Fluidization Engineering

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Lecture - 27
Gas Dispersion and Interchange

Welcome to massive open online course on Fluidization Engineering. So, this lecture should include the Gas Dispersion and Interchange. In previous lectures we have discussed about the gas within the different gas flow rate, how the solid a will be dispersed and also segregated with different dispersion phenomena, and also how the particle size and it is a density affect the segregation effect a on the particle as well as its mixing.

In this lecture we will discuss about that gas dispersion and also how to estimate the different estimation parameter of the gas dispersion, when it will be moving through the bed a with different type of solids and its also physical properties. And this lecture of course will be discussed about the gas dispersion only in the gas solid fluidized bed system and in future we will also discuss about the gas dispersion in the gas liquid solid three phase system also.

Now, the gas dispersion not only the main factor in the fluidized bed, but you will see during the dispersion of the interchange between the different phases that like a emulsion phase and the bubbling phase in case of bubbling fluidized bed, there will be some extent of interchange of that gas within these a phases. So, that also will be discussed about the interchange how and what extent of that interchange of the gas will be dispersed within the phases there.

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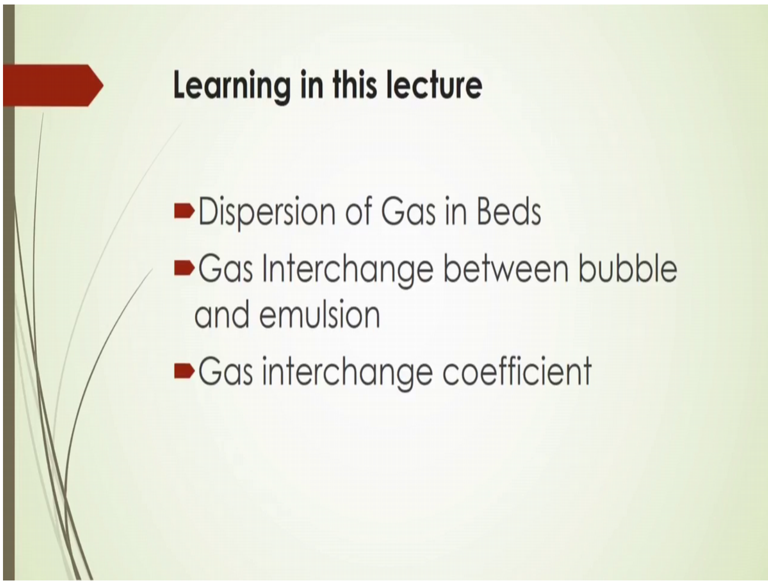

So, we will discuss about the dispersion of gas in beds how to estimate and what is the different methodology and how gas will be interchange between the bubble and emulsion and also what should be the gas interchange coefficient in the bed, when gas will be a exchanged between the bubble and emulsion phases in the fluidized bed of this gas and solid fluidization system.

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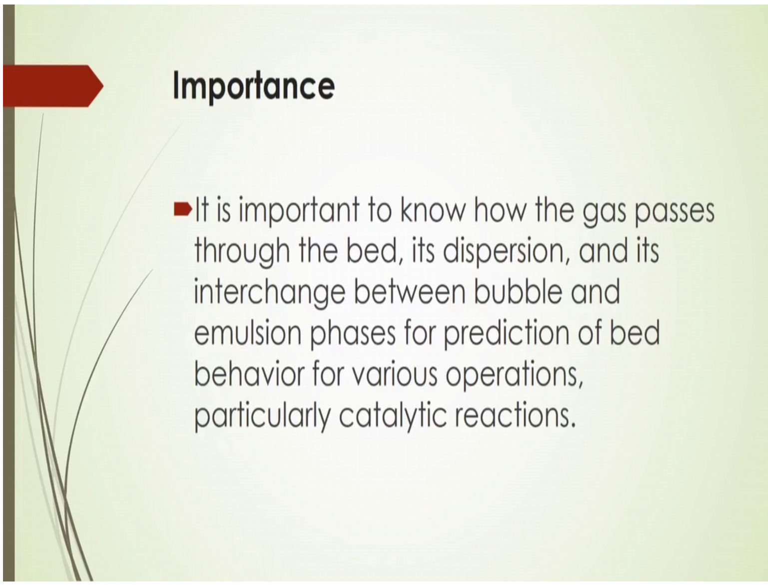

Of course, it is very interesting to know because, you have to know the extent of gas dispersion in the fluidized bed because, you know that any reactive fluidization or reactive particles whenever it will be dispersed in the fluidized bed carrying with that gaseous particles.

Then you have to know how much gas is dispersed in the fluidized bed and also how this gas will be adsorbed in the fluidized particles they are and how then particularly the reactions will be enhanced or that change during that flow of a gas by the adsorption characteristics of the particles.

And so, gas (Refer Time: 03:56) processed through the bed it is a dispersion and it is interchange between the bubble and the emulsion phases for the prediction of bed behavior for the various operations also to be known, for that analysis of that various even catalytic reactions and other selectivity processes they are.

Now, you know that whenever we are talking about the dispersion of gas in the beds; of course you have to quantify how much dispersion of the gas will be happened for a particular operating condition there.

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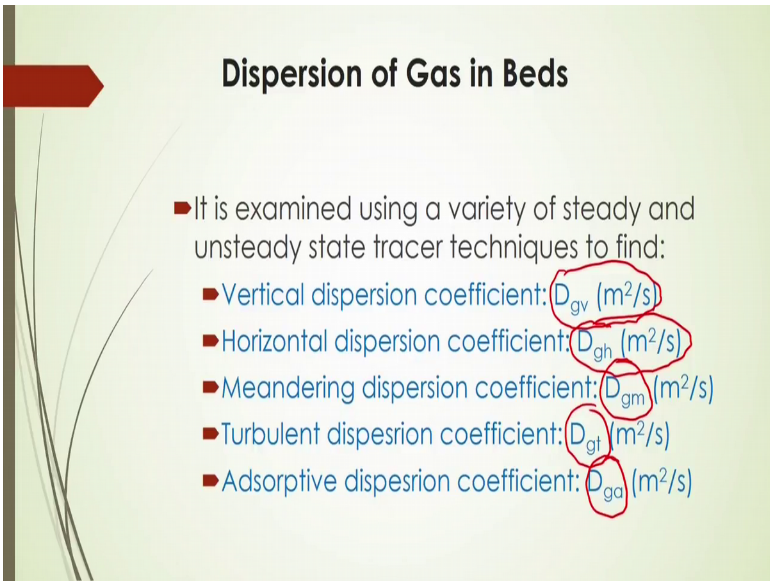

Now, how to estimate that dispersion extent or you can said you can say that intensity of the dispersion of the gas in the bed. Now, there are different off course parameters or you can say one parameter that is divided into different actually way to describe that gas dispersion there.

So, you will see whenever gas should be moving straight upward then this dispersion phenomena will be actually described or quantified by the vertical dispersion coefficient, which is denoted by this D gv the slides it is shown and when gas will be horizontally dispersing in the fluidized bed by certain energy distribution, then the dispersion coefficient will be measured and what extent of that dispersion coefficient could be described by this parameter D gh, h here horizontal and d denotes for vertical and also when there will be no other conviction inside the bed and convection effect radially or you can say whenever fluid is flowing through the fluidized bed.

Then you will see there will be a mean during dispersion of the gas will happen inside the bed; that means, you are to the excel location there will be certain extent of dispersion compare to the other part of the fluidized bed and that is why there if there is no disturbance or other action a acting on that during the flowing of that gas there.

So, without effect of other movement of that solid particles there will be a certain extent of dispersion that is called meandering a dispersion. So, those dispersion will be characterized by the factor like called meandering dispersion coefficient and this will be represented by this D gm and also whenever the fluid is operating under high velocity of that gas then the bed will be categorized as a flow pattern of that turbulent flow pattern.

Then this turbulent or you can say the first fluidized bed condition, the dispersion of the gas should be represented by the turbulent dispersion coefficient and it will be characterized by this parameter of this D gt, g means gas and t means turbulent. And another important factor during that dispersion of gas some gas may be adsorbed by the solid particles, because some particles whichever it will be used they are they may have some tendency to gas adsorption.

So, whenever adsorption will be there inside the bed during that mixing or dispersion, then dispersion of that gas may be will be affected by that adsorption characteristics.

Now, that adsorption a dispersion will be characterized by that adsorptive dispersion are coefficient which is denoted by the D ga, a for adsorption here and g for gas and this capital D actually dispersion coefficient to represent. So, will examine how this vertical, horizontal, a meandering and turbulent adsorption dispersion coefficient are happening inside the bed and how it can be quantified it will be discussed in the lecture here.

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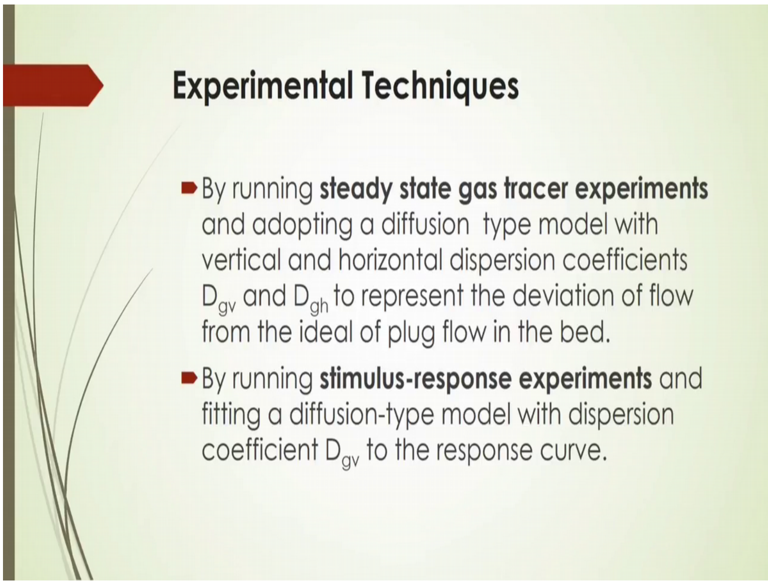

Now, what is the technical way that or you can say experimental method, how a you can actually quantify or estimate the gas dispersion inside the bed; you will see there are several methods to quantify this dispersion effect inside the bed. You know that already we have discussed in previous lectures also, that there will be some tracer techniques by which you can estimate the dispersion phenomena inside the bed.

Now, by running these steady state gas tracer experiments we are in this case what to have to do that some amount of tracer gas. So, shall not be actually taking part for the reaction in the bed and also it will not be some extent if it is not adsorbed then that will be non absorptive condition.

So, basically whenever you are ascending any gas which is not reactive with the existing gas which is to be supplied and if it is used as a tracer gas then how this tracer gas will be changing, it is concentration along this axis at a certain operating condition. Now, if it is measuring this concentration change of that tracer gas with respect to time, then it will be called as unsteady state operations.

So, at that unsteady state tracer concentration if you measure by certain measuring devices and how it will be changing with respect to time and after that if you feed this tracer concentration a profile with respect to time with the experimental data, then you will see how this dispersion coefficient can be measured or estimated. Now, you have to feed this experimental data of this concentration profile of that tracer gas with some with some fundamental models and from that model you will be able to calculate that a efficiency or you can say that intensity of the dispersion of the gas inside the bed.

Now, in this case steady state gas tracer experiments are more common are to estimate this dispersion coefficient in the bed, now for this one fundamental model that I told that you have to feed some fundamental model with the experimental data of that tracer gas concentration profile with respect to time, what is that fundamental model that is actually called a diffusion type model.

So, this diffusion type model it is given in details in reaction engineering a books in chemical engineering chemical reaction engineering books are given by Levenspiel and so that will be more helpful to start in that way, but still you will see here how to estimate that one that by diffusion type model and how it can be utilized to calculate the a dispersion coefficient.

Now, this diffusion type model with vertical and horizontal dispersion coefficients this D gv and D gh are respectively to represent the deviation of flow from the ideal of plug flow in the bed. So, this is one method another important that to that by running stimulus response experiments and fitting a diffusion type model, also will give you the extent of a dispersion of the gas vertically or horizontally from this response of that stimulus response of experiment.

