STUDIES OF BOILING HEAT TRANSFER FROM PLAIN AND INTEGRAL-FIN TUBES TO BINARY AND TERNARY LIQUID MIXTURES

A THESIS

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CANDIDATE'S DECLARATION

I hereby certify that the work which is being presented in the thesis entitled "*STUDIES OF BOILING HEAT TRANSFER FROM PLAIN AND INTEGRAL-FIN TUBES TO BINARY AND TERNARY LIQUID MIXTURES*" in fulfilment of the requirements for the award of the degree of **DOCTOR OF PHILOSOPHY**, submitted in Department of Chemical Engineering of the University is an authentic of my own work carried out during a period from January, 1991 to June, 1994 under the supervision of Professor B.S.Varshney and Dr. B. Monanty.

The matter embodied in this thesis has not been submitted by me for the award of any other degree.

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ABSTRACT

The present investigation deals with experimental and theoretical studies related to nucleate pool boiling heat transfer of single component liquids and their binary and ternary mixtures on a single horizontal plain as well as integral-fin tubes at atmospheric pressure with a view to understand their boiling characteristics. The single component liquids investigated are acetone, isopropanol and water. As regards the binary liquid mixtures, they include aqueous mixtures namely; acetone-water and isopropanol-water and hydrocarbon mixtures of acetone-isopropanol, whereas ternary liquid mixtures consisting of acetone, isopropanol and water.

In fact, the experimental data have been obtained for boiling of 22 binary and 12 ternary compositions as well as their pure components on three horizontal heating tubes one by one. One of the tubes is of plain surface and the remaining ones are of integral-fin surfaces of 1 mm height trapezoidal shape, but of different fin densities, of 748 and 1024 fpm. Each heating tube has an axial hole of 18 mm diameter for its cartridge electric heater to supply heat for boiling of liquid pool. The heated length of each tube is 108.0 mm, and the outer/ envelope diameter of plain/integral-fin tube is 31.2 mm.

The experimental set-up has been carefully designed, fabricated and commissioned to ensure reliable and accurate experimental data. For this, surface area of the condenser, liquid pool height over the heating tube, placement of wall and liquid thermocouples, deaeration of test liquid, thermal stabilization of heating tube surface, homogeneity of heating surface, and various precautions during experimentation have been the main considerations. The experimental data were obtained for heat fluxes varying from 57,624 W/m^2 to 10,391 W/m^2, in decreasing order at atmospheric pressure.

While carrying out experiments with liquid mixtures, the composition of mixtures were determined before and after a given experimental run. It was found that two values were almost the same within $\pm$0.01 mole fraction.

Analysis of experimental data shows that heat transfer coefficients of binary and ternary liquid mixtures depend upon their compositions in addition to heat flux. The heat transfer coefficients for boiling of single component liquids and also of their binary and ternary liquid mixtures on plain tube get enhanced when the surface of tube is finned and the enhancement is more pronounced when fin density is 1024 fpm as compared to 748 fpm.

Further, the data obtained and reported as part of this investigation provide valuable and unique information about nucleate pool boiling heat transfer coefficients. It is revealed that addition of a component leads to degradation in boiling heat transfer coefficient of mixtures on plain as well as integral-fin tubes, having approximately the similar trend. This degradation, expressed as degradation factor, (h/h_{id}), happens to be a function of $|Y-X|$ for a given binary/ternary liquid mixture, irrespective of heating surface.

The heat transfer coefficient prediction accuracies of the existing correlations have been evaluated by employing the earlier and present experimental data of liquid mixtures boiling on plain tube. It is noted that correlation proposed by Schlunder (1982) provides the best predictions of heat transfer coefficient, if the Stephan-Abdelsalam correlation is employed to calculate ideal heat transfer coefficient, h_{id}. However, predictions due to Palen-Small (1964) and Thome-Shakir (1987) are also reasonably well.

A new correlation for the prediction of nucleate pool boiling heat transfer coefficients of binary mixtures on horizontal plain tube, based on the postulated mechanism of heat and mass diffusion to growing vapour bubbles on heating surface, has been derived. It is important to emphasize that this correlation is simple and convenient since it is free from such empirical constants, whose determination, as is noted in available correlations, generally involves a tedious exercise. Thus, this new correlation is unique. It is also important to underline that this correlation accounts for the variation in physico-thermal properties of the mixtures through the Stephan-Abdelsalam correlation which is employed to estimate ideal heat transfer coefficient, h_{id}. The influence of the phase equilibrium on mixture boiling due to preferential evaporation of the more volatile component is taken care of through the quantity known as 'boiling range'. An excellent agreement is noted between experimental and predicted heat transfer coefficients from the new correlation. For this, the experimental data of present and earlier investigations available in open literature have been employed. Further, it is also established that the derived correlation predicts better than the best of the existing correlations.

Another new correlation has also been derived for the boiling of mixtures on integral-fin tubes by incorporating surface correction factor. This correlation also correlates reasonably well the experimental data of the present investigation and those from open literature.

It is important to note that the derived correlations in the present investigation can be successfully extended to predict boiling heat transfer coefficient of ternary liquid mixtures with a good agreement between experimental and predicted values.

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(Abdul-Ameer S. Kadhum)

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NOMENCLATURE

A	heat transfer surface area, m^2
Az.	azeotrope composition
B_o	empirical scaling factor, Eq.(2-26)
B_L	proportionality constant in Levy equation, dimentionless, Table 2.4
Cp	liquid specific heat, J/kg K
C_{sf}	proportionality constant in Rohsenow equation,dimentionless, Table 2.4
D	diameter of heating tube, m
d	bubble departure diameter, m
F_c	mixture boiling correction factor
F_o	fin efficiency
F_s	surface correction factor at a given heat flux
f	bubble departure frequency, s^{-1}
f	fugacity
fpm	fins per meter
G	Gibbs free energy, J
G_m	mass velocity of liquid, defined for Gilmour correlation
g	gravitational acceleration, m/s^2
H	fin height, m
H	enthalpy, kJ/kg

h	average heat transfer coefficient, W/m² K
h_{fg}	differential latent heat of vaporization, J/kg
J	molar flux, moles/m² s
k	thermal conductivity, W/m K
L	effective heating tube length, m
MW	molecular weight, kg/kg mole
P	fin pitch, m
p	pressure, kPa
p_r	reduced pressure
Q	heat rate, W
q	heat flux, W/m²
q_r	reference heat flux, W/m²
R	gas constant
R	cavity site radius, m
R_p	surface roughness parameter, mm
r	bubble radius, m
T	temperature, °C, or K
ΔT	wall superheat, °C, or K
ΔT_{BR}	boiling range of mixture, K
t	time, s

t_t	tip fin thickness, m
t_b	base fin thickness, m
V	volume, m^3
v	specific volume, m^3/kg
X	liquid mole fraction
Y	vapour mole fraction
Z	compressibility factor
z	distance, m

Greek Letters

α	thermal diffusivity, m^2/s
β	contact angle, degree
β_l	mass transfer coefficient, m/s
δ	liquid mass diffusivity, m^2/s
ρ	density, kg/m^3
σ	surface tension, N/m
μ	dynamic viscosity, N s/m^2
Ω	binary interaction parameter
Φ	superficial volume fraction
Φ_G	proportionality constant in Gilmour equation,dimentionless, Table 2.4
ω	acentric factor

Subscripts

cr	critical
eff	effective
expl	experimental
F	integral-fin tube
g	bubble growth
i	inner
i, j,..n	component
id	ideal
l	liquid
o	outer
P	plain tube
ph	phase interface
pred	predicted
r	root
ref	reference
s	bulk liquid, sensible
s,t	bulk liquid at the top
s,s	bulk liquid at the side
s,b	bulk liquid at the bottom

sub	subcooling
t	tube
th	pitch circle diameter of wall thermocouples
v	vapour
W	waiting period
w	wall
wo	outer wall
wo,t	outer wall at the top
wo,s	outer wall at the side
wo,b	outer wall at the bottom
1,2	more-,and less- volatile components in binary mixtures, respectively
1,2,3	more-, intermediate- and less- volatile components in ternary mixtures, respectively

Dimensionless group

Nusselt number, $Nu = (hD/k)$, Table 2.2

Prandtl number, $Pr = (Cp\ \mu/k)$, Eq.(2-32)

Jakob number, $Ja = (\rho_l\ Cp\ /\Delta T)/(\rho_v\ h_{fg})$, Table 2.2

Modefied Jakob number, $Ja^* = (\rho_l\ Cp\ /\Delta T_{eff})/(\rho_v\ h_{fg})$, Table 2.2

Scriven number, $Sn = [1 - (y-x)\ (\alpha/\delta)^{0.5}\ (Cp/h_{fg})\ (dT/dx)]^{-1}$, Eq. (2-8)

CHAPTER 1

INTRODUCTION

Boiling is a physical process of wide industrial applications and thus has been the subject of intensive research for the past several decades. As it stands today, however, most of the research efforts have been confined to the boiling characteristics of single component liquids on horizontal plain tubes. At this stage, it is appropriate to emphasize that boiling of binary and ternary mixtures has engineering relevance in the design of reboilers and evaporators, which have wide application in the chemical, petrochemical, refrigeration, pharmaceutical, and other allied process industries.

The mechanism of boiling of liquid mixtures is significantly different from that of single component liquids. This has been explained by van Stralen and Cole (1979) and Shock (1982). In fact, the rate of vapour bubble growth in single component liquids is governed by the heat flow from the heating surface towards the vapour bubble boundary to satisfy the heat requirements of the growing bubbles. However, during boiling of liquid mixtures the heat diffusion is linked with the mass diffusion of the more volatile component of the mixture. Thus, this component gets exhausted rapidly in the liquid phase adjacent to the vapour bubble. Consequently, a low concentration of the more volatile component results thereby slowing down the bubble growth.

Boiling of binary mixtures on a tube as emphasized by Thome and Shock (1984), and Bajorek et al.(1989) is of complex nature and therefore has been the subject of study by a lesser number of investigators. As regards boiling of ternary mixtures, it has received still lesser attention despite the fact that the boiling of mixtures with more than two components is a common industrial practice. However, a few studies have been devoted to nucleate pool boiling of mixtures having three and more components on plain tube at atmospheric pressure. Grigor'ev et al.(1968b) have performed an experimental investigation for acetone-methanol-water, and acetone-ethanol-water mixtures whereas Stephan and Preusser (1979) have measured nucleate pool boiling heat transfer coefficients in acetone-methanol-water mixtures. Their data have not indicated the presence of a minimum heat transfer coefficient as the Grigor'ev data had shown. Recently, Bajorek et al.(1989) have also measured boiling heat transfer coefficient for acetone-methanol-water mixtures. They found a significantly low heat transfer coefficient of ternary mixtures than those estimated by a linear mixing law.

Several methods have been proposed to predict heat transfer coefficients for the boiling of binary and multicomponent mixtures on plain tube. However, none of these predictive methods is completely satisfactory for boiling of aqueous mixtures, as tested and reported by Thome and Shakir (1987). Evaluation of these methods for ternary and multicomponent mixtures has also been hampered due to the lack of suitable experimental data of such mixtures.

It is important to mention that Sardesai et al.(1986) have studied nucleate pool boiling heat transfer of hydrocarbon mixtures up to five components on plain tube. Such mixtures consisted of n-pentane, n-heptane, cyclohexane, p-xylene, and 1-tetradecene. Based on their experimental data, they have modified the Schlunder correlation (1982), employing several empirical coefficients. However, the values of these coefficients have not been reported. At the same time, the experimental data are also missing.

Thermal design of conventional reboilers is based on boiling heat transfer data on plain tubes, which are the typical heating surfaces. Such reboilers serve as a reference in order to appreciate the performance of reboilers, if their plain tubes are replaced by some enhanced tubes, such as integral-fin tubes. Hence, boiling heat transfer data on such tubes are important.

In fact, integral-fin tubes have potential applications in many areas of thermal engineering. One area of their applications is in the design of reboilers and evaporators dealing with boiling of binary and multicomponent liquid mixtures. Such reboilers are widely used in chemical, petrochemical, pharmaceutical and other similar industries. However, lack of thermal design data including heat transfer coefficients as a function of composition of mixture, heat flux, and tube surface geometry is an impediment in the design of reboilers and evaporators using such tubes. Also, the increased application of integral-fin tubes to practical design of reboilers and evaporators is dependent in part on having an accurate and relatively easily implementable correlation to predict mixture boiling heat transfer coefficient. However, review of the literature by Webb et al. (1989), and Kumar (1992) has amply established that correlations are nearly non-existent for liquid boiling on integral-fin tubes even for single component liquids, except the correlation proposed by Palen and Yang (1983). Even for this correlation the empirical coefficients required to calculate heat transfer coefficients have not been reported. Hence, this correlation suffers from the lack of detailed information necessary to perform the needed calculations of heat transfer coefficients. This calls for a research effort to fill the existing gaps with regard to suitable correlation in the area of nucleate pool boiling of binary, and multicomponent liquid mixtures on integral-fin tubes.

The understanding of nucleate pool boiling of binary and ternary mixtures is important in order to obtain a reliable predictive correlation of heat transfer coefficients for multicomponent mixtures. For successful application of integral-fin tubes in reboilers used in chemical process industry, the concerted research efforts related to the boiling of binary and ternary mixtures on these tubes is an important step.

Quite a wide variety of integral-fin surface geometries have been tested with single component liquids, such as boiling of water, alcohols, hydrocarbons, and refrigerants by Westwater 24 (1973), Yilmaz and Westwater 17 (1981), Gorenflo and Fath 25 (1987), Chen et al. 26 (1988), Hahne et al. 4 (1991), Webb and Pais 21 (1992), and Kumar 8 (1992). However, the boiling of liquid mixtures on such tubes has been reported recently only by Bajorek et al. 27 (1989).

Kadhim has studied

With the above-stated facts in view, the present investigation has been undertaken for nucleate pool boiling of single component liquids and their binary and ternary liquid mixtures on a single horizontal tube of either plain or integral-fin surface at atmospheric pressure with the following distinct objectives:

1. To establish generalized correlation for the prediction of nucleate pool boiling heat transfer coefficients of binary and ternary mixtures, based on the postulated mechanism of heat and mass diffusion into growing vapour bubbles on horizontal heating tubes

2. To obtain experimental data of boiling heat transfer from a single horizontal tube to the pool of single component liquids of widely differing physico-thermal properties

3. To determine the effect of mixture composition on heat transfer coefficient for the boiling of binary and ternary liquid mixtures on a single horizontal tube

4. To evaluate the accuracy of the existing correlations for prediction of heat transfer coefficient for the boiling of binary and ternary liquid mixtures on plain tube

5. To find the relative thermal performance of integral-fin tubes vis-a-vis plain tube with regard to the boiling of single component liquids and their binary and ternary mixtures of wide range of compositions and physico-thermal properties.

CHAPTER 2

LITERATURE REVIEW

Nucleate boiling is one of the few efficient modes of heat transfer of wide industrial importance, whereby liquid undergoes a change of phase as a result of bubble formation at preferred nucleation sites on a given heating surface. Boiling heat transfer is distinct from single-phase forced convection in that the boiling heat transfer coefficient is a strong function of the temperature difference between the heating surface and the bulk liquid. During boiling, a large amount of heat is transported from the heating surface as latent heat in vapour bubbles along with sensible heat to the liquid. Another differing characteristic of boiling heat transfer is that the micro-surface geometry of the heating surface affects the boiling process. Thus, boiling heat transfer is a quite complex process, and hence much more research effort is required to obtain a basic understanding of the involved phenomena, heat transfer mechanisms, and predictive methods.

The study of nucleate pool boiling of liquids began as early as in 1934, with the research work of Nukiyama (1934). He carried out experiments on heat transfer from an electrically heated platinum wire, submerged in the pool of distilled water. Using the experimental data, he obtained a curve between heat flux and wall superheat which is popularly termed as ' boiling curve '. Since then many investigators have studied the various facets of boiling heat transfer and thereby contributed to better understanding of the mechanism of boiling.

This Chapter describes the physical processes that are important in nucleate pool boiling of pure liquids and their mixtures over wide ranges of compositions. Accordingly, a review of available studies that are relevant to physical processes and objectives of the present investigation has been carried out in the following sections. The emphasis in this Chapter is on the physical mechanisms of nucleate pool boiling and also the predictive methods to estimate nucleate pool boiling heat transfer coefficients of pure liquids as well as binary and ternary liquid mixtures.

2.1 NUCLEATE POOL BOILING OF SINGLE COMPONENT LIQUIDS

This has been an area of research interest of many investigators for the last over six decades. As a result, several correlations have been recommended to calculate bubble nucleation sites, bubble growth rate, bubble departure diameter, and bubble emission frequency, along with the prediction methods for heat transfer coefficient. These aspects of boiling heat transfer have been briefed in the following sections.

2.1.1 BUBBLE NUCLEATION SITES

Of all the problems in boiling heat transfer, the proper characterization of the boiling surface in terms of the distribution of bubble nucleation sites is the most difficult task. At this juncture, it is important to point out that without this information it is simply not possible to predict, within reasonable accuracy the heat transfer coefficient corresponding to a given wall superheat in the nucleate boiling regime. The presence of nucleation sites at the heating surface and their shape and size distribution are responsible for the birth and growth of the vapour bubbles, this in turn, play a dominant role in almost all of the proposed mechanisms of heat transfer from the heating surface as emphasized by van Stralen and Cole (1979). It is worth-mentioning that most of the real heating surfaces possess machine-formed pits, scratches, grooves, etc. However, their sizes on heating surface vary from microscopic to the macroscopic.

Nucleation of vapour bubbles occurs on the preexisting gas or vapour phase in these surface cavities. Clark et al. (1959) and Cornwell (1977) are among the investigators, who have shown that vapour bubbles growing on a heated wall originate from small vapour nuclei trapped in pits and cracks of the wall having their sizes of the order of 0.1-10 μm in diameter. The criteria for activation of these vapour nuclei are useful for the prediction of the wall superheat required to initiate boiling. The expression for the equilibrium wall superheat between a vapour bubble and its surrounding liquid is obtained by considering a force balance on the bubble.

A force balance on a bubble for mechanical equilibrium leads to the Laplace equation as follows:

$$(p_v - p_s)\,\pi r^2 = 2\,\pi r\,\sigma \qquad (2\text{-}1)$$

From Eq.(2-1), it is clear that the vapour pressure inside the bubble, p_v is greater than that of the bulk. Hence, the saturation temperature inside the bubble must also be higher than that in the bulk. To satisfy this, the bulk liquid has to be uniformly superheated above its own saturation temperature. Using the Clausius-Clapeyron equation, one gets:

$$(dp/dT)_s = h_{fg}/[T_s(V_v - V_l)] \qquad (2\text{-}2)$$

This equation is integrated from (p_s,T_s) to (p_v,T_{wo}), and using the Kelvin equation

$$p_v/p_s = \exp\,(2\sigma V_v/rRT_s) \qquad (2\text{-}3)$$

one can relate the vapour pressure inside a curved interface, i.e., vapour bubble, p_v to that at a planar interface, p_s. The resulting expression for equilibrium wall superheat,($\Delta T = T_{wo} - T_s$) is obtained as follows:

$$\Delta T = 2\,\sigma/r(dp/dT)_s \qquad (2\text{-}4)$$

Bubble growth begins when the equilibrium superheat for a bubble with radius of curvature, r in a single component liquid is approximated by the following equation:

$$\Delta T = 2\,\sigma/(\rho_v\,h_{fg}r) \qquad (2\text{-}5)$$

At a heated surface, the superheat requirement depends on the effective radius of curvature of the vapour bubble.

Keeping in view the importance of nucleation sites, there has been due interest amongst the investigators to probe deeper and deeper into this aspect of study. As a result, much insight has been obtained. Table 2.1 describes some of the important investigations related to bubble nucleation sites for ready reference.

Table 2.1 Nucleation site correlations for boiling of single component liquids

Investigator(s)	**Correlation**	**Remark**
Nishikawa and Yamagata(1960)	$h \alpha n^{1/3}$	a. Water boiling on a brass tube b. $q < 13{,}600$ Btu/hr ft^2
Kurihara and Myers (1960)	$h \alpha n^{1/3}$	a. Water and organic liquids b. $q < 19{,}000$ Btu/hr ft^2
Gaertner (1965)	$h \alpha n^{2/3}$	a. Water boiling on copper surface b. $q < 58{,}000$ Btu/hr ft^2
Kirby and Westwater (1965)	$q = n^b$, $b = 0.33\text{-}0.5$	
Wiebe and Judd (1971)	For $nf < 55000$ bubble/(inch)^{2}s $q \alpha (n f)^{1/2} (T_{wo}\text{-}T_s)$ For $nf > 55000$ bubble/(inch)^{2}s $q \alpha (n f)^{1/3} (T_{wo}\text{-}T_s)$	Boiling of water
Hsu and Graham (1976)	$n = 0.012\, q^2\, p$	

2.1.2 BUBBLE GROWTH RATE

The bubble growth in a liquid has been classified into three stages, the initial hydrodynamic stage, the trasient stage, and the advanced asymptotic stage. During the first stage of bubble growth, excess pressure in the bubble expands the vapour bubble with a constant radial velocity, and the liquid inertia dominates. As regards the asymptotic stage of bubble growth in a pure liquid, liquid evaporates at the vapour-liquid interface and the necessary heat is supplied by the superheated liquid layer surrounding the vapour bubble. However, the bubble growth is controlled by diffusion process and the rate of bubble growth gradually decreases. The asymptotic stage has the longest duration of the three stages, and it is during this growth phase that the bubble departure takes place. The transient stage takes place between the hydrodynamic and asymptotic stages. During this stage, bubble growth is governed by a combination of liquid inertia and thermal diffusion.

2.1.3 BUBBLE DEPARTURE DIAMETER

After nucleation of vapour bubbles on preferential sites whose radius of curvature is equal to or greater than that obtained from Eq.(2-4), the bubbles begin to grow till they attain a size for which the buoyancy force just exceeds the surface tension force. Then, the bubbles detach from their nucleation sites. The bubble diameter at the time of departure has been studied by many investigators. Some of the important correlations for this have been summarized in Table 2. 2.

Table 2.2 Bubble departure diameter correlations for boiling of single component liquids

Investigator(s)	**Correlation**
Fritz and Ende (1935)	$d = 0.0208\ \beta\ [g\ \sigma/(\rho_l-\rho_v)]^{0.5}$
Staniszewski (1959)	$d = 0.0071[2g\sigma/(\rho_l-\rho_v)][1+0.435(\bar{\partial}d/dt)]$
Zuber (1959)	$d = [12\ \{r\ g\ \sigma/(\rho_l-\rho_v)\}]^{1/3}$
Cole (1963)	$d = 4 \times 10^{-2}\ Ja\ [\ \sigma/(\rho_l-\rho_v)]$
Cole and Shulman (1966)	$d = [133.3/p]\ [\ \sigma/(\rho_l-\rho_v)]$
Cole and Rohsenow (1969)	$d = C\ (Ja^*)^{5/4}\ [\ \sigma/(\rho_l-\rho_v)]^{0.5}$ $C = 1.5 \times 10^{-4}$ for water $C = 4.65 \times 10^{-4}$ for other liquids

2.1.4 BUBBLE EMISSION FREQUENCY

It is important to determine the emission frequency as it influences the heat transported by the bubbles originating on the sites of heating surface. The prediction correlation of bubble emission frequencies at a heating surface have been recommended by several investigators as shown in Table 2.3.

Table 2.3 Bubble emission frequency correlations for boiling of single component liquids

Investigator(s)	**Correlation**	**Remark**
Jincina et al.(1950)	$f\ d = C_{sf}\ [g\rho_l\ \sigma/(\rho_v)]^{0.5}$	
Zuber (1963)	$f\ d = 0.59\ [g\ \sigma\ (\rho_l-\rho_v)/(\rho_l)^2]^{0.25}$	
Cole (1963)	$f\ V_b\ \alpha\ [\ \sigma^{3/5}/g\ (\rho_l-\rho_v)^{3/4}\ (\rho_l)^{2/3}]^{3/4}[Ja]^2$	
Hatton and Hall (1966)	$f\ d^2 = [3/\pi\alpha\ ]\ [16\ k_l\ \sigma\ T_s/h_{fg}{}^2\ (\rho_l)^2\ d]^2$	
Ivey (1967)	$f\ d^{0.5} = 0.90\ q^{0.5}$	for hydrodynamic region
	$f\ d^{0.75} = 0.44\ q^{0.5}$	for transition region
	$f\ d^2 = \text{Constant}$	for thermodynamic region
Wiebe and Judd (1971)	$q\ \alpha\ (nf)^{0.5}\ \Delta T$	for $nf < 55 \times 10^3$ bubbles/(inches)^2 s
	$q\ \alpha\ (nf)^{1/3}\ \Delta T$	for $nf > 55 \times 10^3$ bubbles/(inches)^2 s

2.1.5 CORRELATIONS FOR SINGLE COMPONENT LIQUIDS BOILING ON PLAIN TUBE

Attempts have been made by various investigators to correlate nucleate pool boiling heat transfer data. Most of the proposed correlations were developed by analyzing a simplified model of boiling leading to some dimensionless groups. Table 2.4 summarizes some of the important correlations for the prediction of nucleate boiling heat transfer coefficient of single component liquids on plain tube.

Table 2.4 Correlations of boiling heat transfer coefficient of single component liquids on plain tube

Investigator(s)	Correlation
Rohsenow (1952)	$(Cp\,\Delta T/h_{fg}) = C_{sf}\ [(q/\mu\, h_{fg})\{g_o\,\sigma/g(\rho_l-\rho_v)\}^{0.5}]^{0.33}\ (Cp\,\mu/k_l)$
McNelly (1953)	$(hD_o/k_l) = 0.225\ [q\,D_o/h_{fg}\,\mu]^{0.69}[p\,D_o/\sigma]^{0.31}[(\rho_l-\rho_v)/\rho_v]^{0.33}\ [\mu Cp/k_l]^{0.69}$
Gilmour (1958)	$(h/Cp\,G_m)(Cp\mu/k_l)^{0.6}(\sigma\,\rho_l/p^2)^{0.425} = \Phi/(D\,G_m/\mu)^{0.3}$
Levy (1959)	$q = (1/B_L)[k_l\,Cp\,\rho_l^2/\{\sigma\,T_s(\rho_l-\rho_v)\}]\ (\Delta T)^3$
Borishanskiy & Minchenko(1963)	$(h/k_l)\ [\sigma/(\rho_l-\rho_v)]^{0.5} = 8.7\times10^{-4}[(q/\rho_v\,h_{fg}\alpha)\{\sigma/g(\rho_l-\rho_v)\}^{0.5}]^{0.7}[p/\sigma\{\sigma/g(\rho_l-\rho_v)\}^{0.5}]^{0.7}$
Kutateladze (1963)	$(h/k_l)[g_o\sigma/g(\rho_l-\rho_v)]^{0.5} = 7.0\times10^{-4}[(q/\rho_v\,h_{fg}\alpha')\ \{\sigma/g(\rho_l-\rho_v)\}^{0.5}]^{0.7}$ $\times\ [p/\sigma\{g_o\sigma/g(\rho_l-\rho_v)\}^{0.5}]^{0.7}\ (Cp\mu/k_l)^{-0.35}$
Alam and Varshney (1973a)	$(h/k_ld) = 0.084\times[(q/\rho_v\,h_{fg}\alpha)\ \{\sigma/(\rho_l-\rho_v)\}^{0.5}]^{0.6}\ [1+(\rho_l/\rho_v)(\Delta T_{sub}/T_s)]^{-0.5}$ $\times[(\rho_v h_{fg})^2/\{CpT_s\,\rho_l\,\{\sigma\,(\rho_l-\rho_v)^{0.5}\}]^{0.37}$
Stephan&Abdelsalam (1980)	**For water:** $h = 0.246\times10^7(k_l/d)\ [(qd/k_lT_s)]^{0.673}\ (CpT_sd^2/\alpha^2)^{1.26}\ (h_{fg}d^2/\alpha^2)^{-1.58}$ $\times\ [(\rho_l-\rho_v)/\rho_l]^{5.22}$ **For hydrocarbons:** $h = 0.0546\ (k_l/d)\ [(qd/k_lT_s)\ (\rho_v/\rho_l)^{0.5}]^{0.67}\ ((\rho_l-\rho_v)/\rho_l)^{-4.33}\ (h_{fg}d^2/\alpha^2)^{0.248}$ **For refrigerants:** $h = 207\times(k_l/d)\ (q\,d/k_lT_s)^{0.745}\ (\rho_v/\rho_l)^{0.581}(\mu/\rho_l)^{0.53}$
Cooper (1984)	**For water:** $h = 95\ (q^{0.67})[(p/p_{cr})^{0.12-2\log Rp}](-\log(p/p_{cr}))^{-0.55}\ MW^{-0.5}$ **For other liquids:** $h = 55\ (q^{0.67})[(p/p_{cr})^{0.12-.2\log Rp}](-\log(p/p_{cr}))^{-0.55}MW^{-0.5}$

REMARK ON EXISTING CORRELATIONS

A large number of correlations are available in the literature. These have been mentioned in several references, e.g., Palen et al.(1972), Sharma (1977), and Mehrotra (1980). However, they have been found to give unsatisfactory results over the wide range of physical properties which is encountered in by process industries.

The comprehensive studies have been reported by Stephan and Abdelsalam (1980) and Cooper (1984), who have developed new correlations. These correlations are empirical and are based on a large data bank of about 5000 points collected from 72 research papers for the boiling of a number of liquids having differing physico-thermal properties. Stephan and Abdelsalam have obtained separate correlations for different fluid categories as is noted from Table 2.4.

Cooper's correlation is for all liquids. The Cooper correlation contains a ' surface roughness parameter ', R_p which has a value in the range of 0.3 to 0.9. Cooper has not recommended the value of R_p to be used for a particular surface type.

In recent years, extensive research efforts have been devoted to the mechanisms of nucleate pool boiling. However, it is still difficult to predict boiling heat transfer coefficients with satisfactory accuracy as underlined by Hinrichs et al.(1981), Thome (1990), Jamialahmadi et al.(1991). This is partly because the related experiments are difficult, and due to this, the results show wide scatter, which, in turn, may reflects the insufficient understanding of this crucial physical phenomena.

2.2 NUCLEATE POOL BOILING OF BINARY LIQUID MIXTURES

As it stands today, many aspects of nucleate pool boiling heat transfer with single component liquids are well-explored and reasonably good insight towards their understanding these has been developed. However, boiling heat transfer of liquid mixtures is more complicated. Due in part to this complexity, the heat transfer problems in liquid mixture boiling have received less attention than those in single component liquids. Most of the available research related to mixture boiling is limited to binary mixtures, since these are the easiest to study. Therefore, the approach to the subject is to understand first binary mixture boiling and then to extend this knowledge to multicomponent mixture boiling. The purpose of the following survey is to review the significant advances in the understanding of the mixture boiling process.

2.2.1 BUBBLE NUCLEATION SITES

It is well established that in nucleate pool boiling the effect of the micro-structure of the heating surface is an important factor in determining the heat transfer rate from the surface. This is due to the fact that the vapour bubbles first originate as minute vapour nuclei trapped in naturally available pits and cracks in the heating surface. There exists a large number of papers on bubble nucleation sites regarding the boiling of single component liquid. Table 2.1 briefs them.

For binary liquid mixtures, the only published experimental results are those due to van Stralen and Cole (1979) for boiling on very thin wires, Eddington and Kenning (1979), and Hui and Thome (1985) for boiling on brass disk in vertical orientation.

The tests of van Stralen and Cole have covered several aqueous mixtures. Only one study included data for each of the constituent single component liquids to permit the comparison with the one mixture composition tested, 4.1 wt% methyl ethyl ketone (MEK) in water. The test section in that study was 0.2 mm diameter wire. At a heat flux of 0.3 MW/m², it was observed that the number of active boiling sites cm^{-2} in pure water was 30 and in MEK over 200, but for the 4.1 wt% MEK mixture only one site was active. As substantial effect of composition was evident.

Eddington and Kenning (1979) have investigated the effect of contact angle on the bubble nucleation density in ethanol-water mixtures using a gas diffusion method on two metallic surfaces of copper and brass. The results show that the nucleation density rises as the contact angle increases, and decreases with increasing ethanol composition.

Hui and Thome (1985) have conducted an experimental study to measure bubble nucleation densities and heat transfer coefficients. They have employed binary liquid mixtures of ethanol-water and ethanol-benzene on a heated vertical brass disk at 1.01 bar .They have reported a strong effect of composition on the nucleation density. This is attributed to the nature of the activation of the heating surface and mass diffusion. They have also reported that the heat transfer coefficient is quite insensitive to the very large increases in nucleation site densities as compared to the pure water and the ethanol-water azeotrope mixture results at the same heat flux.

2.2.2 BUBBLE GROWTH RATES

The understanding of the mechanics of bubble growth is one of the keys to the eventual understanding of the boiling process. Consequently, concerted effort has been made to study the vapour bubble growth rates in binary liquid mixtures for the furtherance of the fundamental understanding of multicomponent boiling.

The growth of a vapour bubble in a binary mixture is more complicated compared to that in pure liquid. This is due to the difference in the compositions of vapour and liquid phases. For a mixture, excluding azeotrope mixture, the composition of the more volatile component in the vapour phase Y_1 is greater than that in the bulk liquid phase X_1 surrounding the superheated layer as shown in Figure 2.1. Thus, as the evaporation process proceeds, and the bubble grows, the more volatile component in the liquid layer adjacent to the bubble diffuses to the interface to provide the additional more volatile component in the vapour and simultaneously produces a local composition gradient around the bubble. As the local value of X decreases from X_1 to $X_{1,ph}$, the bubble point at the bubble interface rises from T_1 to T_{ph}, as is clear from the above figure. Eventually, the bubble point at the interface reaches its maximum value when the rate of diffusion of the more volatile component to the interface matches with the rate of its excess evaporation at the interface, which is proportional to IY-XI.This was first explained by van Wijk et al.(1956).

The models for the growth of vapour bubbles in binary liquid mixtures fall into two categories, namely; those for bubbles growing homogeneously in a superheated liquid, and those growing heterogeneously at a heated wall.

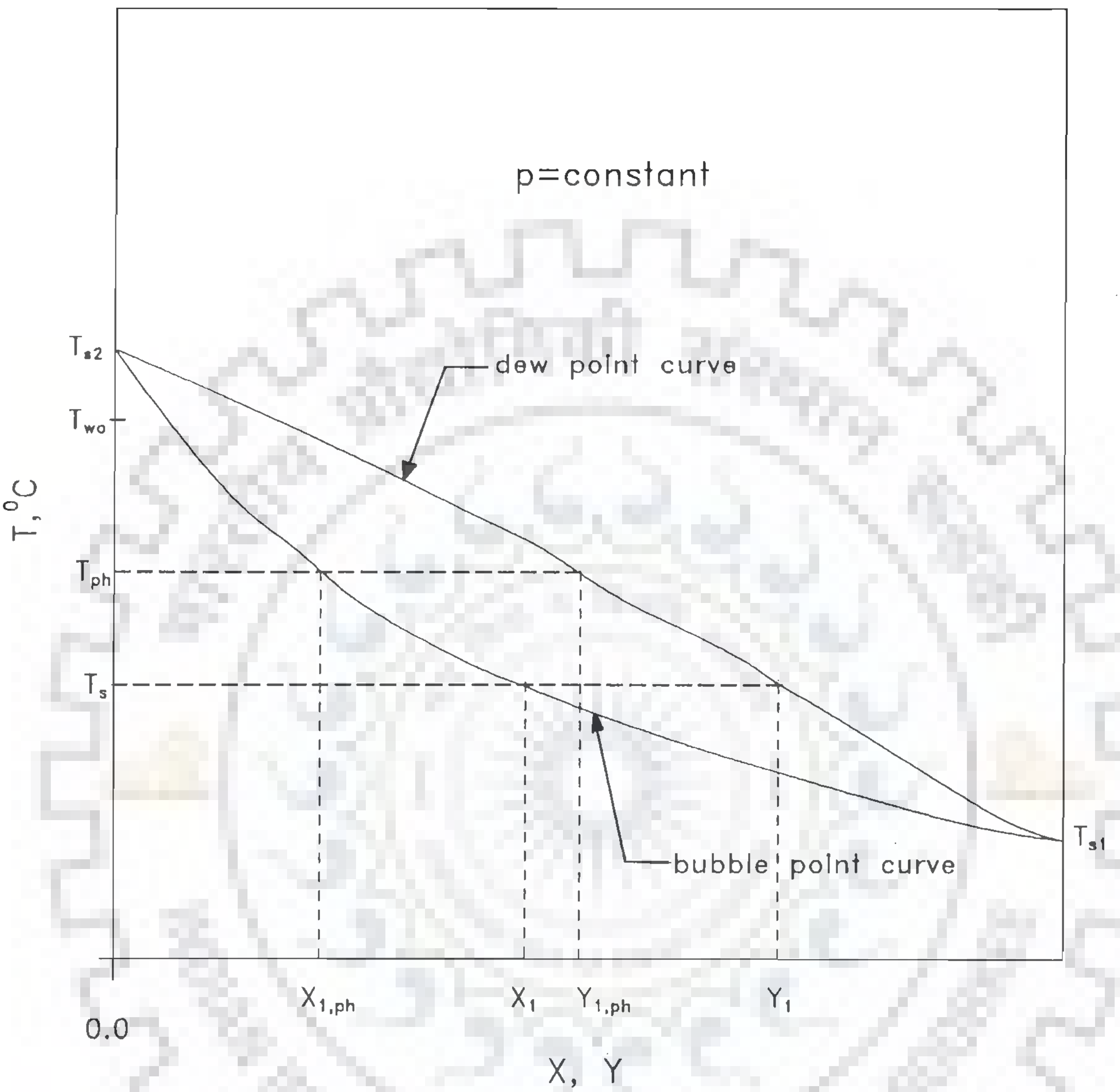


Fig. 2.1 Phase equilibrium diagram for binary mixture

The majority of previous work has been concerned with the asymptotic bubble growth period. Bruijin(1960) deduced the theoretical growth rate for a bubble in a binary liquid mixture with a spherically symmetrical and initially uniform superheat. He extended the study to the case where neighbouring bubbles exhibited mutual interference and showed that this could still further reduce the bubble growth rate. Scriven(1959) performed a similar study for the same initial conditions and accounted for the changes in temperature and concentration fields around the growing bubble that are caused by the radial liquid motion due to the difference in phase densities. Skinner and Bankoff (1964a) analyzed the growth of bubbles in pure fluids in spherically symmetric but nonuniform temperature fields. They (1964b) extended this work to the case of binary liquid mixtures considering a nonuniform concentration field of arbitrary nature. Since all these studies assume a spherically symmetrical system, they are more truly relevant to bubble growth far from a heating surface.

The development of bubble growth rate equation has been given by Scriven(1959). Although it is less general than that by Skinner and Bankoff (1964a) and (1964b), it gives a deeper insight into the physical picture and has been found to be satisfactory in describing the behaviour of real bubbles in several experimental studies.

Scriven considered heat and mass transfer by one-dimensional radial conduction and convection to the bubble interface and perfect mixing in the vapour phase. Starting from the energy and mass balances on the two components, he arrived at a set of equations governing bubble growth that required numerical solution. To make the analysis of more practical importance, he then derived two simplified expressions governing bubble growth, one for small superheats and the other for large superheats. For nucleate pool boiling superheats of practical interest, the expression related to large superheat is more valuable and is as follows:

$$R = \frac{[\Delta T\ (12\ k_l\ t)^{0.5}]}{(\pi)^{0.5}(\rho_v/\rho_l)(h_{fg}/Cp)[1-(y-x)(\alpha/\delta)^{0.5}(Cp/h_{fg})(dT/dx)]} \qquad (2\text{-}6)$$

or $$R = R_{pure}\ Sn \qquad (2\text{-}7)$$

where $$Sn = [1 - (y-x)\ (\alpha/\delta)^{0.5}\ (Cp/h_{fg})\ (dT/dx)]^{-1} \qquad (2\text{-}8)$$

where (dT/dx) is the slope of the bubble point line and the bracketed term is simplification of the original equation due to Calus and Rice (1972). This equation reduces to the Plesset and Zwick (1954) solution, that is R_{pure}, for single components when (y-x) equals zero.

van Stralen (1959) extended the bubble growth model of Plesset and Zwick for a spherical bubble growing remote from a wall in an initially uniformly superheated single component liquid to binary liquid mixtures. He obtained an expression for bubble growth using this approach that was identical to Scriven's results, given by Eq.(2-6).

Cooper and Stone (1981) have studied bubble growth rates in binary liquid mixtures. A single bubble has been grown on a wall in uniformly superheated liquid mixtures of hexane and octane. They have found that the rate of bubble growth of the mixtures conform closely to that of pure liquids, provided the bubble interfacial temperature is used in evaluating the wall superheat rather than the bulk liquid boiling point.

All of the bubble growth models for mixtures that have been discussed in this section have been found to include Scriven number, Sn. It accounts for the effects of heat and mass diffusion on bubble growth rate. Since the terms, (y-x) and (dT/dx) in Eq.(2-6) always have opposite signs, $Sn < 1$. Thus, the bubble growth rate in liquid mixtures is predicted to be less than that in an equivalent pure liquid. This theoretical conclusion has been confirmed experimentally for bubble growth on a heated surface as well as for remotely growing bubbles by Benjamin and Westwater (1961), Florschuetz, and Thome and Davey (1981). Agreement with the theoretical models however, has been only qualitative. For example, Thome and Davey have observed the exponent in $r = a\,t^n$ to be a function of composition rather than being a fixed value of (1/2).

2.2.3 BUBBLE DEPARTURE DIAMETER

Perhaps, the first study is due to Tolubinskiy and Ostrovskiy (1966a) and (1969). They have reported bubble departure diameters for the binary systems ethanol-water, methanol-water. ethanol-butanol, and ethanol-benzene over the wide composition range at atmospheric pressure. Tolubinskiy et al.(1970) later have also studied water-glycerine mixtures. Their data showed that a minima in the departure diameter was observed for each mixture system corresponding to the maximum in the value of IY-XI except for water-glycerine. The conflicting result for water-glycerine has been attributed to the large decrease in the surface tension over the 200 °C rise in the saturation temperature from pure water to pure glycerine.

Isshiki and Nikai (1973) have also measured bubble departure diameters for the ethanol-water system. They have found a minimum in the bubble departure diameter at $X_1 = 0.02$ and a maximum at $X_1 = 0.5$, strikingly different results from those of Tolubinskiy and Ostrovskiy.

Thome and Davey (1981) have carried out similar experiments for nitrogen-argon mixtures at 1.3 bar, combining it with the study on bubble growth rates. Both manual and computerized image analysis methods have been used to analyze the high speed cine films which allowed a great number of consecutive bubble growth cycles at each boiling site. Thus, a statistically average bubble diameter have been determined. A minimum in the departure diameter and a maximum in the departure frequency have been found by them.

In order to investigate the physical explanation for the smaller bubble departure diameter in the binary mixtures, Thome (1981) rederivded the Keshock and Siegel (1964) bubble departure equation to include the effect of a volatile component on bubble growth.

For inertia controlled growth, the expression obtained is as follows:

$$(d/d_{id}) = Sn^{4/5} \qquad (2\text{-}9)$$

since Sn < 1.0 for mixtures, Eq.(2-9) predicts a minimum in departure diameter when Sn is at a minimum, which, as a matter of fact, is near the composition at which |Y-X| is maximum.

For surface tension controlled departure, Thome has obtained an equation for the ratio (d/d_{id}) as follows:

$$(d/d_{id}) = Sn\,[\sigma \sin\beta/(\sigma_{id} \sin\beta_{id})]^2 \qquad (2\text{-}10)$$

Equations (2-8) and (2-10) for inertia and surface tension controlled departure correlate all the data of the Tolubinskiy and Ostrovskiy, but not all of the data due to Isshiki and Nikai.

2.2.4 BUBBLE EMISSION FREQUENCY

The frequency of bubble departure f is defined as follows:

$$f = 1/(t_g + t_w) \qquad (2\text{-}11)$$

where t_g is the time period when the bubble grows from its nucleation size to its departure diameter and t_w is the waiting time period during which the vapour nucleus left behind waits to be reactivated. The effect of composition on the bubble growth time period and the waiting time period determines the overall variation in the bubble emission frequency.

Tolubinskiy and Ostrovskiy(1966a) have measured frequencies for ethanol-water and ethanol-butanol mixtures experimentally. For ethanol-water, there was a maximum value at 30% ethanol composition. But the ethanol-butanol data exhibited relatively no change.

Thome and Davey(1981) have also measured frequency for nitrogen-argon liquid mixtures and have found that the bubble departure frequencies increased for mixtures relative to the single component liquids.

The length of the bubble growth time t_g in the mixtures is affected by two factors. First, the bubbles do not have to grow as larger as in the single components in order to departure. Second, their rate of growth is slower, as has been shown earlier. The variation in the ratio of the growth times for inertia controlled growth can be shown to be

$$[t_g/(t_g)_{id}] = Sn^{2/5} \qquad (2\text{-}12)$$

Thus the growth time is predicted to decrease at the more volatile compositions. Hence, the bubble growth time in the mixture will be shorter than for the equivalent ideal mixture.

For surface tension controlled growth the ratio $[t_g/(t_g)_{id}]$ is obtained as

$$[t_g/(t_g)_{id}] = Sn[\sigma \sin\beta/(\sigma_{id} \sin\beta_{id})]^{\theta} \qquad (2\text{-}13)$$

Here, there is more effect of Sn than in Eq.(2-12) and the potential of large effects by the surface tension and contact angle.

The bubble waiting time, t_w is the time interval required for the thermal boundary layer to get stripped from the surface at the departure of the previous bubble. The bubble waiting time is therefore not only a function of the variation in the nucleation superheat with composition, but also of the thermal diffusivity of the liquid mixture and the effective wall superheat, ΔT_{eff}.

van Stralen and Cole (1979) have considered the effect of surface tension on bubble departure frequency in binary mixtures. They have concluded that the ratio of the frequencies is given as:

$$(f/f_{id}) = [(t_g)_{id} / t_g] [\sigma_{id} / \sigma] \qquad (2\text{-}14)$$

where the surface tension becomes involved via the bubble nucleation criterion. Equation (2-12) shows that $[(t_g)_{id}/t_g]$ will be greater than 1.0. The surface tension ratio (σ_{id}/σ) depends on the particular mixture system but in general is also greater than 1.0. Thus, boiling departure frequency in a binary mixture is predicted to be higher than that for its equivalent pure liquid. Experimental results on aqueous, organic, and cryogenic mixtures tend to support this fact.

2.3 SURFACE STRUCTURE OF HEATING TUBE

Boiling heat transfer coefficient is known to depend strongly on the surface characteristics of a heating tube. Heat transfer increases when the tube surface is not smooth. Many methods have been proposed to enhance nucleate pool boiling heat transfer, which involve a change in the heating surface structure. These methods have been described in details by Webb (1981), and Bergles (1988). One of the methods involves the provision of low-finned heating tube. Low-finned tubes are widely used in industrial shell-and-tube and other types of reboilers to enhance heat transfer coefficient of boiling liquids. Low-finned tubes generally have fin heights less than 6.35 mm. The fins are transverse, solid annular, and spaced at equal intervals along the tube axis.

Rabas and Taborek (1986) have discussed the applications and types of low-finned tubes. The majority of commercial low-finned tubes are of integral type with tube diameter under 38.1 mm, fin thickness 0.2 to 0.7 mm ,and fin density 350 to 1200'fpm.

There has been a great deal of research related to boiling of single component liquids on integral-fin tube as reported by Westwater(1973), Hahne and Muller(1983), Gorenflo and Fath (1987), Chen et al.(1988), and recently by Hahne et al.(1991) and Kumar (1992). However, investigations related to the boiling of liquid mixtures on integral-fin tubes are few inspite of the fact that this is of industrial significance. It seems that there is only one investigation due to Bajorek et al. (1989) which deals with boiling of liquid mixtures on finned tube. They have concluded that provision of fins on plain tube enhances boiling heat transfer coefficient of mixtures

2.4 NUCLEATE POOL BOILING OF MULTICOMPONENT LIQUID MIXTURES

Nearly all the studies of mixture boiling deal with nucleate pool boiling of binary liquid mixtures. The study of multicomponent mixtures boiling is much more tedious because of the large number of experiments required to cover the composition range of all the components. However , boiling of multicomponent liquid mixtures is encountered much more frequently than boiling of binary liquid mixtures in industrial practice. It seems there are three studies in the published literature for boiling of ternary liquid mixtures on plain tube. These are described as follows:

Grigor'ev et al.(1968b) have reported the boiling data for ternary liquid mixtures of acetone-methanol-water and acetone-ethanol-water on 7.7 mm diameter horizontal stainless steel tube at atmospheric pressure. Their results show a minimum heat transfer coefficient for a ternary composition. This minimum value for ternary composition is less than any of the three binary minima.

Stephan and Preusser (1979) have studied the boiling of acetone-methanol-water mixtures.Their data do not indicate the presence of minimum heat transfer coefficient as has been obtained by Grigor'ev et al.(1968b). The data are contradictory to those of Grigor'ev et al. in that the reduction in heat transfer coefficients for ternary mixtures has been less than that for the binary mixtures. To explain their finding, they have argued that the local rise in the saturation temperature has been smaller in the ternary liquid mixtures because of flattening of the bubble point curve on addition of a third component. Hence the ternary liquid mixtures perhaps experience less degradation in the effective wall superheat than the corresponding binary mixture systems.

Recently, Bajorek et al.(1989) have studied nucleate pool boiling heat transfer of acetone-methanol-water mixtures on copper plain heating tube at atmospheric pressure. Their data have revealed two binary minima for acetone-water and methanol-water, respectively. However, they have not mentioned a minima from the data of ternary liquid mixtures. They have also concluded that the effect of addition of the third component on the heat transfer coefficient has been found to depend on the value of IY-XI of the third liquid component.

As regards the investigations of boiling of liquid mixtures involving more than three components, it is important to mention that Sardesai et al.(1986) have studied nucleate pool boiling heat transfer of hydrocarbon mixtures up to five components (n-pentane, n-heptane, cyclohexane, p-xylene, and 1-tetradecene) on plain tube. They have modified the Schlunder correlation (1982) employing several empirical coefficients. However, the values of the correlating coefficients have not been reported. They have also not published their experimental data.

2.5 CORRELATIONS FOR BINARY/ MULTICOMPONENT LIQUID MIXTURES BOILING ON PLAIN TUBE

Several correlations for boiling heat transfer from a horizontal single plain tube to binary liquid mixtures have been published in recent years. These have been reviewed by Sardesai et al.(1982), and Thome and Shock (1984). Most of these correlations are of the general form as follows:

$$(h/h_{id}) = \varphi(X_i, Y_i, T, p, q^n, \delta, \ldots.) \qquad (2\text{-}15)$$

where $\varphi(X_i, Y_i, T, p, q^n, \delta, \ldots.)$ is a function, representing the thermodynamic variables. This function, indeed, accounts for the degradation in heat transfer coefficient due to mass transfer process. The ideal heat transfer coefficient, h_{id} is defined as the heat transfer coefficient required to support the same heat flux in a single component liquid with the physico-thermal of the mixture in question. The value of h_{id} is evaluated by one of the two methods. The most common method is based on linear mixing law as follows:

$$(1/h_{id}) = \sum_{i=1}^{i=n} (X_i/h_{id}) \qquad (2\text{-}16)$$

where h_i represents the heat transfer coefficient of the pure component liquid at the same heat flux as the mixture. However, for azeotropic mixtures, Eq.(2-16) should be modified, as suggested by Thome (1989). For compositions of mixtures lying to the left and the right of the azeotrope composition, the respective wall superheat is given by the following equation:

$$\Delta T_{id} = (X_1/X_2)\Delta T_{s\,Az} + [(X_{1\,Az}-X_1)/X_{1Az}]\Delta T_2 \qquad (2\text{-}17)$$

$$\Delta T_{id} = [(X_1-X_{Az})/(1-X_{Az})]\Delta T_1 + [(1-X_1)/(1-X_{Az})]\Delta T_{Az} \qquad (2\text{-}18)$$

After calculating ΔT_{id} from the relevant equation, the ideal heat transfer coefficient, h_{id} is determined as follows:

$$h_{id} = q/\Delta T_{id} \qquad (2\text{-}19)$$

Another method for calculation of h_{id} of a liquid mixture makes use of correlations meant for boiling of pure liquids but with properties of the boiling liquid mixture as reported by Palen and Small (1964), Sardesai et al.(1982), and Thome and Shakir (1987). Thus, the non-linear variation of mixture properties with the composition is taken into account. In this context, it is worth-mentioning that several investigators including Sardesai et al. (1986), Palen et al. (1986), Thome and Shakir (1987), and Kadhum et al. (1993) have shown that correlation for the boiling of pure liquid due to Stephan and Abdelsalam (1980) can be employed to calculate the heat transfer coefficients of pure hydrocarbons and azeotropes with minimum error. The same are reproduced below for ready reference.

$$h_{id} = 0.0546\,(k_l/d)[(qd/k_lT_s)(\rho_v/\rho_l)^{0.5}]^{0.67}\{(\rho_l-\rho_v)/\rho_l\}^{-4.33}(h_{fg}\,d^2/\alpha^2)^{0.248} \qquad (2\text{-}20)$$

where d is bubble departure diameter and is calculated as below:

$$d = 0.0146\,\beta\,[2\sigma/\{g(\rho_l-\rho_v)\}]^{0.5} \qquad (2\text{-}21)$$

Here β is contact angle and is assumed to be 45^0 for water and 35^0 for all organic compounds and mixtures. As a result of concerted research efforts, several correlations have been recommended. Some of the more popular correlations for predicting heat transfer coefficients of boiling mixtures are presented in the following subsections.

2.5.1 PALEN AND SMALL CORRELATION

One of the earliest practical correlations for boiling of multicomponent liquid mixtures on a plain tube bundle in reboilers has been published by Palen and Small (1964), which is as follows:

$$h = h_{id} \exp(-0.027\Delta T_{BR}) \tag{2-22}$$

where ΔT_{BR} represents the boiling range. It is the difference between the dew point and the bubble point temperatures for the liquid phase mixture composition. Palen and Small have recommended the McNelly (1953) correlation using average properties to evaluate ideal heat transfer coefficient, h_{id}. This correlation has been derived by plotting (h/h_{id}) against boiling range, ΔT_{BR} for the boiling of mixtures. Thome (1983) has tested this correlation for the data of Bonilla and Perry (1941), Cichelli and Bonilla(1945), Cichelli and Bonilla (1946), Grigor'ev et al.(1968a), Tolubinskiy and Ostrovskiy(1969), Valent and Afgan(1973) and Shakir(1987) for the boiling of ethanol-water mixtures at atmospheric pressure. This correlation predicts the results within an error of ±30%.

2.5.2 CALUS AND LEONIDOPOULOS CORRELATION

Calus and Leonidopoulos (1974) have obtained a correlation for predicting the variation in the wall superheat with composition. Their correlation is devoid of empirical coefficients. They have calculated the rise in the local saturation temperature for the growth of a single spherical bubble in an infinite uniformly superheated binary liquid mixture using Scriven's and van Stralen's solutions and have assumed that this situation was equivalent to that for an actual boiling surface. Their correlation is finally of the following functional form:

$$\Delta T = (x_1\,\Delta T_1 + x_2\,\Delta T_2)[1 + (y-x)\,(\alpha/\delta)^{0.5}(Cp/h_{fg})\,(dT/dx)] \tag{2-23}$$

Calus and Leonidopoulos have tested their correlation with their experimental data for boiling of n-propanol-water mixtures at atmospheric pressure and have reported that about 87% per cent of data points compare with the predictions within ±16%.

2.5.3 STEPHAN AND PREUSSER CORRELATION

Stephan and Preusser (1979) have used non-linear regression analysis to derive correlation for liquid mixture boiling. An additional term $[|\Sigma(Y-X)(dY_1/dX_1)|]^{-0.0733}$ has been included to account for the effect of mass diffusion as follows:

$$\begin{aligned} h = {} & [0.0871(k_l/d)(qd/k_lT_s)^{0.674}(\rho_v/\rho_l)^{0.156}(h_{fg}\,d^2/\alpha^2)^{0.371}(\alpha^2/\sigma d)^{0.35} \\ & \times\ (\mu Cp/k_l)^{-0.162}][1+|\Sigma(Y-X)(dY_1/dX_1)|]^{-0.0733} \end{aligned} \tag{2-24}$$

where d is bubble departure diameter and is calculated from Eq.(2-21).

Thome (1983) has tested the Stephan-Preusser correlation, Eq.(2-24), for the data due to Bonilla and Perry (1941), Cichelli and Bonilla(1945), Cichelli and Bonilla(1946), Grigor'ev et al.(1968a), Tolubinskiy and Ostrovskiy(1969), Valent and Afgan(1973) and Shakir(1987) for the boiling of ethanol-water mixtures at atmospheric pressure. He has concluded that the correlation overpredicts the values by about 40%. However, the data of Stephan-Preusser compare with their correlation reasonably well.

2.5.4 THOME CORRELATION

Thome (1981) has derived the following analytical expression based on the effect of composition on the thermal boundary layer stripping mechanism for mixture boiling on a heating surface:

$$(h/h_{id}) = Sn^{7/5} \qquad (2\text{-}25)$$

where Sn is the Scriven number, Eq.(2-8).

Thome has recommended linear mixing law for calculating ideal heat transfer coefficient, h_{id}.

2.5.5 SCHLUNDER CORRELATION

Schlunder (1982), considering the mass transfer process in the liquid surrounding the bubble and using the film theory of mass transfer, has derived the following correlation:

$$(h/h_{id}) = [1 + (h_{id}/q)\,(T_{s2}-T_{s1})\,(Y_1-X_1)\,\{1-\exp(-B_o q/\rho_l\,\beta_l\,h_{fg})\}]^{-1} \qquad (2\text{-}26)$$

His theoretically-derived correlation includes the effect of heat flux on mass transfer process. In Eq.(2-26) β_l is mass transfer coefficient in the liquid phase (assumed to be a constant equal to 0.0002 m/sec by Schlunder), and B_o is an empirical scaling parameter accounts for the fraction of heat consumed for the bubble formation and has been set equal to 1.0 by Schlunder. Schlunder has recommended linear mixing law for calculating ideal heat transfer coefficient.

When mixture boiling form azeotrope, Eq.(2-26) is modified as follows:

(i) For a mixture composition to the left of azeotrope composition of a binary mixture, the following correlation is to be employed:

$$(h/h_{id}) = [1+(h_{id}/q)\{(T_{s2}-T_{sAz})/X_{1Az}\}(Y_1-X_1)\,\{1-\exp(-B_o q/\rho_l\,\beta_l\,h_{fg})\}]^{-1} \qquad (2\text{-}27)$$

(ii) For a mixture composition to the right of azeotrope composition of a binary mixture, the following correlation is to be employed:

$$(h/h_{id}) = [1+(h_{id}/q)\,\{(T_{s1}-T_{sAz})/(1-X_{1Az})\}\,(Y_1-X_1)\,\{1-\exp(-B_o q/\rho_l\,\beta_l\,h_{fg})\}]^{-1} \qquad (2\text{-}28)$$

The Schlunder correlation has been tested by Uhlig and Thome (1985) for their data for the boiling of acetone-water mixtures. 95% of the data are predicted within ±20% by the correlation.

Shakir et al.(1985) have also tested Schlunder correlation for boiling of methanol-water mixtures. The values predicted by the Schlunder correlation are within ±35% for 90% of the data.They have changed the value of empirical scaling factor, B_o in the Schlunder correlation to 2.0. The maximum deviation have reduced to within ±20% and the predicted values are in good agreement with the experimental data.

Gorenflo et al.(1988) have compared experimental data for the boiling of R-22 and R-114 mixtures with the Thome correlation (1981) and Schlunder correlation (1982). The comparison shows a good agreement with the correlated method proposed by Schlunder. Schlunder(1986) has also extended his correlation, Eq.(2-26) to multicomponent mixtures. For boiling of multicomponent mixture of n components, this equation has the form:

$$(h/h_{id}) = [1 + (h_{id}/q) \{ \sum_{i=1}^{i=n-1} (T_{sn}-T_{si}) (Y_i-X_i)\} \{1-\exp(-B_o q/ \rho_l \beta_l h_{fg})\}]^{-1} \quad (2\text{-}29)$$

As regards ideal heat transfer coefficient, h_{id} it is calculated using linear mixing law, Eq.(2-16).

2.5.6 THOME AND SHAKIR CORRELATION

Thome and Shakir (1987) have modified the Schlunder correlation, Eq.(2-26) by using boiling range to provide a better approximation of the slope of bubble point curve. This slope has been approximated as $(dT/dX) = \Delta T_{BR} / (X_{1,ph}-Y_{1,ph})$ where ΔT_{BR} is boiling range and $(X_{1,ph}-Y_{1,ph})$ is the difference between the vapour/liquid equilibrium compositions at the bubble interface. Finally, they have obtained the following expression:

$$(h/h_{id}) = [1 + \{(h_{id}/q)\Delta T_{BR}\} \{1-\exp(-B_o q/ \rho_l \beta_l h_{fg})\}]^{-1} \quad (2\text{-}30)$$

The Stephan and Abdelsalam correlation (1980) has been used to calculate the value of h_{id}. Thome and Shakir have reported good agreement between their correlation and experimental data of four binary aqueous mixtures, i.e., acetone-water, methanol-water, ethanol-water and n-propanol-water with mean absolute error of 17.3%.

2.6 CORRELATION FOR MIXTURE BOILING ON FINNED TUBE

Palen and Yang(1983) have recommended the following correlation for the prediction of nucleate pool boiling heat transfer coefficient of liquid mixtures on horizontal finned tubes:

$$h = F_e\, F_s\, F_c\, h_{nb} + h_{nc} \quad (2\text{-}31)$$

In this expression, the factor F_e is fin efficiency, F_c is a mixture boiling correction factor, h_{nb}, and h_{nc} are nucleate pool boiling and natural convection heat transfer coefficients of mixture on plain tube, respectively, and F_s is a surface correction factor that models the effect of the close proximity of the fins on the boiling process. According to Yilmaz and Palen (1984), the value of this factor varies from tube to tube depending on the type of the fin geometry, fluid, and heat flux.

