Chapter 5 Condensation Heat Transfer

: The removal of heat from a system in such a manner that vapor is converted into liquid.

o homeogenous condensation

o heterogeneous condensation
- drop wise condensation ↔ nucleate boiling
- film wise condensation ↔ film boiling

1. Drop wise condensation

Drop wise condensation occurs when the vapor is cooled sufficiently below the saturation temperature to induce the nucleation of droplets.

- o by contamination of non wettable surfaces
- o incidental reduction of interfacial tension between condensate and surface
- o occur homogeneously within vapor or heterogeneously on the entrained particular matter

A reliable mechanism of drop wise condensation have not yet been established. However, the heat transfer by the drop wise condensation is several times larger than that of the film wise condensarion.

Using $\Delta p (= p_v - p_l = \frac{2\sigma}{R})$ for vapor formation, and Clausius Clapeyron equation, Hill. Witting and Demetri derived the rate of nucleation as,

$$\frac{dn}{dt} = N\frac{v_f}{v_g}\lambda\exp[-4\pi r^2/3kT_g] \tag{1.1}$$

where $\lambda = (2\sigma/\pi m)^{1/2}$ and N is the number of vapor mole per unit volume. Also, Oswatitsch derived the droplet growth rate as,

$$\frac{dr}{dt} = \frac{3p_\infty}{8h_{fg}\rho_f}\sqrt{\frac{3R}{T_g M}}(T_\infty - T_g) \tag{1.2}$$

◆ Two fundamentally different model to describe the dropwise condensation

1) By Jakob (1936)

: In the beginning, the condensation starts as a thin unstable film on whole or a part of surface. When the film has a critical thickness to break up, the liquid forms a droplet due to the surface tension.

In 1964, Silver has quantitatively analyze the drop wise condensation based on the radial inward drainage of condensate.

$$\frac{G_D}{G_f} = \left[\frac{\rho_l^2 D^3 g}{21.2 \mu_l G_f} \right]^{\frac{1}{3}} \qquad \text{D : tube diameter} \qquad (1.3)$$

여기서 G_D : the drop-wise condensation mass flux

G_f : the film wise condensation mass flux for ideal condition

For steam at 1atm for $G_F = 1.36 \cdot 10^{-3} k_g/m^2 s$.

$G_d/G_F = 6.5$ and mean thickness of film ~ $2\mu m$

★ <u>This model is proper when the temperature difference is large.</u>

2) heterogeneous nucleation process By Eucken (1937)

: In 1963, McCormick and Baer suggested that a lot of submicroscopic droplets were randomly nucleated at active sites like minute groove on the surface and collapsed to make big drop which can fall down from the surface.

For hemispherical droplet, the growth rate is assumed to be only dependent on the heat condensation through the drop as,

$$\left(\frac{dr}{dt}\right)^2 = 4.15 \frac{k_l (T_{gi} - T_s)}{h_{fg} \rho_l} \qquad (1.4)$$

where T_{gi} : interfacial gas temperature

By Gose, Mucciardi and Baer (1969),

minimum nucleation site density = $5 \cdot 10^4 sites/cm^2$.

diameter of drop to be chiefly collapsed ~ 0.01 cm

★ <u>This model is proper when the temperature difference is small.</u>

☞ For small temperature difference, the drop wise heat transfer coefficient a little increases as the temperature increases, but significantly decreases in the case of large temperature difference due to change in mechanism and presence of non condensible gas in vapor phase.

☞ influence of pressure

☞ influence of vapor velocity

: maximum at 2m/sec

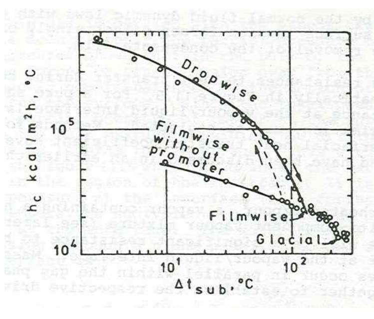


그림 1 HTC in dropwise vs filmwise condensation

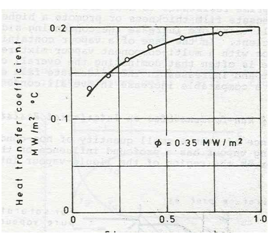


그림 2 Influence of pressure

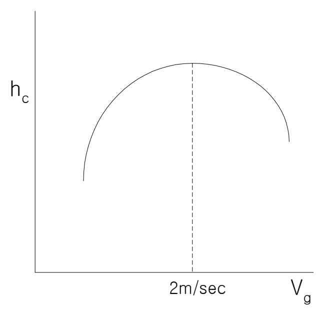


그림 3 Influence of vapor velocity

2. Film wise Condensation

The condensate wets the surface and forms a continuous liquid film.

