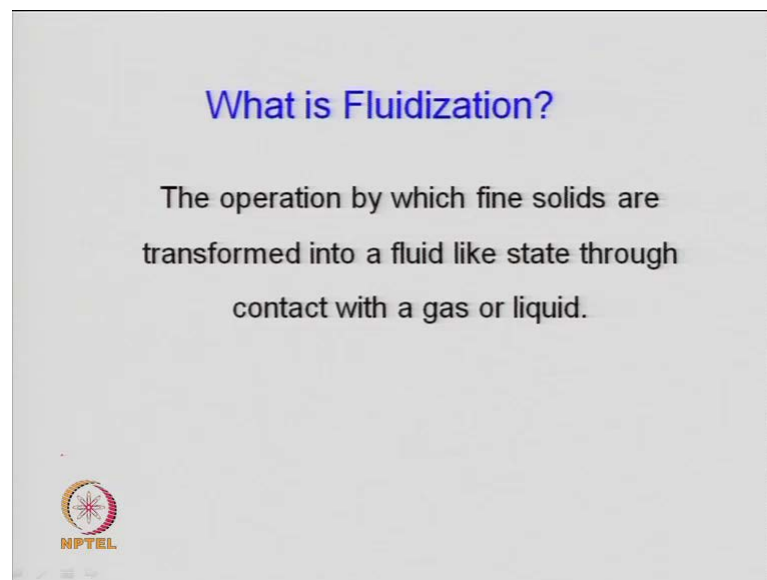


Heterogeneous Catalysis and Catalytic Processes
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Lecture - 39
Fluidized Bed Reactor

So, good afternoon to all of viewers. So, today, I will talk some brief outlines of the fluid bed reactor and try to cover some design aspects of a fluidized bed reactor, especially, the bubbling fluidized bed reactor.

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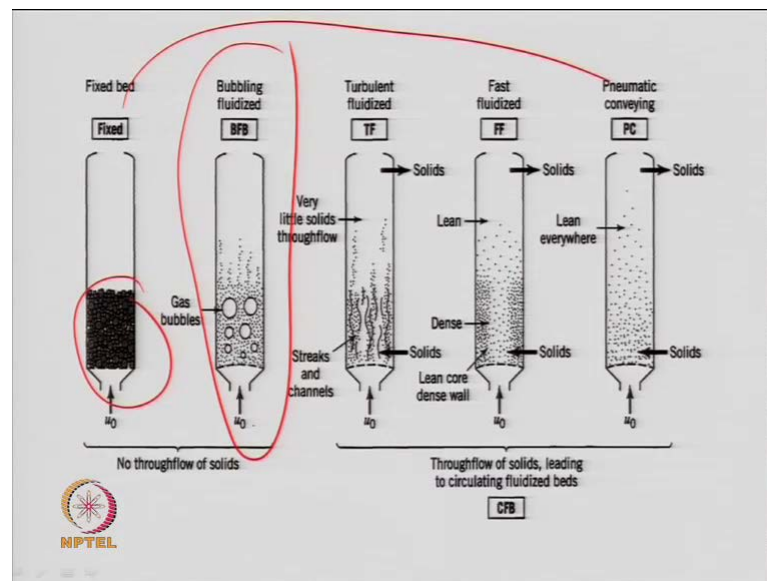


All of you know about the fluidization, because you must have read in your undergrad mass transfer courses; and, you know that, gas – if it is a packed bed and you give the flow at different velocities; then, up to a certain superficial gas velocity, there would not be any kind of the bubbling, or the bed will be static. The solids will remain at a stable condition in the bed. But, when you increase the velocity, then there will be a kind of more and more voids and then the bed starts something like a fluid kind of a state, or there will be a kind of bubbling inside the bed.

So, this principle, which is based on some gravity force acting downwards; then, buoyancy force is acting; and, some drag force is acting on a particle. So, this principle has been used in the reactioning or reactor design or catalytic reactioning, where the

solid is the catalyst. And depending upon the operation, we can have a kind of a CSTR or mix bed reactor, and the gas can be bubbled through this catalyst, or vapor can be bubbled through this catalyst; and, there can be good contact between the gas and solid; and, there is a kind of back mixing in the reactor. So, that basic concept is something like where you have a fine powder or fine solid particles; and, they are transformed into some kind of fluid-like state by using some gas or vapors or liquid.

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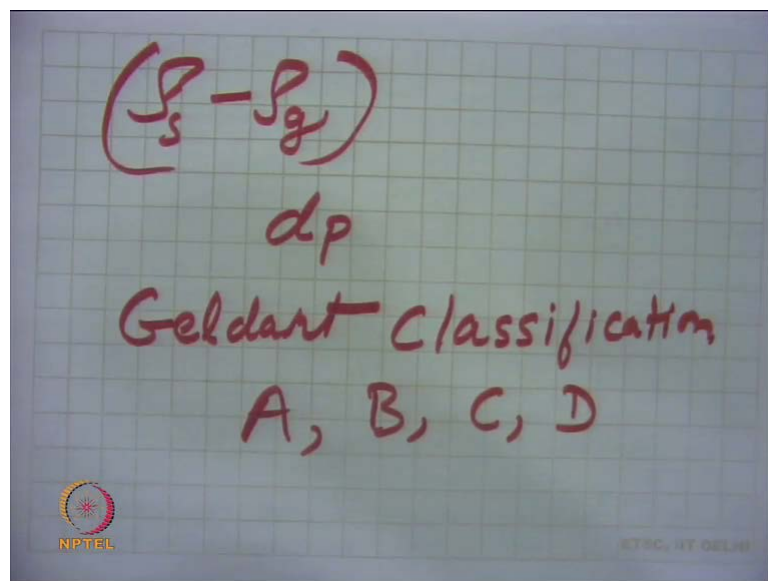
So, depending upon the flow, you can have different situations in a bed; so which can be seen here that, this is just like a fixed bed when the velocity is relatively low – superficial gas velocity is relatively low. So, that time, the weight of the solid is more dominating compared to the buoyant force and viscous force. So, all these what you have done in the Stokes law; and, depending upon that regimes, Reynolds number and pressure drop equations in the fixed bed and fluidized bed equation. So, this is a static condition, where you have a minimum voidage in the bed.

So, bed voidage is almost... And if it is a very fine powder, it will be low. But, when you increase the velocity, the bed starts like this – bubbling – a kind of where the particles are moving out. And you will find more and more voidage between the beds. So, gas fixed interstitial space between the solid particle, there is now a gas, and the particle moves up and down. So, that is a kind of bubbling just like water boils; and, you see the droplet goes up and coming down because of the convective currents. So, this is a

situation here that, you have a kind of bubbling bed, but solids are not leaving the reactor or the column.

But, when you have still a higher velocity, you can have some through-flow of solids like... So, that is the more and more moments. So, the voidage between the particles is increasing. So, the pressure drop will be low now; the contact between gas and solid may be better. But, there may be a kind of situation, where you have a bit slug flow condition; the pockets of solids and then gas leaves, and kind of channeling may appear. So, that is a kind of turbulent fluidized bed. And when you have a still, the particles are very fine. So, all will depend on the size of the particle, that is, first thing. And I told that, there is a buoyant force, there is a viscous force, gravity force. So, it means it will depend on the viscosity of the medium; it will also depend on the density of the gas; and of course, the superficial gas velocity, which is very important.

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So, it means something like a $\rho_s - \rho_g$ if I talk – density of the solid and density of the gas; so that will also be a parameter, which will affect. And size of the particle is very important. So, you need to characterize these things before looking the type of fluidization behavior. And also, it will depend on the property of the solid. So, the solid may be cohesive type solid just like flour; it may not fluidize easily. So, that is what a different characterization or classifications of the solids, what is known as Geldart chart. So, that is known as Geldart classification basically; something a classification given by

this Geldart. So, known as a Geldart classification. So, basically, in this classification, you have something like A, B, C, D. So, we have categorized the solid in different category A, B, C and D; so which may be difficult to fluidize; which may be a cohesive nature; and, that will depend.... So, depending upon that, if you wish to fluidize, you need to have a $\rho_s - \rho_g$ and as a function of diameter of the particle. So, as the diameter of the particle increase, the fluidization will become still difficult. So, this is one.


So, it means depending upon the superficial gas velocity, you can have a regime from fix bed and to a kind of system when the particles leave the bed. So, this fundamental thing can be used in a spray tower unit, where the powder – fine powder is moved from the bottom to the top. And it can be used in kind of a conveying also. So, powder material can be transported through the unit. So, that is solid fine; and, you have very high velocity. So, it is a kind of pneumatic conveying system; where, the solids are leaving the bed continuously.

So, continuously, their solids are coming along with the gas and then they are leaving. So, it can be a kind of elutriation also. So, it means when you look this kind of system – a bubbling fluidized bed or a reactor, which may be a kind of circulating fluidized bed; one needs to control the velocity of the solid; one needs to control the velocity of the gas. So, basically, velocity of gas will decide the particle velocity. So, that is important in this case when you have a fluidization. So, the size of the particle is important; viscosity density of the gas – all these becomes very important in order to give a definite kind of fluidization. So, these are also known as circulating fluidized beds, because the particle moves up and down and one can have a circulation of the solids within the bed.