Palen et al. (1986) have suggested a method to determine F_s by using the following equation:

$$Fs = C\ (q/q_r)^{m1}\ (Pr)^{m2}\ (F_c)^{m3} \qquad (2\text{-}32)$$

where C is an empirical constant, m1, m2 are negative exponents, and m3 is a positive exponent. Values for these quantities have not been reported in the available literature.

The surface correction factor, F_s does not include fluid properties and consequently is insensitive to their effects on the boiling process between adjacent fins, which would probably become most evident at higher reduced pressures and under vacuum conditions as reported by Thome (1990). However, the above method provides a good framework for further improvements.

2.7 MOTIVATION FOR THE PRESENT INVESTIGATION

Boiling of binary/multicomponent liquid mixtures has wide industrial applications in reboilers and evaporators used in chemical, petro-chemical and many allied process industries. However, research related to this area has been limited. Therefore, it demands more research efforts to probe into many aspects required for the design of reboilers.

As regards the published experimental data for the boiling of binary liquid mixtures on a horizontal single plain tube at atmospheric pressure, they are not sufficient enough to derive conclusions. Some of the important ones are given in Table 2.5. As such, more data are called for.

Table 2.5 Summary of some of the experimental investigations for boiling of binary mixtures

Investigator(s)	Liquid mixture	Composition X_1	Heat flux, q W/m^2	Heating tube
van Wijk et al.(1956)	Water-Methyl ethyl ketone	0.01-0.95	——	Platinum
	Acetone-Water	0.01-0.90	——	
	Water-1-Butanol 0.01-0.95	——		
	Water-Ethanol	0.01-0.75	——	
Grigor'ev et al.(1968a)	Ethanol-Benzene 0.1-0.9	58,000-232,000	Stainless steel	
	Ethanol-Water	0.1-0.9	58,000-232,000	
Alam (1972)	Acetone-Water	0.04-0.7	8,000-46,480	Stainless steel
	Water-Acetic acid	0.05-0.7	8,000-46,480	
	Ethylene Glycol-Water	0.05-0.75	8,000-46,480	
	Glycerine-Water		8,000-46,480	

Table 2.5 contd...

Table 2.5 contd...

Isshiki and Nikai (1973)	Ethanol-Water	0.03-0.5	———	Nickel
	n-Butanol-Water	0.002-0.1	———	
	Glycol-Water	0.02-0.9	———	
Calus and Leonidopoulos (1974)	n-Propanol-Water	0.029-0.8	80,000-400,000	Nickel-Aluminum
Happel and Stephan(1974)	Benzene-Toluene	0.22-0.82	———	Nickel
	Ethanol-Benzene	0.1-0.9	———	
Stephan-Preusser(1979)	Methanol-Water	0.05-0.95	50,000-200,000	Nickel
	Acetone-Water	0.05-0.90	50,000-200,000	
	Ethanol-Water	0.04-0.95	50,000-200,000	
Jungnickel et al.(1980)	(R-12)-(R-113)	0.1-0.9	4,000-100,000	Copper
	(R-22)-(R-12)	0.1-0.9	4,000-100,000	
Uhlig and Thome (1985)	Acetone-Water	0.025-0.85	6,600-190,000	Copper
Shakir et al.(1985)	Methanol-Water	0.05-0.85	9,300-193,000	Copper
Shakir and Thome (1986)	Ethanol-Water	0.05-0.85	———	Brass
	Ethanol-Benzene	0.05-0.9	———	
Pandey (1982)	Ethanol-Water	0.05-0.50	9,500-30,534	Stainless steel
	Methanol-Water	0.05-0.50	9,500-30,534	
	Isopropanol-Water	0.05-0.50	9,500-30,534	
Gorenflo et al.(1988)	(R-22)-(R-114)	0.05-0.95	———	Copper
Bajorek et al.(1989)	Methanol-Water	0.05-0.8	14,000-300,000	Copper
	Acetone-Water	0.1-0.8	14,000-300,000	
	Ethanol-Water	0.1-0.3	14,000-300,000	
Alpay and Balkan(1989)	Acetone-Ethanol	0.25-0.75	10,000-40,000	Stainless steel
	Methylene Chloride-Ethanol	0.25-0.75	10,000-40,000	
Tarrad and Burnside(1991)	Pentane-Tetradecene	0.5	10,000-50,000	90 Cu-10 Ni

A review of Table 2.5 reveals that most of the experimental data have been obtained for high heat fluxes. At this juncture, it is important to emphasize that experimental data at low heat fluxes (10,000 W/m²< q < 60,000 W/m²) are of primary importance in chemical process industries. Therefore, it is necessary that data should be generated at low heat fluxes for their use in the design of reboilers.

A further review of literature shows that the research related to boiling of ternary liquid mixtures on a horizontal single plain tube at atmospheric pressure is quite limited. Table 2.6 gives the parametric details of these investigations.

Table 2.6 Summary of the experimental investigations for boiling of ternary mixtures

Investigator(s)	Liquid mixture	Composition	Heat flux, q W/m²	Heating Tube
Grigor'ev et al. (1968b)	Acetone-Methanol-Water Acetone-Ethanol-Water	———	58,000-233,000	Stainless steel D_o = 7.72 mm
Stephan & Preusser (1979)	Acetone-Methanol-Water	———	50,000-200,000	Nickel D_o = 14 mm
Bajorek et al.(1989)	Acetone-Methanol-Water	———	20,000-200,000	Copper D_o = 22.2 mm

It is clear from this table that only three research papers dealing with boiling of ternary liquid mixtures have been published. Thus, these are not sufficient to provide any authentic conclusions with regard to boiling heat transfer of ternary mixtures. It is therefore, necessary that data for more ternary liquid mixtures should be obtained. As can be seen in Table 2.6, the respective investigators have not reported the compositions of the ternary liquid mixtures employed by them. Hence, their data do not throw further insight into the issue, as far as the effect of composition on boiling heat transfer is concerned. Another point which merits mention is that the ternary systems selected in these investigations are of special nature. Any two of the components of these mixtures do not form azeotropes. However, ternary azeotropic systems are frequently used in industrial situations. Therefore, it is worthwhile, if data for ternary liquid mixtures, whose two liquid components form azeotropes, are generated.

For the convenience of the design engineer, correlations have been proposed from time to time for the prediction of heat transfer coefficients of binary and/or multicomponent mixtures.

Table 2.7 summarizes the important correlations which are generally employed for the design calculations.

Table 2.7 Summary of the important correlations for boiling of binary/multicomponent mixtures on a plain tube

Investigator(s)	**Correlation**	**Applicable to Boiling of Mixtures**
Palen and Small (1964)	Eq.(2-22)	Binary & Multicomponent
Calus&Leonidopoulos (1974)	Eq.(2-23)	Binary
Stephan-Preusser (1979)	Eq.(2-24)	Binary & Multicomponent
Thome (1981)	Eq.(2-25)	Binary
Schlunder (1982)	Eq.(2-29)	Binary & Multicomponent
Thome and Shakir (1987)	Eq.(2-30)	Binary & Multicomponent

It is important to mention that the several investigators have claimed superiority of their correlations over others. Hence, it is necessary to scrutinize all these correlations in order to choose the one which provides the best predictions.

It is also noted from literature that boiling of mixtures on integral-fin tubes has not been studied to the extent necessary to obtain a generalized correlation confidently for its use in the design of reboilers and evaporators. In this regard, perhaps the only published literature is due to Bajorek et al.(1989) for the boiling of binary mixtures, i.e., ethanol-water, acetone-water, and methanol-water on copper low-finned tube having fin density of 750 fpm. However, no experimental data for the boiling of ternary liquid mixtures on integral-fin tube are available. But such data are needed for the design of reboilers employing integral-fin tubes.

From the above description it is quite clear that there are definite gaps in the area of liquid mixture boiling which need to be filled by further research investigations for the furtherance of knowledge. Due to this, the present investigation has been motivated and has been accordingly planned.

CHAPTER 3

EXPERIMENTAL SET-UP

This Chapter describes the experimental set-up used in the present investigation related to boiling heat transfer from a horizontal single tube to pool of liquid mixtures at atmospheric pressure.

3.1 DESIGN CONSIDERATIONS

In order to obtain accurate and reliable experimental data, the following considerations have been taken into account in the design, fabrication and commissioning of the experimental set-up.

3.1.1 TEST VESSEL

It is used for holding liquid in which a heating tube is submerged horizontally. Shape of the vessel is an important consideration. A cylindrical shape is considered to be the best, as any other shape is likely to develop mechanical failure more easily when such vessel is subjected to pressure or vacuum.

The height of a test vessel should be enough to ensure a sufficient liquid pool over the submerged heating tube. Consequently, tube surface remains undisturbed by the flow of down-coming mass of condensate from a condenser placed at the top of the vessel, and therefore the vapour bubble dynamics on the heating tube is not disturbed, otherwise boiling data are likely to be inaccurate. Hence, an appropriate height of the test vessel is to be kept.

3.1.2 HEATING TUBE

The heating tube dimensions should be such that a necessary heat flux to cause nucleate pool boiling is possible. At the same time, the diameter should be closer to that of the tubes which are employed in the design of industrial reboilers. In addition, it is important to place a given heating tube in horizontal position in a liquid pool in the vessel.

There are several ways for achieving this. The one, which has been considered to be the most suitable, is to make a hole in the vessel wall and weld a socket over it. It is also necessary to ensure no leakage between the heating tube and the socket. To ensure this, a suitable number of threads have been provided with socket alongwith a suitable packing.

3.1.3 CONDENSER

Function of the condenser is to condense all the vapours coming up from the pool of boiling liquid. At the same time, there should be no hold-up of the condensate in it. Therefore, the condenser should be oriented vertically and mounted at the top of the test vessel. Also, surface area of the condenser should be adequate enough to condense vapours at the same rate at which they are generated, otherwise the following problems are likely to arise :(i) appreciable decrease in the liquid level above the heating tube surface, (ii) variation in the composition of liquid mixture forming the boiling pool, and (iii) fluctuation in the system pressure.

3.1.4 LIQUID POOL HEIGHT

As already discussed in section 3.1.1, there should be sufficient liquid pool height above the top of a given heating tube. From earlier investigations due to Alam (1972) and Wall and Park (1978), it has been established that a liquid pool height of 100 mm is adequate enough to ensure undisturbed vapour bubble dynamics on a heating tube. This height indeed must be maintained during the experimentation.

3.1.5 WALL- AND LIQUID- TEMPERATURES

Present investigation is an attempt to determine experimental values of nucleate pool boiling heat transfer coefficients from horizontal tube to liquid mixtures. As such, values of temperatures of outer wall of the heating tube and the bulk liquid around it are required, in addition to the heat flux. Hence, accurate measurement of temperatures at these positions is of paramount importance.

From the physics of vapour bubble generation on a horizontal tube, it is obvious that wall temperature varies circumferentially. Therefore, a knowledge of this circumferential temperature distribution of heating tube is essential. The greater is the number of wall temperature measuring points on a tube, the more is the accuracy in the determination of average wall temperature. For measuring wall temperature, thermocouple beads should be fixed on the outer surface of the heating tube. However, it is not desirable to install them as above, or else they would become nucleation sites for the bubbles. This, in turn, would affect the boiling phenomenon at these points. Therefore, the wall thermocouples should be placed in axial holes drilled in the wall thickness of heating tube. But there are practical difficulties in drilling holes at more than four points when tube outer diameter is small as used in the present investigation. Accordingly, only four longitudinal holes have been considered reasonable at an interval of $(\pi/2)$ radian.

It has been thought equally desirable to measure circumferential temperature at several cross-sections over entire length of heating tube. Therefore, it seems appropriate that longitudinal thermocouple holes should pass through entire length of the tube. However, when drill bit penetrates deeper and deeper, to more and more lateral movement occurs. Accordingly, the value of pitch circle diameter is likely to change from cross-section to cross-section over tube length. Therefore, penetration length of thermocouple holes should be such to avoid above phenomenon. Keeping in view the symmetry of a heating tube about its mid cross-section, it has been considered sufficient to drill the holes only upto one-half the tube length.

As pointed out earlier, an accurate temperature measurement of bulk liquid surrounding the heating tube is essential. It is a known fact that during boiling of liquid, superheated liquid layer appears between heating tube and bulk liquid. To obtain conservative values of boiling heat transfer coefficients, it is necessary that liquid temperature should be measured in its bulk. Accordingly, thermocouples should be placed in the bulk liquid which, of course, is outside superheated liquid layer.

3.2 DESCRIPTION OF EXPERIMENTAL SET-UP

Realizing the significance of the above-mentioned design considerations, the components of the experimental set-up have been designed, fabricated, and finally the set-up has been commissioned as shown schematically in Figure 3.1 and photographically in Figure 3.2. It essentially consists of test vessel (1), heating tube (2), electric heater (3), condenser(8), vacuum pump (12), and measuring instruments. These are described in the following sections.

3.2.1 TEST VESSEL

A 3 mm thick 304 ASIS stainless steel sheet, rolled in the form of a hollow cylinder of 150 mm inside diameter and 450 mm height, having a flanged top cover and a dished bottom with a valve V_2 to drain out the test liquid as and when required represents the test vessel.

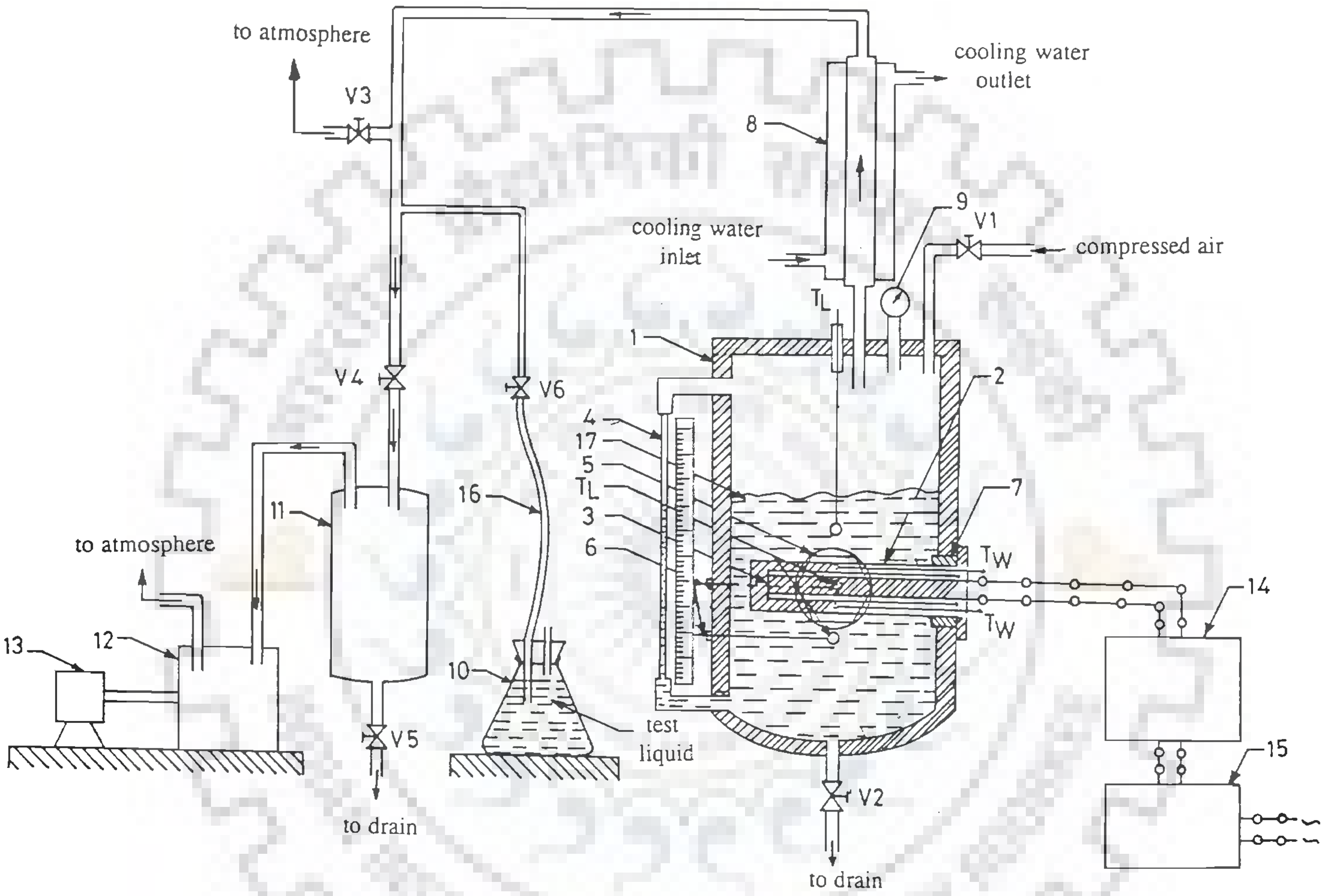
Various fittings, provided at the top cover of the vessel, are for condenser(8),vacuum/pressure gauge(9), and liquid thermocouple(T_L). A liquid level indicator (4) is attached to the side of the vessel. Socket (7) is for holding the heating tube(2) in horizontal position. To facilitate a visual observation of the boiling process on heating tube surface, two diametrically opposite view-ports (5) are there at the front and rear side of the test vessel. Liquid thermocouple probes enter the bulk liquid through the holes provided in wall thickness of the vessel corresponding to the top-, the sides-,and the bottom- positions on the heating tube. These probes are tightened on vessel wall in their positions with respective gland and nut arrangement (6). The vessel is thermally insulated to minimize heat losses to surroundings by covering it with asbestos rope, followed by a thick layer of 85% magnesia powder and finally a layer of glass wool.

3.2.2 HEATING TUBES

Experimental data reported in this investigation have been obtained using three heating tubes one by one. One of the heating tubes is a plain tube, while the others are integral-fin tubes. These data are for the tubes submerged in liquid pool and kept in horizontal orientation.

Figures 3.3 and 3.4 represent schematic diagram and photographic view of plain and integral-fin tubes employed in this investigation, respectively. The fins are of trapezoidal shape with fin height of 1 mm, and density of 748 and 1024 fpm.

The plain and integral-fin tubes used in this investigation were fabricated out of a solid brass rod (70% Cu and 30% Zn) in the workshop of Mechanical and Industrial Engineering Department, University of Roorkee, Roorkee. Table 3.1 shows relevant dimensions of the heating tubes.



1. test vessel
2. heating tube
3. heater
4. liquid level indicator
5. view ports
6. gland & nut assembly
7. socket
8. condenser
9. pressure gauge
10. bubbler
11. surge tank
12. vacuum pump
13. motor
14. auto-transformer
15. voltage stabilizer
16. PVC tube
17. liquid pool

T_w wall thermocouples

T_L liquid thermocouples

V_1- V_6 valves.

Fig. 3.1 Schematic diagram of experimental set-up

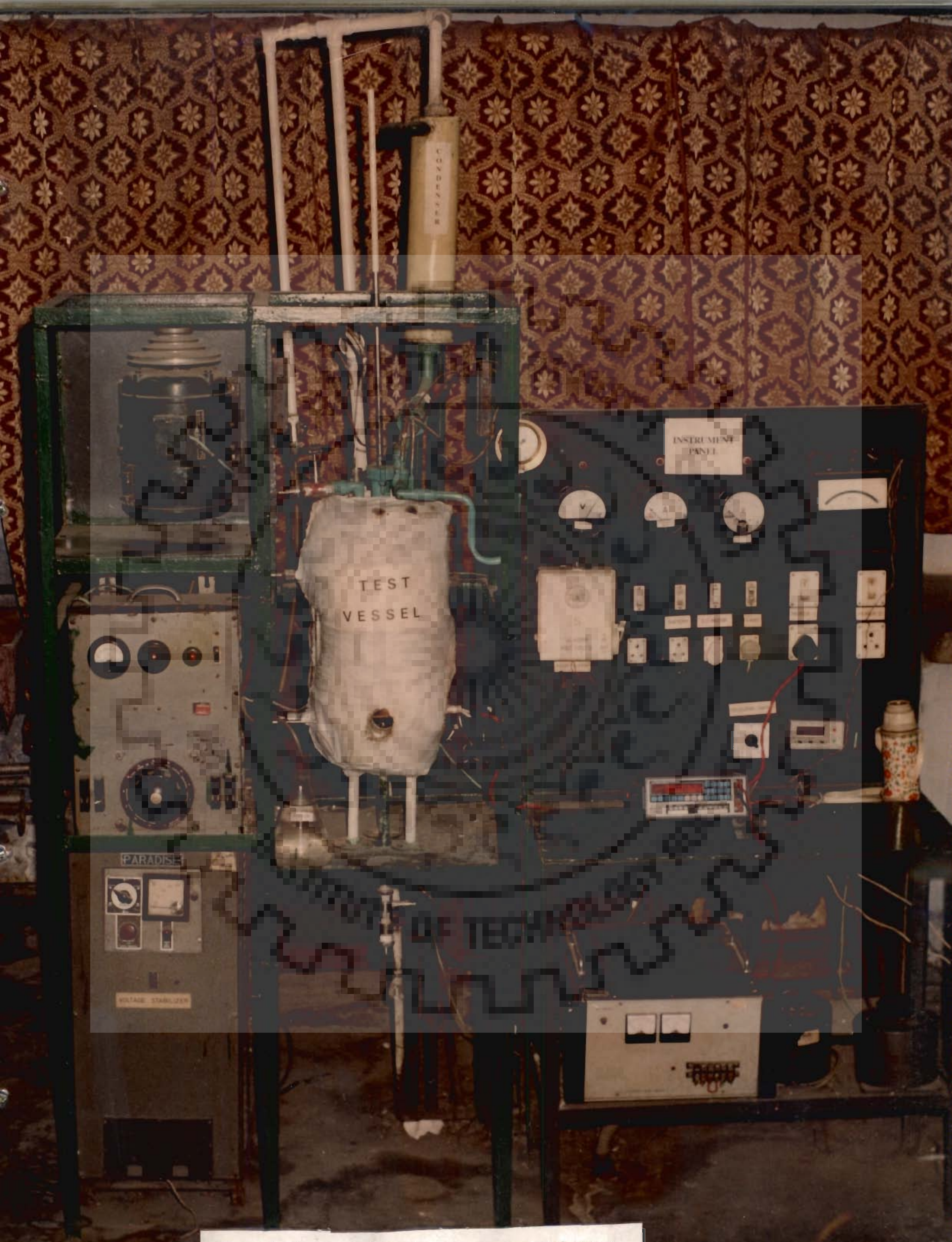
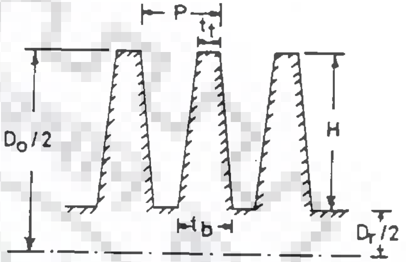
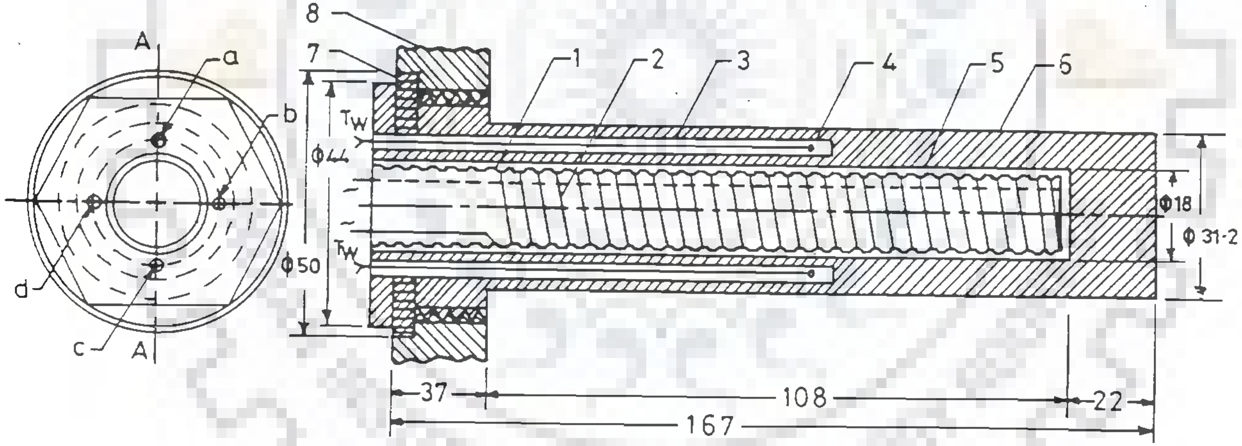


Fig. 3.2 Photographic view of experimental set-up



Fin profile



1.porcelain rod
2.nichrome wire
3.thermocouple holes
4.wall thermocouples
5.mica sheet & glass tape
6.surface (plain/finned)
7.teflon gland
8.socket

Fig. 3.3 Details of heating tubes tested

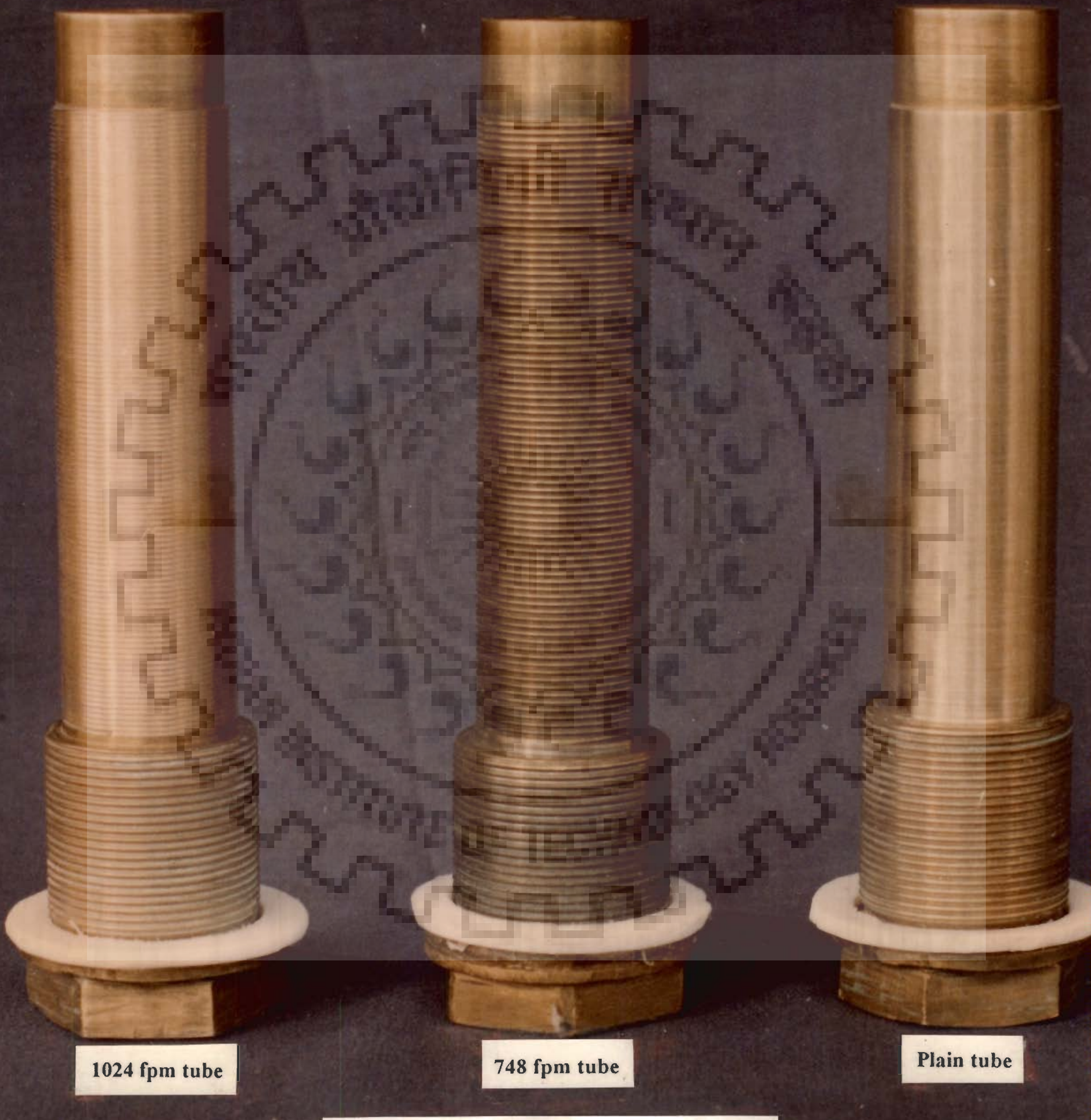


Fig. 3.4 Photographic view of heating tubes

Table 3.1 Specification and dimensions of heating tubes

Parameter	Heating tube geometries		
	Plain	Trapezoidal integral-fin	
		748 fpm	1024 fpm
Effective heated length, L	108.0	108.0	108.0
inner diameter, D_i	18.0	18.0	18.0
outside diameter, D_o	31.2	31.2	31.2
root diameter, D_r	----	29.2	29.2
fin height, H	----	1.0	1.0
fin thickness (base), t_b	----	0.8	0.7
fin thickness (tip), t_t	----	0.35	0.3
fin pitch, P	----	1.35	1.02

* all dimensions are in mm.

As is clear from Figure 3, each tube has an axial hole of 18 mm diameter, drilled from one end of a rod upto a length of 145 mm, leaving a portion of 22 mm undrilled to avoid longitudinal heat flow. In this context, it is important to mention that Gupta (1979) had measured wall temperature of this length of a heated tube in his experimentation. He had reported that wall temperature was almost the same as that of the surrounding liquid pool, implying that heat did not travel axially through the undrilled length of the tube. Heating tube has been secured with vessel wall with teflon seal to avoid leakage of liquid from test vessel. A teflon plug, provided at a hexagonal end of the heating tube, helps to minimize the heat loss to the atmosphere.

Thermocouple holes in wall thickness of each heating tube on a pitch circle diameter of 24.6 mm have been designated as a, b, c and d as shown in Figure 3.3. Each hole is of 2 mm diameter and a length of 91 mm. Calibrated copper-constantan thermocouples, (T_w) placed in each of the holes in such a manner that their beads touch the dead ends of their respective holes. The thermocouples have been electrically insulated from heating tube wall by means of teflon tape a 0.05-1.0 mm thick. Outer surface of a given tube was finally finished by rubbing it with 400 grade emery paper to ensure consistent surface structure.

To provide a desired heat flux from a given tube, a home-made electric cartridge heater was been employed. Details of heater have been shown in Figure 3.3. This essentially consists of a 15 mm outer diameter threaded porcelain rod having 22 gauge nichrome wire wound on it. To safeguard against any electrical leakage between heater and heating tube, the former has been properly covered with a thin layer of mica sheet, followed by a layer of glass tape. The two ends of the nichrome wire have been connected to a-c power mains through an auto-transformer (14). This, in turn, is connected to a 3 kVA voltage stabilizer(15), as shown in Figure 3.1.

3.2.3 CONDENSER

Condenser(8), employed for the condensation of vapours from the boiling liquid, is shown in Figure 3.1. It is a double pipe type heat exchanger having 550 mm long inner and outer tubes of 50 mm and 100 mm diameter, respectively.

It is mounted vertically over the test vessel. Cooling water flows through the outer pipe, whereas the vapours from the liquid pool pass through inner tube. The condensate from condenser returns to the liquid pool by gravity. Thus, it flows counter-currently to the vapours from the pool to condenser, thereby the condensate gets heated up. Thus, it eliminates the possibility of subcooling of the condensate, if any. Air vent valve, V_3 installed at the top of inner pipe of the condenser, helps in the removal of non-condensables to atmosphere.

3.2.4 HEIGHT OF LIQUID POOL IN TEST VESSEL

For the reasons mentioned in section 3.1.4, a liquid pool height of about 100 mm above the upper surface of the heating tube has been found adequate. In fact, it ensures that when condensate from condenser (8) joins back the liquid pool in vessel (1) it does not disturb the superheated liquid layer around the heating tube.

3.2.5 VACUUM PUMP AND ACCESSORIES

Vacuum pump (12) is used to create vacuum in the system. It is a single stage oil immersed type rotary pump driven by 0.5 hp motor having maximum speed of 500 rpm. As is clearly seen in Figure 3.1, the pump is connected to the experimental rig through a surge tank (11) and needle valve V_3. The valve helps in the regulation of vacuum in the vessel.

3.2.6 INSTRUMENTATION

To conduct a series of experiments, it is required to measure variables, such as power input to the heating tube, wall-and liquid-temperatures and pressure in the test vessel. Accordingly, experimental set-up has been provided with suitable instruments, as described below:

(a) Power Input to the Heating Tube

Heater (3) in Figure 3.1, used for supplying power to heating tube, is connected to a-c power mains through an auto-transformer (14) and a voltage stabilizer (15) so that stabilized current is supplied to the heater. Power to the heating tube is measured by means of a precision grade wattmeter of 1% accuracy. The wattmeter was calibrated against a standard wattmeter in the Measurements Laboratory of Electrical Engineering Department of University of Roorkee, Roorkee. The range of wattmeter is 0-650 W. Different values of power input are obtained by adjusting auto-transformer as already mentioned above.

(b) Wall-and Liquid-Temperatures

Wall-and liquid-temperatures were measured by calibrated copper-constantan thermocouples of 26 gauge, whose electro motive force values were measured by a Digital Multi-Meter (D.M.M, model No. 177) manufactured by M/s Keithley Instruments Inc., Ohio, U.S.A. with ±1 mV least count in 20 mV range. Thermocouple leads are connected to the D.M.M. through a 12 point selector switch and cold junction, (a bath of melting ice to give the reference temperature of 0 °C).

(c) Vacuum/Pressure Gauge

Pressure in the vessel has been measured by vacuum/pressure gauge (9) of ±1.0 kPa accuracy calibrated against a mercury barometer.

CHAPTER 4

EXPERIMENTAL PROCEDURE

This Chapter discusses operating procedure along with various precautions exercised to obtain experimental data for nucleate pool boiling of binary and ternary liquid mixtures from a horizontal single tube.

4.1 PRECAUTIONS

To obtain reliable and accurate experimental data, several precautions were observed during the course of experimentation, the important ones are summarized in the following sections.

4.1.1 MECHANICAL AND ELECTRICAL LEAKAGE

All the valves except V_1 were fully closed and the set-up was pressurized to about 200 kPa (2 atm) by compressed air through valve V_1. Soap water solution was applied at all joints of the vessel as well as the pipe-lines connecting one component to other of the set-up. Each joint was carefully examined for the possible formation of air bubbles, as it represented a leaky joint. The faulty joints, so detected, were suitably attended. This procedure was repeated till the set-up became completely leak-proof. It was finally filled with compressed air at a pressure of 200 kPa and observed over a period of 48 hours. No fall in the reading of pressure gauge (9) was noted. Finally, the set-up was subjected to a vacuum of about 31 kPa with the help of vacuum pump (12). No drop in the reading of the vacuum gauge even after a period of about 48 hours was observed, indicating that the set-up was leak-proof.

All electrical connections were earthed for safe operation of the experimental facility. Tests were also conducted to check against any electrical leakage.

4.1.2 PREPARATION OF LIQUID MIXTURES

Reagent grade liquids and double distilled water were used in the preparation of all mixtures. Liquid mixture compositions were prepared on a weight basis using a precision balance of accuracy ±1.0 gm.

4.1.3 CLEANING, RINSING AND CHARGING OF TEST VESSEL

Prior to charging the set-up with a test liquid, it was thoroughly cleaned for the traces of previous liquid. This was accomplished by flushing all the components of the experimental facility with compressed air. Heating tube and test vessel were then rinsed with distilled water, acetone, and finally with liquid under investigation. Now, cooling water supply was started to condenser. Test vessel was then filled with the liquid upto a height of 100 mm above the top surface of the heating tube. The test liquid was drawn into vessel from a closed storage container by creating vacuum in the vessel. This procedure minimized the amount of air that could have entered the vessel along with test liquid.

4.1.4 DEAERATION OF TEST LIQUID

Removal of air dissolved in the test liquid was quite important. Its presence would have affected boiling phenomenon significantly, as explained by Berenson (1962), Alam and Varshney (1973b), Pandey et al.(1986) and Fisenko et al. (1988) amongst many other researchers. Deaeration was done by heating the test liquid to its saturation temperature followed by continuous boiling. Consequently, dissolved air came out of the test liquid, as evinced by the appearance of bubbles in the bubbler (10). Stoppage of bubbling indicated that no more air is coming out with vapours. This exercise was carried out every time before starting a set of experimental runs. It took about two hours for complete deaeration of the liquid pool.

4.1.5 STABILIZATION OF HEATING TUBE SURFACE

In order to obtain reproducible experimental data, it was essential that heating tube was first of all thermally stabilized. If this were not done, data would have been changing with boiling duration. This has been well-documented by Alam and Varshney (1973b), Happel and Stephan (1974), and Webb and Pais (1992) amongst several researchers. For this, the vessel was filled with test liquid upto height of 100 mm above the top of heating tube surface. Then, electric heater (3) was switched on at a maximum power of about 620 W. After a prolonged boiling of 72 hours, no change in the readings of wall thermocouples was noticed. This was indicative of the fact that the heating tube was thermally stabilized.

The above procedure for stabilization was always employed before experiments were carried out with a new test liquid and also heating tube. Data were collected over a period of ten months after one month of preliminary testing.

4.1.6 CONDENSER

There was an obvious need to estimate whether surface area of condenser (8) was sufficient enough to condense all the vapours generated as a result of boiling. For this, first of all, cooling water supply was started to the condenser. The heating tube was energized at a heat flux of 57,624 W/m². This, as a matter of fact, was the largest possible heat flux to the heating tube in the present investigation. This, of course, contributed to the maximum possible vapour generation rate. Vapours condensed inside the inner tube of condenser and thus generated condensate thus generated fell back to the liquid pool in the vessel. No changes in the readings of the pressure gauge (9) and liquid thermocouples were observed. This confirmed that vapours were condensed at the same rate at which they were generated from the liquid pool, indicating that the surface area of condenser and cooling water supply were adequate for the present investigation.

4.1.7 OTHER PRECAUTIONS

a. The bulk liquid temperature was compared with saturation temperature corresponding to the measured pressure inside the test vessel. An excellent matching was observed, which ensured that there were no non-condensables present in the liquid pool. It also verified that there was no subcooling in the pool.

b. The heating tube was tested for circumferential uniformity of heat flux. This was done by rotating the heating tube by $(\pi/2)$ radian. The readings of wall thermocouples did not vary by more than 0.05 ^{0}C. It was further rotated by another $(\pi/2)$ radian in steps. Various readings did not exhibit any appreciable change in wall thermocouples' readings.

c. To validate the thermocouples' installation, the heating tube was rotated, and the temperature was measured at different circumferential locations by two thermocouples. As these two thermocouples registered equal values at a given angular location, it confirmed that there were no installation error, and that heat flux was constant at tube circumference.

4.2 OPERATING PROCEDURE

In this section, the procedure followed in obtaining experimental data of nucleate pool boiling of single component liquids, and their binary and ternary mixtures on plain and integral-fin tubes at atmospheric pressure is explained.

Before obtaining a series of experimental data for the boiling of different test liquids, experimental set-up was commissioned and checked against mechanical and electrical leakages. Other steps such as preparation of liquid mixtures, supplying of cooling water to condenser, cleaning, rinsing, and charging the vessel with the test liquid were carried out.

First of all, experiments were conducted for the pool boiling of single component liquids, viz., distilled water, acetone, and isopropanol on plain tube. For this, the vessel was filled with a test liquid to a height of 100 mm above the top of the heating tube surface. Then, electric heater was energized so that the heating tube transferred heat to the liquid pool at a heat flux of 57,624 W/m^2, by modulating auto-transformer (14).

Deaeration of the liquid pool was carried out following the procedure described in subsection 4.1.4. It continued for two hours till thermal equilibrium was attained ensuring that air trapped in the nucleation sites was removed. The vessel was carefully maintained at atmospheric pressure to avoid the effect of variation in pressure on boiling process.

Thus, the set-up was ready for obtaining the series of experimental data for boiling of single component liquid. Reading of wattmeter was kept under constant vigil and maintained constant. Once wall and liquid thermocouples' readings became constant with the time, it indicated that the system had attained steady state. Then, the readings of wall and liquid thermocouples, wattmeter, and barometer were recorded. To be further sure of the existence of steady state, boiling was continued for another hour and the readings of thermocouples were found to remain unchanged. Experiments were always conducted by decreasing the values of heat flux. This procedure was used to avoid the possibility of hysteresis in the boiling curve, as reported by Kartsounes (1975), Jensen and Hsu (1988), and Hahne et al. (1991). Experiments were carried out for various values of heat flux, viz., 52901, 44399, 35897, 27395, 18893, and 10391 W/m². During experimentation, the pressure was maintained at atmospheric and liquid pool level in the vessel was 100 mm above the top of the heating tube surface.

Experimental data for boiling of liquid mixtures on plain tube were obtained following the procedure similar to that for the boiling of single component liquids. First of all, valve V_2 was opened and test liquid in question was drained off completely from the vessel. This was followed by rinsing the heating tube with acetone and then by test liquid mixture. After this, compressed air was passed into the vessel by opening valve V_1 and was forced out of the vessel through valve V_2. This helped in removing all the traces of previous liquid adhering to inner surface of the vessel. Now, the vessel was filled with test liquid mixture and the experiments were carried out in a manner similar to that for boiling of single component liquid.

While conducting experiments with binary and ternary liquid mixtures, samples of these liquid mixtures before and after each experiment were taken to determine their densities in order to check if any changes in the composition of liquid mixture had taken place during the boiling. The results showed negligible variation, within ±0.01 mole fraction. The samples were compared in the Instrumentation Laboratory at room temperature maintained between 17-20 °C.

After obtaining data for boiling of pure liquids, binary mixtures of acetone-water, isopropanol-water, and acetone-isopropanol, and ternary mixtures of acetone-isopropanol-water on the plain tube, the heating tube was replaced by integral-fin tubes of 748 and 1024 fpm one by one. The set-up was commissioned and experiments were conducted as described above for the plain tube. Table 4.1 shows the operating variables such as composition of liquid mixtures, heat fluxes, and also the surface geometry of heating tubes, employed in the present investigation.

Experimental data, obtained in this investigation, have been listed in Appendix D. In all the tables of Appendix D column 2 lists values of heat flux, whereas outer wall temperatures, T_{wo} appear in columns 3, 4, 5, and 6. Corresponding bulk liquid temperatures, T_s around the heating tube are reported in columns 7, 8, 9 and 10. The last column lists average values of heat transfer coefficients. Calculation procedure for boiling heat transfer coefficients is detailed in Appendix A.

4.3 REPRODUCIBILITY CHECK OF EXPERIMENTAL DATA

It is essential that data obtained are reliable. Hence, several experimental data were checked frequently by repeating them after regular interval of time. It was noticed that the data were reproducible within a maximum of ±5% experimental error.

Table 4.1 Operating variables and heating tube surface geometries

Liquid	Composition, mole fraction						Range of heat flux, W/m^2	Heating tube surface geometries
Single component								
a. Acetone	------------------						10391-57624	Plain, 748 fpm, 1024 fpm
b. Isopropanol	------------------						10391-57624	Plain, 748 fpm, 1024 fpm
c. Distilled water	------------------						18893-57624	Plain, 748 fpm, 1024 fpm
Binary mixture								
a. Acetone-Water	X_1	0.0,0.05,0.10,0.15,0.25 0.40,0.60,0.80,1.0					10391-57624	Plain, 748 fpm, 1024 fpm
b. Isopropanol-Water	X_1	0.0,0.05,0.10,0.20,0.40, 0.60,0.68,0.80,0.90,1.0					10391-57624	Plain, 748 fpm, 1024 fpm
c. Acetone-Isopropanol	X_1	0.0,0.10,0.20,0.30,0.40, 0.50,0.60,0.80,1.0					10391-52901	Plain, 748 fpm, 1024 fpm
Ternary mixture								
a. Acetone-Isopropanol-Water	X_1	0.000	0.000	0.011	0.020	0.050	10391-57624	Plain, 748 fpm, 1024 fpm
	X_2	0.000	1.000	0.074	0.019	0.040		
	X_1	0.083	0.124	0.155	0.212	0.266		
	X_2	0.036	0.290	0.150	0.182	0.282		
	X_1	0.294	0.389	0.524	0.699	1.000		
	X_2	0.484	0.212	0.267	0.105	0.000		

CHAPTER 5

A MODEL FOR BOILING OF LIQUID MIXTURES ON A HORIZONTAL TUBE

Design of reboilers for boiling of binary/multicomponent mixtures is quite cumbersome. This is ascribed to non- availability of a generalized correlation for the prediction of nucleate pool boiling heat transfer coefficients. Hence, a generalized correlation for the prediction of heat transfer coefficients has been derived considering the analysis of the process of nucleate pool boiling of binary and ternary liquid mixtures on a horizontal plain as well as integral-fin tubes at atmospheric pressure.

5.1 REMARKS ON EXISTING CORRELATIONS ON PLAIN TUBE

A review of the existing correlations for the boiling of binary liquid mixtures on plain tube has been carried out in sub-sections 2.5.1 through 2.5.5 of Chapter 2. It reveals that there are analytical correlations due to Calus and Leonidopoulos (1974) and Thome (1981). However, these correlations require a pre-knowledge of physico-thermal properties of liquid mixtures and also the liquid mass diffusivity. But, there is no simple and general equation for predicting liquid mass diffusivity over a wide range of composition of a given mixture, especially for aqueous mixtures, as emphasized by Calus and Rice (1972) and Uhlig and Thome (1985). Further, these correlations are limited to the boiling of binary liquid mixtures only and hence can not be extended to multicomponent mixtures.

Among semi-empirical correlations for the boiling of liquid mixtures on plain tube, the Schlunder correlation (1982) has been recommended by many investigators, namely; Uhlig and Thome (1985), Shakir et al.(1985), Gorenflo et al.(1988), and Tarrad and Burnside (1991). This correlation is based on the film theory of mass transfer and has an empirical scaling factor, B_o which accounts for the fraction of heat required for bubble formation. However, Schlunder has assumed the value of B_o equal to 1.0, which is not always true. This has been pointed out by Shakir et al., Gorenflo et al., and Tarrad and Burnside. Further, Schlunder has also assumed that the liquid mass transfer coefficient has a constant value equal to 0.0002 m/s. However, in actual practice a variation in its value from 0.0001 to 0.0005 m/s has been reported by Nagel (1981) and Gropp (1981). The dependence of Schlunder's correlation on mass diffusivity poses a problem in estimation of heat transfer coefficient of liquid mixtures, as mentioned in the preceding paragraph.

It is important to point out that Thome and Shakir (1987) have tested a number of semi-empirical correlations for the prediction of nucleate pool boiling heat transfer coefficient for liquid mixtures.They concluded that none of these correlations is completely satisfactory for the boiling of aqueous mixtures.

For all the existing correlations, except those due to Palen and Small (1964) and Thome and Shakir (1987), ideal heat transfer coefficient, h_{id} for mixtures has been invariably determined from linear mixing law. However, there are serious reservations in this regard as observed by Thome and Shock (1984).

Hence, there is a need to select a method which is logical and realistic for the determination of h_{id}.

From the above discussion and available correlations, it is logical to conclude that a correlation based on pertinent physical mechanism of the boiling of binary liquid mixtures should be developed which should be simple and of general applicability to mixtures of widely differing physico-thermal properties over their wide ranges of composition.

In the following section, a correlation satisfying the above requirements for the prediction of heat transfer coefficients for nucleate pool boiling of binary liquid mixtures on plain tube has been derived.

5.1.1 ANALYSIS OF BINARY MIXTURE BOILING ON PLAIN TUBE

In this section, physical mechanism related to nucleate pool boiling of a binary liquid mixture on plain tube is analyzed in order to obtain correlation for the prediction of boiling heat transfer coefficients.

van Wijk et al.(1956) were apparently the first to explain the theory of heat transfer during nucleate pool boiling of liquid mixtures on plain tube. Unlike the case of single component liquids, boiling of binary/multicomponent liquid mixtures is characterized by mass diffusion of its components at the interface between liquid phase and vapour bubbles on a heating tube. Let us consider the case of a typical binary liquid mixture, whose vapour-liquid phase equilibrium diagram is as shown in Figure 5.1. For a bulk liquid the composition of more volatile component is equal to X_1, the bubble point temperature is T_s, and its equilibrium composition in vapour phase is Y_1. Evidently Y_1 is greater than X_1, indicating that the vapour in equilibrium with the liquid is richer with respect to the more volatile component. Now , let us consider a vapour bubble on the heating surface but lying in the adjacent superheated liquid layer. The fact, that the composition of the more volatile component in vapour bubble is greater than that in the bulk liquid, can be possible only when this component in the liquid layer immediately adjacent to vapour bubble interface evaporates more in quantity than the less volatile component. Accordingly, the composition of the more volatile component near the interface should decrease from X_1 to a some lower value say, $X_{1,ph}$. Thus, a concentration difference develops around the vapour bubble with respect to the bulk liquid. Hence, mass diffusion of the more volatile component from the bulk liquid to the bubble interface sets in. The equilibrium vapour mole fraction corresponding to $X_{1,ph}$ is denoted by $Y_{1,ph}$. A reduction in concentration of the more volatile component at the vapour bubble interface results in a rise in bubble point temperature from its initial value of T_s to T_{ph}, as shown in Figure 5.1. Consequently, effective wall superheat reduces from (T_{wo} -T_s) to (T_{wo} - T_{ph}), where T_{wo} is outer wall temperature of the heating tube. For an electrically heated system (q = constant), the steady state condition demands that outer wall temperature T_{wo} should rise to a value such that the heating surface transfers the heat at a rate at which it receives it from the electric heater. It is important to remember that in heat transfer studies the wall superheat is conventionally calculated based on the bulk liquid bubble point temperature, T_s instead of the temperature at the interface, T_{ph} for the estimation of heat transfer coefficient,since it is quite difficult to measure T_{ph}.

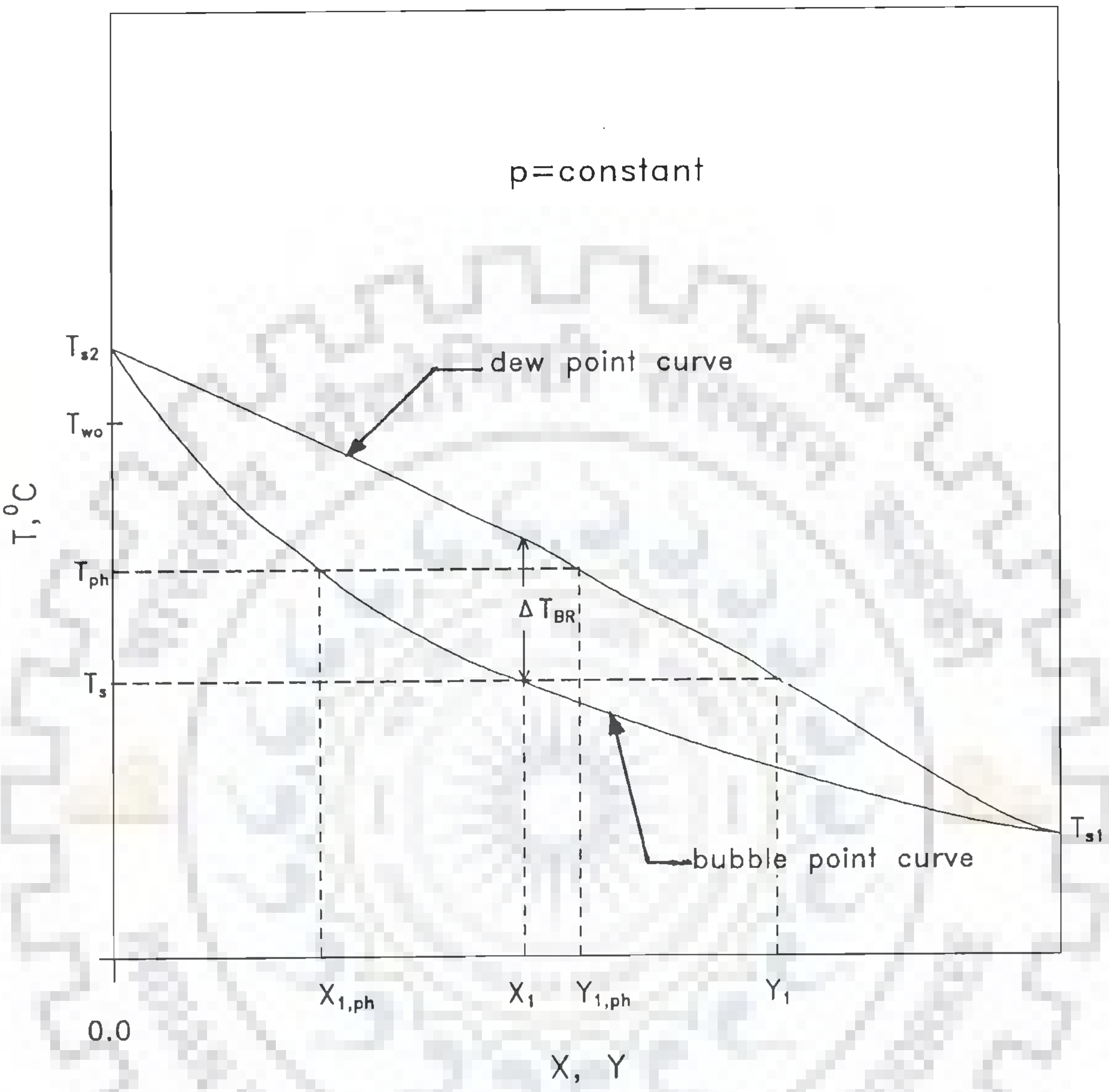


Fig. 5.1 Vapour–liquid phase equilibrium diagram for a binary mixture

Further, values of wall superheat so calculated are always higher.

The value of T_{ph} is dependent on the rate of mass diffusion from the bulk liquid to the bubble interface, and T_{ph} affects the value of wall superheat, $(T_{wo} - T_{ph})$. The latter, in turn, affects the heat transfer coefficient. The mass transfer through diffusion is a slower process than the heat transfer taking place in boiling of liquid mixtures. Thus, the mass diffusion controls the value of heat transfer coefficient of boiling liquid mixtures.

It is now clear that due to vapour bubble formation on a heating tube the liquid composition there-on is different from that in the bulk liquid. As regards the heat from the heating tube, it is partly transferred to growing vapour bubbles directly as latent heat of vaporization and the remainder goes as free convective heat to liquid from that part of the heating surface where there are no vapour bubbles formed . But the temperature difference, responsible for both modes of heat transfer, is equal to $(T_{wo}-T_{ph})$. Hence, heat transfer rate per unit area of heating tube surface is obtained by the following equation:

$$q = h_{ph} (T_{wo}-T_{ph}) \qquad (5\text{-}1)$$

where h_{ph} is heat transfer coefficient at the phase interface accounting for both the modes of heat transfer.

Based on the fact that heat flowing from outer wall of a heating tube is finally received by the bulk liquid mixture at a temperature of T_s, the following equation is understandable:

$$q = h (T_{wo}-T_s) \qquad (5\text{-}2)$$

where h represents average heat transfer coefficient during boiling of liquid mixture on heating tube. Now, from Eqs.(5-1) and (5-2), one gets the following equation:

$$\begin{aligned}(h/h_{ph}) &= (T_{wo}-T_{ph})/(T_{wo}-T_s)\\ &= 1/[(T_{wo}-T_s)/(T_{wo}-T_{ph})]\\ &= 1/[1+(T_{ph}-T_s)/(T_{wo}-T_{ph})] \qquad (5\text{-}3)\end{aligned}$$

Substituting the value of $[1/(T_{wo}-T_{ph})]$ from Eq.(5-1) into Eq.(5-3), the following equation is obtained:

$$(h/h_{ph}) = 1/[1+(h_{ph}/q)(T_{ph}-T_s)] \qquad (5\text{-}4)$$

Eq.(5-4) in its present form contains two quantities h_{ph} and T_{ph}, whose direct measurement is difficult. As a first approximation, h_{ph} may be assumed equal to $(h_{id})_p$, because it can be determined from available correlations for nucleate pool boiling of liquids of differing physico-thermal properties on plain tube. In fact, this is generally assumed so, as suggested by Schlunder (1986). Correlation proposed by Stephan-Abdelsalam (1980) is considered to be the best out of the available ones for the calculation of $(h_{id})_P$. This has been discussed in section 2.1.5. However, this correlation is reproduced below:

$$(h_{id})_P = 0.0546 (k_l /d)[(qd / k_l T_s)(\rho_v /\rho_l)^{0.5}]^{0.67} (\rho_v/ \rho_l)^{-4.33}(h_{fg}\, d^2/\alpha^2)^{0.248} \qquad (5\text{-}6)$$

where d is bubble departure diameter and is calculated from the following equation:

$$d = 0.0146\, \beta\, [(2\sigma/\{g(\rho_l-\rho_v)\})]^{0.5} \qquad (5\text{-}7)$$

Contact angle β is taken to be 35^0 for all the mixtures.

Term $(T_{ph}-T_s)$ in Eq.(5-4) is determined by following equation, as the bubble point curve is almost linear for liquid compositions ranging from $X_{1,ph}$ to X_1 (Figure 5.1):

$$(T_{ph}-T_s) \cong [(X_1-X_{1,ph})\ dT/dX] \tag{5-8}$$

Further, as shown by Stephan (1981), the following relationship holds true:

$$(X_1-X_{1,ph}) \cong (Y_1-X_1) \tag{5-9}$$

Now from Eqs.(5-4), (5-8), (5-9) and the assumption that h_{ph} is equal to $(h_{id})_P$, the following relationship for boiling of binary mixture on plain tube is obtained:

$$(h/h_{id})_P \cong 1/[1-\{(h_{id})_P/q\}\ (Y_1-X_1)(dT/dX)] \tag{5-10}$$

Thome and Shakir (1987) have shown that the value of $[-(Y_1-X_1)(dT/dX)]$ can be approximated as 'boiling range', ΔT_{BR} which is the difference between the dew and bubble point temperatures at a given bulk liquid mixture composition. Using this in Eq.(5-10), the following functional relationship is obtained:

$$(h/h_{id})_P \cong 1/[1+\{(h_{id})_P/q\}\Delta T_{BR}] \tag{5-11}$$

It is important to mention that Eq.(5-11) is not rigorous, as some approximations have been made in its derivation. Hence, resulting error in the prediction of heat transfer coefficient is likely to be large. However, it is convincingly established that $(h/h_{id})_P$ is a function of $[1/\{1+((h_{id})_P/q)\ \Delta T_{BR}\}]$. To obtain a generalized functional relationship between them, least square curve fitting has been carried out using the present experimental data of boiling binary liquid mixtures of acetone-water, isopropanol-water, and acetone-isopropanol on a horizontal plain tube at atmospheric pressure, which are listed in Tables D.2, D.3, and D.4 of Appendix D. Accordingly, the following correlation has been derived:

$$(h/h_{id})_P = 1/[1+\{(h_{id})_P/q\}\ \Delta T_{BR}]^{0.158} \tag{5-12}$$

Thus, Eq.(5-12) is a semi-empirical correlation, which is capable of predicting heat transfer coefficient for the boiling of binary liquid mixtures on plain tube at atmospheric pressure. At this juncture, it is worth-mentioning that correlation, Eq.(5-12) is based on 132 data points. Figure 5.2 shows a comparison between the present experimental data of binary liquid mixtures boiling on plain tube at atmospheric pressure and predicted values from the correlation, Eq.(5-12). From the plot, it is evidently clear that an excellent agreement exists with standard deviation (S.D.) and mean absolute error (M.A.E.) values as given in Table 5.1.

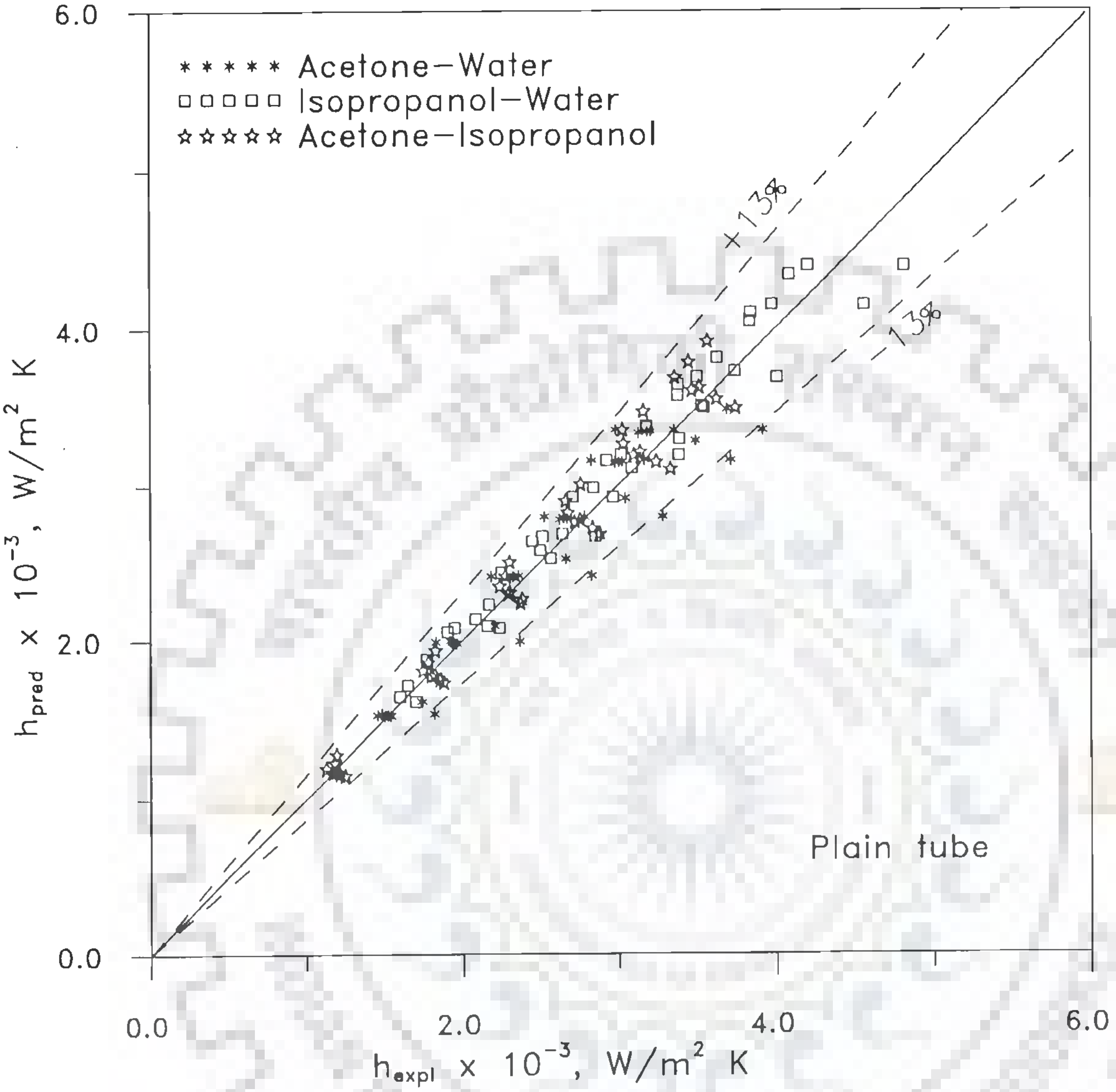


Fig. 5.2 Comparison of present data of binary mixtures with proposed correlation, Eq.(5–12)

Table 5.1 Standard deviation and mean absolute error for the predictions from Eq.(5-12) of present data on plain tube

Binary mixtures	Boiling range ΔT_{BR},K	Composition X_1	Data points	(S.D.)%	(M.A.E.)%
Acetone-Water	8.5-31.3	0.05-0.80	42	7.57	5.52
Isopropanol-Water	0.2-13.6	0.05-0.90	48	5.45	4.99
Acetone-Isopropanol	3.3-8.0	0.10-0.80	42	5.75	5.14

From Table 5.1, it is clear that Eq.(5-12) correlates all the data points of binary liquid mixtures over their wide ranges of compositions and high boiling range. These mixtures are also of widely differing physico-thermal properties. Since, Eq.(5-12) is equally applicable to azeotrope and non-azeotrope systems, it can be claimed to be of general applicability.

