Principles and Practices of Process Equipment and Plant Design
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Module - 02
Lecture - 33
Packed bed adsorber design

Hello everybody. Today we have come to the second lecture on packed bed adsorber design. Initially I had given you an introduction on adsorption, the mechanisms, the types of adsorber beds and then we had discussed in details regarding the packed bed adsorber.

We had discussed that it is an unsteady state operation and I had also given you the concepts of the breakthrough curve, breakthrough point, saturation point, exhaustion bed exhaustion, etc.

So, today we will be using those concepts and we will be discussing about the design of a packed bed adsorber.

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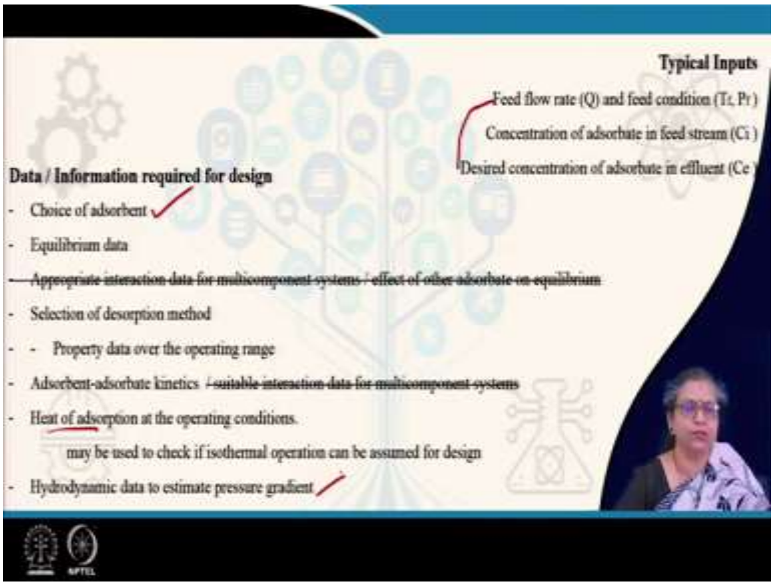

Typical Inputs

Feed flow rate (Q) and feed condition (Tt, Pr)

Concentration of adsorbate in feed stream (Ci)

Desired concentration of adsorbate in effluent (Ce)

Data / Information required for design

- Choice of adsorbent ✓
- Equilibrium data
- ~~Appropriate interaction data for multicomponent systems / effect of other adsorbate on equilibrium~~
- Selection of desorption method
- - Property data over the operating range
- Adsorbent-adsorbate kinetics ~~/ suitable interaction data for multicomponent systems~~
- Heat of adsorption at the operating conditions.
  may be used to check if isothermal operation can be assumed for design
- Hydrodynamic data to estimate pressure gradient ✓

Now, tell me one thing generally what we find? Whenever we go for the packed bed adsorber just like all other problems we find that the problems are usually ill posed, normally in this particular case the feed flow rate is given, the feed conditions are given definitely, the concentration of the adsorbate in the feed stream has to be given. Now, this particular adsorbate has to be removed for purification of the feed stream or for recovery of the adsorbate from the feed stream. So, naturally there has to be some particular desired concentration of the adsorbent in the effluent, this is also specified.

Now, based on these inputs you are supposed to design an adsorber bed. What are the basic information that you require? Definitely first, for this particular adsorbate you would want to have the maximum possible adsorption, first thing naturally is you have to choose a suitable adsorbent. Now, once you have chosen a suitable adsorbent then we have to look for the equilibrium data or we have to generate the equilibrium data in the laboratory. After that if we assume that there is only a single adsorbate which is being adsorbed on the solid then after that we have to select a suitable desorption method.

And then we have to we have to find out the property data over the operating range and naturally see for most of the designs we assume isothermal operation. So, just to check whether it can be isothermal or not we need to find out the heat of adsorption and also we need hydrodynamic data to estimate the pressure gradient.

If it is multi component adsorption then in that case definitely we would need additional data on interaction parameters, etc. Since we are not going to deal with multi component adsorption I am not going to discuss those.

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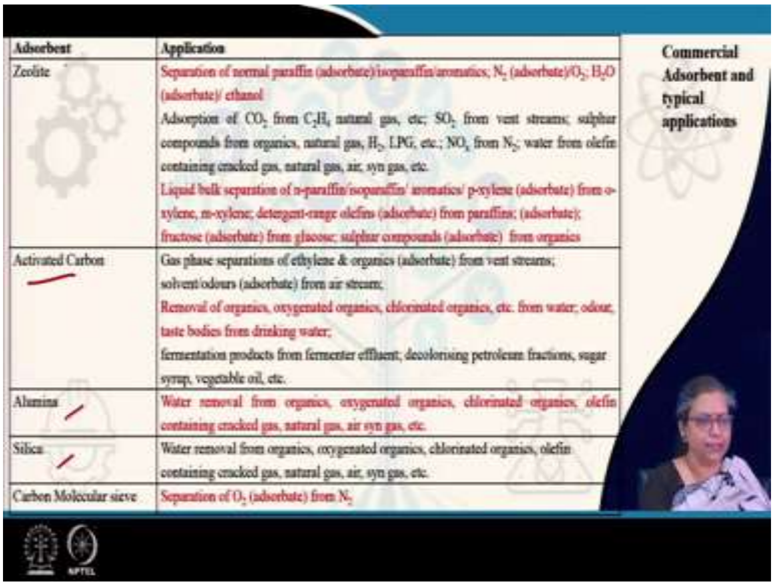

Commercial Adsorbent and typical applications

Adsorbent	Application
Zeolite	Separation of normal paraffin (adsorbate)/isoparaffin/aromatics; N_2 (adsorbate)/O_2; H_2O (adsorbate)/ ethanol Adsorption of CO_2 from C_2H_4 natural gas, etc; SO_2 from vent streams; sulphur compounds from organics, natural gas, H_2, LPG, etc.; NO_x from N_2; water from olefin containing cracked gas, natural gas, air, syn gas, etc. Liquid bulk separation of n-paraffin/isoparaffin/ aromatics/ p-xylene (adsorbate) from o-xylene, m-xylene; detergent-range olefins (adsorbate) from paraffins; (adsorbate); fructose (adsorbate) from glucose; sulphur compounds (adsorbate) from organics
Activated Carbon	Gas phase separations of ethylene & organics (adsorbate) from vent streams; solvent/odours (adsorbate) from air stream; Removal of organics, oxygenated organics, chlorinated organics, etc. from water; odour, taste bodies from drinking water; fermentation products from fermenter effluent; decolorising petroleum fractions, sugar syrup, vegetable oil, etc.
Alumina	Water removal from organics, oxygenated organics, chlorinated organics, olefin containing cracked gas, natural gas, air syn gas, etc.
Silica	Water removal from organics, oxygenated organics, chlorinated organics, olefin containing cracked gas, natural gas, air, syn gas, etc.
Carbon Molecular sieve	Separation of O_2 (adsorbate) from N_2

So, let us start with the choice of adsorbent, now here I have given a list of the commonly used adsorbent for different sort of applications. For example, suppose it will be water treatment, we go for activated carbon, if you want to remove water from say organics, etc., we either opt for alumina or silica or may be a combination of the two. For air separation it is carbon molecular sieve and definitely for a large number of operations we can we also go for the zeolite.

So, depending upon the feed, depending upon the adsorbate which has to be removed, we are first going to select the adsorbent, once we have selected the adsorbent what next we will have to do.

