Kn_{eff} is an effective Knudsen number, defined by

$$Kn_{eff} = \frac{2\tau_r}{a_p\sqrt{2\pi m_p/kT}} = \left(\frac{2a_p\rho_p kT}{27\pi^2\mu^2}\right)^{1/2} C_s/a_p \tag{A9}$$

and a is the probability that a particle-particle collision does not result in agglomeration. The probability a may be quite significant for very small particles possessing sufficient thermal energy to escape the attractive London-van der Walls force (Dahneke, 1976). Here we assume that the particles are sufficiently large for a to be nil.

The flux density on the wall is given by

$$j_k(\theta') = -D\frac{\partial n_k}{\partial r'}\bigg)_{r'=2a_p} = -\frac{Dn_{ok}A}{2a_p}\,\frac{\left(1-\frac{1}{2}St_{pk}\cos\theta'\right)}{(1-St_{pk}\cos\theta')} \tag{A10}$$

where St_{pk} is defined as

$$St_{pk} = \frac{2a_pU_k\rho_p}{9\mu}C_s \tag{A11}$$

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Forced Convective Boiling in Vertical Tubes for Saturated Pure Components and Binary Mixtures

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Over 1 000 sets of data for forced convective boiling of distilled water, ethylene glycol and aqueous mixtures of ethylene glycol are reported. Most of these data were taken in the annular flow regime. These data indicate a previously unrecognized Prandtl number effect on the boiling heat transfer for both pure components and mixtures. A significant reduction in the heat transfer coefficient is observed for mixtures attributable to mass transfer effects. An expression is developed which accounts for both of these effects and correlates the experimental data to within a mean deviation of 14.9%. This correlation reduces to the standard Chen correlation for pure fluids with Prandtl numbers close to unity.

SCOPE

Internal forced convective boiling of high Prandtl number liquids and mixtures of liquids is commonly encountered in industry. Previous studies in pure component heat transfer have concentrated on lower Prandtl number liquids, especially water. Furthermore, there are essentially no published data or correlations to account for any effects resulting from the forced convective boiling of mixtures. The objectives of this study were to obtain forced convective boiling heat transfer data for a relatively high Prandtl number liquid and for aqueous mixtures, and to develop a correlation based on these data which hopefully is applicable to other pure components and binary mixtures.

CONCLUSIONS AND SIGNIFICANCE

The results from over 1 000 sets of data are reported for forced convective boiling of distilled water, ethylene glycol and aqueous solutions of ethylene glycol. These results are not correlated well by the standard Chen correlation. The deviations are attributed to an unrecognized Prandtl number effect and a significant reduction in the heat transfer coefficient resulting from mixture effects. The proposed correlation retains the two heat transfer mechanisms postulated by Chen. A modified Reynold's analogy is used which accounts for the Prandtl number effect. For mixtures, the heat transfer coefficient associated with the local turbulence induced by bubble nucleation and evaporation is modified by the Florschuetz and Khan expression. For this datum base, the heat transfer associated with the bulk movement of the vapor and liquid dominates. Mixtures effects for this mechanism are correlated by postulating local composition profiles in the liquid, near the evaporating layer. The proposed expressions correlate the data to within a mean error of 15%. For pure components with Prandtl numbers around unity, the proposed correlation reduces to the standard Chen correlation.

The improved predictive capability of this correlation was not obtained from empirically adjusting correlation coefficients but was obtained by extending phenomenological models. Therefore, it is hoped that this correlation may be generally applicable to other systems.

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BACKGROUND

Many experimental works have been reported concerning internal forced convective boiling of single-component fluids. Some of the original works are Dengler and Addoms (1956), Bennett et al. (1959), Guerrieri and Talty (1956), Sani (1960), Shrock and Grossman (1962), Write (1961), Davis and David (1964) and Collier et al. (1964). In 1966, Chen showed that no available correlation predicted all the data and developed a general correlation based on the data reported in the first six references cited above. Chen's correlation is

$$h_{2\phi} = h_{Lonly} \cdot F(1/XTT) + 0.00122 \frac{k_L^{0.79} C \rho_L^{0.45} \rho_L^{0.49} g_c^{0.25}}{\sigma^{0.5} \mu_L^{0.29} h_{fg}^{0.24} \rho_v^{0.24}} \Delta T_S^{0.24} \cdot \Delta P_S^{0.75} \cdot S(Re_{2\phi}) \quad (1)$$

The Chen correlation is based on two postulated heat transfer mechanisms. The heat transfer associated with the bulk movement of the vapor and liquid was called macroscopic heat transfer, and its contribution to the total heat transfer coefficient is given by the first term of Equation (1). The heat transfer associated with the turbulence induced by the conception, growth and departure of vapor bubbles was called microscopic heat transfer, and its contribution to the total heat transfer coefficient is given by the last term of Equation (1). Chen used the data from the first six cited references to empirically evaluated $F(1/XTT)$ and $S(Re_{2\phi})$. Chen also showed through a Reynold's analogy that $F(1/XTT)$ was equal to $[(dP/dz)_{2\phi}/(dP/dz)_{Lonly}]^{0.444}$. Using the Lockhart Martinelli correlation for the pressure drop for internal annular flow, Chen derived

$$F(1/XTT) = (\phi_{Ltt})^{0.89} \quad (2)$$

This expression was shown to be essentially equivalent to the empirical F function for values of $1/XTT$ greater than about 5 but to diverge for smaller values.

Many papers have discussed the effect of mass transfer on boiling of multicomponent fluids. The effect of mass transfer on the growth of uniformly superheated vapor bubbles has been recognized since the theoretical work of Scriven (1959) and the experimental and analytical work of Florschuetz and Khan (1970). The effect of mass transfer during pool boiling of binary mixtures has been discussed by Calus and Leonidopoulous (1974), Shock (1977), Write et al. (1971), Calus and Rice (1972), Van Stralen and Sluyter (1969) and Chou (1975). The heat transfer of subcooled boiling over a heated tube has been discussed by Kadi (1975). Film boiling of binary mixtures has been discussed by Calus and Leonidopoulous (1974) and Marschall and Maresco (1977), Van Stralen and Joosen (1972) and Yue and Weber (1973). The only recent paper in the general literature concerning mass transfer effects for internal forced convective boiling is a brief Russian paper, Zizyulcin and Aerov (1975).