5.1.2 VERIFICATION OF DERIVED CORRELATION, EQ.(5-12)

It has been thought desirable to compare derived correlation, Eq.(5-12), for the prediction of heat transfer coefficient for nucleate pool boiling of binary liquid mixtures on plain tube, with experimental data reported by earlier investigators: Calus and Leonidopoulos (1974) for n-propanol-water; Pandey (1982) for isopropanol-water; Uhlig and Thome(1985) for acetone-water; Shakir et al.(1985) for methanol-water; and Bajorek et al.(1989) for acetone-water on single horizontal tubes. This has been carried out in Figure 5.3. Operating parameters for these experimental data are summarized in Table 5.2.

Table 5.2 Operating parameters for liquid mixtures boiling on plain tube at atmospheric pressure

Investigators	Binary mixture	Boiling range ΔT_{BR}, K	Compositions X_1	Heat flux,	Heating tube W/m^2	Data points
Calus et al.(1974)	n-Propanol-Water	0.1-8.5	0.03-0.80	100000-400000	Nickel-Aluminum	77
Pandey (1982)	Isopropanol-Wate	0.2-13.6	0.05-0.50	9974-30840	Stainless steel	25
Uhlig et al.(1985)	Acetone-Water	7.3-31.3	0.025-0.80	25800-190000	Copper	48
Shakir et al.(1985)	Methanol-Water	4.6-13.3	0.05-0.80	28300-217000	Copper	36
Bajoreketal.(1989)	Acetone-Water	8.4-31.3	0.05-0.80	25000-300000	Copper	35

The physico-thermal properties required to calculate $(h_{id})_P$ from Eq.(5-6)have been determined by the procedure given in Appendix C.

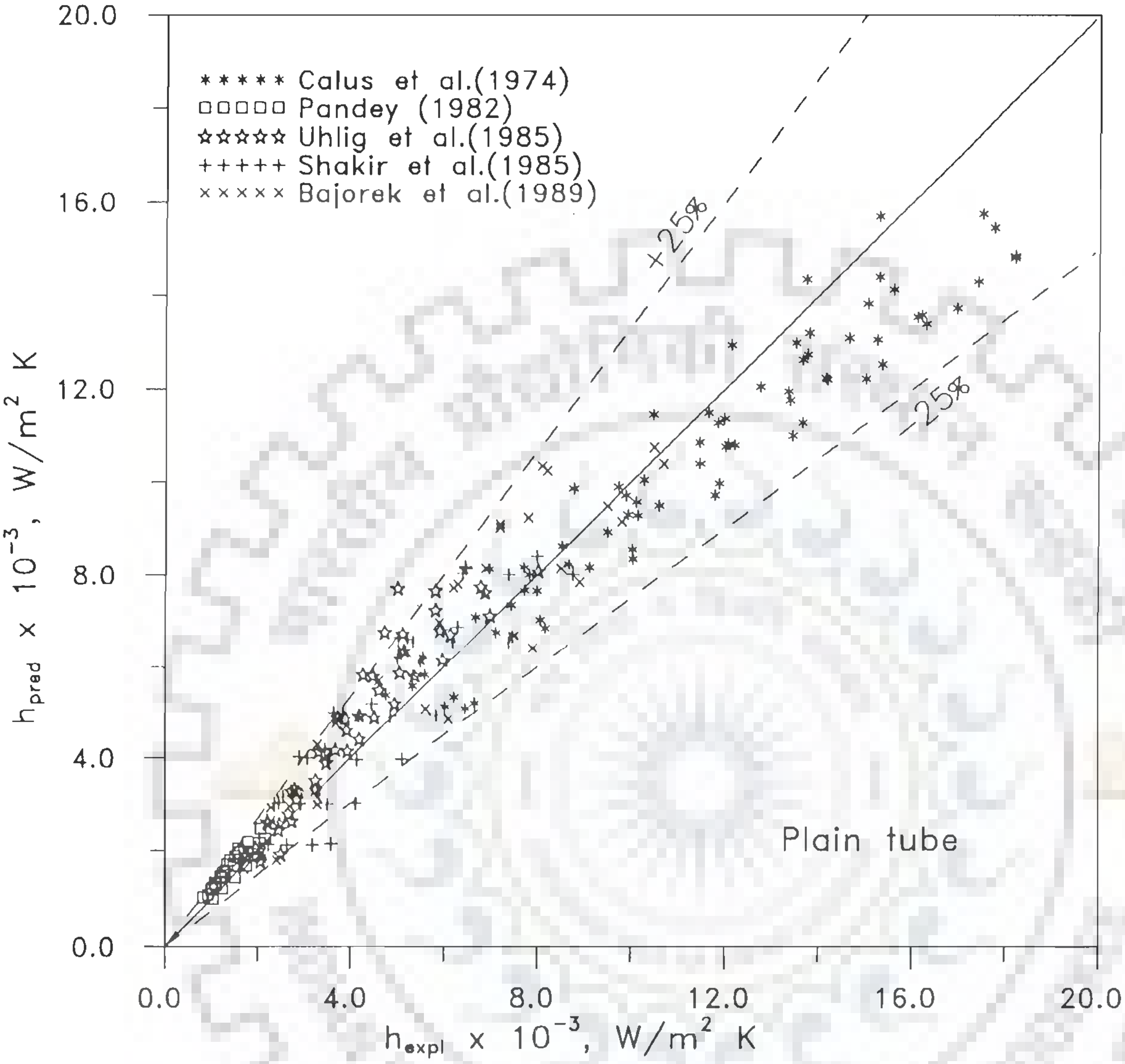


Fig. 5.3 Comparison of earlier binary boiling data with proposed correlation, Eq.(5−12)

An examination of Figure 5.3 shows that there is an good agreement between the experimental and the predicted values within maximum error of ±25 per cent. The standard deviation(S.D.) and mean absolute error(M.A.E.) are given in Table 5.3.

Table 5.3 Standard deviation and mean absolute error for the predictions from Eq.(5-12) of existing data on plain tube

Investigators	(S.D.)%	(M.A.E.)%
Calus et al.(1974)	11.88	10.23
Pandey (1982)	18.36	16.71
Uhlig et al.(1985)	18.15	14.49
Shakir et al.(1985)	20.37	18.26
Bajorek et al.(1989)	17.70	16.06

From the (S.D.) and (M.A.E.) values, it is evident that the derived correlation, Eq.(5-12) is in good agreement with data obtained by earlier investigators on a plain single heating tube of differing materials. Thus, it is confirmed that Eq.(5-12) can be used as a generalized correlation.

5.2 REMARKS ON EXISTING CORRELATIONS ON INTEGRAL-FIN TUBE

As regards the correlation for nucleate pool boiling of liquid mixtures on integral-fin tube, it seems that the correlation due to Palen and Yang (1983) is the only one available in the open literature. This is well-documented by Webb et al.(1989), Thome (1990), and Kumar (1992). A serious drawback of this correlation lies in the fact that it has many empirical coefficients, whose values are neither reported, nor a detailed procedure to evaluate them is available.

5.2.1 DEVELOPMENT OF A CORRELATION

Heat transfer coefficient for boiling of binary liquid mixtures on an integral-fin tube, h_F is related to boiling heat transfer coefficient of the same mixture on plain tube, h_P by the following equation:

$$h_F = F_s\, h_P \qquad (5\text{-}13)$$

where F_s is 'surface correction factor ' for h_P due to the presence of fins on the outer surface of a plain tube. Using linear mixing law for the prediction of h_P and h_f for mixtures one can easily derive the following expression for F_s as given below:

$$F_s = [\{X_1/(h_1)_F\}+\{X_2/(h_2)_F\}]^{-1}/[\{X_1/(h_1)_P\}+\{X_2/(h_2)_P\}]^{-1} \qquad (5\text{-}14)$$

where $[(h_1)_P$ and $(h_2)_P]$ and $[(h_1)_F$ and $(h_2)_F]$ are experimental heat transfer coefficients of the more and less volatile components of the mixture in their pure states on plain and integral-fin tubes, respectively, at the same heat flux.

Now, by substituting the value of h_P from Eq.(5-12) in Eq.(5-13), a final correlation for boiling of liquid mixtures on a horizontal integral-fin tube at atmospheric pressure is obtained as follows:

$$h_F = F_s\,(h_{id})_P\,[1/\{1+\{(h_{id})_P/q\}\Delta T_{BR}\}]^{0.158} \qquad (5\text{-}15)$$

Here F_s is calculated from Eq.(5-14).

5.2.2 VERIFICATION OF DERIVED CORRELATION, EQ.(5-15)

In the present investigation, experimental data have also been generated at atmospheric pressure for nucleate pool boiling of acetone-water, isopropanol-water, and acetone-isopropanol mixtures on integral-fin tubes having fin densities of 748 and 1024 fpm. These are listed in Tables D.7, D.8, and D.9 for the 748 fpm tube and D.12, D.13, and D.14 for 1024 fpm tube. Other sets of similar data available in the literature are due to Bajorek et al.(1989). These pertain to boiling of ethanol-water and methanol-water mixtures at atmospheric pressure on a horizontal finned copper tube of fin density of 750 fpm.

To provide a comparison between experimental values of heat transfer coefficient and those predicted from Eq.(5-15), Figure 5.4 has been drawn. Experimental data are of the present investigation and those of Bajorek et al.(1989) are used for this purpose. From the plot it can be seen that there is a good agreement between the experimental and the predicted values from Eq.(5-15) within maximum error of ±22 per cent. Table 5.4 summarizes the values of standard deviations,(S.D.) and mean absolute errors, (M.A.E.) for the mixtures.

Table 5.4 Standard deviation and mean absolute error for the predictions from Eq.(5-15) of data on integral-fin tube

Investigation	Binary mixtures	Heating tube(fpm)	Composition X_1	Data points	(S.D.)%	(M.A.E.)%
Present	Acetone-Water	748	0.05-0.80	42	11.76	7.70
Present	Isopropanol-Water	748	0.05-0.90	48	9.12	8.22
Present	Acetone-Isopropanol	748	0.10-0.80	42	19.47	22.99
Present	Acetone-Water	1024	0.05-0.80	42	5.57	5.29
Present	Isopropanol-Water	1024	0.05-0.90	48	8.42	7.53
Present	Acetone-Isopropanol	1024	0.10-0.80	42	14.61	16.08
Bajorek et al.(1989)	Ethanol-Water	750	0.125-0.57	18	13.96	15.61
Bajorek et al.(1989)	Methanol-Water	750	0.05-0.85	36	16.41	17.11

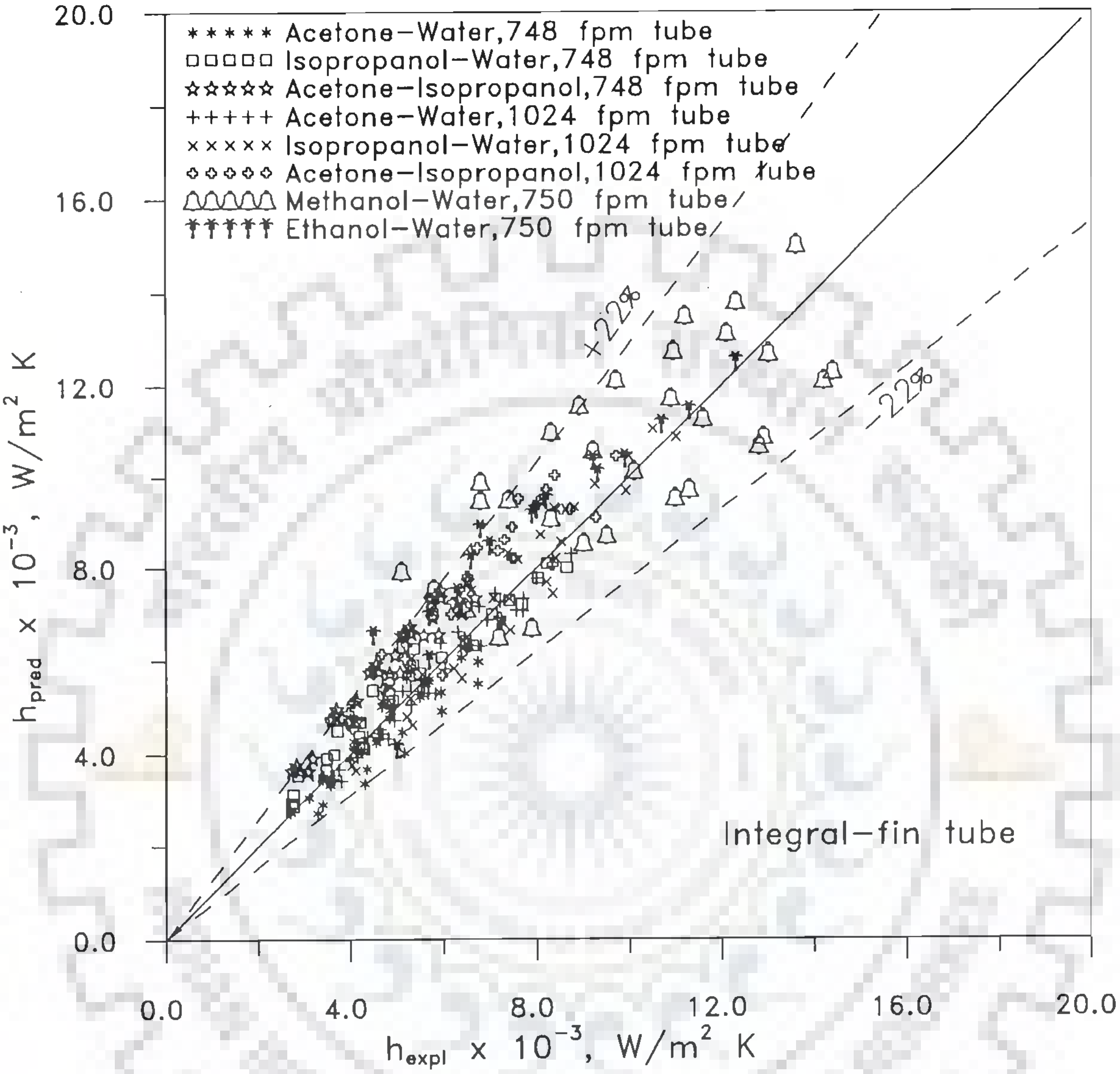


Fig. 5.4 Comparison of binary mixtures boiling data with proposed correlation, Eq.(5−15)

5.3 EXTENSION OF THE CORRELATIONS TO TERNARY MIXTURES

In sections 5.1 and 5.2, correlations, Eqs.(5-12) and (5-15) have been derived for the prediction of heat transfer coefficients for the nucleate pool boiling of binary liquid mixtures on plain and integral-fin tubes, respectively. An introspection of Eq.(5-12) for boiling of liquid mixtures on plain tube clearly reveals that it requires the knowledge of boiling range temperature, ΔT_{BR}, ideal heat transfer coefficient, $(h_{id})_P$ and heat flux, **q** for the determination of heat transfer coefficient. Intuitively, it has been thought to extend Eq.(5-12) to the boiling of ternary liquid mixtures on plain tube. In this case, the value of ΔT_{BR} for binary mixture is substituted by ΔT_{BR} for ternary mixture using its phase equilibrium diagram employing a procedure detailed in section C.2 of Appendix C. The value of $(h_{id})_P$ is evaluated from Eq.(5-6) using the physico-thermal properties of ternary liquid mixtures from Table C.9 of Appendix C.

If Eq.(5-15) is to be employed for the prediction of boiling heat transfer coefficient of ternary mixture on integral-fin tube, the values of F_s, ΔT_{BR}, $(h_{id})_P$, and q are required. Procedure for calculations of these quantities is the same as adopted in the case of binary mixture boiling on the plain tube. However, for F_s the following equation is to be used:

$$F_s = [\{X_1/(h_1)_F\} + \{X_2/(h_2)_F\}+\{X_3/(h_3)_F\}]^{-1}/ [\{X_1/(h_1)_P\} + \{X_2/(h_2)_P\}+\{X_3/(h_3)_P\}]^{-1} \qquad (5\text{-}16)$$

where $[(h_1)_P, (h_2)_P$ and $(h_3)_P]$ and $[(h_1)_F, (h_2)_F$ and $(h_3)_F]$ are the experimental heat transfer coefficients of the more, intermediate and less volatile components of the mixture in their pure states on plain and integral-fin tubes, respectively, at the same heat flux.

5.3.1 VERIFICATION OF THE CORRELATIONS FOR TERNARY MIXTURES

Experimental data related to the boiling of ternary liquid mixtures of acetone-isopropanol-water have been obtained in the present investigation and are listed in Tables D.5, D.10, and D.16 of Appendix D for plain as well as integral-fin tubes of 748 and 1024 fpm, respectively. From these data, average heat transfer coefficients are calculated, as detailed in Appendix A and the predicted values of heat transfer coefficient from Eqs.(5-12) and (5-15) for the boiling of ternary liquid mixtures on plain and integral-fin tubes, respectively. The values of F_s to be used in Eq.(5-15) are obtained from Eq.(5-16). Now, Figures 5.5 and 5.6 compare the experimental and predicted heat transfer coefficients. From the plots, it is noted that an excellent agreement exists between the experimental and predicted values within maximum error of ± 10 per cent for boiling on plain tube. Whereas, the maximum error is within + 20 and - 10 per for integral-fin tubes. The standard deviation, (S.D.) and mean absolute error, (M.A.E.) values are listed in Table 5.5.

Table 5.5 Standard deviation and mean absolute error for the predictions from Eqs.(5-12) and (5-15) of present data for ternary mixtures boiling

Ternary mixture	Heating tube	Data points	(S.D.)%	(M.A.E.)%
Acetone-Isopropanol-Water	Plain	60	7.12	5.83
Acetone-Isopropanol-Water	748 fpm	50	10.35	10.04
Acetone-Isopropanol-Water	1024 fpm	50	15.11	16.02

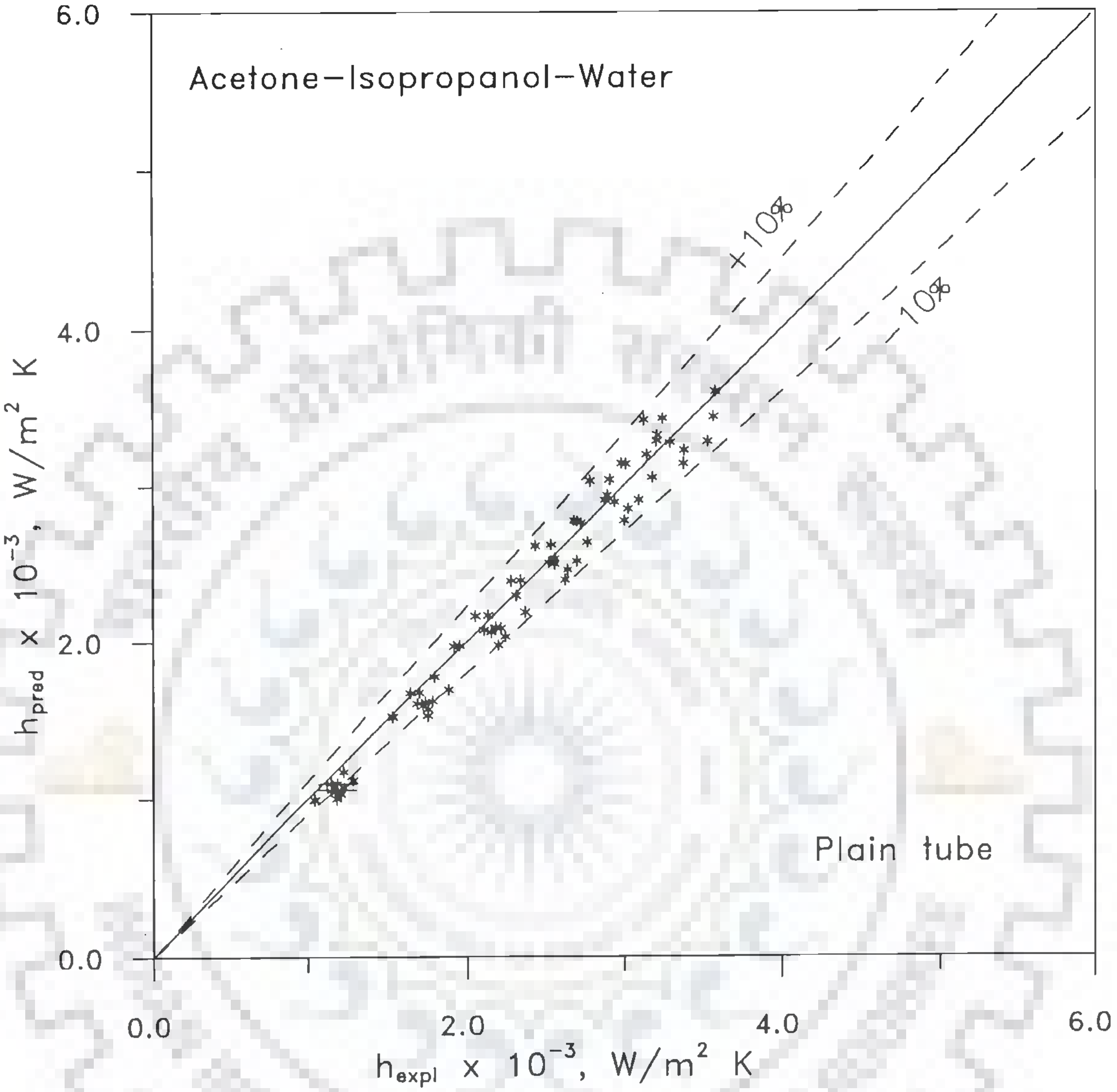


Fig. 5.5 Comparison of ternary boiling data with proposed correlation, Eq.(5–12)

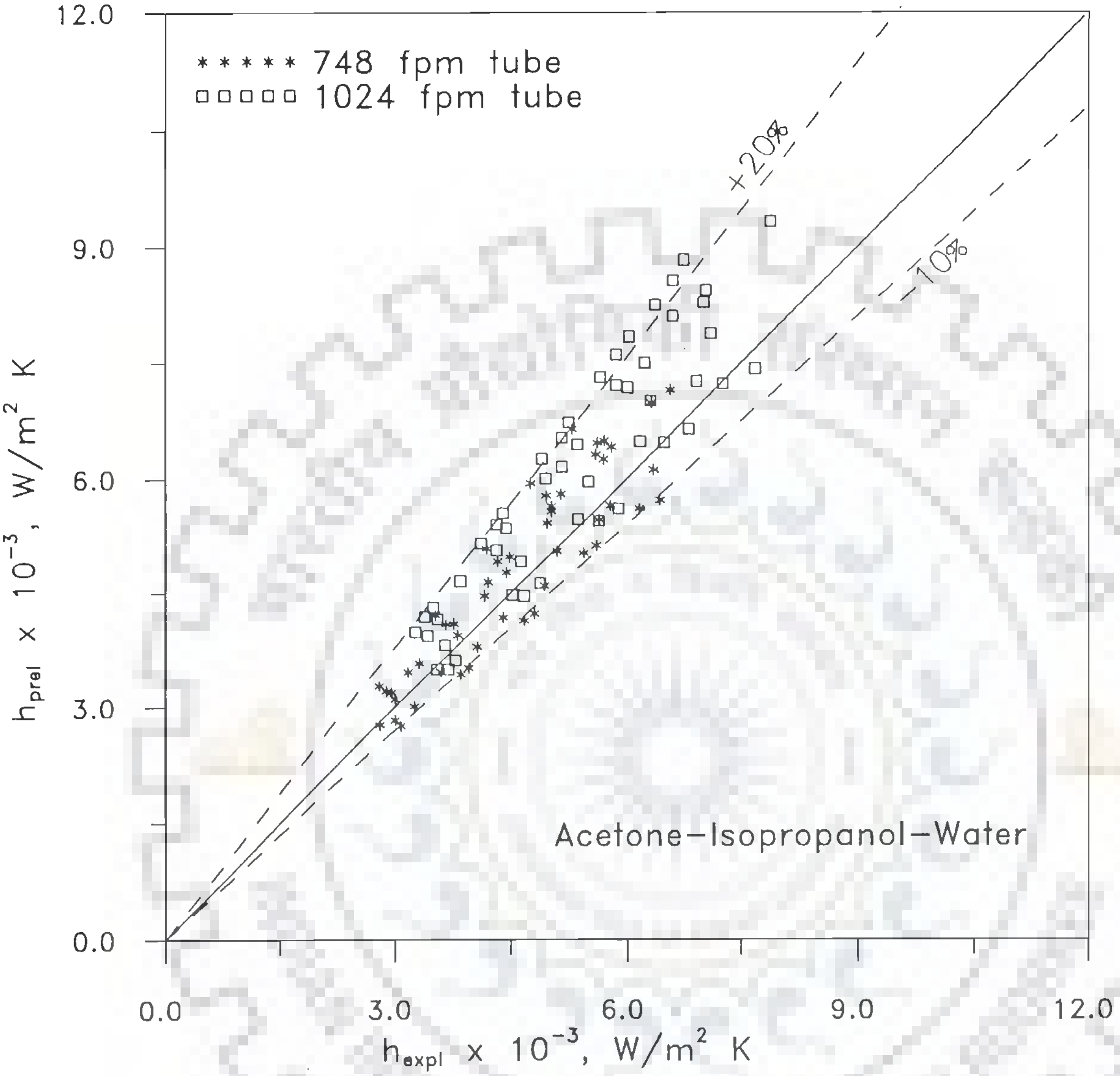


Fig. 5.6 Comparison of ternary boiling data with proposed correlation, Eq.(5-15)

From these values, it is concluded that Eqs.(5-12) and (5-15) are equally applicable to ternary liquid mixtures boiling on plain and integral-fin tubes, respectively. However, it is necessary that experimental data related to boiling of ternary mixtures other than acetone-isopropanol-water mixtures should be generated to extend the general applicability of the correlations further.

5.3.2 VERIFICATION OF SURFACE CORRECTION FACTOR, F_s

In section 5.2.1, the value of F_s is determined by Eq.(5-14) for evaluation of heat transfer coefficient, h_F for boiling of binary liquid mixtures on integral-fin tube at atmospheric pressure. It is necessary to test the validity of this equation. For this, it was thought desirable to compare the predicted values of F_s from Eq.(5-14) with the experimentally-determined values. The latter are calculated from the experimental values of h_F and h_P using Eq.(5-13).

Figure 5.7 has been drawn between $(F_s)_{pred}$ and $(F_s)_{expl}$ for the boiling of acetone-water, isopropanol-water, and acetone-isopropanol mixtures on integral-fin tubes having fin density of 748 and 1024 fpm. From the plot, it is clearly noted that there is a good agreement between the two sets of values. The standard deviation (S.D.) and mean absolute error values are 18.2 and 14.73 respectively.

5.4 COMMENTS ON THE PROPOSED CORRELATIONS

The derived correlations, Eqs.(5-12) and (5-15) for the prediction of heat transfer coefficients for the boiling of binary liquid mixtures on plain and integral-fin tubes, respectively fall under the category of semi-empirical correlations. These correlations possess superiority over others as they are not as complex in their functional forms as others. Further, they require phase equilibrium data only and do not possess coefficients and quantities whose determination is difficult, as is encountered when employing the Schlunder correlation. Besides, values predicted from them are also reasonably accurate and acceptable. However, it is important to underline these correlations are applicable for low heat fluxes.

It is also important to mention that Eq.(5-12) is capable to predict boiling heat transfer coefficients for plain tube for boiling of ternary liquid mixtures having wide ranges of compositions, widely differing physico-thermal properties, azeotrope and non-azeotrope mixtures.

As regards correlation, Eq.(5-15), it is also equally applicable for the boiling of ternary liquid mixtures on integral-fin tubes at atmospheric pressure with the values of F_s evaluated from Eq. (5-16).

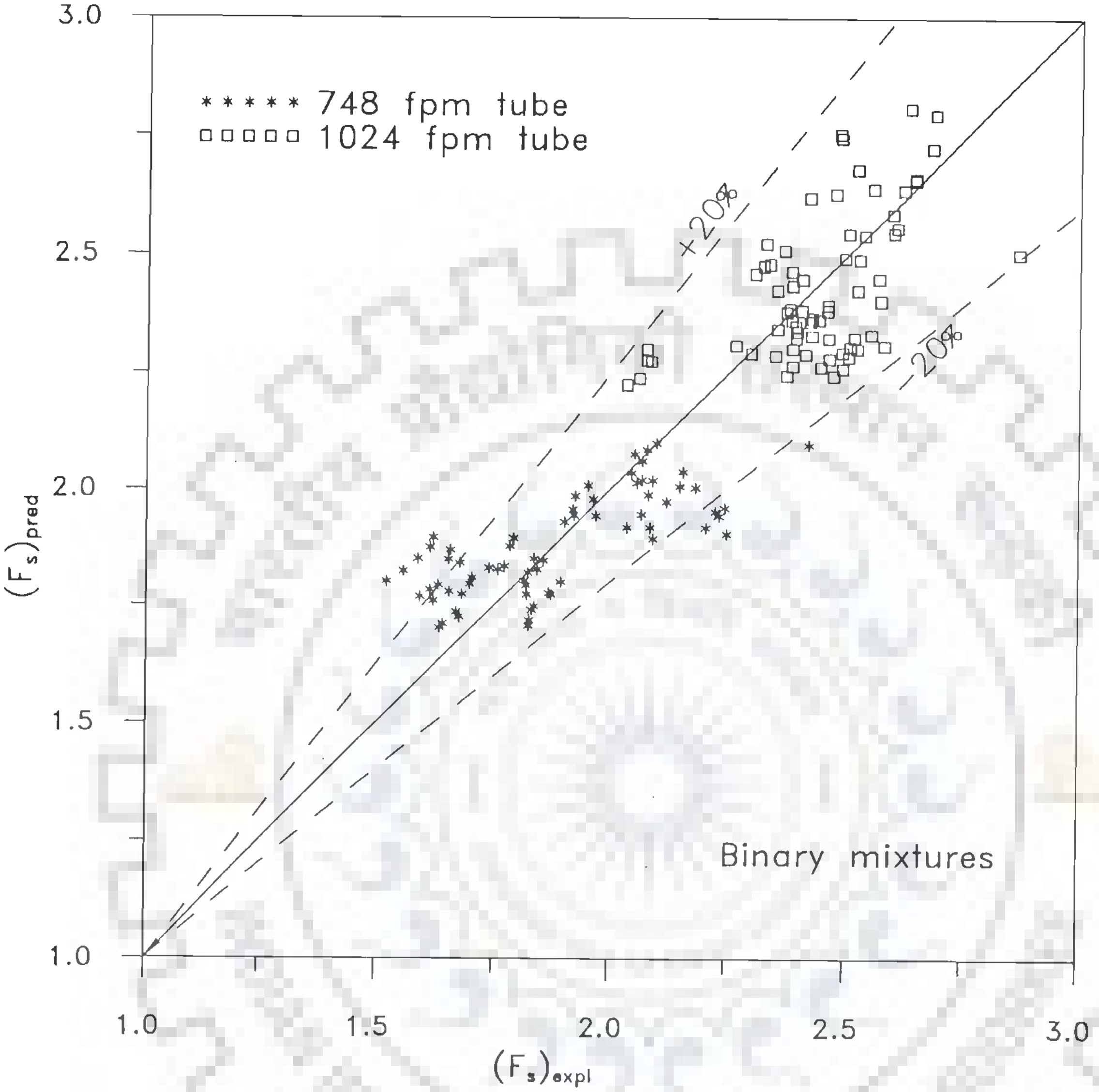


Fig. 5.7 Comparison of surface correction factor for integral-fin tubes

CHAPTER 6

RESULTS AND DISCUSSION

This Chapter describes the salient results emerging out from the analysis of experimental data of present investigation and also those of earlier investigations along with their interpretations.

6.1 NUCLEATE POOL BOILING OF SINGLE COMPONENT LIQUIDS ON PLAIN TUBE

Experimental data related to nucleate pool boiling heat transfer from a horizontal plain tube to single component liquids have been listed in Table D.1 of Appendix D. As mentioned earlier in section 4.2 of Chapter 4, the last column of the table lists the values of average heat transfer coefficient, the procedure for determination of these values is detailed in Appendix A. It was felt necessary first to check the reliability of these data. This is carried out in the following sections as discussed below.

6.1.1 VERIFICATION OF PRESENT EXPERIMENTAL DATA

It is important to mention that earlier investigators have also obtained experimental data related to nucleate pool boiling of pure liquids on a horizontal plain tube at atmospheric pressure. For verifying the reliability of a given set of experimental data, it is generally preferred to compare them against the already published data with average heat transfer coefficient as ordinate and heat flux as abscissa on log-log scale. Accordingly, Figure 6.1 has been drawn to represent experimental data of boiling of water on a horizontal plain tube at atmospheric pressure.

As regards experimental data for the boiling of water, earlier investigators have also published their data. Some of the representative data are due to Cryder and Finalborgo (1937), Charter et al.(1982), Jamialahmadi et al.(1991), and Kumar (1992). They have also been plotted in Figure 6.1. From the plot, it is evidently clear that individual experimental data are represented by separate parallel straight lines, revealing that h varies with q linearly. In other words, the effect of q on h is the same, irrespective of investigations. Mathematically, h is related to q by the following relationship:

$$h = C_{sf}\, q^{0.68} \qquad (6\text{-}1)$$

Since experimental data of individual investigators are for ' nucleate pool boiling ' region of boiling curve, they are expected to follow the same relationship as given in Eq.(6-1) . It is pertinent to point out that for a given heat flux the value of h differs from investigation to investigation. This is obviously due to marked difference in the values of ' C_{sf} ' of Eq.(6-1). In fact, C_{sf} represents ' surface-liquid combination factor '. This factor depends upon heating surface characteristics and the nature of boiling liquid in its contact. It is well-documented by Rohsenow (1952), that C_{sf} has a significant effect on boiling heat transfer coefficient. Hence, for a given boiling liquid, surface characteristics of heating tube should have a noticeable effect on nucleate pool boiling heat transfer coefficient. As regards

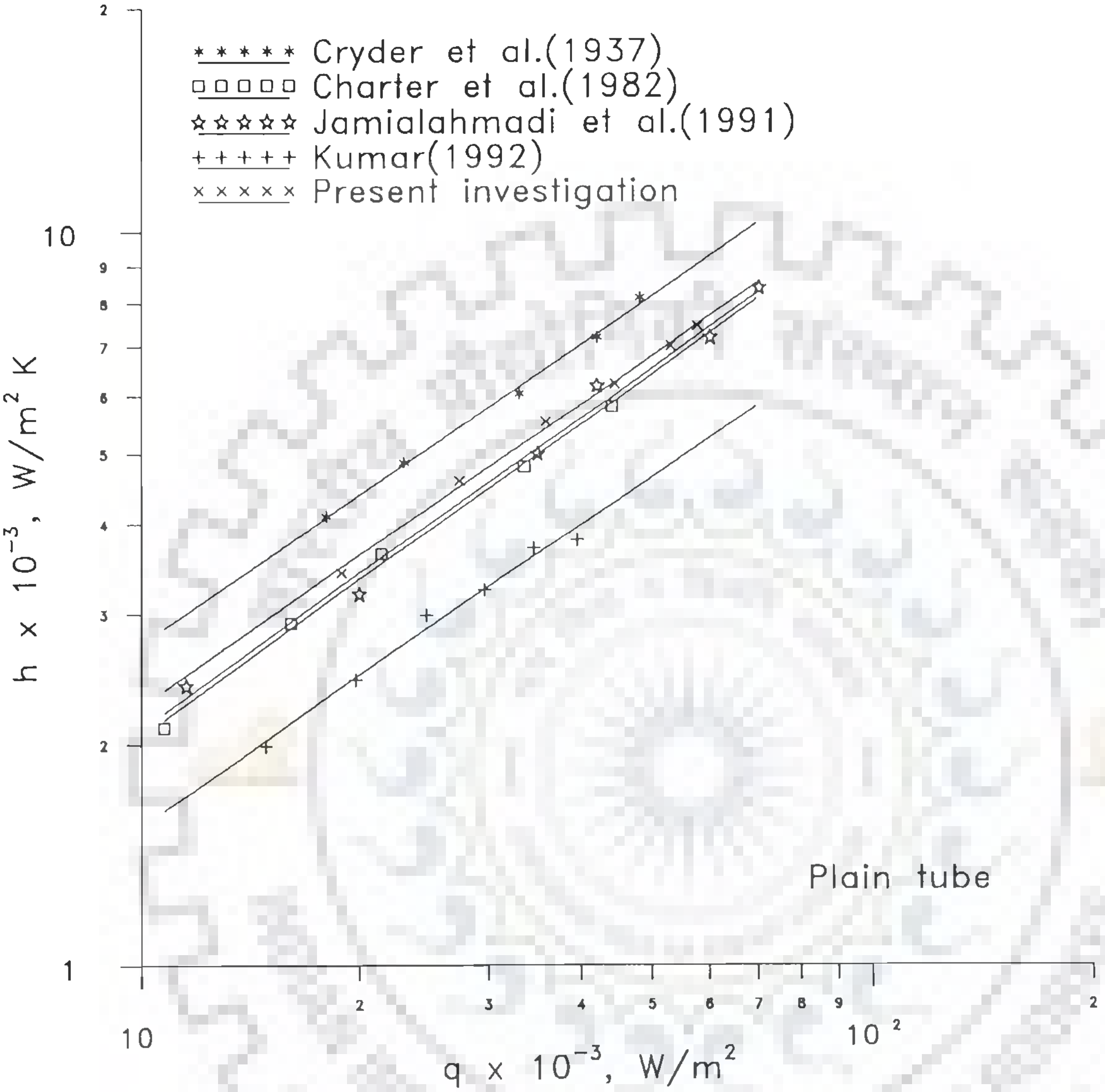


Fig. 6.1 Comparison of boiling heat transfer coefficient of water with earlier investigators

the surface, it is different for different investigators as they employed separate heating tubes whose surface characteristics are, of course, different. Therefore, the differing values of C_{sf} are logical and expected.

From the above discussion, it is implied that experimental data of present investigation are also for ' nucleate pool boiling ' region, as they show the same trend as those of earlier investions which have been conducted in nucleate pool boiling region.

Figures 6.2 and 6.3 represent the same plot as that of Figure 6.1, but for nucleate pool boiling of acetone and isopropanol, respectively at atmospheric pressure. Experimental data due to Uhlig and Thome (1985) and Bajorek et al.(1989) for the boiling of acetone and those due to Sharma (1977), Yilmaz and Westwater (1981) and Kumar (1992) for the boiling of isopropanol have also been plotted in these figures. The plots in individual figures possess the same characteristic features as regards the dependence of h on q, i.e., $h = C_{sf}\, q^{0.68}$. The value of C_{sf} differs from investigation to investigation for the reason mentioned above in relation to Eq.(6-1). Hence, it can be concluded that present experimental data related to nucleate pool boiling of acetone and isopropanol are also reliable.

Another method to verify the reliability of experimental data is to compare them with the predictions from well-known and widely-accepted generalized correlations. This has been carried out in the following section.

6.1.2 COMPARISON OF PRESENT EXPERIMENTAL DATA WITH EXISTING CORRELATIONS

Table 2.4 of Chapter 2 describes some of the important correlations which are available in open literature for predicting boiling heat transfer coefficients. Till the correlations due to Stephan and Abdelsalam (1980) and Cooper (1984), it was well-considered conclusion that none of the correlations has been known possessing the general applicability. Hence, a design engineers have been confronting with the problem at which of the correlations they should employ for the calculation of heat transfer coefficient to boiling liquid side of a reboiler. In an attempt to solve this problem, Stephan and Abdelsalam (1980) have recommended correlations for the prediction of boiling heat transfer coefficient for boiling of liquids on a plain tube, using regression analysis of more than 5000 data points collected from 72 research papers. Therefore, it was decided to compare the experimental data with the predictions from these correlations. For ready reference the correlations due to Stephan and Abdelsalam for hydrocarbons and water are reproduced as follows:

a. For hydrocarbons

$$h = 0.0546\,(k_l/d)[(qd/k_lT_s)(\rho_v/\rho_l)^{0.5}]^{0.67}\{(\rho_l-\rho_v)/\rho_l\}^{-4.33}\,(h_{fg}\,d^2/\alpha^2)^{0.248} \qquad (6\text{-}2)$$

b. For water

$$h = 0.246 \times 10^7\,(k_l/d)\,[(qd/k_lT_s)]^{0.673}\,(Cp\,T_s\,d^2/\alpha^2)^{1.26}\,(h_{fg}\,d^2/\alpha^2)^{-1.58}[(\rho_l-\rho_v)/\rho_l]^{5.22} \qquad (6\text{-}3)$$

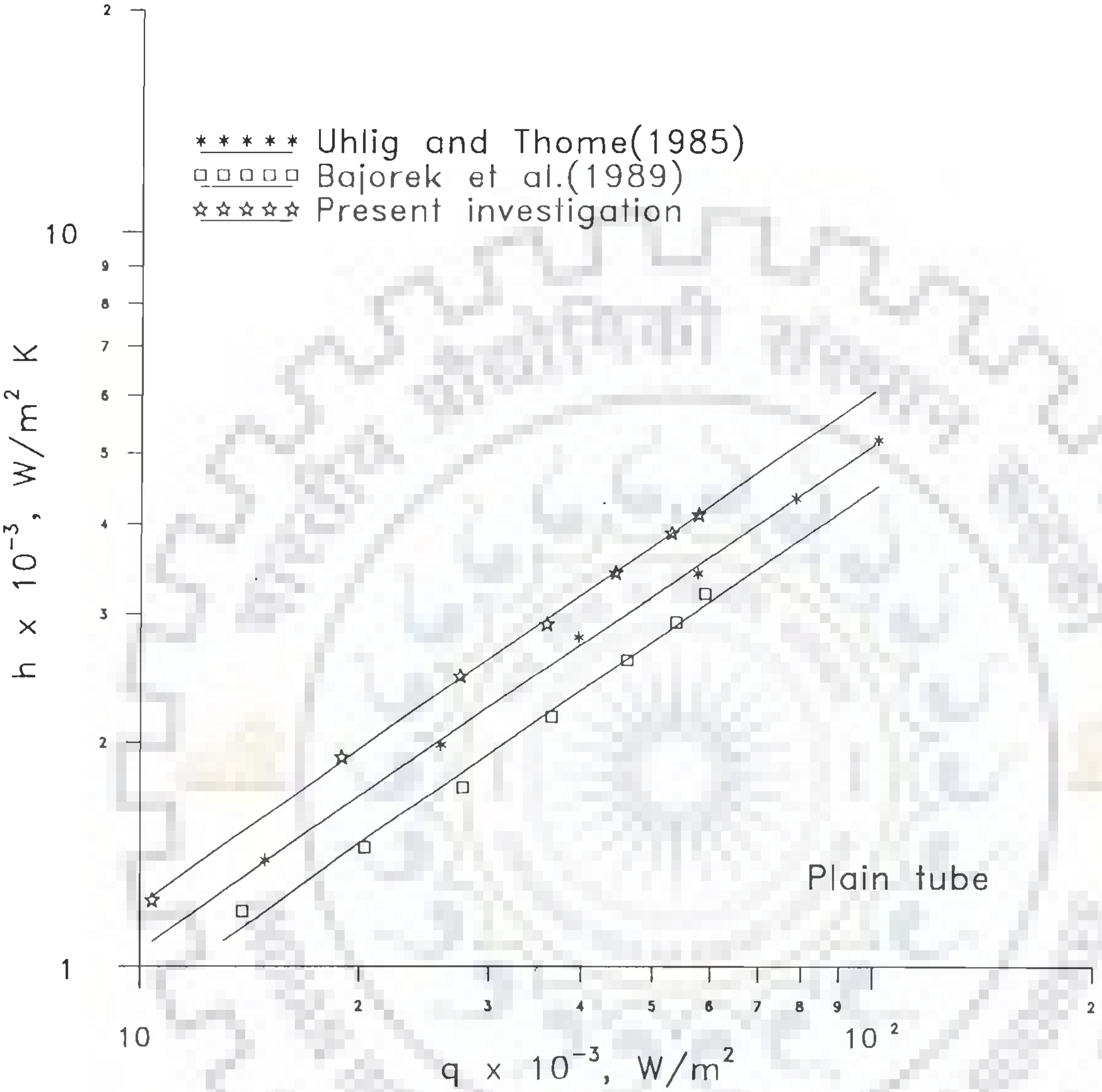


Fig. 6.2 Comparison of boiling heat transfer coefficient of acetone with earlier investigators

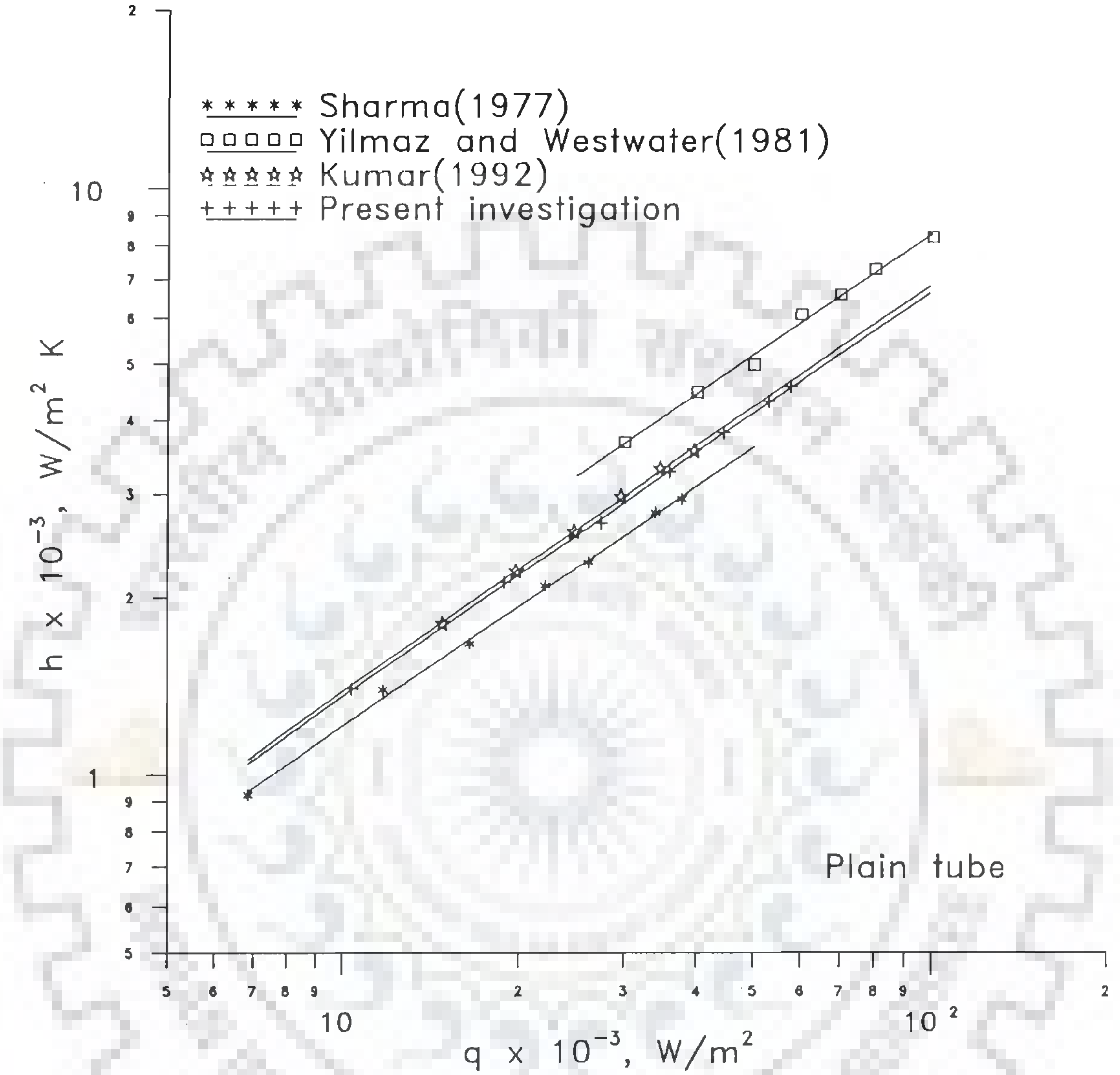


Fig. 6.3 Comparison of boiling heat transfer coefficient of isopropanol with earlier investigators

where d is bubble departure diameter and is calculated by the following equation:

$$d = 0.0146\,\beta\,[2\sigma/\{g(\rho_l-\rho_v)\}]^{0.5} \qquad (6\text{-}4)$$

where β is contact angle being 35° for organic liquids, whereas its value is 45° for water.

Figure 6.4 has been drawn between experimental heat transfer coefficients from Table D.1 of Appendix D and predicted values from Eqs.(6-2) and (6-3). An examination of Figure 6.4 shows that an excellent agreement between experimental and predicted values for the boiling of isopropanol is noted. However, for other remaining liquids, i.e., acetone, distilled water and isopropanol-water (Az.), there is also reasonably good agreement between experimental and predicted values of heat transfer coefficient, indicating that the present data compare well with Eqs.(6-2) and (6-3). An implication of this would be that the present experimental data are reliable.

As mentioned earlier in section 2.5 of Chapter 2, that the correlation proposed by Cooper (1984) is one which is also claimed to be of general applicability. Accordingly, average heat transfer coefficients were calculated for the boiling of acetone, isopropanol, and water on plain tube at atmospheric pressure using Cooper's correlation. It is a simple correlation for predicting nucleate pool boiling heat transfer coefficient of saturated liquids as given below:

$$h = C\,[(p/p_c)^{0.12-0.2\log_{10}R_p}](-\log(p/p_c))^{-0.55}\;MW^{-0.5}\,(q^{0.67}) \qquad (6\text{-}5)$$

where, C = 95 for water, and C = 55 for other liquids.

In Eq.(6-5), R_p is average surface roughness in mm and MW is the liquid molecular weight. Cooper applied the above equation to a large number of experimental data and found that it correlated the data well. Recently, Webb et al.(1989) and Webb and Pais (1992) have recommended the Cooper's correlation for the prediction of boiling heat transfer coefficients of pure liquids with the values of R_p varying from 0.3 to 0.6 μm. The best prediction has been obtained for the present data with R_p equal 0.5 μm. It is important to remember that the Cooper's correlation is not applicable for the boiling of azeotropic mixtures.

The predictions from Eq.(6-5) have been compared against experimental values of average heat transfer coefficient from Table D.1 of Appendix D in Figure 6.5. From the plot, it is clear that there is an excellent agreement for the data of acetone. As regards data of water and isopropanol, the correlation overpredicts and underpredicts them respectively, but quite marginally.

Table 6.1 summarizes the standard deviations and mean absolute errors for each of the above correlations, employed for the verification of present experimental data.

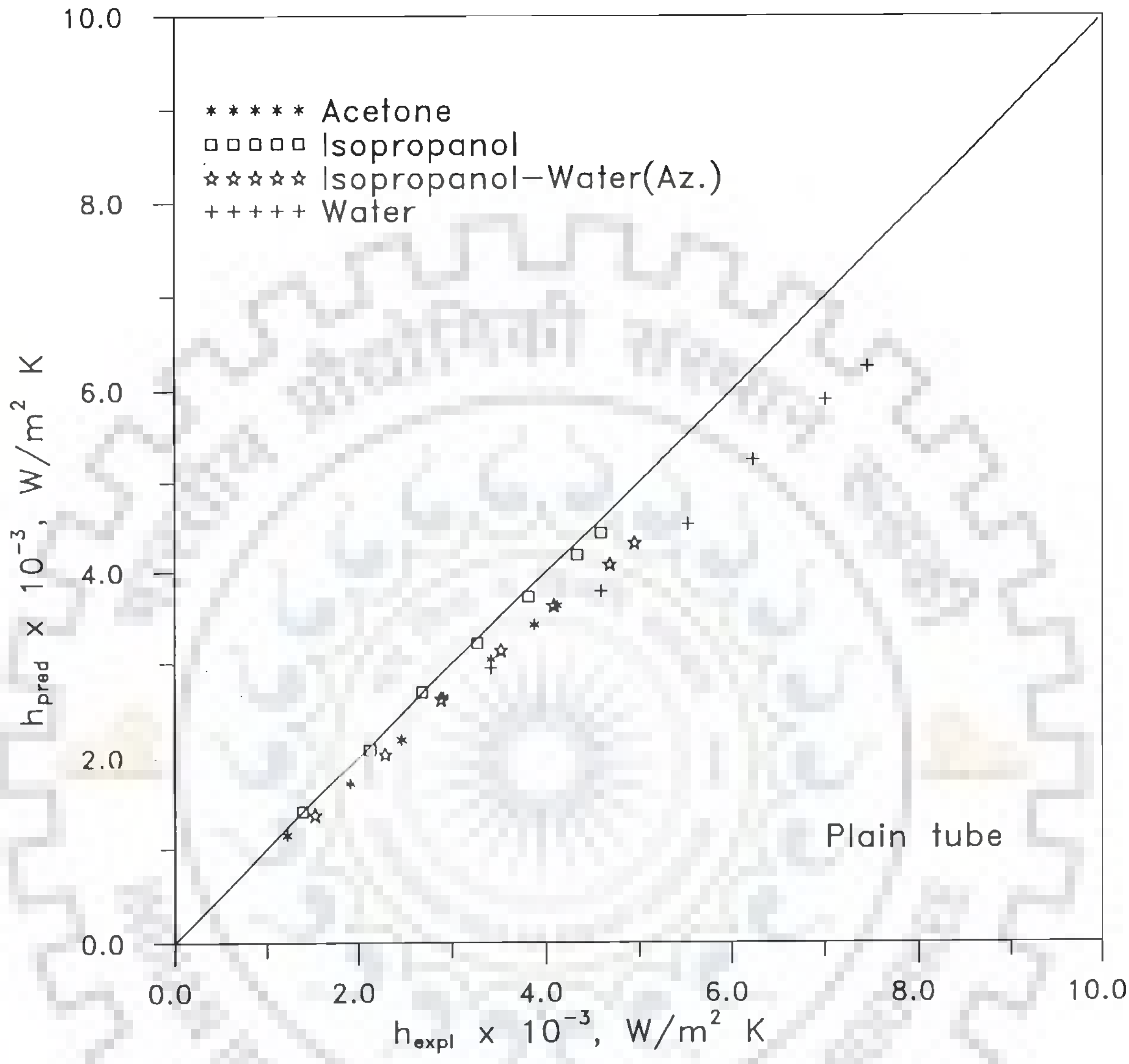


Fig. 6.4 Comparison with the Stephan−Abdelsalam correaltion

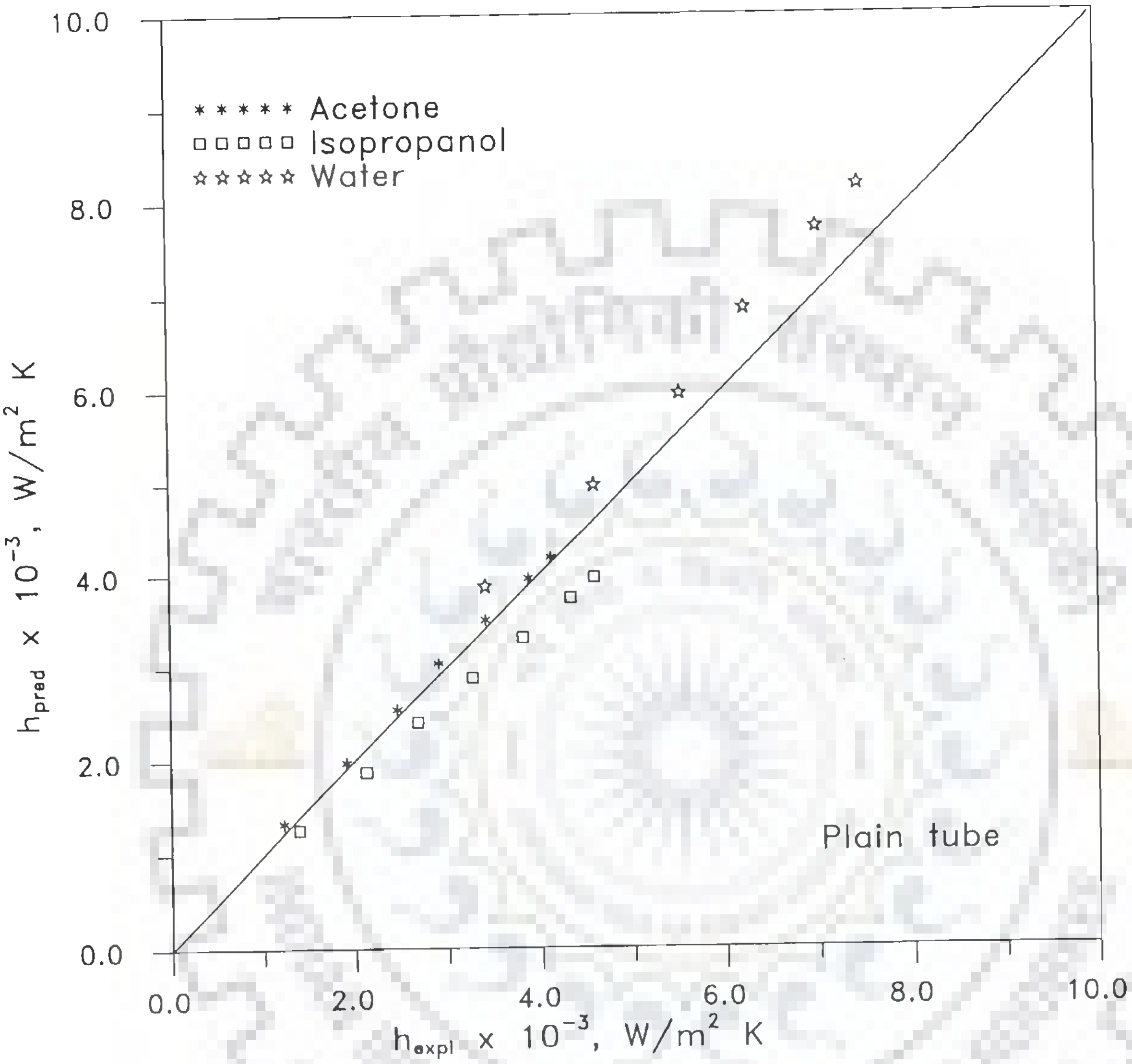


Fig. 6.5 Comparison with the Cooper correlation

Table 6.1 Standard deviation and mean absolute error of various correlations for the prediction of heat transfer coefficient for boiling of pure components on a plain tube

Investigator(s)	Correlation	(S.D.)%	(M.A.E.)%
Stephan-Abdelsalam(1980)	Eqs.(6-2)and (6-3)	12.11	9.10
Cooper (1984)	Eq.(6-5)	9.94	8.72

From Table 6.1 it is noted that the above correlations correlate the present data excellently. However, the Cooper's correlation, Eq.(6-5) predicts better than the Stephan and Abdelsalam correlations, Eqs.(6-2) and (6-3). An indirect implication of the results of Table 6.1 is that present experimental set-up provides reliable data and hence experimental data for mixtures boiling can considered to be reliable.

6.1.3 NUCLEATE POOL BOILING CURVE

Nucleate pool boiling curve has a unique significance as regards boiling heat transfer. In fact, this represents experimental data of heat transfer in terms of **h vs. q** or **q vs. ΔT** plots on log-log scale covering free convection, nucleate pool boiling, transition, and film boiling regions. Since experimental data of present investigation pertain to ‘ nucleate pool boiling ‘ region. Therefore, the plot of experimental data of present investigation between h and q represents the boiling curve for this region.

Figure 6.6 represents nucleate pool boiling curve for the boiling of acetone, isopropanol, isopropanol-water (Az.) and water on plain tube at atmospheric pressure. The plot is on log-log scale. In fact, it is a part of boiling curve obtained by Nukiyama (1934). An inspection of the plot of Figure 6.6 reveals that ‘ boiling curves ‘ of different liquids possess the same characteristic features. Mathematically, h is related to q by the following functional relationship:

$$h = C_1 \, q^{0.68} \qquad (6\text{-}6)$$

where C_1 is a constant.