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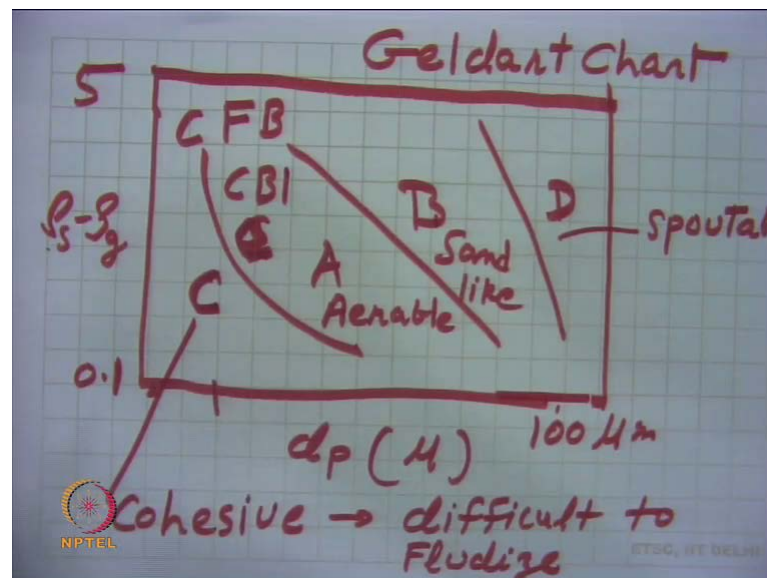
Some Key Terminology

- **Attrition:** breakdown of particles
- **Choking:** collapse of a dilute-phase suspension into a dense-phase flow as the gas velocity is reduced at constant solids flow
- **Circulating fluidized bed:** Unit designed to send particles around in a loop continuously, with no upper interface within the bed.



So, there are some key terminals here or some key terminology in order to define the different characteristics of the solid.

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And, what I was talking a Geldart chart. So, Geldart chart is something; briefly, I can just discuss here in this, which is a different category here. So, we have $\rho_s - \rho_g$ here; and, this is as a function of the diameter of the particle. So, you can have different size of the particles like say 50 microns. So, this is in generally in the micron unit; and, it can go up to a level of 100 micron or even higher to 1000 micron; something like that. So,

these are small sized particles. And this side, you have a density of the particle, which can be difference of the density of solid minus gas. So, 0.1; and, here it can be up to 5 also. So, very... That the solid density is high and gas density is always low. So, effectively, it is the density of the solid, which is dominating. So, you can have different kinds of classifications or standard curves. So, this is known as a type C, which is a cohesive material – this zone. So, this is showing just like a floor type of material, which is difficult to fluidize basically. So, lumps go along with the gas here.

And then, in between, we have a kind of B classification. Then, there is a kind of category D classification. So, overall, if you see this, you have the range now. So, this is the D type; this is the B type; and, this is the C type; in between, this is the A type. So, this is A type. So, this is the cohesive material here; so which is generally for very fine powder just like floor type of material. So, very small particle size may leave the bed easily if you have high superficial gas velocity. And this side, it is the density of the solid and density of the gas. So, depending upon this, one can select a definite condition for required for fluidization. So, it means particle size is important; density or physical property of the solid material is important; gas is important; and, other viscosity of the medium – that is important.

So, this is what I am talking a bubbling or a type of aerable particles here. So, aerable particle – this is one classification, which is known as where the bed expands and then forms the bubble. So, this is what a good region for classification CFB, what is known as a bubbling fluidized bed or a circulating fluidized bed. So, I will write it CFB – circulating fluidized bed. So, here in this region, at a... So, it means some definite size of particle and the nested difference; you will have a better situation. Whereas, this is a kind of cohesive system here. So, cohesive systems. So, that is difficult to fluidize. So, these sized particles, which are very fine in one way; and, they are very difficult to fluidize.

Here in this D type, this is known as a spoutable; and, this is a kind of a spouted bed, which is for large sized particles especially. So, basically, it is not able to fluidize easily again. So, that is, just the particles are larger in size; then, density is also high. So, they are very difficult to fluidize again. This is what the zone, where we feel that, this is a better way of fluidizing. This is for sand-like particles here; which can be fluidized under certain condition. So, this is the category, where bubbles grow up and size increase. So, gas bubbles will be there and solids will be surrounded by that. So, this is what – the

bubbles increase in size. So, when you have higher flow velocity; so bubble will go up. So, gas bubble will increase in the size. So, there are different classifications. And this is what is known as Geldart chart. And this is generally used for the classification of particles depending upon their degree of difficulty of fluidization. So, A, B, C, D, like that be categorized them. And depending upon the size of particle, depending upon the density difference; and then, we categorize them in different categories. And this is the basis for selecting the superficial or minimum superficial velocity required for fluidization also.

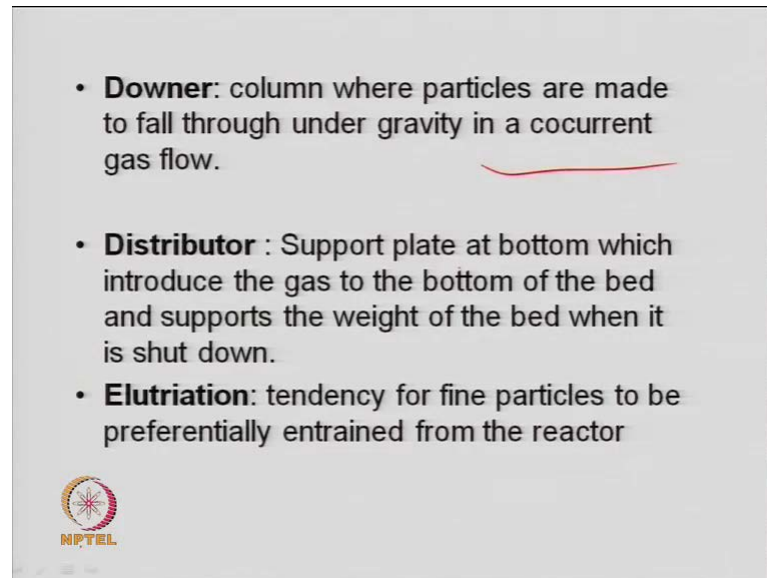
So, in that, the terminology, which is generally used; so as I said, attrition, because in the fluidization, the particles are just moving up and there are the forces acting between the particles. So, particles should have good kind of attrition resistance. If the material is soft, then it will get crumbled; and, because of that, the particle will become smaller in size and they will leave the bed; or, you will have a loss of the catalyst material. So, that is again important. So, attrition generally refers to the breakdown of particles. The second thing is choking. So, choking is again the collapse of a dilute-phase suspension into a dense-phase flow as the gas velocity is reduced at constant solid flow.

So, you have a solid, which is moving up; you have a gas, which is moving up. So, when you reduce the flow, what will happen, all the particle will come to one point. And that is what you call the dilute phase has gone down. So, it is now, has become a dense phase, where more particles or cluster of particles are located at one point. So, we have a kind of dilute phase, where the gas can pass the through that and there is a good contact between gas and the solid catalyst and that the product forms. So, dense phase and dilute phase; so these will be there.


So, if you need a good contact between... the diameter of the column should also be specified, because there should be a good contact between gas and liquid or the gas and liquid or the catalyst gas and catalyst or vapors and the catalyst when they are moving in the bed. If you are circulating fluidized bed, because I told that, degree of mixing is again important; and, fluidized bed when you have, it behaves like a CSTR. So, the conversion may be lower or productivity may be lower. So, sometimes we just enhance the productivity by just recirculating the catalyst in the bed. So, that is again required when you have a cracking reaction; and then, catalyst activity decrease because of the carbon deposition. So, you need to have a unit, which is designed to send particles around in a

loop continuously with no upper interface within the bed. So, the solids are going up; taken to a unit, where they are regenerated; and then, sent back into the system. And this is known as a circulating fluidized bed.

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- **Downer:** column where particles are made to fall through under gravity in a cocurrent gas flow.
- **Distributor :** Support plate at bottom which introduce the gas to the bottom of the bed and supports the weight of the bed when it is shut down.
- **Elutriation:** tendency for fine particles to be preferentially entrained from the reactor

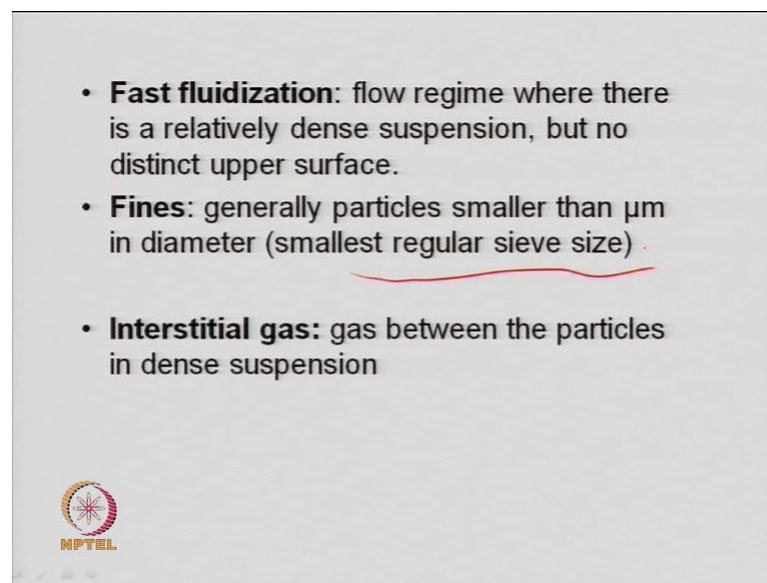


Then, downer, because the solids are moving up and coming down also. So, downer again in the column, a point, where the particles are made to fall through under the gravity in a cocurrent. So, gas is moving up. And because of the gravity, the particles will have a tendency to fall down. So, you have the velocity. So, you are trying to come the pressure drop and find the velocity, where the pressure or the weight of the particle should overcome with that velocity; and then, the particle will start bubbling or moving up. So, this is the point, which is required here in this case and that... So, downer is in the column, where the particles are made to fall through under the gravity in a cocurrent gas. So, gas is moving up and particles are falling down. And that is known as downer. And especially, again in the fluid catalytic cracking unit, there are riser and there is a known as downer also.

