# Convective Heat and Mass Transfer

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## Module No.# 01

## Lecture No. # 34

## Reynolds Flow Model

So, now we will look at the Reynolds flow model, which is an algebraic model as we said earlier, unlike the Stefan flow model which is a differential equation model but one-dimensional differential equation model, and the coquette flow model, which again was a one-dimensional model.

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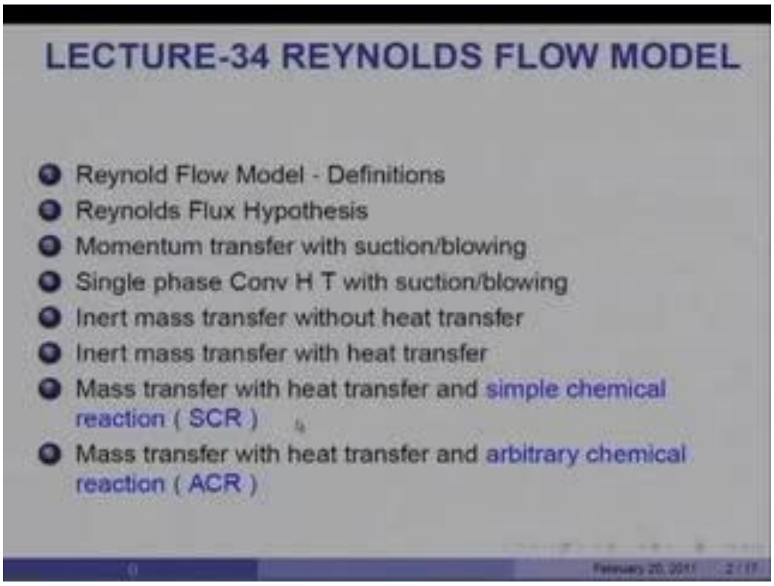

LECTURE-34 REYNOLDS FLOW MODEL

- Reynold Flow Model - Definitions
- Reynolds Flux Hypothesis
- Momentum transfer with suction/blowing
- Single phase Conv H T with suction/blowing
- Inert mass transfer without heat transfer
- Inert mass transfer with heat transfer
- Mass transfer with heat transfer and simple chemical reaction ( SCR )
- Mass transfer with heat transfer and arbitrary chemical reaction ( ACR )

Our interest now is to develop the features of the Reynolds flow model and what I will do is, first of all, define Reynolds flow model, again go over the Reynolds flux hypothesis. Reynolds flow model is supposed to mimic the boundary layer flow model. We will also consider momentum transfer with suction and blowing, we will also consider single phase convective heat transfer with suction and blowing, and then, the four types of mass transfer problems in which the species mass transfer equation as well as the energy equation will be invoked.

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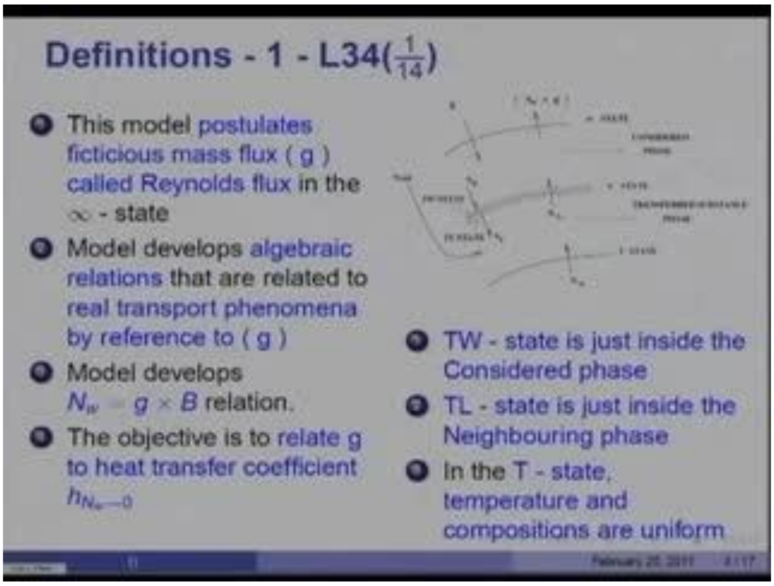

### Definitions - 1 - L34($\frac{1}{14}$)

- This model postulates ficticious mass flux ( g ) called Reynolds flux in the $\infty$ - state
- Model develops algebraic relations that are related to real transport phenomena by reference to ( g )
- Model develops $N_w = g \times B$ relation.
- The objective is to relate g to heat transfer coefficient $h_{N_w \to 0}$
- TW - state is just inside the Considered phase
- TL - state is just inside the Neighbouring phase
- In the T - state, temperature and compositions are uniform

I am now going to define the Reynolds flow model, as you recall this is the interface, this is the considered phase to which the mass transfer n w is taking place. This is the transfer substance phase from where the mass transfer is taking place to the considered phase. And this is the infinity state which is far away from the interface in the considered phase. So, the principle idea of the Reynolds flow model revolves around postulation of a fictitious flux. It is a notional flux kilograms per meter square second. So, it is a notional mass flux.

Defined in the infinity state and its direction is towards the interface, and in doing so, as it flows from the infinity state towards the interface or the w state it carries with it the properties of the infinity state. So, that is the first postulate of the model. And the model develops algebraic relations that are related to the real transport phenomena by reference to this flux g. And further the model will result into n w equal to g times b. So, instead of the logarithmic relations we saw earlier, the model will develop n w equal to g times has the appropriate relationship. Now, the objective of course, in order to solve the problem, we first need to define b of course, which we have done in the previous two lectures for a variety of mass transfer problems. And the second thing is to evaluate g, but the way we will do it is our route will be that g will be related to the value of the heat transfer coefficient in the absence of mass transfer. So, we are going to relate g.

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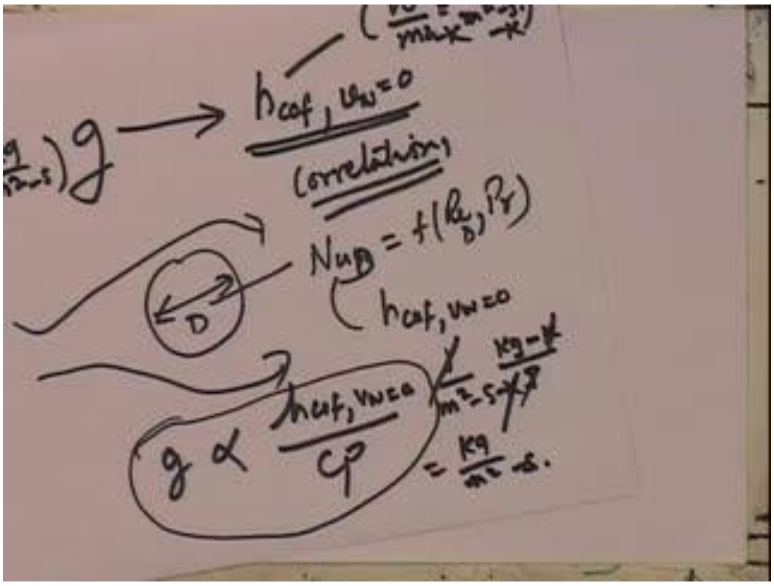

We are going to relate g to heat transfer coefficient for v w equal to 0 and the reason we do this is because, these values are available for wide variety of situations through correlations. So, given a mass transfer situation, let us say, flow over a cylinder like this, then I would know what the nusselt number is as a function of Reynolds number and prandtl number nusselt number based on diameter d.

I would know what it will be as a function of Reynolds number and prandtl number and for which I can divide h cof w equal to 0. If I can relate this value to g which is in kilograms per meter square second, whereas, h is in watts per meter square or essentially joules per meter square second. So, obviously, we expect relationship between g will be proportional to h cof w equal to 0 divided by specific heat.

Now, this will have the same units because h cof is joules per meter square second and c p is joules per kg per kelvin divided by meter square kelvin k divided by k divided by k. So, k k gets cancelled, Joules and joules get cancelled and again you will get kg per meter square second. So, this is what we are aiming at. What therefore, this will do is the following. That given a mass transfer situation of any kind, this can be a mass transfer flow cylinder or it can be flow through a tube or anything like that, we calculate the appropriate value of b and evaluate g from the knowledge of the heat transfer coefficient in the absence of mass transfer in that same situation. So, basically no differential equations are solved the entire model is algebraic.

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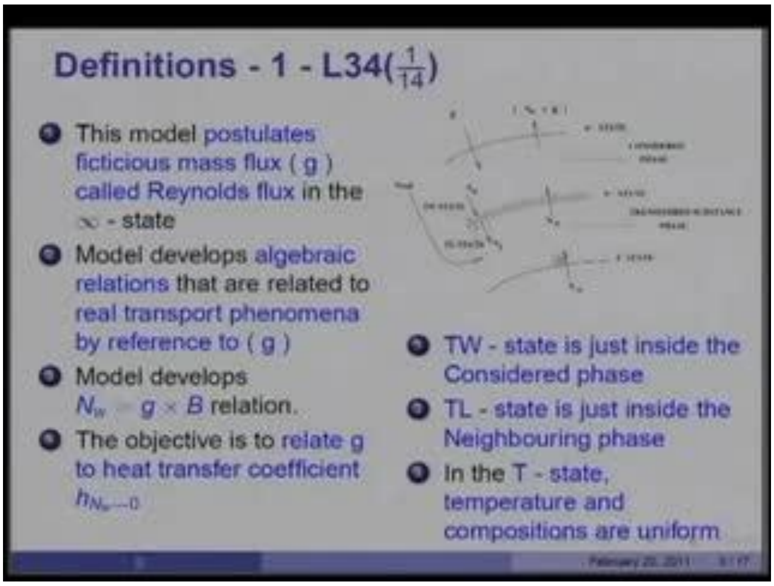

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Now, in order to facilitate further discussion, I am going to define in the w state, two more surfaces, which are almost coincident with the w state interface. One of them is the t w state, which is just inside the considered phase and t l state which will be just inside the transferred substance phase. So, this is what t w and t l are. The t in the t state as you remember there are no temperature or composition gradients and they call they are uniform.

In addition, we shall also postulate that there will be conduction heat transfer from the considered phase to the w state and there can also be possibly some temperature variation from deep inside the transferred substance phase to the interface in which case there would be some conduction heat transfer even in the transferred substance phase which I have called q l. This is q w. In the considered phase, there can also be radiation heat transfer to the wall in the considered phase.

And in this model for convenience we will account for it as though the knit the radiation flux actually penetrates the interface and appears in the transferred substance phase. The reason for doing this would be very obvious as we go along. So, there are few new definitions introduced t w state, t l state, heat flux q w and heat flux q l in the considered n transferred substance phases respectively and the q radiation which is accounted in the transferred substance phase.

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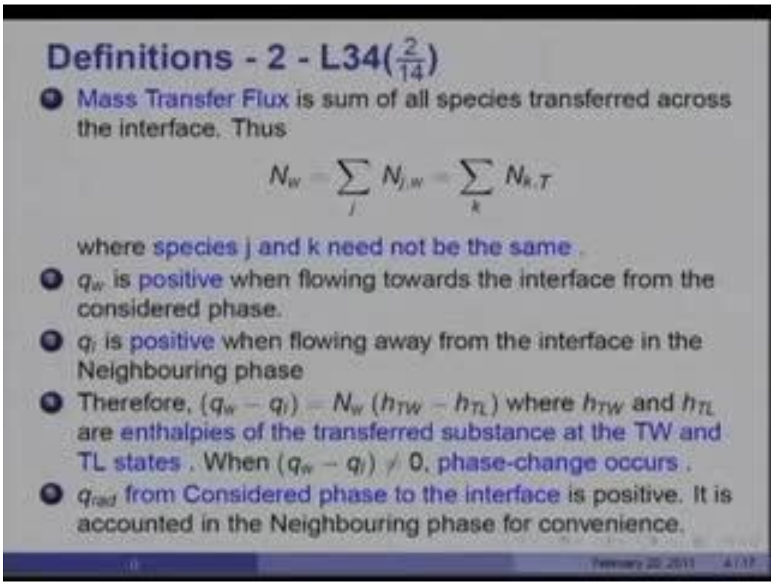

## Definitions - 2 - L34($\frac{2}{14}$)

- **Mass Transfer Flux** is sum of all species transferred across the interface. Thus

$$N_w = \sum_j N_{j,w} = \sum_k N_{k,T}$$

where species j and k need not be the same .

- $q_w$ is positive when flowing towards the interface from the considered phase.
- $q_l$ is positive when flowing away from the interface in the Neighbouring phase
- Therefore, $(q_w - q_l) = N_w\,(h_{TW} - h_{TL})$ where $h_{TW}$ and $h_{TL}$ are enthalpies of the transferred substance at the TW and TL states . When $(q_w - q_l) \neq 0$, phase-change occurs .
- $q_{rad}$ from Considered phase to the interface is positive. It is accounted in the Neighbouring phase for convenience.

The mass transfer flux is the sum of all species transferred across the interface. So, n w can be sum of n j w. Let us say, in the transferred substance phase and n k w in the considered phase. The species j in the transferred substance phase need not be the same as species j species k in the considered phase.

