

Chapter II



LITERATURE REVIEWS AND THEORETICAL CONSIDERATION

2.1 Evaporators

According to Foust et al⁽¹⁾, evaporation is the operation of concentrating a solution by boiling away the solvent. The concentration is normally stopped before the solute begins to precipitate from solution. So the purpose of the evaporator is to permit the transfer of heat, the evaporation of solvent, and the removal of fluid droplets from the vapor before it leaves the evaporator.

Earles⁽²⁾, Biilig, and Werner⁽³⁾ stated that the evaporator had two principal functions, to exchange heat and to separate the vapor that was formed from the liquid. However, it usually consists of three functional units:

- 1) the heat exchanger, to supply sensible and latent heat to the liquid.
- 2) the evaporating unit, where the fluid boils and evaporates,
- 3) the separator in which the vapor leaves the liquid and passes off to the condenser.

Bennett⁽⁴⁾ explained that the simplest device is an open pan or kettle which receives heat from a coil or jacket or by direct firing underneath the pan. A some

what more complicated apparatus is the horizontal tube evaporator in which a liquid in the shell side of a closed, vertical, cylindrical vessel is evaporated by passing steam or other hot gas through a bundle of horizontal tubes contained in the lower part of the vessel. In both devices nucleate boiling occurs in the solution at the heated surface, and any circulation of the liquid that occurs is a result of the heating. The liquid level in the horizontal-tube evaporator is usually at less than half the height of the vessel; the empty space permits disengagement of entrained liquid from the vapor passing overhead.

A more efficient device is the vertical tube evaporator. This consists of a vertical, cylindrical vessel with a bundle of vertical, steam-heated tubes at the base of the vessel. However, the bundle of tubes operates with steam around the tubes and the liquid to be evaporated within the tubes. Essentially, the bundle acts as a vertical shell and tube heat exchange in which steam is admitted to the shell side and the evaporating liquid is on the tube side. After disengagement, the vapor is removed from the top of the vessel and the liquid remaining flows down in an annular space between the tube bundle and the wall of the vessel to the space beneath the tube bundle, where it begins the cycle once again. These evaporator are fed continuously, and the concentrated liquor is removed continuously from the bottom of the vessel. The apparatus described above is

often referred to as a short-vertical-tube, basket-type evaporator and is one of the common types. The flow is caused entirely by natural convection.

An evaporator which works in similar fashion is the long-tube, natural circulation evaporator. Operating on somewhat the same principle as a smokestack, the long-tube evaporator gives higher fluid velocities than the short tube evaporator. This high velocity is desirable because it promotes higher heat-transfer coefficients between the evaporating fluid and the tube walls. A further improvement is sometimes made by pumping the circulating liquid around to give even higher velocities and higher heat transfer coefficients than can be obtained by natural convection. These forced-circulation evaporators are especially useful for evaporating viscous liquids for which natural circulation rates would be low.

Perry⁽⁵⁾ concluded that evaporators may be classified as follows:

1. Heating medium separated from evaporating liquid by tubular heating surfaces.
2. Heating medium confined by coil, jackets, double walls, flat plates, etc.
3. Heating medium brought into direct contact with evaporating liquid
4. Heating by solar radiation

He also classified various types of evaporator and

its diagram was shown in Fig 2.1. Among the most important of recent designs is the falling-film evaporator, in which a film of solution flows down the surface of a tube and receives heat through the tube wall. The short exposure time in the falling-film evaporator makes it useful for concentrating solutions of heat-sensitive materials.

The selection of equipment for particular service is governed by the characteristics of the dilute and concentrated liquids like heat sensitiveness, crystallisation, salting, corrosion, scaling, foaming and viscosity etc.

2.2 Film Evaporators

Liquid films are of importance in a wide variety of industrial processes. Falling films occur, for instance, in a number of types of counter current mass transfer equipment and climbing films are of importance in the context of heat transfer in evaporating systems. Liquid films are clearly important in the lubrication of bearings, gears and other moving parts. There are also a number of processes such as enamelling and painting where the deposition of thin layers of liquid is necessary.

On evaporation of heat sensitive materials, essential operating conditions are the lowest possible evaporation temperature and the minimum time of retention. The minimum evaporation temperature is fixed by the temperature of

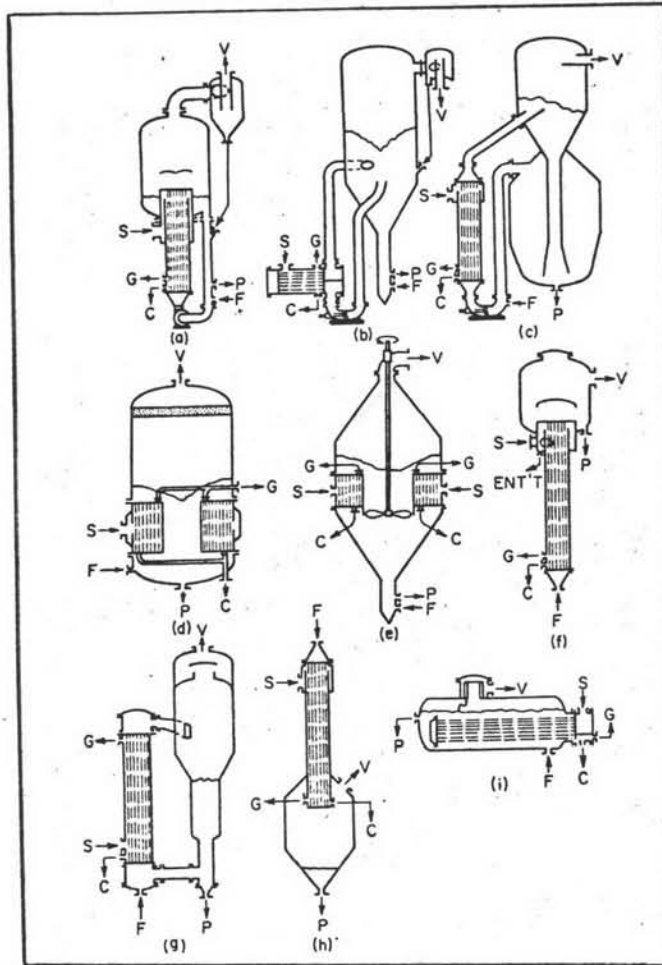


Fig. 2.1 Evaporator types. (a) Forced circulation. (b) Submerged-tube forced circulation. (c) Oslo-type crystallizer. (d) Short-tube vertical. (e) Propeller calandria. (f) Long-tube vertical. (g) Recirculating long-tube vertical. (h) Falling film. (i) Horizontal-tube evaporator. C, condensate; F, feed; G, vent; P, product; S, steam; V, vapor; ENT'T, separated entrainment outlet.