EXPERIMENTAL RESULTS

The experimental apparatus used in this study is illustrated in Figure 1. The test section has two heated parts. One part, the test section preheater, brings the subcooled liquid, which flows inside the inconel heating tube, to the desired quality. This section has three electrically heated, 0.36 m long, chambers. The saturated, two-phase mixture passes through a final heating tube which is the primary test section. The unshorted length of the inconel tubing between the 0.10 m (4.0 in.) diameter copper flanges is 0.0254 m (1.0 in.).

Each flange of the primary test section has an electrically isolated pressure transducer. A sheathed ungrounded thermocouple penetrates each flange and measures the fluid stream temperature. The outer surface of the unshorted section of the primary test section inconel tube is covered with a thin coat of electrically insulating high temperature epoxy. A coated thermocouple is cemented at the edge of the inconel tube and the flange. Another coated thermocouple is mounted at the center line of the tubing. Each thermocouple is wrapped 0.02 m (0.75 in.) circumferentially around the tube to minimize conduction through the thermocouple.

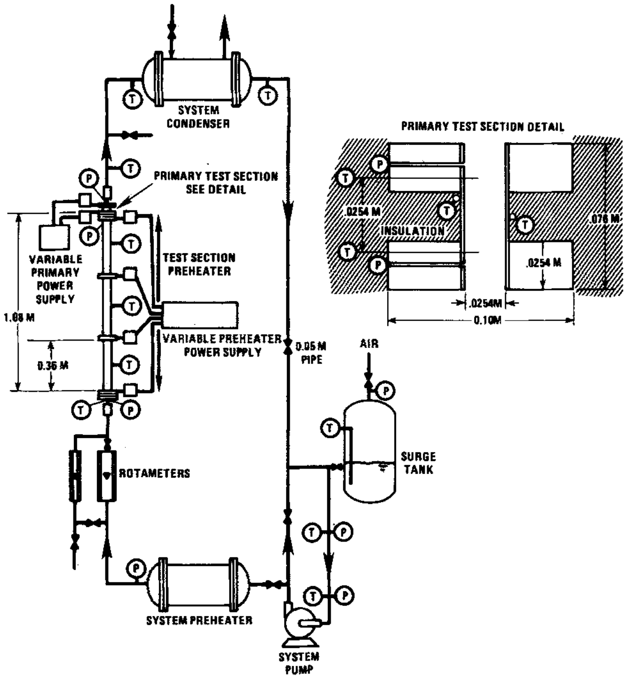

Figure 1. Experimental apparatus.

All instruments were independently verified. Total heat loss to the surroundings was computed to be negligible and verified by a heat balance with an average agreement within 0.04%. The experimental integrity of the equipment was verified by measuring both single- and multiphase heat transfer coefficients and comparing these values to values predicted from accepted correlations. Agreement was good, and these tests were documented by Bennett (1976).

The test loop was filled by evacuating the system of air and filling from the surge tank. The flow rate was adjusted to the desired value and its temperature increased by the system preheater and test section preheater. The test section power supply was adjusted to give a heat flux of about 15×10^5 W/m^2. After the system reached equilibrium, the temperatures, pressures and flow rate were recorded. The power supply was then lowered to give a heat flux of about 7×10^5 W/m^2. As the heat flux was increased, the original heat flux of 15×10^5 W/m^2 was repeated.

Heat transfer coefficients were computed from the heat flux through the primary test section heating tube and the difference of the computed inside surface tube temperature, based on measured outer wall temperatures, and the measured liquid bulk temperature. The heat flux was calculated by subtracting the computed heat loss from the unshorted length of the test section to the copper flanges from the IR^2 value. This correction was almost negligible for all sets of data. The quality was calculated by measuring the inlet temperature to the system preheater and assuming conservation of energy, conservation of species and assuming phase and thermal equilibrium.

The primary experimental parameters were composition, mass flux, heat flux and quality. The compositions tested were binary mixtures of water and ethylene glycol and were selected so that most of the experimental data were taken where the greatest composition effect was expected. The secondary experimental parameters were those parameters that exist at the primary test section as a result of the specific combination of mass flow rate, quality, composition and heat flux. These parameters include the heat transfer coefficient, the Martinelli parameter, the single-component Chen suppression factor and the effective or two-phase Reynolds number. The range of primary and secondary experimental parameters is given in Table 1.

Over one thousand sets of data were obtained. All of the experimental data are plotted in Figure 2a against the calculated

TABLE 1. RANGE OF PRIMARY AND SECONDARY EXPERIMENTAL PARAMETERS

- Range of primary experimental parameters:

Composition:	0 to 99.7% by mass ethylene glycol
Mass flux:	0.16 to 1.6 × 10^3 kg/m^2-s
Heat flux:	7.0 to 30.0 × 10^5 Watts/m^2
Quality:	near zero to 30%

- Range of secondary experimental parameters:

Heat transfer coefficient:	8 to 40 × 10^3 Watts/m$^{2.0}$K
Martinelli parameters:	0.16 to 300
Single component:	0.5 to 0.1
$RE_{2\Phi}$:	9.5 to 60 × 10^5

heat transfer coefficient based on Chen's correlation using the empirical F function [Equation (1)]. As seen from Figure 2*b*, better agreement is obtained when the calculated heat transfer coefficient is based on the Chen correlation using the F function derived from the Reynolds analogy [Equation (2)]. This is due to the divergence of Equations (1) and (2) when $1/X_{tt}$ is less than about 5. Although mass transfer effects could account for much of the existing scatter, the pure ethylene glycol data are the most poorly correlated. Figure 3 plots the normalized experimental heat transfer coefficient vs. the Martinelli parameter for pure ethylene glycol and pure water. The large difference between the data for these two pure components is only partially due to differences in the microscopic boiling contribution. The following development indicates that an additional Prandtl number effect is evident for fluids with high Prandtl numbers, for example, ethylene glycol.