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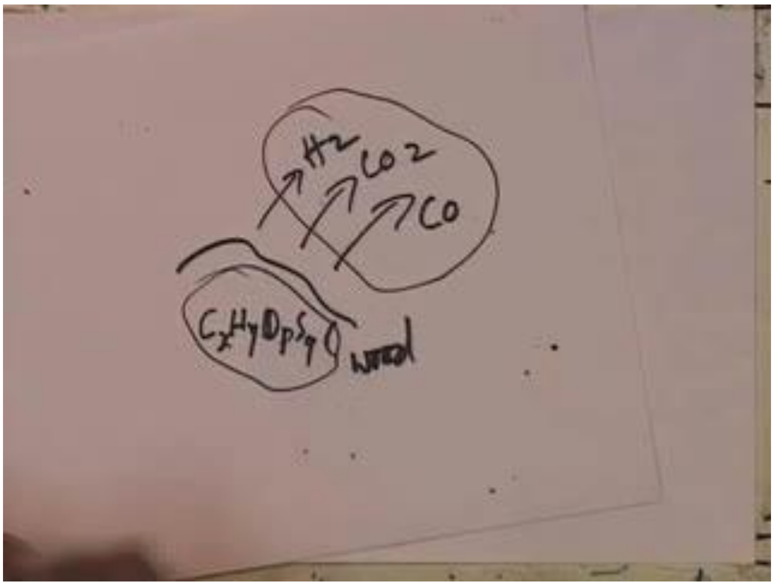

This is typically the case when, let us say, I have mass transfer taking place across a wood particle. Then here, there will be the wood itself has a formula of c x h y s, I mean m ho p s q and so on, so forth. And when the wood burns for example, you will get all sorts of gases like c o two c o or h two depends on the temperature of the surface. So, the species involved inside the wood surface need not be the same.

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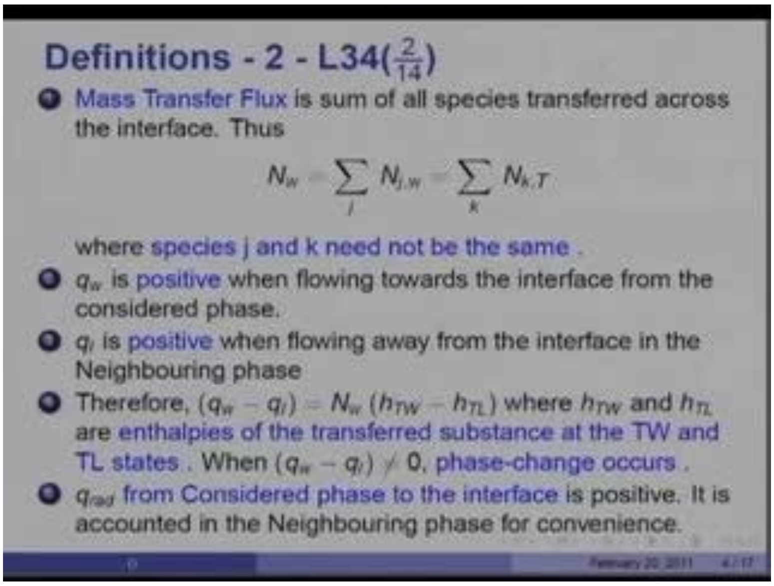

## Definitions - 2 - L34($\frac{2}{14}$)

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- $q_{rad}$ from Considered phase to the interface is positive. It is accounted in the Neighbouring phase for convenience.

As involved in the considered phase, but the sum of those fluxes of these spaces and the sum of the flux of the of the of these species should be the same. We say, q w is positive when flowing towards the interface from the considered phase.

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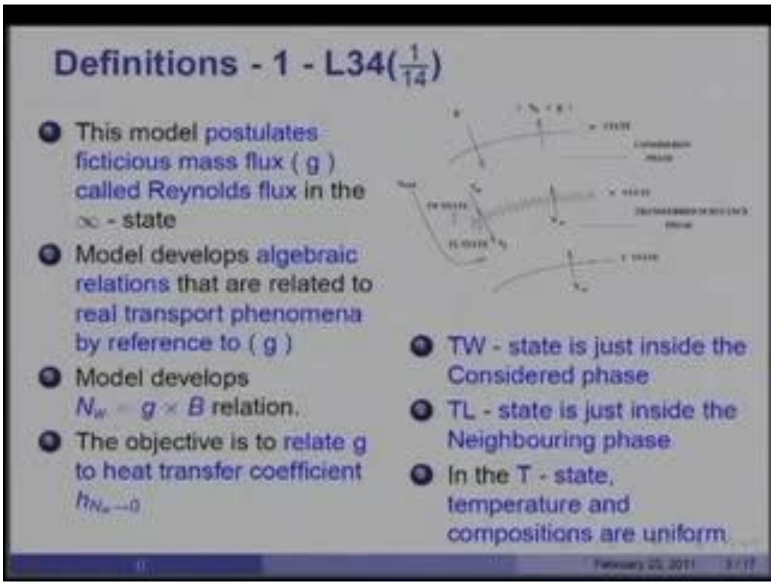

## Definitions - 1 - L34($\frac{1}{14}$)

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And q l is positive when flowing away from the interface in the neighboring phase. So in other words in this figure both q w and q l are shown in the positive direction if I now draw an interface say w w.

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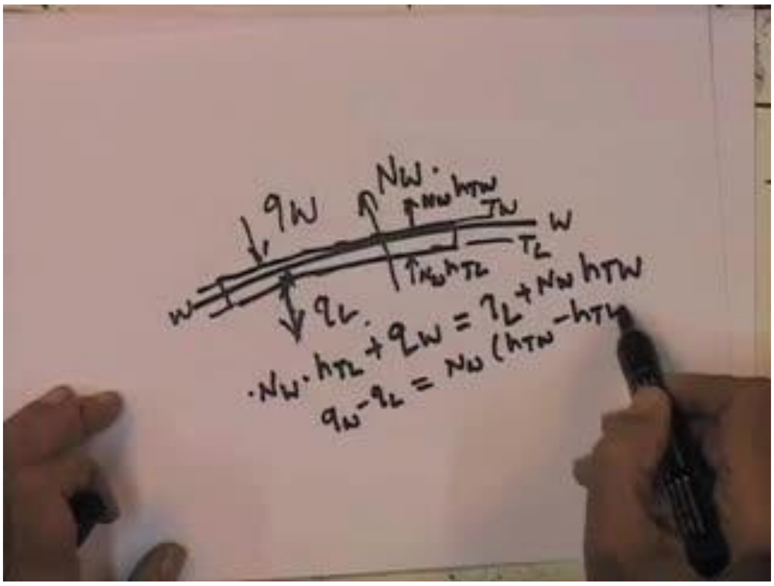

And let us say this is t w and this is t l there is q w here and q sorry q l in this direction then, the mass transfer flux is passing like this n w. So, if I take the control volume like this, then n w into h t l which is coming in. Plus q w which is also coming in must equal what is going out, which is q l plus n w into h t w. So, n w t n wt h t l comes in and what is going out is n w h t w that is what we mean.

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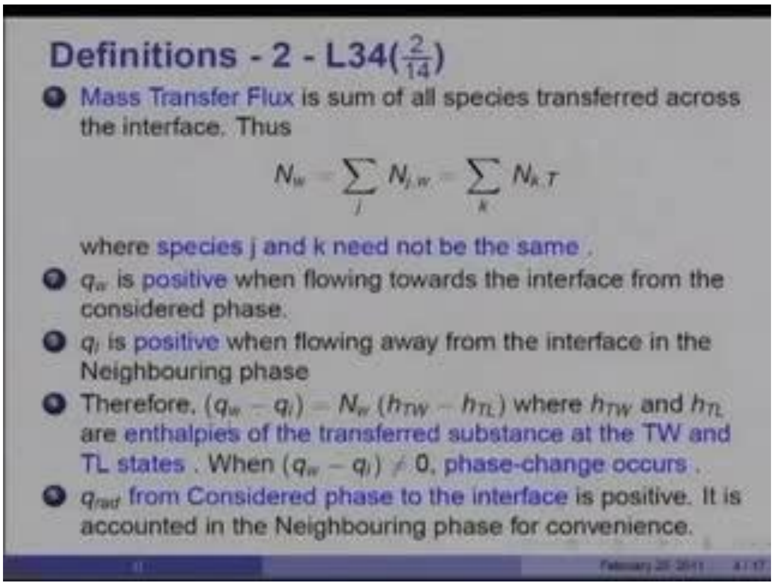

## Definitions - 2 - L34($\frac{2}{14}$)

- Mass Transfer Flux is sum of all species transferred across the interface. Thus

$$N_w = \sum_j N_{j,w} = \sum_k N_{k,T}$$

where species j and k need not be the same .
- $q_w$ is positive when flowing towards the interface from the considered phase.
- $q_l$ is positive when flowing away from the interface in the Neighbouring phase
- Therefore, $(q_w - q_l) = N_w\,(h_{TW} - h_{TL})$ where $h_{TW}$ and $h_{TL}$ are enthalpies of the transferred substance at the TW and TL states . When $(q_w - q_l) \neq 0$, phase-change occurs .
- $q_{rad}$ from Considered phase to the interface is positive. It is accounted in the Neighbouring phase for convenience.

And therefore, you will see that q w minus q l would be equal to n w into h t w minus h t l. Therefore, q w minus q l will be n w times h t w minus h t l where h t w and h t l are enthalpies of the transferred substance, at the t w and t l state. This is very very important, h t w simply implies the enthalpy of the transferred substance, if it was to appear in the considered phase as h t w. h t l on the other hand is the enthalpy of the transferred substance in the neighboring phase, but at the temperature of the interface. So, that it is what we mean by enthalpy here. So, suffix t w and t l are different from suffix w itself, suffix w refers to the mixture enthalpy, whereas suffix t w refers to the transferred substance enthalpy both in the w and l states, that is just inside the considered phase and inside the transferred substance phase difference. Now, if q w minus q l is not equal to 0 then a phase change occur. For example, if h t w minus h t l was equal to the, remember h t w is the vapor phase, for example, in an evaporation problem and h t l will be the liquid enthalpy, then h t w minus h t l would be simply h f g, the latent heat of evaporation and therefore, q w minus q l will not be equal to 0. And that is where when the phase change occurs. So, that is the idea of why we define h t w and h t l as the enthalpies of the transferred substance in the vapor in the considered phase and in the transferred substance phases.

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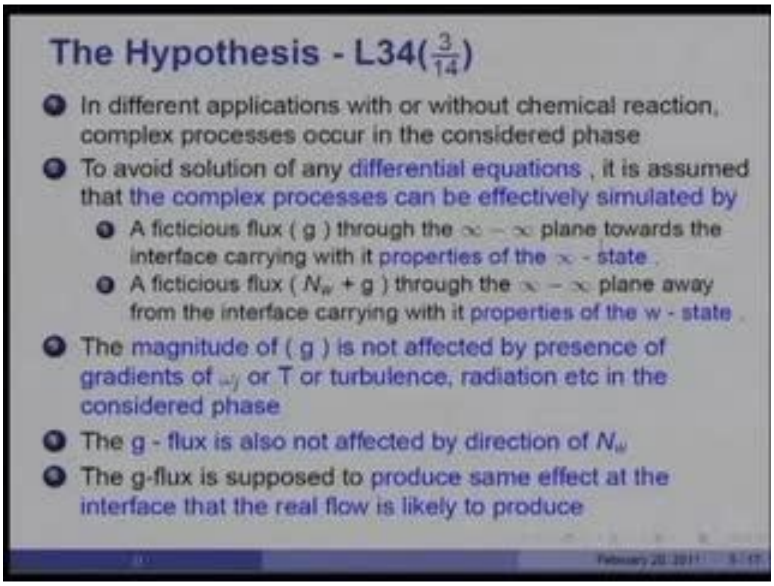

## The Hypothesis - L34($\frac{3}{14}$)

- In different applications with or without chemical reaction, complex processes occur in the considered phase
- To avoid solution of any differential equations , it is assumed that the complex processes can be effectively simulated by
  - A ficticious flux ( g ) through the $\infty - \infty$ plane towards the interface carrying with it properties of the $\infty$ - state .
  - A ficticious flux ( $N_w$ + g ) through the $\infty - \infty$ plane away from the interface carrying with it properties of the w - state .
- The magnitude of ( g ) is not affected by presence of gradients of $\omega_j$ or T or turbulence, radiation etc in the considered phase
- The g - flux is also not affected by direction of $N_w$
- The g-flux is supposed to produce same effect at the interface that the real flow is likely to produce

Q rad from the considered phase to the interface is positive, it is accounted in the neighboring phase. So, with this definitions, we now turn to the main hypothesis of the model. Now in different applications with or without chemical reaction, complex processes occur in the considered phase. That is where you will have effects of property variations, effect of turbulence and variety of other, there may be chemical reaction, there may not be any chemical reaction, so on and so forth.

This sort of situation would require solution of the complete set up boundary layer equations and what the algebraic model or the Reynolds flow model tries to do is to say, that the that to avoid solution of any differential equations, it is assumed that the complex processes can be effectively simulated by a fictitious flux g through the infinity infinity state towards the interface carrying with it properties of the infinity state. In other words, g flux carries with it the properties of the infinity state towards the interface.

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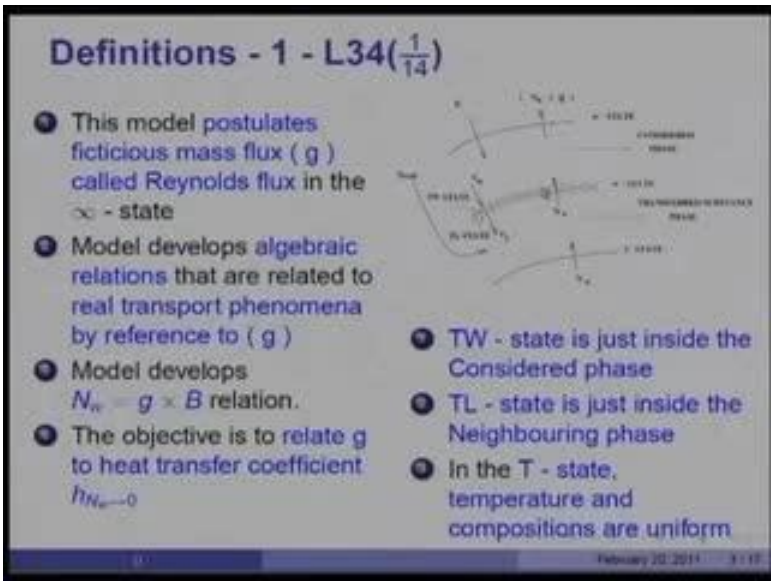

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On the other hand, n w plus g flux carries with it is again postulated at the infinity state, but it carries with it values of the w state, the upstream values if you like, likewise upstream values here in this direction. This is very important distinction between the two fluxes defined that g flux carries with it the properties of the infinite state as it moves towards the interface whereas, n w plus g moves away from the interface and carries with it the values of the w phase, w state.