The film behaves according to the fluid dynamic law which considers the gravity, the vapor shear, the surface tension force and so on.

★ Various resistance to heat transfer during condensation

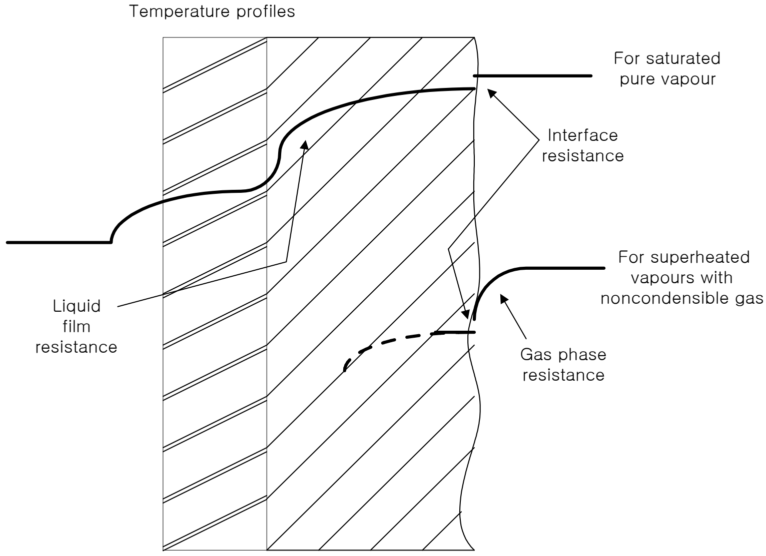
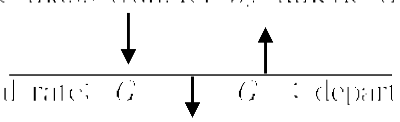


그림 1 Various resistance to heat transfer during condensation

■ The mechanism of condensation at a plane liquid vapor interface

Consider the interface mass transfer by kinetic theory



T_v, p_v

T_i $\qquad G = G - G$

From kinetic theory,

$$|G| = \left(\frac{M}{2\pi R}\right)^{\frac{1}{2}} \frac{P}{T^{1/2}} \qquad (2.1)$$

$$G = G_1 - G \qquad (2.2)$$

Since the interface is actually not in the static thermal equilibrium, it is meaningless to use the thermostatic pressure or the temperature in each side. Thus, we have to solve Boltzman transport equation with proper boundary conditions and the asymptote of the thermal equilibrium in few mfp.

But for the purpose of engineering, a simplified kinetic theory technique and proper correction factor may adopted.

A. Crude theory

For appropriate temperature and pressure of liquid and vapor, from Eqs. (2.1) and (2.2),

$$G = \left(\frac{M}{2\pi R}\right)^{\frac{1}{2}}\left[\frac{p_g}{T_g^{1/2}} - \frac{p_f}{T_f^{1/2}}\right] \tag{2.3}$$

For small difference in pressure and temperature, by differentiating,

$$G = \left(\frac{M}{2\pi RT}\right)^{\frac{1}{2}} p\left[\frac{\Delta p}{p} - \frac{\Delta T}{2T}\right] \tag{2.4}$$

Assume that there is no temperature jump across the interface

$$\therefore\ G = \left(\frac{M}{2\pi RT}\right)^{\frac{1}{2}} \Delta p \tag{2.5}$$

This means that the condensation occurs when $p_g > p_f$ (or $\Delta p > 0$)

Interfacial heat transfer coefficient

$$H_i = q''/\Delta T \tag{2.6}$$

and

$$q'' = h_{fg} G \tag{2.7}$$

From Eqs. (2.4), (2.5) and (2.6),

$$H_i = h_{fg}\left[\frac{M}{2\pi RT}\right]^{\frac{1}{2}} \frac{\Delta p}{\Delta T} \tag{2.8}$$

By Clausius – Clapeyron equation,

$$H_i = \left[\frac{M}{2\pi RT}\right]^{\frac{1}{2}} \frac{J h_{fg}^2}{T v_{fg}}\left[1 - \frac{p v_{fg}}{2 h_{fg}}\right] \tag{2.9}$$

B. Modification of the crude theory

By Silver, Eq. (2.9) is multiplied by a factor, f, called the "**Molecular exchange condensation factor**"

f = 0.036 for water under vacuum condenser conditions.