Again, distributor I told that, there should be a good contact. And that is to be identified through hydrodynamic study. So, we need a good kind of distributor. So, distributor is simply a support plate, which is at the bottom, which introduced the gas to the bottom of the bed. So, through this, the gas is passed from the bottom, so that the good kind of bubble should come – gas bubble should come to introduce the gas to the bottom of the

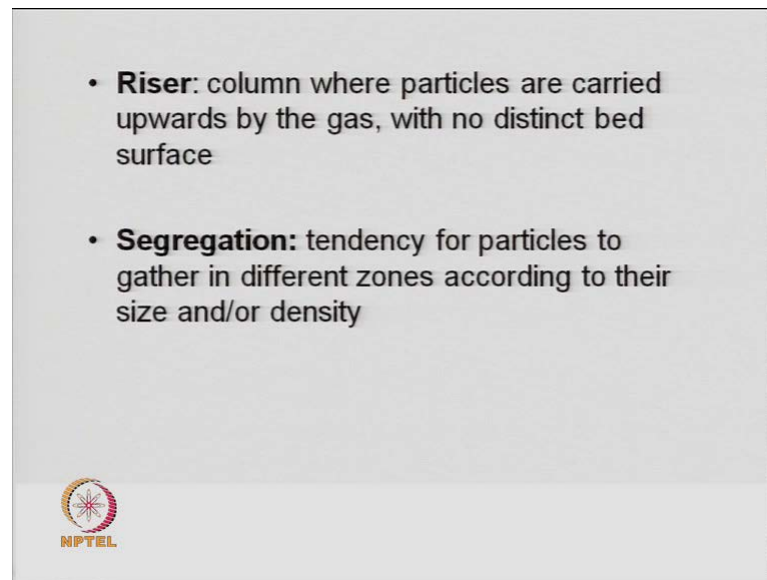
bed and supports the weight of the bed when it is not in operation. So, when it is fixed bed type, through that the solids will stay on that plate. But, when it is in movement or operation, then the gas is sparsed through that. So, it is a sparser basically or distribute. Elutriation – again it is a tendency of fine particles to be preferentially entrained from the reactor. So, during that, when the particles are smaller in size, they will leave the bed. And that is known as elutriation.

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And, again fast fluidization – this is a flow regime where there is a relatively dense suspension, but no distinct upper phase. So, as I said, when I was showing the picture in different types of fluidization, at high superficial gas velocity. So, first, you will... From bubbling fluidized bed, it will transform to a turbulent fluidizing bed and then you have a fast fluidized bed. So, basically, this is again at higher gas velocity. Then, fines are very small particles especially of the submicron level, which are less than a micron in diameter and a smallest regular sieve size; that is what is considered here. So, small size particles and they can leave the surface during the fluidization. So, these are known as fines. Interstitial gas – this is the gas, which is staying between the particles and especially in the dense suspension space. So, that is known as interstitial gas, which is between the particles or the void space.

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Then, again riser; so riser is basically known as... in the column, where the particles are carried upward by the gas with no distinct bed surface. So, in the riser say generally, the vapor and the catalysts are going up through the riser. So, that is why named riser. So, the solids are moving up; gas is moving up; and, the contact point – this is the contact point, because the diameter of this unit is small. And so there is a good contact between the gas and the catalyst. So, in the riser, you will have more and more catalytic cracking when you have the FCC type units. So, most of the cracking takes place in the riser part. So, generally, the height of the diameter is small, but height is very large; they can be of the order say 10 meters or 15 meters; very large column in terms of... and velocity is high. So, that... So, the coke formation is to be minimized especially when you have a higher range of hydrocarbons.

And again, segregation is another term in which we define the tendency for the particles together in different zones according to their size and, or density. So, whatever I was talking in the Geldart chart; so there will be lump of the particles, a crunch of the particle; and, they will be at the localized place as such in the bed column; there will be through and the gas is channeling. So, this is not a desired condition when you look at a good kind of fluid bed reactor design.

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Minimum Fluidizing Velocity: The solids are suspended when the pressure drop exceeds the weight of solids. This happens when the gas velocity exceeds the minimum fluidizing velocity U_{mf} .

Terminal Velocity u_t : Individual particles are blown out of the bed when the gas velocity exceeds what is called the terminal velocity, u_t .



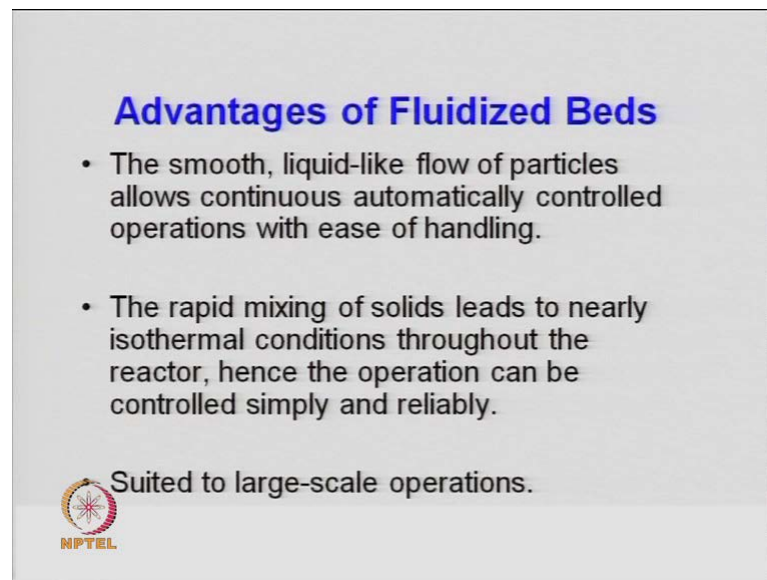
Then, there are some other terms; one is known as minimum fluidizing velocity. So, minimum fluidizing velocity is basically you need to design the reactor. And this will depend on the type of the solid material and the gas. So, basically, minimum fluidization velocity as the name says. So, it is the velocity, which is required to fluidize the bed or to allow the solids to keep them in the suspended form, because the gravity force is always acting vertically downward. So, to overcome with that, buoyant force is required. And that is what superficial gas velocity I was defining. So, depending upon the volumetric flow rate of gas divided by area of cross section of that tower; so this is known as superficial gas velocity. So, this will happen when the pressure drop exceeds the weight of the solids. And this happens when the gas velocity exceeds the minimum fluidization velocity.

So, through the packed bed experiments or fluidization experiments, one needs to calculate the minimum fluidization velocity required for the given size particles in a given column diameter and then the other things. So, this is based on because what will be the pressure drop in the bed at those conditions. So, fixed bed, then fluidizing; and, what conditions are required to make the bed fluidized continuously; what is the elutriation or loss of particle; what is the residence time or hydrodynamic study need to be done, because different size – when you have mixture of particles; so different size particles will have different degree of fluidization. So, to what it is... Suppose there are wide variation in the particle size; then, it may not be good, because fine particles will be

leaving the bed and larger size or coarse particles will not be able to fluidize under the given velocity condition. So, one needs to design or identify the conditions required to fluidize.

And, another term is known as terminal velocity. So, terminal velocity – it is for individual particles are blown out of the bed when the gas velocity exceeds; which is known terminal velocity. So, terminal velocity is basically the velocity at which the particles leave the bed. So, it will again depend on the individual size of the particle and their physical property. So, in lot of applications, if you look at the refinery, petrochemicals or chemical industry or process industry, the fluidized bed columns are being used.


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Advantages of Fluidized Beds

- The smooth, liquid-like flow of particles allows continuous automatically controlled operations with ease of handling.
- The rapid mixing of solids leads to nearly isothermal conditions throughout the reactor, hence the operation can be controlled simply and reliably.

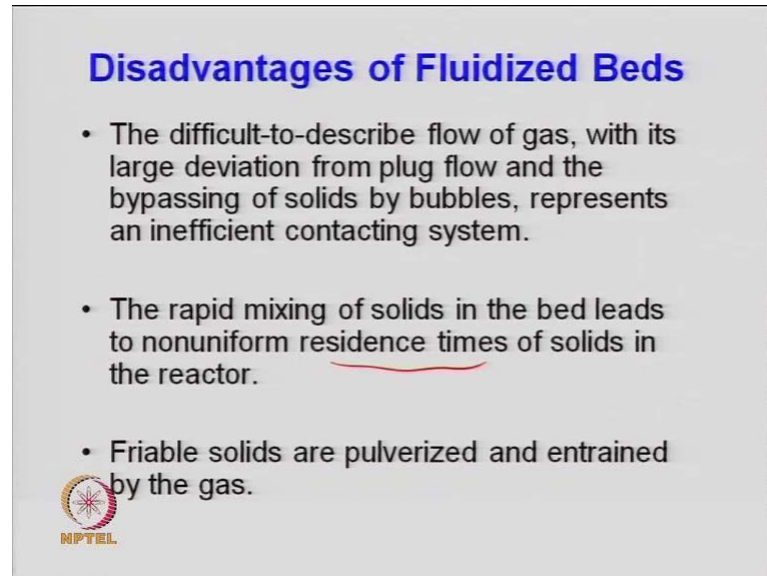
Suited to large-scale operations.



So, operation is a smooth; it behaves like a liquid-like flow of particles. So, it is just like a fluid property, what we are talking; which allows the continuous automatically controlled operations with ease of handling. So, the transportation becomes easy, operation control becomes easy when you have a fluid type of operation. And again I told you that, solids are moving up and down in this. They are going up and they are going down. So, internal kind of circulation is there. So, there will be a good kind of back mixing in this reactor. So, rapid mixing of solid leads to nearly isothermal condition. So, heat transfer; where the heat transfer is an issue, the fluidized bed reactor may be helpful. So, throughout the reactor, you will have good control of heat. And


hence, the operation can be controlled simply and reliably. And these kind of operations can be done at large operations.

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Disadvantages of Fluidized Beds

- The difficult-to-describe flow of gas, with its large deviation from plug flow and the bypassing of solids by bubbles, represents an inefficient contacting system.
- The rapid mixing of solids in the bed leads to nonuniform residence times of solids in the reactor.
- Friable solids are pulverized and entrained by the gas.



