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# BOILING INCIPIENCE AND HEAT TRANSFER

ON SMOOTH AND ENHANCED SURFACES

By

Saleem Shakir

# A DISSERTATION

# Submitted to Michigan State University in partial fulfillment of the requirements for the degree of

# DOCTOR OF PHILOSPHY

Department of Mechanical Engineering

#### ABSTRACT

# BOILING INCIPIENCE AND HEAT TRANSFER ON SMOOTH AND ENHANCED SURFACES

By

#### Saleem Shakir

A comprehensive experimental study in nucleate pool boiling of binary mixtures was carried out to investigate the effects of mixture composition on boiling incipient and deactivation superheats and heat transfer coefficients. All experiments were performed at a pressure of 1.01 bar on conventional smooth surfaces and an enhanced surface (High Flux of Union Carbide Corp.). Contact angles were also measured for the same mixtures on the smooth surfaces of brass and copper. The incipience and deactivation of boiling sites on the enhanced surface occurred at much lower wall superheats than on the smooth ones. For the mixture systems investigated, the incipient superheats were observed to be higher than the corresponding deactivation superheats. The classical boiling nucleation criterion was found to be inadequate in predicting the measured incipient superheats. The variation in the measured contact angles was not sufficient to explain the maximum in the incipient superheats observed at intermediate mixture compositions. Several new factors were identified for boiling nucleation in mixtures which could qualitatively explain these maxima in incipient superheats observed for the mixtures.

The boiling heat transfer coefficients obtained on the smooth surafces showed a deterioration when compared with the values obtained from a simple linear mixing law between the single component values. The enhanced surface heat transfer coefficients for boiling of the same mixtures showed both positive and negative deviations from the linear mixing law between the pure component values. The heat transfer coefficients on the enhanced surface showed appreciable augmentation when compared to those obtained on conventional smooth surfaces under similar conditions. The Schluender correlation was tested against the observed values of heat transfer coefficients and reasonable agreement was observed for only two of the four mixture systems tested on smooth surfaces. Enhanced surface heat transfer coefficients could not be satisfactorily correlated by the Schluender correlation. I dedicate this Dissertation in memory of my father, Shakir Ali Thanvi

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#### NOMENCLATURE

A	empirical parameter
Bo	empirical scaling parameter
C P_e	liquid specific heat $\left[\frac{J}{kg K}\right]$
δ D <sub>d</sub>	mass diffusion coefficient $\left[\frac{m}{\sec}\right]$ bubble departure diameter [m]
f h m	bubble departure frequency [sec <sup>-1</sup> ] mass diffusion shell thickness [m]
h <sub>t</sub>	thermal boundary layer thickness [m]
P q	pressure [bars] heat flux $\left[\frac{kW}{2}\right]$
Q r	power [W] radius of vapor bubble embryo [m]
Ŕ	universal gas constant []
Sn t t	Scriven number time [sec] bubble growth time [sec]
	bubble waiting time [sec]
T T S	temperature [°C] saturation temperature [°C]
T W	wall temperature [°C] 3
v	specific volume $\begin{bmatrix} \frac{m}{kg} \end{bmatrix}$
$\tilde{\mathbf{v}}$	molar volume [m/kg-mole]
V x	volume [m] mass fraction of liquid
x	mole fraction of liquid
$\tilde{\mathbf{x}}_{1}$	liquid mole fraction of component 1 (volatile)
x <sub>2</sub>	liquid mole fraction of component 2 (non-volatile)

х <sub>ъ</sub>	bulk liquid mole fraction of volatile component
х <sub>г</sub>	local liquid mole fraction of volatile component
ў <sub>1</sub>	vapor mole fraction of component 1 (volatile)
ў <sub>2</sub>	vapor mole fraction of component 2 (non-volatile)
<sup>ў</sup> ь	bulk vapor mole fraction
<sup>ÿ</sup> L	local vapor mole fraction
Z	compressibility factor
$\left \frac{d\mathbf{P}}{d\mathbf{T}}\right _{sat}$	slope of saturation curve

# Greek Symbols

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æ	heat transfer coefficient $\left[\frac{kW}{2}\right]$
β	contact angle [degrees]
β <sub>ℓ</sub>	liquid mass transfer coefficient $\left[\frac{m}{s}\right]$
c	2 m
0	mass diffusion coefficient [ <del>]</del> sec]
δc	porous coating thickness [mm]
Δh	heat of vaporization $\left[\frac{J}{mole}\right]$ or $\left[\frac{J}{kg}\right]$
	temperature difference [K] ideal wall superheat [K]
	(-T <sub>w</sub> -T <sub>s</sub> ) wall superheat [K]
	boiling range [K]
	wall superheat of component 1 at given heat flux [K]
	wall superheat of component 2 at given heat flux [K]
γ	cavity half angle [deg.]
ρ	density $\left[\frac{\text{moles}}{3}\right]$ or $\left[\frac{\text{kg}}{3}\right]$
<sup>2</sup> e	liquid thermal conductivity $\left[\frac{W}{m K}\right]$
<sup>7</sup> e	liquid viscosity [ $rac{N \ sec}{2}$ ]
σ	surface tension $\left[\frac{N}{m}\right]$
~ e	liquid thermal diffusion coefficient $\left[\frac{m}{m}\right]$
ົ	acentric factor

# **Subscripts**

az   azeotrope     b   bulk     c   critical     exp   experimental     g   gas     id   ideal     inc   incipience     liquid
b bulk c critical exp experimental g gas id ideal inc incipience l liquid
c critical exp experimental g gas id ideal inc incipience liquid
exp experimental g gas id ideal inc incipience liquid
g gas id ideal inc incipience
id ideal inc incipience
inc incipience
2 liquid
~
L local
m mixture
pred predicted
r reduced
s saturation
sat saturation
v vapor
w wall
1 volatile component
2 non-volatile component

# Superscripts

D	departure
E	excess

#### CHAPTER 1

#### INTRODUCTION

The exchange of heat between a heated surface and the liquid surrounding it is one of the most common phenomena employed in the design of heat exchange equipment. When the temperature difference be tween the heated surface and the surrounding liquid (wall superheat) exceeds a certain minimum the mode of heat exchange progresses from a single-phase to a two-phase convection process. The two-phase process, when accompanied by the formation of vapor bubbles on the heated surface is referred to as nucleate boiling. When the medium surrounding the heated surface is a quiescent pool of stagnant liquid the two-phase Process is called "Nucleate pool boiling." The nucleate pool boiling regime is bounded by the inception of boiling on the heated surface at low heat flux and by the departure from nucleate boiling at the peak heat flux. These two special points are referred to as the "Onset of Nucleate Boiling" (ONB) and "Departure from Nucleate Boiling" (DNB), respectively.

Boiling heat transfer is a two-phase convection process of Significant practical importance. Knowledge of the characteristics of nucleate pool boiling of liquid mixtures is of special interest to the chemical, petrochemical, air separation, and liquid natural gas industries. There are primarily three boiling parameters whose knowledge is desired in the thermal design of any heat exchange equipment involving nucleate boiling. First, is the ability to predict the peak nucleate heat flux at DNB at an arbitrary composition so that

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the process is specified to take place in the nucleate pool boiling regime and not in the less efficient film boiling regime. Secondly, it is desirable to be able to predict the minimum wall superheat that is required to initiate boiling on the heated surface to take advantage of the much higher heat transfer coefficients obtained for boiling as compared to single-phase natural convection. Finally, it is necessary to predict the heat transfer coefficients as a function of either heat flux or wall superheat for the liquid composition of interest.

The objective of the present work is to focus on the wall superheat at ONB and the heat transfer coefficients for the boiling of bimary mixture systems. The inception of boiling on a heated surface is a fundamental aspect of the overall boiling process. Not only is it essential to know the amount of wall superheat required for the start up of heat exchange equipment, but also knowledge of the deactivation superheat at which boiling ceases to occur on the heated surface as the imposed heat flux is lowered is of significance since it specifies the minimum possible wall superheat to sustain boiling during normal operation. Previous studies on boiling nucleation have been directed towards single component liquids. The only published data for the onset of boiling in liquid mixtures, to the author's knowledge, is for the forced convective boiling of ethanol-water and ethanol-cyclohexane mixtures inside heated tubes.

The second parameter of the present study was to investigate the effects of mixture composition on heat transfer coefficients. The boiling heat transfer coefficients for mixtures generally show degradation as compared to their pure component values. Thus, the prediction of heat transfer coefficients for binary mixtures becomes important from an economic and reliability standpoint.

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A pool boiling facility was designed and developed in the present study to investigate the nucleate pool boiling of binary mixtures. Boiling incipience/activation and deactivation superheats and heat transfer coefficients have been experimentally obtained for the binary mixture systems of methanol and water, ethanol and water, n-propanol and water, and ethanol and benzene. All experiments were performed at a pressure of 1.01 bar. Test surfaces studied included smooth surfaces (a roughened disk, a polished tube and a smooth tube), and one enhanced surface (High Flux of Union Carbide Corporation, U.S.A). Since the wetting process was expected to be one of the contributing factors in boiling incipience, contact angles were also measured for all four mixture systems. Mixture physical properties have also been estimated based on the generalized corresponding states method.

#### **CHAPTER 2**

#### LITERATURE SURVEY

The process of boiling can be divided into two main categories, nucleate pool boiling and flow boiling. In convective boiling, the heat exchange process between the heated wall and the liquid is obtained in the presence of either an external force (other than gravity) that is applied to make the fluid flow past the surface or due to buoyancy forces created by the boiling process. This type of boiling occurs for evaporation of a liquid flowing inside a tube or over a bank of tubes. On the other hand, in pool boiling the heat transfer takes place in a pool of otherwise quiescent liquid surrounding the heated surface. The liquid motion is only due to the natural convection currents created by the heat transfer process itself. A general discussion of boiling will be presented before proceeding to nucleate pool boiling in binary mixtures.

#### 2.1 BOILING CURVE

The phenomenon of boiling can be easily understood by considering the boiling curve, as shown in Fig. 2.1. The temperature difference between the heated wall and the liquid surrounding it is plotted on the abscissa. This temperature difference will be referred to as the "wall superheat." The heat flux passing through the heated surface is plotted on the ordinate. A typical curve that is obtained by boiling water is shown in Fig. 2.1. The boiling curve can be divided into four distinct regions:

4



Figure 2.1. Pool boiling curve.

- The first region, AB, is that of single-phase liquid natural convection heat transfer occurring at small wall superheats.
- 2. The second region, BD, is that of nucleate pool boiling. Discrete vapor bubbles form and depart from the heated surface. As the wall superheat is increased, the bubble population increases. This portion, BD, can be further subdivided into two regions. In the lower part, BC, only discrete bubbles are observed. A small portion of the surface experiences bubble generation while the rest of the surface still is in the natural convection mode. With further increase in the wall superheat, the process slowly enters the second portion, CD, of fully developed nucleate boiling. In this portion bubbles from neighboring sites on the surface coalesce and the whole surface experiences vigorous boiling. Heat transfer from the surface continues to improve as the process moves along the curve BD.
- 3. Portion DE on the boiling curve is called the Transition regime. Eventually the vapor bubble generation becomes very intense and starts to restrict the liquid from reaching the heated surface. This results in a deterioration in heat transfer due to the very little contact between the surface and the liquid. Point D represents such a transition. The heated surface is partially covered with patches of vapor and partially with liquid.
- 4. The portion EF is the Film boiling regime. A stable film of vapor is formed between the heated surface and the surrounding

6

liquid. As the wall temperature is increased, the radiant component of heat transfer becomes dominant. The upper limit of this region is usually the melting point of the material of the heated surface or the electrical heater burns out.

The boiling curve shown in Fig. 2.1 is realized if the temperature of the surface can be independently controlled e.g. as can be done by condensing steam for the heat source. For the case where the surface is heated by an electric heater, heat flux becomes the independent parameter. In this situation it is not possible to obtain the transition zone, DE, and the wall temperature suddenly increases to a large value if the heat flux is increased beyond point D. Hence, point D is also referred to as the 'Burnout' point, 'Critical Heat Flux,' and also the point from which 'Departure from Nucleate Boiling' takes place. It is to be noted that the boiling curve for different surfaceliquid combinations remains similar to that shown in Fig. 2.1. However, the position of the curve may shift.

The importance of the nucleate boiling regime is obvious. Fairly high heat transfer coefficients are obtained at relatively low wall superheats. The majority of the processes involving two-phase flow are designed to take place in this portion. The nucleate boiling regime is bounded by the inception of boiling on the heated surface at a relatively low heat flux and the departure from nucleate boiling at the peak heat flux (point D in Fig. 2.1). Therefore, the lower and upper bounds of the nucleate boiling regime are of significant practical importance. A knowledge of these two points is desired to take advantage of the augmentation in heat transfer coefficients as compared to single-phase natural convection.

#### 2.2 INCEPTION OF BOILING

The process of nucleate pool boiling to be considered here involves the transfer of heat from a solid surface to the surrounding liquid. It is now a well established fact that the process of boiling initiates on the solid surface in the form of bubble streams emanating from tiny pits and scratches on the surface. Such imperfections are microscopic in size and are found on any engineering surface. Experimental studies of Clark et. al. [11] confirmed that bubbles did form only in small pits and scratches on an otherwise smooth surface. It was postulated that bubbles emerge from cavities on a surface in which a gas or vapor nucleus preexists. As the surface is heated, the vapor nucleus grows and a bubble emerges and detaches from the surface. After the bubble departs some vapor is left behind in the cavity, which becomes a source for the next bubble. The mechanism of trapping of vapor in such cavities is thus very fundamental to the bubble nucleation from a heated surface.

Bankoff [3] considered the spreading of liquid over a surface containing grooves and cavities. As a simple case, consider the geometry as shown in Fig. 2.2 in which a semi-infinite liquid front advances unidirectionally over the surface. Different situations can be realized depending upon the contact angle  $\beta$  and the cavity half angle  $\gamma$ . If the contact angle  $\beta$  is greater than the angle  $2\gamma$ , the advancing liquid front will first strike the opposite wall of the cavity before reaching the bottom of the cavity. The condition for gas entrapment by this approach is:

$$\beta > 2\gamma \tag{2.1}$$

If the liquid wets the cavity walls, then the remaining vapor pressure will be insufficient to balance the surface tension forces leading to the complete penetration of the liquid to the base of the crevice.

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**(a)** 

Figure 2.2. (a) Vapor nuclei trapped in pits and cracks in a surface; (b) liquid front advancing towards a conical cavity; (c) formation of bubbles from a conical cavity.

Cole [12] has shown that steep, narrow cracks and crevices that are poorly wetted by the liquid are the potential sites for nucleation. Experimental evidence that nucleation indeed results from gas or vapor trapped in such cavities is very promising [32].

#### 2.2.1 Superheat Requirement for Vapor Nucleation

Consider the mechanical equilibrium of a spherical vapor nucleus in a liquid at constant temperature  $T_{\ell}$  and pressure  $P_{\ell}$ . The pressure difference across the interface can be expressed as:

$$P_{v} - P_{l} = \sigma(\frac{1}{r_{1}} + \frac{1}{r_{2}})$$
(2.2)

where  $P_v$  is the vapor pressure inside the nucleus and  $P_{\ell}$  is the imposed liquid pressure corresponding to its saturation temperature.  $r_1$  and  $r_2$ are the principal radii of curvature of the vapor. For  $r_1-r_2-r$ , Eq. (2.2) becomes

$$\Delta P = P_v - P_l = \frac{2\sigma}{r}$$
(2.3)

Equation (2.3) is known as the Laplace equation. For thermal equilibrium, the saturation temperature of the vapor must be equal to that of the surrounding liquid. This implies that the surrounding liquid must be superheated above its saturation temperature  $T_{sat}$ . Curvature of the interface fractionally lowers the vapor pressure,  $P_v$ , inside the nucleus compared to that above a planar interface,  $P_{\infty}$ , for the same liquid temperature. This is given by the Kelvin equation

$$\frac{P_{\mathbf{v}}}{P_{\infty}} = \exp\left[\frac{-2\sigma \tilde{\mathbf{v}}_{\boldsymbol{\ell}}}{r\tilde{\mathbf{R}}T}\right] \approx \left[1 - \frac{2\sigma \tilde{\mathbf{v}}_{\boldsymbol{\ell}}}{rP_{\infty}\tilde{\mathbf{v}}_{\mathbf{v}}}\right]$$
(2.4)

From Eqs. (2.3) and (2.4),

$$P_{\infty} - P_{\ell} = \frac{2\sigma}{r} \left[ 1 + \frac{v_{\ell}}{\tilde{v}_{g}} \right]$$
(2.5)

To calculate the liquid superheat  $(T_v - T_{sat})$  corresponding to the pressure difference  $(P_{\omega} - P_{\ell})$ , the Clausius-Clapeyron equation can be used

$$\frac{\mathrm{dP}}{\mathrm{dT}} - \frac{\Delta h_{\mathbf{v}}}{T(\tilde{\mathbf{v}}_{\mathbf{v}} - \tilde{\mathbf{v}}_{\boldsymbol{\ell}})}$$
(2.6)

Assuming ideal gas behavior for the vapor  $(P\tilde{v}_v = \tilde{R}T)$  and  $\tilde{v}_v >> \tilde{v}_l$ , the above equation can be written as

$$\frac{dP}{dT} - \frac{\Delta h_{v}}{T(\tilde{R}T/P)}$$

$$\frac{dP}{P} - \frac{\Delta h_{v}}{\tilde{R}T^{2}} dT \qquad (2.7)$$

If the vapor is assumed to be at the normal saturation state, Eq. (2.7) may be integrated between  $(P_{\ell}, T_{sat})$  and  $(P_{\infty}, T_{v})$ 

$$\int_{P_{\ell}}^{P_{\infty}} \frac{dP}{P} = \frac{\Delta h_{v}}{\tilde{R}} \int_{T_{sat}}^{T_{v}} \frac{dT}{T^{2}}$$

$$\ln\left(\frac{P_{\infty}}{P_{\ell}}\right) = -\frac{\Delta h_{v}}{\tilde{R}} \left(\frac{1}{T_{v}} - \frac{1}{T_{sat}}\right)$$

$$\ln\left(\frac{P_{\infty}}{P_{\ell}}\right) = +\frac{\Delta h_{v}}{\tilde{R}} \left(\frac{1}{T_{v}} - \frac{1}{T_{sat}}\right)$$

$$T_{v} - T_{sat} = \frac{\tilde{R}T_{v}T_{sat}}{\Delta h_{v}} \ln\left(\frac{P_{\infty}}{P_{\ell}}\right) \qquad (2.8)$$

Substituting Eq. (2.5) into Eq. (2.8),

$$T_{v} - T_{sat} = \frac{\overline{R}T_{v}T_{sat}}{\Delta h_{v}} \ln \left[1 + \frac{2\sigma}{P_{\ell}r} \left(1 + \frac{\overline{v}_{\ell}}{\overline{v}_{v}}\right)\right]$$
(2.9)

If  $\tilde{v}_v > \tilde{v}_l$  and  $\frac{2\sigma}{P_l r} < 1$ , then Eq. (2.9) can be rewritten as:

$$T_{v} - T_{sat} = \Delta T_{sat} = \frac{2\sigma}{r_{c}} \frac{\tilde{R}T_{sat}}{P_{\ell}\Delta h_{v}}$$
(2.10)

The condition along the saturation curve is given by the Clapeyron equation:

$$\left[\frac{\mathrm{dP}}{\mathrm{dT}}\right]_{\mathrm{sat}} = \frac{\Delta h_{\mathrm{v}} P}{\tilde{\mathrm{RT}}^2}$$
(2.11)

Equation (2.11) can be written as

$$\Delta T_{sat} = \frac{2\sigma}{r_c (dP/dT)_{sat}}$$
(2.12)

To maintain equilibrium, the superheat of the liquid needed is a function of the surface tension, cavity radius and the slope of the vapor-pressure curve. It should be noted that Eq. (2.12) has been derived for the case of a uniformly superheated liquid.

In nucleate pool boiling, the process of nucleation takes place on the heated surface which is surrounded by liquid at its saturation temperature or just below its saturation temperature (as in the case of subcooled boiling). A real engineering surface has an abundance of cavities with a large variation in size and shape. The problem is therefore to model and predict the superheat requirement for the vapor nucleus to grow under such conditions.

The criterion (Eq. (2.12)) for the formation of a bubble does not hold when only the solid surface is hot. This was experimentally verified by Griffith and Wallis [20]. The discrepancy has been attributed to the nature of the temperature field of the liquid in the immediate vicinity of the surface.

Hsu [26] proposed a model for the incipience of boiling from a bubble site on a heated surface. A transient one-dimensional heat conduction model was assumed in the liquid layer adjacent to the surface. A hemispherical bubble growing out of a cavity, as shown in Fig. 2.3, was considered. Hsu postulated that the criterion for nucleation from this site is that the temperature of the liquid at the

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Figur


Figure 2.3. Hsu's criterion for inception of boiling [26].

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Ъ t t ċ¢ cr to i.e â7.0 :: is tł. Ç 8 *:*:: ; ; ; ette; lefe to If the <sup>q</sup>∕íva[e: bubble cap is equal to or greater than the bubble interior temperature. This criterion is diagrammatically represented in Fig. 2.3. If the line representing the temperature profile (of the liquid near the surface) intersects the equilibrium bubble curve, then nucleation occurs. The first possible site to be activated corresponds to the point of tangency between the equilibrium bubble curve and the liquid temperature profile. If an active site of size  $r_c$  does not exist, then the wall temperature must be increased to a point where active cavities do exist. Hence, a size range of cavities is predicted by Hsu's criterion. Hsu assumed that the location of the isotherm corresponding to the temperature  $T_v$  is  $2r_c$ . The assumption regarding the location of the isotherm has led to various modifications of this criterion. Han and Griffith [22] suggested the location of the isotherm should be at  $1.5r_c$ .

Howell and Siegel [25] argued that only a favorable heat balance is required for the bubble to grow and that it is not necessary that the thermal layer surrounding the bubble embryo be hotter than the bubble embryo itself at all distances.

Lorenz, Mikic and Rohsenow [34] developed a model to account for the wettability of the surface and the geometrical shape of the cavity. Contact angles were used as a measure of the wettability and the geometry of the cavity was represented by the radius of its mouth and its included angle, Fig. 2.4. A simple vapor trapping mechanism was considered for the idealized case of a conical cavity. Once the vapor is trapped by the liquid front, the interface readjusts to form an embryo with radius of curvature  $r_{eff}$ . Conservation of volume requires  $r_{eff}$  to be a function of  $\beta$  and  $\gamma$ . The model is useful in the sense that if the size range of cavities is known for one liquid, then the equivalent value of  $r_{eff}$  for other liquids with different contact



CONICAL CAVITY

CYLINDRICAL CAVITY



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Vapor trapped for all  $\beta$  ( $\beta \neq 0$ )



Figure 2.4. Vapor trapping mechanism as envisioned by Lorenz et al. [34].

angles can be obtained. An important aspect of their analysis was that the effective radius  $r_{eff}$  may not necessarily be the same as the cavity mouth radius and that  $r_{eff}$  is a function of liquid contact angle and geometrical factors of the cavity.

### 2.2.2 Boiling Incipience in Binary Mixtures

Very few experimental studies have investigated the functional dependence of boiling incipient superheats on the mixture composition. Shock [51] obtained some results for the onset of nucleate boiling for flow boiling of ethanol-water mixtures inside a heated tube. These temperatures were obtained by observing changes in the heated tube's axial temperature profile. He noted that the trends in  $\sigma$  and  $(dP/dT)_{sat}$  at 1.01 bar, Fig. 2.5, predict a decrease in the superheat required for nucleation, Eq. (2.12). Only three mole fractions were tested by Shock, Table 2.1. The increase in the superheat was attributed to the drastic change in the contact angle with increasing composition of ethanol. See the contact angle data of Eddington and Kenning [16] for the ethanol-water mixtures shown in Fig. 2.6, for instance. Shock [52] concluded that for low contact angle fluids it may not be realistic to assume that bubbles on the point of nucleation are hemispheres at the mouth of cavities with circular entrances of radius r. For such cases r is not equal to reff. With low contact angle fluids, the number of sites that trap vapor are smaller in number, and in cavities where vapor is trapped, the volume of the vapor is greatly reduced. The critical nucleus in such conditions may no longer be at the cavity mouth, but deep within the cavity with  $r_{eff} < r_c$ . This decrease in r may be responsible for the greater superheats required for the stability of the nucleus.



Figure 2.5. Phase equilibrium diagram, surface tension, and slope of saturation curve for ethanol-water by Shock [52].



Figure 2.6. Contact angle data for ethanol-water mixtures against nitrogen at 25°C by Eddington and Kenning [16].

x	T w	$\Delta T_{sat}$	$r_{eff}$	P
ethanol	(°C)	(°C)	(µm)	(bar)
0.0	138.2	9.9	1.05	2.61
0.058	143.2	26.6	0.23	2.59
0.197	145.0	36.7	0.095	2.47

Table 2.1.	Shock's data	for the onset of	nucleate boiling in
	flow boiling	of ethanol-water	mixtures

Thome et al. [72] reported on incipient superheats in pool boiling for the cryogenic mixture system nitrogen-argon at 1.0 bar. The activation and deactivation of the same boiling site was obtained for fourteen compositions in the range 0.04-1.0 mole fraction of nitrogen. No effect of composition on the superheats was observed, Fig. 2.7, even though Eq. (2.12), the solid curve in Fig. 2.7(a), predicts a monotonic decrease. The site activated and deactivated at the same wall superheat. Contact angle variation with composition for this mixture system is not known, but the values are assumed to be small. Mercier (in [75]) looked at the effect of prepressurization on the activation superheats for the nitrogen-argon mixture, Fig. 2.8. For this case a maximum was observed. Prepressurization is believed to cause partial or complete condensation of the vapor nuclei, resulting in higher incipient superheats.

### 2.3 NUCLEATE POOL BOILING HEAT TRANSFER IN BINARY LIQUID MIXTURES

Experimental investigations on nucleate pool boiling of binary mixtures have established that the boiling heat transfer coefficients of the mixtures, at a given heat flux, can be considerably lower than would be expected for an "ideal" single-component liquid with the same physical properties as the mixture. To understand the deterioration in the heat transfer coefficients of mixtures, it is necessary to consider the important boiling parameters and their functional dependence on





Figure 2.7. Boiling incipience data for nitrogen-argon mixtures by Thome et al. [72]. (a) Activation superheats; (b) deactivation superheats.



Figure 2.8. The effect of prepressurization on boiling activation superheats for nitrogen-argon mixtures at 1.55 bar [75].

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mixture composition. Various mechanisms that contribute to the augmentation in the heat transfer process in nucleate boiling will also be discussed.

# 2.3.1 Bubble Growth Dynamics

The process of nucleate boiling is identified by the generation of vapor bubbles on the heated surface. Hence, the mechanics of bubble growth is a key to understanding the nucleate boiling process. The growth of a bubble begins once the vapor nucleus attains a size greater than that for stable equilibrium. Stated differently, once the wall superheat exceeds the equilibrium value, the bubble starts to grow. Most of the earlier models of the bubble growth process were developed to describe single component fluids. The same ideas were then extended to explain the bubble growth dynamics for mixtures of fluids.

Bubble growth in a single component fluid can be divided into two periods [19]. The first period of growth is dominated by hydrodynamic and surface tension forces. This early stage of growth, which is relatively short, is hydrodynamically controlled by the imparting of inertia to the surrounding liquid. The inertial stresses rise rapidly after nucleation and then decrease proportional to  $1/r^2$  as the bubble enters the second period of growth, known as the "asymptotic" period. The growth is then limited by the rate at which heat can diffuse from the superheated liquid surrounding the bubble to the bubble interface thus providing the latent heat of vaporization needed to vaporize the liquid. The growth rate of the bubble is therefore governed by the conduction of heat to the bubble interface.

In a binary mixture the phenomenon of vapor bubble growth is much more complex. In general, the composition of the volatile component in the vapor phase,  $\tilde{y}$ , is greater than in the liquid phase,  $\tilde{x}$ . During the

growth of a vapor bubble, the different volatilities of the two components cause a depletion of the volatile component very near the bubble interface. This is due to the fact that the volatile component vaporizes more readily. This results in a local concentration gradient around the bubble. The local value of  $\tilde{x}_1$  decreases and the bubble point rises. Figure 2.9 shows the thermal and concentration gradient around a bubble. The bubble point at the interface reaches its maximum value when the rate of diffusion of the volatile component to the interface balances the rate of its excess evaporation at the interface. The rate of evaporation of the volatile component at the interface is proportional to the composition difference between the vapor phase and the liquid phase  $(\tilde{y}-\tilde{x})$ . This was first explained by Van Wijk et al. [96]. Due to the rise in the local boiling point, the driving force to supply latent heat to the bubble is reduced to  $T_w$ -T\_w-sat, local. This causes a decrease in the bubble growth rate in mixtures as compared to a single-component liquid with physical properties the same as those of the mixture. Such a hypothetical fluid can also be referred to as an "equivalent pure fluid."

Several theoretical and experimental studies have concentrated on bubble growth models. These studies can be separated into two categories, those for bubbles growing homogeneously in a superheated liquid i.e. remote from a heated surface, and those growing heterogeneously at a heated wall.

**Spherical bubble growth remote from a heated surface**. Scriven [49] and Van Stralen [87-93] extended the theory for spherically symmetric bubble growth for single component liquids to binary mixtures. Scriven concluded that the rate of bubble growth, dr/dt, in the asymptotic







Figure 2.9. Bubble growth model of Van Stralen [89] for a spherical bubble growing in a superheated binary mixture.
(a) Temperature and concentration profiles; (b) process shown on a phase-equilibrium diagram.

stage in a binary mixture is always slower than that of an equivalent pure fluid. The expression for the radius is given in the form:

$$r = 2B\sqrt{\kappa_{\ell}t}$$
(2.13)

where B is the bubble growth coefficient and as shown below is a function of the composition of the fluid as well as the wall superheat.

$$B = \frac{\sqrt{3/\pi} \Delta T}{(\rho_v / \rho_\ell) (\Delta h_v / C_{p_\ell}) \left( 1 - (y - x) \sqrt{\kappa_\ell / \delta} (C_{p_\ell} / \Delta h_v) (dT/dx) \right)}$$
(2.14)

Scriven evaluated B for the binary mixture system water-ethylene glycol. The model developed by Van Stralen for bubble growth in a binary mixture is an extension of Plesset and Zwick's [42] model for spherical bubble growth in the bulk of an initially superheated singlecomponent liquid. The wall superheat was modified to take into account the rise in local saturation temperature due to preferential evaporation of the volatile component near the bubble interface. The resulting expression for the radius was the same as that obtained by Scriven Eq. (2.13). The theoretical models proposed have been experimentally verified by Florschuetz et al. [17] for the binary mixture systems of ethanol-water and isopropanol-water.