Or, by Schrage, Eq. (2.3) can be rewritten as,

$$G = \left(\frac{M}{2\pi R}\right)^{\frac{1}{2}}\left[\Gamma \sigma_c \frac{P_g}{T_g^{1/2}} - \sigma_e \frac{P_f}{T_f^{1/2}}\right] \tag{2.10}$$

where σ_c , σ_f : condensation and evaporation coefficient(usually same, σ)

Γ represents net motion of vapor toward surface and is a function of a

a is the ratio of gas overall speed u ($= G/\rho_g$) to characteristic molecular velocity, $(2RT/M)^{1/2}$ as,

$$a = \frac{u}{(2RT/M)^{1/2}} \tag{2.11}$$

For very large a, $\Gamma(a) \approx 2a\pi^{\frac{1}{2}}$

$$\therefore G = \left(\frac{M}{2\pi R}\right)^{\frac{1}{2}}\left[2a\pi^{\frac{1}{2}}\sigma\frac{\rho_g}{T_g^{1/2}} - \sigma\frac{\rho_i}{T_i^{1/2}}\right] \tag{2.12}$$

여기서 $\sigma = 1$ 이면(즉 surface에 닿는 모든 molecule 은 condensing.), evaporation은 condensation에 비해 무시할 수 있다는 것이다. → black body

For very small a (즉 low condensation rate), $\Gamma(a) \approx 1 + a\pi^{\frac{1}{2}}$

$$\therefore G = \left(\frac{M}{2\pi R}\right)^{\frac{1}{2}}\left[\frac{p_g}{T_g^{1/2}} - \frac{p_i}{T_i^{1/2}}\right]\left(\frac{2\sigma}{2-\sigma}\right) \tag{2.13}$$

→ Kucherov – Rikenglaz equation

ㆍ Above kinetic theory neglects the non equilibrium interaction and, moreover, is hard to consider the effect of the non condensable gas.

C. Film condensation on a planar surface

◎ ***<u>Nusselt Equation for a laminar film</u>***

<u>Assumptions</u>

ㄱ. the flow of condensate in the film is laminar

ㄴ. constant fluid properties

ㄷ. negligible subcooling of the condensate

ㄹ. negligible momentum changes through the film

ㅁ. vapor is stationary and exerts no drag on the downward motion of the condensate

ㅂ. heat transfer is by conduction only

Velocity distribution in a liquid film

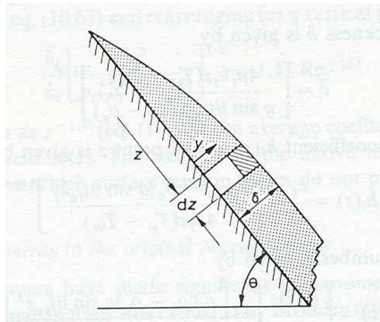


그림 5 laminar flow over a inclined surface(Nusselt)

From force balance

$$(\delta - y)\,dz\,(\rho_l - \rho_v)\,g\sin\Theta = \mu_l\left(\frac{du_z}{dy}\right)dz \qquad (2.14)$$

Integrate with B.C : $u_z = 0$ at $y = 0$

$$\therefore u_z = \frac{(\rho_l - \rho_v)\,g\sin\Theta}{\mu_l}\left[y\delta - \frac{y^2}{2}\right]$$

And the mass flow rate per unit width will be

$$\Gamma = \rho_l\int_0^\delta u_z\,dy = \frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta\,\delta^3}{3\mu_l} \qquad (2.15)$$

Thus, the rate of increase of film flow rate with δ is

$$\frac{d\Gamma}{d\delta} = \frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta\,\delta^2}{\mu_l} \qquad (2.16)$$

If film surface temperature is T_{sat} and wall temperature is T_w, the heat transferred by conduction to an element of dz

$$dq = \frac{k_l}{\delta}(T_{sat} - T_w)\,dz = h_{fg}\,d\Gamma$$

$\therefore$ mass rate of condensation is

$$d\Gamma = \frac{k_l}{\delta\,h_{fg}}(T_{sat} - T_w)\,dz \qquad (2.17)$$

Inserting Eq. (2.16) into (2.17) and separating variables and integrating with $\delta = 0$ at $z = 0$,

$$\mu_l k_l(T_{sat} - T_w)\,z = \rho_l(\rho_l - \rho_v)\,g\sin\Theta\,h_{fg}\left(\frac{\delta^4}{4}\right)$$

