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EVALUATION OF A METHOD FOR PREDICTING HEAT TRANSFER DURING BOILING OF MIXTURES IN PLAIN TUBES

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ABSTRACT

New refrigerants with low GWP and low ODP are being formulated to replace present refrigerants. Heat transfer coefficients of mixtures are lower than of pure fluids due to sensible heat transfer and mass transfer effects. No well-verified method for predicting heat transfer during boiling of mixtures inside tubes is available. The proposed method tested during the present research consists of modifying the predictions of pure fluid correlations by applying a well-known correction factor for pool boiling in the nucleate boiling region and applying a correction factor used for condensation heat transfer to the region in which nucleate boiling has been completely suppressed. For comparison, calculations were also done using only the pool boiling correction factor. These correction factors were used together with five well-known correlations for pure fluids. These were compared to a data base consisting of 695 data points for 42 binary and ternary mixtures of 19 fluids from 18 independent studies. The fluids included halocarbons, hydrocarbons, and carbon dioxide. The range of parameters covered by the data is: tube diameters 2.0 to 14 mm, horizontal and vertical orientations, flow rates from 50 to 930 kg/m^{2}s, reduced pressures from 0.05 to 0.63, temperature glides upto 28 ^{o}C, and boiling numbers from 0.4 x 10^{-4} to 46.2 x 10^{-4}. The proposed method performed better than using the nucleate boiling correction only. Three of the five general correlations tested showed mean deviation of about 19% with all data using the proposed method. Thus the proposed method appears to be generally applicable.

KEY WORDS: Boiling and evaporation, Two-phase/Multiphase flow, Mixtures, Tubes, Heat transfer

1. INTRODUCTION

Due to concerns about ozone layer depletion and global warming, many refrigerants have been phased out and many others are to be phased out once suitable benign substitutes are identified. Therefore continuous efforts are being made to develop new refrigerants with low GWP (Global Warming Potential), low ODP (Ozone Depletion Potential), minimum flammability and toxicity, compatibility with materials, and good COP (Coefficient of Perforformance). Most of the efforts for developing new refrigerants are directed towards formulation of mixtures of two or more fluids to give the desired characteristics. To evaluate the suitability of the proposed new refrigerants, their heat transfer performance has to be measured or predicted. As is well-known, heat transfer coefficients of non-azeotropic mixtures are lower than those of the components due to resistances caused by sensible heat transfer and mass transfer and therefore cannot be calculated by using mixture mean properties in formulas for pure fluids. To perform measurements on each proposed mixture is a very expensive and time consuming task. Therefore a method to predict the heat transfer during boiling is needed. While there have been many experimental

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studies on boiling of mixtures in tubes, no well-verified method for predicting heat transfer of mixtures with significant glide is available. The present research effort was made to try to fulfil this need. A method is proposed in which predictions of pure fluid correlations are modified by published correction factors for pool boiling and condensation heat transfer of mixtures. This method has been validated by comparison with an extensive database consisting of 695 data points for 42 binary and ternary mixtures from 18 independent studies. The range of parameters covered by the data is: tube diameters 2 to 14 mm, horizontal and vertical orientations, flow rates from 50 to 930 kg/m^{2}s, reduced pressures from 0.05 to 0.63, temperature glides upto 28 ^{o}C, and boiling numbers from 0.4 x 10^{-4} to 46.2 x 10^{-4}. So that the results may be viewed in perspective, the entire database was also compared with a method proposed by Thome [1] in which only the nucleate boiling term of pure fluid correlations is modified by a pool boiling correction factor.

In the following, the proposed method is presented and its physical bases are described. The results of data analyses are presented in tabular and graphical forms.

2. PREVIOUS WORK

Many experimental studies have been done on boiling of mixtures in tubes. The review paper by Thome [1] lists 17 studies on boiling of mixtures in plain tubes. Many more studies were identified during the present research. Many of these papers do not give data in an analyzable form.

Numerous correlations, empirical and analytical, have been proposed but none of them has been verified with more than a few data sets. Many of them involve constants which are designed to fit a particular data set. Hence their general applicability is unknown.

Most of the proposed predictive methods modify the nucleate boiling contribution to heat transfer in correlations for single-component fluids. Examples are Palen and Small [2] and Kandlikar [3]. Thome [1] had also suggested the same approach, using the Thome & Shakir [4] correlation for pool boiling of mixtures for this purpose. There have been a few comparisons of this method with data, some showing good agreement and some showing poor agreement. For example Grauso et al. [5] found this method unsatisfactory while Zurcher et al. [6] showed that this method worked well for their R-407C data. Barbosa and Hewitt [7] pointed out that there is also a mass transfer resistance at the liquid-vapor interface in flow without nucleation due to the difference in composition between the liquid and the bulk vapor. They note that this resistance could be accounted for either by the simplified methods such as that of Bell and Ghaly [8] for condensation or by the more rigorous methods such as that of Colburn and Drew [9]. They themselves followed the latter approach and showed reasonable agreement with data from one source, with nucleate boiling correction by the Palen and Small [2] method. Jung et al. [10] have proposed an empirical correction factor for this mass transfer resistance based on their own data.

The correlation of Kandlikar [3] was verified with data from five sources but it is applicable only to binary mixtures. Many of the refrigerants presently being used, considered, or developed, have three or more components. Further, it involves experimentally determined fluid specific factors which are given only for a few fluids. The Kandlikar method does not include correction for mass transfer resistance at interface during evaporation with suppressed nucleation.

