An Improved and Extended General Correlation for Heat Transfer During Condensation in Plain Tubes

M. Mohammed Shah, PhD, PE
Fellow ASHRAE

Received December 2, 2008; accepted April 1, 2009

An improved version of the author's published correlation (Shah 1979), extended to a wider range of parameters, is presented. The new correlation has been shown to be in good agreement with data ranging from highly turbulent flows to the laminar flow conditions of Nusselt's analytical solutions. The data used for the correlation's validation includes 22 fluids (water, halocarbon refrigerants, hydrocarbon refrigerants, and organics) condensing in horizontal, vertical, and downward-inclined tubes. The range of parameters includes tube diameters from 2 to 49 mm, reduced pressure from 0.0008 to 0.9, flow rates from 4 to 820 kg/m^2·s, all liquid Reynolds numbers from 68 to 85,000, and liquid Prandtl numbers from 1 to 18. A total of 1189 data points from 39 sources are predicted with a mean deviation of 14.4%. Comparisons are also made with some other well-known correlations.

INTRODUCTION

Three decades ago, the author presented a general correlation for heat transfer during film condensation inside plain tubes (Shah 1979). It was shown to agree with data for water, refrigerants, and organics covering a wide range of conditions in horizontal, vertical, and inclined tubes. In a later paper (Shah 1981), the author stated that this correlation will fail at very low flow rates, and tentative conservative limits of applicability were provided.

Numerous other researchers have compared this correlation with a wide range of data, and, with very few exceptions, have reported good agreement (examples include Dobson and Chato [1998], Moser et al. [1998], and many others). However, the author decided to further investigate and develop this correlation with the following objectives:

1. Verify/modify the lower limit of applicability.
2. Develop modifications to extend the correlation down to the lowest flow rates (i.e., those in which Nusselt's analytical equations apply).
3. Test the correlation with data for the many new refrigerants that have been developed since the correlation was developed.
4. Test the correlation at reduced pressures higher and lower than those in the original database.
5. Modify the correlation as needed if shortcomings are found.

This paper presents the results of these efforts. As will be seen, the objectives of this research have been substantially met. The improved correlation presented here is shown to be in good agreement with data from 39 sources for 22 fluids that include water, halocarbon refrigerants (chlorofluorocarbons, hydrochlorofluorocarbons, and haloalkane), hydrocarbon refrigerants,

M. Mohammed Shah is with Fletcher Thompson, Inc., Shelton, CT.

and a variety of organics; tube diameters from 2 to 49 mm; flow rates from 4 to 820 kg/m^2·s, and reduced pressures from 0.0008 to 0.9.

THE PUBLISHED SHAH CORRELATION

In 1979, Shah presented the following correlation:

$$h_{TP}/h_{LS} = 1 + 3.8/Z^{0.95} \tag{1}$$

where h_{LS} is the heat transfer of the liquid phase flowing alone in the tube. It is calculated by the following equation:

$$h_{LS} = 0.023 \mathrm{Re}_{LS}^{0.8} \mathrm{Pr}_f^{0.4} \tag{2}$$

Equations 1 and 2 may be combined to give the following equation:

$$h_{TP} = h_{LT}\left[(1-x)^{0.8} + \frac{3.8x^{0.76}(1-x)^{0.04}}{p_r^{0.38}}\right] \tag{3}$$

In Equation 3, h_{LT} is calculated by Equation 2 with the substitution of Re_{LT} for Re_{LS}.

In Shah (1981), the author suggested that until further research provided better criteria, this correlation be used only if all of the following conditions were met:

$$\mathrm{Re}_{LT} > 350,\ \mathrm{Re}_{GT} > 35,000,\ V_{GT} > 3 \text{ m/s}$$

This recommendation was based on several reports. The data satisfactorily correlated by the author were at $V_{GT} \geq 3$ m/s and $\mathrm{Re}_{LT} > 350$. Borchman (1967) reported good agreement of his data with the Nusselt equation at $V_{GT} < 3$ m/s, and Chato (1967) reported that his laminar condensation analysis applies at $\mathrm{Re}_{GT} < 35{,}000$. This recommended limit is very conservative.

THE NEW CORRELATION

Heat Transfer Regimes

Vertical and Inclined Tubes: For vertical and inclined tubes, three heat transfer regimes have been identified, as shown in Figure 1.

The boundary between Regimes I and II is given by the following relation. Regime I occurs when

$$J_g \geq \frac{1}{2.4Z + 0.73}. \tag{4}$$

The boundary between Regimes II and III is given by the following relation. Regime III prevails when:

$$J_g \leq 0.89 - 0.93\exp(-0.087Z^{-1.17}) \tag{5}$$

J_g is the dimensionless vapor velocity defined as:

$$J_g = \frac{xG}{(gD\rho_g(\rho_l - \rho_g))^{0.5}} \tag{6}$$

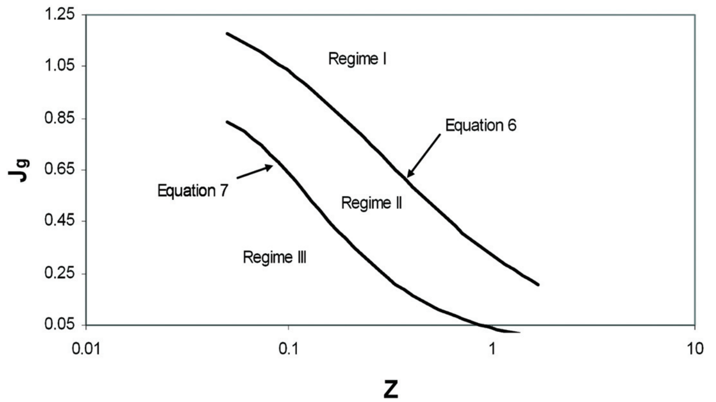

Figure 1. Heat transfer regimes in vertical tubes, according to the present correlation.

Horizontal Tubes: For horizontal tubes, only two regimes have been identified by the present data analysis. These are shown in Figure 2. The boundary between Regimes I and II is given by the following relation. Regime I occurs when.

$$J_g \geq 0.98(Z + 0.263)^{-0.62} . \tag{7}$$

A third regime is expected at very low flow rates. Analyzable data were not available for such conditions.

Heat Transfer Equations

The new correlation uses the following two heat transfer equations:

$$h_I = h_{LT}\left(\frac{\mu_f}{14\mu_g}\right)^n\left[(1-x)^{0.8} + \frac{3.8x^{0.76}(1-x)^{0.04}}{p_r^{0.38}}\right] \tag{8a}$$

where:

$$n = 0.0058 + 0.557p_r \tag{8b}$$

The second equation is

$$h_{\mathrm{Nu}} = 1.32\mathrm{Re}_{LS}^{-1/3}\left[\frac{\rho_l(\rho_l - \rho_g)gk_f^3}{\mu_f^2}\right]^{1/3} . \tag{9}$$

Equation 9 is the Nusselt equation for laminar film condensation in vertical tubes; the constant has been increased by 20% as recommended by McAdams (1954) on the basis of comparison with test data. This equation can also be expressed in terms of heat flux or temperature difference instead of Reynolds number. This form has been preferred as it is more convenient for this correlation and often it is also more convenient for design calculations.