**Bubble growth at the heated surface**. The mechanism of nucleation and bubble growth at a heated surface is far more complex than when the bubble grows homogeneously in a uniformly superheated medium. Experimental studies of mixture bubble growth [4,100] on surfaces found the models proposed by Scriven and by Van Stralen for bubble growth in a uniformly superheated medium to be inadequate. Various studies suggest the following mechanisms to contribute to the bubble growth from a heated surface:

- Microlayer evaporation: A very thin layer of superheated liquid exists between the hemispherical vapor bubble and the heated surface. Evaporation of this microlayer at the interface of the bubble aids in the growth of the vapor bubble [35].
- Relaxation microlayer evaporation: Due to this mechanism, evaporation occurs over that portion of the bubble which is covered by the superheated boundary layer [90,91].

Van Ouwerkerk [86] considered growth of a hemispherical vapor bubble at a heated surface in a binary mixture. His model included evaporation over the hemispherical surface of the bubble and from the microlayer underneath the bubble. Van Ouwerkerk's solution was similar to that of Scriven's except for the growth constant B. Van Ouwerkerk obtained  $(1+\sqrt{3})$  as opposed to  $\sqrt{3}$  obtained by Scriven in Eq. (2.14).

Van Stralen [94] used his model for a bubble growing in an initially uniformly superheated liquid away from a heated surface to study bubble growth on a heated surface. His analysis was based on the evaporation from the relaxation microlayer. A parameter relating the thickness of the thermal boundary layer to the height of the bubble is introduced in his model. This parameter, in general, will be a function of composition and is not known. Van Stralen et al. [95] later developed a model for the entire growth stage of hydrodynamic and asymptotic growth including the combined affect of the relaxation and evaporation microlayers. The same parameter exists in this model which renders it difficult to use.

Another complicating factor in the bubble growth dynamics is the interaction between neighboring boiling sites. To address this issue,

Thome [71] suggested the use of another factor in calculating the effective superheat that is available for the bubble growth, in addition to that due to the local rise in the saturation temperature caused by the bubble itself. The expression for the superheat by Thome is

$$\Delta T_{eff} = \Delta T - \Delta \theta_1 - \Delta \theta_2 \qquad (2.15)$$

where  $\Delta \theta_1$  is the steady state local rise in the boiling point of the liquid adjacent to the heated wall due to the consumption of the volatile component by the neighboring bubbles.  $\Delta \theta_2$  is the rise in the local boiling point due to the growth of the individual bubble in this local liquid layer.

Experimental investigations [13,18,68,70] qualitatively confirm the fact that bubble growth rates are reduced in mixtures compared to an equivalent pure fluid.

# 2.3.2 Bubble Departure from the surface

The augmentation in heat transfer observed in transition from single-phase natural convection to nucleate pool boiling has been attributed to the generation of vapor bubbles on the heated surface. For this reason, bubble departure diameter and frequency become important factors to be considered in an effort to help explain this improvement. The effect of composition on these two parameters is an additional complication introduced in the boiling of mixtures.

**Bubble departure diameter**. Several experimental investigations [96,69,78] confirmed that bubble departure diameters are considerably reduced at intermediate compositions of mixtures as compared to those of the individual fluids. To explain this, Thome and Davey [69] used a

dynamic force balance for bubble departure in a single-component liquid developed by Keshock and Siegel [30]. The force balance takes into account all the forces acting on the bubble at its departure, i.e. a combination of inertia, drag, and excess pressure forces balance the buoyancy force. Smaller bubble departure diameters in mixtures were explained to result from a reduction in the inertia and drag forces.

Thome [71] later modified the Keshock and Siegel bubble departure equation to take into account the effect of the binary mixture bubble growth rate. Two different conditions of bubble growth were identified; (1) inertia controlled, and (2) surface tension controlled growth and departure. The model showed that the actual process is bounded by these two conditions [75]. In addition to the reduction in the bubble inertia force, retardation of the microlayer evaporation rate (due to the preferential evaporation of the volatile component) was also believed to cause smaller bubble departure diameters.

**Bubble departure frequency.** The frequency at which bubbles depart from a nucleation site on the heated surface is also an important factor governing the heat transfer rate from the surface. The frequency of the bubble departure f is defined as

$$f = \frac{1}{t_w + t_g}$$
(2.16)

where  $t_w$  is the waiting time during which a vapor nucleus left behind by a departing bubble waits to be activated and  $t_g$  is the growth time during which the bubble grows from its nucleation size to the departure diameter. Binary mixture experimental investigations show that the bubble departure frequency, in general, is higher than for the equivalent pure fluid. Two opposing factors govern the growth time  $t_g$ . First, in mixtures the bubble departure diameter is smaller and therefore bubbles do not grow as large as in a single component liquid. Secondly, the growth factor is affected by the slower growth rate. The bubble waiting time  $t_w$  depends upon the rate at which the vapor nucleus is provided the necessary superheat for its activation. The rate of formation of the new thermal boundary layer after the departure of the previous bubble is important in this regard. Thome [68] found the bubble departure frequency to increase for the nitrogen-argon mixtures. Tolubinskiy and Ostrovskiy [78,79,80] found a similar trend in the experimental results for ethanol-water, while for ethanol-butanol and water-glycerine no variation was observed.

# 2.4 HEAT TRANSFER COEFFICIENTS IN NUCLEATE POOL BOILING

Nucleate pool boiling of binary mixtures shows degradation of the heat transfer coefficient when compared with that of an "equivalent pure fluid." The first physical explanation for this behavior was presented by Van Wijk et al. [96]. Their explanation was based on the preferential evaporation of the volatile component at the heated surface. This process, shown schematically in Fig. 2.9, causes a rise in the local boiling point of the liquid which in turn reduces the effective superheat. Sterling and Tichacek [59] attributed the deterioration in mixture heat transfer coefficients to the change in: (1) mixture physical properties; (2) bubble growth rates (due to the mass transfer of the volatile component in diffusing to the bubble interface); and (3) the rate of nucleation of new boiling sites on the surface. The last mechanism was suggested to be caused by the relatively high concentration of the non-volatile component near the surface, resulting in the deactivation of the potential sites. This mechanism has no experimental or theoretical basis.

Stephan and Korner [55] compared the work of formation of a bubble in a mixture to that for a bubble in an equivalent pure fluid. They explained the reduction in the mixture boiling heat transfer coefficient to be due to in the work of formation of bubbles in mixtures. Stephan and Preusser [56] attributed the lower heat transfer coefficients in mixtures in part to the non-linear variation in physical properties. Thome [73] examined the major heat transport mechanisms in nucleate pool boiling in mixtures to seek an explanation for the lower values. The principal mechanisms considered by Thome were bubble evaporation and thermal boundary layer stripping at an isolated boiling site. The bubble evaporation mechanism refers to the latent heat carried away by the departing bubble. Thermal boundary layer stripping refers to the cyclic removal of the thermal boundary layer by the departing vapor bubbles. The agitation of the liquid around the departing bubble enhances the convective heat transfer. Using his experimental data for nitrogen-argon mixtures, he showed that these mechanisms are retarded in the mixtures. Thome and Shock [75] conclude that the decrease in the boiling heat transfer coefficient is due to a number of factors including; (1) mixture effects on bubble growth rate and departure; (2) bubble nucleation and boiling site density; (3) nonlinear variation in physical properties. In general, a knowledge of these factors as a function of composition is not available.

To date there is no correlation that adequately predicts the heat transfer coefficients for single-component nucleate pool boiling and hence for the mixtures's "equivalent pure fluid." To avoid this difficulty, a linear molar mixing law was suggested by Stephan and Korner [55] to provide a basis for comparison:

$$\Delta T_{id} = \tilde{x}_1 \Delta T_1 + \tilde{x}_1 \Delta T_2 \tag{2.17}$$

where  $\tilde{\mathbf{x}}_1$  and  $\tilde{\mathbf{x}}_2$  are the liquid mole fractions of the two components in the binary mixture.  $\Delta T_1$  and  $\Delta T_2$  are the respective single-component wall superheats at the given heat flux. Subscript 1 refers to the volatile component. It was suggested that the linear molar mixing law be used as a measure of the deterioration of mixture heat transfer coefficients. However, in this procedure the effect of non-linear variation in the physical properties of the mixtures is not taken into account. Thome and Shock [75] modified the linear molar mixing law to match the wall superheat at the azeotrope. This was necessitated since the azeotrope behaves as a single component. For mixtures exhibiting azeotropy, the wall superheat to the left is prorated as

$$\Delta T_{id} = \left(\frac{\tilde{x}_1}{\tilde{x}_2}\right) \Delta T_{az} + \left(\frac{\tilde{x}_{az} - \tilde{x}_1}{\tilde{x}_{az}}\right) \Delta T_2$$
(2.18)

and to the right of azeotrope

$$\Delta T_{id} = \frac{(\tilde{x}_1 - \tilde{x}_{az})}{(1 - \tilde{x}_{az})} \Delta T_1 + \frac{(1 - \tilde{x}_1)}{(1 - \tilde{x}_{az})} \Delta T_{az}$$
(2.19)

Figure 2.10 shows the difference between the two interpretations of linear molar mixing law. The "ideal" heat transfer coefficient, based on the linear molar mixing law, at a given heat flux is then determined from

$$\alpha_{id} - \frac{\dot{q}}{\Delta T_{id}}$$
(2.20)

where  $\dot{q}$  is the heat flux.

Various conditions that have been observed to influence the heat transfer coefficients in boiling are briefly discussed below.

**Effect of subcooling.** When the bulk of the liquid is maintained at a temperature  $(T_b)$  below its saturation value  $(T_{sat})$ , the process is



Figure 2.10. Differing definitions of linear mixing law for an azeotropic mixture system.

referred to as subcooled boiling. Heat transfer coefficients based on  $(T_w-T_b)$  have been observed to decrease as the amount of subcooling decreased [60,27,28]. However, this decrease is less for a mixture than for its pure components [27,28].

**Effect of pressure**. The heat transfer coefficients for mixtures increase as the pressure increases, the same trend as exhibited by single components. However, the degradation in the heat transfer coefficients of mixtures becomes more pronounced as the pressure increases, because the phase-equilibrium diagram changes substantially with a change in pressure. A number of experimental studies have investigated the effect of pressure on heat transfer coefficients [23,81,99].

**Effect of surface conditions**. Surface roughness has been found to influence the heat transfer coefficients in mixtures as well as single components. The overall effect is an increase in the heat transfer coefficients with increasing roughness. However, the improved performance obtained by artificial surface treatments (excluding "enhanced surfaces") tends to fade with time.

# 2.4.1 Predictive Methods for Boiling Heat Transfer Coefficients of Mixtures on Smooth Surfaces

A wealth of experimental data on heat transfer coefficients exists for nucleate pool boiling in binary mixtures. The influence of various factors on heat transport from the heated surface makes it very difficult to formulate a universal equation or correlation that will predict their values with a reasonable degree of accuracy. It is very difficult to isolate a single parameter and study its influence on the overall boiling process. These problems are reflected in most of the existing correlations which predict the heat transfer coefficients in boiling of binary mixtures. A list of correlations for mixture boiling heat transfer coefficients is given in Ref. [52]. Only the correlations that are more general in nature will be discussed here.

One of the earliest correlations was presented by Palen and Small [38] in 1964. The relation given for mixtures with a wide boiling range is

$$\frac{\alpha}{\alpha_{id}} = \exp\left[-0.027(T_{bo} - T_{bi})\right]$$
(2.21)

The temperature difference  $(T_{bo}-T_{bi})$  is equivalent to the difference between the dew point and bubble point temperatures at the liquid feed composition to a reboiler.

Stephan and Korner [55] expressed the wall superheat as the sum of two contributions:

$$\Delta T = T_{w} - T_{sat} = \Delta T_{id} + \Delta T^{E}$$
 (2.22)

where the ideal superheat,  $\Delta T_{id}$ , is obtained from the linear molar mixing law at the given heat flux, Eq. (2.18), and the excess superheat,  $\Delta T^E$ , was expressed in the form:

$$\Delta T^{E} = A \left| \tilde{y} - \tilde{x} \right| \Delta T_{id}$$
(2.23)

where the empirical constant A is given by

$$A = A_{0}(0.88 + 0.12P)$$
(2.24)

valid over the pressure range of 1-10 bars. The constant  $A_0$  depends on the pure components of the particular binary mixture. A list of their values for different mixture systems is given in Table 2.2. The motivation for correlating the heat transfer coefficients in this form was the observed reduction in bubble growth rates at the maximum in  $|\tilde{y}-\tilde{x}|$ .

Mixture	A
Acetone-ethanol	0.75
Acetone-butanol	1.18
Acetone-water	1.40
Ethanol-acetone	0.75
Ethanol-benzene	0.42
Ethanol-cyclohexane	1.31
Ethanol-water	1.21
Benzene-toluene	1.44
Heptane-methylcyclohexane	1.95
Isopropanol-water	2.04
Methanol-benzene	1.08
Methanol-amyl alcohol	0.80
Methylethyl ketone-toluene	1.32
Methylethyl ketone-water	1.21
n-propanol-water	3.29
Water-glycol	1.47
Water-glycerol	1.50
Water-pyridine	3.56

Table 2.2. Numerical values of A

Happel and Stephan [23] suggested a correlation of the form

$$\frac{\alpha}{\alpha_{id}} = \frac{\Delta T_{id}}{\Delta T} = 1 - K_{st} |\tilde{y} - \tilde{x}|^n \qquad (2.25)$$

where  $K_{st}$  and n depend upon the binary system and  $K_{st}$  also depends upon pressure. Afgan [1] suggested n=1.0 in his similar correlation.

Stephan and Preusser [56] suggested an extension to the Stephan and Korner binary mixture correlation to multicomponent mixtures by expressing the wall superheat as:

$$\Delta T = \Delta T_{id} + \Delta T^{E}$$
 (2.26)

where

and

$$\Delta T^{E} = \sum_{i} K_{in}(\tilde{y}_{i} - \tilde{x}_{i}) \qquad (2.28)$$

Coefficients  $K_{in}$  were estimated by the method of least squares using experimental data from 32 binary and 50 ternary compositions in the heat flux range of 50-200 kW/m<sup>2</sup>. Stephan and Preusser [57] also

extended Stephan and Abdelsalam's [58] correlation for singlecomponents to multicomponent mixtures.

$$Nu = \frac{(\alpha D_{d})}{\lambda_{\ell}} = 0.100 \left[ \frac{\dot{q} D_{d}}{\lambda_{\ell} T_{sat}} \right]^{0.674} \left[ \frac{\rho_{v}}{\rho_{\ell}} \right]^{0.156} \left[ \frac{\Delta h_{v} D_{d}}{\kappa_{\ell}}^{2} \right]^{0.371} \left[ \frac{\kappa_{\ell}}{\sigma D_{d}} \right]^{0.350} \cdot \left[ \frac{\eta_{\ell} C_{p\ell}}{\lambda_{\ell}} \right]^{-0.162} \left\{ 1 + \left| \sum_{i}^{n-1} (\tilde{y}_{i} - \tilde{x}_{i}) \left[ \frac{\partial \tilde{y}_{i}}{\partial \tilde{x}_{i}} \right] \right|_{x_{j}, P} \right\}^{-0.0733}$$
(2.29)

where the bubble departure diameter  $D_d$  is calculated from

$$D_{d} = 0.0146 \ \beta \left[ \frac{2\sigma}{g_{n} \sqrt{\rho_{\ell} - \rho_{v}}} \right]$$
(2.30)

 $\beta$  is the contact angle in degrees. The value of  $\beta$  is assumed to be 45 degrees for water and 35 degrees for organics and for all mixtures. Note that very small values of  $\beta$  result in D<sub>d</sub> approaching zero.

<u>Semi-empirical correlations</u>. Calus and Rice [7] developed a semiempirical correlation for heat transfer coefficients in binary mixtures. Their correlation was based on the Scriven and Van Stralen expressions (Eq. (2.13)) for bubble growth remote from a wall in binary mixtures. Their data for boiling of mixture systems of i-propanol-water and acetone-water on a thin wire were correlated by the expression

$$\frac{\alpha}{\alpha_{id}} = \left[1 + |y-x| \sqrt{\kappa_{\ell}/\delta}\right]^{-0.7}$$
(2.31)

The terms (dT/dx) and  $(C_{p\ell}/\Delta h_v)$  in Eq. (2.14) were dropped without any justification.  $\alpha_{id}$  is evaluated using the Borishankii-Minchenko [33] correlation for single component liquids.

Recently Schluender [48] derived an equation for predicting the heat transfer coefficients in binary mixtures. The equation proposed has the form

$$\frac{\alpha}{\alpha_{id}} = \frac{1}{\left\{1 + \frac{\alpha_{id}}{q}(T_{s_2} - T_{s_1})(\tilde{y}_1 - \tilde{x}_1)\left(1 - \exp\left[-\frac{B_o q}{\rho_\ell \beta_\ell \Delta h_v}\right]\right)\right\}}$$
(2.32)

where  $\alpha_{id}$  is obtained from a linear molar mixing law

$$\alpha_{id} = \left[\frac{\tilde{x}_1}{\alpha_1} + \frac{\tilde{x}_2}{\alpha_2}\right]^{-1}$$
(2.33)

which is equivalent to Eq. (2.17) for non-azeotropic mixtures.  $(T_{s_2} - T_{s_1})$  is the difference between the saturation temperatures of the pure components (subscript 1 refers to the volatile component). For azeotropic mixtures  $(T_{s_2} - T_{s_1})$  has to be replaced by  $(T_{s_2} - T_{s_{az}})/\tilde{x}_{az}$  to the left of azeotrope, and by  $(T_{s_{az}} - T_{s_1})/(1 - \tilde{x}_{az})$  to the right of the azeotrope. B<sub>o</sub> is referred to as an empirical scaling parameter and its value is taken to be 1.0 by Schluender. Bier et al. [5] found good agreement between the predicted values using Schluender correlation and their experimental data obtained for binary mixtures of SF<sub>6</sub> and refrigerants. Uhlig and Thome [84] also tested Eq. (2.32) against their data for the acetone-water mixture system and found the deviations to be within  $\pm 20$ %.

**Analytical expressions for heat transfer coefficients**. Calus and **Leonidoplous** [8] presented the first completely analytical expression for predicting the wall superheats in binary mixtures:

$$\frac{\Delta T}{\Delta T_{id}} = \left[1 - (y-x)\sqrt{\kappa_{\ell}/\delta} \quad \frac{C_{p_{\ell}}}{\Delta h_{v}} \frac{dT}{dx}\right]$$
(2.34)

where  $\Delta T_{id}$  is given by the linear mixing law, i.e.  $\Delta T_{id} = x_1 \Delta T_1 + x_2 \Delta T_2$ . They based their model using only the mixture effect on the bubble growth rate. Thome [71] later derived an analytical expression for binary mixtures that is based on the cyclic thermal boundary layer stripping mechanism. The model also takes into account the effect of mass diffusion controlled bubble growth on bubble departure diameter, thermal boundary layer thickness and bubble departure frequency. Two different expressions are given:

$$\frac{\alpha}{\alpha_{id}} = \frac{\Delta T_{id}}{\Delta T} = Sn^{7/5}$$
(2.35)

for inertia-controlled bubble departure, and for surface-tensioncontrolled departure

$$\frac{\alpha}{\alpha_{id}} = \frac{\Delta T_{id}}{\Delta T} = Sn^{3/2} \cdot \frac{\sigma \sin\beta}{\sigma \sin\beta_{id}}$$
(2.36)

where

$$\operatorname{Sn} = \left[1 - (y-x)\sqrt{\kappa_{\ell}/\delta} \sqrt{C_{p_{\ell}}/\Delta h_{v}} \frac{\mathrm{dT}}{\mathrm{dx}}\right]^{-1}$$
(2.37)

and  $\sigma_{id}$  and  $\beta_{id}$  are the linear molar interpolated values.

Using the phase-equilibrium diagram, Thome [74] later derived an expression that is based on the local rise in the saturation temperature due to the preferential evaporation of the volatile component. Thome hypothesized that this rise is controlled by the total rate of evaporation at the heated surface. The local rise in the saturation temperature caused by the preferential evaporation of the volatile component was used by Thome to derive another expression for heat transfer coefficients in binary mixtures. Thus, this rise increases from zero (for single-phase natural convection) to a maximum at the peak heat flux. This is schematically shown in Fig. 2.11. for the condition at the peak heat flux, where it is assumed that all the liquid arriving at the heated surface is evaporated. Thus, at the peak heat flux  $\tilde{y}_{local} - \tilde{x}_{bulk}$  and for single-phase natural convection  $\tilde{x}_{local}$ 



Figure 2.11. Maximum rise in local boiling point by Thome [74].

 $\frac{\alpha}{\alpha_{id}}$ 

Hgure



Figure 2.12. Comparision of ethanol-water data at 1.0 bar with four predictive methods.
Legend: (●) Valent and Afgan [85] at 300 kW/m<sup>2</sup>;
(X) Valent and Afgan [85] at 190 kW/m<sup>2</sup>; (●) Tolubinskiy and Ostrovskiy [78] at 116 kW/m<sup>2</sup>; (+) Grigorev et al.
[21] at 232 kW/m<sup>2</sup>; (▲) Bonilla and Perry [6] at 95 kW/m<sup>2</sup>;
(♥) Cichelli and Bonilla [10] at 221 kW/m<sup>2</sup>; (■) Preusser [45] at 200 kW/m<sup>2</sup>; (□) present work at 200 kW/m<sup>2</sup>.

-  $\tilde{x}_{bulk}$  . Using the maximum rise in the local saturation temperature,  $\Delta T_{bn}$ , Thome arrived at the equation

$$\frac{\alpha}{\alpha_{id}} = \frac{\Delta T_{id}}{\Delta T} = \frac{\Delta T_{id}}{\Delta T_{id} + \Delta T_{bp}}$$
(2.38)

where  $\alpha_{id}$  is obtained from Eq. (2.33). Equation (2.38) requires only phase-equilibrium data to predict heat transfer coefficients in binary mixtures. However, in this model the local rise in the saturation temperature is independent of the heat flux, which is not true in a real situation. Thus, for heat flux levels below the peak value, Eq. (2.38) will underpredict the heat transfer coefficients. Figure 2.12 compares various predictive methods against experimental data for ethanol-water mixtures at 1.0 bar.

#### 2.5 ENHANCED SURFACES

It has long been recognized that in boiling the surface conditions play a very significant role in the thermal performance of the surface. Various techniques have been employed to improve the heat transfer coefficients obtained from such surfaces. One of the initial steps taken in this regard was the use of roughened surfaces. The improved performance from such superficial treatments lasted for a very short period of time and hence were not of any commercial importance. However, the sustained interest in the surface behavior led to the development of a number of special surfaces that have shown great promise for prolonged commercial use. A detailed survey on the evolution of enhanced surfaces is given by Webb [97].

The high heat transfer coefficients obtained for boiling on enhanced surfaces are an order of magnitude higher than those of a conventional smooth surface. The special geometry of the enhanced surfaces is responsible for this augmentation in heat transfer coefficients. These surfaces are finding increased practical application in many areas of thermal engineering where there is a need to obtain high heat fluxes while maintaining a low temperature difference between the heating fluid and the evaporating fluid.

Enhanced surfaces have been classified into two main groups: (1) porous coated surfaces and (2) "structured" surfaces. This classification is based on the manufacturing processes employed. The factors that influence the commercial acceptance of these surfaces are performance and cost. A surface that has received much attention in this regard is the High Flux surface (Trademark of Union Carbide Corporation). High Flux consists of a porous metallic matrix obtained by coating sintered particles to a metallic substrate. For porous coated surfaces, the geometric variables considered by Webb [98] are the following:

1. the particle size or size distribution,

- 2. the particle shape,
- 3. the coating thickness,

and 4. the particle packing arrangement.

Size or size distribution and the shape of particles used for the coating determine the packing arrangement, which in turn affects the porosity of the matrix. The porous coating of High Flux has a wide size distribution of non-spherical particles. The thickness of the coating has been reported to be about 0.25 mm to 0.50 mm [37] and the porosity between 0.50 and 0.65 [98].

O'Neill et al.[37] suggested a boiling mechanism for the operation of the High Flux surface. On a porous coated boiling surface the vaporization process is assumed to occur within the porous matrix as opposed to a conventional smooth surface where the vaporization process takes place on the outside surface. O'Neill envisioned that vapor

2 e â: θX , Û ٥ţ **L**á (3) ٤od Usi <u>50[]</u> spłe Para <sup>obta</sup>i bubbles exist within the pores of the metallic matrix and that liquid films exist on the surface of the particles. A two resistance theory was thus proposed by O Neill. One of the resistances is the nucleation superheat required for the stability of a vapor nucleus,  $\Delta T$  = sat and the other is the liquid film conduction  $2\sigma/r(dP/dT)$ resistance. Heat is conducted through the particle matrix and then transformed through the liquid film where evaporation occurs at the free interface of vapor and liquid. The structure of the surface allows the pores to be interconnected. The generation of the vapor bubble within a pore results in a pressure increase in the bubble. A large enough pressure inside the bubble overcomes the surface tension force and vapor forces its way out through the interconnected pores to the exterior of the surface. Webb [98] argues the validity of some of O Neill's assumptions, specifically, the assumption of precise stacking of particles and the assumption that each pore contains a vapor bubble.

Czikk and O'Neill [14] later classified the pores within the matrix into four categories; (1) active pores, (2) intermittent pores, (3) liquid filled pores, and (4) non-functional pores. An analytical model was proposed using the two-resistance theory.

Nishikawa and Ito [36] later developed an empirical correlation using a linear multiple regression technique on their data obtained for boiling R-11, R-113 and benzene on a porous surface made from uniform spherical particles. Substituting the values of the geometrical parameters of the High Flux surface in O<sup>'</sup>Neill's model, Webb [98] obtained an expression for the wall superheat as

$$T_{w} - T_{sat} = 0.044 \frac{d_{p}^{2} \dot{q}}{\lambda_{1} \delta_{c}} + 9.66 \frac{\sigma}{d_{p} (dP/dT)_{sat}}$$
 (2.39)

Webb compares the predictions of O'Neill's model and that of an empirical correlation obtained by Nishikawa and Ito [36] with the experimental data of refrigerants at a single heat flux level. The agreement is not very satisfactory. One of the reasons is the error associated with temperature measurement for the wall superheat, which has is a larger percentage of  $\Delta T$  at low heat fluxes.

The performance of the enhanced surfaces has also provoked their use to the boiling of mixtures. Previous work on mixture boiling on enhanced surfaces has been reported by Czikk et al. [15], Ali and Thome [2], Uhlig and Thome [84], and Shakir et al. [50]. The surface used in all of these studies was the High Flux. The boiling heat transfer coefficients were observed to be much higher for mixture boiling on the enhanced surface compared to the results for the conventional smooth surfaces operating under the same conditions. Czikk et al. [15] developed an empirical correlation using their data from eight different mixture systems at various pressures. At a given heat flux the wall superheat is expressed in the form:

$$\frac{\Delta T}{\Delta T_{id}} = 1 + \frac{\Delta T_c}{\Delta T_{id}}$$
(2.40)

where  $\Delta T_{id}$  is obtained from a linear molar mixing law (Eq. (2.33)) at the given heat flux and  $\Delta T_c$  is the deviation from ideality. The ratio  $\Delta T_c / \Delta T_{id}$  is expressed as a function of composition, heat flux, vapor mass velocity, transport properties and physical properties. These properties for mixtures are not easily obtained. The correlation is therefore of very little use unless the required quantities are known. Uhlig and Thome [84] tried using the Schluender correlation for predicting the acetone-water mixture boiling data by varying the value of  $B_o$ . No single value of  $B_o$  was able to predict their data reasonably well at all compositions. Ali and Thome [2] reported the wall superheats required for the initiation of boiling of ethanol-benzene mixtures on the High Flux surface. These superheats were found to be a strong function of composition. To the author's knowledge, no other data has been reported for the boiling incipient superheats of mixtures on enhanced surfaces.
### **CHAPTER 3**

#### EXPERIMENTAL FACILITIES AND PROCEDURES

A facility for the purpose of experimental studies in nucleate pool boiling in binary mixtures was built and developed at Michigan State University. The apparatus shown schematically in Fig. 3.1 was used for obtaining boiling incipient superheats and heat transfer coefficients.

## 3.1 BOILING FACILITY

# 3.1.1 Boiling Vessel and Attachments

The pressure vessel, shown in Figs. 3.1 and 3.2, is a 25.4 mm thick stainless steel chamber with flanged ends. The diameter of each of the four flanged openings is 101.6 mm. The volumetric capacity of the vessel is 4 liters. The top and bottom openings were bolted with 25.4 mm thick stainless steel cover plates. Various openings were provided on these cover plates to install the necessary attachments. A third flanged end was used for mounting the test section. The fourth opening on the vessel was used as a view port for observing the boiling process on the test surface. A high quality glass was used for this purpose. Stainless Steel "Swagelok" fittings were used on the cover plates for mounting attachments. The vessel assembly was supported by a rugged structure fabricated from 50 mm diameter structural pipe. As the experiments required temperatures higher than the ambient, the vessel assembly was insulated on the outside to minimize heat losses from the test liquid inside the vessel to the ambient.

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Figure 3.1. Nucleate pool boiling rig (not to scale). Legend: 1-boiling vessel, 2-test boiling surface, 3-bulk liquid thermocouple, 4-temperature controller thermocouple, 5-immersion heater, 6-sight glass window, 7-liquid feed line, 8-condenser, 9-degassing line, 10-safety relief valve line, 11-pressure gage tap.



Figure 3.2 Nucleate pool boiling rig.

The 1350 watt immersion heater, used to heat up the liquid pool and to maintain it at the saturation conditions, was coiled from a 6.35 mm O.D straight length of a "Chromalox" incoloy sheath heater. The condenser mounted on the inside of the top cover plate was made from 6.35 mm O.D stainless steel tubing. Teflon O-rings were used with the four flanges and the sleeve for sealing purposes. A pressure safety valve was mounted to relieve any accidental overshoot of pressure inside the vessel. Two identical 30-gage copper-constantan thermocouples, shielded by 3.2 mm O.D stainless steel sheath, were employed to measure the temperature of the liquid pool.