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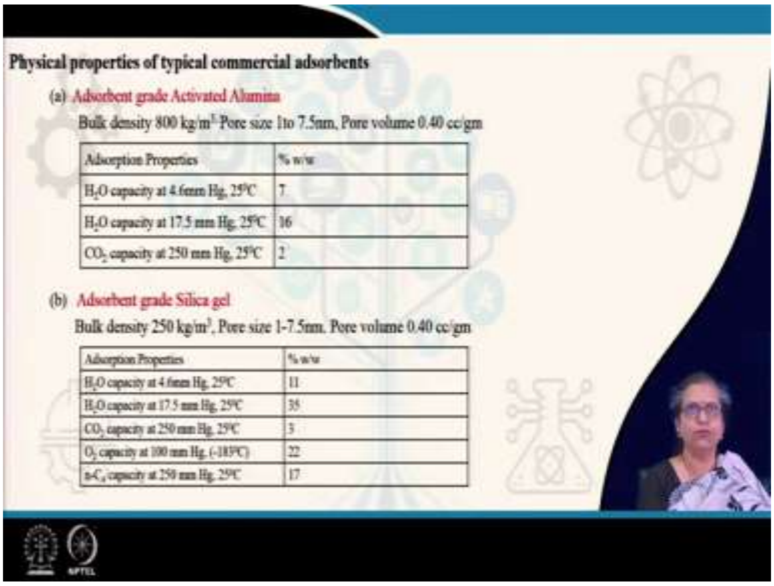

Physical properties of typical commercial adsorbents

(a) Adsorbent grade Activated Alumina

Bulk density 800 kg/m^3 Pore size 1to 7.5nm, Pore volume 0.40 cc/gm

Adsorption Properties	% w/w
H_2O capacity at 4.6mm Hg, 25°C	7
H_2O capacity at 17.5 mm Hg, 25°C	16
CO_2 capacity at 250 mm Hg, 25°C	2

(b) Adsorbent grade Silica gel

Bulk density 250 kg/m^3, Pore size 1-7.5nm. Pore volume 0.40 cc/gm

Adsorption Properties	% w/w
H_2O capacity at 4.6mm Hg, 25°C	11
H_2O capacity at 17.5 mm Hg, 25°C	35
CO_2 capacity at 250 mm Hg, 25°C	3
O_2 capacity at 100 mm Hg, (-183°C)	22
n-C_4 capacity at 250 mm Hg, 25°C	17

We will have to find out the physical properties. So, just for the convenience I have put down the physical properties of the conventional adsorbents which are normally used in packed bed adsorbers in industries.

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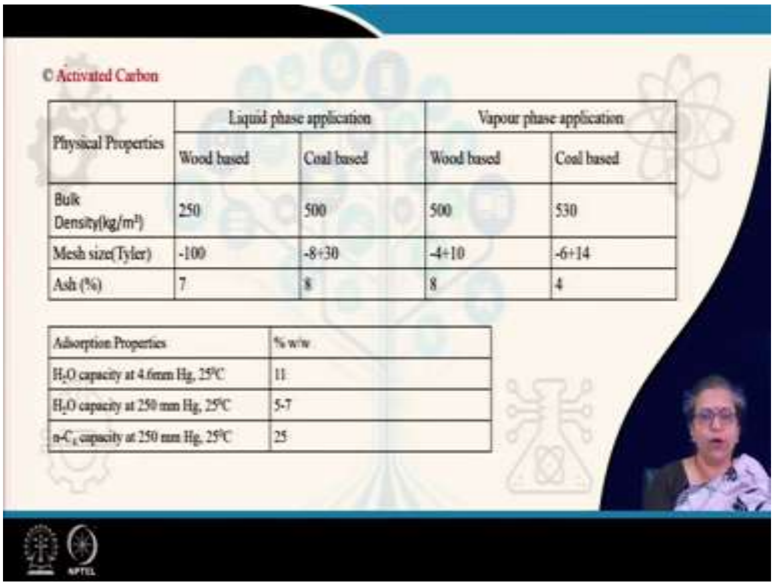

© Activated Carbon

Physical Properties	Liquid phase application		Vapour phase application	
	Wood based	Coal based	Wood based	Coal based
Bulk Density(kg/m^3)	250	500	500	530
Mesh size(Tyler)	-100	-8+30	-4+10	-6+14
Ash (%)	7	8	8	4

Adsorption Properties	% w/w
H_2O capacity at 4.6mm Hg, 25°C	11
H_2O capacity at 250 mm Hg, 25°C	5-7
n-C_4 capacity at 250 mm Hg, 25°C	25

And you will find that in one of the problems we are going to refer to those and design the adsorber.

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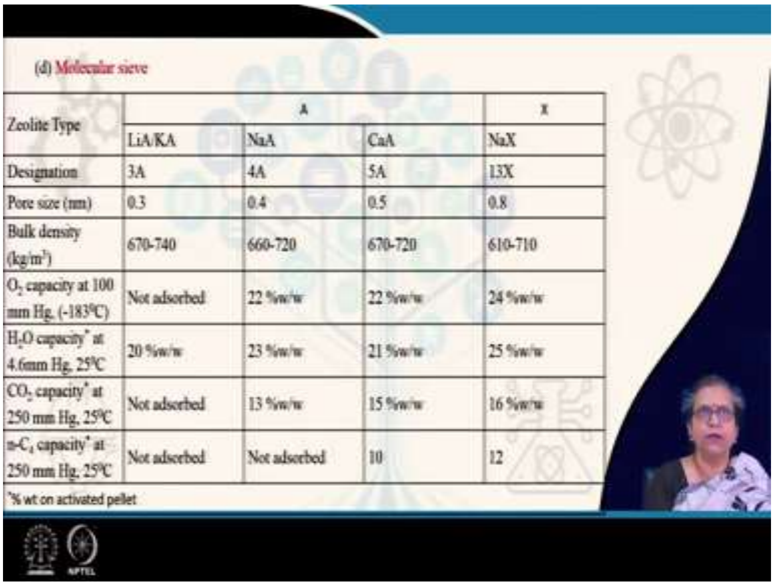

(d) Molecular sieve

Zeolite Type	A			X
	LiA/KA	NaA	CaA	NaX
Designation	3A	4A	5A	13X
Pore size (nm)	0.3	0.4	0.5	0.8
Bulk density (kg/m^3)	670-740	660-720	670-720	610-710
O_2 capacity at 100 mm Hg, (-183°C)	Not adsorbed	22 %w/w	22 %w/w	24 %w/w
H_2O capacity* at 4.6mm Hg, 25°C	20 %w/w	23 %w/w	21 %w/w	25 %w/w
CO_2 capacity* at 250 mm Hg, 25°C	Not adsorbed	13 %w/w	15 %w/w	16 %w/w
n-C_4 capacity* at 250 mm Hg, 25°C	Not adsorbed	Not adsorbed	10	12

*% wt on activated pellet

So, therefore, these are just the typical properties.