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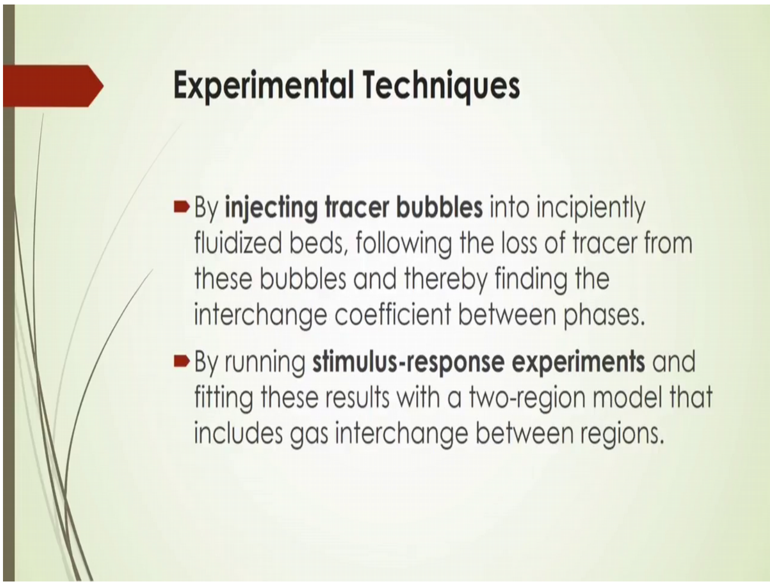

Another important by injecting tracer bubbles if you inject one tracer bubbles, there in the bubbling fluidized bed you will see at incipient fluidized condition and if there is a loss of tracer from these bubbles, and if you are estimating a that loss of tracer from the bubbles and finding that interchange coefficient of this loss of tracer gas from this bubble to that emulsion gas then you will be able to calculate what should be the extent of dispersion is happening inside the bed.

Another important that, by a injecting some other particle like tagged particle inside the bed and then tagged particle height will be changing, with respect to gaseous flow are from one location to another location, and also sometimes the radioactive tracer particles are being used to calculate the dispersion coefficient or dispersion phenomena inside the bed.

So, in that case one particle to be moving from one a position to another position and from it is mean trouble caught and then time averaged mean trouble path, from that will be actually analyzed the dispersion coefficient.

Recently another one important aspects by knowing the conductivity of the tracer gaseous inside the bed with some devices of conductivity meter, and then at different locations of the fluidized bed and with respect to time; if you are measuring that, then from the information entropy of change of that concentration of the tracer gas inside the bed you can calculate the that is maximum mixedness of the phases inside the bed, and from which you will be able to calculate the what extent of mixing or dispersion of gaseous happens inside the bed.

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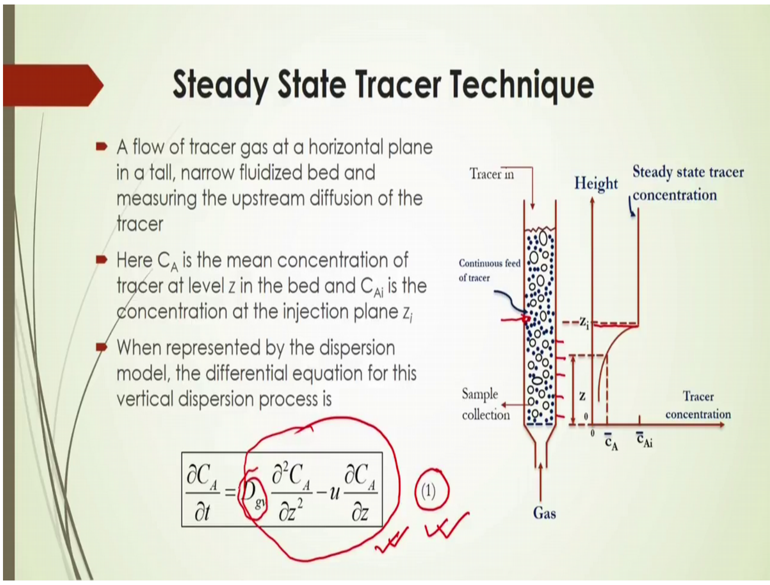

Now, let us see here one important steady state tracer technique here, in this case see the figure described here one a fluidized bed from the bottom gaseous moving upward, and from the top one tracer gaseous allowed to in and allow to mix inside the bed at that particular gas flow rate and with a certain gas particle or and also certain a fluidized particles they are.

And in this case you will see the flow of tracer gas at a horizontal plane in the tall of fluidized bed, and a measuring that upstream diffusion of that tracer, from which you will be able to calculate what should be the extent of mixing. Here if you represent that C A is the mean concentration of tracer at level a Z here inside the bed and C Ai is the concentration at the injection plane Z i here. Here this is the in Z i that is at which this tracer is the tracer concentration to be actually noted down here.

So, if C i is the concentration of the injection plane Z i from who is a you can say that when represented by the dispersion model this the differential equation for this vertical dispersion process will be represented by the equation 1. Now, this equation 1 is nothing, but that at a steady state you will see that only C A will be changing with respect to Z.

Now, experimentally also after inserting that tracer gas at this certain location of Z i and then finding out the tracer concentration from these Z at location Z then you will be able to find out that at this total length of Z, how it will be the changing what is the rate of change of that concentration there.

And also at a different location of course, you have to measure this concentration of the tracer gas and then you have to just feed this fundamental model of this dispersion that is represented by 1, then fitting this experimental data you will be able to find out what should be the dispersion coefficient D gv from this equation there.

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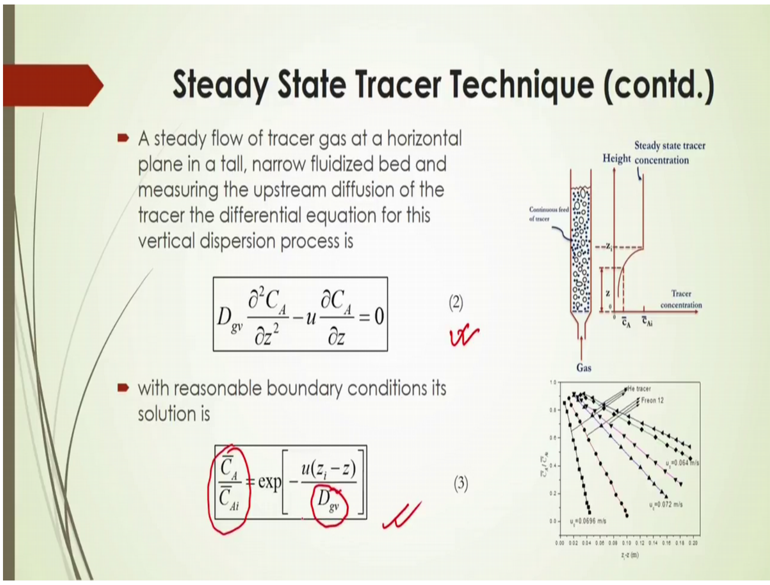

Now, see here at steady state of a tracer gas at a horizontal plane in a tall narrow fluidized bed and measuring the upstream diffusion of the tracer in the different location, and the differentiating equation for that change of concentration with respect to Z a vertical in the vertical direction, it can be represented by equation 2 at steady state condition. Now, if you do this solve of this a equation 2 with certain reasonable boundary condition, then you will get this equation 3. And this equation 3 is representing the average concentration at a different axial location by this

So, C A bar by C A i that will be equal to exponent of a minus u into Z i by minus Z by D gv. Now this will be actually you that if you if you if you draw this concentration data, this is ratio of C A bar by C A i by taking logarithm from each other is on that in on each hand of this equation, and then you will see there will be a straight line formation and from which you will be able to calculate this D gv from the slope of this equation.

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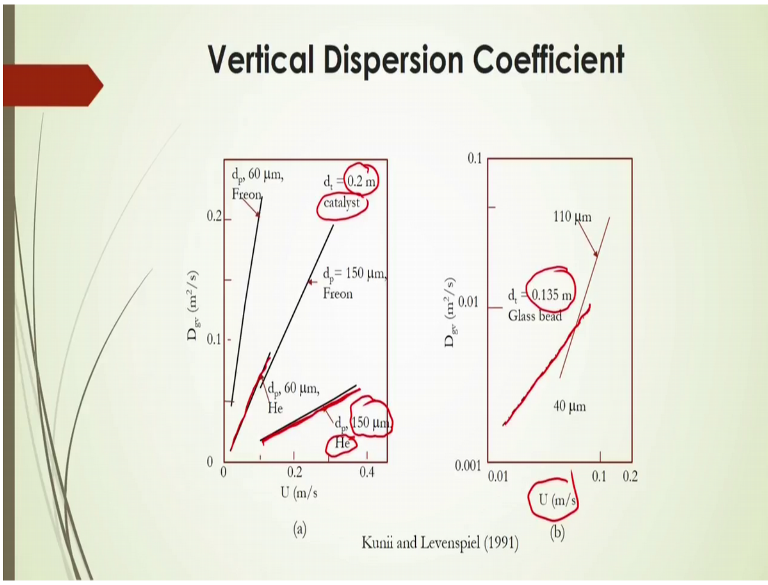

So, from some a observation you will see that the Kunii and Levenspiel he has given some results from their experimental observation for particle dispersion coefficient.

Now, this dispersion coefficient it is seen that that off course the with respect to U; that means, whenever if you increase the gas velocity there inside the bed, then a dispersion coefficient will increase. And it is seen that for different particle size this D gv also will change and that means, this particle dispersion coefficient is a function of particle diameter.

And they did their experiment with different tracer particles and also different particle type and also they have done the experiment with different type of are particles. And they have used Freon gas as a tracer gas or helium gas as a tracer gas. So, whenever they are using that helium gas and particle diameter they are taking 150 micrometer they got this profile here.

So, from this profile also it is seem that that dispersion coefficient the vertical direction is increases with increasing gas velocity. But whenever, this the same helium concentration is used, but the particle size is particle size is reduced to 60 micrometer they got that a higher dispersion coefficient compared to this particle diameter one of 150 micrometer.

So, in this case what we observe that, that if you we decrease the particle size then a of course this a particle dispersion coefficient will increase, why? You will see that for larger particles they are the tendency of that internal circulation of that particles will be lower a relative to the smaller particles there. Because that higher size particles.

So, they have more turbulent they have more terminal velocity and because of which they may go downward immediately without just moving up and then a moving randomly or radially or making no internal circulation a cell inside the bed and because of that this higher particle size will give you the lower range of dispersion coefficient. Similar pattern also they got when they a have used the Freon gas as a tracer gas.

So, in that case they got that the higher dispersion coefficient compared to the helium gas when they use for the same particle size. In this case why then (Refer Time: 21:29) dispersion coefficient will increase in this case very interesting that maybe that Freon gas will be adsorbed by the solid particles by this catalyst particles they are what they have used in their experiment. So, whenever this tracer gas will be adsorbed by the catalyst particles they are interaction of that solid particles with other particles a may not be actually that must intense compared to that whenever it will not be adsorbed in the a solid particle.

So, the dispersion coefficient will increase compared to that non absorbed particles there. And also it is seen that for different type of particles they got different a phenomena of that dispersion coefficient there. For the lower range of this gas velocity they got this type of profile of dispersion coefficient a within a certain range, but after a certain range of; that means, here if gas velocity increased a more than 0.1 meter per second, then it is seen that drastically increase with respect to particle diameter.

Where whether it may be it may be higher or lower, but it is seen that whenever it crosses 0.1 meter per second gas velocity even with the higher particle diameter, the dispersion coefficient will increase. May be a due to the higher gas flow rate there will be a more kinetic energy supplied in the bed and also particle particle interaction and there may be some degree of segregation inside the bed because of which the dispersion may enhance.

And also it is important to note that the tube diameter or you can say the bed diameter fluidized bed diameter also has an intense effect on the fluidized bed gas dispersion there. What that if we increase the tube diameter if you increase the tube diameter, suppose if I use this 0.2 meter instead of 0.135 meter, it is seen that for the same tracer gas and for the same type of particles the dispersion will increase, why?

Because for the larger diameter bed what happened, there will be more space that radial distribution of the gaseous particle as well as the solid particles would be more relative to that what is that smaller diameter tube. They are more internal circulation, that is the inter circulation cell of the solid particles will be more compared to the narrow tube they are.

