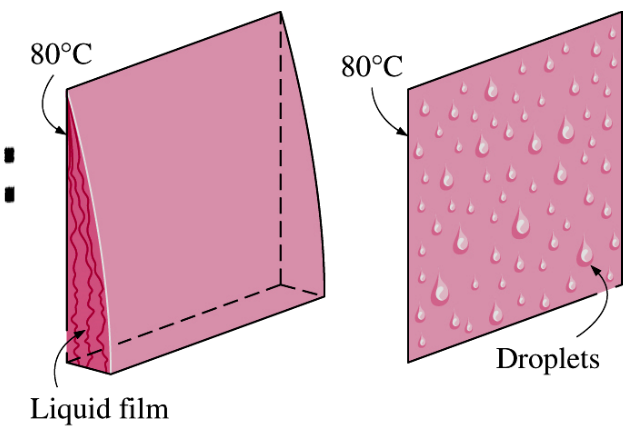
Condensation Heat Transfer

- Condensation **occurs when the temperature of** the vapor is reduced below its saturation temperature.
- **The solid surface whose temperature is below the saturation temperature of the vapor**
- **Two distinct forms of condensation:**
  - Filmwise condensation
  - Dropwise condensation



(*a*) Film condensation

(*b*) Dropwise condensation

Filmwise Condensation

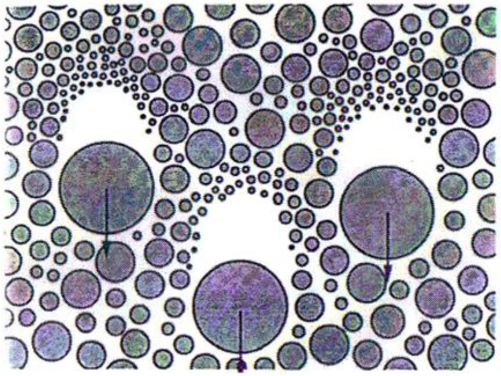
- **When liquid formed by condensation wets the surface. It is a more common type of condensation to occur.**
- **In filmwise condensation liquid condensate forms a continuous film over the surface, this film flows down the surface under the action of gravity, shear force due to vapor flow, or other forces.**
- The layer of liquid condensate acts as a barrier to heat flow due to its very low thermal conductivity and hence low heat transfer rate.

Dropwise condensation

- **Dropwise condensation takes place when the liquid condensate does not wet the solid surface.**
- **The condensate does not spread, but forms separate drops.**
- **These drops in turn coalesce to form large drops and sweeping clean a portion of the surface, where again new droplets are generated.**
- **The average heat transfer coefficient for dropwise condensation is much higher than filmwise condensation.**

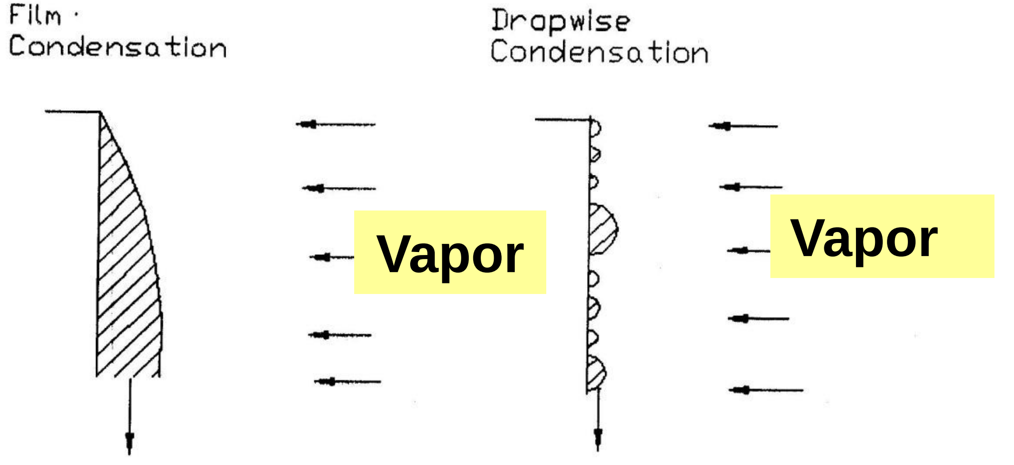
Condensation

- Dropwise condensation:
  - Promoted by making poorly wetted surface:
  - h could be an order of magnitude higher than in the filmwise case
  - Mechanism not well understood
- Film condensation
  - Liquid film covers entire surface
  - May flow due to gravity
  - Charactersitic of clean surfaces