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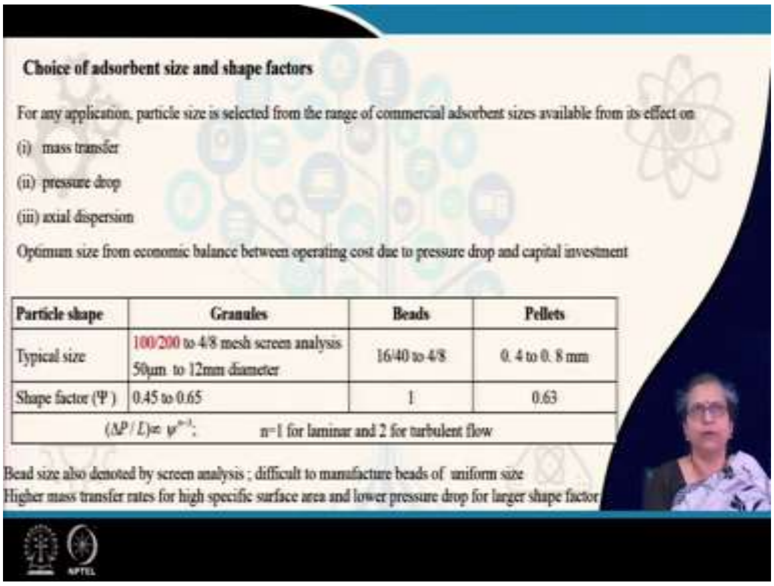

Choice of adsorbent size and shape factors

For any application, particle size is selected from the range of commercial adsorbent sizes available from its effect on

(i) mass transfer

(ii) pressure drop

(iii) axial dispersion

Optimum size from economic balance between operating cost due to pressure drop and capital investment

Particle shape	Granules	Beads	Pellets
Typical size	100/200 to 4/8 mesh screen analysis 50μm to 12mm diameter	16/40 to 4/8	0. 4 to 0. 8 mm
Shape factor (Ψ)	0.45 to 0.65	1	0.63
$(\Delta P / L) \propto \psi^{n-1}$; n=1 for laminar and 2 for turbulent flow			

Bead size also denoted by screen analysis ; difficult to manufacture beads of uniform size

Higher mass transfer rates for high specific surface area and lower pressure drop for larger shape factor

Now, once you have selected the adsorbent you have collected the properties, next we are going to specify the size, the shape and shape factors. Now, you tell me on what should this depend? We will have to select the particle size from the range of sizes that are available, the typical sizes are mentioned in this particular table for granules, beads and pellets. So, from the range of sizes that are available we are going to select a suitable particle size.

What governs the selection of particle size? It is the mass transfer considerations, if suppose we would want an enhanced mass transfer, naturally for that we would be we would want a higher interfacial area and so therefore, we would go for smaller size of the particles. At the same time, with smaller size particles the pressure drop becomes higher, the axial dispersion is also better. So, from axial dispersion and pressure drop point of view, we would like to go for a higher size particle while for mass transfer point of view we would like to opt for a smaller size particle.

So, the optimum size it will be decided from an economic balance between the operating cost due to pressure drop and the capital investment. And here I have given the sizes both in millimeters as well as in screen analysis; very frequently for beads we use screen analysis because you know it is very difficult to manufacture beads of uniform size. And there is one other very interesting thing and just like we know that the higher mass transfer rates give higher specific surface area, same way under certain conditions we find that the pressure drop is lower for higher shapes and higher shape factors as well.

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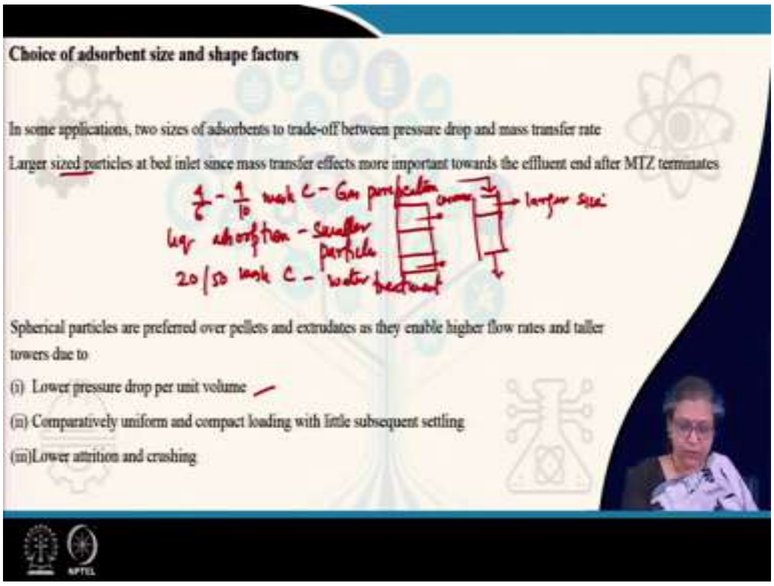

Choice of adsorbent size and shape factors

In some applications, two sizes of adsorbents to trade-off between pressure drop and mass transfer rate

Larger sized particles at bed inlet since mass transfer effects more important towards the effluent end after MTZ terminates

$\frac{4}{6}$ - $\frac{4}{10}$ mesh C - Gas purification

liq adsorption - smaller particle

20/50 mesh C - water treatment

Spherical particles are preferred over pellets and extrudates as they enable higher flow rates and taller towers due to

(i) Lower pressure drop per unit volume

(ii) Comparatively uniform and compact loading with little subsequent settling

(iii)Lower attrition and crushing

And there are certain other things for example, if you select spherical particles over pellets and extrudates we find that they enable higher flow rates and taller towers as because the pressure drop is lower per unit volume and it is possible that we have a comparatively uniform and compact loading with completely lesser amount of subsequent settling and also chances of attrition and crushing. These are also much lower in this particular case.

And we find that in certain applications there can be situations where we select just like I had mentioned in the last class. Say, we might select two sizes of adsorbents, we can have larger size particles at the bed inlets. Suppose we are having a bed and this is the inlet, this is the outlet, then near to the bed we would be having larger sized particles because the mass transfer is much more important at the effluent end. So, at the inlet we might have larger size particles and as we proceed we might have smaller sized particles.

And for most of the cases just as I had mentioned in the last class, normally the adsorbent bent over and above. We have some particular inert particles like ceramics, etc. Wthin these larger sized particles, the actual adsorber bed is confined within these two.

Now, normally there are certain typical sizes for example, 4/6 to 4/10 mesh carbon. There is certain thumb rule sort of things. This is selected for say gas purification, we normally select this. Now, normally for liquid adsorption the particle size is smaller and generally we find, say 20 to 50 mesh carbon can be used for water treatment, etc.

So, these are some of the typical sizes which are normally used in the industries.