Table 6.2 Values of constant C_1 in Eq.(6-6)

Liquid	Acetone	Isopropanol	Isopropanol-Water(Az.)	Water
C_1	0.240	0.275	0.375	0.457

It is important to reason out as what makes the values of constant, C_1 differ from liquid to liquid. As a matter of fact, quantity C_1 represents ‘surface liquid combination factor’. For a given heating

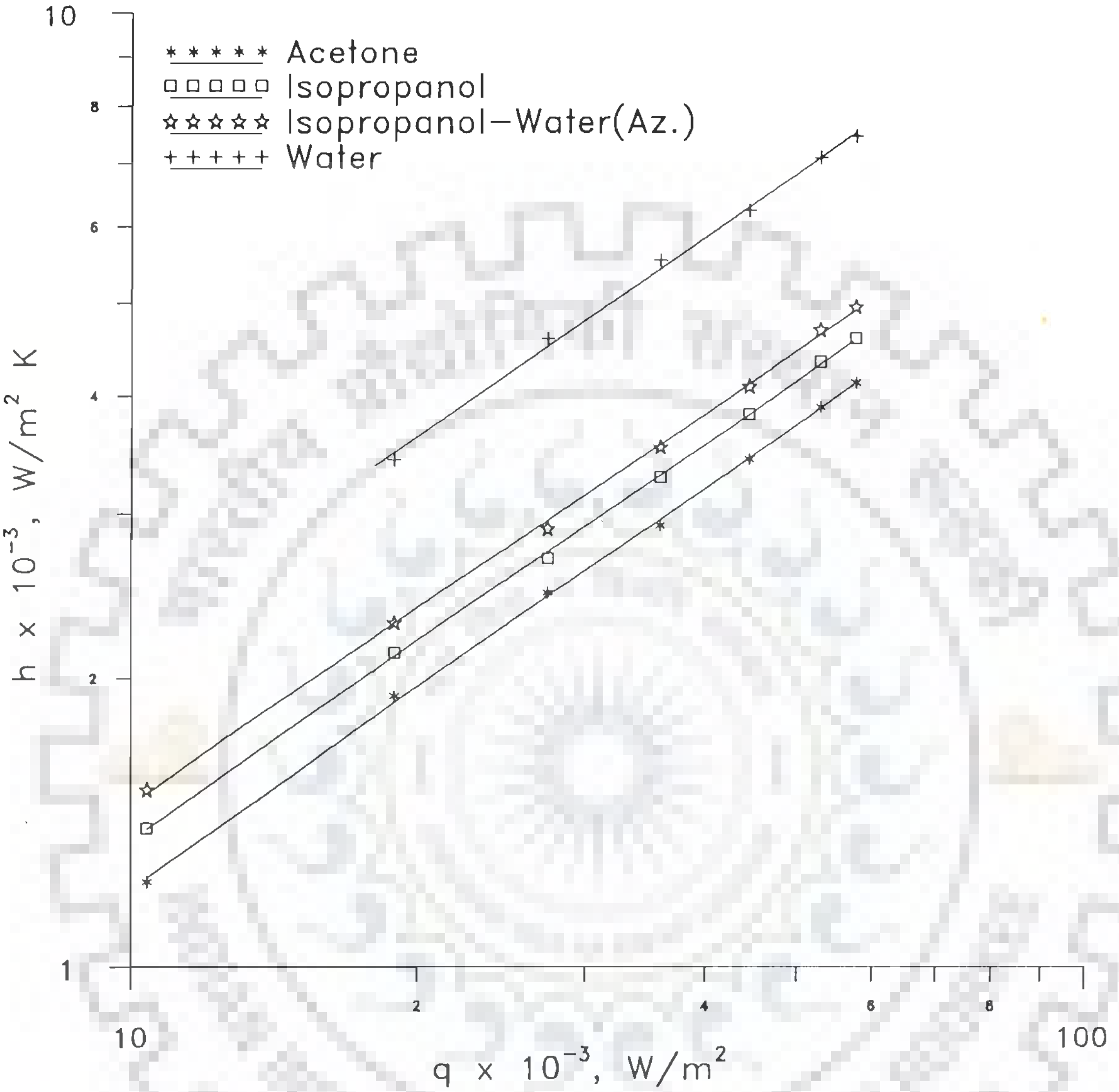


Fig. 6.6 Nucleate boiling curves for pure components on a plain tube

tube, it is an established fact that it depends upon the nature of boiling liquids. Due to this reason, the values of C_1 are found to differ from liquid to liquid as in Table 6.2. The value of **h** increases regularly with **q**. This is understandable from the fact that with a rise in **q** the number of nucleation sites for the birth of vapour bubbles on heating tube increases. Thus, the number of vapour bubbles per unit area increases. These bubbles, in turn, cause induced turbulence in boiling liquid to rise and consequently boiling heat transfer coefficient increases. This, in fact, is also in accordance with the correlation [$\mathbf{h} = (\mathbf{nf})^{13}$] as mentioned in Table 2.1, i.e **h** is proportional to [$\mathbf{n}^{1/3}$] where n represents the number of nucleation sites.

6.2 NUCLEATE POOL BOILING OF BINARY LIQUID MIXTURES ON PLAIN TUBE

The literature survey of Chapter 2 has clearly shown that the pool boiling of binary liquid mixtures is more complex than that for single component liquids. In fact, the mechanism involved in the former case is significantly different. The distinguishing feature lies in bubble growth rate. For boiling of single component liquids, it is governed only by heat diffusion from the heating surface towards bubble boundary to satisfy the heat requirements of a growing bubble. But for binary liquid mixtures, heat diffusion is linked with mass diffusion of the components of a mixture from the 'region' surrounding the bubbles. In fact, the more volatile component evaporates at a faster rate. Due to this reason, it naturally gets exhausted more rapidly near the interface than the less volatile component. This region is just adjacent to the heating surface and is termed as superheated liquid layer. Consequently, a lower concentration of the more volatile component results there-in than in the bulk liquid and thus bubble growth rate slows down. In short, bubble growth rate is affected by the composition of bulk liquid mixture in addition to heat flux and surface characteristics of heating tube.

At this stage, it is worth-mentioning that higher is the bubble growth rate the greater is the boiling heat transfer coefficient. In the following sections, the effects of the heat flux, the heating surface characteristics and the bulk liquid composition on heat transfer coefficient of binary liquid mixtures are discussed.

6.2.1 EFFECT OF HEAT FLUX ON HEAT TRANSFER COEFFICIENT

Figure 6.7 shows the effect of heat flux on heat transfer coefficient for the boiling of acetone-water on plain tube at atmospheric pressure with mixture composition as parameter. As a matter of fact, this figure represents nucleate boiling curves on log-log scale for the boiling of binary liquid mixture similar to Figure 6.6, which is for the boiling of single component liquids. From the plot, it is clearly noted that heat transfer coefficient increases with heat flux linearly for the entire range of the heat flux possessing the following functional relationship:

$$h = C_2 \, q^{0.66} \qquad (6\text{-}7)$$

where C_2 is a constant, whose values depend upon the composition of boiling liquid mixture, irrespective of heat flux. In fact, this constant represents ' surface-liquid combination factor 'and therefore its values are expected to vary with mixture composition for the reason explained in section 6.1.3. From

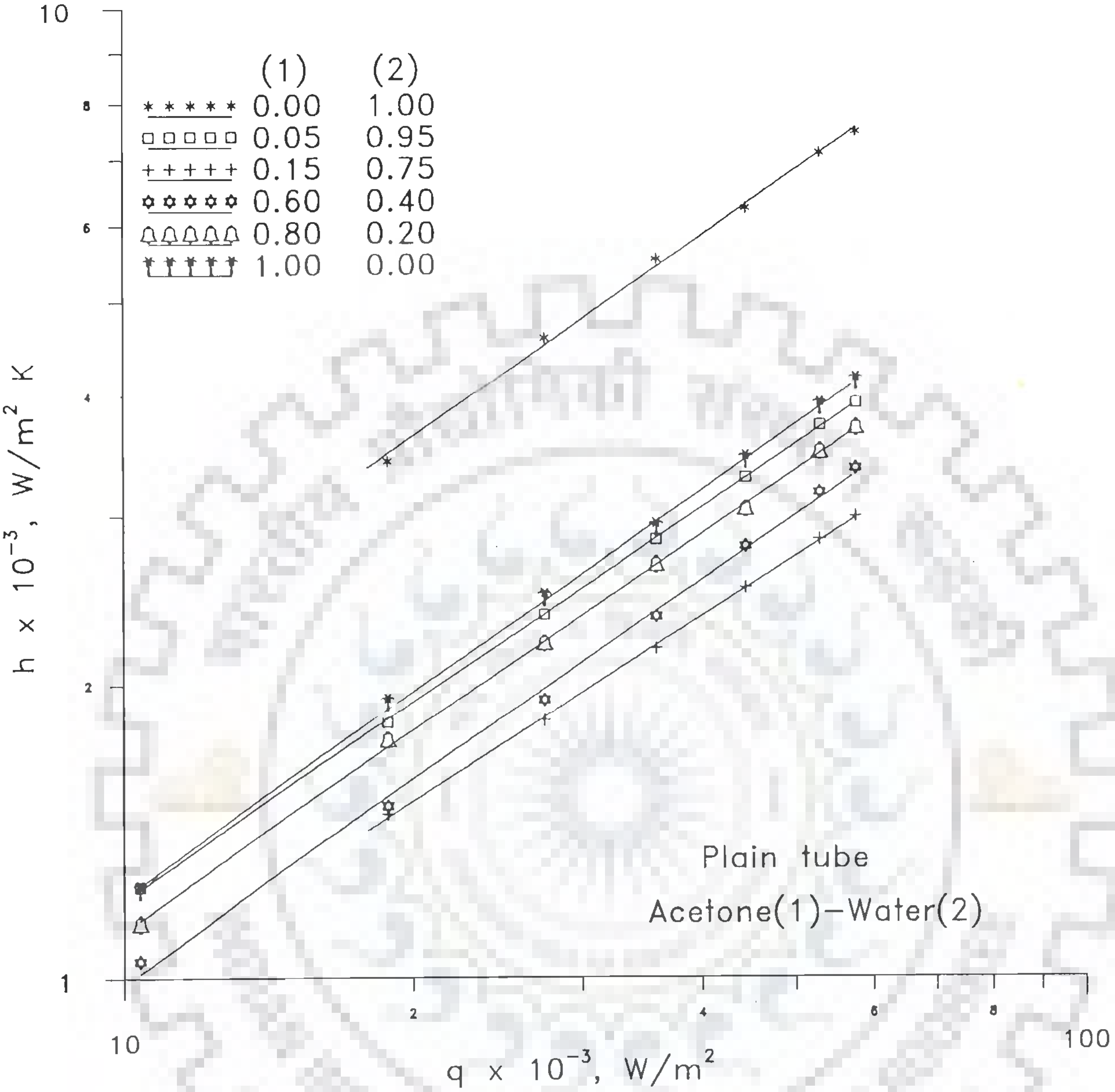


Fig. 6.7 Nucleate boiling curves for acetone–water mixture on a plain tube

Eq.(6-7), it is noted that with the increase in the value of heat flux heat transfer coefficient increases for a particular composition of mixture. This, indeed, is due to the increase in number of active nucleation sites per unit area, which has been well-explained in section 6.1.3 in the case of boiling of single component liquids on plain tube.

Still one more worth-noting observation from Figure 6.7 is that the experimental data for all the mixtures irrespective of their compositions lie below those of the components of binary liquid mixture in their pure states, i.e water and acetone.

Experimental data for the boiling of isopropanol-water mixtures are shown in Figure 6.8, which depict the similar trend as exhibited in Figure 6.7, except data for the mixtures having compositions of 0.68 and 0.8 mole fraction of isopropanol.

Experimental data related to the boiling of organic mixtures containing acetone and isopropanol are shown in Figure 6.9. They are not as much spread over as in Figures 6.7 and 6.8 mentioned above. This differing behaviour seems to be due to the fact that both the components of acetone-isopropanol mixture are of organic origin, whereas in the case of other mixture systems one of the components is water.

6.2.2 EFFECT OF HEATING SURFACE CHARACTERISTICS ON HEAT TRANSFER COEFFICIENT

To demonstrate this, the typical experimental data of Alam (1972), Bajorek et al.(1989), and the present investigation for the boiling of 0.4 mole fraction of acetone in acetone-water liquid mixture at atmospheric pressure on horizontal heating tubes have been employed. These data are for plain tubes, but of differing surface characteristics. Based on these data, a plot representing boiling heat transfer coefficient as ordinate and heat flux as abscissa on log-log scale is plotted in Figure 6.10. It clearly shows that the individual data are represented by separate straight lines but parallel. In fact, these are represented by Eq.(6-7) having different values of C_2 depending upon the investigation. As expected, the values of C_2 should be different. The reason for this is attributed to the fact that these investigators employed heating tubes of differing surface characteristics. This fact is well-supported by the findings of Stephan and Preusser (1978). In short, it is concluded that for a given mixture composition the constant C_2 of Eq.(6-7) depends on the heating surface characteristics, irrespective of heat flux.

After having observed that the mixture composition affects the values of constant C_2 and finally the heat transfer coefficient, it was thought important to find out the behavioural change of the latter with mixture composition. This is discussed in the following section.

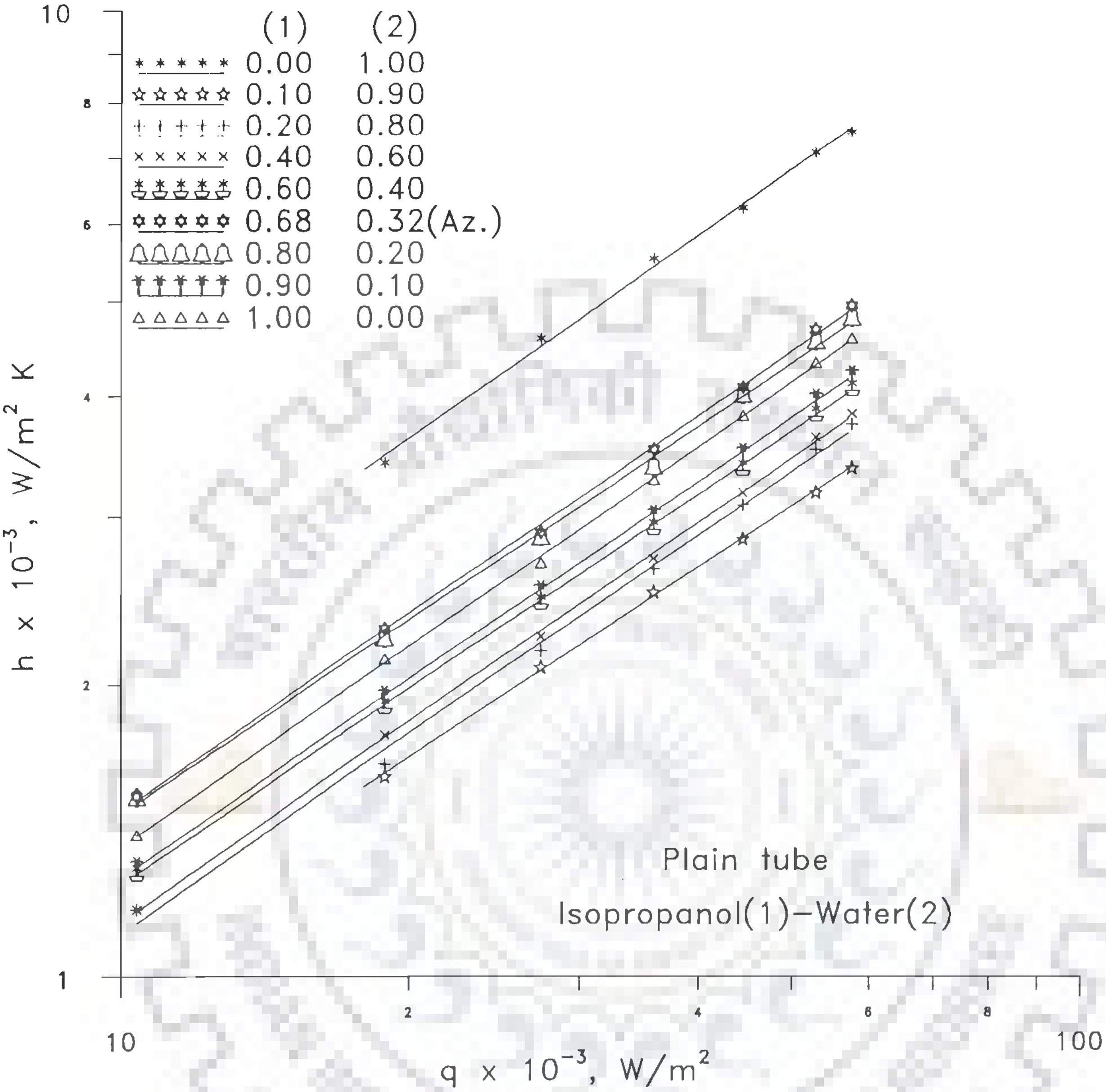


Fig. 6.8 Nucleate boiling curves for isopropanol–water mixture on a plain tube

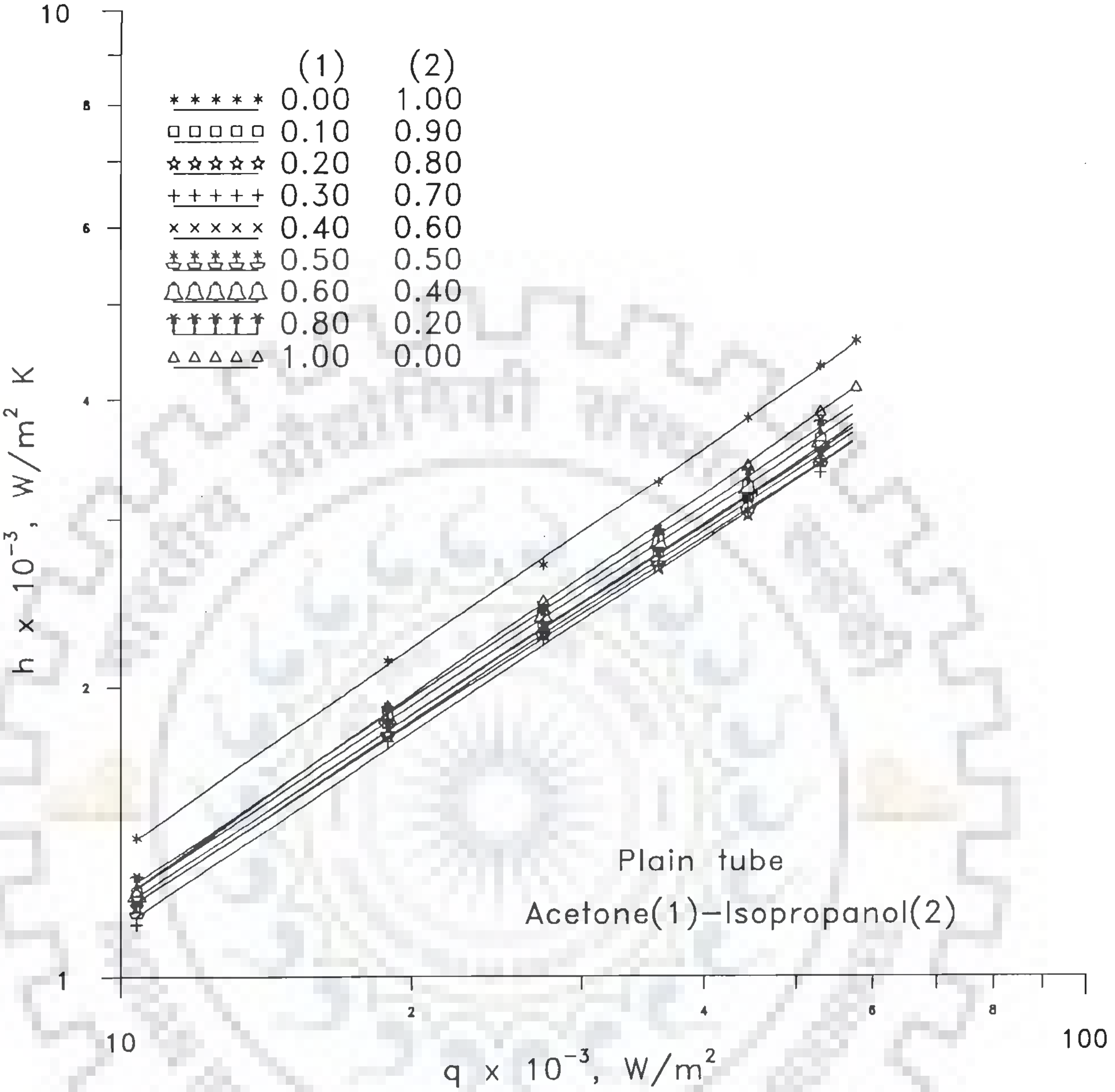


Fig. 6.9 Nucleate boiling curves for acetone-isopropanol mixture on a plain tube

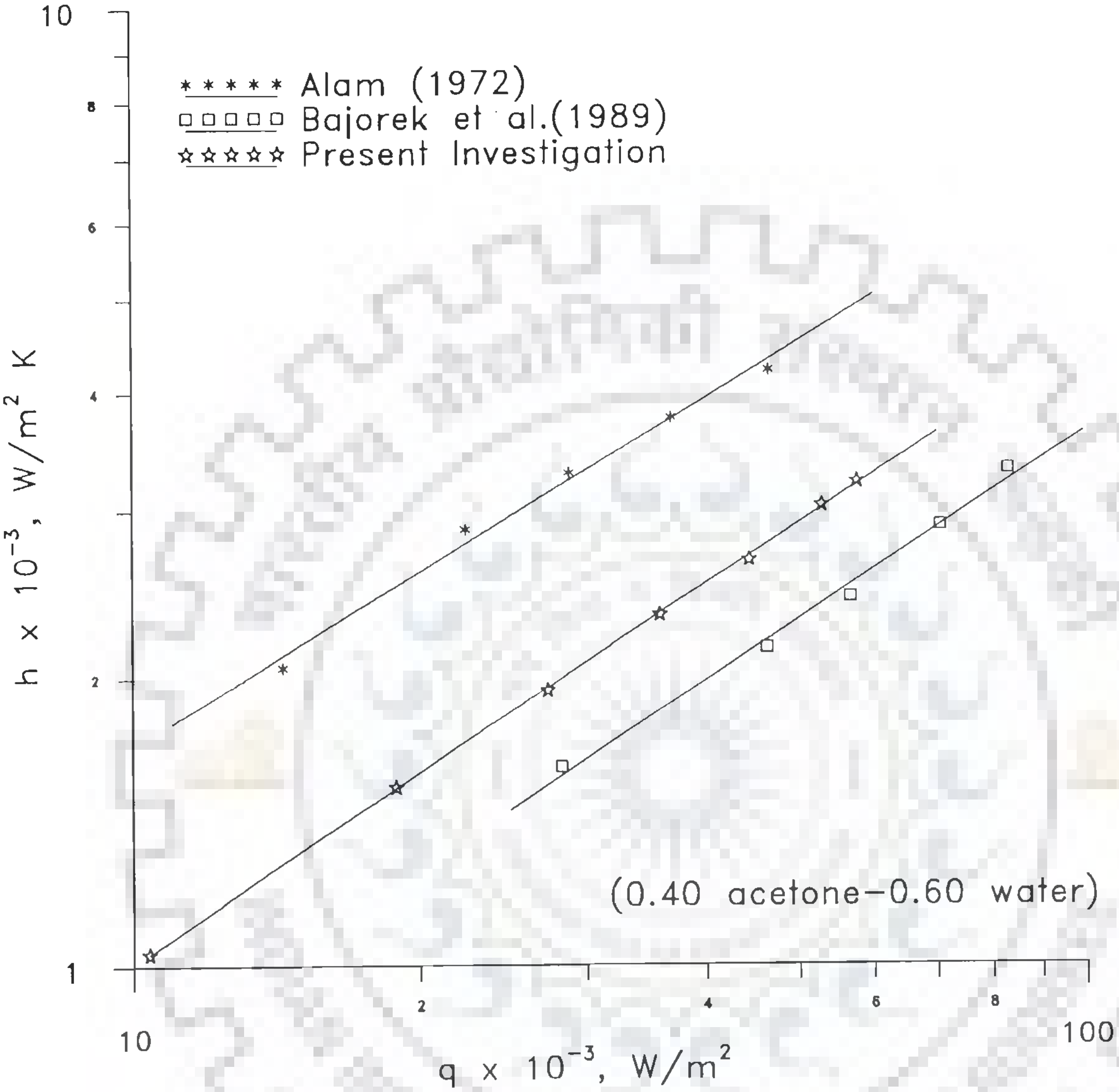


Fig. 6.10 Comparison of heat transfer coefficient for mixture with earlier investigators

6.2.3 VARIATION OF BOILING HEAT TRANSFER COEFFICIENT WITH MIXTURE COMPOSITION

Figure 6.11 shows the variation of boiling heat transfer coefficient with mole fraction of the more volatile component of acetone-water liquid mixtures on the plain tube with heat flux as parameter. The plot clearly depicts that the liquid mole fraction has a distinct effect on the heat transfer coefficient. A further examination of the plot reveals the following not-worthy points:

a. It is observed that for a given heat flux the heat transfer coefficient first decreases with the addition of acetone in distilled water till its composition is about 0.15. Beyond this composition, the heat transfer coefficient continuously increases. This behaviour is repeatedly exhibited for other values of heat flux. An implication of this would be that there is no linear relationship between the heat transfer coefficient and the liquid mixture composition.

At this juncture, it is important to discuss the possible reason which makes the heat transfer coefficient decrease and then increase with composition. It is a known fact that bubble growth rate in the boiling of binary liquid mixture largely depends upon the ability of the more volatile component to diffuse into growing bubbles through their vapour-liquid interfaces. Further, for a given mixture, with the increase in the vapour-liquid mole fraction difference of more volatile component, IY-XI the bubble growth rate decreases and consequently the induced turbulence due to the bubble dynamics gets decreased which results in reduction in the heat transfer coefficient. In addition, Tolubinskiy et al.(1966b) have carried out a photographic study to calculate the growth rate of vapour bubbles in superheated layer of binary liquid mixtures over a heated surface. They have concluded that the liquid concentration at which the bubble growth rate was minimum corresponded to a maximum value of IY-XI. Thus, it is obvious that for mixture composition corresponding to a maximum value of IY-XI the boiling heat transfer coefficient should become minimum. With this in view, a plot between IY-XI and X of the more volatile component for the binary mixture of acetone-water has been drawn in Figure 6.12. It is noted that IY-XI is maximum for an acetone mole fraction lying between 0.10 and 0.20. Hence, for mixture of this mole fraction the value of heat transfer coefficient is expected to be minimum. This, in fact, is what is observed from the plot of Figure 6.11 and thus it explains the behaviour of curves representing boiling heat transfer coefficient and mole fraction of acetone-water mixture.

b. The dotted line in Figure 6.11 joining the heat transfer coefficients corresponding to X = 0.0 and X = 1.0, represents the weighted heat transfer coefficients for the boiling of acetone-water mixture at a heat flux of 52901 W/m². In fact, the weighted heat transfer coefficient, h_{wtd} is calculated by the following equation:

$$h_{wtd} = h_1 X_1 + h_2 X_2 \tag{6-8}$$

where h_1 and h_2 are the respective heat transfer coefficients of the more and the less volatile components

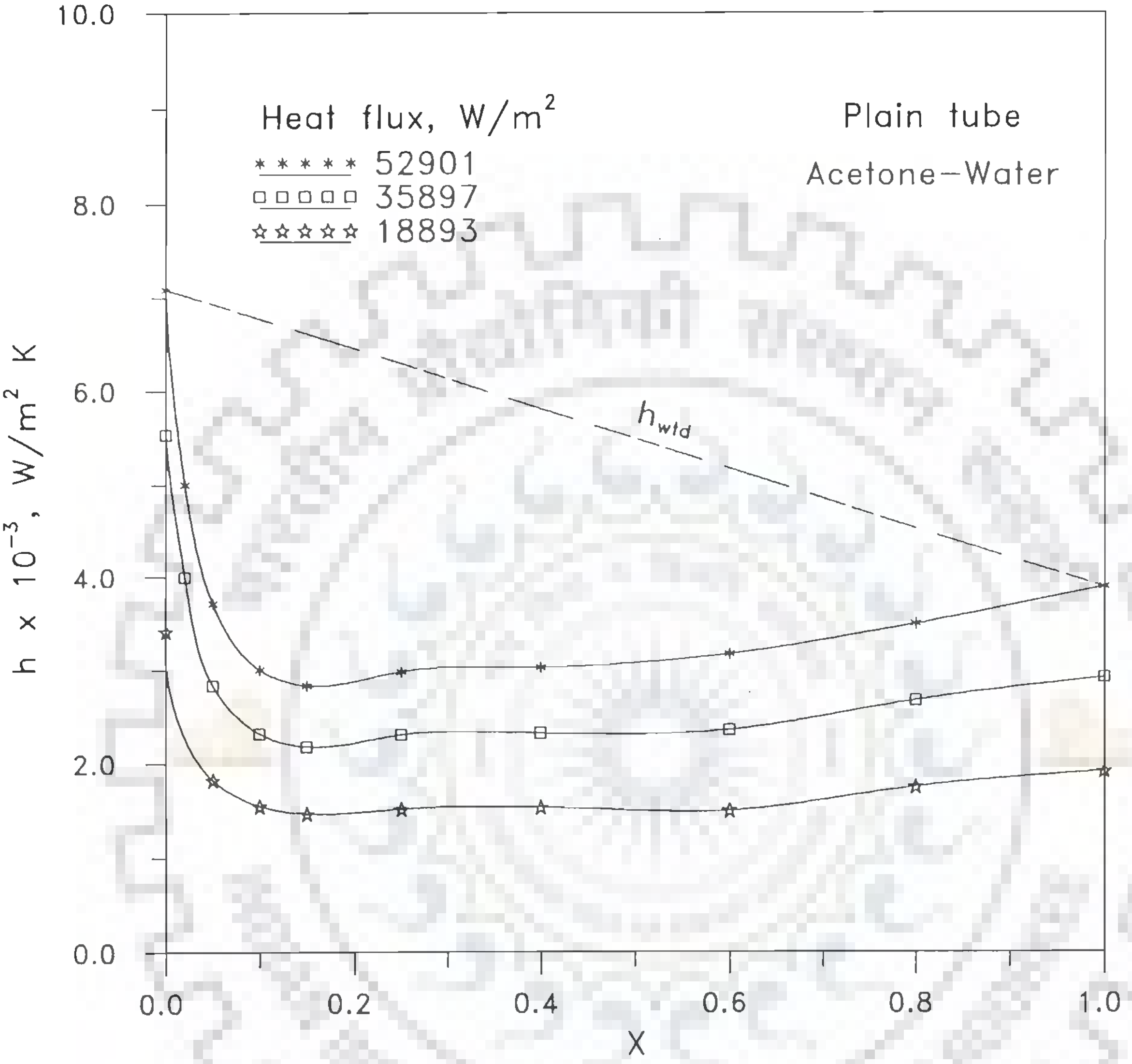


Fig. 6.11 Boiling heat transfer coefficient vs. mole fraction of acetone on a plain tube

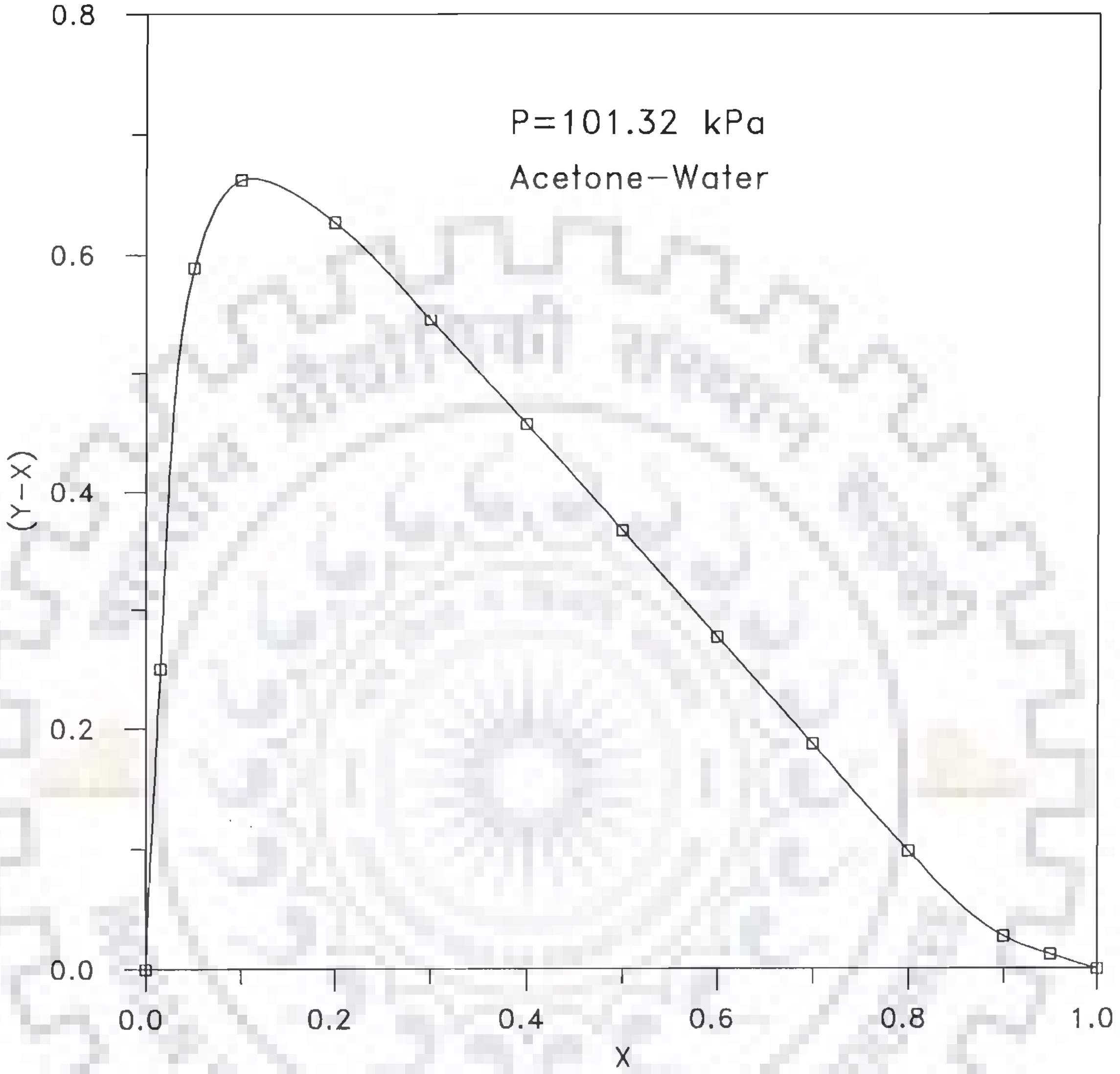


Fig. 6.12 Vapour−liquid mole fraction difference for acetone−water mixture

of the binary mixture and X_1 and X_2 are their respective liquid mole fractions.

From the above plot, it is convincingly clear that h_{wtd} is always greater than the experimental heat transfer coefficient, implying that it is not correct to obtain boiling heat transfer coefficient from the knowledge of h_1, h_2, X_1, and X_2 and using them in the above equation. However, in the past design engineers have been using h_{wtd} for the calculation of heating surface area for the pool boiling of binary liquid mixtures to be employed in the design of reboilers.

Similar to plots in Figures 6.11 and 6.12, the corresponding plots are drawn in Figures 6.13 and 6.14 for the boiling of acetone-isopropanol mixtures on plain tube. From the plot of Figure 6.13, it is clearly noted that h varies with X. However, the variation is not as distinct as in the case of acetone-water mixtures. The minimum value of h, though not quite distinct, appears for the mole fraction of acetone lying between the values of 0.3 and 0.4.

In this context, it is important to refer to plot of Figure 6.14 and to find out that for what value of X the value of IY-XI is the maximum and correspondingly the heat transfer coefficient will be the minimum. It is seen that this corresponds to a value of X between 0.3 and 0.4 mole fraction of acetone. It is also interesting to note that the maximum difference in IY-XI value in case of acetone-isopropanol mixture is about half that of the maximum value of IY-XI in case of acetone-water mixture. Hence, the present data related to the boiling of acetone-isopropanol mixtures are justified.

As regards the experimental data for the boiling of isopropanol-water mixtures, these have been plotted in Figure 6.15 between boiling heat transfer coefficient versus liquid mole fraction of the more volatile component (isopropanol) with heat flux as parameter. From the plot, it is observed that the heat transfer coefficient first nose-dives with the addition of isopropanol to distilled water and then increases with increasing concentration of isopropanol. The turn-around corresponds to a mixture of about 0.1 mole fraction of isopropanol. Further, the addition of isopropanol results in an increase in heat transfer coefficient till a composition of 0.68 mole fraction. In fact, for a composition of 0.68 the liquid forms azeotropic mixture. Therefore, at this composition the mixture behaves as pure single component i.e. Y=X. A further rise in mole fraction of isopropanol beyond a composition of 0.68 lowers the heat transfer coefficient though marginally. In addition to this, it can be observed from Figure 6.16 that, the difference between equilibrium vapour and liquid composition, IY-XI is maximum around the same mole fraction of isopropanol at which the heat transfer coefficient is minimum i.e around 0.1 mole fraction of isopropanol.

6.2.4 VARIATION OF DEGRADATION FACTOR WITH VAPOUR-LIQUID COMPOSITION DIFFERENCE

From Figure 6.11, it has been established that the weighted heat transfer coefficient, h_{wtd} is greater than the experimental heat transfer coefficient of any composition, irrespective of heat flux. Hence, it is clear that heat transfer coefficient gets degraded. However, the degradation is not uniform

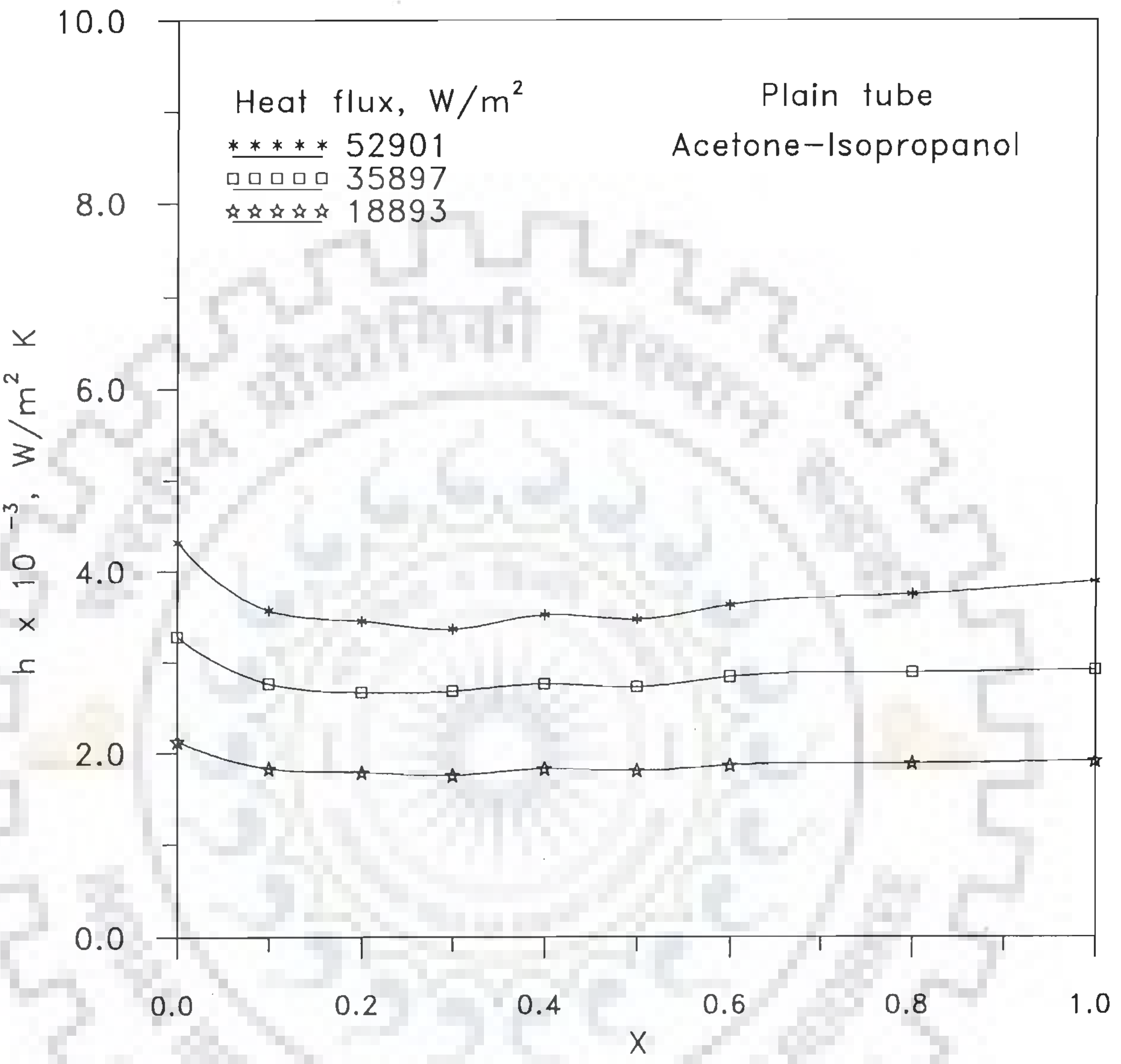


Fig. 6.13 Boiling heat transfer coefficient vs. mole fraction of acetone on a plain tube

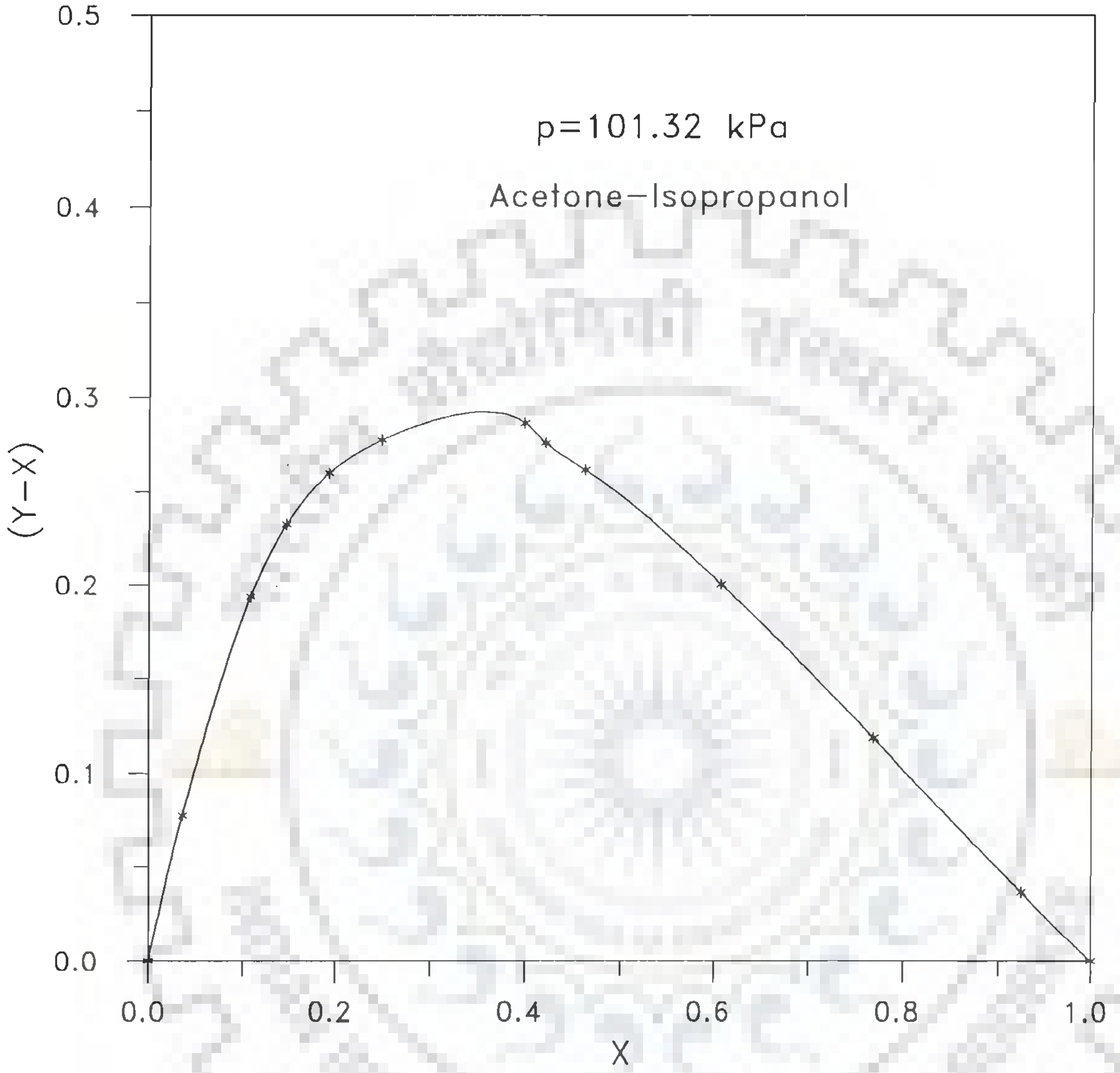


Fig. 6.14 Vapour−liquid mole fraction difference for acetone−isopropanol mixture

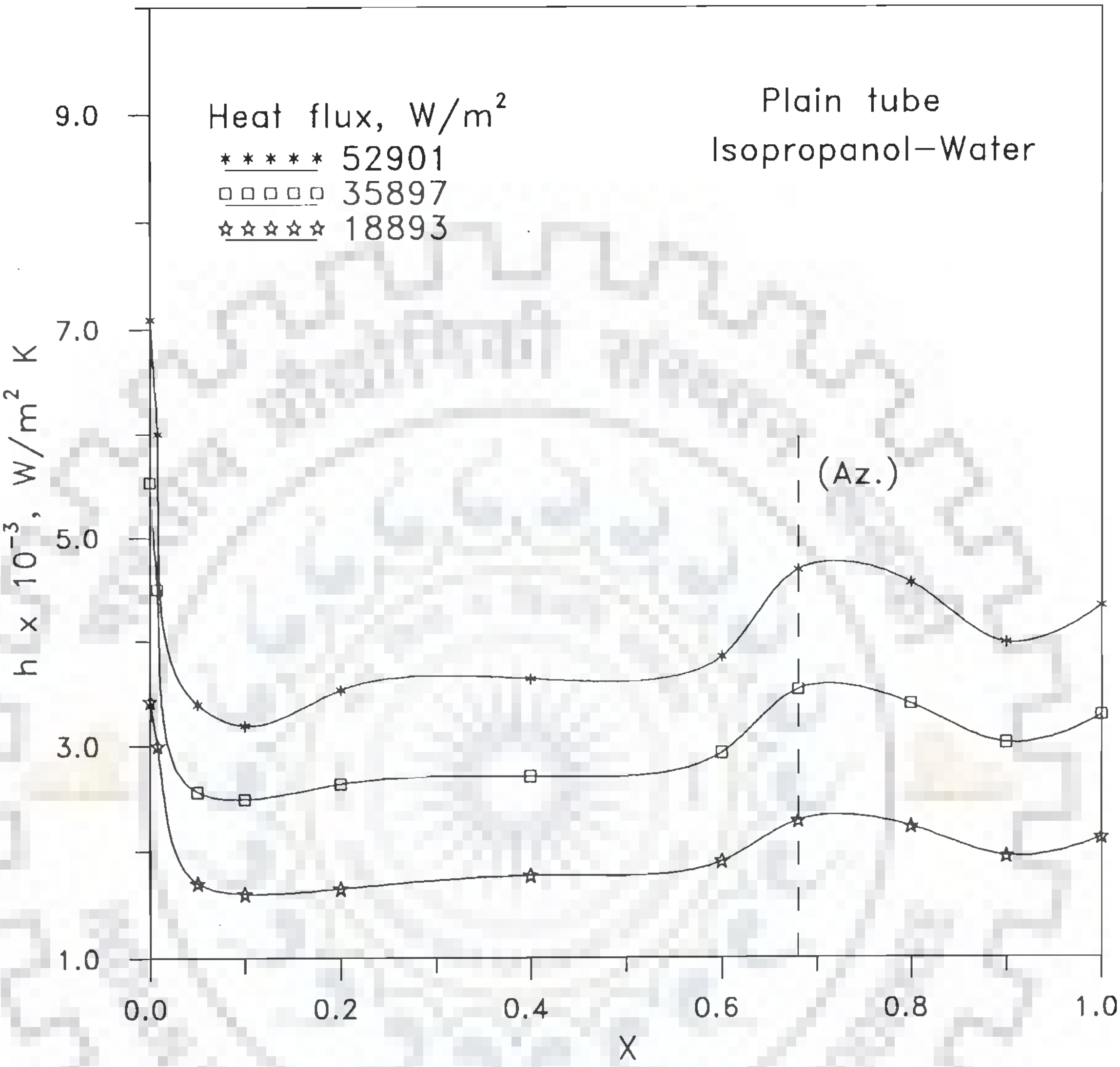


Fig. 6.15 Boiling heat transfer coefficient vs. mole fraction of isopropanol on a plain tube

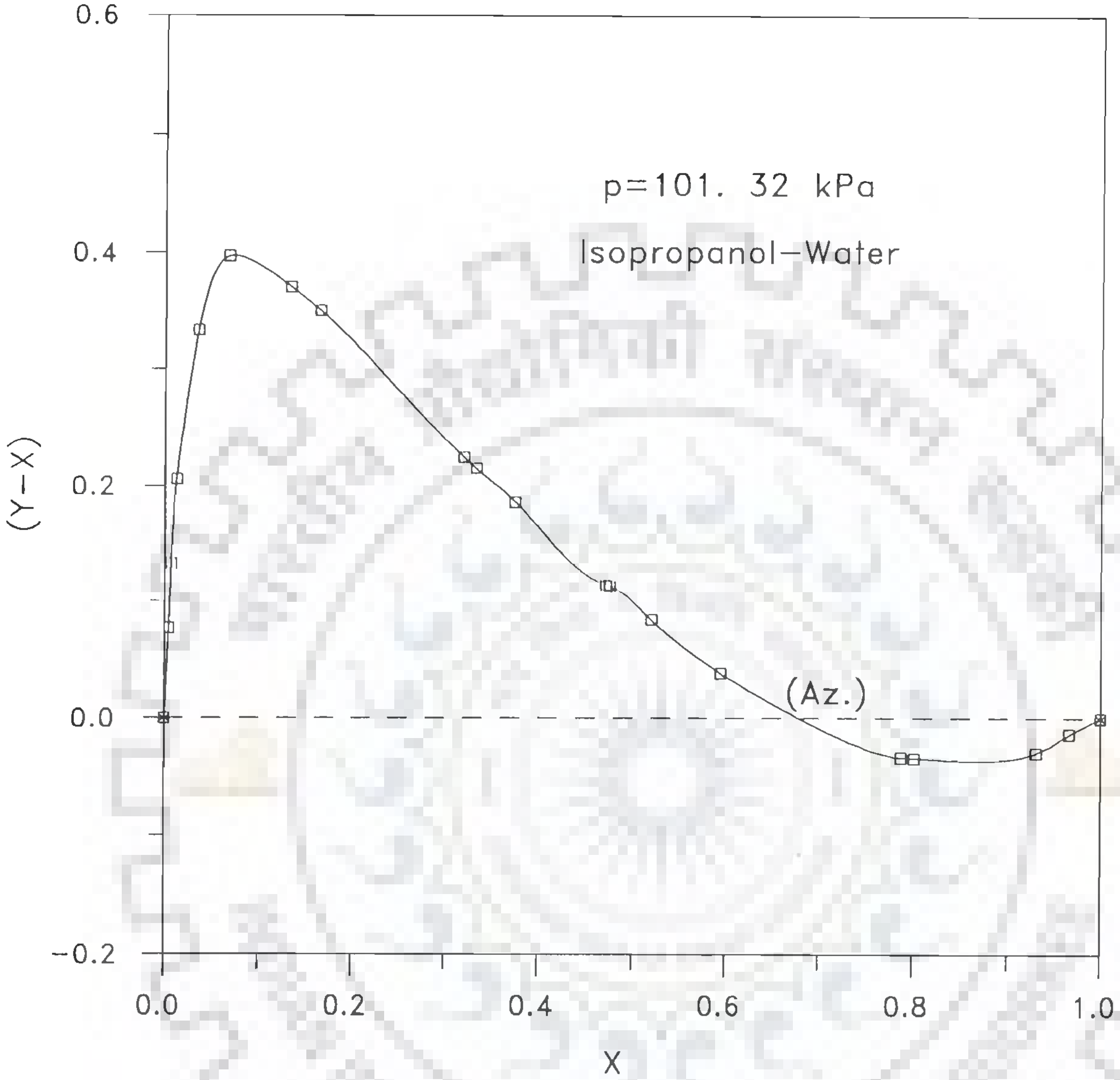


Fig. 6.16 Vapour−liquid mole fraction difference for isopropanol−water mixture

over the entire range of liquid composition. To appreciate this, a plot between (h/h_{id}) and $|Y-X|$ of the more volatile component was considered desirable. Such a plot is drawn in Figure 6.17 for boiling of all three binary liquid mixtures investigated, i.e. acetone-water, isopropanol-water, and acetone-isopropanol systems. The quantity (h/h_{id}) can be termed as ' degradation factor '. The values of h_{id} are to be calculated from Eq.(2-16), which is based on the ' linear mixing law method '. From the plot, the following noteworthy observations are made:

a. Data points of boiling liquids are represented by respective distinct separate curves, indicating that it is not possible to obtain a generalized functional relationship between (h/h_{id}) and $|Y-X|$.

b. However, unlike plots of Figures 6.11 , 6.13, and 6.15, there is a regular variation in values of (h/h_{id}) with $|Y-X|$. In fact, for a given boiling liquid with increase in value of $|Y-X|$, value of (h/h_{id}) decreases continuously, the lowest value corresponds to the largest value of $|Y-X|$.

In case of acetone-water mixture, the largest value of $|Y-X|$ is 0.645 while it is 0.288 and 0.396 for acetone-isopropanol mixture, and isopropanol-water mixture, respectively. It is interesting to point out that these respective values correspond to minimum values of heat transfer coefficient in Figures 6.11, 6.13, and 6.15. Hence, it is concluded that the degradation in heat transfer coefficient due to mass diffusion and variation in physico-thermal properties of mixtures so induced is the maximum for the above respective values of $|Y-X|$ for the respective more volatile component in the mixtures.

c. When $|Y-X|$ is close to zero , the value of (h/h_{id}) is nearer to 1.0, it is an expected behaviour. As a matter of fact, the value of $|Y-X|$ shall be close to zero, when the value of X_1 is nearer to Y_1, representing that the mixture is so much diluted that its physico-thermal properties are almost those of a single component liquid or the mixture composition near the azeotropic point, thereby mass diffusion tends to zero. Hence, degradation in heat transfer coefficient reduces to almost zero value.

d. For a given value of $|Y-X|$, the value of (h/h_{id}) is the highest for the acetone-isopropanol mixture, followed by the acetone-water and isopropanol-water mixtures in decreasing order. An implication of this would be that the degradation in the value of h is the smallest in the case of acetone-isopropanol mixtures, followed by that of acetone-water and isopropanol-water mixtures in increasing order.

6.3 NUCLEATE POOL BOILING OF TERNARY MIXTURE ON PLAIN TUBE

It is important to point out that nearly all studies of mixtures boiling have been confined to binary mixtures. This is due to the obvious reason that the study of ternary/multicomponent mixtures boiling is quite tedious and is more complicated than that of binary liquid mixtures , since the effect of adding a third component to a binary liquid mixture on bubble growth rate is difficult to identify precisely because the additional component changes both the vapour-liquid equilibrium compositions and also the physico-thermal properties of the mixture so obtained.

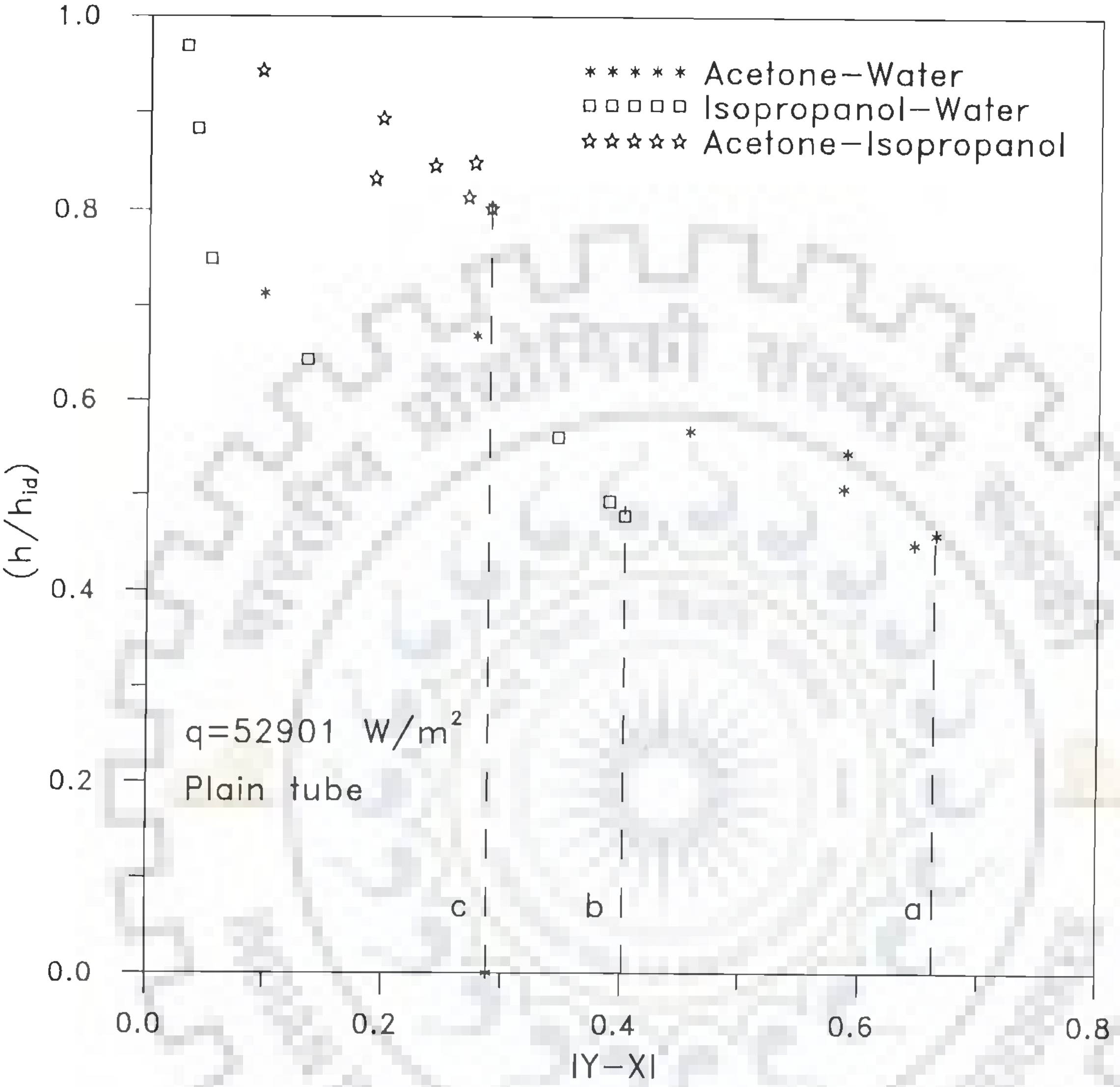


Fig. 6.17 Variation of (h/h$_{id}$) with IY−XI for binary mixtures boiling on a plain tube

In a ternary liquid mixture, one of the components is the more volatile having the less boiling point, the other component is the intermediate volatile having the intermediate boiling point and the still another component is the less volatile having the high boiling point. As a result of it, the vapour-liquid compositions difference dictated by phase equilibrium thermodynamics. The vapour phase will have a higher composition of the more volatile component than in the bulk liquid phase, and the corresponding liquid phase will have less of the more volatile component. As regards the vapour composition of the intermediate volatile component, it can be either higher or lower than the bulk liquid composition depending on the its boiling point i.e. volatility.

As in the case of single component liquids and binary liquid mixtures boiling, the following section is devoted as how does heat transfer coefficient change with heat flux.

6.3.1 EFFECT OF HEAT FLUX ON HEAT TRANSFER COEFFICIENT

Figure 6.18 represents the experimental data for the boiling of acetone-isopropanol-water ternary liquid mixtures on plain tube at atmospheric pressure. The plot depicts the effect of heat flux on nucleate pool boiling heat transfer coefficient with mixture composition as parameter. From the plot it is clearly seen that the boiling heat transfer coefficient changes with heat flux governed by the following functional relationship.

$$h = C_3\ h^{0.66} \qquad (6\text{-}9)$$

The value of constant C_3 depends upon the composition of the mixtures. As regard the value of the exponent of q , it is about 0.66.

All the data points of the ternary liquid mixtures lie below those of pure components, i.e water, isopropanol, and acetone. This established that the degradation in boiling heat transfer coefficients of ternary liquid mixtures, similar to what was observed for the acetone -water binary mixtures, exists.