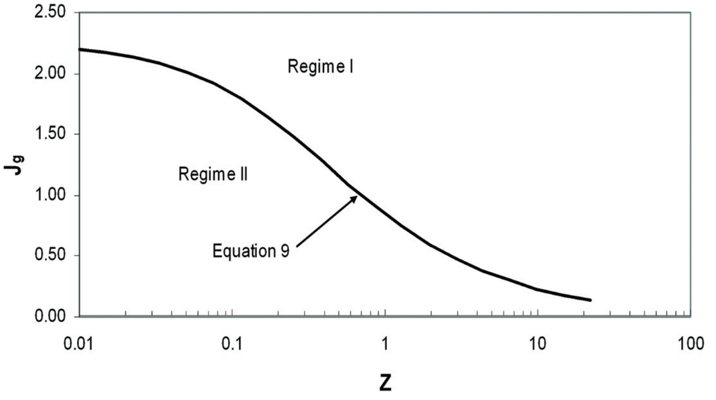

Figure 2. Heat transfer regimes in horizontal tubes, according to the present correlation.

These two heat transfer equations are used as follows:
For all tube orientations (except upward flow):
In Regime I:

$$h_{TP} = h_I \tag{10}$$

In Regime II:

$$h_{TP} = h_I + h_{\mathrm{Nu}} \tag{11}$$

For horizontal tubes, Equation 11 is recommended only if $\mathrm{Re}_{GT} > 35{,}000$.
For vertical tubes in Regime III

$$h_{TP} = h_{\mathrm{Nu}} . \tag{12}$$

DEVELOPMENT OF THE PRESENT CORRELATION

The development of the present correlation given above involved many trials and errors. These efforts are briefly described below.

Comparison of the author's 1979 correlation with a wide range of data showed that it was failing for some fluids at high reduced pressures at moderate to high flow rates. The deviations were found to be related to the viscosity ratio of phases and reduced pressure. A correction factor was developed through data analysis which led to Equation 8a. This equation was found to give good agreement with data at higher flow rates for both horizontal and vertical tubes.

Vertical Tubes

It is well known that at very low flow rates, heat transfer in vertical tubes can be predicted with good accuracy by the Nusselt relation (Equation 9). The author expected that at intermediate flow rates, heat transfer could be predicted by suitably combining Equations 8a and 9. It was found that satisfactory agreement was obtained by simply adding the heat transfer coefficients predicted by these two. Thus, Equation 11 was obtained.

Thus, it was qualitatively established that Equation 8a applied at high flow rates, Equation 9 at the lowest flow rates, and Equation 11 at intermediate flow rates. Then, it was necessary to quantitatively establish the limits of applicability of these equations. Typically, researchers have defined the limits of their formulas in terms of flow patterns (examples include Dobson and Chato [1998], Thome et al. [2003]), and many others. However, the author's correlation (Shah 1979) has been found to agree with a very wide range of data that must have included many flow patterns. Besides, there are significant disagreements among various flow pattern maps. So, it was necessary to determine these limits directly through data analysis using dimensionless parameters. J_g and Z were selected from many parameters. The former, known as the dimensionless gas velocity, has been used in many flow pattern maps including those of Breber et al. (1980) and Tandon et al. (1982). The parameter Z was introduced by the author in his very successful correlation for condensation heat transfer (Shah 1979). Equations 4 and 5 were established by analysis of data that used these parameters. In Figure 1, the curves representing these equations have been drawn only in the range of data analyzed.

Horizontal Tubes

Nusselt has also provided an analytical solution for condensation on the outer surface of tubes. Hence, that relation will appear to be the correct choice instead of Equation 9, which is for vertical tubes. However, the solution for horizontal tubes is based on the condensate being continuously drained from the bottom of the tube. During condensation inside horizontal tubes, condensate accumulates inside the tube, as it does in vertical tubes. Therefore, the author decided to attempt a correlation at intermediate flows using a combination of Equations 8a and 9, in the same way as used for vertical tubes. Available data were satisfactorily correlated in this way. It should be emphasized that this result is empirical; no theoretical merit is claimed.

The boundary between high and intermediate flow still needed to be established. Analysis of data led to Equation 7 becoming the boundary between Regime I and II. In Figure 2, the curve representing this equation has been drawn only in the range of data analyzed.

Analyzable data for horizontal tubes were available only for $\mathrm{Re}_{GT} \geq 15{,}800$. Chato (1962) has provided an analytical solution for horizontal and slightly inclined (downward) tubes with a limit that is stated to be $\mathrm{Re}_{GT} \leq 35{,}000$. It is a modification of Nusselt's solution for condensation outside of horizontal tubes. Kroger (1976) reported agreement of his data with Chato's formula. As the present database contained very few data for Re_{GT} that were well below 35,000, it appears advisable to conservatively set the limit of the present correlation for horizontal tubes at $\mathrm{Re}_{GT} > 35{,}000$. For more discussion on horizontal and slightly inclined tubes with very low flow, see Shah (1981).

Inclined Tubes

The only analyzable data for inclined tubes were from Tepe and Mueller (1947). The data is for tubes inclined downward at 15°. The data show satisfactory agreement with the heat transfer regime relations for vertical tubes. Subject to verification with more data, it is recommended that heat transfer regimes for tubes inclined downward at 15° and greater be calculated as they are for vertical tubes.

COMPARISON OF PRESENT CORRELATION WITH DATA

Data Search and Selection Criteria

A large amount of literature was reviewed in am attempt to obtain data covering as wide a range of parameters as possible. Unfortunately, many of the papers do not present the test in an analyzable form. For comparison with the present correlation, flow rate, pressure, and vapor quality should be known. Many papers give data only in terms of q vs. ΔT, their correlating parameters, or Re vs. h_{TP}/h_{LS}, etc. Such data could not be compared with the present correlation, though the last-mentioned type had been compared to the Shah (1979) correlation.

Only pure fluids, azeotropic mixtures, and near-azeotropic mixture data were considered. The near-azeotropic mixtures included were R-404A and R-410A. The temperature glide for both is less than 0.5 °C, and so they were treated as pure fluids during the calculations. Only those data for refrigerants were considered in which oil content was zero or negligible according to the authors of those papers.

Only data for macrochannels was considered. Macrochannels usually include channels with a diameter that is greater than 2 mm. Here, data for diameters including 2 mm were included. While presently there is great interest in microchannels, the author of this study felt that those needed a separate study, as surface tension effects become important in microchannels.

Only data for horizontal flow and downward flow have been included, as physical phenomena during upward flow are different in many respects.

Where the publications provided a large amount of data, data representative of the range were taken from them. For example, if the data included mass velocities of 100 to 800 kg/m^2·s at interval of 100, the runs at larger intervals (e.g., 200) were used. Similarly, if the data were for qualities from 0 to 1.0 at 0.1 intervals, data were taken at larger intervals, such as 0.2. The purpose was to minimize effort without loss of useful information. It has been the author's experience that samples of data collected in this way are sufficient for the purpose of the development of a correlation. No data points were deleted from any test run analyzed, even if they had large deviations and were suspected to be erroneous.

Fluid Property Data Sources

The primary source of property data was the University of Ottawa Code UO0694 (obtained from the university's mechanical engineering department). The other major source was the *2005 ASHRAE Handbook—Fundamentals* (ASHRAE 2005). The University of Ottawa code provided data for water, R-11, R-12, R-22, R-113, R-123, R-134a, and benzene. ASHRAE (2005) provided data for R-32, R-125, R-404A, R-410A, R-507, propylene, propane, and isobutane. REFPROP Version 8 provided data for R-142b and R-502 (NIST 2007). Beaton and Hewitt (1988) provided data for methanol, ethanol, and toluene. The data for Dowtherm 209 was taken from Blangetti and Schlunder (1979). The program for data analysis was initially prepared using only the University of Ottawa property code; other sources were used as necessary.