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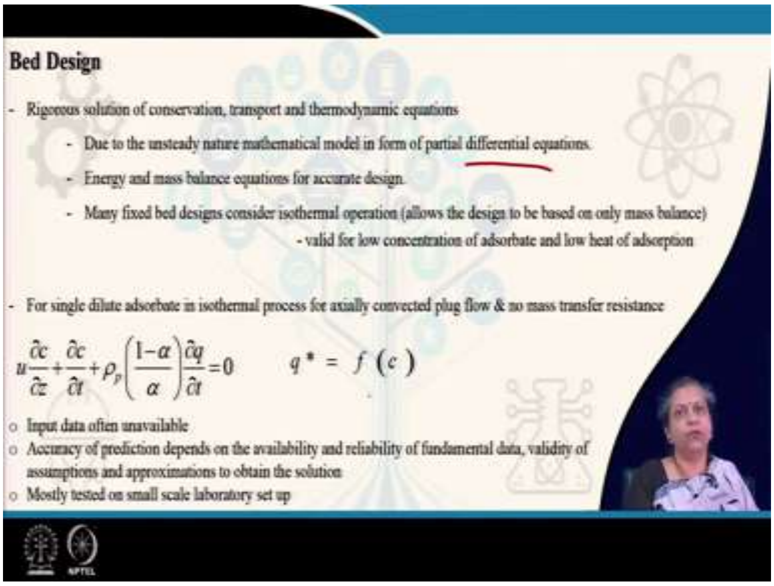

Bed Design

- Rigorous solution of conservation, transport and thermodynamic equations
  - Due to the unsteady nature mathematical model in form of partial differential equations.
  - Energy and mass balance equations for accurate design.
  - Many fixed bed designs consider isothermal operation (allows the design to be based on only mass balance)
    - valid for low concentration of adsorbate and low heat of adsorption
- For single dilute adsorbate in isothermal process for axially convected plug flow & no mass transfer resistance

$$u\frac{\partial c}{\partial z}+\frac{\partial c}{\partial t}+\rho_p\left(\frac{1-\alpha}{\alpha}\right)\frac{\partial q}{\partial t}=0 \qquad q^* = f(c)$$

- ○ Input data often unavailable
- ○ Accuracy of prediction depends on the availability and reliability of fundamental data, validity of assumptions and approximations to obtain the solution
- ○ Mostly tested on small scale laboratory set up

So, well, what we have done, we have selected the adsorbent, we have also considered the saw the particle, size, shape and also the physical properties we are going to collect. After that definitely we will be generating the equilibrium data if equilibrium data is available it is fine for us. Now, we go for the bed design.

Now, the bed design can be done either by a rigorous method. What does the rigorous method mean? It is definitely the conservation equations, transport equations and also the thermodynamic equations. Now, one thing you have to remember that since this is an unsteady state process, the equations will be in the form of partial differential equations. For an accurate design we would require both energy balance as well as mass balance equations.

Several fixed bed designs consider isothermal operation and when they consider isothermal operation naturally we can have the design based on only mass balance itself which is valid when the adsorbate concentration is low and the heat of adsorption is also low. So, accordingly for that we can write down the unsteady state mass balance equation, the equilibrium relationships and we can proceed.

Naturally, one feels that under this condition definitely the design should be much more accurate, but there are certain things also which we need to consider. For example, we find that we need a large number of input parameters in order to solve this equation and the input data is often unavailable and it is quite obvious that the accuracy of the prediction depends upon the availability and the reliability of the fundamental data. And, so naturally for most of the cases we do not opt for the rigorous solution.

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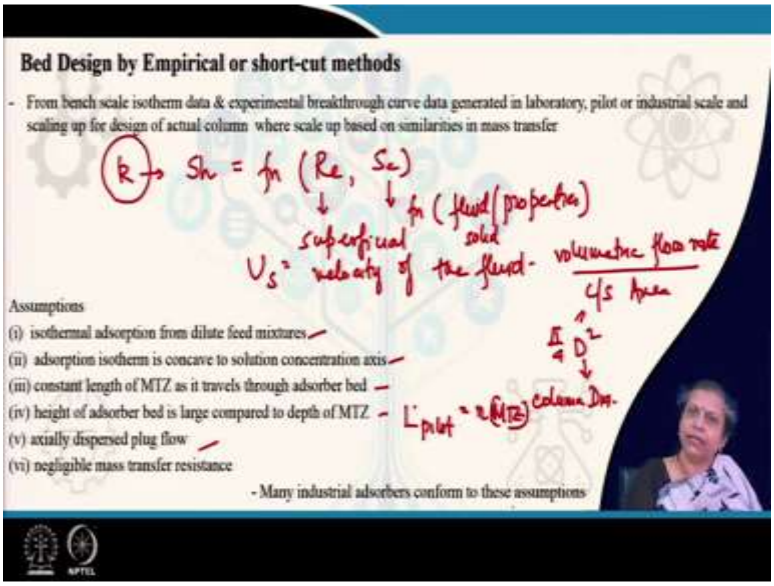

Bed Design by Empirical or short-cut methods

- From bench scale isotherm data & experimental breakthrough curve data generated in laboratory, pilot or industrial scale and scaling up for design of actual column where scale up based on similarities in mass transfer

$\boxed{k} \rightarrow Sh = fn(Re, Sc)$

$Re \downarrow$ superficial velocity of the fluid; $Sc \downarrow$ fn (fluid/solid properties)

V_s = superficial velocity of the fluid = $\frac{\text{volumetric flow rate}}{\text{c/s Area}}$

c/s Area = $\frac{\pi}{4} D^2$, D → column Dia.

Assumptions

(i) isothermal adsorption from dilute feed mixtures

(ii) adsorption isotherm is concave to solution concentration axis

(iii) constant length of MTZ as it travels through adsorber bed

(iv) height of adsorber bed is large compared to depth of MTZ

$L'_{pilot} = 2(MTZ)$

(v) axially dispersed plug flow

(vi) negligible mass transfer resistance

- Many industrial adsorbers conform to these assumptions

On the other hand, we generally in industries the bed design is done by means of an empirical method or a shortcut method. What is this method? In this method what we do is along with equilibrium data we also generate the breakthrough curve either in a pilot scale or maybe in a laboratory scale. We generate the pilot scale data, specifically we generate the breakthrough curve and then we use this particular breakthrough curve and then we try to find out certain scale up factors and based on this scale up factors we scale the design the dimensions of the pilot scale column to the actual column.

Now, under what conditions can we do this? If we really have to take or we if we really have to use the breakthrough curve which we have generated in the pilot scale for designing the actual adsorber, then under what condition we can do it. We can do it. Firstly, definitely it has to be the same adsorbent and the same feed, etc. But, along with that it we have to remember that the mass transfer coefficient should be the same in both the pilot scale as well as in the actual adsorber that we want to design.

Now, what the mass transfer coefficient to be same, what should be ensured to be same in the pilot scale as well as in the actual the plant situation? The mass transfer coefficient if you find this mass transfer coefficient this is usually obtained from a Sherwood number. You will be remembering that this is kl d by d. Sherwood number if you remember this is going to be a function of Reynolds number and Schmidt number. We know that the Schmidt number is a function of properties of the system - fluid and solid properties.

So, therefore, first we have selected the same adsorbent; if suppose we will be using a regenerated adsorbent we will be generating the pilot scale data with the regenerated adsorbent itself. So, once the adsorbent adsorbate pair has been fixed, so naturally the Schmidt number has been fixed.

So, under for a particular system Sherwood number is a function of Reynolds number only which means that the Reynolds number is a function of apart from other things, the superficial velocity of the fluid. I refer to as the fluid because the same thing is going to be applicable for gas phase adsorption as well as for the liquid phase adsorption.

Superficial velocity means the volumetric flow rate of the fluid per unit total cross sectional area - for a cylindrical column this is equals to $\frac{\pi}{4D^2}$ where D is the column diameter. So, therefore, the first thing that we have to ensure is that if the superficial velocity which we have designated as U s provided that it is same in the pilot scale as well as in the actual plant. Only under that condition we can expect that the mass transfer coefficient is going to be the same in both cases and under that condition if we generate the breakthrough curve in the pilot scale we can use it for the actual plant as well.

Well, there are certain other things that we need to remember, even for the pilot scale the height should be such that it comprises of several mass transfer zones. So, in other words the length of the pilot plant should be several lengths of the MTZ zones, this is one thing that we need to consider, the other things of course - they are quite evident, we do not perform any energy balance.