So, there may be residence time we also more because of who is that the dispersion coefficient of the gaseous particles inside the bed will increase.

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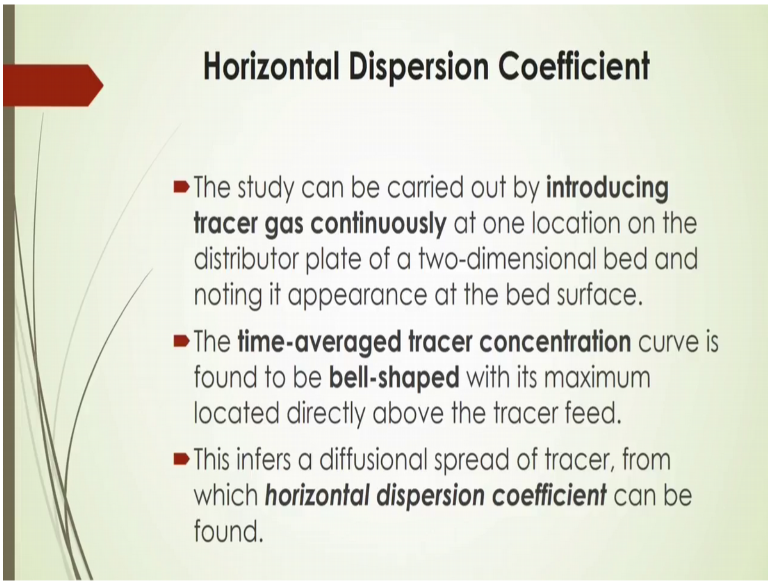

Another important that horizontal dispersion coefficient; what is that horizontal dispersion coefficient? That means, you will see that near the distribution of the gas, the gas immediately will not go to that upper position of the fluidized bed, very near to that distributor high kinetic high kinetic energy distribution will enhance the distribution not only the vertical direction, there may be tendency of that solid particles or gaseous particles to move radially or laterally to the bed.

And also if the bed diameter is high in that case, from that bottom part of the fluidized bed from the distributor due to that higher kinetic energy ejection of the solid particles. May some may change the may change the axial velocity profile of the gas to that radial position.

So, radially there will be a spread of the solid particles or gaseous particles by which that horizontal dispersion may result from that high kinetic energy distribution from the distributor at the bottom side. Whereas, at the top side you will see that there may be a reduction of that lateral dispersion there.

Because in that case the very fine particles those who are coming up due to the increment and the segregation phenomena, what will happen those particles may not become downward and may not be spreading what is that radially as mass as intensity of whatever it is in the lower part. Because in that case the kinetic energy whatever or the streamline of that particles coming out from that bubble position that bubble that is a head, that may not is it that mass of solids from that outside, but it will be horizontally moving up because of that a finer particles there.

So, if there the mixture of particles of larger and smaller particles both will be there, and then there may be tendency of that coarser particles those will be coming out downward at that upper position, they may change the velocity pattern of the gas flow rate at the top position. And because of fluids the dispersion of the gas ratio will change they are 10 they are actually what is that trends.

And important another factor is that the horizontal dispersion coefficient whenever we are going to measure or estimate, you have to introduce the tracer gas continuously in the bed at one at one location on the distributor plate of a two dimensional bed and noting it appearance at the bed surfaces. And in that case time averaged a tracer concentration curve to be found so, that that there will be a maximum what is that tracer feed that would be located directly above the tracer feed and the time averaged the tracer concentration will be looking like a bell shaped trend.

And this actually interpreted the diffusional spread of the tracer and from which you can say what should be the horizontal dispersion coefficient there.

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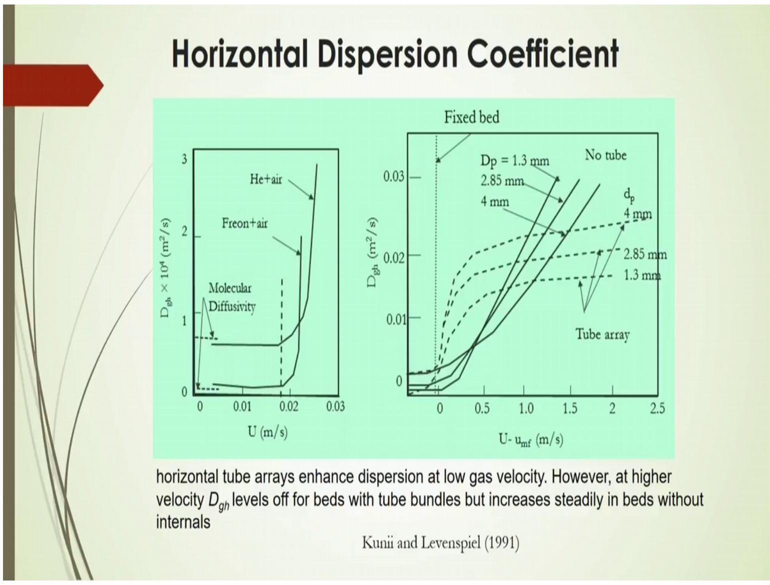

And some results that is given by Kunii and Levenspiel that for that the horizontal dispersion coefficient and for that horizontal dispersion coefficient they have used some horizontal tube arrays for enhancing this dispersion coefficient and they measure quantity of the dispersion coefficient. And they have seen that with increasing gas velocity, initially up to 0.0 or 2 meter per second, the horizontal dispersion coefficient will not change significantly.

Whereas, above this 0.02 meter per second the dispersion coefficient it changed drastically for different air and the tracer gas mixers there. And it is seen that for Freon tracer the molecular diffusivity will be little bit lower than the convective diffusion there.

Whereas the helium tracer the molecular diffusion will be more higher than the convective one. So, in this case maybe they are the physical properties of the gas tracer gas and also is there any adsorption characteristics inside the bed or not that depends on. Another important point that, that if you increase the gas velocity for a fixed bed if there is no fluidization condition, there will be no significant change of horizontal dispersion coefficient.

Whereas, you will see beyond this 0.25 meter per second that effective gas velocity inside the bed, there will be a change of dispersion horizontal dispersion coefficient. And this horizontal dispersion coefficient actually changes as for the particle diameter. And they have done some experiment with the particle diameter a 1.3 millimeter 2.85

millimeter and 4 millimeter. And it is seen that higher a sized particles will give you more horizontal dispersion coefficient because that higher size particles will immediately go downward relative to that upward particles there and they are maybe that is why the spreading of that particles inside the a bed at the adjacent location of the gas distributor there.

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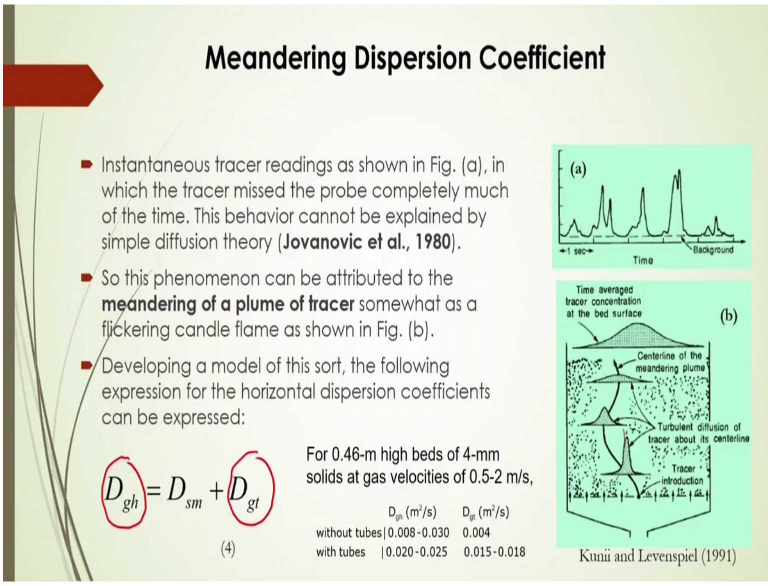

And then meandering a dispersion coefficient what is that? This you will see that if you insert some instantaneous tracer and if you observe the reading of that tracer concentration by sophisticated measuring devices like conductivity meter or other devices, in that case you will see that tracer missed the probe completely mass of the time and this behavior cannot be explained by the simple diffusion theory.

As soon in figure a, how see how this concentration profile with respect to time will change. So, you cannot express this behavior of this concentration change happens early it is what is happening by a standard that is simple diffusion a theory.

So, for this some phenomena to be actually to be introduced so, that that how this type of phenomena can be explained. So, in this case Jovanovic et al 1980 they have observed that as per figure b that they stated that this phenomena can be attributed to the meandering of a plumb of the tracer.

Somewhat as a it is called flickering candle flame that as shown in figure b here. This is the flickering candle flame they are like this. So, in this way particles are moving upward there. So, this pattern of the flow that may change the gas dispersion inside the bed; So, developing a model of this a short, you can you can express the horizontal dispersion coefficient by this equation 4 here. So, the D gh; that means, horizontal dispersion coefficient that will be equal to summation of this meandering dispersion coefficient and the turbulent dispersion coefficient there.

So, this two this two factors that is one is meandering dispersion coefficient of solids, and then turbulent dispersion coefficient of what is that gas. So, both will give you that total a; that means, horizontal dispersion coefficient and Kunii and Levenspiel they have given some results on that for 0.646 meter high fluidized beds of 4 millimeter solids, that the gas velocity of 0.5 to 2 meters per second, they have observed this horizontal dispersion coefficient without inserting any tubes, they got this the range of dispersion coefficient of about 0.0 0.820.03.

Whereas for this turbulent dispersion coefficient will be 0.004 meter square per second. And they have done also the experiment with the a insertion of tubes they are and they got this horizontal dispersion coefficient a within a range of 0.020 0.025 meter square per second whereas, for that turbulent dispersion coefficient it is within the range of 0.0152 0.018 there.

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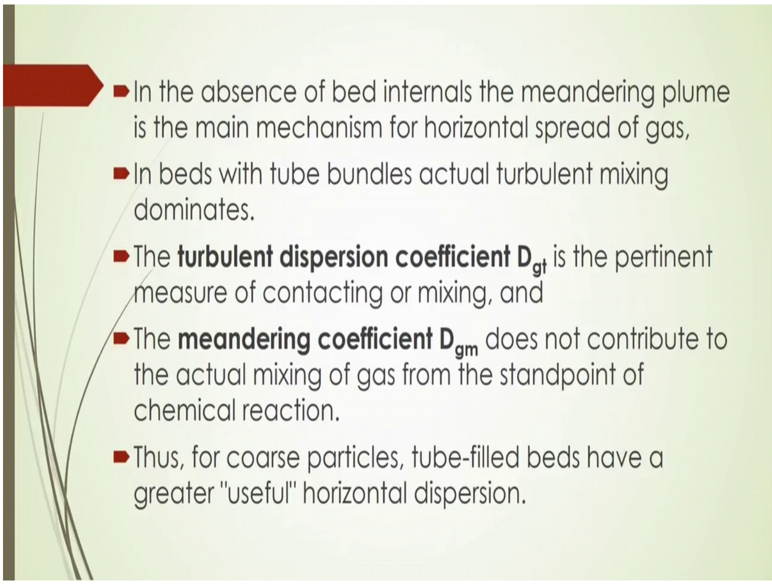

And in the absence of bed internals of course the meandering plumb is the main mechanism for horizontal spread of gas. Now, you can reduce that horizontal spread of gas if you are inserting some tube vertically or you can use the baffle or you can use some other provision. So, that you can decrease a the horizontal dispersion coefficient.

Already we have discussed something about that shroud making up shroud by use you can decrease the what is that attrition mechanism and by the same way, you are you can also reduce that horizontal dispersion coefficient there in the bed. And the bed with two bundles actual turbulent mixing they are to be dominant because there will be some interaction of the solid solid particles in presence of that, due tube bundles of will enhance. The turbulent dispersion coefficient D gt is the pertinent measure of the contacting or mixing inside the bed.

Whereas this meandering coefficient that will not contribute mass to the actual mixing of the gas from the standpoint of the chemical reaction. Does of course, for the higher diameter particles tube filled a beds have a greater I think useful horizontal dispersion they are in the bed.

