Chapter

Boiling and Condensation

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Abstract

This chapter contains a brief overview of both boiling and condensation heat transfer phenomena. Boiling and condensation are the two convective heat transfer phenomena that involve phase change from liquid to vapour and vapour to liquid, respectively. The chapter starts with the basis of heat transfer with an emphasis on the boiling and condensation phenomenon. Next, the overview of the boiling phenomenon and its different classifications like pool, flow, and subcooled and saturated boiling are discussed in detail. Different boiling regimes (natural convection boiling, nucleate boiling, transition boiling and film boiling) with the observed heat transfer rate in the case of pool boiling are mentioned in detail using the boiling curve. The heat transfer aspect and basics of condensation with types (drop and film-wise condensation) and application are also presented. The derivation for the calculation of the rate of heat transfer during film condensation with the correlations for heat transfer coefficient on vertical, horizontal and inclined plates is explained. Some numerical for the calculation of the rate of heat transfer and heat transfer coefficient for condensation phenomena has been also been mentioned. Apart from a basic overview, this chapter also includes information about the advanced heat transfer enhancement techniques available for boiling and condensation.

Keywords: Heat transfer, boiling, condensation, film-wise condensation, boiling regimes, convection, Heat transfer coefficient

1. Introduction

Heat is a type of energy that is in transit between a hot body (source at a higher temperature) and a cool body (receiver at a lower temperature). The driving force for heat energy transport between two points is the temperature difference between them. Calorie and joule are the most frequent units for expressing heat energy [1].

Heat transfer is the branch of science concerned with determining the rates of heat energy transfers. Conduction, convection and radiation are three modes, by which the transfer of heat occurs from a hot source to a cold recipient. In the conduction mode of heat transfer, the heat energy is generally transferred within the substance or to another substance in physical contact and is caused by lattice vibration and free electron movement. Convection is the transmission of heat due to the macroscopic motion of molecules within the medium. In general, conduction heat transfer is observed within solid mediums and convection occurs within fluids (gases or liquids) mediums by the mixing of hot and cold portions of the fluid. In the radiation mode, heat is transferred through electromagnetic waves produced by a hot body. Radiation heat can be transferred over the medium within vacuums and space [2].

Sensible heat is the heat that must be transferred to raise or lower the temperature of a system when no phase change is observed within the medium. The latent heat of phase change is the thermal energy associated with a unit amount of matter at a fixed temperature and pressure when it experiences a phase transition (from solid to liquid and vapour to liquid or vice versa) [2].

2. Boiling and condensation phenomenon

Boiling and condensation both are under the convection heat transmission process in which the system undergoes a phase transition and are opposite to each other. These processes include the involvement of both sensible and latent heat.

Boiling is the process of transferring a medium from a liquid to a vapour state by applying heat. When a liquid medium is applied to heat, then the medium will start to boil at a certain temperature (boiling point temperature). At this particular temperature, the liquid phase changes to the vapour state, known as the boiling phenomenon.

In reverse, condensation is the process of transferring a medium from a vapour to a liquid by removing the heat from the medium. When a medium initially in the vapour state is cooled, then its phase will change from vapour to liquid state, known as the condensation phenomenon. Vapours are generated by boiling, and liquid droplets are formed by condensation [1, 3].

3. Heat transfer to boiling liquids

Boiling phenomenon is generally observed in unit operations such as evaporation, distillation and steam generation and is the opposite of the condensation phenomenon. When the liquid medium is exposed to a surface, at a temperature above the saturation temperature of the liquid, the phase of the medium changes from liquid to vapour.

Suppose a liquid medium is kept within a solid vessel, to which heat is supplied to boil the liquid. Let the temperature of the solid surface be ‘Ts’ and the liquid medium have the saturation temperature of ‘Tsat’. Initially, let the solid surface temperature be below the saturation temperature of the liquid. The boiling will start, when the temperature of the supplied liquid increases from ‘Ts’ to the saturation temperature ‘Tsat’. Further on increasing the temperature of the supplied liquid, the boiling rate will also increase.

According to Newton’s law of convection, heat transferred from a solid surface to the liquid (through convection mode) is

$$Q = hA(T_S - T_{Sat}) = hA\Delta T_{excess}$$

Here, $(T_S - T_{Sat}) = \Delta T_{excess}$ = excess temperature = temperature of the supplied liquid – saturation temperature of liquid = the extra heat supplied in excess above the saturation temperature of the fluid, during boiling of the liquid medium [3, 4].

3.1 Classification of the boiling phenomenon

According to the bulk fluid motion and the bulk fluid temperature, the boiling phenomenon can be classified into two basic categories.

i. Based on the bulk fluid motion within the liquid medium

   Pool boiling: Boiling phenomenon within a liquid medium, which is at a stationary or non-flow condition, is called pool boiling. In pool boiling, heat is generally supplied through a submerged solid surface (by placing a heating coil inside the liquid) or boiling water within a solid container from external heat. Bubbles generate during the heating process travel in the liquid medium, due to buoyance, and the heat gets transferred through the natural convection process.

   Flow boiling: Boiling phenomenon within a liquid medium, which is at flowing condition, is called flow boiling. In the case of flow boiling, heat is generally transferred to a flowing liquid medium through the forced convection process. Boiling in a liquid medium when it is flowing over a hot surface or within a heated pipe is an example of flow boiling [3, 5].

ii. Based on the bulk liquid temperature within the liquid medium

   Subcooled or local boiling: The boiling phenomenon is said to be subcooled or local boiling if the temperature of the bulk liquid medium above the heating surface is less than the saturation temperature of the liquid.

   Saturated boiling: The boiling phenomenon is said to be saturated if the temperature of the liquid medium above the heating surface is about the saturation temperature of the liquid.

   At the early stages of pool boiling, the liquid adjacent to the hot solid surface vaporises and the bubbles are formed by absorbing heat from the hot solid surface. The bubbles contain more heat energy and travel within the liquid medium due to the convection phenomenon.

   But during the initial stage of boiling, the bulk fluid (a certain height above a hot solid surface) will be at a temperature much less than the liquids saturation temperature. These bubbles when they move away from the hot surface and come in contact with cold liquid, they condense and collapse by transferring the absorbed heat (from the hot surface) into the liquid medium. This phenomenon happens when the bulk liquid is at a temperature much lower than the saturation temperature (subcooled or local boiling).

   Further, when the temperature of the entire liquid medium reaches about the saturation temperature, the bubbles will not condense and will start rising to the top (saturated boiling) (**Figures 1** and **2**) [3, 6].

3.2 Boiling regimes and the boiling curve in case of pool boiling

Boiling process in the pool of a liquid medium will start, when the supplied temperature exceeds the saturation temperature (ΔTexcess >0). Depending on this excess temperature ΔTexcess supplied to the liquid medium, different types of boiling regimes are observed in a pool of liquid [1, 3, 4]. Those regimes are