$$\therefore \delta = \left[\frac{4\mu_l k_l z(T_{sat} - T_w)}{g\sin\Theta\,h_{fg}\,\rho_l(\rho_l - \rho_v)}\right]^{\frac{1}{4}} \qquad (2.18)$$

or $H_z(z)$ will be

$$H_z(z) = \frac{k_l}{\delta} = \left[\frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta\,h_{fg}k_l^3}{4\mu_l z(T_{sat} - T_w)}\right]^{\frac{1}{4}} \qquad (2.19)$$

and

$$Nu = \frac{H_z z}{k_l} = \left[\frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta\,h_{fg}z^3}{4\mu_l k_l(T_{sat} - T_w)}\right]^{\frac{1}{4}} \qquad (2.20)$$

The mean value of heat transfer coefficient over the whole surface is

$$\overline{H}_l = \frac{1}{z}\int_0^z H_l(z)\,dz = 0.943\left[\frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta\,h_{fg}k_l^3}{\mu_l z(T_{sat} - T_w)}\right]^{\frac{1}{4}} \quad (2.21)$$

or

$$\overline{H}_l = \frac{\Gamma_z h_{fg}}{z(T_{sat} - T_w)} \quad (2.22)$$

where Γ_z is condensate flow at z from top.

Now, from Eqs. (2.17) and (2.22) ,

$$\delta = \frac{k_l \Gamma_z dz}{\overline{H}_l z\, d\Gamma} \quad (2.23)$$

And from Eqs. (2.15) and (2.23)

$$k_l\left[\frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta}{3\mu_l}\right]^{\frac{1}{3}}\frac{dz}{z} = \frac{\overline{H}_l\Gamma^{-\frac{1}{3}}d\Gamma}{\Gamma_z}$$

Integrate over z

$$\overline{H}_l = 0.925\left[\frac{\rho_l(\rho_l - \rho_v)\,g\sin\Theta\,k_l^3}{\mu_l\Gamma_z}\right]^{\frac{1}{3}} \quad (2.24)$$

Let $Re_\Gamma = 4\Gamma_z/\mu_l$ and $\sin\Theta = 1$

$$\frac{\overline{H}_l}{k_l}\left[\frac{\mu_l^2}{\rho_l(\rho_l - \rho_v)\,g}\right]^{\frac{1}{3}} = 1.47\,Re_\Gamma^{-\frac{1}{3}},\quad Re_\Gamma = \frac{4\Gamma_z}{\mu_l} \quad (2.25)$$

D. Condensation on horizontal tube

By Nusselt

$$H_l(\alpha) = 0.693\left[\frac{\rho_l(\rho_l - \rho_v)\,g\sin\alpha\,k_l^3}{\Gamma_\alpha'\mu_l}\right]^{\frac{1}{3}} \quad (2.26)$$

where α : angle between horizontal tube and the vertical

Γ_α' : the local mass flow rate per unit length

$$\therefore\ \overline{H}_l = 0.725\left[\frac{\rho_l(\rho_l - \rho_v)\,g\,h_{fg}'k_l^3}{D\mu_l(T_{sat} - T_w)}\right]^{\frac{1}{4}} \quad (2.27)$$

where D is outer diameter of tube,

h_{fg}' : modified h_{fg} to improve the original Nusselt theory

by Rohsenow

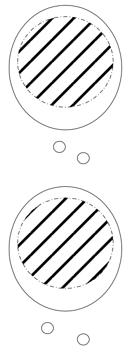
$$h_{fg}\left[1 + 0.68\left(\frac{c_p\Delta T_l}{h_{fg}}\right)\right] \quad (2.28)$$

or $$\frac{\overline{H_L}}{k_l}\left[\frac{\mu_l^2}{\rho_l(\rho_l-\rho_g)g}\right]^{\frac{1}{3}} = 1.51 Re_\Gamma^{-\frac{1}{3}} \quad (2.29)$$

thicker condensate layer on subsequent tube

∴ HT coefficient will be reduced.

▶ Condensation within a horizontal tube



1 Stratified flow

small vapor velocity and low interfacial shear forces

modify Nusselt equation on the outside of a horizontal tube as,

$$\overline{H_i} = F\left[\frac{\rho_l(\rho_l-\rho_g)g h_{fg} k_l^3}{D\mu_l(T_g - T_w)}\right]^{\frac{1}{4}} \quad (2.30)$$

where F is a factor which depends on the angle subtended from the tube center to the chord forming the liquid level.