Thus at the present, there is no well-verified method for the prediction of heat transfer during boiling of mixtures in tubes and there is a need for one.

3. BASIS OF THE PROPOSED CALCULATION METHOD

The calculation method proposed here is to modify the predictions of correlations for pure fluids by the Thome-Shakir [4] correction factor in the nucleate boiling region and the Bell-Ghaly [8] method in the convective boiling region. This methodology is based on the physical phenomena that occur in the process. To explain the basis of this method, the physical phenomena are first discussed and then the Thome-Shakir and Bell-Ghaly methods are described.

3.1 Physical Phenomena

As the mixture gets heated during its passage through the tube, the low boiling point component (called the light component) evaporates faster than the higher boiling point component (called the heavy component). This changes the composition of the liquid and vapor phases and raises the bubble point. This affects nucleate boiling as well as convective boiling, i.e. evaporation without bubble nucleation.

Considering first the nucleate boiling region, the liquid at the interface with a growing bubble is richer in the heavier component and hence its bubble point is higher than that of the bulk liquid. For bubble growth to continue, the lighter component from the bulk liquid has now to diffuse through this heavy component rich layer. This mass transfer resistance causes a decrease in the heat transfer coefficient in the bubble nucleation region.

Consider now the condition in which there is no nucleation. As the mixture flows along the heated tube, the bubble point temperature increases due to faster evaporation of the lighter component. Hence heat has to be supplied for sensible heating of the liquid and vapor phases in addition to the heat for evaporation. Further, the concentrations of the components at the interface are different from those in the bulk vapor and liquid; mass transfer resistance has to be overcome in both liquid and vapor. These two effects lower the heat transfer coefficient compared to an equivalent pure fluid.

3.2 The Thome-Shakir Correction Factor for Nucleate Boiling

To take into account the effect of mass transfer resistance on bubble nucleation described above, Thome and Shakir [4] analytically derived the following formula for pool boiling of mixtures:

$$\frac{h}{h_I} = \left[1 + \left(\frac{h_I}{\dot{q}}\right)(T_{dew} - T_{bub})\left\{1 - \exp\left(\frac{-B\dot{q}}{\rho_f h_{lg} \beta_f}\right)\right\}\right]^{-1} \qquad (1)$$

Where h_I is the ideal heat transfer coefficient calculated by a pool boiling correlation for pure fluids using mixture properties, B is the scaling factor assumed to be 1 (all heat transferred to bubble interface is converted to latent heat), and β_f is the liquid phase mass transfer coefficient which is recommended to be constant at 0.0003 m/s. While Thome and Shakir verified this formula only with the data for a few aqueous mixtures, Thome [1] reports that it has had extensive validation with data for a wide variety of mixtures. This equation is applicable to mixtures with any number of components.

3.3 The Bell & Ghaly Equilibrium Method

As noted earlier, heat transfer during convective boiling of mixtures involves sensible heating of mixture and mass transfer resistance in the liquid and vapor. Condensation of mixtures involves sensible cooling of mixture and mass transfer resistance. Thus the phenomena involved during condensation and boiling without nucleation are very similar and methods used for condensation are applicable to boiling as noted by Barbosa and Hewitt [7]. The rigorous method for mixtures is the film theory of Colburn and Drew [9] but it is very cumbersome. Bell and Ghaly [8] noted that mass transfer is very roughly proportional to heat transfer and heat transfer is much easier to calculate than mass transfer. They assumed equilibrium between the phases, neglected mass transfer resistance, and derived the following equation:

$$\frac{1}{h_{mix}} = \frac{1}{h_c} + \frac{Y}{h_{GS}} \tag{2}$$

Where:

$$Y = xC_{pg}\frac{dT_{DEW}}{dh} \tag{3}$$

h_{mix} is the heat transfer coefficient of the mixture and h_c is the heat transfer coefficient for condensation of an equivalent pure fluid with the properties of the mixture, and h in Eq. (3) is enthalpy. The single phase heat transfer coefficient h_{GS} is to be calculated conservatively so that the error in omitting the mass transfer resistance is compensated for. Thus the Bell and Ghaly method omits mass transfer resistance and compensates for it by over-estimating the sensible heat transfer resistance. Shah et al. [11] evaluated this method by comparison with a wide ranging database for mixtures condensing in tubes and found it to give good agreement with data; the ideal fluid condensing heat transfer coefficient was calculated with the Shah correlation [12].

To apply this method to boiling, the condensing heat transfer coefficient h_c in Eq. (2) is to be replaced by the heat transfer coefficient during convective boiling.

4. DETAILS OF THE PROPOSED CALCULATION METHOD

The proposed method consists of using a general correlation for pure fluids and applying the Thome-Shakir correction factor to the nucleate boiling contribution and the Bell-Ghaly correction factor to the convective boiling contribution.

