Falling Film Evaporation of Binary Mixtures

J. W. Palen, Qi Wang, and J. C. Chen
Dept. of Chemical Engineering, Lehigh University, Bethlehem, PA 18015

The interactive effects of heat and mass transfer in evaporation of binary mixtures flowing as falling films on vertical surfaces was investigated. Evaporative heat-transfer coefficients were measured for aqueous mixtures of ethylene and propylene glycol, with boiling ranges up to 55°C. Tests were carried out at atmospheric pressure with heat fluxes ranging from 3,000 to 25,000 W/m^2, and film Reynolds numbers ranging from 300 to 3,000. Results indicated that the heat-transfer coefficient for mixtures depends weakly on wall superheat and film Reynolds number, but strongly depends on mixture composition. Analysis of the results indicates that mass-transfer resistance in the liquid film causes significant elevation of the interface temperature, causing a reduction of the effective temperature driving force. A semiempirical model for correlation of the interactive heat- and mass-transfer phenomena is proposed.

Introduction

Falling film evaporators are finding increasing use in the chemical process industries for separation of multicomponent mixtures where:

- Fluids are temperature-sensitive
- Low temperature difference driving force is required
- Low pressure drop is necessary.

For these situations, a falling film evaporator is advantageous in achieving minimal static head and pressure drop, minimal fluid residence time and high heat-transfer rates with low temperature differences. In spite of this increasing usage, little information is available in literature on the transport processes in falling film evaporation, particularly, data and predictive models for wide-boiling-range mixtures with large temperature differences between bubble and dew points (boiling range).

The state of the art for design of falling film evaporators was recently summarized in the *Handbook of Evaporation Technology* (Minton, 1986). Research work on falling film heat transfer and hydrodynamics is summarized by Seban (1978), Ganic (1981) and Ganic and Mastanaiah (1981). In another general survey, Smith (1983) cites the work of Whitt (1966) and recommends the correlations of Chun and Seban (1971, 1972) for single-component fluids. For mixtures, the recommendations of Smith (1983, 1986) are not clear-cut, but additional resistance from mass transfer is recognized. It was reported that evaporation rates with 14% brine were about 10% less than for water at the same temperature difference.

Since 1980, relevant work in vaporization of mixtures has been carried out by Schlünder and coworkers (Schnabel, 1980; Gropp, 1981). The original work of Schlünder (1982) applied the film theory to nucleate pool boiling. It was shown that significant mass-transfer effects in pool boiling could be accounted for by use of a heat-transfer factor which depends upon an empirically determined mass-transfer coefficient. Later, proprietary work by Heat Transfer Research, Inc., as reported by Sardesai, Palen and Thome (1986), confirms the applicability of this approach for nucleate pool boiling of wide-boiling-range hydrocarbon mixtures. It was not clear, however, whether the same approach would be required or could be applied for evaporation from falling films. In fact, Schlünder (1983) in a general review of methods for heat and mass transfer gives an order of magnitude argument for falling films which indicates that "in both phases the bulk and interface compositions are equal and in equilibrium."

The first systematic work in falling film vaporization of mixtures by Schlünder et al. was reported in Gropp, Schnabel and Schlünder (1983). That study used a mixture of R11 and R113 vaporizing on a vertical plate with longitudinal grooves. Average mass-transfer coefficients were obtained by measuring vapor and liquid compositions and comparing with calculated equilibrium values. In a later study, Gropp and Schlünder (1986) investigated falling film evaporation outside a vertical

---

Correspondence concerning this article should be addressed to J. C. Chen.

Current address of J. W. Palen: Heat Transfer Research, Inc., P.O. Box 10009, College Station, TX 77842.

Current address of Q. Wang: Rheem Manufacturing Company, P.O. Box 17010, Fort Smith, AR 72917.

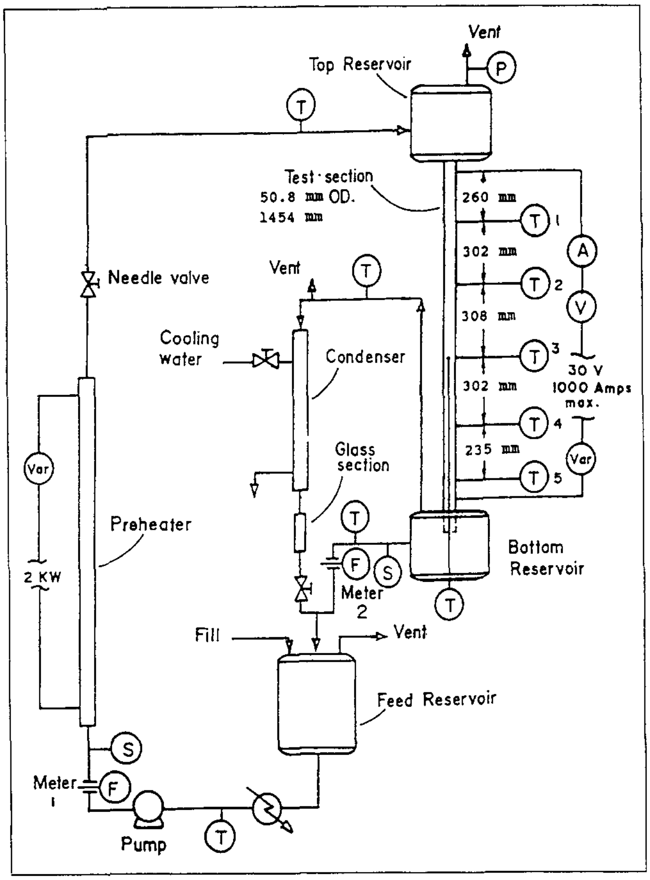

Figure 1. Experimental apparatus.

