Mass Transfer Operations-I
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Lecture - 20
Liquid dispersed: Venture scrubber, wetted wall column, Packed tower

Welcome to the 4th lecture of module 3 of Mass Transfer Operation. In module 3, we are discussing equipment for gas liquid operations. Before going to the next lecture let us have small recap on our previous lecture.

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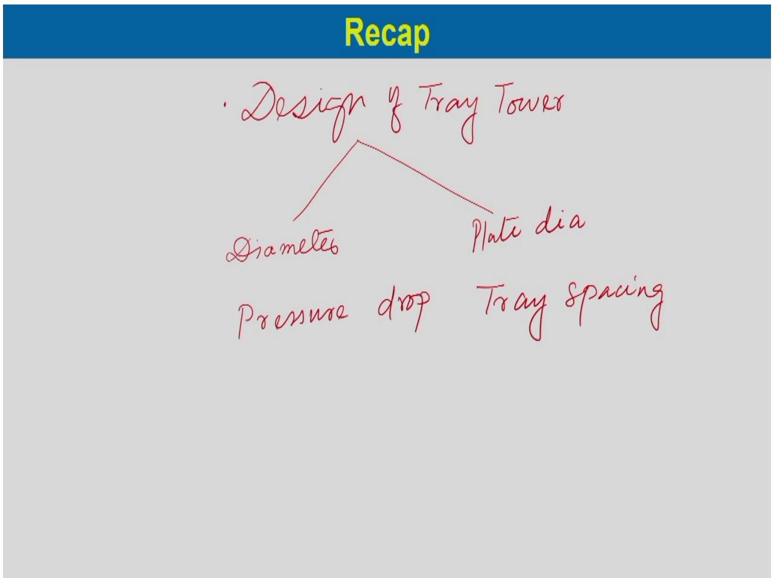

In our previous lecture, we consider mostly design of tray towers or plate tower. So, different aspects of tray towers we have considered mostly a tower diameter, then the plate geometry, plate and we considered the pressure drop in the column and mostly the tray spacing and finally, we have tried to solve a design problems, although it is a very preliminary stage of design, but the digital design will be discussed in your design course chemical engineering design course. So, in this lecture we will discuss the design of or the different aspects of in lecture 4, we will consider design of packed towers.

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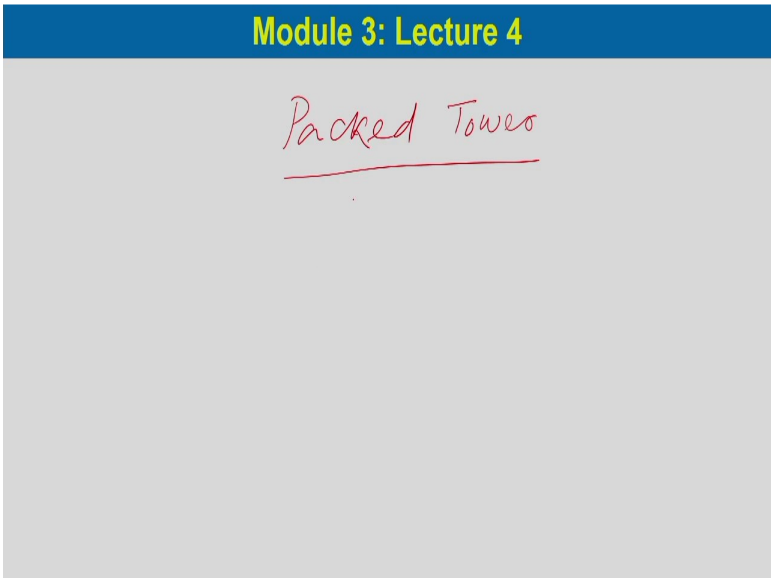

So, different aspects of packed towers we will consider in this lecture.

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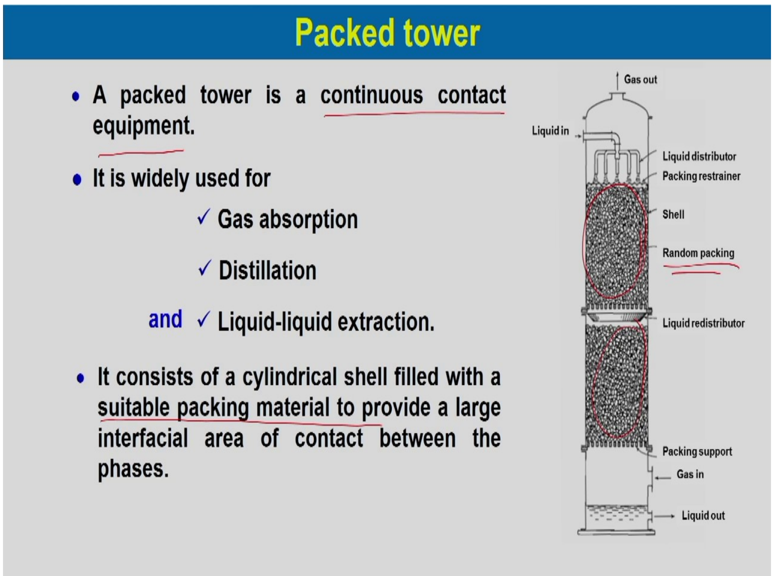

This is no picture in which you can see there are packing. So, this is a continuous contact equipment in a packed towers. It is widely used for gas absorption, distillation and also liquid extraction. As you can see it consists of a cylindrical shell which is filled with a suitable packing material, so which you can see the random packing which is given over here random packing material and the purpose of the packing is to provide no large interfacial area for the contact between the gas and liquid.

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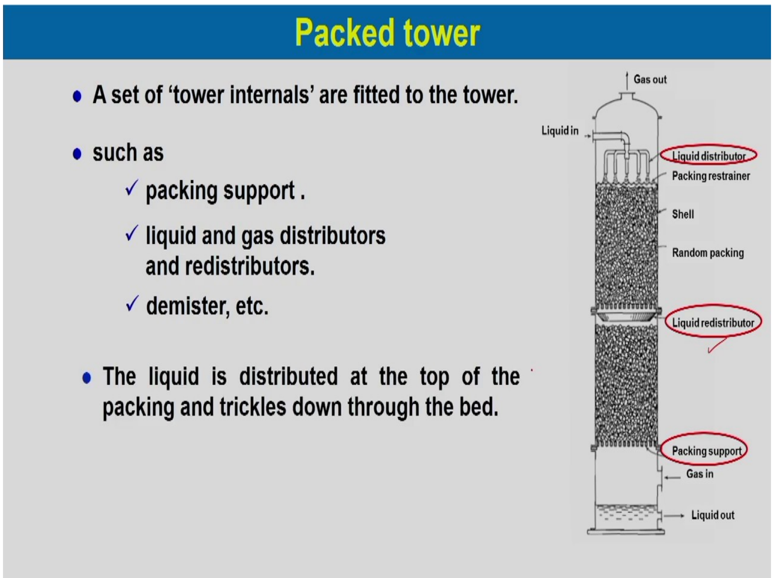

So, a set of tower internals are fitted to the tower you can see the tower internals, one is the packing support over here packing support and then we have liquid distributors. In many situations the liquid distributors are located at different places not only at the top; top is any where it is required, but if you have a tall tower there in that case the single distribution may not help proper flow of the liquid and gas. So, we need to have intermittent liquid distributor inside the column. So, you can see you over here liquid distributor and it is basically redistributors, one is distributor another is called liquid redistributors. Then demister sometimes it is used basically to break the mist which will form inside the column.

The liquid is distributed at the top of the packing and trickles down through the bed, which will fall through the interspace distance or white space between the packing materials and which is trickle down through the bed.

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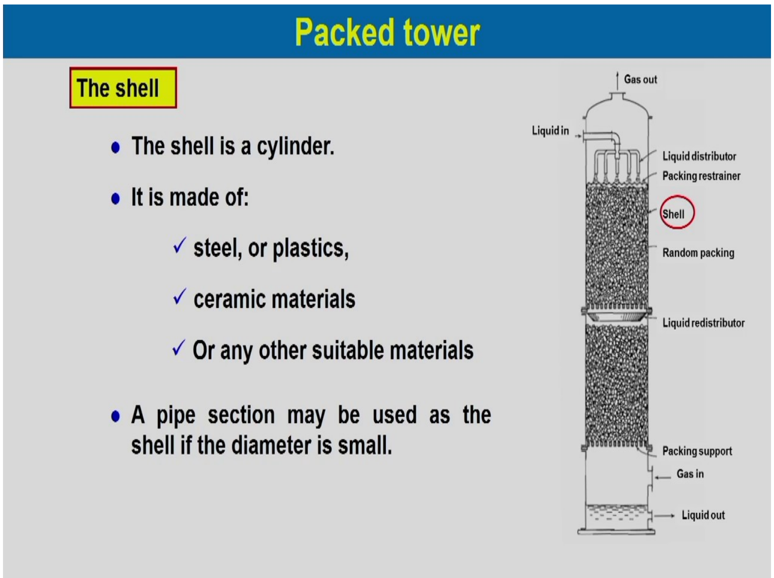

There is a shell in the column this called a no cylinder like a cylindrical shape shell. It is generally made of no steel or plastic and it may be ceramic materials or any other suitable materials depending on the application. A pipe section may be used as a shell if the diameter is small. So, if your diameter of the column required for a particular application is not much, so we can use a pipe as a shell of the packed column.