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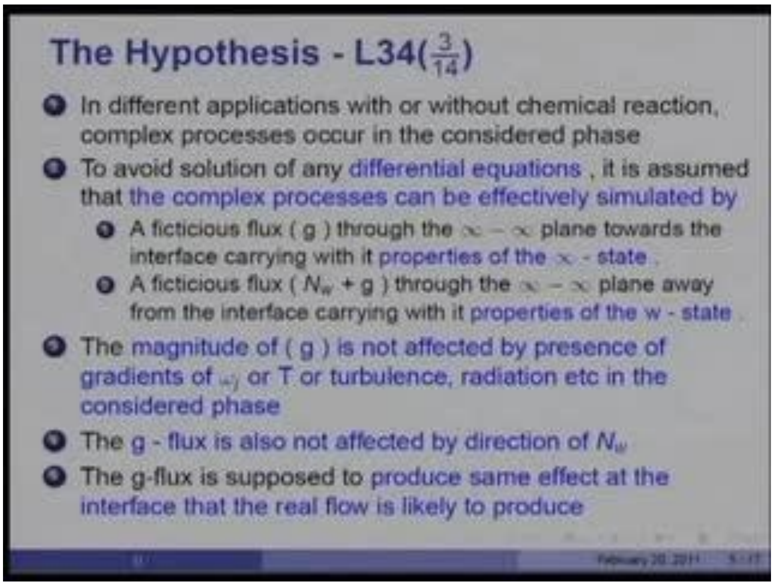

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- To avoid solution of any differential equations , it is assumed that the complex processes can be effectively simulated by
  - A ficticious flux ( g ) through the $\infty - \infty$ plane towards the interface carrying with it properties of the $\infty$ - state .
  - A ficticious flux ( $N_w$ + g ) through the $\infty - \infty$ plane away from the interface carrying with it properties of the w - state .
- The magnitude of ( g ) is not affected by presence of gradients of $\omega_j$ or T or turbulence, radiation etc in the considered phase
- The g - flux is also not affected by direction of $N_w$
- The g-flux is supposed to produce same effect at the interface that the real flow is likely to produce

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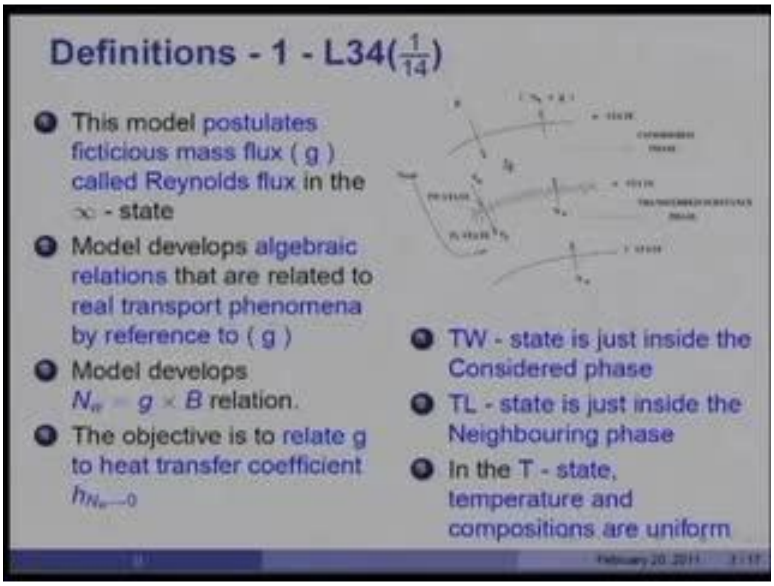

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So, that is what I have said a fictitious flux n w plus g through infinity infinity plane away from the interface carrying with it properties of the w state. Now what this ensures is that in the considered phase, there will be no net generation of mass at all because g is coming in and g is going out.

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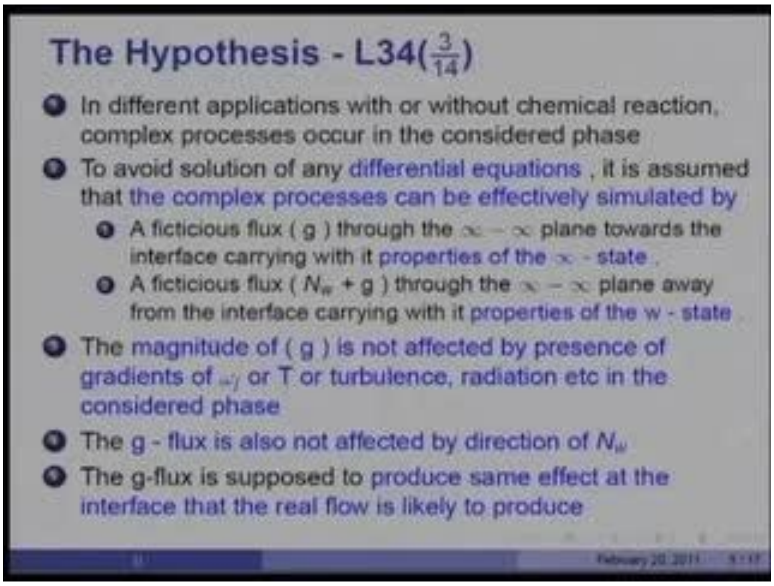

### The Hypothesis - L34($\frac{3}{14}$)

- In different applications with or without chemical reaction, complex processes occur in the considered phase
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- The magnitude of ( g ) is not affected by presence of gradients of $\omega_j$ or T or turbulence, radiation etc in the considered phase
- The g - flux is also not affected by direction of $N_w$
- The g-flux is supposed to produce same effect at the interface that the real flow is likely to produce

So, there is no net generation of mass and that is what is expected even in a real boundary layer flow, there should be no net generation of mass. The magnitude of g is not affected by the presence of gradients of omega j, T or turbulence, radiation etcetera in the considered phase. So, that is the another characteristics of the g g flux.

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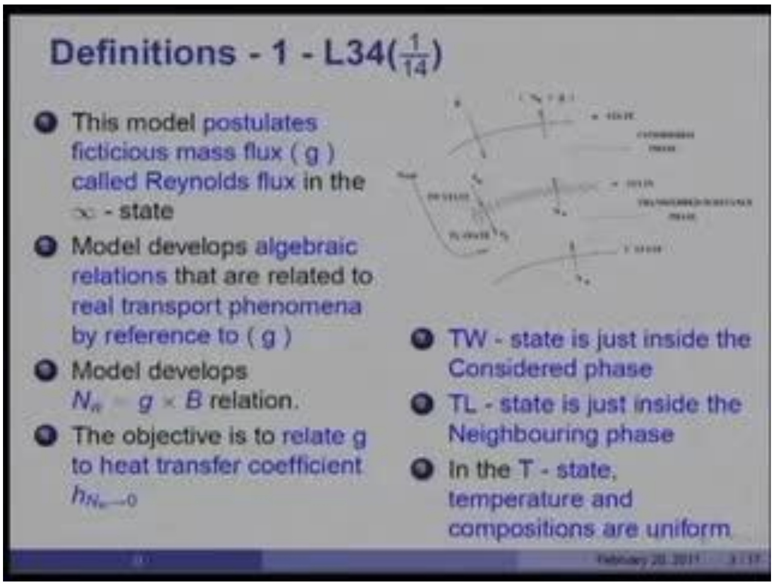

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It is also not affected by the direction of n w, in other words, I may have a problem of evaporation in which n w is this way, I may have a problem in which condensation occurs in which the n w is negative. So, g is not influenced by presence of, by the direction or magnitude of n w and secondly, it is a not affected by gradients of omega j, temperature or turbulence quantities in the considered phase and yet in spite of these assumptions made the g flux is suppose to produce the same effect at the interface that the real boundary layer flow is likely to produce.

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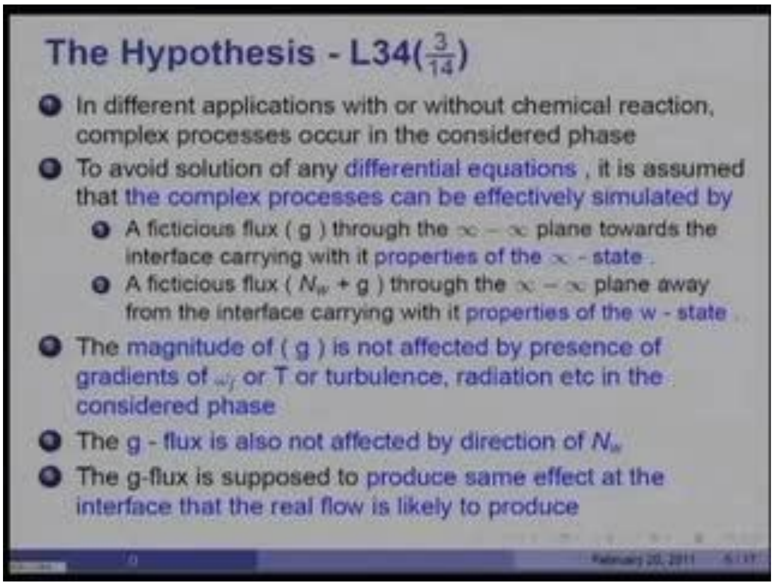

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- The magnitude of ( g ) is not affected by presence of gradients of $\omega_j$ or T or turbulence, radiation etc in the considered phase
- The g - flux is also not affected by direction of $N_w$
- The g-flux is supposed to produce same effect at the interface that the real flow is likely to produce

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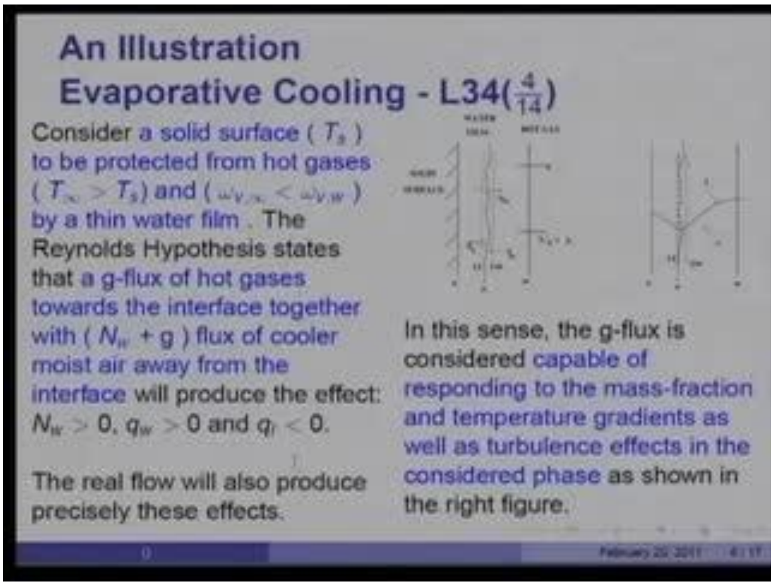

## An Illustration
## Evaporative Cooling - L34($\frac{4}{14}$)

Consider a solid surface ( $T_s$ ) to be protected from hot gases ( $T_\infty > T_s$) and ( $\omega_{v,\infty} < \omega_{v,w}$ ) by a thin water film . The Reynolds Hypothesis states that a g-flux of hot gases towards the interface together with ( $N_w$ + g ) flux of cooler moist air away from the interface will produce the effect: $N_w > 0$, $q_w > 0$ and $q_l < 0$.

The real flow will also produce precisely these effects.

In this sense, the g-flux is considered capable of responding to the mass-fraction and temperature gradients as well as turbulence effects in the considered phase as shown in the right figure.

This is the claim of the model, this is the claim of the hypothesis that it can, in fact, represent all effects of a real boundary layer flow, although the method is very very simple and algebraic. To illustrate this, let me consider the case of evaporative cooling. So, for example, let me say, that I have a very hot surface here at temperature t s and it is hot because very hot gases are flowing over it in the infinity state, very hot gases are flowing and they are dried well before humidity.

Obviously, to protect this surface from the hot gases and to keep its temperature down what one is interest, what one might do is to provide let us say, a water film here, pass a thin film of water then what it will do is that the water film will evaporate. The water film will evaporate and in the process we will receive heat both from the solid surface as well as from the hot gases. So, this is the likely temperature profile. Remember in the absence of the film probably the temperature would have been much higher than what is shown here.

So, in order to cool down, but the liq[uid]- at the liquid film the temperature will drop considerably and that would result into actually sum transfer of heat from through the liquid film from the wall to the film to the edge of the film and likewise there will be heat transfer. So, q w in this case would be positive, but q l which is which would be now in the from the transferred substance phase to the interface would be negative, q l would be negative and n w would be positive.

So, what we expect therefore, the Reynolds flux hypothesis states, that a g flux of hot gases towards the interface together with the n w plus g fluxes of cooler moist air away from the interface, away from the interface hot gases towards interface will produce the effect that n w is greater than 0, q w is also greater than 0 because it flows into the surface and q l will be negative because it is also flowing into the surface from the neighboring phase. Now, this is what we expect the real flow to do.

And the Reynolds flux model claims that it will do precisely that, that it will do precisely that. So, in this sense the g flux is considered capable of responding to the mass-fraction and temperature gradients as well as turbulence effects in the considered phase as shown in this figure on this side. So, it disregards what the temperature gradients here are, what the temperature gradients here are and so on, so forth, or the mass fraction gradients here are, we can remember mass fraction gradient would fall, it disregards any turbulence effects in the considered phase and will produce exactly the same effect, n w greater than 0, q w greater than 0 and q l less than 0. That is the claim of the model. Let us see how we can justify that.

