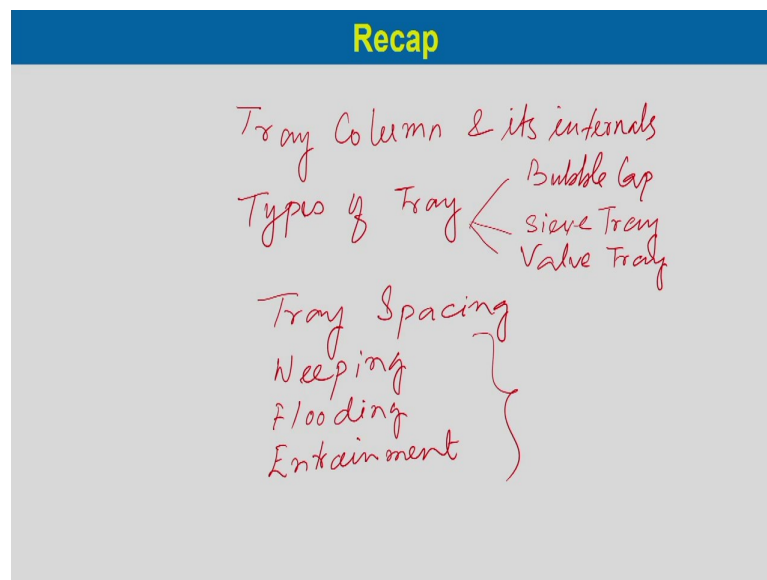


Mass Transfer Operations-I
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Lecture – 19
Sieve Tray

Welcome to the third lecture on module 3 on Mass Transfer Operation. In this module, we are discussing equipment for gas liquid operations. So, before going to the next lecture let us have recap on our previous lecture. In our previous lecture, we have mostly considered the tray column and its different internals.

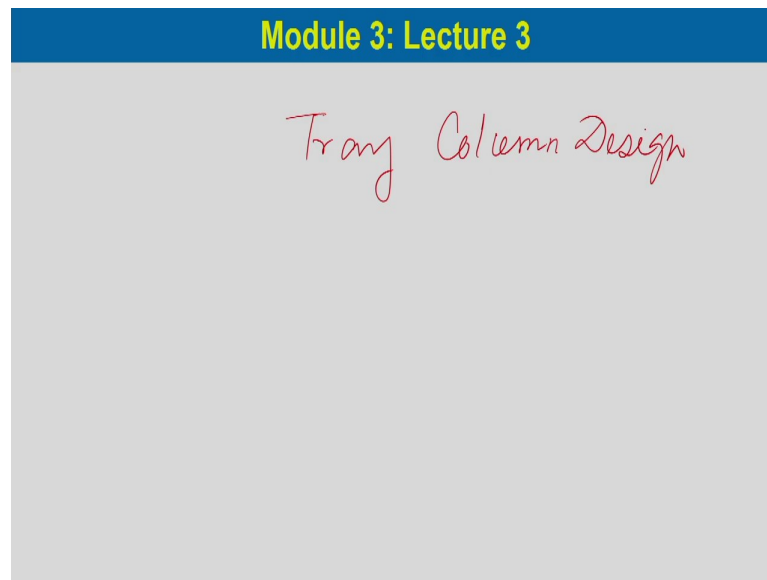
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So, we considered tray column and we have seen how the tray column looks like, its different types of tray and under which we have seen there are three different types of tray generally used one is bubble cap, then sieve tray and then valve tray.

So, valve tray is a proprietary trays which are used for tray column, but bubble cap and sieve tray they have detail information are available in the literature for their design and then we have seen what are the different conditions happens or occurs in a tray column when we change the gas and liquid flow and how the flooding, entrainment that occurs. So, we have also discussed about the tray spacing and weeping, flooding, entrainment so and we have seen how to take care of these parameters.

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In this lecture module 3, lecture 3 we will continue our tray column design. So, we will see how to design a tray column. So, tray column design we will considered in under this know lecture.

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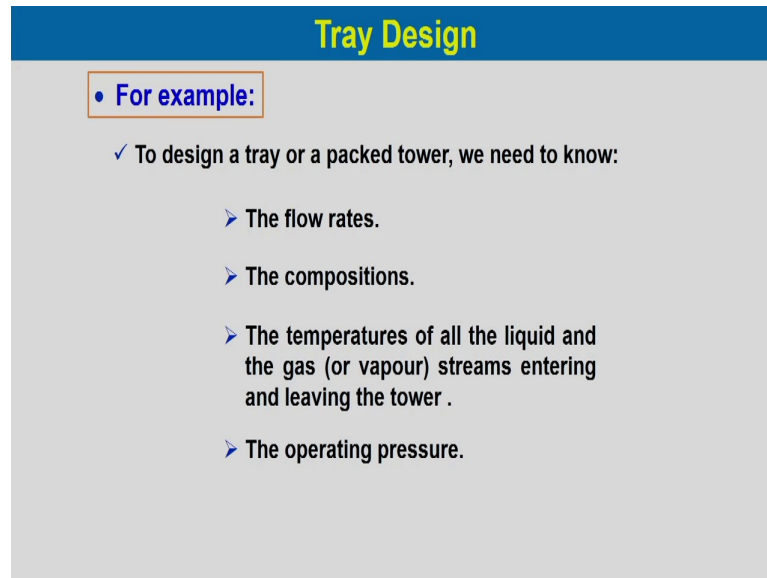
The image shows a slide with a blue header containing the text "Tray Design" in yellow. Below the header, there is a list of design parameters in black text on a light gray background.

- While designing a process plant or a part of it, complete material and energy balance calculations for every piece of equipment or device are done in order to establish the design basis in terms of
 - ✓ The flow rate.
 - ✓ The composition.
 - ✓ The temperature and pressure of each stream.
 - ✓ The amount of heat input, output and generation, if any.

While designing a process plant or a part of it complete material and energy balance calculation for every piece of equipments are required and this can be established for the design only when we have certain parameters in our hand; one of them is the flow rate is known to us. Then we need to know the composition of the mixture both for the liquid

phase composition and the gas phase composition. The temperature and pressure of each stream both the gas and the liquid stream, each stream we should know the temperature and pressure condition. The amount of heat input, output and generation if any because of any chemical reactions or so; so this parameters should be available to do the detail materials and energy balance for a particular process plants.

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Tray Design

- **For example:**
 - ✓ To design a tray or a packed tower, we need to know:
 - The flow rates.
 - The compositions.
 - The temperatures of all the liquid and the gas (or vapour) streams entering and leaving the tower .
 - The operating pressure.

For example, if we take a tray or a packed tower we need to know the flow rates, the compositions, the temperature of all liquids and the gas or the vapour streams entering and leaving the tower and then the operating pressure; so these are required.

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Tray Design

- For example:
 - The physical properties of the streams such as:
 - ❖ density
 - ❖ viscosity
 - ❖ diffusivity
 - ❖ surface tension, etc.
 - These are required to be known or estimated for use in design calculations.