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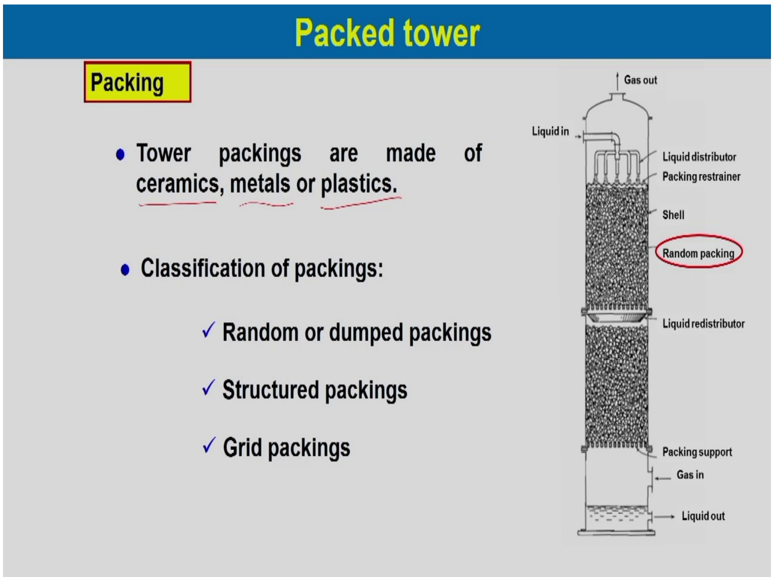

Now, tower packing are made of different materials, one of them is ceramic materials, then metals we can use metals or plastics. So, it also again depends on the material you would like to handle in your tower. There are different designs of know packing, so and they are classified as the random or dumped packing and other one is structured packing.

So, random packing you just take the different materials of different no geometry either ball or in different shapes and load the packing inside the column, so you can just dump the packing material randomly. And structured packing the packing material are stacked in such a way that it helps to proper distribution of the gas and liquid and which increase the also the interfacial area between the gas and liquid, so structured packing. And then also there is grid packing, so you can make a different grid and then you put the packing material grid packing.

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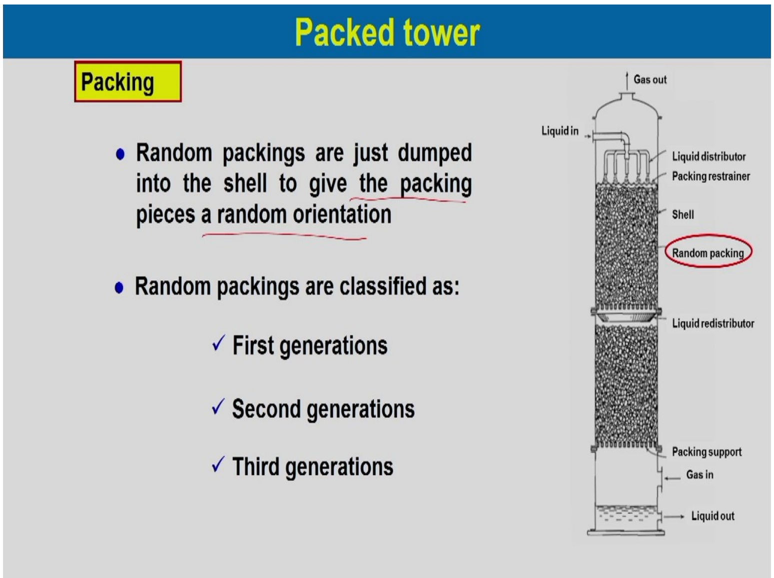

So, the packing random packing are just dumped into the shell to give the packing pieces a random orientation. So, you do not have control on the orientations of the packing material, which is random packing. Random packing are also classified as first generation, second generation and third generation packing.

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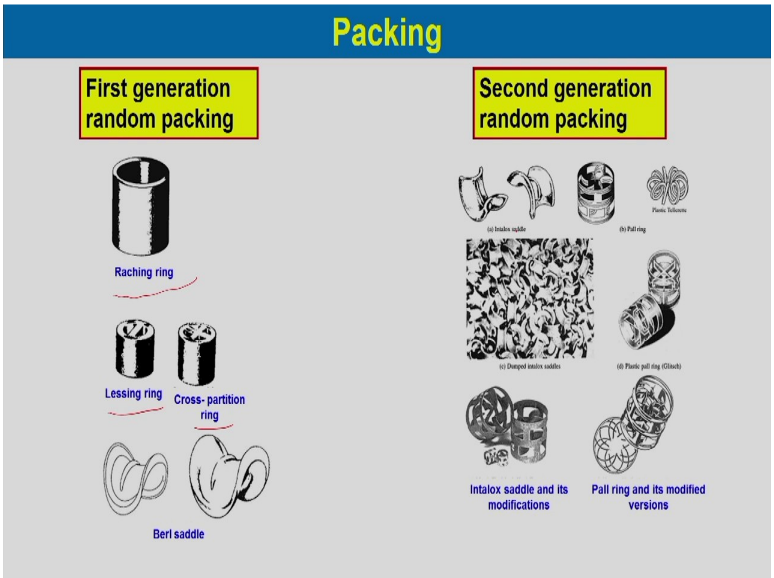

So, the first generation packing is the Raching ring, you can see Raching ring, then Lessing ring and Cross-partition ring, Berl saddle. So, different type of design of the random packing material you could see which times to times design changes depending on the applications you are going to have and then the cost as well (Refer Time: 07:24).

So, the cost parameter also looked into while making this no packing. So, these are called first generation random packing. Then second generation random packing, it has a different shapes one is you can see the interlock saddle which is looks like this and then you have poll ring, then you have dumped interlock saddles plastic rings, then interlock saddle and its modification has been done later. So, different modification poll ring and then their modified versions you could see over here. So, these are a different kind of internal modifications has been made to enhance the surface area and to increase its efficiency of the packing. So, these are called second generation packing.

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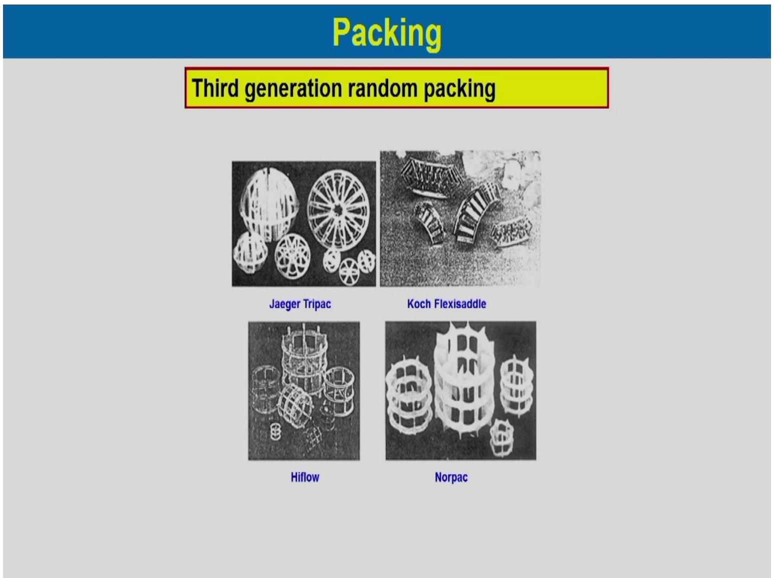

The third generation packing are Jaeger Tripac, Koch Flexisaddle, Hiflow and Norpac. So, you can see the design of the third generation random packing which are much more efficient compared to the first and second generation random packing. As far as random packing is concerned it is generally considered cheaper compared to the structured packing.

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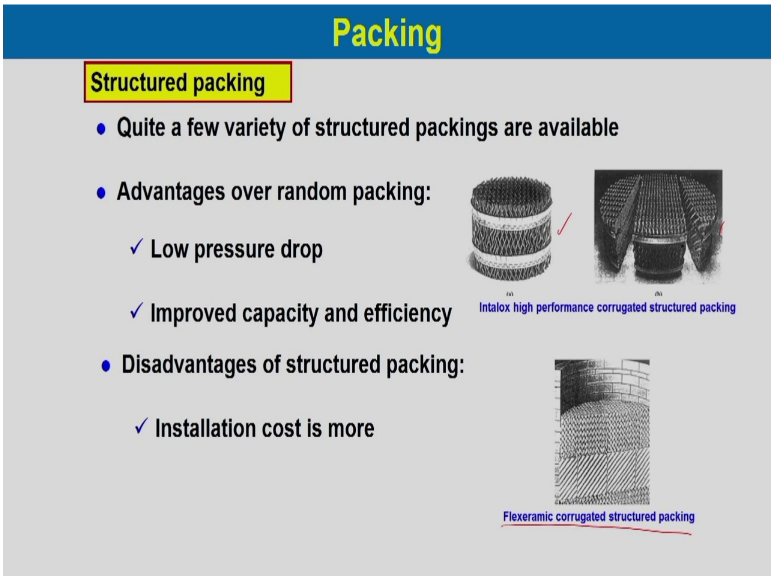

So, structured packing we have quite a few varieties of structured packing are available, one of them is the interlocks high performance corrugated structured packing you can see how it is designed and has very high surface area then flexeramic corrugated structured packing made as no ceramic materials and also it is a structured packing.