tube. The mixture was again R11/R113 (boiling range ≈4°C). It was found that for this mixture there was little effect of mass-transfer resistance at heat fluxes up to about 6,000 W/m^2. At these heat fluxes, no nucleate boiling was observed. However, as the heat flux was increased above 16,000 W/m^2, nucleate boiling was observed and the heat-transfer coefficient became noticeably lower than the expected value for pure components, indicating possible mass-transfer resistance. Gropp and Schlünder concluded that "during surface boiling the reduction of the heat-transfer coefficient is negligible for technical applications because of the minor deviations from evaporation mainly controlled by thermodynamic equilibrium . . during nucleate boiling . . . there is considerable deviation from evaporation controlled exclusively by thermodynamic equilibrium."

The fact that very little effect of mass transfer was noticed at the low heat fluxes for the above tests is not surprising. For this narrow-boiling-range mixture, mass-transfer effects would be expected to be less severe than for mixtures with wide boiling range. We suspected that the impact of mass-transfer effect upon heat transfer is not only a function of the type of heat-transfer mechanism (nucleate boiling or surface evaporation), but also a strong function of the boiling range of the mixture. There is a strong possibility that mass-transfer effects can have a significant effect on heat transfer in many practical cases of evaporation of mixtures in falling films. Yet, all conventional design methods ignore this effect and assume the vapor to be in equilibrium with the bulk liquid (VDI, 1977; Minton, 1986). Presently, there are no models in the literature which account for the potential effect of mass transfer on the heat-transfer coefficient for falling film evaporation of wide-boiling-range mixtures, nor are there representative data which could be used to develop such models.

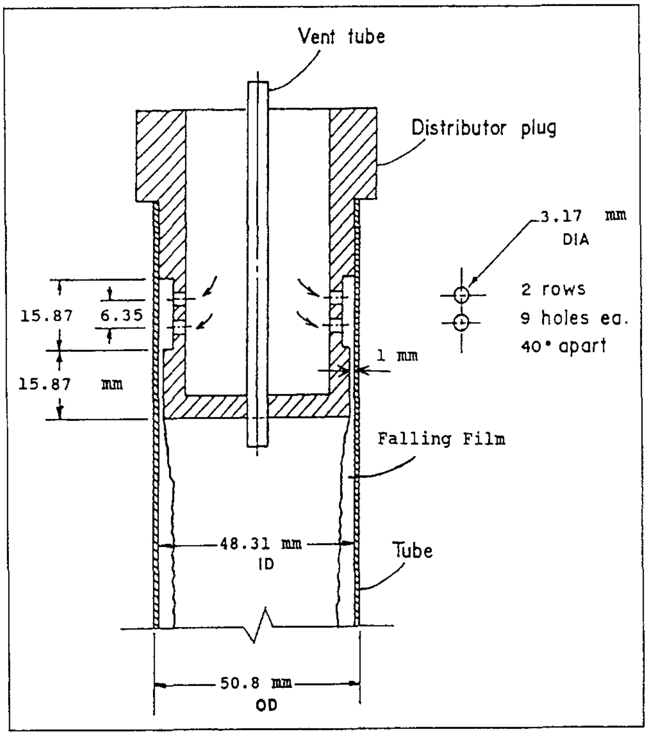

Figure 2. Film distributor.

The objective of this investigation was to obtain experimental measurements of the evaporative heat-transfer coefficients for films of mixtures with wide boiling ranges. These data could then be used to assess the mechanistic question of whether or not liquid-side mass-transfer effects interfacial evaporation in falling films.

Experiment

The experimental equipment is shown in Figure 1. All vessels and piping were stainless steel and insulated with fiberglass. The basic test element was an evaporator tube of 304 stainless steel with 50.8 mm OD, 1.24 mm wall thickness, and an overall length of 1.52 m. Direct Joule heating of the tube was achieved using a DC rectifier with a variable current output of 0 to 1,000 A over a voltage range of 0 to 30 V. Power was supplied to the tube through two copper electrodes spaced to provide a heated length of 1.45 m. The evaporator tube and connecting vessels were electrically isolated from the rest of the piping by Teflon fittings.

The test liquid was pumped from the feed reservoir and entered the top reservoir after being preheated close to saturation temperature. In the top reservoir the liquid flowed through a brass distributor (shown in Figure 2) to form an evenly distributed film (~2.4 mm typical thickness) on the inside surface of the tube. The falling film would be partly vaporized in the tube at a constant heat flux. The vapor-liquid mixture was separated in the bottom reservoir with the vapor flowing to a vertical condenser where it was totally condensed and returned to the feed tank, together with the unevaporated liquid from the test tube.

Figure 1 also indicates instrumentation used. The flow rate to the test section was regulated by a needle valve and measured by a calibrated rotameter. The tube wall temperatures were measured at intervals by thermocouples clamped to the outside wall. These were located over the heated length, as shown in Figure 1, with 90 degree spacings around the tube circumference. Heavy insulation around the tube minimized heat loss and prevented significant temperature gradients in the region of the thermocouple-wall interface. The inside wall temperatures were calculated from the OD temperatures, accounting for heat generation and conduction in the tube wall. The inside vapor bulk temperature was measured by a movable thermocouple which was centered in the tube by an electrically insulated "spider." This traveling thermocouple could be located adjacent to any of the wall thermocouple locations.