The physical properties of the stream are also required such as density, viscosity, diffusivity, surface tension etcetera. These are required to be known or estimated for use in design calculation. So, the physical properties either has to be available in the literature or has to be estimated which will be used in the design calculation. So, if it is a unknown solvent, an unknown gas, unknown liquid mixture this properties has to be measured or a know estimated using certain empirical correlations.

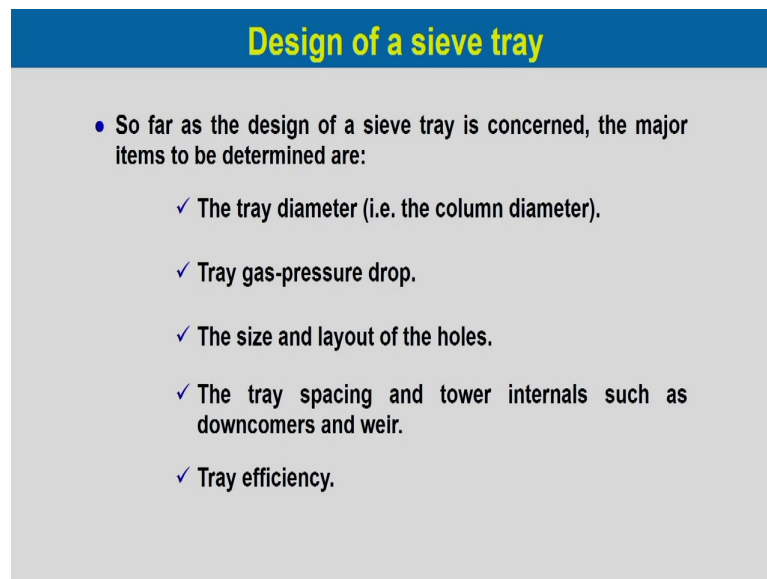
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Tray Design

- The most common types of trays are:
 - ✓ Bubble-cap tray
 - ✓ Sieve tray
 - ✓ Valve tray
- Here we give a brief outline of the procedure of design of a sieve tray (also called 'perforated tray').
- Valve trays are 'proprietary' trays and only limited information on their design is available in the open literature.

The most common type of trays are bubble cap tray, sieve tray and valve tray. So, these we have already introduced in the last class. So, under this trays; we give a brief outline of the procedure of design of sieve tray which is called the perforated tray. Valve tray as you said it is a proprietary trays and only limited information on their design is available in the open literature as far as the design is concerned we will outline on the sieve tray because the design parameters for the proprietary trays like valve trays are not much available in the open literature.

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Design of a sieve tray

- So far as the design of a sieve tray is concerned, the major items to be determined are:
 - ✓ The tray diameter (i.e. the column diameter).
 - ✓ Tray gas-pressure drop.
 - ✓ The size and layout of the holes.
 - ✓ The tray spacing and tower internals such as downcomers and weir.
 - ✓ Tray efficiency.

So, far the design of sieve tray is concerned the major items to be determined are the tray diameter. So, we need to know what would be the tray diameter or that is the column diameter column internal diameter. So, that has to be determined. The tray gas pressure drop so how much pressure drop because of the flow of the gas through the know from one trays to the other through the liquid.

The size and layout of the holes; hole layout and size; the tray spacing and tower internal such as down comers and weir and tray efficiency. All this we will discuss briefly in this lecture, but mostly the detail design of the plate columns or the sieve tray are available in your design course or you will learn in future. So, we will just briefly highlight to give the basic understanding on the sieve tray design.

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Design of a sieve tray

Tray (or Column) Diameter

- The required diameter of a tray or the column for the given flow rates of the gas and the liquid phase is determined from flooding considerations.
- We know that as the gas velocity in a column is gradually increased, a limiting velocity is attained above which entrainment is high enough to cause accumulation of liquid on the trays leading to flooding.
- This velocity corresponds to the theoretical maximum capacity of the column. There are a few methods of calculation of the flooding velocity.

Tray or column diameter the required diameter of a tray of column for a given flow rates of the gas and the liquid phase is determined from the flooding conditions. We know that, the gas velocity in a column is gradually increased. So, if we increase the gas velocity in a column there is a limiting velocity above which entrainments is high enough to cause accumulation of the liquid on the trays leading to flooding after a certain velocity or we call the limiting velocity, so under such circumstances the column will floods. The velocity corresponding to the theoretical maximum capacity of the column that limiting velocity will correspond to the theoretical maximum capacity of the column.

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Tray diameter

- Souders- Brown equation gives the flooding velocity for 'spray entrainment flooding' is as follows:

$$V_{fi} = C_{SB} \left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{1/2}$$

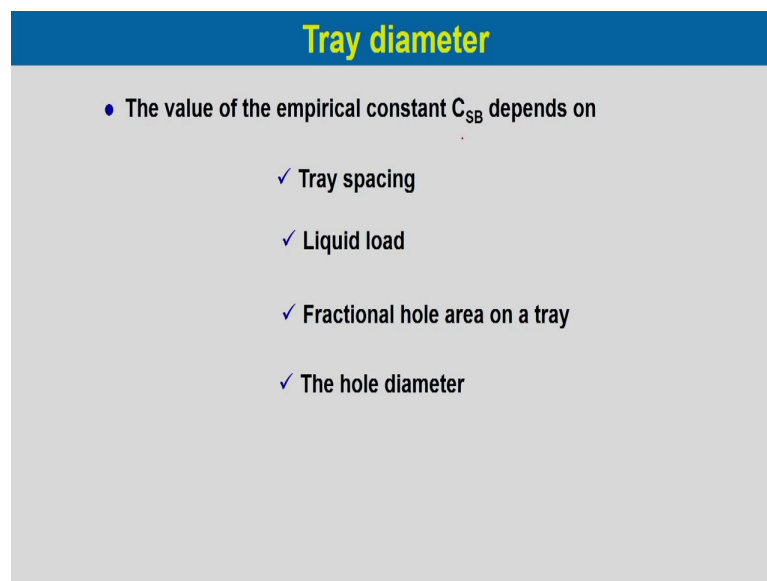
Here, V_{fi} = Superficial velocity at flooding.

C_{SB} = 'Souders-Brown flooding constant' (sometimes called 'capacity factor').

➤ In reality the quantity C_{SB} is not a constant.

There are few methods of calculation of the flooding velocity. Souders-Brown equation gives the flooding velocity for spray entrainment flooding and which is given as follows V_{fl} would be equal to $C_{SB} \sqrt{\rho_L - \rho_G}$ divided by $\sqrt{\rho_G}$, where v_l is the superficial velocity at flooding and C_{SB} is the Souders-Brown flooding constant sometimes called capacity factor and ρ_L and ρ_G they are the density of the liquid and the gas although C_{SB} is called the capacity factor and the Souder-Brown flooding constant, but actually this quantity C_{SB} is not a constant.

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Tray diameter

- The value of the empirical constant C_{SB} depends on
 - ✓ Tray spacing
 - ✓ Liquid load
 - ✓ Fractional hole area on a tray
 - ✓ The hole diameter

The value of empirical constant C_{SB} depends on different parameters like tray spacing, liquid load, fractional hole area on a tray and the hole diameter. So, these are the 4 parameters which will influence the empirical constant values of C_{SB} of the Souder-Brown constant.