Advantage over random packing is that it has a low pressure drop, improved capacity and efficiency. The disadvantage of the structured packing is that installation cost is more. So, for structured packing you should have a know proper installation procedure to be followed, otherwise the distribution until it is properly installed inside the column it will block the flow of the liquid and the gas and pressure drop will be higher. So, proper installation has to be made in case of the structured packing and hence its installation cost is higher compared to the random packing. However, it has no good advantage over the random packing which is know the pressure drop is comparatively low and it has interfacial area is much higher compared to the random packing and its efficiency is also higher.

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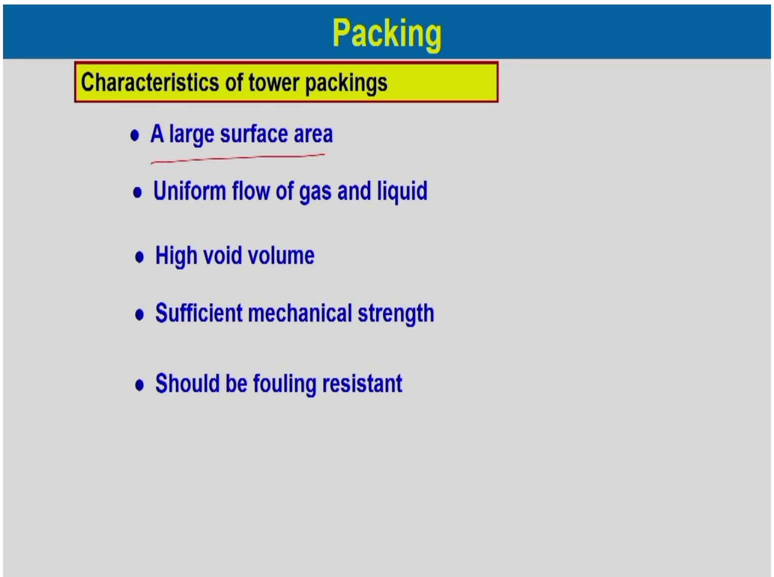

So, as we said the large surface area characteristics of the tower packing which are required uniform flow of gas and liquid and high void volume which is also required, sufficient mechanical strength and should be fouling resistant. So, these are the properties or characteristics of the packing materials required and which mostly supported by the structured packing.

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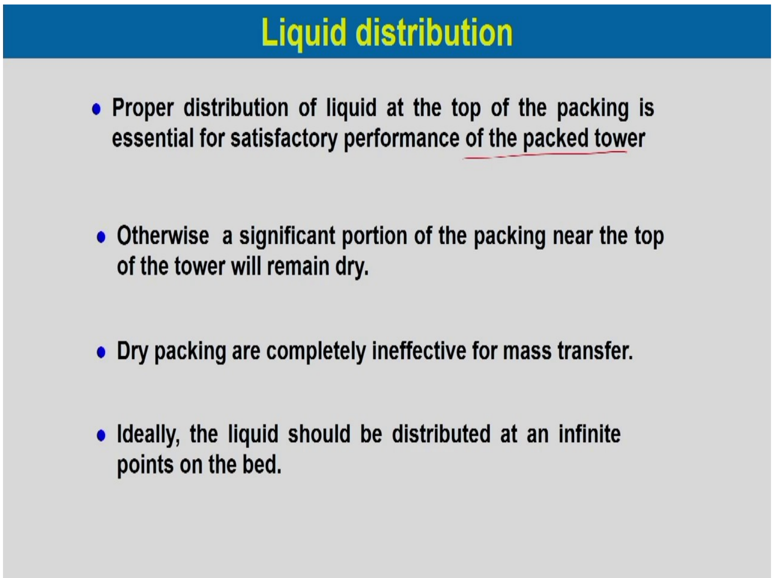

Now, liquid distribution proper distribution of the liquid at the top of the packing is essential for satisfactory performance of the packed towers. Otherwise, what would happen a significant portion of the packing near the top of the tower will remain dry. So, if that happens the efficiency of the tower is going to go down. Dry packing are completely in effective for mass transfers. So, ideally the liquid should be distributed at an infinite points on the bed. So, at all the points the liquid should be properly distributed and then the gas will be you know flowing through this and will have a large interfacial area.

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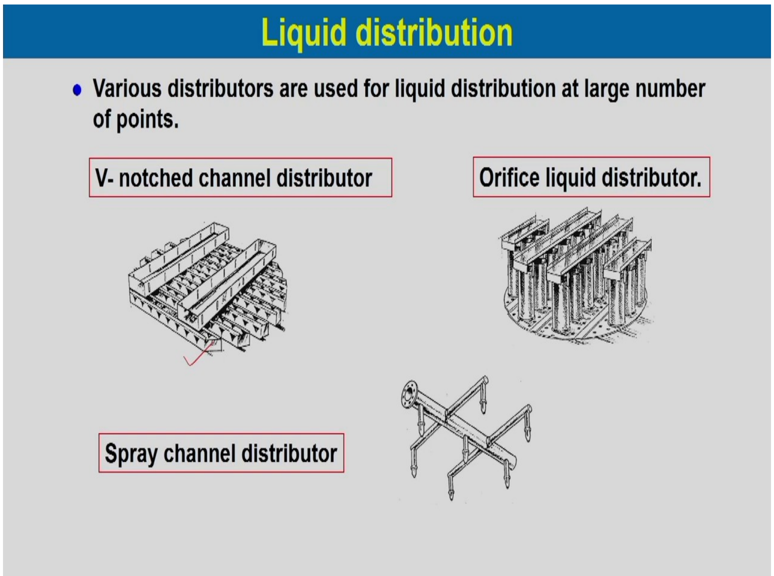

Various distributors are used for liquid distribution at large number of points; one such is that V-notched channel distributor. So, this is the V-notched channel distributor which is used to distribute the liquid at different points. Then Orifice liquid distributor; so, this is you know the design is like a orifice and it you know efficiently distribute the liquid throughout the column internals. Then Spray channel distributor, where know it the liquid is sprayed through a channels, so it distributes know evenly throughout the column.

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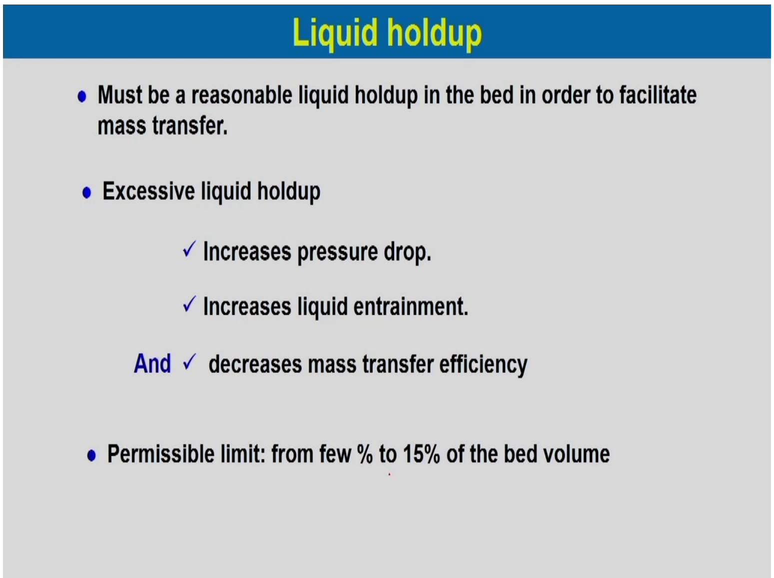

Liquid holdup as you know it should have must have reasonable liquid holdup in the bed in order to facilitate mass transfer. If you could not a no give a proper liquid holdup inside the column there will be less contact between the gas and liquid and hence the mass transfer will not be effective. So, we should have a reasonable liquid hold up in the bed in order to facilitate the mass transfer.

Excessive liquid hold up, so if we have very high liquid hold up then it causes no increase in pressure drop. So, we should not have very high liquid holdup which will increase the pressure drop and also increase the energy cost will be much higher. And it also increases the liquid entrainment, so the liquid holdup is more the entrainment will be more and also decreases the mass transfer efficiency. So, we should not have no excessive liquid holdup; so, permissible limit from few percent to 15 percent of the bed volume. So, the liquid holdup should be within the 15 percent of the total packed bed volume.

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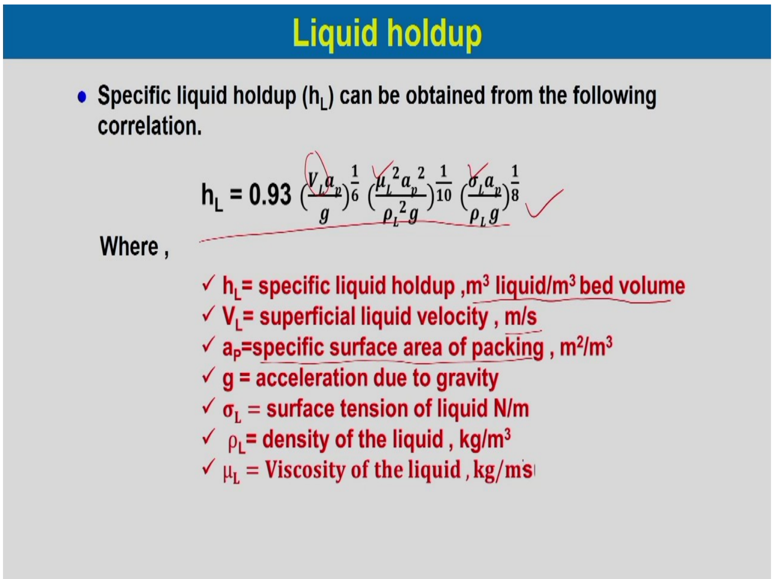

The specific liquid holdup h L can be obtained from the following correlation; h L is equal to 0.93 V L a p divided by g to the power one-six into mu L square a p square divided by rho L square into g to the power one-tenth into sigma L a p divided by rho L g to the power one-eighth. So, with this equation we can calculate the liquid holdup h L.