# 3.1.2 Equipment and Instrumentation

The power to the immersion heater was provided through an Omega model 58 proportioning temperature controller unit which senses the temperature of the liquid pool through one of the bulk thermocouples. Figure 3.3(a) shows the electrical circuit used for this purpose. Power to the test section was supplied through a variable power unit. Figure 3.3(b) shows the electrical circuit for the different test sections used in this experimental study. The power to the test heater was determined by measuring the current in the circuit and the local voltage drop across the test heater. The current was determined by measuring the voltage drop across a known resistance, 1.0 m-ohm ( $\pm$ 1.0%) in this case. A Keithley 177 Microvolt digital multimeter was used to measure the voltages. This multimeter was checked against several other voltmeters to verify its calibration.

Omega Trendicator model 400A digital temperature indicators, accurate to within  $\pm 0.1^{\circ}$ C, were used to measure the thermocouple temperatures. Calibrations of the these devices were periodically checked to assure good results. A Northrup potentiometer was used as a



(a)



Figure 3.3. Electrical circuit for : (a) the immersion heater to heat bulk liquid. (b) supplying power to the test section.

known voltage source, double checked using the Keithley multimeter. Also, temperatures were checked under equilibrium conditions to verify that all thermocouples read the same value.

The vapor pressure in the closed vessel was measured using a 50 psia (3.4 bar) "Heise" dial gage accurate to within  $\pm 0.2$ % of full scale ( $\pm 0.1$  psia). The dial gage was also "zeroed" at atmospheric pressure against the local corrected value of a mercury barometer. The flow rate of cooling water to the condenser was controlled manually using several precision values. Line pressure was sufficient for operation of the condenser.

# 3.2 TEST SECTIONS

Three different test sections were used for the experimental measurements carried out. The first test section was a smooth disk placed horizontally and facing upwards, Fig. 3.4. Ethanol-water and ethanol-benzene mixture systems were tested on this geometry. Two different materials were used for the discs, oxygen-free copper for ethanol-water mixtures and brass for ethanol-benzene mixtures. In each case the surface was treated with a 320 grade emery paper.

An electrical resistance heater ( $\approx$ 1.0 ohm) was specifically designed to provide the necessary heat flux to the disk. A single strip of nichrome was wrapped around a two-piece copper spool. High thermal conductivity epoxy was then filled inside the spool and cured in an oven at high temperatures. The heater assembly was then tightly screwed to the bottom of the disk. A film of high conductivity paste was applied between the heater and disk for a better thermal contact.

The second test section was of a tubular geometry with a Chromalox CIR 3030 electrical resistance heater tightly fitted inside the copper tube, Fig. 3.5. The test surface was positioned horizontally inside the

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Figure 3.4. (a) Smooth disk test surface; (b) Smooth disk test section. Legend: 1-test boiling surface, 2-electrical heater, 3-support section, 4-bottom cover plate, 5-thermocouple holes, 6-power input leads. (Dimensions in mm).



Smooth copper tube test section. Lagenti 1-test bolling surface, 2-cartridge heater, J-stenhiess steel slewer to hold bolling surface, 4-holes for thermocouple leads, 5-power input leads, 6-stainless steel end plug. (Dimensions in mm). Figure 3.5.



Figure 3.6. RMS surface measurement for the smooth disk (R2). (a) near the center; (b) away from the center. (Units: x-axis - micrometer, y-axis - kiloAngstrom).



Figure 3.7. RMS surface measurement for the smooth tube (R1). (a) near the center; (b) away from the center. (Units: xaxis - micrometer, y-axis - kiloAngstrom).



High Flux test boiling tube (not to scale). Ligend: 1-High Flux tube, 2-surface confing, 3-copper alseve, 4-tefion plug, 5-epoxy, 6-sleve to hold boiling surface, 7-thermocouple leads, 8-surtudge heater, 9-power input leads, 10-holes for thermocouple leads, 11-thermocouple junction. (dimensions in mm). Figure 3.8.

vessel. Four thermocouples were inserted axially in the wall of the tube and the temperature of the surface was extrapolated from their readings. Details of the calculations are given in Appendix B. The heater inside the copper tube was powered by an a.c. variac. All four mixture systems, i.e. methanol-water, ethanol-water, propanol-water and ethanol-benzene, were tested on this surface. The surface of this tube was also treated with a 400 grade emery paper. A polished tube (mirror finish) of the same geometry was also used to test only the ethanol-water mixture system.

RMS surface roughness measurements of the disk and the tube (after being roughened with emery paper) are shown in Figs. 3.6 and 3.7. These measurements were taken at University of Notre Dame.

The third test section was an enhanced tube made out of High Flux tubing, Fig. 3.8. Methanol-water, ethanol-water and propanol-water mixture systems were tested on this surface. The ethanol-benzene mixture system was already investigated by Ali and Thome [2] on the same surface and was not repeated.

## 3.3 CONTACT ANGLE FACILITY AND MEASUREMENTS

Contact angles were measured for all four binary mixture systems investigated in this work. Measurements were made on smooth surfaces of brass and copper. The sessile drop method was employed using a NRL (Naval Research Laboratory) Contact Angle Goniometer Model 100 manufactured by Rame Hart, USA. This model is comprised of an optical bench on which are mounted a microscope assembly with a magnification of 23X, a specimen stage and a variable intensity illuminator. Two independently rotatable cross hairs are provided in the goniometer head to allow direct reading of the contact angle in degrees. A stainless steel chamber (similar to that in Fig. 3.1) was specifically fabricated to obtain the controlled conditions (temperature, pressure and composition ) required for contact angle measurements of mixtures. The test surface was made by soldering two semi-circular pieces of brass and copper to form a circular disk 50 mm in diameter and 2 mm thick. It was first thoroughly cleaned with acetone and then with double distilled water before placing on a spool inside the chamber. The chamber was placed on the stage between the light source and the microscope. Using fine adjustment screws the test surface was positioned perfectly horizontal. High quality optical flats (faces parallel to one-millionth of an inch, "Van-Keuren," USA) were mounted on the chamber for viewing purposes. Figure 3.9 shows a picture of the assembly.

The chamber was first sealed, then evacuated using the vacuum pump and finally filled with nitrogen gas. The temperature inside the chamber was maintained at 25°C by using a very small heater (75 watts). Two thermocouples were placed very near the surface to measure the gas temperature. A sensitive diaphragm pressure regulator was used in the nitrogen line to maintain the pressure at 1.01 bar. Any excessive pressure was released manually through a vent valve. A micrometer syringe assembly (2.0 ml capacity) was mounted on the top of the chamber to introduce a liquid droplet of 0.002 ml on the test surface.

When observed on the microscope, the droplet appears as a silhouette against a soft green background. The liquid/solid interface is aligned with the horizontal cross hair, and the contact angle is determined by rotating the read-out crosshair to tangency with the drop profile at its base. The value of the contact angle is directly read on the goniometer scale which is calibrated in 1-degree increments. For each mixture composition an average of 36 measurements were made. The



Figure 3.9. Contact angle measurement apparatus.

error in measurement was  $\pm 0.5$  degrees and the reproducibility of the readings was better than  $\pm 2$  degrees.

## 3.4 MIXTURE PREPARATIONS

The binary mixtures were prepared on a weight basis using a sensitive balance. Double distilled water and reagent grades of methanol, ethanol, propanol, and benzene were used to prepare the mixtures. The density of a sample of each mixture was determined before and after each experiment to check if any changes in the mixture composition took place due to degassing of the system. The results showed negligible variation (within the error of the measurement of density).

# 3.5 EXPERIMENTAL PROCEDURE

The vessel was first evacuated using an Edwards two-stage vacuum pump to a pressure between 0.1 and 0.0 psia. The mixture was then fed in to the vessel through the liquid feed line with all other openings completely closed. As the liquid reached the desired level, 8 to 10 cm above the test section, the valve on the feed line was closed. The set point on the temperature controller was moved to a point corresponding to the saturation temperature of the mixture. Power to the temperature controller and hence to the immersion heater was then turned on. Once the pressure inside the vessel exceeded 1.01 bar the valve on the degassing line was slowly opened at discrete intervals until saturation conditions, temperature and corresponding pressure (1.01 bar), were reached. Power to the test section was then supplied. The heat flux was slowly increased to a moderately high value, approximately 175-200 kW/m<sup>2</sup>, and the surface was allowed to vigorously boil for 15 to 20 minutes. The saturation conditions were maintained by manually

controlling the coolant flow rate to the condenser. The heat flux to the test section was then slowly decreased to zero and the surface was allowed to come to thermal equilibrium with the saturated bulk liquid. Extra care was taken during this cool-down to maintain the pressure at its saturation value (1.01 bar in this case) to avoid any affect of pre-pressurization on the subsequent process. The heat flux was then increased in small increments after steady state was achieved at each step until a single boiling site was observed to be activated or complete surface activation occurred. The temperature readings from the thermocouples embedded in the wall of the test surface were displayed on the temperature readout devices and were recorded at each heat flux setting. The power to the surface was obtained from the voltage and current measurements. Complete surface activation was accompanied by a sudden drop in the surface temperature. After activation occurred, the heat flux was gradually increased to its operating maximum and then lowered to its minimum to obtain the boiling curve and the wall superheat of boiling deactivation on the test surface. This process was repeated at least twice for obtaining heat transfer coefficients and a minimum of three times for incipience and deactivation superheats.

#### **CHAPTER 4**

#### PHASE EQUILIBRIUM AND MIXTURE PHYSICAL PROPERTIES

## 4.1 VAPOR-LIQUID PHASE EQUILIBRIUM

The process of mixture boiling requires an understanding of the thermodynamics of vapor-liquid phase equilibria. Only a summary is given here. The factor that makes thermodynamics of mixtures different from that of pure substances is the differing compositions in the liquid and vapor phase that may still be in equilibrium with each other. The pertinent parameters for mixtures are then pressure, temperature and composition of every component in each phase.Conservation of mass requires

$$\sum_{i} \tilde{x}_{i} = 1.0 \qquad i = 1, 2, \dots, n \qquad (4.1)$$

$$\sum_{i} \tilde{y}_{i} = 1.0 \qquad i = 1, 2, \dots, n \qquad (4.2)$$

here  $\tilde{x}_i$  and  $\tilde{y}_i$  are the mole fractions of component i in the liquid and vapor phases, respectively. For a binary mixture, i=2. The number of independent variables to completely specify the state of the system can be determined from the phase rule for non-reactive components. Consider a system containing C distinct components and let P be the number of phases in which each of the components is present. The degree of freedom of the system, F, i.e. the number of variables that can be independently chosen, is given by:

$$F = C + 2 - P$$
 (4.3)

In a binary mixture system existing in liquid and vapor phases the value for F=2+2-2=2. If pressure and liquid composition are chosen, then the vapor composition and saturation temperature are uniquely defined. The relationship between these parameters is very conveniently expressed in the form of a Phase-Equilibrium diagram. Figure 4.1 is an example of such a diagram for a binary mixture system at any constant pressure. In this diagram the mole fraction of the more volatile component (the one with the lower boiling point) is plotted as the abscissa and the temperature as the ordinate. A constant temperature line, T<sub>sat</sub>, will intersect the bubble point line and the dew point line at two distinct points, labelled as L and V in Fig. 4.1(a). Point L corresponds to the composition of the more volatile component in the liquid phase and point V corresponds to the composition of the same component in the vapor phase. It is noted here that at any intermediate composition the volatile component is above and the non-volatile component is below its boiling point. This results in a vapor phase which is richer in the volatile component since it vaporizes more readily.

Some binary mixture systems exhibit a behavior called "azeotropy," as shown in Fig. 4.1(b). The point where the mole fraction of component i is same in both phases is the azeotropic composition. Hence, at an azeotrope the mixture system behaves as a single component.

Most mixture systems are non-ideal in their behaviour. The measure of ideality of mixtures is determined by Raoult's law. It states that the partial pressure ( $P_i$ ) of component i is related to its mole fraction ( $\tilde{x}_i$ ) and the vapor pressure of pure component i ( $P_i^0$ ) at the same temperature

$$P_{i} - P_{i}^{0} \tilde{x}_{i}$$
(4.4)





Figure 4.1. Phase equilibrium diagram for a binary mixture system. (a) Ideal mixture; (b) azeotropic mixture.



Figure 4.2. Phase equilibrium diagram showing the maximum decrease in the local mole fraction of the volatile component.



Figure 4.3. Phase equilirium diagram for methanol-water at 1.01 bar.



Figure 4.4. Difference in the vapor and liquid mole fractions for methanol-water at 1.01 bar; (a) Equilibrium conditions; (b) local conditions.



Figure 4.5. Phase equilirium diagram for ethanol-water at 1.01 bar.



Figure 4.6. Difference in the vapor and liquid mole fractions for ethanol-water at 1.01 bar; (a) Equilibrium conditions; (b) local conditions.



Figure 4.7. Phase equilirium diagram for n-propanol-water at 1.01 bar.



Figure 4.8. Difference in the vapor and liquid mole fractions for npropanol-water at 1.01 bar; (a) Equilibrium conditions; (b) local conditions.



Figure 4.9. Phase equilirium diagram for ethanol-benzene at 1.01 bar.



Figure 4.10. Difference in the vapor and liquid mole fractions for ethanol-benzene at 1.01 bar; (a) Equilibrium conditions; (b) local conditions.

This is only true for large values of  $\tilde{x}_i$ . In a vapor mixture the partial pressure of a component is also expressed as

$$P_{i} - \tilde{y}_{i} P \qquad (4.5)$$

where P is the system total pressure. Combining Eqs. (4.4) and (4.5)

$$\tilde{y}_{i} = \left(\frac{P_{i}^{0}}{P}\right) \tilde{x}_{i}$$
(4.6)

Using Eq. (4.6), the phase equilibrium diagram for an ideal mixture system can be easily obtained. However, for mixtures that do not obey Raoult's law, i.e. non-ideal mixtures, the methods for predicting equilibrium states is fairly complicated [44].

The phase equilibrium data for all the mixture systems investigated in this work are well documented at 1.01 bar, the pressure at which all experiments were carried out. Figures 4.3 - 4.10 show the phase-equilibrium data for all four binary mixture systems. The data has been obtained from [9,31,77]. The difference between the vapor and liquid mole fractions at equilibrium and local conditions can be interpreted with the help of Fig. 4.2.

# 4.2 MIXTURE PROPERTIES

The need for systematically determined and reasonably accurate mixture properties for use in mixture boiling cannot be overemphasized. Most of the existing sources either report values at temperatures other than saturation or correlations have to be used that have a limited range of applicability. A generalized corresponding states method, proposed by Teja [62,63], has been used to predict binary mixture physical properties.

### 4.2.1 Generalized Corresponding States Method

A pure fluid with critical parameters  $T_c$ ,  $P_c$ ,  $\tilde{v}_c$  and molecular weight M is defined to be in corresponding states with a reference fluid if the compressibility Z and the reduced property of the two substances at the same reduced temperature,  $T_r$ , and reduced pressure,  $P_r$ , are given by

$$Z = Z^{(0)}$$
 (4.7)

and

$$(A\Omega) = (A\Omega)^{(0)}$$
(4.8)

where superscript 'o' refers to the reference fluid, A is a property and (A $\Omega$ ) is the reduced property.  $\Omega$  is a function of critical parameters and is defined for each property.

The method proposed by Teja for estimating mixture properties is based on the known properties of the two reference fluids and requires the critical properties and the acentric factors of the pure components that form the mixture. The reference fluids can be the pure component themselves. Very accurate predictions (well within the accuracy of the experimental data) were reported for mixtures. Only one adjustable coefficient to characterize each binary system is required. The method is very simple to use and can be very easily extended to multicomponent mixtures.

The principle which is extended by Teja was originally proposed by Pitzer et al. [41]. The new proposed equation has the form

$$Z = Z^{(r1)} + \frac{\omega - \omega^{(r1)}}{\omega^{(r2)} - \omega^{(r1)}} \left[ Z^{(r2)} - Z^{(r1)} \right]$$
(4.9)

where Z is the compressibility factor,  $\omega$  is the acentric factor and superscripts rl and r2 refer to the two reference fluids which are chosen so that they are similar to the pure component of interest, or in the case of mixtures to the key components of interest. The reduced property (AΩ) of any (non-spherical) fluid (with critical parameters  $T_c$ ,  $P_c$ ,  $\tilde{v}_c$ , acentric factor  $\omega$  and molecular weight M) may be obtained from a knowledge of the reduced properties  $(A\Omega)^{(r1)}$  and  $(A\Omega)^{(r2)}$  of two reference substances (r1) and (r2) at the same reduced temperature  $T_r$ and reduced pressure  $P_r$  by means of the relationship

$$A\Omega = (A\Omega)^{(r1)} + \frac{\omega - \omega^{(r1)}}{\omega^{(r2)} - \omega^{(r1)}} \left[ (A\Omega)^{(r2)} - (A\Omega)^{(r1)} \right]$$
(4.10)

Equation (4.10) may be extended for mixtures using van der Waals onefluid model to replace  $T_c$ ,  $\tilde{v}_c$ , M and  $\omega$  of a pure fluid by the pseudocritical properties  $T_{cm}$ ,  $\tilde{v}_{cm}$ , M<sub>m</sub> and  $\omega_m$  of a hypothetical equivalent substance:

$$T_{cm} \tilde{v}_{cm} = \sum_{i} \sum_{j} \tilde{x}_{i} \tilde{x}_{j} T_{c_{ij}} \tilde{v}_{c_{ij}}$$
(4.11)

$$\tilde{\mathbf{v}}_{cm} - \sum_{i} \sum_{j} \tilde{\mathbf{x}}_{i} \tilde{\mathbf{x}}_{j} \tilde{\mathbf{v}}_{c_{ij}}$$
(4.12)

$$\omega_{\rm m} = \sum_{\rm i} \tilde{x}_{\rm i} \omega_{\rm i} \qquad (4.13)$$

$$M_{\rm m} = \sum_{i} \tilde{x}_{i} \omega_{i} \qquad (4.14)$$

The one fluid model can be used to obtain the properties of mixtures provided the cross parameters  $T_{c}$  and  $\tilde{v}_{c}$  (i=j) are specified. The mixing rules for this purpose are as follows:

$$\Gamma_{c_{ij}} = \xi_{ij} \sqrt{T_{ii} \tilde{v}_{ii} T_{jj} \tilde{v}_{jj}}$$
(4.15)

$$\tilde{v}_{c_{ij}} = \frac{1}{8} \left[ \tilde{v}_{c_{ii}}^{1/3} + \tilde{v}_{c_{jj}}^{1/3} \right]^{3}$$
(4.16)

where  $\Rightarrow_{ij}$  is a binary interaction coefficient which has to be calculated from experimental data. This coefficient, independent of temperature and composition, is sufficient to characterize each binary mixture. For the case when the two pure components in each binary mixture are used as the reference fluids, this coefficient in part reflects the inability of representing the intermolecular forces via Eqs. (4.10) - (4.16). However, for non-aqueous mixtures, the use of  $=_{ij}$ -1.0 has been claimed to yield good agreement between calculated and experimental values. For binary mixture systems, the simplified forms of Eqs. (4.11) - (4.16) can be written as:

$$T_{cm}\tilde{v}_{cm} = \tilde{x}_{1}^{2} T_{c_{11}}\tilde{v}_{c_{11}} + 2\tilde{x}_{1}\tilde{x}_{2}T_{c_{12}}\tilde{v}_{c_{12}} + \tilde{x}_{2}^{2} T_{c_{22}}\tilde{v}_{c_{22}}$$
(4.17)

$$\tilde{v}_{cm} = \tilde{x}_{1}^{2} v_{c_{11}}^{2} + 2\tilde{x}_{1}\tilde{x}_{2}\tilde{v}_{c_{12}} + \tilde{x}_{2}^{2} \tilde{v}_{c_{22}}$$
(4.18)

$$\widetilde{\omega}_{\rm m} = \widetilde{x}_1 \omega_1 + \widetilde{x}_2 \omega_2 \tag{4.19}$$

$$M_{\rm m} - \tilde{x}_1 M_1 + \tilde{x}_2 M_2 \tag{4.20}$$

The second term on the right side of Eq. (4.10) can also be simplified in the following manner

$$\frac{\omega - \omega^{(r1)}}{\omega^{(r2)} - \omega^{(r1)}} = \frac{\tilde{x}_1 \omega_1 + \tilde{x}_2 \omega_2 - \omega^{(r1)}}{\omega^{(r2)} - \omega^{(r1)}}$$
  
Since  $\omega^{(r1)} = \omega_1$ ,  $\omega^{(r2)} = \omega_2$  and  $\tilde{x}_1 = 1 - \tilde{x}_2$ ,  
$$\frac{\omega - \omega^{(r1)}}{\omega^{(r2)} - \omega^{(r1)}} = \tilde{x}_2$$
(4.21)

Substituting Eq. (4.21) in Eq. (4.10) yields

$$A\Omega = (A\Omega)^{(r1)} + \tilde{x}_{2} \left[ (A\Omega)^{(r2)} - (A\Omega)^{(r1)} \right]$$
$$A\Omega = \tilde{x}_{1} (A\Omega)^{(r1)} + \tilde{x}_{2} (A\Omega)^{(r2)}$$
(4.22)

Hence, the acentric factor for the pure components need not be estimated. The reduced property  $(A\Omega)$  of the <u>pure components</u> has to be obtained from any suitable equation or correlation and Eq. (4.22) is then used to estimate the mixture property at any desired composition.

# 4.2.2 Method of Calculation for Binary Mixtures.

Given  $T_c$ ,  $\tilde{v}_c$  and M for each component, the calculation of physical properties will proceed as follows:

1. Calculate pseudocritical quantities  $T_{cm}$ ,  $\tilde{v}_{cm}$  and  $M_m$  using Eqs. (4.17), (4.18) and (4.20). The value of the binary interaction coefficient  $\xi_{ij}$  may be set initially equal to 1.0.

2. Calculate 
$$T_r = \frac{T}{T_{cm}}$$
.

- 3. Calculate the reduced properties (A\Omega) of the pure components at  ${\rm T}_{\rm r}.$
- 4. Calculate A $\Omega$  for the given mixture system from Eq. (4.22).
- 5. Calculate  $\Omega$  (using mixture pseudocritical parameters  $M_m$ ,  $\tilde{v}_{cm}$ and  $T_{cm}$ ) and obtain the property A.
- 6. If some experimental data are available, the coefficient ≠<sub>ij</sub> can be varied until the difference between the calculated and experimental value is minimized. No iterations are required.

The correlations used for pure components and expressions for  $\Omega$  are given below for each property. The binary interaction coefficients obtained from Ref. [67] are listed in Table 4.1 at the end of the chapter.

**Thermal conductivity-liquid phase**. The correlation used for estimating thermal conductivity of pure components is [65]:

$$\lambda_{\ell} \Omega = A + B(1 - T_{r})^{2/3} + C(1 - T_{r})^{4/3} \qquad [\lambda_{\ell}] = 10^{-3} \frac{J}{s m K} \qquad (4.23)$$
  
and 
$$\Omega = M^{1/2} \tilde{v}_{c}^{2/3} T_{c}^{-1/2} \qquad [\tilde{v}_{c}] = \frac{m}{kmol}, \ [T_{c}] = K$$

Constants A, B and C for the pure components of interest are [67]:



Figure 4.11. Variation in liquid thermal conductivity with composition at saturation conditions.

	METHANOL	ETHANOL	PROPANOL	BENZENE	WATER
A	5.39498	6.33533	11.6854	1.39874	-9.67795
В	11.94140	15.55490	12.7422	3.22771	100.5070
С	0.0	0.0	0.0	0.0	95.6611

For thermal conductivity of aqueous mixtures Teja [65,67] suggests the use of mass fractions instead of mole fractions. Figure 4.11 shows the predicted variation in thermal conductivity with liquid composition for the mixtures. The points on the graphs of physical properties (Figs. 4.11-4.17) represent calculated values and are joined by straight line only for illustrative purposes.

<u>Viscosity-liquid phase</u>. The correlation used for estimating the liquid viscosity of pure components is [64]:

$\ln(\eta_{\ell}\Omega) = A + \frac{B}{T_{r}}$	$[\eta_{\ell}] = cP = 10^{-3} \frac{N-s}{2}$	(4.24)
$\Omega - M^{-1/2} \tilde{v}_c^{2/3} T_c^{-1/2}$	$[\tilde{v}_c] = \frac{m}{kmol}$	

and

Constants A and B for the pure components of interest are [46]:

	METHANOL	ETHANOL	PROPANOL	BENZENE	WATER
A	-11.19289	-11.49610	11.68540	-12.87967	-11.94228
В	2.49172	3.06256	4.07907	2.23317	2.34154

The predicted variations in the values of viscosity are shown in Fig. 4.12.

Heat capacity-liquid phase. The heat capacities of the pure components were obtained from the following correlation [66]:

$$\ln\left[\frac{c_{p_{\ell}}}{R}\right] - A - \frac{B}{T_{r}} \qquad \left[c_{p_{\ell}}\right] - \frac{J}{kg} \qquad (4.25)$$

R is the gas constant (- 8314.0  $\frac{J}{kmol-K}$ ), and the constants are [67]:



Figure 4.12. Variation in liquid viscosity with composition at saturation conditions.


Figure 4.13. Variation in liquid heat capacity with composition at saturation conditions.

	METHANOL	ETHANOL	PROPANOL	BENZENE	WATER
А	2.8410	4.2453	3.6893	3.3498	2.2178
В	0.3315	0.9565	0.4673	0.2985	0.0065

The predicted variations of heat capacity with composition are shown in Fig. 4.13.

**Surface tension**. The reduced surface tension of the reference fluids [47] was correlated by means of the equation:

$\sigma \Omega = A - B T_r$	$[\sigma] - 10^{-3} \frac{M}{m}$	(4.26)
$\tilde{v}_{c}^{2/3}$	3 (~ ) _ <u>m</u>	
$\Omega = \frac{T_{c}}{T_{c}}$	$\begin{bmatrix} v_c \end{bmatrix} = \frac{1}{\text{kmol}}$	

Constants A and B were obtained from the data in Ref. [29]:

and

	METHANOL	ETHANOL	PROPANOL	BENZENE	WATER
А	0.020955	0.027218	0.031569	0.048187	0.026642
В	0.018424	0.024993	0.028329	0.052465	0.021925

Figure 4.14(a) shows the variations in surface tension obtained using Eq. (4.26).

The surface tension values for aqueous mixture systems were also obtained by the method of Tamura, Kurata and Odani [given in Ref. 46]. The final correlation is given in the form:

$$\sigma = s_{w}^{*} \sigma_{w}^{0.25} + s_{o}^{*} \sigma_{o}^{0.25} \qquad [\sigma] = 10^{-3} \frac{N}{m} \qquad (4.27)$$

where subscript w and o refer to water and organic compound, respectively.  $s^*$  is the superficial volume fraction in the surface layer. Figure 4.14(b) shows the surface tension variation using Eq. (4.27) for aqueous mixture systems. It should be noted that Fig. 4.14(b) predicts a drastic drop in surface tension values of aqueous mixtures with the addition of small amounts of an organic to water.



Figure 4.14. Variation in surface tension with composition at saturation conditions. (a) Generalized corresponding states method; (b) Eq. (4.27).

**Liquid density**. The equation proposed for estimating liquid densities of binary mixtures is [61]:

$$\left[\frac{z_{cm}}{\rho_{r}}\right] = \tilde{x}_{1} \left[\frac{z_{c}^{(r1)}}{\rho_{r}^{(r1)}}\right] + \tilde{x}_{2} \left[\frac{z_{c}^{(r2)}}{\rho_{r}^{(r2)}}\right]$$
(4.28)

1,2

where 
$$Z_{cm} = \sum_{i} \tilde{x}_{i} Z_{c}$$
  $i =$ 

The reduced saturated liquid densities,  $\rho_r^{(r1)}$  and  $\rho_r^{(r2)}$ , of the reference fluids were obtained using the Modified Rackett equation [54]:

$$\left[\frac{1}{\rho_{s}}\right] - \left[\frac{RT_{c}}{P_{c}}\right] Z_{RA}^{n} \qquad [\rho_{s}] - \frac{g}{cm} = 10^{-3} \frac{kg}{cm} \qquad (4.29)$$

where  $n = [1 + (1-T_r)^{2/7}]$  and R is the gas constant (=82.06  $\frac{atm-cm}{gmol-K}$ ). Z<sub>RA</sub> is a specified constant for each compound:

	METHANOL	ETHANOL	PROPANOL	BENZENE	WATER
z <sub>ra</sub>	0.23230	0.25041	0.25272	0.26967	0.24091

The predicted variations in liquid density are shown in Fig. 4.15.

Vapor density. In this case the density values correspond to the vapor mole fraction that is in equilibrium with the liquid. The Teja-Patel equation of state [39] was used to estimate vapor density of binary mixtures. This equation of state is cubic in nature and requires the critical temperature and pressure and two additional parameters to characterize each particular fluid. Patel and Teja [39] demonstrated that their equation not only reproduced many of the good features of the Soave [53] and Peng-Robinson [40] equations for non-polar fluids, but also overcomes some of their limitations for polar fluids. The Teja-Patel equation of state is of the form:



Figure 4.15. Variation in liquid density with composition at saturation conditions.

$$P = \frac{\widetilde{R}T}{\widetilde{v} - b} - \frac{a(T)}{\widetilde{v}(\widetilde{v} + b) + c(\widetilde{v} - b)}$$
(4.30)

where  $\tilde{R}$  is the universal gas constant, and a,b and c are given by:

$$a(T) = W_{a} \left[ \frac{\tilde{R}^{2} T_{c}^{2}}{P_{c}} \right] T(T_{r})$$
(4.31)

$$b = W_{b} \left[ \frac{\tilde{R}T_{c}}{P_{c}} \right]$$
(4.32)

$$c - W_{c} \left[ \frac{\tilde{R}T_{c}}{P_{c}} \right]$$
(4.33)

where

$$W_{c} = 1 - 3\Psi_{c}$$
 (4.34)

$$W_{b} = 3\Psi_{c}^{2} + 3(1-2\Psi_{c})W_{b} + W_{b}^{2} + 1 - 3\Psi_{c}$$
(4.35)

and  $W_h$  is the smallest positive root of:

$$W_{b}^{3} + (2 - 3\Psi_{c}) W_{b}^{2} + 3\Psi_{c}^{2} W_{b} - \Psi_{c}^{2} = 0$$
 (4.36)

 $\Psi_{\rm c}$  is the predicted value of the critical compressibility factor. For T(T\_r), the form of the expression is:

$$T(T_r) = [1 + F(1 - \sqrt{T_r})]^2$$
 (4.37)

The optimum values of  $\Psi_{c}$  and F correspond to the minimum deviation in saturated liquid densities and the equilibrium condition of equality of fugacities. For the fluids of interest, the values of  $\Psi_{c}$  and F are given as follows [39]:

	METHANOL	ETHANOL	PROPANOL	BENZENE	WATER
Ψ <sub>c</sub>	0.272	0.300	0.303	0.310	0.269
F	0.972708	1.2303395	1.241347	0.704657	0.689803

Compared with other equations of state, the Teja-Patel equation gave substantially better predictions of saturated liquid and vapor densities (for the compounds above, the maximum deviation is reported to be 3.27%). Equation (4.30) can be used for the estimation of mixture



Figure 4.16. Variation in vapor density with composition at saturation conditions.

properties if a, b and c are replaced by the mixture constants  $a_m$ ,  $b_m$  and  $c_m$  as follows [39]:

$$\mathbf{a}_{m} - \sum_{\mathbf{i}} \sum_{\mathbf{j}} \tilde{\mathbf{x}}_{\mathbf{i}} \tilde{\mathbf{x}}_{\mathbf{j}} \mathbf{a}_{\mathbf{i}\mathbf{j}}$$
(4.38)

$$\mathbf{b}_{m} - \sum_{i} \tilde{\mathbf{x}}_{i} \mathbf{b}_{i}$$
(4.39)

$$c_{m} = \sum_{i} \tilde{x}_{i} c_{i}$$
(4.40)

where  $a_{ij} - \xi_{ij} \sqrt{a_{ii}a_{jj}}$  (4.41)

The predicted variation in vapor density are shown in Fig. 4.16 (these are plotted against the liquid mole fraction).