So, we assume the adsorption is isotherm, it is concave to the c axis - the concentration axis and the other important thing which normally happens is that the length of the MTZ is constant as it travels along the column. If this length varies, but which does not happen, then in that case definitely we cannot go for this particular scaling of process. We will also have to assume that there is no mild distribution, no challenge channeling and the flow is axially dispersed plug flow, negligible mass transfer resistance.

And fortunately we find that most of the industrial adsorbers they conform to these assumptions. So, naturally what is the process by which we proceed further design? Now, what are the things we know, we have all data on the pilot plant.

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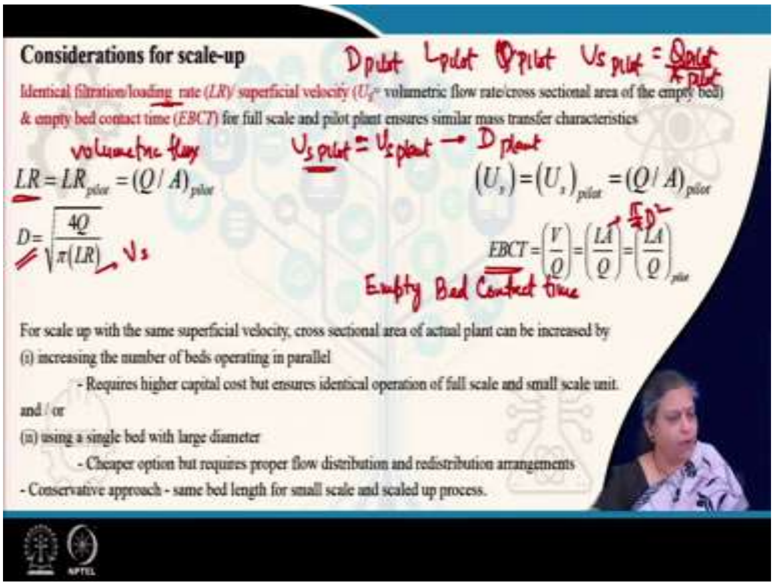

Considerations for scale-up

D_{pilot} L_{pilot} Q_{pilot} $U_{s\,pilot} = \frac{Q_{pilot}}{A_{pilot}}$

Identical filtration/loading rate (*LR*)/ superficial velocity (U_s= volumetric flow rate/cross sectional area of the empty bed) & empty bed contact time (*EBCT*) for full scale and pilot plant ensures similar mass transfer characteristics

volumetric flow

$U_{s\,pilot} = U_{s\,plant} \rightarrow D_{plant}$

$$LR = LR_{pilot} = (Q/A)_{pilot}$$

$$(U_s) = (U_s)_{pilot} = (Q/A)_{pilot}$$

$$D = \sqrt{\frac{4Q}{\pi(LR)}}$$

V_s

$$EBCT = \left(\frac{V}{Q}\right) = \left(\frac{LA}{Q}\right) = \left(\frac{LA}{Q}\right)_{pilot}$$

$\frac{\pi}{4}D^2$

Empty Bed Contact time

For scale up with the same superficial velocity, cross sectional area of actual plant can be increased by

(i) increasing the number of beds operating in parallel

- Requires higher capital cost but ensures identical operation of full scale and small scale unit.

and / or

(ii) using a single bed with large diameter

- Cheaper option but requires proper flow distribution and redistribution arrangements

- Conservative approach - same bed length for small scale and scaled up process.

What are the data that we have on the pilot plant? We know the column the pilot column dimensions, we know the D pilot, we know the L pilot, we know the Q pilot, the flow rate of the feed through the pilot plant. So, we know all these things. So, definitely when we know these things, we can find out the U s pilot. It is nothing but the Q pilot divided by the cross sectional area of the pilot.

Normally what we do, we assume the U s for the pilot plant is equal to the U s for the plant as I have shown here. When that is the case then in that case since we know the D pilot so, from this particular equation we are in a position to find the D plant, let me write it down as D plant that will be better.

So, therefore, we can find out the diameter of the actual column. Now, just one thing I would like to mention in this case for gas phase adsorption we refer to this particular term as the superficial velocity. And when we deal with liquid flow rates then generally it has the same units but the physical significance is slightly different, we refer to as the loading rate. This loading rate is nothing, it is just the volumetric flux which is again nothing but the superficial velocity. But, normally for liquid phase adsorption we refer to this particular term as the loading rate and for gas phase adsorption we refer this as the superficial velocity.

It does not matter whatever it is, first thing that we assume is LR plant equals to LR pilot. And so, therefore, from there we find out D this is LR for liquid phase adsorption or this is U s for gas phase adsorption.

There is one other important parameter also which we consider, that is the empty bed contact time. What is this empty bed contact time? This is also another thing which to start with we keep it equal then as the design proceeds depending on practical considerations, etc. we may have to deviate.

And after we have obtained the actual column dimensions we have to see that they conform to some codes. So, therefore, to confirm to codes we have to change the dimensions, etc. but once we have finalized the design we need to see that the empty bed contact time does not exceed the pilot scale and the volumetric flux or the superficial velocity is also obeyed.

This empty bed contact time is nothing but the volume of the pilot plant by the flow rate of the feed here and this is usually kept equal in the pilot as well as in the actual case. So therefore, we find that from the superficial velocity we can find out the D, once we know the D we know the A and once we know the A we can find out the length of the actual contactor.

Certain suggestions I would like to give you in this particular respect, these are just suggestions, it is not that these have to be strictly followed during the design. In fact, you will find that when we work out two illustration problems you will find that we have deviated for these, but definitely these things will be serving as some thumb rules for you.

The first thing is that see when you make the superficial velocity or the loading rate equal, then in that case we find out the diameter of the actual column from the diameter of the pilot, quite naturally the flow rate is higher in the actual column, the diameter will be more. Now, when this we have to provide for this additional diameter how would you like to provide it, we can either have a larger diameter column or we can have several pilot scale columns which are operating in parallel.

Now, when we have the several pilot scale columns operating in parallel naturally it is going to be expensive. But, in this case the advantage is that the full scale and the small scale units they have identical operation and in the full scale unit also there are no problems of channeling, mal distribution etcetera which is one thing which we have to keep in mind when we are going to design the larger column.

On the other hand, if we have a single bed with a larger diameter then in that case it is definitely a cheaper option, but just the way I have told you. And the other thing is often it can happen that we can use for small scale and scaled up, we can use the same bed length, but normally we operate with vertical columns and when we operate with vertical columns the length by diameter it should be roughly between 3 to 5. So, naturally keeping these things into consideration the final design has to be formed up.

