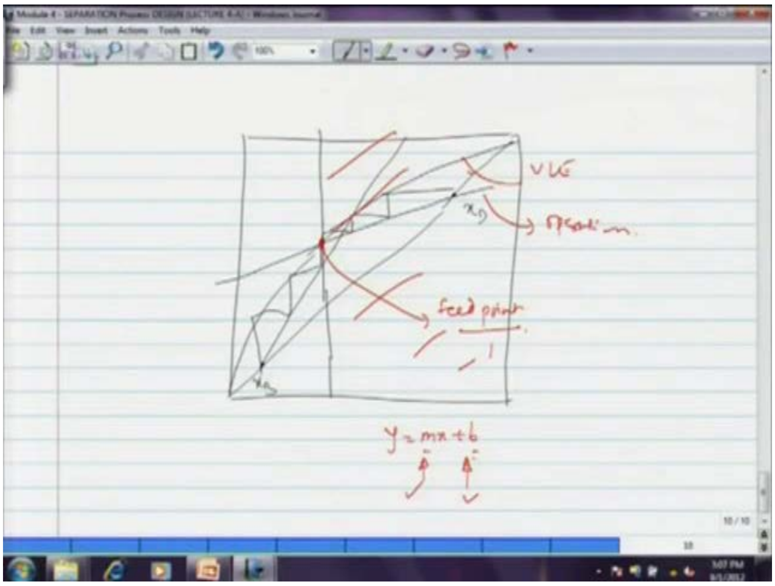

So, in a Mc Café Thiele diagram which typically looks like this is y equal to x line diagonal line this is the V L E curve. And then you start calculations from given x D and given x B and then you make out number of plates. So, we have considered only this portion the portion which I am now sketching in red. Only we have considered only this portion, because we considered the cross section or the point of cross point of intersection of the V L E curve and the operating line.

As our last point as our origin, so we have determined a number of plates in the upper section of the column. This point is essentially the feed point, so we have determined the number of plates in the rectification section. You can apply exactly the same expression in for the lower section of the column the stripping section. However, you have to be careful about values of m and b, here you have learnt the determination of equation of operating line for the stripping section and course of mass transfer.

So, use those values use that m use that b for the stripping section operating line and exactly solve the same procedure and then you will get number of plates. You can very easily code this entire derivation, entire in mat lab or math cad and then that will give you a very handy tool to play with. So, that how does alpha vary how does the number of plates vary with alpha with x B x D. So, this entire algorithm is a very handy tool for optimization of the distillation column.

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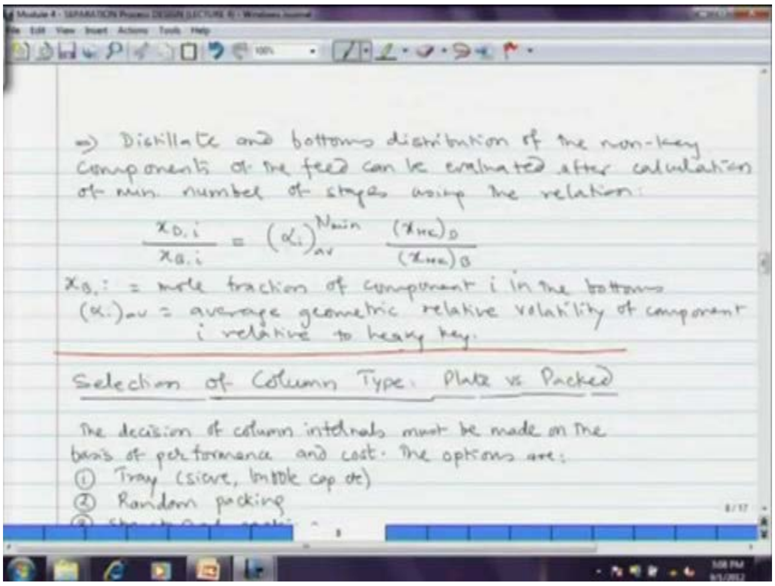

⇒ Distillate and bottoms distribution of the non-key components of the feed can be evaluated after calculation of min. number of stages using the relation:

$$\frac{x_{D,i}}{x_{B,i}} = (\alpha_i)_{av}^{N_{min}} \frac{(x_{HK})_D}{(x_{HK})_B}$$

$x_{B,i}$ = mole fraction of component i in the bottoms

$(\alpha_i)_{av}$ = average geometric relative volatility of component i relative to heavy key.

Selection of Column Type: Plate vs Packed

The decision of column internals must be made on the basis of performance and cost. The options are:

① Tray (sieve, bubble cap etc)

② Random packing

So, this method is given by smoker, this is also quite popular method in determination of number of plates. Now researchers from university of Massachusetts Amherst have further worked on this equation to further simplify it for sharp separation. When x D is close to 1 x D is in the 0.99 or 0.995 all those thing, but you can find these derivations in the book by Douglas conceptual design of chemical processes. Right now I am not covering this because this is outside the scope of this course.

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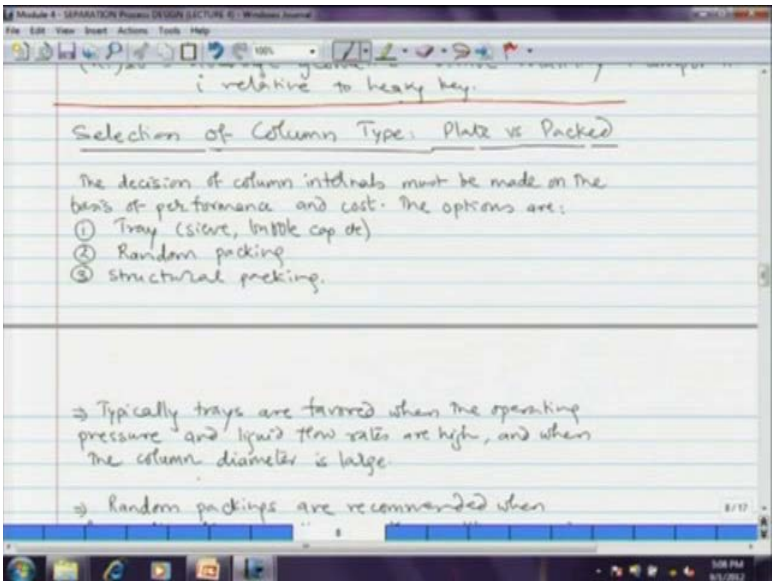

i relative to heavy key.