**Enthalpy of vaporization.** The enthalpy of a mixture of constant composition is a function of both temperature and pressure. At any temperature, T, and pressure, P, the enthalpy of a mixture,  $H_m(T,P)$ , can be represented as:

$$H_{m}(T,P) = \sum_{i} \tilde{x}_{i} H_{i}^{0}(T,0) - H_{m}^{D}(T,P)$$
   
  $i = 1, 2, \dots, n$  (4.42)

where all  $H_i^0(T,0)$  refer to the same reference state, and the term  $H_m^D(T,P)$ , which is known as the isothermal departure, relates the mixture enthalpy at some pressure, P, to the enthalpy of a mixture of ideal gases at zero pressure. To calculate the enthalpy of vaporization of a binary mixture (differential latent heat, i.e. vapor and liquid phase that are in equilibrium with each other) it is necessary to calculate the enthalpy of the mixture corresponding to the vapor and liquid mole fractions separately. For integral latent heat (i.e. vapor and liquid phases corresponding to the same composition), it is sufficient to determine the isothermal enthalpy departures, since the

first term in Eq. (4.42) cancels out. Only differential latent heats are of interest when equilibrium is considered. The ideal gas enthalpy  $H_i^0(T,0)$  is estimated by integrating  $C_p$  from 0 K to the saturation temperature. A fourth order polynomial in T was used for  $C_p$  and constants for each pure component were obtained from Ref. [46]. The isothermal enthalpy departure is calculated using Teja-Patel equation of state, Eq. (4.30). From classical thermodynamics the following relation can be obtained:

$$dH = d(P\tilde{v}) + \left[T(\frac{\partial P}{\partial T})_{v} - P\right]_{T} dT \qquad (4.43)$$

Integration of Eq. (4.43) from zero pressure to the pressure of interest, P, (i.e from infinite volume to volume  $\tilde{v}$ ) gives the isothermal enthalpy departure

$$H_{m}^{D} = RT - P\tilde{v} + \int_{\infty}^{\tilde{v}} [P - T(\frac{\partial P}{\partial T}) - P] d\tilde{v} \qquad (4.44)$$

Now using Eq. (4.30), the enthalpy departure may be computed by performing the integration in Eq. (4.44). The resulting expression is:

$$(H - H^{0}) = f \left\{ RT(Z-1) - \left[ T(\frac{\partial a}{\partial T}) - a \right] \left[ \frac{1}{2N} \ln(\frac{Z+M}{Z+Q}) \right] \right\}$$
(4.45)

where f is a conversion factor, and

$$M = \left[\frac{b+c}{2} - N\right]\frac{P}{RT}$$
(4.46)

$$N = \left[bc + \left(\frac{b+c}{2}\right)^{2}\right]^{0.5}$$
(4.47)

$$Q = \left[\frac{b+c}{2} + N\right] \frac{P}{RT}$$
(4.48)

For mixtures a, b and c need to be replaced by  $a_m$ ,  $b_m$  and  $c_m$  given by Eqs. (4.38)-(4.40). The predicted values of enthalpy of vaporization are plotted in Fig. 4.17.



Figure 4.17. Variation in heat of vaporization with composition at saturation conditions.

Two computer programs were developed to predict the binary mixture properties. The first program, MIXPR1, estimates thermal conductivity, viscosity, heat capacity, surface tension and liquid density. The second program, MIXPR2, is based on the equation of state, Eq. (4.30), and predictions of vapor density and enthalpy of evaporation are obtained from it. Pure component critical data are obtained from [24]. A listing of the computer programs along with the input data files are given in Appendix A. Table A.3 lists all the properties obtained.

Table 4.1. Binary Interaction coefficients  $(\xi_{ij})$  for mixture properties [67]

	<i>ک</i> ر	$\eta_{\ell}$	с <sub>р</sub> г	σ	ρ <sub>v</sub>	${}^{\Delta h}v$
METHANOL-WATER	1.0	1.34	1.0	1.0	1.083	1.083
ETHANOL-WATER	1.4	1.36	1.0	1.0	1.075	1.075
PROPANOL-WATER	1.4	1.37	1.0	1.0	1.0	1.0
ETHANOL-BENZENE	1.0	1.0	1.0	1.0	1.0	1.0

### CHAPTER 5

#### BOILING INCIPIENCE IN BINARY MIXTURES

### 5.1 BOILING INCIPIENCE RESULTS AND DISCUSSION

Boiling incipient (activation) and deactivation superheats were experimentally measured to investigate their variation with mixture composition. Table 5.1 lists the binary mixture systems studied on different surfaces. The rms roughness measurements for the surfaces are given in Chapter 3. The boiling activation and deactivation superheats, Figs. 5.1-5.10, were observed to be complex functions of composition. The superheats were observed to be much smaller for the enhanced (High Flux) surface than for the plain smooth surfaces. Also, for smooth surfaces the activation superheats were much higher than their corresponding deactivation superheats and the two exhibit different trends in their behavior.

# Table 5.1 Summary of mixture systems studied on different surfaces

<u>Mixture System</u>	Surface	<u>Roughness</u>
Methanol-water	Smooth tube	R1
	Enhanced tube	
Ethanol-water	Smooth disk	R2
	Smooth tube	R1
	Polished tube	
	Enhanced tube	
n-Propanol-water	Smooth tube	R1
-	Enhanced tube	
Ethanol-benzene	Smooth disk	R2
	Smooth tube	R1

∆ T<sub>deact</sub>

ΔTine

FIg



Figure 5.1. Variation in measured superheats for methanol-water mixture system on smooth tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.



Figure 5.2. Variation in measured superheats for methanol-water mixture system on enhanced tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.



(Ь)

Figure 5.3. Variation in measured superheats for ethanol-water mixture system on smooth tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.



Figure 5.4. Variation in measured superheats for ethanol-water mixture system on polished tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.



Figure 5.5. Variation in measured superheats for ethanol-water mixture system on smooth disk: (a) boiling incipient superheats; (b) boiling deactivation superheats.



Figure 5.6. Variation in measured superheats for ethanol-water mixture system on enhanced tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.

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Figure 5.7. Variation in measured superheats for n-propanol-water mixture system on smooth tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.



Figure 5.8. Variation in measured superheats for n-propanol-water mixture system on enhanced tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.





Figure 5.9. Variation in measured superheats for ethanol-benzene mixture system on smooth tube: (a) boiling incipient superheats; (b) boiling deactivation superheats.





Figure 5.10. Variation in measured superheats for ethanol-benzene mixture system on smooth disk: (a) boiling incipient superheats; (b) boiling deactivation superheats.

The criterion for nucleation, i.e. the wall superheat required for the growth of a vapor nucleus is given by Eq. (2.12):

$$\Delta T = \frac{2\sigma}{r(dP/dT)_{sat}}$$
(2.12)

The variation of surface tension  $\sigma$  with respect to composition is shown in Fig. 4.14 and that of  $(dP/dT)_{sat}$  in Fig. 5.11(a). The trend in (dP/dT) is almost linear for aqueous mixture systems of methanolwater and ethanol-water. Hence, for n-propanol-water mixtures it is expected to follow a linear variation as well. The combined effect of  $\sigma$ and  $(dP/dT)_{sat}$  is shown in Fig. 5.11(b). For a constant value of radius, as its variation with composition is not known, it was expected that the wall superheat required for initiation of boiling would exhibit the same behaviour as  $2\sigma/[r(dP/dT)_{sat}]$ . According to Fig. 5.11(b) the incipient superheats required for aqueous mixture systems should monotonically decrease with increasing composition of the volatile component. Also, the incipient superheat for water is predicted to be about 3.5 times higher than that of the alcohols. For the ethanol-benzene mixture system, the variation is predicted to be almost linear. On the contrary, the experimental results show an opposite trend. The incipient superheat for water is always observed to be lower than that of the pure alcohols and, in general, a maximum is observed at an intermediate composition. The discrepancy is believed to be associated with the behaviour of radius r.

Considering the vapor trapping mechanism for pure liquids proposed by Lorenz, Mikic and Rohsenow [34] and illustrated in Fig. 2.4, the radius r is proportional to the contact angle. Contact angles were therefore measured for all four mixture systems against nitrogen at 25°C and 1.01 bar, Figs. 5.12-5.15. The contact angle data obtained for ethanol-water mixtures, Fig. 5.13, are very similar to those obtained



Figure 5.11. (a) Variation in the slope of the saturation curve at 1.01 bar; (b) variation in  $\frac{2\sigma}{(dP/dT)}$  with composition.



Figure 5.12. Variation in measured contact angles for methanol-water mixture system at 1.01 bar against nitrogen gas at 25°C.

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Figure 5.13. Variation in measured contact angles for ethanol-water mixture system at 1.01 bar against nitrogen gas at 25°C.



Figure 5.14. Variation in measured contact angles for n-propanol-water mixture system at 1.01 bar against nitrogen gas at 25°C.



Figure 5.15. Variation in measured contact angles for ethanol-benzene mixture system at 1.01 bar against nitrogen gas at 25°C.

by Eddington and Kenning [16] under the same conditions and shown in Fig. 2.6. At very high compositions of ethanol, their photographic method was inadequate to capture the smaller contact angle values. However, no such problem was encountered with the high quality goniometer employed in the present study. The measured contact angles for aqueous mixtures, shown in Figs. 5.12-14, were observed to be smaller than would be expected from a linear molar interpolation, i.e. a straight line drawn between the pure component values. The contact angles for ethanol-benzene mixtures, shown in Fig. 5.15, are seen to vary almost linearly with composition. The contact angle values of aqueous mixtures tend to decrease with increasing composition of the volatile component. Hence, if a trend similar to that observed for contact angle is assumed for radius r, then according to Eq. (2.12) the boiling incipient superheats should increase with the increasing composition of the volatile component. The overall effect of  $\sigma$ ,  $(dP/dT)_{rat}$  and r (assuming a behavior similar to that exhibited by the contact angle data) for any mixture system is still inadequate to explain the the maximum observed at an intermediate composition. However, this does explain why the pure alcohol incipient superheats are higher than those for pure water.

One of the important factors that has to be considered at this point is the experimental procedure adopted in measuring the incipient superheats. Details of this are given in Chapter 3 and only briefly mentioned here. Once the saturation conditions were obtained, the test surface was boiled vigorously at moderately high heat flux (150-200  $kW/m^2$ ) and slowly brought down to zero heat flux and allowed to come to thermal equilibrium with the surrounding liquid at its saturation temperature. Extra care was taken to maintain the pressure at 1.01 bar (at which all the experiments were performed) during this cooldown

process. During the decrease of heat flux, the boiling site density of the surface also decreased and at some finite wall superheat boiling completely ceased to occur on the surface (deactivation). The vapor that is trapped in any cavity on the surface was therefore a residual of the last departing bubble from that particular cavity. Hence, the vapor trapping process here is different from that due to spreading of a liquid front over vapor/gas filled cavities on the surface, as assumed by Lorenz et al. [34]. The reason for boiling the surface prior to taking measurements was to avoid the effect of the "history" of the surface on the boiling incipient superheats. Between any consecutive experimental runs the vessel was drained and evacuated to remove the traces of the previous mixture system. It was also temporarily filled with nitrogen gas when not in use to prevent oxidation, that could have occurred due to exposure to air. The effect of prepressurization has been shown to influence the incipient superheat values [75]. Hence, to provide a uniform basis for all the experimental results, the influence of all these factors was nullified by first boiling the surface.

The volume of the vapor left behind by the departing bubble is believed to be controlled by the height of the break-off point where the bubble separates from the residual vapor left behind (h). This process is illustrated in Fig. 5.16. The complex phenomenon of bubble separation from the boiling surface is governed by the dynamics of departure of the bubble. Surface tension is likely to be the controlling factor in this process. A higher value of the surface tension would result in a larger value of h. The buoyancy force of the bubble will also try to elongate the bubble during this departure process, possibly resulting in a higher value of h. From this reasoning, it follows that  $\sigma \propto h$ ,  $h \propto V$ ,  $V \propto r$ ,  $r \propto \Delta T_{sat}^{-1}$ . According to this proposed vapor trapping process,  $\Delta T_{sat} \propto 1/\sigma$ . The effect of  $\sigma$  here



Figure 5.16. Vapor trapping mechanism at the point of bubble breakoff. (a) bubble detaching from the surface; (b) vapor trapped in the cavity.

is the opposite of that in Eq. (2.12). The radius r of the vapor nucleus is then proportional to the contact angle. All cavities on the surface containing trapped vapor pockets can be potential sites for reactivation.

### 5.2 FACTORS AFFECTING BOILING INCIPIENT SUPERHEATS

## 5.2.1 Condensation Effect

It was pointed out in Chapter 2 that the preferential evaporation of the volatile component in the immediate vicinity of the growing bubble results in the depletion of the volatile component, thus creating a concentration gradient around the bubble. This process is illustrated in Fig. 2.9. The volatile component is preferentially evaporated to provide the vapor phase with the extra volatile component to maintain phase equilibrium. The composition of the volatile component in the vapor phase is thus governed by the rate at which it can diffuse through the liquid to the vapor-liquid interface. As a result, the local conditions near a growing bubble are changed. The extreme condition is achieved when the composition of the vapor is same as that of the liquid reaching the interface. This process is illustrated in Fig. 5.17. Point 1 represents the bulk liquid composition  $\tilde{x}_{b}$  and 1' is the corresponding vapor composition  $\tilde{y}_{b}$ . **Process 1**  $\rightarrow$  2 shows the dese in composition of the volatile component. Point 2' is the local vapor mole fraction  $\widetilde{y}_{\underline{L}}$  that is in equilibrium with local liquid mole fraction  $\tilde{\mathbf{x}}_{\mathrm{L}}$ . In the extreme situation, the composition of the vapor in the last bubble leaving the surface can be assumed to be at  $\tilde{y}_{_{I}}$  and therefore the vapor that is left behind will also be of the same composition. When the last bubble departs from a cavity on the surface, the residual vapor (mole fraction  $\boldsymbol{\tilde{y}}_{T}$  ) left in it is surrounded by the bulk liquid (mole fraction  $\boldsymbol{\tilde{x}})$  which



Figure 5.17. Condensation effect illustrated on a phase equilibrium diagram.

rushes to the surface. It can be seen from the phase-equilibrium diagram in Fig. 5.17 that these two phase are not in equilibrium with each other. A possible mechanism that can lead to a state of equilibrium is the condensation of the non-volatile component in the vapor phase. This process of condensation, shown by  $2' \rightarrow 1'$  in Fig. 5.17, continues until the vapor phase comes to equilibrium with the surrounding liquid phase. Another possible mechanism that would result in the same net effect can be the evaporation of the non-volatile component at the interface. This evaporation process is unlikely to occur since the heat of vaporization is required for evaporation of any amount of liquid and cannot be provided by the surface as it is no longer heated. If condensation in the vapor phase is assumed to be the mechanism for achieving equilibrium, then its effect would be to reduce the volume of the trapped vapor in the cavity. A reduction in volume will be accompanied by a decrease in the radius r of the vapor nucleus. This decrease in volume of the vapor is proportional to the distance between points 2 and 1 which is the difference between the vapor and bulk liquid compositions,  $\tilde{y}_b - \tilde{x}_b$ . Hence, for pure components and at azeotropic compositions this mechanism will not occur as  $(\tilde{y}_h - \tilde{x}_h) = 0$  for such cases.

## 5.2.2 Contact Angle Effect

Experimental data on contact angles obtained by Ponter and Peier [43], shown in Fig. 5.18, suggest that the two can be different when measured under equilibrium and total reflux conditions. Equilibrium refers to the situation when the liquid and vapor are in phase equilibrium with each other. Total reflux corresponds to the case where the vapor and liquid in contact have the same composition. The two situations are represented in Fig. 5.17 by points 1 and 1 and by 1 and





Figure 5.18. Variation in contact angle data reported by Ponter and Peier [43] under conditions of equilibrium and total reflux: (a) methanol-water; (b) n-propanol-water.

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2, respectively. The variation in contact angle values during the process  $2 \rightarrow 1$  will probably have some effect on the radius r. During the present investigation attempts to measure the contact angles under these two conditions proved unsuccessful. The bottom of the vessel specifically designed for measuring contact angles was filled with a liquid mixture to produce vapor of the desired composition. However, when saturation conditions were reached, the condensation on the inside of the optical flats made it impossible to view the surface inside the vessel. This problem was solved by blowing hot air on the outside of the optical flats. The second problem encountered was the difficulty in keeping the test surface dry. A thin film of condensate tended to cover the surface, making it impossible to take any measurements. A heater was provided to heat the surface and evaporate the liquid film from it. However, the slightly higher surface temperature caused the liquid droplet (used for measuring the contact angle) to vaporize when introduced on the surface for measurement. If the temperature of the surface was allowed to decrease, then the liquid film reappeared on it. This heating and cooling process also made it very difficult to maintain the pressure at the desired level. Thus, due to lack of proper conditions, it was not possible to obtain data. In fact, it is not clear if the Ponter and Peier data [43] are reliable since they may have had a problem with a liquid film without knowing it.

#### 5.2.3 Surface Tension Effect

Ponter and Peier [43] also obtained surface tension values corresponding to equilibrium and total reflux conditions and shown in Fig. 5.19. Their data for methanol-water and n-propanol-water mixtures, which are limited to small mole fractions, suggest that surface tension at total reflux can be much smaller than that corresponding to the

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Figure 5.19. Variation in surface-tension reported by Ponter and Peier [43] under conditions of equilibrium and total reflux: (a) methanol-water; (b) n-propanol-water.

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equilibrium condition. The surface tension of the vapor-liquid interface during the process  $2 \rightarrow 1$  will therefore change. It is also probable that a change in value of  $\sigma$ , due to either the temperature gradient over the bubble interface or due to a local change in composition, will have an effect on the height, h, and hence on the radius r.

## 5.3 TRENDS IN THE OBSERVED EXPERIMENTAL INCIPIENT SUPERHEATS

# 5.3.1 Ethanol-benzene mixtures

Of all the mixture systems investigated in this work, the ethanolbenzene mixture system shows the least variation in physical properties, especially surface tension, and measured contact angle. Hence, it can be used as a starting point for testing the validity of the suggested mechanisms and factors influencing boiling incipience. The experimentally measured incipient superheats for this mixture system, Figs. 5.9 and 5.10, which were obtained on two different smooth surfaces show a maximum on either side of the azeotrope. On the left of the azeotrope the maximum, on each surface, is in the vicinity of the maximum in  $(\tilde{y}_h - \tilde{x}_h)$  which occurs at mole fraction 0.15 of the volatile component, Fig. 4.10. On the right of the azeotrope the two surfaces exhibit maxima at different mole fractions which are fairly close to a mole fraction of 0.8 where the maximum in  $(\tilde{y}_b^{}-\tilde{x}_b^{})$  occurs on the right of the azeotrope. As the decrease in radius r is proportional to  $(\boldsymbol{\tilde{y}}_b^{}$  $ilde{\mathbf{x}}_{\mathbf{b}}$ ), the ethanol-benzene results seem to follow the suggested trend to a fair degree, i.e. the effect of condensation on the trapped vapor nucleus. The nature of initiation of boiling on the two surfaces was observed to be different from each other. On the tubular test surface boiling initiated with the emergence of discrete sites whose density increased with heat flux. On the other hand, the initiation of boiling on the disk was generally observed at relatively higher wall superheats with the sudden boiling of the entire surface accompanied by a rapid temperature drop. This type of vapor seeding mechanism is assumed to be associated with the difference in the microgeometry of the surfaces.

#### 5.3.2 Aqueous-alcohol mixtures

Among the three aqueous mixture systems ethanol-water was the most extensively investigated. Three different smooth surfaces were used for measuring the incipient and deactivation superheats as a function of composition, as shown in Figs. 5.3-5.5. The maximum in the incipient superheat on each surface was observed at a different mole fraction. For the polished tube the maximum in incipient superheat, Fig.5.4, coincided with that in  $(\tilde{y}_b - \tilde{x}_b)$ , Fig. 4.6(a), and for the smooth disk, Fig. 5.5, it coincided with that in  $(\tilde{y}_L - \tilde{x}_L)$ , Fig. 4.6(b). The maximum in the incipient superheat for the smooth tube was observed between these two mole fractions, as shown in Fig. 5.3. Ethanol-water is an azeotropic mixture system with the azeotrope occurring at 0.89 mole fraction, as shown in Fig. 4.5. Methanol-water is a non-azeotropic mixture system and the peak in the the incipient superheats, shown in Fig.5.1, coincided with that in  $(\tilde{y}_L - \tilde{x}_L)$ , Fig. 4.4(b). It should be noted that the trend in  $(\tilde{y}_L^{}-\tilde{x}_L^{})$  versus liquid mole fraction  $\tilde{x}_b^{}$  is identical to that in  $(\tilde{y}_b^{}-\tilde{x}_b^{})$  versus vapor mole fraction  $\tilde{y}_b^{}$ . The npropanol-water mixture system is also an azeotropic mixture system with a maximum in  $(\tilde{y}_{b}^{}-\tilde{x}_{b}^{})$  on either side of the azeotrope, as shown in Fig. 4.8. The observed incipient superheats for n-propanol-water mixtures form a maximum, which coincides with that observed in  $(\tilde{y}_b - \tilde{x}_b)$ , to the left of the azeotrope; to the right of the azeotrope no such maximum was observed. This mixture system was repeated in order to confirm the large maximum in the incipient superheats. No difference in the results

was observed. For mixture systems which exhibit a maximum in  $(\tilde{y}_b - \tilde{x}_b)$  on either side of the azeotrope, e.g. n-propanol-water, phase equilibrium dictates that the difference  $(\tilde{y}_b - \tilde{x}_b)$  be negative to the right of azeotrope, Fig. 5.17. This implies that the equilibrium concentration of the volatile component in the vapor phase is smaller than that in the liquid phase. The magnitude of depletion of the volatile component is therefore much smaller as compared to the situation when  $(\tilde{y}_b - \tilde{x}_b)$  is positive. The decrease in the radius due to condensation is proportional to the level of depletion. This means that to the right of the azeotrope the decrease in radius to achieve equilibrium will be relatively smaller. The argument regarding the decreased amount of depletion of volatile component for the case when  $(\tilde{y}_b^{}-\tilde{x}_b^{})$  is negative is further strengthened by looking at the boiling wall superheats on either side of the azeotrope. For the n-propanol-water mixture system, Fig. 6.4, the difference in the boiling wall superheats ( $\Delta T - \Delta T_{id}$ ) is smaller on the right than on the left of the azeotrope.

For mixture compositions to the right of the azeotrope, once vapor is trapped in the cavity it is the volatile component that should condense out of the vapor phase to achieve equilibrium. Such a process may result in a liquid layer rich in the volatile component in the immediate vicinity of the vapor nucleus. Since the heat of vaporization for the volatile component in aqueous mixtures is much smaller than that of water, a relatively smaller wall superheat may be required to vaporize this liquid layer to aid the vapor bubble in its initial nucleation and growth. On the other hand, to the left of the azeotrope, the non-volatile component is condensed out to attain equilibrium. A liquid layer rich in the non-volatile component would require a relatively higher superheat to vaporize. Hence, for azeotropic aqueous mixtures to the right of the azeotrope, the decreased level of depletion and the lower heat of vaporization of the volatile component seem to be the probable cause for smaller incipient superheats observed.

The scatter in data in the measured superheats leads one to question the consistency of the surface. It is observed that the wall superheats required to sustain boiling are much smaller than those required to initiate boiling on the same surface. Once the vapor is trapped in a cavity, then according to the suggested mechanism of condensation, the decrease in volume may be enough to completely deactivate that particular cavity size. In general a particular size range of cavities may be completely snuffed out and the next size range may require higher wall superheats. As the microgeometry of any one surface is usually different from that of any other surface, the same mixture system may exhibit differing incipient superheats.

#### 5.3.3 Enhanced Surface

The incipient superheats on the enhanced surface were measured for the following reasons: (1) to compare its performance with smooth surfaces, and (2) to look for any similarity between the trends in the data.

It was observed that the wall superheats required to initiate boiling on the enhanced surface are much lower than on the smooth surface. The incipient superheats were generally observed to be higher than the corresponding deactivation superheats. For any mixture system, the trend in the incipient superheats for the enhanced surface was different from that observed on the smooth surface. The mechanism of boiling on enhanced surfaces needs to be further explored and more data on incipient superheats may be required before any comprehensive explanation can be presented.

# 5.4 SUMMARY OF EFFECTS IDENTIFIED FOR BOILING INCIPIENCE ON A SMOOTH SURFACE

The following effects on boiling incipience in binary mixtures on smooth surfaces have been identified and reviewed above:

- The vapor trapping mechanism at the point of deactivation of boiling sites.
- Condensation of the non-volatile component within a trapped vapor nucleus.
- Physical properties, especially surface tension and the heat of vaporization.
- 4. The contact angle effect.
- 5. The surface microgeometry effect on the vapor trapping process.

A generalized model of incipience in binary mixtures is a very complex function of some or all of the above factors. The microgeometry of the surface is less amenable to any sort of mathematical formulation since the dimensions of critical importance, e.g. the cavity radii, etc., are on the order of microns. Future investigations may reveal the factors that are dominant under a particular set of conditions.

# **CHAPTER 6**

## BOILING HEAT TRANSFER COEFFICIENTS

## 6.1 EXPERIMENTAL RESULTS AND DISCUSSION

Nucleate pool boiling data for all four binary mixture systems on smooth and enhanced surfaces are tabulated in Appendix C. The ideal wall superheats at any heat flux have been obtained using the linear molar mixing law: Eqs. (2.18) and (2.19) for the azeotropic mixture systems and Eq. (2.17) for the non-azeotropic systems. The ideal heat transfer coefficients are then obtained from the ideal wall superheats. The results on smooth and enhanced surfaces are discussed separately and the performances of the two are compared in regard to mixture boiling.

The values tabulated in Appendix C were obtained from the boiling curves for the decreasing heat flux, as this is more representative of the boiling process. The measurements were taken by first increasing the heat flux to the test surface and then by decreasing it. This procedure was repeated more than once. The reproducibility of the results was very satisfactory, within the accuracy of the measurement errors. For a discussion on the calculations and errors associated with measurements, see Appendix B. In the fully developed boiling regime the wall superheats for the increasing and decreasing heat flux were very close to each other. However, in the lower part of the boiling curve, values of wall superheats for the decreasing heat flux were generally lower than those obtained for the increasing heat flux. This discrepancy is due to the greater number of boiling sites that remain

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active when the heat flux is decreased once it has reached a moderately high heat value. This phenomenon, commonly referred to as hysteresis, is generally observed in boiling.

## 6.2 SMOOTH SURFACE

The variation in the wall superheat as a function of composition at four different heat flux levels is shown in Figs. 6.1-6.6 for all four mixture systems tested. The linear mixing law line, not shown for clarity, is a straight line connecting the pure component 2  $(\tilde{x}_1=0.0)$ , the azeotropic composition (if one exists) and the pure component 1  $(\tilde{x}_1-1.0)$  at the same heat flux. It is observed that, at any constant heat flux, the experimental wall superheats are above their corresponding linear values. This trend implies a deterioration in mixture heat transfer coefficients when compared with the values obtained from linear mixing law. Only one exception is at the 0.30 mole fraction of n-propanol in water, shown in Fig. 6.4. At this point the experimental wall superheat is observed to be lower than the corresponding value from linear mixing law. A similar effect was reported by Calus and Leonidoplous [8] for the same mixture system boiling on a 0.3 mm diameter horizontal wire of nickel-aluminum. The influence of differing surface roughness is reflected in the results of ethanol-water (Figs. 6.2 and 6.3) and ethanol-benzene (Figs. 6.5 and 6.6). It is observed that the wall superheats generally decrease with increasing roughness of the surface.

# 6.2.1 Prediction of Smooth Surface Results

The semi-empirical correlation recently developed by Schluender [48] to predict mixture heat transfer coefficients in pool boiling has not been widely tested against experimental data. However, Uhlig and



Figure 6.1. Wall superheats for methanol-water mixtures boiling on smooth tube (for decreasing heat flux data).

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Figure 6.2. Wall superheats for ethanol-water mixtures boiling on smooth tube (for decreasing heat flux data).



Figure 6.3. Wall superheats for ethanol-water mixtures boiling on polished tube (for decreasing heat flux data).



Figure 6.4. Wall superheats for n-propanol-water mixtures boiling on smooth tube (for decreasing heat flux data).



Figure 6.5. Wall superheats for ethanol-benzene mixtures boiling on smooth tube (for decreasing heat flux data).



Figure 6.6. Wall superheats for ethanol-benzene mixtures boiling on smooth disk (for decreasing heat flux data).