6.3.2 VARIATION OF DEGRADATION FACTOR WITH VAPOUR-LIQUID COMPOSITION DIFFERENCE

In the case of ternary liquid mixture, the degree of degradation in boiling heat transfer coefficient due to mass diffusion and variation in physico-thermal properties should also depend on the thermodynamic requirements for a species to diffuse to the growing bubble interface from the bulk liquid in order to maintain equilibrium. Thus, it is possible to expect that the degradation factor of boiling heat transfer coefficients in ternary mixture, (h/h_{id}) should be a function of $|Y-X|$, which is the summation of $|Y-X|$ of the more and the intermediate volatile components in the mixture as follows:

$$|Y-X| = |Y_1-X_1| + |Y_2-X_2|$$

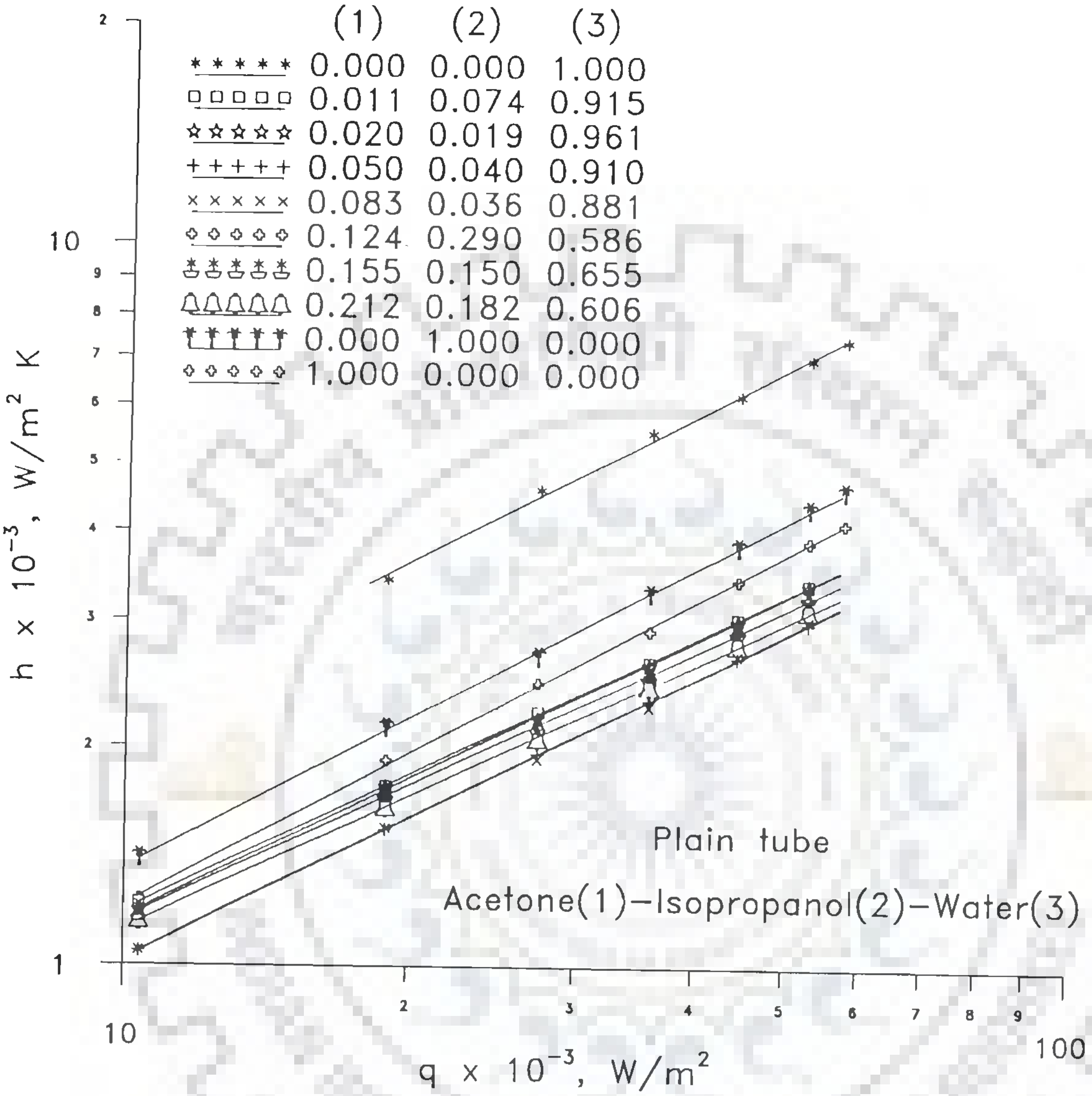


Fig. 6.18 Nucleate boiling curves for ternary mixtures on a plain tube

The indices 1 and 2 represent the more and the intermediate volatile components, respectively. Figure 6.19 has been drawn to represent the variation of (h/h_{id}) against the $|Y-X|$. It is important to mention that the term $|Y-X|$ stands for the summation of the vapour-liquid mole fraction difference of the more volatile component $|Y_1-X_1|$, i.e. acetone and that of the intermediate volatile component $|Y_2-X_2|$, i.e isopropanol in the present ternary mixture investigation at a heat flux of 52901 W/m². The linear mixing law, Eq.(2-16) was used to calculate the ideal heat transfer coefficient, h_{id}. From the plot it can be noticed that for the nucleate pool boiling of mixtures on plain tube the value of (h/h_{id})decreases with the increase in the value of $|Y-X|$. In this respect, the result are in accordance with the findings related to binary liquid mixtures.

6.3.3 A COMPARISON BETWEEN DEGRADATION FACTORS OF BINARY AND TERNARY MIXTURES

In section 6.2.4 and 6.3.2, it has been well-established that addition of a component to a pure liquid or liquid mixtures results in the degradation of boiling heat transfer coefficient. To understand this characteristics of mixture boiling, a new quantity termed as ' degradation factor ', (h/h_{id}) has been defined in section 6.2.4, and the respective plots of Figures 6.17 and 6.19 for binary and ternary mixtures boiling were drawn.

In order to compare ' degradation factor ' of ternary mixture boiling with that of binary mixtures, Figure 6.20 has been redrawn based on experimental data of Figures 6.17 and 6.19. From plots, following observations are apparently noted:

a. The trend of variation of (h/h_{id}) with $|Y-X|$ is the same, irrespective of whether mixture is binary or ternary.

b. As regards the (h/h_{id}) for ternary mixture boiling, it is less than that of acetone-water and still less of acetone-isopropanol mixtures. However, it is more than that of isopropanol-water mixture.

The governing reason for above noted observation appears to be due to the fact is that the addition of water to acetone-isopropanol systems changes the physico-thermal properties and vapour-liquid equilibrium of the so produced ternary mixture, which seemingly retards the mass diffusion and thereby the bubble growth rate. Thus, the value of heat transfer coefficient is further lowered. This, in turn, lowers the value of (h/h_{id}). However, the curve for the ternary system is above that of isopropanol-water. It seems that the properties of ternary system are not deteriorated to such an extent so that its heat transfer coefficient becomes less than that of azeotropic binary system.

6.4 NUCLEATE POOL BOILING OF SINGLE COMPONENT LIQUIDS ON INTEGRAL-FIN TUBES

Like the boiling of single component liquids on plain tube, data were also obtained when single component liquids were boiled on integral-fin tubes at atmospheric pressure. These data are listed in Table

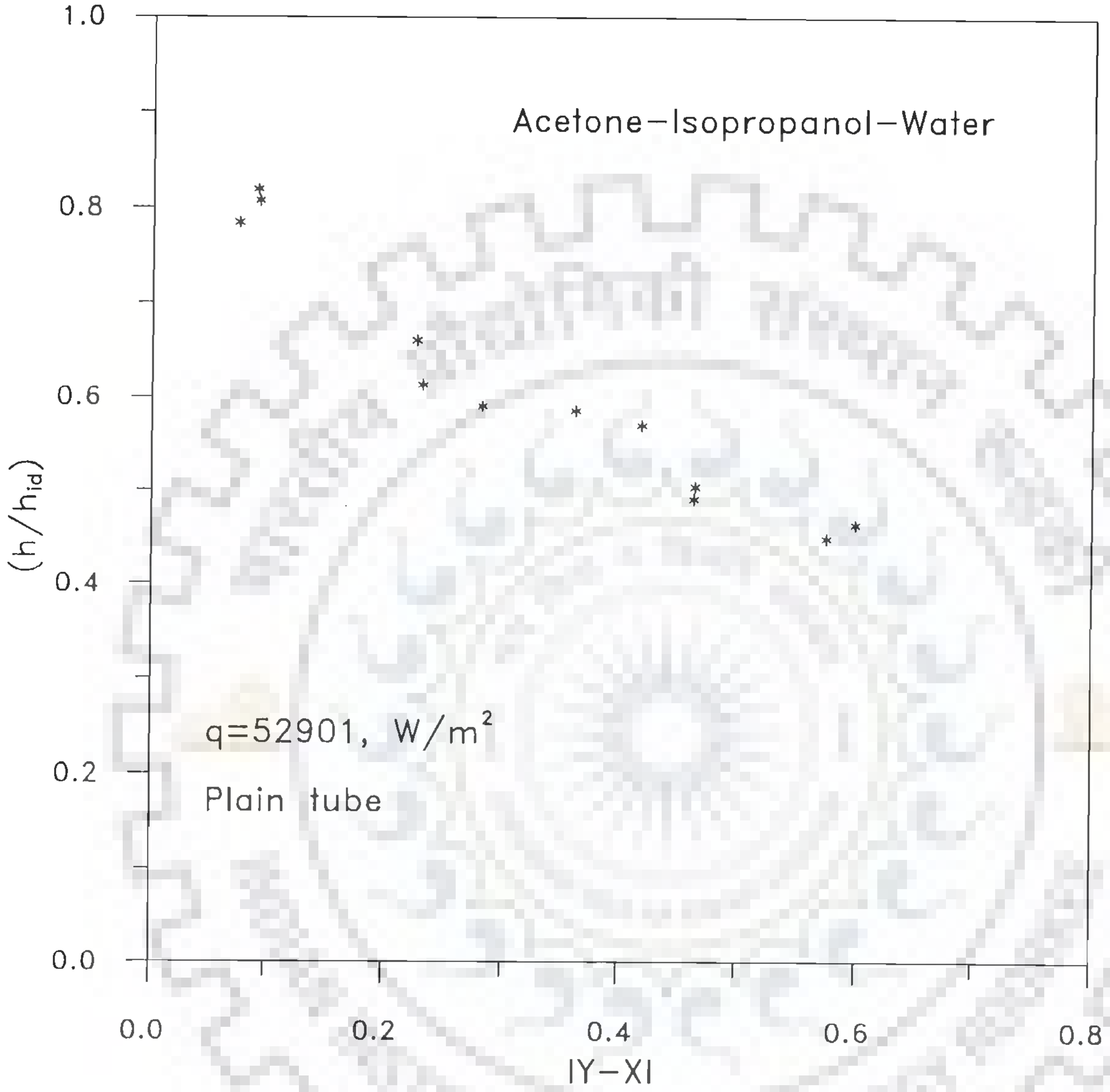


Fig. 6.19 Variation of (h/h_{id}) with IY−XI for ternary mixture boiling on a plain tube

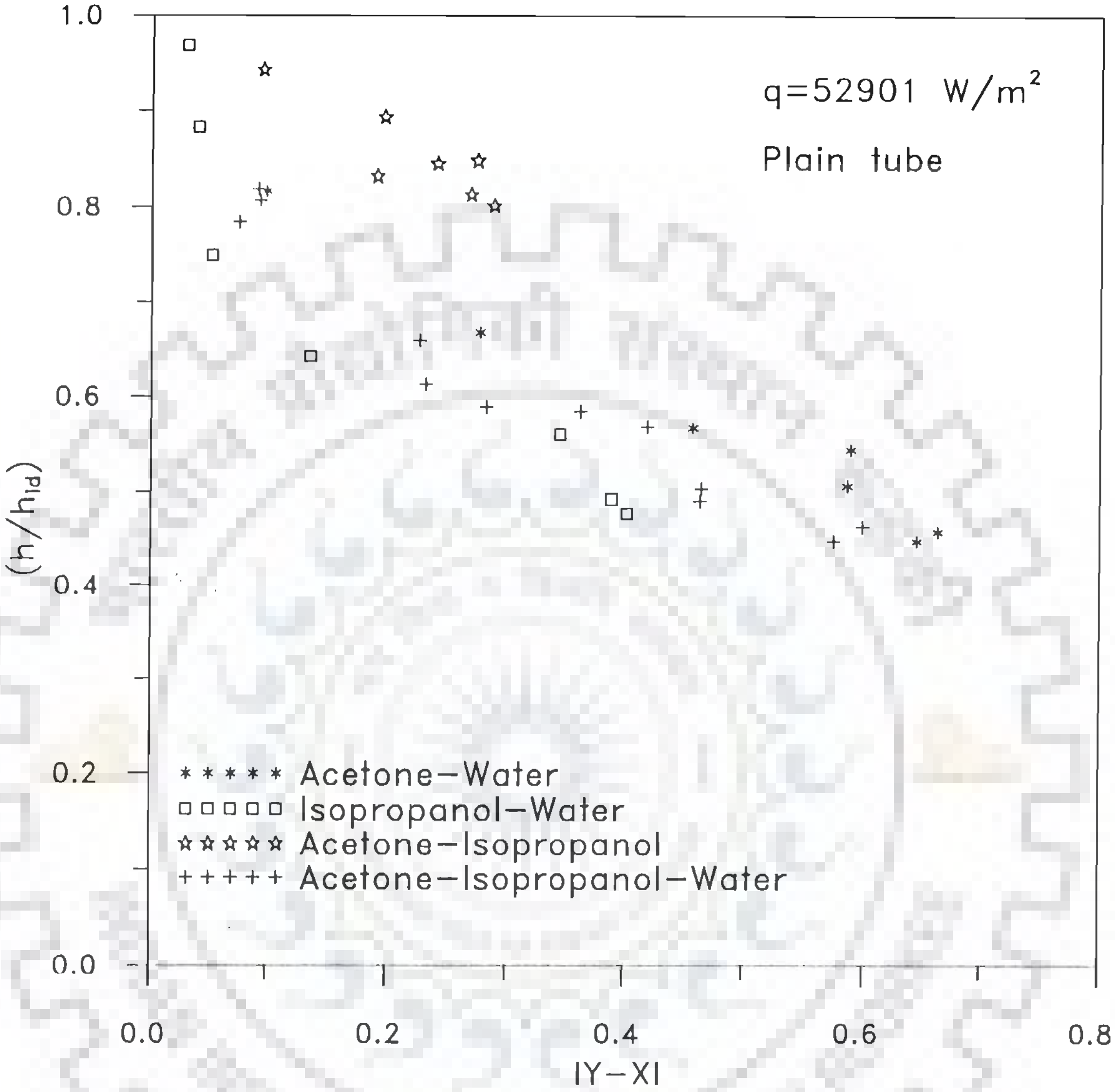


Fig. 6.20 Variation of (h/h_{Id}) with $|Y-X|$ for binary and ternary mixtures boiling on a plain tube

D.6 and D.11 for 748 and 1024 fpm tubes, respectively.

6.4.1 NUCLEATE POOL BOILING CURVES

As explained in section 6.1.3, the ' Nucleate Boiling Curve ' is a plot between h and q. Accordingly, Figures 6.21 and 6.22 have been drawn to represent ' boiling curves ' of acetone , isopropanol, water and isopropanol-water (Az.)when they boil on 748 and 1024 fpm tubes, respectively at atmospheric pressure. An examination of the plots of these figures shows that heat transfer coefficient varies with heat flux, irrespective of boiling liquids and geometry of fins on tubes. Data points of individual liquids are correlated by separate straight lines having the following functional relationship:

$$h = C_4 \, q^{0.71} \qquad (6\text{-}10)$$

where value of constant C_4 depends upon boiling liquid, irrespective of heat flux for a given heating tube. Values of C_4 for single component liquids and isopropanol-water (Az.) have been tabulated in Table 6.3 as below.

Table 6.3 Values of constant C_4 in Eq.(6-10)

Boiling liquid	C_4	
	748 fpm tube	1024 fpm tube
Acetone	0.527	0.591
Isopropanol	0.614	0.851
Isoprpanol-Water(Az.)	0.516	0.646
Water	0.740	0.914

It is interesting to note that Eq.(6-10) is similar to Eq.(6-7) but with different value of exponent of q, being 0.71. As regards the value of constants of these equations, there is a significant difference. For example, it is 0.527 for boiling of acetone on 748 fpm heating tube against a value of 0.240 on plain tube. This value is further enhanced when heating tube has 1024 fpm.

In nutshell, it is easily seen that for a given heat flux and boiling liquid, heat transfer coefficient is the largest when heating tube has 1024 fpm, followed by that of 748 fpm heating tube and of plain tube in decreasing order. In other words, heat transfer coefficient is enhanced when plain tube is replaced by integral-fin tube. For this behaviour , no satisfactory explanation is available. However, it is speculated that this enhancement can be partly due to increase in wetted heat transfer surface area due to integral-fins on heating tube. In fact, an integral-fin tube having fin density of 748 has 1.9 times the wetted surface area of a plain tube, and it is 2.3 times when tube is of 1024 fpm.

Investigations for boiling of pure component liquids on integral-fin tube have been carried out by Westwater (1973), Hahne et al. (1991) and Kumar (1992). They have reported that channel width

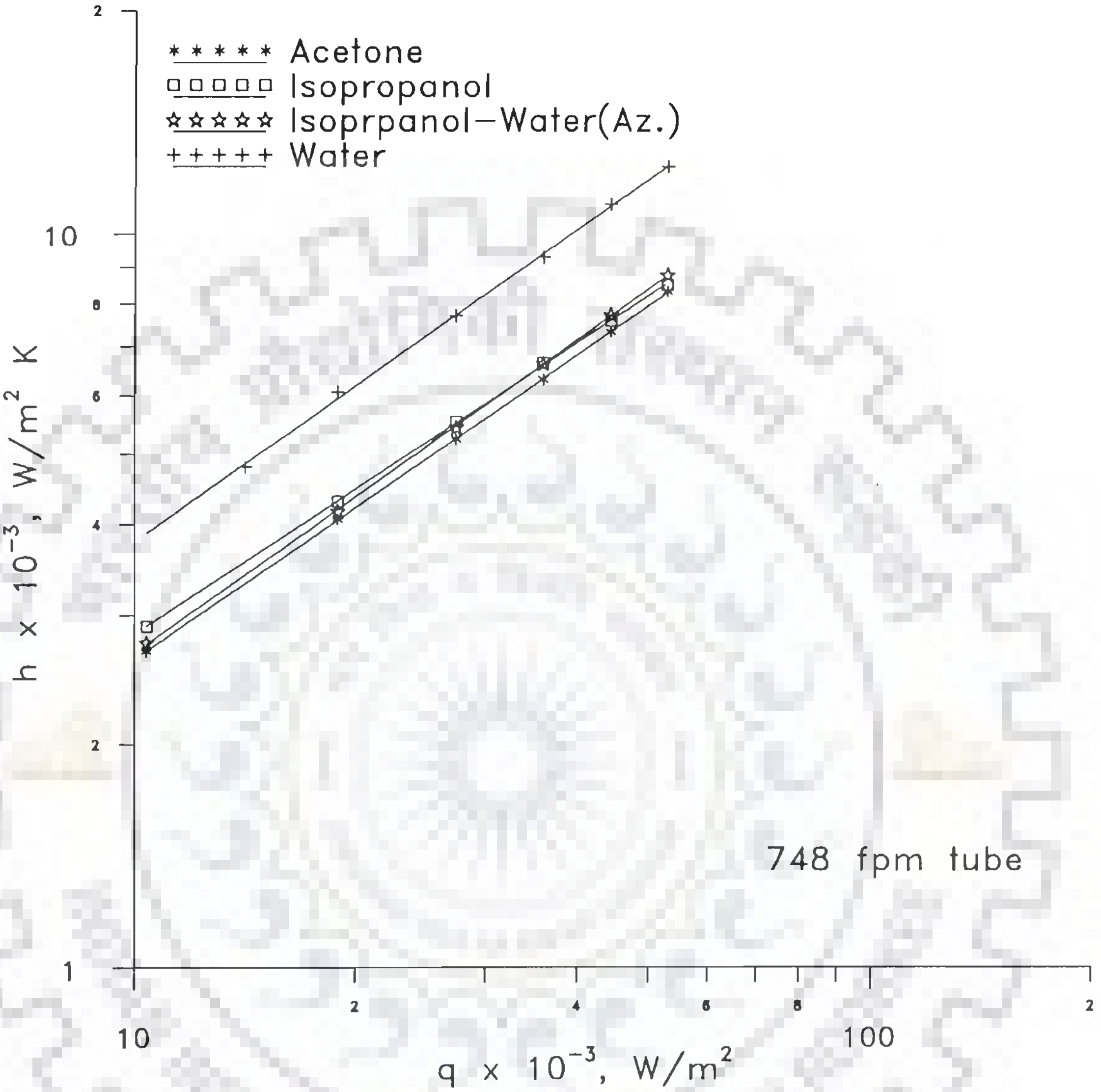


Fig. 6.21 Nucleate boiling curves for pure components on an integral–fin tube

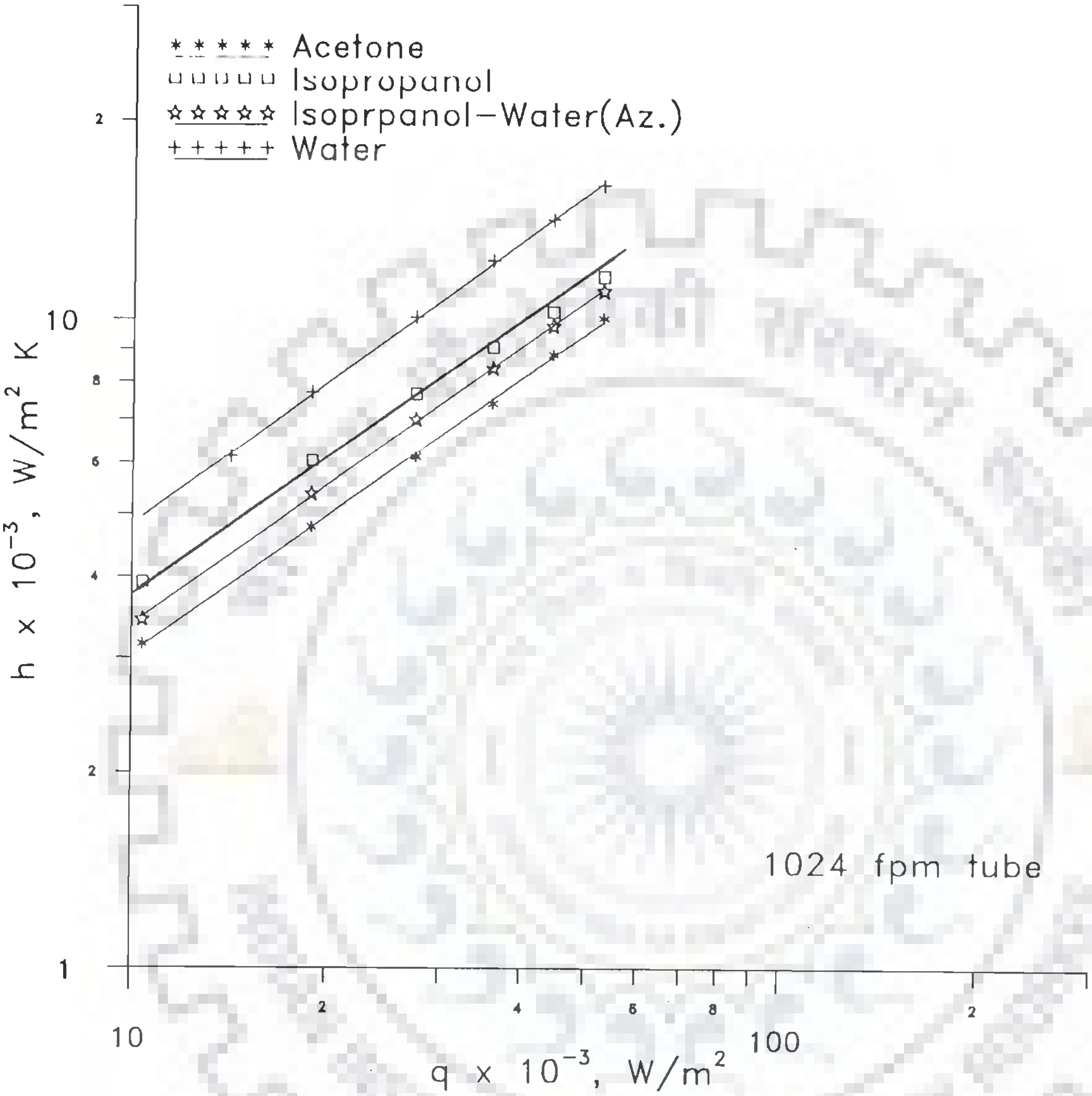


Fig. 6.22 Nucleate boiling curves for pure components on an integral-fin tube

between two adjacent fins is the major factor which influences boiling heat transfer coefficient. In fac with change in width there is a change in fin density. The results of the present investigation supports th above argument and it can be noted from the above plots that fin density is a major factor which influence the heat transfer coefficient.

6.5 NUCLEATE POOL BOILING OF BINARY LIQUID MIXTURES ON INTEGRAL-FIN TUBES

In the following sections, the effects of heat flux and mixture composition on boiling heat transfer coefficien for the boiling of binary liquid mixtures on integral-fin tubes are discussed.

6.5.1 EFFECT OF HEAT FLUX ON HEAT TRANSFER COEFFICIENT

Figures 6.23 and 6.24 represent the respective typical log-log plot to show the effect of heat flu on heat transfer coefficient for the boiling of binary liquid mixtures of acetone-water on integral-fin tube of fin density of 748 and 1024 fpm, respectively with mixture composition as parameter. From these plots the following worth-mentioning points emerge out:

a. For a given liquid mixture and heating tube, heat transfer coefficient increases linearly with the rise in heat flux, being correlated by Eq.(6-7), but with different values of C_2.

It is important to point out that the value of C_2 depends upon the composition of boiling liqui mixture and also heating tube geometries, irrespective of the heat flux. In fact, this observation is justifie as constant, C_2 represents the ' surface-liquid combination factor '. This factor, as a matter of fact, depends upon surface characteristics of heating tube and properties of boiling liquid. Due to this reason, the value of C_2 changes with mixture composition and heating tubes. The detailed account of ' surface-liquid combination factor ' is given in section 6.1.3.

b. Boiling data of all compositions of mixtures lie below those of pure components constituting the mixtures. The reason of this observation is that the addition of a component lowers the boiling heat transfer coefficient. This, of course, is due to the mass diffusion of the more volatile component of the mixture.

The observations at (a) and (b) as above were found applicable to experimental data for the boiling of mixtures of isopropanol-water and acetone-isopropanol systems, supporting that they represent the characteristics of binary mixtures boiling on plain tube.

6.5.2 VARIATION OF HEAT TRANSFER COEFFICIENT WITH MIXTURE COMPOSITION

Figures 6.25 through 6.30 demonstrate variation in the heat transfer coefficient with mole fraction of more volatile component for boiling of acetone-water, isopropanol-water, and acetone-isopropanol

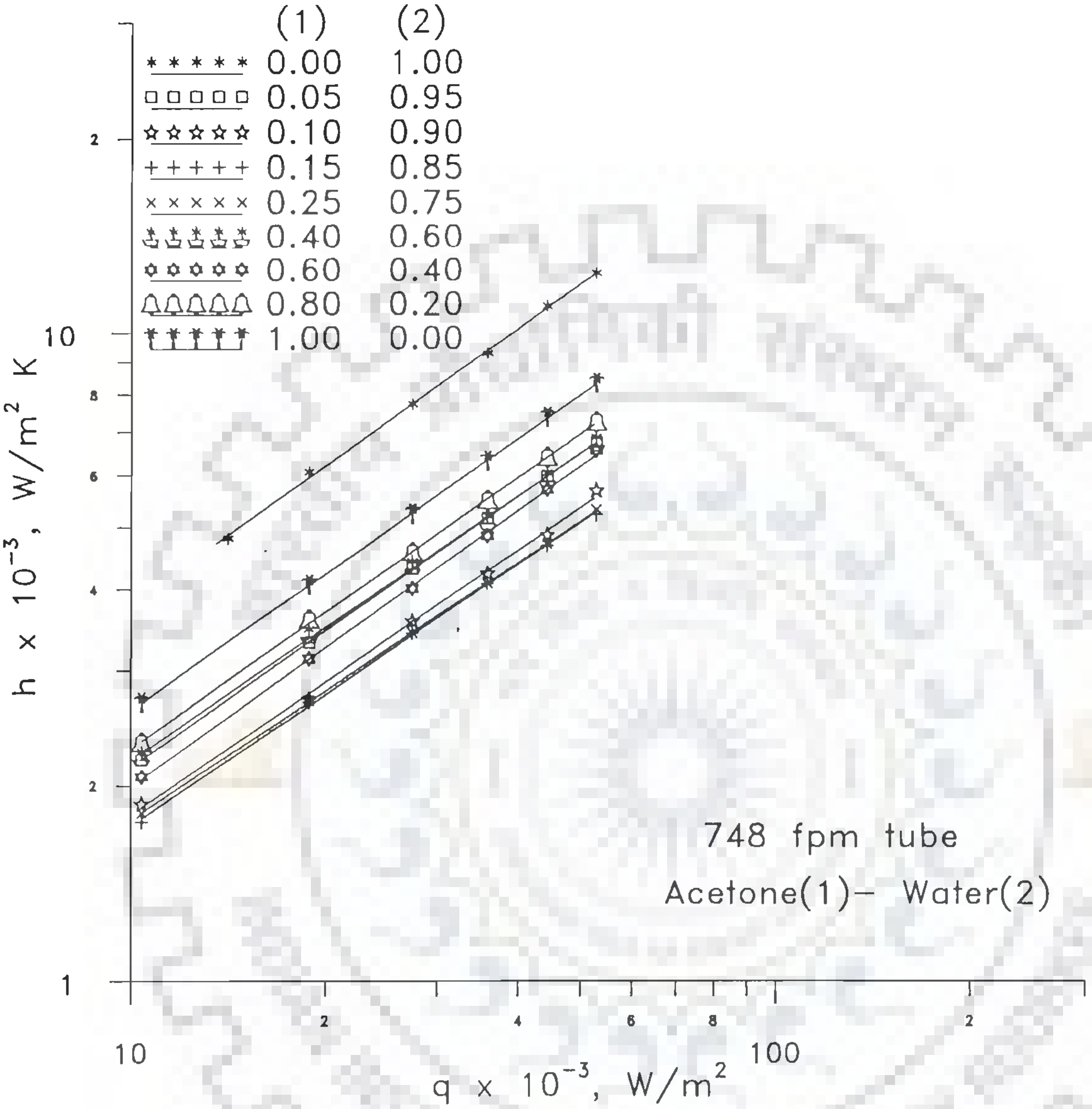


Fig. 6.23 Nucleate boiling curves for acetone–water mixture on an integral–fin tube

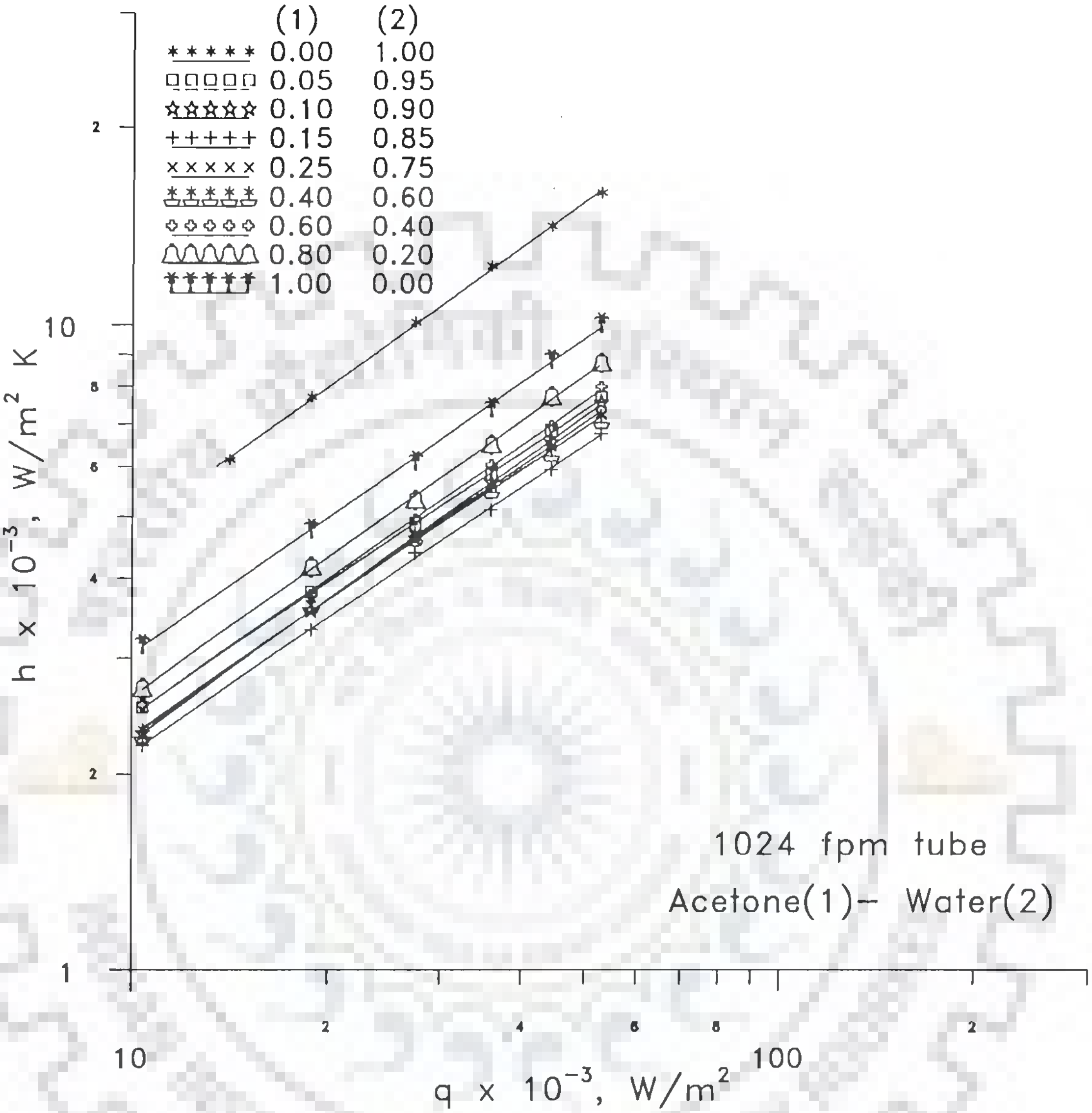


Fig. 6.24 Nucleate boiling curves for acetone–water mixture on an integral–fin tube

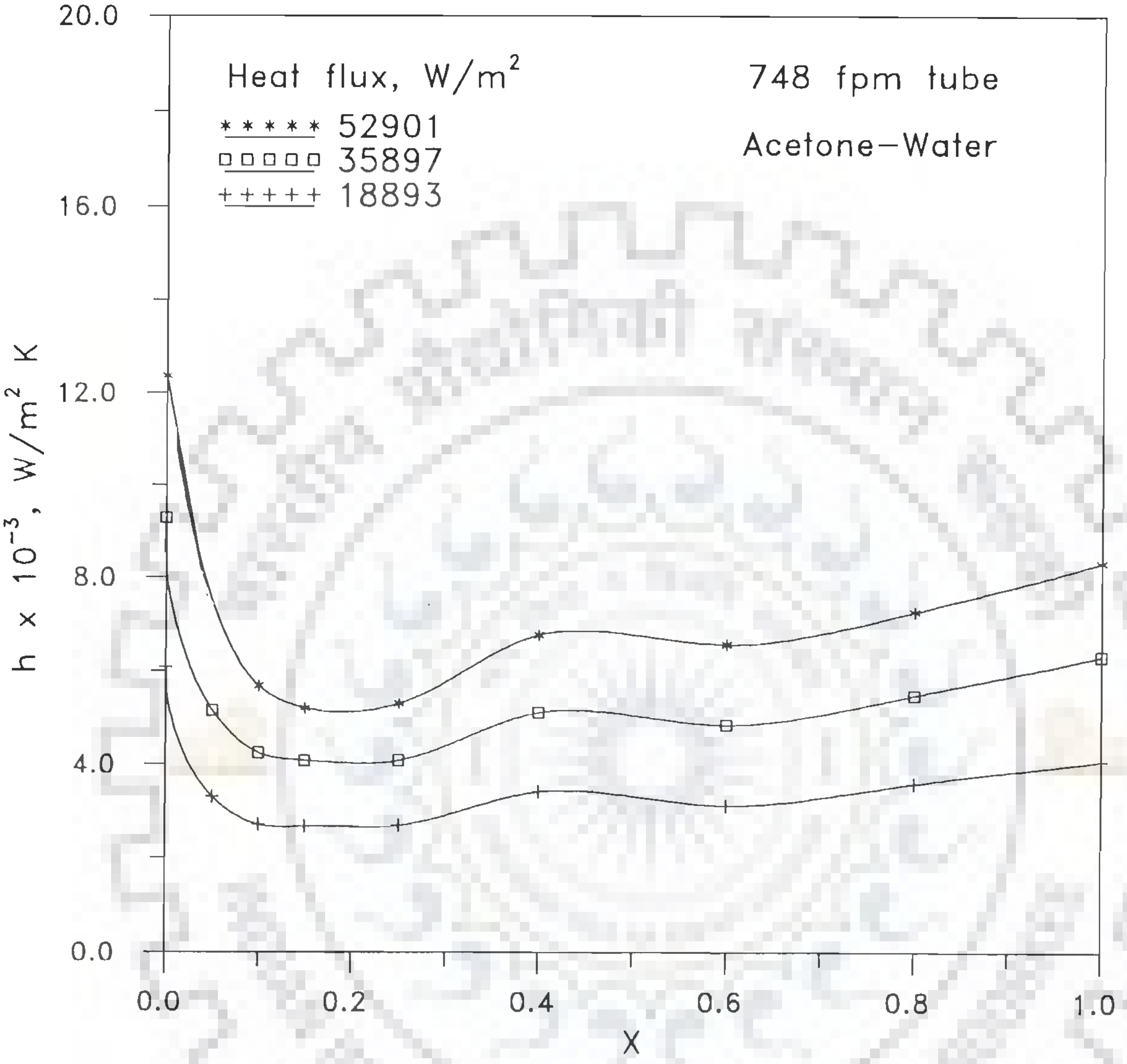


Fig. 6.25 Boiling heat transfer coefficient vs. mole fraction of acetone on an integral–fin tube

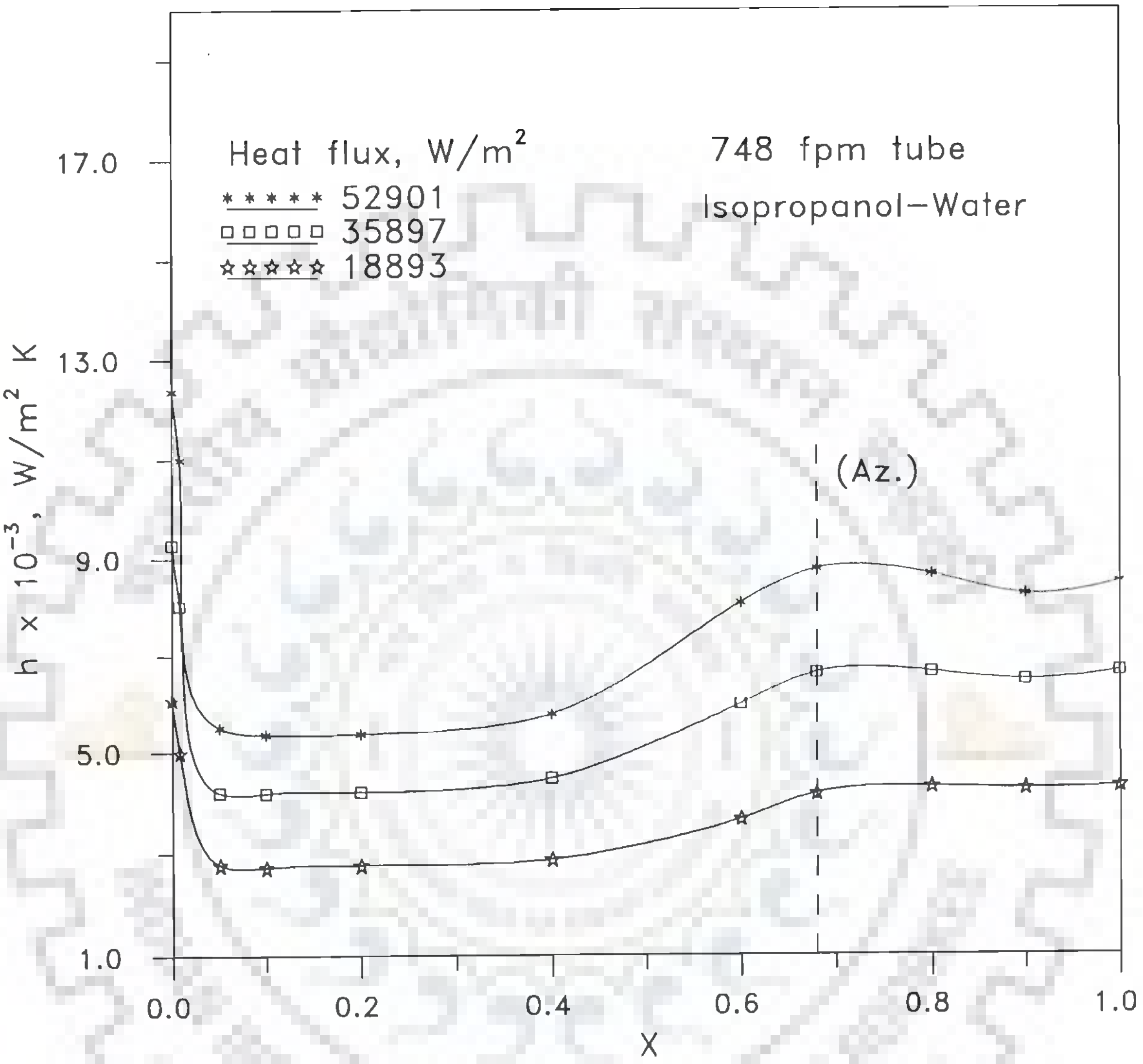


Fig. 6.26 Boiling heat transfer coefficient vs. mole fraction of isopropanol on an integral–fin tube

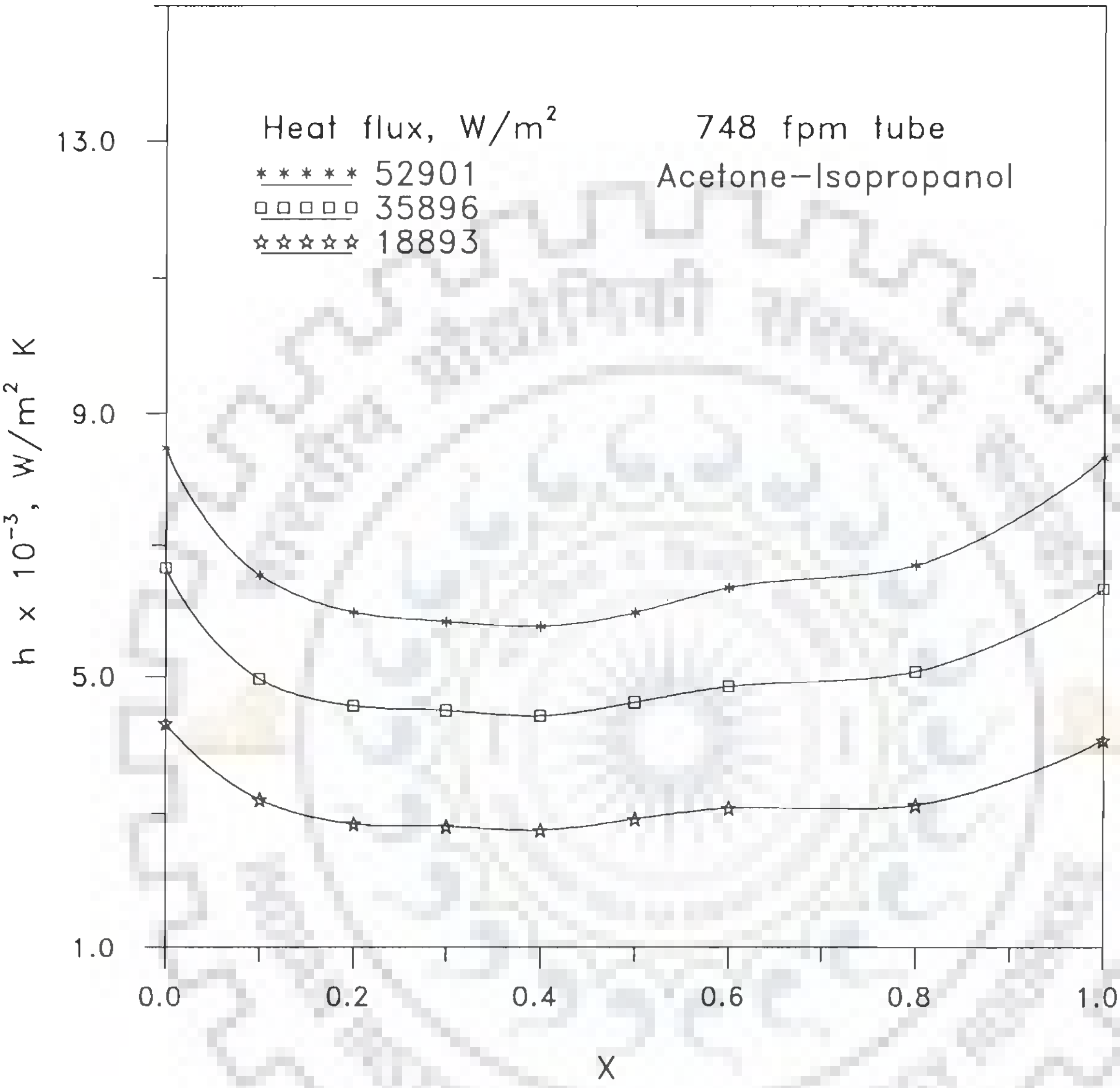


Fig. 6.27 Boiling heat transfer coefficient vs. mole fraction of acetone on an integral-fin tube

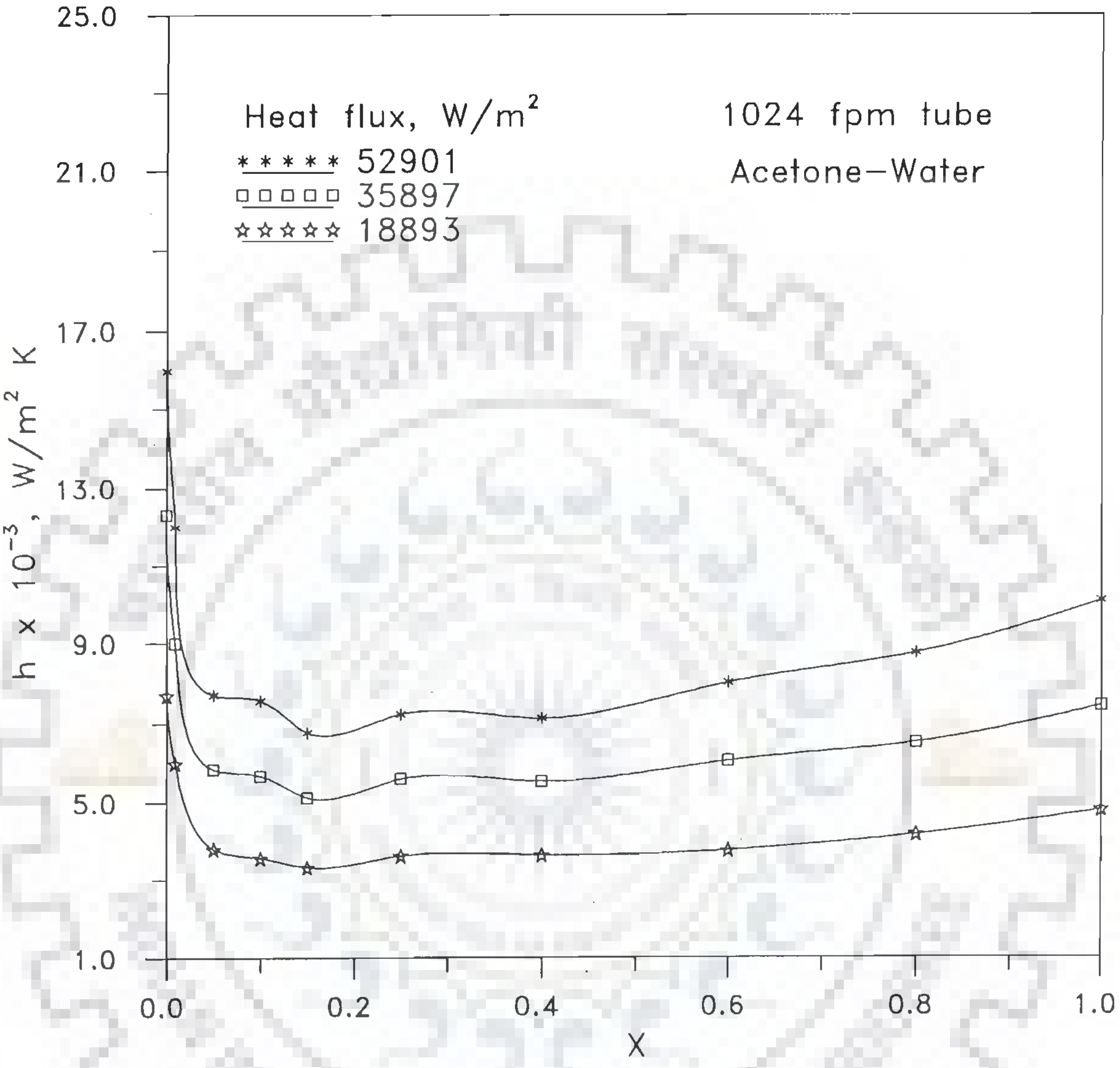


Fig. 6.28 Boiling heat transfer coefficient vs. mole fraction of acetone on an integral-fin tube

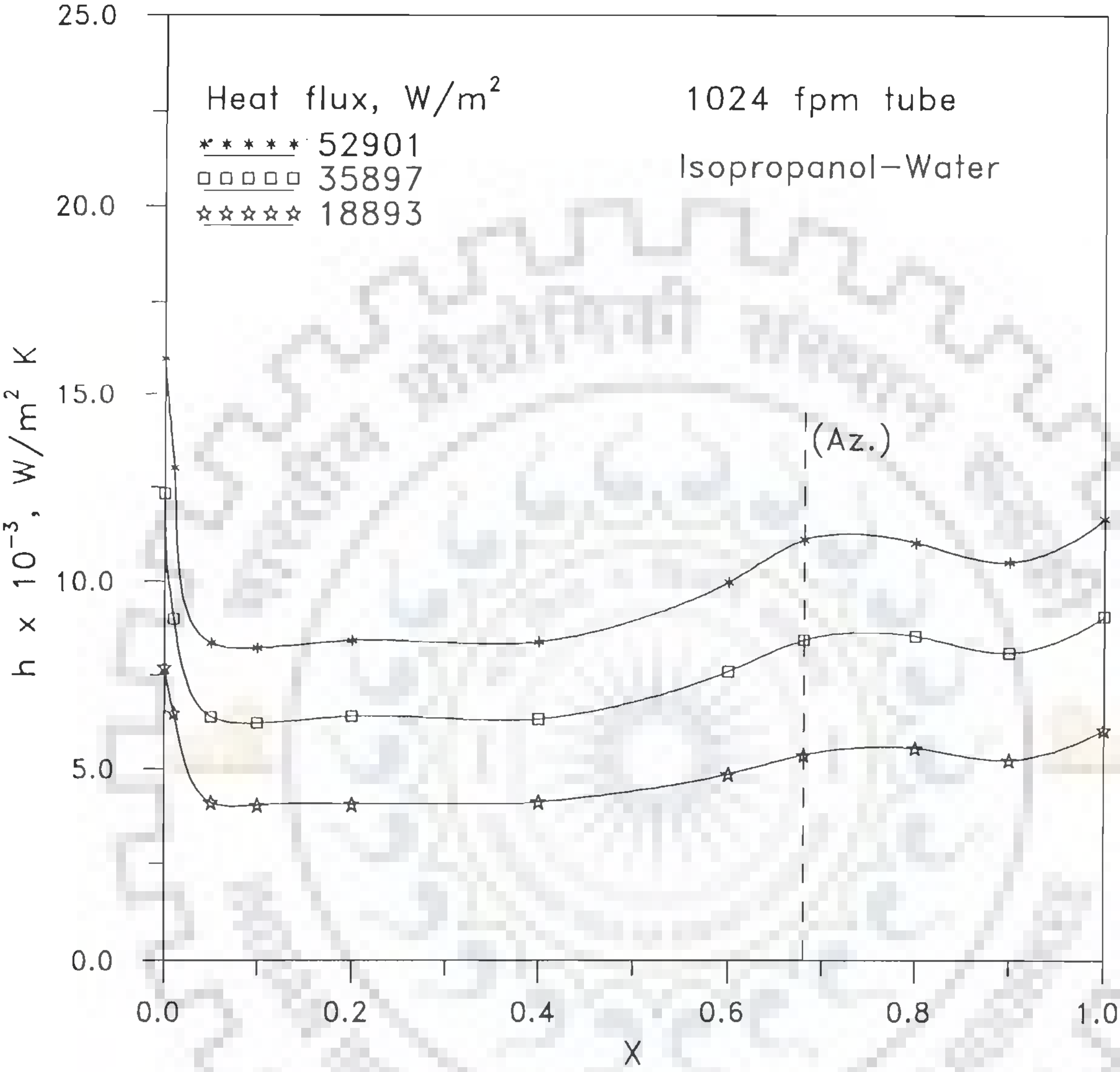


Fig. 6.29 Boiling heat transfer coefficient vs. mole fraction of isopropanol on an integral-fin tube

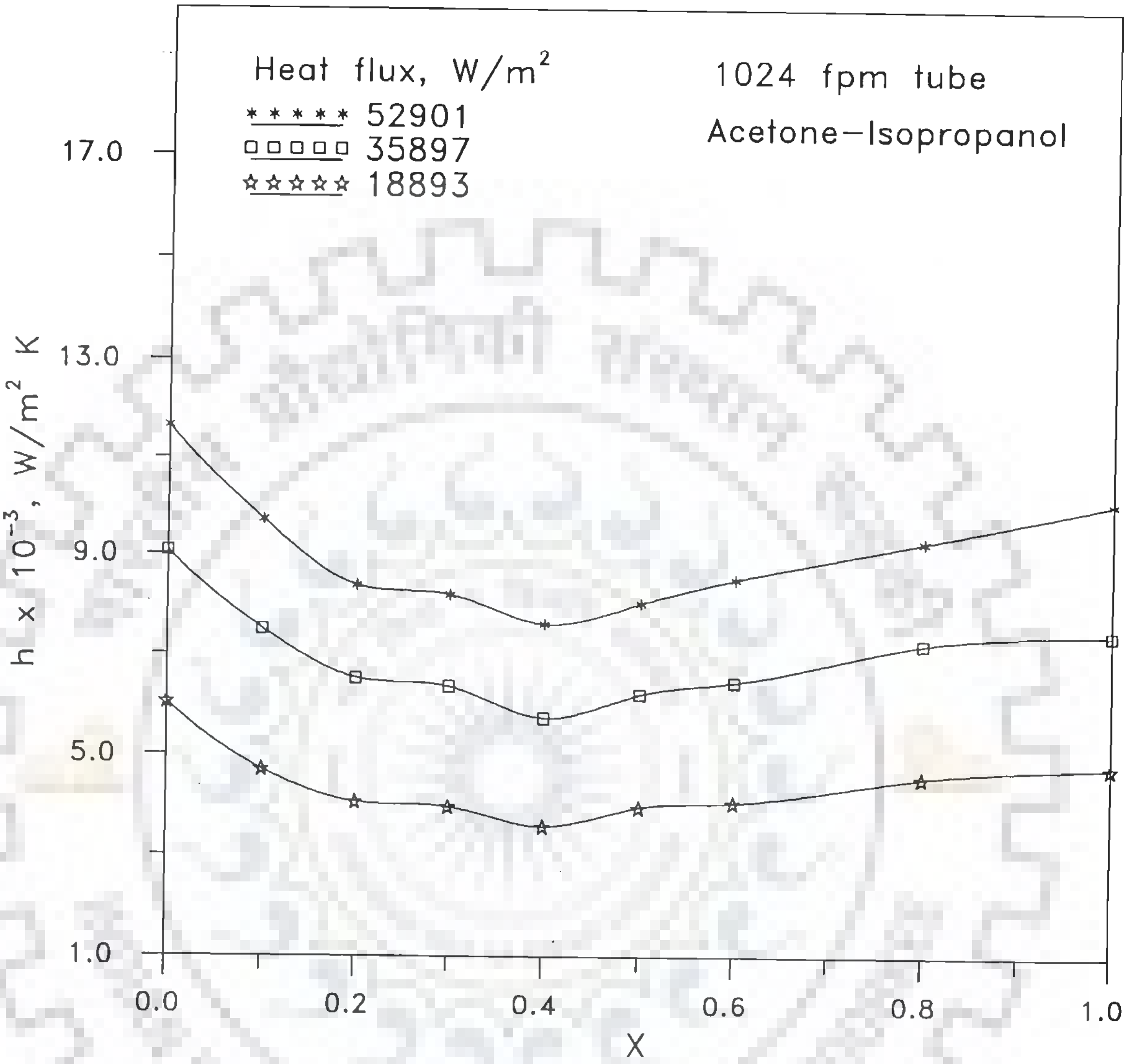


Fig. 6.30 Boiling heat transfer coefficient vs. mole fraction of acetone on an integral–fin tube

mixtures on integral-fin tube having fin density of 748 and 1024 fpm, respectively. A scrutiny of these plots reveals the following characteristic features:

a. Heat transfer coefficient for a given integral-fin tube for the boiling of a mixture depends on the composition of the mixture.

b. The trend of variation in boiling heat transfer coefficient with composition for acetone-water, isopropanol-water and acetone-isopropanol mixtures is the same as it is for boiling of these mixtures on plain tube. The turnaround in heat transfer coefficient occurs corresponding to 0.15 mole fraction of acetone for acetone-water mixtures and about 0.1 mole fraction of isopropanol for isopropanol-water mixtures. Whereas, in the case of boiling of acetone-isopropanol mixture it is between 0.3 and 0.4 mole fraction of acetone.

c. A comparison of plots of Figures 6.25 and 6.28 with those of Figure 6.13 shows that there is a significant enhancement in the boiling heat transfer coefficients of acetone-water mixture when plain tube is replaced by 748 fpm tube and it is still more pronounced for 1024 fpm tube.

d. Observation (c) is equally applicable for isopropanol-water and acetone-isopropanol mixtures, indicating for the boiling of binary liquid mixtures it is more advantageous to employ integral-fin tubes.

6.5.3 VARIATION OF DEGRADATION FACTOR WITH VAPOUR-LIQUID COMPOSITION DIFFERENCE

Figure 6.31 has been drawn to represent the typical effect of $|Y-X|$ of the more volatile component on 'degradation factor', (h/h_{id}) for the boiling of binary liquid mixtures on integral-fin tubes. Data of acetone-water mixtures have been employed for this plot. The values of h_{id} are calculated from Eq.(2-16). From the plots, some of the important observations made are as follows:

a. The value of (h/h_{id}) decreases with increase in $|Y-X|$, irrespective of fin density of heating tube. In this respect, experimental data on integral-fin tube and plain tube are approximately alike.

b. For a given boiling mixture composition, the value of (h/h_{id}) does not change appreciably with fin density of heating tube.

Arising out from observation (b), it was considered desirable to compare data obtained on plain as well as integral-fin tubes for the boiling of acetone-water mixtures. These data have been plotted in Figure 6.32. It is obvious that all data are almost represented by the same curve. Finally, it is inferred that the value of (h/h_{id}) for a given mixture remains unaltered with change in surface geometry of the heating tubes.

