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Review on Heat Transfer in Boiling of Binary Mixtures

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Abstract

Heat transfer is described in nucleate pool boiling and forced convective boiling of pure liquid and their non-azeotropic mixtures at low mass flow rate where the Ocean Thermal Energy Conversion system and the Kalina cycle may operate using a limited temperature difference between a hotter liquid on the sea surface and a cooler liquid at a depth of the sea.

1. Introduction

Though heat transfer in boiling of mixture was first paid attention in many chemical engineering process, it recently has received attention for improvement in a thermal efficiency in a power plant and in a coefficient of performance in a heat pump and refrigeration cycles. In the case of pure liquid, its saturation temperature is fixed at a constant value under a constant pressure, while for a binary mixture, its temperature does change along a boiling point curve on a phase equilibrium diagram with a variation of concentration as shown in Fig.1. In addition to this, its change is responsible for the difference between corresponding saturation temperatures for the liquids. The characteristics that the saturation temperature of the binary mixture can be changed by the concentration, makes it suitable to enhance the thermal efficiency of the Lorentz cycle during which the temperatures of heat source and heat sink vary due to the heat transfer.

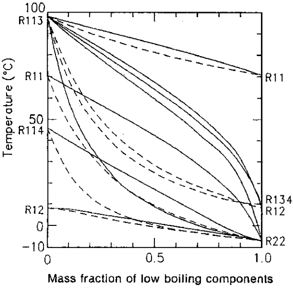


Figure 1 Phase equilibrium diagram of mixtures at p = 4 bar (— Dew point curves; - -Boiling point curves)

More recently, in order to improve the thermal efficiency of the Ocean Thermal Energy Conversion system which is designed to operate at a limited small temperature difference between a hotter liquid on the sea surface and a cooler liquid at a depth of the sea, one conceives a binary mixture as a candidate of test liquid. Kalina proposed the Kalina cycle with a regenerator, a separator of vapor and liquid and a absorber and succeeded in improving its thermal efficiency by using the mixture of water and ammonia.. Since these systems generally make use of a small temperature difference for energy conversion, in these evaporator tubes, wall superheat or heat flux and in addition to these mass velocity(flow rate) naturally become relatively low.

Therefore, the state of knowledge on boiling heat transfer first in pure liquids and then their non-azeotropic mixtures will be described only at a low mass velocity here. It may be necessary to say that although heat transfer problems caused during the operation of the cycles have so far received considerably less attention than those of the single component systems due to a complexity of the heat transfer process, much attention becomes stressed on these problems involved.

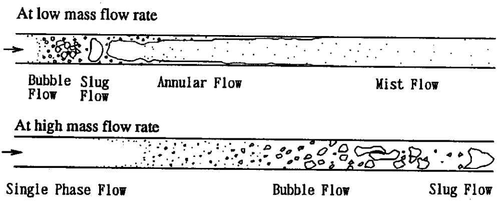
2. Features of Flow Structure in a Heated Tube

The forced convective boiling in an evaporator tube is categorized by the development a flow configuration along the tube length because of the variation in flow state due to phase change by heat addition. Therefore, before discussing heat transfer problems we have to keep the flow situation formed in the tube in mind.

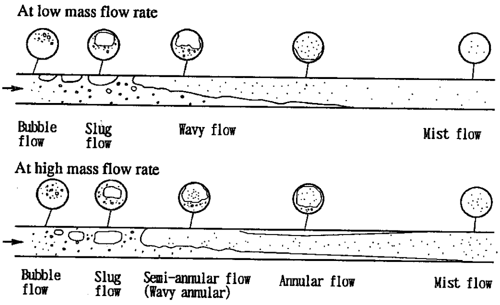
Figure 2 illustrates schematically liquid-vapor flow patterns in a vertical evaporator tube and in a horizontal evaporator tube at low and high mass flow rates, respectively. As shown in Fig.1, the features of the flow are significantly different between vertical and horizontal ones due to an effect of the gravity. This difference becomes of importance in evaluating the heat transfer rates.

In the case of the vertical flow, roughly speaking, the liquid and vapor symmetrically flows along the tube length because of no gravitational effect. The flow pattern changes from bubble flow through slug flow and annular flow to mist flow. The position at which the flow pattern changes, strongly depends both on the mass flow rate and on the vapor generation rate which is controlled by either heat flux or temperature difference. In the case of a low mass flow rate and a small temperature difference, the bubble and annular flow regions becomes relatively longer as compared with the regions of the another flow patterns.

In addition to this, in the case of the horizontal flow since an inertia force is coupled with the gravitational force, the development of the flow pattern strongly depends on the mass flow rate. As a result, the flow patterns are more complicated in the horizontal tube than those in the vertical tube. At a low mass flow rate, or example, the flow pattern changes from bubble flow through slug flow and wavy flow to mist flow. In the wavy flow, the liquid and vapor flow separately in the lower and upper parts of the tube. At a high mass flow rate, the flow pattern develops from bubble flow through slug flow and semi-annular or wavy-annular flow, annular flow, finally to mist flow. In the annular flow, a liquid film covers the whole tube perimeter on which its thickness usually varies from the top to the bottom due to gravity.



For a vertical evaporator tube



For a horizontal evaporator tube

Figure 2 Flow patterns in a vertical and horizontal evaporator tubes at low and high mass flow rates

3. Characteristic of Flow Boiling Heat Transfer

Although the flow boiling heat transfer in the evaporator is strongly affected by the flow pattern, it generally consists of two different mechanisms; one is nucleate boiling of the liquid and the other is forced convection of the liquid with evaporation at the liquid-vapor interface. Since heat transfer is related closely to the flow pattern, it should be separately treated with depending on whether the surface is submerged in a large volume of liquid like a bubble flow or covered with a thin liquid film like an annular flow, and then in case of the thin liquid film on whether the direction of the surface is vertical or horizontal as shown in Fig.2. When our attention is focused on the low mass flow rate, the length of the transition region like a slug flow appearing between bubble and annular flows would be expected to be relatively short. Therefore, the heat transferred at the low mass flow rates is mainly governed by both the nucleate boiling and the forced convection. The heat transfer will be separately described here for the pure liquid and the binary mixture.