In this case, h L is the specific liquid holdup which is in meter cube of liquid per meter cube of bed volume, and V L is a superficial liquid velocity in meter per second, a p is a specific surface area of the packing that means, it is meter square per meter cube of the bed volume or meter cube of the bed and g is the acceleration due to gravity. Sigma L is the surface tension of the liquid it is in newton per meter, rho L is the density of the liquid which is kg per meter cube and mu L is the viscosity of the liquid in kg per meter second. So, all these are in SI unit. With this no, a correlation we can calculate the specific liquid holdup.

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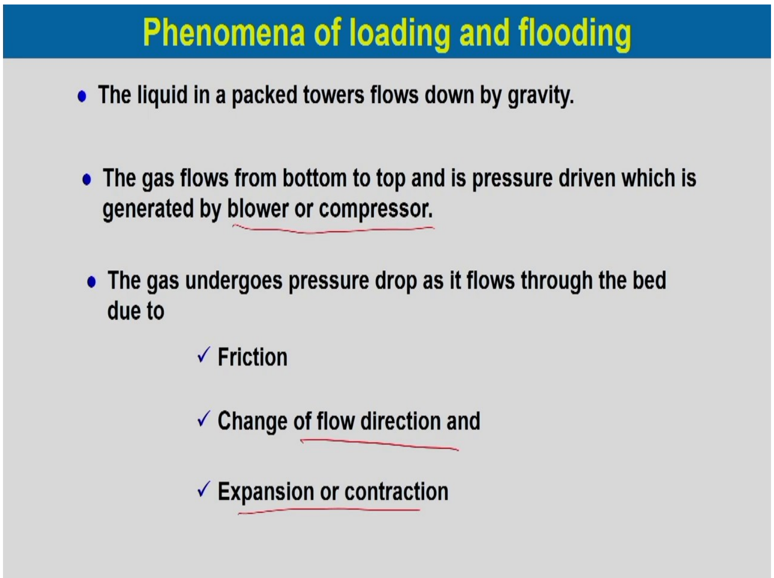

Now, we would like to see the phenomena of loading and flooding in the packed weight. We have discussed this in case of the tray towers or the plate towers now the similar phenomena we will discuss in case of the packed towers. The liquid in a packed towers flows down by gravity. The gas which flows from the bottom to top and is pressure driven which is generated by blower or compressor. Then the gas undergoes pressure drop as it flows through the bed due to the few factors, one is friction and then change of flow direction and expansion or contractions, as you know there will be different size of packing materials and the white space are not uniform throughout the base.

So, if that would happen then the and also there will be channeling and all inside the bed, so one factor would be the friction because of no it is the friction between the solid to you know liquid and the and the gas. And secondly, is the change of flow directions because if the flow path is not no like channel it is not the uniform channels. So, there will be a change in the flow direction, and also as it is not uniform path there will be some places expansion and there will be some places of contraction. So, because of these the know gas undergoes pressure drop when it flows through the bed.

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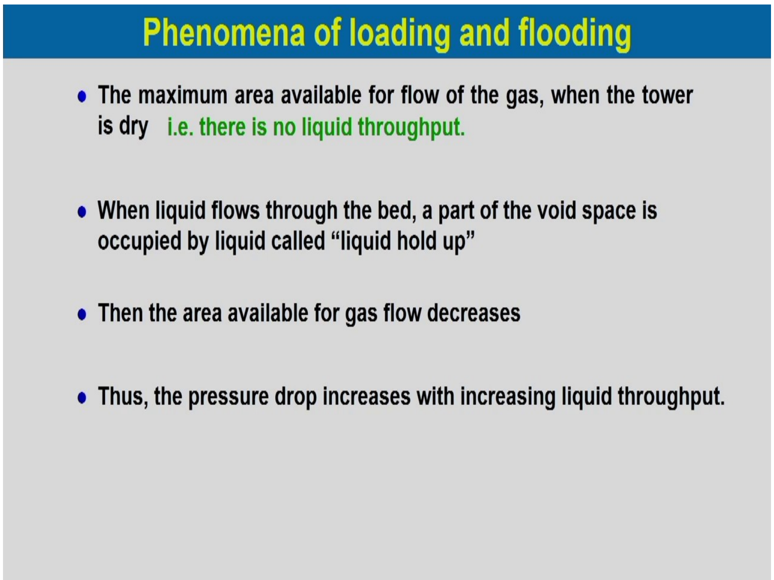

The maximum area available for flow of the gas when the tower is dry that is there is no liquid throughput. So, if the tower is empty there is no liquid flow, so all the void space are available for the gas to flow through that space. So, then we will experience maximum flow of the gas through the tower when it is dry.

When the liquid flows through the bed a part of the void space is occupied by the liquid and we called it is liquid holdup, then the area available for gas flow decreases. Thus, the pressure drop increases with increasing liquid throughput. So, as the area for the gas flow decreases because of increasing liquid throughput the pressure drop also increases.

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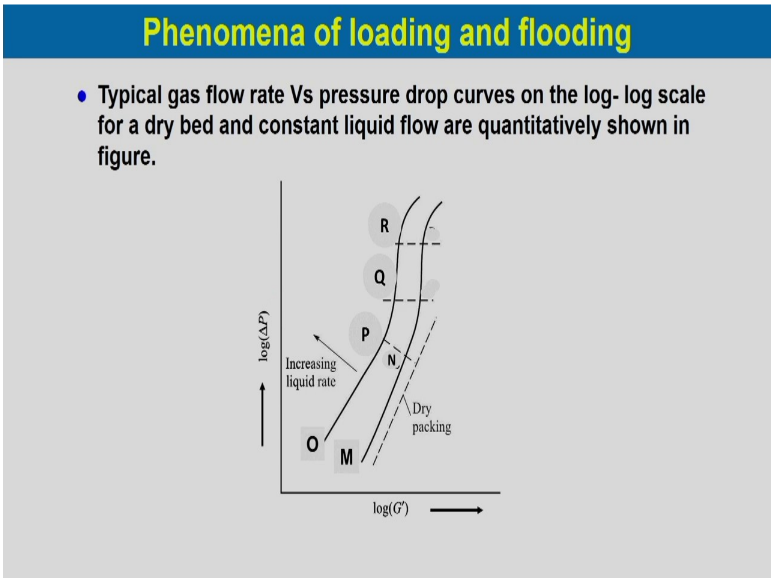

Typical gas flow rate versus know pressure curves on the log-log scale for a dry bed and constant liquid flow rate can be quantitatively shown in the figure over here. So, you can see in the y axis it is log of no log scale log and delta P and the x axis is log G dash the gas flow rate and at a constant liquid flow rate.

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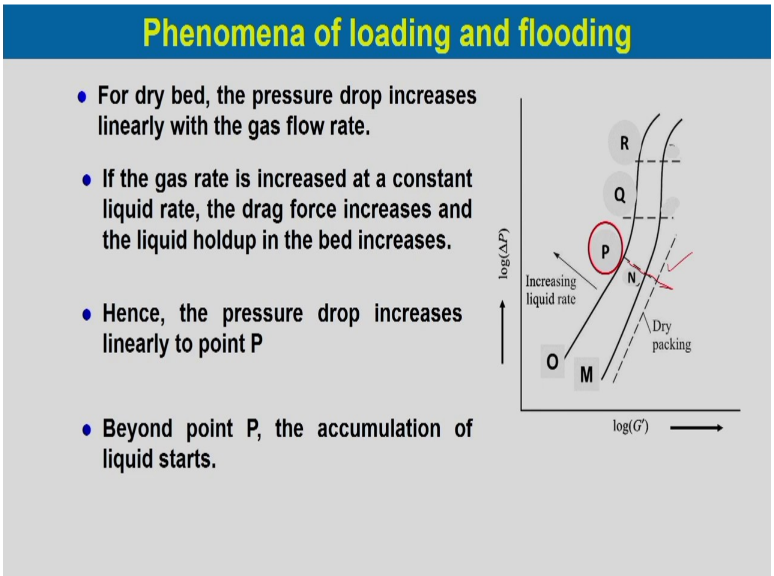

So, for a dry bed the pressure drop increases linearly with the gas flow rate. So, you can see this is the for dry bed and the pressure drop as you can see is you know linearly increases in this case. The gas rate is increased at a constant liquid rate, the drag force increases and the liquid holdup in the bed increases. So, as you, will see the liquid holdup slowly will increase as the drag force increase no because of increased gas flow rate. Hence the pressure drop increases linearly to point P. So, up to this point, so the pressure drop increases linearly.