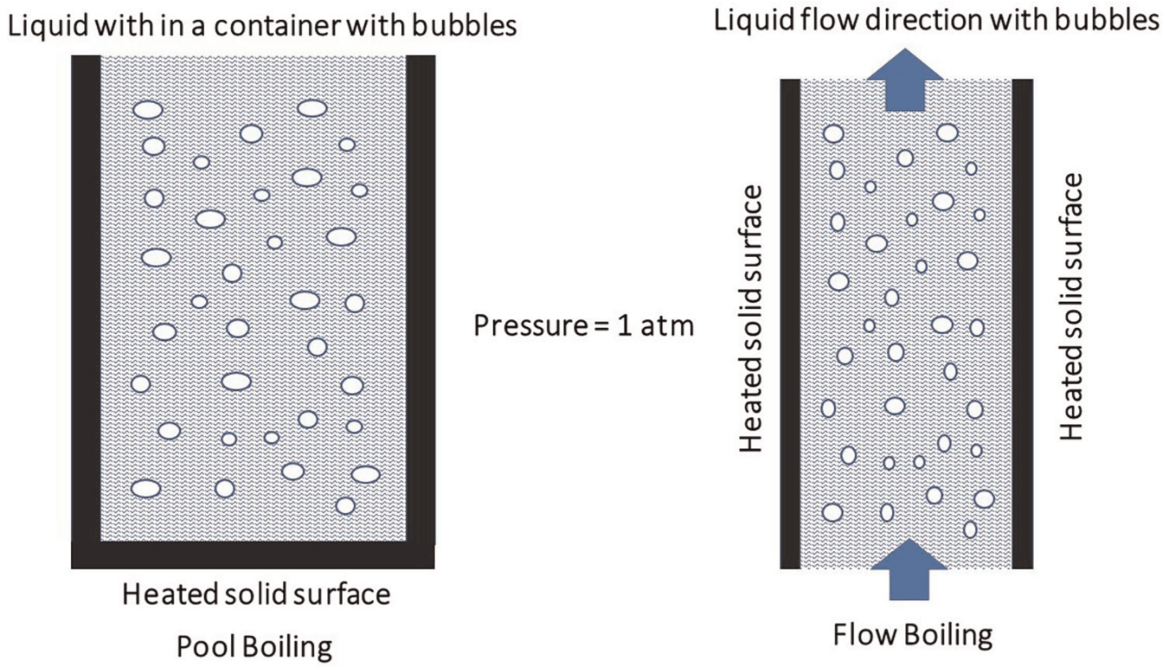

Figure 1.
Pool and flow boiling phenomena [3, 6].

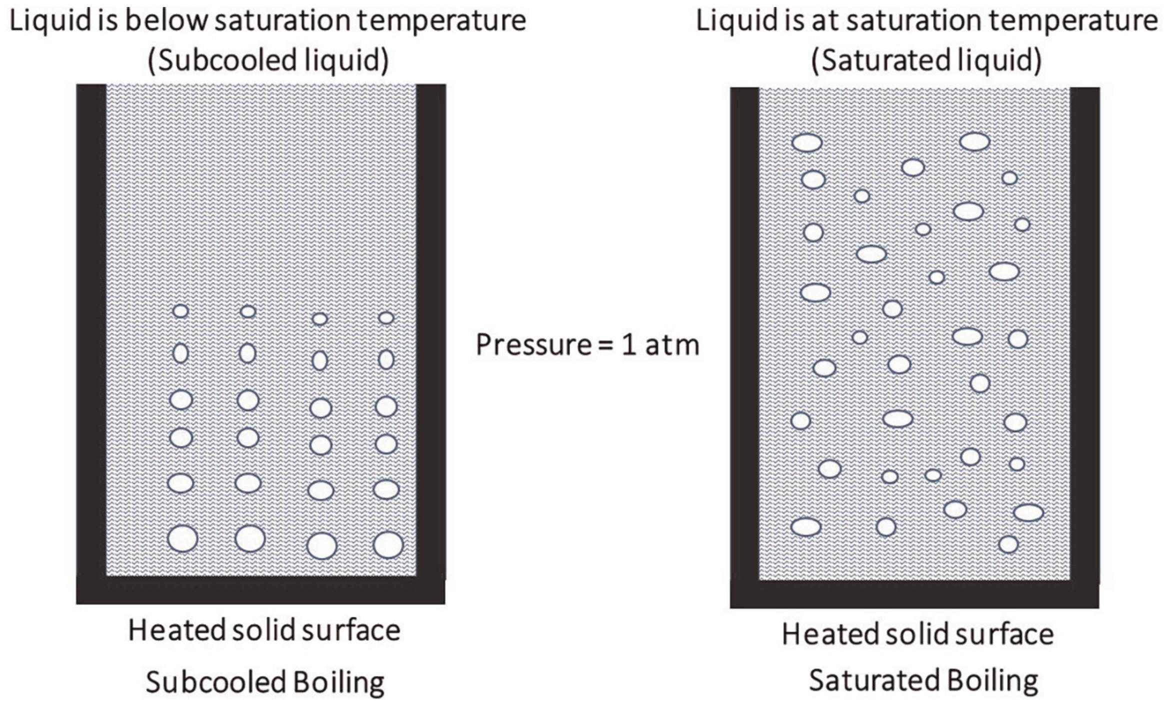

Figure 2.
Subcooled and saturated boiling phenomena [3, 6].

- Natural convection boiling
- Nucleate boiling
- Transition boiling
- Film boiling

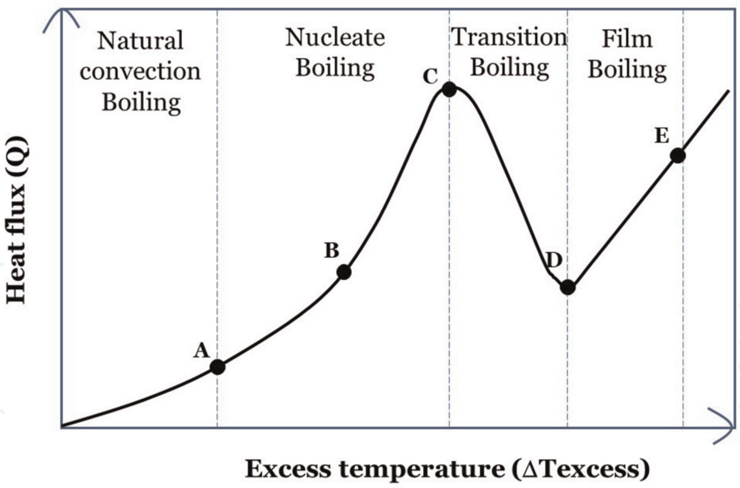

Figure 3.
Boiling curve with the boiling regimes for a pool of liquid [1–5].

To demonstrate these boiling regimes for a pool of liquid, a plot between boiling heat flux (rate of heat transfer per unit area) versus the excess temperature supplied is shown and known as the boiling curve (**Figure 3**).

3.2.1 Natural convection boiling (to point a on the boiling curve)

Boiling or saturation temperature of a pure liquid substance depends on the applied pressure. But in practice, the bubbles are forming on the heating surface only after being heated to a few more degrees above its saturation temperature (up to 6°C for water).

The transfer of heat within the fluid (from the heating surface to the bulk fluid) in this step is by natural convection, and hence the heat flux curve increases slowly. During this condition, the liquid will be at a slightly superheated state and the superheated liquids will evaporate when it rises to the free surface.

3.2.2 Nucleate boiling (between points A and C on the boiling curve)

Upon further increasing the excess temperature (ΔTexcess), the bubbles will start forming at the temperature with respect to point A of the boiling curve. With further increase in excess temperature, the rate of formation of bubbles and hence the heat flux will increase till the point C in the boiling curve.

The nucleate boiling regime (from A to C) can be further divided into two separate regions (from A to B and from B to C). In the region from A to B, with excess heat supplied beyond point A, isolated bubbles will start forming on the heated surface. As soon the bubbles start to move, these bubbles will dissolute in the liquid (subcooled boiling). The formation and dissolution of the bubble will be repeated till the temperature of the liquid reaching to saturated temperature. During nucleate boiling, the movement of the bubbles is responsible for the increase in heat transfer coefficient and heat flux.

In the region from B to C, with excess heat supplied beyond point B, the bubbles form at a great rate and a continuous column of vapour in the liquid will be observed. These bubbles move to the free surface (saturated boiling) where the vapour got released from the bubbles. The heat flux observed in this region will be larger due to the combined effect of liquid entrainment and evaporation.