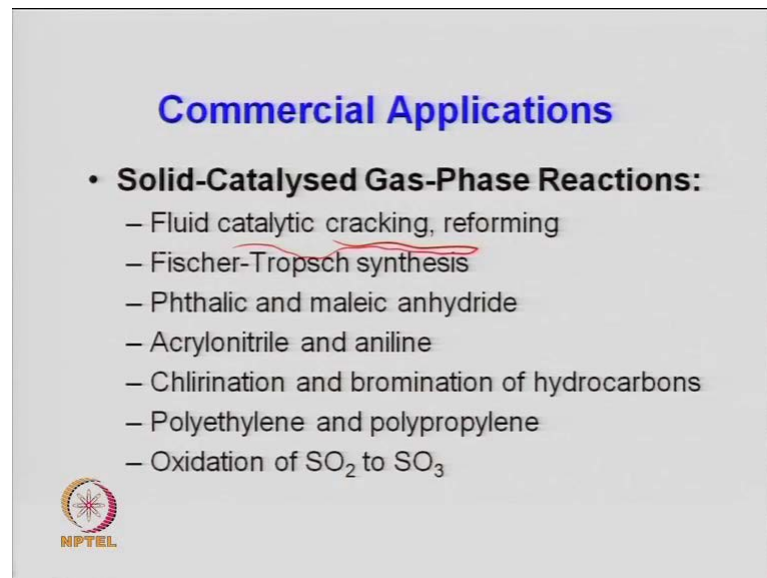
But, there are disadvantage also. The particles that is... It is very difficult to describe the flow of gas with its large deviation from plug flow. As I said that, the fluidized bed reactor generally behaves like a CSTR – mix reactor. But, we try to... because the CSTR – you know the conversion will be lower; and, when conversion is lower, then the productivity will be low. So, what we want? We want to have a more and more plug flow type design. So, it means you need to do the high velocity, higher condition, higher severity of operation. So, the plug flow condition – maintenance of that is difficult. So, we need to find out the RTD; and then, we need to find out what is the exit A's distribution of the particle, which you get generally from their RTD study.

And then, you can correlate the actual residence time inside the reactor. And it is not exactly... It is not behaving like a plug flow reactor. So, there will always be a conversion less than the plug flow reactor. So, this is one thing. So, when you try to high velocity, there may be a bypassing of the solids by bubbles. There may be kind of channeling. So, some unwanted, because more and more voids will come and the gas may channel through those voids. So, this is again a problem when you try to maintain the plug flow condition in a fluidized bed reactor.

And, rapid mixing of solid in the bed – and, that leads to non-uniform residence time of solid in the reactor. So, that is again, as I said, there is a kind of degree of back mixing and there is a downer and the solids are moving up; they are coming down. So, there will be a kind of a degree of mixing. And because of that, there will be always be a non-uniform residence time distribution. So, RTD, which is generally calculated from e to the power minus t divided by t bar, and integrated to the whole length 0 to infinite something. So, dt . So, that is something that concentration profile needs to check based on the radioactive pressure. And then, one needs to utilize that, mean conversion, which I was talking right for the given system.


So, that is again an issue here that the residence time – t bar is not a definite; there is a variation in that residence time; it may vary depending upon the flow condition. And friable solids are pulverized and entrained by the gas, because when you have very fine, the particle which is very softer; so what will happen; it will crumble during the operation. So, attrition because of the attrition; and, it will become finer. And then, it will go along with the gas. So, there will be a loss of the particles. So, these are some kind of disadvantage.

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Commercial Applications

- **Solid-Catalysed Gas-Phase Reactions:**
 - Fluid catalytic cracking, reforming
 - Fischer-Tropsch synthesis
 - Phthalic and maleic anhydride
 - Acrylonitrile and aniline
 - Chlorination and bromination of hydrocarbons
 - Polyethylene and polypropylene
 - Oxidation of SO_2 to SO_3

 NPTEL

But, fluidized bed reactor as I said, they are widely used. So, one is the solid-catalyzed gas-phase reaction; they are widely used. So, one of the most important is the FCC cracking. So, more than 420 FCC reactors are available worldwide and they are

producing a large amount of gasoline from the heavier feed stock. So, larger molecular weight hydrocarbons, which can be a $C_{20}H_{42}$ – something like a paraffin hydrocarbon; crude oil fraction; and, that is cracked to a product like $C_{10}H_{22}$ – something like that. So, which is towards the gasoline. So, we look at something like C_5 C_{15} or C_{18} range of paraffin hydrocarbon. So, by cracking, one can do that. So, FCC reactors are widely employed for the production of gasoline, what is called FCC gasoline.


So, this is one important application of this fluid catalytic cracking. And since these hydrocarbons tend to form coke or carbon; so in this kind of unit, it is possible to regenerate the catalyst time to time. So, that is again another advantage. So, this is the very wide operation. Then, same thing – reforming is done in this kind of reaction – plate forming what I was talking. So, different kind of semi-regenerative process or continuous regeneration of the catalyst is possible. Then, you have this kind of operation. So, something like a moving bed reactor, which was used earlier. And now, this is being replaced by this continuous region reforming by using this fluidized bed reactor system.

Then, Fischer-Tropsch reaction also – they are looking this kind of system, where catalyst is being bubbled or fluidized or something like the slurry bubble column reactor is generally used, because it is a highly exothermic reaction. But, when fluidized bed... then, again the control of heat is better. So, whatever the heat, which is generating; that can be removed. And that is the advantage here. Similarly, there are several other processes in the petrochemical and pharmaceutical industry, where these kind of reactors or fluidized bed reactors are used as oxidation of sulfur dioxide to sulfur trioxide.

Generally, the fixed bed system is used, but reaction is highly exothermic. So, again, this fluidized bed reactor can be a better option. Same thing for acrylonitrile and aniline; and, HDPE – high density polyethylene, low density polyethylene in the petrochemical industry; where, by polymerization of ethylene, these kind of high density polyethylene, low density polyethylene, polypropylene – these are made. So, Ziegler-Natta type catalysts are generally used; and, they are in the fluidized bed reactor. So, wide applications on the catalytic as well as non-catalytic also.

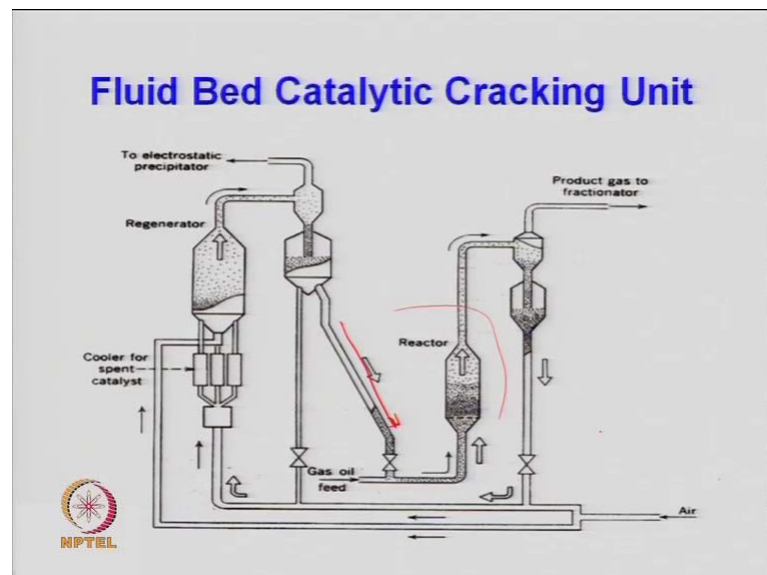
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- **Gas-Solid Reactions:**
 - Roasting of ores (ZnS, Cu₂S, nickel sulphides, etc.)
 - Combustion and incineration
 - Gasification, coking and pyrolysis/carbonization
 - Calcination (limestone, phosphates, aluminium hydroxide)
 - Fluid coking
 - Reduction of iron oxide
 - Catalyst regeneration



You can say here roasting of all the ores, glass furnace; so combustion, incineration, coal gasification, pulverizers. You might have learnt about coal combustion process. So, these are generally done in the fluidized bed reactor. So, these are known catalytic type of operations also.

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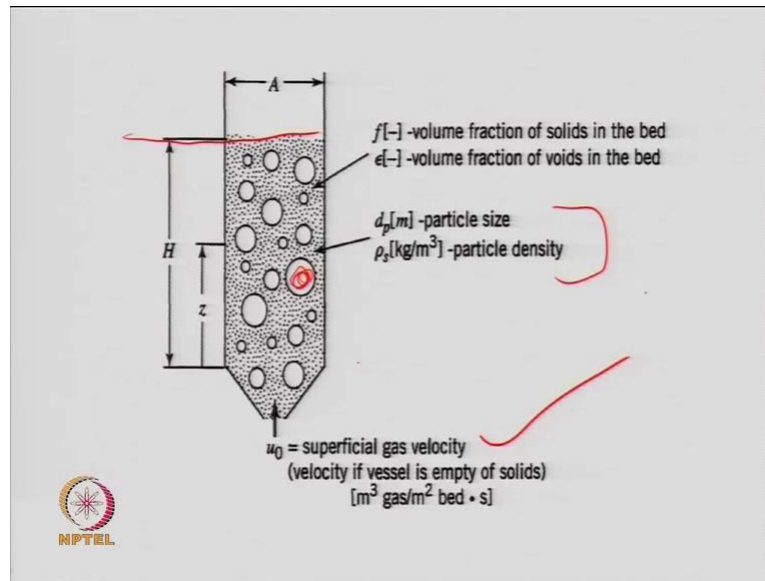
Just a typical example, which has been taken from the reference that, a simple reactor and regenerator combination, which is FCC unit – fluid bed catalytic cracking unit; so you can see here that, you have a reactor system like this. So, this is the reactor part. And

the gasoil, which the feed – gasoil is nothing but the atmospheric gasoil or vacuum gasoil, which is obtained from the bottom of a distillation unit, crude oil distillation unit. So, gasoil comes like this and you have a regenerated catalyst, which is coming from here. So, catalysts circulation rate one can calculate based on this; how much temperature is required in the regenerator section. So, you have a fine powder of the catalyst coming from here, which may be 100 to 150 to 100 micron or up to 150 micron; and, accordingly, depending upon the velocity of this. So, this is acting as a gas for fluidization also; that is, vapors of the hydrocarbon. And sometime, steam is also used to strip out the hydrocarbon fraction along with this. So, this is taken here and this is the reactor part. So, you have the dense bed and dilute phase. So, the product forms, the vapor forms. So, the vapors are taken to fractionator for distillation.