2 Slug, plug and wavy flow

significant vapor shear force

different coefficient from top to bottom of the tube

- Top : $F = 0.31\left[\frac{G_v D}{\mu_v}\right]^{0.12}$

- Bottom : $$\left[\frac{H_i(\pi)D}{k_l}\right] = \frac{\Phi_{g}^{2}\sqrt{8(G_l D/\mu_l)}}{5\left[1+(1/Pr_l)\ln(1+5\,Pr_l)\right]} \quad (2.31)$$

E. The influence of various parameters on the film condensation

1 Turbulence

Experiments shows that the wavy surface has 10 ~ 50% higher HT coefficient than theoretical value

$$\therefore\ H_f^{0.85} = 1.2H_f\,(of\ Eq.\ (2.24)) \quad (2.32)$$

By Colburn, the heat transfer coefficient for the case of turbulent is

$$\frac{h_l(z)}{k_l}\left[\frac{\mu_l^2}{\rho_l(\rho_l-\rho_g)g}\right]^{\frac{1}{3}} = 0.056\left(\frac{4\Gamma_l}{\mu_l}\right)^{0.2}\left(\frac{c_{pl}\mu_l}{k_l}\right)^{\frac{1}{3}} \quad (2.33)$$

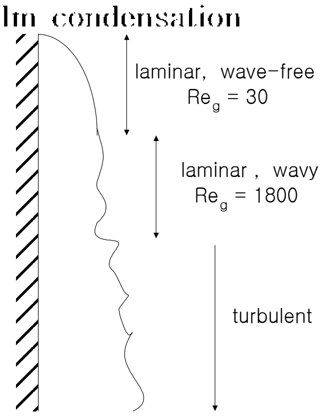


그림 6 Wave development on falling film

2) Influence of interfacial shear. (That is, vapor velocity exists)

For inclined plane surface, Eq.(2.11) will be

$$(\delta - y)\ dz\left(\rho_l g\ \sin\Theta - \left(\frac{dP}{dz}\right)\right) + \tau_i dz = \mu\left(\frac{du_z}{du}\right)dz \qquad (2.34)$$

Define fictitious vapor density, ρ_g^i, as

$$\frac{dP}{dz} = \rho_g^i g\ \sin\Theta \qquad (2.35)$$

If the only pressure gradient is due to vapor head,

then $\rho_g^i = \rho_g$

Inserting Eq. (2.35) into (2.34) and integrating

$$u_z = \frac{(\rho_l - \rho_g^i)g\ \sin\Theta}{\mu_l}\left[y\,\delta - \frac{y^2}{2}\right] + \frac{\tau_i y}{\mu_l} \qquad (2.36)$$

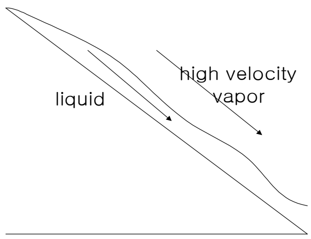


Fig. 7 Interfacial shear on falling film

Then, $\Gamma = \rho_l \int_0^{\delta} u\ dy = \left[\frac{\rho_l(\rho_l - \rho_g^i)g\ \sin\Theta\ \delta^3}{3\mu}\right] + \frac{\tau_i\ \rho_l\ \delta^2}{2\mu_l}$

$$\therefore\ \frac{d\Gamma}{d\delta} = \left[\frac{\rho_l(\rho_l - \rho_g^i)g\ \sin\Theta\ \delta^2}{\mu_l}\right] + \frac{\tau_i\ \rho_l\ \delta}{\mu_l} \qquad (2.37)$$

With same procedure as that from Eq. (2.16) to (2.18),

$$\left[\frac{4\mu_l k_l z(T_{sat} - T_w)}{g\ \sin\Theta\ h_{fg}\rho_l(\rho_l - \rho_g^i)}\right] = \delta^4 + \frac{1}{3}\left[\frac{\tau_i\ \Theta^3}{(\rho_l - \rho_g^i)g\ \sin\Theta}\right] \qquad (2.38)$$

3) non condensable gas effect.

