A General Correlation for Heat Transfer During Saturated Boiling with Flow Across Tube Bundles

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Heat exchangers consisting of bundles of horizontal tubes with boiling on the outer surface of the tubes are widely used in the industry. Examples are refrigerated liquid coolers and kettle reboilers. No well-verified general method for predicting heat transfer during saturated boiling is available in the open literature.

Presented here is a dimensionless correlation for predicting heat transfer on individual tubes of a bundle that can be used with predictive models for bundle heat transfer such as that of Brisbane et al. (1980). The new correlation shows good agreement with data for single tubes and tube bundles from many sources covering a wide range of parameters, including seven fluids (water, n-pentane, R-11, R-12, R-113, R-123, and R-134a), reduced pressures from 0.005 to 0.189, mass velocities from 1.3 to 1391 kg/m^{2}s, heat flux from 1 to 1000 kW/m^2, tube diameters from 3 to 25.4 mm, and pitch to diameter ratios from 1.17 to 1.5. A total of 690 data points are correlated with a mean deviation of 15.2%. The results of comparisons of the new correlation with test data are presented and discussed.

INTRODUCTION

Heat exchangers consisting of bundles of horizontal tubes with boiling on the outer surface of the tubes are widely used in the industry. Examples are refrigerated liquid coolers and kettle reboilers. The boiling liquid may be flowing upwards or downwards. The outer surfaces of the tubes may be plain or enhanced. This paper is concerned only with upward flow across plain tubes.

An evaporator/boiler involves one or more of the following modes of heat transfer:

- subcooled boiling
- saturated boiling prior to dryout
- post-dryout heat transfer

The author has previously presented a general correlation for subcooled boiling heat transfer (Shah 1984, 2005). This paper is concerned exclusively with saturated boiling at vapor qualities from zero upwards, prior to dryout.

Because of the practical importance of such heat exchangers, many experimental studies have been conducted to observe and measure heat transfer on tube bundles and single tubes with cross flow. Further, many correlations for predicting heat transfer have been published. Many of these experimental studies and prediction methods have been reviewed fairly recently by Browne and Bansal (1999) and Cascario and Thome (2001). Study of this literature shows that no well-validated general method for predicting heat transfer in saturated boiling is available in the open literature. The study reported here was undertaken to fill this gap.

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It is agreed by most researchers that correlations for bundle mean heat transfer cannot be generally applicable. For reliable design, one has to use models such as that of Brisbane et al. (1980) that perform calculations of heat transfer coefficients of each tube from bottom to top in terms of the local flow, quality, and heat flux. Hence, the author's efforts were directed toward developing a correlation applicable to individual tubes in bundles.

Presented here is a dimensionless correlation that shows good agreement with data for single tubes and tube bundles from many sources covering a wide range of parameters, including seven fluids (water, n-pentane, R-11, R-12, R-113, R-123, and R-134a), reduced pressures from 0.005 to 0.189, mass velocities from 1.3 to 1391 kg/m^{2}s, heat flux from 1 to 1000 kW/m^2, tube diameters from 3 to 25.4 mm, and pitch to diameter ratios from 1.17 to 1.5. A total of 690 data points are correlated with a mean deviation of 15.2%. The results of comparisons of the new correlation with test data are presented and discussed.

TRENDS SHOWN BY EXPERIMENTAL DATA

The reports on the effect of quality on heat transfer are apparently conflicting. A number of researchers have reported large increases in heat transfer coefficients on a slight increase of quality above zero. Examples are Bitter (1973), Polley et al. (1980), and Burnside and Shire (2005). On the other hand, many researchers report that quality had no effect on heat transfer. Examples are Cotchin and Boyd (1992), Grant et al. (1983), Abbot and Comley (1938), and Webb and Chien (1994). Jensen et al. (1992) found no effect of quality except at very low qualities. Burnside and Shire (2005) reported a modest increase of heat transfer coefficient with quality. Chien and Wu (2004) found a significant increase in heat transfer with increasing quality at higher heat fluxes. Hwang and Yao (1986) found the heat transfer coefficient to increase with quality at low heat flux.

Most of the studies show that at high heat fluxes, the heat transfer coefficient depends on heat flux only and is about the same as that during pool boiling on a single tube; mass flow rate and quality have no effect. Examples are Cotchin and Boyd (1992), Grant et al. (1983), Abbot and Comley (1938), and Webb and Chien (1994). However, methods for determining the heat flux beyond which this occurs are not available. Further, there is the question about which pool boiling correlation to use. The present research proposes answers to these questions.

Many researchers report an increase in the heat transfer coefficient with increasing mass flow rates, for example Hwang and Yao (1986). On the other hand, no effect of mass flow rate is reported by many authors, as noted in the previous paragraph.

From the above, it appears that the different trends reported by various researchers occur under different combinations of parameters. What is needed is to find under what combinations of parameters the various trends occur, i.e., to define the regimes in which particular trends occur. The next requirement is to find methods to predict heat transfer coefficients in the various regimes. This is what the research reported here attempts to do.

THE NEW CORRELATION

This author studied and analyzed test data from many sources. As a result, three regimes of heat transfer were identified and separate equations were developed for heat transfer in each regime, as is given in the following.

Heat Transfer Regimes

Three regimes of heat transfer were identified:

1. *Regime I (Intense Boiling Regime).* In this regime, heat transfer depends only on heat flux; mass velocity and vapor quality have a negligible effect. This regime occurs when

$$Y_{IB} > 0.0008 \,. \tag{1}$$

2. *Regime II (Convective Boiling Regime).* In this regime, both heat flux and mass velocity have an effect on heat transfer; vapor quality has a negligible effect. Thus, both nucleate boiling and convection contribute to heat transfer. This regime occurs when

$$0.00021 < Y_{IB} \le 0.0008 \; . \tag{2}$$

3. *Regime III (Convection Regime).* In this regime, heat transfer is affected by mass velocity and vapor quality; heat flux has a negligible effect. This suggests that bubble nucleation is completely suppressed. This regime occurs when

$$Y_{IB} \le 0.00021 \; . \tag{3}$$

The boiling intensity parameter Y_{IB} is defined as

$$Y_{IB} = F_{pb} Bo \mathrm{Fr}^{0.3} \tag{4}$$

where

$$F_{pb} = h_{pb,\,actual}/h_{cooper} \; . \tag{5}$$

The variable h_{cooper} is the pool boiling heat transfer coefficient calculated by the simplified Cooper correlation, Equation 6:

$$h_{cooper} = 55.1 q^{0.67} p_r^{0.12} (-\log p_r)^{-0.55} M^{-0.55} \tag{6}$$

The variable $h_{pb,actual}$ is the same as h_{cooper} unless pool boiling test data is available for the tubes to be used in the heat exchanger; in that case, $h_{pb,actual}$ is calculated from the test data. Thus, $F_{pb} = 1$ unless test data for the actual tubes used or to be used are available.