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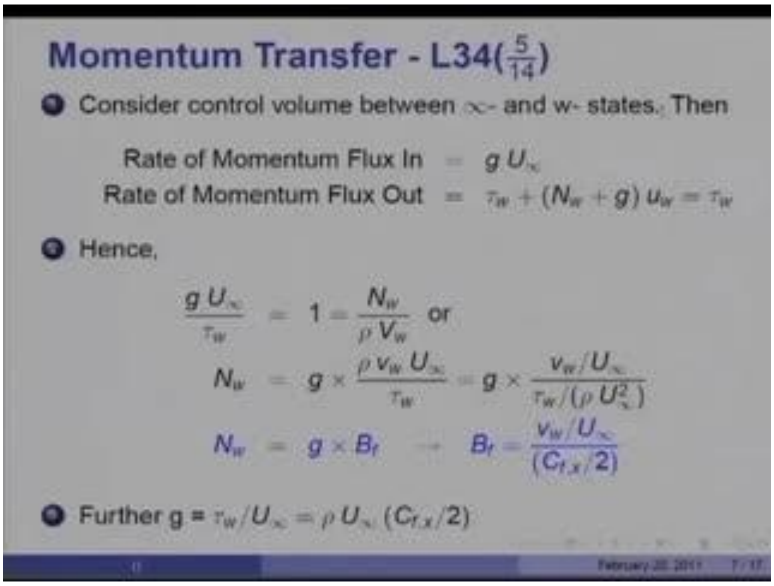

**Momentum Transfer - L34($\frac{5}{14}$)**

- Consider control volume between $\infty$- and w- states. Then

Rate of Momentum Flux In $= g\,U_\infty$
Rate of Momentum Flux Out $= \tau_w + (N_w + g)\,u_w = \tau_w$

- Hence,

$$\frac{g\,U_\infty}{\tau_w} = 1 = \frac{N_w}{\rho\,V_w} \quad \text{or}$$

$$N_w = g \times \frac{\rho\,v_w\,U_\infty}{\tau_w} = g \times \frac{v_w/U_\infty}{\tau_w/(\rho\,U_\infty^2)}$$

$$N_w = g \times B_f \rightarrow B_f = \frac{v_w/U_\infty}{(C_{f,x}/2)}$$

- Further $g = \tau_w/U_\infty = \rho\,U_\infty\,(C_{f,x}/2)$

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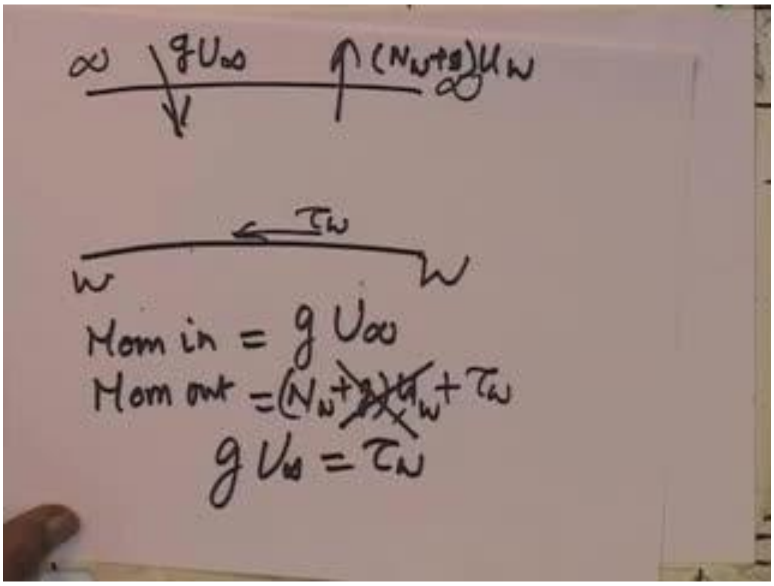

Because it can handle both momentum and heat transfer we are going to consider the control volume between infinities and w states. So, basically what I am going do now is to consider the w state and infinity state and I am going to consider momentum transfer, then the g flux will bring with it g times u infinity. So, momentum in will be g times u infinity, momentum out will be n w plus g u u w n w plus g u at w plus shear stress, tau w, that is also momentum out of the of the cell. But at the surface u w will be 0 therefore, that goes out. So, basically if I quit the two, I get g u infinity equal to tau w that is what the Reynolds flow model implies. So, let us see what does this mean.

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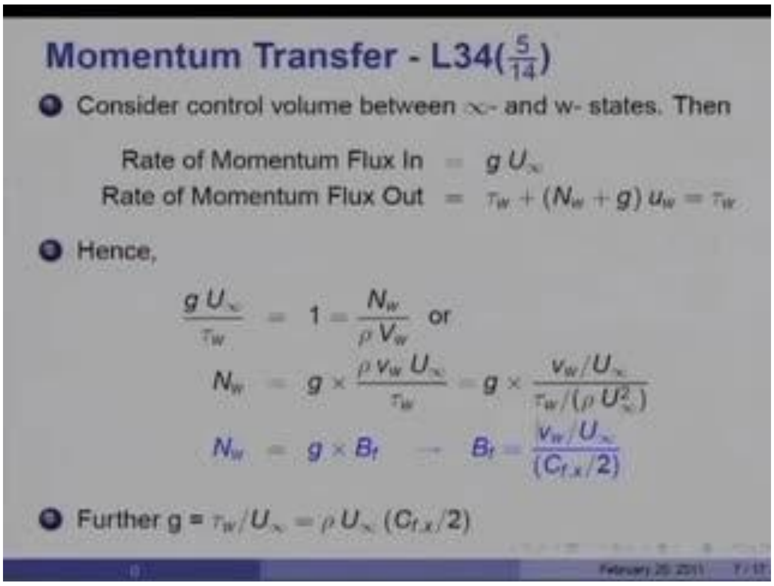

**Momentum Transfer - L34($\frac{5}{14}$)**

- Consider control volume between $\infty$- and w- states. Then

$$\text{Rate of Momentum Flux In} = g\,U_\infty$$
$$\text{Rate of Momentum Flux Out} = \tau_w + (N_w + g)\,u_w = \tau_w$$

- Hence,

$$\frac{g\,U_\infty}{\tau_w} = 1 = \frac{N_w}{\rho\,V_w} \quad \text{or}$$
$$N_w = g \times \frac{\rho\,v_w\,U_\infty}{\tau_w} = g \times \frac{v_w/U_\infty}{\tau_w/(\rho\,U_\infty^2)}$$
$$N_w = g \times B_f \rightarrow B_f = \frac{v_w/U_\infty}{(C_{f,x}/2)}$$

- Further $g = \tau_w/U_\infty = \rho\,U_\infty\,(C_{f,x}/2)$

Well. So, basically as you will see here, I have said n w plus g u w equal to 0 and therefore, g u infinity divided by tau wall would equal 1 and n w divided by rho times v w yes. So, what I have done here is, 1 is equal to n w itself is a rho v w. So, I am replacing 1 by n w by v w or n w can be written as g times rho v w u infinity by tau wall and that would be equal to g into vv w divided by u infinity tau wall over rho u infinity square. Essentially I have divided and multiplied by u infinity and what is this quantity this is essentially v w by u infinity divided by c of x by 2 and therefore, n w is simply g times b f, our claim that the model is able to predict all effects of momentum transfer through this expression n w equal to g b is verified as long as we interpret b in this fashion, vf equal to v w u infinity divided by c f x two. So, in order to get g, what will be g, the meaning of g will be tau wall over rho u infinity which will be rho u infinity c f x by two. So, in the momentum transfer problem, g would be simply rho u infinity c f x by two and c f x by two is available to us from correlations for many many problems.

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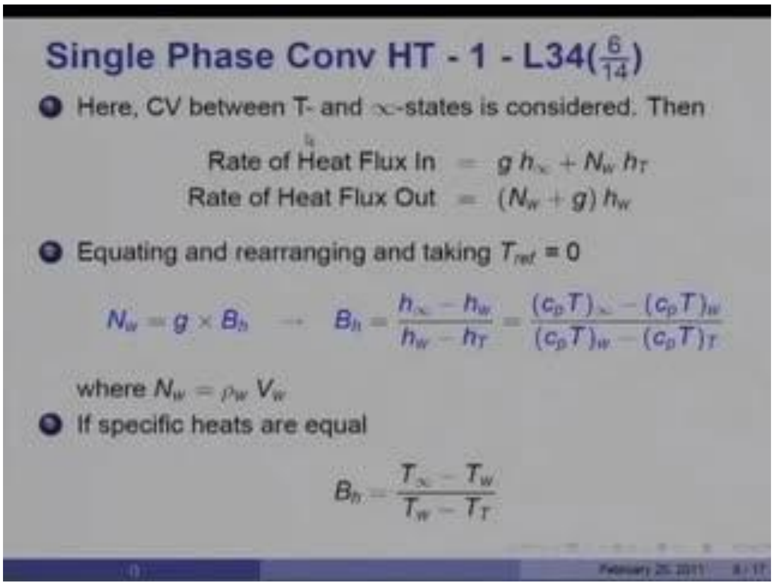

## Single Phase Conv HT - 1 - L34($\frac{6}{14}$)

- Here, CV between T- and $\infty$-states is considered. Then

$$\text{Rate of Heat Flux In} = g\,h_\infty + N_w\,h_T$$
$$\text{Rate of Heat Flux Out} = (N_w + g)\,h_w$$

- Equating and rearranging and taking $T_{ref} = 0$

$$N_w = g \times B_h \rightarrow B_h = \frac{h_\infty - h_w}{h_w - h_T} = \frac{(c_pT)_\infty - (c_pT)_w}{(c_pT)_w - (c_pT)_T}$$

where $N_w = \rho_w V_w$

- If specific heats are equal

$$B_h = \frac{T_\infty - T_w}{T_w - T_T}$$

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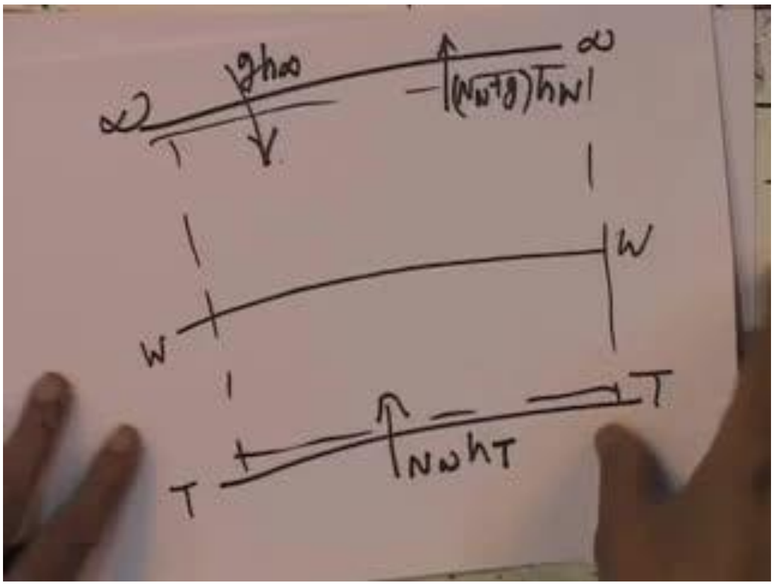

Let us look at single phase convective heat transfer in which they consider the c v between t and infinity state. Now what I am going to do is to consider the, this is the w state, this is the t state and this is the infinity infinity state, then you will see that heat in will be g times h infinity, there will be n plus w n w n w plus g hw and there would be here n w times h t.

This is the. So, I am considering the control volume between infinity infinity and t t states. So, there is a it will bring with it g u h infinity at the enthalp[y]-total [en/energy]

energy flux, it there would be n w into h t will be the total [en/energy] energy flux coming in here also and here and what will be going out is n w plus g h w, and that is going out. So, if I equate what is coming in and what is going out then you will see that I get this relationship, rate of heat flux in is equal to g h infinity plus n w h t plus n w plus g divided by, multiplied by h w.

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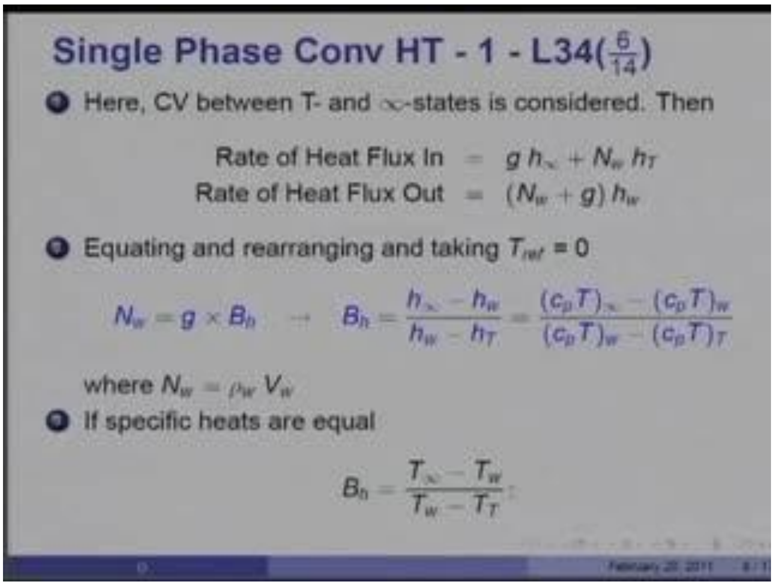

**Single Phase Conv HT - 1 - L34($\frac{6}{14}$)**

- Here, CV between T- and $\infty$-states is considered. Then

$$\text{Rate of Heat Flux In} = g\,h_\infty + N_w\,h_T$$
$$\text{Rate of Heat Flux Out} = (N_w + g)\,h_w$$

- Equating and rearranging and taking $T_{ref} = 0$

$$N_w = g \times B_h \rightarrow B_h = \frac{h_\infty - h_w}{h_w - h_T} = \frac{(c_p T)_\infty - (c_p T)_w}{(c_p T)_w - (c_p T)_T}$$

where $N_w = \rho_w V_w$

- If specific heats are equal

$$B_h = \frac{T_\infty - T_w}{T_w - T_T}$$

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So, if I equate and rearrange and then again you will see, I get n w equal to g times b h where b h is equal to h infinity minus h w divided by h w minus h t or h infinity in this case, because it is a single phase mass transfer, single phase convective heat transfer with suction and blowing may be, this will be simply c p into t in the infinity state minus c p into t into w state divided by c p t w state minus c p t in the transferred substance phase and therefore, n w will be equal to n w v w, if the specific heats are equal of course, b h would be simply t infinity minus t w divided by t w minus t t. If I now consider infinity infinity state n w states, say if I now consider those states, then you will see here you will get n w h w.