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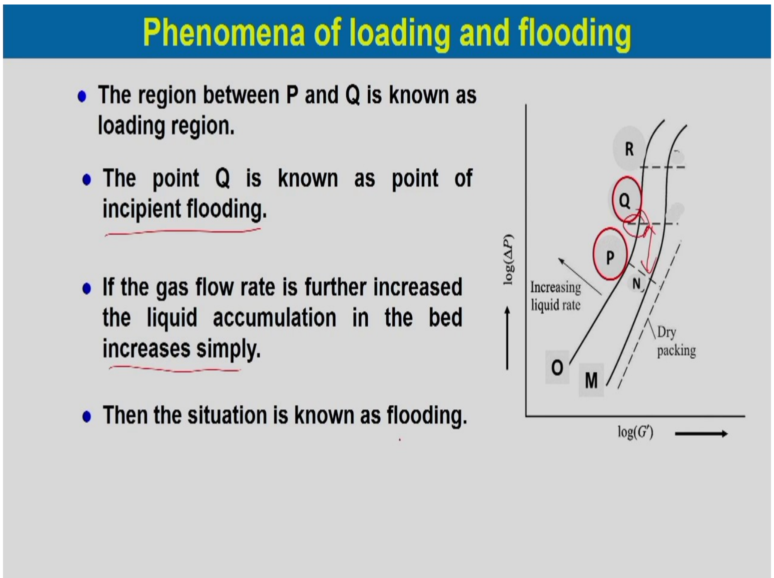

So, what happens beyond this point? The accumulation of the liquid starts inside the column and the region between P and Q is known as the loading region. So, this region, so between this P and Q, in this region they is called the loading region and where the maximum accumulation of that liquid can happen. The point Q is known as the point of incipient flooding. So, this is the incipient flooding at point Q.

If the gas flow rate is again increased the liquid accumulation in the bed increases very sharply because the column is already started flooding, and as you increase the pressure there will be no complete flooding inside the column and the pressure drop will sharply increase. So, you can see the sharp increase of pressure drop with a very small change of the vapor or gas flow. So, then the situation is known as the flooding.

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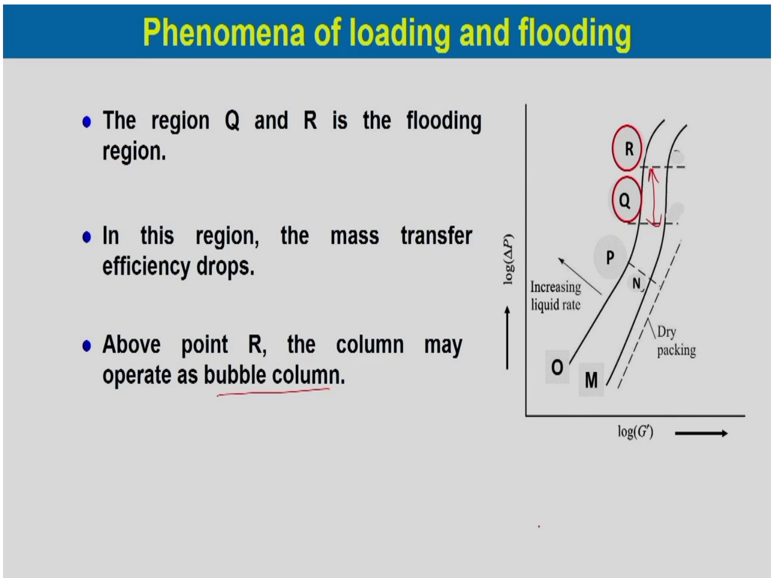

So, the region between Q and R is the flooding region. So, in this region is known as the flooding region. In this region the mass transfer efficiency drops sharply and above for the column may operate as bubble column, as you can see in this case it will run as bubble column.

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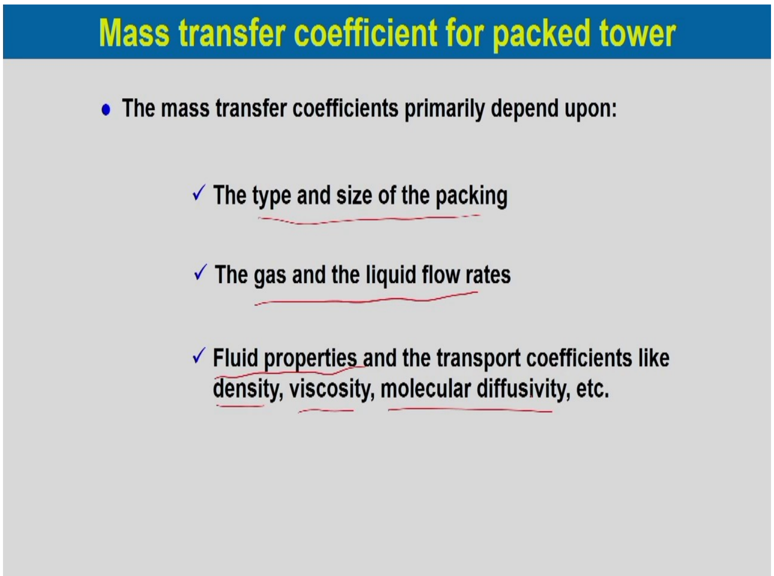

Now, we will discuss the mass transfer coefficient for the packed column. It will depend on the type and size of the packing, the gas and the liquid flow rates, fluid properties and the transport coefficient like density, viscosity, molecular diffusivity etcetera. So, the mass transfer coefficient in packed towers will primarily depend on this parameters type and size of packing we are going to use, gas and liquid flow rates, and other fluid properties like know transport coefficient like density, viscosity, molecular diffusivity, this parameters are required to calculate the mass transfer coefficient.

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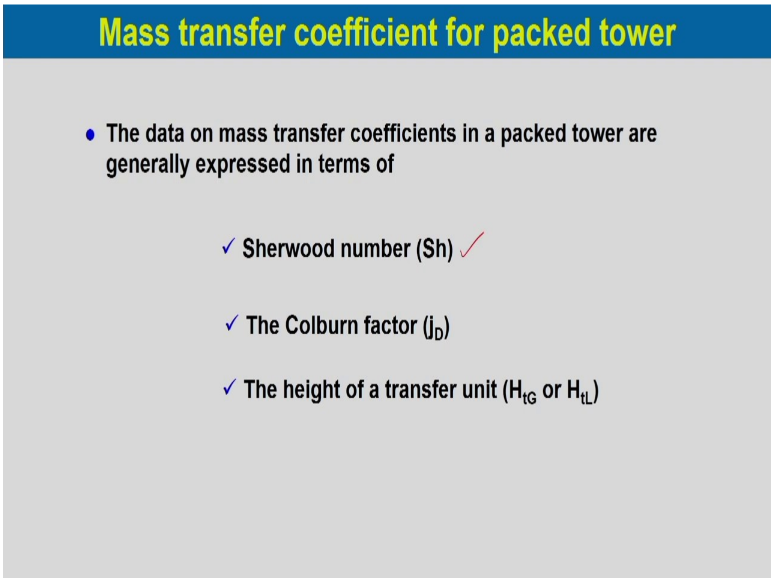

The data and mass transfer coefficient in a packed towers are generally expressed in terms of dimensionless number and one of them is the Sherwood number. This is already we have discussed at the beginning different dimensionless groups. So, in mass transfer we have discussed you know different dimensionless groups in mass transfer. Sherwood number is one of the dimensionless number, and then the Colburn factor j D and the third is the height of transfer unit that is H tG or H tL.

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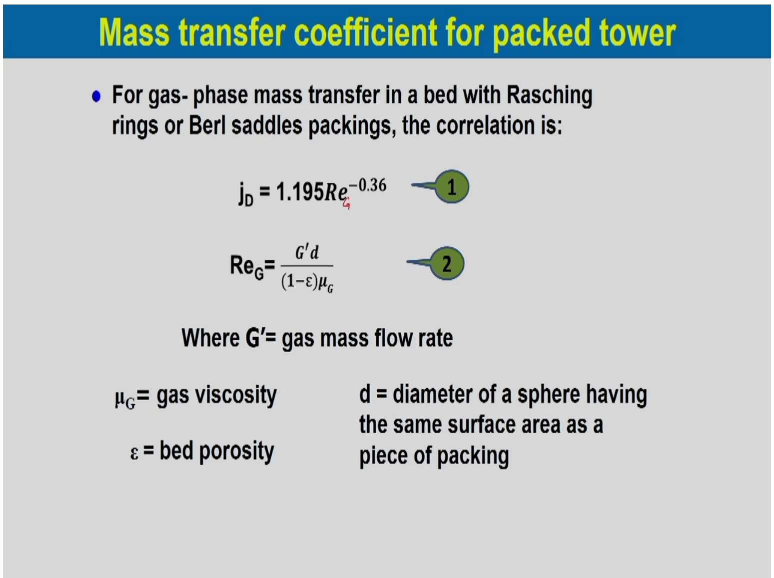

For gas phase mass transfer in a bed with Rasching rings or Berl saddle packing the correlations which are available which is j D is equal to 1.195 into Re G to the power minus 0.36. So, in this case the Re G is equal to G dash into d divided by 1 minus epsilon into mu G. G dash is the gas mass flow rate and mu G is the gas viscosity, epsilon is the bed porosity and D is the diameter of the sphere having the same surface area as a piece of packing. As you know the packing materials are not in uniform in size and the diameter of a sphere, we will consider which have the same surface area of that packing material. So, we can calculate like the equivalent diameter.