CORRELATION DEVELOPMENT

Extension of *F* Function to Higher Prandtl Number Liquids

Following Chen's (1966) application of the Reynold's analogy to two-phase flow

$$\frac{q_W}{T_W - T_B} = + \frac{\tau_W\, g_c\, C_p}{\int_0^{V_B} \bar{D}\, du} = h \qquad (3)$$

For the two-phase case

$$h_{\text{mac}} = \frac{\tau_{W2\Phi}\, g_c\, C_{pL}}{\int_0^{V_{B2\Phi}} \bar{D}_{2\Phi}\, du}$$

For the case when the liquid fraction is flowing alone

$$h_{L\text{only}} = \frac{\tau_{WL\text{only}}\, g_c\, C_{pL}}{\int_0^{V_{L\text{only}}} \bar{D}_{L\text{only}}\, du}$$

Therefore, normalizing h_{mac} by $h_{L\text{only}}$ gives the following relationship:

$$\frac{h_{\text{mac}}}{h_{L\text{only}}} = \frac{\tau_{W2\Phi}}{\tau_{WL\text{only}}} \frac{\int_0^{V_{L\text{only}}} \bar{D}_L\, du}{\int_0^{V_{2\Phi}} \bar{D}_{2\Phi}\, du} = \left[\frac{Re_{2\Phi}}{Re_{L\text{only}}}\right]^{0.8} = F$$

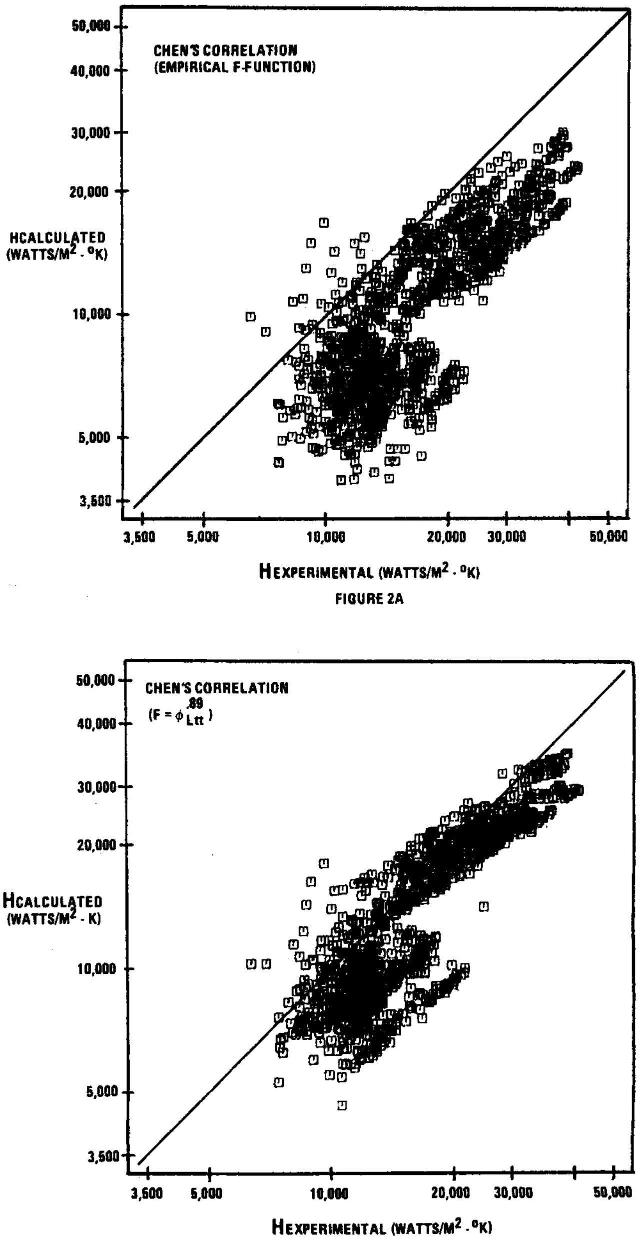

Figure 2a & 2b. Comparison of all the experimental data to predictions from the Chen correlation.

If we use the relationship between shear stress and pressure drop, the expression for F becomes

$$F = \frac{\left.\frac{dP}{dz}\right|_{2\Phi}}{\left.\frac{dP}{dz}\right|_{L\text{only}}} \frac{\int_0^{V_{L\text{only}}} \bar{D}_L\, du}{\int_0^{V_{2\Phi}} \bar{D}_{2\Phi} du}$$

In Chen's original solution, the standard Reynold's approximation that $\bar{D}$ is approximately equal to 1 is used to evaluate the F function. This approach is valid for fluids with a Prandtl number of 1. A more general approach is taken here.

When nucleation is present, any viscous sublayer would be continuously broken up by the formation, growth and departure of the vapor bubbles from the heating surface and through the liquid film. Therefore, it is assumed here that when nucleation is present, $\bar{D}_{2\Phi}$ is equal to the bulk stream value throughout the boundary layer. This simplifies the above equation to

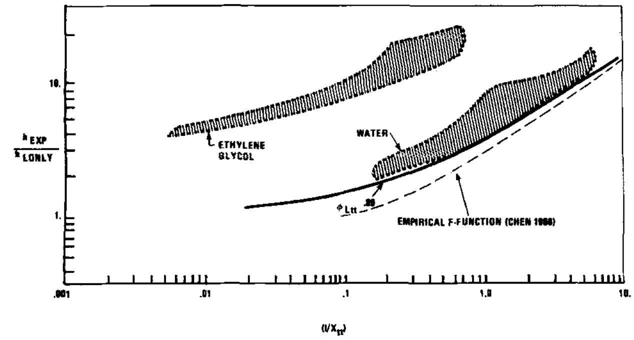

Figure 3. Experimental heat transfer coefficients for pure components.

$$F = \frac{\left.\frac{dP}{dz}\right]_{2\phi}}{\left.\frac{dP}{dz}\right]_{Lonly}} = \frac{\int_0^{V_{Lonly}} \bar{D}_L \, du}{Pr_{t_{2\phi}} V_{2\phi}} = \left[\frac{Re_{2\phi}}{Re_{Lonly}}\right]^{0.8} \tag{4}$$

Several solutions for the F function can be obtained depending on the single-phase boundary layer model.

