Principles and Practices of Process Equipment and Plant Design
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Module - 02
Lecture - 18
Design of absorbers (Contd.)

Good day to you all. Today in this session we are going to talk about and continue on the Design of absorbers that we had started in the last session.

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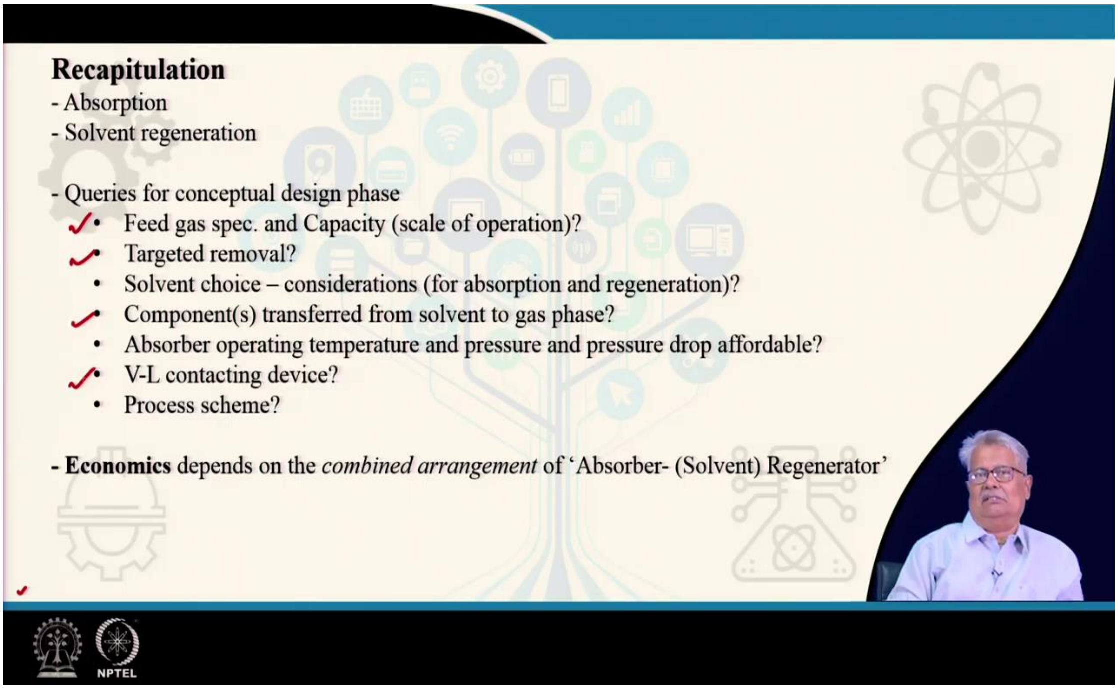

Recapitulation

- Absorption
- Solvent regeneration

- Queries for conceptual design phase
  - ✓ Feed gas spec. and Capacity (scale of operation)?
  - ✓ Targeted removal?
  - Solvent choice – considerations (for absorption and regeneration)?
  - ✓ Component(s) transferred from solvent to gas phase?
  - Absorber operating temperature and pressure and pressure drop affordable?
  - ✓ V-L contacting device?
  - Process scheme?

- **Economics** depends on the *combined arrangement* of 'Absorber- (Solvent) Regenerator'

What we start with is a little bit of recapitulation on the absorption and solvent regeneration. We know that they have to occur in pairs particularly in cases of continuous plants, where solvent circulation is also an important thing and the total cost of the economics depends on the combined arrangement of absorption in the absorber of the solvent and the regeneration of the solvent, the solvent circulating between these two units.

Wherever we are posed with a problem this also we have discussed that we start with a conceptual design phase. So, in the case of absorption the first question comes what is the composition and the flow rate of the feed gas this is also what we have discussed in the last class. We need to know the capacity which is basically the scale of operation whether it's going to be one time or it's going to be a continuous matter for quite some time.

We need to have a target of removal of the component for the gas phase. The choice of solvent is absolutely important because it is the solubility of the solute in the solvent governs the process. But, when you choose the solvent there is not only the consideration for the absorber but the solvent regeneration also. The regeneration often is done by boiling the solute off. So, that consumes a good amount of energy.

It could be by stripping with an inert gas as well or a reboiled stripper. It could also happen that a small distillation column will be used, most of it will be the stripping section. We definitely in our simplified treatment will be considering that it's only the solute component that is transferred for the gas to the liquid phase, but in reality, some small amount of some component or all components of the solvent may also get transferred from the liquid to the gas phase itself.

A common thing is when the gas is unsaturated to the solvent liquid some amount of vapour phase transfer or vaporization of the solvent and joining the gas phase will definitely be happening. It is important to note one thing that is the absorber operating temperature. Normally, the absorption process is associated with the liberation of a small heat of absorption.

So, what we have here is based on the fundamental thermodynamics that absorbers work better at lower temperatures. So, quite often you will find the solvent before entering the absorber is from a cooler. The pressure plays a key role in the solubility of the gas in a liquid. So, quite naturally higher the pressure more will be the carrying capacity of the solvent to the solute.

When we have to talk about the pressure drop in the equipment and definitely the supply pressure of your gas should be sufficient enough so that the affordable pressure drop can be availed from the source. If it is not possible then definitely the pressure has to be boosted by either a blower or a compressor as the case can be.

In the design, we have to decide on the option of the contacting device. In reality, most of the absorbers use either a packed bed or a stage-wise plate tower. Most of the time number of ideal stages required for this contacting is low, typically maybe around 3 or so, it could be even 4, it is not large.

So, absorbers quite often will be using low-efficiency trays like the sieve tray. Because we can always provide quite a bit of an extra number of trays to account for its inefficiency. The typical efficiency of a sieve tray will be of the order of maybe some 40% or so, it could be slightly less than that even.

During the conceptual design phase, we need to have a process scheme and the process scheme should include not only the absorber but the solvent regenerations system also. In the end, we reiterate by saying that the economics of the entire system depends on the combined arrangement of the absorber and solvent regenerator.