the condenser coolant while in order to keep the heat transfer surface reasonably small there must be a sizeable temperature difference between vapor and the cooling medium. However, if the liquid had a high boiling point this minimum temperature may be limited by the solidifying point and on the other hand the use of low pressure for the evaporation will increase capital costs through the expense of the vacuum pumps. These components, apart from adding an element of operational unreliability, will also demand connections and pipes of large diameter and special attention to joints to prevent leakage. There are therefore a number of conflicting demands which eventual compromise in design must recognise; the fixing of pressure and temperature for a given evaporation should result from an assessment of the economic factors, the short residence time condition being especially desirable.

The shortest residence time can be achieved if liquid is spread in the form of a thin layer over the heating surface and vaporised to a desired state in a single pass through the evaporator.

Skocylas⁽⁶⁾ described that vaporization of liquids in the form of thin films was first applied in molecular distillation at the pressure of 10^{-3} mm Hg to 10^{-4} mm Hg. This process is used for evaporation of substances with high molecular weight, boiling at high temperatures,

sensitive to even small increments in temperature and if a high degree of purity is required (for instance, in case of vitamins). The comparison between the process carried out in a distillation kettle, a conventional circulation evaporator and a film evaporator is given in Table 2.1.

Table 2-1

Type of apparatus	Concentration in liquid subjected to evaporation	Residence time
Distillation column	Increases with time	5 hr to 50 hr
Conventional circulation evaporator	Constant	15 min to 60 min
Film-evaporator	Increases along the film	3 sec to 20 sec

It should be noted that the film evaporator besides giving a shorter residence time also makes possible a higher concentration; the higher the concentration the greater is the risk of decomposition. Contrary to the conventional circulation evaporation and flash distillation which are carried out to reach final concentration in film evaporation, as in the case of a simple distillation, the maximum concentration and the highest temperature are reached at the end, at the exit from the apparatus. Further advantages of film evaporation, as

in the case of a simple distillation, the reduced length of path for diffusion of the low boiling component from the bulk to the surface of the liquid (thickness of the liquid layer is of the order of 1 mm only, while in a distillation kettle it may be 1 m or even more).

With respect to the mechanism of spreading of liquid over the heating surface, Skocylas⁽⁶⁾ concluded that film-evaporators can be divided into two groups, static and mechanical evaporators.

The simplest type of static apparatus is the falling film evaporator. In this the liquid, under the action of gravity, flows down along a heating surface which has usually the shape of cylinder. See Fig 2.2. According to Schneider⁽⁷⁾, application of this type of apparatus is justified only at high liquid loads and if high purity of the final product is not essential. It is difficult to achieve uniform distribution of liquid over the surface because through the rapid vaporisation of the liquid, the film is often broken, channelling takes place and some parts of the surface remain unwetted. Jones⁽⁸⁾ describing this type of evaporator concluded that it can be used only for nonviscous liquids, since the higher the viscosity the more difficult it is to achieve uniform liquid distribution over the surface. A viscous liquid flows in the form of rivulets thereby leaving parts of the surface unwetted, the process of vaporisation