Prandtl-Taylor Assumptions If we use the Prandtl and Taylor assumption of a viscous sublayer for single-phase flow, the relationship for F becomes

$$F = \left[\frac{\frac{u_1}{V_{Lonly}}[Pr_L - Pr_{t_{Lonly}}] + Pr_{t_{Lonly}}}{Pr_{t_{2\phi}}}\right]^{0.444} \left[\frac{\left.\frac{dP}{dz}\right|_{2\phi}}{\left.\frac{dP}{dz}\right|_{Lonly}}\right]^{0.444}$$

Using the assumption that the laminar sublayer ends at the location defined as u^+ equals y^+ equals 5, and using the definitions for u^+ and for the skin friction coefficient, we obtain

$$F = \left[\frac{5\sqrt{\frac{1}{2} C_f}\,[Pr_L - Pr_{t_{only}}] + Pr_{t_{Lonly}}}{Pr_{t_{2\phi}}}\right]^{0.444} \left[\frac{\left.\frac{dP}{dz}\right|_{2\phi}}{\left.\frac{dP}{dz}\right|_{Lonly}}\right]^{0.444} \tag{5a}$$

First Alternate Approach: Another approach is taken here which is similar to the Prandtl-Taylor approach but allows a continuous variation in the ratio of the total momentum and total thermal diffusivities $\bar{D}$. Assume that $\bar{D}$ is a continuous functional relationship and is only a function of the velocity. Two values of $\bar{D}$ are known; when the axial velocity is equal to zero (at the wall), $\bar{D}$ becomes equal to the Prandtl number. When the axial velocity is equal to the bulk velocity (in the turbulent core), $\bar{D}$ becomes equal to the turbulent Prandtl number. If it is assumed that within the region of interest a linear relationship can be used, these boundary conditions yield for the boundary layer

$$\bar{D}_{Lonly} = Pr_{Lonly}\left[1 - \frac{u}{V_{BLonly}}\right] + Pr_{t_{only}} \frac{u}{V_{BLonly}}$$

Substituting this into Equation (4) and assuming the turbulent Prandtl number, we get

$$F = \left[\frac{Pr_L + 1}{2}\right]^{0.444} \left[\frac{\left.\frac{dP}{dz}\right|_{2\phi}}{\left.\frac{dP}{dz}\right|_{Lonly}}\right]^{0.444} \tag{5b}$$

Second Alternate Approach Another expression for the F function can be derived from analysis and Colburn's empirical heat transfer coefficient adjustment for higher Prandtl number liquids.

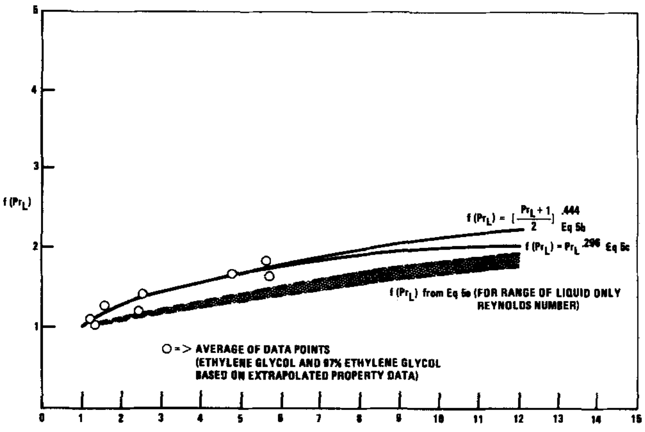

Figure 4. Different formulations of the Prandtl number effect compared to averaged data.

If we use the functional relationship for D that was proposed for the first alternate approach, Equation (3) and the definition of the friction factor, the single phase heat transfer coefficient is given by

$$St_{Lonly}\left[\frac{Pr_L + 1}{2}\right] = \frac{f}{8} \tag{5a}$$

The classical Reynold's analogy approach which is valid for fluids with Prandtl numbers equal to unity is

$$St_{Lonly} = \frac{f}{8}$$

In 1933, Colburn suggested an empirical correction for this equation for fluids with Prandtl numbers not equal to unity. His relationship is

$$St_{Lonly} \, Pr_L^{2/3} = \frac{f}{8} \tag{5b}$$

By analogy with expression a, the following expression for the F function which is consistent with the Colburn assumption is

$$F = Pr_L^{0.296}\left[\frac{\left.\frac{dP}{dz}\right]_{2\phi}}{\left.\frac{dP}{dz}\right]_{Lonly}}\right]^{0.444} \tag{5c}$$

To evaluate the various expressions for F, the microscopic component of the heat transfer coefficient was computed by the standard Chen equation and subtracted from the experimental heat transfer coefficient. The remainder was assumed to be the experimental macroscopic heat transfer coefficient. This number was used to converge on the observed Prandtl number effect. In Figure 4, average values of the observed Prandtl number effect are shown, as are plots of Equations (5a), (5b), and (5c). Although the format of the Prandtl number dependency is different, the results are similar. Over the Prandtl number range from 1 to 6, the average difference between results using (5b) and (5c) are −0.13%. Over the Prandtl number range from 1 to 10, the average difference between results using (5b) and (5c) is +2.23%. Equations (5b) and (5c) diverge slightly at higher Prandtl number values, and additional data in this region would be required to ascertain the more accurate expression.

Mass Transfer Effects on Forced Convective Boiling Heat Transfer

Forced convective boiling heat transfer can be characterized by two heat transfer mechanisms. It is postulated that for mixtures, each of these mechanisms is affected in separate ways by mass transfer.

Mass Transfer Effects on the Microscopic Heat Transfer Mechanism: Foster and Zuber assumed for pool boiling that the vapor bubbles grow in a region of relatively constant superheat. Therefore, the governing temperature potential is equal to the wall superheat. This is reasonable because for the pool boiling case, the temperature gradient is fairly small in the zone where the bubble is being generated.