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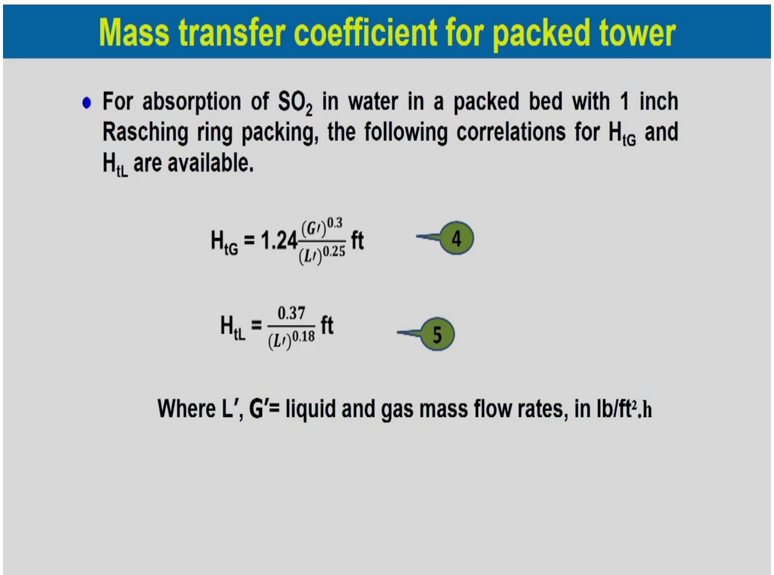

For gas absorption of sulphur dioxide in water in a packed bed with 1 inch no Rasching ring packing, the following correlations for H tG and H tL are available in the literature.

H tG is equal to 1.24 into G dash to the power 0.3 divided by L dash to the power 0.25 in feet and H tL is equal to 0.37 divided by L dash to the power 0.18 feet. So, here L dash and G dash is the liquid and gas mass flow rates in pound per feet square hour.

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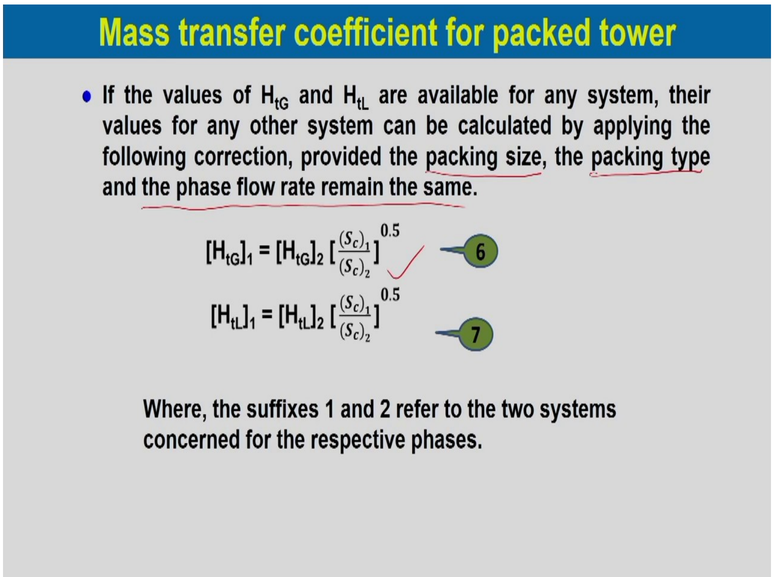

Now, if the values of H tG and H tL are available for any system, their values for any other system can be calculated by applying the following correlations. So, if we know the values for a particular system for H tG or H tL this can be applied to other system provided that the packing size, the packing type and the phase flow rate remains the same. So, if these conditions are satisfied same packing size, packing type and the phase flow rate those are same if we know for a particular system it can be calculated for other system of these values. So, these are very helpful correlations which is H tG 1 is equal to H tG 2 into Schmidt number 1 divided by Schmidt number 2 whole to the power 0.5.

So, using this equation we can calculate the H tG for the new system. And similarly, for the H tL and which also related to the Schmidt number of the 2 systems. The suffix which are used 1 and 2 refer to the two systems concerned for the respective phases.

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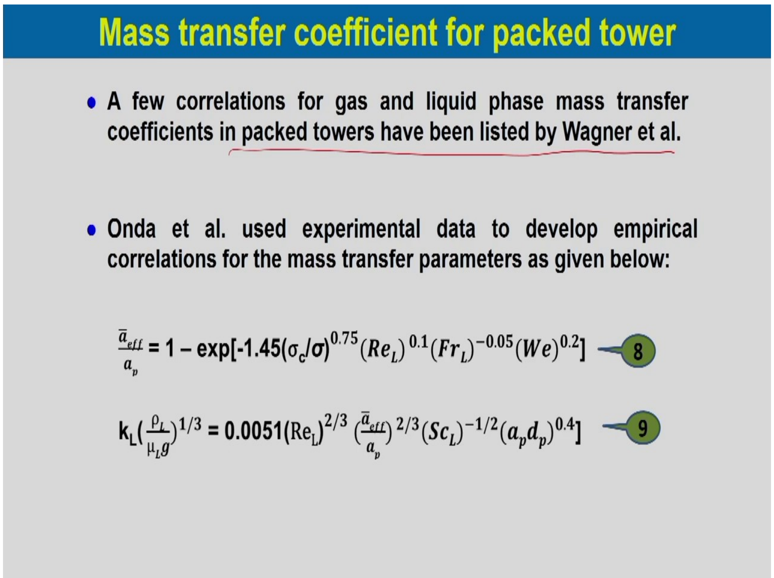

Now, a few correlations for gas and liquid phase mass transfer coefficients in packed towers have been listed by Wagner et al. So, it is given in the literature for few correlations for this systems. And Onda et al used experimental data to develop empirical correlation for the mass transfer parameters as given below. So, one is a bar effective divided by a p would be equal to 1 minus exponential minus 1.45 into sigma c by sigma to the power 0.75 Re L to the power 0.1 into Fr L to the power minus 0.05 into We to the power 0.2. And then k L would be rho L by mu L G to the power one-third would be equal to 0.0051 Re L to the power two-third a bar effective divided by a p to the power two-third Sc L to the power minus half a p d p to the power 0.4.

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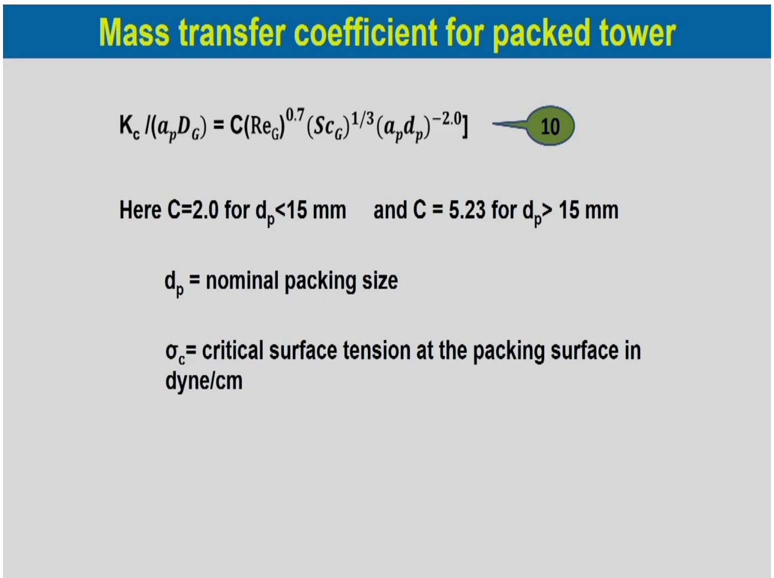

So, here K c by a p D G would be equal to c into reg to the power 0.7, Sc G to the power one-third into ap d p to the power minus 2.0. This equations 8, 9 and 10 these are used to calculate different parameters or the mass transfer coefficients. Here C is equal to 2 for d p less than 15 millimeter and c would be equal to 5.23 for d p greater than 15 millimeter. d p is the nominal packing diameter, sigma c as we in the earlier equation sigma c is the critical surface tension at the packing surface in dyne per centimeter.

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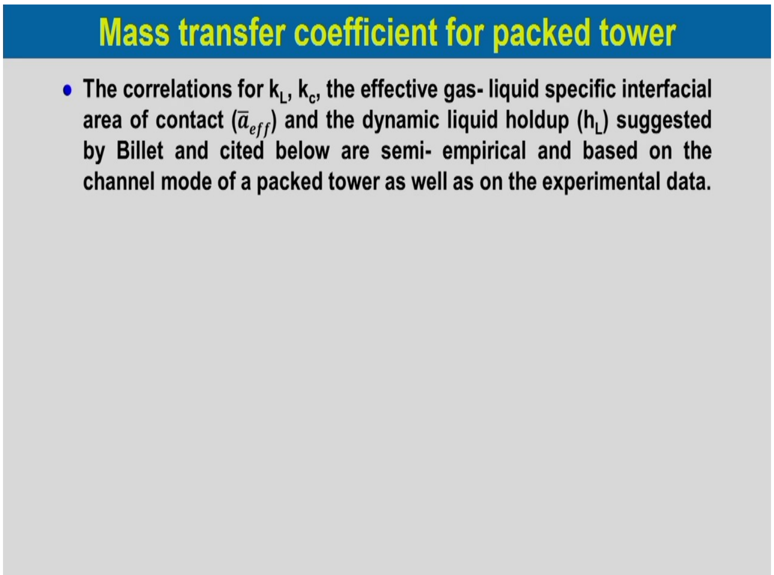

The correlation for k L and k c the effective gas liquid specific interfacial area of contract that is a bar effective and the dynamic liquid holdup h L suggested by billet and cited are semi empirical and based on the channel mode of a packed tower as well as on the experimental data.