Selection of Column Type: Plate vs Packed

The decision of column internals must be made on the basis of performance and cost. The options are:

① Tray (sieve, bubble cap etc)
② Random packing
③ structural packing.

⇒ Typically trays are favored when the operating pressure and liquid flow rates are high, and when the column diameter is large.

⇒ Random packings are recommended when

So, now, we go ahead we have now determine, we have seen an important design aspect of distillation column that is in determination of number of equilibrium stages. Actual number of plates you can calculate by dividing the theoretical number of plates by the efficiency that we shall see subsequently. Now, I told you in the beginning that there are 2 types of distillation columns, first is the plate column and another is packed column.

The decision of selection between these columns depends on various factors decision of column internals must be made on the basis of performance and cost. The options for column internal or tray column sieve cap bubble, sieve tray, bubble cap tray. Then among packing there are 2 types of packing 1 is random packing and second is structured packing.

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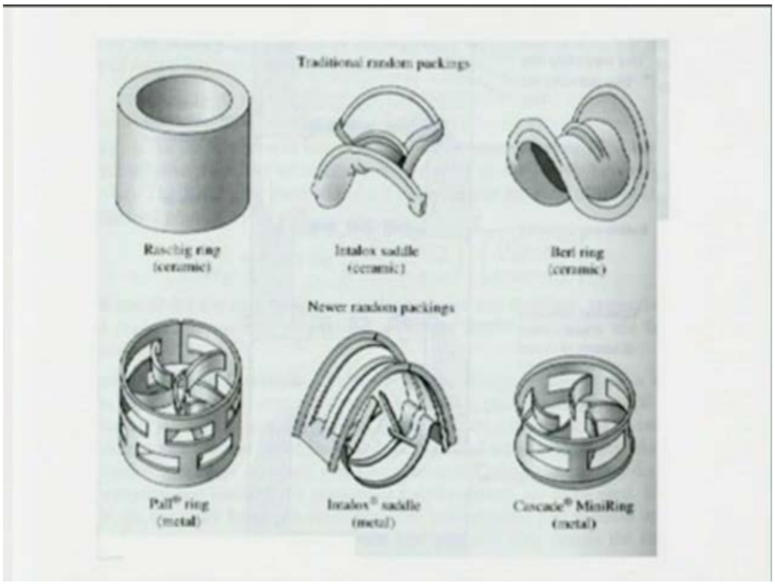

Random packing’s are of various types, I can now I am showing you here on the screen, raschig rings, then interlock saddle raschig rings and interlock saddle made of ceramic. There are 2 factor what is the shape of this tower packing and what is the material of construction of this tower packing? Now in earlier days ceramic was most widely used, because ceramic was very corrosion resistant ceramic essentially an oxide, so it is highly corrosion resistant. So, raschig ring is made of ceramic interlock saddles made of ceramic, burn saddles or burn wheels made of ceramic, then pall rings made of metal interlock’s metal tower packing known as IMTP again made of metal and then cascade Mannering made of metal. So, these are these are some of the common tower packing’s that are used.

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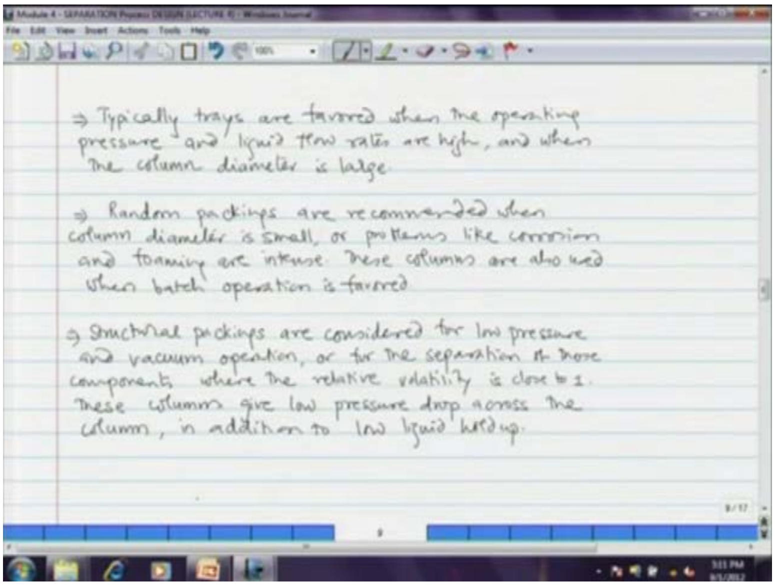

⇒ Typically trays are favored when the operating pressure and liquid flow rates are high, and when the column diameter is large.

⇒ Random packings are recommended when column diameter is small, or problems like corrosion and foaming are intense. These columns are also used when batch operation is favored.

⇒ Structural packings are considered for low pressure and vacuum operation, or for the separation of those components where the relative volatility is close to 1. These columns give low pressure drop across the column, in addition to low liquid hold up.

That tray columns are favoured in the operating pressure and liquid flow rate are high and then the column diameter is large. Random packing’s are recommended when the column diameter is small or problems such as corrosion or foaming are intense. This columns are also used when batch operations are preferred structural packing’s are considered for low pressure and vacuum operation or for separation of those components where, the relative volatility is close to 1.

A typical example in this case is the heavy outlet distillation H 2 O D 2 O where, relative volatility is something around 1.005. You get very large number of plates. So, separating this in plate column is practically impossible. So, the structured packing is used which gives very high interfacial area. And therefore, very less high the if you want to find an analogy between number of tray and the packing there is quantity called as H T P Height equivalent to Theoretical Plate and that we shall see subsequently.