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Tray diameter

- C_{SB} can be estimated using the following relationship:
$$C_{SB} = F_{st} F_f F_{ha} C_f$$

where F_{st} = surface tension factor = $(\sigma/20)^{0.2}$
 σ = surface tension, dyn/cm

F_f = foaming factor = 1.0 for nonfoaming systems for many absorber may be 0.75 or even less

$F_{ha} = 1.0$ for $(A_h/A_a) \geq 0.10$ and
 $= 5(A_h/A_a) + 0.5$ for $(A_h/A_a) < 0.1$

(A_h/A_a) = ratio of vapor hole area to tray active area

C_{SB} can be estimated using the following relationship; C_{SB} equal to F_{st} into F_f into F_{ha} into C_f where F_{st} is the surface tension factor and which is defined by σ by 20 to the power 0.2 σ is the surface tension.

So, which is in know unit of dynes per centimetre and then F_f is the foaming factors and it is usually one for non foaming systems. For many absorber it maybe 0.75 or even less and F_{ha} is 1 for A_h by A_a greater than equal to 1.0 and it would be equal to 5 into A_h by A_a plus 0.5 where for A_h by A_a less than 0.1. This A_h by A_a is the ratio of vapour hole area to tray active area. So, A_a is the tray active area and A_h is the vapour hole area. So, vapour hole area divided by tray active area this ratio is related with the F_{ha} another factor hole to tray area and which is used over here to calculate C_{SB} .

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Tray diameter

$$C_f = \alpha \log\left(\frac{1}{m}\right) + \beta$$

$C_{SB} = F_{st} F_{ra} C_f$

where

$$\alpha = 0.0744t + 0.01173$$
$$\beta = 0.0304t + 0.015$$
$$m = \text{flow parameter} = (L/G) (\rho_G/\rho_L)^{0.5}$$

t = tray spacing, in m

- If the value of m is in the range of 0.01 to 0.1, then use $m = 0.1$ in the above equation.

Now, C_f which can be related with known $\alpha \log 1$ by m plus beta. This alpha can be related term with $0.0744t + 0.01173$ with this equation and then beta is also a constant can be related with $0.0304t + 0.015$. So, with these two relations alpha and beta are related and then we can have m is the flow parameter which also depends on liquid and gas flow and their density.

So, L is the flow rate of the liquid, G is the gas flow rate and then ρ is the density of the gas ρ_G and ρ_L is the density of the liquid. So, L/G into ρ_G/ρ_L to the power 0.5, t is the tray spacing over here in equation of alpha and beta tray spacing which is in metre and the value of m over here is in the range of 0.01 to 0.1 then use m is equal to 0.1 in the above equation. So, we can use as 0.1 at the extreme case.

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Tray diameter

- Typically, the column diameter is based on a specific fractional approach to flooding **f**.
- Then,
$$D_t = \left(\frac{4Q_G}{fV_{fl}(1-\frac{A_d}{A_t})\pi} \right)^{1/2}$$
- It is suggested that A_d/A_t must be chosen based on the value of m , as

$$\frac{A_d}{A_t} = \begin{cases} 0.1 & \text{for } m \leq 0.1 \\ 0.1 + \frac{m - 0.1}{9} & \text{for } 0.1 \leq m \leq 1.0 \\ 0.2 & \text{for } m \geq 1.0 \end{cases}$$

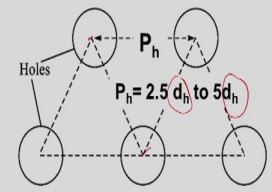
Typically the column diameter is based on a specific fractional approach to flooding. So, fractional approach to flooding means fractional flooding is f and then we can write D_t column diameter would be equal to $4Q_G$ divided by f into V_{fl} 1 minus A_d by A_t into π to the power half. So, it is suggested that A_d/A_t must be chosen based on the values of m as A_d/A_t would be 0.1 when, m is less than 0.1 A_d/A_t would be 0.1 plus m minus 0.1 divided by 9 for this range when m is no greater than equal to 0.1 and it is less than equal to 1 and it would be 0.2 for m greater than 1.0. So, we can use this values over here and V_{fl} and given a particular flooding conditions with a volumetric flow rate we can calculate the tower diameter **ok** or column diameter.

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Perforations and active area

- In most sieve trays, the holes are placed in the corner of equilateral triangles at a distance between centers (pitch, P_h) from 2.5 to 5 hole dia (d_h)
- For this arrangements:

$$\frac{A_h}{A_a} = \frac{\pi}{4 \sin(60)} \left(\frac{d_h}{P_h}\right)^2 = 0.907 \left(\frac{d_h}{P_h}\right)^2$$



Here, A_h = vapor hole area
 A_a = Tray active area

Now, in most sieve trays the holes are placed in the corner of equilateral triangle at a distance between centres from 0.25 d h 2.5 hole dia. So, this pitch length from here to here would be 2.5 d h to 5 d h that is the hole dia. So, it would be in this cases in most of the sieve trays. For this arrangement we can write A_h by A_a would be equal to π by $4 \sin 60$ d_h by P_h whole square which is equal to $0.907 d_h$ by P_h square. So, A_h is the vapour hole area and A_a is the tray active area.

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Downcomer geometry

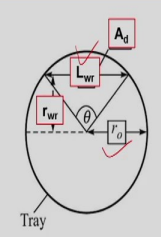
L_{wr} = Weir length

A_d = Downcomer area

A_t = Tower cross sectional area

D_t = Diameter of the tower = $2r_o$

r_{wr} = Distance of the weir from the center of the tower



$$\frac{A_d}{A_t} = \frac{\theta - \sin\theta}{2\pi}$$

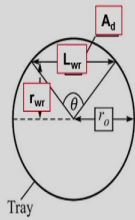
Now, we need to know the down comer geometry. L_{wr} is the weir length which is given over here this is the weir length and A_d is the down comer area here A_t is the tower cross sectional area and D_t is the diameter of the tower which is equal to twice r_o here, r_{wr} is the distance of the weir from the centre of the tower.

So, we can calculate A_d by A_t would be equal to θ minus $\sin \theta$ by 2 pi.

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Downcomer geometry

L_{wr} = Weir length
 A_d = Downcomer area
 A_t = Tower cross sectional area
 D_t = Diameter of the tower = $2r_o$
 r_{wr} = Distance of the weir from the center of the tower



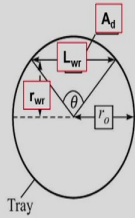
$$\frac{L_{wr}}{D_t} = \sin(\theta/2)$$

So, here L_{wr} by D_t would be equal to $\sin \theta$ by 2.

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Downcomer geometry

L_{wr} = Weir length
 A_d = Downcomer area
 A_t = Tower cross sectional area
 D_t = Diameter of the tower = $2r_o$
 r_{wr} = Distance of the weir from the center of the tower



$$\frac{r_{wr}}{D_t} = \frac{1}{2} \cos(\theta/2)$$

And r wr by D t we can write half into cos theta by 2.