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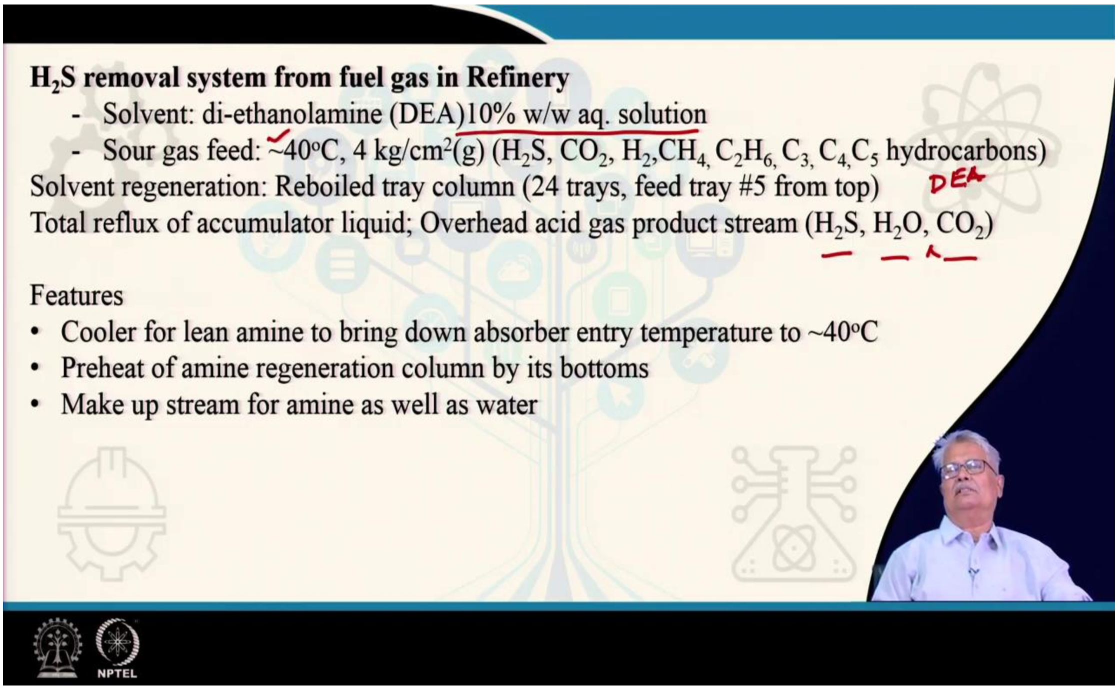

H_2S removal system from fuel gas in Refinery

- Solvent: di-ethanolamine (DEA)10% w/w aq. solution
- Sour gas feed: ~40°C, 4 kg/cm²(g) (H_2S, CO_2, H_2,CH_4, C_2H_6, C_3, C_4,C_5 hydrocarbons)

Solvent regeneration: Reboiled tray column (24 trays, feed tray #5 from top) DEA

Total reflux of accumulator liquid; Overhead acid gas product stream (H_2S, H_2O, CO_2)

Features

- Cooler for lean amine to bring down absorber entry temperature to ~40°C
- Preheat of amine regeneration column by its bottoms
- Make up stream for amine as well as water

We talk about an example of an absorption system. See the fuel gas very often in produced in the refinery which is a hydrocarbon mixture. It contains plenty of things. It contains H_2S, carbon dioxide, hydrogen small amount, methane, ethane, some amount of C_3, C_4 and C_5 hydrocarbons. It is typically called sour due to the presence of H_2S. The available temperature is around 40 °C. Now, this H_2S and carbon dioxide have to be removed. In a refinery the H_2S is sent for recovery of sulfur. So, this is a very important unit not only to remove H_2S and CO_2 but also for providing the raw material which will be used for the production of sulfur. Typically the H_2S and CO_2 are not highly soluble in water or in the aqueous phase. But if you have roughly some alkanol-amine solution in this specific case we are talking about diethanolamine or DEA 10 % aqueous solution. 10% aqueous solution as a solvent. So, the solvent consists of two things; it contains water and 10% by weight of diethanolamine.

Now, DEA 10 % aqueous solution has a high affinity for H_2S and carbon dioxide. So, naturally when the liquid and the gas come in contact with the solvent and the gas feed the H_2S and CO_2 get transferred from the gas to the liquid phase leaving the hydrocarbons and the hydrogen and if there be a small quantity of nitrogen which is also there which is not mentioned here, but it may also be there in some cases leaving in the unabsorbed phase.

The regeneration of the solvent which is rich in H_2S and CO_2 is rather easy. If you boil the solubility decreases and this is done in a reboiled tray column typically with around 24 trays and you have a feed tray which is number 5 from the top. The top product is a gas; primarily H_2S, carbon dioxide and H_2O.

You will notice one thing it will also contain a small quantity of DEA and there is a special configuration which is done to reduce this particular loss that we will see after this. You will notice in the scheme that is presented immediately after this that there is a cooler for lean amine. It brings down the absorber entry temperature to around 40 °C which is very close to the gas supply temperature. In fact, during absorption, the amine temperature rises slightly possibly to around 45 °C or so. Now, to regenerate the rich amine or rather the amine solution which is rich in H_2S and CO_2, it is sent to the next distillation column after preheating it.

We will see there is an arrangement of preheating this particular stream with the column bottom itself. We also have noticed that along with the gas some amount of H_2O will be leaving and we have also mentioned that apart from H_2O you will also have a small quantity of DEA loss from this phase as well.

So, what happens is there has to be a make-up stream of DEA as well as water which can be either continuous or periodic. Typically in refinery, you will find that there will be such make-up once a week. So, this is, in brief, we have an idea of a practical H_2S removal system from the fuel gas.

You will also note here that the choice of the solvent diethanolamine here is for a very specific purpose. It is a weak chemical bond, it forms with H_2S and carbon dioxide which increases the liquid phase or the solvent capacity to absorb much higher. That means, the effective solubility if I use only water and effective solubility that I get if I use 10% the DEA solution will have many times higher capacity to absorb.

That means, the g/m^3 of the solution which is getting circulated would go up substantially. So, this would definitely require less amount of circulation. So, you must choose a solvent that has got a high capacity and you have chosen a solvent to have the high capacity which forms a weak bond, now the question is why weak bond. If I have a solvent that forms a very strong bond with the solute, it becomes very difficult for it to break this bond and release and regenerate the solvent.

So, in this specific case, DEA is found to be quite suitable. In fact, there are quite a few other alkanol-amines like monoethanolamine and MDEA which are also used for removal of H_2S as well as CO_2. This is a very standard absorption regeneration process. Now, what we do is let us look at a practical implementation of this arrangement.