Many correlations for heat transfer to pure fluids are of the form:

$$h_{TP} = h_{nb} + h_{cb} \tag{4}$$

The proposed method is expressed as:

$$h_{TP,mix} = F_{TS}h_{nb} + \left(\frac{1}{h_{cb}} + \frac{Y}{h_{GS}}\right)^{-1} \tag{5}$$

F_{TS} is the Thome-Shakir correction factor which, in accordance with Eq. (1), is:

$$F_{TS} = \left[1 + \left(\frac{\mathrm{h_I}}{\mathrm{q}}\right)(\mathrm{T_{dew}} - \mathrm{T_{bub}})\left\{1 - \exp\left(\frac{-\mathrm{B}\dot{\mathrm{q}}}{\mathrm{\rho_f h_{lg} \beta_f}}\right)\right\}\right]^{-1} \tag{6}$$

For calculation of h_I, a reliable pool boiling correlation is to be used. The following correlation of Cooper [13] is recommended:

$$h_I = 55p_r^{0.12}(-logp_r)^{-0.55}M^{-0.5}\dot{q}^{2/3} \tag{7}$$

This is to be used for all plain tubes of any material including copper. Eq. (7) is dimensional; the units are: h_l in W/m^2K and $\dot{q}$ in W/m^2.

The single phase heat transfer coefficient h_{GS} is calculated conservatively by assuming the vapor phase to be flowing alone in the tube for which the following equation is used:

$$h_{GS} = 0.023\left(\frac{\dot{m}xD}{\mu_g}\right)^{0.8}\frac{\Pr_g^{0.4}k_g}{D} \tag{8}$$

5. EVALUATION OF PROPOSED METHOD

5.1 Pure Fluid Correlations

The proposed method was evaluated by applying it to a number of well-verified general correlations for pure fluids. Some correlations use boiling number for determining nucleate boiling contribution while some directly use pool boiling formulas. The correlations of Shah [14, 15] and Gungor & Winterton [16] use boiling number; these were found to be the most accurate among all correlations by Shah [17] on comparison with a very extensive and varied database. The most successful among those using pool boiling correlations was Liu & Winterton [19]. The pioneering correlation of Chen [18] with Cooper pool boiling correlation was found to perform fairly well by Shah [17]. It was therefore decided to include it together with the other four mentioned in the present evaluation. Another well-verified correlation is that of Gungor & Winterton [20] which also uses Cooper pool boiling correlation.

The correlations of Chen, as well as Gungor & Winterton [16, 20] are of the form of Eq. (4). The calculation procedure for them has already been described. The procedure for the other two is described below.

For the Liu & Winterton correlation, the following is the equation for mixtures using the proposed method:

$$\left(h_{TP,mix}\right)^2 = (F_{TS}h_{nb})^2 + \left(\frac{1}{h_{cb}} + \frac{Y}{h_{GS}}\right)^{-2} \tag{9}$$

The Shah correlation has three regimes: nucleate boiling, convective boiling, and bubble suppression in which both nucleate and convective boiling have influence. The equations for mixture according to the present method are:

$$h_{nb,mix} = h_{LS}F_{TS}(230Bn^{0.5}) \tag{10}$$

$$h_{cb,mix} = \left(\frac{J^{0.8}}{1.8h_{LS}} + \frac{Y}{h_{GS}}\right)^{-1} \tag{11}$$

$$h_{bs,mix} = \left[(h_{bs} - h_{nb,mix})^{-1} + \frac{Y}{h_{GS}}\right]^{-1} \tag{12}$$

Heat transfer coefficient of the mixture is the largest of those given by the above three equations. The equations defining the terms used above are the following.

$$h_{bs} = E h_{nb,mix} \exp\,(aJ^{n}) \tag{13}$$

For J > 0.1, a = 2.74 and n = -0.1. For J $\leq$ 0.1, a = 2.47 and n = -0.15. E = 0.064 for Bn $\geq$ 0.0011, E = 0.064 for Bn < 0.0011.

J = Co for vertical tubes, and for horizontal tubes when $Fr_l > 0.04$. When $Fr_l < 0.04$ for horizontal tubes,

$$J = 0.38 Fr_l^{-0.3} Co \tag{14}$$

$$h_{LS} = 0.023\left(\frac{\dot{m}(1-x)D}{\mu_l}\right)^{0.8} \frac{\mathrm{Pr}_l^{0.4}\, k_l}{D} \tag{15}$$

5.2 Calculation Details

The data collected were compared to the general correlations listed above with the proposed method. The data base was also compared to the method suggested by Thome [1] in which effect of mixture is taken into account only by modifying the nucleate boiling contribution with the Thome-Shakir correction factor; there is no modification to the convective boiling term. Thus for correlations of the form of Eq. (4), the Thome method gives:

$$h_{TP,mix} = F_{TS} h_{nb} + h_{cb} \tag{16}$$

For the Shah correlation, there is no change to the convective boiling term and hence $h_{TP,mix}$ is the highest of Equations 10, 13, and the following equation:

$$h_{TP} = 1.8/J \tag{17}$$

Composition of vapor and liquid was calculated at each value of vapor quality. Dew point, bubble point, and properties of vapor and liquid were calculated using the calculated compositions and temperatures. All mixture properties were calculated with REFPROP version 9.1, Lemmon et al. [21].

5.3 Data Collection

Literature search was done to identify publications giving data for boiling of mixtures. For refrigerants, only oil-free data were accepted as oil can have a profound effect on heat transfer. Data for tubes of diameter smaller than 2 mm were not considered as heat transfer behavior in mini-tubes has been reported to be different from that in larger tubes. Table 1 lists the features and parameters of the data collected. As seen in it, there are 695 data points from 42 data sets from 18 sources. Data for each fluid and each tube diameter from a source is considered a distinct data set. These data are for 41 distinct mixtures of 18 fluids. Note that the listed glides are prior to boiling; these change with composition during boiling.