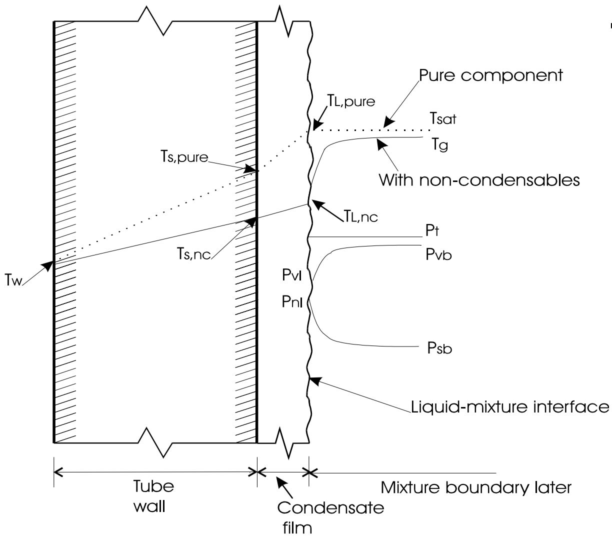


Fig. 8. Boundary Layer Temperature and Pressure Distributions

o The noncondensable gas near the condensate film protects the diffusion of the vapor from the bulk mixture to the film and thus reduce the mass and energy transfer.

simultaneously solve conservation equations for both condensate film and vapor gas layer

o In nuclear plants, N_2 from accumulator or FP during LOCA

o hamper the performance of the pressurizer.

1) For stagnant vapor gas mixture

- By Sparrow and Lim, for free convection arising from density differences, the condensing rate is proportional to

  - bulk gas mass fraction $(\rho\ \sigma\ /\rho\ \sigma\ )^{\frac{1}{2}}$ (2.39)
  - vapor gas mixture Schmidt $(c_p\ (T_i - T_w)/P_r\ \rho_v)$ (2.40)

∴ non condensable gas effect increase with increase of

√ Sc

√ $(\rho\ \sigma\ /\rho\ \sigma\ )^{\frac{1}{2}}$

2) Moving vapor with noncondensable gas

In this case, the non condensable gas effect is smaller than the case of stationary vapor gas mixture. That is, the gas concentration at the interface is low relatively due to the "sweep effect" of the moving vapor gas mixture.

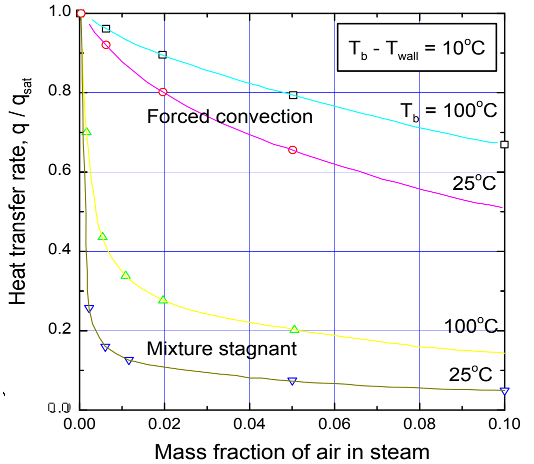


그림 9 The Influence of Noncondensable Air in Steam

◆ **Diffusion Layer Theory for Turbulent Vapor Condensation**

Colburn and Hougen(1934) first proposed that condensation mass transport is controlled by diffusion across a thin layer or film by the difference in the bulk and interface gas partial pressured.

As shown in Fig.10, non condensable gas accumulates at the liquid vapor interface, and reduce the T_i^s below T_b^s

$$q_t'' = h_w\left(T_i^s - T_\infty\right) = q_c'' + q_s''$$

$$= -h_{fg} c M_v \tilde{v}_i + k_v \left(\frac{\partial T}{\partial Y}\right)_i$$

(2.41)

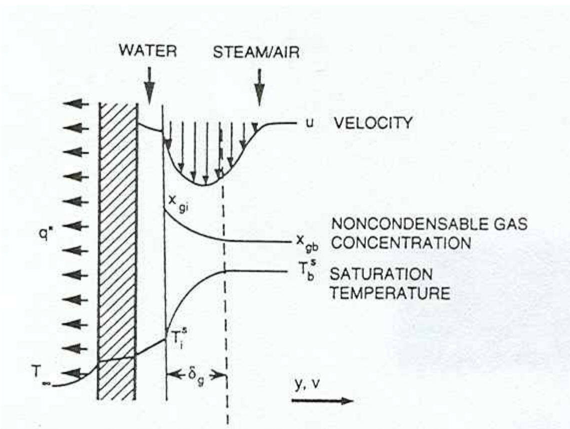


그림 10 Condensable diffusion layer on a vertical plate

The average molar velocity away from the interface, $\widetilde{v_i}$, is related to the X_g by Ficks law.

$$c v_{gi} = c\, x_{gi} \widetilde{v_i} - cD \frac{\partial x_g}{\partial y} \qquad (2.12)$$

Since the interface is impermeable to the noncondensable gas and so the absolute gas velocity at the interface equals zero, $v_{gi} = 0$.