4 Correlation of heat transfer for pure liquid

4.1 Nucleate boiling heat transfer

The nucleate boiling heat transfer plays in essential role in heat transfer in the bubble flow region and then some contributions are always drawn into the another flow pattern. Therefore, it is of importance to know the characteristic of nucleate boiling even in dealing with heat transfer in flow boiling.

Many correlations are so far proposed to predict the heat transfer coefficients for the nucleate boiling as listed in Table 1. The nucleate heat transfer is subject to the properties of any particular liquid-surface combination but is hardly affected by the forced convection(the mass flow rate) and the orientation of the surface that would be easily understood from the fact that these effects are not included into any correlation in Table 1. It is very difficult to recommend which correlation is the best appropriate because each one inherits from what effects is mainly taken into account in its derivation.

4.2 Convective heat transfer in the vertical evaporator tube

There exist two ways of how to predict the heat transfer the annular flow. One is based on a relation of the form:

$$\frac{h_{TP}}{h_{Lo}} = f\,(X_{tt}, Bo) \tag{1}$$

where h_{LO} is the heat transfer coefficient if the pure liquid was flowing through the tube, X_{tt} is the Lockhart-Martinelli parameter, which plays an important role in determining the flow pattern:

$$X_{tt} = \left(\frac{1-x}{x}\right)^{0.9} \left(\frac{\rho_v}{\rho_v}\right)^{0.5} \left(\frac{\mu_L}{\mu_v}\right)^{0.1} \tag{2}$$

and

$$Bo = \frac{q}{G\,H_{fg}} \tag{3}$$

Table 2 gives an example of this types of correlations.

The second one is

$$h_{TP} = S\,h_{NB} + F\,h_{con} \tag{4}$$

where h_{TP} is the local heat transfer coefficient, S h_{NB} is the contribution due to the nucleate boiling, F h_{CON} the contribution due to convection with evaporation, h_{NB} is the heat transfer coefficient of ordinary pool boiling and hCON is the heat transfer coefficient for the two phase flow. The mostly used correlation for saturated boiling in the annular flow is the Chen correlation[14] and later his original correlation was modified by Bennett and Chen[15] which succeeded in correlating all the forced convective boiling heat transfer data for water and organic systems. Since this correlation stands on the assumption that both mechanisms occur to some degree over the entire range of the correlation and the contributions made by the two mechanisms are additive, the multiplier, F, means the effect of the increase in liquid velocity, which is caused by the existence of vapor flow and then the other factor, S, means the effect of the suppression of nucleate boiling due to the steeping of the temperature gradient in the liquid film on the wall, which is caused by the forced convection. Chen proposes F hCON using the multiplier F and the Dittus-Boelter correlation as:

$$h_{CON} = 0.023\,\frac{k_L}{D}\left(\frac{G(1-x)D}{\mu_L}\right)^{0.8}\left(\frac{c_L\mu_L}{k_L}\right)^{0.4} \tag{5}$$

and

$$F = \left[\frac{\left(\frac{dp}{dz}\right)_{TP}}{\left(\frac{dp}{dz}\right)_{L}}\right]^{0.444}\left(\frac{\mathrm{Pr}_L+1}{2}\right)^{0.444} = \begin{cases} 1 & 1/X_{tt} \le 0.1 \\ 2.35\left[\frac{1}{X_{tt}}+0.213\right]^{0.736} & 1/X_{tt} \ge 0.1 \end{cases}$$

(6)[16]

and S h_{NB} using the Forster and Zuber[10] analysis and the suppression factor as:

$$h_{NB} = 0.00122\left(\frac{k_L^{\,0.79}\,c_{pl}^{\,0.45}\,\rho_L^{\,0.49}}{\sigma^{0.5}\,\mu_L^{\,0.29}\,H_{fh}^{\,0.24}\,\rho_v^{\,0.24}}\right)\Delta T_{sat}^{\,0.24}\,\Delta p_{sat}^{\,0.75} \tag{7}$$

where $\Delta T_{sat} = T_w - T_{sat}$ and $\Delta p_{sat} = p_{sat}(T_w) - p_{sat}(T_{sat})$ and

$$S = \left(\frac{1}{1+2.53x10^{-6}\,F^{1.46}\,\mathrm{Re}_L^{\,1.17}}\right) \tag{8}$$

It might be expected that the suppression factor, S, would approach unity at low flows and zero at high mass flows. Chen[14] suggests that S can be first represented as a function of the local two-phase Reynolds number, $Re_{TP} = F^{1.25}\,Re_L$, $Re_L = G(1\text{-}x)D/\mu_L$ and the Bennet and Chen[15] improved it to give equation(8). The functions F and S were determined empirically from experimental data using an iterative procedure to obtain the best solutions. It is necessary to say finally that the other correlation for the nucleate boiling is available though Chen employed only the Forster and Zuber correlation, while most of correlations employed the Dittus-Boelter correlation for convective heat transfer.

Recently, Gungor and Winterton[17], for example, derived a new generalized correlation using the Cooper correlation[9] in Table 1 for the nucleate boiling and the Dittus-Boelter correlation and determined the two factors F and S as:

$$F = 1 + 24000 Bo^{1.16} 1.37 (1/X_{tt})^{0.86} \tag{9}$$

$$S = \left(\frac{1}{1 + 2.53 \times 10^{-6} F^{1.46} \mathrm{Re}_L^{1.17}} \right) \tag{8}$$

It is concluded that flow boiling heat transfer, for saturated and subcooled conditions, vertical and horizontal flow, tubes and annuli, can be predicted with reasonable accuracy for over 4300 data points for waters, refrigerants and ethylene glycol and so on by their correlation.