Thome [84] have reported good agreement, within  $\pm 20$ %, between their experimental results and those predicted by the Schluender correlation for the acetone-water mixture system. The experiments were performed on a horizontal tube of 19.05 mm diameter at 1.01 and 3.03 bars. The Schluender correlation is attractive because it incorporates the functional dependence of the degradation of the heat transfer coefficient on heat flux (the experimental observations do indeed confirm this dependence). The Schluender correlation for azeotropic mixture systems is:

$$\frac{\alpha}{\alpha_{id}} = \frac{1}{\left\{1 + \frac{\alpha_{id}}{q}(T_{s_2} - T_{s_1})(\tilde{y}_1 - \tilde{x}_1)\left(1 - \exp\left[-\frac{B_o \tilde{q}}{\rho_{\ell} \beta_{\ell} \Delta h_v}\right]\right)\right\}}$$
(6.1)

$$T_{s_2} - T_{s_1} = \frac{T_{s_2} - T_{s_{az}}}{\tilde{x}_{az}} \qquad \tilde{x}_1 < \tilde{x}_{az} \qquad (6.2a)$$

where

and

 $T_{s_2} - T_{s_1} = \frac{\frac{T_{s_1} - T_{s_1}}{1 - \tilde{x}_{az}}}{1 - \tilde{x}_{az}}$   $\tilde{x}_1 > \tilde{x}_{az}$  (6.2b)

In terms of wall superheat (since  $\alpha = \dot{q}/\Delta T$ ), Eq. (6.1) can be rewritten as:

$$\frac{\Delta T_{id}}{\Delta T} = \frac{1}{\left\{1 + \frac{\alpha_{id}}{q}(T_{s_2} - T_{s_1})(\tilde{y}_1 - \tilde{x}_1)\left(1 - \exp\left[-\frac{B_o \dot{q}}{\rho_\ell \beta_\ell \Delta h_v}\right]\right)\right\}}$$
$$\Delta T = \Delta T_{id} + (T_{s_2} - T_{s_1})(\tilde{y}_1 - \tilde{x}_1)\left(1 - \exp\left[-\frac{B_o \dot{q}}{\rho_\ell \beta_\ell \Delta h_v}\right]\right)$$
(6.3)

 $(T_{s_2} - T_{s_1})$  and  $(\tilde{y}_1 - \tilde{x}_1)$  are always of the same sign and their product is therefore positive. The second term on right hand side of Eq. (6.3) is the amount by which the actual superheat,  $\Delta T$ , is different from its linear value (or ideal value),  $\Delta T_{id}$ . The difference  $\Delta T - \Delta T_{id}$ , referred to as "excess superheat,"  $\Delta \theta$ , is generally positive as the observed superheats are always above their corresponding linear values in mixture boiling. As the heat flux approaches its critical value,  $\dot{q}_{cr}$ , the exponential term in Eq. (6.3) approaches zero. Hence, for this condition Eq. (6.3) can be written as:

$$\Delta \theta = (\mathbf{T}_{s_2} - \mathbf{T}_{s_1})(\tilde{\mathbf{y}} - \tilde{\mathbf{x}}) \qquad \text{as } \dot{\mathbf{q}} \to \dot{\mathbf{q}}_{cr} \qquad (6.4)$$

Thome (Eq. (2.38)), using the phase equilibrium diagram expresses the excess superheat at this condition as:

$$\Delta \theta = \Delta T_{bp}$$
 as  $\dot{q} \rightarrow \dot{q}_{cr}$  (6.5)

The variation in the mass transfer coefficient,  $\beta_1$ , with composition is generally not known. In the absence of any data,  $\beta_1$  has been assumed by Schluender to be 2 x 10<sup>-4</sup> m/s (this value corresponds to the metyhylethylketone-toluene mixture system at 1.0 bar [48]). The factor B<sub>o</sub>, according to Schluender, includes all the improbabilities and has to be adjusted to experimental results. In general B<sub>o</sub> depends upon heat flux and pressure. Schluender [48] reports B<sub>o</sub> to be of the order unity and obtains good agreement between the experimental and predicted values using B<sub>o</sub>-1. Uhlig and Thome [84] (acetone-water mixtures) and Bier et al. [5] (binary mixtures of sulfurhexafloride and refrigerants) obtained reasonable agreement using B<sub>o</sub>-1. Hence, B<sub>o</sub>-1 is taken to be unity in Schluender correlation.

The experimental and predicted values from Schluender correlation are compared in Figs. 6.7-6.18 at three different heat flux levels for each mixture system and the overall deviations for the entire data are also shown. The correlation works very well for ethanol-benzene, Fig. 6.15-18, and reasonably well for the n-propanol-water mixture system, shown in Figs. 6.13 and 6.14. The deviations between the experimental and predicted values are quite large for the methanolwater, Figs. 6.7 and 6.8, and ethanol-water mixture systems, shown in



Figure 6.7. Experimental and predicted (Eq. 6.3) wall superheats for methanol-water mixtures boiling on smooth tube.



Figure 6.8. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using entire data for methanol-water mixtures boiling on smooth tube.



Figure 6.9. Experimental and predicted (Eq. 6.3) wall superheats for ethanol-water mixtures boiling on smooth tube.



Figure 6.10. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using entire data for ethanol-water mixtures boiling on smooth tube.



Figure 6.11. Experimental and predicted (Eq. 6.3) wall superheats for ethanol-water mixtures boiling on polished tube.



Figure 6.12. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using entire data for ethanol-water mixtures boiling on polished tube.



Figure 6.13 Experimental and predicted (Eq. 6.3) wall superheats for n-propanol-water mixtures boiling on smooth tube.



Figure 6.14. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using entire data for n-propanol-water mixtures boiling on smooth tube.



Figure 6.15. Experimental and predicted (Eq. 6.3) wall superheats for ethanol-benzene mixtures boiling on smooth tube.



Figure 6.16. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using entire data for ethanol-benzene mixtures boiling on smooth tube.



Figure 6.17. Experimental and predicted (Eq. 6.3) wall superheats for ethanol-benzene mixtures boiling on smooth disk.



Figure 6.18. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using data for ethanol-benzene mixtures boiling on smooth disk.

Figs. 6.9-6.12. However, the scatter in the data is small and better agreement can be obtained using higher values of  $B_0$ .  $B_0$ -2 yielded much better predictions, within ±25%. There is no rationale for choosing a different value of  $B_0$  for each particular mixture system since data would be needed to determine the best value of  $B_0$  and the correlation would be more difficult to apply and less general in form.

#### 6.3 ENHANCED SURFACE RESULTS

Experimentally observed variations in wall superheats for the boiling of methanol-water, ethanol-water and n-propanol-water mixture systems on High Flux are shown in Figs. 6.19-6.21 at four different heat flux levels. The linear molar mixing law lines are again not shown for clarity. For the methanol-water mixture system, the observed wall superheats, shown in Fig. 6.19, exhibit positive deviations (hence negative deviation in  $\alpha$ ) from those predicted using the linear mixing law approach. The other two mixture systems show both positive and negative deviations in wall superheats, as shown in Figs. 6.20 and 6.21. This implies that for boiling on an enhanced surface the heat transfer coefficients can be higher or lower than expected from a linear mixing law approach. It should be kept in mind that the linear mixing law does not incorporate any non-linear variations in mixture physical properties. To demonstrate the augmentation in heat transfer on the enhanced surface (High Flux) as compared to a conventional smooth surface, experimental wall superheats from the two surfaces under the same conditions are plotted in Figs 6.22-6.24 for methanolwater, ethanol-water and n-propanol-water mixture systems, respectively. The enhancement is obtained for both the pure components and the mixtures. In general a 2 to 4-fold increase in heat transfer coefficients is observed. These results support the earlier



Figure 6.19. Wall superheats for methanol-water mixtures boiling on enhanced tube (for decreasing heat flux data).



Figure 6.20. Wall superheats for ethanol-water mixtures boiling on enhanced tube (for decreasing heat flux data).



Figure 6.21. Wall superheats for n-propanol-water mixtures boiling on enhanced tube (for decreasing heat flux data).



Figure 6.22. Comparision of enhanced surface wall superheats to the smooth tube for methanol-water mixtures.


Figure 6.23. Comparision of enhanced surface wall superheats to the smooth tube for ethanol-water mixtures.



Figure 6.24. Comparision of enhanced surface wall superheats to the smooth tube for n-propanol-water mixtures.

observations made by Ali and Thome [2] and Uhlig and Thome [84] for boiling of mixtures on the High Flux surface. The enhanced surface shows a reduced "mixture effect," as the deviation between the actual and linear mixing law predictions is much smaller on the enhanced surface than on the conventional smooth surface.

The mixture boiling results obtained on the enhanced surface were used to test the validity of Eq. (2.39) derived by Webb [98] for pure components and tested on refrigerants only. Eq. (2.39) is:

$$T_{w} - T_{sat} = 0.044 \frac{d_{p}^{2} q}{\lambda_{l} \delta_{c}} + \frac{9.66 \sigma}{d_{p} (dP/dT)_{sat}}$$
 (2.39)

where  $\delta_c$  is the porous layer thickness and  $d_p$  is the particle size used for the coating. For High Flux tubing  $\delta_c = 0.25$  mm and  $d_p = 0.042$  mm are reported [37,98]. Webb [98] also reports in-line cubic packing for High Flux surface. Shakir et al. [50] have compared the experimental and predicted values of wall superheats using Eq. (6.6), and the results are shown in Table 6.1. It is obvious from the predicted wall superheats that even with mixture properties the correlation is far from being satisfactory when used for mixture boiling. It is not surprising since the mass transfer effects are not taken in to consideration in Eq. (6.6). Also, the nucleation superheat term in Eq. (6.6) is too dominant, such that  $(T_w - T_{sat})$  is only a slight function of the heat flux  $\dot{q}$ .

Schluender correlation was also tested for mixture boiling on enhanced surfaces. Fig. 6.25 compares the results for the n-propanolwater mixture system data to the predicted results. The deviations are observed to be higher than those obtained for the same mixture system boiling on smooth surfaces (Fig. 6.14) using the same  $B_0$  and  $\beta_l$ .



Figure 6.25. Comparision of experimental and predicted (Eq. 6.1) boiling heat transfer coefficients using entire data for n-propanol-water mixtures boiling on enhanced surface.

Thome [76] has recently considered the nature in which mixture boiling on enhanced surfaces differs from that on a conventional smooth surface. The reduced mixture effect observed on enhanced surfaces is explained on the basis of a superposition boiling model. The model proposed by Thome comprises contributions from boiling heat transfer and boiling induced liquid-phase convection. Thome suggests that due to the depletion of the more volatile component a deterioration in boiling heat transfer occurs, as is true for boiling on a conventional smooth surface. The factor that augments heat transfer on an enhanced surface is the liquid-phase convection induced by the special geometry of such surfaces. Thome draws an analogy between the convective boiling inside of tubes and that in the narrow passages of the enhanced surface. A qualitative analysis shows very promising results. Future efforts, according to Thome, for modeling boiling on enhanced surfaces should include both the mass transfer and the augmented liquid-phase convection effects.

Table 6.1. Comparison of experimental and predicted [Eq. (6.6)] wall superheats for the enhanced surface

	ΔT	(K)	
at	ġ	(kW/m <sup>2</sup>	? )

x methanol	9	9	8	51	5	2
	EXP	PRED	EXP	PRED	EXP	PRED
0.00	3.5	3.9	3.1	3.9	2.5	3.9
0.05	6.2	3.2	5.6	3.2	4.7	3.2
0.25	6.0	2.1	5.4	2.1	4.3	2.0
0.55	5.6	1.5	5.1	1.5	4.1	1.4
0.85	5.7	1.3	5.2	1.3	4.0	1.2
1.00	4.6	1.2	4.3	1.2	3.7	1.2

#### CHAPTER 7

#### CONCLUSIONS

#### 7.1 BOILING INCIPIENCE AND DEACTIVATION

- 1. On smooth surfaces, the incipient superheats for the mixtures were larger than is predicted from a simple linear mixing law between the single components values. The corresponding incipient and deactivation superheats on the enhanced surface showed both positive and negative deviations from the linear mixing law. The incipient and deactivation superheats for all four mixture systems investigated were strong, non-linear functions of mixture composition.
- 2. Deactivation superheats for boiling on smooth surfaces were much smaller on value than the corresponding incipient superheats on the same surface. The incipient and deactivation superheats for any mixture system showed a different trend for boiling on the same surface.
- 3. On smooth surfaces, the incipient superheats exhibit a maximumfor azeotropic and non azeotropic aqueous mixture systems. For the ethanol-benzene mixture system, a maximum was observed onboth sides of the azeotrope.
- 4. The incipient superheats were observed to vary with the type of surface finish and enhancement. The incipient superheats on the enhanced surface (High Flux) were much lower than for the smooth surfaces. The maximum in the incipient superheat was observed to shift with the type of surface.

5. The classical boiling nucleation equation  $\Delta T_{inc} = \frac{2\sigma}{r(dP/dT)_{sat}}$  was demonstrated to be inadequate for predicting incipientsuperheats for mixtures. Several new factors were identified forboiling nucleation on smooth surfaces which could qualitativelyexplain the maxima in incipient superheats observed for themixtures.

## 7.2 CONTACT ANGLES

- 1. The contact angles for the pure organic components were observed to be much smaller than for pure water. The contact angles foraqueous mixtures showed a non-linear variation with the mixture composition and were smaller than the linear mixing law prediction.
- 2. The material of the surface showed a negligible effect on the contact angles for the same mixture system.

## 7.3 HEAT TRANSFER COEFFICIENTS

- The wall superheats (and heat transfer coefficients) for boiling of mixtures on smooth and enhanced surfaces were observed to be non-linear function of mixture composition.
- 2. On smooth surfaces the wall superheats were observed to be much higher than expected from a linear mixing law. This implies a deterioration in mixture heat transfer coefficients when compared to the pure component values.
- For boiling of mixtures on the enhanced surface, the wall superheats showed both positive and negative deviations from the linear mixing law.

4. The enhanced surface results show appreciable augmentation of the heat transfer coefficients for the mixtures compared to those for the smooth surface.

## RECOMMENDATIONS

Suggestions for future work in the area of nucleate pool boiling of mixtures and for further advancement of the experimental facility are as follows:

- The experimental studies be extended to other enhanced surfaces, e.g. Thermoexcel, Gewa-T, etc. The experiments may be designed to obtain:
  - (a) boiling nucleation and decativation superheats as a function of mixture composition;
  - (b) nucleate pool boiling heat transfer coefficients as a function of mixture composition and heat flux;
  - (c) peak nucleate heat flux as a function of mixture composition;
  - (d) functional dependence of (a), (b) and (c) on pressure and subccooling will also be of interest.

The experimental results thus obtained may help explain the mechanism of boiling on enhanced surfaces and eventually lead to the development of a physical model in the form of a empirical or semi-empirical correaltion. The results may also help in determining the suitability of enhanced surfaces to multicomponent mixture boiling.

- 2. A data acquisition acquisition system should be interfaced with the existing instrumentation to record and process the experimental data. this will also aid in creating a data bank of all the research work carried out in the laboratory.
- A flow boiling facility be designed to study forced convective boiling inside tubes.

APPENDIX A

### APPENDIX A

### COMPUTER PROGRAMS AND MIXTURE PROPERTIES

Table A.1.1. Listing of Program MIXPR1.FOR

PROGRAM MIXPR1 С С C THIS PROGRAM CALCULATES THE PROPERTIES OF BINARY MIXTURES С С С LOGICAL\*1 FNAME(15), FTAME(15), FMAME(15), FPAME(15) DIMENSION X(15), T(15), Y(15, 10), STRO(16), C1(5), C2(5), C3(5)1 C4(5),C5(5),C6(5),D1(5),D2(5),D3(5),D4(5),D5(5),D6(5),D7(5), 1 D8(5), D9(5), W11(15) 100 TYPE \*, 'ENTER THE MIXTURE SYSTEM' ACCEPT 4, (STR0(L), L=1, 16) 4 FORMAT(16A1) OPEN(UNIT=2,NAME='CRIDAT.MX1',TYPE='OLD',FORM='FORMATTED') DO 1 MM=1,5M-MM READ(2,\*)C1(M),C2(M),C3(M),C4(M),C5(M),C6(M) 1 CONTINUE CLOSE(UNIT=2) С OPEN(UNIT-2, NAME-'CONDAT.MX1', TYPE-'OLD', FORM-'FORMATTED') DO 7 KK-1,5 K-KK READ(2, \*)D1(K), D2(K), D3(K), D4(K), D5(K), D6(K), D7(K), D8(K), D9(K)7 CONTINUE CLOSE(UNIT=2) TYPE \*,'1-METHANOL, 2-ETHANOL, 3-PROPANOL, 4-BENZENE, 5-WATER' TYPE \*,' TYPE \*, 'ENTER THE # FOR THE VOLATILE COMPONENT' ACCEPT \*, I TYPE \*, 'ENTER THE # FOR THE NON-VOLATILE COMPONENT' ACCEPT \*,J WM1=C1(I)WM2=C1(J)C WM IS THE MOLECULAR WEIGHT PC1-C2(I)PC2=C2(J)C PC IS THE CRITICAL PRESSURE IN BARS

TC1=C3(I)TC2=C3(J)C TC IS THE CRITICAL TEMPERATURE IN DEG K DC1=C4(I)DC2=C4(J)C DC IS THE CRITICAL DENSITY IN KG/CU-M ZC1-C5(I)ZC2=C5(J)C ZC IS THE CRITICAL COMPRESSIBILITY FACTOR ZRA1=C6(I)ZRA2=C6(J)C ZRA IS THE CONSTANT IN MODIFIED RACKETT EQ. FOR LIQUID DENSITY CALL GTLIN(FTAME, 'ENTER INPUT FILENAME FOR X-TSAT') OPEN(UNIT=2, NAME=FTAME, TYPE='OLD', FORM='FORMATTED') READ(2, 11)MF11 FORMAT(12) DO 12 KK=1,MF K-KK READ(2,13)X(K),T(K)13 FORMAT(2(F7.3))12 CONTINUE CLOSE(UNIT=2) PRINT 18, (STRO(L), L=1, 16)18 FORMAT(1H1,16A1,//) PRINT 19 19 FORMAT(3X,'X1',5X,'TSAT',7X,'TCOND',8X,'VIS',9X, 1 'SP HT', 7X, 'S TEN', 5X, 'LIQ DEN', /) F1 = 1.0/3.0F2=2.0/3.0F3-4.0/3.0F4=2.0/7.0R-82.06 C R IS UNIVERSAL GAS CONSTANT IN ATM-(CU CM)/GMOL-K VC1=WM1/DC1 VC2=WM2/DC2 C VC IN CU-M/KG MOL VC12=(((VC1\*\*F1)+(VC2\*\*F1))\*\*3.0)/8.0 ATC12=(SQRT(TC1\*VC1\*TC2\*VC2))/VC12 TYPE \*, 'DO YOU WANT TO CALCULATE THERMAL CONDUCTIVITY?' TYPE \*, 'Y OR N' ACCEPT 21, IANS 21 FORMAT(A1) IF(IANS.EQ.'N')GO TO 31 TYPE \*, 'CALCULATION OF THERMAL CONDUCTIVITY NOW BEGINS' TYPE \*, 'ENTER THE BINARY INTERACTION CONSTANT PAR' ACCEPT \*, PAR A1=D1(I)A2=D1(J)B1=D2(I)B2-D2(J)CC1-D3(I) CC2=D3(J)TYPE \*, 'AQUEOUS MIXTURE - Y OR N'

ACCEPT 21, IANS IF(IANS.EQ.'N')GO TO 28 C CHANGE MOLE FRACTION TO MASS FRACTION FOR AQUEOUS MIXTURES DO 23 KK-1.MF W11(KK) = (X(KK) \* WM1) / (X(KK) \* WM1 + (1.0 - X(KK)) \* WM2)23 CONTINUE GO TO 29 28 DO 24 KK-1,MF W11(KK) = X(KK)24 CONTINUE 29 CONTINUE DO 22 K-1,MF W1 = W11(K)W2-1.0-W1 WMM=W1\*WM1+W2\*WM2 TC12-PAR\*ATC12 VCM=(W1\*W1\*VC1)+(2.0\*W1\*W2\*VC12)+(W2\*W2\*VC2) TCM=((W1\*W1\*TC1\*VC1)+(2.0\*W1\*W2\*TC12\*VC12)+ 1 (W2\*W2\*TC2\*VC2))/VCM PM=(SQRT(WMM/TCM))\*(VCM\*\*F2) TR = (T(K) + 273.15) / TCMTKP1=A1+B1\*((1.0-TR)\*\*F2)+CC1\*((1.0-TR)\*\*F3)TKP2=A2+B2\*((1.0-TR)\*\*F2)+CC2\*((1.0-TR)\*\*F3) TK=((W1\*TKP1+W2\*TKP2)/PM)\*0.001  $Y(K, 1) = TK \times 1000.0$ C THE VALUE OF THERMAL CONDUCTIVITY HAS BEEN MULTIPLIED BY 1000.0 C Y(I,1) IS IN 0.001\*W/M-K 22 CONTINUE GO TO 33 31 DO 32 K-1,MF Y(K, 1) = 0.032 CONTINUE 33 TYPE \*, 'DO YOU WANT TO CALCULATE LIQUID DYNAMIC VISCOSITY?' TYPE \*.'Y OR N' ACCEPT 34, IANS 34 FORMAT(A1) IF(IANS.EQ.'N')GO TO 42 TYPE \*.'CALCULATION OF LIQUID DYNAMIC VISCOSITY NOW BEGINS' **TYPE \*.'ENTER THE BINARY INTERACTION CONSTANT PAR'** ACCEPT \*, PAR A1=D4(I)A2 - D4(J)B1=D5(I)B2=D5(J)DO 35 K=1,MF X1=X(K)X2-1.0-X1 WMM-X1\*WM1+X2\*WM2 TC12=PAR\*ATC12  $VCM = (X1 \times X1 \times VC1) + (2.0 \times X1 \times X2 \times VC12) + (X2 \times X2 \times VC2)$ **TCM=((X1\*X1\*TC1\*VC1)+(2.0\*X1\*X2\*TC12\*VC12)+** 1 (X2\*X2\*TC2\*VC2))/VCM PM=(VCM\*\*F2)/(SQRT(TCM\*WMM)) TR = (T(K) + 273.15) / TCMVIS1-A1+(B1/TR)VIS2=A2+(B2/TR)

```
C VIS1 = ALOG(VIS1*P1) AND VIS2 = ALOG(VIS2*P2)
     VISM-(EXP((X1*VIS1)+(X2*VIS2)))/PM
C VISM IS THE VISCOSITY OF THE MIXTURE IN CENTI POISE
     Y(K,2)=VISM*1000.0
C Y(K,2) IS THE LIQUID DYNAMIC VISCOSITY IN E-06*(N-S/SQ M)
35
     CONTINUE
     GO TO 44
42
     DO 43 K=1,MF
     Y(K, 2) = 0.0
43
      CONTINUE
44
      TYPE *, 'DO YOU WANT TO CALCULATE LIQUID HEAT CAPACITY?'
      TYPE *, 'Y OR N'
     ACCEPT 45, IANS
45
     FORMAT(A1)
      IF(IANS.EQ.'N')GO TO 53
TYPE *, 'CALCULATION OF HEAT CAPACITY NOW BEGINS'
      TYPE *, 'ENTER THE BINARY INTERACTION CONSTANT PAR'
     ACCEPT *, PAR
     A1 = D6(I)
     A2=D6(J)
      B1-D7(I)
      B2=D7(J)
      TC12=PAR*ATC12
      DO 46 K=1, MF
     X1 - X(K)
     X2-1.0-X1
     WMM=X1*WM1+X2*WM2
      TC12=PAR*ATC12
      VCM=(X1*X1*VC1)+(2.0*X1*X2*VC12)+(X2*X2*VC2)
      TCM=((X1*X1*TC1*VC1)+(2.0*X1*X2*TC12*VC12)+
    1 (X2*X2*TC2*VC2))/VCM
      TR = (T(K) + 273.15) / TCM
      RR=8314.0
C RR IS GAS CONSTANT IN J/KMOL-K
      CP1=(RR/WM1)*(EXP(A1-(B1/TR)))
      CP2=(RR/WM2)*(EXP(A2-(B2/TR)))
      CP=X1*CP1+X2*CP2
C CP IS THE SPECIFIC HEAT IN J/KG-K
      Y(K,3) = CP/1000.0
C Y(K,3) IS SPECIFIC HEAT IN 1000.0*J/KG-K
46
      CONTINUE
      GO TO 55
53
      DO 54 K-1,MF
      Y(K,3)=0.0
      CONTINUE
54
55
      TYPE *.'DO YOU WANT TO CALCULATE SURFACE TENSION'
      TYPE *, 'Y OR N'
      ACCEPT 56, IANS
56
      FORMAT(A1)
      IF(IANS.EQ.'N')GO TO 64
TYPE *, 'CALCULATION OF SURFACE TENSION NOW BEGINS'
      TYPE *, 'ENTER THE BINARY INTERACTION CONSTANT PAR'
      ACCEPT *, PAR
      TC12=PAR*ATC12
```

```
A1-D8(I)
      A2 = D8(J)
      B1-D9(I)
      B2-D9(J)
      DO 57 K-1,MF
      X1 = X(K)
      X2=1.0-X1
      VCM=(X1*X1*VC1)+(2.0*X1*X2*VC12)+(X2*X2*VC2)
      TCM=((X1*X1*TC1*VC1)+(2.0*X1*X2*TC12*VC12)+
     1 (X2*X2*TC2*VC2))/VCM
      TR = (T(K) + 273.15) / TCM
      PM-((VCM)**F2)/TCM
      ST1=A1-(B1*TR)
      ST2=A2-(B2*TR)
      ST=(((X1*ST1)+(X2*ST2))/PM)*0.001
      Y(K,4) = ST + 1000.0
C THE VALUE OF SURFACE TENSION HAS BEEN MULTIPLIED BY 1000.0
C Y(K,4) IS IN 0.001*N/M
57
      CONTINUE
      GO TO 66
      DO 65 K-1,MF
64
      Y(K,4) = 0.0
      CONTINUE
65
TYPE *, 'DO YOU WANT TO CALCULATE LIQUID DENSITY?'
66
      TYPE *, 'Y OR N'
      ACCEPT 67, IANS
      FORMAT(A1)
67
      IF(IANS.EQ.'N')GO TO 75
TYPE *, 'CALCULATION OF LIQUID DENSITY NOW BEGINS'
      TYPE *.'ENTER THE BINARY INTERACTION CONSTANT PAR'
      ACCEPT *, PAR
      TC12=PAR*ATC12
      DO 68 K-1,MF
      X1 = X(K)
      X2-1.0-X1
      VCM=(X1*X1*VC1)+(2.0*X1*X2*VC12)+(X2*X2*VC2)
      TCM=((X1*X1*TC1*VC1)+(2.0*X1*X2*TC12*VC12)+
     1 (X2*X2*TC2*VC2))/VCM
      ZCM=X1*ZC1+X2*ZC2
      TR = (T(K) + 273.15) / TCM
      FAC=1.0+((1.0-TR)**F4)
      RH01=1.0/(((82.06*TC1*0.001)/(PC1*WM1))*(ZRA1**FAC))
      RH02=1.0/(((82.06*TC2*0.001)/(PC2*WM2))*(ZRA2**FAC))
      RHO-ZCM/((X1*ZC1/RHO1)+(X2*ZC2/RHO2))
      Y(K,5)-RHO
C Y(K,5) IS THE DENSITY IN KG/CU-M
68
      CONTINUE
      GO TO 77
      DO 76 K-1,MF
75
      Y(K, 5) = 0.0
      CONTINUE
76
77
      CONTINUE
CALL GTLIN(FMAME, 'ENTER OUTPUT FILENAME FOR PROPERTIES')
      OPEN(UNIT=2, NAME=FMAME, TYPE='NEW', FORM='FORMATTED')
      DO 111 M-1,MF
```

		X2=1.0-X(M)
		WRITE(2, 110)X(M), X2, T(M), (Y(M, N), N-1, 5)
110		FORMAT(F5.3,2X,F5.3,2X,F6.2,2X,F5.1,2X,F5.1,2X,F5.3,2X,F5.2,
	1	2X,F5.1)
		TYPE $113, X(M), T(M), (Y(M,N), N=1, 5)$
113		FORMAT(2X,F5.3,2X,F6.2,5(1X,F10.3),/)
		PRINT 112, $X(M)$ , $T(M)$ , $(Y(M,N)$ , $N=1$ , 5)
112		FORMAT(1X,F5.3,3X,F6.2,5(2X,F10.3),/)
111		CONTINUE
		CLOSE(UNIT=2)
С		
****	**>	***************************************
		TYPE *, 'DO YOU WANT TO CONTINUE FOR ANOTHER MIXTURE SYSTEM?'
		TYPE *, 'ENTER Y TO CONTINUE'
		ACCEPT 120, IANS
120		FORMAT(A1)
		IF(IANS.EQ.'Y')GO TO 100
		STOP
		END

CRIDAT.MX1
Å
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2
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р.
, ag

ZRA	0.23230 0.25041 0.25272 0.26967 0.24091	
Zc	0.21969 0.24834 0.25236 0.27622 0.23827	
$\rho_{\rm c} \frac{\rm kg}{\rm m}$	275.00000 280.00000 273.00000 301.60000 315.00000	
T <sub>c</sub> (K)	513.15000 516.25000 536.85000 562.60000 647.30000	
P <sub>c</sub> (bars)	79.50000 63.90000 50.50000 49.24000 221.29000	
Mol. Wt.	32.00000 46.10000 60.10000 78.10800 18.01560	
	°.43.0.1	

Table A.1.3. Listing of CONDAT.MX1

()	В	)18424 )24993 )28329 )52465 )21925
I. (4.26		200000
E	A	0.02095 0.02721 0.03156 0.04818 0.02664
.25)	ß	0.33150 0.95650 0.46730 0.29850 0.00650
Eq. (4.	A	2.84100 4.24530 3.68930 3.34980 2.21780
24)	В	2.49172 3.06256 4.07907 2.23317 2.34154
Eq. (4	A	-11.19289 -11.49160 -12.87967 -10.98099 -11.94228
Eq. (4.23)	U	0.00000 0.00000 0.00000 0.00000 0.00000
	В	11.94140 15.55490 12.74220 3.22771 100.50700
	A	5.39498 6.33533 11.68540 1.39874 -9.67795
		5 4 3 5 H