Table 1 Salient features of the data analyzed and the results of comparison with various correlations using only the Thome-Shakir correction factor (designated TS) as well as using Tome-Shakir and Bell-Ghaly correction factors together (designated TSBG).

Source & (uncertainty in measured h_{TP}, %)	Dia. mm	Fluid (Note 1)	Glide °C	p_r	$\dot{m}$ Kg/m^{2}s	Bn x 10^4	No. of Data	Deviation, Percent, for Listed Correlations (Mean Absolute Above, Average Below)									
								Shah with TS	Shah with TSBG	LW with TS	LW with TSBG	Chen with TS	Chen with TSBG	GW-86 with TS	GW-86 with TSBG	GW-87 with TS	GW-87 with TSBG
Shin et al. [36] (5.8-7.3)	7.7	Propane – isobutane 75-25	5.4	0.133	424	2.0	7	6.2 6.2	6.4 -3.4	24.32 4.3	17.8 15.6	29.8 29.8	20.0 20.0	34.7 34.7	27.0 24.1	18.8 18.8	18.3 8.8
		Propane – Isobutane 25-75	6.4	0.090	424	2.0	7	15.5 15.5	6.3 6.3	37.2 37.2	27.3 27.3	34.3 34.3	25.4 25.4	36.8 36.8	27.5 27.5	18.8 18.8	17.0 17.0
		Propane – Isobutane 50-50	7.5	0.111	424	2.0	6	8.2 8.2	4.7 -2.2	30.53 0.5	20.0 19.5	31.9 31.9	21.2 21.2	38.5 38.5	28.9 26.6	17.3 16.2	17.9 11.6
		R32-R134a 75-25	3.0	0.191	583	1.8 1.9	6	5.0 3.4	4.8 -2.1	13.8 13.8	13.3 9.7	28.2 28.2	21.6 21.6	33.3 33.3	26.0 26.0	16.1 16.1	16.0 13.6
		R32-R134a 50-50	5.4	0.176	583	2.0	6	5.1 4.0	4.3 -4.3	17.41 7.4	12.7 10.6	29.3 29.3	20.0 20.0	35.6 35.6	25.6 25.6	16.5 16.3	15.3 12.5
		R32-R134a 25-75	5.9	0.152	583	2.2 2.4	7	0.9 -0.2	7.6 -7.5	13.6 13.6	11.9 6.6	20.9 20.9	13.2 12.8	26.1 26.1	21.2 16.7	14.0 9.3	15.0 5.7
Jung et al. [10] (5-10)	9.0	R12-R152a 88-12 Mole	1.0	0.08	530	1.1 5.2	41	10.6 -3.3	11.3 -4.1	13.3 -1.2	13.5 -2.2	12.0 2.2	12.1 1.1	19.7 12.7	19.2 11.2	14.5 4.1	14.6 3.6
		R12-R152a 21-79 Mole	2.4	0.08	251 724	0.8 4.5	77	9.0 -6.7	12.3 -10.1	12.9 3.2	13.6 -0.3	11.3 6.2	11.8 2.6	18.8 9.9	18.6 5.9	14.8 -0.3	15.6 -1.9
Takamatsu et al. [37] (15-25)	7.9	R22-R114 51-49 Mole	18.1	0.132	299	3.0 8.9	14	16.4 -16.4	32.8 -32.8	6.4 -2.0	18.5 -18.5	13.6 13.6	17.1 -5.1	32.5 26.5	25.9 -1.9	14.4 -7.4	22.7 -16.7
Jung et al. [29] (5-10)	9.0	R22-R114 89-11 mole	5.9	0.080	510	2.6	3	31.9 31.9	8.5 8.5	29.3 29.3	11.7 11.5	39.2 39.2	18.6 18.6	35.9 35.9	15.3 15.3	24.8 24.8	14.4 13.9
		R22-R114 78-22 Mole	11.1	0.08	355 510	2.5 2.8	6	11.6 11.6	16.9 -16.9	11.1 11.1	14.1 -14.1	18.9 18.9	11.2 -6.1	16.3 16.3	14.0 -9.1	6.6 3.4	13.2 -10.1
		R22-R114 23-77 Mole	17.4	0.08	355 510	3.3 3.7	6	3.9 3.4	15.0 -15.0	13.4 13.4	6.4 -4.1	13.5 13.5	7.7 -2.6	15.9 15.9	8.4 -2.7	7.1 3.5	7.9 -4.8
		R22-R114 47-	19.2	0.080	503	0.8	30	20.0	8.6	35.8	15.4	29.7	13.4	33.9	21.1	18.1	16.7