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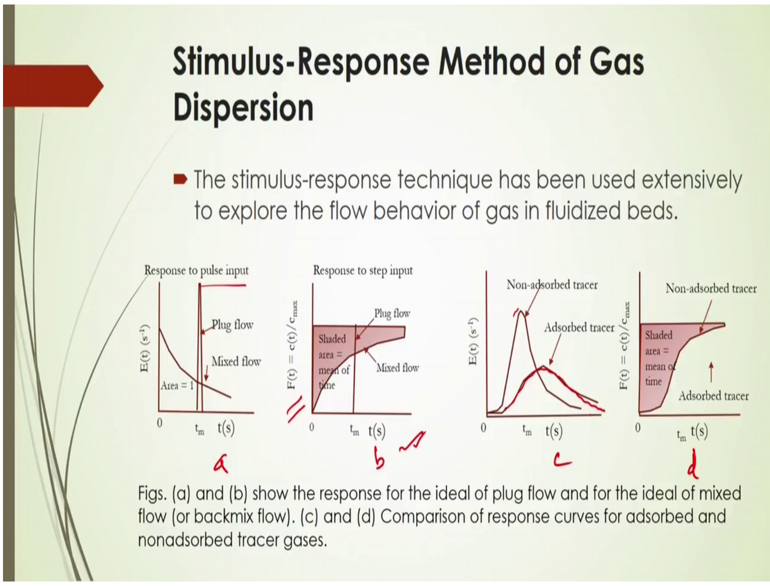

And there are different methods of estimation of gas dispersion that what you told. So, we have discussed one method and then another method is called stimulus response method of gas dispersion. So, this stimulus response technique has been used a extensively to explore the flow behavior of the gas in the fluidized bed.

Here in this figure you see that sometimes the tracer is being introduced inside the bed as a pulse input and as a step input. And as a pulse input within a short period of time a as mass as possible it is short, then you have to insert the tracer gas in the bed and immediately you have to you have to collect the sample or by online measurement you can measure that how this concentration of the tracer gas will change with respect to time.

Once you know that concentration of that tracer gas with respect to time, you may expect this type of profile here shown in this figure here. And also for a step input what happen here, you have to input the you have to insert the tracer gas a within a certain interval of type time continuously in the bed, and you will see this type of tracer ah; that means, here distribution a function of the tracer inside the bed will observe with respect to time.

Here this line this vertical line is shown whether it is plug flow or mix flow and with respect to mean residence time there. And you will see if there is no adsorption of tracer gas, then you will see this type of concentration profile you can get whereas, for adsorbed tracer you will see this type of profile.

So, from this a distribution of this concentration, you will be able to find out what will be the exactly the dispersion coefficient for adsorbed tracer or non adsorbed tracer in the bed. So, for pulse input response these are the this type of response you may expect. So, it is very easy to actually calculate that dispersion phenomena inside the bed by this pulse input method.

Only thing you have to calculate the tracer concession with respect to time after immediately inserting that tracer gas in the bed. Now, figure this here this is a, this is b, c and d if we represent then figures a and b show the response for the ideal of plug flow and for the ideal of mixed flow, and c and d that will give you the comparison of the response curves for the adsorbed and non adsorbed a tracer gases in the bed.

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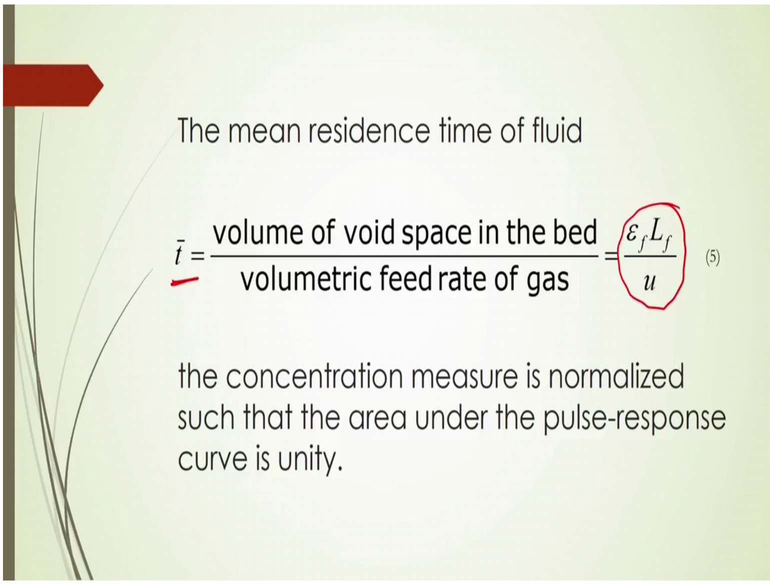

Now, one important parameter for actually identifying the behavior of the solid particles in the bed or fluid particles in the bed so, that how long it will be residing inside the bed, that is called mean residence time. So, this mean residence time is defined as that volume of void space, what is in the bed and what will be the volumetric a feed rate of the gas.

So, by which you can a calculate what will be the mean residence time of the fluid; that means, how long this gas will be residing inside the bed. So, this t bar this is denoted for this mean residence time, this will be is equal to epsilon f L f by u.

What is the epsilon f? Epsilon f is nothing, but the void fraction of the fluid and L f is called the length of the fluidized bed they are and u is called the velocity of the fluid at which the fluidization operation a is being done. And the concentration measure is the normalized a way to represent such that the area under the pulse response curve should be unity as for this figure here given. So, this the this here the area under this curve should be is equal to 1 here also in this case the area under this curve should be is equal to 1 it is shown in this figure.

So, the concentration measure that will be a of course, normalized to represent the profile or distribution of this concentration of the tracer with respect to time. So, that after measurement so, anywhere you will plot it, you will get some profile like this and then a what would be the area under that curve you can measure from that graph. And there may be the measurement from the graph you can do otherwise you can do it by theoretically ah; that means, by seems of one third rule or some other techniques or integration techniques by which you can get that what is that area under the curve.

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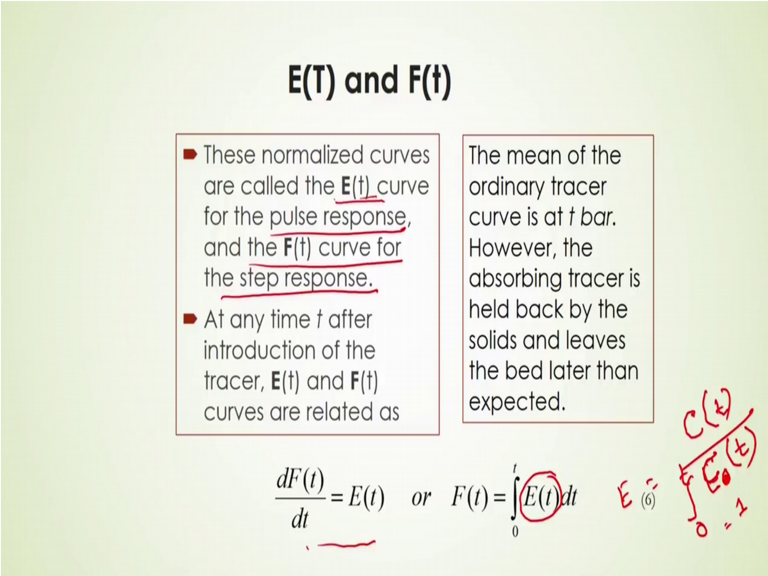

So, once you know this area under the curve then you will be able to calculate what should be the mean residence time they are in the bed. Now, what is that relationship between this E t and F t? E t is the concentration function here normalized concentration function. So, these normalized curves for all curves that is called E t curve for the pulse response and F t curve for the step response here. At any time after introduction of the tracer, this E t and F t curves are will be interrelated a by this equation here.

This d F t by d t that will be is equal to E t or F t you can say that integration within this 0 to 1 what should be that value this E t is the are normalized concentration graph what is the normalized concentration graph? This so, if you are measuring the concentration a with respect to time at a certain time, and then what will be the initial concentration of the tracer before inserting to that fluidized bed, then what the initial concentration of the tracer it is represented by C 0, then C e C t by C 0 t that will be is equal to normalized function.

But generally this C 0 t is nothing, but that here 0 to t here what will be the curve a. So, this will be represented by E t here and this will be is equal to 1. So, in this way this E will be represented.

So, this description is given in the chemical reaction engineering by Levenspiel, I think you will get more information about this E curve and F F t curve and also there is another a textbook that is called Fogler you can get more information about this E t and F t. Once you know this E t curve and then observing that train and from which if you fit with experimental data with that standard diffusion model you will be able to calculate what will be the dispersion coefficient.

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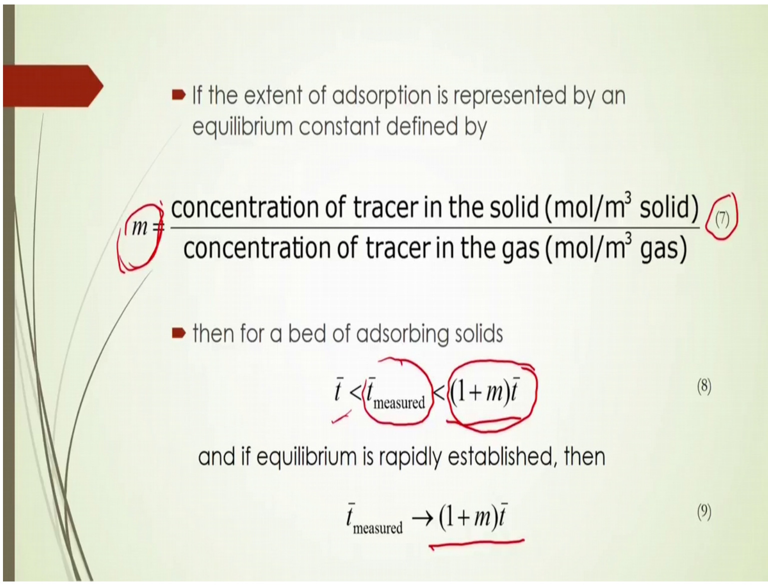

Now, if the extent of adsorption is represented by an equilibrium constant defined by if you if you use some tracer gas that is being observed by the a particles what are being used in the fluidized bed, then you have to consider those the a adsorption phenomena also.

So, in that case at equilibrium condition now, what should be that equilibrium parameter for that adsorb gas there inside the bed so, that will be represented by that m. M should be defined as the concentration of tracer in the solid divided by concentration of tracer in the gas they are by this equation 7. Then for a bed of adsorbing solids then t bar; that means, here mean residence time will be less than is equal to t measured there and what will be off course less than is equal to 1 plus m into t bar what does it mean here?.

This whenever adsorbed absorption will be there you will see there will be more time for dispersion of the gas inside the bed. And also if equilibrium is rapidly established then this 2 ah; that means, measured mean residence time will be is equal to; that means, adsorb tracer a mean residence time there.

So, in this case why it is happening because they rapidly whenever it will; that means, it will not take more time to adsorb the that gas and whenever dispersion will take more time. So, this adsorption a time contact time will be relatively very a small compared to that total mean residence time there or; however, is a residence time they are inside the bed.

So, that is why if the adsorption it will take within a very short residence time or you can say that very rapidly if the equilibrium of this adsorption happened then the measured mean residence time will be exactly with that that mean residence time for the adsorption condition.

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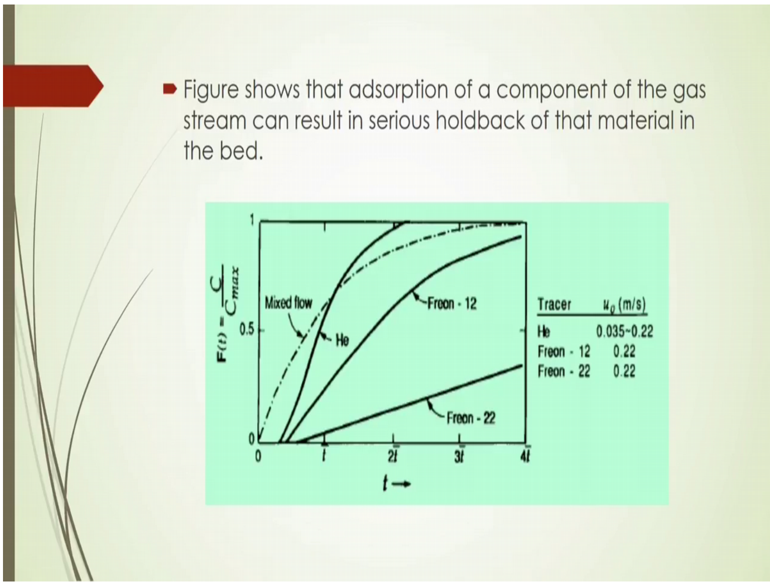

This figure shows that the adsorption of a component a of the gas stream that can results in the series of holdback of that material in the bed.