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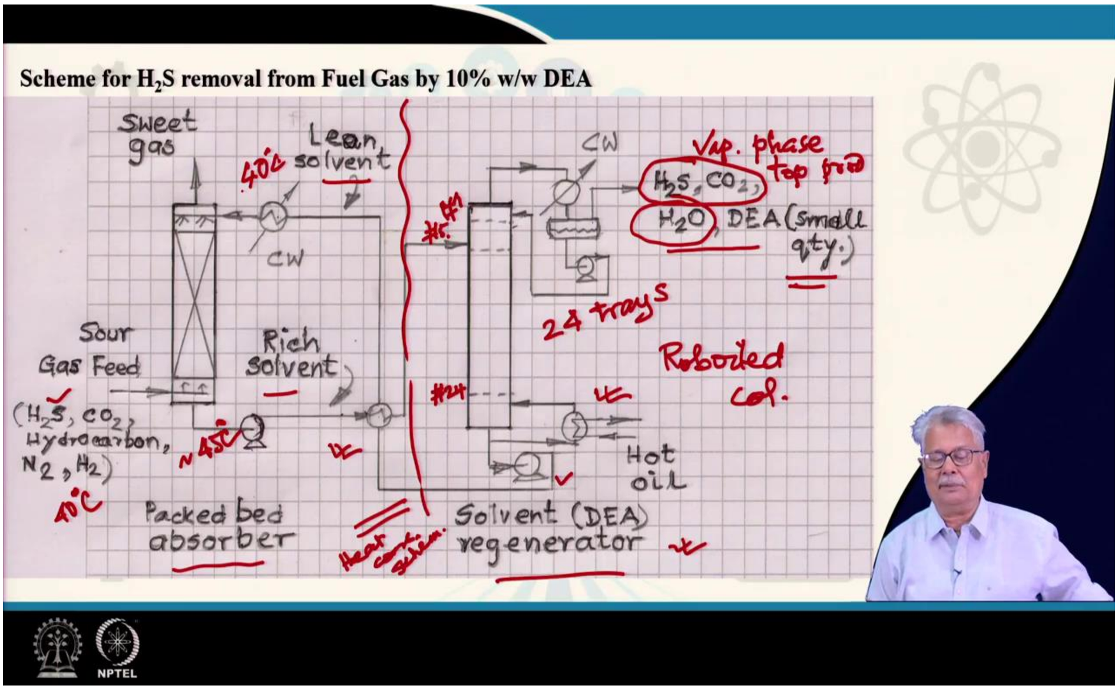

Scheme for H_2S removal from Fuel Gas by 10% w/w DEA

This is the scheme. What you have here is a sour gas feed. It is soured because of the presence of H_2S and it contains these components it has got carbon dioxide, it has got hydrocarbon, it has got a small quantity of nitrogen and hydrogen as well. Here you have a packed tower that is being used as the absorber.

Now, once this H_2S is removed, it will not be sour. So, the exit is a sweet gas. The sweet gas will be containing a very small quantity of residual H_2S and carbon dioxide and the rest amount will be the unabsorbed material which is hydrocarbon, nitrogen and hydrogen. I told you that this feed gas is available at a temperature close to 40 °C. The solvent which has got minimum quantity of H_2S and CO_2 after regeneration in this particular section. So, that is what is coming here to the top of the tower for absorption. Now, this is cooled in a cooler to get to a temperature of close to 40 °C itself.

Often this (exit liquid) temperature will be close to around 45 °C. These are typical operating values for this type of tower. I iterate once more that the solvent is lean in the solute and this solvent after absorption becomes rich in the solute. So, this is the rich solvent, this is a lean solvent and the solvent keeps on circulating between these two units.

Now, let us look at the process which we have crossly described in the previous slide and we have talked about it is here the regenerator. In this regenerator, as it looks like what you have here is a reboiled column. What we have mentioned? There are 24 trays, the feed is on number 5, the overhead reflux is on number 1. So, the bottom tray is number 24.

You will see here that below the feed we have a large number of trays that is from 5 to 24 these are the stripping trays. So, you have a good amount of stripping done which means, as the rich amine falls through these trays the vapour which is generated here by the reboil strip out the H_2S and CO_2 and it finally, goes out. Possibly you have a query now that why is the feed on number 5 and not just like an ideal stripper it is on number 1 that is not on the top tray.

Here it is very important to the engineering of this type of column. One thing you will find is that this is a column in which there is only a vapour draw off. It is a vapour phase that is drawn off as the top product. The vapour comes here, it goes to the condenser, it condenses and comes to the accumulator, the entire liquid is refluxed back to the top tray.

So, this is in industrial parlance, but it is not exactly the way the textbook says. This is called a total reflux column operation. But in fact, that is also a misnomer in the textbook terms, but what is meant here is that the top product is only a vapour. So, what you have at the top is a partial condenser.

Now, what you have in addition is the question to be still answered is why is the feed on number 5. The purpose that we have the top product ideally would have been the H_2S and CO_2 only. But, what is refluxed here is primarily H_2O.

I could have given the feed on the top tray itself, but in that case, some amount of DEA would have gone to the top also. It would have vaporized and would have gone to the top. That is the reason you have the feed on the 5th tray and not on the top primarily to reduce the loss of DEA.

So, with this configuration, you have only a small quantity of DEA which is lost as an overhead vapour. Now it is a reboiled column. Typically it will be heated with hot oil. Possibly some hot oil stream that requires cooling within the refinery will be used as a source of heat in the reboiler itself.

Here what you find is another thing which you have learnt in case of your distillation column also that you need to preheat your feed. That means that your distillation requires energy to be supplied and one source of energy is your reboiler. The other source of energy is your preheating heat exchanger.

In your distillation column, the bottom temperature is higher than the feed temperature. Quite naturally there can be some amount of heat that can be transferred from the bottom stream here to the column feed and which is done in this particular heat-saving scheme; basically, it is a heat conservation scheme.

Anyway still you will not be reaching 40 °C because at the bottom of the temperature will be around 115, 120 °C, but it cannot just come down to around 40 °C. So, you usually will be having a cooling water heat exchanger which will be cooling down the temperature to around 40 °C. So, what you do now is we have an idea that how an absorber regenerator system works in the industry. We have a typical industrial example.