4.3 Convective heat transfer in the horizontal evaporator tube

Figure 2 shows that the flow patterns in the horizontal tube are different from those in the vertical tube, especially in the annular flow where most of the heat exchange is done at the low mass flow rate, liquid and vapor separately flow each other. As the result, the difference in heat transfer between the upper and lower parts becomes significantly large. Taking account for this respect, Yoshida et al.[18] developed their method to predict the heat transfer in the vertical tube based on the Chen concept in which the circumferential averaged or axially local heat transfer coefficient for the separated and annular flows is treated with respectively during the flow boiling inside horizontal smooth tubes. This method shows the best precision in predicting flow boiling heat transfer of pure refrigerants among correlation proposed hitherto.

The procedure proposed by Yoshida et al.[18] is introduced as follows:

(1) Calculate a separation angle ϕ_s with equation (11)

$$\frac{\phi_s}{\phi_o} = 1 + 0.72 \left[\left(\frac{1-x}{x} \right)^{0.9} \left(\frac{\rho_v}{\rho_v} \right)^{0.5} \right]^n \left(\frac{\mu_L}{\mu_v} \right)^{0.17} \tag{9}$$

where

$$n = 0.22 \left(\frac{G^2}{gD\rho_v(\rho_L - \rho_v)} \right)^{0.38} \left(\frac{q}{GH_{fg}} x10^4 \right)^{-0.06} \left(\frac{gD}{H_{fg}} x10^3 \right)^{-0.27} \tag{10}$$

and ϕ_o is a separation angle estimated by assuming that there is no slip between the liquid and vapor flows and that the liquid-vapor interface is a horizontal plane.

(2)When $\phi_s > 162$ deg, set ϕ_s equal to 180 deg. In this case the flow regime is annular. Then, calculate the local heat transfer coefficient by equation(11) which is applicable to the annular flow.

$$h_{TP} = S\, h_{NB} + F\, h_{con} \tag{11}$$

where

$$h_{CON} = 0.023 \frac{k_L}{D} \left(\frac{G(1-x)D}{\mu_L} \right)^{0.8} \mathrm{Pr}_L^{0.4} \tag{12}$$

$$F = 1 + 2 X_{tt}^{-0.88} \tag{13}$$

$$h_{NB} = 207 \frac{k_L}{D_b} \left(\frac{qD_b}{k_L T_b} \right)^{0.745} \left(\frac{\rho_v}{\rho_L} \right)^{0.581} \mathrm{Pr}_L^{0.533} \tag{14}$$

$$D_b = 0.51 \left[\frac{2\sigma}{g(\rho_L - \rho_v)} \right]^{0.5} \tag{15}$$

$$S = \frac{1}{1 + 0.9 \left[\frac{G(1-x)D}{\mu_L} x F^{1.25} \Big/ 10^4 \right]^{0.5} \left(\frac{q}{GH_{fg}} x10^4 \right)^{-0.5} X_{tt}^{-0.5}} \tag{16}$$

(3)When $\phi_s < 162$ deg, the flow regime is a separated one. In this case, the circumferential averaged heat transfer coefficient must be calculated by taking account of the circumferential heat conduction within the tube wall.

In the case of a material with a high thermal conductivity such as copper, the variation of the wall temperature is approximately compensated by the circumferential heat flow so that the heat transfer rate in the dry part of the tube perimeter can be ignored. The circumferential averaged or axially local heat transfer coefficient can be calculated from equation (17), where only the heat transfer in wet part is considered

$$h_{TP} = \frac{\phi_s}{180} h_{wet} \tag{17}$$

where h_{wet} is the heat transfer coefficient for the wet part and is calculated from equation(18)

$$h_{wet} = F\, h_{CON} + S_{wet}\, h_{NB} \tag{18}$$

$$S_{wet} = \frac{1}{1 + 1.2 \left[\frac{G(1-x)D}{\mu_L} x F^{1.25} \Big/ 10^4 \right]^{0.3} \left(\frac{q}{GH_{fg}} x10^4 \right)^{-0.3}} \tag{19}$$

$$q_{wet} = \frac{180}{\phi_s} q \tag{20}$$

In the case of other material, it is needed to take into account of the heat transfer in the dry part due to the large temperature difference between the wall and vapor. The heat transfer coefficient in the dry part hdry is given as follows:

$$\frac{h_{dry}}{h_v} = 1 + 1.53 \left[\left(\frac{GxD}{\mu_v} \right) \Big/ 10^4 \right]^{-1.62} \left[\frac{G^2}{gD\rho_v(\rho_L - \rho_v)} \right]^{-0.98} \tag{21}$$

where

$$h_{CON} = 0.023 \frac{k_L}{D} \left(\frac{G(1-x)D}{\mu_L} \right)^{0.8} \mathrm{Pr}_L^{0.4} \tag{22}$$

On the other hand, Gungor and Winterton[17] revised their generalized correlation for the vertical flow by taking the flow effects into consideration using the Froud number, Fr, leading to the result that each factor for the vertical tube multiplied by $F_{hori} = Fr^{(0.1 - 2 Fr)}$ and $S_{hori} = Fr^{1/2}$, respectively, is enough to predict the heat transfer coefficient in the horizontal tubes.

Jung et al.[47] also proposed their correlation using the Stephan and Abdelsalam in place of the Forster and Zuber correlation in Table 1 for the nucleate boiling and the Dittus-Boelter correlation and determined the two factors F and S as:

$$F = 2.37 \left(0.29 + \frac{1}{X_{tt}} \right)^{0.85} \tag{23}$$

$$S = 4048\, X_{tt}^{1.22}\, Bo^{1.13} \tag{24}$$

They reported that their correlation was better in predicting the heat transfer than either the Bennett and Chen and the Gungor and Winterton correlations

5. Correlation of heat transfer for binary mixture

5.1 Nucleate boiling heat transfer

It is generally known that heat transfer coefficient in nucleate boiling of binary mixture is much lower than that expected from the individual components of the mixture. This is illustrated as an example in Fig.3, where the heat transfer coefficients in a boiling of R12-R113 and R22-R11 mixtures, the temperature difference between boiling and dew points on phase equilibrium at constant pressure and the difference of the mass fraction in liquid and vapor phase, are plotted over the liquid mass fraction of R12 and R22. The heat transfer coefficients are surprisingly much lower than the values, obtained when linearizing between the pure components. The same behavior is found to appear in another mixtures. The mechanism that heat transfer coefficients are sharply reduced is considered: Individual components pass from the liquid to the vapor phase in different proportions, and the faster evaporation of the more volatile component causes an enrichment of the bubble forming boundary layer with the less volatile component, so that the local boiling temperature increases. There are so far two different ways of how to estimate this effect. One is related to the temperature differences between boiling and dew points and the other one to the difference of the mass fraction in liquid and vapor phase. Many correlations proposed so far can be roughly categorized into two from the physical point of view as listed in Table 3. The reduction in the heat transfer coefficient is considered to be attributed to the difference of the mass fraction as pointed out by the most of the correlations in Table 3.