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Tower diameter

- If the tower diameter calculated from equations is less than 60 cm, then a packed column is generally used
- As per the equation given for calculation of tower diameter, tray spacing must be specified.
- On the other hand, the tray spacing is also dependent on tower diameter.

If the tower diameter calculated from the equation is less than 60 centimetre then, a packed column is generally used. As per the equation given for the calculation of the tower diameter, tray spacing must be specified. On the other hand, the tray spacing is also depend on the tower diameter.

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Tower diameter

- The recommended values are given below:

Tower diameter, D_t , m	Tray spacing, t_s , m
1 or less	0.5
1 to 3	0.6 ✓
3 to 4	0.75 ✓
4 to 8	0.9 ✓

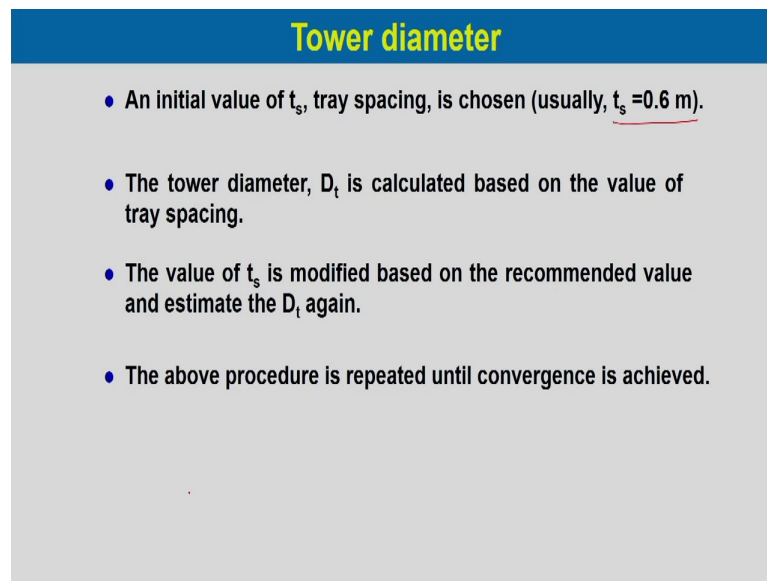
- Therefore, the calculation of tower diameter is an iterative procedure.

So, these are the recommended values for the tower diameter and you can see over here the tower diameter which is in metre is 1 or less when the tray spacing you can set as 0.5

metre and when it is between 1 to 3, the tower diameter between 1 to 3 the tray spacing is 0.6 metre and it between 3 to 4 metre tower diameter the tray spacing is 0.75 metre and 4 to 8 metre the tray spacing is 0.9 metre.

So, these are the recommended values for the tower diameter of which need to be used for the design; therefore, the calculation of tower diameter is an iterative procedures.

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Tower diameter

- An initial value of t_s , tray spacing, is chosen (usually, $t_s = 0.6$ m).
- The tower diameter, D_t is calculated based on the value of tray spacing.
- The value of t_s is modified based on the recommended value and estimate the D_t again.
- The above procedure is repeated until convergence is achieved.

So, we have to use the iterative procedures and an initial value of t_s that is tray spacing is chosen when t_s is 0.6 m at their lower values and then gradually we can iterate with the higher values of tray spacing.

The tower diameter D_t is calculated based on the values of tray spacing. The value of t_s is modified based on the recommended values and estimate the D_t again. This above procedure is repeated until the convergence is achieved.

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Tray pressure drop

Pressure drop for a sieve tray → Dry tray pressure drop (frictional loss for vapour flow through the short tube formed owing to the plate thickness + an exit loss) + pressure drop due to liquid holdup on the tray + pressure drop due to surface tension

$\therefore h_t = h_d + h_l + h_s$

where,

- ✓ h_t = total pressure drop/ plate, cm of liquid
- ✓ h_d = dry-plate pressure drop, cm of liquid
- ✓ h_l = hydraulic head, cm of liquid
- ✓ h_s = head loss due to the surface tension, cm of liquid

Now, there is a pressure drop in the tray. So, pressure drop of a sieve tray we can calculate tray pressure drop fractional loss of vapour flow through the short tube formed owing to the plate thickness and an exit loss. So, this know short tube means when we make a hole with a particular thickness of the tray there will be a sort no tube like arrangement.

So, owing to that when the vapour flow is through that, you consider a vapour flowing through a short tube due to the thickness of the plate plus an exit loss plus pressure drop due to the liquid holdup on the tray and plus the pressure drop due to surface tension. So, all this has to be accounted for dry tray pressure drop, then we have to consider pressure drop due to liquid holdup and pressure drop due to surface tension. Under dry tray pressure drop, we have two types due to the frictional loss which is due to the forming the tube short tube due to the plate thickness another one is an exit loss. So, these has to be consider under the dry tray pressure drop.

So, we can write the equation h_t is equal to h_d plus h_l plus h_s where h_t is the total pressure drop per plate of centimetre of liquid it is in centimetre of liquid and h_d is the dry plate pressure drop centimetre of liquid and h_l is the hydraulic head of that is centimetre of liquid. So and h_s is the head loss due to the surface tension that is centimetre in of liquid; so, total would be the drive plate, pressure drop, hydraulic head

and the head loss due to the surface tension, this 3 pressure drop are added to get the total pressure drop in a sieve tray.

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Dry- plate pressure drop

Given as modified orifice type equation (Ludwig, 1979) :

$$h_d = 0.051 \left(\frac{v_h}{C_o} \right)^2 \rho_G \left(\frac{\rho_w}{\rho_L} \right) \left[1 - \left(\frac{A_h}{A_a} \right)^2 \right]$$

where,

v_h = gas velocity through the hole, m/s
 ρ_w = density of water
 C_o = orifice - coefficient

Now, given as modified orifice type equation, Ludwig in 1979 this is the equation h_d would be equal to 0.051 into v_h by C_o to square into ρ_G into ρ_w by ρ_L whole into 1 minus A_h by A_a whole square. So, with this we can calculate h_d where, v_h is the gas velocity through the holes in metre per second, ρ_w is the density of water, C_o is the orifice coefficient and A_h and A_a is already mentioned earlier it is a hole area A_h and A_a is the active tray area. The ρ_G , ρ_w and ρ_L are conventionally used.

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Surface tension pressure drop

Can be obtained from force balance between the internal force of a bubble and the surface tension force:

$$h_s = \frac{6\sigma_L}{g\rho_L d_h} \quad \checkmark$$

where it is assumed that maximum bubble size may be taken as the hole diameter, d_h

Then surface tension pressure drop we can obtain from the force balance between the internal force of bubble and the surface tension force. So, from this h_s would be equal to $6\sigma_L / g\rho_L d_h$; so using these we can calculate the surface tension force where it is assumed that maximum bubble size may be taken as the hole diameter. So, this is the assumptions if you wanted to calculate the pressure drop due to surface tension force and where we consider d_h is the maximum bubble diameter which can be achieved.