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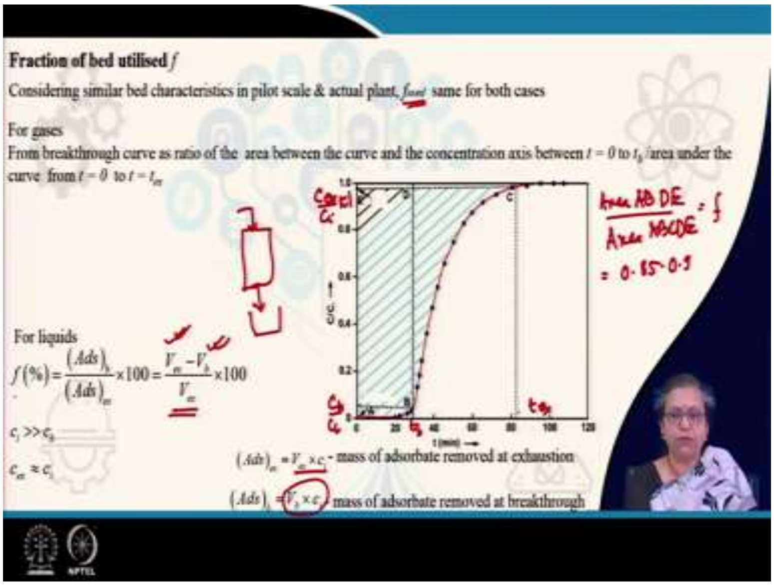

Fraction of bed utilised f

Considering similar bed characteristics in pilot scale & actual plant, f_{bed} same for both cases

For gases

From breakthrough curve as ratio of the area between the curve and the concentration axis between $t = 0$ to t_b /area under the curve from $t = 0$ to $t = t_w$

For liquids

$$f(\%) = \frac{(Ads)_b}{(Ads)_w} \times 100 = \frac{V_w - V_b}{V_w} \times 100$$

$c_i >> c_b$

$c_w \approx c_i$

$(Ads)_w = V_w \times c_i$ - mass of adsorbate removed at exhaustion

$(Ads)_b = V_b \times c_i$ - mass of adsorbate removed at breakthrough

So, the once the contactor dimensions are more or less roughly estimated you know the next thing is to find out the amount of adsorbent. How to find the amount of adsorbent?

We have to find out the fraction of the adsorbent bed that is used because just as we have found out yesterday that we can allow adsorption till the breakthrough point. And this particular breakthrough point is decided based on the outlet specification which has been provided by the customer to the designer.

So, therefore, for finding out the f used you need to have the breakthrough curve. Now, this breakthrough curve you either generate and if you are fortunate enough then for the adsorbent that you have selected, there might be some particular literature data where this breakthrough curve is available.

Now, suppose the breakthrough curve is available how to find out f, can you tell me? Just if you see, in this particular figure this is a typical breakthrough curve that has been drawn and this corresponds to c b by c I, this corresponds to t b the breakthrough time and then we operated till the bed exhaustion. So, therefore, this corresponds to t ex and this part it corresponds to c ex by c i.

So, therefore, from here what do we find, we find that once this curve is there then the area is A B C D E. This particular area naturally corresponds to the fraction of adsorbate which has been adsorbed till the breakthrough time. This is A B D E divided by the area A B C D E, what should this give you? So, this should give you the fraction of the bed that has been used for adsorption. So, from there you can find out the fraction that has been used.

And as I had mentioned in the last class normally we would like to keep it between 0.85 to 0.9, you are going to find that in the pilot scale column often it can also exceed 0.9, but for actual columns it is usually much lower. And if it is really low then what has to be done operating two beds in series that we had discussed in the last class itself. So, this is one particular way by which we can find out the f used.

Normally, for gas phase adsorption we do it, but for liquid phase adsorption it often happens that the entire breakthrough curve is not available for us. On the other hand, during liquid adsorption often what happens, the liquid is coming in here, it flows out from here, this liquid is collected and the cumulative volume of the liquid is measured.

So, very often we find that to find out the adsorbate which has been adsorbed till breakthrough as a ratio of the adsorbate adsorption till the exhaustion point, normally what we do is - for situations where this particular curve is very steep and also at the exhaustion point more or less this is almost equal to 1 and this c b is much less than c ex or in other words if the curve is quite steep under that particular case we can find out that at exhaustion what is the amount of adsorbate that has been removed.

This is nothing but the cumulative volume of the liquid collected at exhaustion into the into c i and for the amount of adsorbate which has been collected till breakthrough this is the expression. So, very frequently for finding out f we can use this expression and from the breakthrough experiments we have data on this, from where we can find out f.

Now, I am telling you this because you do not know when you go for the adsorber design, what will be the data, which are going to be available to you. If the breakthrough curve is available the area under the curve can be 1, if other than that the cumulative volume of the liquid which has been collected during the generation of the breakthrough curve then we can use this particular equation.

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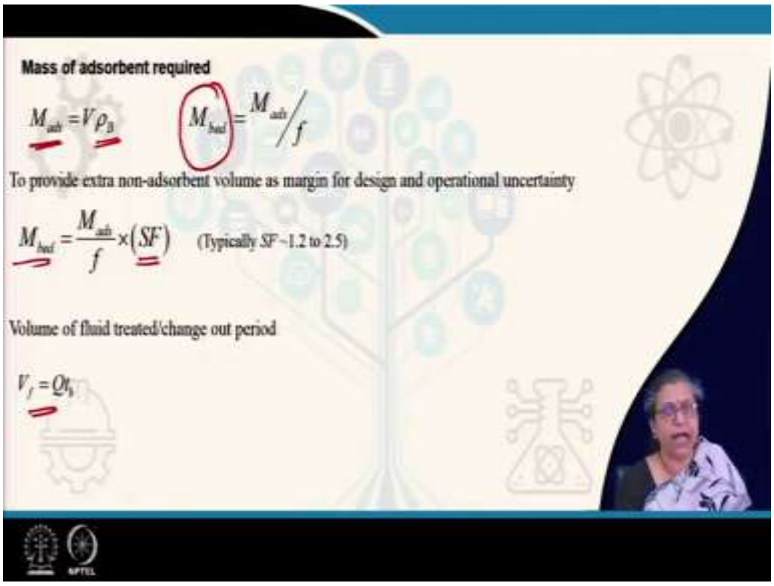

Mass of adsorbent required

$M_{ads} = V\rho_B$ $\quad$ $M_{bed} = M_{ads}/f$

To provide extra non-adsorbent volume as margin for design and operational uncertainty

$M_{bed} = \frac{M_{ads}}{f} \times (SF)$ (Typically $SF \sim 1.2$ to 2.5)

Volume of fluid treated/change out period

$V_f = Qt_b$

So, from here once you have found out the amount which has been used, naturally from there you can find out the volume of the bed and you can find out the mass of the bed. You must have seen in all those property data that I have shown you the bulk density of their adsorbent is usually available to us. Now, normally what we do, along with the mass of the actual bed we generally augment it by a safety factor just keeping into account certain uncertainties, etc.

So, normally we augment it by a safety factor which varies from 1.2 to 2.5 and using that we can find out the actual mass of adsorbent in the bed and you can find out the total amount of fluid that has been treated during this entire adsorption cycle. So, that of course comes out from the volumetric flow rate and the breakthrough time.