Structural packing column give low pressure across the column in addition to low liquid hold up, typically the tray efficiencies are between 60 to 80 percent for the sieve trays. Typical examples of random packing’s I already shown raschig rings, interlock’s saddle burl, saddle I M P T. Now there is some discussion regarding the material of construction, the popular material of construction for tower packing is either ceramic or metal.

And in the modern days even plastic is being used, but all of these have some advantages and disadvantages. Now let us see those stone ware is susceptible to attack the alkali and acids such as hydro fluoric acid. Metals have issue of vet ability of surface and possibility of high corrosion rates, plastic packing’s cannot be used for high temperature operation because they may decompose. The random packing’s are characterised on the basis of void fraction surface area per volume and the quantity which I justified height equivalent to theoretical plate.

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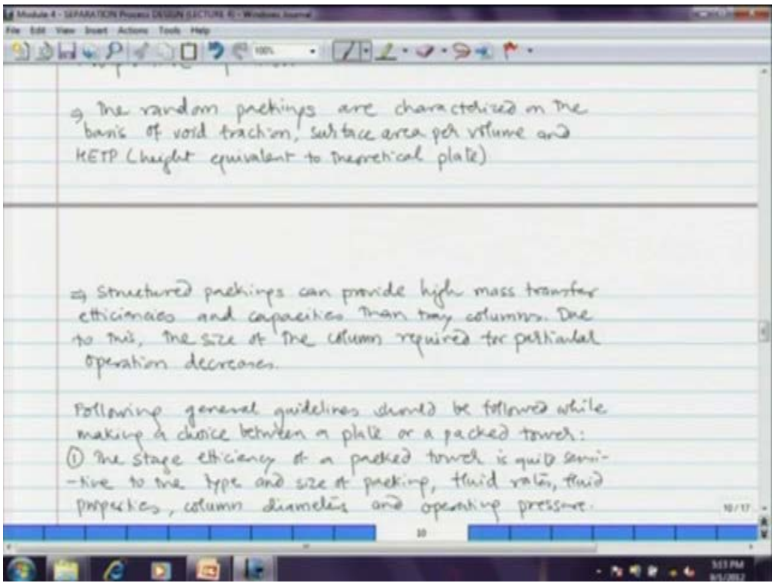

⇒ The random packings are characterized on the basis of void fraction, surface area per volume and HETP (height equivalent to theoretical plate)

⇒ Structured packings can provide high mass transfer efficiencies and capacities than tray columns. Due to this, the size of the column required for particular operation decreases.

Following general guidelines should be followed while making a choice between a plate or a packed tower:

① The stage efficiency of a packed tower is quite sensitive to the type and size of packing, fluid rates, fluid properties, column diameter and operating pressure.

Now, the extent of separation that is achieved in a particular height of the tower and those between 2 plates is the analogy between this 2. If the two plates of a distillation column give like enrichment in vapour as let us say 2 percent. Then how much height of the random packing is required to achieve the same separation is essentially what is H E T P Height Equivalent to Theoretical Plate.

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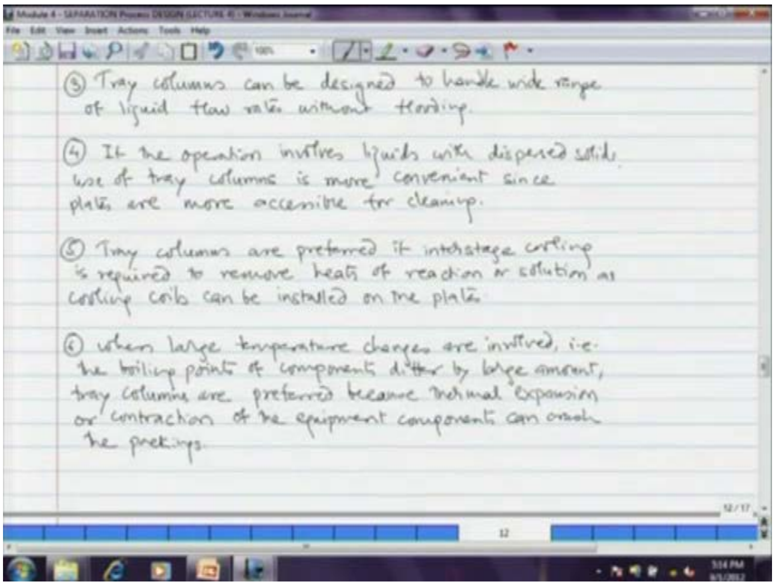

③ Tray columns can be designed to handle wide range of liquid flow rates without flooding.

④ If the operation involves liquids with dispersed solids use of tray columns is more convenient since plates are more accessible for cleanup.

⑤ Tray columns are preferred if interstage cooling is required to remove heats of reaction or solution as cooling coils can be installed on the plates.

⑥ when large temperature changes are involved, i.e. the boiling points of components differ by large amount, tray columns are preferred because thermal expansion or contraction of the equipment components can crush the packings.

The structured packing’s can provide high mass transfer efficiencies and capacities than the tray column, due to this the size of the column required for particular operation decreases. Following general guidelines should be used while making the choice between a plate and a packed tower. First the stage efficiency of a pack tower is quite sensitive to the type and size of the packing, the fluid rates the fluid properties column diameter and operating pressure. So, the manufacturer will report some value of void fraction and some value of operating that the sorry the pressure drop and some value of H E T P, but you will have to conduct your own experiments to verify these values. This value may change from operation to operation, these are not constant.