Note that the measurement of bulk vapor temperature was not necessary for calculation of the evaporative heat-transfer coefficient. By tradition, this coefficient is based on a bulk saturation temperature which is determined by equilibrium flash calculation. Therefore, care was required in the energy balance of the test section. For the test conditions, latent heat of phase change was dominant over the negligible sensible heat effect. We measured the vapor temperature just to obtain verification of the equilibrium state at system pressure.

Other important temperature measurements were: inlet fluid temperature to the top reservoir, which had to be maintained somewhat less than saturation, inlet fluid temperature to the pump, which was kept at least 15°C less than saturation to prevent cavitation and the vapor temperature to the condenser. All thermocouples were read directly from a digital 10 channel temperature logger. This instrument was calibrated at the ice point and checked at the boiling point. Consistency of the thermocouples was checked at room temperature and determined to be better than 0.5°C. Power to the test section was regulated by a rheostat on the DC rectifier unit, and determined from direct measurements of voltage and current.

Accuracy in measuring wall temperatures was of particular importance. We elected Joule heating of the wall in order to obtain accurately determined heat fluxes. However, this required special considerations in method of measuring ID wall temperature. Inaccuracies could be introduced by insertion of thermocouples in the wall due to disruption of either thermal or electrical fields in the local zone. Our choice was to heavily insulate at the OD of the wall, and place electrically-isolated thermocouples on the OD surface. This places the thermocouple junction on an adiabatic surface, and there was no temperature drop between the junction and the wall OD. The calculation of the ID temperature was then just a simple calculus exercise.

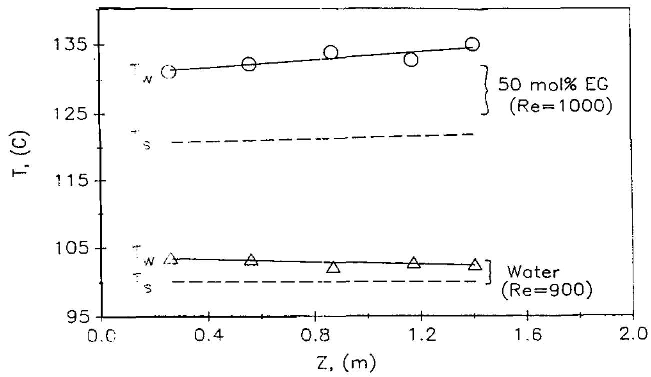

Figure 3. Examples of axial temperature profiles.

Fluid samples were taken at the inlet to the pump and at the outlet of the bottom reservoir. The compositions were determined from refractive index. The system was always degassed by venting for approximately 30 min before taking data. In a typical run, the selected fluid mixture would be introduced to the top reservoir at close-to-saturation temperature and flow rate adjusted to obtain a desired film Reynolds number ($4\Gamma/\mu$). Heat flux to the test section was established, and all data then recorded upon attaining steady-state conditions. Wall temperatures were kept sufficiently low to avoid bubble nucleation. Uncertainty in the resulting heat-transfer coefficient was determined from error analysis to be approximately 5%.

Results and Discussion

Over a hundred experimental runs were carried out using water (pure fluid), and various mixtures of water with propylene and ethylene glycol. For the binary mixtures, the boiling range (difference between bubble and dew points) varied up to a maximum of 55°C. All tests were carried out at atmospheric pressure. Heat fluxes ranged from 3,000 to 25,000 w/m^2, and film Reynolds numbers ranged from 300 to 3,000. Some sample results are presented and discussed in this section.

Figure 3 shows axial temperatures along the length of the test tube at a heat flux of 15.5 kw/m^2 for one case of pure water film and a second case of 50 mole percent ethylene glycol solution. The axial coordinate Z is taken to be 0 at the top of the heated section. With negligible pressure drop, the bulk saturation temperature for water remained constant along the tube, but increased slightly with progressive evaporation of the volatile component for the case of ethylene glycol solution. The measured wall temperatures are shown by respective symbols in Figure 3, indicating the degree of temperature scatter in the measurements. Axial profiles of the wall temperatures were calculated by regression fit to the measured points, illustrated by the solid lines in Figure 3. Several qualitative points are worth noting from this figure. First, it is seen that the temperature difference ($T_w - T_s$) is noticeably smaller for water than for the water/glycol mixture, indicating a diminished heat-transfer coefficient for the mixture at essentially equal values of heat flux and Reynolds number. Second, the temperature difference in the case of water decreased very slightly along the length of the tube, but increased slightly for the case of the binary mixture. The first effect can be attributed to thinning of the water film due to evaporation, while the second effect is most likely associated with the increasing effect of mass transfer upon heat transfer as water was preferentially evaporated from the glycol solution.