High heat transfer rates are observed in the case of the nucleate boiling regime compared to other regimes; hence, it is the most desirable boiling regime in practice. For water, it can be achieved with ΔTexcess within about 30°C. The correlation for the boiling heat flux for this region was proposed by Rohsenow, which is [6–9].

$$q = \mu_l h_{fg} \left[\frac{g(\rho_l - \rho_v)}{\sigma} \right]^{½} \left[\frac{C_{pl} \Delta T_{excess}}{C_{sf} h_{fg} (Pr_l)^n} \right]^3$$

3.2.3 Transition boiling (between points C and D on the boiling curve)

WithΔ Texcess value near reaching the point C, the rate of evaporation of bubbles at the heater surface will be at a very high rate throughout the entire solid surface. These bubbles may cover the heater surface; hence, the contact between solid surface and liquid will be difficult. This formed vapour film acts as an insulation due to the low thermal conductivity of the vapour relative to that of the liquid, and hence the heat flux decreases beyond point C.

The heat flux till point C reaches a maximum value and is called the *critical (or maximum) heat flux,* q_{max}. The correlation for the boiling critical heat flux for this region for boiling was proposed by Lienhard et al. and is expressed as [6–8]:

$$q_{max} = 0.149 \rho_l h_{fg} \left[\frac{\sigma g(\rho_l - \rho_v)}{\rho_v^2} \right]^{¼}$$

3.2.4 Film boiling (beyond point D on the boiling curve)

During the transition boiling, the heater surface will be completely covered by a continuous stable layer of vapour film (with increasing ΔTexcess). As the vapour film separates the liquid from the heater surface and will be responsible for transferring less heat flux, the heat flux will reach a minimum value (point D), called the *Leidenfrost point*. The correlation for the boiling minimum heat flux at the *Leidenfrost point* for a horizontal plate was proposed by Zuber, which is [7].

$$q_{max} = 0.09 \rho_v h_{fg} \left[\frac{\sigma g(\rho_l - \rho_v)}{(\rho_l + \rho_v)^2} \right]^{¼}$$

As the ΔTexcess further increases, the heat transfer will start due to the radiation from hot solid surface to liquid through the vapour medium. The heat flux will again rise slowly and as the heating takes place over a film of vapour, it is known as film boiling. The correlation for the boiling heat transfer coefficient for this region for boiling above a horizontal tube was proposed by Bromley, which is given as follows [6–8]. However, for a vertical plate, the constant 0.62 and D will be replaced by 0.7 and L, respectively:

$$h = 0.62 \left[\frac{k_v^3 \rho_v (\rho_l - \rho_v) g \left(h_{fg} + 0.4 C_{pv} \Delta T_{excess} \right)}{D \mu_v \Delta T_{excess}} \right]^{¼}$$

Burnout point: As the heat flux is decreasing beyond point C, the boiling process will not be advised to continue further. Beyond point C, the power that needs to be provided to the heater surface will be more (as the heat flux decreases). However, in this condition of excess power supply, the temperature of the nichrome wire (heater) immersed in the liquid will abruptly rise to the melting point of the wire, resulting in burnout. Burnout can be avoided by using platinum wire, which has a much higher melting point.

4. Heat transfer in condensation

Condensation is a convection process of changing a vapour medium to a liquid state and generally occurs when a saturated vapour comes into contact with a cold solid surface at a temperature less than the saturation temperature of the vapour. The latent heat of vaporisation must be removed during the condensation.

Condensers are widely used in the chemical industry. The process of condensation occurs by two distinct mechanisms/modes and at different rates of heat transfer. Those are film-wise and drop-wise condensation [2, 3].

4.1 Difference between drop-wise and film-wise condensation

4.1.1 Drop-wise condensation

When a saturated vapour comes into contact with a cold solid surface (a surface at a lower temperature than the saturated temperature of vapour), it condenses to liquid form. If condensate does not wet the surface, then the droplets of liquid are formed on the surface.

The size of droplets expands with time and eventually drops down the surface in a random pattern (due to the effect of gravity), leaving the metal surface bare on which further condensation develops.

4.1.2 Film-wise condensation

When a saturated vapour comes into contact with a cold surface (a surface at a lower temperature than the saturated temperature of vapour), it condenses to liquid form. If condensate wets the surface, then it forms a continuous film of condensate.

These condensates completely cover the solid surface, and then heat must be transported through the condensed liquid layer. Then the vapours have to condense into the liquid film rather than direct contact with the surface. Under the action of gravity, the condensate eventually flows down the surface. The condensation caused by this technique is termed as film-wise condensation [2, 3].

The film covering the solid surface serves as a heat transmission barrier in film-wise condensation, but in drop-wise condensation, a considerable section of the surfaces is exposed directly to the vapour. Hence, the heat transfer coefficients (and thus the heat transfer rates) in drop-wise condensation are generally four to eight times greater than in film-wise condensation. The presence of dirt on the surfaces (where condensate drops develop), which appear to favour drop-wise condensation, is known as nucleation sites. Because most surfaces become wet after being exposed to the condensing vapours, film-wise condensation is very common (**Figure 4**) [2, 3, 6].

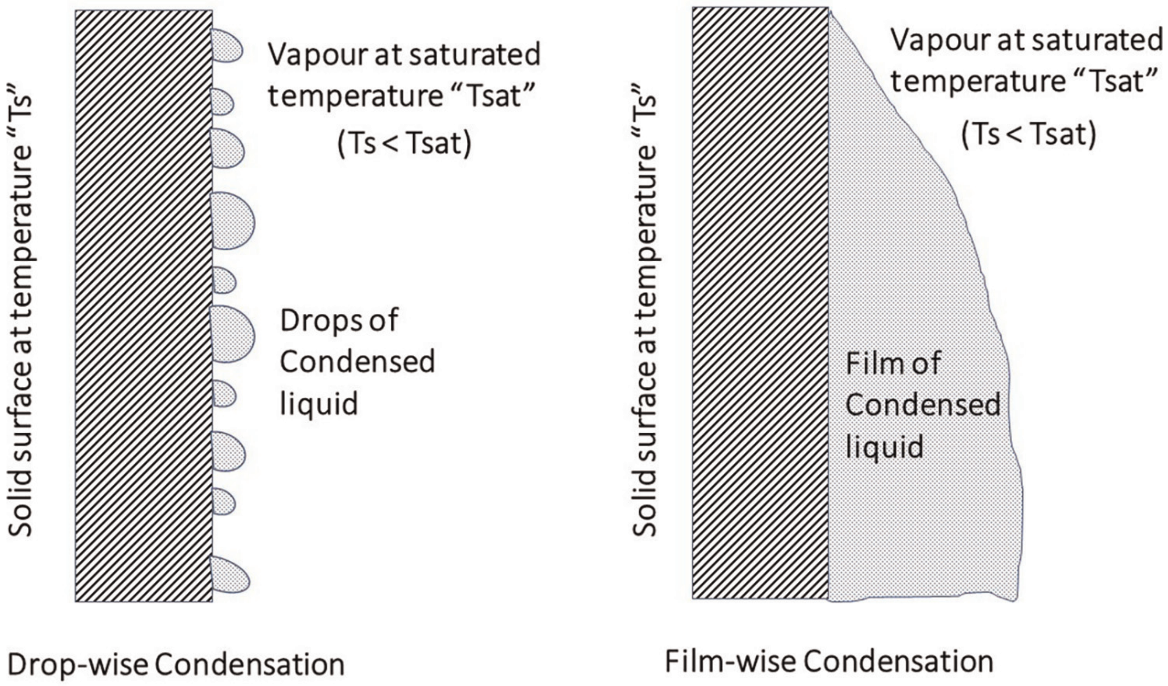

Figure 4.
Drop-wise and film-wise condensation phenomena [2, 3, 6].