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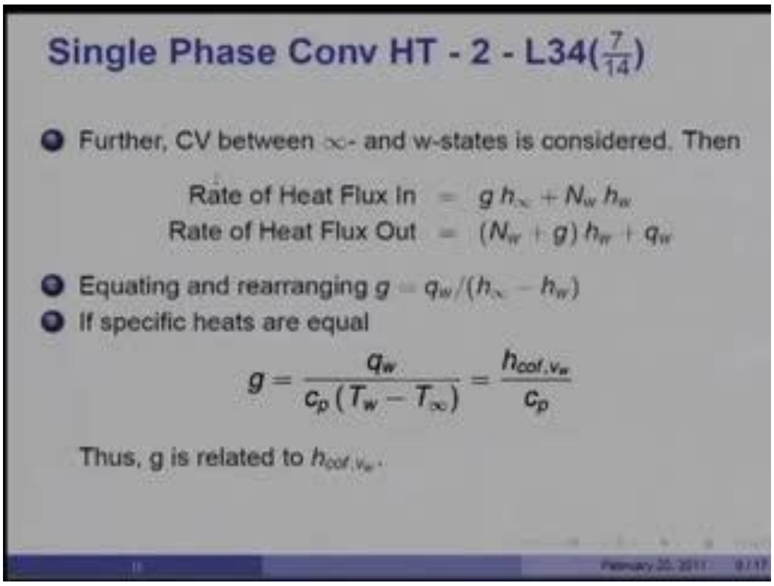

## Single Phase Conv HT - 2 - L34($\frac{7}{14}$)

- Further, CV between $\infty$- and w-states is considered. Then

$$\text{Rate of Heat Flux In} = g\,h_\infty + N_w\,h_w$$
$$\text{Rate of Heat Flux Out} = (N_w + g)\,h_w + q_w$$

- Equating and rearranging $g = q_w/(h_\infty - h_w)$
- If specific heats are equal

$$g = \frac{q_w}{c_p\,(T_w - T_\infty)} = \frac{h_{cof,v_w}}{c_p}$$

Thus, g is related to $h_{cof,v_w}$.

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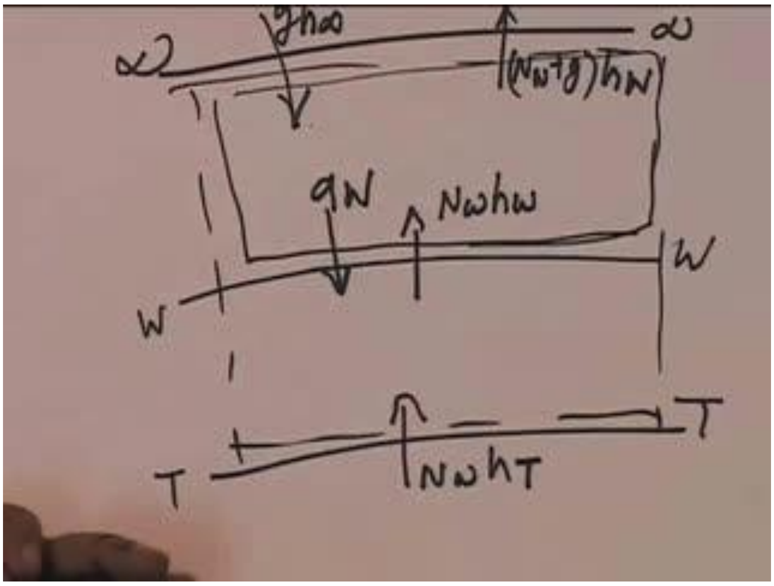

So, my equations would now read as g h infinity coming in, n w h w coming in and n w plus g h w going out. So, that is what I have written, plus there would be q w also which will go out between in this control volume, for this control volume.

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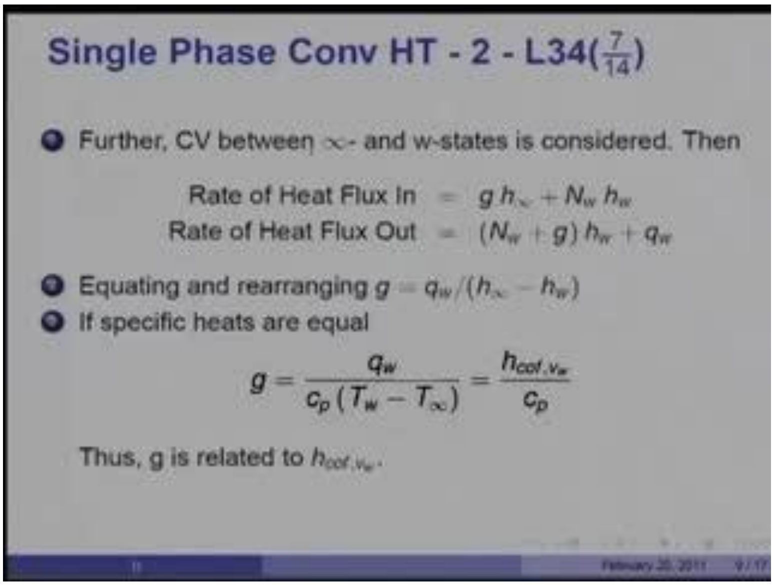

## Single Phase Conv HT - 2 - L34($\frac{7}{14}$)

- Further, CV between $\infty$- and w-states is considered. Then

$$\begin{aligned}\text{Rate of Heat Flux In} &= g\, h_\infty + N_w\, h_w \\ \text{Rate of Heat Flux Out} &= (N_w + g)\, h_w + q_w\end{aligned}$$

- Equating and rearranging $g = q_w/(h_\infty - h_w)$
- If specific heats are equal

$$g = \frac{q_w}{c_p\,(T_w - T_\infty)} = \frac{h_{cof,v_w}}{c_p}$$

Thus, g is related to $h_{cof,v_w}$.

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So, you will see that I get rate of heat flux in as g h infinity plus n w h w, rate of heat flux out will be n w g h w plus q w and if I equate and rearrange then I will get g is equal to h times, the g q w divided by h infinity minus h w. And if specific heats are equal in the infinity and w states, then g would be simply q wall divided by c p t infinity minus t w and as you know, this would be simply minus h cof v w by c p.

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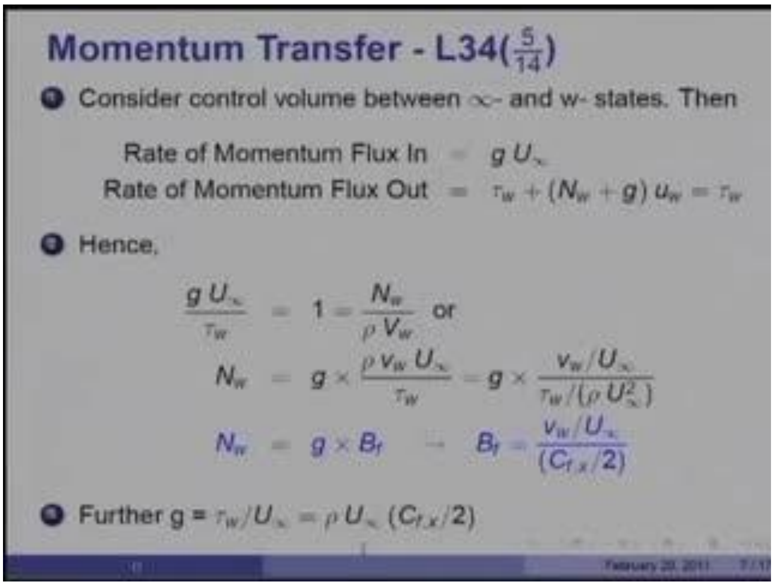

## Momentum Transfer - L34($\frac{5}{14}$)

- Consider control volume between $\infty$- and w- states. Then

$$\begin{aligned}\text{Rate of Momentum Flux In} &= g\, U_\infty \\ \text{Rate of Momentum Flux Out} &= \tau_w + (N_w + g)\, u_w = \tau_w\end{aligned}$$

- Hence,

$$\begin{aligned}\frac{g\, U_\infty}{\tau_w} &= 1 = \frac{N_w}{\rho\, V_w} \quad \text{or} \\ N_w &= g \times \frac{\rho\, v_w\, U_\infty}{\tau_w} = g \times \frac{v_w/U_\infty}{\tau_w/(\rho\, U_\infty^2)} \\ N_w &= g \times B_f \quad \rightarrow \quad B_f = \frac{v_w/U_\infty}{(C_{f,x}/2)}\end{aligned}$$

- Further $g = \tau_w/U_\infty = \rho\, U_\infty\, (C_{f,x}/2)$

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So, therefore, g is indeed related to heat transfer coefficient in the presence of suction and blowing. So, that is what the single phase convective heat transfer shows, convective momentum transfer shows that g is equal to rho u infinity c f x by 2 and here h cof is equal to.

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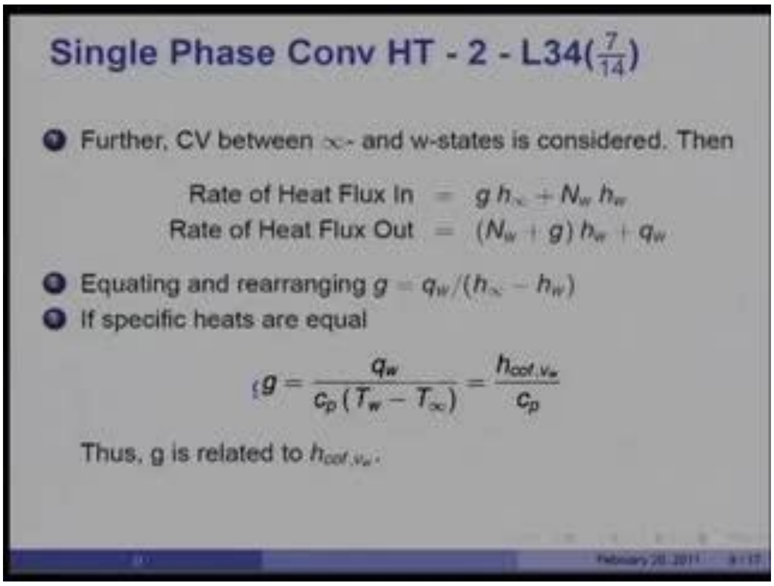

## Single Phase Conv HT - 2 - L34($\frac{7}{14}$)

- Further, CV between $\infty$- and w-states is considered. Then

$$\text{Rate of Heat Flux In} = g\,h_\infty + N_w\,h_w$$
$$\text{Rate of Heat Flux Out} = (N_w + g)\,h_w + q_w$$

- Equating and rearranging $g = q_w/(h_\infty - h_w)$
- If specific heats are equal

$$g = \frac{q_w}{c_p\,(T_w - T_\infty)} = \frac{h_{cof,v_w}}{c_p}$$

Thus, g is related to $h_{cof,v_w}$.

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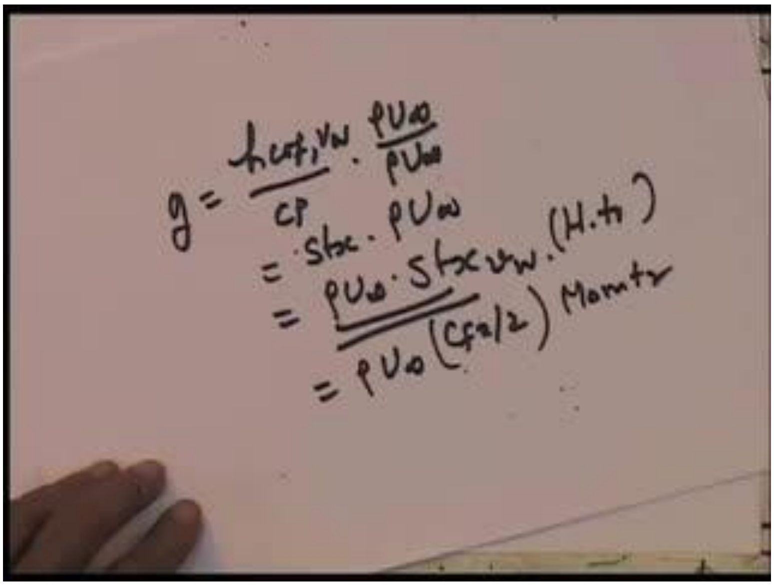

$$g = \frac{h_{cof,v_w}}{cp} \cdot \frac{\rho U_\infty}{\rho U_\infty}$$
$$= St_x \cdot \rho U_\infty$$
$$= \underline{\underline{\rho U_\infty \cdot St_{x\,v_w}}} \cdot (\text{H.tr})$$
$$= \rho U_\infty \left(Cf_x/2\right) \text{Mom tr}$$

Now you can see very well that if I define, if I, since g is equal to minus h cof v w divided by c p, if I divide and multiply this by rho u infinity rho u infinity, then you will see this will be stanton x into rho u infinity. So, rho u infinity stanton x for v w, analogous to rho u infinity c f x by 2, that was from momentum transfer and this is for heat transfer, very similar formulas which essentially implies that the analogy between heat transfer and momentum transfer through the coefficient g, the convective mass transfer coefficient.