Most of the correlations can be roughly given in the following form:

$$\frac{h}{h_{id}} = \frac{1}{C_{mix}} \quad , \quad h_{id} = \frac{1}{\dfrac{x_1}{h_1} + \dfrac{x_2}{h_2}} \tag{25}$$

except for the case of Jungnickel[28] employing $h_{id} = h_1 x_1 + h_2 x_2$. The constant C_{mix} is always larger than unity whose magnitude largely depends on the mass fraction. The constant C_{mix} means a level of the reduction in heat transfer coefficient due to the effect of mass transfer.

Some experiments for the nucleate boiling are, for reference, summarized in Table 4.

Recently, Inoue and Monde[32] and Inoue et al. [57]have measured boiling heat transfer coefficients for binary mixtures with a large between the corresponding pure substances(see Fig.1) and pointed out that the Jungnuckel et al.[28] and the Schlunder[29] correlations which stand on the difference of the mass fraction are rather different in a character from the measured data as shown in Fig.4. while the Thome correlation[26] has a similar trend, but underpredicts the data as shown in Fig.5. Therefore, they have devised the Thome correlation by taking into account of the effect of the heat flux on the basis of observation of boiling behavior(ref. see Fig.6 in Inoue and Monde[32]) on a horizontal wire and succeeded in correlating their data well.

Figures 6 and 7 show a comparison of their new correlation[57] and the measured data for mixtures with a large difference in saturation temperatures and with a small difference in saturation temperatures, respectively.

Fujita and Tsutsui[33] separately propose the correlation from the similar concept as estimated from comparison of both equations listed in Table 3. Their correlation can predict their data for methanol-water, methanol-ethanol, ethanol-n-butanol, and methanol-benzene mixtures with a good agreement. It is worth to say that the correlations in which only the values determined from phase equilibrium diagram are used, are easier for calculating the heat transfer coefficient than those using the physical properties depending on the mass fraction.

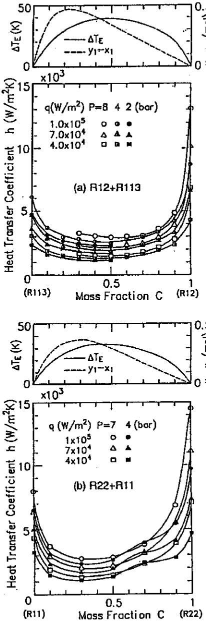


Figure 3 Effect of the mass fraction on the heat transfer coefficients together with relations of DTE and (y_1 -x_1)

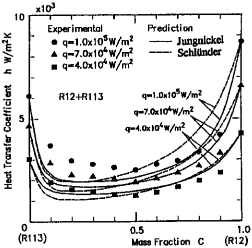


Figure 4 Comparison of the the Jungnuckel et al. [28] and the Schlünder[29] correlations and their data

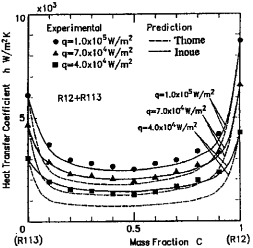
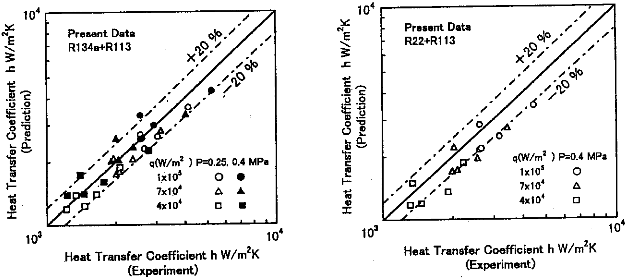


Figure 5 Comparison of the Thome et al. [26] correlation and their data



(a) R134a + R113

(b) R22 + R113

Figure 6 Comparison of the Inoue and Monde correlation with their measured data

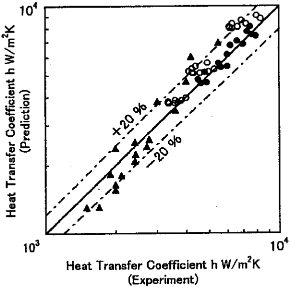


Figure 7 Comparison of the Inoue and Monde correlation with existing data

- ○ Data from Jungnickel et al.: R22+R12; C = 0.1 - 0.9; p = 2, 4, 10 bar; q = 4 x 104 W/m2
- ▲ Data from Gorenflo et al.: R22+R114; C = 1 - 0.9; p/pc = 0.2, 0.4, 0.6, q = 104 W/m2
- ● Data from Fujita et al.: R11+R113; C = 0.2 - 0.8; p = 1, 2, 4, 8 bar, q = 105 W/m2

5.2 Convective heat transfer in the vertical evaporator tube

The heat transfer in forced convective boiling is also much lower for the mixtures than that for the pure liquid as estimated from the result obtained in the nucleate boiling. Bennett and Chen[15] developed the Chen correlation for the pure liquids by taking into account of mass transfer effects, so as to extend to the aqueous mixtures of ethylene glycol and then gave the following equation for the heat transfer of the mixtures.