Drop-wise condensation	**Film-wise condensation**
In drop-wise condensation, the condensate liquids partially wet the complete solid surface by forming droplets of condensate on the surface.	In film-wise condensation, the condensed liquid wets the solid surface by forming a continuous film of condensate on the surface.
These droplets then fall down the surface under the action of gravity, leaving the bare solid surface to condense further.	Condensate flows down the surface under the action of gravity by forming a continuous film, and further heat transfer takes place through this layer.
As the bare solid surface is further available to condense, the heat transfer coefficients and thus heat transfer rate are very high compared to film-wise condensation. The heat transfer coefficient value is difficult to predict.	Due to the presence of a continuous liquid film of condensate between the vapour and solid surface, the heat transfer coefficients and thus heat transfer rate are very low compared to drop-wise condensation. The heat transfer coefficient value can be predicted easily.
Drop-wise condensation is difficult to achieve and generally occurs on oily or greasy surfaces.	Film-wise condensation is easily obtainable and generally occurs on smooth, clean and uncontaminated surfaces.
Drop-wise condensation condition is difficult to maintain and is unstable. Drop-wise condensation is commonly not used industrially.	Film-wise condensation conditions can be easily maintained and stable. Film-wise condensation is commonly used industrially.

4.2 Heat transfer for film-wise condensation on vertical plate [3, 4]

Let us consider a vertical plate maintained at a constant surface temperature 'Ts' with a height 'L' and width 'b'. Let us consider a single vapour medium, at the saturation temperature 'Tsat', exposed to this surface. The surface temperature of the solid surface is below the saturation temperature (Ts < Tsat). When this saturated vapour comes in contact with the cold surface, then the vapour will condense on it.

In the case of film-wise condensation, it will form a continuous film of condensate on the surface of the vertical plate. The condensate liquid film layer ultimately flows down and will obtain a state as shown in the figure under the influence of viscosity and gravity.

Let the downward direction is taken as the positive x-direction with the origin placed at the top of the plate where condensation initiates, as shown in **Figure 5**. The film thickness of condensate 'δ' and thus the mass flow rate of the condensate increase with respect to the length of plate 'x'.

Heat must be transferred from the vapour to the plate through the film, which provides heat transfer resistance. The greater the thermal resistance of the film, the slower the rate of heat transfer will be. Nusselt first derived the analytical relationship between the heat transfer coefficient throughout the length of the plate in film condensation on a vertical plate in 1916, using the following assumptions:

- The surface temperature of solid surface ('Ts') and the vapour is at saturation temperature ('Tsat') and is more than solid surface temperature (Ts < Tsat).
- The temperature of condensate liquid varies linearly across the liquid film.
- The flow within the condensate liquid layer is laminar and the acceleration of the condensate liquid layer is negligible.
- Heat transfer within the condensate liquid film is through conduction mode only (no convection phenomenon in the film).
- All the properties of the condensate liquid are constant throughout the film.

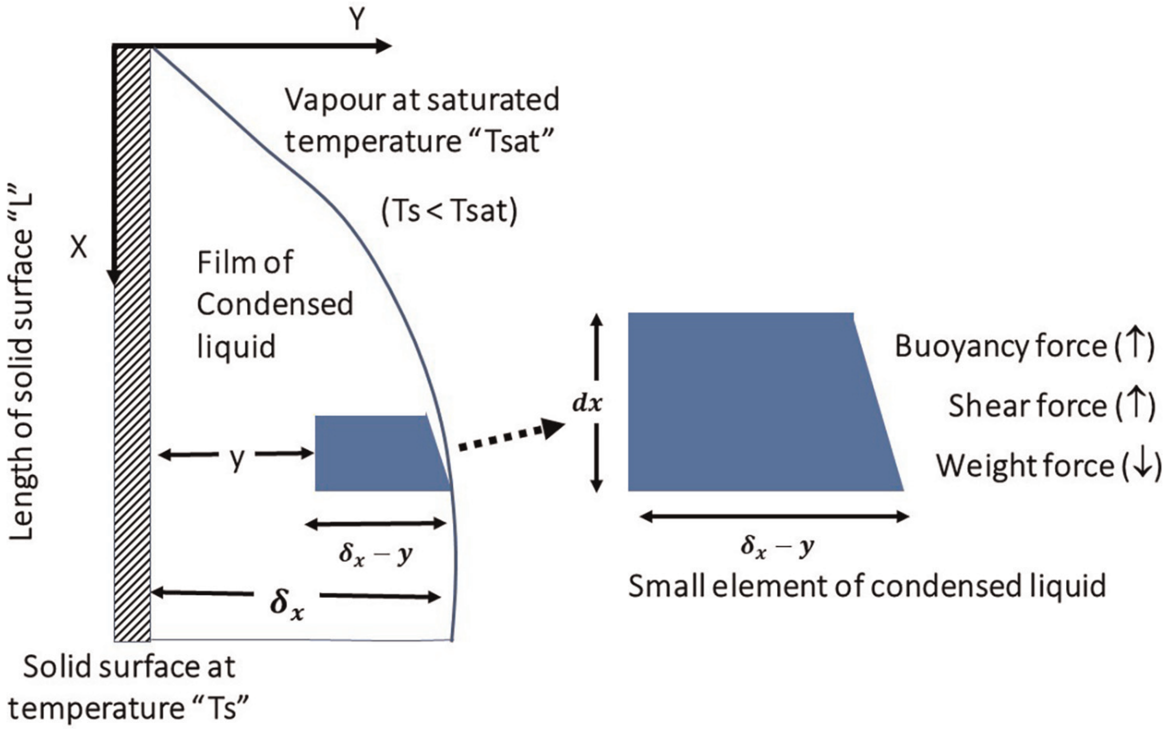

Figure 5.
Heat transfer phenomena for film condensation on vertical plate [3, 4].

- The viscous shear on the liquid–vapour interface is negligible (velocity of the saturated vapour is maintained very low to avoid drag on the condensate film).

The rate of heat transfer from the vapour phase to the solid surface with respect to vertical direction x can be expressed as:

$$Q = h_x A(T_{sat} - T_s) = k_l A \frac{(T_{sat} - T_s)}{\delta_x}$$

$$Then, Heat\ flux = q = \frac{Q}{A} = h_x(T_{sat} - T_s) = k_l \frac{(T_{sat} - T_s)}{\delta_x}$$

$$h_x = \frac{k_l}{\delta_x} \tag{1}$$

The heat transfer coefficient value for the heat transfer from the vapour to the plate is changing along the length of the plate due to the thermal resistance offered by the varying thickness of condensate liquid film. If the thickness of the film is more, then more will be the thermal resistance for the flow of heat from vapour to solid and thus lower the rate of heat transfer.