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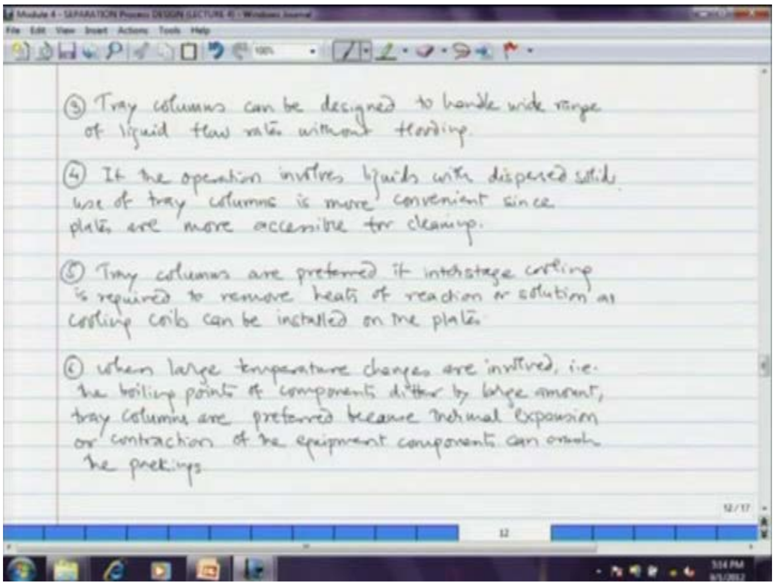

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Secondly, if the ratio of liquid mass velocity to gas mass velocity is low, then tray columns are more reliable than pack column due to liquid separation problems in this columns. Then the tray columns can be designed to handle wide range of liquid flow rates without flooding, but that is not possible in case of pack column. If the operation involves liquid with dispersed solids, then use of tray column is more convenient. Since, the plates are more accessible for cleaning, tray columns are preferred if the inter stage cooling is required to remove heat of reaction or solution.

As a cooling coils can be installed on the plates, another important aspect tray columns have a liquid pool over each tray, when you have a multi component distillation, when you want to draw more than 2 products from a column. As we have seen in previous lecture, you can draw 3, 4, 5 products, in that case also tray columns are preferred because the composition on each tray varies, in case where the feed is dominated either by middle distillate or higher heavy component or middle component. Then depending on the feed composition you can draw more than 1 product and in that case that tray columns are most suitable.

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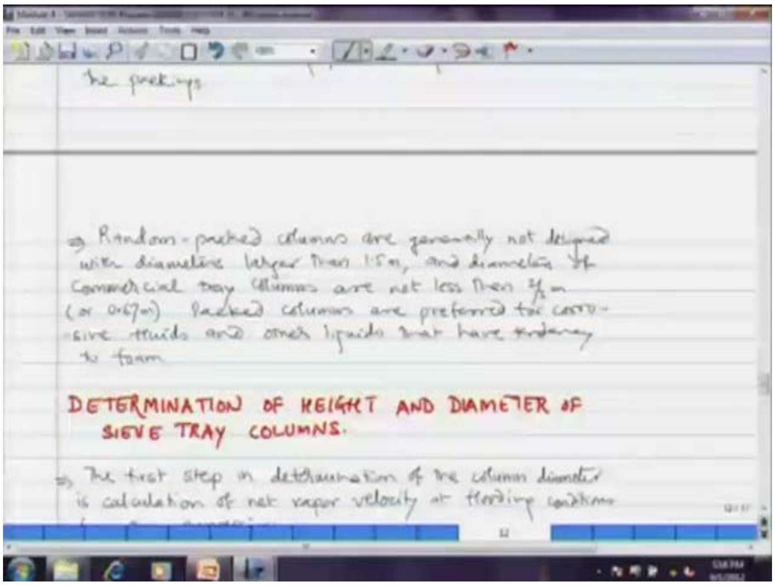

the packings

⇒ Random-packed columns are generally not designed with diameters larger than 1.5 m, and diameters of commercial tray columns are not less than $\frac{2}{3}$ m (or 0.67 m). Packed columns are preferred for corrosive fluids and other liquids that have tendency to foam

DETERMINATION OF HEIGHT AND DIAMETER OF SIEVE TRAY COLUMNS.

⇒ The first step in determination of the column diameter is calculation of net vapor velocity at flooding conditions

Pack tower cannot be used for such purpose when the last temperature differences are encountered in the column; that means, when the boiling points of component defer by large amount. The tray columns are preferred because thermal expansion or contraction of the equipment components can crush the packing. So, packing’s are not used when you encounter large temperature differences. And finally, random pack columns are generally not designed with the diameters larger than 1.5 meter and diameters of the commercial tray columns are not less than two-thirds of a meter or 0.67 metres.

Pack columns are preferred for corrosion fluids and other liquids that have tendency to foam. Why? Reason is very easy. The tower packing’s can be removed very easily, they can be washed, they can be regenerated and put back, but that facility is not possible for trays which are fixed to the column shell. So, dismantling a plate column and taking off the tower packing’s from a pack tower are for different things.

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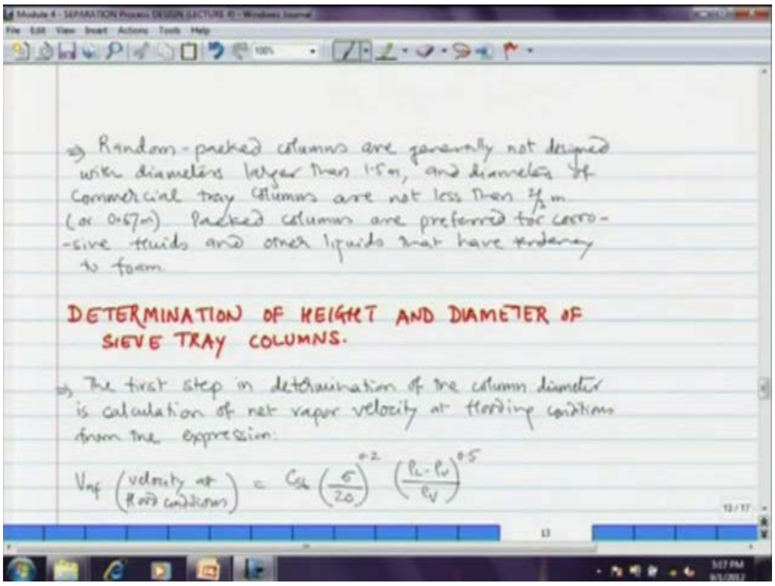

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DETERMINATION OF HEIGHT AND DIAMETER OF SIEVE TRAY COLUMNS.