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Entrainment

At high vapour rate → Entrainment is significant
→ Decreases tray performance

The fractional entrainment [E_f] can be defined as:

$$E_f = \frac{\text{mass flow rate of entrainment liquid}}{\text{mass flow rate of upward gas}}$$

Can be calculated using the following correlation:

$$E_f = 0.00335 \left(\frac{h_B}{t_s}\right)^{1.1} \left(\frac{\rho_L}{\rho_G}\right)^{0.5} \left(\frac{h_L}{h_B}\right)^k$$

Now, at very high vapour rate entrainment is significant that we know. So, decreases tray performance and the fractional entrainment E_f can be defined as mass flow rate of the entrainment liquid divided by the mass flow rate of upward gas. So, it can be calculated using the correlation E_f would be equal to $0.00335 h \beta t s$ to the power 1.1 ρL by ρG to the power 0.5 $h L$ by $h \beta$ to the power k .

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Tray efficiency

- Tray efficiency defined as the fractional approach by a real tray to an equilibrium stage.

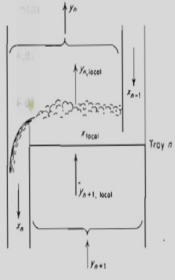
Point efficiency

- The contact among the phases is not uniform at all locations

$Y_{n+1, local}$ = Local concentration of the gas leaving $(n + 1)^{th}$ tray ✓

$Y_{n, local}$ = Local concentration of the gas leaving n^{th} tray ✓

$Y_{n, local}^*$ = Concentration in equilibrium with $X_{n, local}$



The diagram illustrates a tray with gas flowing upwards and liquid flowing downwards. It shows the local concentration of gas leaving the (n+1)th tray ($Y_{n+1, local}$), the local concentration of gas leaving the nth tray ($Y_{n, local}$), and the concentration in equilibrium with the local liquid concentration ($Y_{n, local}^*$). The tray is labeled 'Tray n'.

The tray efficiency which is defined as the fractional approach by a real tray to an equilibrium tray. So, we need to know the tray efficiency. So, in tray efficiency we will discuss few terms one is point efficiency. You can see in this case the contact among the phases is not uniform at all locations. So, the point efficiency will vary $Y_{n+1, local}$ is the local concentration of the gas leaving $n + 1$ th tray and $Y_{n, local}$ is the local concentration of the gas living at n th tray. So, these 2 are local concentration and $Y_{n, local}^*$ is the concentration in equilibrium with $X_{n, local}$. So, this $Y_{n, local}^*$ is in equilibrium with $X_{n, local}$.

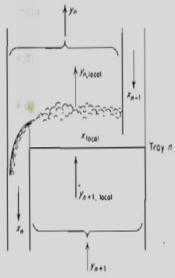
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Point efficiency

- The point efficiency is defined as

$$E_{OG} = \frac{Y_{n,local} - Y_{n+1,local}}{Y_{n,local}^* - Y_{n+1,local}}$$
$$= \frac{\text{Actual enrichment of the gas}}{\text{Maximum enrichment}}$$

- Assumptions taken:
 - ✓ The local concentration $X_{n,local}$ is constant in the vertical direction.
 - This means that the liquid is vertically well mixed.
 - Thus the concentration is uniform in the vertical direction at any point on the tray.



The diagram illustrates a distillation tray labeled 'Tray n'. It shows a cross-section of the tray with gas flowing upwards and liquid flowing downwards. Key variables are labeled: y_n at the top, $x_{n,local}$ at the liquid-gas interface, x_n at the bottom left, $x_{n+1,local}$ at the bottom right, and y_{n+1} at the bottom. A dashed line represents the equilibrium curve.

So, with this nomenclature we can write the point efficiency as E_{OG} equal to $Y_{n,local} - Y_{n+1,local}$ divided by $Y_{n,local}^* - Y_{n+1,local}$, numerator is the actual enrichment of the gas divided by the maximum enrichment possible. So, this will give the point efficiency of that particular tray.

Now, what are the assumptions taken over here, the local concentration $X_{n,local}$ is constant in the vertical direction. This means that, the liquid is vertically well mixed. Thus the concentration is uniform in the vertical direction at any point on the tray. So, this local concentration X_n is constant in the vertical direction and then uniform throughout the tray.

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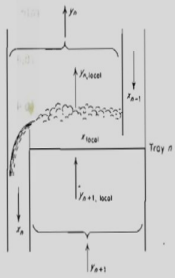
Point efficiency

- The point efficiency is defined as

$$E_{OG} = \frac{Y_{n,local} - Y_{n+1,local}}{Y_{n,local}^* - Y_{n+1,local}}$$

$$= \frac{\text{Actual enrichment of the gas}}{\text{Maximum enrichment}}$$

- Assumptions taken:**
 - ✓ The gas phase is in plug flow i.e., the gas changes along the depth.

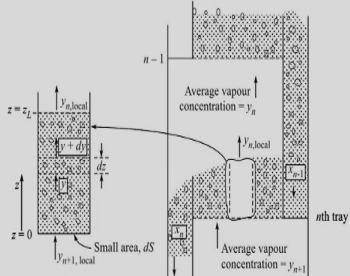


The gas phase is in plug flow, that is the gas changes along the depth it is a plug flow. So, its concentration will change along the depth of the flow.

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Point efficiency

- In order to achieve high tray efficiency, mass transfer coefficient and the interfacial area must be higher.
- Now we will relate the point efficiency with these parameter



In order to achieve high tray efficiency, mass transfer coefficient and interfacial area must be higher. Now we will relate the point efficiency with this parameters interfacial area and the mass transfer coefficient.

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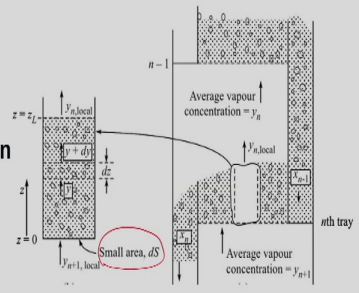
Point efficiency

Let

G_S = Molar gas flow rate and assumed to remain constant.

dL = Small thickness of dispersion

dS = Small area of the tray



- Steady state mass balance over the small thickness dL
- Rate of mass transfer to the gas phase = $G_S dS dy$ 1

Let G_s is the molar gas flow rate and assume to remain constant, dL is the small thickness of dispersion, dS is the small area of the tray which is shown over here. Steady state mass balance over the small thickness dL , then we can write the rate of mass transfer of gas would be equal to gas flow rate into area into thickness. So, this is the rate of mass transfer G_s is the molar flow rate into the differential area of the tray into the change of concentration dy .

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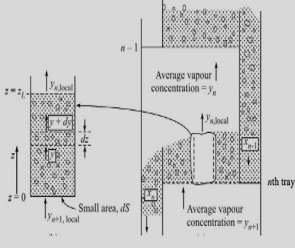
Point efficiency

Assumption:

- The concentration of gas y undergo change in concentration dy in the height dL
- So, it is assumed that the mass transfer occur from gas phase to the liquid phase.
- Again, The rate of mass transfer = $K_y a dS dL (y_{n,local}^* - y)$ 2

Overall gas phase mass transfer coefficient

Specific interfacial area



Now, assuming the concentration of gas y undergoes change in concentration dy in the height dL . So, it is assumed that the gas mass transfer occurs from the gas phase to the liquid phase and again the rate of mass transfer we can write $K_y a$ into dS into dL and $Y_{n,local}^* - Y$. So, capital $K_y a$ is the overall gas phase mass transfer coefficient and a is the specific interfacial area it is area per unit volume of the liquid.