Let us consider a small selected volume element of condensate in the vertical x-direction. Since the acceleration of the small section fluid is assumed zero, then according to Newton's second law of motion, it can be written as:

$$\sum F_x = ma_x = 0$$

$$\sum F_x(Downward\ direction, \downarrow) = \sum F_x(Upward\ direction, \uparrow)$$

$$Weight\ force\ (\downarrow) = \text{Viscous shear force } (\uparrow) + \text{Buoyancy force } (\uparrow)$$

The forces that act on this small elemental volume will be the weight of the liquid element (acting downward), viscous shear or fluid friction force (acting upward) and buoyancy force (acting upward).

$$Weight\ (\downarrow)\ or\ gravity\ force\ on\ small\ liquid\ element = \rho_l g(\delta_x - y)dx$$

$$\text{Viscous shear force } (\uparrow)\ on\ small\ liquid\ element = \mu_l \frac{du}{dy} dx$$

$$\text{Buoyancy force } (\uparrow)\ on\ small\ liquid\ element = \rho_V g(\delta_x - y)dx$$

Thus,

$$\rho_l g(\delta_x - y)dx = \mu_l \frac{du}{dy} dx + \rho_V g(\delta_x - y)dx$$

$$\mu_l \frac{du}{dy} dx = \rho_l g(\delta_x - y)dx - \rho_V g(\delta_x - y)dx$$

$$\frac{du}{dy} = \frac{(\rho_l - \rho_V)g(\delta_x - y)}{\mu_l}$$

$$du = \frac{(\rho_l - \rho_V)g}{\mu_l}(\delta_x - y)dy$$

Integrating the aforementioned equation from y = 0 to y, we will have the relationship between the velocity along the length of the vertical pipe:

[At y = 0; u = 0 (no-slip boundary condition) and at y = y; u = u (y) (not zero)]

$$\int_0^{u(y)} du = \frac{(\rho_l - \rho_V)g}{\mu_l}\int_0^y (\delta_x - y)dy$$

$$u = \frac{(\rho_l - \rho_V)g}{\mu_l}\left(y\delta_x - \frac{y^2}{2}\right)$$

Then, the mass flow rate of the condensate with the boundary layer thickness is ‘δ’, at any location ‘x’ over the solid surface will be

$$\dot{m} = \int \rho_l u dA = \int_0^{\delta} \rho_l \frac{(\rho_l - \rho_V)g}{\mu_l}\left(y\delta_x - \frac{y^2}{2}\right) bdy$$

$$\dot{m} = \frac{\rho_l(\rho_l - \rho_V)gb}{\mu_l}\int_0^{\delta}\left(y\delta_x - \frac{y^2}{2}\right) dy$$

$$\dot{m} = \frac{\rho_l(\rho_l - \rho_V)gb}{\mu_l}\left(\frac{\delta_x^2}{2}\delta_x - \frac{\delta_x^3}{2\times 3}\right) = \frac{\rho_l(\rho_l - \rho_V)gb\delta_x^3}{3\mu_l} = \frac{\rho_l(\rho_l - \rho_V)gb\delta_x^3}{3\mu_l}$$

$$Then$$

$$\frac{d\dot{m}}{dx} = \frac{d}{dx}\left(\frac{\rho_l(\rho_l - \rho_V)gb\delta_x^3}{3\mu_l}\right) = \frac{\rho_l(\rho_l - \rho_V)gb}{3\mu_l}\frac{d}{dx}\delta_x^3 = \frac{\rho_l(\rho_l - \rho_V)gb\delta_x^2}{\mu_l}\frac{d\delta_x}{dx} \quad (2)$$

Again, the rate of heat transfer from the vapour to the solid surface through the liquid film layer will be equal to the amount of heat released when vapour is condensed and is expressed as:

$$dQ = d\dot{m}\ \mathrm{x}\lambda = k_l A\frac{(T_{sat} - T_s)}{\delta_x} = h_x A(T_{sat} - T_s)$$

$$dQ = d\dot{m}\ \mathrm{x}\lambda = k_l(b\ \mathrm{x}dx)\frac{(T_{sat} - T_s)}{\delta_x} = h_x(b\ \mathrm{x}dx)(T_{sat} - T_s)$$

$$\frac{d\dot{m}}{dx} = k_l b\frac{(T_{sat} - T_s)}{\delta_x\lambda} \quad (3)$$

Equating the aforementioned two Eqs. (2) and (3), we have

$$\frac{d\dot{m}}{dx} = k_l b\frac{(T_{sat} - T_s)}{\delta_x\lambda} = \frac{\rho_l(\rho_l - \rho_V)gb\delta_x^2}{\mu_l}\frac{d\delta_x}{dx}$$

$$k_l b\frac{(T_{sat} - T_s)}{\delta_x\lambda} = \frac{\rho_l(\rho_l - \rho_V)gb\delta_x^2}{\mu_l}\frac{d\delta_x}{dx}$$

$$\frac{k_l b\mu_l(T_{sat} - T_s)}{\rho_l(\rho_l - \rho_V)gb\lambda}dx = \delta_x^3 d\delta_x$$

$$\delta_x^3\, d\delta_x = \frac{k_l\mu_l(T_{sat} - T_s)}{\rho_l(\rho_l - \rho_V)g\lambda}dx$$

The liquid film thickness at any location x can be determined by integrating the aforementioned equation from x = 0 (δ_x = 0 at the top of the plate) to x = x (δ_x = x):

$$\int_0^{\delta_x} \delta_x^3 d\delta_x = \frac{k_l \mu_l (T_{sat} - T_s)}{\rho_l(\rho_l - \rho_V) g\lambda} \int_0^x dx$$

$$\frac{\delta_x^4}{4} = \frac{k_l \mu_l (T_{sat} - T_s)}{\rho_l(\rho_l - \rho_V) g\lambda} x$$

$$\delta_x = \left(\frac{4k_l \mu_l (T_{sat} - T_s)\mathrm{x}}{\rho_l(\rho_l - \rho_V) g\lambda} \right)^{1/4}$$

Then the heat transfer rate from the vapour to the solid plate surface at any location x along the length of the plate can be expressed as (from Eq. (1)):

$$h_x = \frac{k_l}{\delta_x}$$

$$h_x = \frac{k_l}{\left(\frac{4k_l \mu_l (T_{sat} - T_s)\mathrm{x}}{\rho_l(\rho_l - \rho_V) g\lambda} \right)^{1/4}} = \left(\frac{\rho_l(\rho_l - \rho_V) g\lambda k_l^4}{4k_l \mu_l (T_{sat} - T_s)\mathrm{x}} \right)^{1/4}$$

$$h_x = \left(\frac{\rho_l(\rho_l - \rho_V) g\lambda k_l^3}{4\mu_l (T_{sat} - T_s)\mathrm{x}} \right)^{1/4}$$

Upon integrating the aforementioned equation for the local heat transfer coefficient over the entire length of the plate (L), the value average heat transfer coefficient value is determined:

$$h_{avg} = h_{Vertical} = \frac{1}{L} \int_0^L h_x dx = 0.943 \left(\frac{\rho_l(\rho_l - \rho_V) g\lambda k_l^3}{\mu_l (T_{sat} - T_s)\mathrm{L}} \right)^{1/4}$$

Note: In general, the density of the vapour medium will be negligible compared to the density of the liquid:

$$\rho_l \gg \rho_V \; or \; \rho_l \approx (\rho_l - \rho_V)$$

$$h_{Vertical} = 0.943 \left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l (T_{sat} - T_s)\mathrm{L}} \right)^{1/4}$$

Note 1: In case, the vertical plate is inclined at an angle Θ. The heat transfer coefficient is

$$h_{Inclined} = h_{Vertical} (cos\theta)^{1/4}$$

Note 2: In the case of the horizontal tube or sphere with diameter D, the heat transfer coefficient is

$$h_{Horizontal} = 0.729 \left(\frac{\rho_l(\rho_l - \rho_V) g\lambda k_l^3}{\mu_l (T_{sat} - T_s)\mathrm{D}} \right)^{1/4} = 0.729 \left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l (T_{sat} - T_s)\mathrm{D}} \right)^{1/4}$$

Note 3: The aforementioned equations are applicable for a single-tube system, in case N number of tubes are arranged in the system or a stack of tubes are present. h_1 is the heat transfer coefficient for the top tube. Then the heat transfer coefficient for N tubes (h_N) is

$$h_N = h_1(\mathrm{N})^{1/4}$$

5. Research and techniques of enhancement of boiling and condensation heat transfer

Enhancement of boiling heat transfer: Different advanced techniques that can be used to improve heat transfer in pool boiling are classified as active or passive techniques. Active approaches regulate the fluid movement by different techniques such as mixing the fluid using mechanical agitation, pumping the fluid, vibrating the surface of the container, rotating the container continuously, and adding an external electrostatic or magnetic field [6, 7, 10]. On the other hand, passive heat transfer enhancement techniques focus on changing fluid characteristics and/or heat transfer surfaces, such as increasing the number of active nucleation sites and the rate of bubble formation at each site [11, 12]. A rough, dirt-covered surface produces more nucleation sites than a smooth surface. The rate of nucleation can also be promoted by applying a thin porous layer to the surface or constructing mechanical voids on the surface to allow for continuous vapour production.