		53 Mole			720	4.6		18.6	-6.5	35.8	8.0	29.7	6.5	33.9	8.3	17.7	6.4
Kattan et al. [30] (0-7.5)	12.0	R402A	1.3	0.167	102 317	1.8 9.7	25	13.0 -10.8	13.3 -11.5	18.9 18.6	18.0 17.6	38.5 38.5	36.9 36.9	41.8 41.8	38.8 38.8	12.1 2.9	12.1 2.2
Greco [27] (3.5-12.5)	6.0	R404A	0.5	0.125 0.212	483	1.4 3.2	11	21.8 -0.8	21.9 -1.9	25.8 15.7	25.2 14.6	31.4 25.0	30.5 23.8	30.9 24.7	30.2 23.4	23.2 23.2	23.2 6.3
		R417A	3.3 4.1	0.094 0.238	200 360	2.4 3.2	39	34.6 9.4	31.1 1.4	37.1 23.3	31.8 15.3	51.3 48.6	44.5 40.4	50.5 47.2	43.7 38.3	31.4 12.5	29.6 8.1
		R407C	5.6 6.4	0.076 0.216	200 360	0.9 2.3	59	35.6 25.5	27.5 11.9	41.7 38.3	32.1 26.7	68.7 68.5	56.2 55.9	64.4 64.4	53.3 52.5	31.5 24.3	28.3 18.5
Rabah & Kabelac [34] (Not Stated)	10.0	Propane-R134a 65-35 Mole	3.4	0.15	100 300	1.9 5.8	3	23.6 -23.6	23.6 -23.6	10.5 -10.5	10.0 -10.0	10.6 -2.5	10.8 3.5	10.9 3.4	10.8 3.5	16.3 -16.3	16.4 -16.4
		Propane-R134a 30-70 Mole	11.3	0.049	100 300	2.0 6.1	3	5.7 -2.2	12.5 -12.5	26.4 26.4	11.6 -11.6	32.7 32.7	22.4 22.4	44.0 44.0	26.8 26.8	7.5 7.5	3.1 3.1
		Propane-R134a 12-88 Mole	12.5	0.048	100 300	2.3 6.9	3	7.4 6.4	9.3 1.9	34.8 34.8	24.4 24.4	45.4 45.4	38.2 38.2	60.4 60.4	46.4 46.4	20.1 20.1	17.1 17.1
Zhang et al. [26] (0-7)	6.0	R32-R134a 30-70	5.6	0.199	250	1.8	5	9.1 -6.4	12.1 -12.1	16.0 16.0	10.5 10.5	49.5 49.5	42.3 42.3	55.3 55.3	47.0 47.0	8.1 8.1	5.6 5.5
		R32-R125-R134a 23-25-52	5.6	0.224	250	2.1	6	12.7 -12.7	19.7 -19.7	7.0 4.3	5.3 -1.6	34.6 34.6	26.3 -26.3	36.7 36.7	28.6 27.4	6.1 -3.9	8.7 -7.4
Grauso et al. [5] (0-10)	6.0	CO2-Propane 70-30	17.3 18.4	0.52 0.58	200 350	1.6 5.4	47	20.1 3.8	29.0 -29.0	17.9 14.2	16.7 -15.4	54.4 54.2	13.9 9.5	80.7 80.7	22.4 17.3	15.8 12.1	15.0 -4.8
		CO2-Propane 83-17	6.6 7.4	0.55 0.63	200 350	1.5 2.8	26	24.7 -19.5	40.3 -40.3	14.7 -1.7	19.7 -18.3	31.7 31.7	15.1 4.2	35.2 35.2	16.4 3.7	17.1 -13.5	24.9 -24.9
Celata et al. [24] Note 2 (Not Stated)	7.57	R12-R114 77-23 mole	7.1	0.049	500 930	1.9 4.7	8	26.2 -23.2	27.9 -27.9	23.7 -20.4	26.7 -26.7	24.8 -20.8	26.0 -26.0	13.8 -0.2	12.9 -9.4	16.3 -13.0	16.0 -15.5
		R12-R114 59-41 mole	9.4	0.051	500 930	2.0 5.0	8	27.1 24.7	29.9 -29.9	24.6 -19.5	27.1 -27.1	25.8 -21.2	27.1 -27.1	18.6 1.4	16.1 -9.7	17.4 -13.3	16.1 -16.1
		R12-R114 37-63 mole	9.1	0.054	500 930	2.0 5.0	6	24.6 -24.6	29.9 -29.9	24.2 -20.6	27.9 -27.9	23.8 -21.8	27.4 -27.4	21.4 -0.1	18.9 -10.7	16.7 -14.1	16.7 -16.7
Li et al. [31] (5.2-17.2)	2.0	R1234yf-R32, 81-19	8.6	0.202	100 404	1.4 6.4	33	18.6 -14.2	24.8 -24.8	11.1 5.6	12.0 -3.5	50.9 50.9	38.4 38.4	53.7 53.7	39.2 38.3	14.5 8.1	18.3 -14.1
		R1234yf-R32, 50-50	2.7	0.21	100 400	1.2 5.1	15	21.7 -21.5	26.5 -26.5	13.2 -2.4	16.2 16.0	30.3 30.3	24.2 24.1	33.6 33.1	27.4 25.7	16.3 -15.5	18.5 -18.0
Choi et al. [25]	7.75	R407C	6.2 5.7	0.1120 .199	427 477	1.1 2.7	21	15.8 9.8	13.6 -1.5	27.5 24.5	21.5 14.6	35.8 34.4	27.2 23.6	39.2 36.7	32.7 24.8	24.1 15.9	24.2 10.6