Figures 1–4 illustrate the data in the three regimes.

Heat Transfer Equations

In regime I, $$h_{TP} = F_{pb} h_{cooper} \; . \tag{7}$$

In regime II, $$\varphi = \varphi_0 \; . \tag{8}$$

In regime III, $$\varphi = \frac{2.3}{Z^{0.08} \mathrm{Fr}^{0.22}} \; . \tag{9}$$

The parameter Z was introduced by this author to correlate heat transfer during film condensation in tubes (Shah 1979). As heat transfer during film condensation in tubes is due to convective effects only, it was felt that it may be applicable in this regime. It is defined as

$$Z = \left(\frac{1-x}{x}\right)^{0.8} p_r^{0.4} \; . \tag{10}$$

In the heat transfer equations above,

$$\varphi = h_{TP}/h_{LT} \; , \tag{11}$$

where φ_0 is the value of φ when $x = 0$. It is the highest of that calculated by the following relations:

$$\varphi = 443 Bo^{0.65} F_{pb} \tag{12}$$

$$\varphi_0 = 31 Bo^{0.33} F_{pb} \tag{13}$$

$$\varphi_0 = 1 \tag{14}$$

With $F_{pb} = 1$, Equations 12 and 13 are the same as that developed by this author through analysis of varied data from many sources for flow across single tubes at zero vapor quality (Shah 2005).

The variable h_{LT} is the heat transfer coefficient for all mass flowing as a liquid. It is calculated by the following equation (Shah 1984):

$$h_{LT}D/k = 0.21 \mathrm{Re}_L^{0.62} \mathrm{Pr}^{0.4} \tag{15}$$

All fluid properties are calculated at the saturation temperature.

It will be noted that Z cannot be calculated with Equation 10 at $x = 0$ and hence Equation 9 is inapplicable at $x = 0$. This does not pose a problem, as there are no two-phase convective effects at zero vapor quality; the heat transfer coefficient is calculated with Equations 12–14, which apply at $x = 0$. There may be a question about the continuity in regime III between $x = 0$ and qualities slightly higher than 0. The available data in this range showed adequate agreement with the present correlation. Further study is desirable when more data become available, though its practical interest will be limited, as only a very short length is involved.

DEVELOPMENT OF THE CORRELATION

The development of this correlation involved many trials and errors. A brief description of the process is given here.

As noted previously, study of experimental data from many sources indicated the three regimes of boiling. In regime I, heat transfer depends only on heat flux and is the same as in pool boiling. In regime II, both heat flux and mass flux have an effect. As nucleate boiling is suppressed by convective effects of velocity, the author initially thought that the boundary between these regimes would depend on the boiling number, Bo, being the dimensionless ratio of heat flux to mass flux. Examination of data showed that the boundary between the regimes has additional dependence on the mass flux. The Froude number, Fr, was chosen as the dimensionless number to represent the influence of mass flux. Data points showing dependence only on heat flux and those showing dependence on both heat flux and mass flux were plotted on a Bo vs. Fr graph. The boundary between the two regimes was found to be represented by the following equation:

$$Bo \mathrm{Fr}^{0.3} = 0.0008 \tag{16}$$

The left-hand side of this equation is the intensity of boiling parameter Y_{IB} at $F_{pb} = 1$. The factor F_{pb} was added after considering all data, specially those of Robinson and Thome (2004). Thus, Equation 1 was established as the boundary between regimes I and II.

In regime III, heat flux has no effect. Examination of data showed that the boundary between regimes II and III is at $Y_{IB} = 0.00021$.

Having established the boundaries between the three regimes, equations for heat transfer in these regimes were sought. The choice for regime I has been thoroughly discussed in the foregoing. In regime II, heat transfer depends on heat flux and mass flux while vapor quality has no effect. The author's correlation for boiling in tube bundles at zero vapor quality (Shah 2005), Equations 12–14, is therefore applicable.

In regime III, heat transfer increases with increasing mass flux and vapor quality while heat flux has no effect. Thus, the increase of heat transfer above the single-phase heat transfer is purely due to convective effects. The author's first attempt was to calculate two-phase heat transfer using a single-phase heat transfer correlation with the velocity in the Reynolds number of the two-phase mixture based on homogeneous flow. This attempt failed. The next attempt was to plot j against Z, the parameter used in the author's successful correlation for condensation in tubes (Shah 1979), the thinking being that condensation heat transfer is also purely convective. The plots showed that an additional parameter involving mass flux was needed. The Froude number was tried. This approach was successful. After several iterations, Equation 9 was obtained as the correlation for regime III.

COLLECTION OF TEST DATA

Efforts were made to collect data for as many fluids and as wide a range of parameters as possible for horizontal tube bundles as well as single tubes with upward flow of boiling fluid. While there are a large number of such studies, most of them do not provide analyzable data, as has been noted by Casciaro and Thome (2001). For the data to be analyzable, mass flow, heat flux, and quality at the tube location should be known. Most of the studies provide only the heat flux and heat transfer coefficient.

The analyzable studies found and used here are listed in Tables 1 and 2 for single tubes and bundles, respectively. A couple of studies that provided analyzable data were not considered due to reasons discussed later in this paper.