Methanol Ethanol n Propanol Benzene Water 

Table A.2.1. Listing for Program MIXPR2.FOR

```
PROGRAM MIXPR2
С
С
С
C MARCH 4, 1985.
С
C THIS PROGRAM CALCULATES THE PROPERTIES OF BINARY MIXTURES
С
       LOGICAL*1 FNAME(15), FTAME(15)
       DIMENSION X(15),Y(15),T(15),AY(5),V(5),STR0(16),G1(5),G2(5),
     1 G3(5), G4(5), G5(5), G6(5), G7(5), DE(5), A(5), B(5), C(5), D(5)
100
       TYPE *, 'ENTER THE MIXTURE SYSTEM'
       ACCEPT 4, (STR0(L), L=1, 16)
4
       FORMAT(16A1)
       OPEN(UNIT-2, NAME-'CRIDAT.MX2', TYPE-'OLD', FORM-'FORMATTED')
       DO 7 KK-1,5
       K-KK
       READ(2, *)G1(K),G2(K),G3(K),G4(K),G5(K),G6(K),G7(K)
7
       CONTINUE
       CLOSE(UNIT-2)
       OPEN(UNIT-3, NAME-'CONDAT.MX2', TYPE-'OLD', FORM-'FORMATTED')
       DO 77 KKK-1,5
       K-KKK
       READ(3,89)A(K),B(K),C(K),D(K)
89
       FORMAT(4(E15.7))
77
       CONTINUE
       CLOSE(UNIT=3)
       TYPE *, '1-METHANOL, 2-ETHANOL, 3-PROPANOL, 4-BENZENE, 5-WATER'
       TYPE *,'
       TYPE *, 'ENTER THE # FOR VOLATILE COMPONENT'
       ACCEPT *, I
       TYPE *.'ENTER THE # FOR NON-VOLATILE COMPONENT'
       ACCEPT *.J
       WM1=G1(I)
       WM2=G1(J)
C WM IS THE MOLECULAR WEIGHT
       PC1=G2(I)
       PC2=G2(J)
C PC IS THE CRITICAL PRESSURE IN BARS
       TC1-G3(I)
       TC2=G3(J)
C TC IS THE CRITICAL TEMPERATURE IN DEG K
       DC1-G4(I)
       DC2=G4(J)
C DC IS THE CRITICAL DENSITY IN KG/CU-M
       EC1=G5(I)
       EC2=G5(J)
C OB IS OMEGA SUB B
       F1=G6(I)
       F2=G6(J)
C EC IS THE CRITICAL COMPRESSIBILITY FACTOR
       OB1=G7(I)
       OB2=G7(J)
C F IS A FACTOR FOR CALCULATING ALPHA
```

```
CALL GTLIN(FTAME, 'ENTER INPUT FILENAME FOR FOR X-TSAT-Y')
     OPEN(UNIT=2, NAME=FTAME, TYPE='OLD', FORM='FORMATTED')
     READ(2, 11)MF
11
     FORMAT(12)
     DO 12 KK-1,MF
     K-KK
     READ(2,13)X(K),T(K),Y(K)
13
     FORMAT(3(F7.3))
12
     CONTINUE
     CLOSE(UNIT-2)
PRINT 18, (STRO(L), L-1, 16)
18
     FORMAT(1H1,/,1X,16A1,//)
     PRINT 19
19
     FORMAT(5X,'X1',6X,'X2',7X,'TSAT',5X,'LIQ DEN',4X,
      'VAP DEN', 4X, 'ENT VAP', /)
    1
P=1.0
C P IS THE PRESSURE IN ATM
     R-0.08206
C R IS THE GAS CONSTANT IN (ATM - CU-M)/(KG-MOL - K)
     PI=4.0*ATAN(1.0)
     FAC=1.0/3.0
     OA1=(3.0*EC1*EC1)+3.0*(1.0-2.0*EC1)*OB1+(OB1*OB1)+
    1 1.0 - (3.0 \pm EC1)
     OA2=(3.0*EC2*EC2)+3.0*(1.0-2.0*EC2)*OB2+(OB2*OB2)+
    1 1.0 - (3.0 + EC2)
     OC1-1.0-(3.0*EC1)
     0C2=1.0-(3.0*EC2)
     B1-OB1*R*TC1/PC1
     B2-OB2*R*TC2/PC2
     C1=OC1*R*TC1/PC1
     C2-OC2*R*TC2/PC2
VC1=WM1/DC1
     VC2-WM2/DC2
C VC IN CU-M/KG MOL
     VC12=(((VC1**FAC)+(VC2**FAC))**3.0)/8.0
     ATC12=(SQRT(TC1*VC1*TC2*VC2))/VC12
C
TYPE *, 'ENTER THE BINARY INTERACTION CONST - PAR'
     ACCEPT *, PAR
     DO 75 KK-1,MF
     K-KK
     PX1-X(K)
     PX2=1.0-PX1
     DO 88 IX-1,2
     DELTA1 = Y(K) - X(K)
     X1-X(K)
     IF(IX.EQ.2)X1-Y(K)
     X2-1.0-X1
     WMM=(X1*WM1)+(X2*WM2)
     TT-T(K)
     TA = T(K) + 273.15
     TC12-ATC12
     VCM=(X1*X1*VC1)+(2.0*X1*X2*VC12)+(X2*X2*VC2)
```

170

```
TCM-((X1*X1*TC1*VC1)+(2.0*X1*X2*TC12*VC12)+
     1 (X2*X2*TC2*VC2))/VCM
       TR-TA/TCM
       ALP1=((1.0+F1*(1.0-SQRT(TR)))**2)
       ALP2=((1.0+F2*(1.0-SQRT(TR)))**2)
       A1=OA1*((R*TC1)**2)*ALP1/PC1
       A2=OA2*((R*TC2)**2)*ALP2/PC2
       A12=PAR*(SORT(A1*A2))
       AM = (X1 \times X1 \times A1) + (2.0 \times X1 \times X2 \times A12) + (X2 \times X2 \times A2)
       BM = (X1 * B1) + (X2 * B2)
       CM = (X1 + C1) + (X2 + C2)
С
C SOLUTION OF CUBIC EQ. NOW BEGINS
CC2=CM-(R*TA/P)
       CC1=(AM/P) - (2.0*BM*CM) - (BM*BM) - (R*TA*(BM+CM)/P)
       CCO=(BM*BM*CM)+(R*TA*BM*CM/P)-(AM*BM/P)
       PP=(3.0*CC1-(CC2*CC2))/3.0
       QQ=((27.0*CC0)-(9.0*CC1*CC2)+(2.0*CC2*CC2*CC2))/27.0
       RR = (PP * PP * PP / 27.0) + (QQ * QQ / 4.0)
       IF(RR-0.0)101,102,103
C RR IS NEGATIVE
101
       IF(PP-0.0)111,112,112
112
       PRINT 115.X1
115
       FORMAT(10X, F5.3, 5X, 'PP IS EITHER ZERO OR POSITIVE')
       GO TO 140
111
       ARG=SQRT((27.0*QQ*QQ)/(-4.0*PP*PP*PP))
       AL-1.0E-15
       IF(ABS(ARG).GT.AL)GO TO 25
       PHI = PI/2.0
       GO TO 35
25
       ARG2=SQRT((1.0/(ARG*ARG))-1.0)
       PHI-ATAN(ARG2)
       IF(ARG.LT.0.0)GO TO 39
       GO TO 35
39
       PHI-PI-PHI
35
       CONTINUE
       DO 94 LL-1,3
      L-LL
       AY(L)=2.0*SQRT(-1.0*PP/3.0)*COS((PHI/3.0)+(2.0*PI/3.0)*(L-1))
94
       CONTINUE
       IF(QQ.GT.0.0)GO TO 96
       GO TO 120
96
       DO 97 LL-1,3
       L-LL
      AY(L) = -1.0 * AY(L)
97
       CONTINUE
       GO TO 120
C RR=0.0
102
       CONTINUE
       IF(QQ.GT.0.0)GO TO 105
       AA=((-0.5*QQ)**FAC)
       GO TO 107
105
       AA=-1.0*((0.5*QQ)*FAC)
107
      AY(1) = 2.0 * AA
      AY(2) = -1.0 * AA
       AY(3) = -1.0 * AA
```

```
GO TO 120
C RR IS POSITIVE
103
      PRINT 108.X1
108
      FORMAT(10X, F5.3, 5X, 'R IS POSITIVE; ONE REAL AND TWO
    1 COMPLEX ROOTS')
      GO TO 140
120
      CONTINUE
      DO 125 LL-1,3
      L-LL
      V(L) = AY(L) - (CC2/3.0)
      DE(L) = WMM/V(L)
125
      CONTINUE
      TYPE 131, X1, X2, DE(1), DE(2), DE(3)
131
      FORMAT(/, 3X, F5.3, 3X, F5.3, 3(5X, F15.10))
C CALCULATION OF ENTHALPY NOW BEGINS
VV=V(1)
      DV - DE(1)
      VL=V(2)
      DL = DE(2)
      IF(IX.EQ.1)DLL-DL
      IF(IX.EQ.2)DVV-DV
      DA1DT=(-1.0*OA1*F1*R*R*TC1/PC1)*SQRT(ALP1/TR)
      DA2DT = (-1.0 \times OA2 \times F2 \times R \times R \times TC2/PC2) \times SQRT(ALP2/TR)
      DAMDT = (X1 \times X1 \times DA1DT) + ((X1 \times X2 \times PAR2/SQRT(A1 \times A2)) \times
     1 (A1*DA2DT+A2*DA1DT))+(X2*X2*DA2DT)
      ZL=P*VL/(R*TA)
      ZV = P*VV/(R*TA)
      HN=SQRT((BM*CM)+((0.5*(BM+CM))**2))
      HM=(0.5*(BM+CM)-HN)*P/(R*TA)
      HQ=(0.5*(BM+CM)+HN)*P/(R*TA)
C 1.0 BAR = (10**5) J/CU-M; 1.0 ATM = 1.01325 BARS
      CF-101.325
      HV=(CF/WMM)*(R*TA*(1.0-ZV)+(TA*DAMDT-AM)*(0.5/HN)*
     1 ALOG((ZV+HM)/(ZV+HQ)))
      HL=(CF/WMM)*(R*TA*(1.0-ZL)+(TA*DAMDT-AM)*(0.5/HN)*
     1 ALOG((ZL+HM)/(ZL+HQ)))
      IF(IX.EQ.1)HLL=HL
       IF(IX.EQ.2)HVV=HV
      PRINT 63, DL, DV, HL, HV
63
      FORMAT(5X, 4(1X, F9.4))
88
      CONTINUE
      TREF-0.0
С
*
C TREF IS THE REFERENCE TEMP. [DEG. K] FOR IDEAL GAS ENTHALPY
С
HI1=((A(I)*(TA-TREF))+((B(I)/2.0)*((TA**2)-(TREF**2)))+
     1 ((C(I)/3.0)*((TA**3)-(TREF**3)))+
     1 ((D(I)/4.0)*((TA**4)-(TREF**4))))/WMM
      HI2=((A(J)*(TA-TREF))+((B(J)/2.0)*((TA**2)-(TREF**2)))+
     1 ((C(J)/3.0)*((TA**3)-(TREF**3)))+
     1 ((D(J)/4.0)*((TA**4)-(TREF**4))))/WMM
      ENT=DELTA1*(HI1-HI2)+HLL-HVV
       PRINT 61, DELTA1, HI1, HI2
```

61 FORMAT(5X, F7.3, 3X, F9.4, 3X, F9.4)C ENTHALPY IN KJ/KG TYPE 54, PX1, PX2, TT, DLL, DVV, ENT 54 FORMAT(2X,2(2X,F5.3),3X,F7.3,5X,F8.3,3X,F8.5,3X,F9.4,/) PRINT 55, PX1, PX2, TT, DLL, DVV, ENT 55 FORMAT(1X,2(3X,F5.3),3X,F7.3,3X,F8.3,3X,F8.5,3X,F9.4,/) 75 CONTINUE TYPE \*, 'DO YOU WANT TO CONTINUE FOR ANOTHER MIXTURE SYSTEM?' TYPE \*,'Y OR N' ACCEPT 114, IANS 114 FORMAT(A1) IF(IANS.EQ.'Y')GO TO 100 140 STOP END

Table A.2.2. Listing of CRIDAT.MX2

	M	P <sub>c</sub> (bars)	T <sub>c</sub> (K)	$   \rho_{c} \left(\frac{kg}{m}\right) $	Ψ	F	W <sub>b</sub>
1.	32.0000	79.50000	513.150	275.0000	0.2720	0.972708	0.0660761744
2.	46.1000	63.90000	516.250	280.0000	0.3000	1.230395	0.0753109306
3.	60.1000	50.50000	536.850	273.0000	0.3030	1.241347	0.0763160884
4.	78.1080	49.24000	562.600	301.6000	0.3100	0.704657	0.0786728710
5.	18.0156	221.29000	647.300	315.0000	0.2690	0.689803	0.0651030242

```
Table A.2.3. Listing of CONDAT.MX2
(Constants for estimating C [46]).
```

	Α	В	С	D
1.	0.2155779E+02	0.7093802E-01	0.2587521E-04	-0.2852177E-07
2.	0.9015913E+01	0.2141122E+00	-0.8391960E-04	0.1373534E-08
3.	0.2470687E+01	0.3325796E+00	-0.1855528E-03	0.4296482E-07
4.	0.3392378E+02	0.4744556E-01	-0.3017588E-03	0.7131491E-07
5.	0.3224874E+02	0.1924204E-02	0.1055695E-04	-0.3597152E-08

- 1. Methanol
- 2. Ethanol
- 3. n Propanol
- 4. Benzene
- 5. Water

[T <sub>sat</sub> ] - °C	$[\lambda_{\ell}] = 10^{-3} \frac{J}{s m K}$
$[\eta_{l}] = 10^{-6} \frac{M s}{2}$	$\begin{bmatrix} C \\ p_{\ell} \end{bmatrix} = 10^3 \frac{J}{kg K}$
$[\sigma] - 10^{-3} \frac{\text{N}}{\text{m}}$	$[\rho_{\ell}, \rho_{v}] = \frac{kg}{m}$
$[\Delta h_v] - 10^3 \frac{J}{kg}$	

Table A.3. Mixture Properties from MIXPR1.FOR and MIXPR2.FOR.

# METHANOL-WATER

x1	x2	T sat	٦	7 L	с Р	σ	° 1	٩ ر	Δh <sub>v</sub>
0.000	1.000	100.00	669.5	275.0	4.192	61.06	948.7	0.594	2260.0
0.050	0.950	92.80	594.1	342.2	4.108	57.50	941.4	0.629	2065.3
0.100	0.900	87.90	533.8	405.7	4.024	54.00	932.3	0.662	1907.0
0.250	0.750	80.10	407.9	548.2	3.781	44.55	900.3	0.751	1572.2
0.400	0.600	75.60	329.2	607.9	3.549	37.01	868.4	0.837	1383.9
0.550	0.450	72.60	275.9	587.4	3.326	30.98	838.0	0.920	1274.4
0.700	0.300	69.80	238.1	517.0	3.110	26.16	810.2	1.005	1210.4
0.850	0.150	67.20	210.1	422.7	2.899	22.24	784.7	1.090	1174.7
1.000	0.000	64.50	188.8	326.6	2.690	19.01	761.5	1.178	1155.3

## ETHANOL-WATER

<del>x</del> 1	x2	T sat	<sup>۲</sup>	71	с Р	σ	۶	٩ ر	∆h ▼
0.000	1.000	100.00	669.5	275.0	4.192	61.06	948.7	0.594	2260.0
0.050	0.950	90.10	546.2	369.9	4.102	56.51	941.4	0.658	1964.3
0.100	0.900	86.00	467.5	451.9	4.016	51.95	929.1	0.714	1735.6
0.200	0.800	83.20	371.3	580.5	3.865	44.07	901.8	0.818	1423.0
0.350	0.650	81.20	288.8	701.0	3.674	35.38	864.9	0.971	1159.4
0.500	0.500	79.90	237.7	733.6	3.513	29.21	833.3	1.124	1019.0
0.650	0.350	79.00	202.5	695.0	3.371	24.64	806.0	1.278	940.2
0.800	0.200	78.20	176.8	613.4	3.240	21.15	782.5	1.432	894.3
0.890	0.110	78.00	164.4	552.2	3.167	19.41	769.5	1.524	876.2
1.000	0.000	78.35	151.5	471.3	3.088	17.53	754.2	1.634	859.4

Table A.3. (continued)

# n PROPANOL-WATER

π̃ <sub>1</sub>	x2	T sat	λ	۳£	с <sub>р</sub>	σ	° <u>1</u>	٩ ۲	∆h v
0.000	1.000	100.00	669.5	275.0	4.192	61.06	948.7	0.594	2260.0
0.060	0.940	89.60	464.9	401.1	4.089	54.08	934.1	0.697	1767.4
0.150	0.850	88.50	340.7	539.4	3.948	45.00	902.0	0.829	1329.3
0.300	0.700	88.10	251.7	712.4	3.728	35.17	857.9	1.048	983.3
0.430	0.570	87.70	212.4	792.9	3.546	29.73	827.8	1.240	844.2
0.550	0.450	87.60	189.1	805.3	3.381	26.12	804.5	1.417	773.7
0.650	0.350	88.45	174.9	767.7	3.248	23.71	786.5	1.561	736.7
0.770	0.230	89.60	162.0	691.3	3.088	21.39	767.3	1.733	708.9
0.900	0.100	92.90	150.9	569.7	2.926	19.23	746.4	1.909	689.0
1.000	0.000	97.00	143.7	465.9	2.811	17.72	729.8	2.033	677.4

# ETHANOL-BENZENE

ĩ,	x2	T sat	<sup>ک</sup> <b>ا</b>	<sup>7</sup> 1	с Р	σ	° <u>1</u>	ρ <sub>v</sub>	∆h v
0.000	1.000	80.00	20.3	308.1	1.885	21.12	829.9	2.775	391.7
0.070	0.930	71.70	26.6	348.9	1.925	21.83	834.0	2.765	395.0
0.140	0.860	69.50	33.0	371.2	1.980	21.74	831.4	2.700	397.9
0.250	0.750	68.90	43.8	395.7	2.080	21.25	823.9	2.574	406.9
0.350	0.650	68.50	54.6	417.5	2.177	20.81	817.0	2.457	422.8
0.450	0.550	68.50	66.2	437.3	2.281	20.34	809.5	2.338	447.5
0.550	0.450	68.50	78.9	456.8	2.392	19.90	801.9	2.219	483.3
0.700	0.300	69.30	100.2	480.1	2.577	19.22	789.4	2.035	563.1
0.800	0.200	70.50	116.1	490.2	2.716	18.75	780.1	1.909	638.6
0.900	0.100	72.80	133.4	491.7	2.877	18.24	769.3	1.779	736.6
1.000	0.000	78.30	151.5	471.6	3.088	17.54	754.2	1.634	859.5

APPENDIX B

## APPENDIX B

# CALCULATIONS TO OBTAIN WALL SUPERHEATS AND HEAT TRANSFER COEEFICIENTS B.1\_\_\_\_\_Smooth\_Disk

A sketch of the test surface and specimen is shown in Fig. 3.4. Three thermocouples are located at a distance of 1, 5, and 9 mm from the the top (z=0), Fig. B.1.1. The surface temperature,  $T_w$ , is obtained by extrapolating these three thermocouple readings. At steady state, where all measurements were taken, the temperature gradient resulting from these readings was observed to be almost linear. The saturation temperature of the bulk liquid,  $T_{sat}$ , was obtained by averaging the



Fig. B.1.1.

$$r_i = 12.7 \times 10^{-3} m$$
,  $A_i = \pi r_i^2 = 0.507 \times 10^{-3} m^2$ ,  $\lambda$  (copper) = 391  $\frac{W}{m-K}$ ,  
 $r_o = 15.9 \times 10^{-3} m$ ,  $A_o = \pi r_o^2 = 0.792 \times 10^{-3} m^2$ ,  $\lambda$  (brass) = 111  $\frac{W}{m-K}$ ,  
 $t = 0.8 \times 10^{-3} m$ ,  $A_x = 2\pi r_m t = 3.175 \times 10^{-3} m^2$ ,  
 $P = 0.115 m$ ,  $L = 3.2 \times 10^{-3} m$ 

readings from two separate thermocouples,  $T_1$  and  $T_2$ , denoted by 4 and 5 in Fig. 3.1. The wall superheat  $\Delta T$  was calculated by taking the difference of  $T_w$  and  $T_{sat}$ , i.e.,

where 
$$\Delta T = T_w - T_{sat} [K]$$
(B.1.1)  
$$\Delta T = \frac{1}{2}(T_1 + T_2) [K]$$

The objective was to determine the wall superheat and corresponding heat transfer coefficient from the central portion of the disk. The boiling process was observed only from this central portion of radius 12.7 mm. Hence, the thin fin around the surface proved to be an effective way for confining the boiling process to the main surface and thus minimizing losses from it. However, a small percentage of the total power was still expected to be lost through the fin. (Note: for the incipience tests this fin was needed; any joint formed by using an insulation packing around the radius  $r_i$ , instead of using the fin, would have become the point at which incipience preferentially occur. At low heat flux levels, i.e. before the initiation of boiling, the entire surface is in the natural convection mode and as a first approximation the heat transfer coefficient  $\alpha$  is obtained using the relation

$$\alpha = \frac{\dot{q}}{\Delta T} \qquad \begin{bmatrix} \frac{W}{2} \\ m \\ K \end{bmatrix} \qquad (B.1.2)$$

where

 $\frac{1}{q} = \frac{VI}{A}$ 

$$\begin{bmatrix} \frac{W}{2} \\ m \end{bmatrix}$$
(B.1.3)

V is the voltage drop (volts) across the heater and I is the current (amperes) in the circuit. Due to the dimensions, the circular fin can be treated as a rectangular fin as shown in Fig. B.1.1. As the thickness of the fin is very small, a one-dimensional variation was considered. The boundary conditions are:

$$T(x=0) = T_{w}$$
 (B.1.4)

$$\left(\frac{\mathrm{dT}}{\mathrm{dx}}\right)_{\mathbf{x}=\mathbf{L}} = 0 \tag{B.1.5}$$

The second boundary condition assumes that the tip of the fin is at the saturation temperature. Using the above boundary conditions the expression for the heat loss from the fin is

$$Q_{L} = \sqrt{\alpha P \lambda A_{x}} (T_{w} - T_{sat}) \tanh(mL) [W]$$
(B.1.6)

where

 $m = \frac{\alpha P}{\lambda A_x}$ 

The losses from the underside of the fin are neglected since it is on the same side as the heater and only in contact with stagnant air. The total power through the central portion of area  $A_i$  was then obtained from

$$Q = V I - Q_{I} \qquad [W] \qquad (B.1.7)$$

and the heat transfer coefficient given as

$$\alpha = \frac{\dot{q}}{\Delta T} \qquad \begin{bmatrix} \frac{W}{2} \\ m \end{bmatrix} \qquad (B.1.8)$$

where

$$\dot{q} = \frac{Q}{A_{i}}$$
  $[\frac{W}{2}]$  (B.1.9)

Once the boiling starts on the central portion of the surface, the procedure for calculating the heat transfer coefficients was different. As the fin was always in the natural convection mode, the losses from its surface were calculated using natural convection heat transfer coefficients. These coefficients were obtained by extrapolating those at lower temperatures to the temperature of interest. The losses were calculated using Eq. (B.1.6) and heat transfer coefficient for the main surface obtained from Eq. (B.1.8).

# B.2 Smooth Tube

The smooth tube test specimen is shown in Fig. 3.5. The saturation temperature of the liquid pool was measured by two separate thermocouples denoted by 4 and 5 in Fig. 3.1. The wall temperature of the test section was obtained by extrapolating the readings of four thermocouples located inside the wall of the tube. The voltage V (volts) and current I (amperes) to the test section were measured to obtain the power input to the test section. To estimate the losses from the unheated length of the tube, two thermocouples,  $T_{L1}$  and  $T_{L2}$ , were placed at the end of the copper tube, Fig. B.2.1.



Fig. B.2.1.

$$\begin{aligned} \mathbf{r_{i}} &= 6.3 \times 10^{-3} \text{ m}, \quad \mathbf{r_{t}} = 8.7 \times 10^{-3} \text{ m}, \quad \mathbf{r_{o}} = 11.1 \times 10^{-3} \text{ m}, \\ \mathbf{L_{b}} &= 76.2 \times 10^{-3} \text{ m}, \quad \mathbf{L_{nc}} = 12.7 \times 10^{-3} \text{ m}, \quad \mathbf{L} = \mathbf{L_{b}} + \mathbf{L_{nc}}, \\ \mathbf{A_{b}} &= 2 \pi \mathbf{r_{o}} \mathbf{L_{b}}, \quad \mathbf{A_{nc}} = 2 \pi \mathbf{r_{o}} \mathbf{L_{nc}}, \\ \mathbf{A_{b}} &= 2 \pi \mathbf{r_{o}} \mathbf{L_{b}}, \quad \mathbf{A_{nc}} = 2 \pi \mathbf{r_{o}} \mathbf{L_{nc}}, \\ \lambda \text{ (copper)} &= 391 \frac{W}{m - K} , \\ \mathbf{T_{sat}} = \frac{1}{2} (\mathbf{T_{1}} + \mathbf{T_{2}}), \quad \mathbf{T_{t}} = \frac{1}{4} (\mathbf{T_{3}} + \mathbf{T_{4}} + \mathbf{T_{5}} + \mathbf{T_{6}}), \quad \mathbf{T_{L}} = \frac{1}{2} (\mathbf{T_{L1}} + \mathbf{T_{L2}}) \quad [^{\circ}C] \end{aligned}$$

$$Q = \frac{(T_w - T_{sat})}{\frac{1}{(2\pi\lambda_c L_b)} \ln(\frac{r_o}{r_t})}$$
[W]

$$T_{w} = T_{t} - Q \left(\frac{1}{2\pi\lambda_{c}L_{b}}\right) \ln(\frac{r_{o}}{r_{t}})$$
 [°C] (B.2.1)

$$\Delta T = T_{w} - T_{sat}$$
 [K] (B.2.2)

$$\alpha = \frac{\dot{q}}{\Delta T} \qquad \left[\frac{W}{m}\right] \qquad (B.2.3)$$

where  $\dot{q} = \frac{Q}{2\pi r_0 L_b} \qquad \left[\frac{W}{m}\right]$ 

For the natural convection regime it is assumed that the entire surface is at the uniform temperature,  $T_w$ . As a first approximation with Q-VI in Eq. (B.2.1),  $\Delta T$  and  $\alpha$  were obtained from Eqs. (B.2.2) and (B.2.3). The losses from area  $A_{nc}$  were then estimated from this value of  $\alpha$  from the expression

$$Q_{L} = -\lambda_{c} A_{x} \left( \frac{\Delta T_{L}}{L_{nc}} \right) + \alpha A_{nc} \Delta T_{m} \qquad [W] \qquad (B.2.4)$$

[K]

[K]

where

and

$$\Delta T_{m} = \frac{1}{2} (T_{w} - T_{L})$$

 $\Delta T_{I} = T_{W} - T_{I}$ 

A new value of Q-VI-Q<sub>L</sub> is then substituted in Eq. (B.2.1) to obtain  $\Delta T$ and  $\alpha$  from Eqs.(B.2.2) and (B.2.3).

When boiling started on the heated length the unheated length was observed to be in the natural convection mode. To estimate losses from the unheated length,  $L_{nc}$ , the value of  $\alpha$  used in Eq. (B.2.4) was obtained by extrapolating the heat transfer coefficients in natural convection regime to the mean temperature  $\Delta T_m$ . From this estimate of losses a new value of Q-VI-Q<sub>L</sub> was substituted in Eq. (B.2.1) and  $\Delta T$  and  $\alpha$  were obtained from Eqs. (B.2.2) and (B.2.3).

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## **B.3** Enhanced Tube

The sketch of the test section is shown in Fig. 3.8. The procedure for estimating the wall superheat and heat transfer coefficients was similar to that described for the smooth tube. The only difference is the composite nature of the enhaned surface test section. Also, when this surface was constructed, the thermocouples for measuring the losses from the end were not incorporated in the design. Hence, for all heat flux levels the temperature at the end of the tube was assumed to be at the saturation value. This assumption, in the case of enhanced surface is justified for the following reasons: (1) the wall superheats are very small, (2) very vigorous boiling was observed on the main surface above the heater resulting in very high heat transfer coefficients, and (3) the unheated length of the tube experiences natural convection even at the highest heat flux.



Fig. B.3.1

$$r_{i} = 4.8 \times 10^{-3} \text{ m}, r_{t} = 6.6 \times 10^{-3} \text{ m}, r_{c} = 8.5 \times 10^{-3} \text{ m}, r_{o} = 9.3 \times 10^{-3} \text{ m},$$

$$L_{b} = 0.152 \text{ m}, L_{nc} = 12.5 \times 10^{-3} \text{ m}, L = L_{b} + L_{nc}$$

$$A_{b} = 2\pi r_{o}L_{b}, A_{nc} = 2\pi r_{o}L_{nc} \qquad [m^{2}]$$

$$\lambda_{c} \text{ (copper)} = 391 \frac{W}{m-K}, \lambda_{e} \text{ (enhanced)} = 242 \frac{W}{m-K}$$

$$T_{sat} = \frac{1}{2} (T_{1} + T_{2}), T_{t} = \frac{1}{4} (T_{2} + T_{3} + T_{5} + T_{6}) \qquad [^{\circ}C]$$

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$$Q = \frac{(T_{w} - T_{sat})}{\frac{1}{(2\pi\lambda_{c}L_{b})} \ln(\frac{r_{c}}{r_{t}}) + \frac{1}{(2\pi\lambda_{e}L_{b})} \ln(\frac{r_{o}}{r_{c}})} [W]$$

$$T_{w} = T_{t} - Q \left[ \frac{1}{2\pi\lambda_{c}L_{b}} \ln(\frac{r_{c}}{r_{t}}) + \frac{1}{2\pi\lambda_{e}L_{b}} \ln(\frac{r_{o}}{r_{c}}) \right] \quad [^{\circ}C] \quad (B.3.1)$$

$$\Delta T = T_{w} - T_{sat}$$
 [K] (B.3.2)

$$\alpha = \frac{\dot{q}}{\Delta T} \qquad \begin{bmatrix} \frac{W}{2} \\ m & K \end{bmatrix}$$
(B.3.3)

where  $\dot{q} = \frac{Q}{2\pi r_o L_b} \qquad [\frac{W}{2}]$ 

ΔT

For the natural convection regime it was assumed that the entire surface was at the uniform temperature,  $T_w$ . As a first approximation with Q-VI in Eq. (B.3.1),  $\Delta T$  and  $\alpha$  were obtained from Eqs. (B.3.2) and (B.3.3). The losses from area  $A_{nc}$  were then estimated from the value of  $\alpha$  from from Eq. (B.3.3)

$$Q_{L} = \alpha A_{nc} \Delta T_{m} \qquad [W] \qquad (B.3.4)$$

where

$$m = \frac{1}{2} (T_w - T_{sat}) [K]$$

A new value of Q=VI-Q<sub>L</sub> was then substituted in Eq. (B.3.1) to obtain  $\Delta T$ and  $\alpha$  from Eqs.(B.3.2) and (B.3.3).