So, sometimes whenever you will see that some tracer particles will be used to represent that concentration profile by F t and then if is there any adsorption happens, then sometimes you will see that gas stream sometimes it will be changed without adsorption particles also.

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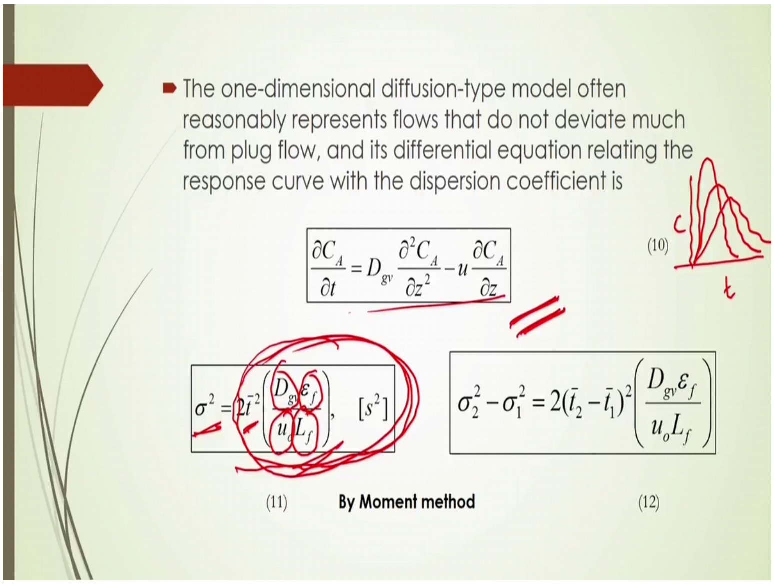

Now, the one dimensional diffusion type model very often reasonably represents the flow characteristics of the fluid inside the any inside any unit of fluid flow.

So, in that case you will see that diffusion model if you are considering one dimensional, there may be other dimension also, but if you are fluidized bed if you are fluidized bed is may be a narrow or you can say the diameter will not be more than 5 centimeter, then you can easily use that one dimensional diffusion model there.

Because the radial distribution will not be that much effective compared to that axial dispersion coefficient there. So, in that case the one dimensional diffusion type model reasonably a gives good results to represent the dispersion phenomena inside the bed, and also this one dimensional model is being a widely used actually for a calculating or estimating the dispersion coefficient of the fluid a medium inside the bed.

So, what is that one dimensional dispersion model? This is the final form of this equation, this equation basically derived from this material balance. So, the material balance is not given details in this lecture, but these final equations you can get in any standard reaction engineering book, there it is I am just giving you 2 books that Levenspiel and Fogler you just follow then you will get this final form of the equation and this is called one dimensional diffusion a equation.

And you will see this equation can be represented by this profile here like this t c versus t and this will be coming like this way, this type of profile you can expect from this model equation 10.

Now, from this profile you will see whether this profile will be widely spread or the spread is not that much wide. So, from the spreadness of this concentration versus time profile, you will be able to measure what should be the dispersion coefficient there.

Now, this spreadness; that means, called a sigma that is called variance, variance this sigma square of this model can be represented by this equation 11. Actually this variance is coming by this moment method and there are maybe first moment, second moment and third moment are there. And basically first two moments are very important to represent this equation to solve for that particular parameter.

So, in that case first moment is called mean residence time and the second moment is called the variance. So, from this second moment and the mean residence time, you will be able to express this residence time in terms of this here this dispersion coefficient and the porosity of the other a variables there.

So, from this sigma square, this sigma square a value you will be able to calculate D g v. Once you know this epsilon f, u 0 and L f there. What is u 0? U 0 as the gas velocity at which the fluidized bed is being operated and t bar is the mean residence time that can be obtained from the first moment of this equation and sigma is square this sigma square can be calculated from this momentum method once you know that concentration, you will be able to calculate what will be the variance.

So, from this variance what is the dispersion coefficient it can be easily calculated.

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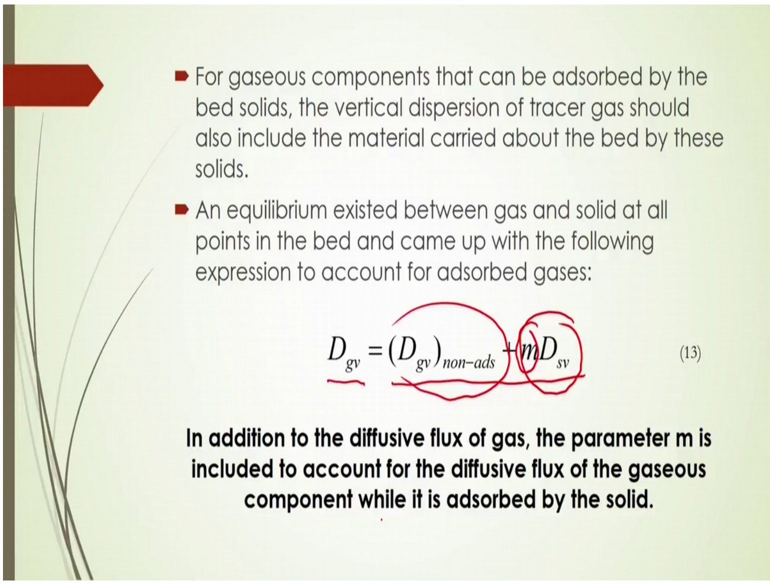

Now, for gaseous components that can be adsorbed by the bed solids; the vertical dispersion of the tracer gas should also include the material that will be carried about the bed by this solids. Now, an equilibrium existed between a gas and solid at all positions in the bed and that came up with the following expression to account for adsorbed gaseous here.

So, this D gv can be expressed the expressed in terms of the a dispersion coefficient of non adsorbed condition and what is that, dispersion coefficient of that vertical dispersed solids and if you multiply it by that adsorption equilibrium, then you will get what should be the vertical dispersion coefficient of gas in the bed.

In this case you have to remember in addition to that diffusive flux of gas the parameter m is included to account for the diffusive flux of the gaseous component while it is adsorbed by the solid.

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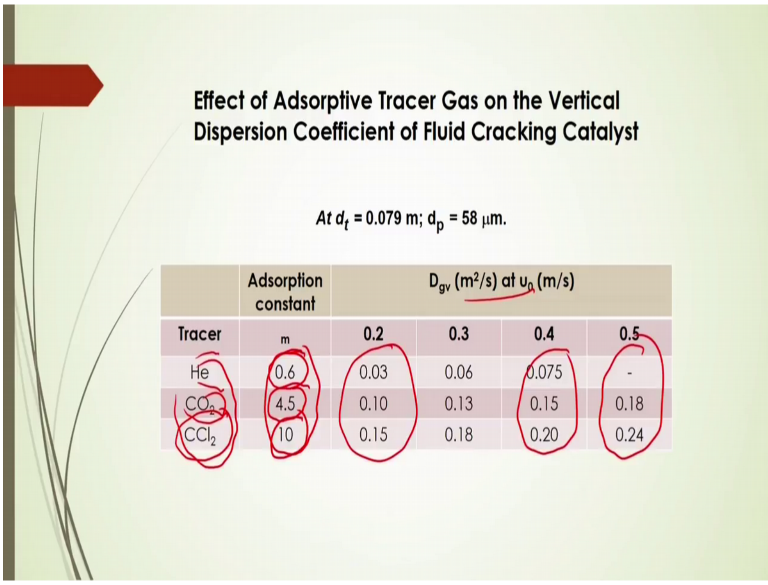

Now, effect of adsorb tracer gas on the vertical position and denoted by the dispersion coefficient of fluid catalytic cracking. So, in that case some observation is given by Kunii and Levenspiel there, if you are using that different tracer concentration then m value will be like this. So, for helium this m value is generally 0.6 for carbon dioxide tracer gas it will be 4.5 and carbon a dichloride it will be is equal to 10.

So, in that way for different a m value, you will see that D gv how it will be changing that is given in this table here. So, this D gv will change accordingly with respect to the equilibrium comes equilibrium of the adsorb of equilibrium parameter of that adsorption of the gas.

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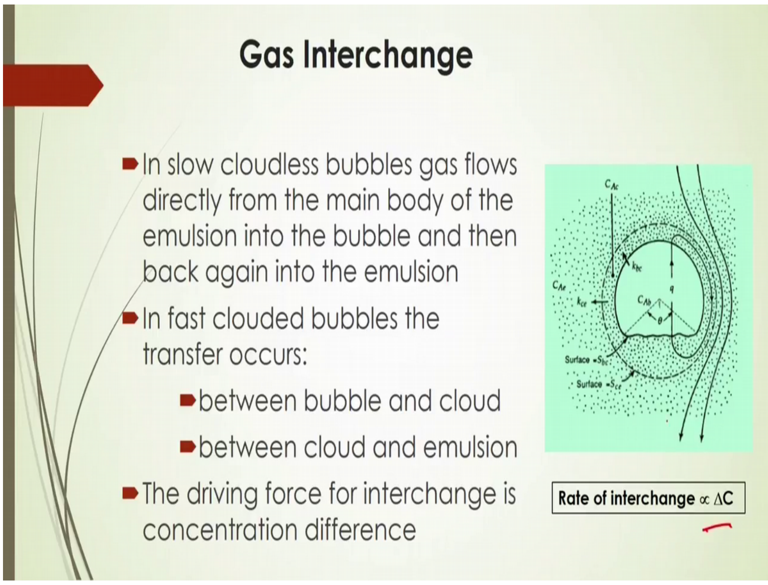

Now, another important factor is called the gas interchange. It is seen that for slow cloudless bubbles gas flows directly from the main body of the emulsion into the bubble, and then back again into the emulsion. So, in first clouded bubble by in the bubbling fluidized bed, the tracer occurs between bubble and cloud and also the exchanger of this tracer gas will be within the cloud and emulsion.

So, as for this figure surrounding this bubbles there will be the cloud and emulsion that has already been discussed in the bubbling fluidized bed phenomena. So, we know that there will be cloud region the cloud radius and also that what is that bubble radius and then emulsion, surface cloud surface all these things we have already discussed earlier. So, this driving force for the interchange between this bubble and cloud and cloud and emulsion that will be called as concentration difference that is delta C.

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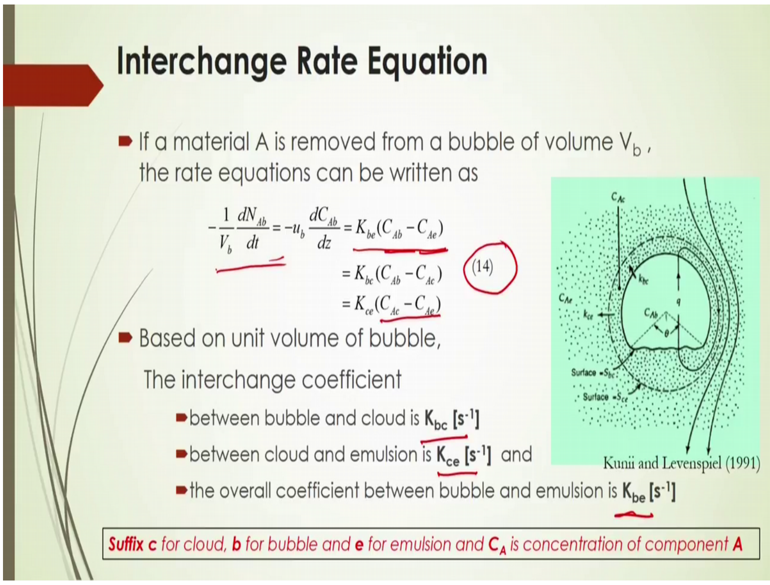

Now, if a material A is removed from a bubble of volume V b, the rate equation can be expressed by this equation 14 here. So, in this case you will see that the number that is number of moles of tracer gas that is changes in the bubbling in the bubble with respect to time, that will be depending on the bubble velocity as well as the concentration of the bubble concentration of the tracer gas in the bubble. And then it will be actually proportional to the driving force of that concentration difference here C Ab minus C Ae; that means, between a bubble and emulsion there.