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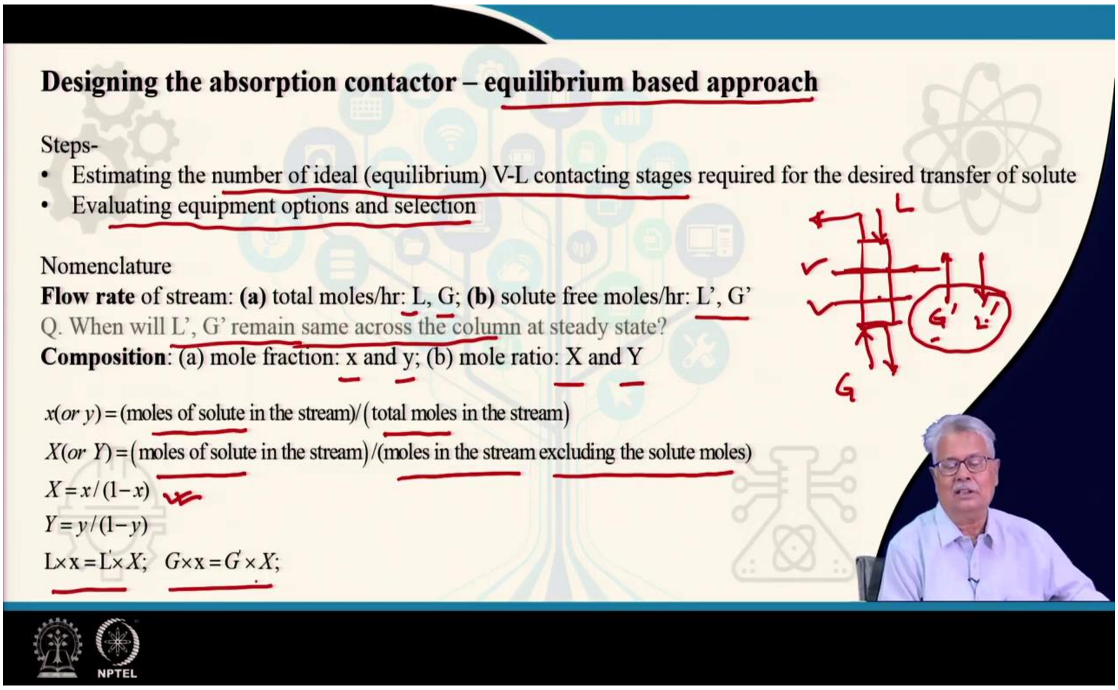

Designing the absorption contactor – equilibrium based approach

Steps-
- Estimating the number of ideal (equilibrium) V-L contacting stages required for the desired transfer of solute
- Evaluating equipment options and selection

Nomenclature

Flow rate of stream: **(a)** total moles/hr: L, G; **(b)** solute free moles/hr: L’, G’

Q. When will L’, G’ remain same across the column at steady state?

Composition: (a) mole fraction: x and y; (b) mole ratio: X and Y

$x(or\ y) = $(moles of solute in the stream)/(total moles in the stream)

$X(or\ Y) = $(moles of solute in the stream)/(moles in the stream excluding the solute moles)

$X = x/(1-x)$

$Y = y/(1-y)$

$L\times x = L'\times X;\quad G\times x = G'\times X;$

We now move to the actual design. We look at the steps of designing the contactor. We already have said that right at the beginning of the course that we are going to have an equilibrium based approach. That means, what we intend to do is to find out the number of ideal stages which means that the vapour and the liquid streams leaving will be at equilibrium. We find out the number of such stages which is the number of ideal stages then we implement it.

So, we have are two steps. The first is the estimation of the number of ideal vapour-liquid contacting stages required for the separation, after all, it is a separation equipment and the separation process. Then we evaluate the equipment options and select and then after we go for the detailed design of this.

Before we move further there is something which I want you to note here at this particular moment that is we are going to use a specific convention here. We are going to say that your L and G. These are going to be the total molar flow rates in mol/h.

Similarly, the corresponding number of moles present in that stream except the solute are denoted by L' and G'. I have a question here. When will the L' and G' remain the same across a column? That means if I have my column here if I have the gas entering here my liquid entering here and the liquid coming out from here and my gas going out from here.

In that case, if I take any section I have a stream of gas $G^{'}$ going up and I have a stream of $L^{'}$ which is coming down.

Now, my question to you is under which condition or which assumption $G^{'}$here and the $G^{'}$ here is the same and if it is the same here and here your $L^{'}$ here and $L^{'}$ at this particular point will also have to be the same. So, what you do now is have an idea that when can this happen. What is $G^{'}$? $G^{'}$ is the number of moles in that particular gaseous phase that is moving up except the solute component.

So, quite naturally if any of these components do not get transferred to the liquid phase my $G^{'}$will be the same at this location and this location also? So, that is the condition and the same thing is true in case of $L^{'}$ also. If no component of the liquid gets transferred I mean except the solute there is no component of L that gets transferred to the gas phase during its travel from the top to the bottom of my column, in that case, indeed $L^{'}$ will also remain constant.

We define the compositions now. The conventional composition we have been using for mass transfer is the mole fraction. x denoting the liquid phase mole fraction and y denoting the vapor phase mole fraction. The mole ratio capital X and capital Y means it's defined here very explicitly. x or y means moles of solute in the stream divided by the total moles. Capital X or capital Y means moles of solid in the stream, it is the same divided by the moles in a stream excluding the solute moles.

So, quite naturally just by the solute balance.

You can say: $X = {}^{x}/_{(1-x)}$ and $Y = {}^{y}/_{(1-y)}$

It can also be said in that particular case that L into x what does it give? L into x gives basically the moles of solute in my stream and L dash into capital X will also be giving you the same.

$$L \times x = L' \times X$$

$$G \times y = G' \times Y$$

So, naturally, this is equality. We are going to use these equalities after this when we frame up the design equations.

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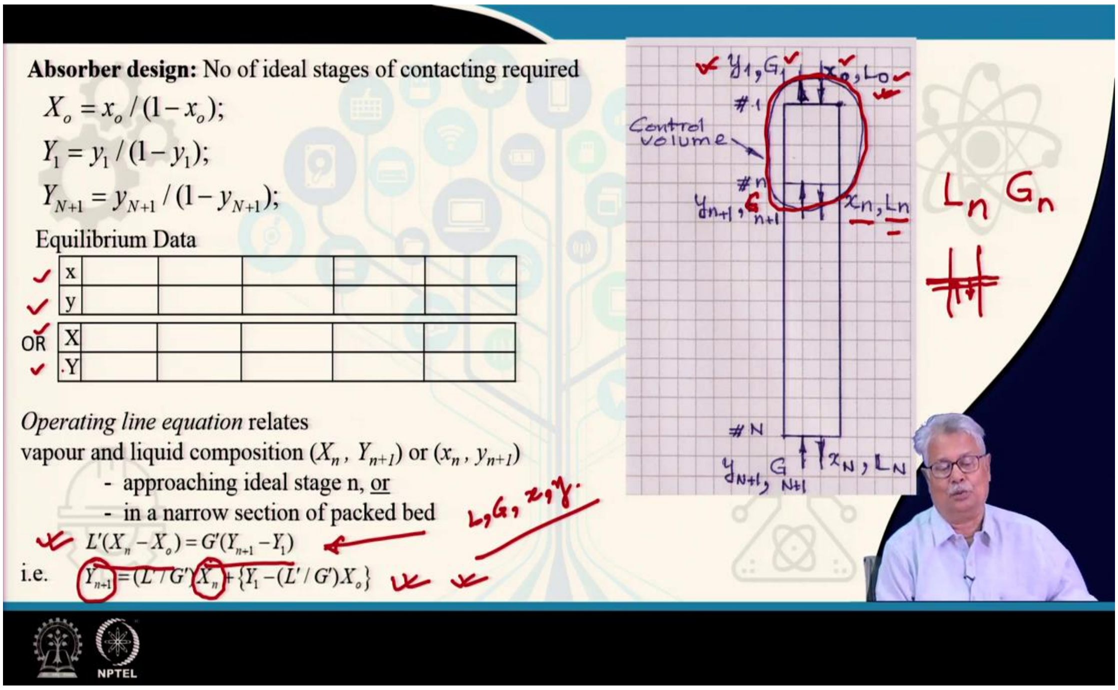