Different surface modification approaches for improving channel flow boiling heat transfer were reviewed by Liang et al. [12] and Kim et al. [13]. It covers macroscale (the use of cylindrical pins, macro ribs and twisted tape inserts), microscale (the use of micro-fins, micro-pin-fins, artificial cavities porous coating) and nanoscale (the use of nanotubes or nanowires to coat a heating surface) [13, 14] techniques to enhance the rate of heat transfers. Nanostructure approaches are reported to be less effective than macroscale and microscale improvement techniques.

Shah et al. [15] investigated the flow pattern, nucleate boiling, bubble growth, void fraction, liquid layer thickness, critical heat flux, pressure drop and heat transfer models for boiling fluid in microchannels. Adnan et al. [16] investigated the usage of nanofluids (h-BN/DCM and SiO_2/DCM) to improve heat transmission in pool boiling. These nanoparticles were found to greatly improve the thermal properties of the base fluid, with a 27.59% improvement in the rate of heat transfer coefficient for saturation boiling. Heat transfer enhancement employing ZnO-water, TiO_2-water, and Al_2O_3-water nanofluids has also been reported [17–19]. Amiri et al. [20] reported that multi-walled carbon nanotubes treated with cysteine, silver nanoparticles and Gum Arabic exhibited significant enhancement in the pool boiling heat transfer coefficients and critical heat fluxes when added with different concentrations to the aqueous media. Chen et al. [21] reported a recent and detailed review on the boiling heat transfer enhancement using different nanofluid solutions [22].

Enhancement of condensation heat transfer: Many recent reviewed research articles [23–25] are reported for advanced condensation phenomenon techniques. For condensation, the two major conditions that are nucleation on the surface and departure of liquid droplets are greatly influenced by the hydrophilicity and hydrophobicity properties of solid surfaces. Similar to boiling, numerous surface modification techniques are used to improve the heat transfer rate for condensation. Among them, constructing low free energy surfaces and building micro-nano structure surfaces are two possibilities to enhance the nucleation of liquid droplets [26].

Metallisation, ion implantation and organic polymer coating are available reported methods for lowering surface free energy. Some precious metals, such as gold, silver, palladium, rhodium, and chromium, can be plated on solid surfaces to produce metallisation. Ion implantation procedures involve the ionisation of gases such as nitrogen, argon, helium, and hydrogen using a high-voltage electric field before bombarding them into a metal surface. A thin covering of organic polymer such as fluorocarbon polymer, silica gel polymer, hexamethyldisiloxane polymer, fluorinated propylene polymer, polyhexafluoropropylene polymer and poly(p-xylene) polymers can also be used to induce drop-wise condensation on a solid metal surface [27–30].

In micro-nano structure surfaces techniques, different micro-nano structures like nanowires, nanocons, nanosheets, nanoblocks of carbon nanotube, nanographene particles, ZnO, Ni and polystyrene are fabricated on the solid surfaces that provide the nucleation sites and promoting drop-wise condensation [22, 26, 30, 31].

6. Conclusion

This chapter provides a comprehensive overview of the boiling and condensation phenomenon. These are the two opposite phenomena related to convective heat transfer which is the heat transfer involved during changing phase from liquid to vapour and from vapour to liquid, respectively. Boiling occurs when the temperature of liquids raises above its saturation temperature. Boiling can be classified as pool and flow boiling and as subcooled and saturated boiling. Depending on the value of the excess temperature supplied (above saturation temperature) to the liquid medium, different types of boiling regimes are observed in a pool of liquid. Those regimes include natural convection boiling, nucleate boiling, transition boiling, and film boiling.

Again, the condensation process deals with changing a vapour to a liquid state with two distinct mechanisms, that is, film-wise condensation and drop-wise condensation. The heat transfer coefficients (and thus the heat transfer rates) in drop-wise condensation are greater than in film-wise condensation. Drop-wise condensation is difficult to achieve and generally occurs on oily or greasy surfaces. Film-wise condensation is easily obtainable and generally occurs on smooth, clean uncontaminated surfaces.

7. Numerical related to condensations

Example 1: Consider a vertical tube within which hot gas is flowing at 80°C. The tube has a diameter of 40 mm and 1 m in length. This tube is now utilised to condense steam at atmospheric pressure. Determine the mass of condensate or the rate of condensation per hour that will generate in this system. Given the properties of condensate:

$k = 0.67$ W/m.K; $\rho = 972$ kg/m^3; $\lambda = 2310$ kJ/kg; $\mu = 3.55 \times 10^{-04}$ (N.s)/m^2.

Answer:

We know that the heat transfer coefficient over a vertical surface will be

$$h_{Vertical} = 0.943\left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l(T_{sat} - T_s)\mathrm{L}}\right)^{1/4}$$

L = height of the vertical plate = 1 m; D = diameter of the vertical plate = 40 mm = 0.04 m; g = gravitational acceleration = 9.81 m/s^2; ρ_l = densities of the condensed liquid = 972 kg/m^3; μ_l = viscosity of the condensed liquid = 3.55 x 10^{-04} (N.s)/m^2 = 355 x 10^{-06} kg/m·s; λ = latent heat of vaporisation = 2310 kJ/kg = 2310 x 10^3 J/kg; k_l =thermal conductivity of the condensed liquid = 0.67 W/m.K; Ts = surface temperature of the plate = 60°C = 273.15 + 60 = 333.15 K; Tsat = saturation temperature (at atmospheric pressure) of the condensing fluid = 100° C = 273.15 + 100 = 373.15 K.

Then

$$h_{\,Vertical} = 0.943\left(\frac{972^2\ \mathrm{x}9.81\ \mathrm{x}2310\ \mathrm{x}10^3\ \mathrm{x}0.67^3}{355\ \mathrm{x}10^{-6}\mathrm{x}(373.15-333.15)\ \mathrm{x}1}\right)^{¼} = 4352\frac{W}{m^2.K}$$

We know that the rate of heat transfer from the vapour to the solid surface through the liquid film layer will be equal to the amount of heat released when vapour is condensed and is expressed as:

$$dQ = d\dot{m}\ \mathrm{x}\lambda = k_l A\frac{(T_{sat}-T_s)}{\delta_x} = h_x A(T_{sat}-T_s)$$
$$Q = \dot{m}\ \mathrm{x}\lambda = h_x A(T_{sat}-T_s)$$

Area of heat transfer = π x D x L = π x 0.04 x 1 = 0.12564 m^2

$$Mass\ of\ condensate = \dot{m} = \frac{h_x A(T_{sat}-T_s)}{\lambda}$$
$$= \frac{4352\ \mathrm{x}0.12564\ \mathrm{x}(373.15-333.15)}{2310\ \mathrm{x}10^3} = 9.46\ \mathrm{x}10^{-03}\frac{kg}{s}$$
$$= 34.08\frac{kg}{hr}$$

Example 2: Consider a horizontal tube within which hot gas is flowing at 80°C. The tube has a diameter of 40 mm and 1 m in length. This tube is now utilised to condense steam at atmospheric pressure. Determine the mass of condensate that will generate in this system. Given the properties of condensate:

k = 0.67 W/m.K; ρ = 972 kg/m^3; λ = 2310 kJ/kg; μ = 3.55 x 10^{-04} (N.s)/m^2.