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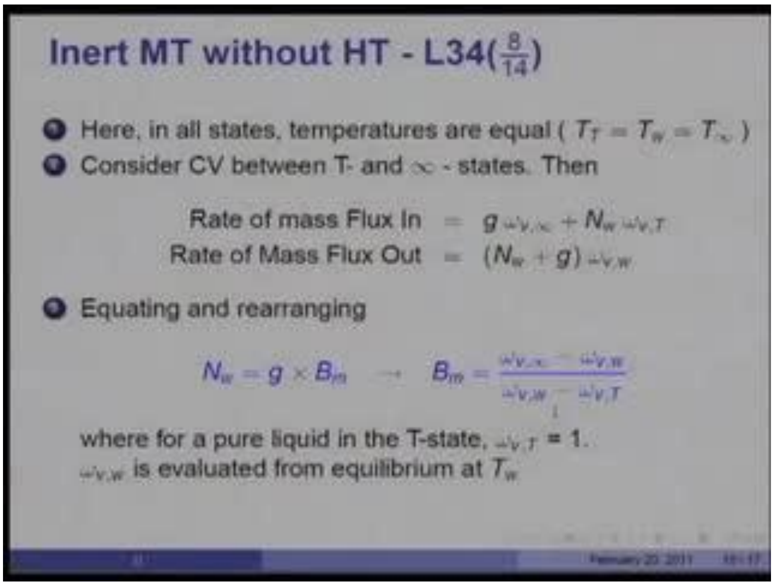

**Inert MT without HT - L34($\frac{8}{14}$)**

- Here, in all states, temperatures are equal ( $T_T = T_w = T_\infty$ )
- Consider CV between T- and $\infty$ - states. Then

$$\text{Rate of mass Flux In} = g\,\omega_{v,\infty} + N_w\,\omega_{v,T}$$
$$\text{Rate of Mass Flux Out} = (N_w + g)\,\omega_{v,w}$$

- Equating and rearranging

$$N_w = g \times B_m \rightarrow B_m = \frac{\omega_{v,\infty} - \omega_{v,w}}{\omega_{v,w} - \omega_{v,T}}$$

where for a pure liquid in the T-state, $\omega_{v,T} = 1$.
$\omega_{v,w}$ is evaluated from equilibrium at $T_w$

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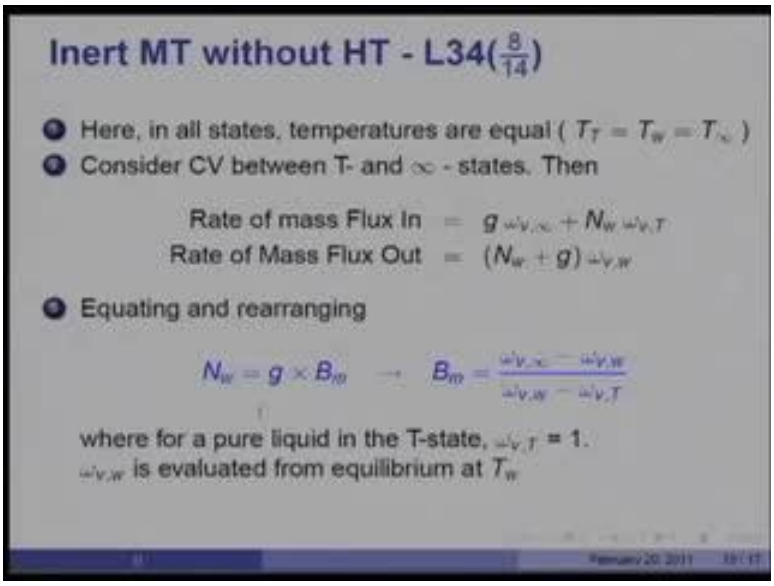

**Inert MT without HT - L34($\frac{8}{14}$)**

- Here, in all states, temperatures are equal ( $T_T = T_w = T_\infty$ )
- Consider CV between T- and $\infty$ - states. Then

$$\text{Rate of mass Flux In} = g\,\omega_{v,\infty} + N_w\,\omega_{v,T}$$
$$\text{Rate of Mass Flux Out} = (N_w + g)\,\omega_{v,w}$$

- Equating and rearranging

$$N_w = g \times B_m \rightarrow B_m = \frac{\omega_{v,\infty} - \omega_{v,w}}{\omega_{v,w} - \omega_{v,T}}$$

where for a pure liquid in the T-state, $\omega_{v,T} = 1$.
$\omega_{v,w}$ is evaluated from equilibrium at $T_w$

So, inert mass transfer without heat transfer, so, now we consider, now let us say, in all states the temperatures are equal. So, there is no heat transfer at all, but there would be mass transfer. So, again if I consider t control volume between t v t state and infinity state, then there will be rate of mass flux in will be g v omega v infinity plus n w into omega v t which is from the neighboring phase and the transferred substance rate and what would go out would be n w plus g omega v w, because the outgoing flux carries with it properties of the w state, again if I rearrange, equate in and out, then you will get n w equal to g b m, where b m now is omega v infinity minus omega v w omega v w minus omega v t and for a pure liquid in the t state omega v t will be equal to 1.

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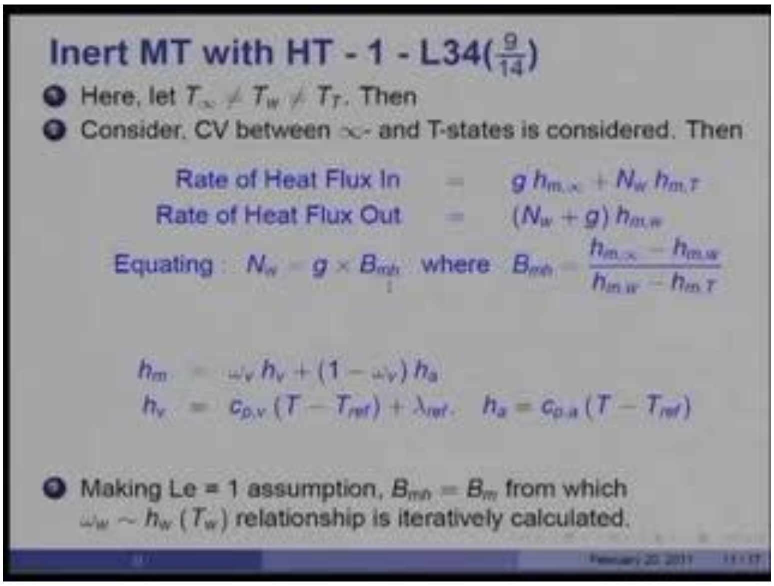

## Inert MT with HT - 1 - L34($\frac{9}{14}$)

- Here, let $T_\infty \neq T_w \neq T_T$. Then
- Consider, CV between $\infty$- and T-states is considered. Then

$$\text{Rate of Heat Flux In} = g\, h_{m,\infty} + N_w\, h_{m,T}$$

$$\text{Rate of Heat Flux Out} = (N_w + g)\, h_{m,w}$$

$$\text{Equating : } N_w = g \times B_{mh} \quad \text{where} \quad B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,T}}$$

$$h_m = \omega_v\, h_v + (1 - \omega_v)\, h_a$$

$$h_v = c_{p,v}\,(T - T_{ref}) + \lambda_{ref}, \quad h_a = c_{p,a}\,(T - T_{ref})$$

- Making Le = 1 assumption, $B_{mh} = B_m$ from which $\omega_w \sim h_w\,(T_w)$ relationship is iteratively calculated.

And how would we evaluate omega v w? It would be evaluated from equilibrium relation at t w, yes, this is what the Reynolds flow model shows, even for inert mass transfer without heat transfer. Now let us consider inert mass transfer with heat transfer. Now, in this case, let us say infinity state is not temperature is not equal to t w and not equal to t t, then the c v between infinity and t states will show that the rate of heat flux in will be g times h m infinity plus n w h m t and what would go out will be n w plus g into h m w, again equating you get n w equal to g b m h, where b m h is equal to h m infinity minus h m w divided by h m w minus h m t. In an inert mass transfer case, let us say, air and water vapor mixture, then h m would be omega v times enthalpy of the vapor and omega a times which is the mass fraction of a into enthalpy of a and since there are only 2 species omega a, I would be written as 1 minus omega v and enthalpy of the vapor would be c p v t minus t f plus lambda ref which is the latent heat defined at t ref and h a would have simply sensible heat c p a t minus t ref.

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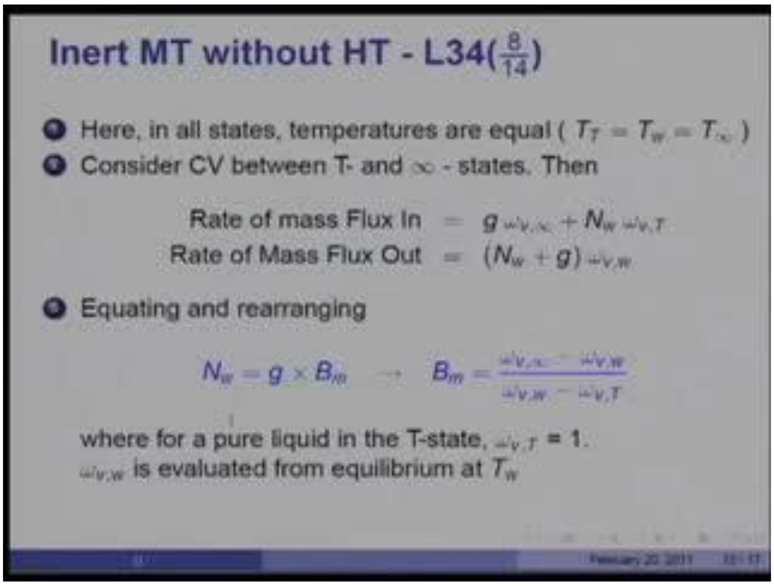

## Inert MT without HT - L34($\frac{8}{14}$)

- Here, in all states, temperatures are equal ( $T_T = T_w = T_\infty$ )
- Consider CV between T- and $\infty$ - states. Then

$$\text{Rate of mass Flux In} = g\,\omega_{v,\infty} + N_w\,\omega_{v,T}$$
$$\text{Rate of Mass Flux Out} = (N_w + g)\,\omega_{v,w}$$

- Equating and rearranging

$$N_w = g \times B_m \rightarrow B_m = \frac{\omega_{v,\infty} - \omega_{v,w}}{\omega_{v,w} - \omega_{v,T}}$$

where for a pure liquid in the T-state, $\omega_{v,T} = 1$.
$\omega_{v,w}$ is evaluated from equilibrium at $T_w$

As you recall if you make the Louis number equal to 1 assumption then b m h would be equal to b m from which omega v h w t w relation is iteratively calculated. So, we can since n w is equal to g b m h and n w is equal to g b m.

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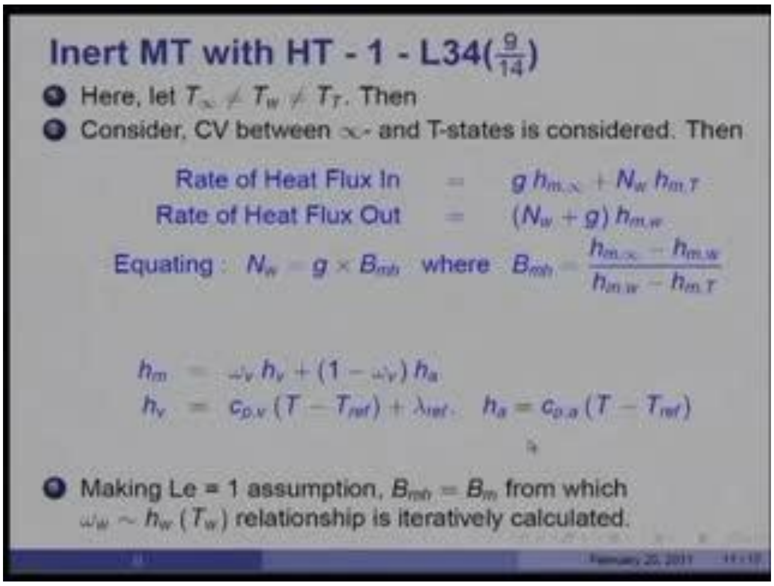

## Inert MT with HT - 1 - L34($\frac{9}{14}$)

- Here, let $T_\infty \neq T_w \neq T_T$. Then
- Consider, CV between $\infty$- and T-states is considered. Then

$$\text{Rate of Heat Flux In} = g\,h_{m,\infty} + N_w\,h_{m,T}$$
$$\text{Rate of Heat Flux Out} = (N_w + g)\,h_{m,w}$$

$$\text{Equating : } N_w = g \times B_{mh} \text{ where } B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,T}}$$

$$h_m = \omega_v\,h_v + (1 - \omega_v)\,h_a$$
$$h_v = c_{p,v}\,(T - T_{ref}) + \lambda_{ref}, \quad h_a = c_{p,a}\,(T - T_{ref})$$

- Making Le = 1 assumption, $B_{mh} = B_m$ from which $\omega_w \sim h_w\,(T_w)$ relationship is iteratively calculated.

So, b m is equal to b m h and we can calculate the, in case, unknown t w state and omega w states can be evaluated from the equilibrium relationship iteratively. This is exactly what we did even in Stefan flow and Couette flow models. Now, a few more interesting results.