Absorber design: No of ideal stages of contacting required

$X_o = x_o/(1-x_o)$;

$Y_1 = y_1/(1-y_1)$;

$Y_{N+1} = y_{N+1}/(1-y_{N+1})$;

Equilibrium Data

x					
y					

OR

X					
Y					

Operating line equation relates
vapour and liquid composition (X_n, Y_{n+1}) or (x_n, y_{n+1})

- approaching ideal stage n, or
- in a narrow section of packed bed

$L'(X_n - X_o) = G'(Y_{n+1} - Y_1)$

i.e. $Y_{n+1} = (L'/G')X_n + \{Y_1 - (L'/G')X_o\}$

We come to the state of designing now. The number of contact ideal stages for contacting required is the purpose of the design step. We have simply written here. We have drawn here the column itself. I have the top stage is number 1, the bottom stage is capital N. Any general stage is small n and what the convention that we follow is if it is coming from a particular tray number n, if the liquid falling from this will be L_n and G_n will be telling us the gas which is leaving this particular stream. So, quite naturally if n is this particular ideal stage and I have one more stage below which is going to be number (n + 1).

This liquid stream is going to be L_n, its composition x_n and similarly since the vapour here is going to come from here having flow rate G_{n+1}, and having composition y_{n+1}. So, what I do here is that I have an external liquid that is coming from the top which is my solvent and it is coming into the column at stage number 1. So, 1 ideally if it exists at all we have denoted this particular stream with a subscript of 0 and its flow rate is L_0 and its composition is x_0. Exactly in a similar way since the top stage is 1, the gas leaving is having a flow rate of G_1 and the corresponding composition is y_1. Now, we simply draw a control volume. We draw a control volume and try to do a solute balance here.

What we find here is if I say my L' is a solute free flow rate in that case my L' when it falls from a particular external point to a particular point L_n, the composition corresponding to x_n will be capital X_n. So, the change in composition from here to here for liquid solvent is going to be $L'(X_n - X_0)$ and in case of gas it is going to be $G'(Y_{n+1} - Y_1)$. Here X_n will be higher than X_0 because it picks up the solute.

Similarly, Y_{n+1} is also higher than Y_1 because it's more close to the entry where the concentration of the solute or the gas phase is higher. Similarly, the same thing I could have also written in terms of L, G, small x and small y also. In fact, in the case of absorption, both the conventions and the units are used. I just rearrange my upper equation here and this becomes this particular expression.

$$Y_{n+1} = \left({}^{L'}/_{G'}\right) X_n + \left\{Y_1 - \left({}^{L'}/_{G'}\right) X_0\right\}$$

This has a very interesting point here right now. It shows the relationship between X_n and the relationship and the corresponding vapour. That means if I am talking about the column at any particular point it's relating the liquid composition with the vapour composition in this thin section. That means Y_{n+1} is the vapour approaching the nth ideal stage and X_n is basically the composition of the liquid leaving that particular stage and falling down. This is a generalized relationship. So, this definitely will be valid all through. So, I could also remove this Y and X and I could write here this as a generalized expression also. Now, I move forward and we will look at the data which is available to me.

I can have that equilibrium data at the average operating condition of that column within terms of either small x and y or in terms of the capital X and Y. We know what is an interchangeable formula. If I have small x, small y, I could convert it to capital X capital Y or I could do the other way also if one is available the other is also known.

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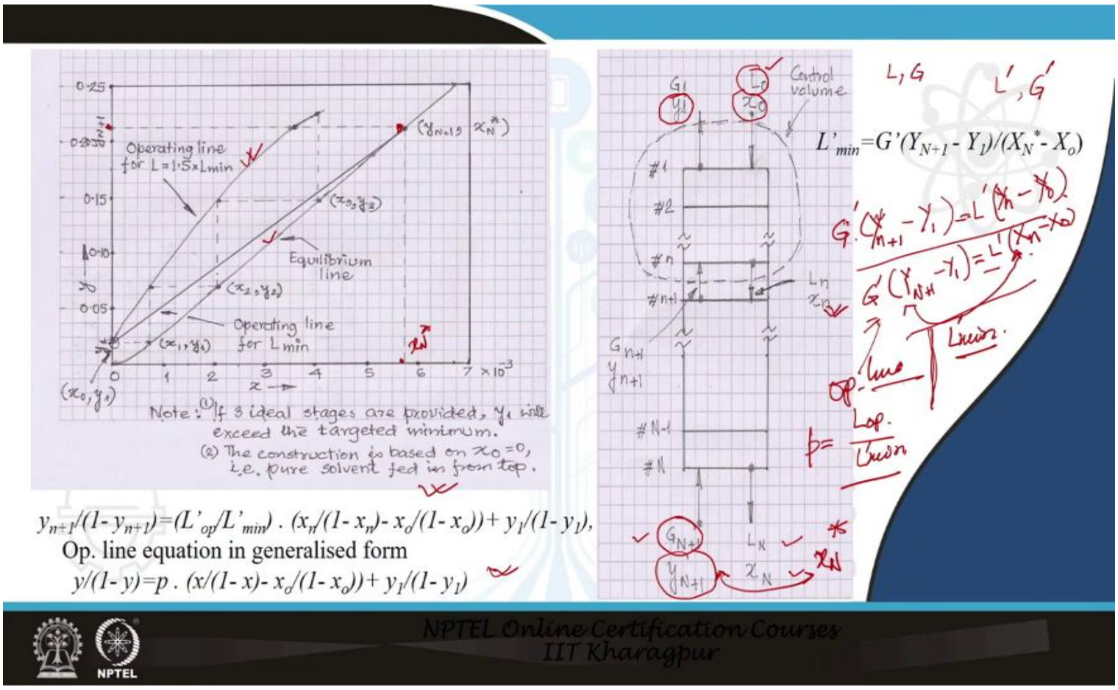

$$L'_{min}=G'(Y_{N+1}-Y_1)/(X_N^*-X_o)$$

$$y_{n+1}/(1-y_{n+1})=(L'_{op}/L'_{min})\,.\,(x_n/(1-x_n)-x_o/(1-x_o))+y_1/(1-y_1),$$