⇒ The first step in determination of the column diameter is calculation of net vapor velocity at flooding conditions from the expression:

$$V_{nf}\left(\begin{matrix}\text{velocity at}\\ \text{flood conditions}\end{matrix}\right) = C_{sb}\left(\frac{\sigma}{20}\right)^{0.2}\left(\frac{\rho_L-\rho_V}{\rho_V}\right)^{0.5}$$

So, the later is always easier and therefore, whenever you have corrosive fluids or foaming tendency we would like to go for pack column. Now we shall see the last topic of today's lecture that is determination of height and diameter of a sieve tray column. We are not going to the pack column, because those are not very frequently used, the most frequently used columns are plate columns. So, the sieve tray columns the first step in determination of the column diameter is the calculation of net vapour velocity at flooding condition.

From the expression given by fair this is this was given J R fair and that equation says that V n f which is a velocity at flooding condition. Now flooding and loading conditions of a pack tower are or any column you have learnt in solid fluid operation on some operation. When does flooding occur, when does loading occur when there is the gas flow rate is, so high that the liquid cannot go out that is loading point and flooding point is when the column is filled with liquid. So, those things are going to be treated in that greater detail in other courses by NPTEL. So, I do not go much deeper into it, at I suggest that you read through book of and McCabe smith to learnt about the loading and flooding velocity.

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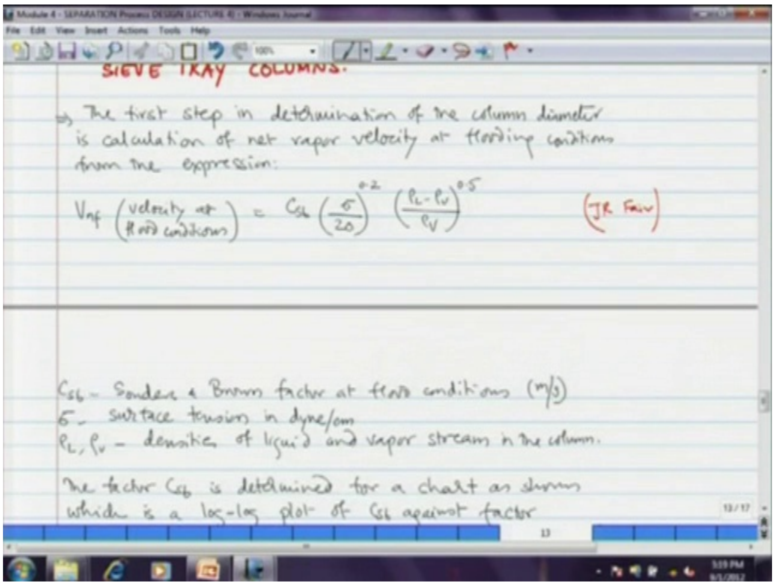

SIEVE TRAY COLUMNS:

⇒ The first step in determination of the column diameter is calculation of net vapor velocity at flooding conditions from the expression:

$$V_{nf} \left(\begin{array}{c}\text{velocity at}\\ \text{flood conditions}\end{array}\right) = C_{sb} \left(\frac{\sigma}{20}\right)^{0.2} \left(\frac{\rho_L - \rho_V}{\rho_V}\right)^{0.5}$$

(JR Fair)

C_{sb} - Souders & Brown factor at flood conditions (m/s)

σ - surface tension in dyne/cm

ρ_L, ρ_V - densities of liquid and vapor stream in the column.

The factor C_{sb} is determined for a chart as shown which is a log-log plot of C_{sb} against factor

So, the velocity at flooding condition V n f is equal to C s b into sigma by 20 raise to 0.2 into rho L minus rho V divided by rho V raise to 0.5. Now, let us see the notations C s b is the Sounders and Brown factor at flooding condition. Sounders and Brown where the first people or first scientist to propose that the flooding occurs due to entrainment of liquid droplets through the raise in vapour in a column or raise in gas in a column. So, based on that analysis, they have given this particular expression. That analysis is quite involved quite complicated it involves essentially the balance of forces on a liquid droplet. So, when can improvement occur under what circumstances it can occur, and so on and so forth. So, that derivation is again out of syllabus, so we do not cover it.

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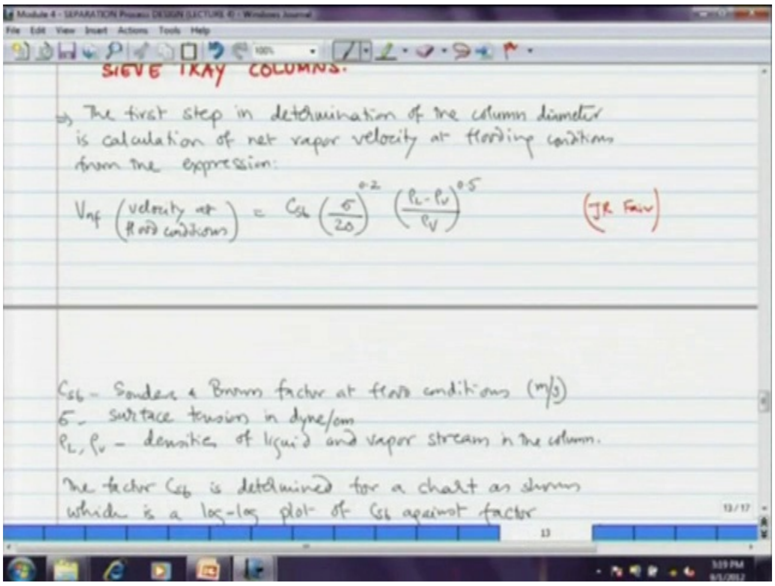

SIEVE TRAY COLUMNS:

⇒ The first step in determination of the column diameter is calculation of net vapor velocity at flooding conditions from the expression:

$$V_{nf}\left(\begin{array}{c}\text{velocity at}\\ \text{flood conditions}\end{array}\right) = C_{sb}\left(\frac{\sigma}{20}\right)^{0.2}\left(\frac{\rho_L-\rho_V}{\rho_V}\right)^{0.5}$$ (JR Fair)

C_{sb} - Souders & Brown factor at flood conditions (m/s)
σ - surface tension in dyne/cm
ρ_L, ρ_V - densities of liquid and vapor stream in the column.