The almost similar boiling characteristics were exhibited by the remaining binary liquid mixtures inves-

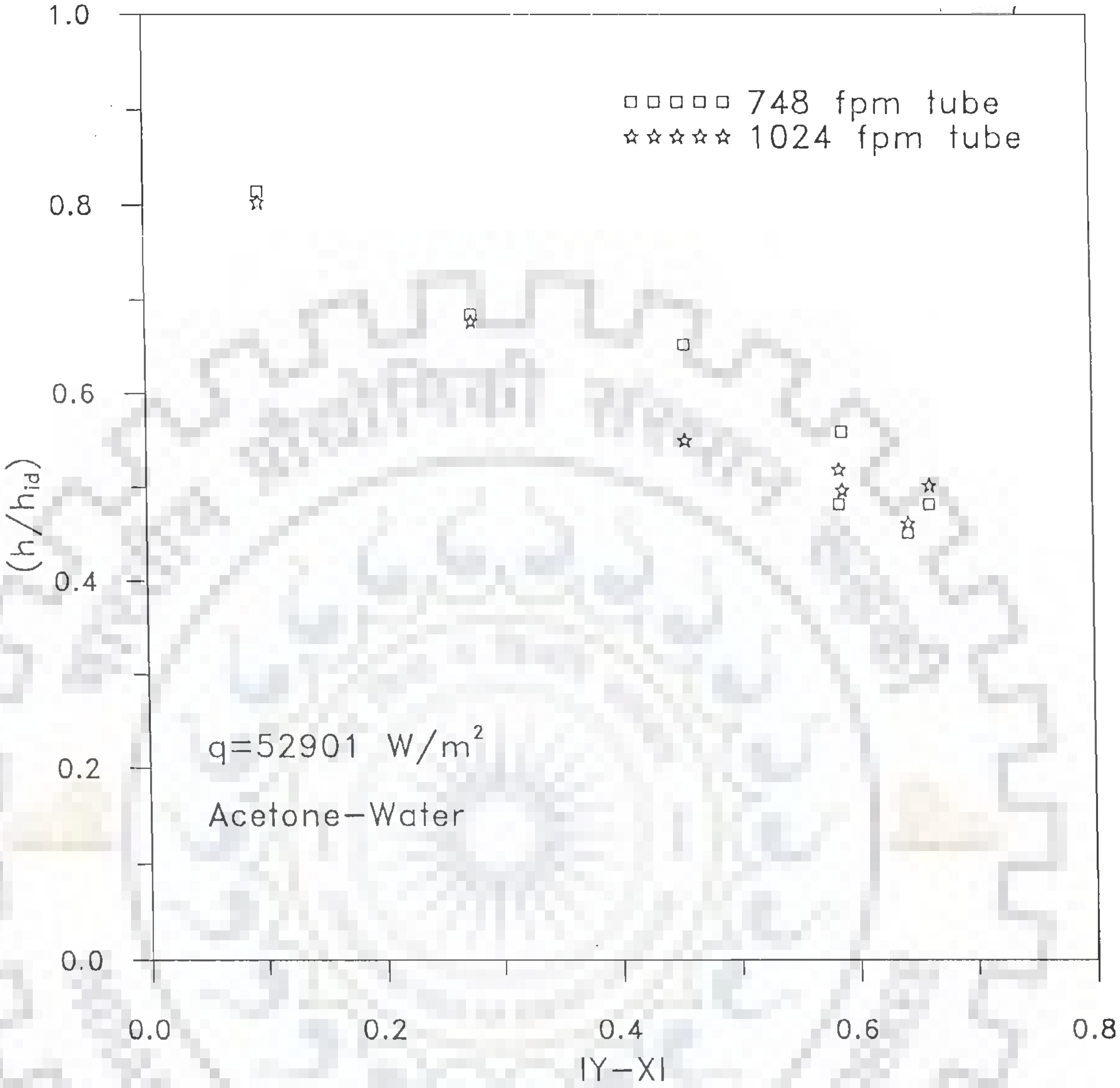


Fig. 6.31 Variation of (h/h$_{id}$) with IY−XI for acetone−water mixture boiling on integral−fin tubes

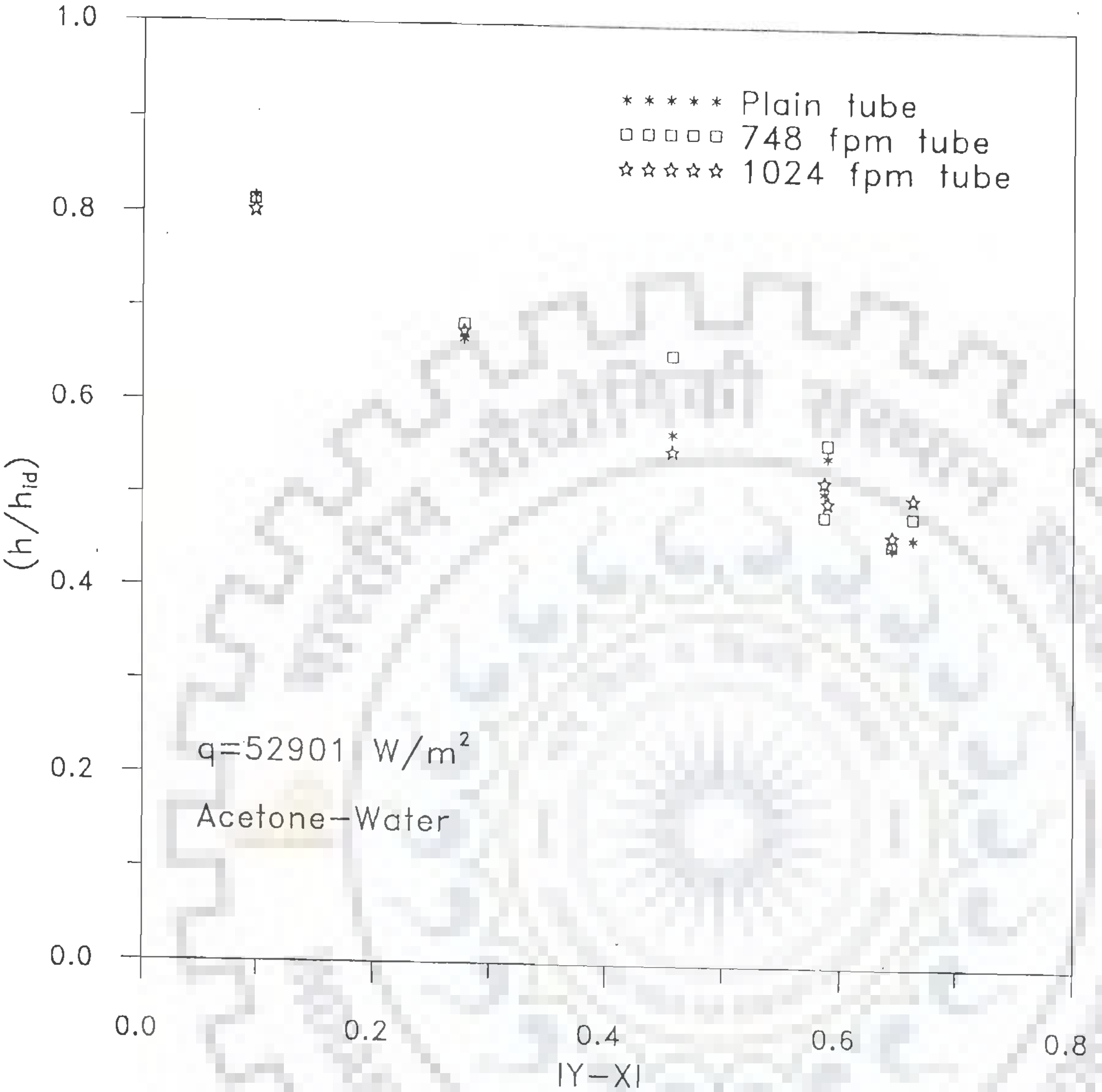


Fig. 6.32 Variation of (h/h$_{id}$) with IY−XI for acetone−water mixture boiling on tubes

tigated, namely; isopropanol-water, and acetone-isopropanol.

6.6 NUCLEATE POOL BOILING OF TERNARY MIXTURES ON INTEGRAL-FIN TUBES

In the following subsection experimental data for nucleate pool boiling heat transfer from a horizontal integral-fin tube to ternary liquid mixtures at atmospheric pressure have been discussed.

6.6.1 EFFECT OF HEAT FLUX ON HEAT TRANSFER COEFFICIENT

Figures 6.33 and 6.34 represent the effect of heat flux on heat transfer coefficient for the boiling of acetone-isopropanol-water ternary mixtures on 748 and 1024 fpm tubes, respectively at atmospheric pressure. From plots, it is clearly seen that boiling heat transfer coefficient changes with heat flux and is represented by an equation similar to Eq.(6-9). Value of constant C_3 depends upon the composition of liquid mixtures and tube geometries. As regards the value of the exponent n, it is found to be 0.68.

All boiling curves (**h vs. q**) of ternary liquid mixtures lie below boiling curves of pure components of the mixture, i.e. acetone, isopropanol, and water. This fact is valid for both integral-fin heating tubes. This shows that in the case of boiling of ternary mixtures also there is a degradation in boiling heat transfer coefficient.

6.6.2 VARIATION OF DEGRADATION FACTOR WITH VAPOUR-LIQUID COMPOSITION DIFFERENCE

Figure 6.35 represents a typical variation in degradation factor, (h/h_{id}) with $|Y-X|$. It is important to mention that term $|Y-X|$ stands for the summation of the vapour-liquid mole fraction difference of the more volatile component $|Y_1-X_1|$, i.e. acetone and that of the intermediate volatile component $|Y_2-X_2|$, i.e isopropanol. The plot is based on experimental data obtained for boiling of acetone-isopropanol-water mixtures on an integral-fin tube having 748 fpm and also on other tube of 1024 fpm at a heat flux of 52901 W/m². Linear mixing law, Eq.(2-16) was used to define ideal heat transfer coefficient, h_{id}.

From the plot, it is easily noticed that for nucleate pool boiling of ternary mixtures on integral-fin tube the value of (h/h_{id})decreases as $|Y-X|$ increases. There is a close agreement in this regard with the boiling characteristics of binary liquid mixtures.

6.7 COMPARISON BETWEEN EARLIER AND PROPOSED CORRELATIONS

Section 2.5 of Chapter 2 details several correlations for the boiling of binary liquid mixtures on a horizontal plain tube, which have been recommended time to time. The important ones are due to Palen-Small (1964), Calus-Leonidopoulos (1974), Stephan-Preusser (1979), Thome (1981), Schlunder (1982), and Thome-Shakir (1987). They have been summarized in following section for ready reference.

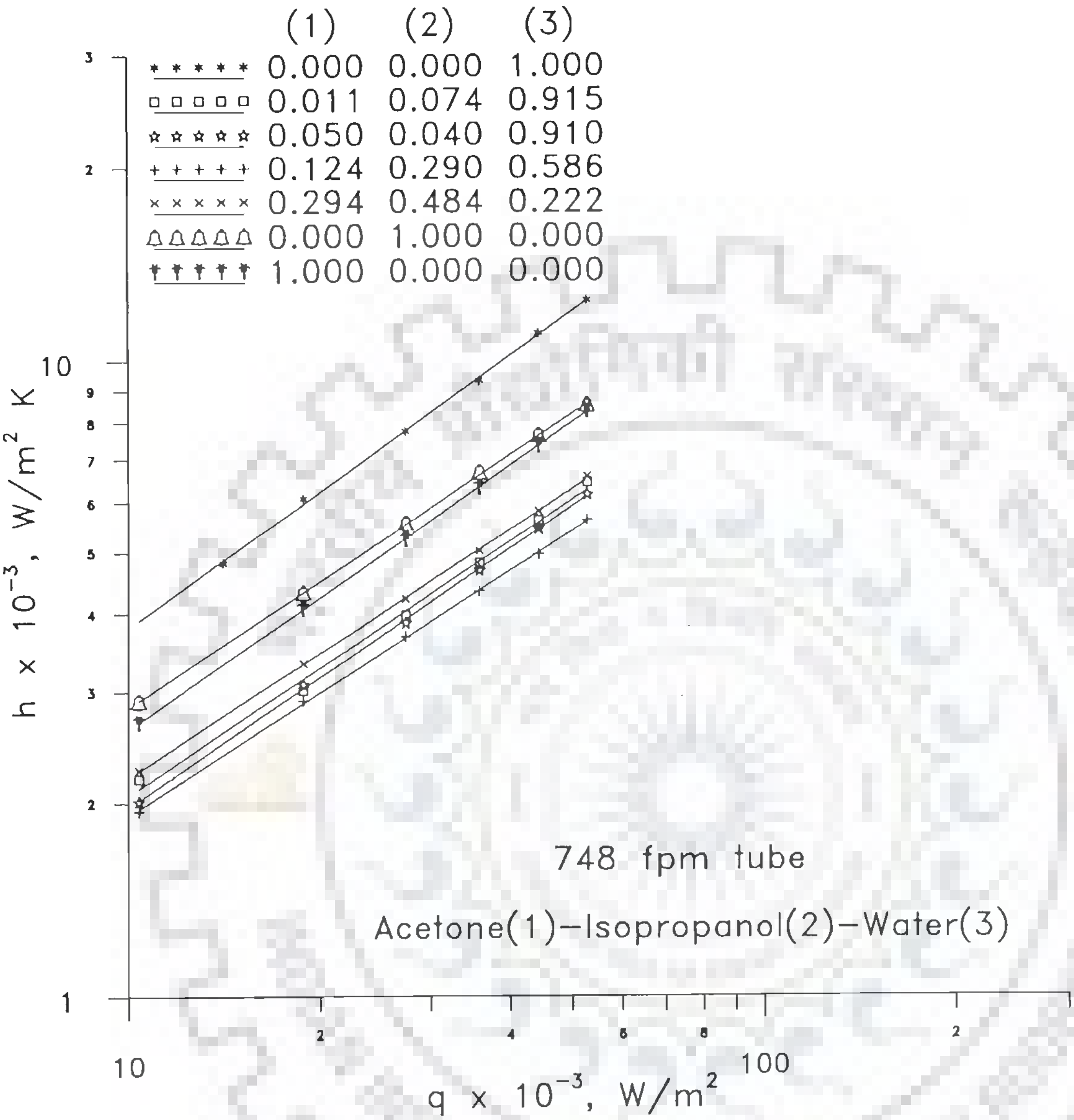


Fig. 6.33 Nucleate boiling curves for ternary mixture on an integral-fin tube

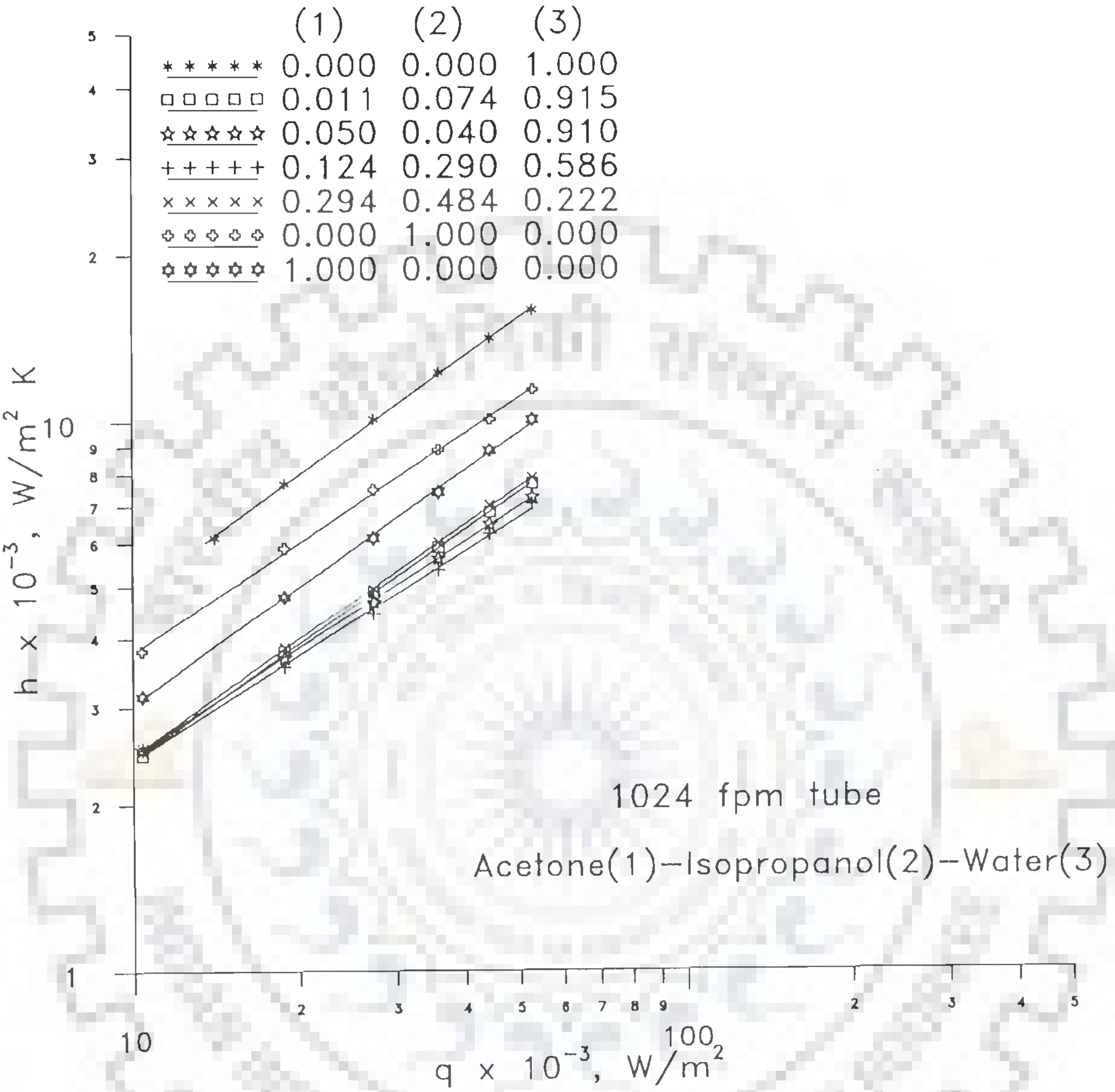


Fig. 6.34 Nucleate boiling curves for ternary mixture on an integral-fin tube

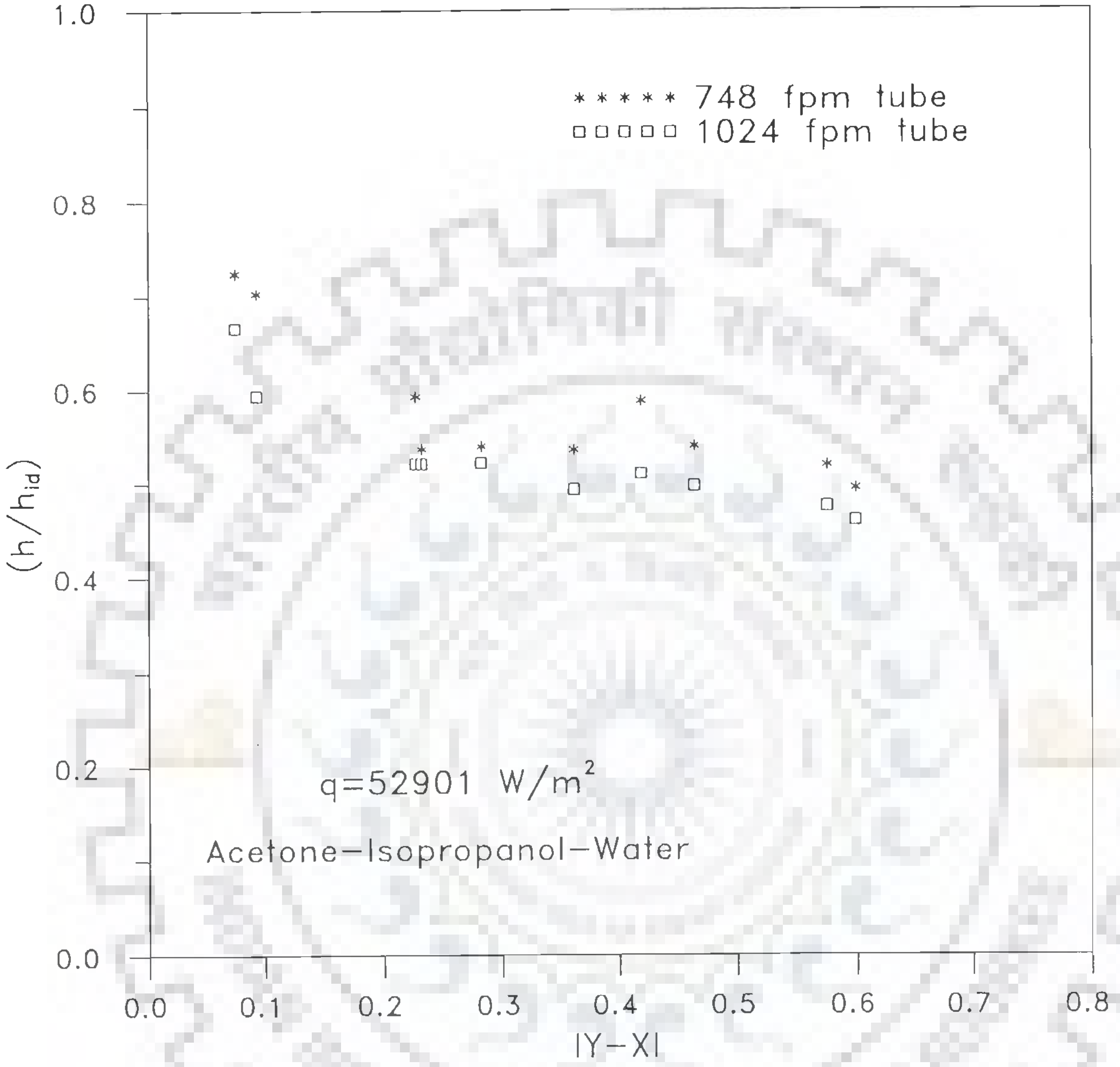


Fig. 6.35 Variation of (h/h_{id}) with $|Y-X|$ for ternary mixture on integral−fin tubes

Palen-Small Correlation

This is the earliest correlation and is considered to be the most practical one, represented by the following equation:

$$h/h_{id} = \exp(-0.027\Delta T_{BR}) \qquad (2\text{-}22)$$

This correlation makes use of the McNelly correlation (1953) for the calculation of h_{id}.

Calus-Leonidpoulos Correlation

Their correlation is based on the fact that there is a rise in the local saturation temperature for the growth of a single spherical bubble in an infinite uniformity superheated binary liquid mixture. Finally, they obtained the following equation for wall superheat:

$$\Delta T/\Delta T_{id} = [1 + (y_1 - x_1)\,(\alpha / \delta)^{0.5}(C_p/h_{fg})\,(dT/dx)] \qquad (2\text{-}23)$$

where ΔT_{id} is $[x_1\Delta T_1 + x_2\,\Delta T_2]$. Knowing the value of ΔT, heat transfer coefficient is calculated using $h = (q/\Delta T)$.

Stephan-Preusser Correlation

They employed dimensional analysis to obtain the following correlation for boiling of liquid mixtures:

$$\begin{aligned} h = {} & [0.0871(k_l/d)(qd/k_lT_s)^{0.674}(\rho_v/\rho_l)^{0.156}(h_{fg}d^2/\alpha^2)^{0.371} \\ & \times(\alpha^2/\sigma d)^{0.35}(\mu Cp/k_l)^{-0.162}]\,[1 + |\Sigma(Y_1 - X_1)(dY/dX)|]^{0.0733} \end{aligned} \qquad (2\text{-}24)$$

where d is bubble departure diameter and is calculated from Eq.(2-20), described in Chapter 2.

Thome Correlation

Accounting for the effect of composition on the stripping of thermal boundary layer due to boiling on a heated surface, the following correlation has been recommended by Thome:

$$(h/h_{id}) = Sn^{7/5} \qquad (2\text{-}25)$$

where Sn is the the Scriven number, Eq.(2-8).

The value of h_{id} is calculated using linear mixing law.

Schlunder Correlation

This is the theoretically-derived correlation accounting the effect of heat flux on mass diffusion during mixture boiling and is as follows:

$$(h/h_{id}) = [1 + (h_{id}/q)\,(T_{s2} - T_{s1})(Y_1 - X_1)\,\{1 - \exp(-B_oq/\rho_l\,\beta_l\,h_{fg})\}]^{-1} \qquad (2\text{-}26)$$

where β_l is mass transfer coefficient in the liquid phase , and B_o an empirical scaling parameter.

Thome-Shakir Correlation

Their correlation is a modification of the Schlunder correlation and its final form is as follows:

$$(h/h_{id}) = [1 + \{(h_{id}/q)\,\Delta T_{BR}\}\{1 - \exp(-B_oq/\rho_l\,\beta_l\,h_{fg})\}]^{-1} \qquad (2\text{-}30)$$

For the calculation of h_{id}, they recommend that the Stephan and Abdelsalam correlation (1980) should be employed.

A Remark on Correlations

An examination of various correlations as above clearly shows that for the calculation of heat transfer coefficients from a horizontal plain tube to pool of saturated binary mixtures a knowledge of ideal heat transfer coefficient, h_{id} is essential. For all the correlations, except those due to Palen-Small, Stephan-Preusser, and Thome-shakir, the value of h_{id} is based on linear mixing law as follows:

$$(1/h_{id}) = \sum_{i=1}^{i=n} (X_i / h_i) \qquad (2\text{-}16)$$

where h_i represents the heat transfer coefficient of the pure component liquid at the same heat flux as the mixture.

The most convenient method to carry out a comparison between existing correlations and the one derived in the present investigation, Eq.(5-12) perhaps lies in assessing their relative accuracy as regards the prediction capability of heat transfer coefficient. This can be best appreciated by comparing the predicted values of heat transfer coefficient with the experimental ones. Experimental data available in literature are due to Calus and Leonidopoulos (1974), Pandey (1982), Uhlig and Thome (1985), Shakir et al.(1985), and Bajorek et al.(1989).

6.7.1 PREDICTION OF BINARY MIXTURE BOILING DATA ON PLAIN TUBE

First of all, the values of boiling heat transfer coefficient were calculated by using various correlations for above mentioned and present investigation. For determination of physico-thermal properties, the procedure described in Appendix C was employed.

Predicted values so calculated from Eqs.(2-22), (2-23), (2-24),(2-25),(2-26), and (2-30) have been plotted as ordinate against the experimental ones as abscissa in Figures 6.36 through 6.41, respectively. Table 6.4 provides respective values of standard deviation (S.D.) and mean absolute error (M.A.E.). Essence of comparison is that the Thome-Shakir correlation yields the best result within +25% and -35%, error followed by Palen-Small correlation within ±30%. The Stephan-Preusser correlation also predicts equally well, however, within ±35%. As regards correlation due to Calus-Leonidopoulos, it over predictes. Similar is the situation with correlation of Thome and that of Schlunder

At this juncture it is important to refer back to plots of Figures 5.2 and 5.3. These plots have been drawn using those experimental data points, which have been used for Figures 6.36 through 6.41 and predicted values from the present correlation, Eq.(5-12). Data are correlated within ±25%, proving that the present correlation is perhaps the better than the best amongst all the existing correlations as regard the prediction capability.

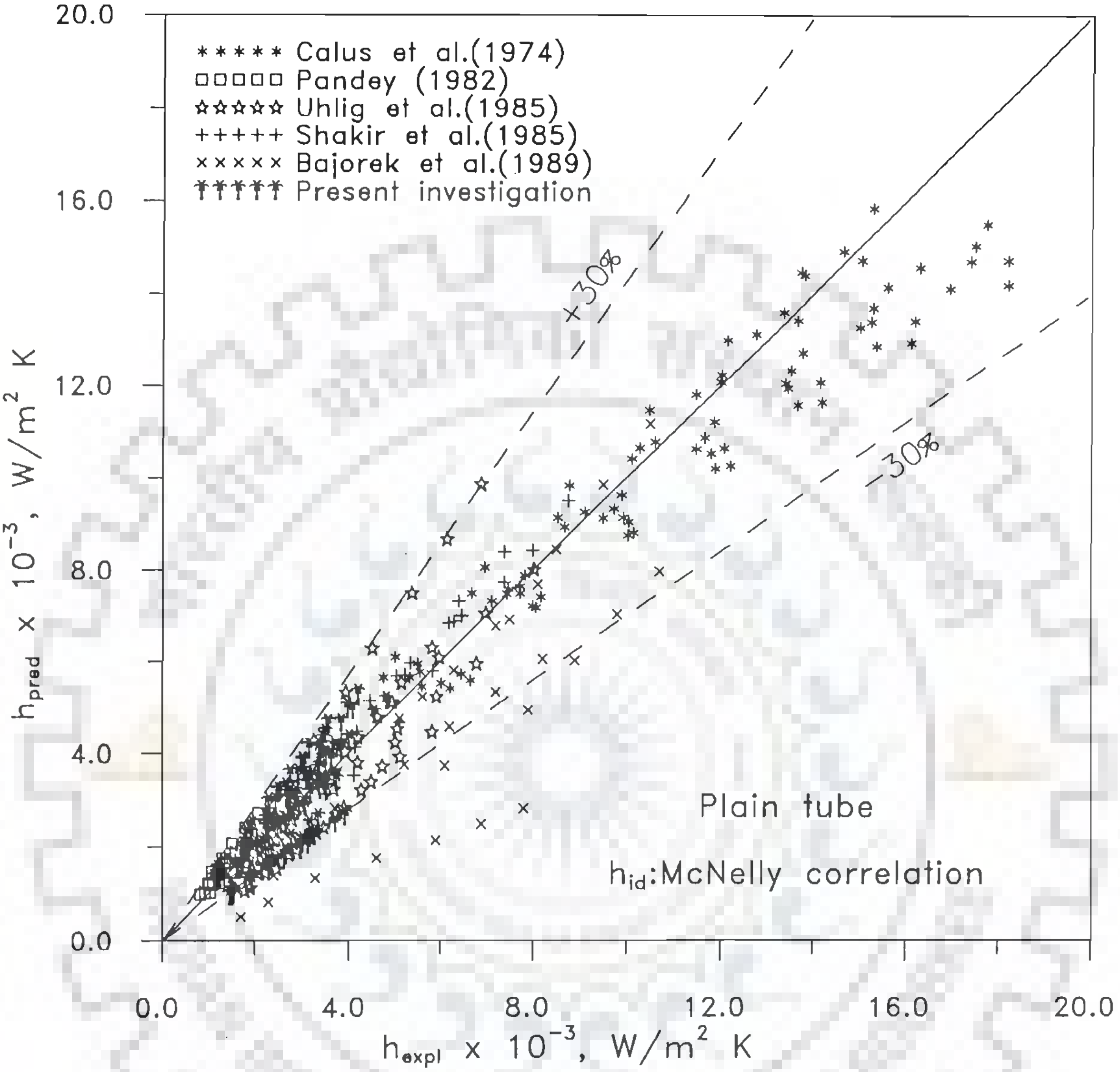


Fig. 6.36 Comparison of binary mixtures boiling data with the Palen–Small correlation

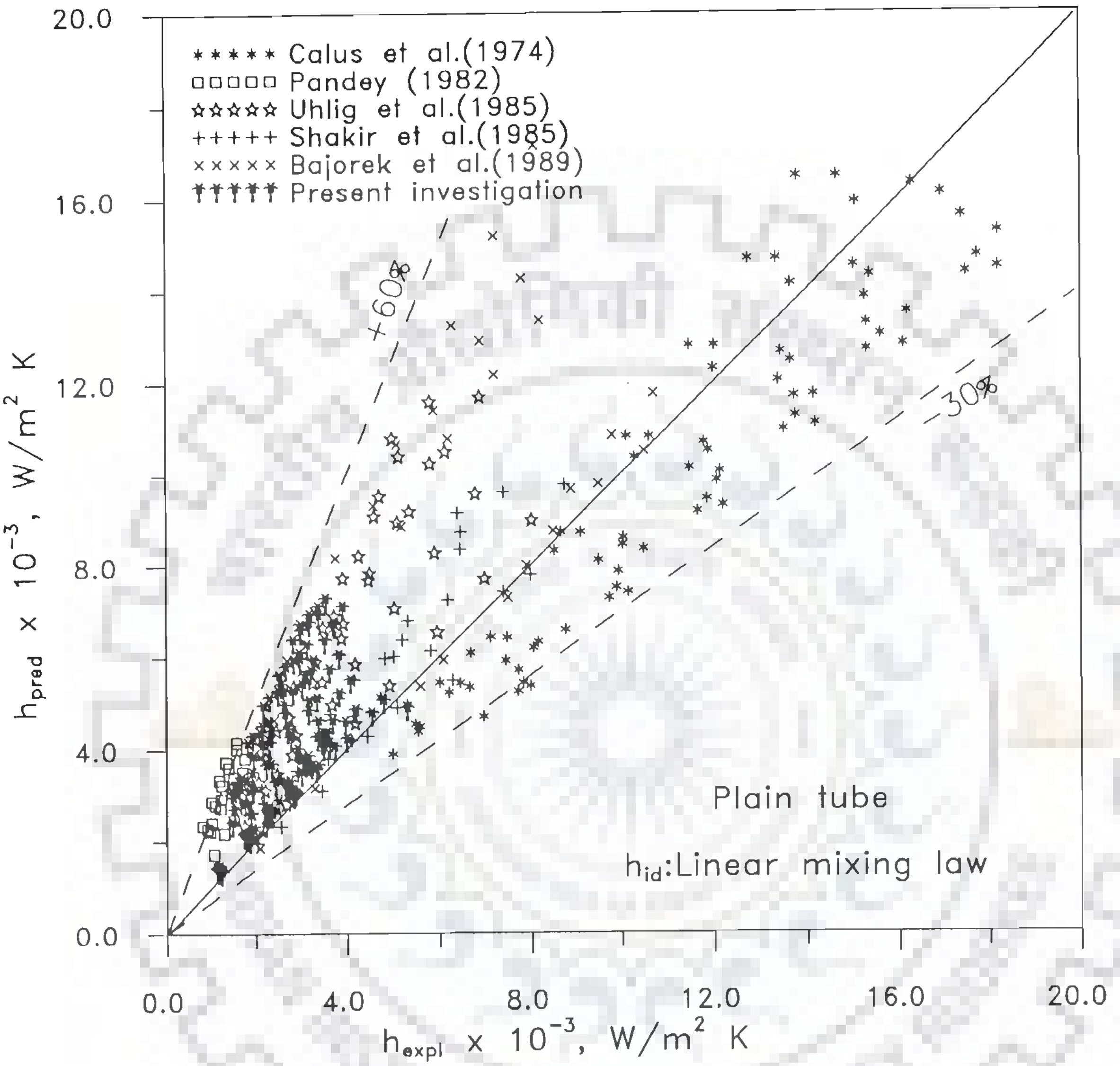


Fig. 6.37 Comparison of binary mixtures boiling data with the Calus–Leonidopoulos correlation

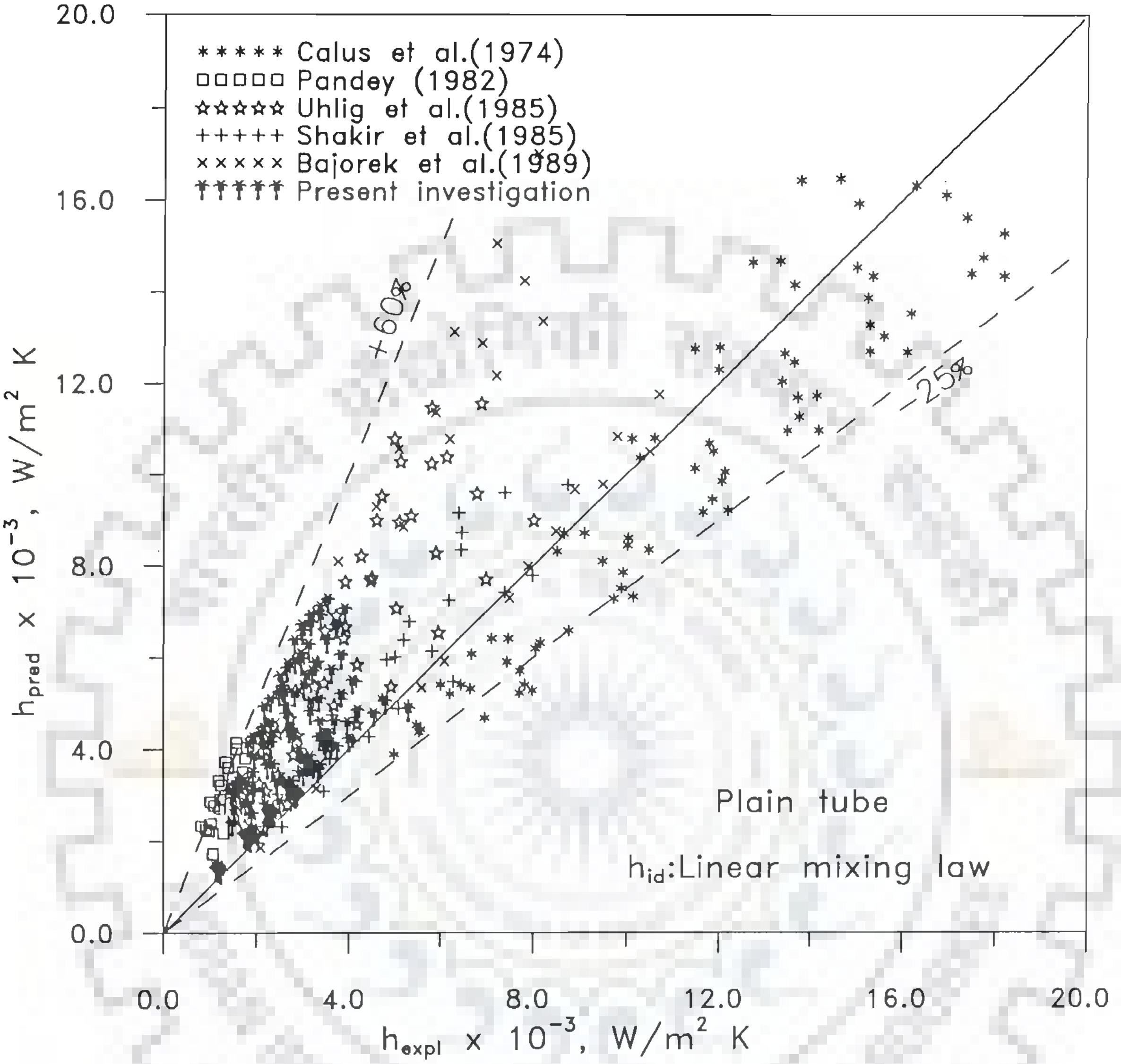


Fig. 6.38 Comparison of binary mixtures boiling data with the Thome correlation

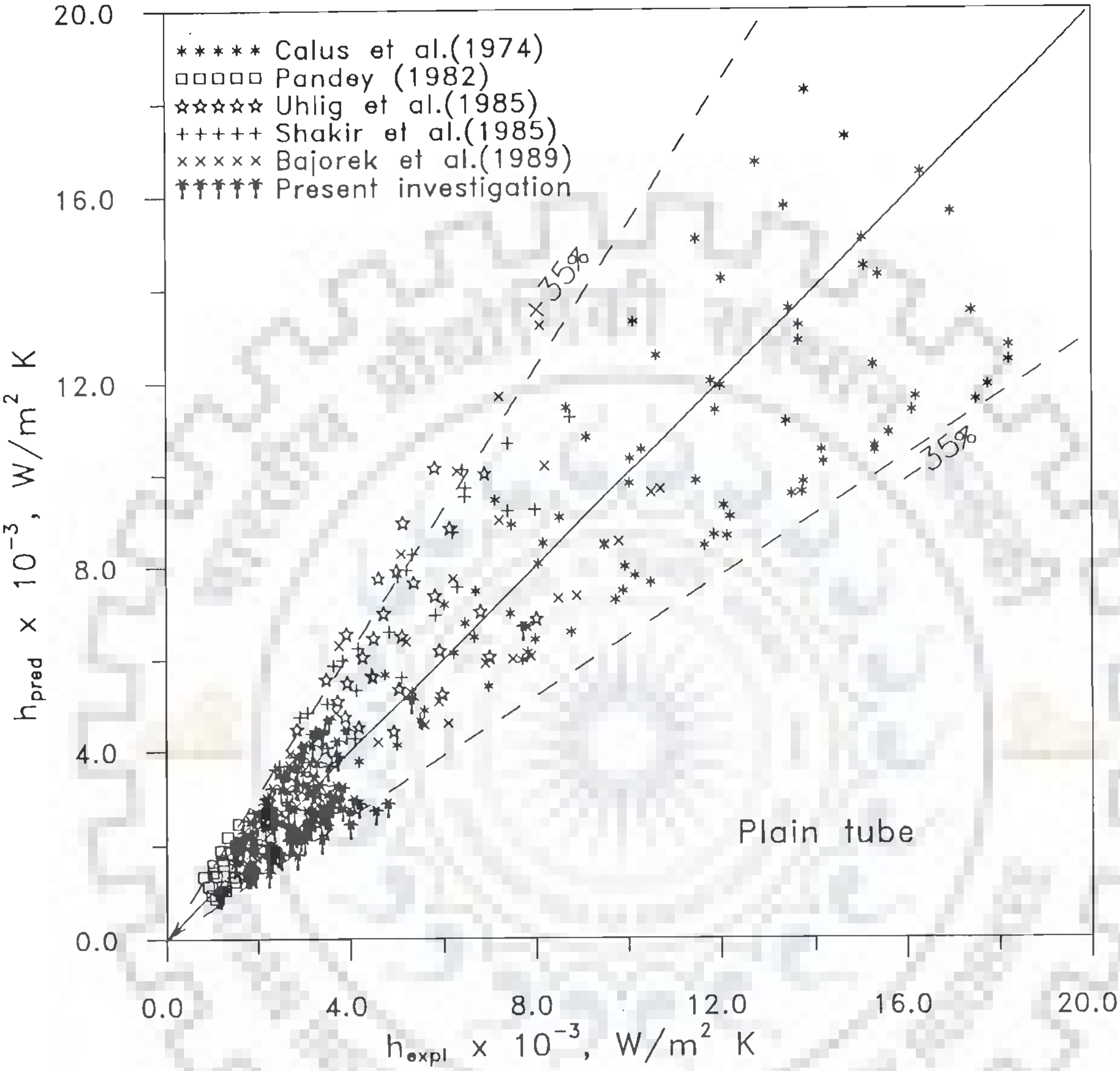


Fig. 6.39 Comparison of binary mixtures boiling data with the Stephan–Preusser correlation

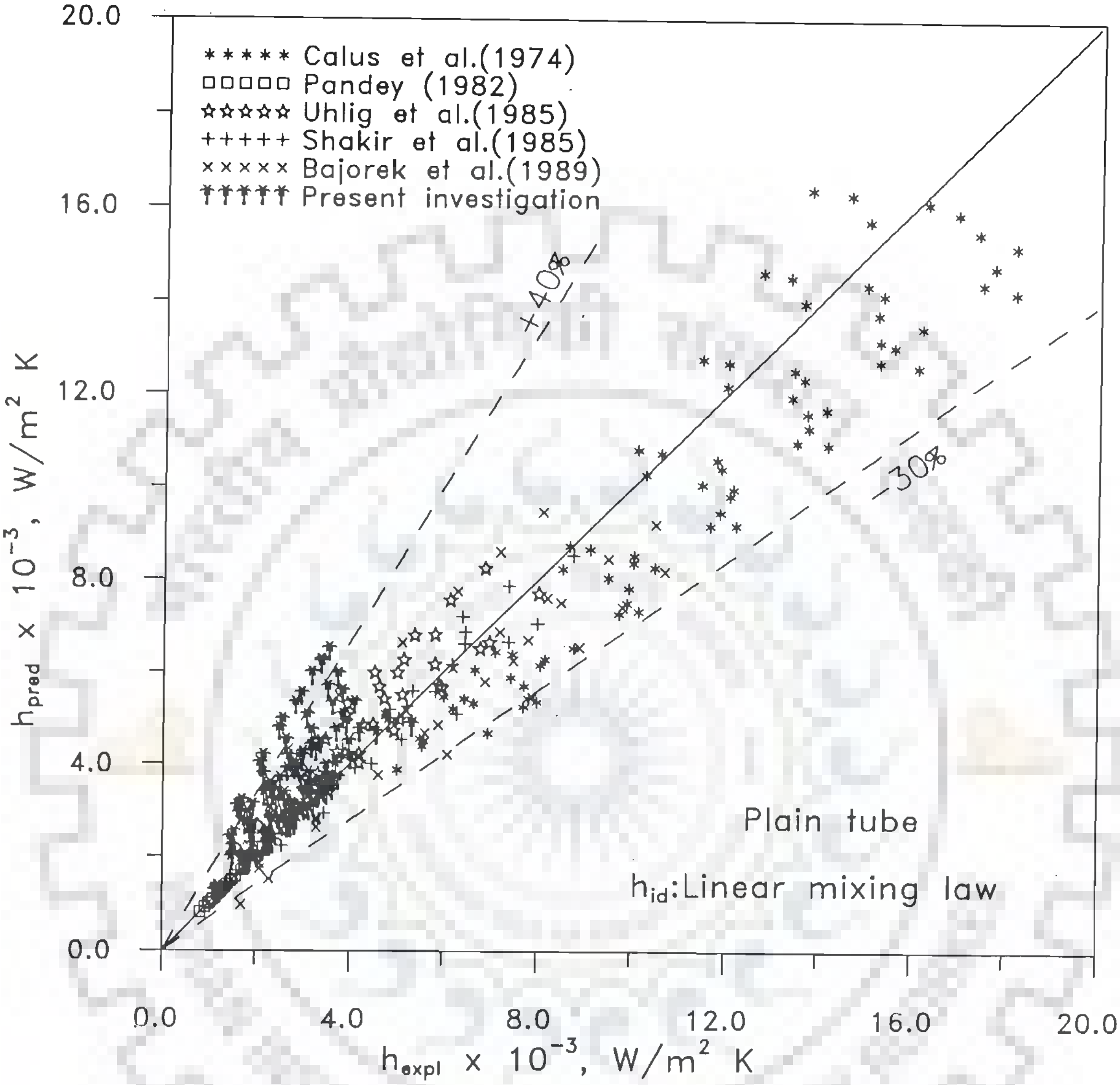


Fig. 6.40 Comparison of binary mixture boiling data with the Schlunder correlation

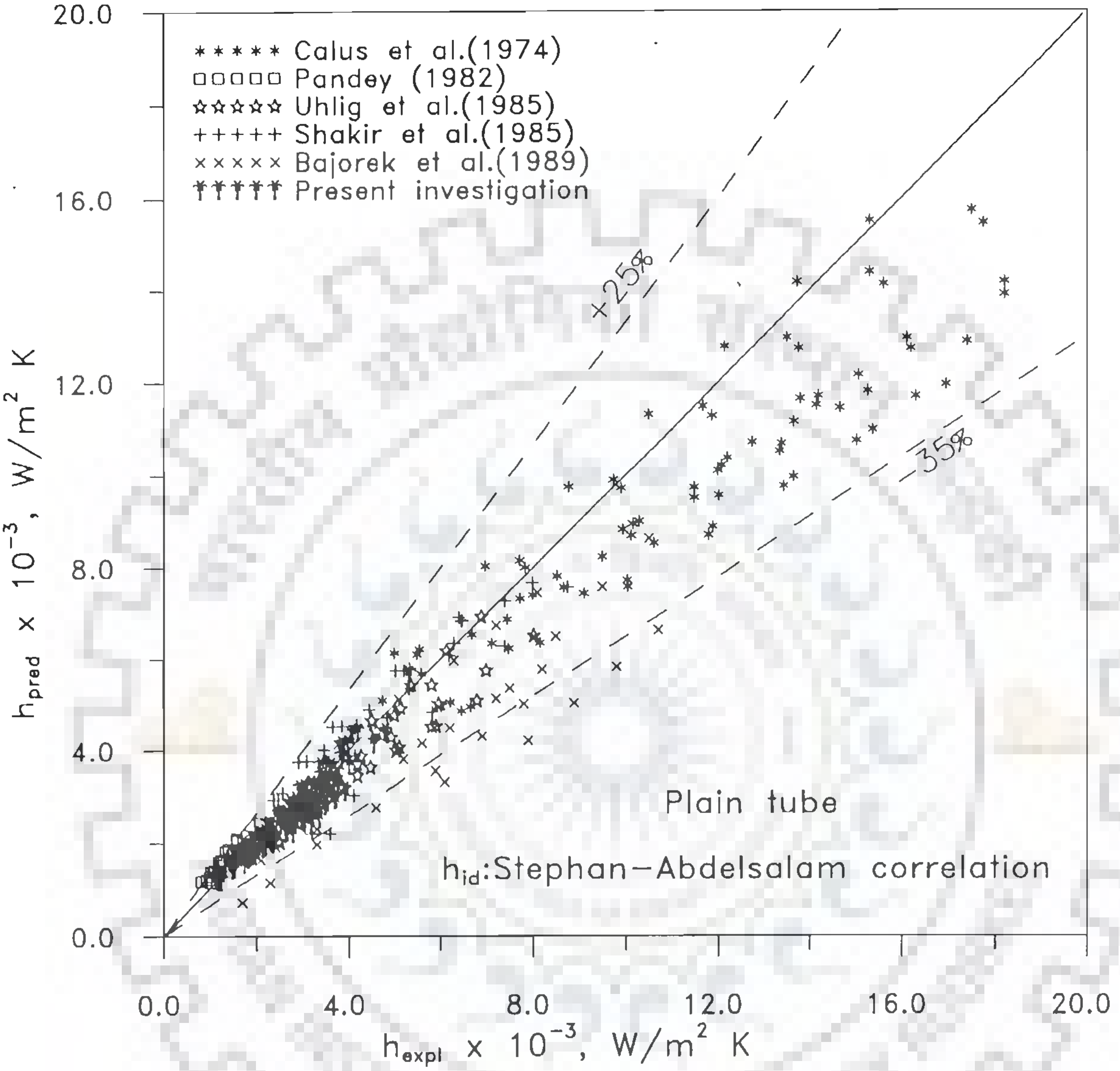


Fig. 6.41 Comparison of binary mixtures boiling data with the Thome–Shakir correlation

Table 6.4 Standard deviation and mean absolute error values for binary mixtures boiling data on plain tube

Correlation	Eq.	Experimental data	Mixtures	Points	(S.D.)%	(M.A.E.)%
Palen-Small	(2-22)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	10.57	9.23
		Pandey (1982)	Isopropanol-Water	25	20.95	18.02
		Uhlig-Thome(1985)	Acetone-Water	48	23.73	19.14
		Shakir-Thome(1985)	Methanol-Water	36	14.41	12.58
		Bajorek et al.(1989)	Acetone-Water	35	35.76	28.47
		Present investigation	Acetone-Water	42	28.70	26.82
		Present investigation	Isopropanol-Water	48	13.88	11.17
		Present investigation	Acetone-Isopropanol	42	16.06	14.09
Calus-Leonidopoulos	(2-23)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	21.59	15.32
		Pandey (1982)	Isopropanol-Water	25	54.55	126.69
		Uhlig-Thome(1985)	Acetone-Water	48	37.20	58.32
		Shakir-Thome(1985)	Methanol-Water	36	16.78	19.49
		Bajorek et al.(1989)	Acetone-Water	35	37.84	58.45
		Present investigation	Acetone-Water	42	43.80	78.02
		Present investigation	Isopropanol-Water	48	36.41	56.41
		Present investigation	Acetone-Isopropanol	42	13.06	14.09
Stephan-Preusser	(2-24)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	20.89	19.85
		Pandey (1982)	Isopropanol-Water	25	34.56	30.23
		Uhlig-Thome(1985)	Acetone-Water	48	34.50	27.46
		Shakir-Thome(1985)	Methanol-Water	36	38.98	34.81
		Bajorek et al.(1989)	Acetone-Water	35	30.17	24.28
		Present investigation	Acetone-Water	42	19.56	16.06
		Present investigation	Isopropanol-Water	48	27.98	25.55
		Present investigation	Acetone-Isopropanol	42	24.95	24.63
Thome	(2-25)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	21.89	15.85
		Pandey (1982)	Isopropanol-Water	25	54.47	126.52
		Uhlig-Thome(1985)	Acetone-Water	48	36.50	57.44
		Shakir-Thome(1985)	Methanol-Water	36	16.78	19.07
		Bajorek et al.(1989)	Acetone-Water	35	37.76	57.96
		Present investigation	Acetone-Water	42	41.48	78.06
		Present investigation	Isopropanol-Water	48	35.32	56.11
		Present investigation	Acetone-Isopropanol	42	12.39	14.77

Table 6.4 , continued

Table 6.4, continued

Schlunder	(2-26)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	17.92	15.75
		Pandey (1982)	Isopropanol-Water	25	117.14	110.92
		Uhlig-Thome(1985)	Acetone-Water	48	19.72	15.55
		Shakir-Thome(1985)	Methanol-Water	36	9.17	7.32
		Bajorek et al.(1989)	Acetone-Water	35	26.76	22.96
		Present investigation	Acetone-Water	42	31.48	28.06
		Present investigation	Isopropanol-Water	48	47.32	38.11
		Present investigation	Acetone-Isopropanol	42	4.96	4.64
Thom-Shakir	(2-30)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	17.52	14.85
		Pandey (1982)	Isopropanol-Water	25	24.84	21.95
		Uhlig-Thome(1985)	Acetone-Water	48	12.74	10.05
		Shakir-Thome(1985)	Methanol-Water	36	16.96	14.09
		Bajorek et al.(1989)	Acetone-Water	35	30.64	25.71
		Present investigation	Acetone-Water	42	15.67	13.88
		Present investigation	Isopropanol-Water	48	6.39	5.84
		Present investigation	Acetone-Isopropanol	42	9.9	8.73

Arising out from above result, it was thought intuitively that the existing correlations might yield better predictions, if h_{id} is calculated by using the Stephan-Abdelsalam correlation rather than using linear mixing law method. In fact, the Stephan-Abdelsalam correlation is based on pertinent physico-thermal properties of boiling mixtures. Accordingly, an exercise was undertaken using the Stephan-Abdelsalam correlation, Eq.(5-6) for estimation of h_{id}. Predicted values of heat transfer coefficient from Calus-Leonidopoulos (1974), Thome (1980), and Schlunder (1982) are compared in Figures 6.42 through 6.44 for individual correlations. The corresponding values of standard deviation, (S.D.) and mean absolute error, (M.A.E.) so obtained for all data points are given in Table 6.5

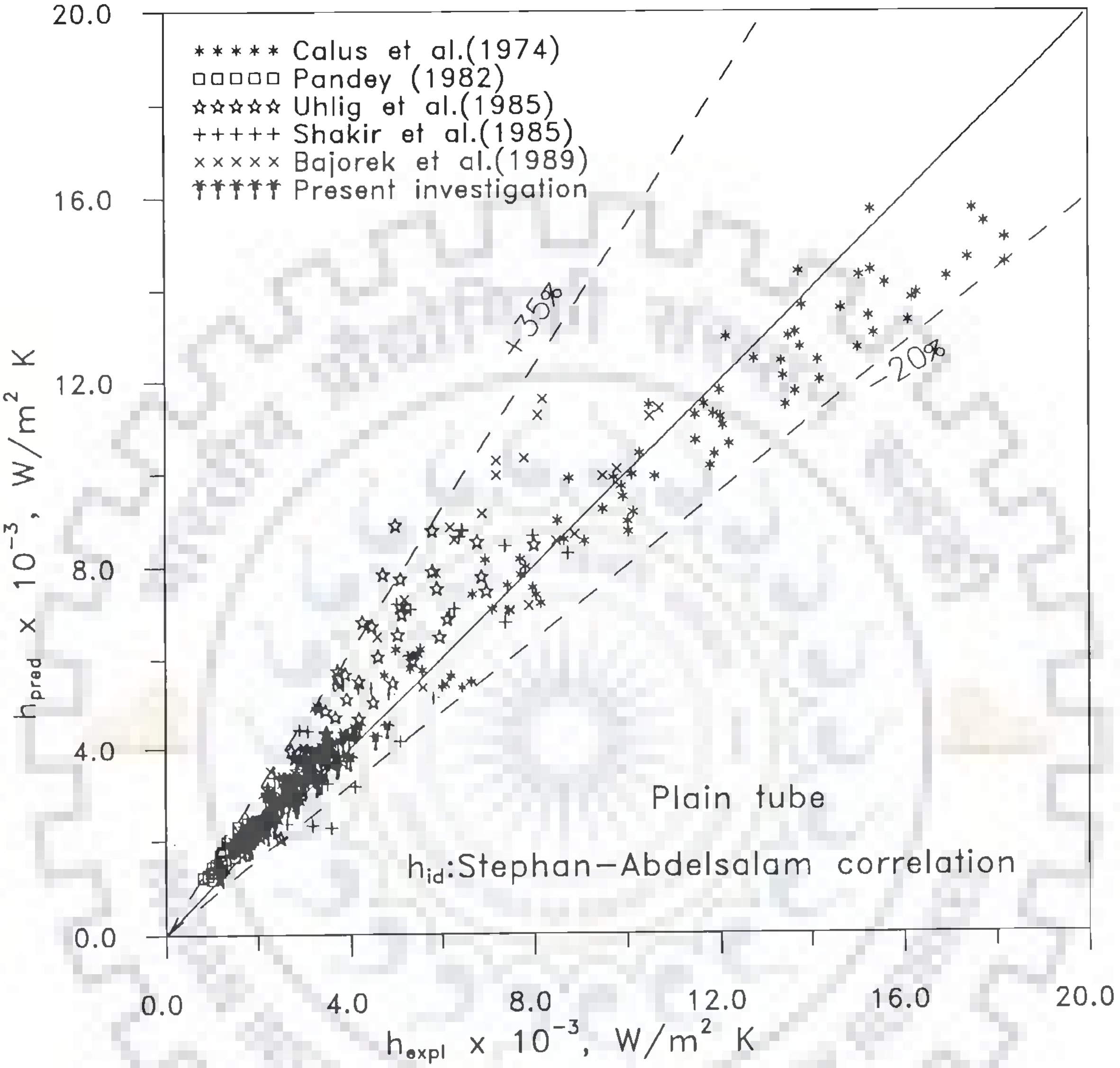


Fig. 6.42 Comparison of binary mixture boiling data with the Calus–Leonidopoulos correlation

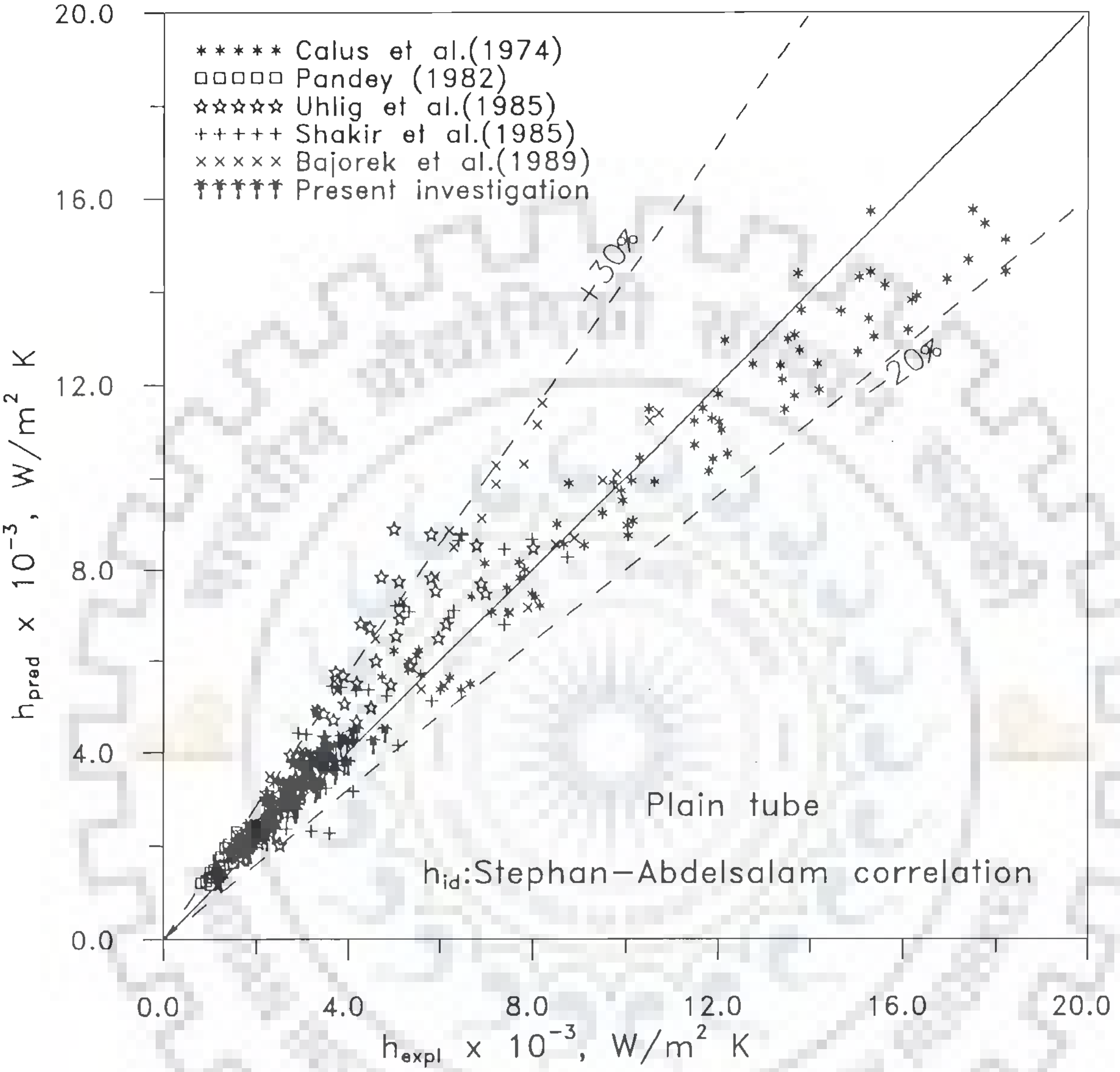


Fig. 6.43 Comparison of binary mixtures boiling data with the Thome correlation

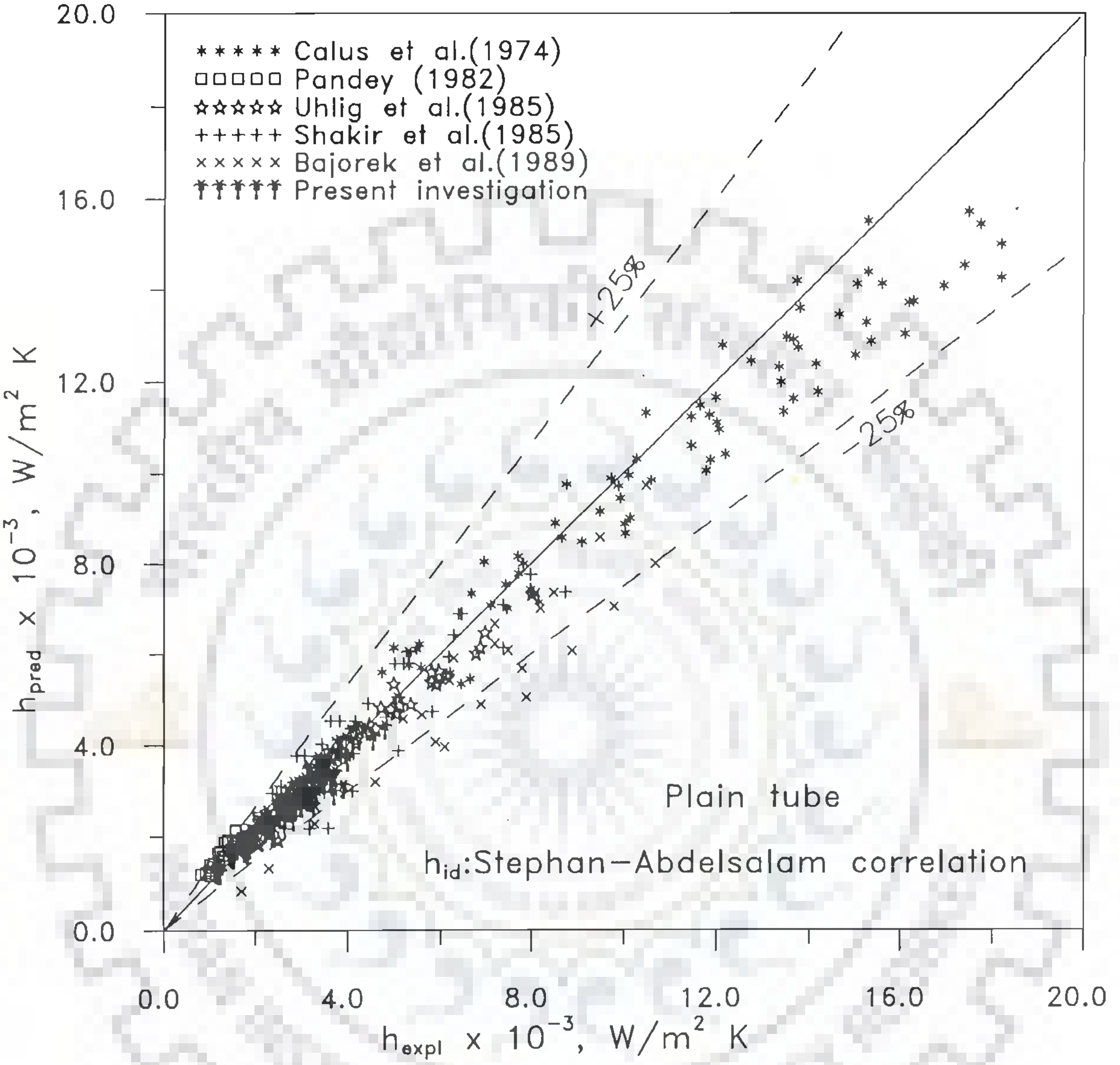


Fig. 6.44 Comparison of binary mixtures boiling data with the Schlunder correlation

Table 6.5 Standard deviation and mean absolute error values for binary mixtures boiling data on plain tube using Stephan-Abdelsalam correlation for h_{id}

Correlation	Eq.	Experimental data	Mixtures	Points	(S.D.)%	(M.A.E.)%
Calus-Leonidopoulos	(2-23)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	11.52	9.23
		Pandey (1982)	Isopropanol-Water	25	22.55	28.91
		Uhlig-Thome(1985)	Acetone-Water	48	23.21	28.28
		Shakir-Thome(1985)	Methanol-Water	36	23.45	24.83
		Bajorek et al.(1989)	Acetone-Water	35	21.57	24.79
		Present investigation	Acetone-Water	42	15.38	16.16
		Present investigation	Isopropanol-Water	48	10.50	10.22
		Present investigation	Acetone-Isopropanol	42	9.45	9.27
Thome	(2-25)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	11.65	9.04
		Pandey (1982)	Isopropanol-Water	25	22.54	28.85
		Uhlig-Thome(1985)	Acetone-Water	48	23.00	27.91
		Shakir-Thome(1985)	Methanol-Water	36	23.35	24.79
		Bajorek et al.(1989)	Acetone-Water	35	21.27	24.56
		Present investigation	Acetone-Water	42	15.33	16.15
		Present investigation	Isopropanol-Water	48	10.50	10.29
		Present investigation	Acetone-Isopropanol	42	9.20	9.49
Schlunder	(2-26)	Calus-Leonidopoulos(1974)	n-Propanol-Water	77	11.12	9.78
		Pandey (1982)	Isopropanol-Water	25	27.33	24.64
		Uhlig-Thome(1985)	Acetone-Water	48	6.96	6.15
		Shakir-Thome(1985)	Methanol-Water	36	17.10	14.11
		Bajorek et al.(1989)	Acetone-Water	35	23.53	18.76
		Present investigation	Acetone-Water	42	11.95	10.31
		Present investigation	Isopropanol-Water	48	6.23	5.12
		Present investigation	Acetone-Isopropanol	42	8.77	7.94

From the values of standard deviation and mean absolute error as recorded in Tables 6.4 and 6.5, it is observed that the prediction capability of individual correlations is improved. Further, from the plots of Figures 6.42 through 6.44, it is noted that amongst the existing correlations, the Schlunder correlation proves to be the best as it correlates data within ±25%. Thus, the Schlunder correlation acquires the same predictive capability as that of the present correlation, Eq.(5-12), provided that ideal heat transfer coefficient, h_{id} is calculated from the Stephan-Abelsalam correlation, Eq.(5-6). Hence, the use of the Stephan-Abdelsalam correlation instead of the linear mixing law is recommended to calculate ideal heat transfer coefficient, h_{id}. This, in fact, is expected as the Stephan-Abdelsalam correlation does employ pertinent physico-thermal properties of boiling mixtures in the estimation of ideal heat transfer coefficient.,

6.7.2 PREDICTION OF TERNARY MIXTURE BOILING DATA ON PLAIN TUBE

The objective of this subsection is to evaluate the existing correlations for nucleate pool boiling heat transfer in ternary mixtures by comparing their predictions with the experimental data.