COMPARISON OF CORRELATION WITH DATA

The data listed in Tables 1 and 2 were compared to the present correlation. Where the researchers had also performed pool boiling measurements on the same tubes used for flow boiling tests, F_{pb} was calculated based on these pool boiling data. The average values of F_{pb} from such studies are included in Table 3. Where the researchers did not do pool boiling tests, $F_{pb} = 1$ was used in analyzing their data. Vapor quality at the middle of the tube was used in calculations.

Figures 1–7 show the comparison of some representative data with the present correlation. Figures 1 and 2 show some data of Webb and Chien (1994) in regime I ($Y_{IB} > 0.0008$)—Figure 1 has data for R-113 while Figure 2 has data for R-123. It is seen that the heat transfer coefficient depends only on heat flux; neither vapor quality nor mass velocity affect heat transfer even though they vary over a considerable range. Figure 3 includes data from two test runs with R-113, one in regime II and one in regime III. Figure 4 has data for R-11 at low heat flux, showing the sharp increase in heat transfer coefficient with a small increase in quality characteristic of regime III. It is seen that the new correlation predicts the trends correctly and the quantitative agreement is also good. While Figures 1–4 have data only for halocarbon refrigerants, Figure 5 has data for pentane, an organic chemical. This gives confidence in the correlation's applicability to organics. Figure 6 shows comparison with data for R-12 from two sources. The good agreement here gives further confidence in the general applicability to all halocarbon refrigerants. Figure 7 is of special interest as it shows comparison of the present correlation with data for water. This good agreement and the fact that the properties of water differ vastly from those of refrigerants and organics encourages the hope that the correlation may be applicable to all

Newtonian, non-metallic fluids. However, this optimism should be tempered with much caution, as all these water data are from tests on single tubes and at zero to slightly positive qualities.

Tables 1 and 2 summarize the results of comparison of the correlation with all data for single tubes and tube bundles, respectively. The definition of deviation is

$$\delta = \frac{predicted\ h_{TP} - measured\ h_{TP}}{measured\ h_{TP}} \ . \tag{17}$$

The mean absolute deviation, δ_{mean}, of a data set is defined as

$$\delta_{mean} = \frac{\Sigma(Absolute(\delta))}{N} \ , \tag{18}$$

where N is the number of data points in the set. The average deviation of a data set is calculated using the actual deviations of the data points.

It is seen that a very wide range of data from numerous sources is predicted with a mean absolute deviation of only 15.2%. These results and some aspects of the new correlation are discussed in the following section.

DISCUSSION

Choice of Single-Phase Correlation

In the article on subcooled boiling during cross flow (Shah 2005), the author reported that several correlations for single-phase heat transfer were tried. Among these, the present Equation 15, the Holman correlation (Holman 1968), and the correlation of Churchill and Bernstein (1977) gave comparable mean deviations considering all data sets. However, the results were best when using Equation 15 at a Reynolds number above 700 and the Holman correlation below 700. In the present study, it was found that the mean deviations are minimized by using Equation 15 throughout, including at Reynolds numbers lower than 700.

The calculation of the single-phase heat transfer coefficient is required only when $Y_{IB} < 0.0008$. The number of data points for $Re_L < 700$ together with $Y_{IB} < 0.0008$ is small. Therefore, the present recommendation to use Equation 15 at all Reynolds numbers needs to be verified with more data.

Calculation of Pool Boiling Heat Transfer Coefficient

The new correlation for flow boiling presented here requires the calculation of the pool boiling heat transfer coefficient and uses the simplified Cooper correlation for this purpose. The reasons for this choice are discussed below.

A very large number of correlations have been proposed for pool boiling heat transfer. Most of them are based on small amounts of data and have had little validation. However, those of Stephan and Abdelsalam (1980), Mostinski (1963), and Cooper (1984) have had considerable validation. Among these three, that of Cooper was developed from the largest database covering a very wide range of liquids and operating conditions. The full Cooper correlation is

$$h_{pb} = 55.1 C_{material} q^{0.67} p_r^{(0.12 - 0.211 \log R_p)} (-\log p_r)^{-0.55} M^{-0.55} \ . \tag{19}$$

The variable $C_{material}$ is 1.7 for copper cylinders and is equal to 1 for all other materials and shapes. Cooper stated that there was only indirect evidence supporting this factor of 1.7 and it was not conclusive. He further stated that this factor appears to be illogical and may be changed by new data. R_p is the surface roughness in µm; a value of 1 µm is recommended if the surface roughness is not known.

With $C_{material} = 1$ and $R_p = 1$ µm, Equation 19 becomes

$$h_{cooper} = 55.1 q^{0.67} p_r^{0.12} (-\log p_r)^{-0.55} M^{-0.55} \quad . \tag{6}$$

In this paper, this is called the *simplified Cooper correlation*. This is the form in which the Cooper correlation is most commonly used.

Many of the flow boiling studies whose data have been analyzed here also contain pool boiling data on the same tubes. These data were compared with Equation 6. Some readily available studies on bundles with recirculation also included pool boiling data on the same tubes. These data were also analyzed. The results are shown in Table 3. It is noted that most of the data are in satisfactory agreement. The one notable exception is the data of Robinson and Thome (2004) for copper tubes, whose values are much higher. These data agree with Equation 18 when using the 1.7 factor for copper cylinders and assuming a surface roughness of 5.4 µm; the actual values measured varied from less than 1 to greater than 7. However, all other data sets for copper tubes shown in Table 3 are in reasonable agreement with the simplified Cooper correlation, Equation 6. Thus, it appears that the 1.7 factor for copper cylinders is applicable only rather rarely. It will also be noted in Table 3 that deviations from the Cooper correlation are not related to the tube material or the type of fluid.

From the foregoing, it may be concluded that the simplified Cooper correlation, Equation 6, can be relied upon to make reasonable predictions of heat transfer for all materials with a high probability. There is a low probability that some commercial surfaces may have exceptionally favorable microstructure for nucleation and it may underpredict heat transfer for such cases. Designs should be based on the most probable case; hence, Equation 6 should be preferred to Equation 18. The use of Equation 18 would have resulted in gross overprediction of all data for copper tubes except those of Robinson and Thome (2004).