But, the catalyst is separated here by using a cyclone separator and then taken back into this unit, where the air is used for oxidation, because I told you regeneration of the catalyst is done by burning this catalysis – spent catalyst in the presence of air at around 550 degree centigrade for 5 hours. So, here roughly, the carbon reacts with the oxygen and converts it into carbon monoxide, carbon dioxide. So, this is being done and in a regenerator unit here. So, this is along with this catalyst comes here. And since the reaction is exothermic here; so proper heat control is required.

So, sometime this heat is being transferred here to vaporize the hydrocarbon depending upon the unit. And here it is a cooler for a spent catalyst, which is just to cool it down below, because the temperature should not exceed above 750 degree centigrade here. So, the regeneration is done here and whatever the particle – they are taken into electrostatic precipitate or so. You can separate along with CO-CO₂ through that unburned oxygen and nitrogen. And the catalyst is taken again here, separated, and then sent back. So, that is a kind of circulating fluidized bed also – catalytic ((Refer Time: 30:59)) So, this is very important or wide application in petroleum refinery for the production of FCC gasoline for gasoline production.

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The design and the hydrodynamic study; that is, when we look at a reactor design; so in general, it is a very simple; you know that, we are talking something like a CSTR if you just...

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Handwritten equations on a grid background:

$$\frac{W}{F_{A0}} = \frac{X}{(-r_A)}$$

$$(-r_A) = \frac{k' C_A}{RTD + \tau}$$

Simplest case, which say we assume that, it behaves like a CSTR, then one can write simply this thing W by F_{A0} is equal to X upon minus r_a . And one can calculate if this is known. So, rate of reaction is to be calculated; which will be function of the hydrodynamic, because there will be a mass transfer of the gas. So, there may be

different phases. So, that is important that, how to find out the rate. But, once that is known, then this is simply the mass of the catalyst required for a given conversion X in the given feed rate of F_A . But, simultaneously, one needs to check the distribution of the particle, because... But, what we are talking is not a fixed bed reactor; it is a moving bed reactor; and, not a moving bed reactor, but the particles are in bubbling condition. So, there will be voids. So, we need to look at those conditions. And depending upon that, you have to find out the degree of mass transfer and degree of chemical reaction, because every step will have its own resistance.

So, overall rate – one can write it something like that – $-r_A$ is equal to some constant K times whatever the kinetic; say if it is first order with respect to say this; or, whatever the n -th order C_A to the power n ((Refer Time: 32:36)) But, this is not just like the kinetic rate constant; it will be a contribution of the mass transfer resistances also. So, that is important that, how to find out this rate of reaction in the case of a fluidized bed reactor; that is one thing. Simultaneously, one needs to find out because this conversion will depend on the residence time. So, the smaller particle may spend lower time and larger particle may spend higher time; or, depending upon the superficial gas velocity and the size of the particle, there will be a distribution of the particle. So, the conversion will also be a function of residence time distribution. If it is plug flow, it is fine; it will be just like τ . But, most of the time, it is not a plug flow condition.

So, one needs to find out the mean residence time – \bar{t} . And this mean residence time – \bar{t} will depend on the age of the particle. So, we need to find out some kind of exit age distribution for each size range of particle. So, conversion will be now mean conversion of the solid. So, what I am talking here? This is something like... or, if it is non-catalytic solid; otherwise, it is conversion of the gas. So, one needs to calculate mean conversion, which is not constant; which will depend on the residence time, that is, \bar{t} . So, one has to calculate this concentration as a function of time on a stream or the mean average concentration throughout the reactor and then calculate. So, simplest is one can take a simple CSTR and take the rate equation and can find out. But, this is not a correct way.

So, the purpose of showing this here is based on that, depending upon this superficial gas velocity, which is being used to fluidize the solid particles; so that is important. So, this is a superficial gas velocity. So, that is all of you know; this is the velocity if the vessel is

empty of solid. So, whatever the area of cross section of this pipe multiplied by the velocity; that is known as the volumetric flow rate. So, if you know the volumetric flow rate of gas or if you know the mass flow rate of gas; so mass flow rate is nothing but density of the gas multiplied by, that is, the volume of that volumetric feed rate. So, one can very easily find out this thing.

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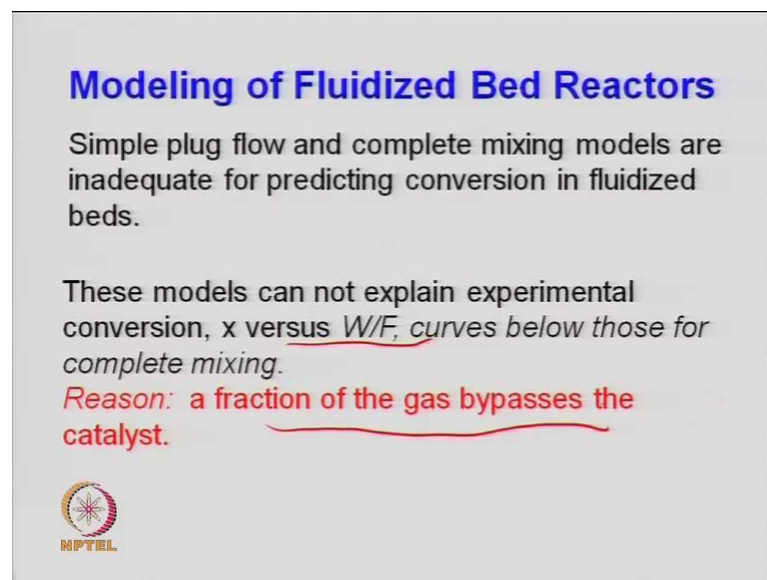
$$\frac{\pi D^2}{4} u = V_g$$

So, if I know that diameter of the column – simply D; so pi D square by 4. This is the area of cross section and multiplied by the velocity. And that will be the volumetric flow rate of the gas V_g . So, one can very easily find out this superficial gas velocity. So, this is called superficial gas velocity, because it is calculated based on empty bed. Then, diameter of the particle size; so different sized particles may be there; single sized particle. So, these are small dots, are representing the particles – particle size and their density – both are important as I discussed. Fraction set – the volume fraction of the solid in bed. So, basically, what is the mass of the solid in the bed? Or, we define it in terms of the volume fraction of the solids in the bed. So, that is again important.

And, the void fractions – volume fraction of the voids in the bed – that is again important in each phase. And this is what is shown here – this one; this is the gas bubble. So, there are gas bubbles you can see here, because they are moving up; the gas is moving up; which is surrounded by a solid now – some crunch of the solid particle and some solid, which is there in the front positions.

And, that is known as kind of emulsion form. So, let... And this height of the bed can be calculated simply from the mass balance. So, one is the height of the packed-bed, which is the... when there the bed is not fluidized. And when the bed is being fluidized, but mass will remain same. So, one can very easily calculate it based on the solid balance. So, something like H_f into $1 - \epsilon_f$. If I say H_f into $1 - \epsilon_f$. So, this is a fraction, because ϵ_f is the void fraction. So, $1 - \epsilon_f$ will be the solid fraction. So, H_f into $1 - \epsilon_f$ will be equal to H_m into $1 - \epsilon_m$, because that is the fraction when the bed is packed. So, that time, height will be smaller. So, overall mass will remain same. So, this is what the mass balance.

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Modeling of Fluidized Bed Reactors

Simple plug flow and complete mixing models are inadequate for predicting conversion in fluidized beds.

These models can not explain experimental conversion, x versus W/F , curves below those for complete mixing.

Reason: a fraction of the gas bypasses the catalyst.



So, when you do the modeling or when you design these kind of fluidized bed reactor; so I am not going in detail, because that is a separate course. But, very quickly, I will just give you some glimpse – what is the approach to design it. So, modeling of the fluidized bed reactor means related to something like hydrodynamics; and, something is used to find out the overall rate of reaction or resistance, which is offered. So, simplest as I said, a combination of the plug flow and a CSTR. So, simple plug flow and complete mixing models; they may not be good to define the fluidized bed, because there is some kind of back mixing; there is some kind of the plug flow condition.

So, exactly, by considering a plug flow, that is also not correct. And completely, a back mix reactor CSTR is also not correct, because there is a kind of residence time


distribution, which is between the plug flow and the CSTR. So, in general, what I discussed is the conversion versus W by F naught curves. And you need to look at that degree of back mixing. But, they are not well to define a plug flow reactor. The reason is that, the fraction of the gas bypasses the catalyst. So, some of the gas is leaving the bed without contacting the solids. And this will happen, because there will be pockets of the solid and the gas. So, there are the voids between these solid particles; and, the gas leaves them. So, it may not be good to define it like a CSTR or like a plug flow reactor.

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Hydrodynamic Flow Models.
The behavior of rising gas bubbles is important ,
since they probably cause much of the difficulty.

Two developments :

The first is Davidson's (Two Phase model)
theoretical development and experimental
verification of the flow in the vicinity of a single rising
bubble in a fluidized bed which is otherwise at
minimum fluidizing conditions.