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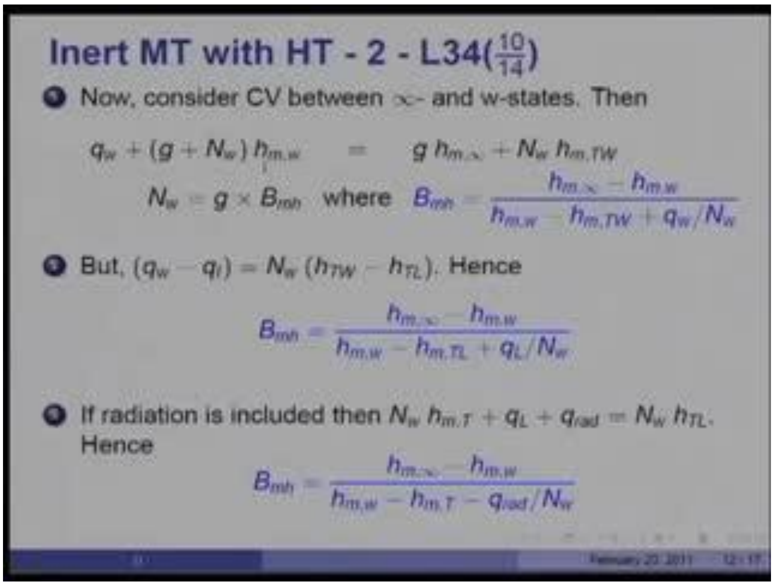

## Inert MT with HT - 2 - L34($\frac{10}{14}$)

- Now, consider CV between $\infty$- and w-states. Then

$$q_w + (g + N_w)\, h_{m,w} = g\, h_{m,\infty} + N_w\, h_{m,TW}$$

$$N_w = g \times B_{mh} \quad \text{where} \quad B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TW} + q_w/N_w}$$

- But, $(q_w - q_l) = N_w\,(h_{TW} - h_{TL})$. Hence

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TL} + q_L/N_w}$$

- If radiation is included then $N_w\, h_{m,T} + q_L + q_{rad} = N_w\, h_{TL}$. Hence

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,T} - q_{rad}/N_w}$$

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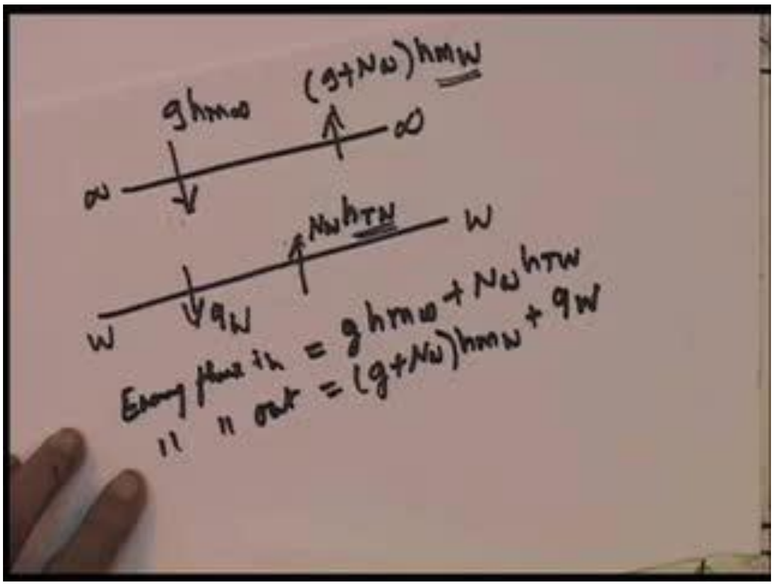

Energy flux in $= g h_{m\infty} + N_w h_{TW}$

" " out $= (g+N_w)h_{mW} + q_W$

If I consider the c v between w state infinity state and w w state and write, then you know this is q w, this is, here I will get n w times h t w. This will be g times h m infinity and there would be g times plus n w into h m w. You can see now the difference between h m w and h t w, h t w is the, is the enthalpy of the substance, which has been transferred from the neighboring phase into the interface then therefore, if I energy in will be equal to g h m infinity plus n w h t w and energy flux out will be equal to g plus n w h m w plus q w plus q w and that is what I have written here.

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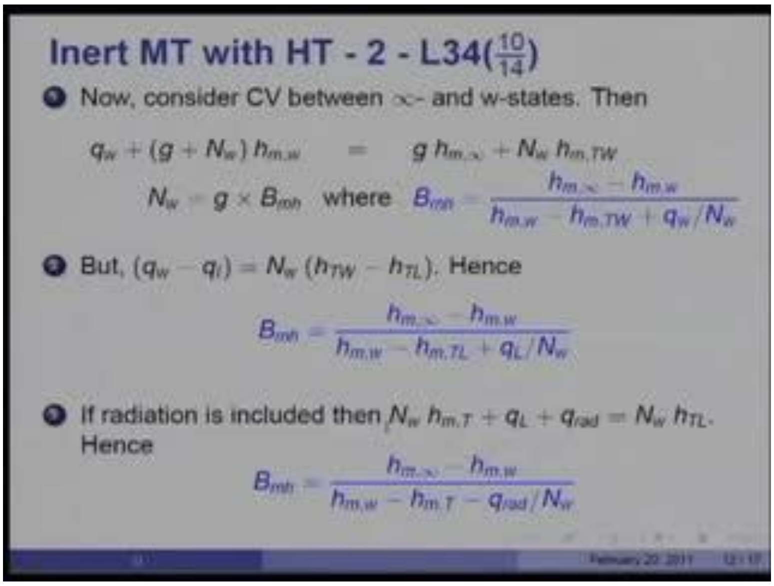

## Inert MT with HT - 2 - L34($\frac{10}{14}$)

- Now, consider CV between $\infty$- and w-states. Then

$$q_w + (g + N_w)\, h_{m,w} \;=\; g\, h_{m,\infty} + N_w\, h_{m,TW}$$

$$N_w = g \times B_{mh} \quad \text{where} \quad B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TW} + q_w/N_w}$$

- But, $(q_w - q_l) = N_w\,(h_{TW} - h_{TL})$. Hence

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TL} + q_L/N_w}$$

- If radiation is included then $N_w\, h_{m,T} + q_L + q_{rad} = N_w\, h_{TL}$. Hence

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,T} - q_{rad}/N_w}$$

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So, on the left hand side, you have the fluxes going out and on the right hand side, you have the fluxes going in and therefore, if I now rearrange these equals equality then you will see b m h would now get defined as h m infinity minus h m w, h m w minus h t w plus q wall by n w. But you will recall we showed that q w minus q l where q l is the heat transfer in the transferred substance phase equal to n w t w minus h t l. So, I can replace q w from this relationship and I get another expression for b m h in terms of q liquid divided by n w.

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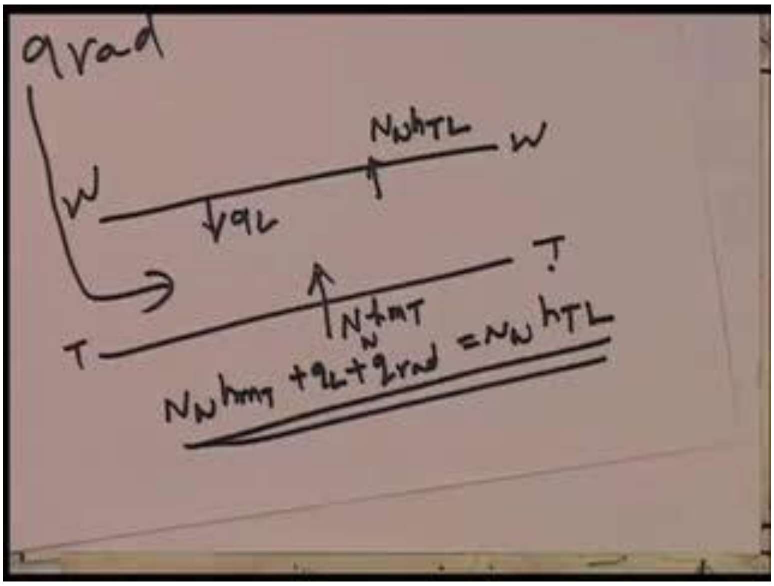

On the other hand, if I also consider the radiation in the transferred substance phase, then you will see that I get, if I consider radiation in the transferred substance phase, then this is the w w surface, this is the t t surface. Then q l is so, q rad is so, this is n times h m t and here you have n w times h t l. So, I would get essentially n w h m t plus q l plus q rad equal to n w h t l. This would be the energy balance between w and t surfaces in the presence of radiation and then you will see that I can even replace q l divided by n w h m t l by minus h m t minus q rad by n w. So, variety of expressions for b m h can be found when inert mass transfer with heat transfer.

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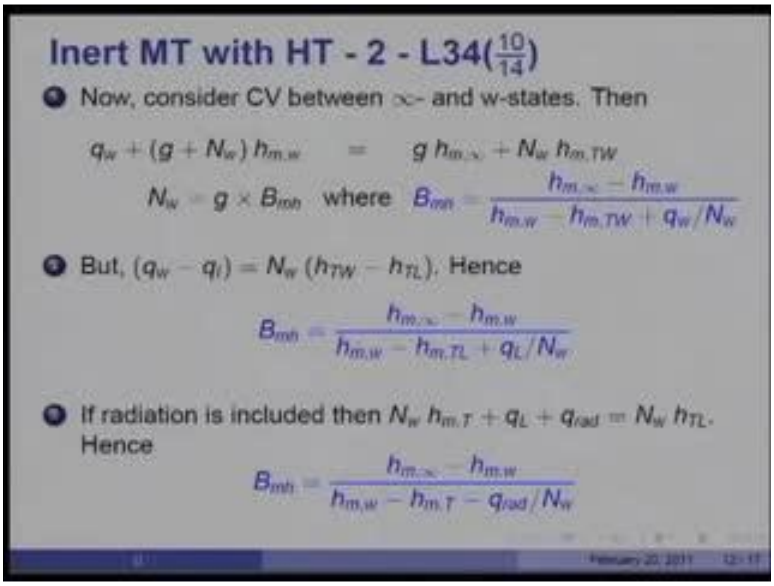

## Inert MT with HT - 2 - L34($\frac{10}{14}$)

- Now, consider CV between $\infty$- and w-states. Then

$$q_w + (g + N_w)\,h_{m,w} = g\,h_{m,\infty} + N_w\,h_{m,TW}$$

$$N_w = g \times B_{mh} \quad \text{where} \quad B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TW} + q_w/N_w}$$

- But, $(q_w - q_l) = N_w\,(h_{TW} - h_{TL})$. Hence

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TL} + q_L/N_w}$$

- If radiation is included then $N_w\,h_{m,T} + q_L + q_{rad} = N_w\,h_{TL}$. Hence

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,T} - q_{rad}/N_w}$$

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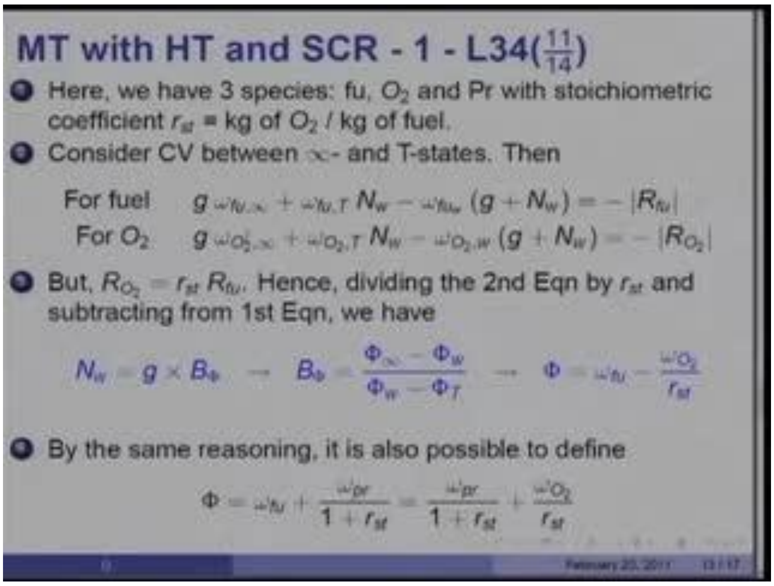

### MT with HT and SCR - 1 - L34($\frac{11}{14}$)

- Here, we have 3 species: fu, $O_2$ and Pr with stoichiometric coefficient $r_{st}$ = kg of $O_2$ / kg of fuel.
- Consider CV between $\infty$- and T-states. Then

For fuel $\quad g\,\omega_{fu,\infty} + \omega_{fu,T}\,N_w - \omega_{fu,w}\,(g + N_w) = -\,|R_{fu}|$

For $O_2$ $\quad g\,\omega_{O_2,\infty} + \omega_{O_2,T}\,N_w - \omega_{O_2,w}\,(g + N_w) = -\,|R_{O_2}|$

- But, $R_{O_2} = r_{st}\,R_{fu}$. Hence, dividing the 2nd Eqn by $r_{st}$ and subtracting from 1st Eqn, we have

$$N_w = g \times B_\Phi \;\rightarrow\; B_\Phi = \frac{\Phi_\infty - \Phi_w}{\Phi_w - \Phi_T} \;\rightarrow\; \Phi = \omega_{fu} - \frac{\omega_{O_2}}{r_{st}}$$

- By the same reasoning, it is also possible to define

$$\Phi = \omega_{fu} + \frac{\omega_{pr}}{1 + r_{st}} = \frac{\omega_{pr}}{1 + r_{st}} + \frac{\omega_{O_2}}{r_{st}}$$

This is of great relevance because different expressions come handy depending on the problem at hand, the mass transfer problem at hand as we will discover, when we solved practical problems, we turn to mass transfer with heat transfer and simple chemical reaction. Now, as you know in this case, 1 kg of fuel reacts with r s t kg of oxygen to produce 1 plus r s t kg of product with where r s t is the kilograms of oxygen to a kilogram of fuel is the stoichiometric coefficient.

So, I can write between infinity and t states, an equation for the fuel where this is the amount of fuel mass flux in from the bottom n w and this is the amount of fuel that will go out of the considered phase, carrying with it value of omega f u w and all of that must equal minus r f u for oxygen, likewise, same expression equal to minus r o 2 because in the considered phase, in the presence of chemical reaction both fuel and oxygen would be depleted.