In forced convective boiling, the thermal gradient in the vapor generating zone near the wall can be significantly greater. Chen suggested the use of an effective superheat which accounts for the variation of the interfacial temperature in the vapor generation zone near the wall. Chen postulated that the effective superheat divided by the wall superheat was equal to a suppression factor which is a function of the two-phase Reynold's number. Chen defined the suppression factor such that

$$T_E - T_B = S(Re_{2\Phi})^{1/.99}\,(T_W - T_B)$$

This expression is limited to single component liquids. Analytical and experimental works by Scriven (1959) and by Florshuetz and Khan (1970) show that for binary liquids, the interfacial temperature of the vapor bubble is affected by mass transfer. For the case of constant superheat with binary mixtures and assuming only molecular transport of mass, heat and momentum, the thermal driving potential was determined by Florshuetz and Khan to be approximately

$$\Delta T_b = \frac{\Delta T_S}{1 - \dfrac{C_{pL}(F_V - F_L)}{h_{fg}} \dfrac{\partial T_{\text{sat}}}{\partial F_L} \left[\dfrac{\alpha}{D}\right]^{1/2}}$$

A reasonable expression for the thermal driving potential governing the growth of a vapor bubble during the forced convective boiling of a liquid is

$$\Delta T_{TE} = \frac{1}{1 - \dfrac{C_{pL}(F_V - F_L)}{h_{fg}} \dfrac{\partial T_{\text{sat}}}{\partial F_L} \left[\dfrac{\alpha}{D}\right]^{1/2}} \cdot S\,(Re_{2\Phi})^{1/.99}\,(T_W - T_B)$$

This expression reduces to the Chen expression for a single-component liquid and reduces to the Florshuetz and Khan solution if there exists a uniform interfacial temperature profile in the vapor generation zone.

Therefore, for a binary mixture, the microscopic heat transfer coefficient for forced convective boiling is given by

$$h_{\text{mic}} = 0.00122\,\frac{k_L^{0.79} C_{pL}^{0.45} \rho_L^{0.49} g_c^{0.25}}{\sigma^{0.5} \mu_L^{0.29} h_{fg}^{0.24} \rho_v^{0.24}}\,(\Delta T_{TE})^{0.24}\,(\Delta P_{TE})^{0.75}$$

Analogous to Chen's single component derivation, a binary suppression factor can be defined as

$$S_{\text{Binary}} = \left[\frac{\Delta T_{TE}}{T_W - T_B}\right]^{0.99}$$

If we follow Chen's procedures and use the Clausius Clapeyron equation, the expression for the microscopic heat transfer coefficient for binary mixtures becomes

$$h_{\substack{\text{mic}\\\text{Binary}}} = 0.0012\,\frac{k_L^{0.79} C_{pL}^{0.45} \rho_L^{0.49} g_c^{0.25}}{\phi^{0.5} \mu_L^{0.29} h_{fg}^{0.24} \rho_v^{0.24}}\,(\Delta T_S)^{0.24} \quad (\Delta P_S)^{0.75} S_{\text{Binary}}$$

where

$$S_{\text{Binary}} = \frac{1}{1 - \dfrac{C_{pL}(F_V - F_L)}{h_{fg}} \left.\dfrac{\partial T_{\text{sat}}}{\partial F_L}\right|_{P_B} \left[\dfrac{\alpha}{D}\right]^{1/2}}\,S\,(Re_{2\Phi})$$

It should be noted that this datum base is dominated by the macroscopic heat transfer mechanism; therefore, this study did not allow a sensitive test of the proposed mixture effects on nucleate boiling during forced convective boiling. However, some additional data taken with another system have provided a more sensitive test, and agreement with this equation appears reasonable.

Mass Transfer Effects on the Macroscopic Heat Transfer Mechanism It is postulated that mass transfer does not affect the heat transfer coefficient but affects the driving force. Therefore

$$\dot{q}_{\substack{\text{mac}\\\text{1-comp}}} = h_{\text{mac}}\,(T_W - T_B)$$

and

$$\dot{q}_{\substack{\text{mac}\\\text{Binary}}} = h_{\text{mac}}\,(T_W - T_i)$$

The ratio of the thermal driving force for binary liquids to the single component driving force is denoted by

$$\left[\frac{\widetilde{\Delta T}}{\Delta T_s}\right]_{\text{mac}} = \frac{T_W - T_i}{T_W - T_B}$$

If it is desirable to maintain $(T_W - T_B)$ as the driving force, then a binary heat transfer coefficient can be defined as

$$h_{\substack{\text{mac}\\\text{Binary}}} = \frac{\dot{q}_{\text{mac}}}{T_W - T_B}$$

From this definition, it can be shown that

$$h_{\substack{\text{mac}\\\text{Binary}}} = h_{\substack{\text{mac}\\\text{1-comp}}} \left[\frac{\widetilde{\Delta T}}{\Delta T_s}\right]_{\text{mac}}$$

It is postulated that T_i is equal to the equilibrium temperature of the interface between the liquid film and vapor core and is a function of the local interface composition. Therefore

$$\left[\frac{\widetilde{\Delta T}}{\Delta T_s}\right]_{\text{mac}} = \frac{T_W - T_{\text{sat}}\,(P_B, F_{LS})}{T_W - T_{\text{sat}}\,(P_B, F_L)}$$

A mass transfer coefficient can be defined as

$$j = \rho_{\text{avg}}\,h_m\,(F_{LS} - F_L)$$

If we assume that the vapor is generated in equilibrium with the liquid composition F_L, and if essentially all of the heat flux results in evaporation at the interface,* a mass and energy balance yields

$$j = \frac{(1 - F_V)\,\dot{q}}{h_{fg}}$$

If we use the definition of the mass transfer coefficient, the solution for the mass fraction of the less volatile component at the interface is

$$F_{LS} = F_L + \frac{(1 - F_V)\,\dot{q}}{h_{fg}\,h_m}$$

If $\partial T_{\text{sat}}/\partial F_L$ is assumed constant over this concentration change, the macroscopic heat transfer correction factor for binary mixtures can be determined from

$$\frac{\Delta T}{\Delta T_{S_{\text{mac}}}} = 1 - \frac{(1 - F_{VB})\,\dot{q}}{\rho_{\text{avg}}\,h_{fg}\,h_m \Delta T_S} \left.\frac{\partial T_{\text{sat}}}{\partial F_L}\right|_{P_B}$$

The only unknown is the mass transfer coefficient h_m. By dimensional analysis, the following is expected:

$$Sh_d = C_1\,(Re_{2\Phi})^{C_2}\,Sc^{C_3}$$

By analogy with heat transfer, C_2 is assumed to be 0.8, and C_3 is assumed to be 0.4. The coefficient C_1 was experimentally evaluated from the data by selecting the value which gave the minimum mean deviation. Equation (5*b*) was used for the Prandtl number effect. The best value for the coefficient C_1 was determined to be 0.023. These data are plotted in Figure 5. It is very interesting that the value of the coefficient for heat transfer data is given by the Dittus-Boelter correlation to be 0.023.