(Not Stated)		R32-R134a 50-50	5.7	0.118	427	1.8	2	9.0 9.0	3.7 -1.3	20.72 0.7	14.2 12.5	35.3 35.3	25.7 25.7	36.7 36.7	26.6 26.6	16.7 16.7	14.5 12.9
		R32-R134a 75-25	3.2	0.129	427	1.6	2	13.6 13.6	6.0 6.0	22.2 22.2	17.0 17.0	42.6 42.6	35.3 35.3	43.2 43.2	35.7 35.7	23.6 23.6	20.7 20.7
		R32-R134a 25-75	6.2	0.101	427	2.0	2	9.5 -6.7	15.0 -15.0	6.5 6.5	0.9 -0.9	14.1 14.1	6.7 6.7	15.6 15.6	7.5 7.5	0.3 -0.3	3.4 -3.4
Wetterman & Steiner [28] (Not Stated)	14.0	R116-R134a 80-20 Mole	5.0 8.5	0.327 0.634	240 400	0.45 5.5	6	60.6 -33.3	53.7 -42.0	66.5 -9.3	61.5 -19.8	67.9 40.5	47.8 -18.6	121.6 99.6	47.5 -16.3	56.5 -8.4	53.1 -12.3
		R116-R134a 40-60 Mole	17.9 27.7	0.254 0.621	240 400	0.39 46.2	6	31.1 -24.1	44.4 -44.4	20.7 -4.0	30.8 -25.7	17.3 13.0	13.8 -13.3	210 210	15.4 -12.0	17.9 4.9	18.3 -5.4
		R116-R134a 10-90 Mole	10.1 18.0	0.253 0.620	80 400	0.3 31.4	6	20.4 -10.7	19.8 -15.9	36.4 14.2	30.6 6.7	41.8 29.8	32.3 19.5	75.2 75.2	29.3 22.9	23.9 13.3	22.6 9.6
Murata & Hashizume [32] (Not Stated)	10.3	R11- R114 25-75	2.4	0.054	100 300	2.4 22.1	10	20.3 -14.2	22.9 -7.1	29.3 -26.1	32.1 -30.9	12.4 3.5	13.1 0.1	22.4 6.6	20.3 -3.1	18.6 -7.7	20.0 8.6
		R11-R114 50-50	3.7	0.050	100 300	4.6 20.7	24	15.3 -14.5	22.9 -22.9	21.12 0.5	25.8 -25.8	11.3 3.1	11.7 -1.7	19.7 5.8	17.6 -4.8	14.9 -9.8	16.8 -11.9
Murata & Hashizume [33] (Not Stated)	10.3	R123-R134a 90-10 Mole	15.0	0.051	100 300	2.0 17.8	28	11.3 -2.6	15.0 -8.9	19.0 5.5	16.0 -3.8	28.4 25.6	25.0 18.7	28.2 20.1	21.1 7.1	18.2 1.4	19.1 -2.4
Ross et al. [35] (0-10)	9.0	R13B1-R152a 75-25	8.9	0.126	510	6.6 7.1	8	59.4 59.4	38.7 38.7	63.6 63.6	43.9 43.9	81.6 81.6	53.5 53.5	172. 172.	98.5 98.5	100.7 100.7	88.0 88.0
		R13B1-R152a 18-82	9.5	0.100	334 354	4.7 12.5	23	24.2 24.0	22.9 22.2	44.3 44.3	38.7 38.7	48.8 48.8	41.1 41.1	106. 106.	76.2 76.2	56.8 56.8	53.8 53.8
Wattalet et al. [38] (3-12.5)	10.9	R22-R124-R152a 52-33-15	4.9	0.092	50 300	1.0 5.0	44	11.2 2.4	11.5 5.0	9.9 -3.4	44.9 44.8	36.0 34.0	13.0 9.8	10.8 0.7	8.6 0.0	7.7 -2.5	10.0 -7.9
All data	2.0 14.0		0.5 27.7	0.0480 0.634	50 930	0.4 46.2	695	18.8 1.4	19.7 -8.8	23.41 2.8	20.5 12.9	36.5 32.3	27.9 20.4	44.3 40.5	30.5 21.7	20.3 8.3	20.3 3.0

Note 1: Compositions listed are mass percent except where noted as mole. Note 2: Vertical tube. All others were horizontal.

5.4 Results of Calculations

Results of calculations are listed in Table 1. Mean absolute deviation is defined as:

$$\delta_m = \frac{1}{N}\sum_1^N ABS\left((h_{predicted} - h_{measured}) / h_{measured}\right) \tag{18}$$

verage deviation is defined as:

$$\delta_{avg} = \frac{1}{N}\sum_1^N \left((h_{predicted} - h_{measured}) / h_{measured}\right) \tag{19}$$

For each of the tested general correlations, mean absolute and average deviations are listed for the following cases:

1. When only the Thome-Shakir correction factor is applied (marked TS). This is the method proposed by Thome [1].
2. When both Thome-Shakir and Bell-Ghaly correction factors are applied (marked TSBG). This is the present proposed method.

The names of the tested correlations have been abbreviated in Table 1. GW-86 is Gungor & Winterton [20], GW-87 is Gungor & Winterton [16], LW is Liu & Winterton [19].