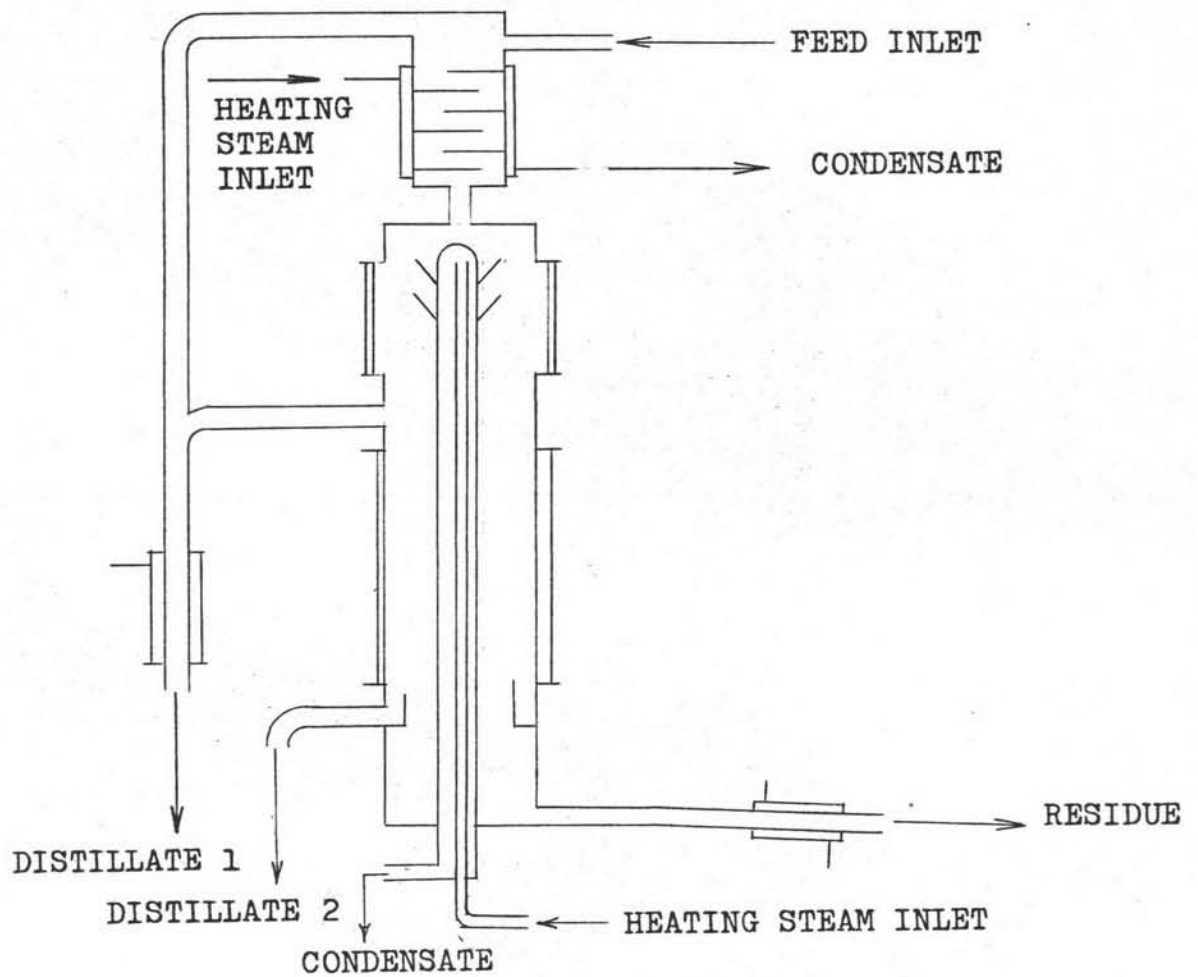


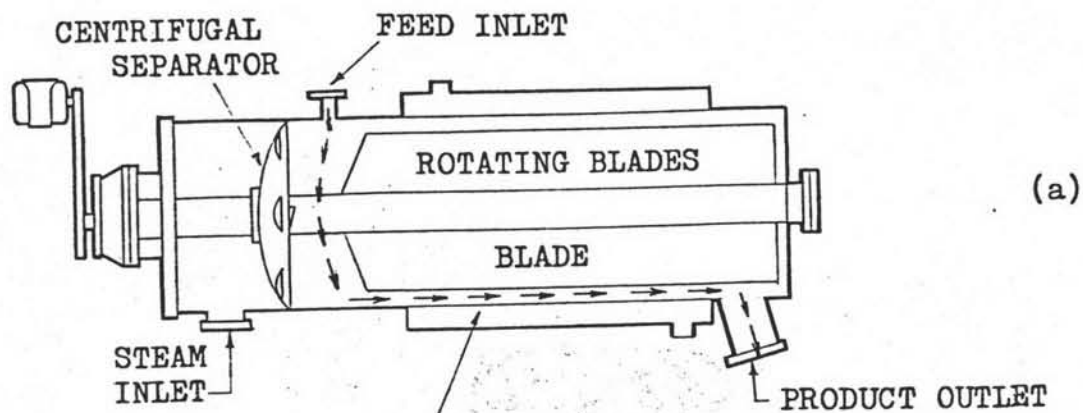
Fig. 2.2 Static film evaporator with gravitational liquid flow along the heating surface

becomes irregular, resulting in variations in the composition of the final product.

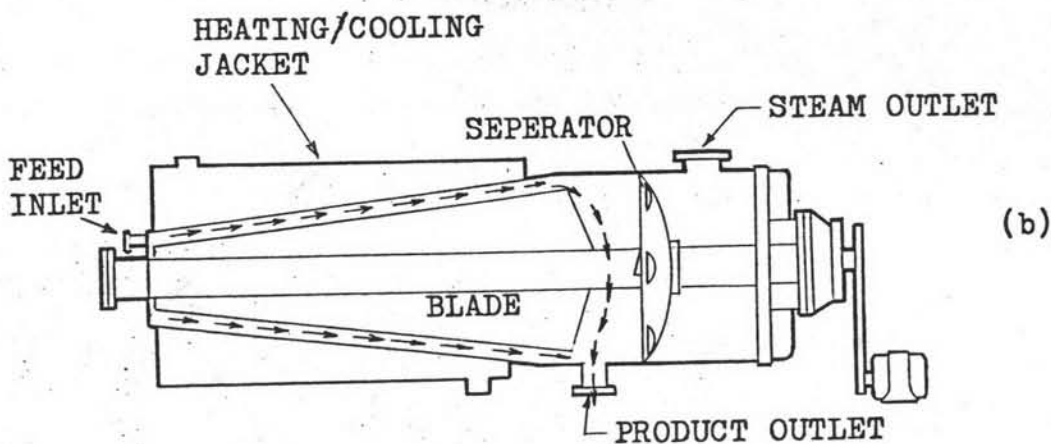
Because of difficulties arising from maldistribution of liquid over the heating surface, static film apparatus cannot compete with mechanical evaporators. There are several types of mechanical evaporators differing from each other in the manner of introducing the feed or withdrawing the product. Some of them was shown schematically in Fig 2.3. 005106

The apparatus uses turbulent mixing of the film during evaporation generated by the action of rotor blades revolving inside a cylinder or frustum of cone; centrifugal force created by the rotation produces the turbulent mixing of the film on the heating surface. Liquid hold up in the apparatus is small and the residence time is a few seconds only. A vertical cylindrical evaporator is particularly suitable for evaporating medium of high viscosity liquids. The design with the shape of a frustum of cone can be operated either in vertical or horizontal position. In the first evaporator the centrifugal force accelerates the liquid while in the second apparatus it has a retarding effect. The effect of viscosity which is significant in the case of a static film evaporator is reduced to a minimum because of intensive mixing of the liquid film.