Dropwise condensation

Filmwise and dropwise Condensation



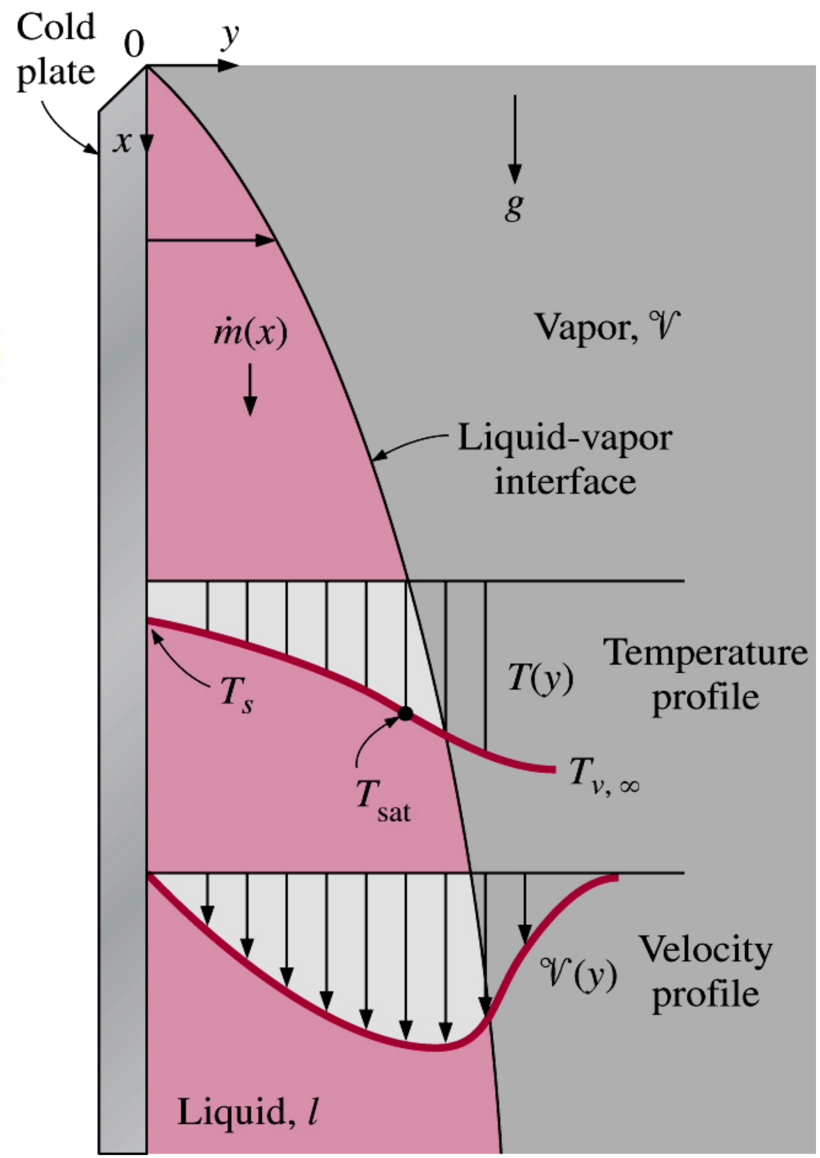
Heat transfer coefficient

$h_{film} << h_{drop}$

Filmwise Condensation: Nusselt Analysis

Idealizations:

- Laminar flow
- Constant properties
- Negligible subcooling of liquid
- Inertia effects negligible in momentum balance
- Vapor stationary, no drag
- Smooth liq-vap interface
- Heat transfer across film by conduction only