So, what will be the proportional constant which is called K be. Similarly, whenever you are representing this tracer gas transfer from this bubble to cloud then it will be K bc into C Ab minus C Ac. And if it is from cloud to emulsion region then it will be K ce into C Ac minus C Ae.

So, different are interchange coefficients of this b K be K bc and K ce are represented to express this exchange rate of the tracer gas from one phase to that another phase; based on unit volume of bubble the interchange coefficient between bubble and cloud will be is equal to K bc and cloud and emulsion it will be K ce and overall coefficient between bubble and emulsion will be is equal to K be and all those coefficients will be unit as second inverse.

Now, this c suffix c b and e will be represented as cloud bubble and emulsion respectively and C A is the concentration of the component A which is being used for that transfer from one phase to the another phase.

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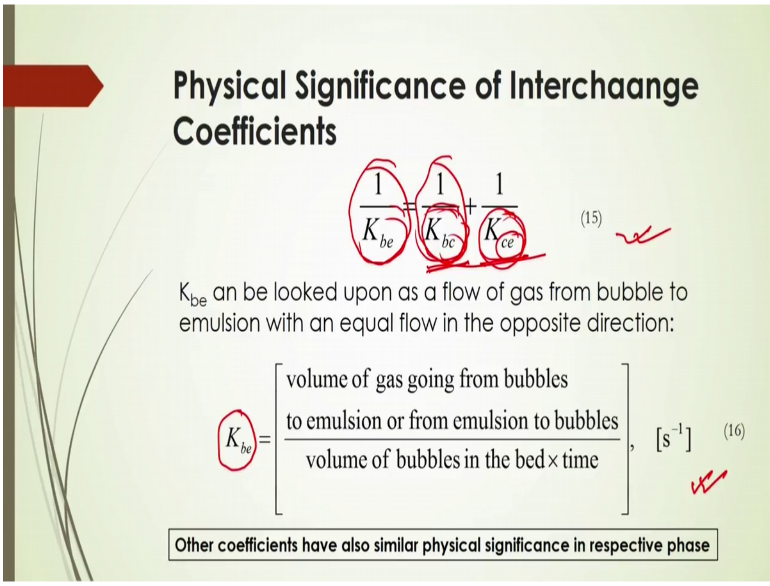

And you will see that the overall the transfer coefficient that is capital K be, you can obtain from this individual a transfer coefficient of a K ce and K b c. So, in this case this 1 by K be will be is equal to 1 by K bc plus 1 by K ce, once you know this K ce and K be then it is a easy to calculate that overall transfer coefficient from directly from bubble to emulsion.

So, whenever transfer is coming from bubble to emulsion that two actually zone should be crossed 1 is bc zone and another is ce zone. So, that is why overall transfer coefficient from this bubble to emulsion will be calculated as per equation 15. Now, this K be can be looked upon as the flow of gas from bubble to emulsion with an equal flow in the opposite direction also. So, this can be defined physically by this volume of gas going from bubbles to emulsion or from emulsion to bubbles by volume of bubbles in the bed per unit time.

So, in this case it is unit is coming a second inverse other coefficients also have to be similar physical significance with respect to the phase.

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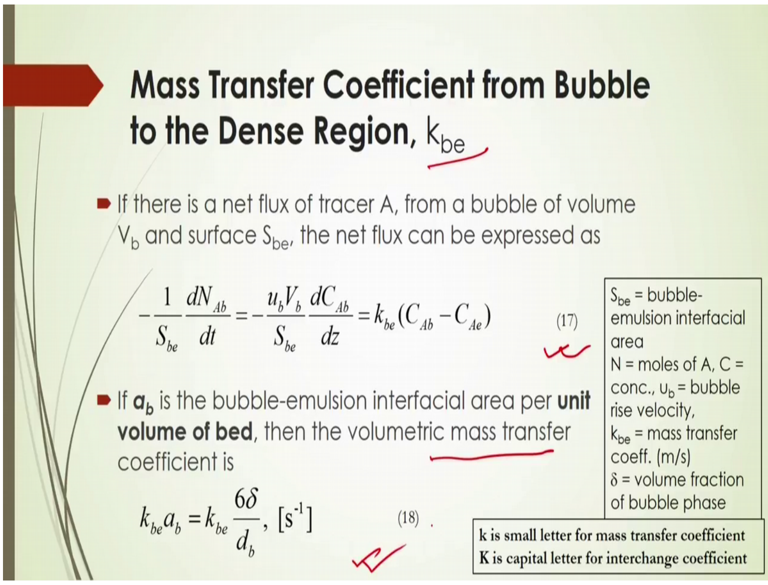

Now, mass transfer coefficient from bubble to the dense region this will be represented by small k be, there it is the transfer coefficient here mass transfer coefficient, so one will be discussed later on also in details.

So, in this case here you will see that if there is a net flux of tracer A, from a bubble of volume V b and surface S be then the net flux can be expressed as by equation 17 here. If a if a b is the bubble emulsion interpolation area per unit volume of bed, then volumetric mass transfer coefficients can be represented by this equation 18.

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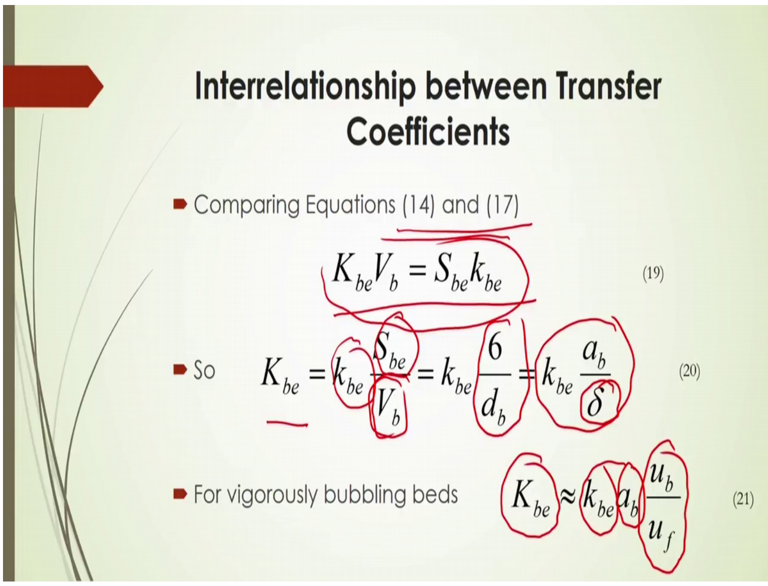

And another interrelationship between that transfer coefficient of this mass transfer and the interchange coefficient. Comparing this equation 14 and 17 you can get this a equation 19 for this relationship between this K be, overall transfer coefficient between bubble and emulsion and the mass transfer coefficient bubble to emulsion there once you know that interfacial area between bubble and emulsion.

So, this K b e will be is equal to here a small k b e into S be by V b. Now, this S be by V b that will be equal to 6 by d b, this is very easy to actually derive what is the surface area under to the volume, just you substitute you will get this 6 by d b and then finally you can get it what is the delta here? Delta is nothing, but that whatever the volume fraction of the bubble in the bed.

And for vigorously bubbling beds use this interchange coefficient between bubble and emulsion will be is equal to this small K be into a b into u v by u f. This u b is nothing, but the bubble rise velocity whereas; u f is called the fluid velocity.

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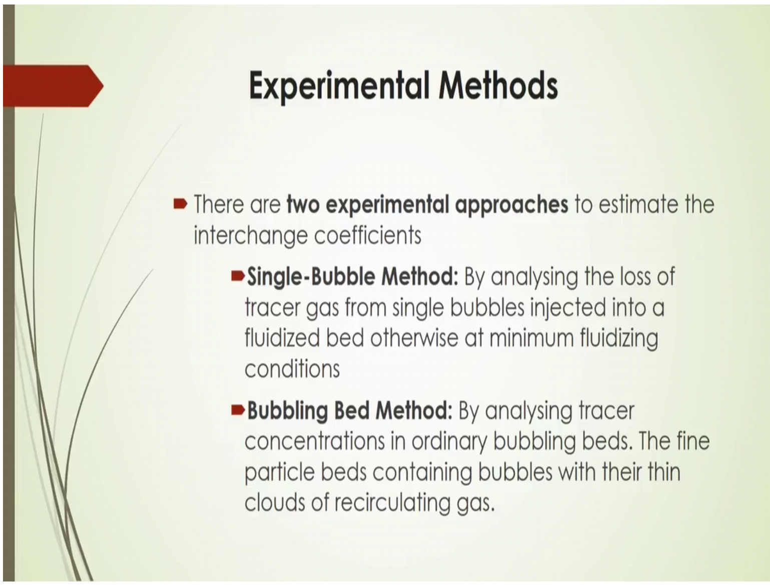

Now, also you can get this interchanges coefficient by experimental method also, there may be single bubble method and bubbling bed methods are there, from which you will be able to calculate that interchange coefficient.

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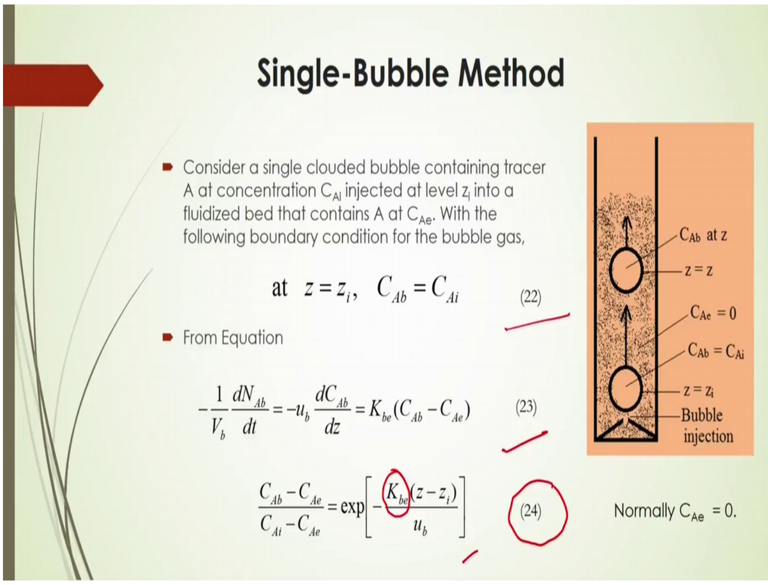

Now, as for single bubble method a, if you are considering that a single clouded bubble and that will contain the tracer A at concentration C Ai which is injected at the level Z, here I into a fluidized bed that contains a at concentration C Ae. Then with the following boundary conditions a given by equation 22, you can obtain this equation 23 and after solution of this 22 this boundary condition you can represent this by this 24.

So, from this equation 24 once you know that concentration at this different phases, then you will be able to calculate what will be the coefficient of this K be; that means, interchange changes coefficient for bubble to emulsion.

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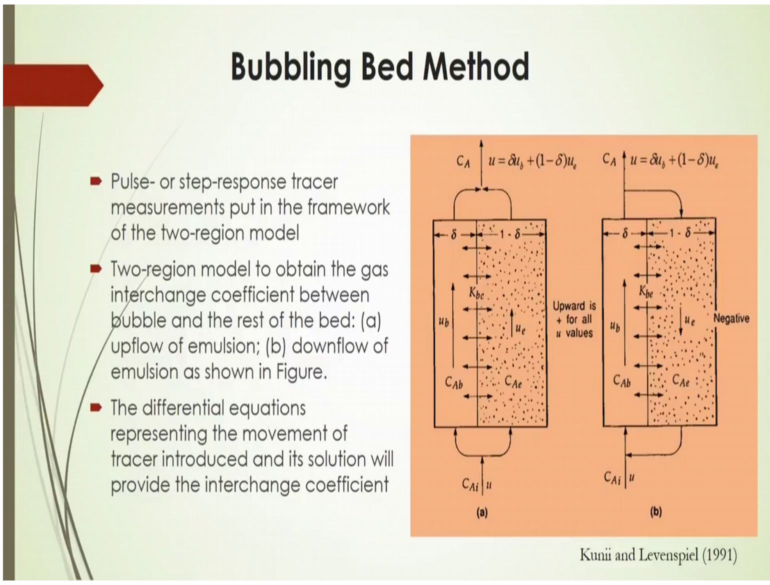

Similarly, bubbling the method also can be employed to estimate that bubble size interchanges coefficient. So, in this case you can use the pulse or step response tracer technique for this two region model, and you will get that after mass balance the differential equation from which you will be able to calculate the dispersion coefficient.