So, that is why I have shown these are the area this is the volume average reaction rates r f u and r o 2, but r o 2 is equal to r s t r f u and therefore, if I divide this equation by r s t and subtract from the first equation, I would again get essentially n w equal to g phi where b phi is phi infinity minus phi w and phi w minus phi t where phi is, as you know is a conserved property omega f u minus omega o 2 by r s t. By the same reasoning, I can also form write an equation for product here and I would get phi equal to omega f u omega product plus 1 plus r s t or omega product 1 plus r s t plus omega o 2 by r s t. Now these we have gone over this type of material before. The important thing is, even in the presence of chemical reaction, we are able to show that n w equal to g b holds, where b is now found from appropriate conserved property.

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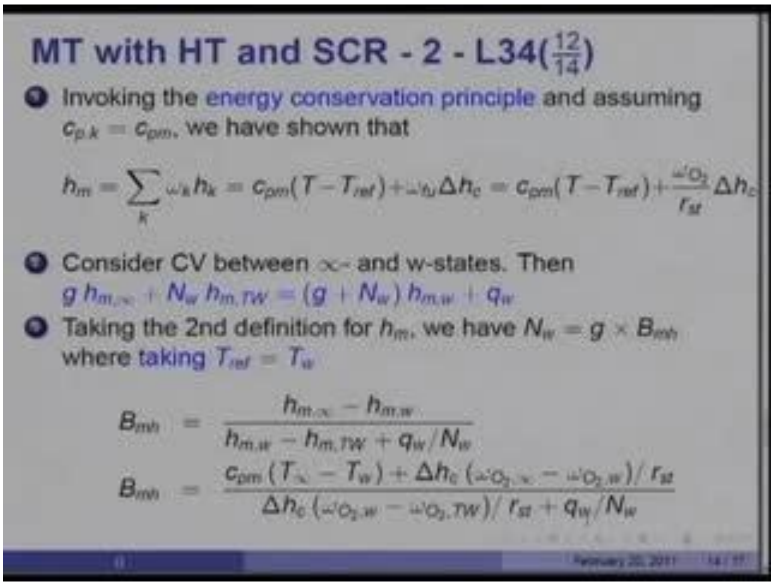

## MT with HT and SCR - 2 - L34($\frac{12}{14}$)

- Invoking the energy conservation principle and assuming $c_{p,k} = c_{pm}$, we have shown that

$$h_m = \sum_k \omega_k h_k = c_{pm}(T - T_{ref}) + \omega_{fu} \Delta h_c = c_{pm}(T - T_{ref}) + \frac{\omega_{O_2}}{r_{st}} \Delta h_c$$

- Consider CV between $\infty$- and w-states. Then $g\, h_{m,\infty} + N_w\, h_{m,TW} = (g + N_w)\, h_{m,w} + q_w$
- Taking the 2nd definition for $h_m$, we have $N_w = g \times B_{mh}$ where taking $T_{ref} = T_w$

$$B_{mh} = \frac{h_{m,\infty} - h_{m,w}}{h_{m,w} - h_{m,TW} + q_w/N_w}$$

$$B_{mh} = \frac{c_{pm}\,(T_\infty - T_w) + \Delta h_c\,(\omega_{O_2,\infty} - \omega_{O_2,w})/\,r_{st}}{\Delta h_c\,(\omega_{O_2,w} - \omega_{O_2,TW})/\,r_{st} + q_w/N_w}$$

Here, we had invoked the mass conservation principle. Now, we will invoke the energy conservation principle and assume c p k equal to c p m as we have done earlier also, then h m will be omega k h k c p m t minus t ref minus omega f u del h c and that I can also associate del h c with omega o 2 by r s t. So, if I consider c v between infinity and w states, then g h m infinity plus n w h m t w would equal g plus n w h m w plus q w.

If I take the second definition of h m, then again we would have n w equal to g b m h, where if I now take t ref equal to t w itself, then b m h would be h m infinity minus h m w h m w minus h m t w plus q w n w, divided by n w, that would result into c p m t infinity minus t w plus del h c omega o 2 infinity minus omega o 2 w divided by r s t del h c omega o 2 w minus omega o 2 t w divided by r s t plus q w by n w. Now of course, you can imagine, supposing I was burning a liquid fuel, it would have no oxygen. So, that would be, it would have no oxygen and therefore, that would be 0. If the liquid was at its boiling point, then no oxygen can survive there and therefore, that would be 0 and I would get simply c p m t infinity minus t w del h c omega o 2 at infinity divided by r s t and for a liquid droplet burning in air, I would readily know what omega o 2 infinity is, the only thing that I now need to do is to interpret q w.

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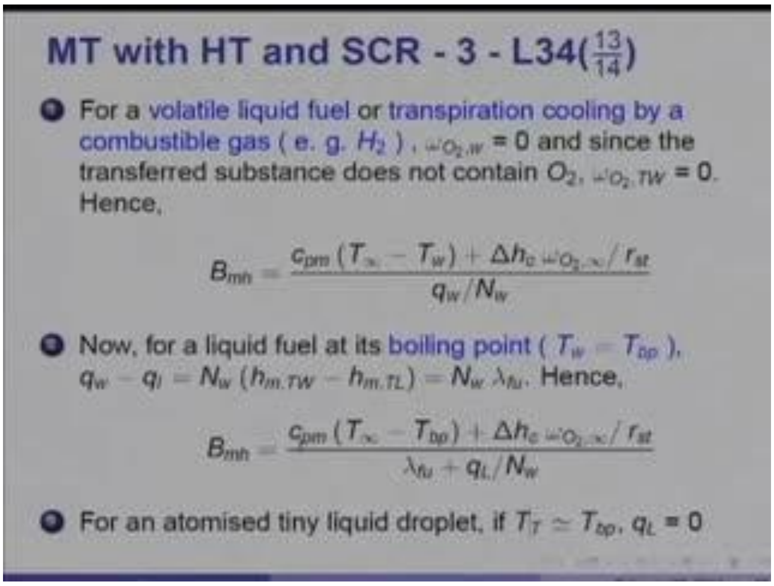

MT with HT and SCR - 3 - L34($\frac{13}{14}$)

- For a volatile liquid fuel or transpiration cooling by a combustible gas ( e. g. $H_2$ ) , $\omega_{O_2,w} = 0$ and since the transferred substance does not contain $O_2$, $\omega_{O_2,TW} = 0$. Hence,

$$B_{mh} = \frac{c_{pm}\,(T_\infty - T_w) + \Delta h_c\, \omega_{O_2,\infty} / r_{st}}{q_w / N_w}$$

- Now, for a liquid fuel at its boiling point ( $T_w = T_{bp}$ ), $q_w - q_l = N_w\,(h_{m,TW} - h_{m,TL}) = N_w\,\lambda_{fu}$. Hence,

$$B_{mh} = \frac{c_{pm}\,(T_\infty - T_{bp}) + \Delta h_c\, \omega_{O_2,\infty} / r_{st}}{\lambda_{fu} + q_L / N_w}$$

- For an atomised tiny liquid droplet, if $T_T \simeq T_{bp}$, $q_L = 0$

So, for a volatile fuel or transpiration cooling by combustible gas like hydrogen, omega o 2 w is 0 and since the transferred substance does not contain oxygen, I would get this expression that is what I have shown. If the liquid fuel is at its boiling point t w is equal to t b p, then q w minus q l would be equal to n w h f u and therefore, q w divided by n w will be simply c p m t infinity minus t b p del h c omega o 2 infinity by r h t lambda f u plus q l divided by n w.

Now, for an atomized tiny droplet, we can say that there is hardly any temperature difference between w and t state. It is like saying that we have injected inside the diesel engine, the liquid droplet already at its boiling point. In which case, there can be no conduction heat transfer in the transferred substance phase and therefore, you will see that with q l equal to 0 and t t equal to t b p, you will get c p t infinity minus t d p delta h c o omega plus lambda f u, divided by lambda f u and that quantity would be 0. So, you can see b m h can now be evaluated very easily, because boiling point of a liquid is known, delta h c the heat of combustion of the fuel is known, omega o 2 infinity would be typically known and the stoichiometric coefficient in the simple chemical reaction would also be known, along with the latent heat of the fuel. So, all quantities are known and b m h can be evaluated very very easily.

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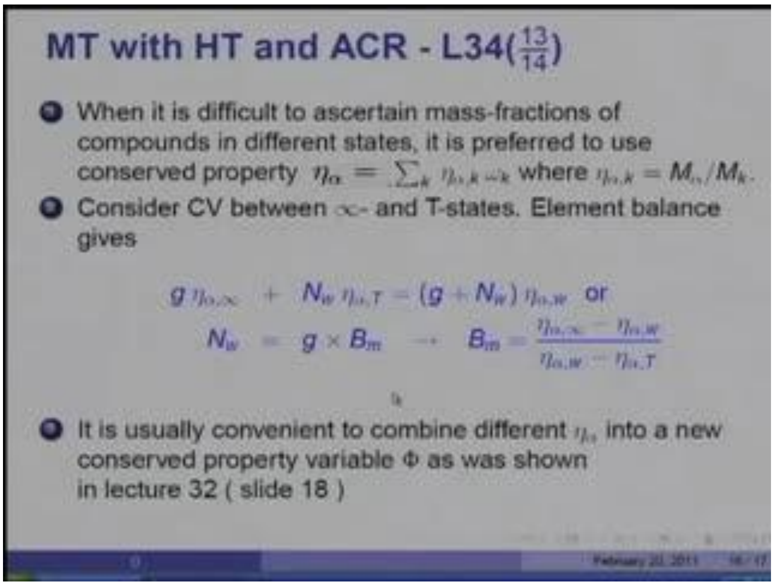

## MT with HT and ACR - L34($\frac{13}{14}$)

- When it is difficult to ascertain mass-fractions of compounds in different states, it is preferred to use conserved property $\eta_\alpha = \sum_k \eta_{\alpha,k}\,\omega_k$ where $\eta_{\alpha,k} = M_\alpha / M_k$.
- Consider CV between $\infty$- and T-states. Element balance gives

$$g\,\eta_{\alpha,\infty} + N_w\,\eta_{\alpha,T} = (g + N_w)\,\eta_{\alpha,w} \quad \text{or}$$

$$N_w = g \times B_m \rightarrow B_m = \frac{\eta_{\alpha,\infty} - \eta_{\alpha,w}}{\eta_{\alpha,w} - \eta_{\alpha,T}}$$

- It is usually convenient to combine different $\eta_\alpha$ into a new conserved property variable $\Phi$ as was shown in lecture 32 ( slide 18 )

Now, when it is difficult to ascertain mass fractions of compounds in different states, it is preferred to use the conserved property n alpha equal to n alpha k omega k n alpha k being equal to the molecular weight of alpha, the element alpha divided by molecular weight of the species in which it occurs. So, again if I do c v between infinity and t states, then element balance would give me n w equal to g b m, really easy to see this and b m would be again, eta alpha in infinity w w and t states. Many a times, we will find that it is much more convenient to combine different eta alpha into a new conserved property variable phi, as was shown in lecture number 32, where we consider the burning of graphite.

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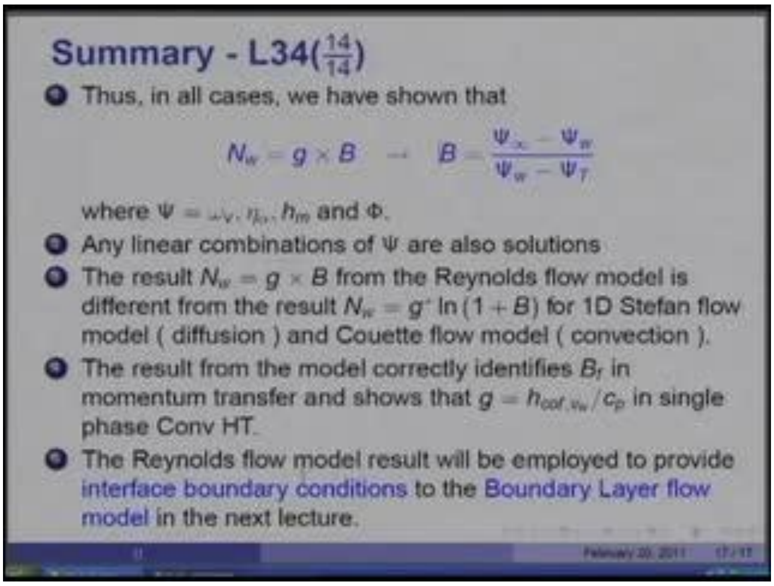

## Summary - L34($\frac{14}{14}$)

- Thus, in all cases, we have shown that

$$N_w = g \times B \rightarrow B = \frac{\Psi_\infty - \Psi_w}{\Psi_w - \Psi_T}$$

where $\Psi = \omega_V, \eta_\alpha, h_m$ and $\Phi$.

- Any linear combinations of $\Psi$ are also solutions
- The result $N_w = g \times B$ from the Reynolds flow model is different from the result $N_w = g^* \ln(1 + B)$ for 1D Stefan flow model ( diffusion ) and Couette flow model ( convection ).
- The result from the model correctly identifies $B_f$ in momentum transfer and shows that $g = h_{cof,V_w}/c_p$ in single phase Conv HT.
- The Reynolds flow model result will be employed to provide interface boundary conditions to the Boundary Layer flow model in the next lecture.

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In all cases, we now find that the Reynolds flow model results into n w equal to g b, where b is equal to psi infinity minus psi w divided by psi w minus psi t, psi can be in only omega v, it can be element fraction eta alpha, it can be mixture enthalpy h m or it can be an appropriately defined phi from the participating species. Any linear combination of psi's are also solutions. So, we can add any of these types of psi for our convenience.

The result n w into g from the Reynolds flow model is different from the result n w g star l n 1 plus b from 1 d Stefan flow model, which was essentially diffusion model and the Couette flow model. Both of them had logarithmic form whereas, here it is a g into b, it is much more like the ohm's law type of model. The result from the model correctly identifies b f in momentum transfer and shows that g would be equal to h cof v w by c pin single phase convective heat transfer.

So, the Reynolds flow model result will be employed to provide the interface boundary conditions to the boundary layer flow model in the next lecture and we will see how this can be done in the next lecture.