$$h_{mix} = h_{CON} F_{mix} + h_{NB} S_{mix} \tag{26}$$

where

$$F_{mix} = F\, f(\mathrm{Pr}_L) \left[\frac{\Delta T}{\Delta T_s}\right]_{CON} \tag{27}$$

$$f(\mathrm{Pr}_L) = \left(\frac{\mathrm{Pr}_L + 1}{2}\right)^{0.444} \tag{28}$$

$$\left[\frac{\Delta T}{\Delta T_s}\right]_{CON} = 1 - \frac{(1-y_1)q}{\rho_L H_{fg} h_{mix} \Delta T_s} \left[\frac{dT_{sat}}{dx}\right]_{p,bulk} \tag{29}$$

$$h_{mx} = 0.023\, \frac{D_{dif,L}}{D} \mathrm{Re}_L{}^{0.8} \left[f(\mathrm{Pr}_L) F \right] \left(\frac{\mu_L}{D_{dif,L}}\right)^{0.4} \tag{30}$$

$$S = \left(\frac{1}{1 + 2.53x10^{-6} \left[f(\mathrm{Pr}_L) F \right]^{1.46} \mathrm{Re}_L{}^{1.17}}\right) \Big/ C, \tag{31}$$

$$C = \left[\frac{1}{1 - \frac{c_{pL}(y_1 - x_1)}{H_{fg}} \left[\frac{dT_{sat}}{dx}\right] \left(\frac{a_L}{D_{dif,L}}\right)^{1/2}}\right]$$

Recently, Celata et al.[42] show that the Bannett and Chen correlation can applicable for the binary mixtures of R12-R114 and is superior to the type of correlation such as proposed by Mishra et al.[43].

Table 5, for reference, listed some recent experimental studies of heat transfer for mixtures in flow boiling.

5.3 Convective heat transfer in the horizontal evaporator tube

A significant reduction in the heat transfer coefficient appears for binary mixtures attributable to mass transfer effects.

Jung et al.[49] proposed their following correlation which is derived based on the Chen concept but different for the Bannett and Chen correlation, and reported that it can predict their data for the mixture of R22-R114 with a mean deviation of 9.6 %.

$$h_{mix} = h_{CON} F_{mix} + h_{NB,mix} S_{mix} \tag{32}$$

where $h_{NB,mix}$ is the Ünal correlation, h_{CON} is also the Dittus-Boelter equation, $S_{mix} = S/C_{\ddot{u}n}$, $C_{\ddot{u}n}$ is the constant in the Ünal correlation, S is given by equation(24) and $F_{mix} = F\, C_{me}$, F is given by equation(23) and C_{me} is proposed as:

$$C_{me} = (1 - 0.35 |y_1 - x_1|^{1.56}) \tag{33}$$

Jung et al.[49] recommend to use the Stephan and Abdelsalam correlation for the heat transfer coefficients for the individual components needed when calculating $h_{NB,mix}$. They concluded that their correlation is the best one in correlating their data among the other correlations[15][17][46].

It may be noticed finally that most of the correlations predicting heat transfer for the annular flow derived based on the Chen criterion in which the Dittus-Boelter equation is commonly employed for heat transfer in forced convective flow, while the correlation for the nucleate heat transfer seems to be arbitrary chosen from any correlation in Table 3. The reduction in the heat transfer for the mixture due to the mass transfer effects seems to be treated with from the two different points of view not only in nucleate boiling but also in convective boiling: one is focused on the temperature difference, and the other one the concentration difference.

6.Concluding Remarks

The interest in flow boiling heat transfer in mixtures has been increased. Further experimental works with different mixtures must be done on a wide range of compositions to verify the known correlations and to improve them. It would be hoped to make it clear how the mass transfer influences the heat transfer, especially in the field of nucleate boiling.

Nomenclature

a :Thermal diffusivity
B_o :Boiling number (= $q/\rho_L H_{fg}$)
C :Mass fraction of a more volatile component in a system
C_p :Specific heat at constant pressure
d_o :Diameter of bubble departure
D :Tube diameter of Mass diffusivity
f_p :Pressure factor
f_s :Surface factor
f_ζ :Nucleation factor
F :Enhancement factor
Fr :Froude number {= $G^2/(\rho_L{}^2 gD)$}
g :Acceleration of gravity
G :Mass flux
h :Heat transfer coefficient
h_{id} :Ideal heat transfer { = $1/(x_1/h_1 + 1/(x_2/h_2)$ }
H_{fg} :Latent heat in vaporization
l :Representative length of heated surface
M :Molecular weight
P :Pressure
P_c :Critical pressure
Pr :Prandtl number
q :Heat flux
R_s :Smoothness degree of heated surface
Re_L:Reynolds number (= $G(1 - x)D/\mu_L$)
S :Suppression factor
T :Temperature

T_c :Critical temperature
T_s :Saturation temperature
$T_{w,id}$:Ideal wall temperature { = ($T_{bulk} + \Delta T_{id}$) }
ΔT_E :Temperature difference between dew and bubble point
{ =($T_{local,max} - T_{bulk}$)
ΔT_{id} :Ideal wall superheat{ = ($\Delta T_{sat,1}\, x_1 + \Delta T_{sat,2}\, x_2$)}
ΔT_{sat} :Wall superheat
x, y :Quality or Mole fractions of liquid and vapor
x_m, y_m :Mass fraction of liquid and vapor
X_{tt} :Lockhart-Martinelli parameter

Greek symbol

α :Volume expansion coefficient
β :Mass transfer coefficient
κ :Thermal conductivity
λ :Thermal conductivity
μ :Viscosity
ν :Kinematic viscosity
ρ :Density
σ :Surface tension
σ

Suffix

1 :More volatile component
2 :Less volatile component
bulk :Bulk
CON :Convective flow
dry :Dry surface
id :Ideal
L :Liquid
local :Local
mix:Mixture
NB :Nucleate
Two phase flow
v :Vapor
wet :Wet surface