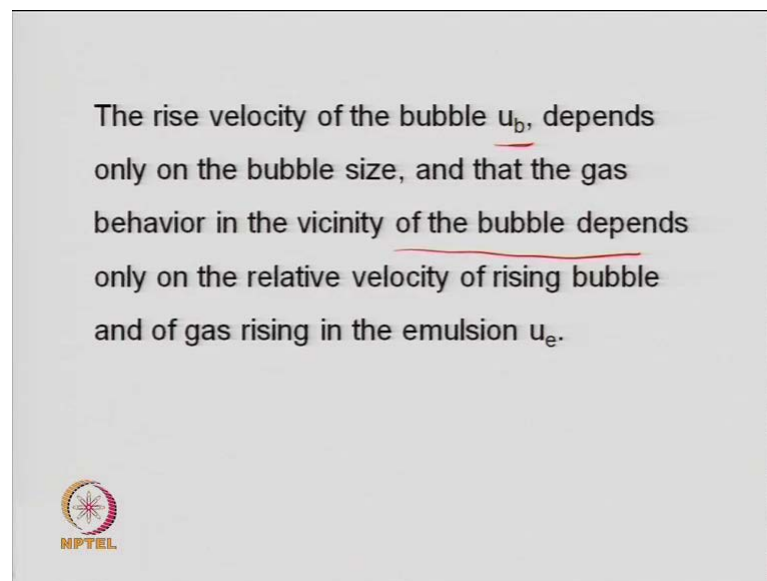


So, we need to do hydrodynamics. So, hydrodynamics of this fluidized bed reactor is very important, very crucial to define the residence time distribution. So, let us look quickly on that. So, we have to look at the behavior of a rising bubble, because as I said, the gas moves up like this. So, what is the behavior of this gas? It will depend on many factors. It will depend on the particles; it will depend on the density; it will depend on the size; and, depending upon the reaction also – kinetics. So, this is the first one, which is known as two phase model. So, there are generally two models, which have been used. But, this is the simplest one – a two phase model; where, we are saying that, there is a bubble phase and there is the emulsion phase.

So, emulsion phase means solid phase; and, bubble phase means the gas – the gas bubble, which has some solid surrounding it. And experimentally, under certain condition, this is valid. But, most of the time, a three phase model, which are also known

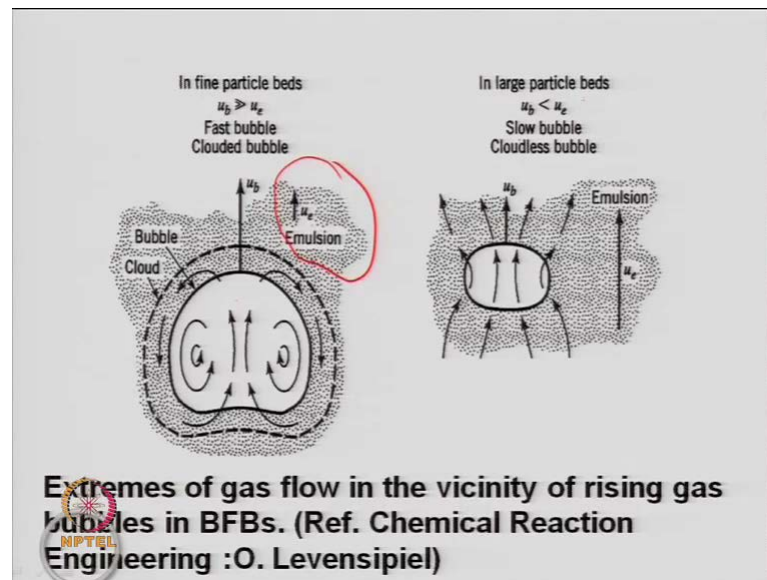
as Kunii-Levenspiel model. So, this is the model, which is developed by Davidson. So, sometime we call it Davidson model also – two phase model. So, it simply says that, in the vicinity of a single bubble in a fluidized bed, which is something like in the... There is a solid and there is a gas. So, bubble and the emulsion phase. So, solid – clusters of the solid – I am calling emulsion, which are chunks of the solid. So, there is less reaction. But, most of the reaction when we are saying, the gas, which is in contact with the solid and there is a reaction. So, let us just talk on that.

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So, it says that, the rise velocity of the bubble, which is I am defining now u_b ; which has something like a unit of meter per second. So, this will depend on the bubble size. So, these are just the hydrodynamic correlations and the mass balance and the force balance equation; one can calculate this. So, the bubble size is important – velocity of the bubble. So, that is very important because that bubble size will decide the rise velocity of the bubble. In the bed, at what velocity the bubble is going up, because residence time of the gas bubble is important. And for how much time that gas bubble is in contact with the solid, will decide the conversion or average conversion. So, that is important. So, the gas behavior in the vicinity of the bubble depends on the relative velocity of the rising bubble and the gas rising in the emulsion. So, gas is in the emulsion phase; gas is in the bubble phase. So, let us see that.

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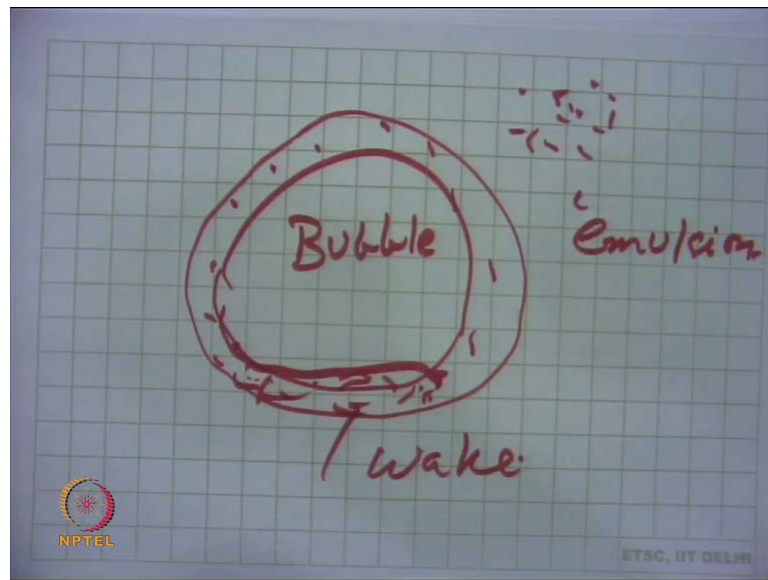


So, this is what the picture of a bubble. So, here we have a bubble. So, this is a gas bubble. The white one is showing a gas bubble, which is surrounded by a solid. So, basically, this is a three-phase model, but very easy to understand; and then, one can convert it into two-phase model. So, you have a bubble here; you have a cloud. So, we are calling the film, the solid is... The bubble is surrounded by a solid – thin film of the solid. And here we will have more reaction, because there is a good contact between the gas and the solid. So, this is the bubble; this is the cloud. Here is again some more crunches of the solid. So, basically, this is known as a wake phase – another phase. And this is where the gas is low or some stagnant gas. And this is known as emulsion phase here.

So, now, we have mainly – if I say the two-phase, then there will be a bubble and the emulsion. If I say three-phase, then there will be a bubble, then there is a cloud, and then emulsion. So, cloud and wake – these two are combined together. So, bubble phase, cloud and wake phase, and emulsion phase. So, that is known as a three-phase model. So, what is there; in fine particles bed, the u_b , which is the velocity of the bubble is much more higher than the emulsion phase velocity. So, fast bubble and clouded bubbles here – something like that; the bubbles are moving up. When the particle size is larger, then bubble velocity is a smaller. So, it will depend on the particle size. And the bubble velocity is smaller than the emulsion velocity. So, these are... I am not going in detail of them; but what I mean to say; when I discuss a fluidized bed reactor, we are just trying to

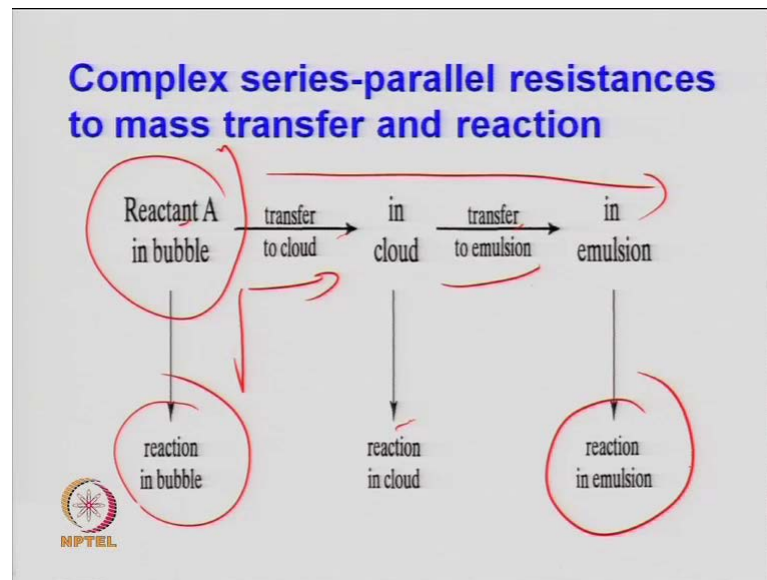
visualize it something that, what is the degree of mixing; what is the mass transfer in each zone; and, what is the chemical reaction contribution. And that will decide the overall resistance, which may be a combination of some are in series and some are in parallel.

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But, just as a simplest, what we assume? That there is a bubble – something like a bubble; this is a gas bubble; on this, there is surrounded by a film of the solid. So, there are solid particles around this; and, there is some kind of wakes, because here at this point, you will have something more concentrated solids here. So, this is known as a wake sphere – wake phase. And then, outside of this bubble, this is something like a emulsion phase. So, this is the bubble and this is the emulsion when I say a two-phase model.

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And, this is what the complex series-parallel resistance to mass transfer. So, when you have a packed bed reactor, you can easily eliminate these mass transfer resistances. But, when you have a system like a fluid bed reactor, the system in that system – it is not easy to eliminate them. So, one needs to control them. But, if it is a plug flow; of course, we can minimize them under the plug flow condition. So, basically a complex series-parallel resistance to mass transfer and reaction. So, we have a reactant A in bubble. That is what I said that, reactant A in bubble – it will react, because it has a solid surrounded. So, there will be some reaction; there is some concentration of the solid. So, depending upon that, there will be some reaction in the bubble. The rest, which could not react – it will transfer to the cloud. So, this is something here – the bubble. So, there is some solid here; there will be a reaction; and, rest will transfer to the cloud phase.