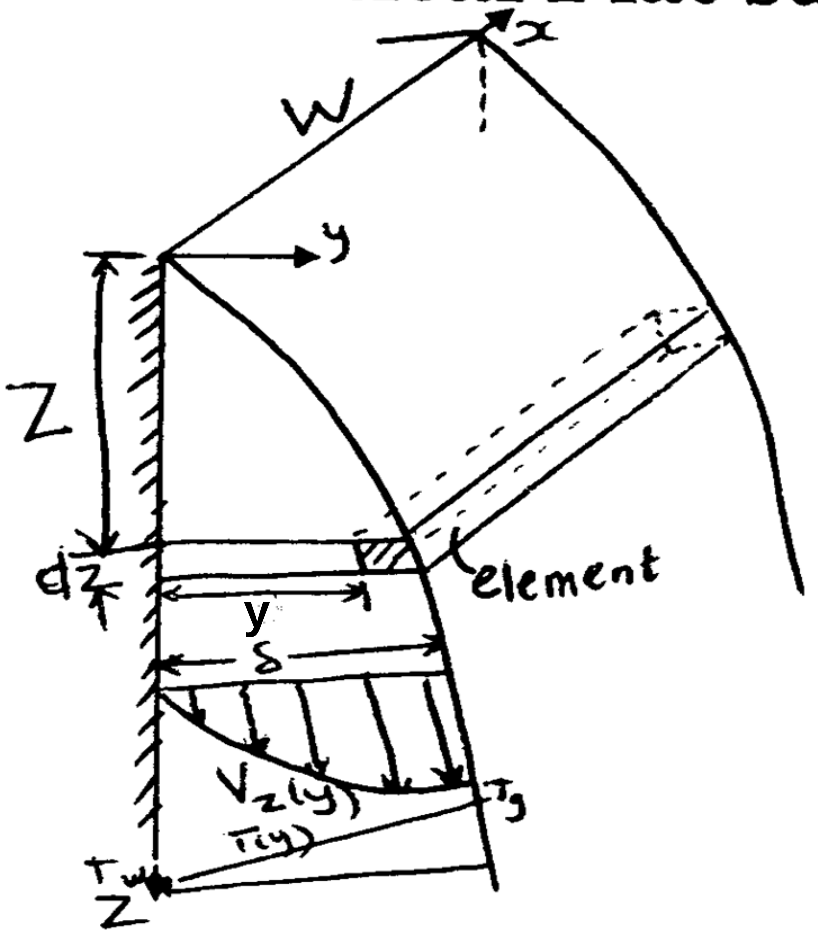


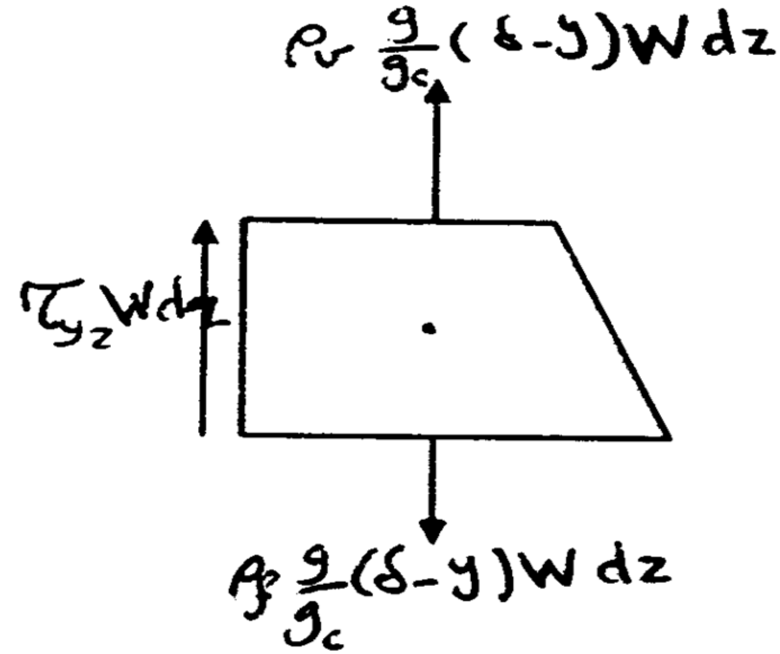
Smooth Film Analysis

- The first attempt to analyze the film-wise condensation problem was done by Nusselt in 1916.
- By making certain assumptions;
  - The flow of condensate in the film is laminar,
  - Fluid properties are constant,
  - Sub cooling of the condensate may be neglected,
  - Momentum changes through the film are negligible,
  - The vapor is stationary and exerts no drag on the condensate,
  - Heat transfer is by conduction only, and
  - Surface is isothermal.