(Refer Slide Time: 27:33)

Point efficiency

The rate of mass transfer = $K_y a dS dL (y_{n,local}^* - y)$ 2

$dS dL =$ Small volume of dispersion

$a dS dL =$ Gas liquid contact area in the small volume of the dispersion

$(y_{n,local}^* - y) =$ Local mass transfer driving force at any height L from the tray floor

The diagram illustrates a distillation tray with height L from the tray floor. A small volume of dispersion $dS dL$ is shown at height z . The local vapor concentration is $y_{n,local}$, and the average vapor concentration is y_n . The average vapor concentration at the next tray level is y_{n+1} . The local mass transfer driving force is $(y_{n,local}^* - y)$.

The rate of mass transfer as we have now defined earlier. So, this is over here and dS into dL is the small volume of dispersion $a dS dL$ is the gas liquid contact area in the small volume of the dispersion and $Y_{n,local}^* - Y$ is the local mass transfer driving force at any height L from the tray floors.

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Point efficiency

- Rate of mass transfer to the gas phase = $G_s dS dy$ 1
- The rate of mass transfer = $K_y a dS dL (y_{n,local}^* - y)$ 2
- Equating (1) and (2) $G_s dS dy = K_y a dS dL (y_{n,local}^* - y)$ ✓
- Integrating over the height
$$\int_{y_{n+1,local}}^{y_{n,local}} \frac{dy}{(y_{n,local}^* - y)} = \int_0^L K_y a \frac{dL}{G_s}$$

$$-\ln \frac{y_{n,local}^* - y_{n,local}}{y_{n,local}^* - y_{n+1,local}} = \frac{K_y a L}{G_s} = N_{tOG}$$

The rate of mass transfer to the gas phase is this one which is already mentioned and then rate of mass transfer we can also correlate with this equation and then if we equate equation 1 and 2, we can get $G_s dy = K_y a dL (y_{n,local}^* - y)$. So, if we integrate over the height. Then we will get integral $y_{n+1,local}$ to $y_{n,local}$ $\frac{dy}{(y_{n,local}^* - y)}$ would be equal to integral 0 to L $\frac{K_y a dL}{G_s}$.

So, from here we can get $-\ln \frac{y_{n,local}^* - y_{n,local}}{y_{n,local}^* - y_{n+1,local}}$ would be equal to $\frac{K_y a L}{G_s}$. So, this is defined as N_{tOG} number of transfer unit of the gas phase N_{tOG} .

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Point efficiency

$$-\ln \frac{y_{n,local}^* - y_{n,local}}{y_{n,local} - y_{n+1,local}} = \frac{K_y a L}{G_s} = N_{tOG}$$

$$-\ln \left[1 - \frac{y_{n,local} - y_{n+1,local}}{y_{n,local}^* - y_{n+1,local}} \right] = -\ln (1 - E_{OG}) = N_{tOG}$$

$$E_{OG} = 1 - e^{-N_{tOG}}$$

N_{tOG} = number of overall gas phase transfer

If N_{tOG} is very high, E_{OG} approaches unity

i.e., the tray approaches ideal tray performance.

So, if we just subtract both sides with minus 1 and then we can write this equation minus ln 1 minus Y n local minus Y n plus 1 local divided by Y n star local minus Y n plus 1 local would be equal to minus ln 1 minus E OG would be equal to N tOG. So, E OG we can write would be equal to 1 minus e to the power N tOG, N tOG as we said number of overall gas phase transfer unit and if N tOG is very high, E OG approaches unity that is; the tray approaches ideal tray performance when N tOG approaches infinity.

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Point efficiency

- In order to obtain E_{OG} , $K_y a$ is needed
- Alternatively, the following correlation may be used to calculate E_{OG} for sieve tray.

$$E_{OG} = 1 - \exp \left[- \frac{-0.0029}{1 + m \frac{C_G}{C_L} \sqrt{\frac{D_G(1-\beta)}{D_L \left(\frac{A}{A_s}\right)}}} Re_f^{0.4136} \left(\frac{h_L}{d_n}\right)^{0.6074} \right]$$

Here, $Re_f = \frac{\rho_G V_h h_L}{\mu_G \beta}$

C_G, C_L = Molar concentration of gas and liquid.
 m = Local slope of equilibrium curve.
 D_G, D_L = Diffusivity of gas and liquid.

In order to obtain E OG, K y a is needed. Alternatively the following correlations may be used to calculate the E OG for sieve tray. So, one of them is E OG would be equal to 1 minus exponential minus 0.0029 divided by 1 plus m C G by C L in to root over D G into 1 minus beta divided by D L into A h by A a whole into R e f to the power point 0.4136 into h L by d h to the power 0.6074 here, R e f is the Reno's number of the fluid which is defined as rho G V h into h capital L divided by mu G beta. C G and C L is the molar concentration of the gas and liquid, m is the locals slope of equilibrium curve and D G and D L are the diffusivity of the gas and liquid.

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Murphree efficiency

- **Assumption:**
- The gas leaving the dispersion at different locations of a tray get mixed up before entering to the upper tray.
- The concentration of the liquid changes as it flows over the tray.
- Let x_n = Concentration of liquid leaving n^{th} tray.

y_n^* = Equilibrium concentration corresponding to x_n

$$E_{MG} = \frac{y_n - y_{n+1}}{y_n^* - y_{n+1}}$$

y_n, y_{n+1} = Average concentration of the gas leaving n^{th} and tray $(n + 1)^{th}$ respectively

Now, there is another efficiency is called the Murphree efficiency, define Murphree efficiency these are the following assumptions taken the gas living the dispersion at different locations of a tray get mixed up before entering to the upper tray. So, the concentration of the liquid changes as it flows over the tray. Let x_n is the concentration of the liquid leaving of the n^{th} tray and y_n^* is the equilibrium concentration co corresponding to x_n .

So, E_{MG} is defined as y_n minus y_{n+1} divided by y_n^* minus y_{n+1} . Here, y_n and y_{n+1} is not the local concentration it is the average concentration of the gas leaving n^{th} and n plus oneth tray respectively. So, in case of the point efficiency we have considered the local concentration, but in this case we have consider the average concentration in case of Murphees tray efficiency.

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Murphree efficiency

- Two causes will be considered depending on the degree of mixing between vapor and liquid phase.

- Case 1:

- ✓ If the liquid on the tray is completely backmixed, the liquid concentration x_n is uniformly everywhere.

$$E_{MG} = E_{OG}$$

Cases will be considered depending on the degree of mixing between vapour and liquid phase. Case 1, if the liquid on the tray is completely back mixed, the liquid concentration x_n is uniform everywhere. So, know for complete back mixing systems the liquid concentration x_n will be uniform throughout the tray. So, E_{MG} would be equal to E_{OG} . So, point efficiency would be equal to the Murphree efficiency.