Answer:

We know that the heat transfer coefficient over a horizontal surface will be

$$h_{Horizontal} = 0.729\left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l(T_{sat}-T_s)\mathrm{D}}\right)^{¼}$$

L = height of the horizontal plate = 1 m; D = diameter of the vertical plate = 40 mm = 0.04 m; g = gravitational acceleration = 9.81 m/s^2; ρ_l = densities of the condensed liquid = 972 kg/m^3; μ_l = viscosity of the condensed liquid = 3.55 x 10^{-04} (N.s)/m^2 = 355 x 10^{-06} kg/m·s; λ = latent heat of vaporisation = 2310 kJ/kg = 2310 x 10^3 J/kg; k_l =thermal conductivity of the condensed liquid = 0.67 W/m.K; Ts = surface temperature of the plate = 60°C = 273.15 + 60 = 333.15 K; Tsat = saturation temperature (at atmospheric pressure) of the condensing fluid = 100°C = 273.15 + 100 = 373.15 K.

Then

$$h_{Horizontal} = 0.729\left(\frac{972^2 x9.81\ x2310\ x10^3\ x0.67^3}{355\ x10^{-6}\ x(373.15 - 333.15)\ x0.04}\right)^{1/4} = 7522\frac{W}{m^2.K}$$

We know that the rate of heat transfer from the vapour to the solid surface through the liquid film layer will be equal to the amount of heat released when vapour is condensed and is expressed as:

$$dQ = d\dot{m}x\lambda = k_l A\frac{(T_{sat} - T_s)}{\delta_x} = h_x A(T_{sat} - T_s)$$
$$Q = \dot{m}x\lambda = h_x A(T_{sat} - T_s)$$

Area of heat transfer = π x D x L = π x 0.04 x 1 = 0.12564 m^2

$$Mass\ of\ condensate = \dot{m} = \frac{h_x A(T_{sat} - T_s)}{\lambda} =$$
$$= \frac{75222\ x0.12566\ x(373.15 - 333.15)}{2310\ x10^3} = 0.016\frac{kg}{s} = 58.91\frac{kg}{hr}$$

Example 3: Consider a vertical tube with a temperature at 96°C and is exposed to steam at saturation temperature at atmospheric pressure. The tube has a diameter of 20 mm and 30 cm in length. Obtain the outside film heat transfer coefficient and the rate of heat transfer. Given the properties of condensate:

k = 0.57 kcal/hr-m-°C; ρ = 950 kg/m^3; λ = 540 kcal/kg; μ = 1.02 kg/m.hr.

Answer:

We know that the heat transfer coefficient over a vertical surface will be

$$h_{Vertical} = 0.943\left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l(T_{sat} - T_s)\mathrm{L}}\right)^{1/4}$$

L = height of the vertical plate = 30 cm = 0.3 m; D = diameter of the vertical plate = 20 mm = 0.02 m; g = gravitational acceleration = 9.81 m/s^2; ρ_l = densities of the condensed liquid = 950 kg/m^3; μ_l = viscosity of the condensed liquid = 1.02 kg/m · hr. = 2.83 x 10^{-4} kg/m.s; λ = latent heat of vaporisation = 540 kcal/kg = 540 x 10^3 x 4.184 J/kg = 2,259,360 J/kg; k_l =thermal conductivity of the condensed liquid = 0.57 kcal/hr-m-°C = (0.57 x 10^3 x 4.184/3600) J/s-m-°C = 0.66 W/m. °C; Ts = surface temperature of the plate = 96°C; Tsat = saturation temperature (atmospheric pressure) of condensing fluid =100°C.

Then

$$h_{Vertical} = 0.943\left(\frac{950^2 x9.81\ x2259360\ x0.66^3}{2.83\ x10^{-4}\ x(100 - 60)\ x0.3}\right)^{1/4} = 6049.28\frac{W}{m^2.^\circ C}$$

Area of heat transfer = π x D x L = π x 0.002 x 0.3 = 0.018846 m^2.

We know that

$$\begin{aligned} &Rate\ of\ heat\ transfer \\ &\quad = Q = h_x A(T_{sat} - T_s) \\ &\quad = 6049.28\ x0.018846\ x(100 - 96) = 456.018\ W \end{aligned}$$

Example 4: A tube 40 mm in diameter and 1 m in length is used to condense steam at 100°C. The tube surface is at 60°C. Determine for which arrangement of the tube, the rate of heat transfer and the mass of condensate will be maximum. (a) Vertical, (b) horizontal, (c) inclined at an angle of 45°C and (d) ten number of horizontal tubes in the vertical direction. Properties of condensate are.

k = 0.67 W/m.K; ρ = 972 kg/m^3; λ = 2310 kJ/kg; μ = 3.55 x 10^{-04} (N.s)/m^2.

Answer:

Given data, L = height of the vertical plate = 1 m; D = diameter of the vertical plate = 40 mm = 0.04 m; g = gravitational acceleration = 9.81 m/s^2; ρ_l = densities of the condensed liquid = 972 kg/m^3; μ_l = viscosity of the condensed liquid = 3.55 x 10^{-04} (N. s)/m^2 = 355 x 10^{-06} kg/m·s; λ = latent heat of vaporisation = 2310 kJ/kg = 2310 x 10^3 J/ kg; k_l =thermal conductivity of the condensed liquid = 0.67 W/m.K; Ts = surface temperature of the plate = 60°C = 273.15 + 60 = 333.15 K; Tsat = saturation temperature (atmospheric pressure) of condensing fluid =100°C = 273.15 + 100 = 373.15 K.

Area of heat transfer = A = π x D x L = π x 0.04 x 1 = 0.12564 m^2

a. We know that the heat transfer coefficient over a vertical surface will be

$$h_{Vertical} = 0.943\left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l(T_{sat} - T_s)\mathrm{L}}\right)^{1/4}$$

$$h_{Vertical} = 0.943\left(\frac{972^2 \mathrm{x} 9.81\ \mathrm{x} 2310\ \mathrm{x} 10^3\ \mathrm{x} 0.67^3}{355\ \mathrm{x} 10^{-6} \mathrm{x}(373.15 - 333.15)\ \mathrm{x} 1}\right)^{1/4} = 4352\frac{W}{m^2.K}$$

We know that

$$\begin{aligned} Rate\ of\ heat\ transfer &= Q = h_x A(T_{sat} - T_s) \\ &= 4352\ \mathrm{x} 0.12564\ \mathrm{x}(373.15 - 333.15) = 21871.42\ W \end{aligned}$$

$$Mass\ of\ condensate = \dot{m} = \frac{h_x A(T_{sat} - T_s)}{\lambda} = \frac{21871.42}{2310\ \mathrm{x} 10^3} = 9.46\ \mathrm{x} 10^{-3}\frac{kg}{s} = 34.08\frac{kg}{hr}$$

b. Heat transfer coefficient over a horizontal surface will be

$$h_{Horizontal} = 0.729\left(\frac{\rho_l^2 g\lambda k_l^3}{\mu_l(T_{sat} - T_s)\mathrm{D}}\right)^{1/4}$$

$$h_{Horizontal} = 0.729\left(\frac{972^2 \mathrm{x} 9.81\ \mathrm{x} 2310\ \mathrm{x} 10^3\ \mathrm{x}\ 0.67^3}{355\ \mathrm{x}\ 10^{-6}\ \mathrm{x}\ (373.15 - 333.15)\ \mathrm{x}\ 1}\right)^{1/4} = 3364.37\frac{W}{m^2.K}$$