All fluid properties were calculated at the saturation temperature.

Results of Comparison of Data with the Present Correlation

The salient features of the data that were analyzed are listed in Tables 1 and 2. Table 1 lists the data for horizontal tubes, and Table 2 includes vertical and inclined tubes. Many of the data were for mean heat transfer coefficients over the length of the tubes. Such data were analyzed by using the arithmetic average quality in calculations. This is an approximation because actual mean quality can be lower than the arithmetic mean quality as was discussed in Shah (1979). Hence the author would have preferred to use only local heat transfer data but included mean heat transfer data, as local heat transfer data were not available in that range.

Tables 1 and 2 list the mean and average deviations of the present correlation. Mean deviation δ_m is defined as

$$\delta_m = \frac{1}{N}\sum_{N}^{1} ABS(h_{predicted} - h_{measured})/h_{measured}\,. \tag{13}$$

Table 1. Salient Features of Data for Horizontal Tubes and Results of Comparison with the Present Correlation

Source	Diameter, mm	Fluid	p_r	G, kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Varma (1977)	49.0	water	0.0023	12.6	0.95 0.58	1808	54,415	4	6.3 1.6
Tang et al. (2000)	8.8	R-134a	0.25	260 820	0.81 0.09	11,573 36,500	181,808 573,395	24	8.2 –2.0
		R-410A	0.495	320 720	0.81 0.091	29,822 73,624	191,929 473,824	16	16.8 16.8
		R-22	0.308	270 790	0.91 0.09	11,591 33,914	165,849 485,263	28	8.1 –7.8
Bae et al. (1969)	12.5	R-22	0.235 0.325	210 634	0.90 0.09	12,579 38,430	193,612 569,436	27	15.2 –3.0
Bae et al. (1968)		R-12	0.197 0.211	344 634	0.91 0.03	17,721 32,932	327,303 599,510	29	17.9 –16.8
Powell (1961)	12.8	R-11	0.035	258	0.24	8689	283,628	1	3.5 3.5
Lambrecht et al. (2006)	8.1	R-22	0.308	300 800	0.5	11,854 31,611	169,619 452,317	6	21.3 21.3

*These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 1. Salient Features of Data for Horizontal Tubes and Results of Comparison with the Present Correlation *(Continued)*

Source	Diameter, mm	Fluid	p_r	G, kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Jung et al. (2003)	8.0	R-32	0.428	100 300	0.5	8430 25,290	55,402 166,205	3	9.9 –3.2
		R-12	0.127	100 300	0.93 0.10	4253 12,759	63,431 190,294	14	20.4 –15.2
		R-125	0.559	100 300	0.90 0.15	7306 21,918	42,781 128,342	13	15.8 –15.8
		R-123	0.042	100 300	0.90 0.15	2675 8024	70,573 211,720	15	14.5 12.7
		R-142b	0.126	100 300	0.92 0.2	4073 12,220	72,727 218,182	13	10.1 0.9
Infante-Ferreira et al. (2003)	8.0	R-404A	0.491	250 600	0.88 0.14	19,605 47,053	150,036 360,086	16	13.4 –9.8
Park et al. (2008)	8.8	propylene	0.354	100 300	0.91 0.10	10,784 32,355	90,072 270,215	28	32.6 32.6
		isobutane	0.146	100 300	0.89 0.10	6882 20,646	110,913 332,739	21	11.2 10.0
		propane	0.322	100 300	0.88 0.1	10,643 31,930	93,739 281,217	27	16.4 16.4
		R-22	0.308	100 300	0.90 0.10	4293 12,879	61,426 184,277	27	9.2 –6.6

*These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 1. Salient Features of Data for Horizontal Tubes and Results of Comparison with the Present Correlation *(Continued)*

Source	Diameter, mm	Fluid	p_r	G, kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Jiang and Garimella (2003)	9.4	R-404A	0.805 0.907	200 500	0.88 0.20	28,415 84,827	96,507 275,264	40	9.0 –5.1
Lee et al. (2006)	10.9	propylene	0.354	150	0.88 0.01	20,074	167,656	10	17.2 –17.2
		isobutane	0.146	150	0.88 0.01	12,810	206,450	10	13.7 –13.7
		propane	0.32	150	0.90 0.01	19,811	174,483	10	15.2 –15.2
		R-22	0.308	150	0.91 0.01	7991	114,336	10	24.2 –14.2
Jung et al. (2004)	8.8	R-134a	0.250	100 300	0.98 0.05	4461 13384	70,085 210,255	27	13.3 –12.0
		R-410A	0.495	100 300	0.94 0.03	9341 28022	60,114 180,342	27	5.9 –2.0
		R-22	0.308	100 300	0.96 0.08	4303 12908	61,565 184,696	26	20.7 –19.2

*These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 1. Salient Features of Data for Horizontal Tubes and Results of Comparison with the Present Correlation *(Continued)*

Source	Diameter, mm	Fluid	p_r	G, kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Eckels and Tesene (1993)	8.0	R-507	0.505	251 599	0.80 0.10	19844 47455	147,434 352,565	23	15.5 7.8
		R-502	0.411	600	0.75 0.13	38989	342,547	8	21.0
Eckels et al. (1993)	8.0	R-12	0.233	134 374	0.47* 0.43	4560 12726	79,488 221,742	5	7.1 0.8
	8.0 11.0	R-134a	0.245	87 3368	0.49* 0.43	3511 14851	55,531 234,889	12	5.7 0.6
Nan and Infante Ferreira (2000)	8.8	propane	0.286	150 250	0.59 0.10	15132 25220	144,510 240,849	6	10.5 –9.6
Dobson and Chato (1998)	7.0	R-410A	0.438	75 650	0.90 0.09	5172 44,827	37,258 322,900	18	9.3 –3.5
		R-22	0.272	75 650	0.90 0.16	2558 22,171	37,768 327,323	18	16.3 –14.7
		R-134a	0.219	75 650	0.9 0.09	2622 22,725	42,961 372,331	19	15.4 –14.9
Wijaya and Spatz (1995)	7.7	R-22	0.272 0.405	481 495	0.80 0.21	18,138 18,587	245,041 274,408	18	12.5 –11.1
		R-410A	0.573 0.652	481	0.79 0.25	43,405 47,297	231,147 242,447	13	6.6 –6.6
Shao and Granyrd (1995)	6.0	R-134a	0.191	183	0.92 0.10	5351	90,935	6	7.3 –4.2

*These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 1. Salient Features of Data for Horizontal Tubes and Results of Comparison with the Present Correlation *(Continued)*