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* See appendix.

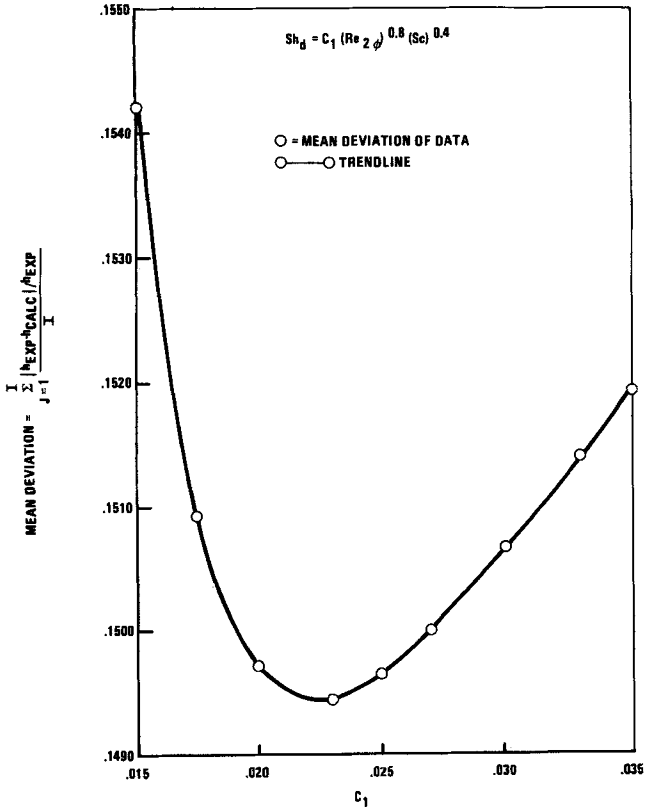

Figure 5. Mean deviation of equation 6 vs. coefficient for mass transfer correlation.

Summary Statement of the Correlation for Binary Mixtures:

$$h_T = \left[0.023\,(Re_{L\text{only}})^{0.8}\,Pr_L^{0.4}\,\frac{k_L}{D_I}\right]\left[\frac{\left.\frac{dP}{dz}\right|_{2\phi}}{\left.\frac{dP}{dz}\right|_{L\text{only}}}\right]^{0.444} f(Pr_L)\left[\frac{\widetilde{\Delta T}}{\Delta T_s}\right]_{\text{mac}} + 0.00122\,\frac{k_L^{0.79}C_{pL}^{0.45}\rho_L^{0.49}g_c^{0.25}}{\sigma^{0.5}\mu_L^{0.29}h_{fg}^{0.24}\rho_v^{0.24}}(\Delta T_S)^{0.24}\,(\Delta P_S)^{0.75}S_{\text{Binary}}(Re_{2\phi}) \qquad (6a)$$

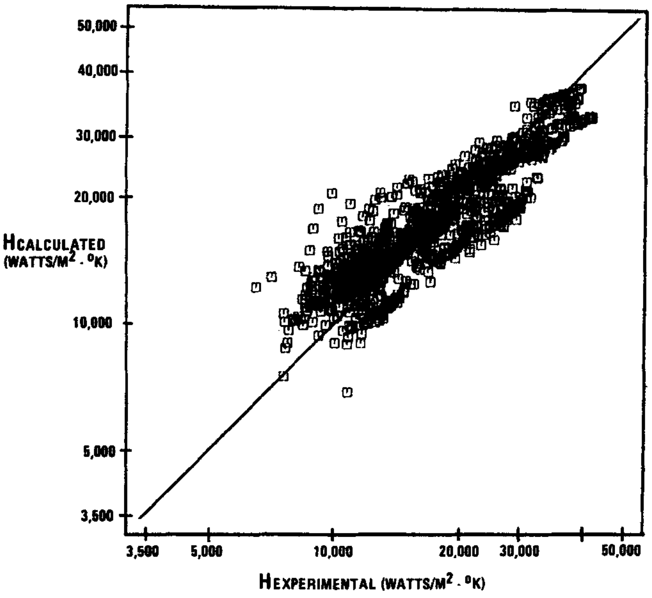

Figure 6. Comparison of all the experimental data to predictions from equation 6.

where

$$S_{\text{Binary}}(Re_{2\phi}) = \frac{1}{1 - \frac{C_{pL}\,(F_V - F_L)}{h_{fg}}\,\frac{\partial T}{\partial F_L}\left[\frac{\alpha}{D}\right]^{1/2}}\,S(Re_{2\phi}) \qquad (6b)$$

$$f(Pr_L) = \left[\frac{Pr_L + 1}{2}\right]^{0.444} \text{ is preferred, or}$$
$$= Pr_L^{0.296}$$

$$\left[\frac{\widetilde{\Delta T}}{\Delta T_s}\right]_{\text{mac}} = 1 - \frac{(1 - F_V)\,\dot{q}}{\rho_{\text{avg}}\,h_{fg}\,h_m\,\Delta T_S}\,\left.\frac{\partial T_{\text{sat}}}{\partial F_L}\right|_{P_B} \qquad (6c)$$

and

$$h_m = 0.023\,(Re_{2\phi})^{0.8}(Sc)^{0.4}\,D/DI \qquad (6d)$$

$$Re_{2\phi} = Re_{L\text{only}}\left[f(Pr_L)\cdot\left[\frac{\left.\frac{dP}{dz}\right|_{2\phi}}{\left.\frac{dP}{dz}\right|_{L\text{only}}}\right]^{0.444}\right]^{1.25} \qquad (6e)$$

TABLE 2. MEAN DEVIATION AND AVERAGE DEVIATION (ALL DATA)