Laminar Film Condensation

- **Vertical Flat surface**





2-dimension

Definitions

Film thickness = δ at any z location

Width of the plate = W

Vol of element = $(\delta-y)$ W dz

Types of forces acting on element

1. **gravity force "+ve downward"**
2. **buoyancy force "-ve"**
3. **friction due to viscosity "-ve"**

Force Balance on Element

Gravity Force =Buoyancy + Friction

$$\rho_f \frac{g}{g_c}(\delta - y)Wdz = \rho_v \frac{g}{g_c}(\delta - y)Wdz + \tau_{yz}(Wdz) \qquad (1)$$

T_g is the sat. temp. may be < ambient temp "T_∞" or = ambient temp "T_∞"

Since $\tau_{yz} = \mu \frac{dV_z}{dy}$ **. Substitute in eq. 1**

and rearranging

$$\frac{dV_z}{dy} = \frac{g}{\mu g_c}(\delta - y)(\rho_f - \rho_v)$$

Integrating gives

$$V_z = \frac{g}{\mu g_c}(\rho_f - \rho_v)(\delta y - y^2/2) + C_1$$

$$B.C \quad y = 0 \quad , V_z = 0 \quad \therefore C_1 = 0$$

$$\therefore velocity\,\text{Profile}\; V_z = \frac{\delta^2 g}{v_f}(1 - \rho_v/\rho_f)\left(\frac{y}{\delta} - \frac{y^2}{2\delta^2}\right) \qquad (2)$$

Note

Eq. 2 can be integrated over the cross section ' normal to the direction of the Z-directed velocity V_z' to get the mass flow rate of condensate, $\dot{m}_f$

$$\therefore \dot{m}_f = \iint \rho_f V_z \, dA$$

$$= \int_0^w \int_0^\delta \frac{\rho_f g}{\nu_f}\left(1 - \frac{\rho_v}{\rho_f}\right)\left(\delta y - \frac{y^2}{2}\right) dy \, dx$$

$$= \frac{w \delta^3 g \rho_f}{3 \nu_f}\left(1 - \frac{\rho_v}{\rho_f}\right) \quad \text{.........(3)}$$

Heat transfer equation

The heat flow at the wall within the area W dz is given by

$$q_y = -k_f (W dz) \left.\frac{dT}{dy}\right|_{y=0}$$

Assume a linear temp. profile within the film varies from Tw at the wall to Tg, the saturation temp. at the liq-vap interface.

$$q_y = -k_f W dz \frac{T_w - T_g}{\delta}$$

The amount of condensate bet. z and $z+dz$ is given by

Added Condensate in dz $= \dot{m}_f \big|_{z+dz} - \dot{m}_f \big|_z$

condensat in dz $= \frac{d\dot{m}_f}{dz} dz = \frac{d\dot{m}_f}{d\delta} \cdot \frac{d\delta}{dz} \cdot dz$

$= \frac{d\dot{m}_f}{d\delta} \cdot d\delta$

from eq. 3

$$\frac{d\dot{m}_f}{d\delta} \cdot d\delta = \frac{W \rho_f \delta^2 g}{\nu_f} (1 - \rho_v/\rho_f)\, d\delta \qquad \ldots\ldots (4)$$

Multiplying eq. 4 by h_{fg} gives q_y

$$\frac{W \rho_f \delta^2 g}{\nu_f}\left(1-\frac{\rho_v}{\rho_f}\right) h_{fg}\, d\delta = k_f W\, dz\, \frac{T_g - T_w}{\delta}$$

OR

$$\delta^3\, d\delta = \frac{k_f \nu_f (T_g - T_w)}{\rho_f\, g\, h_{fg} (1-\rho_v/\rho_f)}\, dz$$

Integrating

$$\delta^4 = \frac{4 k_f \nu_f Z (T_g - T_w)}{\rho_f g h_{fg} (1 - \rho_v/\rho_f)} + C_1$$

B.C at $Z = 0$, $\delta = 0$ $\therefore C_1 = 0$

$$\therefore \delta = \left[\frac{4 k_f \nu_f Z (T_g - T_w)}{\rho_f g h_{fg} (1 - \rho_v/\rho_f)} \right]^{1/4} \quad \text{-----(5)}$$

Express H. transfer in terms of Local conv. coeff. "h_z"

$$h_z \, w \, dz \, (T_w - T_g) = -k_f \, w \, dz \, \frac{T_g - T_w}{\delta}$$

OR

$$h_z = k_f / \delta$$

Substituting in eq. (5)

$$h_z = k_f \left[\frac{\rho_f \, g \, h_{fg} (1 - \rho_v/\rho_f)}{4 \, k_f \, \nu_f \, Z \, (T_g - T_w)} \right]^{1/4} \quad \ldots\ldots (6)$$