While the results with the simplified Cooper correlation given in Table 3 are fairly good, further evaluation of this and other pool boiling correlations is desirable to find/develop the most reliable correlation. If a correlation better than Equation 6 is found or developed, it should be used in the present correlation instead of Equation 6.

Consequence of Using $F_{pb} = 1$ in Data Analysis

If F_{pb} is taken to be 1 for all data analyzed, instead of the values from pool boiling tests listed in Table 3, the mean deviations of all data sets will still be in acceptable range except for three data sets. The data of Robinson and Thome (2004) for R-134a will be greatly underpredicted. Indeed, it was the study of this data set that prompted this author to introduce F_{pb} into the correlation. The other two not in acceptable range are the data of Chien and Wu (2004) for R-134a and Singh et al. (1983) for water, which will be underpredicted by about 40%.

Thus, only three of the many data sets analyzed here show large deviation from the present correlation using $F_{pb} = 1$. This is very satisfying, as heat exchanger manufacturers are unlikely to be able to perform pool boiling tests on the tubes to be used in the heat exchanger. These results show that reliable designs can be done in most cases by using the Cooper correlation, Equation 6.

It is also to be noted that in Table 3, the predictions of Equation 6 are always equal to or lower than the test measurements. Hence, any errors resulting from its use are likely to be on the safe side.

Analyzable Data Sets Not Considered

Cornwell et al. (1980) performed tests on a 25 mm wide section of a 17-row tube bundle. These data have also been published in other papers and have been widely quoted. However, Andrews and Cornwell (1987) concluded that these data and other data from such narrow test sections are not accurate, being too high. As the person who performed those tests considers the data unreliable, the data were not included in the present study.

The other analyzable data not considered are those of Fujita and Hidaka (1998). These show varying results with the present correlation, some data points being in agreement while others show large deviations. This lack of good agreement may indicate some shortcoming of the present correlation. However, it may be mentioned that there is some indication of measurement problems. Their single-phase measurements often show a large decrease in heat transfer coefficient along the height, those at the top tube being up to 70% lower than those at the bottom tube. This unusual trend suggests the possibility of instrumentation errors. Hence, these data were not included in the results given in Table 1.

Other Predictive Techniques

A very large number of heat transfer correlations have been proposed. Most of these have been reviewed by Casciaro and Thome (2001) and Browne and Bansal (1999). Only a few are mentioned below. Various theories about mechanisms of heat transfer are not discussed.

Many of the correlations attempt to predict the mean heat transfer coefficient for the bundle as a whole. An early example is that of Palen and Taborek (1962), which was based on data from many heat exchangers for chemical processing. According to it, mean bundle heat transfer is always lower than that during pool boiling on a single tube. There are numerous test data now that show bundle heat transfer coefficients to be higher than those of single tubes. This illustrates the futility of attempting general correlations for mean bundle heat transfer coefficients.

Many dimensional correlations have been proposed for particular fluids that include parameters like pressure and number of rows. Examples are the correlations of Danilova et al. (1972) for ammonia and R-22. These correlations as well as other similar correlations have had very little validation.

Many predictive techniques for local heat transfer based on the correlation of Chen for boiling inside tubes (Chen 1966) have been proposed. An example is the computer program developed by Webb et al. (1989), which was validated with performance data on three commercial chillers provided by the manufacturer. No other validation of this method has come to this author's notice. Chen-type correlations have also been presented by Polley et al. (1980), Hwang and Yao (1986), Webb and Chien (1994), Gupta et al. (1995), Fujita and Hidaka (1998), and Wege and Jensen (1984). These were validated with one or two data sets, each covering a small range of parameters. Their general applicability is unknown. Webb and Chien (1994) also developed an asymptotic model that agreed with their own data as well as R-113 data from two other sources; its general applicability is unknown.

A two-dimensional recirculation model together with correlating equations has been given by Kumar et al. (2003) but has been validated with only one data set. Hence, its general applicability is unknown.

Palen and Yang (1983) presented a correlation based on a wide range of proprietary data for reboilers used in chemical processing. While it seems to be a good rational correlation, it is not available for general use as the authors have not revealed the constants and exponents in their correlation.

It is thus clear that no well-validated method for predicting heat transfer during saturated boiling in tube bundles is available in the open literature.

On the Heat Transfer Regimes

The three heat transfer regimes in the present correlation were arrived at empirically by the study of experimental data. While these have resulted in reasonably satisfactory correlation of data, there seems to be room for improvement. A number of data points identified in a particular regime show better agreement with the heat transfer correlation in another regime. It is also to be noted from Table 4 that the correlation shows significantly higher deviations in regimes II and III compared to regime I. Hence, further efforts are desirable for improving the correlations for these two regimes.

Type of Fluid

The data analyzed are for seven fluids. Five of the fluids are halocarbon refrigerants; their data show satisfactory agreement. The properties of these vary to some extent but are generally similar. Another fluid is n-pentane. Satisfactory agreement with these data suggests general applicability to organics. The seventh fluid is water. Its properties differ vastly from those of the other fluids. There are three data sets for water. All of these are at zero or near-zero vapor quality and are for single tubes. While these are reasonably well correlated, applicability to water at higher qualities and to tube bundles remains unverified. Hence, at present, the new correlation can be recommended only for halocarbon refrigerants and organics.

Tube Arrangement and Pitch

Data on inline and staggered tube bundles are included. No significant difference in deviations from the present correlation is apparent. This is in agreement with experimental studies in which the effect of tube arrangement was tested. An example is the study by Andrews and Cornwell (1987) in which an inline bundle with square pitch was rotated 90° to form a staggered bundle. The difference in heat transfer coefficients in the two cases was found to be negligible. Palen et al. (1972) also did not find any difference between the performance of inline and staggered bundles.

The tube bundle data analyzed include pitch to diameter ratios from 1.17 to 1.5. Deviations do not appear to be related to P/D. Thus, in this range, P/D does not appear to have any effect. Application to smaller P/D ratios should be made with much caution, as tests by Liu and Qiu (2004) show that behavior changes drastically as the gap between adjacent tubes approaches zero. They found that as the gap between tubes approaches zero, heat transfer coefficients at lower heat fluxes increase and dryout heat flux is lowered.