Again in the cloud phase, we will have some reaction in the cloud, and then... which is not reacted – it will transfer to the emulsion phase. And finally, the emulsion phase – there will be nothing to transfer. So, whatever there – that will react in the emulsion phase. So, we have now a combination; that is, in series, if you look at this side, it is in series, and a combination of series and parallel. So, mass balance – one can very easily write. Whatever input here; that is transfer to the cloud plus reacted. Whatever input here; that is transferred to the emulsion and reacted. Whatever here; that is transferred to the product. So, that is a very complicated series-parallel reaction, is used to find out the overall rate of reaction.

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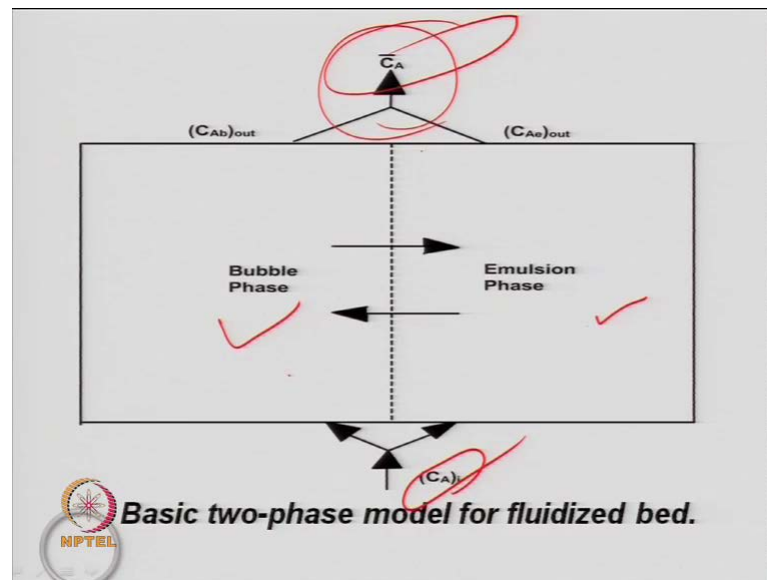
A two-phase model contains three parameters:

- (i) the cross section of the bed taken by either the bubble or the emulsion phase,
- (ii) the interchange coefficient, and
- (iii) the axial effective diffusivity in the emulsion phase.



So, if I just talk only a two-phase model; so in the two-phase model, we just talked a cross section of the bed, which is taken either by the bubble or by the emulsion. So, what fraction has been taken by the emulsion phase and what fraction of the bed has been taken by the bubble, because we are not considering any other phase. So, this is one parameter – the fraction of the bubbles in each phase. Then, second – interchange coefficient, because there is cross flow basically. What I discussed here; so there is a kind of cross flow – transfer and chemical reaction. So, it means the interchange coefficient. So, that is, from bubble to emulsion, how much is the mass transfer and how much is the reaction. So, interchange coefficient – that is the second parameter. And third is the axial effective diffusivity in the emulsion phase, because the reaction will depend on that effective diffusivity of this gas into the solid. So, that is again important here. So, it is basically, there are three parameters when you look at a two-phase.

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And, this is what is shown here. So, when an actual picture is different; but when we are assuming a two-phase model; so we have a bubble phase; we have a emulsion phase here. So, something is coming here in terms of the concentration or mass. So, one can very easily write a mass balance here, which is partly... So, this mass gas is going partly into the bubble phase and partly coming into the emulsion phase. And then, when it leaves, part of the concentration is coming from the bubble phase C_{Ab} out and part of the concentration is coming from the emulsion side of a concentration of gas A in the bubble, concentration of gas A in emulsion; and then, this is coming at the exit. So, we get the average concentration.

So, we have modeled it like this. But... So, this is the black box, where you need to evaluate the parameter. And to match it with, this is the input and this is the output. And that will decide the conversion, whatever \bar{C}_A by C_{Ai} ; or, one can calculate the X or mean conversion based on that. But, actual conditions inside the bed will depend on the gas velocity, superficial gas velocity, density of the solid, and size distribution of the solid.

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Two Phase Model


- A fraction of the total flow rate through the bed is considered to be in the bubble phase, the rest in the emulsion phase.
- Between both phases there is a certain interchange or crossflow.
- At the outlet both streams, with their respective conversions, are hypothetically mixed to give the exit stream, with its mean conversion or concentration.



So, just if you look at here – the two-phase model; so in this, the fraction of the total flow rate through the bed is considered to be in the bubble phase and rest in the emulsion phase. That is what I discussed. And between both phases, there is certain interchange or cross flow. So, it is a kind of transfer coefficient. We need to calculate that either from some empirical correlation or some from the hydrodynamic condition. So, it is a kind of mass transfer coefficient or overall transfer coefficient, which can be determined experimentally or from the model. So, at the outlet, both streams with their respective conversions are hypothetically mixed to give the exit stream with its mean conversion or concentration. So, this is what I was discussing here. So, all these are discussed here.

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- Since there is no reaction in the bubble phase and because of its high velocity the flow through that phase can be taken to be of the plug flow type.
- In the emulsion phase various degrees of deviation from plug flow can be postulated.



So, since there is no reaction... So, we assume that, there is negligible reaction in the bubble phase in the two-phase; and, because of its high velocity, the flow through that phase can be taken to be of the plug flow. So, bubble phase is assumed to be plug flow. But, in the emulsion phase, the various degrees of deviation from plug flow can be postulated, because now, solids are moving up and down. So, there is a kind of back mixing in this. So, that cannot be considered as a plug flow.

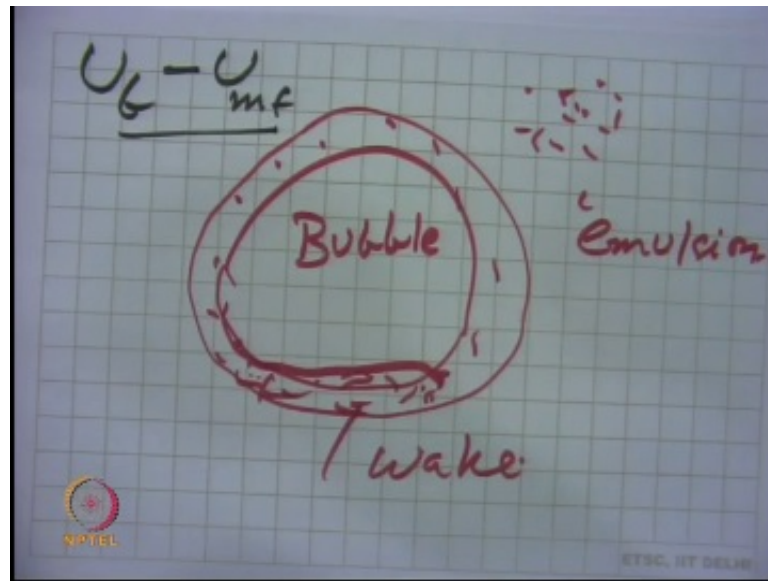
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The cross section of the **volume fraction of the bed occupied by bubbles** is derived from the observation that essentially all the gas flowing through the bed in excess of that required for minimum fluidization **goes into the bubble phase**.



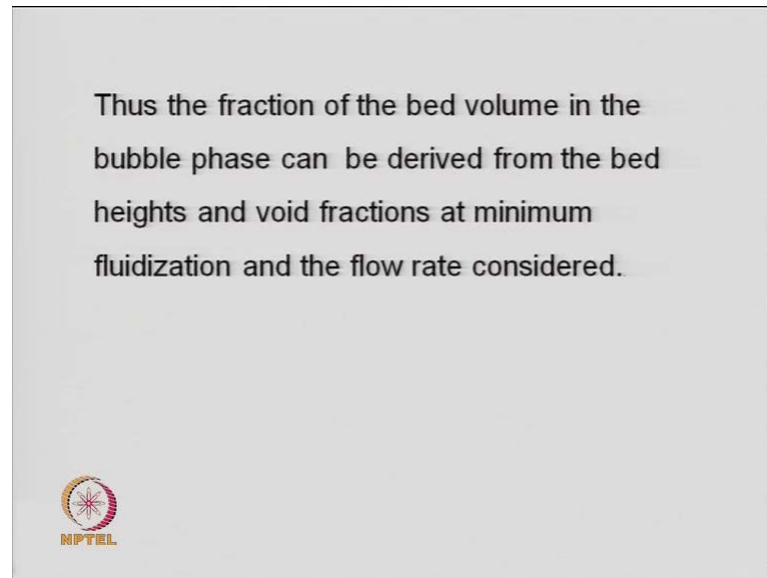
So, cross section from bubble to emulsion – transport; the cross section of the volume fraction of the bed occupied by the bubbles is derived from the observation that, essentially, all gases flowing through the bed in excess of that required for the minimum fluidization goes into the bubble phase. So, basically, what you get? That you get from the minimum fluidization condition. So, gas flowing through the bed, which is in excess is nothing but what goes to the bubble phase.