Calculation Scheme	$\frac{\sum_{j=1}^{I}\|(h_{\text{exp}} - h_{\text{calc}})/h_{\text{exp}}\|}{I}$	$\frac{\sum_{j=1}^{I}[(h_{\text{exp}} - h_{\text{calc}})/h_{\text{exp}}]}{I}$
Chen's correlation with empirical F function (for pure components)	0.37557	0.36523
Chen's correlation with f function determined by Reynold's Analogy (for pure components)	0.20833	0.15997
This correlation, but for pure components (corrected for Prandtl number)	0.17259	0.10476
$\left(\frac{\widetilde{\Delta T}}{\Delta T_s}\right)_{mic} = \left(\frac{\widetilde{\Delta T}}{\Delta T_s}\right)_{mac} = 1$		
This total correlation including mass transfer effects.	0.14945	0.03612

TABLE 3. MEAN DEVIATION AND AVERAGE DEVIATION (PURE COMPONENTS)

	$\frac{\sum_{j=1}^{I}\|(h_{\text{exp}} - h_{\text{calc}})\|/h_{\text{exp}}}{I}$		$\frac{\sum_{j=1}^{I}(h_{\text{exp}} - h_{\text{calc}})/h_{\text{exp}}}{I}$	
Calculation scheme	H_2O	99.7% Ethylene glycol	H_2O	99.7% Ethylene glycol
Chen's correlation with empirical F function	0.32	0.55	0.31	0.55
Chen's correlation with F function determined by Reynold's analogy	0.13	0.46	0.09	0.44
This correlation, but for single components	0.10	0.176	0.0036	0.11

DISCUSSION

The experimentally evaluated heat transfer coefficients are plotted in Figure 6 against calculated values using the total correlation. Equation (5*b*) was used to define *F*, and the friction pressure drop gradient ratio was obtained from the Martinelli correlation (ϕ^2_{Ltt}). Table 2 reports, for various calculation schemes, the mean deviation and average deviation for all sets of data. The mean deviation has improved from 37 to 15%, and the average deviation has improved from +37 to −3.6%. Table 3 reports the mean deviation and average deviation for the pure components. The mean deviation has improved from 31 to 10% for the water data and from 55 to 17% for the ethylene glycol data. The average deviation has decreased from +31 to +0.4% for the water data and from +55 to +11% for the ethylene glycol data.

To predict the heat transfer coefficient, the best available correlation for the two-phase friction pressure drop should be used. If no better correlation is used, the Martinelli equation is suggested.

It is expected that this correlation would be valid for boiling binary mixtures anticipated to be in the annular and annular dispersed flow regimes. It should also be valid for boiling liquids with Prandtl numbers under saturated conditions between unity and 10. Furthermore, since this correlation reduces to the standard Chen correlation for pure liquids with Prandtl numbers around unity, it should be applicable for the same range of flow conditions as the Chen equation. The derivation extending the *F* function to higher Prandtl number liquids assumes the viscous sublayer is disrupted by vapor bubbles, and its effect is, therefore, negligible. All data accumulated for this study confirmed this assumption, but there may be cases (such as condensation) when this assumption is not valid. It can be shown that under these conditions, the Prandtl number effect reduces to unity.

The works of Sani (1960), Wright (1961) and Somerville (1962) provide an additional and very consistent test of the Prandtl number effect derived in this paper. These three works were conducted at the Lawrence Radiation Laboratory using essentially the same test loop. All studies concerned downward forced convective boiling. Sani's and Wright's studies utilized water. Somerville's study utilized *n*-butanol. Each author presented correlations in consistent formats which allow the separation of the heat transfer which the authors attributed to nucleation and convective mechanisms. The correlation for the macroscopic heat transfer coefficient for downward flow of saturated water proposed by Sani is

$$\frac{h_{\text{mac}}}{h_{L\text{only}}} = 2.22(1/X_{tt})^{2/3}$$

The correlation proposed by Wright is

$$\frac{h_{\text{mac}}}{h_{L\text{only}}} = 2.09(1/X_{tt})^{2/3}$$

These data suggest that nucleation is present, and, even though the direction of flow is downward, an equation of the format

$$\frac{h_{\text{mac}}}{h_{L\text{only}}} = C\left[\frac{Pr_L + 1}{2}\right]^{0.444} (1/X_{tt})^{2/3}$$

would be expected. If we use a typical value for the Prandtl number of saturated water, the correlations of Sani and Wright can be used to solve for the coefficient. The expression is

$$\frac{h_{\text{mac}}}{h_{L\text{only}}} = 1.99\left|\frac{Pr_L + 1}{2}\right]^{0.444} (1/X_{tt})^{2/3}$$

This expression agrees with the Sani and Wright correlations to within about ±3%. *n*-butanol has a Prandtl number of around 7 for the tested pressures. Therefore, the above equation would predict that the macroscopic heat transfer coefficient for downward boiling flow of *n*-butanol would be correlated by

$$\frac{h_{\text{mac}}}{h_{L\text{only}}} = 3.68(1/X_{tt})^{2/3}$$

This is in exact agreement with the correlation proposed by Somerville and supports the mechanisms proposed in this paper. Recent work has shown that the internal flow binary data for a cryogenic system supports this correlation and that freon data over a tube bank exhibit the same sort of interactive micro-macro mechanisms. The extension of the correlation to multicomponent systems is also underway.