Effect of Heating Mode

Robinson and Thome (2004) measured pool boiling and flow boiling heat transfer on liquid-heated tubes. As seen in Table 3, their pool boiling heat transfer coefficient measurements are much higher than the simplified Cooper correlation, Equation 6. All other pool boiling studies listed in Table 3 were on electrically heated tubes and show reasonable agreement with Equation 6. It makes one wonder whether this difference is due to the difference in heating mode. However, there is a lot of evidence against such a view. First of all, the flow boiling measurements of Roser et al. (1999) were also done on a bundle of liquid heated tubes. These data are in good agreement with the present correlation using Equation 6 to calculate the pool boiling heat transfer coefficient. The pioneering correlation of Chen (1966) for boiling inside tubes directly used a pool boiling correlation to account for the nucleate boiling component. Yet it was based on and verified with data that included electric heating, heating by condensing steam, and heating by hot liquid. The same is the situation with the Gungor and Winterton (1986) correlation for boiling in tubes. The correlation of Shah (1982) for boiling in tubes has been verified with data for 30 fluids including all modes of heating (Shah 2006). Note that the 1.7 factor for copper cylinders was proposed by Cooper (1984) as the best interpretation of pool boiling data examined by him. Most of those data were from electrically heated test sections. The fact that most experimental studies are done with electric heating while actual heat exchangers involve heating by fluids indicates a general consensus that nucleate boiling phenomena are independent of mode of heating. Hence, it may be concluded that the present correlation is applicable to all modes of heating.

Range of Various Parameters

The complete range of dimensional and nondimensional parameters covered by the data analyzed is listed in Table 5. This range is very wide except that of reduced pressure, which is limited to a maximum of 0.18. However, the heat exchangers involving boiling outside tubes normally operate at low pressures. Hence, the range of reduced pressures covered is adequate for most practical purposes.

SUMMARY AND CONCLUSIONS

1. A general correlation for heat transfer during saturated boiling during upflow across plain tubes has been presented. It agrees with most of the published data for single tubes as well as tube bundles. The data include seven fluids over a wide range of parameters. The new correlation is recommended for halocarbon refrigerants and organics. For other fluids, more validation is needed.
2. None of the presently available nonproprietary predictive techniques has had much validation. Hence, the new correlation may be useful in the design of heat exchangers.
3. Further evaluation and development of the new correlation is desirable. Specially desirable is comparison with more data for water and other fluids whose properties differ widely from halocarbon refrigerants and organics—for example, ammonia. New experimental studies are needed for this purpose.

NOMENCLATURE

All equations are dimensionless except the Cooper correlation, Equations 6 and 18. Any consistent system of units may be used in the dimensionless equations.

A	=	total surface area of tube
Bo	=	boiling number, $q/(G \cdot i_{fg})$
D	=	outside diameter of tube
F_{pb}	=	parameter defined by Equation 5
Fr	=	Froude number, $\dfrac{G^2}{\rho_L^2 g D}$
G	=	mass velocity, based on the narrowest gap between tubes of bundle
h	=	heat transfer coefficient
h_{cooper}	=	pool boiling heat transfer by simplified Cooper correlation, Equation 6, $W/m^2 \cdot K$
h_{LT}	=	single-phase heat transfer coefficient assuming all mass flowing as liquid
h_{pb}	=	heat transfer coefficient during pool boiling
$h_{pb,actual}$	=	heat transfer coefficient from pool boiling tests on the actual tube used
h_{TP}	=	heat transfer coefficient with forced convection boiling
i_{fg}	=	latent heat of vaporization
k	=	thermal conductivity of liquid
M	=	molecular weight
Nu	=	Nusselt number, hD/k
p	=	absolute pressure
p_c	=	critical pressure
p_r	=	reduced pressure, p/p_c
P	=	pitch of tube bundle, i.e., center-to-center distance between adjacent tubes
Pr	=	Prandtl number of liquid
q	=	total heat flux on the tube, W/m^2 in Cooper correlation
R_p	=	roughness of surface, μm
Re_L	=	Reynolds number assuming all mass flowing as liquid, GD/μ
x	=	thermodynamic vapor quality
Y_{IB}	=	boiling intensity parameter, defined by Equation 4
Z	=	parameter defined by Equation 10

Greek Letters

φ	= parameter defined by Equation 11	μ	= dynamic viscosity of liquid
φ_0	= value of φ at $x = 0$	ρ	= density of liquid

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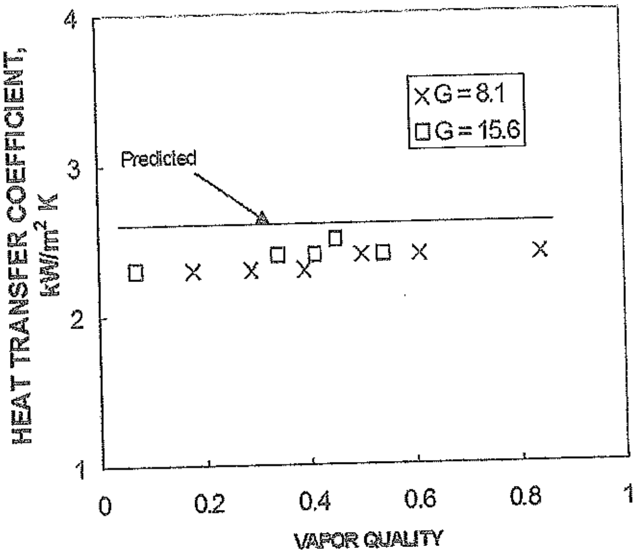

Figure 1. Data of Webb and Chien (1994) for R-113 compared with the present correlation: $q = 54$ kW/m^2, $T_{SAT} = 19$°C, $G = 8.1$ and 15.6 kg/m^{2}s, and $Y_{IB} > 0.0008$ (regime I).