When boiling started on the heated length, the unheated length remained in the natural convection mode. To estimate losses from the unheated length,  $L_{nc}$ , the value of  $\alpha$  used in Eq. (B.3.4) was obtained by extrapolating the heat transfer coefficients in natural convection to the mean temperature  $\Delta T_m$ . From this estimate of losses a new value of Q=VI-Q<sub>L</sub> was substituted in Eq. (B.3.1) and  $\Delta T$  and  $\alpha$  obtained from Eqs. (B.3.2) and (B.3.3), respectively.

# **B.4** Estimation of Errors in Experimental Data

The experimental error in the wall superheats measured are estimated to be  $\pm 0.2$  K. The error in the heat flux is estimated to be  $\pm 6$ %. The error in the heat transfer coefficients thus range from  $\pm 25$ % to  $\pm 8$ % in going from low end of the enhanced surface boiling curve to the top end of the smooth surface boiling curve. APPENDIX C
## APPENDIX C

## EXPERIMENTAL DATA OBTAINED ON SMOOTH AND ENHANCED SURFACES

$$\dot{q} \qquad heat flux \qquad \left[\frac{kW}{m}\right]$$

$$T_{sat} \qquad saturation temperature [°C]$$

$$T_{w} \qquad wall temperature [°C]$$

$$\Delta T_{id} \qquad ideal wall superheat, Eqs. (2.18) and (2.19) \qquad [K]$$

$$\Delta T \qquad wall superheat (-T_{w} - T_{sat}) \qquad [K]$$

$$\alpha \qquad heat transfer coefficient (-\frac{\dot{q}}{\Delta T}) \qquad \left[\frac{kW}{m}\right] \\ \alpha_{id} \qquad ideal heat transfer coefficient (-\frac{\dot{q}}{\Delta T_{id}}) \qquad \left[\frac{kW}{m}\right]$$

$$Ratio \qquad = \frac{\Delta T_{id}}{m} = \frac{\alpha}{m}$$

Ratio = 
$$\frac{\Delta T_{id}}{\Delta T} - \frac{\alpha}{\alpha_{id}}$$

METHANOL-WATER	ENHANCED TUBE		MOLE FR 0.00 (PURE WATER)		T <sub>sat</sub> - 100.0°C		
P	ΔT	a					
99.1	4.6	21.5					
82.0 66.4	4.3	18.9 16 7					
52.7	3.7	14.3					
40.4	3.4	12.0					
29.9	3.1	9.8 7.2					
13.6	2.4	5.6					
7.6	2.1	3.6					
3.5	1.3	2.6					
METHANOL WATER	ENHAN	CED TUBE	MOLE FR.	- 0.05	T <sub>sat</sub> = 92.6°C		
• q	∆T <sub>id</sub>	ΔT	a id	α	Ratio		
98.3	4.5	6.2	21.6	15.8	0.73		
81.2	4.3	5.6	19.1	14.4	0.75		
66.0	3.9	5.1	16.8	13.0	0.77		
52.3 40.3	3.8	4.7	14.4	9.3	0.78		
29.8	3.0	3.9	9.9	7.6	0.77		
20.9	2.9	3.5	7.3	6.0	0.82		
13.3	2.4	3.1	5.6	4.2	0.76		
7.0	2.1	2.6	3.6	2.8	0.79		
	1.5	1.0	2.0	2.1	0.01		
METHANOL WATER	ENHAN	CED TUBE	MOLE FR.	- 0.10	T <sub>sat</sub> - 87.8°C		
ŗ	∆T <sub>id</sub>	ΔΤ	a <sub>id</sub>	a	Ratio		
98.3	4.5	6.2	21.9	15.8	0.72		
81.3	4.2	5.7	19.4	14.3	0.74		
66.0	3.9	5.0	17.0	13.1	0.77		
52.1	3.6	4.6	14.6	11.4	0.78		
40.5 29 R	3.Z	4.1 3.6	10 1	9.9 8 0	0.00		
20.6	2.8	3.2	7.4	6.3	0.86		
13.2	2.3	2.8	5.6	4.6	0.83		
7.6	2.0	2.5	3.7	3.0	0.82		
3.4	1.3	1.8	2.6	1.9	0.72		

Table C.1.1. Methanol-water mixtures - Enhanced tube (High Flux)

METHANOL-WATER	ENHANC	ED TUBE	MOLE FR.	- 0.25	T <sub>sat</sub> - 80.0°C
ġ	۵T <sub>id</sub>	ΔΤ	a <sub>id</sub>	a	Ratio
99.1	4.3	6.0	22.9	16.6	0.72
81.5	4.0	5.4	20.3	15.1	0.74
66.2	3.7	4.8	17.9	13.8	0.77
52.5	3.4	4.3	15.5	12.3	0.79
40.3	3.1	3.8	13.1	10.6	0.81
29.8	2.0	3.4	10.7	8.8 6.9	0.82
20.8	2.0	5.0 2 7	6.0	5.0	0.80
7.6	1.9	2.3	3.9	3.3	0.82
3.5	1.3	1.7	2.7	2.0	0.74
METHANOL-WATER	ENHANC	CED TUBE	MOLE FR.	- 0.40	T75.5°C
ġ	∆T <sub>id</sub>	ΔТ	a id	a	Ratio
98.6	4.1	5.9	23.8	16.6	0.70
81.5	3.8	5.3	21.3	15.4	0.72
66.0	3.5	4.7	18.6	13.9	0.75
52.3	3.2	4.1	16.3	12.6	0.77
40.3	2.9	3./	13.9	10.9	0.78
29.6	2.0	3.3	11.5	8.9 7 1	0.77
13 3	2.4	2.9	6.5	7.1 5 1	0.85
7.6	1.8	2.4	4.2	3.2	0.76
3.4	1.2	1.9	2.7	1.7	0.64
METHANOL-WATER	ENHAN	CED TUBE	MOLE FR.	. = 0.55	T = 72.4°C
ġ	∆T <sub>id</sub>	ΔΤ	a <sub>id</sub>	α	Ratio
98.3	4.0	5.6	24.7	17.6	0.712
81.5	3.6	5.1	22.5	16.1	0.717
66.0	3.4	4.5	19.5	14.5	0.744
52.3	3.0	4.1	17.2	12.8	0.744
40.3	2.7	3.7	14.8	10.8	0.732
29.8	2.4	3.2	12.3	9.2 7 2	U./49 0 700
20./ 13 /	2.2	2.8 2.7	9.2 7 0	/.J 5 0	U./89 0 79%
7 6	1 7	2.7	45	3.0	0.724
3.5	1.2	1.8	2.9	2.0	0.671

Table C.1.1. (continued)

METHANOL-WATER	ENHAN	CED TUBE	MOLE FR.	- 0.70	T69.6°C
• q	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
98.6	3.8	5.3	25.9	18.5	0.71
81.5	3.4	5.0	23.7	16.4	0.69
66.4	3.2	4.5	20.7	14.7	0.71
52.3	2.9	4.1	18.3	12.7	0.69
40.3	2.5	3.8	15.8	10.7	0.67
29.9	2.2	3.4	13.4	8.7	0.65
20.7	2.1	3.0	10.0	6.9	0.69
13.3	1.8	2.7	7.4	4.9	0.66
7.6	1.6	2.3	4.8	3.3	0.69
3.5	1.2	1.9	2.9	1.8	0.62
METHANOL-WATER	ENHAN	CED TUBE	MOLE FR.	- 0.85	T <sub>sat</sub> - 67.3°C
ġ	ΔΤ	ΔТ	<b>a</b>	a	Ratio
T	1d		-id	_	
98.6	3.6	5.7	27.2	17.3	0.64
81.5	3.2	5.2	25.1	15.6	0.62
66.2	3.0	4.6	21.7	14.3	0.66
52.3	2.7	4.0	19.5	13.0	0.66
40.3	2.4	3.5	17.0	11.6	0.68
29.8	2.0	3.0	14.5	9.9	0.68
20.9	1.9	2.6	11.1	7.9	0.71
13.4	1.7	2.2	8.1	6.1	0.74
7.6	1.5	1.8	5.1	4.1	0.79
3.4	1.2	1.5	2.9	2.3	0.77
METHANOL-WATER	ENHAN	CED TUBE	MOLE FR. (PURE ME	- 1.00 THANOL)	T <sub>sat</sub> - 64.8°C
ġ	ΔT	a			
98.3	3.4	28.5			
81.5	3.1	26.6			
66.2	2.9	23.0			
52.3	2.5	20.9			
40.4	2.2	18.5			
29.9	1.9	16.0			
20.9	1.7	12.3			
13.4	1.5	8.9			
7.6	1.3	5.6			
3.5	1.1	3.0			

Table C.1.1. (continued)

METHANOL-WATER	SMOO!	TH TUBE	MOLE FR (PURE WAT)	0.00 ER)	T <sub>sat</sub> - 100.6°C
ġ	ΔT	α			
216.4	16.2	13.4			
192.4	15.6	12.3			
142.8	14.4	9.9			
94.5	12.4	7.6			
69.8	11.0	6.3			
46.1	9.0	5.1			
27.6	6.3	4.4			
18.3	4.5	4.1			
9.1	2.5	3.6			
MERIANOT HATTER	CHOOL	<b>11 111</b> 100		0.05	T 02 880
METHANOL-WATEK	2400	LH TUBE	MULE FR	0.05	sat = 92.8°C
ŗ	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
215.9	16.4	22.8	13.2	9.5	0.72
191.3	15.8	22.1	12.1	8.6	0.71
143.0	14.6	19.4	9.8	7.4	0.75
93.6	12.7	16.2	7.3	5.8	0.78

Table C.1.2. Methanol-water mixture - Smooth tube

				sat 52.00	
<sup>∆T</sup> id	ΔT	a <sub>id</sub>	a	Ratio	
16.4	22.8	13.2	9.5	0.72	
15.8	22.1	12.1	8.6	0.71	
14.6	19.4	9.8	7.4	0.75	
12.7	16.2	7.3	5.8	0.78	
11.3	13.7	6.1	5.0	0.82	
9.4	11.2	4.9	4.1	0.84	
6.7	7.8	4.1	3.5	0.85	
4.8	5.8	3.9	3.2	0.83	
2.7	3.2	3.5	2.9	0.84	
	<b>AT</b> id 16.4 15.8 14.6 12.7 11.3 9.4 6.7 4.8 2.7	ΔT         ΔT           16.4         22.8           15.8         22.1           14.6         19.4           12.7         16.2           11.3         13.7           9.4         11.2           6.7         7.8           4.8         5.8           2.7         3.2	ΔT idΔT 22.8a id16.422.813.215.822.112.114.619.49.812.716.27.311.313.76.19.411.24.96.77.84.14.85.83.92.73.23.5	ΔT idΔT a idα16.422.813.29.515.822.112.18.614.619.49.87.412.716.27.35.811.313.76.15.09.411.24.94.16.77.84.13.54.85.83.93.22.73.23.52.9	

T<sub>sat</sub> - 87.9°C METHANOL-WATER SMOOTH TUBE MOLE FR. = 0.10ġ ∆T<sub>id</sub> ΔT a<sub>id</sub> α Ratio 215.4 16.6 27.9 12.9 7.7 0.60 192.2 16.0 26.2 12.0 7.3 0.61 142.5 9.6 6.1 23.2 14.9 0.64 93.7 7.2 4.8 13.0 19.5 0.67 69.4 11.6 16.9 6.0 4.1 0.69 3.5 45.9 9.7 13.1 4.7 0.74 27.2 7.0 8.8 3.9 3.1 0.80 21.7 5.1 6.3 4.3 2.6 0.61

3.3

3.5

9.3

2.8

2.7

0.82

METHANOL-WATER	SMOO'	TH TUBE	MOLE FR	0.25	T = 80.1°C sat
ġ	∆T <sub>id</sub>	ΔΤ	a <sub>id</sub>	a	Ratio
216.0	17.3	31.5	12.5	6.9	0.55
192.3	16.7	30.2	11.5	6.4	0.55
143.0	15.6	26.9	9.1	5.3	0.58
94.0	13.9	22.6	6.7	4.2	0.62
69.8	12.6	19.9	5.5	3.5	0.63
46.0	10.8	15.7	4.3	2.9	0.69
27.9	8.1	10.6	3.4	2.6	0.76
18.3	6.0	/.6	3.0	2.4	0.79
9.2	3.4	4.1	2.7	2.2	0.83
METHANOL-WATER	SMOOT	TH TUBE	MOLE FR	0.40	T <sub>sat</sub> - 75.6°C
ġ	∆T <sub>id</sub>	ΔΤ	<i>a</i> id	α	Ratio
216.3	18.0	31.1	12.0	6.9	0.58
192.4	17.4	29.9	11.1	6.4	0.58
143.6	16.4	27.5	8.8	5.2	0.59
93.9	14.8	24.6	6.3	3.8	0.60
69.3	13.6	22.6	5.1	3.1	0.60
45.4	11.9	18.6	3.8	2.4	0.64
27.6	9.2	12.5	3.0	2.2	0.73
18.3	6.9	8.9	2.6	2.1	0.78
9.2	3.9	4.0	2.3	2.0	0.85
METHANOL-WATER	SMOO?	TH TUBE	MOLE FR	0.55	T <sub>sat</sub> = 72.6°C
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
217.1	18.7	30.7	11.6	7.1	0.607
192.5	18.1	30.0	10.6	6.4	0.602
142.8	17.1	28.5	8.3	5.0	0.600
93.6	15.7	25.9	5.9	3.6	0.607
69.3	14.6	24.0	4.7	2.9	0.610
45.9	13.0	19.4	3.5	2.4	0.668
27.5	10.2	13.0	2.7	2.1	0.785
18.3	7.8	9.4	2.3	1.9	0.828
9.0	4.4	5.3	2.0	1.7	0.839

Table C.1.2. (continued)

Table C.1.2. (continued)

METHANOL-WATER	SMOOT	TH TUBE	MOLE FR. =	0.70	r <sub>sat</sub> - 69.8°C
ġ	∆T <sub>id</sub>	ΔT	<i>a</i> id	a	Ratio
218.3	19.3	28.9	11.3	7.5	0.67
193.4	18.8	28.3	10.3	6.8	0.66
144.2	17.9	27.3	8.1	5.3	0.65
94.5	16.6	25.8	5.7	3.7	0.64
69.7	15.6	24.3	4.5	2.9	0.64
45.6	14.0	21.7	3.2	2.1	0.65
27.7	11.3	15.2	2.5	1.8	0.75
18.4	8.7	10.5	2.1	1.7	0.83
9.2	5.0	5.7	1.8	1.6	0.87
METHANOL-WATER	SMOOT	TH TUBE	MOLE FR	0.85 T	sat - 67.2°C
ġ	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
218.5	20.0	25.0	10.9	8.7	0.80
193.7	19.4	24.2	10.0	8.0	0.80
144.3	18.6	22.8	7.7	6.3	0.82
93.7	17.5	21.2	5.3	4.4	0.83
69.7	16.6	20.2	4.2	3.4	0.82
45.8	15.1	18.0	3.0	2.5	0.84
27.2	12.4	13.7	2.2	2.0	0.90
17.8	9.6	9.9	1.8	1.8	0.97
9.0				4 7	4 ^ /
	5.5	5.3	1.6	1./	1.04

METHANOL-WATER	SMOO?	TH TUBE	MOLE FR 1.00 (PURE METHANOL)	T <sub>sat</sub> - 64.5°C
ġ	ΔT	a		
218.6	20.7	10.6		
194.1	20.1	9.6		
144.1	19.4	7.4		
94.7	18.4	5.1		
70.1	17.6	4.0		
45.8	16.2	2.8		
27.7	13.5	2.1		
18.2	10.5	1.7		
9.2	6.0	1.5		

ETHANOL-WATER	ENHANCED TUBE		MOLE FR. = 0.00 (PURE WATER)		T <sub>sat</sub> = 100.3°C	
• P	ΔΤ	a				
99.7 75.0 49.6 25.1	5.0 4.6 4.0 3.3	20.1 16.2 12.3 7.7				
ETHANOL-WATER	ENHANC	ed <b>T</b> ube	MOLE FR.	- 0.06	T <sub>sat</sub> - 89.1°C	
• q	∆T <sub>id</sub>	ΔТ	a <sub>id</sub>	a	Ratio	
99.6 75.0 49.8 25.0	4.9 4.6 4.0 3.2	5.6 4.9 4.2 3.5	20.2 16.4 12.6 7.8	17.9 15.4 11.9 7.2	0.88 0.94 0.95 0.92	
ETHANOL-WATER	ENHANC	ED TUBE	MOLE FR.	- 0.14	T <sub>sat</sub> - 85.0°C	

Table	C.2.1.	Ethanol-water	mixtures	-	Enhanced	tube	(High	Flux)	•
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• q	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
99.5	4.9	4.8	20.4	20.9	1.02
75.1	4.5	4.1	16.6	18.3	1.10
49.8	3.9	3.6	12.8	13.9	1.09
24.9	3.1	2.9	8.0	8.5	1.07

ETHANOL-WATER	ENHANCED TUBE		MOLE FR.	= 0.25	T <sub>sat</sub> = 82.9°C	
• P	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio	
99.9	4.8	5.3	20.8	18.8	0.90	
75.2	4.4	4.4	17.0	17.2	1.01	
49.9	3.8	3.7	13.2	13.5	1.02	
25.0	3.0	3.0	8.3	8.3	1.00	

ETHANOL-WATER	ENHANC	ED TUBE	MOLE FR.	- 0.35	$T_{sat} = 81.6$ °C
ġ	∆T <sub>id</sub>	ΔT	<sup>a</sup> id	a	Ratio
99 4	4.8	4.6	20.9	21.7	1.04
75.2	4.3	3.9	17.4	19.1	1.10
49.9	3.7	3.3	13.6	15.1	1.11
25.0	2.9	2.6	8.6	9.5	1.10
ETHANOL-WATER	ENHANC	KD TUBE	MOLE FR.	- 0.50	T <sub>sat</sub> - 80.1°C
• q	∆T <sub>id</sub>	ΔT	<b>a</b> id	α	Ratio
99.6	4.7	4.0	21.3	24.9	1.17
75.1	4.2	3.4	17.9	21.8	1.22
49.8	3.5	2.9	14.2	17.3	1.22
25.0	2.7	2.3	9.1	10.8	1.19
ETHANOL-WATER	ENHANC	ED TUBE	MOLE FR.	- 0.65	T <sub>sat</sub> = 79.2°C
• q	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
99 9	4.6	38	21 7	26 5	1,22
75.1	4.1	3.3	18.5	22.5	1.22
49.8	3.3	2.8	14.9	17.6	1.19
25.0	2.6	2.4	9.6	10.5	1.08
ETHANOL-WATER	ENHANC	ED TUBE	MOLE FR.	- 0.80	T <sub>sat</sub> - 78.7°C
ġ	∆T <sub>id</sub>	ΔТ	a <sub>id</sub>	a	Ratio

4.5

4.0

3.1

2.4

22.1

19.0

15.6

10.3

22.2

18.8

15.8

10.6

1.00

0.99

1.01

1.03

4.5

3.9

3.2

2.4

99.6

75.0

49.8

25.0

Table C.2.1. (continued)

Table C.2.1. (continued)

ETHANOL-WATER	ENHANC	ED TUBE	MOLE FR 0.89 (AZEOTROPE)	T <sub>sat</sub> - 78.3°C
ġ	ΔT	a		
99.4	4.5	22.3		
75.1	3.9	19.5		
49.7	3.1	16.0		
24.8	2.3	10.6		

ETHANOL-WATER	ENHANCED	TUBE	MOLE FR 1.00 (PURE ETHANOL)	T <sub>sat</sub> = 78.75°C
ġ	ΔT	a		
99.4	3.6	27.9		
75.1	3.2	23.8		
49.8	2.5	19.5		
25.0	2.0	12.6		

ETHANOL-WATER	SMOOTH	TUBE	MOLE FR 0.00 (PURE WATER)	T <sub>sat</sub> - 100.2°C
ġ	ΔΤ	a		
218.9	15.7	14.0		
193.9	15.2	12.8		
144.6	14.0	10.3		
95.4	12.3	7.8		
71.2	10.9	6.5		
46.9	9.4	5.0		
28.5	6.8	4.2		
18.8	4.8	3.9		
9.4	2.8	3.4		

Table C.2.2. Ethanol-water mixtures - Smooth tube.

ETHANOL-WATER	SMOOTH	TUBE	MOLE FR	0.05 T	at - 90.1°C
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
218.4	15.8	24.1	13.8	9.1	0.65
193.6	15.3	23.4	12.7	8.3	0.65
144.3	14.1	21.3	10.2	6.8	0.66
94.8	12.4	18.5	7.7	5.1	0.67
70.8	11.0	16.6	6.4	4.2	0.66
46.5	9.6	13.1	4.9	3.5	0.73
27.7	7.0	8.7	3.9	3.2	0.80
18.7	5.1	6.2	3.7	3.0	0.82
9.5	3.0	3.5	3.2	2.7	0.85

T<sub>sat</sub> - 86.0°C ETHANOL-WATER SMOOTH TUBE MOLE FR. -0.10ġ ∆T<sub>id</sub> ΔT Ratio α a<sub>id</sub> 217.4 15.9 27.8 13.7 7.8 0.57 193.2 15.4 26.0 12.5 7.4 0.59 14.2 144.3 23.1 10.2 6.2 0.62 94.4 12.5 20.0 7.6 4.7 0.62 70.7 11.2 17.8 6.3 4.0 0.63 3.0 46.3 9.7 15.2 4.8 0.64 27.7 7.2 10.5 3.8 2.6 0.69 5.3 7.3 18.6 3.5 2.5 0.72 9.4 3.2 4.1 2.9 2.3 0.77

ETHANOL-WATER	SMOOTH	I TUBE	MOLE FR.	- 0.20	T <sub>sat</sub> - 83.2°C
• q	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
216.7	16.1	24.5	13.5	8.8	0.66
192.8	15.6	22.1	12.4	8.7	0.71
143.6	14.4	19.3	10.0	7.4	0.75
94.6	12.7	17.5	7.4	5.4	0.72
70.6	11.4	15.7	6.2	4.5	0.73
46.2	10.0	13.4	4.6	3.4	0.75
27.7	7.7	10.4	3.6	2.7	0.73
18.4	5.7	7.8	3.2	2.4	0.74
9.2	3.6	4.4	2.5	2.1	0.83
ETHANOL-WATER	SMOOTE	I TUBE	MOLE FR.	- 0.35	T <sub>sat</sub> - 81.2°C
ġ	∆T <sub>id</sub>	ΔТ	<b>a</b> id	a	Ratio
217.4	16.4	22.6	13.2	9.6	0.73
193.2	15.9	21.7	12.1	8.9	0.73
143.8	14.7	19.2	9.8	7.5	0.76
95.0	13.0	17.5	7.3	5.4	0.75
70.5	11.8	16.4	6.0	4.3	0.72
46.0	10.5	14.8	4.4	3.1	0.71
27.7	8.3	12.1	3.3	2.3	0.68
18.3	6.4	9.5	2.8	1.9	0.68
9.1	4.3	5.3	2.1	1.7	0.81
ETHANOL-WATER	SMOOTH	I TUBE	MOLE FR.	- 0.50	T <sub>sat</sub> = 79.9°C
ġ	∆T <sub>id</sub>	ΔT	α <sub>id</sub>	a	Ratio
217.1	16.8	20.6	13.0	10.5	0.81
192.9	16.2	19.8	11.9	9.8	0.82
144.5	15.0	17.6	9.6	8.2	0.85
94.7	13.4	16.1	7.1	5.9	0.83
70.8	12.2	15.3	5.8	4.6	0.80
45.9	10.9	14.3	4.2	3.2	0.76
27.3	8.9	12.1	3.1	2.3	0.74
18.3	7.1	10.3	2.6	1.8	0.69
9.2	4.9	6.0	1.9	1.5	0.82

Table C.2.2. (continued)

Table C.2.2. (continued)

144.4

94.7

70.4

45.8

27.3

18.4 9.2

ETHANOL-WATER	SMOOTH	TUBE	MOLE FR (	0.65	T <sub>sat</sub> = 79.0°C
ŗ	۵T <sub>id</sub>	ΔΤ	<b>a</b> id	α	Ratio
217.8	17.1	17.2	12.7	12.7	0.99
193.7	16.6	16.6	11.7	11.7	1.00
144.0	15.3	15.3	9.4	9.4	1.00
94.6	13.7	13.9	6.9	6.8	0.98
70.5	12.6	13.0	5.6	5.4	0.97
45.9	11.4	11./	4.0	3.9	0.97
2/./	9.6	10.2	2.9	2.1	0.94
10.4 Q 3	7.0	59	2.5	1 6	0.92
					<b>m</b> 70.08 <i>a</i>
ETHANOL-WATER	SMOOTH	TUBE	MOLE $FR. = 0$	0.80	T = /8.2°C sat
ġ	∆T <sub>id</sub>	ΔΤ	<b>α</b> id	a	Ratio
218.3	17.4	17.3	12.5	12.6	1.01
193.7	16.9	16.8	11.5	11.5	1.00
144.6	15.6	15.4	9.3	9.4	1.01
94.9	14.0	14.0	6.8	6.8	1.00
/0./	13.0	13.1	5.4	5.4	1.00
45.7 27 /	10.2	12.1	3.9	3.8	0.98
27.4	85	9 1	2.7	2.5	0.95
9.2	6.1	6.2	1 5	1 5	0.95
ETHANOL-WATER	SMOOTH	TUBE	MOLE FR ( (AZEOTROP)	0.89 E)	T <sub>sat</sub> - 78.0°C
ġ	ΔT	α			
218.7	17.6	12.4			
193.6	17.1	11.3			

9.1

6.7 5.3

3.8

2.6

2.1

1.4

15.8

14.2

13.3

12.1

10.6

8.9

6.5

1	9	6

Table C.2.2. (continued)

ETHANOL-WATER	SMOOTH	I TUBE	MOLE FR 1.00 (PURE ETHANOL)	T <sub>sat</sub> = 78.3°C
• Q	ΔT	a		
218.8	16.8	13.0		
194.5	16.2	12.0		
144.6	15.0	9.6		
95.1	13.7	6.9		
70.7	12.9	5.5		
46.1	11.8	3.9		
27.8	10.4	2.7		
18.5	8.9	2.1		
9.3	6.5	1.4		

qΔTα217.316.912.8193.216.112.0144.514.69.994.612.57.670.710.96.527.37.53.6	
$\begin{array}{cccccccccccccccccccccccccccccccccccc$	
193.2       16.1       12.0         144.5       14.6       9.9         94.6       12.5       7.6         70.7       10.9       6.5         27.3       7.5       3.6	
144.5       14.6       9.9         94.6       12.5       7.6         70.7       10.9       6.5         27.3       7.5       3.6	
94.6         12.5         7.6           70.7         10.9         6.5           27.3         7.5         3.6	
70.7       10.9       6.5         27.3       7.5       3.6	
27.3 7.5 3.6	
ETHANOL-WATER POLISHED TUBE MOLE FR 0.14 T <sub>sat</sub> - 8	35.0°C
q ΔΤ <sub>id</sub> ΔΤ α <sub>id</sub> α Rati	Ĺ <b>o</b>
217 / 18 3 32 7 11 8 6 6 0 56	:
	, ,
1/2.0 $1/.4$ $30.0$ $11.1$ $0.3$ $0.3/$	
	<i>)</i> )
95./ 15.0 10.0 /.U 5.1 U./2	2
/1.1 12.1 1/.3 5.9 4.1 0.70	)
ETHANOL-WATER POLISHED TUBE MOLE FR. = 0.25 T = 8 sat	32.7°C
q ΔT <sub>id</sub> ΔT α <sub>id</sub> α Rati	Ĺo
216.1 19.5 28.0 11.1 7.7 0.70	)
	, ,
143 9 16 7 22 8 8 6 6 3 0 7	, X
94 6 14 5 18 8 6 5 5 0 0.73	, ,
70.1         13.0         16.9         5.4         4.1         0.77	7
ETUANOL LATED DOLLCUED TIDE MOLE ED $-0.35$ T $-0.65$	01 / °C
sat	JI.4 U
q ΔT <sub>id</sub> ΔT α <sub>id</sub> α Rati	Lo
216.4 20.5 28.8 10.6 7.5 0.71	L
216.4       20.5       28.8       10.6       7.5       0.71         192.7       19.5       27.5       9.9       7.0       0.71	L
216.420.528.810.67.50.71192.719.527.59.97.00.71144.417.625.38.25.70.70	L L )
216.420.528.810.67.50.71192.719.527.59.97.00.71144.417.625.38.25.70.7094.715.322.16.24.30.69	L L ) )

Table C.2.3. Ethanol-water mixtures - Polished tube.