So, now we will discuss the design of Sparged Vessel.

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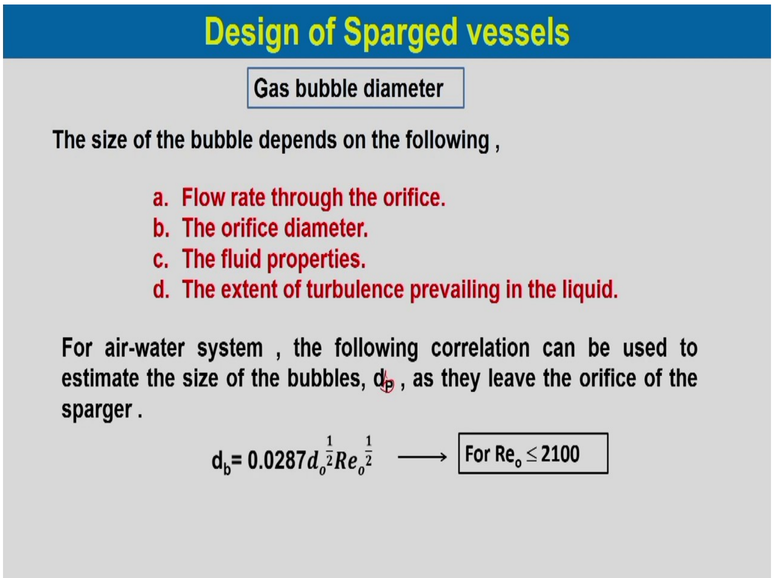

So, in the sparged vessel gas bubble diameter is very important. The size of bubble depends on the following factors. As you know the flow rate through the orifice the orifice diameter the fluid properties then the extent of turbulence prevailing inside the liquid. So, for air water system the following correlations can be used to estimate the size of the bubble that is d p as they leave the orifice of the sparger. So, d b equal to 0.287 d naught to the power half into Re naught to the power half. So, for Re naught less than 2100 we can use this correlation this is basically for the Reynolds number which is in the laminar range, so within 2100.

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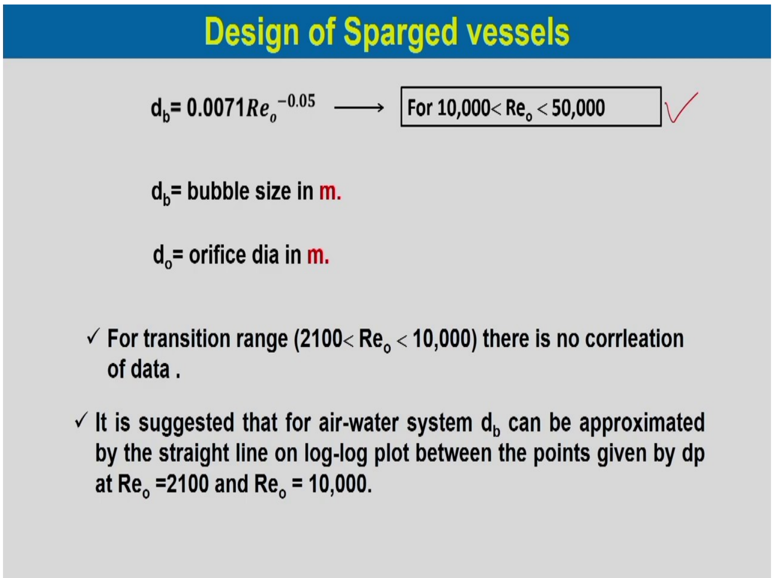

And then this d b bubble diameter we can use this correlation for Reynolds number which is greater than 10,000 and less than 50,000 which is d b equal to 0.0071 Re naught to the power minus 0.05. So, this is for this Reynolds number. And d b is the bubble diameter in meter and d o is the orifice dia in meter.

Now for the transition range, so which is between 2100 to around 10,000 there is no correlations available for the data. So, it is suggested for air water system d b can be approximated by the straight line on log-log plot between the points given by d p at Re naught is equal to 2100 and Re naught equal to 10,000. So, if you wanted to calculate the d b in case of the transition region we can have a log-log plot having the other two data, two extreme data, so we can intrapolate and extrapolate and get the values for know d b bubble diameter, in that range.

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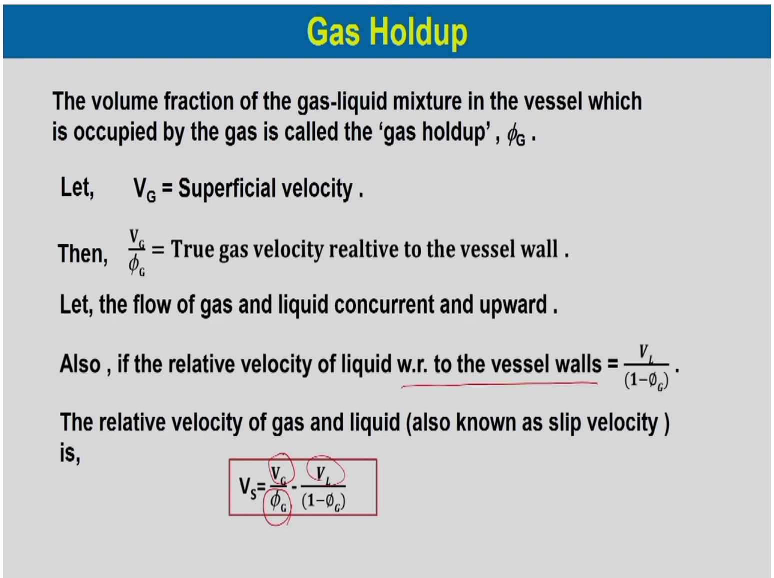

Now, gas holdup the volume fractions of the gas liquid mixture in the vessel which is occupied by gas is called the gas holdup which is defined by small phi G and V G is the superficial velocity. Then this V G by phi G is the true gas velocity relative to the vessel wall, and let the flow of gas and liquid concurrent and upward. Then, if the relative velocity of the liquid with respect to the vessel walls is equal to V L by 1 minus phi G.

The relative velocity of gas and liquid also known as slip velocity can be calculated V s would be equal to V G by phi G minus V L by 1 minus phi G. So, with this equation we can calculate. So, knowing the value of superficial velocity and then the gas holdup phi G and also the liquid velocity we can calculate the slip velocity or relative velocity. So, this equation will also give the slip velocity for counter current flow of a liquid if V L for the downward liquid flow is assigned a negative value.

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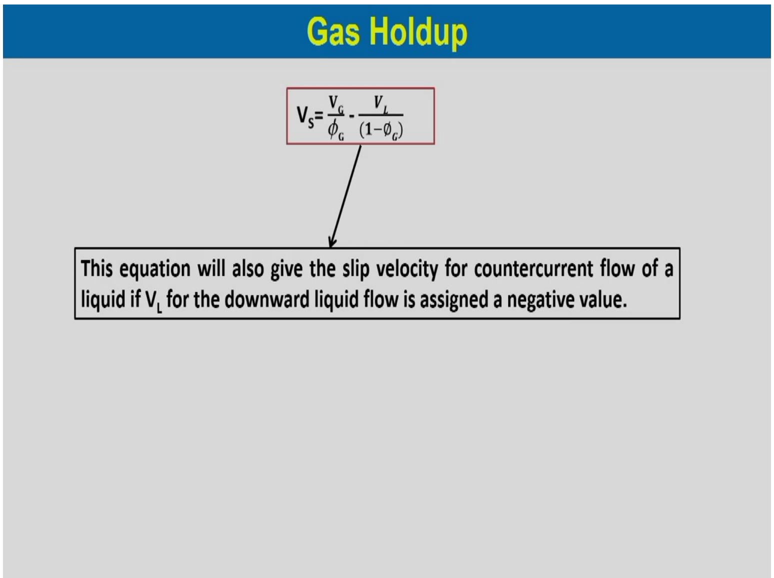

So, if we just put negative value to V L the same equations can be used to calculate the slip velocity for the counter current operation. This is for the co-current operation or concurrent operation and we can have the same slip velocity for the counter current operation as well.

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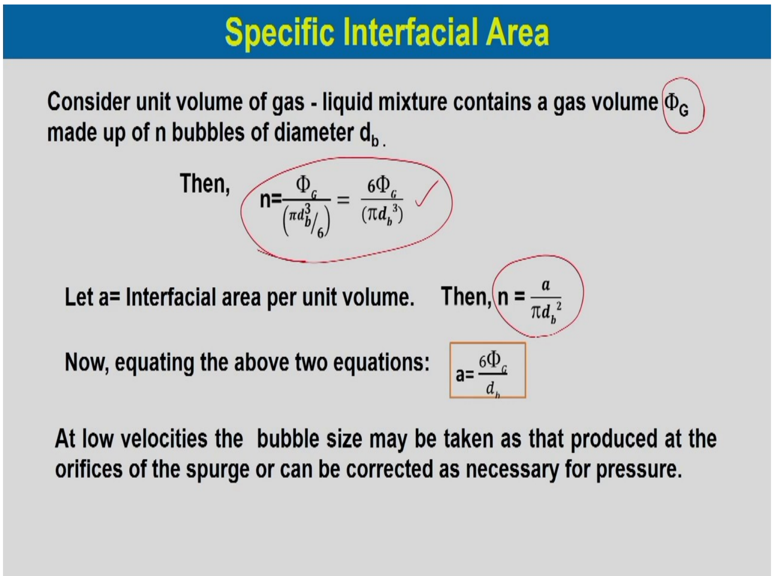

Now, specific interfacial area consider unit volume of gas and liquid mixture which contains a gas volume of capital phi G made of n bubbles of diameter d b. So, in this case we can write the number of bubbles would be equal to the gas volume phi G, capital phi

G divided by the pi d b cube divided by 6, the volume of the bubble. So, you can calculate the number of bubbles which is equal to 6 phi G divided by pi d b cube.