Op. line equation in generalised form

$$y/(1-y)=p\,.\,(x/(1-x)-x_o/(1-x_o))+y_1/(1-y_1)$$

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So far, what we have understood is we are going to use two definitions of the flow rate. L' typically or G' which means it's solute free and otherwise it's either L or G which is the total moles per hour flow rate.

So, we have a look at the design using graphical procedure. What we have here is the gas coming in. This one is a Nth tray and this one is the top one. So, naturally G_{N+1} comes in and G_1 goes out as per our convention of leaving and entering a particular ideal stage.

Similarly, L_0 is the external solvent in flow with a corresponding composition of x_0 and whatever comes out after picking up the solute is L_N and x_N correspondingly. What we do now is basically we take a control volume and we make a solute balance. If you make a solute balance here, what we find is the solute entering from this is G'. The G' remains all through the same because without the solute and if we assume there is no liquid component going to the gas phase your G' will be all through same. The moles of solute free gas which goes up the column is the same in every stage.

So, naturally G' multiplied by whatever is a corresponding composition is basically capital Y. We know here the capital Y denotes the solute free composition in n plus 1 minus capital Y 1 is the amount of solute transfer in this particular section. And what is where does it go? It goes to L dash which is basically the liquid which falls and it gets enriched up to Y_n and the enrichment starts from the composition of Y_0.

$$G'(Y_{n+1} - Y_1) = L'(X_n - X_0)$$

So, this is the balance.

This is an equation that relates X_n and Y_{n+1}, which are the composition of the two components the gas and the liquid phases here and here. We know also that my G' and L' is the same all through.

In your design problem, what are the things known? My L_0 will be known, my x_0 will be known, my gas in flow rate will be known, my gas composition will be known, the desired gas composition at the outlet will also be known. So, here if you see if we somehow know $G'$and L' the relationship between Y_{n+1} and X_n is fixed.

We know the relationship between the capital Y and small y and the capital X and small x. So, quite naturally we also have an equivalent relationship between this small x and small y in a general sense as well. So, what you have here is that basically the equation of the operating line which is also shown here in the generalized form.

Now, there is one more thing that we have to keep in mind, this is for a particular value of your L operating. Here we are using a parameter L' operating by L'_{min} which is equal to ‘p’, which is also a design input.

$$p = \frac{L'_{op}}{L'_{min}}$$

So, to have this we need to find what exactly is the L'_{min}.

L'_{min} is a minimum flow of solvent required to remove this much quantity of solute. When can this happen? This can happen only when my X_N^* here is an equilibrium composition corresponding to the value Y_{N+1}.

$$L'_{min} = G'(Y_{N+1} - Y_1)/(X_N^* - X_0)$$

So, if I have my x-y diagram here on which this is my equilibrium line that we have already marked here as an equilibrium line.

We have marked here the y_{N+1} point and in the corresponding point of X_N^* is going to be this. Now, we have this with me. In this G' is same. My L' this time is L'_{min} and here all parameters are known and this x_N is basically X_N^*. So, we can find out the only unknown parameter in this particular equation utilizing the same equation or the solute balance which is L'_{min}.

If you have decided typically on a p-value which is about 1.5.

$$L'_{op} = p \times L'_{min} = 1.5 \times L'_{min}$$

The corresponding equation is also possible to plot here.

$$y/(1-y) = p(x/(1-x)) - x_0/(1-x_0) + y_1/(1-y_1)$$

You will notice one thing that this operating line equation is above the equilibrium line equation in the case of the absorption phenomena and in the case of stripping it will be below this. The procedure is the same.

With this, I think I will stop here and we will illustrate the details of the procedure with a numerical example in the next class.

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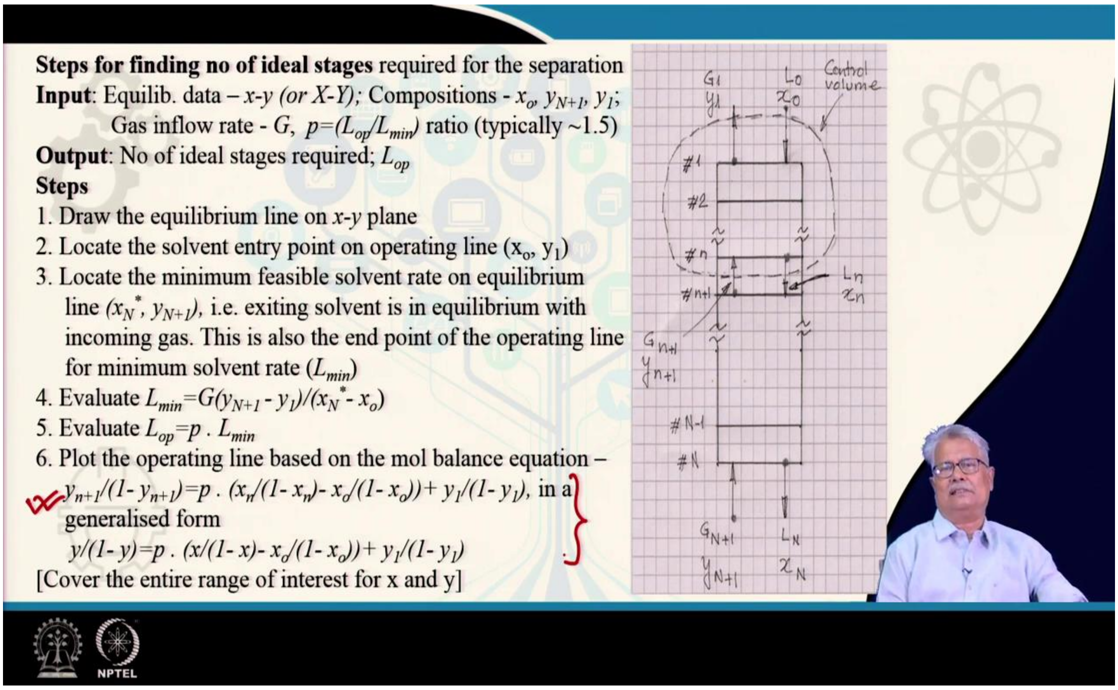