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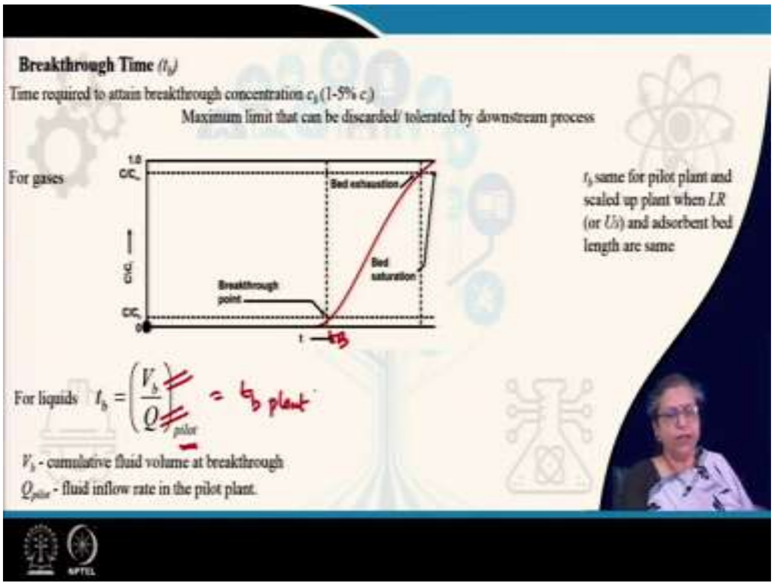

Breakthrough Time *(t_b)*

Time required to attain breakthrough concentration c_b (1-5% c_i)

Maximum limit that can be discarded/ tolerated by downstream process

For gases

t_b same for pilot plant and scaled up plant when *LR* (or *Us*) and adsorbent bed length are same

For liquids $t_b = \left(\frac{V_b}{Q}\right)_{pilot}$ = t_b plant

V_b - cumulative fluid volume at breakthrough

Q_{pilot} - fluid inflow rate in the pilot plant.

Now, for the breakthrough time again if you have this particular data it is not at all a problem, this corresponds to the breakthrough time. On the other hand, if you are having data on the cumulative volume collected you do not have this particular data, then in that case definitely the cumulative fluid volume which is collected at breakthrough divided by the volumetric flow rate that gives you the breakthrough time. This particular data you can get from the pilot scale and then you can use it for the actual plant. So, in this particular way we can proceed with the design.

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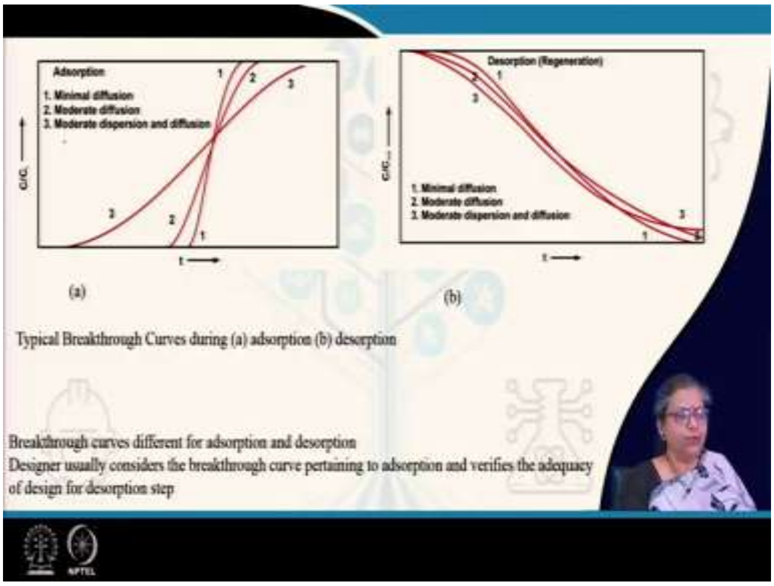

Typical Breakthrough Curves during (a) adsorption (b) desorption

Breakthrough curves different for adsorption and desorption

Designer usually considers the breakthrough curve pertaining to adsorption and verifies the adequacy of design for desorption step

There is just one thing which I would like to mention, see when you are having the breakthrough data you will find that different curves are generated for adsorption and for desorption. So, for the design what would you have to do? For the design we would generally consider the breakthrough curve which has been generated for adsorption pertaining to our specific situation and then we will verify the same for the desorption step.

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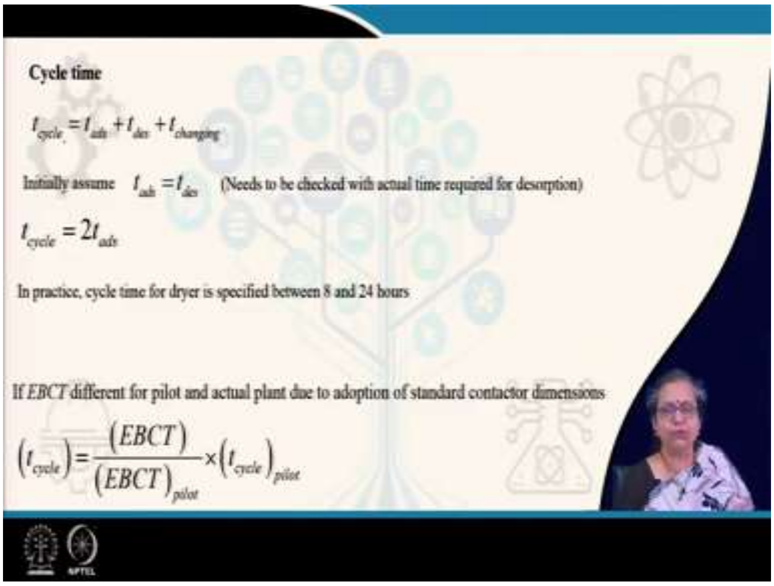

Cycle time

$t_{cycle} = t_{ads} + t_{des} + t_{changing}$

Initially assume $t_{ads} = t_{des}$ (Needs to be checked with actual time required for desorption)

$t_{cycle} = 2t_{ads}$

In practice, cycle time for dryer is specified between 8 and 24 hours

If *EBCT* different for pilot and actual plant due to adoption of standard contactor dimensions

$$\left(t_{cycle}\right) = \frac{(EBCT)}{(EBCT)_{pilot}} \times \left(t_{cycle}\right)_{pilot}$$

So, therefore, in this particular way we proceed with the design and we fix up the contactor dimensions, we find out the total amount of adsorbent that we have to use and we also find out the total amount of volume of the fluid which has been treated, etc. Now, as you know that since this is an unsteady state process the adsorber operates in cycles. For one part of the cycle adsorption goes on, as long as adsorption goes on, say we call it the t adsorption the duration for which adsorption goes on.

After sometime we find that the breakthrough concentration is reached, once the breakthrough concentration is reached we cannot use this adsorber anymore. Assuming that more or less at the breakthrough concentration more or less the amount of adsorbent used is roughly about 85 to 90 percent. So, therefore, after that we send it for the desorption stage, while it gets desorbed there is another column where adsorption takes place. So, therefore, we find that the entire cycle it comprises of the time for adsorption, the time for desorption and also some particular time to allow for the changeover.

Generally, we try to reduce the time of desorption as less as possible, but to start with we can assume that this time we can assume it to be negligible or in other words we can assume that this entire time is equal to the time for which the bed was operated for adsorption. So, to start with we assume t cycle equals to t adsorption, where this t adsorption is equals to t breakthrough point.

And also there are some particular typical cycle times say for example, for dryer it is between 8 to 24 hours, etc. and at the end of the design as I have said for practical considerations, it can happen that finally, the empty bed contact time that we have selected for the plant that is not equal to the empty bed contact time which was obtained for the pilot plant. In that case, the cycle time for the adsorber can be obtained from this particular equation $\left(t_{cycle}\right)=\frac{\left(EBCT\right)}{\left(EBCT\right)_{pilot}}\times\left(t_{cycle}\right)_{pilot}$.