Correlations due to Palen-Small (1964), Eq.(2-22) Schlunder (1982), Eq.(2-26) and Thome-Shakir (1987), Eq.(2-30) applicable to binary mixture boiling, have been claimed for estimation of boiling heat transfer coefficients of ternary mixtures also. However, they have not been tested satisfactory due to lack of sufficient data for boiling of ternary mixtures.

The respective predictions were obtained from correlations, represented by Eqs.(2-22), (2-26), and (2-30).Predicted values, h_{pred} have been compared with the experimental ones, h_{expl} of the present investigation for the boiling of acetone-isopropanol-water ternary mixtures on plain tube at atmospheric pressure in Figures 6.45 through 6.47. From plots it is clearly noted that correlation due to Palen and Small predicts reasonably well within +15% and -25%, whereas correlation due to Schlunder overpredicts within a maximum of +40%, and that of Thome and Shakir underpredicts within -15% The standard deviation, (S.D.) and mean absolute error, (M.A.E.), for these correlations are listed in Table 6.6.

Table 6.6 Standard deviation and mean absolute error values for ternary mixtures boiling dat on plain tube

Correlation	Eq.	Points	(S.D.)%	(M.A.E.)%
Palen-Small	(2-22)	72	19.30	16.65
Schlunder	(2-29)	60	37.14	29.91
Thom-Shakir	(2-30)	72	8.63	7.51

At this stage, it is important to point out that correlation due to Schlunder makes use of linear mixing law, Eq.(2-16) to calculate h_{id} rather than the Stephan-Abdelsalam correlation, Eq.(5-6). The prediction capability of correlation due to Schlunder got improved when h_{id} was calculated from Eq.(5-6) in case of binary mixture boiling. This is clearly explained in section 6.7.1. Hence, an attempt was made to predict heat transfer coefficient of ternary mixture from the Schlunder correlation using h_{id} from Eq.(5-6). Pedictions from this correlation are found to be greatly improved as shown in Figure 6.48 within -15%. The standard deviation and mean absolute error have become 10.74% and 8.34 %, respectively.

It is important to point out that the proposed correlation, Eq.(5-12) has been extended for predicting heat transfer coefficient of ternary mixture boiling, as described in section 5.3 and predicted values have been compared with experimental values in Figure 5.5 within ±13%.

To decide which correlation between proposed correlation, Eq.(5-12) and existing correlation predicts better. For this plots of Figures 5.5, 6.45, 6.46, 6.47 and 6.48 were re-examined. It is easily seen that present correlation seems to be the best for prediction of heat transfer coefficient for boiling of ternary mixture of acetone-isopropanol-water on a horizontal plain tube as it predicts with minimum error, less complicated, and less computational efforts.

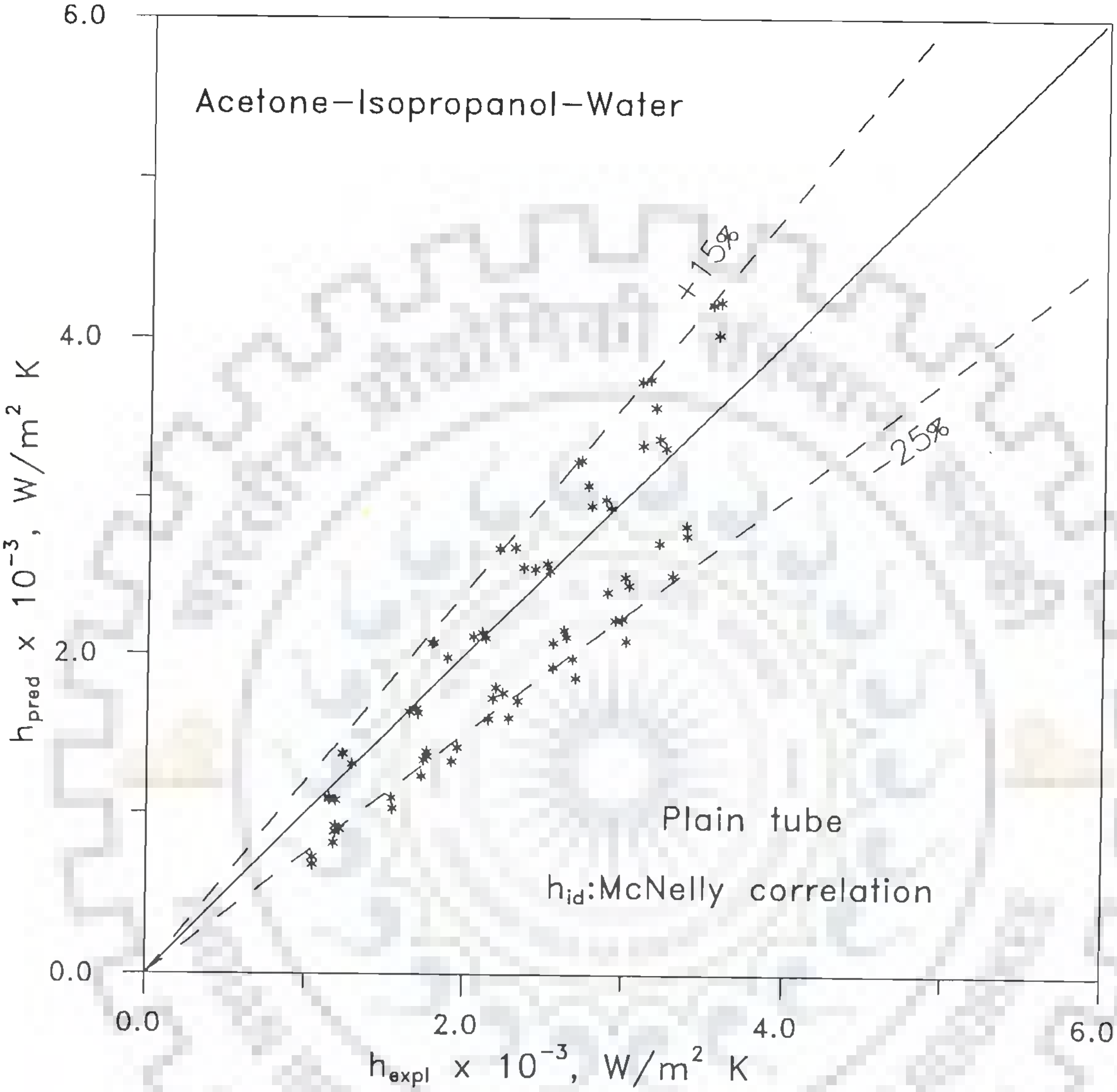


Fig. 6.45 Comparison of ternary mixture boiling data with the Palen–Small correlation

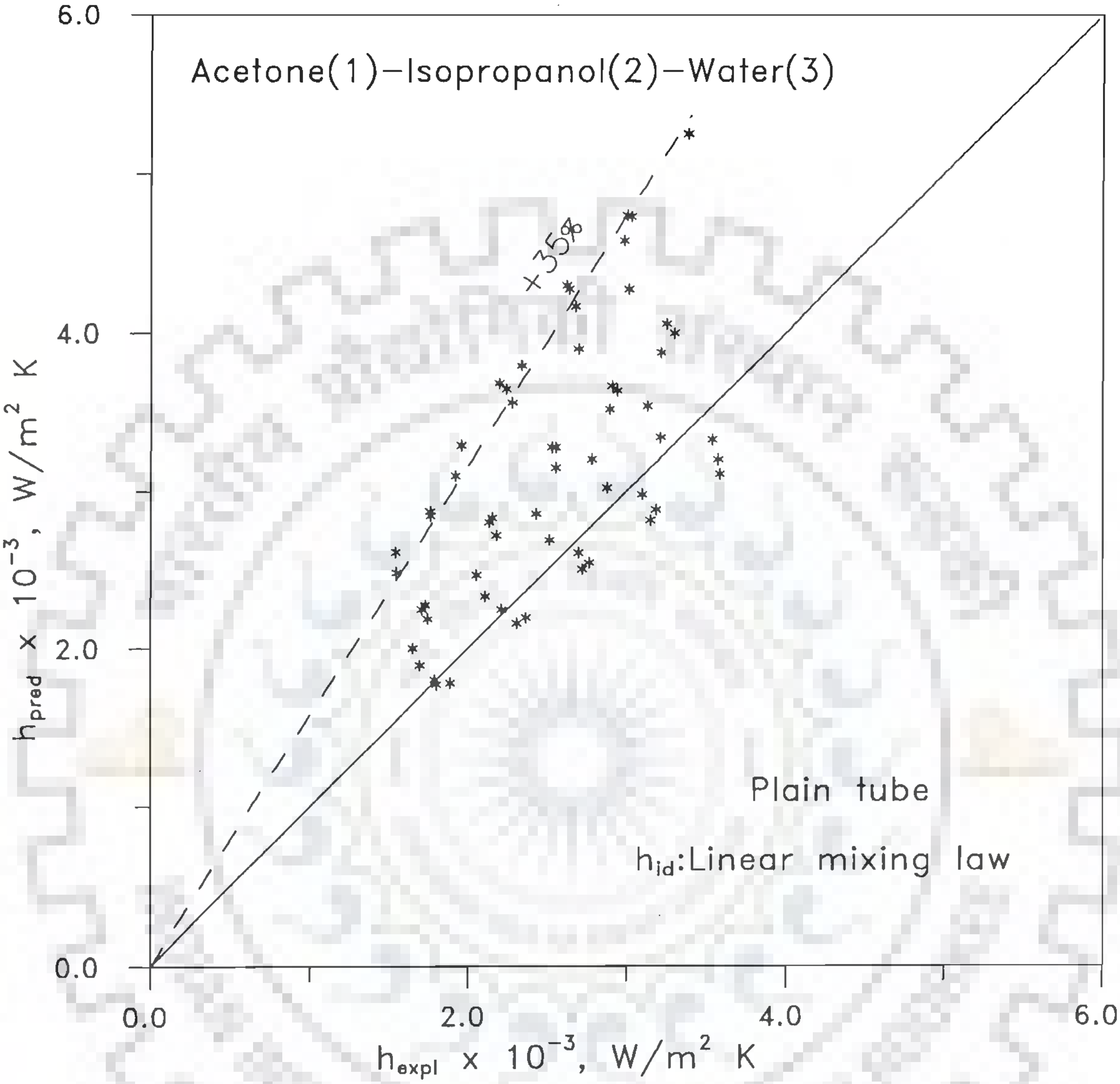


Fig. 6.46 Comparison of tenary mixture boiling data with the Schlunder correlation

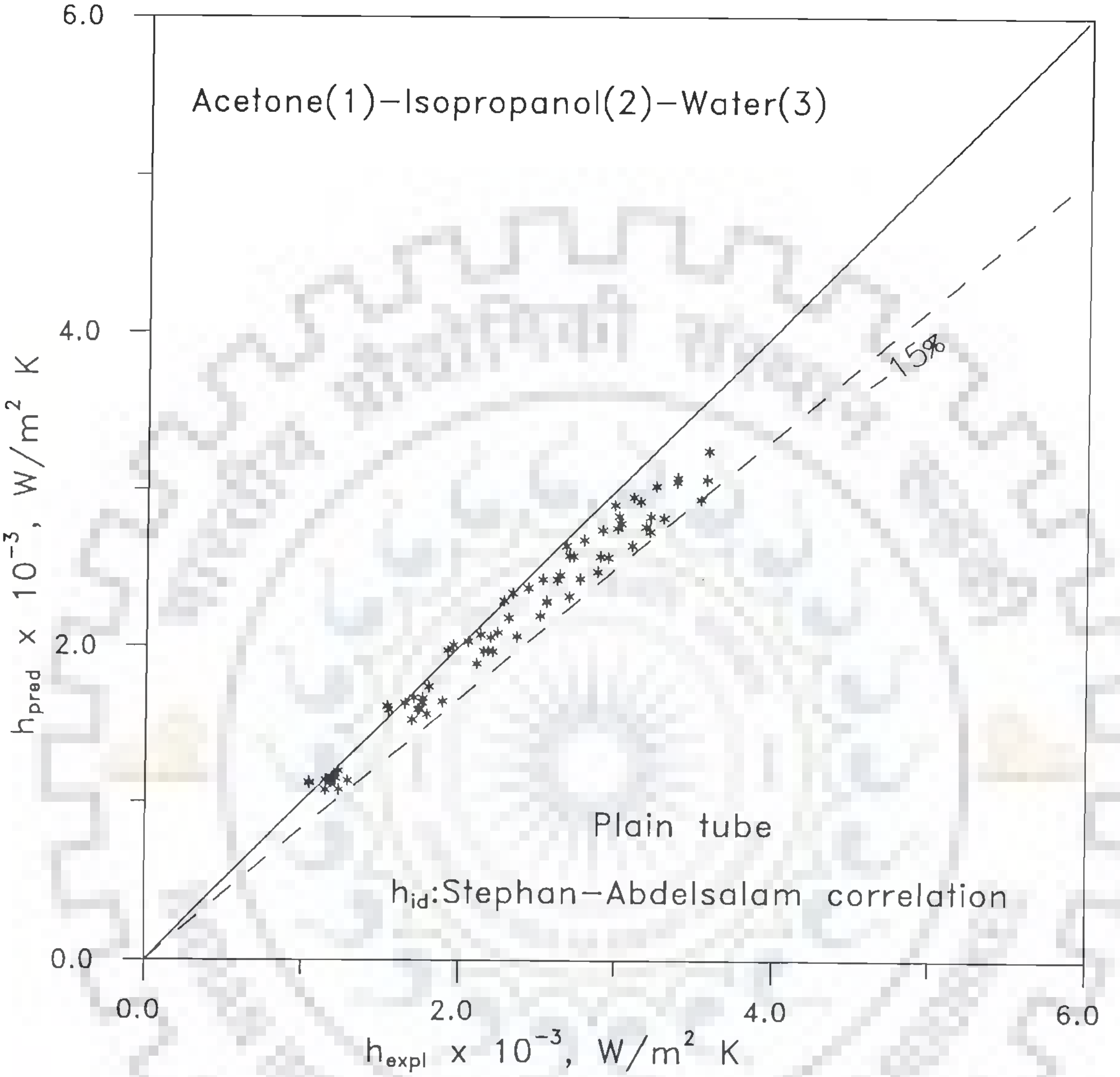


Fig. 6.47 Comparison of ternary mixture boiling data with the Thome−Shakir correlation

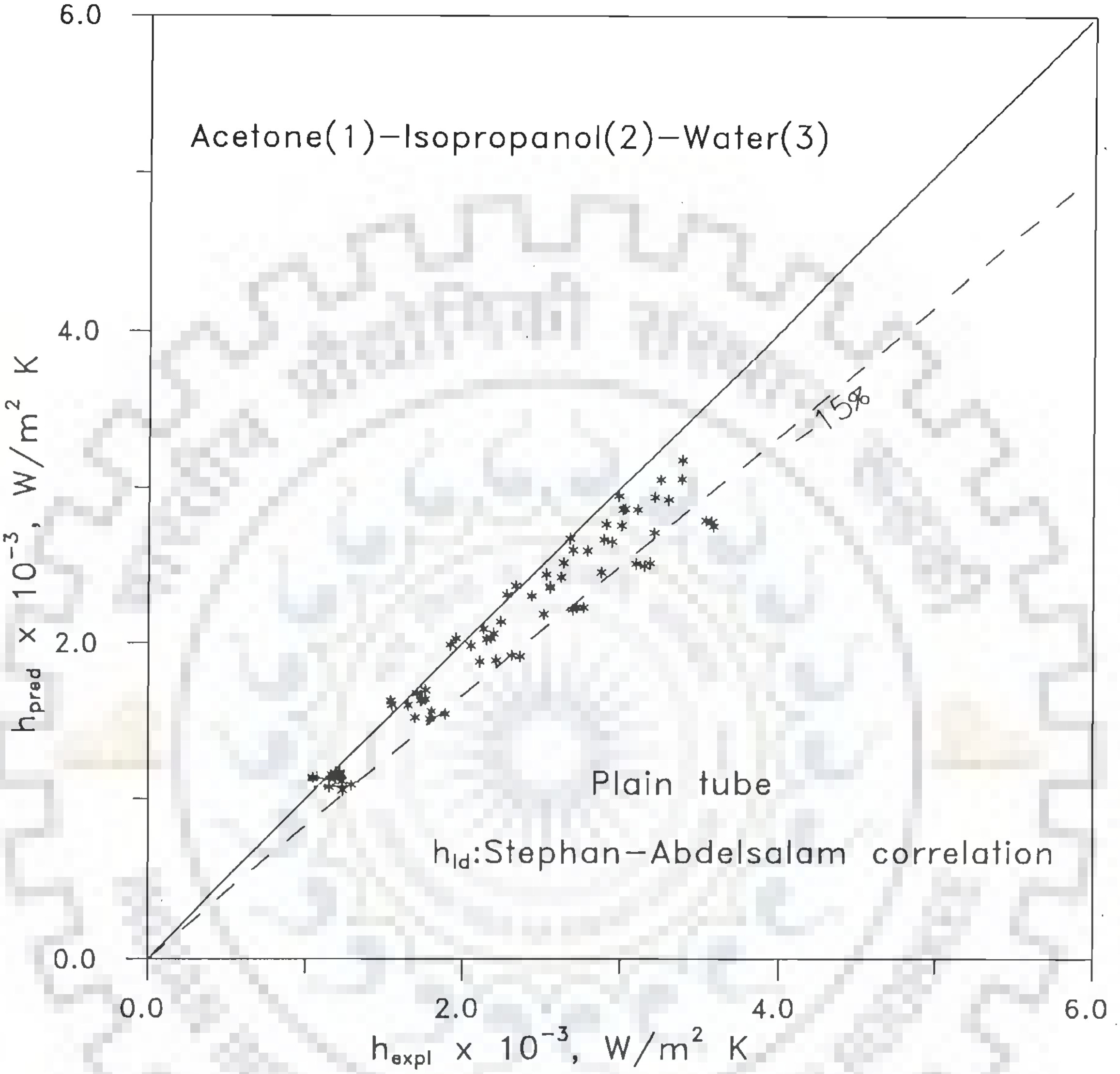


Fig. 6.48 Comparison of ternary mixture boiling data with the Schlunder correlation

CHAPTER 7

CONCLUSIONS AND RECOMMENDATIONS

Some of the salient conclusions based on experimental and theoretical investigations, related to nucleate pool boiling of single component liquids and their binary and ternary mixtures on a single horizontal plain as well as integral-fin tubes at atmospheric pressure are summarized as follows:

1. Based on the . present experimental data of boiling of saturated single component liquids: acetone, isopropanol, and distilled water on a single horizontal plain tube, it is established that the predicted values of heat transfer coefficients from the respective correlations due to Stephan and Abdelsalam (1980), and Cooper (1984) match excellently with the experimental average heat transfer coefficients. An implication of this is that these correlations, represented by Eqs.(6-2), (6-3), and (6-5), are recommended for predicting nucleate pool boiling heat transfer coefficients from a single horizontal plain tube to pool of a saturated single component liquids.

   As regards the above stated experimental data, it is also found that they possess the same boiling heat transfer characteristics as those of earlier investigators. The variation of heat transfer coefficient with heat flux is given by the following equation:

   $$h = C_1 q^{0.68} \qquad (6\text{-}6)$$

   where C_1 is a constant as given in Table 6.2.

2. Experimental data for the boiling of non-azeotropic binary liquid mixtures namely; acetone-water and acetone- isopropanol and azeotropic mixture of isopropanol-water on a horizontal plain tube exhibit the same functional relationship between h and q as that for the boiling of pure components of the mixtures, described by Eq.(6-6). However, the value of constant C_1 is different depending upon the composition of a given binary mixture and exponent of q is 0.66 instead of 0.68.

   Further, the present experimental data of 0.40 mole fraction of acetone in acetone-water mixture compare well with the available data from open literature due to Alam (1972) and Bajorek et al.(1989) with regard to functional dependence of h on q, implying that present experimental data are reliable.

3. For the boiling of non-azeotropic binary liquid mixtures, it is conclusively noted that heat transfer coefficient varies with mixture composition as follows:

   With the addition of the more volatile components, a degradation in heat transfer coefficient occurs, being maximum when acetone concentration is about 0.15 mole fraction in acetone-water system, whereas it is 0.4 mole fraction in acetone-isopropanol system

4. As regards the boiling of azeotropic mixtures of isopropanol-water, there appear two minima corresponding to isopropanol concentrations of 0.10 and 0.90 mole fraction.

5. The experimental data related to boiling of all the binaries investigated establish that variation of degradation factor of heat transfer coefficient, (h/h_{id}) with IY-XI of the more volatile component exhibits a regular decreasing trend unlike the variation of h with X.

6. Another important conclusion is that the value of (h/h_{id}) depends on IY-XI irrespective of plain or integral-fin heating surfaces for a given binary system. Whereas, (h/h_{id}) for a given value of IY-XI differs from one binary to another binary liquid mixtures.

7. The variation of (h/h_{id}) with IY-XI for the boiling of ternary liquid mixture of acetone-isopropanol-water, a similar trend as that in the case of binary mixtures is noted, implying that characteristics of binary mixture boiling are the same as those of ternary mixture boiling. It is to be noted that here IY-XI is different than in the case of binaries. As a matter of fact, it is a summation of IY-XI of the more and the intermediate volatile components.

8. Heat transfer coefficient for boiling of pure liquids, binary and ternary liquid mixtures on plain tube gets enhanced when the tube-surface is finned integrally. This enhancement is more pronounced when fin density is raised from 748 fpm to 1024 fpm.

9. Using the mechanism of heat and mass diffusion during the boiling of binary liquid mixtures on plain tube, a predictive correlation for heat transfer coefficient has been derived, which has the following functional form:

$$(h/hid)_P = 1/[1 + \{(h_{id})_P /q\} \Delta T_{BR}]^{0.158} \qquad (5\text{-}12)$$

This expression correlates experimental data of the present investigation and those from earlier five investigations within ±25 per cent.

The above correlation is also found valid to experimental data of ternary liquid mixtures of acetone-isopropanol-water within a maximum error of +10 per cent. Thus, it implies that the boiling characteristics of the ternary mixture and those of the binary mixtures are the same. This aspect makes this new correlation unique as compared to existing correlations due to Palen and Small (1964), Schlunder (1986), and Thome and shakir (1987).

10. Amongst the existing correlations related to the boiling of binary liquid mixtures on a horizontal plain tube, it is found that correlations due to Palen-Small and Thome-Shakir yield the best

predictions of boiling heat transfer coefficient of binary mixtures. However, in the present investigation it has been established that the Schlunder correlation acquires the best prediction capability, if h_{id} to be employed in this correlation is determined from the Stephan-Abdelsalam correlation, instead of linear mixing law method as proposed by Schlunder.

11. This investigation has also succeeded in extending Eq.(5-12) to the data of nucleate pool boiling of mixtures on integral-fin tubes by incorporating a term known as surface correction factor, F_s in it. The final form of this correlation is as follows:

$$(h/hid)_F = (F_s)\,(h_{id})_P\,[1/\{1 + \{(h_{id})_P/q\}\Delta T_{BR}\}]^{0.168} \qquad (5\text{-}15)$$

The experimental data of the present investigation and those available in the literature for the boiling of binary and ternary liquid mixtures on integral-fin tubes have been correlated within ±22 per cent.

RECOMMENDATION FOR FUTURE WORK

(1) In process industries, the reboilers deal with boiling of binary and multicomponent mixtures on tubes under sub-atmospheric pressures also. Hence, it would be important if experimental data are obtained under sub-atmospheric pressures.

(2) More experimental data for the boiling of ternary systems other than acetone-isopropanol-water especially with regard to integral-fin tubes covering wide range of mixture compositions, fin geometry, and heat flux should be carried out to further establish the applicability of Eq.(5-15).

(3) It would be of more practical significance if the experimental data related to boiling of liquid mixtures on separate bundles of plain as well as integral-fin tubes are also carried out.

APPENDIX A

CALCULATION OF HEAT TRANSFER COEFFICIENT

This Appendix describes the procedure followed for the determination of heat transfer coefficients from plain and integral-fin tubes to boiling binary and ternary liquid mixtures at atmospheric pressure. Experimental data are listed in Appendix D. Columns 3 through 6 list the outer wall temperatures, T_{wo} at the top-, the side-, the bottom-, and the other side- positions, respectively. These values have been calculated by using Eq.(A-3) or Eq.(A-4) depending upon heating tube surface geometry. Bulk liquid temperatures, T_s are also measured at four positions corresponding to wall thermocouple positions. The power supplied to the heating tube, Q is measured by a wattmeter. The outer wall temperature T_{wo} at the top-, the sides-, and the bottom- positions of heating tube have been estimated by extrapolating the corresponding temperature readings of the wall thermocouples. For this, it is reasonable to assume one dimensional radial heat conduction through the heating tube wall between the thermocouple locations and outer wall of the heating tube. The rate of radial heat transfer have been determined using the following equations as reported by Hahne et al. (1991), and Webb and Pais (1992):

a. For plain tube

$$Q =. \frac{(T_{th} - T_{wo})}{(1/2\pi L k_t)\ \ln(D_o/D_{th})} \qquad \text{(A-1)}$$

b. For integral-fin tube

$$Q = \frac{(T_{th} - T_{wo})}{(1/2\pi L k_t)\ \ln(D_r/D_{th})} \qquad \text{(A-2)}$$

where T_{th} is the temperature of a given thermocouple, D_{th} the wall thermocouple pitch circle diameter, L the effective heated length of the tube, k_t the thermal conductivity of the heating tube, D_o the outer diameter of plain tube, and D_r root diameter of the integral-fin tube.

Solving Eqs. (A-1) and (A-2) for T_{wo} one obtain:

a. For plain tube

$$T_{wo} = T_{th} -(Q/2\pi L k_t)\ \ln(D_o/D_{th}) \qquad \text{(A-3)}$$

b. For integral-fin tube

$$T_{wo} = T_{th} - (Q/2\pi L k_t)\ \ln(D_r/D_{th}) \qquad \text{(A-4)}$$

The average outer wall temperatures, (T_{wo}) are calculated as the arithmetic average of the top-, the sides- , and the bottom- positions as follows:

$$T_{wo} = [(T_{wo})_{top} + 2(T_{wo})_{side} + (T_{wo})_{bottom}] /4 \qquad \text{(A-5)}$$

The average bulk liquid temperature, T_s is taken as the average temperatures from four liquid thermocouples located in the bulk liquid as follows:

$$T_s = [(T_s)_{top} + 2(T_s)_{side} + (T_s)_{bottom}] /4 \qquad \text{(4-6)}$$

The wall superheat temperature difference for heat transfer from heating tube to boiling liquid is given by:

$$\Delta T = (T_{wo} - T_s) \qquad \text{(A-7)}$$

For the determination of average boiling heat transfer coefficient, the following equation is employed:

$$h = q/(\Delta T) \qquad \text{(A-8)}$$

where the heat flux, q is defined as:

$$q = Q/A \qquad \text{(A-9)}$$

where A is the surface area of heating tube ($= \pi D_o L$), for plain as well as integral-fin tubes, D_o represents the outer diameter in case of plain tube and envelope diameter for integral-fin tube.

Based on Eqs.(A-1) through (A-9) a computer program in FORTRANT 77 was developed to calculate the outer wall temperatures, wall superheats, and heat transfer coefficients. The name of the program is AMEERH. FOR. The listing of the program is given in Appendix E.

APPENDIX B

ERROR ANALYSIS

Elimination of experimental errors completely from experimental results is unattainable. However, due care has been exercised to select accurate and reliable instruments to minimize the measurement errors.

The instruments used for measurement are: the vernier calipers for heating tube dimensions, the wattmeter for power supplied to the heater, standard mercury in glass thermometer for calibrating thermocouples, and the digital multimeter for the measurement of thermo-electro motive force of thermocouples of the heating tube and the bulk liquid. These instruments also have inherent inaccuracies which contribute to the error in the value of average heat transfer coefficient.

In order to appreciate the accuracy of the experimental data and the heat transfer coefficients based on them, an error analysis has been carried out for several experimental runs using the standard method recommended by Schultz and Cole (1979) as described hereafter:

Let the dependent variable x be expressed in terms of independent measured quantities y_1, y_2, y_3, ... y_n; in the following functional form:

$$x = f\,(y_1, y_2, y_3, \ldots\ldots y_n)$$

Then, error in quantity x is defined as follows:

$$E_x = \Big[\sum_{i=1}^{i=n} \{(\delta x/\delta y_i)\, Ey_i\}^2\Big]^{0.5} \qquad \text{(B-1)}$$

where E_{yi} is the error in a quantity y_i.

Using Eq.(B-1), error in average heat transfer coefficient is calculated as detailed below:

The average heat transfer coefficient, h is defined by the following equation:

$$h = Q/A(T_{wo}-T_s) \qquad \text{(B-2)}$$

Now, applying the definition of error in any given quantity from Eq.(B-1) into Eq.(B-2), one obtains an expression for error in average heat transfer coefficient, E_h as follows:

$$E_h = [\{E_Q/(A(T_{wo}-T_s))\}^2 + \{D\,Q\,E_A/(A^2(T_{wo}-T_s) + \{D\,Q\,E_{(Two)}/(A(T_{wo}-T_s)^2)\}^2 + \{Q\,E_{(Ts)}/(A(T_{wo}-T_s))\}^2]^{0.5} \qquad \text{(B-3)}$$

where EQ, E_A, $E_{(Two)}$, and $E_{(Ts)}$ are the errors in the measurements of power supplied to heater, surface area of the heating tube, average outer wall temperature, and average bulk liquid temperature, respectively.

To estimate the error in the values of average heat transfer coefficient, Run No. 6 of Table D.6 of Appendix D for boiling of distilled water is considered, whose basic data are as follows:

Inside diameter of heating tube, $D_i = 0.018$ m
Outside diameter of heating tube, $D_o = 0.0312$ m
Thermocouple pitch circle diameter, $D_{th} = 0.0246$ m
Effective heated length of tube, L = 0.108 m
Surface area, $A = 1.058591 \times 10^{-2}$ m^2
Power supply to heater, Q = 560 W
Outer wall and bulk liquid temperatures: are listed in Table B.1

Table B.1 Outer wall and bulk liquid temperatures, ^{0}C

Temperature	Thermocouple position			
	Top	Side	Bottom	Side
Outer wall temperature, T_{wo}	105.357	103.382	104.305	104.606
Bulk liquid temperature, T_s	100.172	100.129	100.107	100.129

Average outer wall temperature, $T_{wo} = [105.357 + 103.382 + 104.305 + 104.606]/4$
$= 104.41$ ^{0}C

Average bulk liquid temperature, $T_s = [100.172 + 100.129 + 100.107 + 100.129]/4$
$= 100.13$ ^{0}C

Average heat transfer coefficient, $h = Q/[A (T_{wo} - T_s)]$
$= 560/[1.058591 \times 10^{-2} (104.41-100.13)]$
$= 12365$ W/(m^2 ^{0}C)

B.1 ERROR IN POWER SUPPLY, E_Q

The wattmeter used to measure the power has an error of 1% full scale. Hence, maximum possible error in the measurement of power supply is 5.6 W.

Therefore, $E_Q = 5.6$ W

B.2 ERROR IN HEATING SURFACE AREA, E_A

$$A = \pi D_o L$$

hence,

$$E_A = [(\pi L E_{Do})^2 + (\pi D_o E_L)^2]^{0.5} \qquad \text{(B-4)}$$

where E_{Do} and E_L are the errors in the measurement of outside diameter and length of the heating tube, respectively. A vernier caliper, having a least count of (1×10^{-4}) m has been used to measure diameter and length of the heating tube.

Hence, $E_{Do} = E_L = E_{th} = 1 \times 10^{-4}$ m

Using the values of E_{Do} and E_L, in Eq. (B-4), one gets:

$$E_A = [(\pi \times 0.108 \times 1 \times 10^{-4})^2+(\pi \times 0.0312 \times 1 \times 10^{-4})^2]^{0.5}$$

$$E_A = 0.3532 \times 10^{-4} \text{ m}^2$$

B.3 ERROR IN AVERAGE OUTER WALL TEMPERATURE OF TUBE, $E_{(Two)}$

Outer wall temperature of the heating tube is determined by subtracting the temperature drop across the wall thickness of the heating tube from the wall temperature, T_w measured by thermocouples as follows:

$$T_{wo} = T_w - (q_{Do}/2k_t) \ln (D_o/D_{th}) \tag{B-5}$$

$$E_{Two} = [\{E_{(Tw)}\}^2 + \{E (-q_{Do}/2k_t) \ln (D_o/D_{th}) \}^2]^{0.5} \tag{B-6}$$

Therefore, error in T_{wo} is given by:

$$E_{Two} = [(E_{Tw})^2 + [\{(D_o/2k_t) \ln (D_o/D_{th}).E_q \}^2+ \{(q/2k_t)\ln (D_o/D_h) E_{Do}\}^2+\{(q/2k_t).E_{Do}\}^2$$

$$+ \{(-qD_o/2k_t^2)\ln (D_o/D_{th}).E_{kt}\}^2 + \{(-qD_o/2k_t)(1/D_{th}) .E_{Dth}\}^2]]^{0.5} \tag{B-7}$$

where E_{Tw}, E_q, E_{kt}, and E_{Dth} represent the errors in the measurements of temperature of wall, heat flux, thermal conductivity, and the thermocouple pitch circle diameter, respectively.

Since $q = Q/A$

Hence,

$$E_q = [\{DE_Q/(\pi D_o L)\}^2 + \{- Q/(\pi L D_o^2).E_{Do}\}\} + \{- Q/\pi D_o L^2).E_L\}]^{0.5} \tag{B-8}$$

$$E_q = [279846.32 + 28748.18 + 2399.23]^{0.5}$$

$$E_q = 557.668 \text{ W/m}^2$$

Thermal conductivity of heating tube material, $k_t = 133.5$ W/(m ^{0}C)

Error in thermal conductivity of material, $E_{kt} = 0.0$

Error in temperature of wall, $E_{TW} = 0.1$ ^{0}C

Substituting the values of E_q, E_{Do}, E_{kt} and E_{Dth} in Eq. (B-7)

$$E_{Two} = [(0.1)^2 + [\{0.0312/(2 \times 133.5) \ln (0.0312/0.0246) \times 2179.855\}^2+ \{(52901/(2 \times 133.5))\ln(0.0312/0.0246) \times 1 \times 10^{-4}\}^2+ \{ 52901/(2 \times 133.5) \times 1 \times 10^{-4}\}^2+ \{(-52901/(2 \times (133.5)^2) \ln (0.0312/0.0246) \times 0.0\}^2+ \{(-52901 \times 0.0312/ 2 \times 133.5 \times 0.0246 \times 1 \times 10^{-4}) \}^2]]^{0.5}$$

$$= [0.01+ 0.00129]^{0.5}$$

$$E_{Two} = 0.106 \ ^0\text{C}$$

Average outer wall temperature of the heating tube is calculated as follows:

$$(T_{wo}) = [T_{wo,t} + 2(T_{wo,s}) + T_{wo,b}]/4 \qquad \text{(B-9)}$$

where $T_{wo,t}$, $T_{wo,s}$, and $T_{wo,b}$ denote the local outer wall temperatures at the top-the sides-, and the bottom- positions, respectively. Now, employing the definition of error in a quantity from Eq.(B-1) into Eq.(B-9), the error in the average outer wall-temperature of the heating tube is:

$$E_{(Two)} = [(E_{Two,t}/4)^2 + (E_{Two,s}/2)^2 + (E_{Two,b}/4)^2]^{0.5} \qquad \text{(B-10)}$$

where, $E_{Two,t} = E_{Two,s} = E_{Two,b} = E_{Two} = 0.106\,^0C$

Substitution of the values of $E_{Two,t}$, $E_{Two,s}$, and $E_{Two,t}$ in Eq.(B-10) gives the value of $E_{(Two)}$:

$$E_{(Two)} = [(0.106/4)^2 + (0.106/2)^2 + (0.106/4)^2]^{0.5}$$

$$= 0.0747\,^0C$$

B.4 ERROR IN AVERAGE BULK LIQUID TEMPERATURE, $E_{(Ts)}$

Average bulk liquid temperature is calculated by:

$$T_s = [T_{s,t} + 2(T_{s,s}) + T_{s,b}]/4 \qquad \text{(B-11)}$$

where $T_{s,t}$, $T_{s,s}$, and $T_{s,b}$ refer to the local bulk liquid temperatures at the top-, the sides-, and the bottom- positions around the heating tube.

So,

$$E_{(Ts)} = [(E_{Ts,t}/4)^2 + (E_{Ts,s}/2)^2 + (E_{Ts,b}/4)^2]^{0.5} \qquad \text{(B-12)}$$

Now, $E_{Ts,t} = E_{Ts,s} = E_{Ts,b} = 0.1\,^0C$

Inserting the values of $E_{Ts,t}$, $E_{Ts,s}$, and $E_{Ts,b}$ in Eq.(B-12), the error in average bulk liquid temperature is as follows:

$$E_{(Ts)} = [(0.1/4)^2 + (0.1/2)^2 + (0.1/4)^2]^{0.5}$$

$$= 0.0612\,^0C$$

B.5 ERROR IN AVERAGE HEAT TRANSFER COEFFICIENT, E_h

Substitution of the values of the quantities in Eq.(B-3) yields the value of error in average heat transfer coefficient as:

$$E_h = [\,[(5.6)/\{1.058591\text{x}10^{-2}(4.278)\}]^2 + [-560\text{x}0.3532\text{x}10^{-4}/\{(1.058591\text{x}10^{-2})^2(4.278)\}]^2$$
$$+ [(-560\text{x}0.0649)/\{1.058591\text{x}10^{-2}\text{x}(4.278)^2\}]^2 + [(560\text{x}0.0612)/\{1.058591\text{x}10^{-2}(4.278)^2\}]^2\,]^{0.5}$$

$$E_h = 288.928\ W/(m^2\ {}^oC)$$

The per cent error in the measurement of average heat transfer coefficient is given by:

= (Error/ Experimental heat transfer coefficient) x 100

= (288.928/12365) x 100

= 2.337 %

Similarly, calculations for errors in the average heat transfer coefficients were made for several other runs. It is found that the error is of the same order.

APPENDIX C

THERMODYNAMIC AND TRANSPORT PROPERTIES OF MIXTURES

The ability of correlations employed for predicting boiling heat transfer coefficients depends to a large extent on the accuracy of thermodynamic and transport properties of pure components and their mixtures used. This Appendix discusses the methods of estimation of thermodynamic and transport properties of binary and ternary mixtures. These properties include liquid and vapour densities, differential heat of vaporization, liquid thermal conductivity, liquid viscosity, and liquid specific heat. Information related to vapour-liquid phase equilibrium plays a vital role for bubble growth rate and estimation of boiling heat transfer coefficient of binary and ternary mixtures. Hence, a discussion in this regard is also included in this section. The methods used to estimate physico-thermal properties of mixtures were chosen mainly based on their ability to predict the properties reliably and partly on the capability of these methods to be extended to ternary mixtures. Besides, the thermodynamic and transport properties, the technique used to estimate binary liquid mixture mass diffusivity is also included.

C.1 VAPOUR-LIQUID EQUILIBRIUM

Vapour-liquid equilibrium, **VLE** data for binary mixtures and various multicomponent mixtures have been compiled by Hirata et al.(1975) and Behrens et al. (1984). Methods generally adopted for the prediction of these data are discussed by Walas (1985), and Reid et al.(1987).

It is important to point out , that experimental **VLE** data for three and more component liquid mixtures are relatively scarce. The **VLE** data for acetone-water and acetone-isopropanol-water mixtures, have been taken from Behrens and Eckermann (1984), whereas, the **VLE** data for the isopropanol-water and acetone-isopropanol mixtures have been obtained from Kojima et al.(1969) and Freshwater and Pike (1967), respectively. Figures C.1 through C.3 present binary phase equilibrium diagrams for acetone-water, isopropanol-water, and acetone-isopropanol mixtures, respectively. For the sake of convenience, the **VLE** data of acetone-isopropanol-water ternary mixture at 101.32 kPa is reproduced in Table C.1.

C.2 BOILING RANGE

The boiling range of mixture, ΔT_{BR} is the difference between the dew point and the bubble point temperatures for a given liquid phase mixture composition. In the present investigation the boiling range for binary as well as ternary mixtures were estimated using experimental vapour-liquid equilibrium data available in the literature.

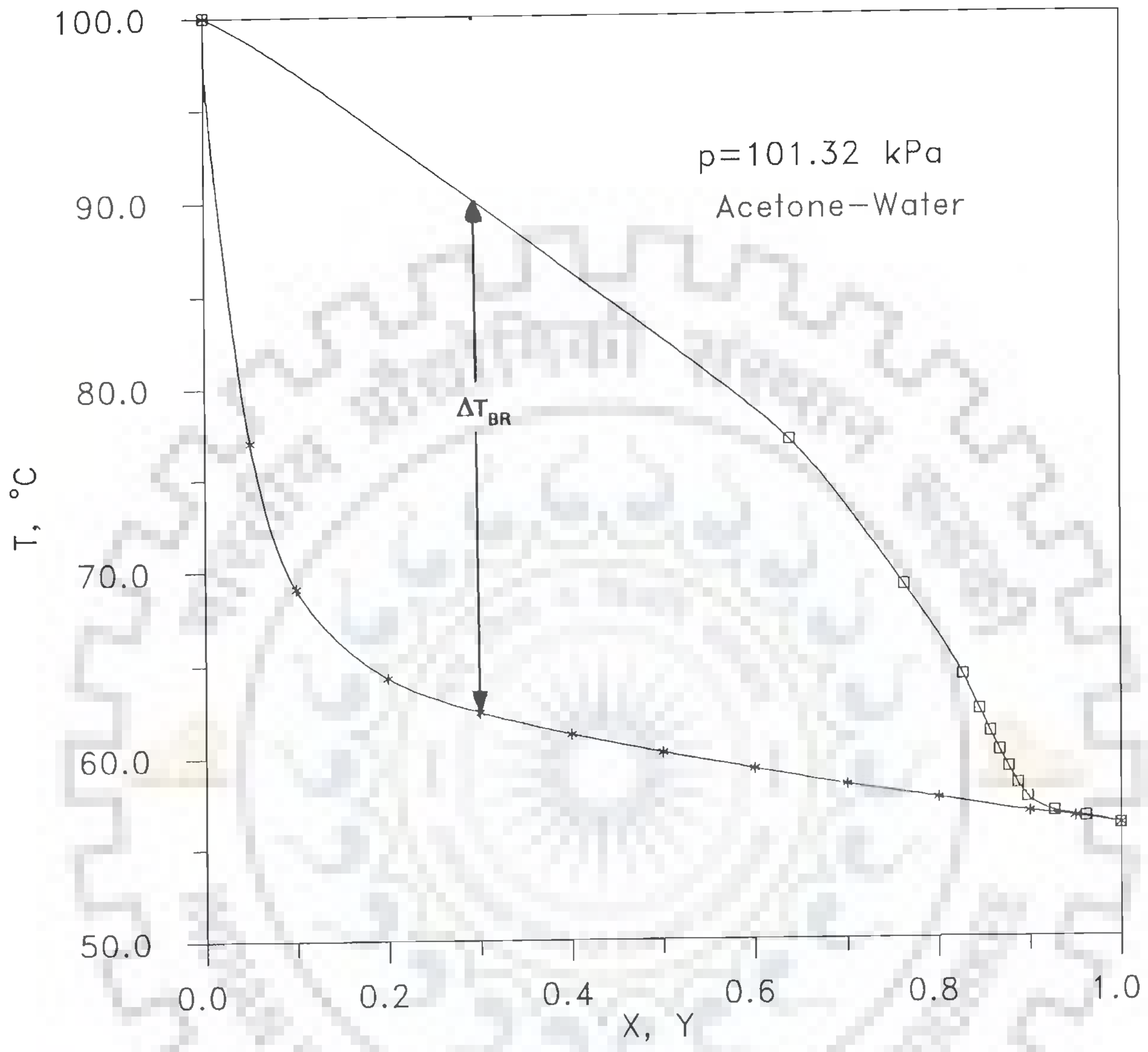


Fig. C.1 Vapour-liquid phase equilibrium diagram for acetone-water mixture

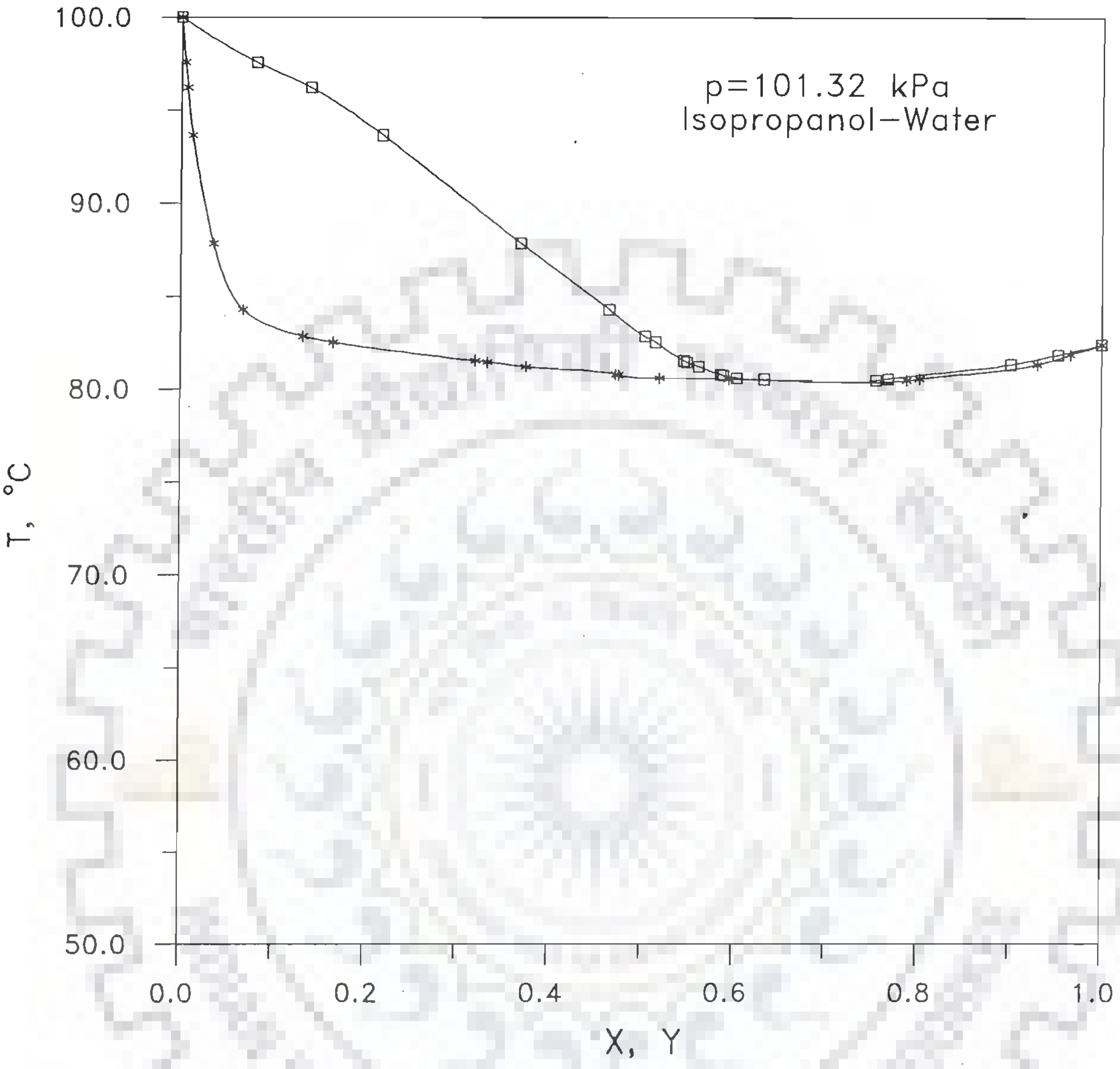


Fig. C.2 Vapour-liquid phase equilibrium diagram for isopropanol-water mixture

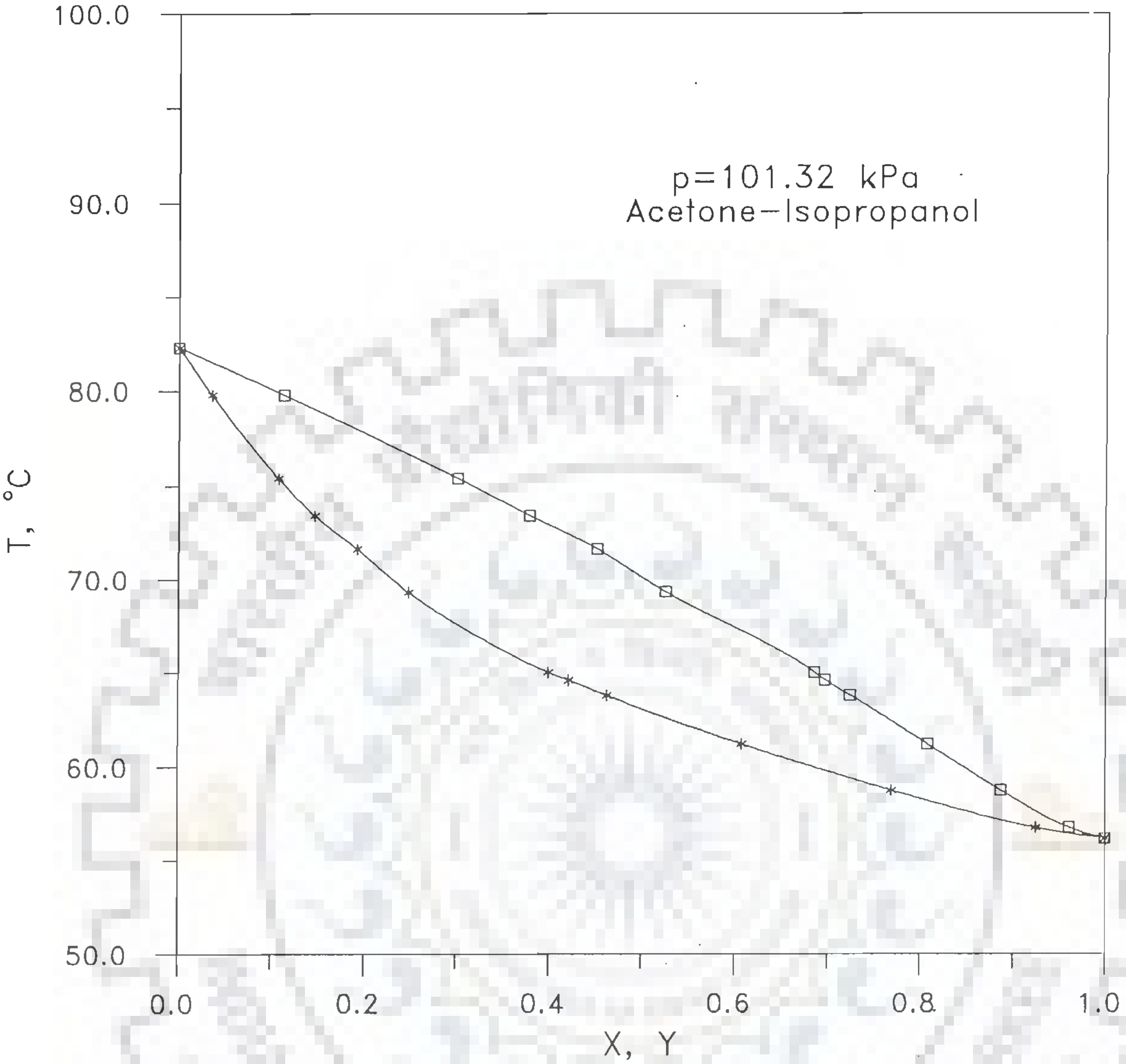


Fig. C.3 Vapour-liquid phase equilibrium diagram for acetone-isopropanol mixture

Table C.1 Vapour-liquid phase equilibrium data for acetone(1)- isopropanol(2)-water(3) mixture

No.	T ^{0}C	X_1	X_2	Y_1	Y_2
1	90.0	.0030	.0120	.1400	.1780
2	82.0	.0040	.0750	.0970	.4050
3	80.0	.0040	.5900	.0150	.6140
4	80.0	.0050	.6900	.0150	.6850
5	80.0	.0050	.8020	.0150	.7700
6	80.0	.0080	.7440	.0190	.7150
7	80.0	.0090	.2430	.1020	.4640
8	80.0	.0090	.3860	.0530	.5300
9	80.0	.0090	.7660	.0150	.7440
10	90.0	.0100	.0880	.2120	.1020
11	80.0	.0100	.5200	.0300	.5900
12	80.0	.0110	.0740	.2240	.3250
13	82.0	.0120	.0620	.1380	.3720
14	80.0	.0130	.0720	.2180	.3360
15	82.0	.0140	.0440	.2180	.2930
16	80.0	.0160	.1510	.1350	.4350
17	80.0	.0160	.2780	.0900	.4800
18	80.0	.0170	.9410	.0880	.8580
19	80.0	.0175	.1750	.1220	.4480
20	82.0	.0180	.0450	.2140	.2920
21	82.0	.0200	.0190	.3490	.1530
22	80.0	.0260	.0200	.4100	.1420
23	77.0	.0340	.0360	.4520	.1650
24	75.0	.0420	.0190	.5770	.0780
25	75.0	.0470	.0220	.5510	.1010
26	77.0	.0470	.1780	.2860	.3440
27	73.0	.0500	.0400	.5250	.1400
28	77.0	.0590	.4660	.1760	.4780
29	75.0	.0650	.1300	.3770	.2660
30	73.0	.0780	.0950	.4600	.2110
31	75.0	.0820	.2940	.3280	.3420
32	70.0	.0830	.0360	.6320	.0860
33	75.0	.0900	.4800	.2840	.4210

Table C.1 contd...

34	75.0	.1050	.6100	.2700	.4960
35	75.0	.1080	.6750	.2730	.5310
36	69.0	.1100	.0500	.6380	.1080
37	73.0	.1240	.2900	.4000	.2960
38	68.0	.1320	.0530	.6420	.1090
39	73.0	.1400	.5800	.3380	.4560
40	69.0	.1480	.1140	.5890	.1570
41	70.0	.1550	.1500	.5300	.1940
42	66.0	.1660	.0240	.7280	.0410
43	68.0	.1900	.1200	.6000	.1500
44	70.0	.1920	.2810	.4920	.2550
45	65.0	.1980	.0340	.7310	.0490
46	66.0	.2050	.0680	.6800	.0970
47	69.0	.2120	.1820	.5370	.2190
48	70.0	.2260	.5320	.4770	.3560
49	70.0	.2270	.5500	.4740	.3660
50	69.0	.2360	.3360	.5200	.3560
51	65.0	.2460	.0570	.7370	.0370
52	66.0	.2530	.1100	.6520	.1270
53	69.0	.2620	.4920	.4970	.3330
54	68.0	.2660	.2820	.5600	.2200
55	64.0	.2690	.0520	.7470	.0460
56	68.0	.2940	.4840	.5450	.3080
57	66.0	.3240	.1970	.6400	.1640
58	66.0	.3440	.1750	.6870	.1170
59	63.0	.3560	.0640	.7450	.0590
60	66.0	.3760	.4210	.6160	.2590
61	65.0	.3820	.2740	.6390	.1890
62	64.0	.3850	.1750	.7020	.1140
63	64.0	.3890	.2120	.6740	.1540
64	65.0	.3990	.4150	.6570	.2290
65	62.0	.4100	.0390	.7660	.0510
66	65.0	.4150	.4650	.6730	.2500
67	64.0	.4150	.3860	.6830	.2120
68	64.0	.4360	.3000	.6790	.1750
69	63.0	.4380	.1750	.7160	.1250

Table C.1 contd...

Table C.1 contd...

70	62.0	.4580	.0650	.7700	.0600
71	63.0	.4600	.2280	.6980	.1430
72	62.0	.4660	.3100	.7050	.1730
73	62.0	.5020	.1610	.7460	.0950
74	61.0	.5230	.0670	.7990	.0430
75	62.0	.5240	.2670	.7310	.1530
76	62.0	.5510	.2970	.7400	.1690
77	60.0	.5690	.500 0	.8260	.0260
78	61.0	.5850	.1770	.7780	.1030
79	61.0	.6200	.2610	.7950	.1440
80	60.0	.6250	.1260	.8000	.0700
81	59.0	.6520	.0620	.8370	.0310
82	60.0	.6730	.2110	.8380	.1030
83	59.0	.6990	.1050	.8550	.0400
84	59.0	.7360	.1730	.8600	.0930
85	58.0	.7570	.0590	.8650	.0360
86	58.0	.8080	.1300	.9010	.0520
87	56.1	1.0000	.0000 1	.0000	.0000
88	82.3	.0000	1.0000	.0000	1.0000
89	1.000	.0000	.0000	.0000	.0000

C.2.1 Boiling Range of Binary Mixture

Figure C.1 shows the vapour-liquid phase equilibrium diagram of acetone-water mixture. From this diagram the boiling range, ΔT_{BR} of the binary mixture can be readily estimated for any liquid mixture composition. For computational purposes, polynomials of $T_b = f(X)$, and $T_d = g(Y)$ can be fitted for bubble and dew point curves of the vapour-liquid equilibrium diagram of any system. Using these polynomials, the boiling range at a specified liquid composition, say X_1 can be calculated as the difference in the dew point and bubble point temperatures as follows:

$$\Delta T_{BR} = g(X_1) - f(X_1) \qquad (C\text{-}1)$$

C.2.2 Boiling Range of Ternary Mixture

Boiling range of ternary mixtures is more difficult to determine due to the non-availability of required number of experimental **VLE** data points. Thus, more complex methods are called for to determine the boiling range in ternary mixture system as detailed below:

In the case of ternary mixtures, the boiling range can be conceptualized as the difference between dew and bubble point temperatures for a given liquid mixture composition. For a ternary system the loci of dew and bubble points form surfaces and ΔT_{BR} is calculated by drawing a perpendicular from composition plane and estimating the distance between the points created by the intersection of dew and bubble planes with the perpendicular. The boiling range for ternary mixture in the present investigation was estimated using the method given below:

For a given liquid mixture composition, the bubble point temperature can be determined from **VLE** data directly or from Figure C.4, which represents the contour map of $\mathbf{X_1}$, $\mathbf{X_3}$, and Temperature, **T** as parameter, using the SURFER Access System Version 3.0 Copyright (C) Golden Software Inc. 1987. The above figure is plotted using **VLE** data from Table C.1. In the same way, the dew point temperature can be determined from Figure C.5, which represents the contour map of $\mathbf{Y_1}$, $\mathbf{Y_3}$, and **T** for the same liquid mixture composition. The difference of above estimated temperatures provides ΔT_{BR}. For example, let us calculate the boiling range temperature of mixture composition: $X_1 = 0.05$, $X_2 = 0.04$, and $X_3 = 0.91$. The step-wise procedure for the estimation of ΔT_{BR} is given below:

Step 1: Determine the bubble point temperature corresponding to $X_1 = 0.05$ and $X_3 = 0.91$ from Figure C.4. The value is found to be 73 °C.