Table 1 Correlations for pool boiling in pure liquid

Author	Correlation
Kutateladze[1]	$N_{ub}=7.0\times10^{-4}R_{eb}^{0.7}P_{rL}^{0.35}f_P^{0.7}$ $N_{ub}=\frac{hb}{\lambda_L}$, $R_{eb}=\frac{qb}{H_{fg}\rho_v\nu_L}$ $f_P=\frac{P}{\sqrt{\sigma g(\rho_L-\rho_v)}}$, $b=\sqrt{\frac{\sigma}{g(\rho_L-\rho_v)}}$
McNelly[2]	$N_u=0.225\left(\frac{qd}{H_{fg}\mu_L}\right)^{0.69}\left(\frac{pd}{\sigma}\right)^{0.31}\left(\frac{\rho_L-\rho_v}{\rho_v}\right)^{0.33}P_{rL}^{0.69}$
Rohsenow[3]	$N_{ub}=\frac{P_{rL}^{-0.7}}{C_{sf}}\left(R_{eb}\frac{\rho_v}{\rho_L}\right)^{0.667}$, $C_{sf}=0.00225\sim0.0154$
Forster and Greif[4]	$N_{ub}^{*}=0.0346\sqrt{R_{eb}^{*}}\left(\frac{\sigma}{\rho_L\nu_L}\frac{b^*}{\nu_L}\right)^{0.125}P_{rL}^{0.6}$ $N_{ub}^{*}=\frac{hb}{\lambda_L}$, $R_{eb}^{*}=\frac{qb^*}{H_{fg}\rho_L\nu_L}$, $b^*=\frac{\rho_L C_{PL}T_s}{\rho_v H_{fg}}\frac{\sigma}{\rho_v H_{fg}}$
Zuber[5]	$N_u=\text{const.}\left\{\frac{gl^3}{\nu_L a_L}\left(\alpha\Delta T_{sat}+\frac{nb}{A}\frac{\pi}{6}d_0^2\frac{\rho_L-\rho_v}{\rho_L}\right)\right\}^{1/3}$ $d_0=0.021\theta\left\{\frac{\sigma}{g(\rho_L-\rho_v)}\right\}^{1/2}$
Borishanskii[6]	$N_{ub}=8.7\times10^{-4}R_{eb}^{0.7}P_{rL}^{0.7}f_p^{0.7}$
Nishikawa and Fujita[7]	Laminer Flow : $Y=6.24(f_\zeta f_P X)^{2/3}$ Turburent Flow : $Y=0.661^{-2/5}(f_\zeta f_P X)^{4/5}$ $Y=\frac{hl}{\lambda_L}$, $X=\left\{\left(\frac{1}{M^2N}\right)\frac{C_{PL}\rho_L^2 g}{\lambda_L\sigma H_{fg}\rho_v}\right\}^{1/2}ql^{3/2}$ $N=1.976\ W$, $M=900\ m^{-1}$, $f_P=\left(\frac{P}{P_s}\right)^{0.7}\left\{1+3\left(\frac{P}{P_c}\right)^3\right\}$
	Low Heat Flux : $h_L=1140\frac{P_C^{1/3}}{\mathfrak{M}^{1/6}T_C^{5/6}}f_{PL}(f_\zeta q)^{2/3}$ $f_{PL}=\left(\frac{P}{P_C}\right)^{0.19}\left\{1+2\left(\frac{P}{P_C}\right)^2+\left(\frac{P}{P_C}\right)^8\right\}$ High Heat Flux : $h_H=492\frac{P_C^{1/5}}{\mathfrak{M}^{1/10}T_C^{9/10}}f_{PH}(f_\zeta q)^{4/5}$ $f_{PH}=\left(\frac{P}{P_C}\right)^{0.23}\left\{1+2\left(\frac{P}{P_C}\right)^2+8\left(\frac{P}{P_C}\right)^8\right\}$

Table 1 Correlations for pool boiling in pure liquid(Continued)

Author	Correlation
Stephan and Abdelsalam[8]	Water : $N_u=0.246\times10^{7}\left(\frac{qd}{\lambda_L T_{sat}}\right)^{0.673}\left(\frac{C_p d^2 T_{sat}}{a_L{}^2}\right)^{1.26}$ $\times\left(\frac{H_{fg}d^2}{a_L{}^2}\right)^{-1.58}\left(\frac{\rho_L-\rho_v}{\rho_L}\right)^{5.22}$ $10^{-4}<\frac{P}{P_c}<0.886\ ,\quad \theta=45^{\circ}$
	Hydrocarbons : $N_u=0.0546\left(\frac{qd}{\lambda_L T_{sat}}\right)^{0.67}\left(\frac{H_{fg}d^2}{a_L{}^2}\right)^{0.248}$ $\times\left(\frac{\rho_v}{\rho_L}\right)^{0.335}\left(\frac{\rho_L-\rho_v}{\rho_L}\right)^{-4.33}$ $5.7\times10^{-3}<\frac{P}{P_c}<0.9\ ,\quad \theta=35^{\circ}$
	Cryogenic fluids : $N_u=4.82\left(\frac{qd}{\lambda_L T_{sat}}\right)^{0.624}\left(\frac{C_{PL}d^2 T_{sat}}{a_L{}^2}\right)^{0.374}\left(\frac{H_{fg}d^2}{a_L{}^2}\right)^{-0.329}$ $\times\left(\frac{\rho_v}{\rho_L}\right)^{0.257}\left(\frac{\rho_W C_{PW}\lambda_W}{\rho_L C_{PL}\lambda_L}\right)^{0.117}$ $4\times10^{-3}<\frac{P}{P_c}<0.97\ ,\quad \theta=1^{\circ}$
	Refrigerants : $N_u=207\left(\frac{qd}{\lambda_L T_{sat}}\right)^{0.745}\left(\frac{\rho_v}{\rho_L}\right)^{0.581}P_{rL}{}^{0.533}$ $3\times10^{-3}<\frac{P}{P_c}<0.78\ ,\quad \theta=35^{\circ}\ ,\quad d=0.16\,\theta\sqrt{\frac{2\sigma}{g(\rho_L-\rho_v)}}$
Cooper[9]	$h=55\left(\frac{P}{P_c}\right)^{(0.12-0.21\log R_a)}\left\{-\log\left(\frac{P}{P_c}\right)\right\}^{-0.55}\mathfrak{M}^{-0.5}f_s q^{0.67}$ $f_s=1\sim1.7$
Forster and Zuber[10]	$\frac{q}{\rho_v H_{fg}}\left(\frac{\pi}{\alpha_L}\right)^{1/2}\left(\frac{\rho_L r^{*3}}{2\sigma}\right)^{1/4}=0.0015\left\{\frac{\rho_L}{\mu_L}\left(\frac{\Delta T_{sat}\lambda_L}{\rho_v H_{fg}}\right)^2\frac{\pi}{\alpha_L}\right\}^{5/8}\left(\frac{C_p\mu}{k_B}\right)_L^{1/3}$ $r^*=\frac{2\sigma R T_{sat}^2}{H_{fg}\mathfrak{M}p_L\Delta T_{sat}}$