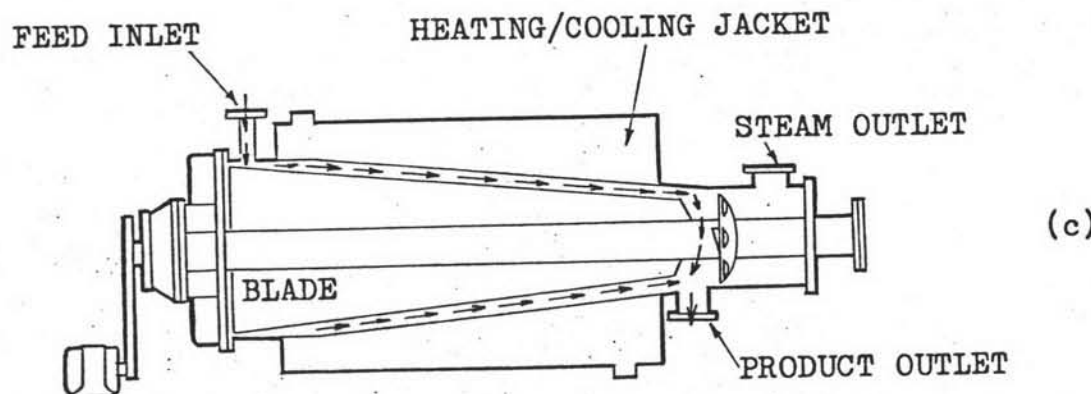
The selection of the appropriate evaporator is ultimately a matter of economics. The static film evaporators



(a)



(b)



(c)

Fig. 2.3 Mechanical film evaporator

- (a) cylindrical apparatus with gravitational liquid flow
- (b) conical apparatus with accelerated liquid flow
- (c) conical apparatus with slowed-down liquid flow

would be selected because of their simplicity of construction, compactness and generally high heat transfer coefficient. In the mechanical film evaporators, the cost is often believed to be prohibitive and the technique is considered too complex to be effective and reliable, due to moving parts and the high vacuum generally associated with it. However, the properties of the material being evaporated may sharply limit the choice for instance, a highly viscous solution might not move readily past the heating surface by the action of natural convection alone. In this case the use of the mechanical film evaporator might be necessary.

2.3 Vertical Falling Film Evaporator

2.3.1 Principle of Vertical Falling Film Evaporator

Agarwal⁽⁹⁾ explained that in these evaporators, the dilute liquids flow downward on a heating plate or heating tube in the form of a thin film as well as the influence of the descending vapor created by the evaporation of the liquid. There is no static height in the tube, but only an acceleration of the downward flow of the liquid film by the force of gravity. The vapours formed in the tubes pass in the same downward direction at very high speeds and thus greatly accelerate the downward flow of the liquid film. The high speed of both vapour and liquid

results in shorter exposure of the product of heat, usually less than a minute.

McCabe⁽¹⁰⁾ added that vapor evolved from the liquid is usually carried downward with the liquid, and leaves from the bottom of the unit. In appearance these evaporators resemble long, vertical, tubular exchangers with a liquid vapor separator at the bottom and a distributor for the liquid at the top.

2.3.2 Design Consideration

Wiegand⁽¹¹⁾ explained that, in the evaporators, the liquid was caused to flow down the heating surface, covering the surface of the tube with a thin layer of liquid. It was usually to operated under reduced pressure.

Bielig⁽³⁾ pointed out that the liquid to be concentrated must enter the evaporator at a temperature very close to the boiling point in the evaporator.

Wiegand⁽¹¹⁾ added that if this condition was not fulfilled, evaporating could not occur, but only preheating took place.

Chun and Seban⁽¹²⁾ pointed out that the minimum feed rate was limited by the occurrence of circumferentially nonuniform wetting, with the formation of accompanying rivulets on the surface of the tube. Non wetting would lead to serious local overheating. The maximum feed rate was limited by overshooting of the feed at the distributor,

so that all the liquid would not attach itself to the tube surface. Gray⁽¹³⁾ added that it was necessary to design the evaporators properly for a certain minimum liquid rate per unit width of heat transfer surface to be maintained.

Wiegand⁽¹¹⁾ stated that the size of heated surface, required for predetermined evaporating output, was proportional to the difference between the heating and boiling temperature. He also pointed out that the narrower the surface, the greater was the so-called "liquid load". Wiegand⁽¹¹⁾ and Gray⁽¹³⁾ found that this liquid load must never be too small, as otherwise there was a danger of dry spots forming on the heating surface, with consequent burning of the liquid, increasing fouling, and resulting poor heat transfer coefficients. Moreover Weigand⁽¹¹⁾ pointed out that, higher liquid loads resulted in a better heat transfer from the heating surface to the liquid to be evaporated. The temperature difference between the heating and boiling temperature had necessarily to be of a certain degree to avoid the cease of evaporation.

2.3.3 Theoretical Consideration

A liquid solution is being fed on a vertical heated tube at its boiling temperature. As the solution is flowing

downward in a form of thin film, under the gravitational force, it gains heat from the heated tube, water in the solution vaporised, the concentration of the solution increases, and the thickness of the film might change.

Let the consideration of the system be assigned as shown in Figure 2.4. It is assumed that the temperature gradient in y direction is negligible. Therefore, all the heat used for vaporisation of water is being transferred in x direction from the plate surface to the vapor-liquid interface. If the liquid film is very thin, the flow is laminar, consequently, the heat is conducted in x direction.