Source	Diameter, mm	Fluid	p_r	G, kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Cavallini et al. (2001)	8.0	R-134a	0.250	65 750	0.80 0.28	2630 30,349	41,320 476,769	37	8.6 –6.6
		R-410A	0.495	750	0.75 0.20	63,542	408,939	7	29.8
		R-125	0.559	100 750	0.80 0.23	7306 54,795	42,781 320,856	23	11.0 –8.6
		R-32	0.429	100 600	0.80 0.24	8430 50,580	55,402 332,410	24	10.5 5.7
		R-22	0.308	100 750	0.85 0.20	3903 29,270	55,842 418,812	31	11.3 –10.6
Altman et al. (1959)	8.7	R-22	0.268 0.441	300 618	0.92 0.23	12,725 26,166	184,687 379,779	15	14.2 –14.2
Azer et al. (1972)	12.7	R-12	0.219 0.296	210 446	0.99 0.35	115,362 4690	195,269 411,239	39	22.7 9.7
Chitti and Anand (1995)	8.0	R-22	0.272 0.356	149 437	0.75 0.20	5793 17,124	84,608 236,958	12	22.2 –22.2
Berrada et al. (1996)	8.9	R-134a	0.278	170 214	0.79 0.25	7765 9774	117,866 148,373	14	18.0 17.2
		R-22	0.312	114 214	0.80 0.12	4963 9317	70,769 132,846	12	12.1 –2.3
Jassim et al. (2007)	8.9	R-134a	0.164	100 300	0.94 0.04	75,125 12,663	75,125 225,375	25	21.6 –21.6

*These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 1. Salient Features of Data for Horizontal Tubes and Results of Comparison with the Present Correlation *(Continued)*

Source	Diameter, mm	Fluid	p_r	G, kg/m²·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Akers et al. (1959)	15.7	R-12	0.662	78 418	0.94* 0.63	6786 36,356	67,301 360,575	32	6.9 1.6
		propane	0.657	13 162	0.83* 0.51	3899 48,103	17,473 215,578	15	20.5 20.5
Tepe and Mueller (1947)	18.5	benzene	0.021	54 82	0.57* 0.51	3264 4991	106,965 163,546	6	10.3 –4.6
Yan and Lin (1999)	2.0	R-134a	0.16 0.32	100 200.	0.94 0.10	1012 2076	15,892 33,764	21	15.0 –7.0
All data	2.0 49.0		0.0023 0.907	13 820	0.98 0.01	1012 84,827	15,892 476,789	931	14.3 –2.5

*These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 2. Range of Data in Vertical and Downward-Inclined Tubes, and Comparison with the Present Correlation

Source	Diameter, mm	Fluid	p_r	G kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Jakob et al. (1932)	40.0	water	0.0046	24 48	0.96* 0.82	3427 6854	79,438 158,877	29	7.8 1.2
Al-Shammari et al. (2004)	28.2	water	0.0008	3	0.9 0.4	173	8210	6	11.8 10.6
Kuhn et al. (1997)	47.5	water	0.023	10	0.94 0.12	2554	32,642	8	18.8 –4.9
Borishanskiy et al. (1978)	10.0 19.3	water	0.036 0.308	12 598	0.5*	763 58,546	8284 333,119	24	14.9 –1.1
Lee and Kim (2008)	12.0	water	0.0046	27 45	0.75 0.06	1183 1944	27,421 45,071	14	18.3 7.9
Goodykoontz and Dorsch (1967)	7.4	water	0.002 0.0062	131 264	0.92 0.06	3827 6567	78,853 167,186	25	14.0 4.5
Blagetti and Schlunder (1978)	30.0	water	0.0046	4 69	0.75 0.04	408 7474	91,732,524 48	19	23.1 0.4
Blagetti and Schlunder (1979)	30.0	Dowtherm 209	0.008	4 81	0.98 0.04	68 1464	9534 20,5932	24	19.9 –15.9

These are inclined tube data. All others are for vertical tubes.
* These are mean heat transfer data. Range of mean quality for the tube length is listed.

Table 2. Range of Data in Vertical and Downward-Inclined Tubes, and Comparison with the Present Correlation *(Continued)*

Source	Diameter, mm	Fluid	p_r	G kg/m^2·s	x	Re_{LT}	Re_{GT}	Number of Data	Deviation Percent
Carpenter (1948)	11.6	ethanol	0.017	11 147	0.75* 0.50	307 3891	14,294 181,405	12	24.1 –10.2
		toluene	0.025	32 154	0.50*	1505 7141	41,976 97,587	9	24.1 –10.1
		methanol	0.016	23 148	0.72* 0.50	874 5533	24,396 154,522	6	22.7 22.7
		water	0.0048	16 140	0.66* 0.50	692 5934	15,686 134,474	10	17.3 13.8
Lilburne and Wood (1982)	12.8	R-113	0.030 0.034	18 50	0.98 0.63	1205 1541	50,850 141,042	12	13.7 7.5
Mochizugi et al. (1984)	13.9	R-11	0.042	80	0.9 0.1	3109	93234	8	5.0 0.7
Cavallini and Zecchin (1971)	20.0	R-11	0.025 0.028	85 303	0.92* 0.65	4232 15,905	152,816 523,317	28	4.1 –2.4
Tepe and Mueller (1947)	18.5	benzene	0.021	25 66	0.62* 0.52	1513 3996	49,576 130,954	11#	13.7 –11.7
				52 88	0.60* 0.51	3174 5369	104,001 175,940	4	10.3 –10.3
		methanol	0.016	16 30	0.71* 0.53	970 1819	27,100 50,813	4#	16.7 –0.6
For All Sources Above	7.4 47.5		0.0008 0.308	4 598	0.98 0.04	68 58,406	9534 523,317	253	

These are inclined tube data. All others are for vertical tubes.
* These are mean heat transfer data. Range of mean quality for the tube length is listed.

Average deviation is defined as

$$\delta_{avg} = \frac{1}{N}\sum_{N}^{1}(h_{predicted} - h_{measured})/h_{measured}\,. \tag{14}$$

The mean deviation of all of the horizontal tube data is 14.3% and that of the vertical and inclined tubes is 15.9%. The mean deviation of all 1189 data points for all tube inclinations is 14.4%.

Table 3 gives a breakdown of the data in the three heat transfer regimes. It is seen that the agreement with data is satisfactory in all regimes. However, the mean deviation for vertical tubes in Regime II is the highest (21.8%). This could be partially attributed to the fact that a few data points have very high deviations. Higher deviations also occur near the boundaries between the heat transfer regimes.

Table 4 lists the complete range of data over which the present correlation has been verified.

COMPARISON WITH OTHER PREDICTIVE TECHNIQUES

Besides the author's correlation (Shah 1979), numerous predictive techniques have been proposed, most of them for horizontal tubes. Many of them are analytically derived (examples include Moser et al. [1998], Thome et al. [2003], Dobson and Chato [1998], and Traviss et al. [1973]). Some are entirely empirical (examples include Cavallini et al. [2006], Akers et al. [1959], and Ananiev et al. [1961]). The last mentioned is often called the *Boyko-Kruzhilin correlation*, which is based on the co-authors of that paper. Among these predictive methods, only that of Cavallini et al. has been based on and verified with a wide variety of fluids covering a very wide range of parameters. The correlation is intended to be applied to all flow rates, from the highest to the lowest. The Dobson-Chato method is also applicable to all flow rates but has only been validated with data for halocarbon refrigerants.

The objective of this research was not to evaluate various correlations. But, the data compared with the author's correlation have also been compared with a few others. The results are presented here so that it may be viewed in perspective. The correlations chosen are those of Cavallini et al. (2006), Moser et al. (1998), Traviss et al. (1973), Ananiev et al. (1961), and Shah (1979). Except for Shah's correlation, all are stated to be only for horizontal tubes. No well-validated correlation for vertical tubes was found. Hence, comparison has been made only with horizontal tube data.