Steps for finding no of ideal stages required for the separation

Input: Equilib. data – x-y (or X-Y); Compositions - x_o, y_{N+1}, y_1;
Gas inflow rate - G, $p=(L_{op}/L_{min})$ ratio (typically ~1.5)

Output: No of ideal stages required; L_{op}

Steps

1. Draw the equilibrium line on x-y plane
2. Locate the solvent entry point on operating line (x_o, y_1)
3. Locate the minimum feasible solvent rate on equilibrium line (x_N^*, y_{N+1}), i.e. exiting solvent is in equilibrium with incoming gas. This is also the end point of the operating line for minimum solvent rate (L_{min})
4. Evaluate $L_{min}=G(y_{N+1}-y_1)/(x_N^*-x_o)$
5. Evaluate $L_{op}=p\,.\,L_{min}$
6. Plot the operating line based on the mol balance equation – $y_{n+1}/(1-y_{n+1})=p\,.\,(x_n/(1-x_n)-x_o/(1-x_o))+y_1/(1-y_1)$, in a generalised form
$y/(1-y)=p\,.\,(x/(1-x)-x_o/(1-x_o))+y_1/(1-y_1)$

[Cover the entire range of interest for x and y]

The steps are written here very clearly. The operating line equation is also here in a generalized form using that particular p. It could also be written in terms of capital X, capital Y if you are going to use the coordinates capital X and capital Y also and the steps are very simple. Draw the equilibrium line. Locate the solvent entry point on the operating line which is the (x_0, y_1). The bottom left graph point that we have started with.

Locate the minimum feasible solvent rate on the equilibrium line which is that x_N^*. We find out x_N^*. Based on that we calculate and find out what exactly is a L_{min}. Then evaluate what based on the value we have for p what exactly is going to be my L_{op} and based on the p value we have adapted we find out the equation or the operating line for L_{op}.

Now, here you will also notice something. If you are using a small x, small y coordinate your operating line will also be non-linear. The generalized formula expression is given over here and basically it is to be drawn covering the entire range of interest for x and y.

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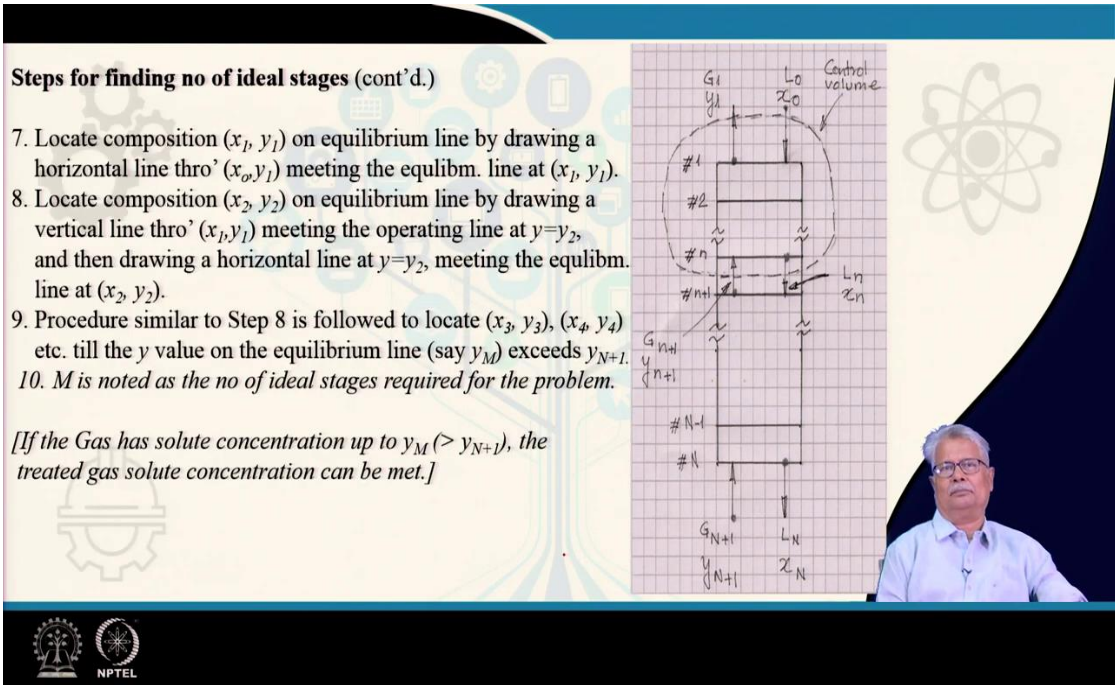

Steps for finding no of ideal stages (cont'd.)

7. Locate composition (x_1, y_1) on equilibrium line by drawing a horizontal line thro' (x_o,y_1) meeting the equlibm. line at (x_1, y_1).
8. Locate composition (x_2, y_2) on equilibrium line by drawing a vertical line thro' (x_1,y_1) meeting the operating line at $y=y_2$, and then drawing a horizontal line at $y=y_2$, meeting the equlibm. line at (x_2, y_2).
9. Procedure similar to Step 8 is followed to locate (x_3, y_3), (x_4, y_4) etc. till the y value on the equilibrium line (say y_M) exceeds y_{N+1}.
10. *M is noted as the no of ideal stages required for the problem.*

[If the Gas has solute concentration up to y_M (> y_{N+1}), the treated gas solute concentration can be met.]

Now, the procedure is simple and we have found out that with three stages which are noted to be the number of ideal stages required for this problem which is M or number 3 in this particular case.