The energy balance of a differential element dy , is

$$dq = k \pi D dy \frac{T_s - T_v}{x_e} \quad (2.1)$$

where q is the rate of heat transfer

k is the thermal conductivity of the liquid

D is the diameter of the tube

T_s is the surface temperature of the tube

T_v is the vaporization temperature of the liquid

x_e is the liquid film thickness at $y=y$

However, from the conventional definition of heat transfer coefficient h ,

$$dq = h \pi D dy (T_s - T_v) \quad (2.2)$$

$$\text{Therefore } h = \frac{k}{x_e} \quad (2.3)$$

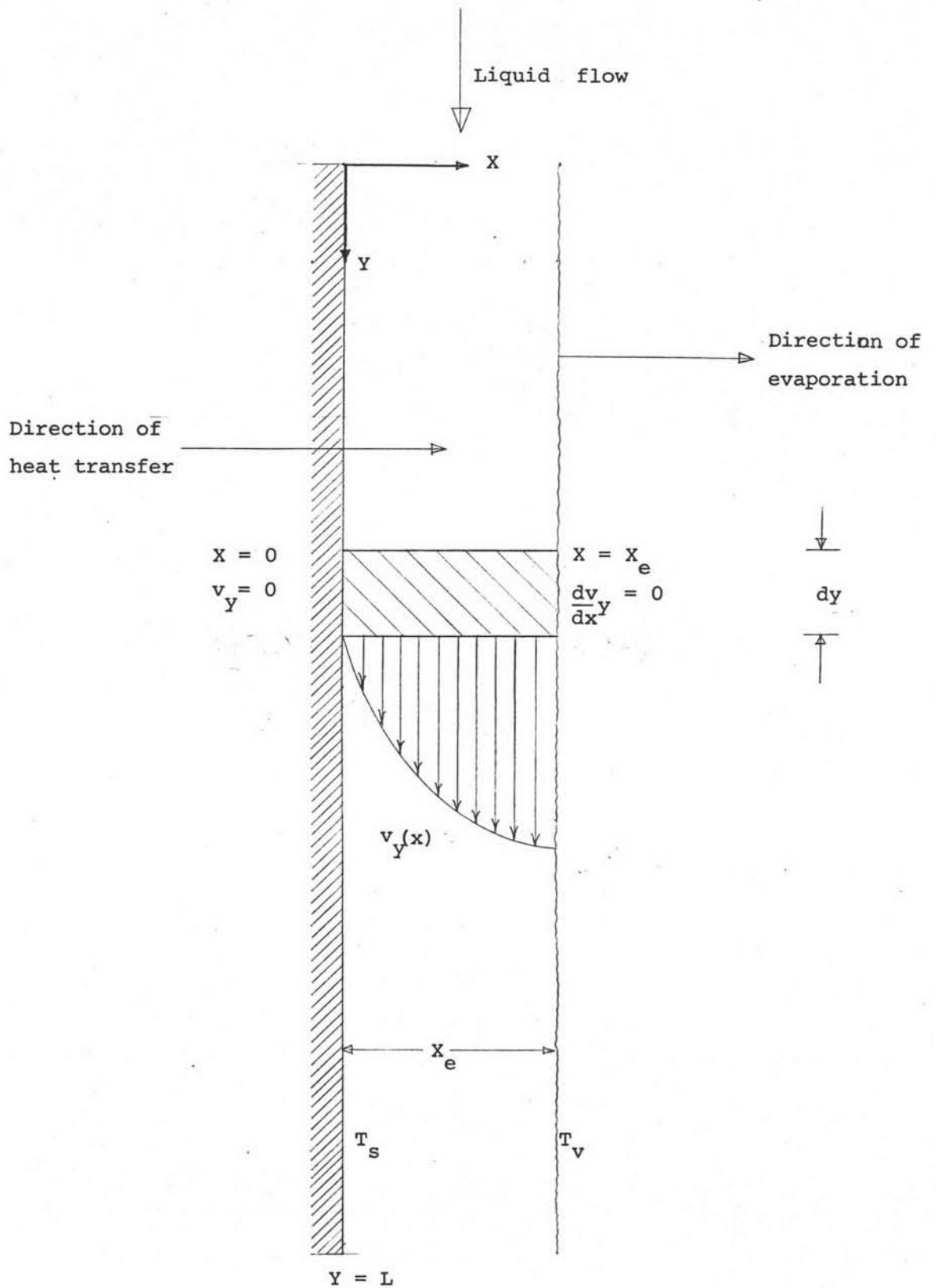


Fig. 2.4 Evaporation on a vertical heated tube

The amount of heat transferred may be determined from the change in liquid flow rate

$$dq = -\lambda d\omega \quad (2.4)$$

where ω is the liquid flow rate at $y = y$

λ is the heat of vaporization per unit mass

Then eq. 2.2 becomes

$$\begin{aligned} -\lambda d\omega &= h \pi D dy (T_s - T_v) \\ h &= -\frac{\lambda}{T_s - T_v} \frac{1}{\pi D} \frac{d\omega}{dy} \\ \frac{k}{x_e} &= h = -\frac{\lambda}{T_s - T_v} \frac{d\Gamma}{dy} \quad (2.5) \end{aligned}$$

where Γ is the flow rate of liquid per unit length of the tube perimeter

$$T_s - T_v = -\frac{x_e \lambda}{k} \frac{d\Gamma}{dy} \quad (2.6)$$

For the entire length L of the tube, let a mean heat transfer coefficient, h_m , be defined as

$$\begin{aligned} q &= h_m \pi DL (T_s - T_v) = \lambda (\omega_o - \omega_L) \\ &= \lambda \Delta \omega \quad (2.7) \end{aligned}$$

Where ω_o is the flow rate of the feed solution

ω_L is the flow rate of the concentrated solution

$$T_s - T_v = \frac{\lambda \Delta \omega}{h_m \pi DL} = \frac{\lambda \Delta \Gamma}{h_m L} \quad (2.8)$$

Combining equation 2.6 + 2.8 gives

$$d_y = \frac{-x_e h_m L}{k \Delta \Gamma} d\Gamma \quad (2.9)$$

For mass balance of the differential element, dy , rectangular coordinates can be used instead of cylindrical coordinate because the liquid film thickness is very small as compared to the radius of the tube.