The effective heat-transfer coefficient (HTC) for falling-film evaporation is commonly defined as:

$$h = \frac{q}{(T_w - T_s)} \tag{1}$$

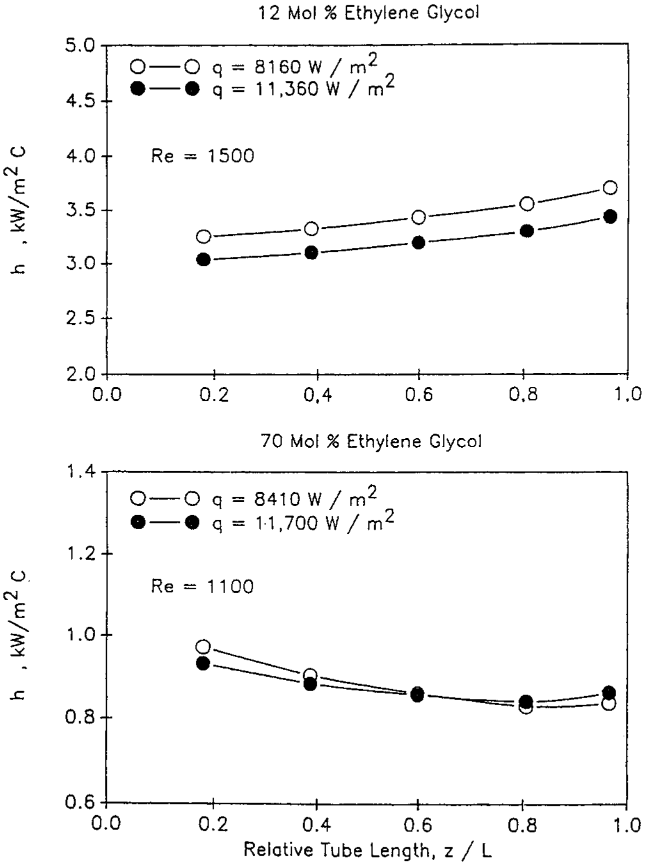

Figure 4. Local heat-transfer coefficients for two binary mixtures.

For the case of mixtures, the saturation temperature is defined as the bubble point for the local mixture bulk composition, as determined from energy balance and flash calculation. In this study, the experimentally measured profiles of T_w were used to determine effective heat-transfer coefficient. Figure 4 shows sample results, where local heat-transfer coefficients are plotted against axial position along the heated length (downward) for two compositions of ethylene glycol solution. The qualitative observations from Figure 3 are confirmed here, namely that increasing concentration of the less volatile component (glycol) caused:

a) Decrease in magnitude of the HTC

b) Decrease in heat-transfer coefficient along axial length.

We feel that both consequences are associated with mass-transfer effects in the liquid film, as discussed in the section on analysis.

Figure 5 shows the heat-transfer coefficients at a given axial position ($Z/L = 0.6$) plotted against the film Reynolds number for a narrow range of compositions of ethylene glycol solutions. There is only a slight dependence of the heat-transfer coefficient upon the Reynolds number, consistent with the expectation that film thickness varies only as the cube root of film flow rate. It is also seen that varying heat flux by an order of magnitude caused only slight scatter in the results. This confirms our observation that test conditions in this investigation did not involve bubble nucleation and boiling within the film, since the heat-transfer coefficient for boiling processes would be highly sensitive to magnitude of heat flux.

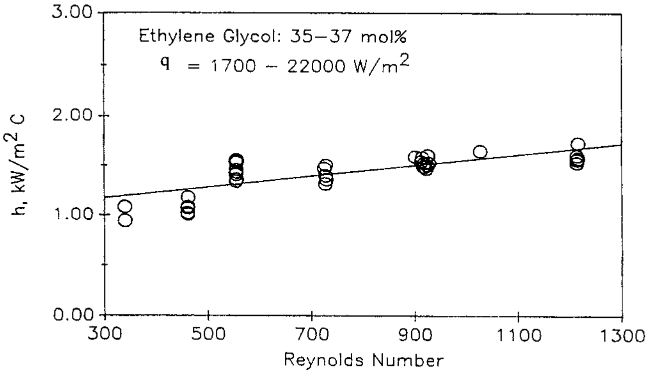

Figure 5. Variation of heat-transfer coefficient with Reynolds number.

Figure 6 plots heat-transfer coefficients vs. composition for various mixtures of ethylene glycol, over selected ranges of Reynolds number and heat flux. While the effect of both flow rate and heat flux on the heat-transfer coefficient was small, it is seen that mixture composition had a very significant effect upon the evaporative heat-transfer coefficient. For the data presented in Figure 6, increasing ethylene glycol from 0 to 0.75 mole fraction caused a decrease of over 80% in the effective heat-transfer coefficient. Due to temperature limitation of the experimental facility, no data were obtained for higher concentration of ethylene glycol. However, it is expected that the heat-transfer coefficient recovers to the value for pure ethylene glycol, as indicated by the dashed portion of the curve in Figure 6. Thus, it appears that the effective heat-transfer coefficient for evaporation of binary mixtures is nonlinearly dependent upon composition, with a minimum approximately at the composition corresponding to maximum spread between bubble and dew points. This behavior is similar to that observed in nucleate pool boiling of mixtures where mass diffusion resistance is known to cause an elevation of the vapor/liquid interface temperature. In our later analysis, this same mechanism is hypothesized for the current case of evaporation at the interface of a falling film.

The next question is whether the dependence of the heat-transfer coefficient on mixture composition can be attributed to change in physical properties, that is, in magnitude of the

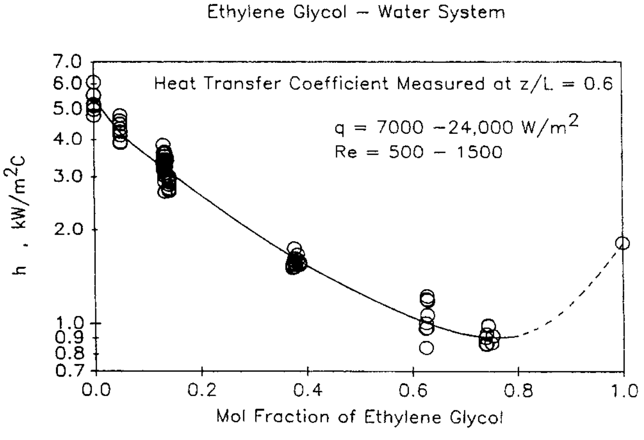

Figure 6. Variation of heat-transfer coefficient with mixture composition.