The factor C_{sb} is determined for a chart as shown which is a log-log plot of C_{sb} against factor

We take this expression directly and sigma is a surface tension, now sigma by 20 that 20 number is very tricky. If you see surface tensions are most of organic liquids, it is in the range of 20 to 30 damper cell centimetre. So, that sigma by 20 factor appears.

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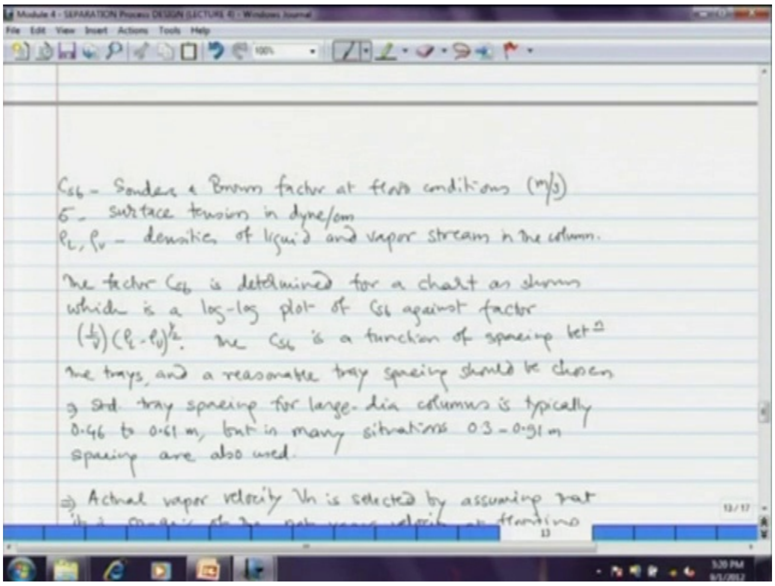

C_{sb} - Souders & Brown factor at flood conditions (m/s)
σ - surface tension in dyne/cm
ρ_L, ρ_V - densities of liquid and vapor stream in the column.

The factor C_{sb} is determined for a chart as shown which is a log-log plot of C_{sb} against factor $\left(\frac{L}{V}\right)(\rho_L-\rho_V)^{1/2}$. The C_{sb} is a function of spacing betⁿ the trays, and a reasonable tray spacing should be chosen

⇒ Std. tray spacing for large-dia columns is typically 0.46 to 0.61 m, but in many situations 0.3 – 0.91 m spacing are also used.

⇒ Actual vapor velocity V_n is selected by assuming that [illegible] flooding

And then what we have is the ratio of densities rho L minus rho V divided by rho V raise to 0.5. The factor C s b is determined for by chart which is log-log plot of C s b against factor L by V into rho L minus rho V divided by half and the C s b is a function of spacing between the trays.

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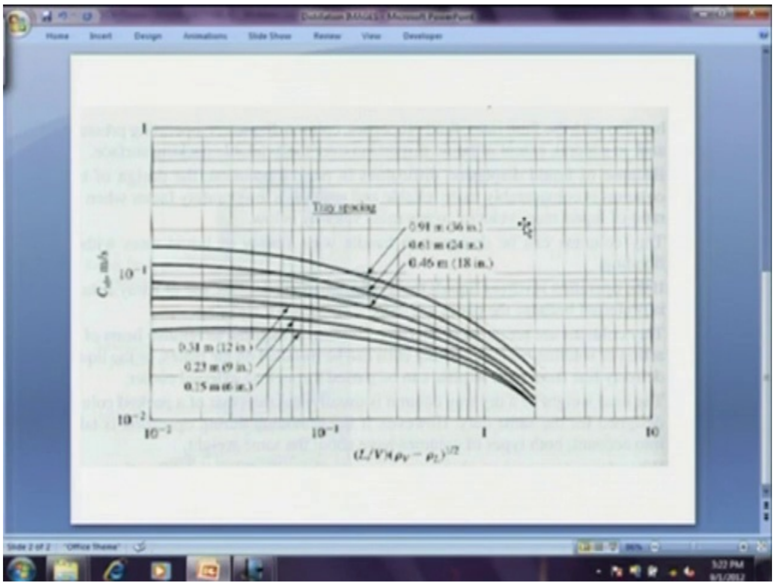

Now, I will show you that chart, chart appears on the screen now. The function C s b the Sounders and Brown factor meter per second is plotted against L by V into rho V minus rho L raise to half. L is the molar flow rate of liquid, V is the molar flow rate of sorry mass flow rate, L is the mass flow rate of liquid divided by mass flow rate of vapour into rho V minus rho L. rho V is the density of vapour, rho L is the density of rho L is the density of liquid and rho V is the density of vapour. And the difference raise to half and what you have is another factor here that is a number of sorry the tray spacing.