According to the continuity equation

$$\frac{\partial \rho}{\partial t} + \frac{\partial (\rho v_x)}{\partial x} + \frac{\partial (\rho v_y)}{\partial y} + \frac{\partial (\rho v_z)}{\partial z} = 0 \quad (2.10)$$

For incompressible fluid and one dimensional steady flow,

$$\rho \text{ is constant, } \frac{\partial \rho}{\partial t} = 0 \text{ and } v_x = v_z = 0$$

$$\text{Therefore } \frac{\partial v_y}{\partial y} = 0 \quad (2.11)$$

According to the equation of motion (Navier stoke's equation in y direction

$$\begin{aligned} \frac{\partial v_y}{\partial t} + v_x \frac{\partial v_y}{\partial x} + v_y \frac{\partial v_y}{\partial y} + v_z \frac{\partial v_y}{\partial z} \\ = \frac{g_c}{\rho} \frac{\partial p}{\partial x} + \nu \left(\frac{\partial^2 v_y}{\partial x^2} + \frac{\partial^2 v_y}{\partial y^2} + \frac{\partial^2 v_y}{\partial z^2} \right) + g_y \end{aligned} \quad (2.12)$$

For steady state one-dimensional flow without an external forces except the gravity, equation 2.12 becomes

$$\frac{d^2 v_y}{dx^2} = \frac{-g_y}{\nu} = \frac{-g}{\nu} \quad (2.13)$$

To solve this equation, two boundary conditions are required. They are

$$\text{B.C 1 : at } x = x_e \quad \frac{dv_y}{dx} = 0$$

$$\text{B.C 2 : at } x = 0 \quad v_y = 0$$

Integrating equation 2.13 gives

$$\frac{dv_y}{dx} = \frac{-gx + c_1}{\nu} \quad (2.14)$$

Apply B.C 1 yields

$$c_1 = \frac{g x_e}{\nu} \quad (2.15)$$

$$\text{Then } \frac{dv_y}{dx} = \frac{g}{\nu} (x_e - x) \quad (2.16)$$

Integrating equation (2.16) gives

$$v_y = \frac{g}{\nu} (x_e x - \frac{x^2}{2}) + c_2 \quad (2.17)$$

Apply B.C 2 yields

$$v_y = \frac{g}{\nu} (x_e x - \frac{x^2}{2}) \quad (2.18)$$

The mass flow rate of the solution at any value of y is

$$\begin{aligned} \omega &= \int_0^{x_e} v_y \rho \, \delta x \\ &= \frac{\rho g}{\nu} \int_0^{x_e} (x_e x - \frac{x^2}{2}) dx \\ &= \frac{\rho g}{\nu} \frac{x_e^3}{3} \end{aligned} \quad (2.19)$$

$$\text{Therefore } \Gamma = \frac{\omega}{\delta} = \frac{\rho g x_e^3}{3\nu} \quad (2.20)$$

$$x_e = \left(\frac{3\nu\Gamma}{\rho g} \right)^{1/3} \quad (2.21)$$

Substituting x_e into equation 2.9 yields

$$dy = \frac{-h_m L}{k \Delta T} \left(\frac{3\nu T}{\rho g} \right)^{1/3} dT \quad (2.22)$$

Integrating gives

$$\int_0^L dy = \frac{-h_m L}{k \Delta T} \left(\frac{3\nu}{\rho g} \right)^{1/3} \int_{T_0}^{T_L} T^{1/3} dT$$

$$L = \frac{h_m L}{k \Delta T} \left(\frac{3\nu}{\rho g} \right)^{1/3} \cdot \frac{3}{4} (T_0^{4/3} - T_L^{4/3})$$

$$\begin{aligned} h_m &= \frac{4}{3} \frac{k \Delta T}{T_0^{4/3} - T_L^{4/3}} \left(\frac{\rho g}{3\nu} \right)^{1/3} \\ &= 0.925 \frac{\Delta T}{T_0^{4/3} - T_L^{4/3}} \left(\frac{\rho g k^3}{\nu} \right)^{1/3} \quad (2.23) \end{aligned}$$

From equation 2.8

$$h_m = \frac{\lambda \Delta T}{L(T_s - T_v)} \quad (2.24)$$

Combining equation 2.23 + 2.24 gives

$$\begin{aligned} T_0^{4/3} - T_L^{4/3} &= 0.925 \frac{L(T_s - T_v) (\rho k^3 g)^{1/3}}{\lambda \nu} \\ &= 0.925 \frac{L(T_s - T_v) (\rho^2 k^3 g)^{1/3}}{\lambda \mu} \quad (2.25) \end{aligned}$$

Equation 2.25 is derived based on the assumption that the physical properties of liquid, density, viscosity, thermal conductivity, and heat of vaporization, temperature different $T_s - T_v$ are constant. In fact, these values are not constant but vary to a certain extent. However, if their mean values are used the result might be agreeable with the equation.

2.3.4 Heat Transfer in Vertical, Film Evaporator

Vapor production from a thin film of liquid on a vertical heated surface may occur as bubble formation at the liquid-solid interface or directly by evaporation from the liquid-vapor interface. Available observations support the latter mechanism for modest temperature difference between the solid and vapor ($\Delta T < 16^\circ\text{F}$). If bubble formation makes negligible contribution to vapor evolution, the principal resistance to heat transfer will occur within the liquid film; several empirical and theoretical relations have been presented for heat transfer by conduction through this film of fluid.

McAdams⁽¹⁴⁾ et al showed that heat transfer to subcooled falling water films was a function of the water flow rate substantially independent of tube length, tube diameter, temperature difference, and inlet fluid temperature. The recommended correlation was

$$h = 120 \Gamma^{1/3} \quad (2.26)$$

for values of film Reynolds number (Γ/μ), from 400 to 5000 and maximum water temperature of 200°F.

Nusselt⁽¹²⁾ proposed an equation for local heat transfer coefficients in liquid film for laminar flow as follows:

$$h = \frac{k}{\delta} = \left(\frac{4}{3}\right)^{1/3} \left(\frac{k^3 g}{\mu^2}\right)^{1/3} \left(\frac{4\Gamma}{\mu}\right)^{-1/3} \quad (2.27)$$

Kapitza⁽¹⁵⁾ predicted that capillary waves would form on the laminar layer surface when the Reynolds number exceeded the value of

$$\left(\frac{4\Gamma}{\mu}\right)_i = 2.43 \left(\frac{\mu^4 g}{\rho \delta^3}\right)^{-1/11} \quad (2.28)$$

$\mu^4 g / \rho \delta^3$ is the Kapitza number, Ka.