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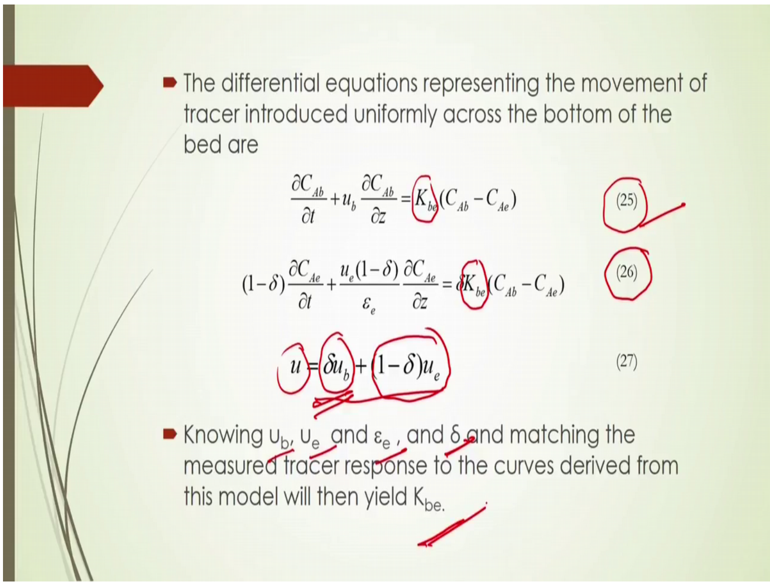

And that exchange coefficient and the differential equation are representing the movement of the tracer introduced uniformly across the bottom of the bed. If you are just considering that two regions of that this is bubbling and that emulsion region there.

So, based on these you will see that how concentration will be changing, with respect to time and also a across the axis of that bed, then can be represented by this equation 25 and 26. Whereas, this u should be the effective velocity, which will be the what will be the bubble rise velocity and what will be the emulsion velocity.

So, the effective velocity will be is equal to that is what will be the fraction of the bubbles inside, here see in this figure this portion. If you are representing this a fluidized bed here, a total fluidized bed and then if we a divide this fluidized bed into a two parts one is the bubbling here there is no particle and here if it is that emulsion region that is gas and particles are mixed here.

So, two regions and this regions is called delta and this regions will be called 1 minus delta. So, in this case this delta is called bubble fraction 1 minus delta is called emulsion fraction. So, based on these a bubble affection and emulsion volume fraction they are, that you will get the effective velocity by which this solid particles will be transferred and also the flow pattern of the gas since the bed will change.

And because, of this two region change of this concentration of that tracer particles you can a represent this change of concentration a based on this two region a by exchange of that gaseous particles from one region to another region and representing by equation 25 and 26 by mass balance. Then a from after solving these two equations you will be able to calculate for should be the interchange coefficient a represented by capital K be.

So, knowing this here bubble rise velocity, emulsion velocity and the porosity for emulsion and bubble volume fraction inside the bed and massing these two equations to the tracer response as per experiment and then fitting with that model you will get this interchange coefficient of bubble to emulsion there.

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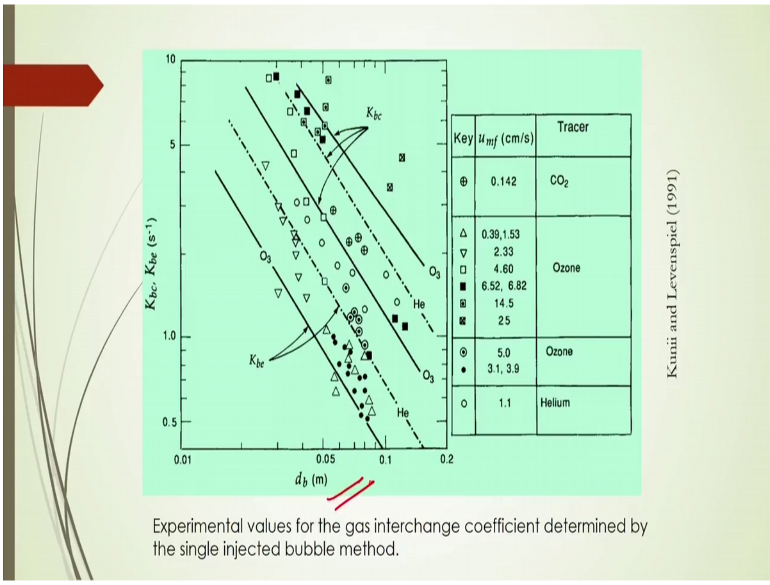

So, this is the one trained you can have that this K be and K bc will be changing with respect to diameter of this bubble.

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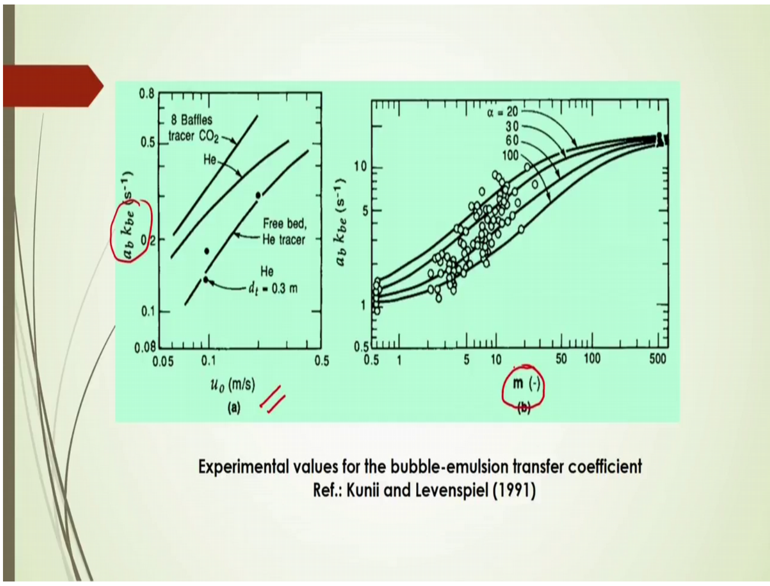

And also here see that this overall mass transfer coefficient that is small k be also will change with respect to what is that gas velocity and also the equilibrium constant of the gas adsorption.

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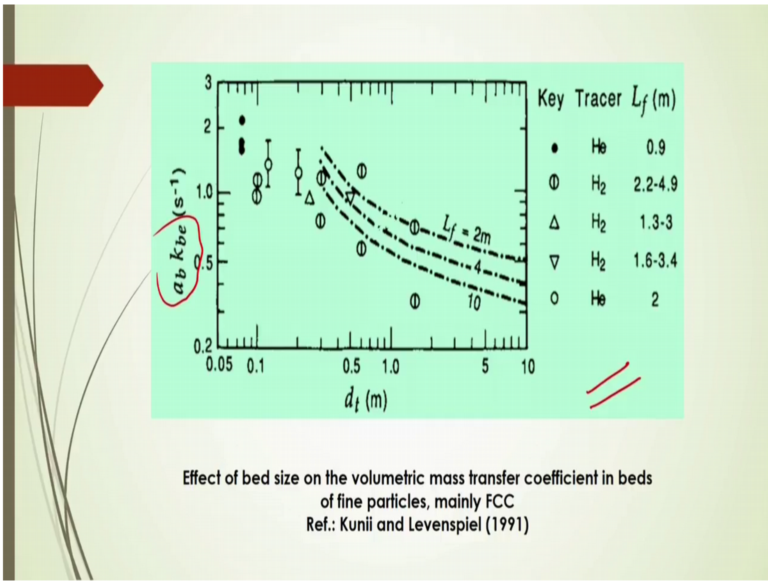

In this case you will see also that mass transfer coefficient will change with respect to a tube diameter.

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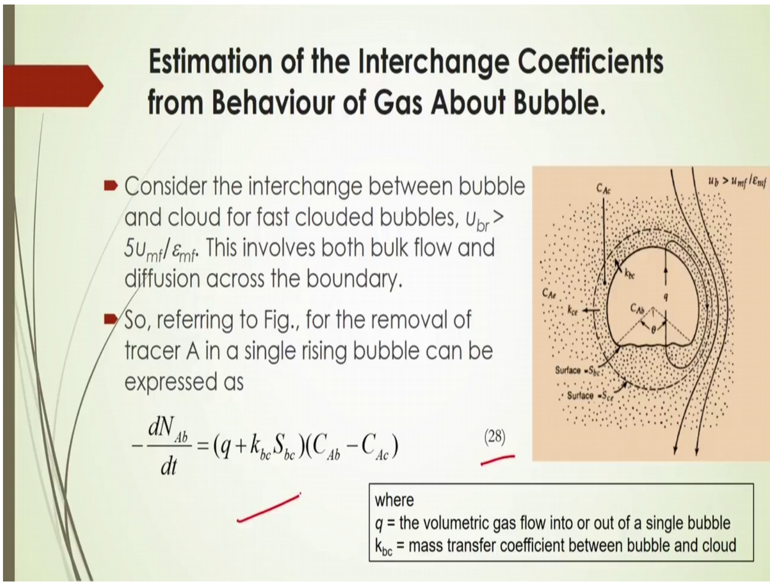

And for this FCC catalyst, Kunii and Levenspiel, they have given this data for how this overall mass transport coefficient will be changing with respect to tube diameter. And estimation of the interception from the behavior of the gas about the bubble also you can calculate from this equation 28 here given in this equation.

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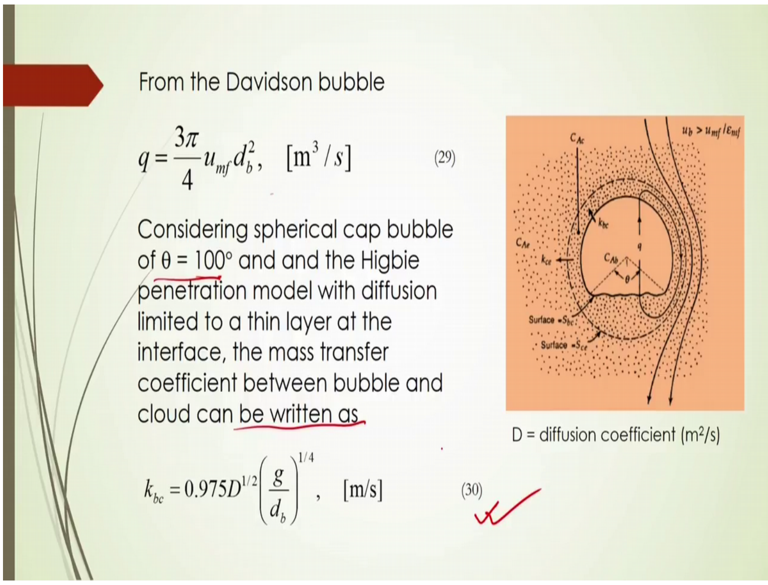

And you will see that from the Davidson model that q, I think we have already discussed about this how these particles or gas will be flowing over this bubble, and considering that bubble will have that cap of theta 100 degree and the Higbie penetration model. If we used there the mass transport coefficient between bubble and cloud can be written by this equation 30.