We know that

$$\begin{aligned} Rate\ of\ heat\ transfer &= Q = h_x A(T_{sat} - T_s) \\ &= 3364.37\ \mathrm{x}\ 0.12564\ \mathrm{x}(373.15 - 333.15) = 16907.98\ W \end{aligned}$$

$$Mass\ of\ condensate = \dot{m} = \frac{h_x A(T_{sat} - T_s)}{\lambda} = \frac{16907.98}{2310\ \mathrm{x}\ 10^3} = 7.31\ \mathrm{x}\ 10^{-3}\frac{kg}{s} = 26.35\frac{kg}{hr}$$

c. Heat transfer coefficient for the inclined angle of 45°C will be

$$h_{Inclined} = h_{Vertical}(cos\theta)^{1/4}$$

$$h_{Inclined} = 4352 \text{ x } (cos\ \ 45)^{1/4} = 3990\frac{W}{m^2.K}$$

We know that

$$\begin{aligned}&Rate\ of\ heat\ transfer\\&= Q = h_x A(T_{sat} - T_s)\\&= 3990 \text{ x } 0.12564 \text{ x}(373.15 - 333.15) = 20052.15\ W\end{aligned}$$

$$Mass\ of\ condensate = \dot{m} = \frac{h_x A(T_{sat} - T_s)}{\lambda} = \frac{20052.15}{2310 \text{ x } 10^3} = 8.68 \text{ x } 10^{-3}\frac{kg}{s} = 31.25\frac{kg}{hr}$$

d. Heat transfer coefficient for 10 number horizontal tubes in the vertical direction will be

$$h_N = h_1(\mathrm{N})^{1/4}$$

$$h_N = 3364.37 \text{ x } (10)^{1/4} = 5982\frac{W}{m^2.K}$$

We know that

$$\begin{aligned}&Rate\ of\ heat\ transfer\\&= Q = h_x A(T_{sat} - T_s)\\&= 5982 \text{ x } 0.12564 \text{ x}(373.15 - 333.15) = 30063.14\ W\end{aligned}$$

$$Mass\ of\ condensate = \dot{m} = \frac{h_x A(T_{sat} - T_s)}{\lambda} = \frac{30063.14}{2310 \text{ x } 10^3} = 0.014\frac{kg}{s} = 46.85\frac{kg}{hr}$$

Mass of condensate and the heat transfer rate will be maximum for the arrangement with 10 horizontal tubes in the vertical direction.

Example 5: Consider a stainless steel pan with water at atmospheric pressure. Externally heat is supplied, to boil the liquid, through the bottom of pan (diameter is 40 cm) and a temperature of 106°C is maintained at the inner surface of the bottom of the pan. Considering steady state condition, determine the rate of heat transfer to the water and the rate of evaporation of water.

Data: The saturation temperature of Tsat = 100°C and the properties of water at this condition are.

$\sigma = \mathbf{0.0589}$ **N/m**	$\mathbf{h_{fg}}$ **= 2257 x $\mathbf{10^3}$ J/kg**	$\mathrm{Pr}_l = \mathbf{1.75}$
ρ_l =957.9 kg/m^3	ρ_v =0.6 kg/m^3	$\mu_l = 0.282 \text{ x } 10^{-3}$ kg · m/s
$C_{pl} = 4217$ J/kg · °C	$C_{sf} = 0.013$	n = 1

Answer:

The rate of heat transfer will be equal to the product of heat flux and heat transfer area.

Here, $\Delta T_{excess} = (T_S - T_{Sat})$ = excess temperature = temperature of the supplied liquid – saturation temperature of liquid = 106–100 = 6°C.

Excess temperature resembles the stage of nucleate boiling of liquid. Hence, the heat flux at this nucleate boiling condition will be (Rohsenow, [9])

$$q = \mu_l h_{fg} \left[\frac{g(\rho_l - \rho_v)}{\sigma} \right]^{½} \left[\frac{C_{pl} \Delta T_{excess}}{C_{sf} h_{fg} (Pr_l)^n} \right]^3$$

$$q = 0.282\text{x}10^{-3}\text{x}2257\text{x } 10^3 \left[\frac{9.81 \text{ x } (957.9 - 0.6)}{(0.0589)} \right]^{½} \left[\frac{4217 \text{ x } 6}{(0.013)\text{x } \left(2257 \text{ x } 10^3\right)\text{x}(1.75)^1} \right]^3$$

$$q = 30409.35 W/_{m^2}$$

Heat transfer area or the surface area of the bottom of the pan:

$$A = \frac{\pi}{4} D^2 = \frac{\pi}{4} \times (0.4)^2 = 0.125 m^2$$

The rate of heat transfer will be = Q = q x A = 30409.35$W/_{m^2} \times 0.12 m^2 = 3821\ W$.

Again, the rate of evaporation of water ($\dot{m}$) can be determined from the formula:

$$Q = \dot{m} * \lambda$$

$$\dot{m} = \frac{Q}{\lambda} = \frac{3821}{2257 \text{ x } 10^3} = 1.692 \text{ x } 10^{-3} \frac{kg}{s} \approx 1.7\, gram\ of\ water\ evaporates\ per\ sec$$

Nomenclature

Q	Amount of heat transferred from a solid surface to the fluid, W
a_x	Acceleration of small selected volume element of condensate, m/s^2
h_N	Average heat transfer coefficient for N tubes, W/m^2.K
h_1	Average heat transfer coefficient for top horizontal tube, W/m^2.K
h_{avg} *or* h	Average heat transfer coefficient value, W/m^2.K
$h_{Inclined}$	Average heat transfer coefficient value for inclined surface, W/m^2.K
$h_{Horizontal}$	Average heat transfer coefficient value for horizontal surface, W/m^2.K
$h_{Vertical}$	Average heat transfer coefficient value for vertical surface, W/m^2.K
q_{max}	Critical (or maximum) heat flux, W/m^2
D	Diameter of the horizontal tube or sphere, m
x	Distance along the length of the solid surface, m
y	Distance along the width of the condensate, m
ρ_l	Densities of the condensed liquid, kg/m^3
ρ_V	Densities of the condensed vapour, kg/m^3
h_{fg} or λ	Enthalpy or latent heat of vaporisation, J/kg
C_{sf}	Experimental constant depends on the fluid-surface combination
n	Experimental constant depends on the fluid
ΔT_{excess}	Excess temperature supplied to fluid, °C, K

δ	Film thickness of condensate, m
δx	Film thickness of condensate along the L direction, m
F_x	Force on small selected volume element of condensate, kg/m.s^2
g	Gravitational acceleration, m/s^2
q	Heat flux, W/m^2
h_x	Heat transfer coefficient along the L direction, W/m^2·°C, W/m^2·K
L or H	Height of the solid surface, m
Θ	Inclination angle of a solid surface, ^{0}
$\dot{m}$	Mass flow rate of condensate along the length of condensate, kg/s
m	Mass of small selected volume element of condensate, kg
N	Number of horizontal tubes arranged in stacks
Pr_l	Prandtl number of liquid
C_{pl}	Specific heat of the liquid, J/kg. ^{o}C
C_{pv}	Specific heat of the vapour, J/kg. ^{o}C
A	Surface area of solid surface, m^2
σ	Surface tension in the vapour-liquid interface, N/m
Tsat	Saturation temperature of the condensing fluid, °C or K
Ts	Surface temperature of the solid surface, °C or K
k_l	Thermal conductivity of the condensed liquid, W/m·°C or W/m·K
k_v	Thermal conductivity of the vapour, W/m·°C or W/m·K
u	Velocity rate of condensate along the length of condensate, m/s
μ_l	Viscosity of the condensed liquid, kg/m·s
μ_v	Viscosity of the vapour, kg/m·s
b	Width of the solid surface, m

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