The analysis of Kapitza, presuming wavelengths long compared to layer thickness, indicated that the average layer thickness for a given flow rate is reduced by the action of the ripples, so that the heat transfer coefficient is increased.

Zazuli⁽¹⁵⁾ examined condensation data and formulated an empirical correction factor for condensation data in terms of average heat transfer coefficients.

$$(\bar{h})_{\text{ripples}} = 0.8 \left(\frac{\Gamma}{\mu}\right)^{0.11} \bar{h}_{\text{Nusselt}} \quad (2.29)$$

By assuming, for the condensation problem, $h \sim \left(\frac{\Gamma}{\mu}\right)^n$, it is found that the local coefficient to be consistent with equation (2,29) must be

$$h = 0.606 \left(\frac{k^3 g}{\nu^2}\right)^{1/3} \left(\frac{\Gamma}{\mu}\right)^{-0.22} \quad (2.30)$$

or assuming that the increase in local heat transfer coefficient arises entirely from a reduction in the average film thickness,

$$\frac{\delta_{\text{ripples}}}{\delta_{\text{Nusselt}}} = 1.33 \left(\frac{4\Gamma}{\mu}\right)^{-0.11} \quad (2.31)$$

For the turbulent regime, initiated at some postulated Reynolds number, theoretical predictions of Chun and Seban⁽¹²⁾ have been used to estimate the local heat transfer coefficient

$$h \left(\frac{\nu^2}{gk}\right)^{1/3} = 3.8 \times 10^{-3} \left(\frac{4\Gamma}{\mu}\right)^{0.4} \left(\frac{\nu}{\alpha}\right)^{0.65} \quad (2.32)$$

Dukler⁽¹⁶⁾ numerically integrated the equations for the transfer of heat and momentum across the falling film. Results for film thickness and local and average heat transfer coefficients were presented graphically,

Fig 2.5 contains curves showing the prediction value of the heat transfer coefficient for the Nusselt layer and for turbulent flow according to Dukler. Equation 2.30 and 2.32 is curve B and C of Fig 2.5. The results for $\frac{\nu}{\alpha} = 5.7$ imply "turbulent" flow at Reynolds number

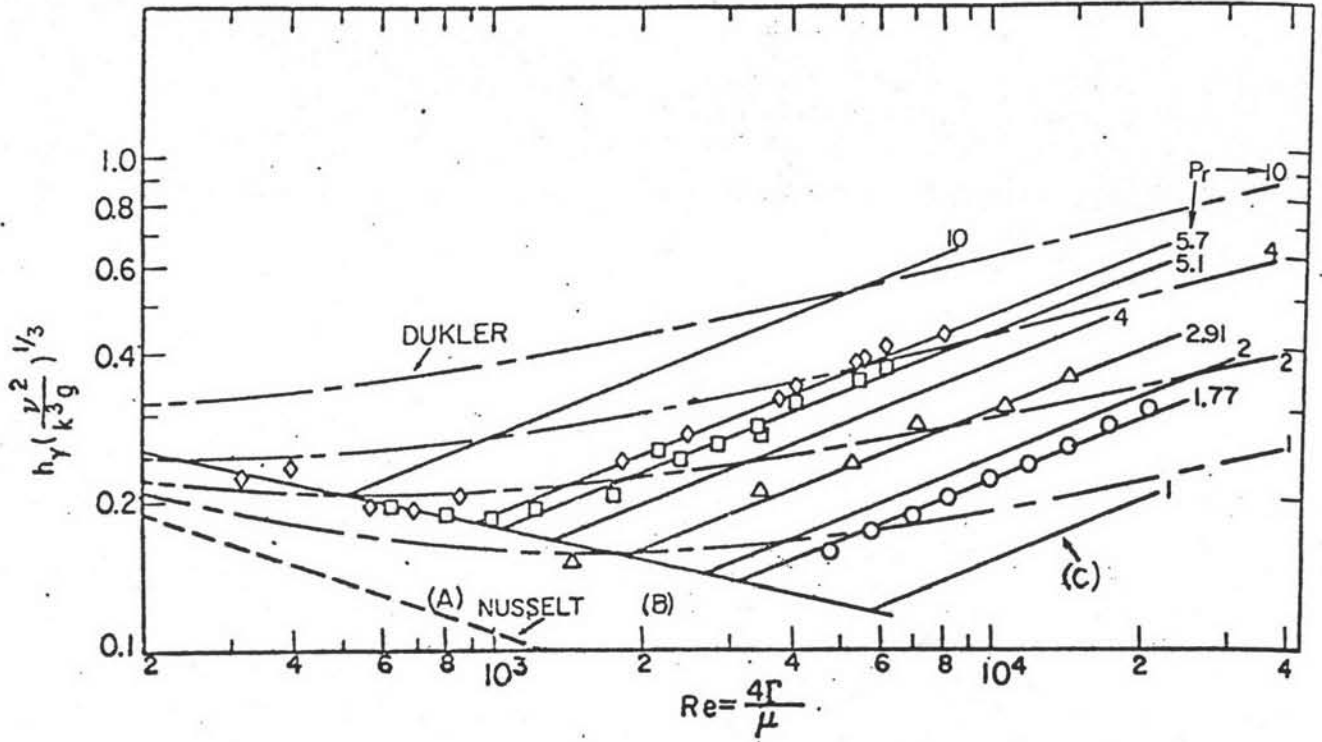


Fig.2.5 Local heat transfer coefficient as function of Reynolds number; curve A is the Nusselt solution, equation(2-27); curve B is the "wavy laminar" solution, equation(2-30); for turbulent flow the solid curves are the empirical fit to equation(2-32), shown for various integral values of the Prandtl number, the broken curves are the turbulent prediction of Dukler.

down to the order of 1500 and those for the lowest Prandtl number of 1.77 imply the possibility of laminar flow up to $\frac{4\Gamma}{\mu} = 3200$, or somewhat less if a more gradual transition is considered. The intersection of equation 2.30 and 2.32 indicates a departure from the "wavy laminar" regime at the Reynolds number

$$\left(\frac{4\Gamma}{\mu}\right)_{tr} = 5800 \left(\frac{\nu}{\alpha}\right)^{-1.06} = 0.215(K_a)^{-1/3} \quad (2.33)$$

the latter form arising from the fact that the Kapitza number is a power function of the Prandtl number in the temperature range involved, with the viscosity being the controlling element of both Kapitza and Prandtl numbers.