$$\therefore \widetilde{v_i} = \left(D \frac{1}{x_g} \frac{\partial x_g}{\partial y} \right)_i = \left(D \frac{\partial}{\partial y} ln(x_g) \right)_i \qquad (2.13)$$

Or

$$\widetilde{v_i} = \frac{D}{\delta_g} \{ \ln(x_{gb}) - \ln(x_{gi}) \} \qquad (2.14)$$

Define a log mean mole fraction as,

$$x_{avg} = \frac{x_b - x_i}{\ln(x_b / x_i)} \quad . \qquad x_b < x_{avg} < x_i \qquad (2.15)$$

Then $$\widetilde{v_i} = \frac{D}{x_{g,avg} \delta_g} (x_{gb} - x_{gi}) \qquad (2.16)$$

Assuming the ideal gas,

$$\widetilde{v_i} = \frac{D}{P_i X_{g,avg} \delta_g} (P_{vi} - P_{vb}) \qquad (2.50)$$

Since the difference of partial pressure in not convenient for heat transfer calculation, use Clausius Clapeyron equation to express the pressure with the temperature. Then,

$$\widetilde{v_i} = \frac{D h_{fg} M_v x_{v,avg}}{R T_{avg}^2 x_{g,avg} \delta_g} \left(T_b^s - T_i^s \right) \qquad (2.51)$$

where $T_{avg} = \left(T_i^s + T_b^s \right)/2$: average temperature in the diffusion layer

The Sherwood number, from Eq.(2.11), (2.33) and (2.51) ,

$$Sh_L = \frac{L}{\delta_g} = \frac{q_c''}{\left(T_b^s - T_i^s \right)} L \, \phi \left(\frac{R^2 T_{avg}^3}{h_{fg}^2 P_t M_v^2 D} \right) \qquad (2.52)$$

$\downarrow$ $\qquad$ $\downarrow$

h_c $\qquad$ inverse unit of thermal conductivity

where $\phi = \dfrac{x_{g,avg}}{x_{v,avg}} = - \dfrac{\ln\left[(1 - x_{gb}) / (1 - x_{gi}) \right]}{\ln\left[x_{gb} / x_{gi} \right]}$

Introduce the effective condensation of thermal conductivity

$$k_c = \frac{1}{\phi T_{avg}} \left(\frac{h_{fg}^2 P_0 M_v^2 D_0}{R^T T_0^2} \right) \tag{2.53}$$

Then the Sherwood number for condensation takes a simple form.

$$Sh_L = \frac{h_c L}{k_c} \tag{2.54}$$

Finally $$q_t^{''} = \frac{h_c\left(T_b^s - T_\infty\right) - h_s\left(T_b - T_\infty\right)}{1 + \dfrac{h_c + h_s}{h_w}}$$

We can solve the condensation heat transfer with

1) an appropriate correlation for Sh

2) determination of T_i , X_{gi}

① iteration for T_i

② fixed T_i

3) Pressure

increasing pressure (or saturation temperature) lower the effect of noncondensables due to the less effect on vapor saturation temperature at higher pressure the heat transfer rate must be smaller at lower pressure than at higher pressure

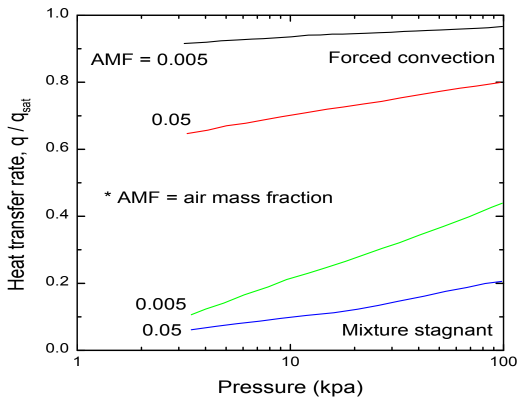


그림 11 Effect of Pressure on Condensation of Steam-Air Systems

4) Superheat in vapor

larger density difference between bulk mixture and mixture near the interface with noncondensables than in pure vapor for larger superheat

∵ larger buoyancy effects make much stronger convective flow to sweep past the cooled surface, and thus reduce the interfacial concentration & the thickness of noncondensable film

∴ The heat transfer rate increases.