Prandtl numbers. In the opinion of these investigators, changes in physical property could not explain the mixture effect as the Prandtl numbers varied by less than 25% for the conditions represented in Figure 6. Since the heat-transfer coefficient would be expected to vary with Prandtl numbers to some power less than unity, the 80% variation of heat-transfer coefficient in Figure 6 cannot adequately be explained by changes in physical properties.

A better way to examine this question is to compare the measured heat-transfer coefficients with that calculated for single-component films, using some correlation that accounts for variations of physical properties. From initial tests with water alone, it was determined that the well-known correlation of Chun and Seban (1971, 1972) adequately represented the thermal transport process for single-component films in the range of Prandtl numbers of interest. Figure 7 shows results for ethylene glycol mixtures where the Chun-Seban correlation was used to calculate the heat-transfer coefficient at test conditions, using mixture properties. The ratio of calculated to experimental values of the heat-transfer coefficient range from unity (for pure water) to over two for 75% ethylene glycol solution. Since the Chun-Seban correlation adequately represents single-component evaporation with fluids of various Prandtl numbers, this result confirms that some mechanism other than changes in physical properties must be the cause for the decrease in the heat-transfer coefficient with changing mixture composition.

Analysis and Correlation

The experimental results of this investigation clearly indicate that heat-transfer coefficients for falling film evaporation of binary mixtures are significantly lower than those for evaporation of single-component fluids at corresponding conditions. We contend that this reduction in heat-transfer coefficient is a result of the mass-transfer phenomenon in the liquid film. Figure 7 presents a qualitative representation of the various transport processes in the falling film. Flow of the liquid is of course governed by the interaction of gravitational body force with viscous shear, resulting in a transverse velocity profile with zero velocity at the wall and maximum velocity at the vapor/liquid interface (for negligible interfacial shear). Heat transfer occurs by convection/conduction through the liquid film with the local fluid temperature decreasing almost linearly from the wall temperature to the interfacial temperature. In the case of negligible sensible heat effects (that is, saturated film evaporation), essentially all the heat flux leaving the solid wall passes through the liquid film to supply the latent heat of vaporization at the interface. In the case of single component films, the interface temperature is identical with the saturation temperature at system pressure. This is the situation represented by the various models and correlations for single-component falling film evaporation. In the case of binary liquid films, the more volatile component would be preferentially evaporated at the interface, causing a depression in the local concentration of that component. In the steady-state condition, axial convection would interact with transverse diffusion to supply a net molar flux of the volatile component toward the interface. This process results in a convex local concentration profile, as indicated in Figure 8, with maximum concentration at the solid surface and minimum concentration at the interface (X_i). A bulk composition (X_b) could be calculated as the integral average of the product of local concentration and velocity. By definition, the bulk saturation temperature at any axial position corresponds to the bubble point of the bulk mixture. For design purposes, the heat-transfer coefficient would be based upon the temperature difference ($T_w - T_s$). Since the interface composition of the volatile component (X_i) would be lower than the bulk composition (X_b), it follows that the true interface temperature (T_i) would be higher than the bulk saturation temperature (T_s). As indicated in Figure 8, this represents a net reduction in the available temperature driving force ($T_w - T_i$) for evaporative heat transfer.

In the case of laminar film flow without waves, it is possible to derive an exact solution for the interactive momentum-mass- and heat transfer in the falling film. Hoke and Chen (1992)

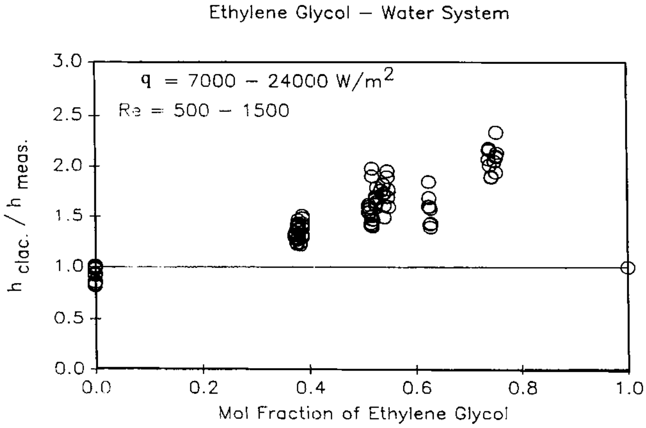

Figure 7. Comparison of experimental heat-transfer coefficient to heat-transfer coefficient calculated by correlation (Chun and Seban, 1971, 1972).