Wilke⁽¹⁷⁾ presented results for the heating, essentially without evaporation, of falling films of water and mixtures of water and ethylene glycol on the exterior of a vertical tube. He gave a correlation for "turbulent" heat transfer coefficient as

$$h_H \left(\frac{\nu^2}{\rho k^3}\right)^{1/3} = 8.4 \times 10^{-3} \left(\frac{4\Gamma}{\mu}\right)^{0.4} \left(\frac{\nu}{\alpha}\right)^{0.344} \quad (2.34)$$

valid for $\frac{4\Gamma}{\mu} > 3200$. For "non-laminar" flows for $\frac{4\Gamma}{\mu} < 3200$, Wilke presents separate correlations which, only for water, $\frac{\nu}{\alpha} = 5.4$, depart less than 15 percent from equation (2.32). Wilke defined a "departure" Reynolds number in his experiments in terms of the first departure

of the heat transfer coefficient from laminar performance. Because the transfer from the laminar to the turbulent line is actually less abrupt than given by the intersection of equations like 2.30 and 2.32, his departure Reynolds numbers are relatively lower; they were given as:

$$\left(\frac{4\mu}{\rho}\right)_{tr} = 2400 \left(\frac{\nu}{\alpha}\right)^{-0.65} \quad (2.35)$$

Equation 2.30 and 2.32 are representative of expectation as long as no nucleate boiling occurs.

Chun and Seban⁽¹²⁾ was suggested that boiling could be made to occur in some conditions of operation, usually at a location near the bottom of the tube. This could be determined visually and also by a sudden reduction of wall temperature in the region where bubbles were observed. This indicates that once nucleation is initiated it will continue at the lower superheat associated with the higher heat transfer coefficient for boiling. Chun⁽¹⁸⁾ has also applied the elementary liquid superheat model to specify the incidence of nucleate boiling and equation 2.36 is the most appropriate for these results

$$T_g - T_s = \frac{2T_s^2 R \delta}{\lambda_{pr}} \quad (2.36)$$

Chen⁽¹⁹⁾ correlated available data on flow boiling in tubes; for high flow rates he postulated that nucleate boiling was

suppressed and vapor production occurred at the vapor liquid interface. For this condition Chen's correlation reduced to

$$h = F(0.023)(N_{Re})_l^{0.8} (N_{Pr})^{0.4} \frac{k_l}{D_e} \quad (2.37)$$

where the Reynolds number was evaluated assuming the liquid filled the cross section of the tube. Value of the empirical correction factor, F, were presented graphically:

Penman and Tait⁽²⁰⁾ have measured the value of local heat transfer coefficient of liquid vapor mixtures flowing inside vertical tubes at atmospheric pressure. They derived an equation to calculate the local heat transfer coefficient.

This equation was expressed as below :

$$\frac{h}{c_{p1}} \sqrt{\frac{D}{\rho_1 g}} = 0.012 \left[v_v \sqrt{\frac{D \rho_v}{g}} \right]^{0.5} \quad (2.38)$$

They also recommended that their equation have been derived irrespective of whether the film be moving upward or downward in vertical tubes, or on the inside of a horizontal tube.

Wallgren⁽²¹⁾ integrated an empirical equation for the local heat transfer coefficient in liquid film flowing presented by Penman and Tait and obtained the following expression for the average heat transfer coefficient

$$\frac{\bar{h}}{c_{p1}} \sqrt{\frac{D}{\rho_1 g}} = 2.88 \times 10^{-4} \frac{c_{p1} \Delta T}{\lambda} \frac{L}{D} \left(\frac{\rho_1}{\rho_v} \right)^{0.5} \quad (2.39)$$

2.3.5 Special Features of Vertical Film Evaporator

The four different evaporator characteristics that are of interest in connection with this type of evaporator are briefly recommended by Agarwal⁽⁹⁾ as follows:

1. Steam Consumption

The gravity flow evaporator offers the advantage of low steam consumption. Since these evaporators do not require energy for any upward movement of liquid in the tubes, they can be operated with exceptionally small temperature differences. Hence, it becomes possible to increase the number of stages without changing the pressure of the exhaust steam or vapor pressure in the last stage, i.e. without changing the total temperature difference, and thereby reducing the steam consumption. The steam consumption is less with a great number of stages of evaporation.

2. Heat transfer area

Heat transfer values in these evaporators, are better than those of the conventional evaporators because of high velocity of flow in the tubes. In other words, the heat transfer area requirements are less, and since this represents the largest part of evaporator cost, the falling film evaporators are cheaper per unit evapora-

tion capacity. Moreover, since these are built with long tubes, valuable floor space is saved.

3. Entrainment and production losses

Good vapor-liquid separation is important for a number of reasons. Most important is the prevention of entrainment because of the value of product lost, contamination of condensing water and corrosion of surfaces on which vapors are condensed. The entrainment and production losses are low with this type of evaporator, on account of using centrifugal separators of high efficiency.

4. Cleaning cycles

Because of the high falling speed of both vapor and liquid in this evaporator, scaling is reduced by a considerable extent. Hence cleaning required is less frequent resulting in much less downtime. These evaporators normally do not require mechanical cleaning and are easily cleaned by rinsing with chemical cleaning agents dilute acid or dilute alkali as the requirement may be. When need arises the tubes and the distribution device can be cleaned with brushes having bristles of stainless steel or suitable material.

The gravity flow evaporators are most suitable for the evaporation of heat sensitive products, for example: in the chemical industry pharmaceuticals, glycol, gelatine tannin extract etc; in the food industry-sugar, glucose, steep water, fruit juices, etc.