As noted earlier, Shah (1979) recommended his correlation only for higher flow rates. Traviss et al. (1973) derived their formulas using the annular flow pattern, and hence, should be expected to apply only at higher flow rates. The Cavallini et al. (2006) correlation gives two sets of formulas: one for higher flow rates and one for lower flow rates. Their formulas for lower flow rates require heat flux (or ΔT). For most of the data sets in Tables 1 and 2, heat flux was not known. Hence comparison could be made only with their correlation for higher flow rates. They call it the *heat flux-independent regime*. This regime occurs when the following condition is met:

$$J_g \geq [(7.5/(4.3X_{tt}^{1.11} + 1))^{-3} + C^{-3}]^{-1/3} \tag{15}$$

where

Table 3. Breakdown of the Results of the Present Correlation for Various Tube Orientations and Heat Transfer Regimes

Tube Orientation		Heat Transfer Regime I N	I Deviation Percent	II N	II Deviation Percent	III N	III Deviation Percent
Horizontal	Mean	726	13.5	205	17.0	N/A	N/A
	Average		–0.4		–9.7		
Vertical	Mean	169	15.0	33	21.8	41	15.8
	Average		4.0		9.1		–4.5
Inclined	Mean	10	17.4	5	8.7	0	
	Average		–17.4		8.7		

$$X_{tt} = \left(\frac{1-x}{x}\right)^{0.9}\left(\frac{\rho_g}{\rho_f}\right)^{0.5}\left(\frac{\mu_f}{\mu_g}\right)^{0.1} \tag{16}$$

where $C = 1.6$ for hydrocarbons and $C = 2.6$ for all other fluids. It may be noted that Regime I of the present correlation is also heat flux independent, but it differs significantly from that of Equation 15.

All prediction methods were tested within the range defined by Equation 15 to ensure that all were within their applicable range. Results of this comparison are presented in Table 5. While all of the tested correlations performed reasonably well, the Cavallini et al. (2006) correlation has the least mean deviation (12.6%). The present correlation has a mean deviation of 13.8%. The deviations of other prediction methods are significantly higher.

Table 4. Complete Range of Parameters in the Data Showing Satisfactory Agreement with the Present Correlation

Parameter	Range
Fluids	Water, R-11, R-12, R-22, R-32, R-113, R-123, R-125, R-134a, R-142b, R-404A, R-410A, R-502, R-507, isobutane, propylene, propane, benzene, ethanol, methanol, toluene, and dowtherm 209
Tube diameter, mm	2 to 49
Tube orientations	Horizontal, vertical downwards, 15° downward
Reduced pressure	0.0008 to 0.905
G, kg/m^2·s	4 to 820
Pr_f	1 to 18
Re_{LT}	68 to 84827
Re_{GT}	9534 to 523317
x	0.01 to 0.99
Z	0.005 to 20
J_g	0.06 to 20

Comparisons of some test data with these correlations are shown in Figures 3 through 10. Figure 6 is especially interesting, as it features a comparison of various correlations with data at a reduced pressure of 0.9. In this figure, it's obvious that the Ananiev et al. (1961) and Cavallini et al. (2006) give good agreement. Other predictive schemes, shown in this figure grossly over-predict. Data in Figures 3 through 6 display the heat flux-independent regime defined by Equation 15, as well as Regime I of the present correlation. Figures 7 through 9 display data in Regime II and show the contributions of Equations 8a and 9 to the predicted heat transfer coefficients.

DISCUSSION

Type of Fluids

Data for 22 fluids have been analyzed including halocarbon refrigerants, water, hydrocarbon refrigerants, and organics. The properties of these fluids differ so greatly that applicability to most fluids is likely. The fluids included several that did not exist when the original Shah correlation was developed in 1979 (R-32, R-123, R-125, R-134a, R-142b, R-404A, R-410A, and R-507). The data for Dowtherm 209, which has a Prandtl number of 18, was especially interesting, since it was the highest of the 22 fluids.

Efforts were made to find data for cryogenic fluids, as they are a distinct group. While some papers reporting experimental studies were found, none of them provided mass flow rate and vapor quality, and so they could not be analyzed.

Data for fluid mixtures that have large temperature glide were not analyzed. It is likely that they would be in agreement with the present correlation if correction for mass transfer effect was applied. The well-known method for correcting mass transfer effects proposed by Bell and Ghaly (1973) was successfully used by Cavallini et al. (2006) for adjusting the predictions of their correlation for condensation of mixtures. This could work for the present correlation, as well.

Various Parameters

The range of parameters over which the present correlation was verified was extremely wide, as seen in Table 4. The range of reduced pressures (0.0008 to 0.9) covered almost all practical applications. The tube diameters varied from 2 to 49 mm. Larger diameters were rarely used, and 2 mm was the lower limit of the macrochannels. The only limitations of the data were tube inclinations of less than 15°, and an Re_{LT} of less than about 16,000 for horizontal tubes. While many tests have been completed under those conditions, the publications did not provide analyzable data; those data have probably been irretrievably lost to future researchers. Hopefully, more data will be forthcoming with which this correlation may be tested and further extended.

Table 5. Deviations of Various Correlations for Horizontal Tube Data in the Heat Flux-Independent Regime as Given by Equation 15

No. of Data		Moser et al. (1998)	Ananiev et al. (1961)	Traviss et al. (1973)	Shah (1979)	Cavallini et al. (2006)	Present
444	Mean	18.6	19.9	29.8	23.2	12.6	13.6
	Average	–4.0	–16.1	22.1	12.8	–5.4	1.9

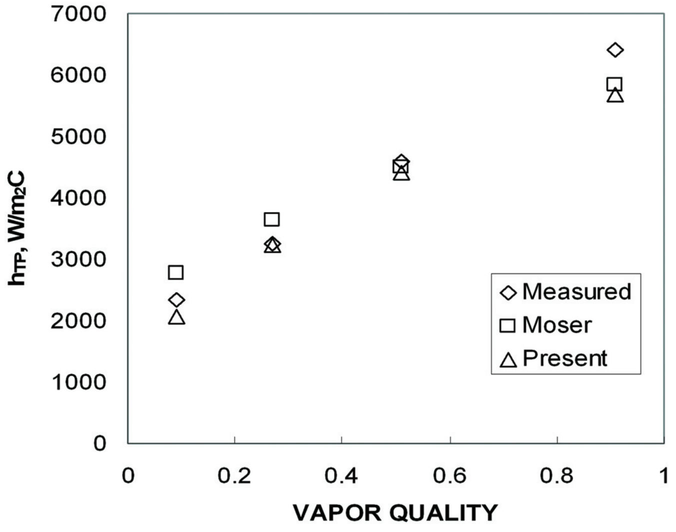

Figure 3. Comparison of the present correlation and that of Moser et al. (1998) with data from Tang et al. (2000). R-22 at 40*C* in a horizontal 8 mm diameter tube. G = 560 kg/m^2·s.

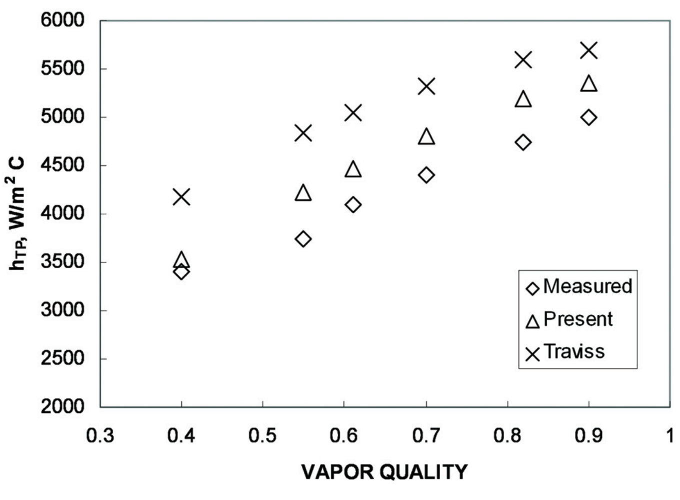

Figure 4. Comparison of the present correlation and that of Traviss et al. (1973) with data from Jung et al. (2003). R-32 in a horizontal 8.8 mm diameter tube. T_{SAT} = 40*C*, G = 300 kg/m^2·s.