Table C.2.3. (continued)

ETHANOL-WATER	POLISH	ED TUBE	MOLE FR	0.50	T <sub>sat</sub> - 80.1°C
ġ	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
216.7	22.0	29.3	9.8	7.4	0.75
192.4	20.9	28.2	9.2	6.8	0.74
143.2	18.9	25.7	7.6	5.6	0.74
94.5	16.5	22.7	5.7	4.1	0.73
70.2	15.2	21.0	4.6	3.3	0.72
ETHANOL-WATER	POLISH	ED TUBE	MOLE FR	0.65	T <sub>sat</sub> - 79.3°C
ġ	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
215.9	23.6	27.7	9.2	7.8	0.85
193.1	22.4	26.6	8.6	7.2	0.84
143.1	20.2	24.5	7.1	5.8	0.82
93.7	17.7	22.2	5.3	4.2	0.80
69.6	16.4	20.6	4.2	3.4	0.80
ETHANOL-WATER	POLISE	IED TUBE	MOLE FR. =	= 0.80	T 78.4°C sat
•	۸ T	A 77	_	_	Dette
q	٥lid	Δι	aid	α	KALIO
216.6	25.1	26.0	8.6	8.3	0.96
192.4	23.8	25.1	8.1	7.7	0.95
143.2	21.5	23.0	6.6	6.2	0.94
93.8	19.0	21.1	5.0	4.4	0.90
69.6	17.7	19.8	3.9	3.5	0.89
ETHANOL-WATER	POLISH	IED TUBE	MOLE FR	- 0.89	T <sub>sat</sub> - 78.2°C
			(AZEUIK	/ <b>C</b> BJ	
ġ	ΔΤ	a			
216.2	26.0	8.3			
191.8	24.7	7.8			
143.0	22.3	6.4			
94.0	19.7	4.8			
69.5	18.5	3.8			

Table C.2.3. (continued)

ETHANOL-WATER	POLISH	ED TUBE	MOLE FR 1.00 (PURE ETHANOL)	T <sub>sat</sub> = 78.6°C
ġ	ΔT	a		
216.1	26.2	8.2		
192.6	24.8	7.8		
142.8	22.6	6.3		
93.6	20.6	4.5		
69.4	19.6	3.5		
45.4	17.9	2.5		
26.9	15.0	1.8		
17.9	12.6	1.4		
8.8	8.1	1.1		

PROPANOL-WATER	ENHANCED TUBE		MOLE FR. (PURE)	- 0.00 WATER)	T <sub>sat</sub> - 100.3°C
ġ	ΔΤ	a			
99.1	4.6	21.5			
82.0	4.3	19.0			
66.4 52 7	4.0	16.7			
40.4	3.4	12.0			
29.9	3.1	9.8			
21.0	2.9	7.2			
13.6	2.4	5.6			
3.5	1.3	2.6			
					_
PROPANOL-WATER	ENHAN	CED TUBE	MOLE FR.	- 0.06	T 89.3°C sat
ġ	∆T <sub>id</sub>	ΔT	<sup>a</sup> id	a	Ratio
98.8	4.4	5.9	22.2	16.8	0.76
81.7	4.1	5.0	19.7	16.2	0.82
66.4 52.5	3.8	4.3	1/.3	15.3	0.89
40.4	3.2	3.3	12.6	12.2	0.95
29.8	2.9	2.8	10.2	10.5	1.03
20.9	2.8	2.5	7.5	8.4	1.11
13.4	2.3	2.3	5.8	5.9	1.03
/.6	2.0	1.9	3./	4.1	1.09
3.5	1.5	1.4	2.0	2.4	0.95
PROPANOL-WATER	ENHAN	CED TUBE	MOLE FR.	- 0.15	T <sub>sat</sub> - 88.5°C
• q	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
98.6	4 2	4 2	23 4	22 2	0 99
81.5	3.9	3.7	20.9	22.3	1.07
66.4	3.6	3.2	18.4	20.6	1.12
52.5	3.3	2.9	15.9	18.1	1.14
40.3	3.0	2.6	13.6	15.7	1.16
29.8	2.7	2.3	11.0	13.1	1.19
20.7	∠.⊃ 2 2	2.0	0.2 6 2	8 A	1.24 1.28
7.5	1.9	1.5	3.9	4.9	1.23
3.4	1.3	1.3	2.6	2.7	1.03

Table C.3.1. Propanol-water mixture - Enhanced tube (High Flux).

PROPANOL-WATER	ENHAN	CED TUBE	MOLE FR 0.30		T <sub>sat</sub> = 87.7°C	
ġ	∆T <sub>id</sub>	ΔТ	a <sub>id</sub>	a	Ratio	
98.6	3.8	3.4	26.0	28.8	1.11	
81.5	3.5	3.2	23.3	25 2	1 08	
66 2	3 2	29	20 4	22.8	1 12	
52 7	29	2.7	18 0	19 7	1 10	
40 4	2.5	2.7	15.7	16 3	1.10	
20.4	2.0	2.5	10.7	13 1	1 03	
29.0	2.5	2.5	0 7	20	0.01	
20.9	2.1	2.4	<b>7</b> 1	0.9	0.91	
13.4	1.9	1.9	/.1	/.0	0.99	
7.6	1./	1.8	4.5	4.1	0.91	
3.4	1.3	1.7	2.7	2.0	0.77	
PROPANOL-WATER	ENHAN	CED TUBE	MOLE FR 0.43 (AZEOTROPE)		T <sub>sat</sub> - 87.8°C	
ġ	ΔΤ	a				
98.6	3.4	28.6				
81.7	3.1	26.1				
66.2	2.9	22.6				
52.5	2.6	20.2				
40.4	2.2	18.1				
29.8	2.0	14.7				
20.9	1.8	11.5				
13.3	1.7	8.0				
7.5	1.5	5.0				
3.4	1.3	2.6				
PROPANOL-WATER	ENHAN	CED TUBE	MOLE FR.	= 0.60	T = 88.1°C sat	
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	α	Ratio	
99 2	37	4 8	26 7	20 6	0 77	
81 5	3 4	4 4	24 1	18 5	0 77	
66 2	3 1	3 0	27.1	16 9	0.70	
52 5	2.1	3.5	10 0	1/ 0	0.75	
JZ.J 40 5	2.0	2.2	16 7	12 0	0.73	
40.5	2.4	2.1	10./	13.2	0.79	

Table C.3.1. (continued)

29.9

20.8

13.3

7.6

3.4

2.2

1.9

1.8

1.6

1.5

2.7

2.3

2.1

1.7

1.5

13.6

10.7

7.4

4.6

2.3

11.2

9.0 6.5

4.3

2.2

0.83

0.84

0.87

0.94

0.96

PROPANOL-WATER	ENHANCED TUBE		MOLE FR 0.77		T <sub>sat</sub> - 89.6°C	
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio	
98.8	4.0	6.1	24.9	16.1	0.65	
81.8	3.6	5.7	22.5	14.4	0.64	
66 2	3 3	53	20 2	12 6	0.62	
52 7	29	45	17 9	11 7	0.65	
40 5	2.5	4.0	15 5	10 1	0.65	
40.5	2.0	4.0	10.5	9 /	0.05	
29.9	2.4	3.5	12.5	0.4	0.67	
20.9	2.1	3.1	10.1	0./	0.87	
13.4	1.9	2.7	6.9	5.0	0.72	
7.6	1.8	2.3	4.2	3.3	0.78	
3.5	1.6	2.1	2.1	1.6	0.78	
PROPANOL-WATER	ENHANC	CED TUBE	MOLE FR.	- 0.90	T <sub>sat</sub> - 92.4°C	
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio	
98.6	4.2	6.3	23.6	15.6	0.66	
81.6	3.8	5.7	21.3	14.4	0.67	
66.2	3.4	5.2	19.4	12.7	0.65	
52 6	3 1	47	17 1	11 2	0 66	
40 4	2.1	4.7	14 7	9 /	0.60	
20.9	2.7	3.0	11 0	י. ד ד	0.65	
29.0	2.5	3.3	0 6	6 1	0.05	
20.9	2.2	5.4	9.0	0.1	0.63	
13.4	2.1	3.1	6.5	4.3	0.66	
7.5	1.9	3.0	3.9	2.5	0.64	
3.5	1.8	2.5	1.9	1.4	0.71	
PROPANOL-WATER	ENHANC	CED TUBE	MOLE FR 1.00 (PURE PROPANOL)		T <sub>sat</sub> - 96.8°C	
• q	ΔΤ	a				
98.5	4.3	22.8				
81.5	4.0	20.5				
66 2	3 5	18 9				
52 6	3 2	16 6				
40 3	2 9	14 1				
-0.5 20 R	2.5	11 2				
29.0	2.0	11.5				
20.7 12 c	2.2	7.3				
13.3	2.1	2.0				
1.0	2.0	J. Ö				
3.4	I.7	τ.δ				

Table C.3.1. (continued)

PROPANOL-WATER	SMOOT	TH TUBE	MOLE FR 0.00 (PURE WATER)	T <sub>sat</sub> - 100.6°C
ġ	ΔT	a		
216.9	14.9	14.5		
192.3	14.8	13.0		
143.7	13.7	10.5		
94.3	12.3	7.6		
69.9	11.1	6.3		
46.5	9.0	5.2		
27.6	6.4	4.3		
18.5	4.6	4.0		
9.5	2.7	3.6		

Table C.3.2. Propanol-water mixture - Smooth tube.

PROPANOL-WATER	SMOOTH TUBE		MOLE FR 0.06		T <b>=</b> 89.6°C	
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	α	Ratio	
216.7	15.2	24.1	14.2	9.0	0.63	
192.2	15.0	22.5	12.8	8.5	0.67	
143.5	13.9	19.7	10.3	7.3	0.70	
94.2	12.6	17.4	7.5	5.4	0.72	
69.9	11.4	15.7	6.1	4.4	0.72	
45.8	9.5	13.9	4.8	3.3	0.68	
27.5	7.1	9.8	3.9	2.8	0.72	
18.3	5.3	6.6	3.4	2.8	0 81	
9.3	3.1	3.7	3.0	2.5	0.84	

T<sub>sat</sub> - 88.5°C PROPANOL-WATER SMOOTH TUBE MOLE FR. = 0.15ġ ∆T<sub>id</sub> ΔT a<sub>id</sub> Ratio α 218.5 15.6 16.6 14.0 0.94 13.2 193.9 15.3 16.3 12.7 11.9 0.93 144.5 14.2 15.5 10.2 9.3 0.92 94.7 12.9 15.0 7.3 6.3 0.86 70.2 11.8 14.4 5.9 4.9 0.82 45.8 10.2 4.5 3.5 13.1 0.78 27.5 8.1 10.8 3.4 2.5 0.75 18.3 6.4 7.3 2.9 2.5 0.88 10.1 3.8 4.1 2.7 2.4 0.92

PROPANOL-WATER	SMOOTH TUBE		MOLE FR 0.30		T <sub>sat</sub> - 88.1°C	
ġ	∆T <sub>id</sub>	ΔΤ	<b>a</b> id	a	Ratio	
218.9	16.2	15.1	13.5	14.4	1.07	
194.2	15.7	14.8	12.3	13.1	1.06	
144.4	14.7	14.1	9.8	10.2	1.04	
95.2	13.5	13.5	7.1	7.1	1.00	
70.2	12.6	13.3	5.5	5.3	0.95	
45.9	11.5	12.8	4.0	3.6	0.90	
27.4	9.8	11.8	2.8	2.3	0.84	
18.3	8.2	9.8	2.2	1.9	0.83	
9.2	4.9	5.4	1.9	1.7	0.92	

PROPANOL-WATER	SMOO	TH TUBE	MOLE FR 0.43 (AZEOTROPE)	T <sub>sat</sub> - 87.7°C
ġ	ΔT	α		
218.3	16.8	13.0		
192.9	16.2	11.9		
143.6	15.1	9.5		
94.4	14.0	6.8		
69.6	13.3	5.2		
45.6	12.6	3.6		
27.7	11.3	2.4		
18.5	9.7	1.9		
9.3	5.9	1.6		

PROPANOL-WATER	SMOOTH TUBE		MOLE FR 0.55		T <sub>sat</sub> - 87.6°C	
• q	∆T <sub>id</sub>	ΔΤ	<b>a</b> id	α	Ratio	
217.6	17.1	18.2	12.7	12.0	0.94	
192.8	16.4	17.6	11.7	10.9	0.93	
143.8	15.3	16.1	9.4	8.9	0.95	
93.7	14.2	14.9	6.6	6.3	0.96	
69.6	13.6	14.2	5.1	4.9	0.96	
45.7	12.8	13.2	3.6	3.5	0.97	
27.2	11.5	11.8	2.3	2.3	0.97	
18.0	9.9	10.2	1.8	1.8	0.97	
9.1	6.2	6.7	1.4	1.3	0.93	

Table C.3.2. (continued)

PROPANOL-WATER	SMOOTH TUBE		MOLE FR 0.65		T <sub>sat</sub> - 88.45°C	
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio	
217.5	17.3	17.9	12.6	12.2	0.97	
192.9	16.6	17.4	11.6	11.1	0.96	
143.5	15.5	16.3	9.2	8.8	0.95	
94.0	14.4	15.1	6.5	6.2	0.96	
69.5	13.8	14.2	5.0	4.9	0.97	
45.3	13.0	13.1	3.5	3.4	0.99	
26.7	11.6	11.7	2.3	2.3	1.00	
17.9	10.1	9.9	1.8	1.8	1.02	
9.2	6.5	6.8	1.4	1.3	0.95	
PROPANOL-WATER	SMOOT	TH TUBE	MOLE FR	0.77	T <sub>sat</sub> - 89.6°C	
• q	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	α	Ratio	
217.6	17.5	19.2	12.4	11.3	0.91	
193.2	16.9	18.7	11.5	10.3	0.90	
143.9	15.8	17.4	9.1	8.3	0.91	
94.3	14.7	16.0	6.4	5.9	0.92	
69.8	14.0	15.0	5.0	4.6	0.93	
45.2	13.2	13.8	3.4	3.3	0.96	
27.0	11.8	12.3	2.3	2.2	0.96	
18.1	10.3	10.7	1.7	1.7	0.97	
8.9	6.8	7.1	1.3	1.3	0.96	
PROPANOL-WATER	SMOO	TH TUBE	MOLE FR	0.90	T = 92.9°C sat	
ģ	∆T <sub>id</sub>	ΔΤ	a <sub>id</sub>	a	Ratio	
218.1	17.8	19.4	12.3	11.3	0.92	
193.3	17.1	18.7	11.3	10.3	0.91	
143.5	16.0	17.7	8.9	8.1	0.91	
94.4	15.0	16.4	6.3	5.8	0.92	
69.5	14.3	15.7	4.9	. 4.4	0.91	
45.6	13.4	14.6	3.4	3.1	0.92	
27.3	12.0	13.1	2.3	2.1	0.91	
18.1	10.5	11.4	1.7	1.6	0.93	
9.2	7.1	7.4	1.3	1.2	0.96	

Table C.3.2. (continued)

Table C.3.2. (continued)

PROPANOL-WATER	SMOOTI	I TUBE	MOLE FR 1.00 (PURE PROPANOL)	T <sub>sat</sub> - 97.0°C
• P	ΔT	a		
217.7	18.0	12.1		
193.2	17.3	11.1		
143.3	16.2	8.8		
94.0	15.2	6.2		
69.3	14.5	4.8		
45.3	13.6	3.3		
26.9	12.1	2.2		
17.9	10.7	1.7		
9.0	7.4	1.2		

ETHANOL-BENZENE	SMOOTH 7	TUBE	MOLE FR 0.00 (PURE BENZENE)	T <sub>sat</sub> - 80.0°C
P	ΔT	a		
217.4	23.4	9.3		
193.2	21.9	8.8		
143.8	20.1	7.2		
94.3	18.5	5.1		
69.6	17.6	4.0		
45.7	16.6	2.8		
27.2	14.9	1.8		
18.2	13.4	1.4		
8.3	8.6	1.0		

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR	- 0.07	T <sub>sat</sub> = 71.7°C	
• q	∆T <sub>id</sub>	ΔT	<i>a</i> id	α	Ratio	
143.9	19.5	23.2	7.4	6.2	0.84	
93.9	18.0	20.9	5.2	4.5	0.86	
69.6	17.1	19.7	4.1	3.5	0.87	
45.5	16.2	18.3	2.8	2.5	0.88	
27.2	14.6	16.3	1.9	1.7	0.89	
18.2	13.1	14.6	1.4	1.2	0.90	
9.1	8.6	9.3	1.1	1.0	0.92	

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR	T <sub>sat</sub> = 69.5°C	
ġ	∆T <sub>id</sub>	ΔT	a id	a	Ratio
143.2	18.9	22.5	7.6	6.4	0.84
94.2	17.5	19.7	5.4	4.8	0.89
69.5	16.7	18.7	4.2	3.7	0.89
45.6	15.8	17.6	2.9	2.6	0.90
27.0	14.3	16.0	1.9	1.7	0.89
18.2	12.8	14.6	1.4	1.2	0.88
9.1	8.5	9.9	1.1	0.9	0.86

Table C.4.1. Ethanol-benzene mixture - Smooth tube.

ETHANOL-BENZEN	e smoo	TH TUBE	MOLE FR	T - 68.9°0 sat	
• P	∆T <sub>id</sub>	ΔT	<sup>a</sup> id	a	Ratio
144.1	18.1	19.3	8.0	7.5	0.93
94.7	16.7	17.5	5.7	5.4	0.95
70.0	15.9	16.6	4.4	4.2	0.96
45.7	15.2	15.8	3.0	2.9	0.96
27.3	13.8	14.4	2.0	1.9	0.96
18.3	12.3	12.7	1.5	1.4	0.97
9.3	8.4	8.5	1.1	1.1	0.99

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR	T <sub>sat</sub> - 68.5°C	
ġ	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
144.2	17.2	17.5	8.4	8.2	0.98
94.7	15.9	16.2	5.9	5.8	0.98
70.2	15.3	15.5	4.6	4.5	0.99
46.0	14.6	14.7	3.1	3.1	0.99
27.3	13.3	13.5	2.0	2.0	0.98
18.3	11.9	12.1	1.5	1.5	0.99
9.2	8.3	8.2	1.1	1.1	1.02

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR 0.45 (AZEOTROPE)	T <sub>sat</sub> - 68.5°C
ġ	ΔT	a		
144.2	16.4	8.8		
94.7	15.2	6.2		
70.0	14.6	4.8		
45.8	14.0	3.3		
27.5	12.9	2.1		
18.1	11.5	1.6		
9.3	8.2	1.1		

Table C.4.1. (continued)

Table C.4.1. (continued)

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR	T <sub>sat</sub> - 68.5°C	
• q	∆T <sub>id</sub>	ΔT	<b>a</b> id	α	Ratio
144.4	16.1	16.5	9.0	8.7	0.98
95.3	14.9	15.5	6.4	6.1	0.96
70.4	14.3	14.8	4.9	4.7	0.96
45.6	13.7	14.0	3.3	3.2	0.98
27.4	12.5	12.8	2.2	2.1	0.98
18.2	11.2	11.7	1.6	1.6	0.96
9.3	8.2	8.3	1.1	1.1	0.99

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR	T <sub>sat</sub> - 69.3°C	
ġ	∆T <sub>id</sub>	ΔT	° id	a	Ratio
144.2	15.6	16.7	9.2	8.6	0.93
94.3	14.4	15.4	6.5	6.1	0.93
69.6	13.8	14.8	5.0	4.7	0.93
45.7	13.1	13.7	3.5	3.3	0.95
27.3	12.0	12.3	2.3	2.2	0.97
18.2	10.7	11.1	1.7	1.6	0.96
9.2	8.2	8.3	1.1	1.1	0.99

ETHANOL-BENZENE	SMOOTH TUBE		MOLE FR.	- 0.80	T <sub>sat</sub> - 70.5°C	
ġ	∆T <sub>id</sub>	ΔΤ	a <sub>id</sub>	a	Ratio	
94.0	14.1	15.4	6.6	6.1	0.92	
69.8	13.5	14.7	5.2	4.7	0.92	
45.6	12.7	13.7	3.6	3.3	0.93	
27.3	11.6	12.2	2.3	2.2	0.95	
18.2	10.4	11.0	1.7	1.6	0.95	
9.2	8.2	8.9	1.1	1.0	0.92	

Table C.4.1. (continued)

ETHANOL-BENZENE	SMOOTH TUBE		$\mathbf{E} \mathbf{F} \mathbf{R} \cdot \mathbf{-} 0$	.90 T sat	T <sub>sat</sub> - 72.8°C	
ġ 4	AT <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio	
143.6 1	15.0	18.3	9.6	7.8	0.82	
94.4 1	L <b>3.8</b> 1	L6.0	6.8	5.9	0.86	
70.0 1	<b>13.1</b>	15.0	5.3	4.6	0.87	
45.6 1	12.3	4.2	3.7	3.2	0.87	
27.4 1	1.3 1	L <b>2.</b> 7	2.4	2.1	0.88	
18.2 1	10.1	1.6	1.8	1.6	0.87	
9.1	8.2	9.1	1.1	1.0	0.90	

ETHANOL-BENZENE	SMOOTH	TUBE	MOLE FR 1.00 (PURE ETHANOL)	T <sub>sat</sub> - 78.3°C
• q	ΔT	a		
217.2	16.2	13.4		
192.1	15.8	12.2		
143.5	14.7	9.8		
94.1	13.5	6.9		
69.9	12.8	5.4		
45.7	11.9	3.8		
27.2	10.9	2.5		
18.0	9.8	1.8		
9.3	8.1	1.1		

qΔTα238.015.715.2201.315.413.1168.415.011.2138.114.49.6110.613.98.083.113.26.364.412.65.144.311.83.730.310.32.9
$\begin{array}{cccccccccccccccccccccccccccccccccccc$
$\begin{array}{cccccccccccccccccccccccccccccccccccc$
168.4 $15.0$ $11.2$ $138.1$ $14.4$ $9.6$ $110.6$ $13.9$ $8.0$ $83.1$ $13.2$ $6.3$ $64.4$ $12.6$ $5.1$ $44.3$ $11.8$ $3.7$ $30.3$ $10.3$ $2.9$
$\begin{array}{cccccccccccccccccccccccccccccccccccc$
83.1       13.2       6.3         64.4       12.6       5.1         44.3       11.8       3.7         30.3       10.3       2.9
64.4       12.6       5.1         44.3       11.8       3.7         30.3       10.3       2.9
44.3       11.8       3.7         30.3       10.3       2.9
30.3 10.3 2.9
ETHANOL-BENZENE SMOOTH DISK MOLE FR. $-0.05$ T $-74.0^{\circ}$ C
sat sat
q ΔT <sub>id</sub> ΔT α <sub>id</sub> α Ratio
234.2 15.5 18.6 15.1 12.6 0.83
<b>199.0 15.2 17.5 13.1 11.4 0.87</b>
165.9 14.7 17.1 11.3 9.7 0.86
136.3 14.3 16.5 9.5 8.2 0.86
62 9 12 7 15 1 5 0 4 2 0 84
44.7         11.9         14.5         3.7         3.1         0.82
29.6 10.3 12.5 2.9 2.4 0.82
$\mathbf{v}$
EINANOL-BENZENE SHOOIN DISK HOLE FR. = 0.12 I sat
q ΔT <sub>id</sub> ΔT α <sub>id</sub> α Ratio
234.2 15.2 18.0 15.4 13.0 0.85
199.0 14.9 17.3 13.4 11.5 0.86
<b>166.1 14.5 16.7 11.5 9.9 0.87</b>
136.1         14.1         16.1         9.6         8.4         0.87
108.6 13.6 15.9 8.0 6.8 0.86
84.4         13.3         15.4         6.3         5.5         0.86           60.0         10.7         15.0         5.0         0.00         0.00
$\begin{array}{cccccccccccccccccccccccccccccccccccc$

2.8

29.3

10.3

13.0

2.3

0.80

Table C.4.2. Ethanol-benzene mixture - Smooth disk.

ETHANOL-BENZENE	SMOOTH	DISK	MOLE FR.	- 0.24	T <sub>sat</sub> = 69.0°C
ė	∆T <sub>id</sub>	ΔT	<b>a</b> id	a	Ratio
234.5	14.7	16.6	15.9	14.1	0.89
199.1	14.5	15.9	13.7	12.5	0.91
166.3	14.1	15.5	11.8	10.7	0.91
136.3	13.8	15.2	9.9	8.9	0.90
109.0	13.4	14.9	8.1	7.3	0.90
84.5	13.1	14.7	6.4	5.7	0.89
63.0	12.6	14.2	5.0	4.4	0.89
44.4	11.9	13.7	3.7	3.2	0.87
29.5	10.3	11.9	2.9	2.5	0.87
	CNOOTU	NTCK	MOLE FR -	- 0.35	T - 68.3°C
EINANOL-BENZENE	540011	DISK	Inter In.		sat
einanol-benzene q	ΔT <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
<b>q</b> 234.3	ΔT <sub>id</sub> 14.3	<b>ΔT</b> 14.7	<b>a</b> id 16.4	<b>a</b> 16.0	Ratio
<b>q</b> 234.3 198.1	ΔT <sub>id</sub> 14.3 14.0	<b>AT</b> 14.7 14.5	<b>a</b> id 16.4 14.1	<b>a</b> 16.0 13.6	<b>Ratio</b> 0.97 0.97
<b>q</b> 234.3 198.1 164.3	ΔT <sub>id</sub> 14.3 14.0 13.8	ΔT 14.7 14.5 14.2	<b>a</b> id 16.4 14.1 11.9	<b>م</b> 16.0 13.6 11.6	<b>Ratio</b> 0.97 0.97 0.97 0.97
<b>q</b> 234.3 198.1 164.3 133.6	AT id 14.3 14.0 13.8 13.4	ΔT 14.7 14.5 14.2 14.1	<b>a</b> id 16.4 14.1 11.9 9.9	<b>م</b> 16.0 13.6 11.6 9.5	<b>Ratio</b> 0.97 0.97 0.97 0.97 0.95
<b>q</b> 234.3 198.1 164.3 133.6 105.3	AT id 14.3 14.0 13.8 13.4 13.7	ΔT 14.7 14.5 14.2 14.1 13.9	<b>a</b> id 16.4 14.1 11.9 9.9 7.7	<b>م</b> 16.0 13.6 11.6 9.5 7.6	<b>Ratio</b> 0.97 0.97 0.97 0.97 0.95 0.98
<b>q</b> 234.3 198.1 164.3 133.6 105.3 80.5	AT id 14.3 14.0 13.8 13.4 13.7 13.0	ΔT 14.7 14.5 14.2 14.1 13.9 13.8	<b>a</b> id 16.4 14.1 11.9 9.9 7.7 6.2	a 16.0 13.6 11.6 9.5 7.6 5.9	<b>Ratio</b> 0.97 0.97 0.97 0.95 0.98 0.94
<b>q</b> 234.3 198.1 164.3 133.6 105.3 80.5 58.9	AT <sub>id</sub> 14.3 14.0 13.8 13.4 13.7 13.0 12.6	AT 14.7 14.5 14.2 14.1 13.9 13.8 13.4	<b>a</b> id 16.4 14.1 11.9 9.9 7.7 6.2 4.7	a 16.0 13.6 11.6 9.5 7.6 5.9 4.4	<b>Ratio</b> 0.97 0.97 0.97 0.95 0.98 0.94 0.94
<b>q</b> 234.3 198.1 164.3 133.6 105.3 80.5 58.9 41.4	AT <sub>id</sub> 14.3 14.0 13.8 13.4 13.7 13.0 12.6 11.5	AT 14.7 14.5 14.2 14.1 13.9 13.8 13.4 12.4	aid 16.4 14.1 11.9 9.9 7.7 6.2 4.7 3.6	a 16.0 13.6 11.6 9.5 7.6 5.9 4.4 3.3	Ratio 0.97 0.97 0.97 0.95 0.98 0.94 0.94 0.93
<b>q</b> 234.3 198.1 164.3 133.6 105.3 80.5 58.9 41.4 27.2	AT id 14.3 14.0 13.8 13.4 13.7 13.0 12.6 11.5 10.3	ΔT 14.7 14.5 14.2 14.1 13.9 13.8 13.4 12.4 10.1	<b>a</b> id 16.4 14.1 11.9 9.9 7.7 6.2 4.7 3.6 2.6	<b>a</b> 16.0 13.6 11.6 9.5 7.6 5.9 4.4 3.3 2.7	Ratio 0.97 0.97 0.97 0.95 0.98 0.94 0.94 0.93 1.02

 $T_{sat} = 68.3$ °C MOLE FR. -0.45SMOOTH DISK ETHANOL-BENZENE (AZEOTROPE) ġ ΔΤ α 238.8 14.0 17.0 199.1 13.8 14.5 168.7 13.6 12.4 135.5 13.3 10.2 108.4 13.1 8.3 85.7 12.9 6.6 63.6 12.6 5.1 44.4 11.9 3.7 29.8 10.4 2.9

Table C.4.2. (continued)

ETHANOL-BENZENE	SMOO	OTH DISK	MOLE FR	- 0.57	T <sub>sat</sub> - 68.3°C
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
236.9	13.6	14.5	17.5	16.4	0.94
200.1	13.2	13.9	15.2	14.4	0.95
167.3	13.1	13.5	12.8	12.4	0.97
137.4	12.9	13.2	10.7	10.4	0.98
110.2	12.5	12.8	8.8	8.6	0.98
85.7	12.2	12.3	7.0	7.0	0.99
64.2	11.9	11.9	5.4	5.4	1.00
45.6	11.4	11.3	4.0	4.0	1.01
30.3	10.0	10.0	3.0	3.0	1.00
ETHANOL-BENZENE	SMOO	OTH DISK	MOLE FR	= 0.70	$T_{sat} = 69.0^{\circ}C$
•					
P	∆T <sub>id</sub>	ΔT	<b>α</b> id	a	Ratio
234.5	12.9	14.1	18.1	16.7	0.92
198.8	12.6	13.6	15.7	14.6	0.93
165.0	12.3	13.6	13.4	12.1	0.91
134.1	12.4	12.9	10.8	10.4	0.96
107.4	11.8	12.4	9.1	8.6	0.95
82.5	11.6	11.9	7.1	6.9	0.97
60.8	11.2	11.5	5.4	5.3	0.97
43.7	10.7	10.8	4.1	4.0	0.99
28.6	9.6	9.6	3.0	3.0	1.00
ETHANOL-BENZENE	SMOO	OTH DISK	MOLE FR. =	- 0.80	T <sub>sat</sub> - 70.4°C
ġ	∆T <sub>id</sub>	ΔT	<i>a</i> id	α	Ratio
235 0	12 6	14 0	18 6	16.8	0 90
198 4	12.3	13 4	16 1	14 8	0.92
165 6	12.0	13 1	13 7	12 6	0 92
135 0	12.0	12.1	11 2	10 8	0.96
107 4	11 5	12.5	9 3	8 7	0 94
82 2	10 9	11 6	7 5	7 1	0 94
60 8	10.7	11 1	5 7	7.1 5 5	0 97
42 1	10 0	10 5	4 2	4 0	0.95
27.4	9.1	9.4	3.0	2.9	0.97

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Table C.4.2. (continued)

ETHANOL-BENZENE	SMOOTH	DISK	MOLE FR. =	0.90	T <sub>sat</sub> = 73.0°C
ġ	∆T <sub>id</sub>	ΔT	a <sub>id</sub>	a	Ratio
234.6	12.3	14.0	19.1	16.8	0.88
198.4	11.8	13.4	16.8	14.8	0.88
165.4	11.5	13.1	14.3	12.7	0.88
135.0	11.4	12.5	11.8	10.8	0.91
107.8	10.8	12.0	10.0	9.0	0.90
83.5	10.5	11.5	7.9	7.2	0.91
62.0	10.2	10.3	6.1	6.0	0.99
43.6	9.8	10.4	4.5	4.2	0.94
30.5	8.9	9.5	3.4	3.2	0.94
ETHANOL-BENZENE	SMOOTH	DISK	MOLE FR. ( PURE ETH	- 1.00 ANOL)	T <sub>sat</sub> = 78.3°C
ġ	ΔT	α			
240.1	11.9	20.2			
203.2	11.5	17.7			
167.9	11.2	15.0			
137.5	11.4	12.1			
110.5	10.5	10.5			
86.1	10.1	8.5			
64.7	9.8	6.6			
46.3	9.3	5.0			
30.6	8.7	3.5			

Table C.4.2. (continued)

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