6. DISCUSSION

6.1 Comparison of Correction Methods

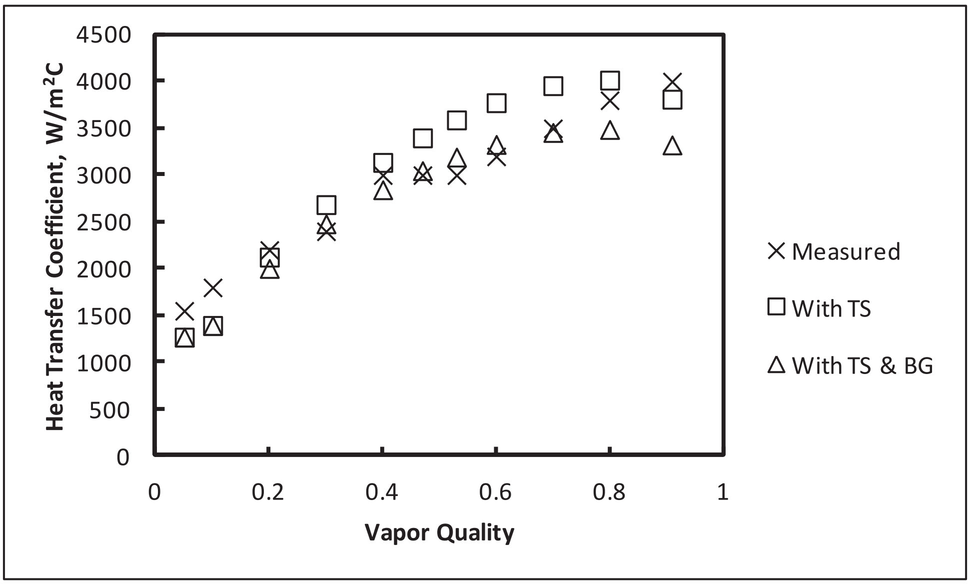

Fig. 1 Comparison of measured heat transfer coefficients with predictions of Shah correlation with Thome – Shakir correction factor alone (TS) and with both Thome Shakir & Bell-Ghaly Correction factors (TS & BG). Data of Wattalet et al for the mixture R22-R-124-R152a 52-33-15 percent by mass

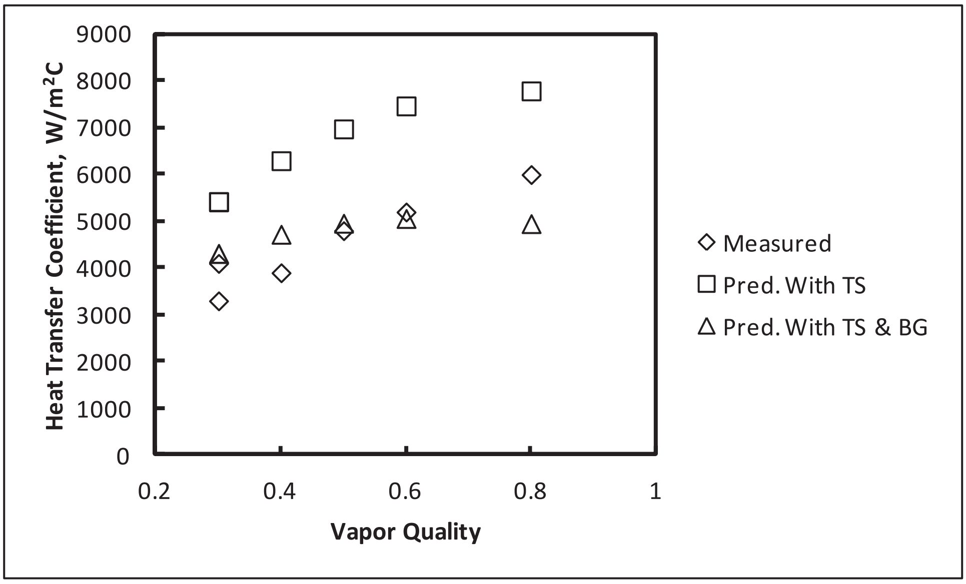

Fig. 2 Comparison of measured heat transfer coefficients with predictions of Shah correlation with Thome–Shakir correction factor alone (TS) and with both Thome-Shakir & Bell-Ghaly correction factors (TS & BG). Data of Park et al. for mixture of R123-R134a-R22 19-62-19 Mole percent. Glide = 16.8 °C.

Figs. 1 and 2 show the comparison of data with the Shah correlation using the correction according to the present method as well as when only the nucleate boiling correction is applied. The performance of the proposed method is seen to be better.

Comparison with all data sets is given in Table 1. For the correlations of Gungor-Winterton [20], Chen, and Liu & Winterton, it is seen that the proposed method gives significantly better agreement with data compared to the method proposed by Thome [1]. For the correlation of Gungor & Winterton [16], the two methods give the same mean absolute deviation though the average deviation is significantly less for the present method. With the Shah correlation, the Thome method performs a little better than the present method (mean deviations 18.8 and 19.7 percent respectively). Thus four out of the five general correlations perform better with the present proposed correction method. It is also more consistent with the physical phenomena involved. Hence the proposed correction method appears to be the better choice.

It may be mentioned several authors have given modified forms of the Bell and Ghaly method, for example McNaught [22]. Shah et al. [11] had found it to perform a little better than the Bell and Ghaly method at low flow rates. Hence while the results obtained with the Bell-Ghaly method are satisfactory, evaluation of the alternative forms may be worth wile.

6.2 Comparison of Correlations

The deviations of the correlations of Chen and Gungor-Winterton [20] are significantly higher than the other three. The performance of the correlations of Shah, Liu and Winterton, and Gungor & Winterton (1987) are fairly close to each other when using the proposed correction method, the mean deviations being 19.7, 20.5, and 20.3 percent respectively. Using the Thome–Shakir correction factor alone, the Shah correlation performs significantly better with mean deviations of these three correlations being 18.8, 23.4, and 20.3 percent respectively. If using the present proposed correction method, any of these three correlations may be used. If using the Thome-Shakir correction factor only as suggested by Thome [1], the Shah correlation is preferable.