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Murphree efficiency

- Case 2:

- ✓ If the liquid flows on the tray as plug flow and no axial mixing in the liquid phase.

$$E_{MG} = A \left[\exp\left(\frac{E_{OG}}{A}\right) - 1 \right]$$

Where $A = \frac{L}{mG}$

m = Henry's law constant

Now, for case 2, if the liquid flows on the tray as plug flow and no axial mixing in the liquid phase, in this case, E_{MG} would be equal to A into exponential E_{OG} by A minus

where, A is L by mG it is absorption factor m is the Henry's law constant and L is the liquid flow rate and G is the gas flow rate.

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Overall tray efficiency

$$E_o = \frac{\text{number of ideal trays}}{\text{number of real trays}}$$

If

1. Gas and liquid flow rate G , L and slope of the equilibrium curve remain constant over the section of the column.
2. E_{MG} remains same for all trays.

- Then the relation between E_o and E_{MG} is as follows

$$E_o = \frac{\ln \left[1 + E_{MG} \left(\frac{1}{A} - 1 \right) \right]}{\ln \left(\frac{1}{A} \right)}$$

If $A \sim 1$, then $E_o \approx E_{MG}$

So, overall tray efficiency E_o is the number of ideal trays divided by the number of real trays. If no point 1, if gas and liquid flow rates G and L and the slope of the equilibrium curve remain constant over the section of the column, E_{MG} remains same for all trays. So, then the relation between E_o and E_{MG} is as follows E_o would be equal to $\frac{\ln [1 + E_{MG} (1/A - 1)]}{\ln (1/A)}$ here, A if A tends to 1 then E_o would be equal to no approximately equal to E_{MG} . So, when this absorption factor tends to 1 then, the overall tray efficiency would be equal to the Murphree stage efficiency or tray efficiency.

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Problem : Solution

Ammonia is absorbed by pure water from air-ammonia mixture using a sieve-tray tower. The mixture contains 10% NH₃ and 90% air. It is desired to remove 90% NH₃. The gas enters at the bottom of the tower at a flow rate of 150 kmol/h at 298K and 1atm. The water is fed at the top of the tower at flow rate of 150kmol/h. Assume surface tension of liquid is 72 dyn/cm. The diameter of the sieve is 2 mm which is on an equilateral-triangular pitch of 10mm. The density of the liquid is 1000 kg/m³. The recommended foaming factor is 0.8. Design the tower for an 75% approach to the flooding velocity.

Solution:

$$M_{g,avg} = 0.1 \times 17 + (0.9 \times 29) = 27.8$$

$$G' = (150 \times 27.8) / 3600 = 1.158 \text{ kg/s} \quad \checkmark$$

$$\rho_G = \frac{(P \cdot M_{g,avg})}{RT} = \frac{101.3 \times 27.8}{8.314 \times 298} = 1.137 \text{ kg/m}^3 \quad \checkmark$$

$$\text{NH}_3 \text{ absorbed, } 150 \times 0.1 \times 0.9 \times 17 = 229.5 \text{ kg/hr.}$$

$$L' = \frac{150 \times 18 + 229.5}{3600} = 0.814 \text{ kg/s}$$

Now, let us consider a small problem to look after the design aspect which we have discussed for the tray column. Ammonia is absorbed by pure water from air ammonium mixture using a sieve tray tower. The mixture contains 10 percent ammonia and 70 percent air. It is desired to remove 90 percent ammonia.

The gas enters at the bottom of the tower at a flow rate of 150 kilo mole per hour at 298 Kelvin and 1 atmosphere pressure. The water is fed at the top of the tower at flow rate of 150 kilo mole per hour. Assume surface tension of liquid is 72 dynes per centimetre. The diameter of the sieve tray is given 2 millimetre which is on an equilateral triangular pitch of 10 millimetre. The density of the liquid is 1000 kg per metre cube. The recommended foaming factor is 0.8. Now design the tower for an 75 percent approach to the flooding velocity.

Now, let us move forward to solve this problem step by step. So, average molecular weight we can calculate for the no gas mixtures which are given. We have 10 percent ammonia, 90 percent air; so, 0.1 into 17 the molecular weight of ammonia plus 0.9 into 29. So, it is 27.8. Now, G dash gas flow rate which is given 150 kilo mole per hour. So, we can calculate in terms of kg per second G dash would be no 150 into average molecular weight 27.8 divided by 3600 seconds. So, it would be 1.158 kg per second. Now rho G would be equal to P t M G average divided by RT. So, it is 101.3 total

pressure into the average molecular weight divided by R into T so, which is coming around to be 1.137 kg per metre cube.

Now, ammonia **absorbed**. So, we have a gas flow rate of 150 kilo mole per hour. So, 10 percent of that is 150 into 0.1 and out of that 90 percent has to be **absorbed** so into 0.9 into the molecular weight of the ammonia. So, would be 229.5 kg per hour. So, L dash is also over here is 150 kilo mole per hour for water which is fed at the top. So, it would be 150 into 18 plus 229.5 divided by 3600. So, the water flow rate would be point 1 dash would be 0.814 kg per second.

(Refer Slide Time: 39:03)

Problem : Solution

Ammonia is absorbed by pure water from air-ammonia mixture using a sieve-tray tower. The mixture contains 10% NH₃ and 90% air. It is desired to remove 90% NH₃. The gas enters at the bottom of the tower at a flow rate of 150 kmol/h at 298K and 1atm. The water is fed at the top of the tower at flow rate of 150kmol/h. Assume surface tension of liquid is 72 dyn/cm. The diameter of the sieve is 2 mm which is on an equilateral-triangular pitch of 10mm. The density of the liquid is 1000 kg/m³. The recommended foaming factor is 0.8. Design the tower for an 75% approach to the flooding velocity.

Solution:

$$\rho_L = 1000 \text{ kg/m}^3 \text{ at } 298 \text{ K}$$

$$m = \frac{L'}{G'} \left(\frac{\rho_G}{\rho_L} \right)^{1/2} = \frac{0.814}{1.158} \left(\frac{1.137}{1000} \right)^{1/2} = 0.024$$

$$\frac{A_n}{A_a} = 0.97 \left(\frac{d_n}{P_n} \right)^2 = 0.97 \left(\frac{2}{10} \right)^2 = 0.037$$

Now , assume tray spacing $t_s = 0.6 \text{ m}$

The density rho L is 1000 kg per metre cube at 298 Kelvin, m would be equal to L dash by G dash into rho G by rho L to the power half. So, if we just substitute of values of L dash then G dash into rho G 1.137 which we have calculated and then the density of no liquid which is given 1000 kg per metre cube to the power half. So, will get is around 0.024. A n by A h would be equal to 0.97 d n by P n whole square which is equal to 0.97 2 by 10 whole square. So, which is coming out to be 0.037.

(Refer Slide Time: 40:05)

Problem : Solution

Ammonia is absorbed by pure water from air-ammonia mixture using a sieve-tray tower. The mixture contains 10% NH₃ and 90% air. It is desired to remove 90% NH₃. The gas enters at the bottom of the tower at a flow rate of 150 kmol/h at 298K and 1atm. The water is fed at the top of the tower at flow rate of 150kmol/h. Assume surface tension of liquid is 72 dyn/cm. The diameter of the sieve is 2 mm which is on an equilateral-triangular pitch of 10mm. The density of the liquid is 1000 kg/m³. The recommended foaming factor is 0.8. Design the tower for an 75% approach to the flooding velocity.