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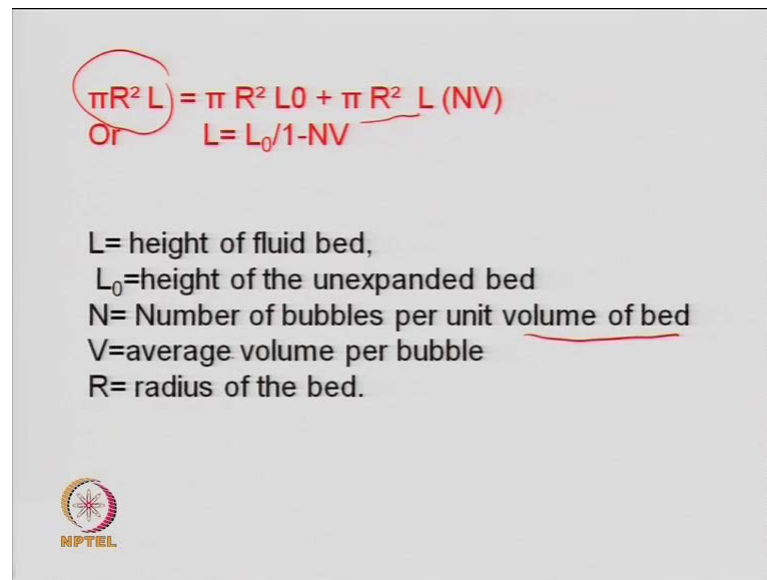
So, basically, when you write a mass balance, everything is calculated above this – that is, u_b minus u_{mf} , or if I calculate some velocity in the bubble or in the cloud phase. So, whatever excess of the minimum fluidization; that will go into the bubble phase. So, that is one essence that, the mass flow rate, which we are calculating – a volumetric flow rate, which we are calculating from the bottom of the column in terms of the superficial gas velocity. And that is allowing the bubble to rise. And that will happen... Bubble will rise only when it exceeds the minimum fluidization velocity. So, that is what the bubble rise velocity also we have to define accordingly in this. So, I am not going in detail of that. But, one should have some idea when he looks for the fluidized bed type of reactor design; and, details can be in the subsequent courses.

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The fraction of the bed volume in the bubble phase can be derived from the bed heights and the void fractions at the minimum fluidization and flow rate. So, basically, it is a mass balance. So, you know that, the fraction of the solids in the bubble phase, fraction of the solids in the emulsion phase; and, if there is some other wake phase. So, all these will be total fraction of the... void fraction of the bed. So, one can... Total fraction of the solid in the bed. So, one can calculate that simply. And same thing – the height of the bed, which I was talking – the fluidized bed height; so minimum fluidized bed height; and, related to the void fraction of the bed. So, it is just a mass balance for the solid. So, one can calculate that. So, a fraction of the bed volume in the bubble phase can be derived from the bed heights and the void fractions, which is at the minimum fluidization and the flow rate considered at that point.


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$$\pi R^2 L = \pi R^2 L_0 + \pi R^2 L (NV)$$

Or
$$L = L_0 / (1 - NV)$$

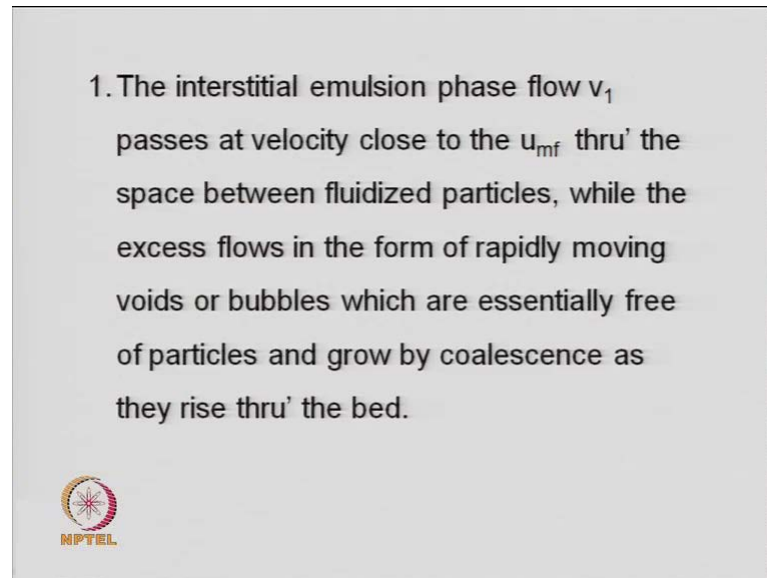
L = height of fluid bed,
L₀ = height of the unexpanded bed
N = Number of bubbles per unit volume of bed
V = average volume per bubble
R = radius of the bed.



So, this is simply written here in terms of that pi R square L. So, this is the radius of the bed or diameter, which I was discussing – pi D square by 4 into L, which is the height of the bed – height of the fluid bed. So, this is nothing but the volume of the bed. So, then, it is fluidized. And pi R square L₀ if you say, this is the height of the fixed bed or a packed bed when it is not expanded. So, area is same; height is now H₀ or L₀. So, this is the initial volume. And whatever the rise – that will be in terms of the how many bubbles have come, because the bed is rising, because in between in the interstitial space, there are gas bubbles now.

So, nothing but pi R square L; which is now the volume of the bed in the fluidized condition times whatever NV; which is nothing but representing the number of bubbles per unit volume multiplied by the average volume per bubble. So, basically, bubble volumes. So, pi R square L into N into V; where, N is the number of bubbles per unit volume of the bed; and, V is the average volume per bubble. So, this product will give you the total volume of the bubbles. So, that is nothing but expansion, which has arrived or bed height has been increased because of these gas bubbles. So, simple material balance or a volumetric balance in terms of the bed height.

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Some salient... Again, features when you look at that design; the interstitial emulsion phase flow v_1 – that pass at velocity close to the minimum fluidization velocity. So, because I told you that, in the emulsion phase, the solids are together altogether; in the bubble phase, the gas is there – more gas; and, reaction is not with the solid; reaction is when the gas contact with the solid. So, this is a one important thing when you have a three-phase model. So, more reaction is in the bubble phase – bubble phase, cloud phase, because it has the solid and the gas. And the other – in the emulsion phase, solids are more; the gas is just at the almost zero velocity there.

So, this is one thing that, interstitial emulsion phase flow v_1 at velocity close to the minimum fluidization velocity through the space between the fluidized particle; while the excess flows in the form of rapidly moving voids or bubbles, which are essentially free of particles and grow by coalescence as they rise through the bed. So, when the bubbles move up, they will now increase in the size; so that is, they are growing in size; the bubble size is increasing when it goes up. So, this is one thing when you look at the velocity v_1 or interstitial velocity in the emulsion phase.

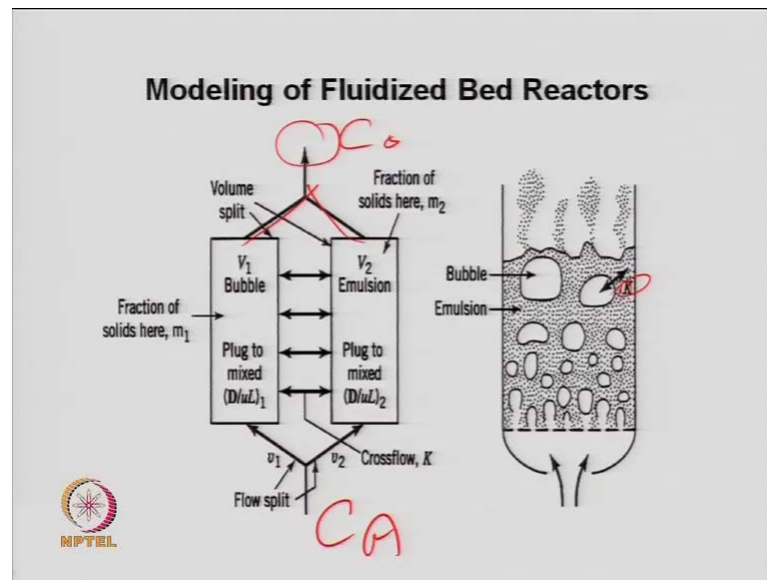
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2. Bubbles phase gas does not make good contact with soild → reaction in bubble phase is inefficient.
3. Gas flowing interstitially is in intimate contact with particles and if chemical kinetics are favorable the efficiency of chemical conversion in this phase will be high.
4. Exchange between two phases is determined by the size of the bubbles and decreases as they become higher.



Second thing is the bubble phase. So, bubble gas phase – this does not make good contact with the solid. So, reaction in the bubble phase is inefficient in the two-phase case. And the gas, which is flowing interstitially, is in intimate contact with the particle. And if the chemical kinetics are favorable, the efficiency of chemical conversion in this phase will be high. So, that is again within the particle; interstitially, the gas, which is present between the particles. So, there if the kinetics is favorable, temperature is ok; then, the reaction will be in the interstitial space; that is, where the gas is there and the solid is there. And exchange between these two phases is determined based on the size of the bubble; and, it decreases as the size of the bubble increase. So, this is what I was talking; I have already discussed. And this is what the modeling when you do.

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So, you have a two-phase flow system. The v_1 is the flow towards the bubble and v_2 is towards the emulsion phase. This is the volume v_1 – volume in the bubble phase; and, v_2 is the volume... So, what the whole system has been just divided into two phases. So, one is the bubble phase; another is the emulsion phase. And there is a interchange of mass between the two. So, cross flow what we are talking. So, this is a cross flow. So, flow splits into two: bubble, emulsion.

And, this is something like a plug to mix flow. So, you will have something like v upon u_1 kind of dispersion number. And this is again a plug to mix flow. So, again here you have a kind of degree of non-ideality, that is, a dispersion. So, fraction of the solid we are assuming here is m_1 ; and, fraction of the solid in this section is m_2 . And then, these two are taken together; what has been discussed earlier. And then, again it is a concentration at the exit. So, something is in here and something is out here; which is defined in terms of the inlet and exit. So, one can calculate the conversion.

So, this is a simplest model – a two-phase model; and, which is actually shown here as I told that, bubbles go. As the bubbles go up and up, the size is growing. And this is the emulsion phase here. And then, there is a cross flow here. So, bubble. And then, this is the k here, which is a cross flow coefficient. And that is whatever the gas, which is in the interstitial space here; that is reacting and transforming into the product. So, this is what. So, this is for the two-phase model. And... So, due to the shortage of time, I have to just

close this; and, maybe in my last lecture, I will just discuss briefly this and some other points for the heterogeneous catalytic applications. And then, if time permits, then I will just look at some applications in refinery and petrochemicals. So, with this, I close it today here.

And, thank you for your attention.