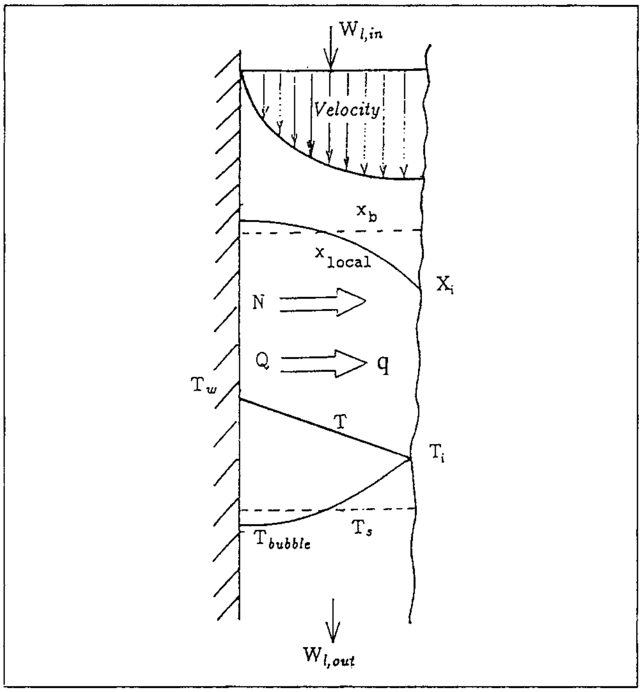

Figure 8. Transverse profiles in falling film.

present a solution for the developing mass-transfer coefficient for this limiting situation. For the cases considered in this study, the film Reynolds number is typically greater than 300, sufficient to have turbulent film flow with interfacial waves. This regime of film flow has resisted analytical solutions to date. Nevertheless, the phenomena represented in Figure 8 are still pertinent and provide a basis for seeking a phenomenological correlation. The development presented below utilizes the concept of film theory, as originally proposed in Stephen's extension of the work of Maxwell, and summarized by Bird et al. (1962). The specific application to phase-change problems also follows the work of Schlünder et al. (1982) who was concerned with nucleate boiling of binary mixtures.

Consider a binary mixture with components 1 and 2 being the more and less volatile components, respectively. A boundary layer of thickness b is defined in the film between the mixed bulk and the interface. The molar flux in the boundary layer for component 1 may be written as:

$$N_1 = -(\rho_1/MW_1)D_m\left(\frac{dX}{dy}\right) + NX \qquad (2)$$

where y is measured from interface into the liquid film, and N is the mass flux in $-y$ direction.

At the interface, for case of negligible vapor-phase resistance:

$$N_1 \simeq NY_1 \simeq NY_b \qquad (3)$$

Integrating across the composition boundary layer:

$$\frac{Y_b - X_b}{Y_b - X_i} = \exp\left[-\frac{N(MW_1)}{\rho_1\beta}\right] \le 1 \qquad (4)$$

where the mass-transfer coefficient β is defined as:

$$\beta \equiv \frac{D_m}{b} \qquad (5)$$

With negligible subcooling or superheating, the heat flux, q, is given by:

$$q = N\lambda(MW)_{va} \simeq N\lambda(MW)_1 \qquad (6)$$

and

$$N\frac{(MW_1)}{\rho_1\beta} = \frac{Q}{\beta} \qquad (7)$$

where:

$$Q \equiv \frac{q}{\rho_1\lambda} \qquad (8)$$

for binary mixtures with components having significantly different boiling points ($T_2 > T_1$). Therefore,

$$(X_b - X_i) = (Y_b - X_b)(e^{Q/\beta} - 1) \qquad (9)$$

Taking a difference approximation for the temperature-composition gradient:

$$T_i - T_s = \left[\frac{dT_s}{dX}\right](X_i - X_b) \qquad (10)$$

and combining with Eq. 9:

$$T_i - T_s = \frac{dT_s}{dX}(Y_b - X_b)(1 - \exp(Q/\beta)) \qquad (11)$$

The heat-transfer coefficient defined in Eq. 1 is based on the temperature difference ($T_w - T_s$). In the case of single-component fluids, the saturation temperature (T_s) is identical to the interface temperature (T_i). It is hypothesized that in the case of binary mixtures, the effective heat-transfer coefficient would be the same as for single-component films, but with the true interface temperature (T_i) in place of the bulk saturation temperature (T_s). Thus, a mixture correction factor can be defined as:

$$F_c = \frac{T_w - T_i}{T_w - T_s}$$

$$= \frac{h}{h_i} \qquad (12)$$

where h_i = coefficient for single-component films at same *Re* and *Pr*:

$$h_i = \frac{q}{T_w - T_i} \qquad (13)$$

Combining Eqs. 11, 12 and 13:

$$F_c = \left[1 + \frac{h_i}{q}\frac{dT_s}{dX}(Y_b - X_b)(1 - \exp(Q/\beta))\right]^{-1} \qquad (14)$$

Schlünder (1983) has suggested an approximation for the saturation temperature gradient, dT_s/dX. It is recognized that the saturation temperature of the heavy component is its boiling point (T_2) at $X = 0.0$ and the saturation temperature of the light component is its boiling point (T_1) at $X = 1.0$; then, overall temperature gradient is approximated as:

$$\frac{dT_s}{dX} = T_1 - T_2 \qquad (15)$$

The simplified expression for F_c then becomes:

$$F_c = (1 - (h_i/q)(T_2 - T_1)(Y_b - X_b)\theta_m)^{-1} \qquad (16)$$

$$\theta_m = 1 - \exp(Q/\beta) \qquad (17)$$

Equations 16 and 17 permit the calculation of the mixture correction factor F_c if the mass-transfer coefficient β is known. Without an analytical solution, it was necessary to rely on experimental data to develop an empirical correlation for this mass-transfer coefficient, using the following procedure:

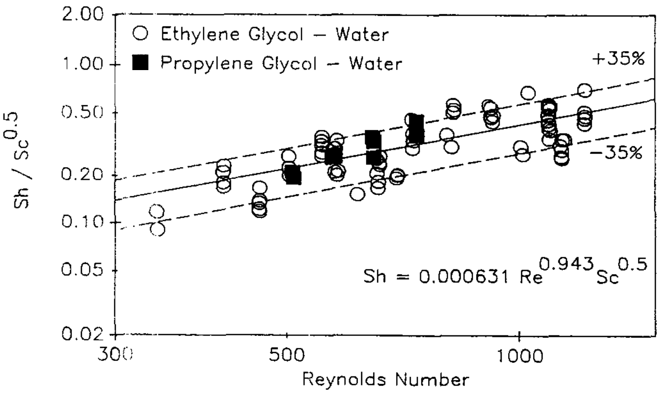

Figure 9. Correlation for Sherwood number.