The Local Nusselt No is

$$Nu_z = \frac{h_z Z}{k_f} = \left[\frac{\rho_f \, g \, Z^3 h_{fg} (1 - \rho_v/\rho_f)}{4 \, k_f \, \nu_f \, (T_g - T_w)} \right]^{1/4} \quad \ldots\ldots (7)$$

Average Coeff. over the entire surface

$$\bar{h} = \frac{1}{LW} \int_0^W \int_0^L h_z \, dz \, dx$$

$$= \frac{1}{L} \int_0^L k_f \left[\frac{\rho_f g h_{fg} (1 - \rho_v/\rho_f)}{4 k_f \nu_f Z (T_g - T_w)} \right]^{1/4} dZ$$

Solving, we get

$$\bar{h} = \frac{4}{3} k_f \left[\frac{\rho_f g h_{fg} (1 - \rho_v/\rho_f)}{4 k_f \nu_f L (T_g - T_w)} \right]^{1/4} = \frac{4}{3} h_L \quad \ldots\ldots (8)$$

$$= 0.943 \left[\frac{\rho_f g k_f^3 h_{fg} (1 - \rho_v/\rho_f)}{\nu_f L (T_g - T_w)} \right]^{1/4} \quad \ldots (8)$$

Nu based on the average coeff.

$$Nu = \frac{\bar{h} L}{k_f} = \frac{4}{3} \left[\frac{\rho_f g h_{fg} (1 - \rho_v/\rho_f) L^3}{4 k_f \nu_f (T_g - T_w)} \right]^{1/4} \quad \ldots\ldots (9)$$