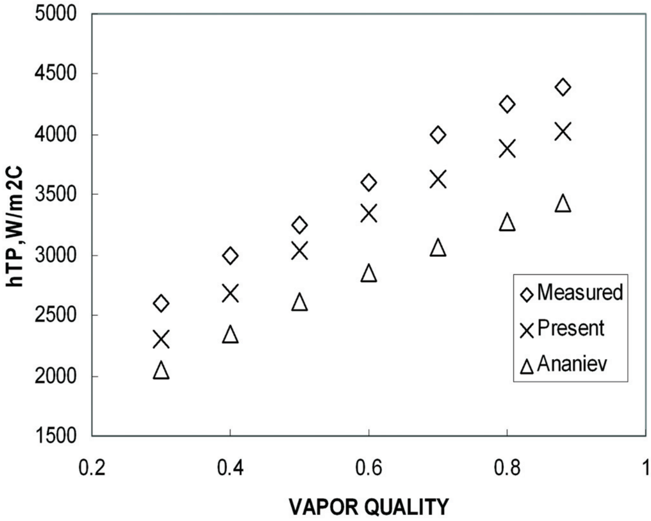

Figure 5. Comparison of the present correlation and that of Ananiev et al. (1961) with data from Lee et al. (2006) for isobutane in a horizontal tube. $T_{SAT} = 40C$, $G = 150$ kg/m^2·s.

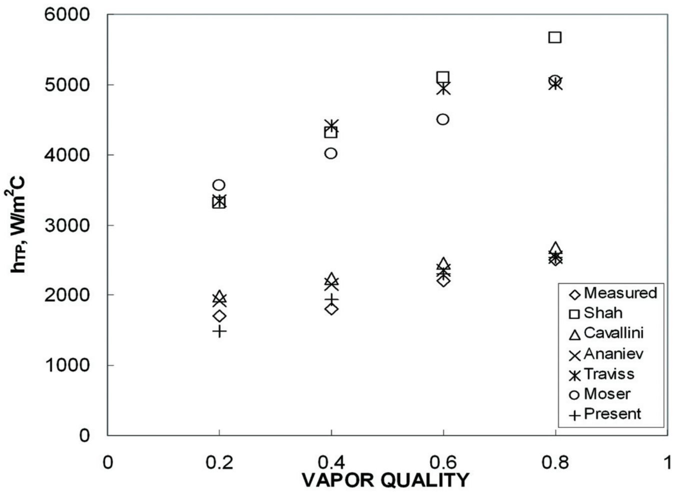

Figure 6. Comparison of the present correlation and those of Ananiev et al. (1961), Shah (1979), Traviss et al. (1973), Moser et al. (1998), and Cavallini et al. (2006) with data from Jiang and Garimella (2003) for R-404A in a horizontal tube. $G = 400$ kg/m^2·s, $p_r = 0.9$.

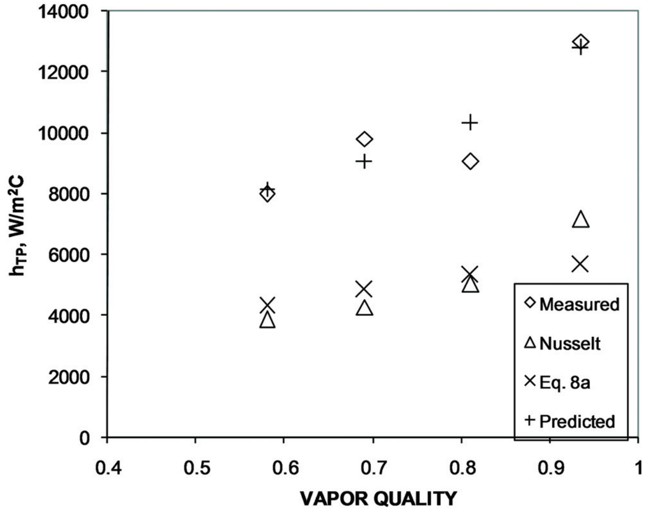

Figure 7. Comparison of the data from Varma (1977) for water in a horizontal 49 mm diameter tube. With the present correlation, T_{SAT} = 82.2*C*, *G* = 12.6 kg/m^2·s. Data are in Regime II. Hence, predictions are the sum of those by Equations 8a and 9 (the Nusselt equation).

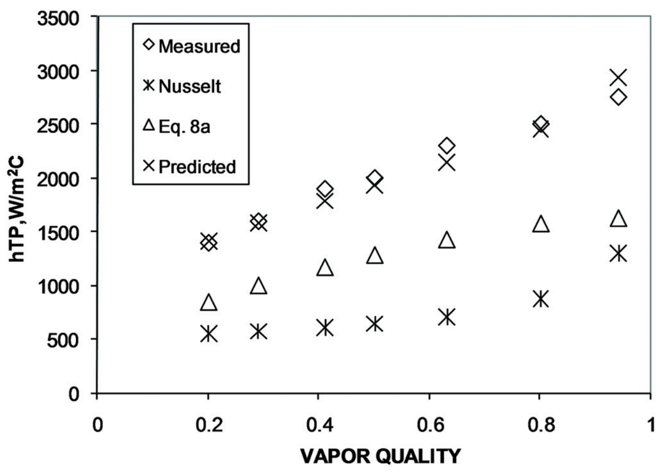

Figure 8. Comparison of the present correlation with data from Jung et al. (2004) for R-410A in a horizontal 8.8 mm diameter tube. T_{SAT} = 40*C*. Data are in Regime II. Hence, predictions are the sum of Equations 8a and 9 (the Nusselt equation).

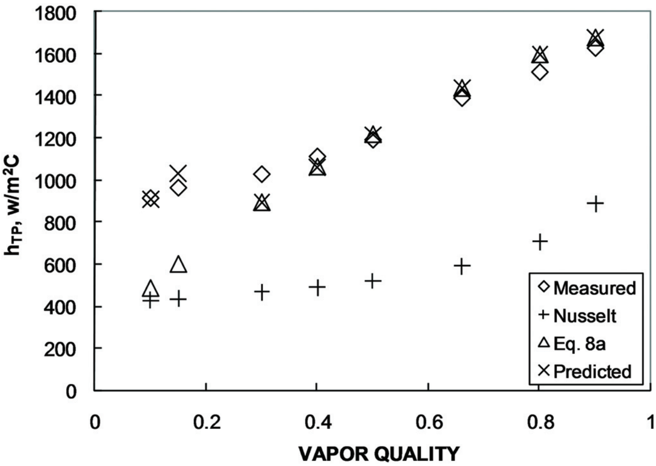

Figure 9. Comparison of the present correlation with data from Mochizuki et al. (1984) for R-11 in a vertical 13.9 mm diameter tube. $G = 80.4$ kg/m^2·s, $T_{SAT} = 42.4C$.