(a) For every experimental run, determine the effective heat-transfer coefficient (h) from measured wall temperature and heat flux at a position of Z/L equal to 0.6.

(b) Calculate the equivalent single-component heat-transfer coefficient (h_i) by the Chung-Seban correlation, using bulk values of the Reynolds and Prandtl numbers.

(c) Calculate the value for F_c as the ratio h/h_i.

(d) Substitute into Eqs. 16 and 17 to determine the value of the mass-transfer coefficient β.

The resulting values of β for our test conditions ranged from 3×10^{-5} to 2×10^{-4} m/s.

From experience with other interfacial mass-transfer problems, one would expect that these mass-transfer coefficients could be represented by the dimensionless Sherwood number and correlated against Schmidt and Reynolds numbers. The experimental data obtained in this investigation for aqueous solutions of ethylene glycol and propylene glycol covered a range of 125 to 500 in Schmidt number and 330 to 3,000 in Reynolds number. Figure 9 shows the regression correlation of the data from this study, resulting in the equation:

$$Sh = 0.000631\ Re^{0.94}\ Sc^{0.5} \qquad (18)$$

For this correlation, the dimensionless groups are defined as:

$$Sh = \beta\delta/D_m$$

$$Re = 4\Gamma/\mu$$

$$Sc = \mu/\rho D_m$$

where:

$$\delta = \left[\frac{3\mu\Gamma}{\rho^2 g}\right]^{1/3}$$

the equivalent laminar film thickness.

The degree of scatter in Figure 9 is more than desirable. Nevertheless, if the mass-transfer coefficient (β) determined from this correlation is used in Eqs. 1, 12 and 16, a significant improvement is obtained in the prediction of the falling film heat-transfer coefficient for binary mixtures. Figure 10 shows a parity plot of the heat-transfer coefficient calculated by this model compared to experimental measurements. This approach permits calculation of the evaporative heat-transfer coefficient with an average deviation of less than 15%, significantly better than the current practice of using single-component correlations.

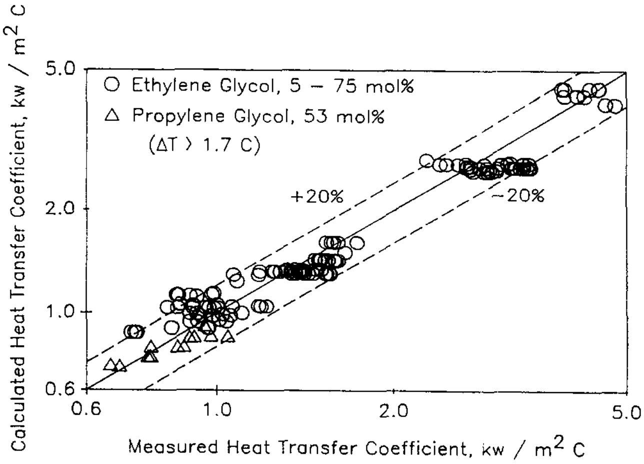

Figure 10. Comparison of calculated and measured heat-transfer coefficient.

Summary and Conclusions

Experimental measurements of falling film evaporation show that the effective heat-transfer coefficients for binary mixtures of wide boiling ranges can be as much as 80% lower than the coefficients calculated for single-component films at corresponding conditions. A mass-transfer induced elevation of the interface temperature was hypothesized as the cause for this apparent decrease in effective heat-transfer coefficient. Using an approximate film-theory formulation, mass-transfer coefficients for the falling films were determined from the experimental data, and correlated in terms of Sherwood, Schmidt and Reynolds numbers. The resulting correlation can be used in the film-theory formulation to estimate elevation of interface temperature in binary films. Improved estimation of the effective heat-transfer rate in falling-film evaporation of binary mixtures is expected.

Acknowledgment

The authors gratefully acknowledge the financial support of Heat Transfer Research, Inc. (HTRI) for this work.

Notation

b = boundary layer thickness at interface of film
D_m = molar diffusivity in liquid
F_c = correction factor, Eq. 13
g = gravitational acceleration
h = heat-transfer coefficient, Eq. 1
h_i = heat-transfer coefficient for single-component fluid
k = thermal conductivity of liquid
L = length of heated section
MW = molecular weight
N = molar flux
Pr = Prandtl number of liquid
q = heat flux
Re = Reynolds number
Sc = Schmidt number
Sh = Sherwood number
T = temperature

T_s = saturation temperature at bulk composition
T_w = wall surface temperature
T_1 = boiling point of volatile component
T_2 = boiling point of less volatile component
X = mol fraction of volatile component in liquid
Y = mol fraction of volatile component in vapor
y = transverse coordinate from interface in liquid film
Z = axial distance from top of heated section

Greek letters

β = mass-transfer coefficient
Γ = film mass-flow rate per unit width
δ = equivalent film thickness for laminar flow
θ_m = parameter, Eq. 17
λ = latent heat of vaporization
μ = viscosity of liquid
ρ = density of liquid
σ = surface tension

Subscripts

b = bulk
i = interface
s = saturation for bulk mixture
w = wall
1 = volatile component
2 = less volatile component

Literature Cited

Chun, K. R., and R. A. Seban, "Heat Transfer to Evaporating Films," *J. Heat Transf.*, **93**, 391 (1971).