$$\therefore Re = \frac{V D_h}{\nu_f}$$

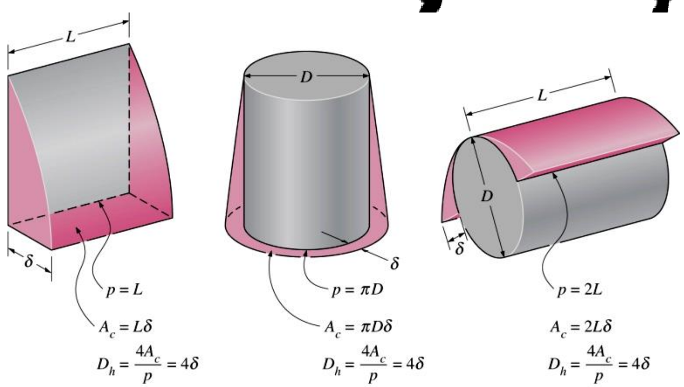
V : avg. film velocity

D_h : Hydraulic diam

$$D_h = \frac{4A}{p} = \frac{4\delta W}{W} = 4\delta$$

$$\therefore Re = \frac{V(4\delta)}{\nu_f}$$

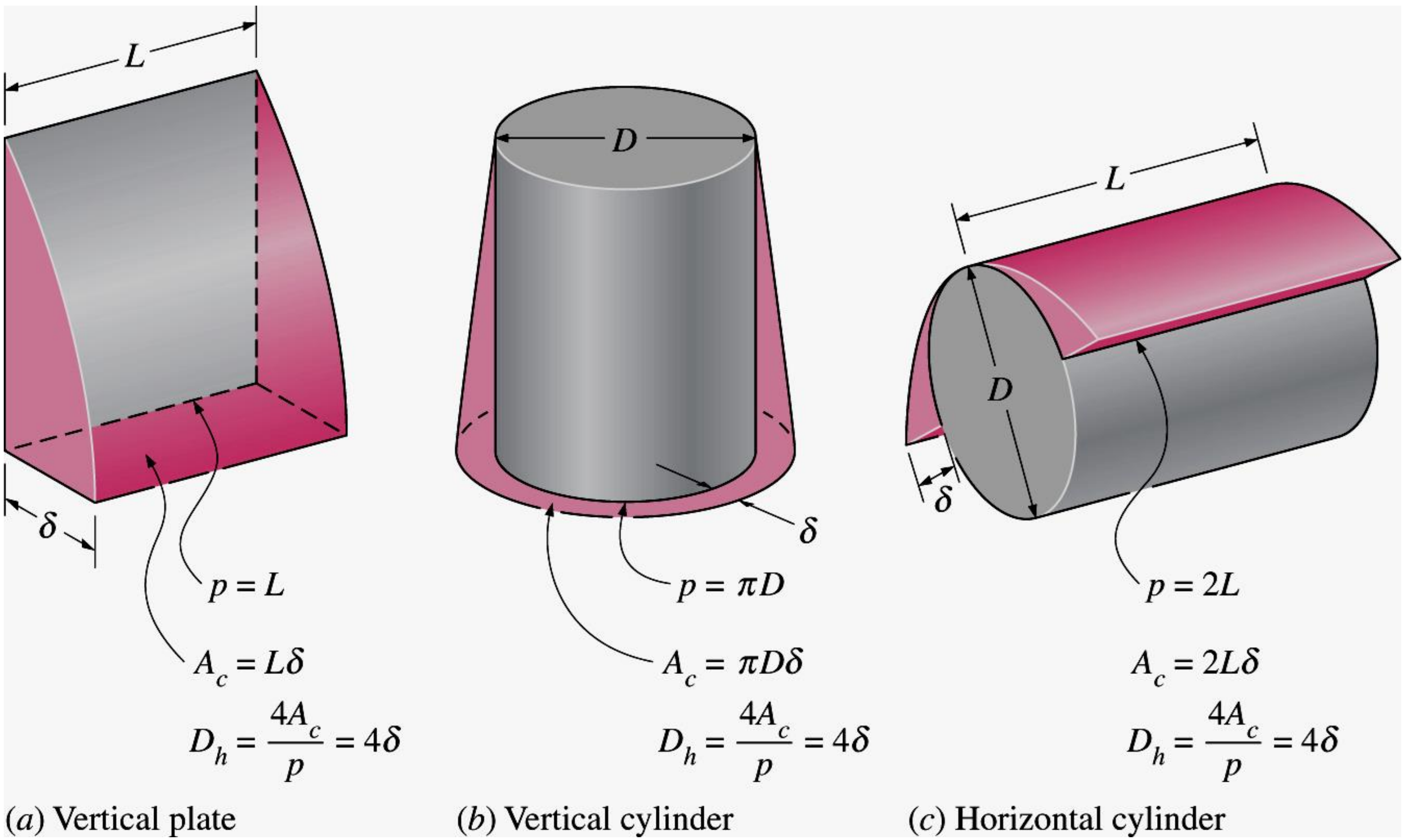
$$= \frac{\dot{m}_f}{\rho_f A} \cdot \frac{4\delta}{\nu_f} = \frac{4\dot{m}}{\rho_f(\delta W)} \frac{\delta}{\nu_f} = \frac{4\dot{m}_f}{W \rho_f \nu_f} \quad \text{......(1g}$$



(a) Vertical plate (b) Vertical cylinder (c) Horizontal cylinder

for laminar flow $R_e < 1800$

The wetted perimeter p, the condensate cross-sectional area A_c, and the hydraulic diameter D_h for some geometries



Note 1

Re can be written in terms of Conv. coef

$$\therefore q = h_L A_s (T_g - T_w) \quad \cdots\cdots\cdots (11)$$

where

A_s : Surface area of plate in contact with the film

but

$$q = \dot{m}_f h_{fg}$$

$$\therefore \dot{m}_f = \frac{h_L A_s (T_g - T_w)}{h_{fg}} \quad \cdots\cdots\cdots (12)$$

Substituting into eq. (10)

$$\therefore Re = \frac{4 h_L A_s (T_g - T_w)}{W h_{fg} \rho_f \nu_f} \quad \cdots\cdots\cdots (13)$$

Note 2

- Experimental values of the convective coefficient can be as much as 20 % higher than those predicted by eq. (8) for laminar flow;