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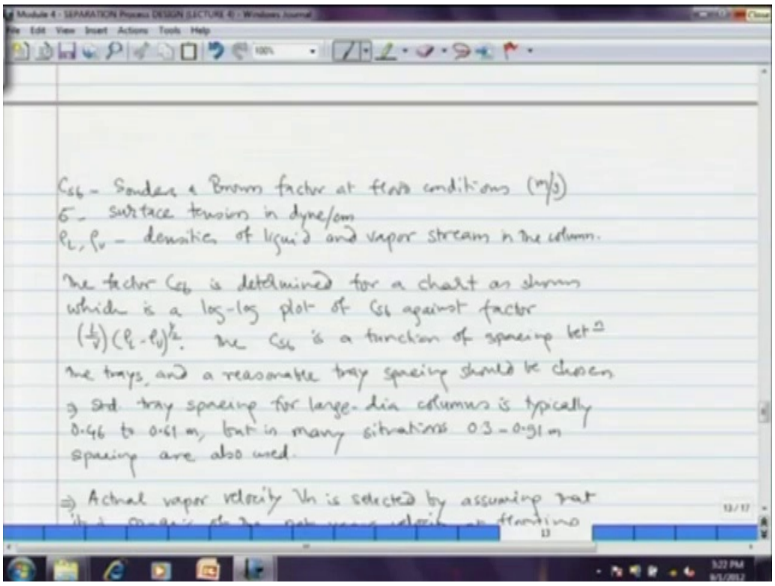

C_{sb} - Sonders & Brown factor at flood conditions (m/s)
σ - surface tension in dyne/cm
ρ_L, ρ_V - densities of liquid and vapor stream in the column.

The factor C_{sb} is determined for a chart as shown which is a log-log plot of C_{sb} against factor $\left(\frac{L}{V}\right)(\rho_L - \rho_V)^{1/2}$. The C_{sb} is a function of spacing betn the trays, and a reasonable tray spacing should be chosen

⇒ Std. tray spacing for large-dia columns is typically 0.46 to 0.61 m, but in many situations 0.3 - 0.91 m spacing are also used.

⇒ Actual vapor velocity V_n is selected by assuming that

What should be the distance between the 2 trays, the standard spacing between the space for a large diameter column is typically 0.46 to 0.61metres. So, let us say approximately 1 and half to 2 feet, but in many situations 0.3 to 0.91metre spacing is also used, somewhat larger spacing. So, that there is enough pace to enter the column and clean the column at a time of start up. The actual vapour velocity V n is selected by assuming that it is 50 to 90 percent of the net vapour velocity at the flooding condition and after determination of this.

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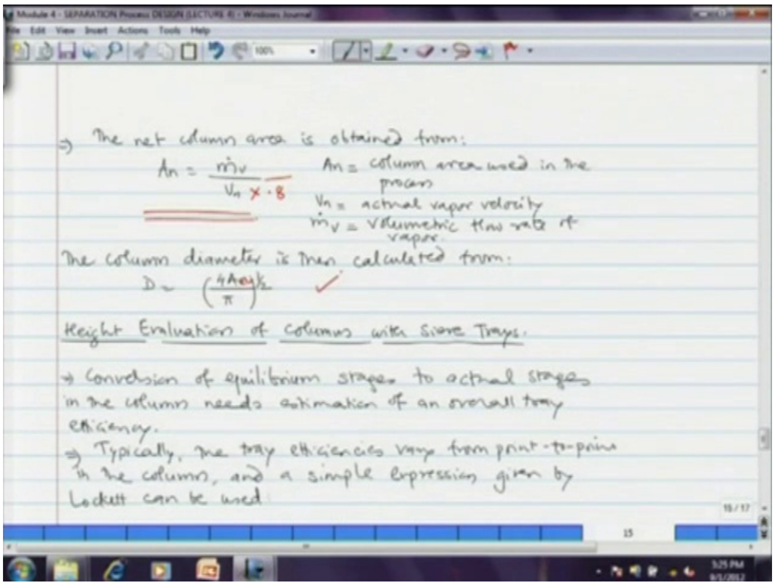

⇒ The net column area is obtained from:

$$A_n = \frac{\dot{m}_v}{V_n \times .8}$$

A_n = column area used in the process
V_n = actual vapor velocity
$\dot{m}_v$ = volumetric flow rate of vapor.

The column diameter is then calculated from:

$$D = \left(\frac{4A_n}{\pi}\right)^{1/2} \checkmark$$

Height Evaluation of Columns with Sieve Trays.

⇒ Conversion of equilibrium stages to actual stages in the column needs estimation of an overall tray efficiency.

⇒ Typically, the tray efficiencies vary from point-to-point in the column, and a simple expression given by Lockett can be used

So, C s b you get from that plot for given surface tension and vapour and liquid densities. And that C s b into this factor sigma by 20 raise to 0.2 and this density ratio raise of 0.5 will give you V n f. An actual vapour velocity V n is selected assuming that it is 50 to 90 percent of the net vapour velocity at the flooding condition. And once you select the vapour velocity, then the net column area is n is equal to m dot v divided by V n, where n is the column area used in the process, n is the actual vapour velocity, m dot v is the volumetric flow rate of the vapour.

Now this expression is assumed under unless this expression is assumes that the entire column area is available for vapour flow rate, but that is not the case. For every plate there is a down comer through which the liquid will flow, the down comer will occupy about 20 percent of the column area and rest 80 percent area is available of the vapour flow rate. So, while calculating the net column area you have to again divide here by 0.8.

So, that you account for the area for the down comer, so after getting the area of the column you can calculate the diameter of the column in straight forward way. D is equal to 4 into A n divided by pi raise to half. So, we have seen as how the diameter is determined, then comes the height. Conversion of equilibrium stages to actual stages in the column needs estimation of overall tray efficiency.

A few moments before we saw as how theoretical number of plates can be determined for different methods, whether is Smoker method or F-U-G method Fenske Underwood Gilliland method, but those are theoretical number of stages to get the actual number of stages. You have to determine the tray efficiency, you have learned efficiency called as Murfree efficiency in mass transfer operation.

For similar efficiency, but is given by a locket, Murfree efficiency equation is more theoretical, so it is difficult to apply in terms of practical quantities. So, locket has given a simpler expression in practical terms, for efficiency E naught is equal to 0.49 into mu L into alpha L K H K that is random volatility between light key component and heavy key component average relative volatility. So, E naught is equal to 0.492 into mu L, mu L is the liquid viscosity of the mixture into alpha average raise to minus 0.245.