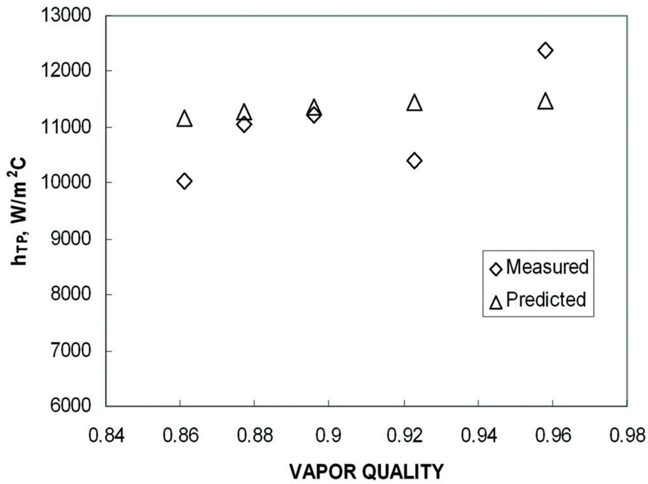

Figure 10. Comparison of the present correlation with data from Jakob et al. (1932) for water at atmospheric pressure condensing in a vertical 40 mm diameter tube. All data are in Regime I.

Physical Interpretation of Heat Transfer Regimes

In Regime III, the Nusselt equation applies. As it is based on the assumption of laminar flow, it may be appropriate to call it the *laminar regime*. In Regime I, Equation 8a was used, which incorporates Equation 2 and which is based on data for fully turbulent flow. Thus, Regime I may be considered to be the turbulent regime. In Regime II, the contributions of the laminar and turbulent equations were added. So, it may be appropriate to call it a *transition regime*. These interpretations are, of course, purely empirical. Analytical studies are needed for validation.

CONCLUDING REMARKS

1. The objectives of this research effort have been substantially fulfilled. The author's published correlation (Shah 1979) has been tested, modified, and its range of applicability has been widened. The present correlation has been shown to be applicable to vertical tubes at all flow rates and to horizontal tubes down to $Re_{GT} \geq 16{,}000$. It has been shown to agree over a reduced pressure range of 0.0008 to 0.9, with data for 22 fluids that include water, halocarbon refrigerants, hydrocarbon refrigerants, and various organics.
2. The present correlation is the only well-validated general correlation for vertical tubes. For horizontal tubes, it provides strong agreement with data over the entire range. Hopefully, this correlation will be helpful in the design and analysis of heat exchangers.
3. Further research is needed for validating/extending this correlation to horizontal and slightly inclined tubes at $Re_{GT} < 16{,}000$. Analyzable data from earlier studies are not available. Further checking and refinement of the boundaries between the heat transfer regimes is desirable.

NOMENCLATURE

D = inside diameter of tube
G = total mass flux (liquid + vapor)
g = acceleration due to gravity
h = heat transfer coefficient
h_I = heat transfer coefficient given by Equation 10
h_{LS} = heat transfer coefficient assuming liquid phase flowing alone in the tube
h_{LT} = heat transfer coefficient assuming all mass flowing as liquid
h_{Nu} = heat transfer coefficient given by Equation 11, the Nusselt relation
h_{TP} = two-phase heat transfer coefficient
J_g = dimensionless vapor velocity defined by Equation 6
N = number of data points
p_r = reduced pressure
Re_{GT} = Reynolds number assuming total mass flowing as vapor, $= GD/\mu_g$
Re_{LS} = Reynolds number assuming liquid phase flowing alone, $= G(1-x)D/\mu_f$
Re_{LT} = Reynolds number assuming total mass flowing as liquid, $= GD/\mu_f$
T_{SAT} = saturation temperature
V_{GT} = vapor velocity assuming all mass flowing as vapor
X_{tt} = Martinelli's correlating parameter, defined by Equation 18
x = vapor quality
Z = Shah's correlating parameter, $(1/x-1)^{0.8}p_r^{0.4}$

Greek Symbols

μ = Dynamic viscosity
ρ = density

Subscripts

f = of liquid
g = of vapor

REFERENCES

Al-Shammari, S.B., D.R. Webb, and P. Heggs. 2004. Condensation of steam with and without the presence of non-condensable gases in a vertical tube. *Desalination* 169:151–60.

Akers, W.W., H.A. Deans, and O.K. Crosser. 1959. Condensing heat transfer within horizontal tubes. *Chem. Eng. Prog. Symp. Ser.* 59(29):171–76

Altman, M., F.W. Staub, and R.H. Norris. 1959. Local heat transfer and pressure drop for Refrigerant 22 condensing in horizontal tubes. ASME AIChE Conference, Storrs, CT.

Ananiev, E.P., I.D. Boyko, and G.N. Kruzhilin. 1961. Heat transfer in the presence of steam condensation in horizontal tubes. *Int. Developments in Heat Transfer* 2:290–95.

ASHRAE. 2005. *2005 ASHRAE Handbook—Fundamentals*. Atlanta: American Society of Heating, Refrigerating and Air-Conditioning Engineers, Inc.

Azer, N.Z., L.V. Abis, and H.M. Soliman. 1972. Local heat transfer coefficients during annular flow condensation. *ASHRAE Transactions* 78(2):135–43.

Bae, J., L. Maulbetsch, and W.M. Rohsenow. 1968. Refrigerant forced convection condensation inside horizontal tubes. Report DSR-79760-59, Massachusetts Institute of Technology, Cambridge, MA.

Bae, J., L. Maulbetsch, and W.M. Rohsenow. 1969. Refrigerant forced convection condensation inside horizontal tubes. Report DSR-79760-64, Massachusetts Institute of Technology, Cambridge, MA.

Beaton, C.F., and G.F. Hewitt. 1988. *Thermophysical Property Data for the Chemical and Mechanical Engineer*. New York: Hemisphere.

Bell, K.J., and M.A. Ghaly. 1973. An approximate generalized method for multicomponent/partial condenser. AChE Symp. Ser. 69:72–79.

Berrada, N., C. Marviller, A. Bontemps, and S. Daudi. 1996. Heat transfer in tube condensation of a zeotropic mixture of 23/134a in a horizontal smooth tube. *Int. J. Refrig.* 19(7):463–72.

Blangetti, F., and E.U. Schlunder. 1978. Local heat transfer coefficients in condensation in vertical tubes. *Proceedings of the Sixth International Heat Transfer Conference, Toronto, Canada,* pp. 437–42.

Blangetti, F., and E.U. Schlunder. 1979. "Local heat transfer coefficients in fil condensation at high Prandtl numbers," in *Condensation Heat Transfer*, eds. P.J. Marto and P.G. Kroger, 17–25. New York: American Society of Mechanical Engineers.

Borishanskiy, V.M., D.I. Volkov, N.I. Ivashenko, G.A. Makarova, Yu. T. Illarionov, L.A. Vorontsova, I.A. Alekseyev, N.I. Ivashchenko, and O.P. Kretunov. 1978. Heat transfer in steam condensing inside vertical pipes and coils. *Heat Transfer Soviet Research* 10(4):44–58.

Borchman, J. 1967. Heat transfer of high velocity vapor condensing in annuli. *ASHRAE Transactions* 73(VI.2.1–VI.2.13).

Breber, G., J.W Palen, and J. Taborek. 1980. Prediction of horizontal tubeside condensation of pure components using flow regime criteria. *J.heat Transfer* 102(3):471–76.

Carpenter, F.G. 1948. Heat transfer and pressure drop for condensing pure vapors inside vertical tubes at high vapor velocities. PhD disertation, Department of Chemical Engineering, University of Delaware, Newark, DE.