Table 2 Heat transfer correlation for forced convective boiling

Authors	Correlations	Condition
Schrock and Grossman[11]	$\frac{h_{TP}}{h_{LO'}}=0.739\left[Bo\times10^4+1.5\left(\frac{1}{X_{tt}}\right)^{2/3}\right]$	Water, p = 3 - 35 ata Upward flow
Wright[12]	$\frac{h_{TP}}{h_{LO'}}=0.67\left[Bo\times10^4+3.5\left(\frac{1}{X_{tt}}\right)^{2/3}\right]$	Water, p = 1 - 5 ata Downward flow
Pujol and Stenning[13]	$\frac{h_{TP}}{h_{LO}}=0.90\left[Bo\times10^4+4.45\left(\frac{1}{X_{tt}}\right)^{0.37}\right]$ $\frac{h_{TP}}{h_{LO}}=0.53\left[Bo\times10^4+7.75\left(\frac{1}{X_{tt}}\right)^{0.37}\right]$	R113 Upward flow R113 Downward flow
	$h_{LO}=0.023\frac{k_L}{D}\left(\frac{DG}{\mu_L}\right)^{0.8}\left(\frac{c_L\mu_L}{k_L}\right)^{0.4}$	$h_{LO'}=0.023\frac{k_L}{D}\left(\frac{DG}{\mu_L}\right)^{0.8}\left(\frac{c_L\mu_L}{k_L}\right)^{1/3}$

Table 3 Correlations for pool boiling in mixture

Authors	Correlations
Van Stralen[19]	$h=\frac{1}{2}\rho_2 H_{fg} k_1\sqrt{\tau}$ $k_1=\sqrt{\frac{12}{\pi}}\frac{\sqrt{a_L}}{\frac{\rho_2}{\rho_1}\left(\frac{1}{C_{PL}}-C\left(\frac{y_{m1}-x_{m1}}{x_{m1}}\right)\sqrt{\frac{a_L}{D}}\left(\frac{dT}{dx_{m1}}\right)_{x=C}\right)}$
Stephan and Körner[20]	$\frac{h}{h_{id}}=\frac{1}{1+A_0(0.88+0.12P[bar])(y-x)\Delta T_{id}}$
Tolubinsky and Ostrobsky[21]	$N_u=\left[\frac{q}{\rho_v H_{fg}\{x_{m1}(fd_0)_1+x_{m2}(fd_0)_2\}}\right]^{0.7}$ $\times\left\{1-\frac{(y_{m1}-x_{m1})^2}{y_{m1}(1-x_{m1})}\right\}^{1.8}P_{rL}^{-0.2}$
Calus and Rice[22]	$N_u=f_s\left\{\frac{P_e}{1+(y_{m1}-x_{m1})\sqrt{\frac{a_L}{D}}}\right\}^{0.7}\left(\frac{T_s}{T_{sw}}\right)^4 f_p^{0.7}$ $f_p=\frac{P}{\sqrt{g\sigma(\rho_L-\rho_v)}}$
Calus and Leonidopoulos[23]	$\Delta T_{sat}=(x_{m1}\Delta T_{sat,1}+x_{m2}\Delta T_{sat,2})$ $\times\left\{1-(y_{m1}-x_{m1})\sqrt{\frac{a_L}{D}}\left(\frac{C_P}{H_{fg}}\right)\left(\frac{dT}{dx_{m1}}\right)\right\}$
Happel and Stephan[24]	$\frac{h}{h_{id}}=1-k_2(y_1-x_1)^{n1}$
Stephan and Preusser[25]	$\frac{hd_0}{\lambda_L}=0.10\left(\frac{qd_0}{\lambda_L T_{sat}}\right)^{0.674}\left(\frac{\rho_v}{\rho_L}\right)^{0.156}$ $\times\left(\frac{H_{fg}d_0^2}{a_L^2}\right)^{0.371}\left(\frac{a_L^2}{\sigma d_0}\right)^{0.350}\left(\frac{\mu_L C_P}{\lambda_L}\right)^{-0.162}$ $\times\left\{1+\left\|\Sigma(y_1-x_1)\left(\frac{\partial y_1}{\partial x_1}\right)_{x_{J},P}\right\|\right\}^{-0.0733}$ $d_0=0.0146\theta\left\{\frac{2\sigma}{g(\rho_L-\rho_v)}\right\}^{0.5}$

Table 3 Correlations for pool boiling in mixture(Continue)