NOTATION

C_f = skin friction coefficient = $\tau_W/(\rho_L V_B^2/2)$, evaluated for $Re_{L\text{only}}$ (1)
C_p = specific heat, J/kg·K
dP/dz = friction pressure drop gradient, N/m^3
D_I = inside diameter of heating tube, m
$\bar{D}$ = ratio of total momentum diffusivity to total thermal diffusivity (1)
D = molecular diffusion coefficient, m^2/s
f = friction factor = 4 C_f (1)
F = F function = $h_{\text{mac}}/h_{L\text{only}}$ = $(Re_{2\Phi}/Re_{L\text{only}})^{0.8}$ (1)
F_L = local liquid mass fraction of the less volatile component (ethylene glycol) (1)
F_{LS} = local liquid mass fraction of the less volatile component (ethylene glycol) at the liquid-vapor interface (1)
F_V = local vapor mass fraction of the less volatile component (ethylene glycol) (1)
g_c = proportionality constant required for Newton's second law
h_{fg} = latent heat of vaporization, J/kg
h_m = mass transfer coefficient, m/s
h_{mac} = single-component heat transfer coefficient due to macroscopic heat transfer mechanism, W/m^2 · K
h_{mic} = single component heat transfer coefficient due to microscopic heat transfer mechanism, W/m^2 · K
h = total heat transfer coefficient = $\dot{q}/(T_W - T_B)$, W/m^2 · K
j = mass flux of more volatile material (water), kg/m^2 · S
k_L = thermal conductivity of liquid mixture, J/s · m · °K
Pr = Prandtl number (1)
Pr_t = turbulent Prandtl number (1)
P_{sat} = saturation pressure, N/m^2
q = heat flux, W/m^2
$Re_{2\Phi}$ = effective two-phase Reynold's number (1)
S = pure component suppression factor as defined by Chen (1)
Sh_d = Sherwood number, $h_m D_I/D$ (1)
St = Stanton number, $Nu/Pr\ Re$ (1)
Sc = Schmidt number, $\mu/\rho D$ (1)
T = temperature, K
T_B = bulk temperature of liquid at primary test section, K
T_i = temperature at liquid vapor interface at primary test section, K
T_E = effective temperature of single-component zone where vapor bubble is growing, K
T_W = temperature at the wall-liquid interphase, K
u = axial velocity, a function of radial position, m/s
u_1 = velocity at the viscous sublayer, turbulent core transition, m/s
V_B = bulk velocity, m/s
x = quality, ratio of mass flow rate of vapor to total mass flow rate (1)
X_{tt} = Martinelli parameter = $(\mu_L/\mu_V)^{0.1}$ $(\rho_V/\rho_L)^{0.5}$ $(1 - x/x)^{0.9}$ (1)

Greek Letters

ρ = density, kg/m^3
σ = surface tension, n/m
α = thermal diffusivity, m^2/s

τ = shearing stress, N/m^2

ϕ_{Ltt} = defined by Martinelli, $[dP/dz_{2\phi}/dP/dz_{\text{Lonly}}]^{1/2}$, is a function of X_{tt} (1)

ΔT_S = superheat, $T_W - T_{\text{sat}}$, K

$\widetilde{\Delta T}/\Delta T_{S_{\text{mac}}}$ = ratio of binary macroscopic heat transfer coefficient to the heat transfer coefficient evaluated for mixture properties, but neglecting mass transfer effects

ΔP_S = change in saturation pressure determined from ΔT_S, N/m^2

μ = viscosity, N · s/m^2

Subscripts

2ϕ = two phase

L = liquid fraction

Lonly = computed for the liquid fraction flowing alone

TE = true effective, takes into account both flow and composition effect

v = vapor mixture

W = wall

exp = experimental

calc = calculated

MV = more volatile

sat = saturated

M.C. = multicomponent

MAC = macroscopic

MIC = microscopic

1-comp = single component fluid

Binary = binary fluid

AVG = liquid mixture

B = bulk conditions

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APPENDIX: MODIFICATION OF THE MACROSCOPIC MASS TRANSFER EFFECT WHEN THE INCREASE IN LATENT HEAT OF THE LIQUID IS SIGNIFICANT

The derivation of the effect of mass transfer on the macroscopic heat transfer mechanism assumes that all of the heat flux results in evaporation at the interface. For some systems, a major portion of the heat flux is utilized to increase the temperature of the boiling fluid. For these systems, Equation (6c) can overestimate the mass transfer related reduction in the heat transfer coefficient. To account for this effect, Equation (6c) can be modified.

If we assume that the specific heat of the vapor is negligible compared to the specific heat of the liquid, the total heat flux can be expressed as the sum of the flux causing evaporation and flux resulting in a temperature rise of the liquid. Therefore

$$\dot{q}_{\text{total}} = \dot{q}_{\text{vap}} + \dot{q}_{\text{lat}}$$

For a differential control volume with circular cross section, this equation is equivalent to

$$\dot{q}_{\text{total}} = j_T h_{fg} + \frac{1}{\pi D_I} C_{pL} \dot{M}_L \frac{dT_{\text{sat}}}{dL} \tag{A1}$$

where j_T is the total evaporating mass flux, and $\dot{M}_L$ is the liquid flow rate through the heating tube. By the chain rule

$$\frac{dT_{\text{sat}}}{dL} = \frac{dT_{\text{sat}}}{dF_L} \times \frac{dF_L}{dL} \tag{A2}$$

By continuity of species and total mass, Equation (A2) can be evaluated and substituted into Equation (A1). After rearranging, we get

$$\dot{q}_T = J_T h_{fg} + C_{pL} \frac{dT_{\text{sat}}}{dF_L} [F_L - F_V]$$

Therefore, an effective heat of vaporization can be defined such that

$$h'_{fg} = \left[h_{fg} + C_{pL} \frac{dT_{\text{sat}}}{dF_L} [F_L - F_V] \right] \tag{A3}$$

The overall effect over the composition range of interest can be indicated by the integration of h'_{fg}/h_{fg}, or

$$\frac{\overline{h'_{fg}}}{h_{fg}} = \frac{1}{(F_{L_2} - F_{L_1})} \int_{F_{L_1}}^{F_{L_2}} \left[1 + \frac{C_{pL}}{h_{fg}} \frac{dT_{\text{sat}}}{dF_L} (F_L - F_V) \right] dF_L \tag{A4}$$

For the data base reported in this paper, the peak value of the argument of Equation (A4) is about 1.4, but the overall value for Equation (A4) for all the data is about 1.08. Utilization of the effective heat of vaporization does not decrease the error of this correlation and is, therefore, not warranted when Equation (A4) is near unity.

If the basic correlation [Equation (6)] is utilized, conservative (smaller) heat transfer coefficients will be estimated if h'_{fg}/h_{fg} is large, but h_{fg} is used as indicated in Equation (6c). This is due to the smaller volatile component flux at the interface resulting from a portion of the heat flux heating the liquid. It is recommended that if h'_{fg}/h_{fg} is less than 1.5, Equation (6) be used directly and h_{fg} be used in Equation (6c). If h'_{fg}/h_{fg} is greater than 1.5, it is suggested that a more realistic (but less conservative) coefficient may be calculated if the effective heat of vaporization [Equation (A3)] is used in Equation (6c) if the coefficient in Equation (6d) is reduced from 0.023 to 0.015. This value for the coefficient is the weighted average using h'_{fg} computed from a sample of the data for the three compositions which exhibited the largest experimental mass transfer effects.

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