So, dividing the theoretical number of plates n by E naught will give you the actual number of plates. And n actual that is what we denote n actual, and we already know the plate spacing, because when we calculated V n f the flooding velocity. We had already decided the plate spacing, the Sounders and Brown plot was based on the tray space and so we know tray spacing.

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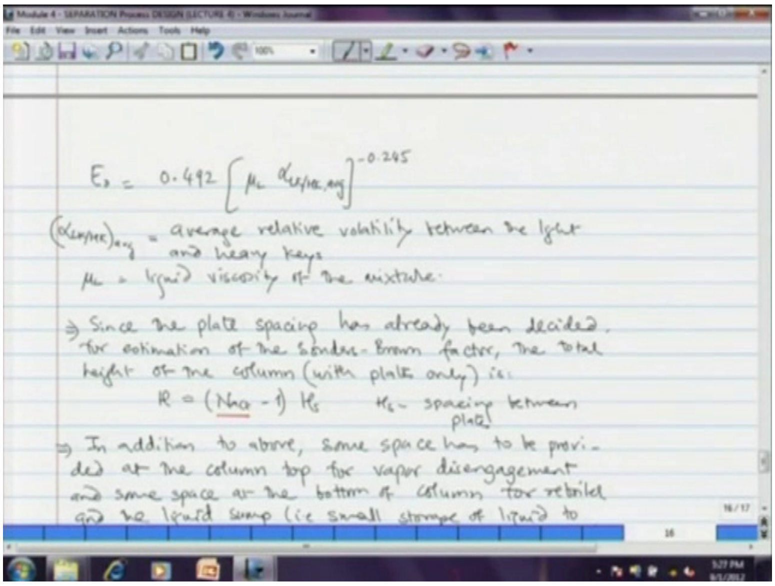

$$E_o = 0.492\left[\mu_L\, \alpha_{LK/HK,avg}\right]^{-0.245}$$

$(\alpha_{LK/HK})_{avg}$ = average relative volatility between the light and heavy keys

μ_L = liquid viscosity of the mixture.

⇒ Since the plate spacing has already been decided for estimation of the Souders-Brown factor, the total height of the column (with plates only) is:

$$H = (N_{act} - 1)\, H_s \qquad H_s - \text{spacing between plates}$$

⇒ In addition to above, some space has to be provided at the column top for vapor disengagement and some space at the bottom of column for reboiler and the liquid sump (i.e small storage of liquid to

And then the height of the column with only trays is obvious that n actual minus 1 into H s, where H s is the spacing between the plates, but the column also has a re-boiler and a condenser. The vapour slicing from the top most plate of the column will have lot of liquid droplets entrained in it. So, to remove those droplets to have the vapour making disengagement, you need some extra space over the top plate.

That space will also account for any raise in the pressure of the column as the distillation proceeds. So, you provide extra space above the top plate similarly, you have provide extra space below the bottom plate, because you have to have a liquid sound that will provide liquid to the re-boiler. Why? Because, having that reservoir is essential, so as to give some kind of a margin for hollow operation in case the previous column or the previous supply is disrupted.

So, that instead of feeding directly to the bottom, you need some liquid reservoir. So, that storage also require space typically about 15 percent of the column height is provided as this margin space at top and bottom. And therefore, the total height H is not just N actual minus 1 into H s plus delta H, which is this margin. Delta H is 0.3 into H 15 percent of the total height on each side top and bottom.

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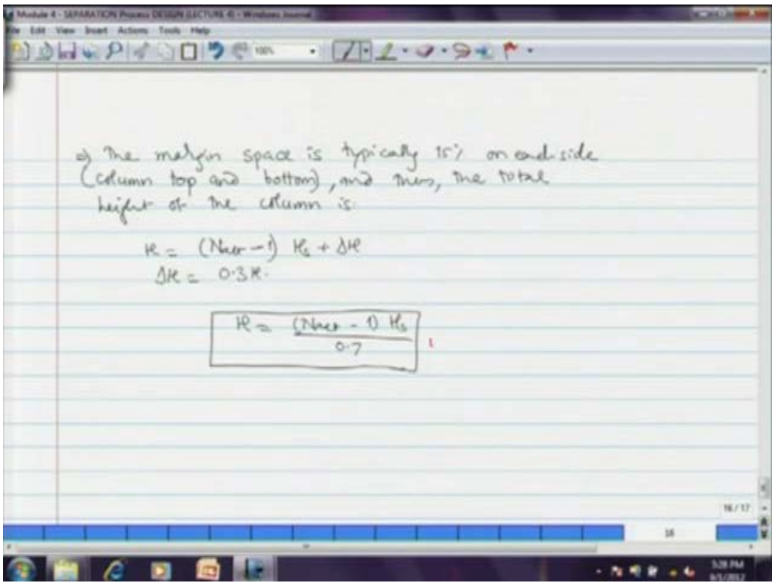

e) The margin space is typically 15% on each side (column top and bottom), and thus, the total height of the column is:

$$H = (N_{act} - 1)\, H_S + \Delta H$$

$$\Delta H = 0.3 H$$

$$\boxed{H = \frac{(N_{act} - 1)\, H_S}{0.7}}$$

And therefore, the total height of the column can be approximated as N actual minus 1 in to H s, which is the spacing between the plates divided by 0.7. Remember, there are many approximations and assumptions involved in this expression, but still this gives a fairly good idea as how tall will be the column for a given separation. So, today we have seen several aspects of design of a distillation column, first we compared the packed and plate columns.

Then we saw what are the steps in design of a distillation column, then we saw 2 methods for determination of number of plates in the column, their F-U-G method Fenske Underwood Gilliland method and Smoker method. And finally, we saw a simple procedure for determination of the column diameter and height. So, this lecture completes module 4 of our course that is design of separation processes. In the next two lectures we shall see several problems on the topics that we have covered in the 4 theory lectures. These topics are essentially the absorber and distillation column design.

Thank you.