Author	Correlations
Thome[26]	$\frac{h}{h_{id}}=N_{sn}^{7/5}$ $N_{sn}=\left\{1-(y_1-x_1)\sqrt{\frac{a_L}{D}}\left(\frac{C_P}{H_{fg}}\right)\left(\frac{dT}{dx}\right)\right\}^{-1}$
Thome[27]	$\frac{h}{h_{id}}=\frac{1}{1+\frac{\Delta T_E}{\Delta T_{id}}}$
Jungnickel et al.[28]	$\frac{h}{h_{id}}=\frac{1}{1+k_0\frac{\rho_v}{\rho_L}\lvert y_1-x_1\rvert q^{(0.48+0.1x1)}}$
Schlünder[29]	$\frac{h}{h_{id}}=\frac{1}{1+\frac{h_{id}}{q}(T_{s2}-T_{s1})(y_1-x_1)\left\{1-\exp\left(-\frac{B_0 q}{\rho_L \beta_L H_{fg}}\right)\right\}}$
Bier et al.[30]	$\frac{h}{h_{id}}=1-\left\{A+B\left(\frac{P}{P_c}\right)+\frac{C}{1-\left(\frac{P}{P_c}\right)}\right\}(y_1-x_1)^m$
Ünal[31]	$\frac{h}{h_{id}}=\frac{1}{C_{un}}$ $C_{un}=\{1+(b_2+b_3)(1+b_4)\}(1+b_5)$ $b_2=(1-X_1)\ln\left(\frac{1.01-X_1}{1.01-Y_1}\right)+X_1\ln\left(\frac{X_1}{Y_1}\right)+\lvert Y_1-X_1\rvert^{1.5}$ $b_3=0$ for $X_1\geq 0.01$ $b_3=(Y_1/X_1)^{0.1}-1$ for $X_1<0.01$ $b_4=152(p/p_{cmvc})^{3.9}$ $b_5=0.92\lvert Y_1-X_1\rvert^{0.001}(p/p_{cmvc})^{0.66}$ $X_1/Y_1=1$ for $X_1=Y_1=0$
Inoue and Monde[32]	$\frac{h}{h_{id}}=\frac{1}{1+k\frac{\Delta T_E}{\Delta T_{id}}}$, $k=0.45\times 10^{-5}q+0.25$ $10^4\leq q[W/m^2]\leq 10^5$
Fujita and Tsutsui[33]	$\frac{h}{h_{id}}=\frac{1}{1+k\frac{\Delta T_E}{\Delta T_{id}}}$, $k=1-0.8\exp(-10^{-5}q)$
Inoue et al.[57]	$\frac{h}{h_{id}}=\frac{1}{1+k\frac{\Delta T_E}{\Delta T_{id}}}$, k = 1 - 0.75 exp[-0.75x10-5 q] $10^4\leq q[W/m^2]\leq 10^5$

Table 4 Some experimental studis of pool boiling heat transfer in mixtures

Authors	Heated Surface	Test Substances	Pressures(MPa)
Bonilla et al. [34]	Horizontal Plate	Acetone+Ethanol Ethanol+Butanol Acetone+Water Ethanol+Water Isobutanol+Water	0.1
Van Wijk et al. [35]	Horizontal Wire	Ethanol+Toluene Isopropanol+Cyclo Hexane Acetone+Water Butanol+Water Pentanol+Water	0.1
Van Stralen[36]		Acetone+Butanol	
Sternling et al. [37]		Ethanol+Benzene Isopropanol+Ethylene Glycol Isopropanol+Cyclo Hexane Methanol+Ethylene Glycol	
Grigor'ev et al. [38]	Horizontal Tube	Ethanol+Butanol Acetone+Butanol Ethanol+Water	0.1
Van Stralen [19]	Horizontal Wire	Methyl-Ethyl-Ketone+Water Methanol+Water	0.1
Tolubinsky et al. [39]	Horizontal Tube	Ethanol+Benzene Ethanol+Water	0.1
Körner [40]		Acetone+Butanol n-Heptane+Methyl-Cyclo-Hexane Methyl Ethyl Ketone+Toluene Acetone+Water Propanol+Water	
Stephan et al. [20]		Acetone+Butanol Heptane+Methyl-Cyclo-Hexane Methanol+Benzol	0.1-1.0
Tolubinsky et al. [21]	Vertical Tube	Ethanol+Water Methanol+Water Ethanol+Butanol Ethanol+Benzene	0.1

Table 4 Some experimental studis of pool boiling heat transfer in mixtures(Continued)

Authors	Heated Surface	Test Substances	Pressures(MPa)
Stephan et al. [41]	Horizontal Tube	Ethanol+Cyclo-Hexane Methanol+Benzol	0.1-0.3
Calus et al. [23]	Horizontal Wire	Isopropanol+Water Acetone+Water Ethylene-Glycol+Water	0.1
Calus et al.	Horizontal Wire	Propanol+Water	0.1
Happel et al. [24]	Horizontal Tube	Benzene+Toluene Ethanol+Benzene Isobutanol+Water	0.05-0.2
Stephan et al. [25]	Horizontal Tube	Acetone+Methanol Acetone+Water	0.1
Jungnickel et al. [28]	Horizontal Plate	R22+R12 R12+R13	0.1-2.0
Bier et al. [30]	Horizontal	TubeR12+SF6 R22+SF6 R13B1+SF6	0.56-0.97 (Reduced Pressure)
Fujita et al. [33]	Horizontal Plate	Methanol+Water Ethanol+Water Methanol+Ethanol Ethanol+n-Butanol Methanol+Benzene	0.1

Table 5 Some recent Experimental studies of heat transfer for mixtures in flow boiling

Authors	year	mixture			
Bennett and Chen[15]	1980	water/ethylene glycol	t	v	e
Toral et al.[44]	1982	ethanol/cychlohexan	t	v	e
Nakanishi et al.[45]	1986	R12/R13	t	h	e
Ross et al.[46]	1987	R152a/R13B1	t	h	e
Hihara et al.[46]	1989	R12/R22 R114/R22	t	h	e
Jung et al.[47]	1989	R22/R113	t	h	e
Jung et al.[48]	1989	R12/R152a	t	h	e
Jung et al.[49]	1989	refrigerants	t	h	e
Murata and Hashizume[50]	1990	R11/R114	t	h	e
Takamatsu et al.[51]	1990	R22/R114	t*	h	w
Fujita et al. [52]	1991	R11/R113	t	v	e
Yoshida et al.[18]	1991	R22/R114	t	h	e
Sami and Duong[53]	1992	R22/R114	a*	h	w
Sami and Schnotale[54]	1992	R22/R114 R22/R152a	t*	h	w
Celata et al.[42]	1993	R12/R114	t	v	e
Takamatsu et al.[55]	1993	R22/R114	t	h	w

t: tube, a: annulus, *:enhanced surface
h: horizontal, v: vertical
e: electrically heated, w: water heated

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