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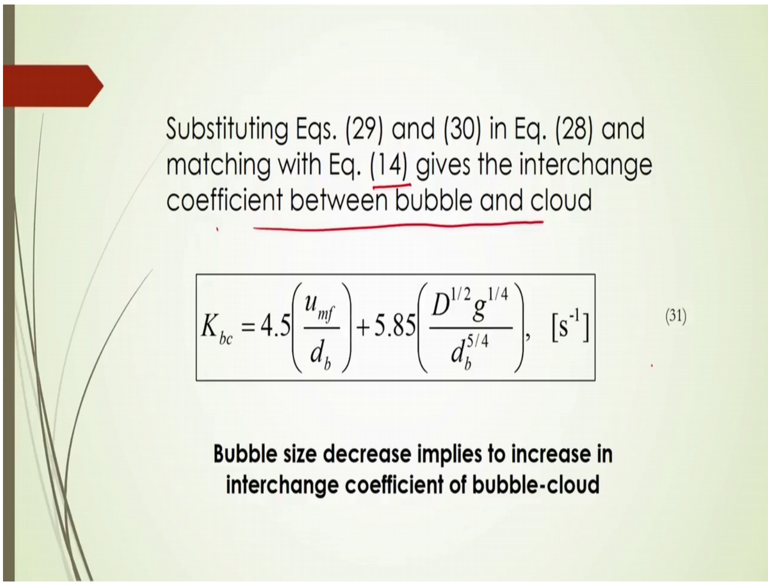

And then substitution of this equation 29 into equation 30 and then with the help of this equation 29 and 30, in equation 28 and matching with equation 14, then you can calculate what should be the intercept coefficient between bubble and cloud by this equation 31.

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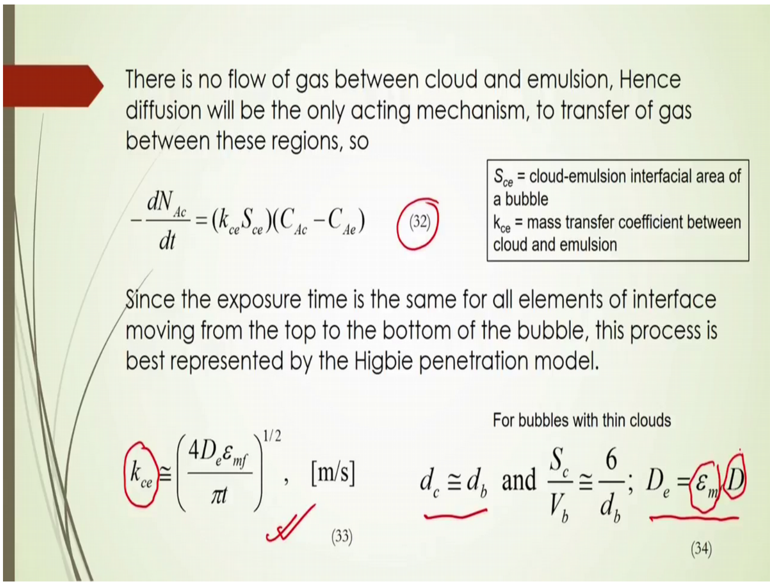

Again if you consider that the flow of gas between cloud and emulsion, hence the diffusion will be the only acting mechanism that to tracer of gas between a this region so that this equation 32 can be expressed for that. And in that case k ce; that means, mass transfer coefficient between a cloud and emulsion will be represented by this.

What is that there by considering that elements of inter phase moving from the top to bottom of the bubble there. And for the bubble within clouds the you can consider is the d c will be equal to d b and from which equivalent diameter a will be is equal to epsilon m f into d here.

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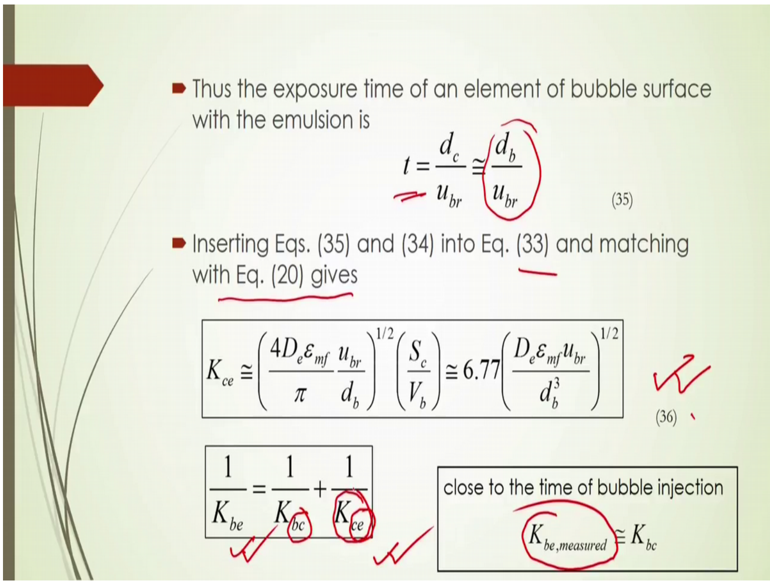

Does the exposure time of an element of bubble surface will be represented by t that will be is equal to d b by u br, once you know this d b by u ah; that means, exposure time and substitution of this equation 35 and 34 into 33 and matching with equation 20 you can calculate what should be the actually interchange coefficient or transfer coefficient between cloud and emulsion that can be represented by equation 36.

And once you know this capital K ce and capital K bc; that means, bubble and cloud and cloud to emulsion by this equation 36 and other what is that here by 31, then after substitution of 31 and 36 here in this equation, then you will get the overall transfer coefficient between bubble and emulsion.

And it is seen that that close to the time of bubble injection this overall coefficient or bubble to emulsion will be is equal to K bc there, there we no effect of cloud emulsion coefficient or interchange inside the bed.

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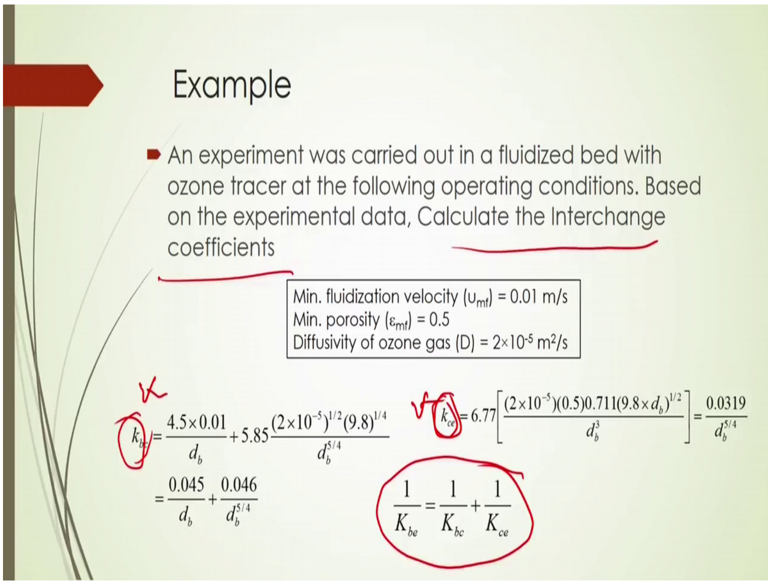

Now, here it is given one example to represent or to calculate this mass transfer coefficient or interchange coefficient there. So, once you know this k bc and k ce as per this example here, see that example it is said that an experiment was carried out in a fluidized bed with ozone tracer at the falling operating condition, based on the experimental data, calculate the interchange coefficient here.

So, this interchange coefficient you can calculate case k bc and k ce this would be correction that is capital K bc and it will be capital K ce, once you know this then what will be the overall bubble a emulsion interchange coefficient.

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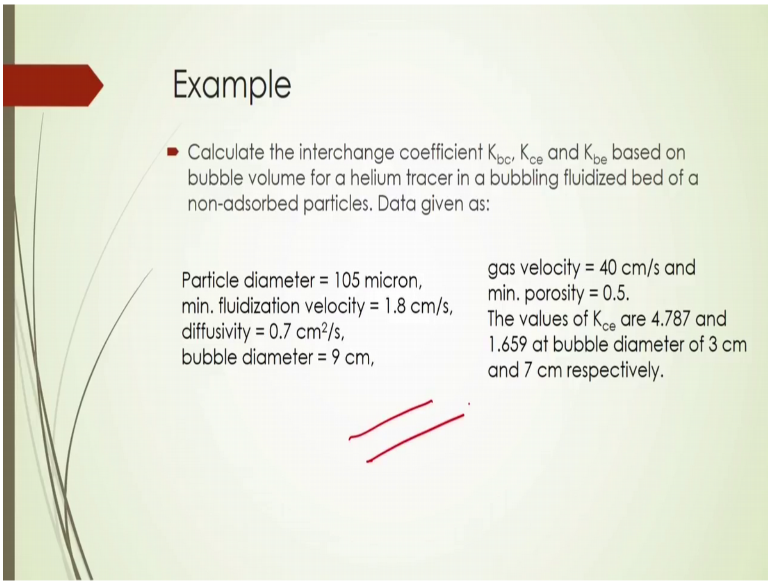

Similarly, another examples you have to calculate what will be the K be, a once you know this K bc and K ce with this a operating conditions.

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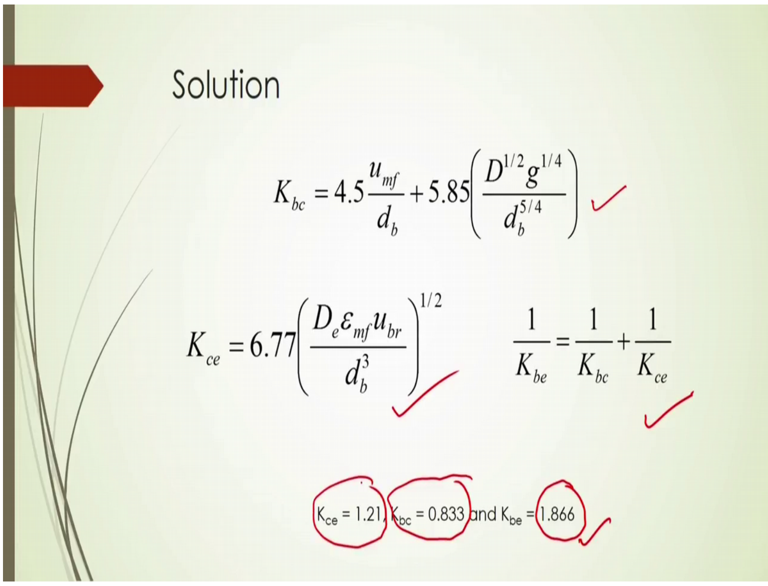

And thus simple you have to calculate K bc and K ce and substituting here and then you will get that K be value is equal to a 1.866 and K bc is equal to 0.833 and K ce is equal to 1.0. So, what will be the interchange coefficient t that you can get.

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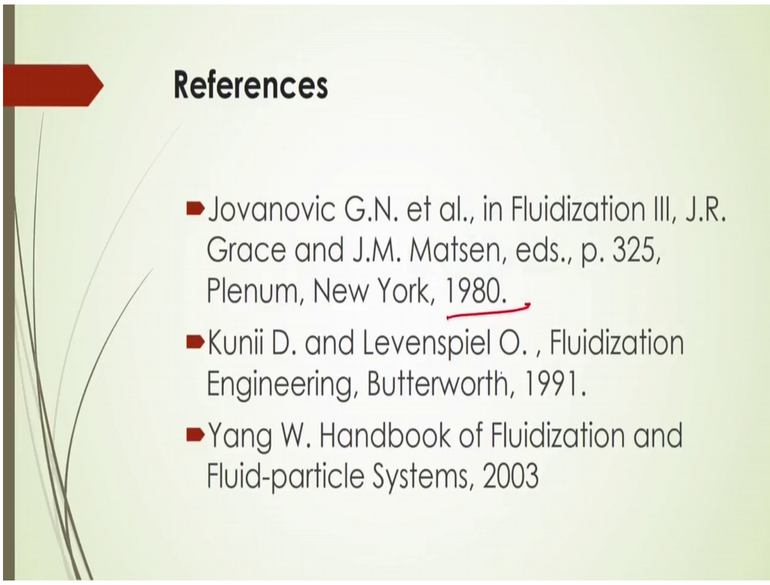

So, more information about this interchange coefficient and the gas mixing inside the bed, how to be actually depending on the particle diameter and also horizontal and vertical dispersion coefficient of the gas that affected by that interchange coefficient and also solute redistribution segregation mixing effect, that can be and how to estimate the dispersion coefficient by the tracer techniques.

I think we have certain extent learned about those phenomena of that gas dispersion and mixing inside the bed. More information about this mixing you can get it from this Jovanovich this book and also here Kunii and Levenspiel and young text book from these references they are. So, next lecture will be on mass transfer in gas solid fluidization so.

Thank you.