Chun, K. R., and R. A. Seban, "Performance Prediction of Falling Film Evaporators," *J. Heat Transf.*, **94**, 432 (1972).

Dukler, A. E., "Fluid Mechanics and Heat Transfer in Vertical Falling-Film Systems," *Chem. Eng. Prog. Symp. Ser.*, **56**(30), 1 (1960).

Ganic, E. N., "On the Heat Transfer and Fluid Flow in Falling Film Shell-and-Tube Evaporators," *Heat Exchangers, Thermal-Hydraulic Fundamentals and Design*, S. Kakac, A. E. Bergles, and F. Mayinger, eds., Hemisphere Publishing, New York, p. 705 (1981).

Gropp, U., "Experimental Determination of Liquid-Side Mass Transfer Resistances in the Falling Film Evaporation of a Binary Mixture," Diss. Univ. of Karlsruhe (1981).

Gropp, U., G. Schnabel, and E. U. Schlünder, "The Effect of Liquid-Side Mass Transfer Resistance on Selectivity During Partial Evaporation of Binary Mixtures in a Falling Film," *Int. Chem. Eng.*, **23**(1), 11 (1983).

Gropp, U., and E. U. Schlünder, "The Effect of Liquid-Side Mass Transfer on Heat Transfer and Selectivity During Surface and Nucleate Boiling of Mixtures in a Falling Film," *Chem. Eng. and Processing*, **20**, 103 (1986).

Hirschburg, R. I., and L. W. Florschuetz, "Laminar and Wave Flow: Part II. Condensation and Evaporation," 20th ASME/AIChE Heat Transfer Conf., Milwaukee, WI, paper 81-HT-14 (1981).

Hoke, B. C., Jr., and J. C. Chen, "Mass Transfer in Evaporating Falling Liquid Film Mixtures," *AIChE J.*, **38**(5), 781 (1992).

Labuntzov, D. A., "Heat Transfer on Film Condensation of Pure Vapors on Vertical Surfaces and Horizontal Pipes," *Teploenergetika*, **4**(7), 7279 (1957).

Le Goff, H., A. Ramadane, M. Barkaoui, Y. Chen, and P. Le Goff, "Modeling the Coupled Heat Mass Transfer in a Falling Film, Application to Absorbers and Desorbers in Absorption Heat Pumps," *Proc. Int. Heat Transfer Conf.*, San Francisco (1986).

Minton, P. E., *Handbook of Evaporation Technology*, Noyes Publications, Park Ridge, NJ (1986).

Mueller, A. C., "Process Heat Exchangers," *Handbook of Heat Transfer Applications*, W. M. Rohsenow, J. P. Hartnett, and E. N. Ganic, eds., McGraw-Hill, New York (1985).

Palen, J. W., "Falling Film Evaporation of Wide-Boiling-Range Mixtures Inside a Vertical Tube," PhD Diss., Dept. of Chemical Engineering, Lehigh University (1988).

Sardesai, R. G., J. W. Palen, and J. G. Thome, "Nucleate Pool Boiling of Hydrocarbon Mixtures," AIChE Meeting, Miami Beach, Paper 127a (1986).

Schlünder, E. U., "Uber den Warmeubergang bei der Blasenverdampfung von Gemischen," *Verfahrenstechnik*, **16**(9), 692 (1982); English Translation in *Int. Chem. Eng.*, **23**(4), 586 (1983).

Schlünder, E. U., "State of the Art of Heat and Mass Transfer," *Heat Exchanger Design Handbook*, Sec. 2.1.7, Hemisphere Publishing, Washington, DC (1983).

Schlünder, E. U., "Heat Transfer in Nucleate Boiling of Mixtures," *Int. Chem. Eng.*, **23**(4), 589 (1983).

Schnabel, G., "Determination of the Local Heat Transfer in Falling Film Evaporation and Condensation on Wavy Surfaces for the Design of High-Performance Evaporators," Diss., University of Karlsruhe (1980).

Seban, R. A., "Transport to Falling Films, Keynote Address," *Int. Heat Trans. Conf.*, p. 417 (1978).

Smith, B. D., *Design of Equilibrium Stage Processes*, McGraw-Hill, New York (1963).

Smith, R. A., "Evaporation," *Heat Exchanger Design Handbook*, Sec. 3.5, Hemisphere Publishing (1983).

Smith, R. A., *Vaporizers*, Wiley, New York (1986).

VDI-Wärmeatlas (*VDI Heat Handbook*), 3rd ed., VDI-Verlag (1977).

Wang, Qi, "Mass Transfer Effects in Falling Film Evaporation of Ternary Mixtures," PhD Diss., Dept. of Chemical Engineering, Lehigh University (1990).

Whitt, F. R., "Performance of Falling Film Evaporators," *British Chem. Eng.*, **11**, 1523 (1966).

Manuscript received Jan. 5, 1993, and revision received July 1, 1993.