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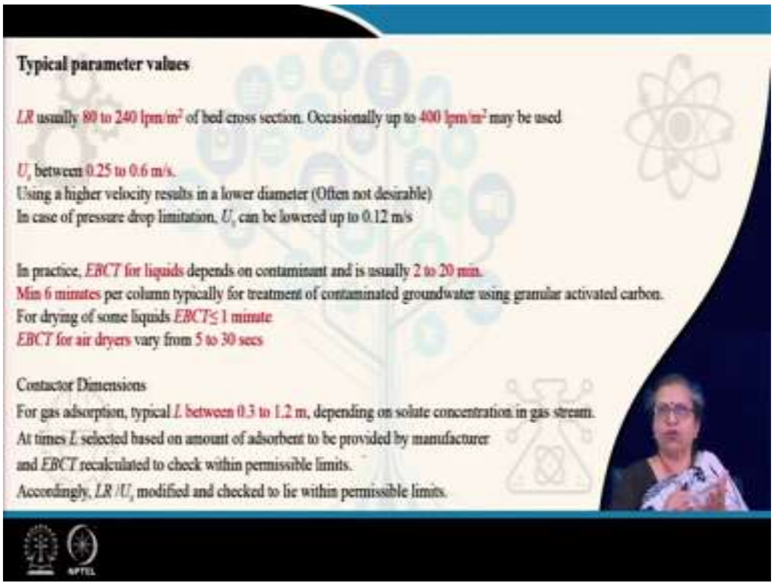

Typical parameter values

LR usually 80 to 240 lpm/m^2 of bed cross section. Occasionally up to 400 lpm/m^2 may be used

U_s between 0.25 to 0.6 m/s.
Using a higher velocity results in a lower diameter (Often not desirable)
In case of pressure drop limitation, U_s can be lowered up to 0.12 m/s

In practice, *EBCT for liquids* depends on contaminant and is usually 2 to 20 min.
Min 6 minutes per column typically for treatment of contaminated groundwater using granular activated carbon.
For drying of some liquids *EBCT* ≤ 1 minute
EBCT for air dryers vary from 5 to 30 secs

Contactor Dimensions
For gas adsorption, typical *L* between 0.3 to 1.2 m, depending on solute concentration in gas stream.
At times *L* selected based on amount of adsorbent to be provided by manufacturer
and *EBCT* recalculated to check within permissible limits.
Accordingly, *LR* /U_s modified and checked to lie within permissible limits.

And we would also like to check up at the end - the empty bed contact time and also the loading rate, the superficial velocities; these particular values normally for industrial adsorbers have been observed to lie within certain ranges.

So, therefore, when we start the design we need to check, whether for the pilot plant LR, U s and EBCT they more or less conform to these limits or not. Finally, when we arrive at the final design we also again need to check whether these particular limits they have been respected or not.

So, just for the convenience of the designer I have simply noted down the typical parameter values for LR, for the superficial velocity. Definitely there is something I would like to mention that the superficial velocity varies between 0.25 to 0.6 meters per second. Quite naturally what we feel, that if we go for a higher superficial velocity then the adsorption can take place in a lower diameter, but this is often not desirable.

If I ask you why; you will say that this will lead to higher pressure drop, but this is not the actual reason for this. When we go for a higher velocity the contact time, the residence time of the liquid inside the adsorber decreases as a result of which adsorption is not so very effective. So, therefore, to increase the residence time also, it is not always necessary that we would opt for a higher velocity.

The empty bed contact time definitely depends on the contaminants and it is definitely higher for liquid adsorption as compared to gas phase adsorption. Some typical EBCT for air dryers, liquid treatment the purposes for which these are mostly used they have been mentioned in this particular table. So, you need to check up all these things finally. When we have completed the design then also we need to just see that they lie within the permissible limits.

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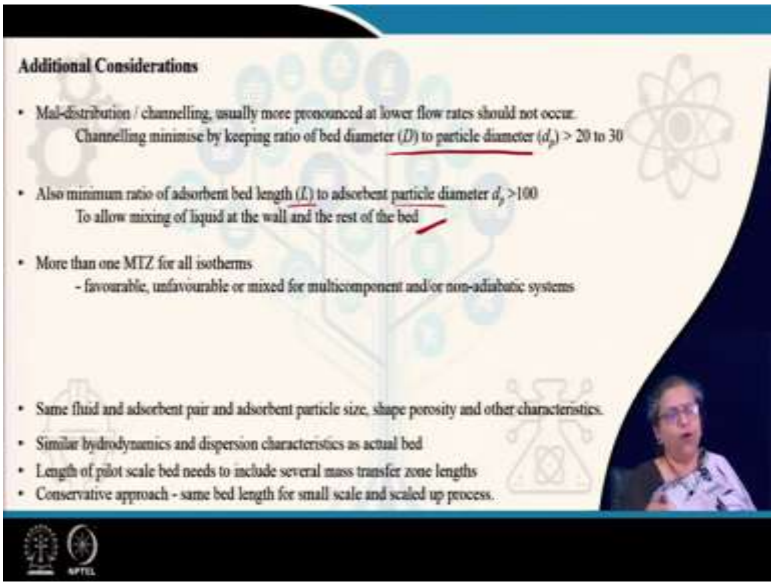

Additional Considerations

- Mal-distribution / channelling, usually more pronounced at lower flow rates should not occur.
  Channelling minimise by keeping ratio of bed diameter (D) to particle diameter (d_p) > 20 to 30
- Also minimum ratio of adsorbent bed length (L) to adsorbent particle diameter d_p >100
  To allow mixing of liquid at the wall and the rest of the bed
- More than one MTZ for all isotherms
  - favourable, unfavourable or mixed for multicomponent and/or non-adiabatic systems
- Same fluid and adsorbent pair and adsorbent particle size, shape porosity and other characteristics.
- Similar hydrodynamics and dispersion characteristics as actual bed
- Length of pilot scale bed needs to include several mass transfer zone lengths
- Conservative approach - same bed length for small scale and scaled up process.

So, with this we have come to the end of the discussions on the design part, we have discussed how the design proceeds, then we have also discussed after the design what are the checks that we need to do particularly with respect to the parameter values, there are certain additional considerations which we need to keep in mind. For example, the first important thing in the adsorption is a proper distribution of the feed into the bed.

So, naturally in order to minimize mal distribution or channeling, there are certain thumb rules, generally these also have to be checked that the column diameter to the particle diameter is greater than 20 to 30. Also it is important to check that the ratio of the bed length to the adsorbent particle diameter that is also greater than 100 which will enable mixing of the liquid at the wall and the rest of the bed. Definitely, both in the pilot plant as well as in actual column there has to be more than one MTZ for all isotherms.

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Pressure Drop

- Important Design Parameter as processing rate can be limited by pressure drop
- Excessive pressure drop - bed compaction or lifting
- Very low pressure drop - uneven distribution and channelling
- Most adsorption plants operate with small pressure drop across adsorbent bed to keep power costs low as large particles are used whenever possible and velocity is also typically low to allow equilibrium to be attained between fluid and adsorbent.

So, once we have done the design more or less then we go for the pressure drop part. The pressure drop is something very important which you need to remember. Why? Because if ressure drop is excessive then it leads to bed compaction or lifting, if the pressure drop is very less then there will be uneven distribution and channeling.

So, what we do in the next class, we discuss the estimation of pressure drop, after that we will be taking up two illustration problems, one on gas adsorption and one on liquid adsorption, we will be discussing just the outlines of the design. So, it will give you an idea regarding the actual design of adsorbers and that is going to end up our discussion on adsorbers. So, this much for today.

Thank you very much.