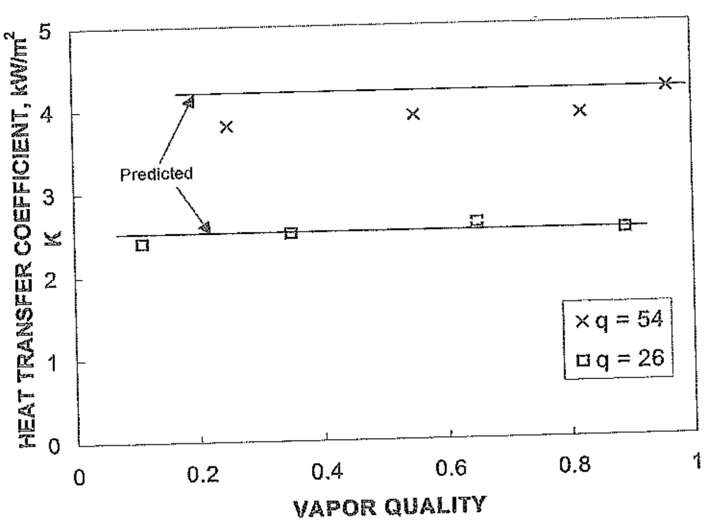

Figure 2. Data of Webb and Chien (1994) for R-123 compared with the present correlation: $G = 8$ kg/m^2, $T_{SAT} = 37$°C, $q = 26$ and 54 kW/m^2, and $Y_{IB} = 17 - 36$ (regime 1).

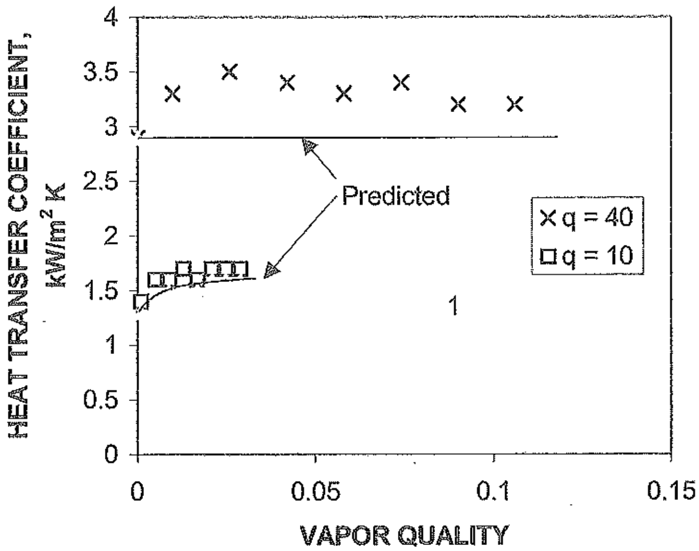

Figure 3. Data of Jensen et al. (1992) for R-113 compared with the present correlation: $G = 217$ kg/m^{2}s, $T_{SAT} = 70$°C, and $q = 40$ (regime II) and 10 (regime III) kW/m^2.

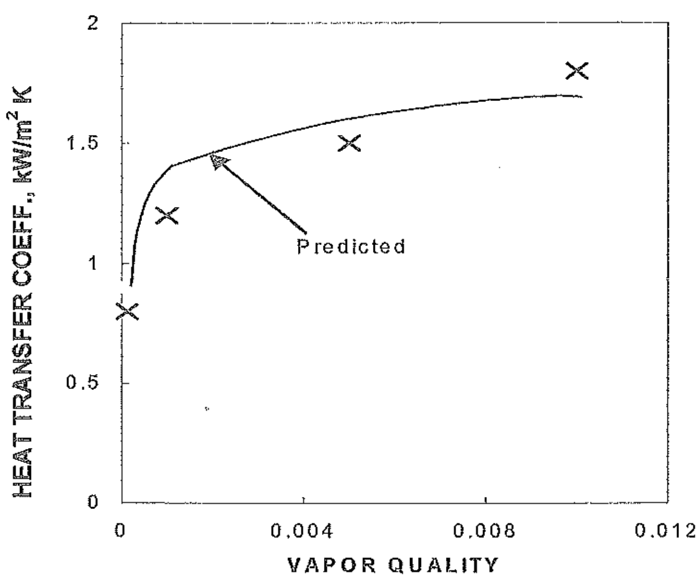

Figure 4. Data of Bitter (1972) for R-11 compared with the present correlation: $G = 190$ kg/m^{2}s, $T_{SAT} = 24.5$°C, $q = 5.1$ kW/m^2, and $Y_{IB} = 0.75 \times 10^{-4}$ (regime III).

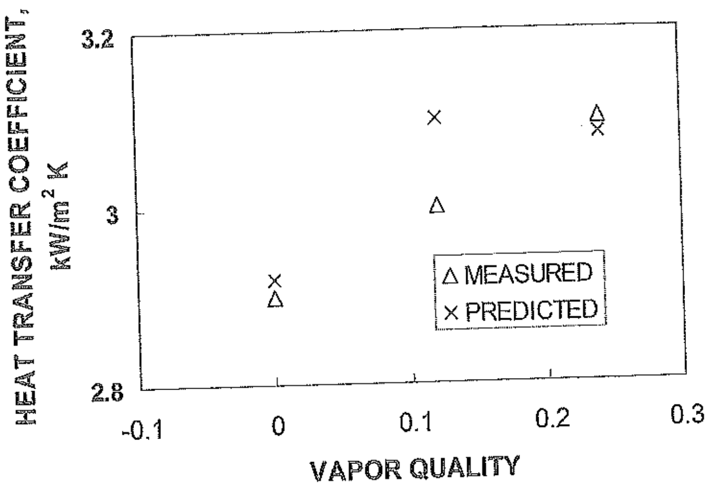

Figure 5. Data of Roser et al. (1999) for n-pentane compared to the present correlation: $T_{SAT} = -50°C$, $G = 30$ kg/m^{2}s, and $q = 42$ kW/m^2.