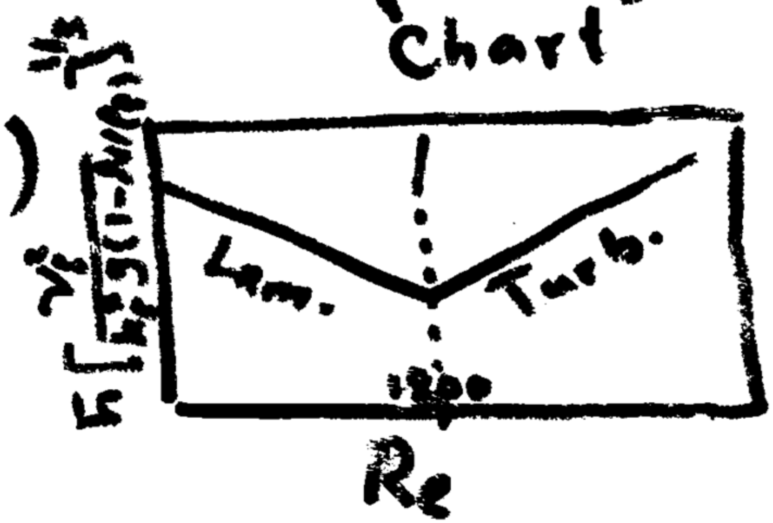
$$\bar{h} = 1.13\left[\frac{\rho_f k_f^3 h_{fg} g(1 - \rho_v / \rho_f)}{\upsilon_f L(T_g - T_w)}\right]^{1/4}$$

Turbulent Film Condensation on a Vertical Flat surface

$$\bar{h} = 0.0077\, k_f \left[\frac{g(1-\rho_v/\rho_f)}{\nu_f^2}\right]^{1/3} Re_\delta^{0.4} \quad \ldots\ldots (14)$$

$$Re_\delta = \frac{4\dot{m}_f}{W \rho_f \nu_f} > 1800$$

"Chart"



$$q = \bar{h} A_s (T_g - T_w)$$

and

$$\dot{m}_f = q/h_{fg}$$

Laminar Film Condensation on an Inclined Flat Surface

use the vertical-plate eqs; but replace g with $g \sin\theta$. 

where θ is the angle of inclination of the plate with horizontal.

Film Condensation on a Vertical Tube

Correlations of vertical plate can be used for vertical tube if the film thickness < outside diam.

$$Re_\delta = \frac{4\dot{m}_f}{\pi D \mu_f}$$

πD is the wetted perimeter

Notes

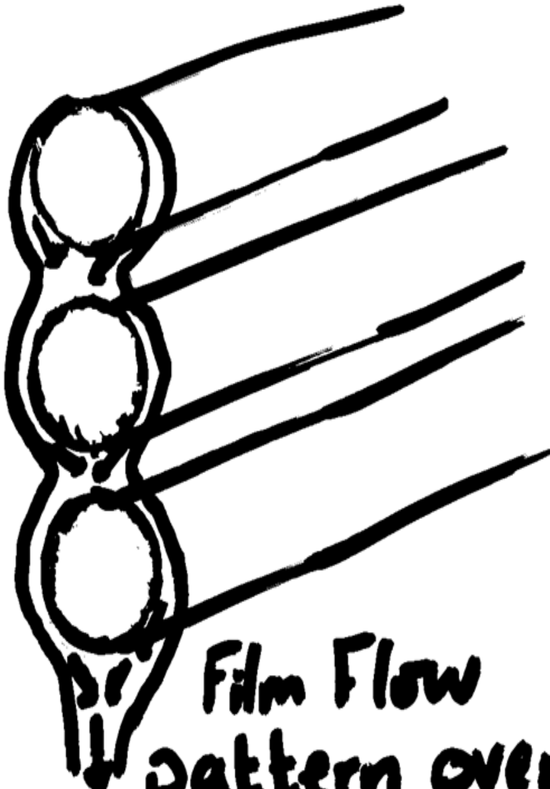
1. Tube loading, $\Gamma = \frac{\dot{m}_f}{\pi D}$ Condensate per unit tube perimeter

2. Plate loading, $\Gamma = \frac{\dot{m}_f}{W}$

$\therefore Re_\delta$ (tube) $= \frac{4\Gamma}{\mu_f}$

3. for a tube bundel, $\Gamma = \frac{\dot{m}_f}{N_t \pi D}$

Film Condensation on a horizontal tube and a horizontal tube bank



Film Flow pattern over tube bank

film flow pattern for Singk tube 'horizontal'