Now, consider a is the interfacial area per unit volume in that case n would be equal to a by pi d b square. Now, if we equate this two relation from this and this we will obtain a is equal to 6 phi G divided by d b. At low velocities the bubble size may be taken as that produced at the orifices of the spurge or can be correlated as necessary for pressure. So, there might be corrections with the pressure or we can take as per the orifice design.

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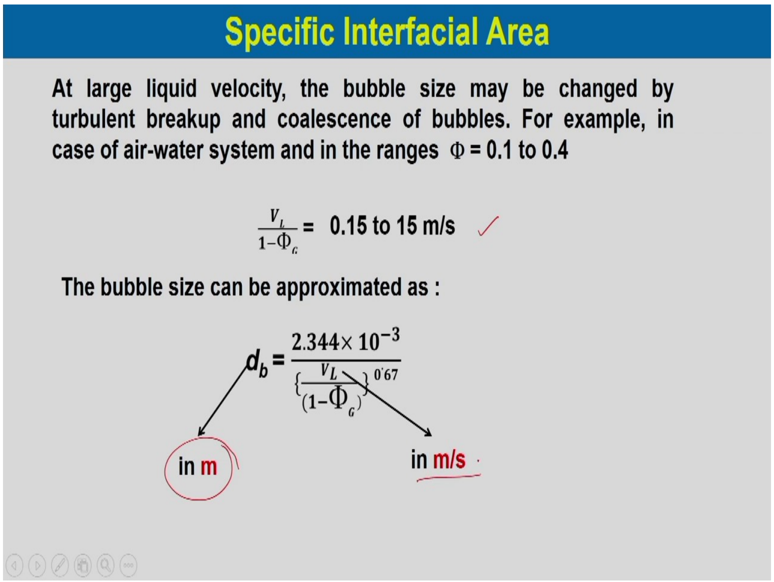

At large liquid velocity the bubble size may be changed by turbulent breakup and coalescence of bubbles. For example, in case of air water system and in the range of phi is equal to 0.1 to 0.4; V L by 1 minus phi G is equal to 0.15 to 15 meter per second. So, the bubble size can be approximated d b would be equal to 2.344 into 10 to the power minus 3 divided by V L by 1 minus phi G to the power 0.67. So, this d b is in meter and V L is in meter per second.

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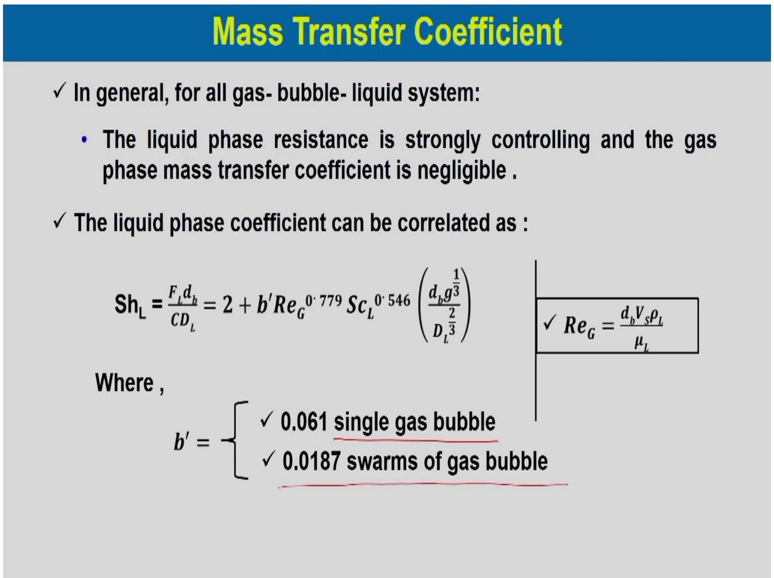

So, in general for gas bubble liquid systems the liquid phase resistance is strongly controlling and the gas phase mass transfer coefficient is negligible. So, because of this know resistance which mostly lies in the liquid phase the liquid phase coefficient we need to obtain for this case.

The liquid phase mass transfer coefficient can be correlated with the Sherwood number which is Sh L in the liquid phase would be equal to F L d b divided by C D L, which is equal to 2 plus b dash Re G to the power 0.779, Schmidt number in the liquid to the power 0.546 whole into d b g to the power one-third divided by D L to the power two-third.

So, here b dash over here is a constant which is 0.061 for the single gas bubbles and it is 0.0187 for know number of gas bubbles. Re G is equal to d b V s into rho L divided mu L. So, this is the Reynolds number of the gas.

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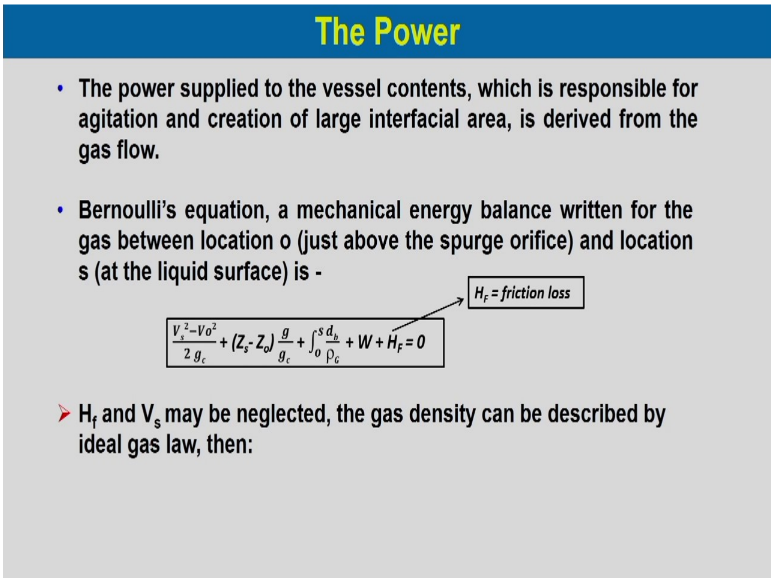

Now, the power supplied to the vessel content which is responsible for agitation and creation of the large interfacial area is derived from the gas flow rate. Now, know Bernoulli's equation a mechanical energy balance know written for the gas between the location o and the location s, location just above the sparge orifice and at the location s at the liquid surface which can be used you have learnt the Bernoulli's equation mechanical energy balance equation in your fluid mechanics course.

So, which can be written over here V s square minus V naught square divided by twice g c plus Z s minus Z naught into g by g c plus integral 0 to s d b by rho G plus W plus H F equal to 0. Here H F is the frictional loss, and H F and V s may be neglected the gas density can be described as ideal gas law.

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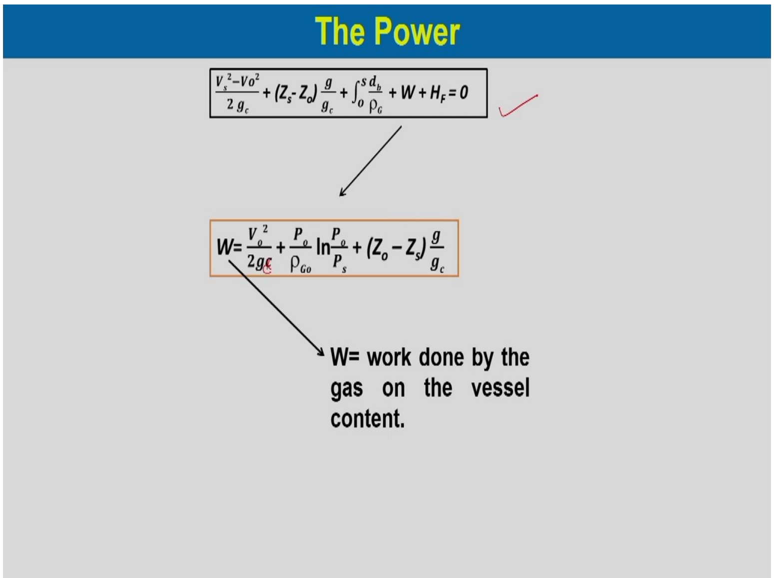

In that case we can write V s square minus v naught square divided by twice g c plus Z s minus Z naught into g by g c plus integral 0 to s d b by rho G plus W plus H F is equal to 0. This equation would be written as W is equal to V naught square divided by twice gc plus rho c, W is equal to V naught square divided by 2 gc plus P naught by rho G naught ln P naught divided by P s plus Z naught minus Z s into g by g c. This W is work done by the gas on the vessel contact.

So, thank you for hearing this lecture. And we will continue our discussion in the next lecture.