Solution: Then , $\alpha = 0.0744 t_s + 0.01173$

$$= 0.0744 \times 0.6 + 0.01173 = 0.0564 \quad \checkmark$$

$$\beta = 0.0304 \times t_s + 0.015$$

$$= 0.0304 \times 0.6 + 0.015 = 0.0332 \quad \checkmark$$

Since , $m < 0.1$, use $m = 0.1$

$$C_F = \alpha \log \frac{1}{m} + \beta = 0.0896 \quad \checkmark$$

Now, assume tray spacing t_s is 0.6 meter, then we can calculate alpha would be equal to 0.0744 into 0.6 plus 0.01173 would be equal to 0.0564. So, alpha we can calculate. Then we can calculate beta using the e_r relations which are provided in the lecture; so which will be around 0.0332. Since, m is less than 0.1, use m is equal to 0.1 as we have said no earlier. So, C_F would be equal to $\alpha \log 1$ by m plus beta. So, beta and alpha we have calculated m which we can use 0.1. So, C_F we can calculate is about 0.0896.

(Refer Slide Time: 41:09)

Problem : Solution

Ammonia is absorbed by pure water from air-ammonia mixture using a sieve-tray tower. The mixture contains 10% NH₃ and 90% air. It is desired to remove 90% NH₃. The gas enters at the bottom of the tower at a flow rate of 150 kmol/h at 298K and 1atm. The water is fed at the top of the tower at flow rate of 150kmol/h. Assume surface tension of liquid is 72 dyn/cm. The diameter of the sieve is 2 mm which is on an equilateral-triangular pitch of 10mm. The density of the liquid is 1000 kg/m³. The recommended foaming factor is 0.8. Design the tower for an 75% approach to the flooding velocity.

Solution: Since $\frac{A_n}{A_t} < 0.1$

$$F_{ha} = 5 \times \left(\frac{A_n}{A_t}\right) + 0.5 = 5 \times 0.037 + 0.5 = 0.618 \quad \checkmark$$

$$F_{st} = \left(\frac{\sigma_l}{20}\right)^{0.2} = \left(\frac{80}{20}\right)^{0.2} = 1.92$$

$$F_f = 0.8$$

Now , $C_{SB} = F_{st} \times F_f \times F_{ha} \times C_f$

Now, since A_n by A_t is less than 0.1, we can use the relation of F_{ha} which is 5 into A_n by A_t plus 0.5 which is equal to 5 into 0.037 plus 0.5 which is equal to 0.618. So, F_{ha} we can calculate and then F_{st} would be equal to a sigma L divided by 20 to the power point 2. So, sigma L which is given over here is taken as 80 times per second. So, 80 divided by 20 to the power 0.2 which is equal to 1.92.

F_f ; F_f is the you know factor flooding factor F_f which is equal to 0.8 which is given and C_{SB} we can calculate using this relation. So, all the parameters F_{st} , F_f , F_{ha} and C_f are known.

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Problem : Solution

Ammonia is absorbed by pure water from air-ammonia mixture using a sieve-tray tower. The mixture contains 10% NH₃ and 90% air. It is desired to remove 90% NH₃. The gas enters at the bottom of the tower at a flow rate of 150 kmol/h at 298K and 1atm. The water is fed at the top of the tower at flow rate of 150kmol/h. Assume surface tension of liquid is 72 dyn/cm. The diameter of the sieve is 2 mm which is on an equilateral-triangular pitch of 10mm. The density of the liquid is 1000 kg/m³. The recommended foaming factor is 0.8. Design the tower for an 75% approach to the flooding velocity.

Solution: Now , $C_{SB} = F_{st} \times F_f \times F_{ha} \times C_f$

$$= 1.292 \times 0.8 \times 0.618 \times 0.0896$$

$$= 0.0572 \text{ m/s}$$

$$Q_G = \frac{G'}{\rho_G} = \frac{1.158}{1.137} = 1.019 \text{ m}^3/\text{s} \quad \checkmark$$

$$V_{fL} = C_{SB} \left(\frac{\rho_L - \rho_G}{\rho_G} \right)^{1/2} = 0.0572 \left(\frac{1000 - 1.137}{1.137} \right)^{1/2}$$

$$= 1.696 \text{ m/s} \quad .$$

So, we can calculate the no values of C_{SB} which is 0.0572 meter per second. Now Q_G would be equal to G' dash by ρ_G , ρ_G we have calculated ρ_G we are given then, G' dash we have calculated. So, Q_G would be equal to 1.019 metre cube per second. So, volumetric flow rate of the gas we can calculate. V_{fL} we can calculate using this relation which is 0.0572 into ρ_L minus ρ_G divided by ρ_G to the power half which is 1.696 meter per second; so V_{fL} we can calculate.

(Refer Slide Time: 43:15)

Problem : Solution

Ammonia is absorbed by pure water from air-ammonia mixture using a sieve-tray tower. The mixture contains 10% NH₃ and 90% air. It is desired to remove 90% NH₃. The gas enters at the bottom of the tower at a flow rate of 150 kmol/h at 298K and 1atm. The water is fed at the top of the tower at flow rate of 150kmol/h. Assume surface tension of liquid is 72 dyn/cm. The diameter of the sieve is 2 mm which is on an equilateral-triangular pitch of 10mm. The density of the liquid is 1000 kg/m³. The recommended foaming factor is 0.8. Design the tower for an 75% approach to the flooding velocity.

Solution: Since $m < 0.1$ m

$$\frac{A_d}{A_t} = 0.1$$

For , 75 % approach of flooding

$$D_t = \left(\frac{4Q_G}{F V_{fL} \left(1 - \frac{A_d}{A_t}\right) \pi} \right)^{1/2} = \left(\frac{4 \times 1.019}{0.75 \times 1.696 (1 - 0.1) \pi} \right)^{1/2} = 1.064 \text{ m}$$

Since the calculated value $1 < D_t < 3$, $t_s = 0.6$ m is the recommended tray spacing and no further interaction needed.

Since m less than 0.1, m we can take A_d by A_t is equal to 0.1. For 75 percent approach of flooding we can calculate D_t which is equal to $4Q_G$ divided by $F V_{fL}$ into $1 - A_d$ by A_t into π to the power half. So, if you substitute the values then we can see 4 into 1.019 divided by 0.75 into 1.696 into $1 - 0.1$ into π to the power half. So, then we will get no the values of D_t is around 1.064 meter.

Now, since the calculated values of diameter is between 1 and 3. So, it is in between 1 and 3 metre and we assume t_s is equal to 0.6 meter which is recommended tray spacing. So, no farther iteration is needed. So, that is within the acceptable limit. So, this way we can do the tower design for the plate towers and with this discussion we will complete our tray tower design discussion and thank you for hearing this lecture and we will continue our design of the towers for packed column in the next lecture.