Cavallini, A., and R. Zecchin. 1971. High velocity condensation of R-11 vapors inside vertical tubes. In "Studies on Heat Transfer in Refrigeration," Proc. IIR Commission 2, Trondheim, Norway, pp. 385–96.

Cavallini, A., G. Censi, D.D. Col, L. Doretti, G.A. Longo, and L. Rossetto. 2001. Experimental investigation on condensation heat transfer and pressure drop of new refrigerants (R134a, R125, R32, R410A, R236ea) in a horizontal smooth tube. *Int. J. Refrig.* 21:73–87.

Cavallini, A., D.D. Col, L. Doretti, M. Matkovic, L. Rossetto, and C. Zilio. 2006. Condensation in horizontal smooth tubes: a new heat transfer model for heat exchanger design. *Heat Transfer Engineering* 27(8)31–38.

Chato, J.C. 1962. Laminar condensation inside horizontal tubes. *ASHRAE J.* 4(2):52–60.

Chitti, M.S., and N.K. Anand. 1995. An analytical model for local heat transfer coefficients for forced convective condensation inside smooth horizontal tubes. *Int. J. Heat Mass Transfer* 2:615–27.

Dobson, M.K., and J.C. Chato. 1998. Condensation in smooth horizontal tubes. *J. Heat Transfer* 120: 193–213.

Eckels, S., T.M. Doerr, and M.B. Pate. 1993. Heat transfer and pressure drop during condensation and evaporation of R-134a/oil mixtures in smooth and micro-fin tubes. Final Report, RP-630, American Society of Heating, Refrigerating and Air-Conditioning Engineers, Inc., Atlanta.

Eckels, S.J., and B. Tesene. 2002. Forced convective condensation of refrigerants R-502 and R-507 in smooth and enhanced tubes. *ASHRAE Transactions* 102(2):627–38.

Goodykoontz, J.H., and R.G. Dorsch. 1967. Local heat transfer coefficients for condensation of steam in vertical downflow within a 5/8 inch diameter tube. TN D-3326, National Aeronautics and Space Administration, Washington, DC.

Infante-Ferreira, C.A., T.A. Newell, J.C. Chato, and X. Nan. 2003. R404A condensing under forced flow conditions inside smooth, microfin and cross-hatched tubes. *Int. J. Refrigeration* 26:433–41.

Jakob, M., S. Erck, and H. Eck. 1932. *Forschungsheft Geb. Ing. Wes.* 3:290–95.

Jassim, E.W., T.A. Newell, and J.C. Chato. 2008. Prediction of two-phase condensation in horizontal tubes using probabilistic flow regime maps. *Int. J. Heat Mass Transfer* 51(3-4):485–96.

Jiang, Y., and S. Garimella. 2003. Heat transfer and pressure drop for condensation of R-404A at near critical pressure. *ASHRAE Transactions* 109(1):677–88.

Jung, D., K. Song, Y. Cho, and S. Kim. 2003. Flow condensation of heat transfer coefficients of pure refrigerants. *Int. J. Refrigeration* 26:4–11.

Jung, D., Y. Cho, and K. Park. 2004. Flow condensation heat transfer coefficients of R22, R134a, R407C, and R410A inside plain and microfin tubes. *Int. J. Refrigeration* 27:25–32.

Kroger, D.C. 1976. Laminar condensation heat transfer inside inclined tubes. *Chem. Eng. Progress Symp. Ser.* 73(164):160–256.

Kuhn, S.Z., V.E. Schrock, and P.F. Peterson. 1997. An investigation of condensation from steam-gas mixtures flowing downwards inside a vertical tube. *Nuclear Engineering and Design* 177:53–69.

Lee, H., J. Yoon, J. Kim, and P.K. Bansal. 2006. Condensing heat transfer and pressure drop characteristics of hydrocarbon refrigerants. *Int. J.Heat Mass Transfer* 49:1922–27.

Lilburne, G.M., and D.G. Wood 1982. Condensation inside a vertical tube. *Proceedings of the Seventh International Heat Transfer Conference, Munich, Germany,* pp. 113–17.

McAdams, W.H. 1954. *Heat Transmission*, 3d ed. New York: McGraw Hill.

Mochizuki, S., Y. Yagi, R. Tandano, and W. Jang. 1984. Convective filmwise condensation of nonazeotropic binary mixtures in a vertical tube. *J.Heat Transfer* 106:531–38.

Moser, K.W., R.L. Webb, and B. Na. 1998. A new equivalent Reynolds number model for condensation in smooth tubes. *J. Heat Transfer* 120:410–16.

Nan, X.H., and C.A. Infante-Ferreira. 2000. In-tube evaporation and condensation of natural refrigerant R 290 (propane). *Preliminary Proceedings of the Fourth IIR Gustav Lorentzen Conference on Natural Working Fluids, Purdue University, West Lafayette, IN,* pp. 248–53.

NIST. 2007. *REFPROP*, Version 8. National Institute of Standards and Technology, Gaithersburg, MD.

Park, K., D. Jung, and T. Seo. 2008. Flow condensation heat transfer characteristics of hydrocarbon refrigerants and dimethyl ether inside a horizontal plain tube. *Int. J. Multiphase Flow* 34:628–35.

Powell, C.K. 1961. Condensation inside a horizontal tube with high vapor velocity. Master's thesis, Department of Mechanical Engineering, Purdue University, West Lafayette, IN.

Shah, M.M. 1979. A general correlation for heat transfer during film condensation in pipes. *Int. J. Heat Mass Transfer* 22:547–56.

Shah, M.M. 1981. Heat transfer during film condensation in tubes and annuli; a literature survey. *ASHRAE Transactions* 87(1):1086–105.

Shao, D.W., and E. Granryd. 1995. Heat transfer and pressure drop of 134a-oil mixtures in horizontal condensing tube. *Int. J. Refrig.* 18(8):524–33.

Tandon, T.N., H.K. Varma, and C.P. Gupta. 1982. A new flow regime map for condensation inside horizontal tubes. *J. Heat Transfer* 104(4):763–68.

Tang, L., M.M. Ohadi, and A.T. Johnson. 2000. Flow condensation in smooth and micro-fin tubes with HCFC-22, -134a and -410A refrigerants. *Enhanced Heat Transfer* 7:289–310.

Tepe, J.B., and A.C. Mueller. 1947. Condensation and subcooling inside an inclined tube. *Chem. Eng. Prog.* 43(5):267–78.

Thome, J.R., J. El Hajal, and A. Cavallini. 2003. Condensation in horizontal tubes, part 2: new heat transfer model based on flow regimes. *Int. J. Heat Mass Transfer* 46:3365–87.

Traviss, D.P., W.M. Rohsenow, and A.B. Baron. 1973. Forced convection condensation inside tubes: A heat transfer equation for condenser design. *ASHRAE Transactions* 79(1):157–65

Varma, V.C. 1977. A study of condensation of steam inside a tube. Master's thesis, Department of Mechanical Engineering, Rutgers University, New Brunswick, NJ.

Wijaya, H., and M.W. Spatz. 1995. Two-phase flow heat transfer and pressure drop characteristics of R-22 and R-32/R125. *ASHRAE Transactions* 101(1):1020–27.

Yan, Y., and T. Lin. 1999. Condensation heat transfer and pressure drop of refrigerant R-134a in a small pipe. *Int. J. Heat Mass Transfer* 42:697–708.