6.3 Effect of Various Parameters on Deviations

Results of data analysis shown in Table 1 were reviewed to see whether there was any systematic trend of increasing or decreasing deviations with parameters such as reduced pressure, glide, boiling number, tube diameter, or tube orientation. No such trend was found. Some authors, for example [3, 13], have suggested that nucleate boiling is more intense in copper tubes compared to other materials. The materials of the tubes used in these tests included copper and several types of stainless steels. No influence of tube material was discernible. Test data included those from electrically heated and liquid heated tubes. The deviations do not appear to be related to the mode of heating.

6.4 Data Uncertainty

All papers have reported the accuracy of the instruments used by the researchers; it is always reported to be excellent. Only some of the researchers have done error propagation analysis to estimate uncertainty in reported heat transfer coefficients. The reported uncertainties are listed in column 1 of Table 1. Most authors have reported uncertainties of 10 % or less. Takamatsu et al. [37] have estimated uncertainty to be upto 25%. The mean deviation of their data from the modified Shah correlation is 32.8 % which is high. For other data, there does not appear to be a relation between uncertainty and deviations. Greco [27] data uncertainty is upto 12.5 %. Of the three fluids tested by him, two have large deviations while one has moderate deviations. Wattelet et al. [38] also have the same uncertainty of upto 12. 5 % but their data show excellent agreement with predictions. Hence reported uncertainties in measurements are of only limited help in interpreting the deviations from correlations.

6.5 Recommended Range of the Proposed Method

The results of data analysis have been presented in Table 1 and the reader may make his own judgment about the range of applicability. The author recommends it in the range of glide, reduced pressure, and boiling numbers included in the data analyzed. The difference in the behavior of mixtures and pure fluids is due to the glide. Hence this is clearly an important parameter. Thome [1] states that the various assumptions made in deriving the Thome- Shakir formula, Eq. (1), are valid upto a glide of about 30 ^{o}C; the maximum glide in Table 1 is 27.7 ^{o}C. The verified range of reduced pressure is 0.048 to 0.634. It will be prudent to keep application within this range until verification outside this range. The maximum boiling number in the data analyzed is 0.0046. It is possible that behavior of the mixture may change at significantly higher boiling numbers. The recommendation is therefore to restrict application to this value until further verification.

7. CONCLUSIONS

1. Two methods for applying general correlations for pure fluids to mixtures were evaluated by comparison to a very wide ranging database. The first method was to apply the Thome-Shakir correction factor to the nucleate boiling terms of the general correlations. The second method (i.e. the proposed method) was to apply the Thome-Shakir correction to nucleate boiling term and the Bell-Ghaly correction to the convective boiling term. These two methods were applied using five well-verified general correlations for pure fluids.

2. The second method gave better agreement with four of the five correlations and with the remaining one, that of Shah, the mean deviation was only slightly higher than with the first method. The mean deviations of the correlations of Shah, Liu & Winterton, and Gungor & Winterton [16] with this method were 19.7, 20.5, and 20.3 percent respectively. The other two correlations gave significantly larger deviations (27.9 and 30.3 percent).

3. The results with the proposed method indicate that it is generally applicable. However, comparison with more test data beyond the presently tested range is desirable. Evaluation of the proposed method using modifications to the Bell-Ghaly correction factor such as that of McNaught is also desirable.

4. It is recommended that the application of the proposed method be limited to the verified range of parameters (glide < 28 ^{o}C, boiling number < 0.0046, and reduced pressures from 0.048 to 0.63) until validated beyond this range.

NOMENCLATURE

Bn	Boiling number = $\dot{q}/(\dot{m}\ h_{lg})$	(-)
C_{pg}	Specific heat of vapor at constant pressure	J/kg K
Co	Convection number = $(1/x - 1)^{0.8}\ (\rho_g/\rho_l)^{0.5}$	(-)
D	Inside diameter of tube	(m)
Fr_l	Froude number for all mass flowing as liquid = $\dot{m}/(\rho_l^2 gD)$	(-)
g	Acceleration due to gravity	(m/s^2)
h	Heat transfer coefficient, also enthalpy	(W/m^{2}K), (J/kg)
h_{cb}	Heat transfer coefficient during convective boiling	(W/m^{2}K)
h_{GS}	Heat transfer coefficient assuming vapor is flowing alone in the tube	(W/m^{2}K)
h_I	Pool boiling heat transfer coefficient with mixture mean properties	(W/m^{2}K)
h_{lg}	Latent heat of vaporization	(J/kg K)
h_{LS}	Heat transfer coefficient with liquid phase flowing alone in the tube	(W/m^{2}K)
h_{nb}	Heat transfer coefficient during nucleate boiling	(W/m^{2}K)
h_{TP}	Two-phase heat transfer coefficient	(W/m^{2}K)
J	= Co, except where Eq. (14) is applicable	(-)
k	Thermal conductivity	(W/m K)
$\dot{m}$	Total mass flux (liquid + vapor)	(kg/m^{2}s)
M	Molecular weight	(-)
N	Number of data points	(-)
p_r	Reduced pressure	(-)
Pr	Prandtl number	(-)
$\dot{q}$	Heat flux	(W/m^2)
T_{BUB}	Bubble point temperature of mixture	(K)
T_{DEW}	Dew point temperature of mixture	(K)
T_{SAT}	Saturation temperature	(K)
x	Vapor quality	(K)
Y	Factor in Bell-Ghaly method, defined by Eq. 3	(-)

Greek

μ	Dynamic viscosity	(kg/m s)
ρ	Density	(kg/m^3)

Subscripts

l	Of liquid
g	Of vapor
mix	Of mixture

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