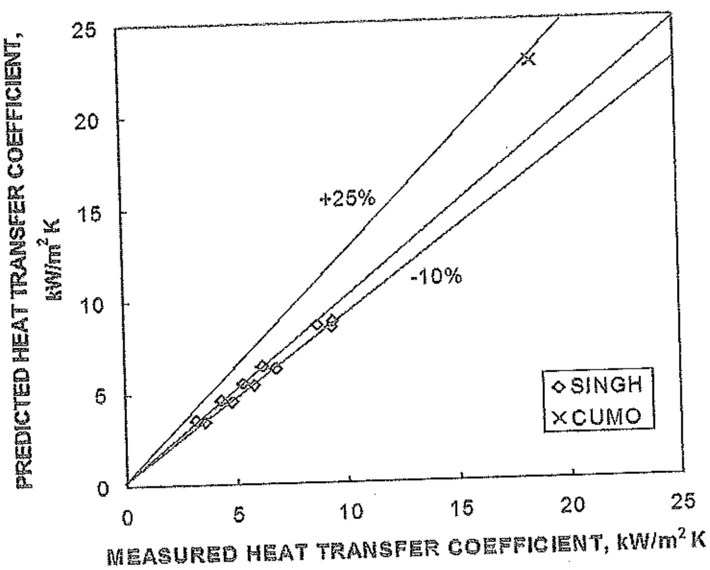

Figure 6. Data of Singh et al. (1985) and Cumo et al. (1980) for R-12 compared with the present correlation.

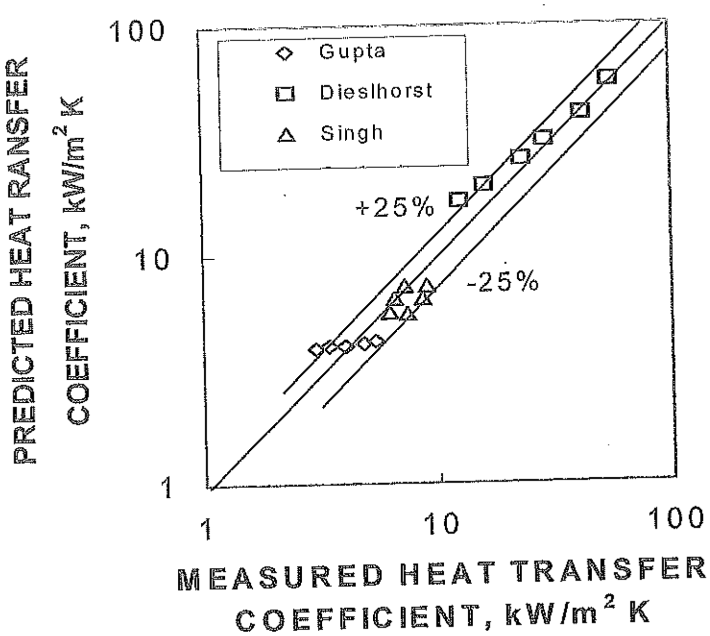

Figure 7. Comparison of the present correlation with the water data of Dieselhorst (1972), Gupta et al. (1995), and Singh et al. (1983).

Table 1. Summary of the Data Analyzed for Flow Boiling on Single Tubes and Results of Comparison with the Present Correlation

Researcher	D,* mm	Tube Material	Fluid	p_r*	G,* kg/m^{2}s	q,* kW/m^2	x,* %	Re$_L$*	1/Z*	$Bo \times 10^4$*	$Y_{IB} \times 10^4$*	Deviation,†‡ %	No. of Data Points
Gupta et al. (1995)	19.05	SS	water	0.005	8.9	10 40	5E-5 22E-5	559	0.003 0.010	5.4 21.6	0.45 2.01	16.2 2.4	5
Singh et al. (1983)	12.7	SS	water	0.005	1.3 2.5	11 44	4E-5 34E-5	58 115	0	40 153	1.5 7.7	15.2 −14.7	17
Dieselhorst (1978)	3.0	SS	water	0.005	300	150 1000	0	3226	0	2.2 14.7	3.2 21.2	14.9 14.0	6
Bitter (1972, 1973)	9.0 15.0	Porcelain, nickel-coated	R-11	0.023	55 1391	1 26	0.00 0.10	1973 49462	0 0.775	0.04 26.0	0.07 5.7	14.6 4.3	34
Singh et al. (1985)	9.5	SS	R-12	0.088	2 5	20 85	0.005 0.052	67 160	0	276 2632	23 132	6.0 −1.6	11
Hwang and Yao (1984)	19.1	SS 304	R-113	0.041	5.9 242	2 100	0.002 0.14	246 10079	0.02 0.86	0.58 710	.3 41.9	13.6 −3.9	55
Wege and Jensen (1984)	12.7	SS 321	R-113	0.050	138 312	30	0.1 0.2	4103 36271	0.57 1.09	1.7 15.5	3.1 7.5	15.0 −6.5	3
All Data	3.0 19.1			0.005 0.088	1.3 1391	1 1000	0.000 0.2	58 49462	0 1.09	1.7 2632	0.07 132	13.7 −3.5	131

* In each cell, the upper line has the minimum value of the parameter and the lower line has the maximum value.

† Mean absolute deviation

‡ Average deviation

Table 2. Summary of the Data for Tube Bundles and Results of Comparison with the Present Correlation