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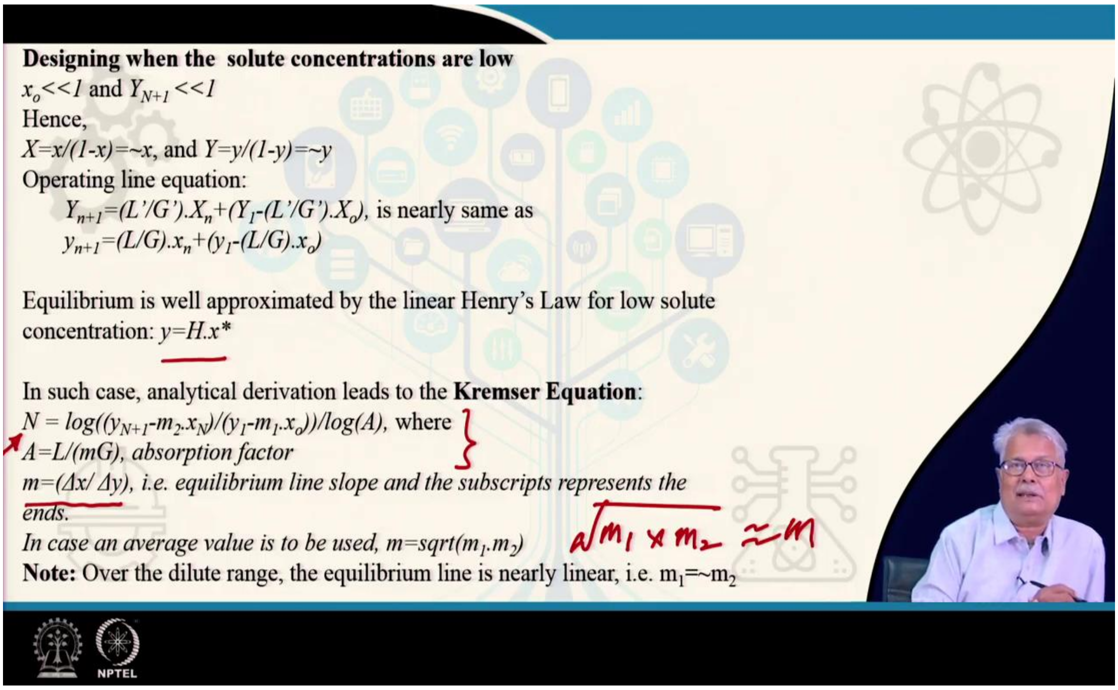

Designing when the solute concentrations are low

$x_o << 1$ and $Y_{N+1} << 1$

Hence,

$X = x/(1-x) =\sim x$, and $Y = y/(1-y) =\sim y$

Operating line equation:

$Y_{n+1} = (L'/G').X_n + (Y_1 - (L'/G').X_o)$, is nearly same as

$y_{n+1} = (L/G).x_n + (y_1 - (L/G).x_o)$

Equilibrium is well approximated by the linear Henry's Law for low solute concentration: $y = H.x^*$

In such case, analytical derivation leads to the **Kremser Equation**:

$N = \log((y_{N+1} - m_2.x_N)/(y_1 - m_1.x_o))/\log(A)$, where

$A = L/(mG)$, *absorption factor*

$m = (\Delta x/\Delta y)$, *i.e. equilibrium line slope and the subscripts represents the ends.*

In case an average value is to be used, $m = \mathrm{sqrt}(m_1.m_2)$ $\sqrt{m_1 \times m_2} \simeq m$

Note: Over the dilute range, the equilibrium line is nearly linear, i.e. $m_1 =\sim m_2$

The next stage is a situation a very special case when your solute concentrations are really low in that case you can derive an analytical expression, based on the linearity of your operating line as well as the equilibrium line. This equation will give you directly the number of theoretical stages as N.

$$N = \log((y_{N+1} - m_2 x_N)/(y_1 - m_1 x_0))/\log(A)$$

It is shown in the above expression where which involves a slope of the operating line which is Δx/Δy that is the equilibrium line slope and the subscript represent the ends.

In case an average value is to be used typically it is the square root of $m_1 * m_2$ which is used as the “m” value. This is valid over a dilute range only the equilibrium line is linear nearly linear when I have m_1 and m_2 same.

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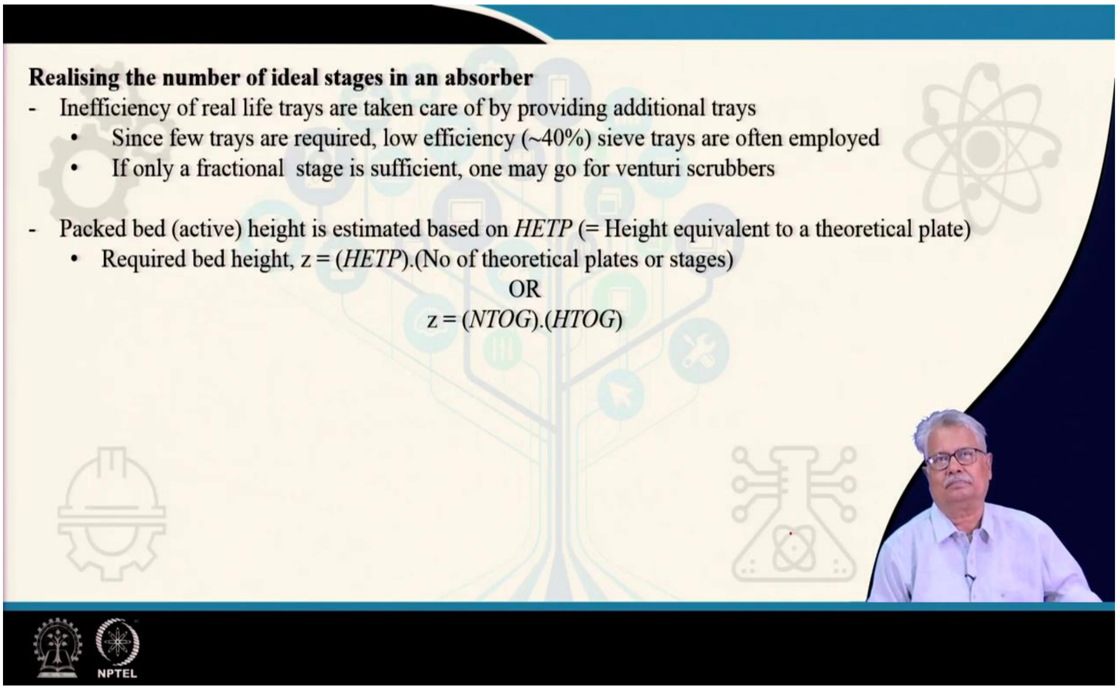

Realising the number of ideal stages in an absorber

- Inefficiency of real life trays are taken care of by providing additional trays
  - Since few trays are required, low efficiency (~40%) sieve trays are often employed
  - If only a fractional stage is sufficient, one may go for venturi scrubbers

- Packed bed (active) height is estimated based on *HETP* (= Height equivalent to a theoretical plate)
  - Required bed height, z = (*HETP*).(No of theoretical plates or stages)

OR

z = (*NTOG*).(*HTOG*)

Next comes the realization of the number of ideal stages in practice. We take care of the inefficiency of the real-life by considering non-equilibrium which means, a factor of efficiency and we carry this on in the next class for finding out the details.