For one tube

$$\bar{h} = 0.728\left[\frac{g\,\rho_f(1-\rho_v/\rho_f)k_f^3\,h_{fg}}{\nu_f\,(T_g-T_w)\,D}\right]^{1/4} \quad \text{laminar filmwise} \quad \cdots\cdots(15)$$

where D: outside diam

A comparison of Eq (8) & (15) for a condensation on a vertical tube of Length L and a horizontal tube of diam D, gives

$$\frac{\bar{h}_{vert}}{\bar{h}_{horz}} = 1.3\left(\frac{D}{L}\right)^{1/4} \qquad \text{For a given } (T_g - T_w)$$

The two coeff.s become equal when L = 2.87 D {vertical tube of length L ≈ 3 times the diam of horizontal tube}.

when $L = 100D \Rightarrow \bar{h}_{horz} = 2.44 \bar{h}_{vert}$

Therefore, horizontal tube arrangements are preferred to vertical tube arrangements in condenser design.

For j tubes vertically above each other

$$\bar{h} = 0.728\left[\frac{g\rho_f(1-\rho_v/\rho_f)k_f^3 h_{fg}}{\nu_f(T_g-T_w)\,j\,D}\right]^{1/4}$$

OR

$$\bar{h}_j = \bar{h}\cdot j^{-1/4} \quad \text{.....(16)}$$

<u>Note</u>

In practice, it is better to use index 1/6 instead of 1/4

$$h\Big|_{\substack{\text{tube}\\ \text{bundle}}} = \bar{h}\cdot j^{-1/6} \quad \text{----(17)}$$

Comparison

Average Heat transfer Coefficient

- For a vertical flat plate:

$$h_{vert} = 0.943\left(\frac{g\rho_l(\rho_l - \rho_v)h_{fg}k_l^3}{\mu_l(T_{sat} - T_s)L}\right)^{\frac{1}{4}}$$

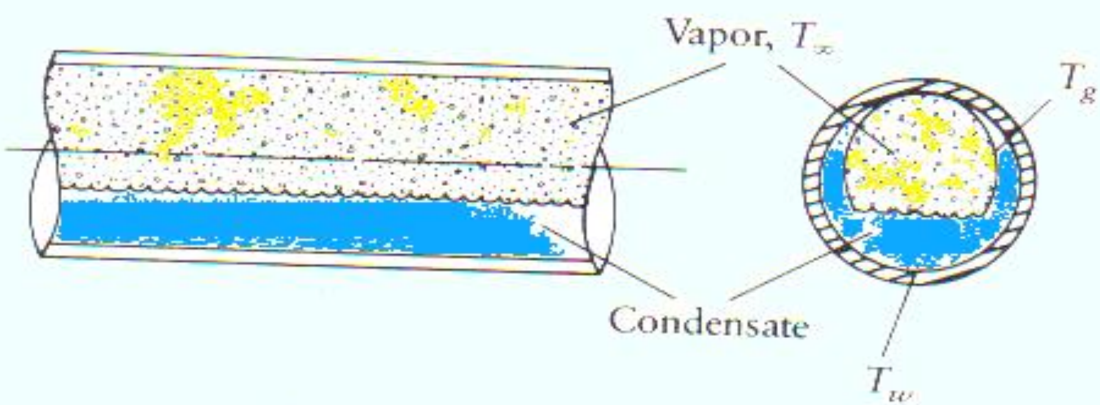
- For a horizontal tube:

$$h_{horiz} = 0.729\left(\frac{g\rho_l(\rho_l - \rho_v)h_{fg}k_l^3}{\mu_l(T_{sat} - T_s)D}\right)^{\frac{1}{4}}$$

Film condensation within Horizontal Tubes

- **For low vapor velocities, the following expression for condensation of refrigerants can be used**

$$\mathrm{Re}_D = \frac{V_v I D_t}{\nu_v} < 35\,000$$



Condensation in a tube conveying a two-phase mixture at a low flow rate the Chato Equation has been derived:

$$\bar{h}_D = 0.555\left[\frac{g\rho_f(1-\rho_v/\rho_f)k_f^3 h'_{fg}}{\nu_f(T_g - T_w)ID_t}\right]^{1/4}$$

where V_v is the average velocity of the vapor, ID_t is the inside diameter of the tube, and

$$h'_{fg} = h_{fg} + \tfrac{3}{8}c_{pf}(T_g - T_w)$$

Reynolds number evaluated at the inlet temperature

Liquid properties at $(T_g + T_w)/2$

For higher flow rates, the flow becomes quite complicated. Correlations are available in the literature, however.