Researcher	D, mm	Bundle Geometry (Data at)	Tube Material (P/D)	Fluid	p_r†	G,† kg/m²s	q,† kW/m²	X†	Re_L†	$1/Z$†	$Bo \times 10^4$†	$Y_{IB} \times 10^4$†	Deviation,** ‡ %	No. of Data Points
Roser et al. (1999)*	19.05	Staggered 17 × 5	Cu (1.33)	n-pentane	0.0089	30	20 42	0 0.26	1276	0 2.9	16 34	4 9	11.1 −10.4	22
Cumo et al. (1980)	13.6	Staggered, 5 × 3 (middle tube heated)	SS (1.25)	R-12	0.189	41	296	0.13	2297	0.42	534	124	22.6 22.6	1
Polley et al. (1980)	25.4	6 × 6 (middle tube in top row)	SS (1.244)	R-113	0.029	450	17	0.002 0.017	22165	.031 .164	2.6	1.9	37.3 −37.3	4
Hwang and Yao (1986)	19.1	16 × 3 (middle tube of row 13)	SS 304 (1.5)	R-113	0.041	132	3 80	0.014 0.062	5497	0.014 0.016	0.12 0.41	0.6 6.8	18.8 −17.1	28
Chien and Wu (2004)	15.9	Staggered 5 × 3 (middle)	(1.5)	R-134a	0.102	10 40	10 50	0.07 0.24	688 2752	0.31 0.99	13.1 261	5.6 30.5	20.8 −20.8	30
				R-123	0.017 0.030	10 40	10 50	0.02 0.10	482 1926	0.18 0.70	14.7 294.1	4.4 27.4	12.7 7.9	25
Jensen and Hsu (1988)	7.9	Inline 27 × 5	SS (1.3)	R-113	0.060	100 675	6.3 37.8	0.001 0.184	1980 13326	0.01 0.93	0.67 7.6	1.2 6.9	15.4 9.6	56
Robinson and Thome (2004)*	18.9	Staggered 8 × 3	Copper (1.17)	R-134a	0.084	5 39	2 35	0.1 0.81	392 3057	0.46 48.0	9 123	4.8 21.1	18.2 −16.5	58
Cornwell and Scoones (1988)	25.4	Inline 15 × 6	Brass (1.25)	R-113	0.029	150	6 36	0 0.35	7354	0 2.5	2.7 16.2	1.0 6.1	13.7 10.5	35
Webb and Chien (1994)	16.8	Staggered	Copper (1.42)	R-113	0.01 0.021	7.9 37.0	13 55	0.07 0.88	184 1077	0.28 23.7	24.0 430	5.9 34.8	11.4 0.2	90
				R-123	0.019 0.038	7.7 15.0	13 54	0.06 0.96	348 819	0.41 47.0	51 408	6.6 36.5	10.3 1.4	63
Burnside and Shire (2005)	19.1	Inline 17 × 5	Cu-Ni (1.33)	R-113	0.03	211 622	10 65	0.004 0.25	7800 23235	0.06 1.7	1.0 12.9	1.3 12.9	22.8 22.1	70
Jensen et al. (1992)	19.1	Staggered 15 × 5 (middle column)	Copper (1.17)	R-113	0.059 0.176	51 500	10 40	0 0.74	2452 34930	0 7.2	3.3 32.8	3.0 8.8	16.4 −3.5	61
			Copper (1.5)	R-113	0.059	217	10 40	0 0.106	10233	0 0.56	3.3 13.3	1.7 16.9	9.2 −9.2	16
	7.9 25.4		1.17 1.5		0.005 0.189	1.7 679	2 296	0 0.88	184 23235	0 48	0.12 534	0.5 124	15.6 −0.03	559

* Tubes heated by fluid; electrically heated tubes in all other tests.
† In each cell, the upper line has the minimum value of the parameter and the lower line has the maximum value.
** Mean absolute deviation
‡ Average deviation

Table 3. Comparison of Data for Pool Boiling on Tubes with the Simplified Cooper Correlation, Equation 6

			Tube			Pool Boiling Test Data		
						Data Correlation*		Average
Data Source	Fluid	Reduced Pressure	Material	Diameter, mm	Rp, μm	a	n	$F_{pb} = \frac{h_{pb,actual}}{h_{cooper}}$
Bitter (1973)	R-11	0.021	Porcelain, nickel coated	15.0	6			1.17
Wallner (1971)	R-11	0.023	Nickel coated ceramic	10.0	3	191	0.72	1.03
Singh et al. (1985)	R-12	0.087	SS	9.5				1.23
Jensen et al. (1992)	R-113	.029	Cu	19.1				1.0
Wege and Jensen (1984)	R-113	0.05 0.079	SS	12.7				1.14
Hwang and Yao (1986)	R-113	0.029	SS	19.1		224	0.67	1.1
Marto and Anderson (1992)	R-113	0.021	Cu	15.9		175	0.745	1.23
Webb and Chien (1994)	R-113	0.010 0.021	Cu	16.8				1.1
Memory et al. (1995)	R-114	0.031	Cu	15.9				1.19
Webb and Chien (1994)	R-123	0.02 0.038	Cu	16.8				1.17
Robinson and Thome (2004)	R-134a	0.084	Cu	18.9	< 1 - > 7	1008	0.66	2.7
Gupta et al. (1995)	Water	0.0046	SS	19.1				1.0
Singh et al. (1983)	Water	0.0046	SS	12.7				1.44
Dieselhorst (1978)	Water	0.0046	SS	6.0				1.0
Cornwell and Scoones (1988)	R-113	0.021	Brass	25.4				1.0
Chien and Wu (2004)	R-123	0.017 0.030	Copper	19.05		453 516	0.48 0.48	1.25
	R-134a	0.102				1010	0.49	1.37

* Correlated by $h = a \cdot q^n$, with h in W/m^2·°C and q in kW/m^2.

Table 4. Comparison of the Present Correlation with Data in the Three Heat Transfer Regimes of This Correlation

Regime	No. of Data Points	Mean Deviation Percent	No. of Data Points with Deviation	
			> 30%	> 40%
I	261	10.7	10	1
II	288	18.2	51	7
III	141	17.6	18	7
All Regimes	690	15.2	79	15

Table 5. Complete Range of Data Correlated

Fluids	Water, n-pentane, R-11, R-12, R-113,R-123, R-134a
Tube diameter	3–25.4 mm
Tube material	Copper, brass, cupro-nickel, stainless steel, nickel-coated porcelain
Tube arrangements	Single tubes and tube bundles. Square inline and equilateral triangular staggered, *P/D* from 1.17 to 1.5.
Pressure	0.3–7.8 bar
Reduced pressure	0.005–0.189
Heat flux	1–1000 kW/m^2
Mass flux	1.3–1391 kg/m^2s
Re_L	58–4949,462
$1/Z$	0–2.9
$Bo \times 10^4$	0.12–2632
$Y_{IB} \times 10^4$	0.07–132