5) RELAP5/MOD3 condensation model with noncondensible gas

noncondensible component is assumed to be in thermal and mechanical equilibrium three correlations for condensation heat transfer calculations

1) laminar film condensation on an inclined plane

: standard Nusselt film condensation correlation

2) laminar film condensation inside a horizontal tube with a stratified liquid surface

: modification of original theory

3) turbulent film condensation inside a vertical tube

: Carpenter and Colburn correlation

the reduction multiplier, FNC, for the presence of noncondensible gas

$$FNC = \frac{(p_s - p_{min})}{p * f(Re_g)} * F \, , \qquad \text{for } 0 < Re_g < 20{,}000 \qquad (2.55)$$

where $f(Re_g) = \dfrac{5}{1 + 0.0001 Re_g}$

$$F = 1 + f(Re_g) * \exp[-5(\frac{P_a}{p})]$$

and p_s, p_a, p are partial pressures of steam and air, and total pressure of mixture, repectively. p_{min} is minimum pressure in the steam table.

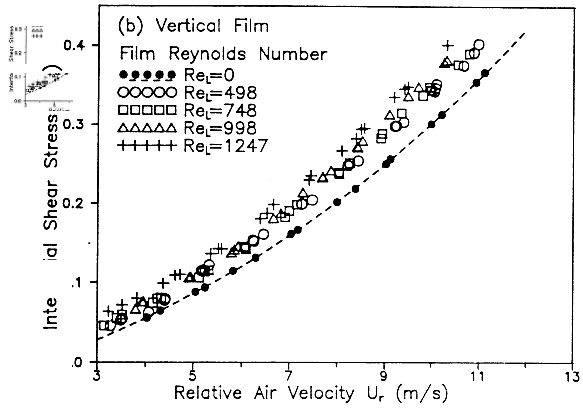
F. Waviness effect

The rate of mass transfer in the liquid phase is enhanced by a factor of 2-3 in the presence of waves.

The entire process of transfer in the liquid is controlled by the large scale wave structure.

o The rate of transfer is high just after the lump passes and decays with time until next lump arrives & a fresh film is generated: renewal process & unsteady mass transfer

The transfer in the gas phase is controlled by the small wave structure : the small waves act as a roughness but, the roughness is of unique shape.



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3. Falling film evaporation on a vertical wall

differences with condensation

1. start with a certain thickness of film
2. reduce of film Re according to evaporation

But, the reduction of RE is usually neglected since the evaporation rate is much smaller than the film flow rate.

Since $T_w > T_{sat}$, the liquid loser to the wall is superheated

∴ boiling or bubble cavitation in the feed liquid.

→ film disruption

But, water evaporates into a vapor air mixture

∴ total pressure >> $P_{sat}(T_w)$ and little possibility of boiling.

include mass transfer process

- film condensation : suction effect
- film evaporation : blowing effect

Since evaporating films are mostly wary laminar or turbulent, by Chun and Seban

$$Nu = 0.822\ Re^{-0.22}, \quad 30 < Re < Re_{tr} \qquad \text{(wavy laminar)}$$

$$Nu = 3.8\ 10^{-3}\ Re^{0.4}\ Pr^{0.65}, \quad Re_{tr} < Re \qquad \text{(turbulent)} \qquad (3.1)$$

where $Re_{tr} = 5800\ Pr^{-1.06}$

: same with condensation

4. Methods of Improving the HT coeff is condensation

(a) changes to geometry of the surface to increase the available area.

(b) treatment of the surface to promote drop wise rather than film wise force

(c) use of force fields

(1) The influence of surface geometry

At low R_e, $h_{rough} < h_{smooth}$ ∵ the condensate retains due to the surface tension.

At high $R_e (>140)$, $h_{rough} < h_{smooth}$

low fin tubing in horizontal tube condenser

∴ The HTC for fined surface is 15% less than that for the smooth surface, but has larger net heat transfer rate due to increased transfer area.

roped or spirally indented tube

by placing wires along vertical tube

: Optimum # of wire $n = \frac{0.18\pi D}{d}$

(2) The promotion of drop wise condensation

make the surface non wetting

1) chemically coated surfaces

2) polymer coated surfaces

3) electroplated coated surfaces

(3) The use of force field to enhance condensation

In the film wise condensation, the condensate film plays a role of the resistance of heat transfer.

∴ use external force, centrifugal, vibrational and electrostatic force to reduce the thickness and to promote drainage

horizontal rotating disc

$$\overline{h_c} = 0.904\left[\frac{k_l^3\rho_l^2\omega^2 h_{fg}}{\mu_l(T_{sat} - T_w)}\right]^{0.25} \qquad (2.1)$$

acoustic vibration

$$\frac{h_{vib}}{h_c} = 1 + 0.0018FA$$

where A : amplitude (mm)

F : frequency (0~150Hz)