Step 2: Determine the dew point temperature for the same composition mixture that of step 1 ($X_1 = 0.05$ and $X_3 = 0.91$) from Figure C.5. The dew point temperature comes out to be 97.1 °C.

Step 3: The boiling range, ΔT_{BR} is equal to 24.1°C, (= 97.1 - 73).

C.3 PHYSICO-THERMAL PROPERTIES OF MIXTURES

Physico-thermal properties have large influence on the process of heat and mass transfer during the boiling of mixtures. Sometimes only a small change in liquid composition affects the properties like; surface tension, latent heat of vaporization, and viscosity. Further, the bubble nucleation, bubble growth rate and bubble departure diameter, which play decisive role on boiling of mixtures, are particularly sensitive to changes in these properties. Hence, this calls for methods which predict accurately the variations in physico-thermal properties with composition and temperature.

Though the non-linear variation in physico-thermal properties with composition are a typicality of aqueous mixture systems, yet it can also occur for other types of mixture systems. In the above situation, the prediction of mixture properties to be used in correlations for the prediction of boiling heat transfer coefficients should be done with utmost care. Reid et al.(1987) have reviewed

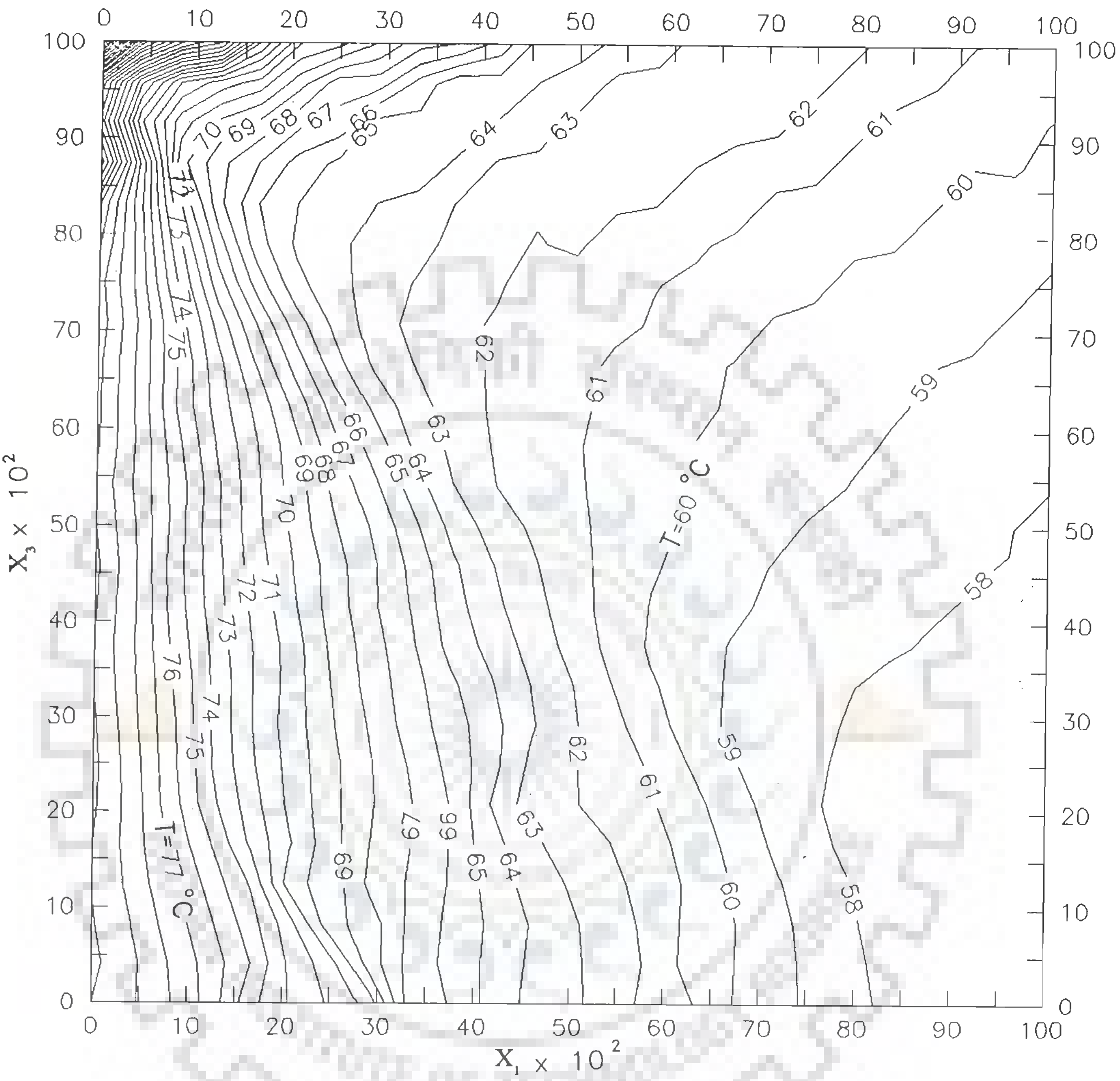


Fig. C.4 Contour map for ternary liquid mixtures X_1, X_3 and T

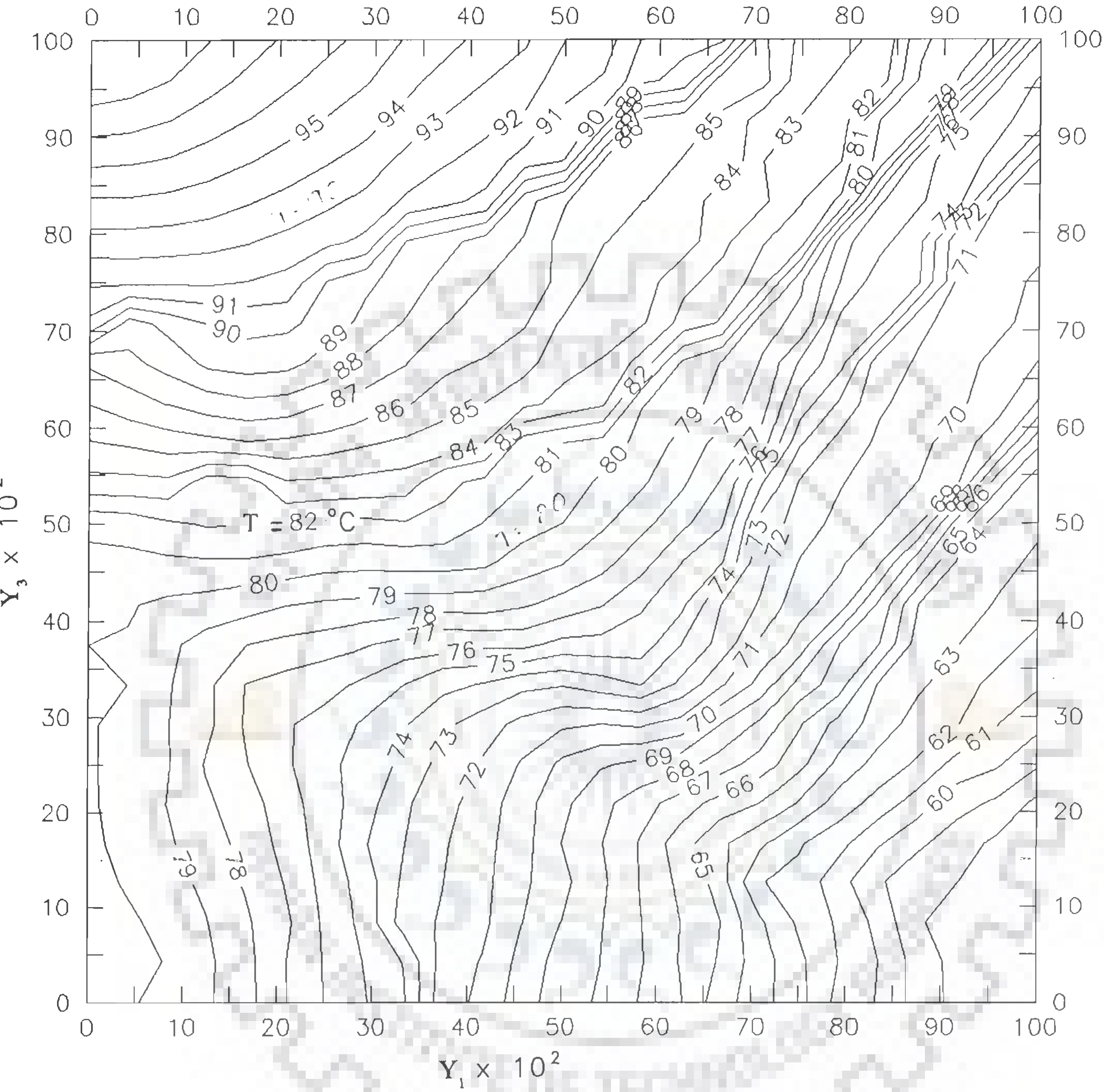


Fig. C.5 Contour map for ternary liquid mixtures Y_1, Y_3 and T available literature critically as regards the accuracy of various methods for prediction of physico-thermal properties of pure liquids and mixtures. Out of these methods the 'multifluid corresponding states method ' as proposed by Teja and Rice (1981) is widely used for prediction of physico-thermal properties of mixtures as reported by Reid et al.(1987).

C.3.1 Multifluid Corresponding States Method

The physico-thermal properties of a mixture are often estimated by determining the properties of the pure components and then by combining the pure component properties using an appropriate mixing rule. The unknown properties of a given fluid can be determined from the known properties of another fluid using the principle of corresponding states. A pure fluid is defined to be in corresponding states with a reference fluid if the compressibility factor Z and the reduced property $(\Phi\varepsilon)$ of both fluids (at the same reduced temperature T_R and reduced pressure P_R) obey the rules as given below by Teja and Rice (1981):

$$Z = Z^{(o)} \qquad \text{(C-2)}$$

and

$$(\Phi\varepsilon) = (\Phi\varepsilon)^{o} \qquad \text{(C-3)}$$

where Φ represents any physical property and ε is a function of the critical parameters T_{cr}, p_{cr}, V_{cr}, and MW. However, Eqs.(C-2) and (C-3) are only valid for fluids with spherically symmetrical molecules like noble gases with negligible intermolecular attraction.

The more general case of non-spherical molecules is treated by introducing the acentric factor, ω in Eq.(C-2). This was first proposed by Pitzer et al.(1955) as follows:

$$Z = Z^{(o)} + \omega Z^{(1)} \qquad \text{(C-4)}$$

where $Z^{(o)}$ is the compressibility factor of a spherical reference fluid with zero acentric factor at the same reduced condition and $Z^{(1)}$ is a function which describes deviation from spherical molecule behaviour. The above concept of Pitzer et al.(1955) has been modified and extended by Lee and Kesler (1975) and recently by Teja and Rice(1981). The generalized corresponding states principle proposed by Teja and Rice uses two non-spherical reference fluids and represents the compressibility factor as:

$$Z = Z^{(r1)} + [(\omega-\omega^{(r1)})/(\omega^{(r2)}-\omega^{(r1)})]\,[Z^{(r2)}-Z^{(r2)}] \qquad \text{(C-5)}$$

For a binary mixture, the reference fluids(r1) and (r2)are the two pure components of interest which are not necessarily spherical. The method is extended to multicomponent mixtures via the vander Waals model where by the properties .T_{cr}, v_{cr}, Z_{cr}, and w of the reference fluid are replaced by their pseudocritical values defined as:

$$v_{cm} = \sum_i \sum_j X_i \; X_j \, v_{c\,ij} \qquad \text{(C-6)}$$

$$T_{cm} \, v_{cm} = \sum_i \sum_j X_i X_j \; T_{c\,ij} v_{c\,ij} \qquad \text{(C-7)}$$

$$Z_{cm} = \sum_i X_i \, Z_{ci} \qquad \text{(C-8)}$$

$$MW_m = \sum_i X_i \, MW_i \qquad \text{(C-9)}$$

The vander Waals model can be used to obtain the properties of mixtures provided values can be assigned to the cross-parameters T_{cij} and v_{cij} $(i = j)$. For simple non-polar mixtures, the most successful mixing rules are:

$$T_{cij} v_{cij} = (T_{cii} V_{cii} T_{cjj} v_{cjj})^{0.5} \qquad \text{(C-10)}$$

$$v_{cij} = [\Omega_{ij} \, (v_{cii}^{\;1/3} + v_{cjj}^{\;1/3})^3] \, /8 \qquad \text{(C-11)}$$

where Ω_{ij} is a binary interaction coefficient which must be obtained from experiment. No additional coefficients are necessary to predict the properties of mixtures with three components or higher than three components.

The multifluid model consists of a series of two reference fluids interpolations. Thus, in a multicomponent mixture, a two fluid expansion involving component 1 and 2 is carried out ignoring all other components.

$$Z_{12} = Z_2 + [(\omega_{12} - \omega_2)/(\omega_1 - \omega_2)] \; [Z_1 - Z_2] \qquad \text{(C-12)}$$

where

$$\omega_{12} = \sum_1^2 X_i \, \omega_i / \sum X_i \qquad \text{(C-13)}$$

If the mixture contains a third component, another expansion is performed using the third and pseudo-component"12"as the reference fluids. Then,

$$Z_{123} = Z_3 + \; (\omega_{123} - \omega_2)/(\omega_{12} - \omega_2)[Z_{12} - Z_3] \qquad \text{(C-14)}$$

where ω_{12} is defined by Eq.(C-13) and ω_{123} is given by:

$$\omega_{123} = \sum_{1}^{3} X_i\,\omega_i / \sum X_i \qquad \text{(C-15)}$$

In general, for n components, the expansion becomes(see Teja and Rice (1981)):

$$Z_{12..n} = Z_{12..n-1} + (\omega_{12..n} - \omega_n)/(\omega_{12..n-1} - \omega_n)[Z_n - Z_{12..n-1}] \qquad \text{(C-16)}$$

with

$$\omega_{12..n} = \sum_{1}^{n} X_i\,\omega_i / \sum X_i \qquad \text{(C-17)}$$

A similar relationship can be written for other physico-thermal properties by starting with Eq.(C-3). The property of a mixture with n component is determined from:

$$(\Phi\varepsilon)_{12..n} = (\Phi\varepsilon)_{12..n-1} + [(\omega_{12..n} - \omega_n)/(\omega_{12..n-1} - \omega_n)]\,[(\Phi\varepsilon)_n - \Phi\varepsilon_{12..n-1}] \qquad \text{(C-18)}$$

C.3.2 DENSITY OF VAPOUR MIXTURES

The following equation of state due to Patel and Teja (1982) was used to estimate the specific volumes, v of pure vapour and their mixtures:

$$p = (RT/\ (v-b)\ ) - [\ a(T)\ /\ (v\ (v+b) + c\ (v-b))] \qquad \text{(C-19)}$$

where R is the universal gas constant , and a(T), b, and c are calculated from the following equations:

$$a(T) = \Lambda_a (R^2 T_{cr}^2/p_{cr})\ \ \alpha(T_R) \qquad \text{(C-20)}$$

$$b = \Lambda_b\ (RT_{cr}/p_{cr}) \qquad \text{(C-21)}$$

$$c = \Lambda_c\ (RT_{cr}/p_{cr}) \qquad \text{(C-22)}$$

where

$$\Lambda_a = 3\,\delta_c^2 + 3\,(1-2\delta_c)\,\Lambda_b + \Lambda_b^2 + 1 - 3\delta_c \qquad \text{(C-23)}$$

and Λ_b is the smallest root of the cubic equation, Eq.(C-24):

$$\Lambda_b^3 + (2 - 3\delta_c)\,\Lambda_b^2 + 3\delta_c^2\Lambda_b - \delta_c^3 = 0 \qquad \text{(C-24)}$$

and where

$$\Lambda_c = 1-3\,\delta_c \qquad \text{(C-25)}$$

The function $\alpha(T_R)$ is given by the following equation:

$$\alpha(T_R) = [1 + F(1 - T_R^{1/2})]^2 \qquad \text{(C-26)}$$

The parameters F and δ_c have been correlated with the acentric factor, and are given as:

$$F = 0.452413 + 1.30982\omega - 0.295937\omega^2 \qquad \text{(C-27)}$$

$$\delta_c = 0.329032 - 0.076799\omega + 0.0211947\omega^2 \qquad \text{(C-28)}$$

Table C.2 lists the pure component parameters used in the present investigation for estimation of physico-thermal properties. The parameters have been taken from Reid et al.(1987).

Table C.2 Pure component parameters used in property estimation

Component	Formula	MW	T_{cr}	p_{cr}	v_{cr}	Z_{cr}	ω	Z_{RA}
Acetone	C_3H_6O	58.08	508.15	47.61	0.0036630	0.239	0.309	0.2477
Isopropanol	C_3H_8O	60.10	508.75	53.70	0.0036496	0.278	0.665	0.2493
Water	H_2O	18.02	647.30	221.29	0.0031746	0.235	0.344	0.2338
Methanol	CH_4O	32.04	513.15	79.50	0.0036364	0.217	0.559	0.2334
Ethanol	C_2H_6O	46.06	516.25	63.90	0.0035714	0.245	0.635	0.2500
n-Propanol	C_3H_8O	60.10	536.85	50.50	0.0036630	0.249	0.624	0.2541

Equation of state recommended by patel and Teja can be extended to mixtures by replacing the constants a(T), b, and c as follows:

$$a_m(T) = \sum_i \sum_j X_i X_j a_{ij} \qquad \text{(C-29)}$$

$$b_m = \sum_i X_i b_i \qquad \text{(C-30)}$$

$$c_m = \sum X_i c_i \qquad \text{(C-31)}$$

The cross-interaction term, a_{ij} in Eq.(C-29) is evaluated using:

$$a_{ij} = \Omega_{ij} (a_i\, a_j)^{1/2} \qquad \text{(C-32)}$$

where Ω_{ij} is a binary interaction coefficient which must be evaluated from experimental data.

Patel and Teja report that their equation of state provides accurate and consistent predictions of liquid and vapour densities for both pure components and mixtures. It was therefore accepted for prediction of the pure component and mixture vapour densities reported in this investigation. Figure C.6 shows saturated vapour densities for acetone-water, isopropanol-water, and acetone-isopropanol binary mixtures.

Algorithm for Estimation of Density of Vapour Mixture

Step 1:Calculate the values of parameter δ_c appearing in Eqs.(C-23) through (C-25) using Eq.(C-28).

Step 2: Calculate the values of Λ_b (the smallest root) from Eq.(C- 24) by substitution the value δ_c calculated in step 1 for each component.

Step 3: Calculate the values of Λ_a in Eq.(C-23) by substitution the values of δ_c and Λ_b from steps 1 and 2, respectively for each component.

Step 4: Calculate the value of the slope F in Eq.(C-26) using Eq.(C-27) for each component.

Step 5: Calculate the value of $\alpha(T_R)$ in Eq.(C-26) by substitution the values of F from step 5 and reduce temperature, T_R

Step 6: Calculate the value of a(T) for each component from Eq.(C-20), by substituting the value of Λ_a from step 4, and critical parameters from Table C.2.

Step 7: Calculate the value of cross-interaction term a_{ij} from Eq.(C-32) by substitution the value of a(T) for each component from step 6, and the values of Ω_{ij} are taken equal 1 for the systems investigated as suggested by Patel and Teja (1982).

Step 8: Calculate the value of of $a_m(T)$ from Eq.(C-29) using the values of a_{ij} from step 7.

Step 9: Calculate the value of constant b from Eq.(C-21) using the value of of Λ_b from step 2 and the critical properties values from Table C.2.

Step 10: Calculate the value of Λ_c from Eq.(C-25) using the value of δ_c from step 1.

Step 11: Calculate the value of constant c from Eq.(C-22) using the value of Λ_c from the step 10.

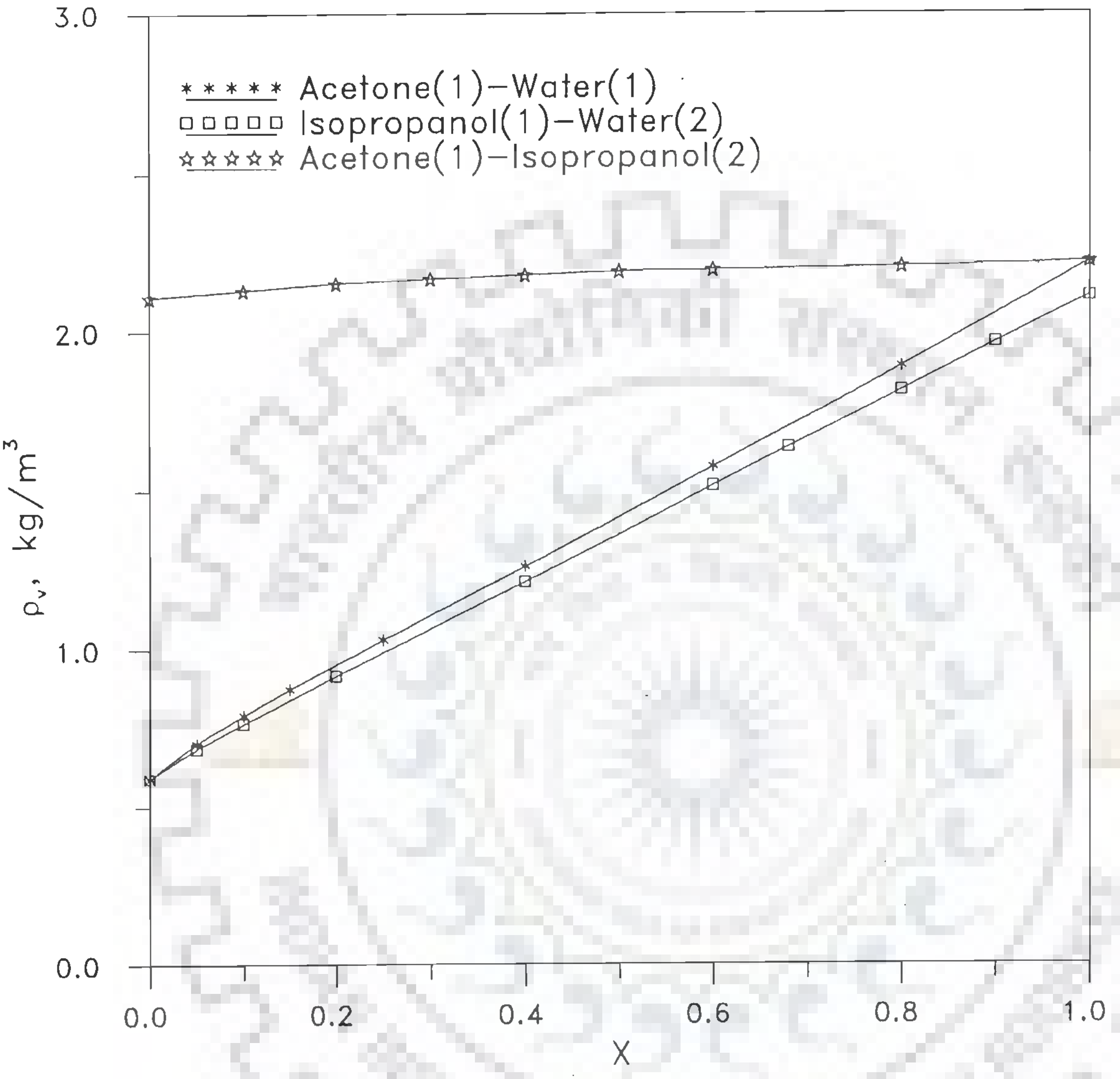


Fig. C.6 Variation of vapour density with mole fraction of more volatile component

Step 12: Calculate the value of specific volume of vapour, v (denotes the maximum positive root of Eq.(C-19)) using the values of $a_m(T)$, b, and c, from the steps 8, 9 and 11, respectively.

Step 13: Calculate the value of MW_m from Eq.(C-9) by substitution the values of molecular weight of each component listed in Table C.2.

Step 14: Calculate the vapour specific volume of mixture by dividing the value of v obtained in step 12 by MW_m from step 13.

Step 15: Calculate vapour density of mixture as follows:

Density = 1/(specific volume).

C.3.3 DENSITY OF LIQUID MIXTURE

Pure component liquid specific volumes at saturation temperature were estimated using modified Rackett equation, as reported by Spencer and Danner (1972) as follows:

$$v = (RT_{cr}/p_{cr})Z_{RA}^{\,n} \tag{C-33}$$

where $n = [1+(1-T_R)^{2/7}]$ and Z_{RA} is a specified constant for each liquid. Table C.2 lists the values of Z_{RA} used in the present investigation. The generalized corresponding states method proposed by Teja (1980) was used to estimate saturated liquid mixture specific volume. The liquid specific volume is obtained from:

$$(Z_{cr}v) = (Z_{cr}v)^{(r1)} + [(\omega-\omega^{(r1)})/(\omega^{(r2)}-\omega^{(r1)})]\ [(Z_{cr}v)^{(r2)}-(Z_{cr}v)^{(r1)}] \tag{C-34}$$

For more general case of a mixture with n components, Eq.(C-18) is used with $\Phi = v$, and $\varepsilon = Z_{cr}$.

Figure C.7 shows saturated liquid densities for acetone-water, isopropanol-water, and acetone-isopropanol binary mixtures.

Algorithm for Estimation of Density of Liquid Mixture

Step 1: Calculate pseudocritical properties of mixture v_{cm}, T_{cm}, Z_{12}, MW_m, and ω_{12} from Eqs. (C-6),(C-7), (C- 8),(C-9),(C-12), and (C-13) respectively. The value of MW_m is calculated from linear mixing law of molecular weight ,MW of pure components listed in Table C.2. The critical property values have been taken from Table C.2.

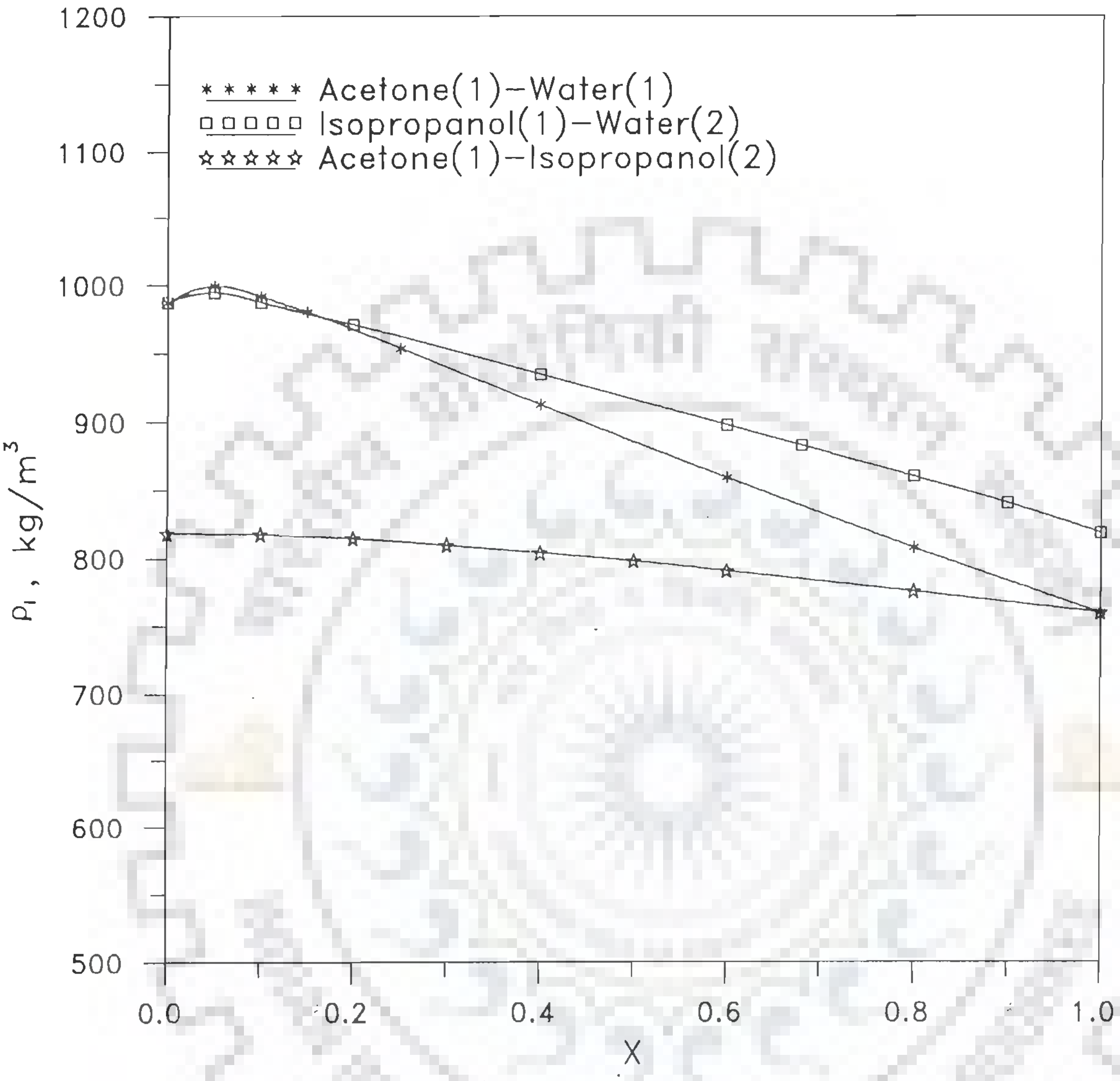


Fig. C.7 Variation of liquid density with mole fraction of more volatile component

Step 2: Calculate specific volume v_i of each pure liquid from Eqs. (C-33), where $T_R = T_i/T_{cm}$.

Step 3: Calculate specific volume of liquid mixture from Eqs.(C-34) and (C-16).

Step 4: Calculate density of liquid mixture as follows:

Density = 1/(specific volume).

C.3.4 LATENT HEAT OF VAPORIZATION

The latent heat of vaporization (enthalpy) of a mixture can be expressed as:

$$H(T,p,X_i) = \Sigma X_i H_i (T,0,X_i) - H^D (T, p, X_i) \tag{C-35}$$

where X_i represents either liquid or vapour mole fraction and where all $H_i(T,0,X_i)$ are at the same reference state. The term $H^D(T, p, X_i)$ is referred to as isothermal enthalpy departure and accounts for the difference between the mixture enthalpy at the pressure p, and the enthalpy of the mixture of ideal gases at zero pressure.

It is important to distinguish between two types of vaporization enthalpies when discussing mixtures. The integral latent heat is the difference between the enthalpies of the vapour and liquid at the same composition. The differential latent heat is the difference between the enthalpy of the liquid mixture and the enthalpy of vapour mixture at the equilibrium compositions. Only the differential latent heat is of interest in vapour-liquid equilibrium calculations, (see Thome(1990)).

The isothermal enthalpy departure is determined from classical thermodynamics and equation of state. From classical thermodynamics,

$$dH = d(pV) + [T\ (dp/dT) - p]_T\ dT \tag{C-36}$$

The isothermal enthalpy departure is determined by integration of this expression from zero pressure to p. Using Patel-Teja (1982) equation of state, which is applicable for both vapour and liquid states, the resulting expression for the enthalpy departure is:

$$\Psi(H - H^o) = RT(Z-1) - [T\ (da/dT) - a]\ [(1/2N)\ \ln((Z + M)/(Z + Q))] \tag{C-37}$$

or

$$\Psi(H - H^o) = H^D (T, p, X_i)$$

Where Ψ is a conversion factor and M, N, and Q are given by:

$$M = [\{(b+c)/2\} - N]\ (p/RT) \tag{C-38}$$

$$N = [bc + \{(b+c)^2\} / 2]^{-1/2} \tag{C-39}$$

$$Q = [\{(b + c)/ 2\} + N]\ (p/RT) \tag{C-40}$$

Equations. (C-20) through (C-22) are the appropriate expressions to be used for determination of values of a, b, and c for pure components, and Eqs.(C-29) through (C-31) are used to calculate a_m, b_m, and c_m in the case of mixtures. H° represents ideal gas enthalpy and can be estimated by integrating Cp from 0 K to the saturation temperature. Figure C.8 shows the differential latent heat for acetone-water, isopropanol-water, and acetone-isopropanol binary mixtures.

Algorithm for Estimation of Latent Heat of Vaporization

Step 1: Calculate the enthalpy departure of the liquid mixture, Hml ($= \psi(H-H^o)$) using Eq.(C-37) from the following steps:

(i) Calculate the value of cross-interaction term $a_m(T)$ from step 8, of subsection C.3.2

(ii) Calculate pseudocritical properties of mixture v_{cm}, T_{cm}, Z_{cm}, WM_m, Z_{12} and w_{12} from step 1 of subsection C.3.3.

(iii) Calculate the value of p_{cm} as follows:

$$p_{cm} = (Z_{cm})\,(R)\,(T_{cm}) \,/\, V_{cm} / MW_m$$

using the pseudocritical properties of mixture from the step 1 of subsection C.3.3.

(iv) Calculate the value of (da/dT) by differentiating Eq.(C-20) by T and putting the values of different parameters.

(v) Calculate the value of Z appearing in Eq.(C-37) from the expression given below:

$$Z = p\,(R)\,(v_l)\,(MW_m) \,/\, (T_s)$$

The value of v_l is calculated based on step 3, of subsection C.3.3, for p equal to 98.1 kPa, and T_s equal to the saturated liquid temperature of mixture.

(vi) Calculate the value of constants M, N, and Q from Eqs.(C- 38), (C-39), and (C-40) respectively, using the value of b and c from the steps 9 and 11 of subsection C.3.2, respectively.

(vii) Calculate the value of $\Psi(H - H^o)$ from Eq.(C-37)

Step 2: Calculate the vapour phase enthalpy departure for the mixture, (Hmv) using Eq.(C-37) following the steps (i - vii) used in the calculation of enthalpy departure for liquid mixture, except the step (v) for the estimation the value of Z. The following equation is used for estimation of

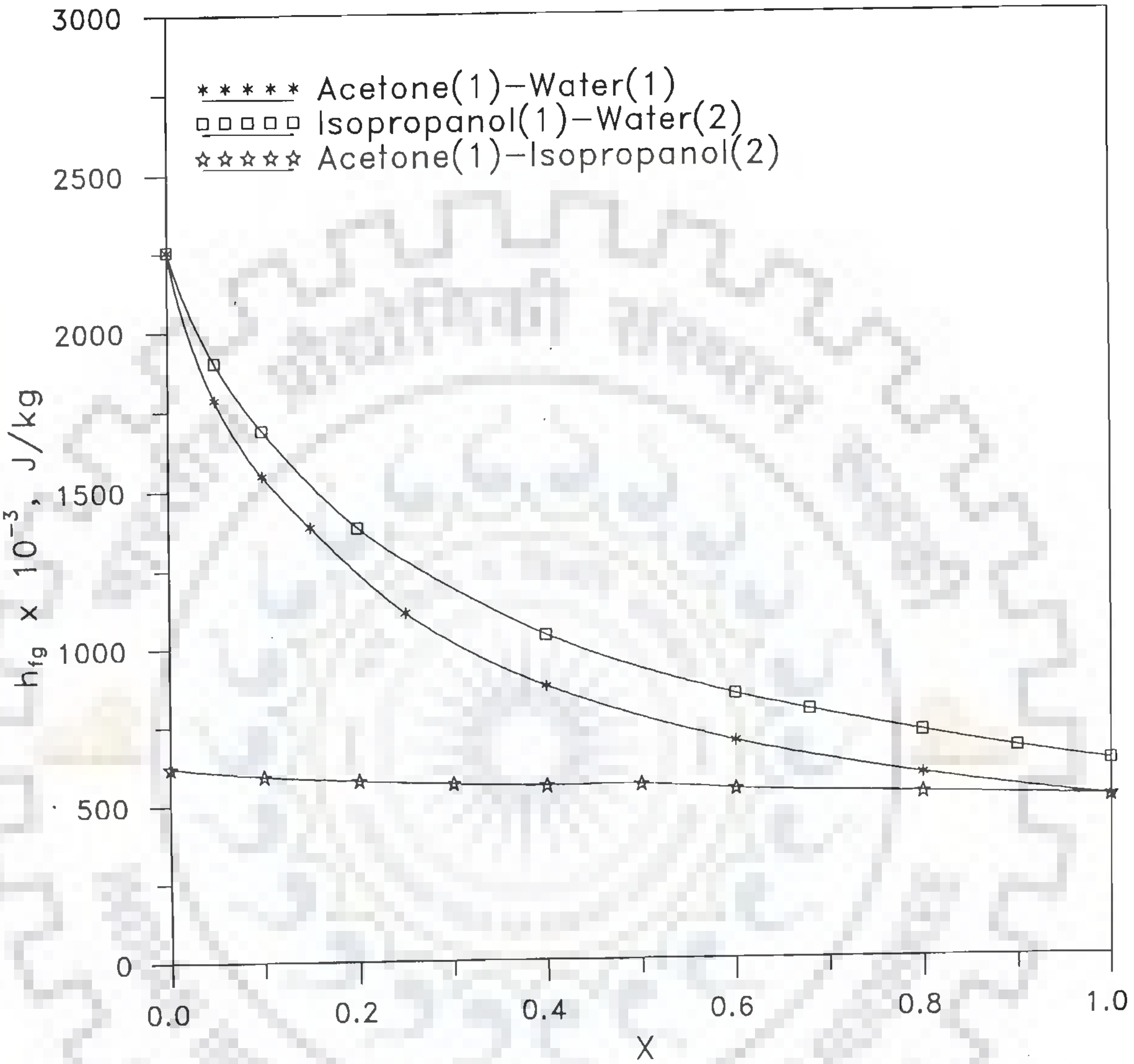


Fig. C.8 Variation of differential heat of vaporization with mole fraction of more volatile component

Z in case of vapour:

$$Z = p\ (R)\ (v_v)\ (MW_m)\ /\ (T_s)$$

using the value of v_v from the step 14, of section C.3.2, for p equal to 98.1 kPa, and T_s equal to the saturated liquid temperature of mixture.

Step 3: Calculate the ideal gas enthalpy (Hov - Hol) as follows:

(i) Calculate the enthalpy H for each component from the following equation:

$$H_i = A\ (T_s - T_{ref}) + B\ (T_s^2 - T_{ref}^2)/2 + C\ (T_s^3 - T_{ref}^3)/3 + D\ (T_s^4 - T_{ref}^4)/4$$

the value of constants A, B, C, and D have taken from Table C.3, and T_{ref} is taken to be 0.

Table C.3 Parameters for calculation of enthalpy, H_i

Component	A	B	C	D
Acetone	6.301E+00	2.606E-01	-1.253E-04	2.038E-08
Isopropanol	3.243E+01	1.885E-01	6.406E-05	-9.261E-08
Water	3.244E+01	1.924E-03	1.055E-05	-3.596E-09
Methanol	2.115E+01	7.092E-02	2.587E-05	-2.852E-08
n-Propanol	2.470E+00	3.325E-01	-1.855E-04	4.296E-08
Ethanol	9.014E+00	2.141E-01	-8.390E-05	1.373E-09

(ii). Calculate the value of Hov as follows:

$$Hov = \Sigma\ Y_i\ H_i$$

(iii). Calculate the value of Hol as follows:

$$Hol = \Sigma\ X_i\ H_i$$

Finally, latent heat of vaporization (Hm)for the mixture is equal

$$Hm = [(Hmv) - (Hml)] + [(Hov) - (Hol)]$$

C.3.5 SPECIFIC HEAT OF LIQUID MIXTURES

The specific heat capacity of pure liquid components was determined using the following expression:

$$Cp(T) = A + BT + CT^2 + DT^3 \qquad (C\text{-}41)$$

The values of A, B, C, and D were obtained from Miller et al.(1976), and Perry et al.(1984). Table C.4 lists the constants used for each fluid in Eq.(C-41).

Table C.4 Constants for calculation of liquid specific heat, Eq.(C-41)

Component	A	B	C	D
Acetone	3.339	-1.031E-02	2.154E-05	0.00E+00
Isopropanol	-1.964	1.961E-02	-1.297E-05	0.00E+00
Water	2.822	1.183E-02	-3.504E-05	3.60E-08
Methanol	3.509	-1.352E-02	3.473E-05	-7.07E-10
n-Propanol	-1.156	3.589E-02	1.432E-04	2.080E-07
Ethanol	-1.465	4.001E-02	-1.585E-04	2.280E-07

The specific heat capacity for liquid mixtures was determined using a method proposed by Teja(1983). For binary mixture, liquid specific heat is determined from the following equation:

$$(Cp/R) = (Cp/R)^{(r1)} + [(\omega-\omega^{(r1)}) / (\omega^{(r2)}-\omega^{(r1)})] [(Cp/R)^{(r2)} - (Cp/R)^{(r1)}] \qquad (C\text{-}42)$$

For the more general case of a mixture with n components, Eq.(C-18) is used, with Φ = Cp and $\varepsilon = (1/R)$. Figure C.9 shows liquid specific heat for acetone-water, isopropanol-water, and acetone-isopropanol binary mixtures.

Algorithm for Estimation of Specific Heat of Mixture

Step 1: Calculate pseudocritical properties of mixture v_{cm}, T_{cm}, Z_{cm}, WM_m, Z_{12} and ω_{12} from step 1 of subsection C.3.3.

Step 2: Calculate Cp from Eq.(C-41) for each liquid using T_s and the values of constant A, B, and C from the Table C.4.

Step 3: Calculate the liquid specific heat of mixture from Eq.(C-42) using the the value of Cp for each liquid from step 2, and the values of ω for each component from Table C.2.

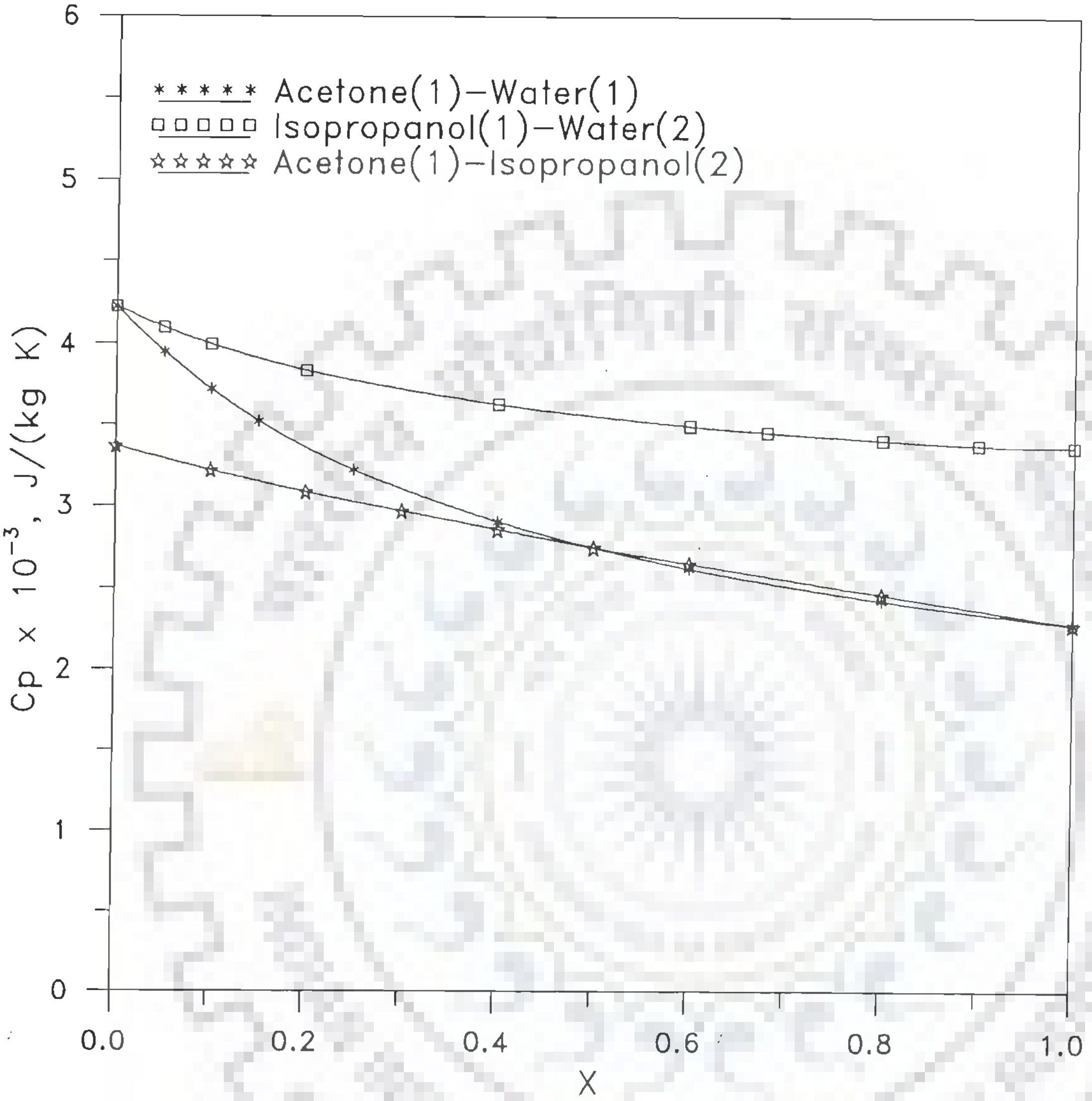


Fig.C.9 Variation of liquid specific heat with mole fraction of more volatile component

C.3.6 VISCOSITY

Pure component liquid viscosities were estimated using the following expression suggested by Yaws et al.(1976):

$$\ln \mu = A + B/T + CT + D\,T^2 \qquad \text{(C-43)}$$

Values for the constants A, B, C,and D were obtained from Reid et al. (1987) as shown in Table C.5.

Table C.5 Values of constants A, B, C, and D for Eq.(C-43)

Liquid	A	B	C	D
Acetone	-4.033E+00	8.456E+02	0.00E+00	0.00E+00
Isopropanol	-8.114E+00	2.624E+03	0.00E+00	0.00E+00
Water	-2.471E+01	4.209E+03	4.527E-02	-3.376E-05
Methanol	-3.935E+01	4.826E+03	1.091E-01	-1.127E-04
n-Propanol	-1.228E+01	2.666E+03	2.008E-02	-2.233E-05
Ethanol	-6.210E+00	1.614E+03	6.180E-08	-1.132E-05

The viscosity of liquid mixtures was estimated using the method proposed by Teja and Rice(1981). The viscosity for a liquid mixture is estimated in a manner analogous to the generalized method of corresponding states. For a binary mixture, the viscosity is given by:

$$\ln(\mu\varepsilon) = \ln(\mu\varepsilon)^{(r1)} + [(\omega-\omega^{(r1)}) / (\omega^{(r2)}-\omega^{(r1)})]\,[\ln(\mu\varepsilon)^{(r2)} - \ln(\mu\varepsilon)^{(r1)}] \qquad \text{(C-44)}$$

where ε is given by:

$$\varepsilon = V_{cr}^{2/3}\,T_{cr}^{-0.5}\,M^{-0.5} \qquad \text{(C-45)}$$

The general expression for liquid viscosity in an n component mixture based on Eq.(C-18) is

$$\ln(\mu\varepsilon)_{12..n} = \ln(\mu\varepsilon)_{12..n-1} + [(\omega_{12..n}-\omega_n) / (\omega_{12..n}-\omega\varepsilon_n)]\;[\ln(\mu\varepsilon)_n - \ln(\mu\varepsilon)_{12..n-1}] \qquad \text{(C-46)}$$

Figure C.10 shows the variation of liquid viscosity with composition in acetone-water, isopropanol-water, and acetone-isopropanol binary mixtures.

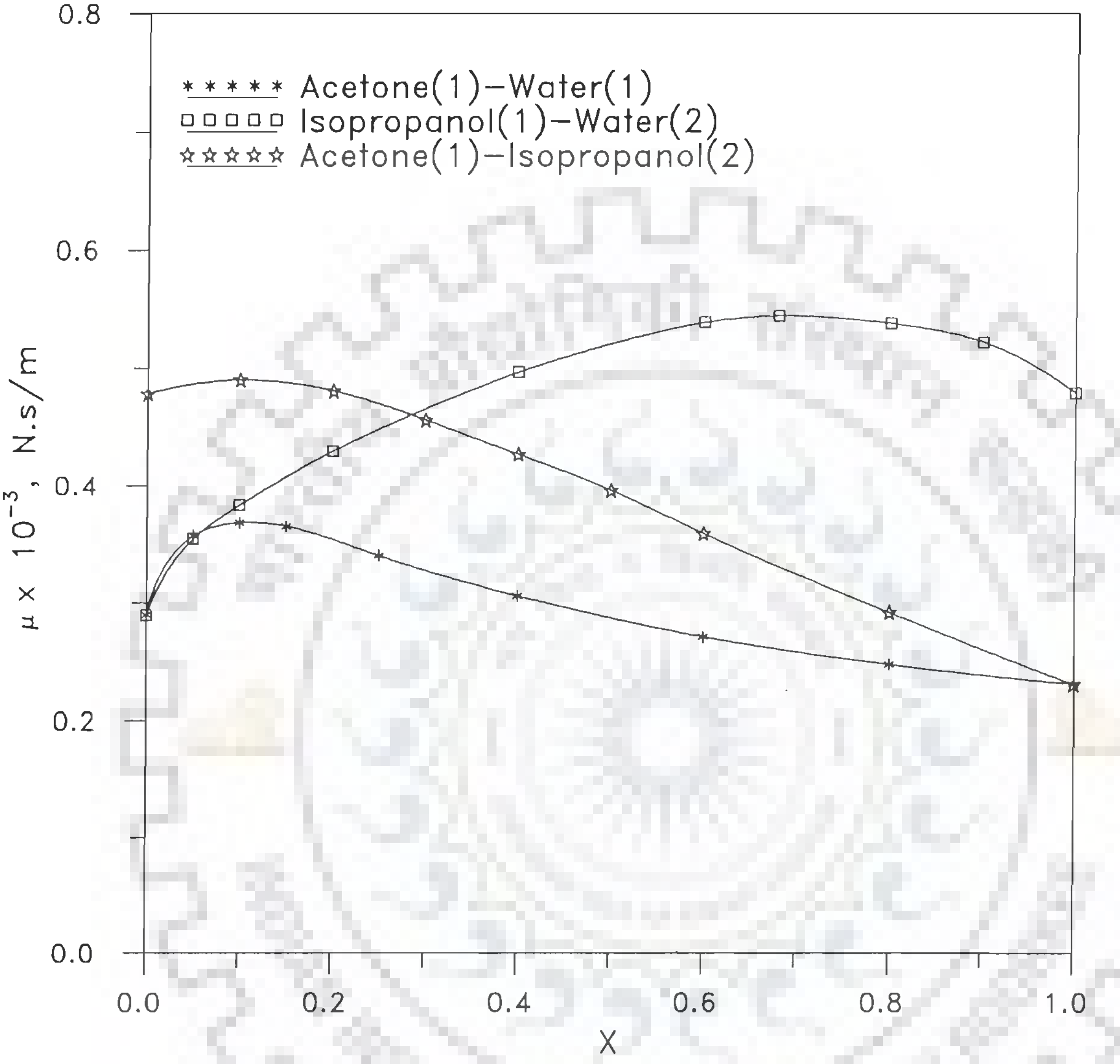


Fig. C.10 Variation of viscosity with mole fraction of more more volatile component

Algorithm for Estimation of Viscosity of Mixture

Step 1: Calculate pseudocritical properties of mixture v_{cm}, T_{cm}, Z_{cm}, WM_m, Z_{12} and ω_{12} from step 1 of subsectio C.3.3.

Step 2: Calculate the viscosity of each liquid from Eq.(C-43) using the values of constants A, B, C, and D from Reid et al.(1987),Table C.5.

Step 3: Calculate the function critical parameter, ε from Eq.(C- 45).

Step 4: Calculate the value of $(\mu\varepsilon)$ for each liquid from steps 2 and 3.

Step 5: Calculate the reduce viscosity $(\mu\varepsilon)$ of mixture from Eq.(C-46).

Step 6: Calculate the viscosity of mixture by dividing the value obtained by step 6 by the value obtained in step 3.

C.3.7 SURFACE TENSION OF LIQUID MIXTURES

The surface tension of a pure liquid decreases with increasing temperature and can be estimated using the following equation:

$$\sigma = A - BT \qquad \text{(C-47)}$$

Values of A and B were taken from the of experimental data for surface tension of liquids by Jasper (1972). Table C.6 lists the values of A and B for the liquids used in this investigation. Aqueous mixtures show pronounced non-linear variation of surface tension with composition. The surface tension decreases dramatically upon the addition of an organic liquid. The corresponding states method proposed by Rice and Teja (1982) was used to estimate the surface tension of liquid mixtures in this investigation. This method is currently the only method which is capable of estimating the surface tension of mixtures with more than two components. For binary mixture, surface tension is calculated from the equation as given below:

$$(\sigma\varepsilon) = \ln(\sigma\varepsilon)^{(r1)} + [(\omega-\omega^{(r1)}) / (\omega^{(r2)}-\omega^{(r1)})] \; [\ln(\sigma\varepsilon)^{(r2)} - \ln(\sigma\varepsilon)^{(r1)}] \qquad \text{(C-48)}$$

where ε is given by:

$$\varepsilon = T_{c\,r}^{-1} \, v_{c\,r}^{2/3} \qquad \text{(C-49)}$$

The general expression for surface tension for an n component mixture based on Eq.(C-18) is as follows:

$$(\sigma\varepsilon)_{12..n} = (\sigma\varepsilon)_{12..n-1} + [(\omega_{12..n} - \omega_n) / (\omega_{12..n} - \omega_n)] \; [(\sigma\varepsilon)_n - (\sigma\varepsilon)_{12..n-1}] \qquad \text{(C-50)}$$

Figure C.11 shows the surface tension for acetone-water, isopropanol-water, and acetone-isopropanol mixtures as a function of the liquid mole fraction of more volatile component.

Table C.6 Values of A and B for Eq.(C-47)

Component	A	B
Acetone	26.26	0.1120
Isopropanol	22.90	0.0789
Water	75.83	0.1477
Methanol	24.00	0.0773
n-Propanol	25.26	0.0777
Ethanol	24.05	0.0832

Algorithm for Estimation of Surface Tension of Mixture

Step 1: Calculate pseudocritical properties of mixture v_{cm}, T_{cm}, Z_{cm}, WM_m, Z_{12} and ω_{12} from step 1 of subsection C.3.3.

Step 2: Calculate the value of function parameter ε from Eq.(C-49) for each liquid.

Step 3: Calculate the value of surface tension , ε of each liquid from Eq.(C-47) using the values of constants A and B from in Table C.6.

Step 4: Calculate the reduce surface tension $(\sigma\varepsilon)$ of each component of mixture from steps 2 and 3.

step 5: Calculate the reduce surface tension $(\sigma\varepsilon)$ of mixture using Eq.(C-50).

Step 6: Calculate the surface tension of mixture by dividing the value of step 5 by the value of ε obtained from step 2.

C.3.8 THERMAL CONDUCTIVITY OF LIQUID MIXTURES

Two methods were used to determine the thermal conductivity of liquid components. For water, methanol, n-propanol, and ethanol, the thermal conductivity was estimated using the method

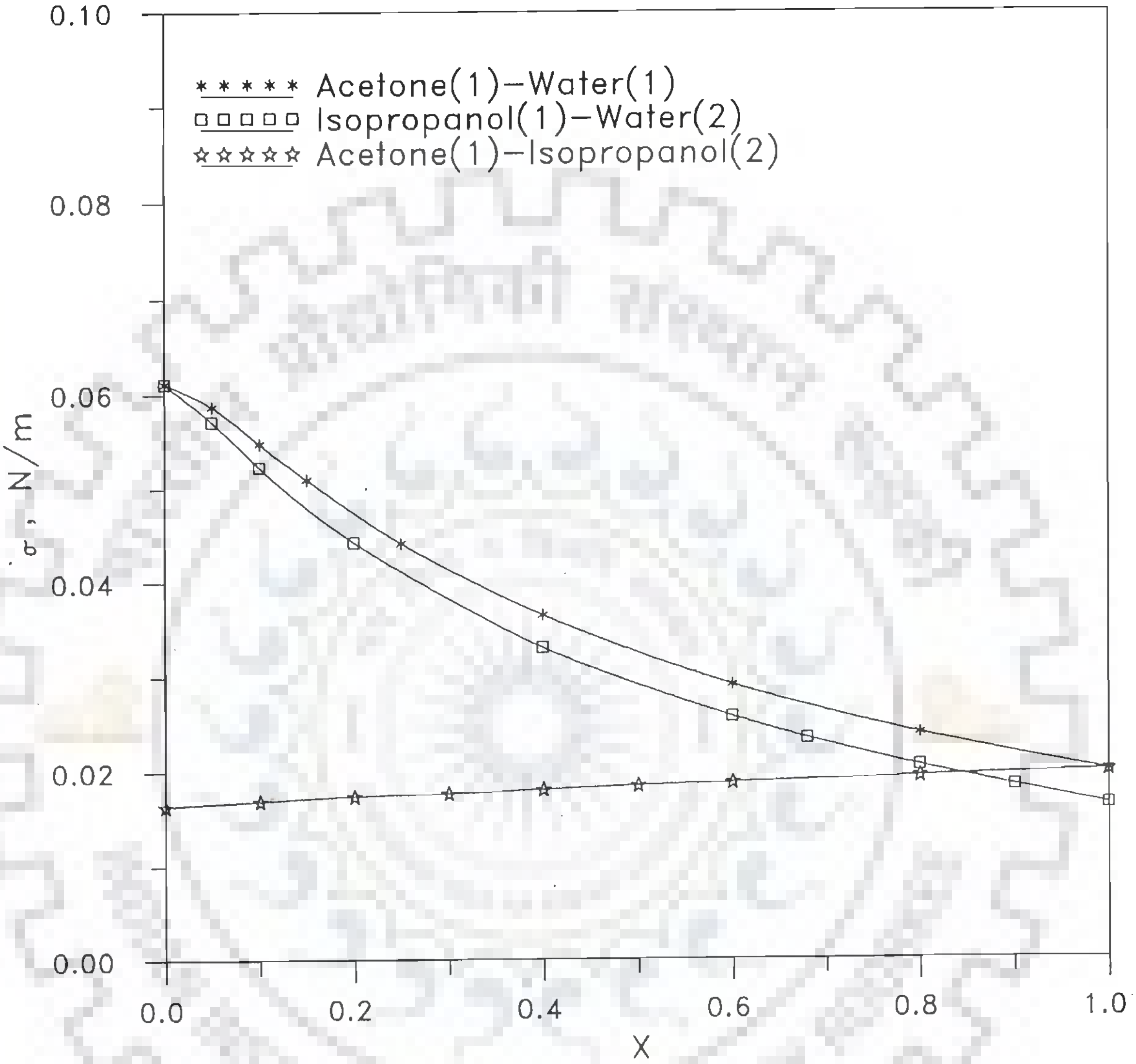


Fig. C.11 Variation of surface tension with mole fraction of more volatile component proposed by Miller et al. (1976):

$$k(T) = A + BT + CT^2 \qquad (C\text{-}51)$$

The values of A, B, and C in Eq.(C-51) are listed in Table C.7.

Table C.7 Values of A, B, and C in Eq.(C-51)

Component	A	B	C
Water	-0.3838	5.254E-03	-6.369E-06
Methanol	0.3225	-4.785E-04	1.168E-07
Ethanol	0.2629	-3.847E-04	2.211E-07
n-Propanol	0.1854	-3.366E-05	-2.215E-07

The values of A, B, and C for the estimation of thermal conductivity of acetone and isopropanol were not readily available and therefore the method of Latini and Pacetti (1977) was used instead of Eq.(C-51). Latini and co-workers had recommended a correlation of the following form:

$$k = E\,(1-T_R)/T_R^{1/6} \qquad (C\text{-}52)$$

where

$$E = A^*\,T^{\tau} / (M^{\beta}\,T_{cr}\,\Gamma) \qquad (C\text{-}53)$$

The values of the constants A^*, τ, β and Γ are listed in Table C.8.

Table C.8 Constants in Eq.(C-53)

Component	A^*	τ	β	Γ
Acetone	3.83E-03	1.2	0.5	0.167
Isopropanol	3.93E-03	1.2	0.5	0.167

The method proposed by Li (1976) was used to estimate the thermal conductivity of liquid mixtures.

$$k = \sum_i \sum_j \Phi_i \Phi_j k_{ij} \qquad (C\text{-}54)$$

where

$$k_{ij} = 2(k_i^{-1} + k_j^{-1})^{-1} \tag{C-55}$$

and

$$\Phi_i = X_i v_i / \Sigma X_j v_j \tag{C-56}$$

In the above equations, Φ_i is the superficial volume fraction of component i and v_i represents the liquid molar volume of the pure component. Reid et al.(1987) have reviewed the above and several other methods for the estimation of liquid thermal conductivity and found that the absolute error of the Li method hardly exceeds 5%. Figure C.12 shows the liquid thermal conductivities for acetone-water, isopropanol-water and acetone-isopropanol.

Algorithm for Estimation of Thermal Conductivity of Mixture

Step 1: Calculate the thermal conductivity of each liquid either from Eq.(C-51) or Eq.(C-52).

Step 2: Calculate the value of Φ for each liquid using the value of specific volume from step 2 in Section C.3.3.

Step 3: Calculate the value of k_{ij} by substituting the value of k in Eq.(C-52) for each liquid.

Step 4: Calculate the liquid thermal conductivity of mixture using Eq.(C-54) by substituting the values of Φ and k_{ij} from steps 2 and 3, respectively.

C.3.9 MASS DIFFUSION COEFFICIENT

The diffusion flux for a binary mixture is given by the following equation:

$$J_1 = - c\, D_{12}\, dX_1/dz \tag{C-57}$$

where c is the total molar concentration and z is the direction in which diffusion occurs. The diffusion coefficient represents the proportionality between the molar flux of component 1 relative to a plain of no net molar flow, and the composition gradient.

Several methods have been proposed for the estimation of binary diffusion coefficients in liquid mixtures. However, none have been proved satisfactory as reported by Tyn and Calus (1975) and Reid et al.(1987). The most widely used correlation for the estimation of binary diffusion

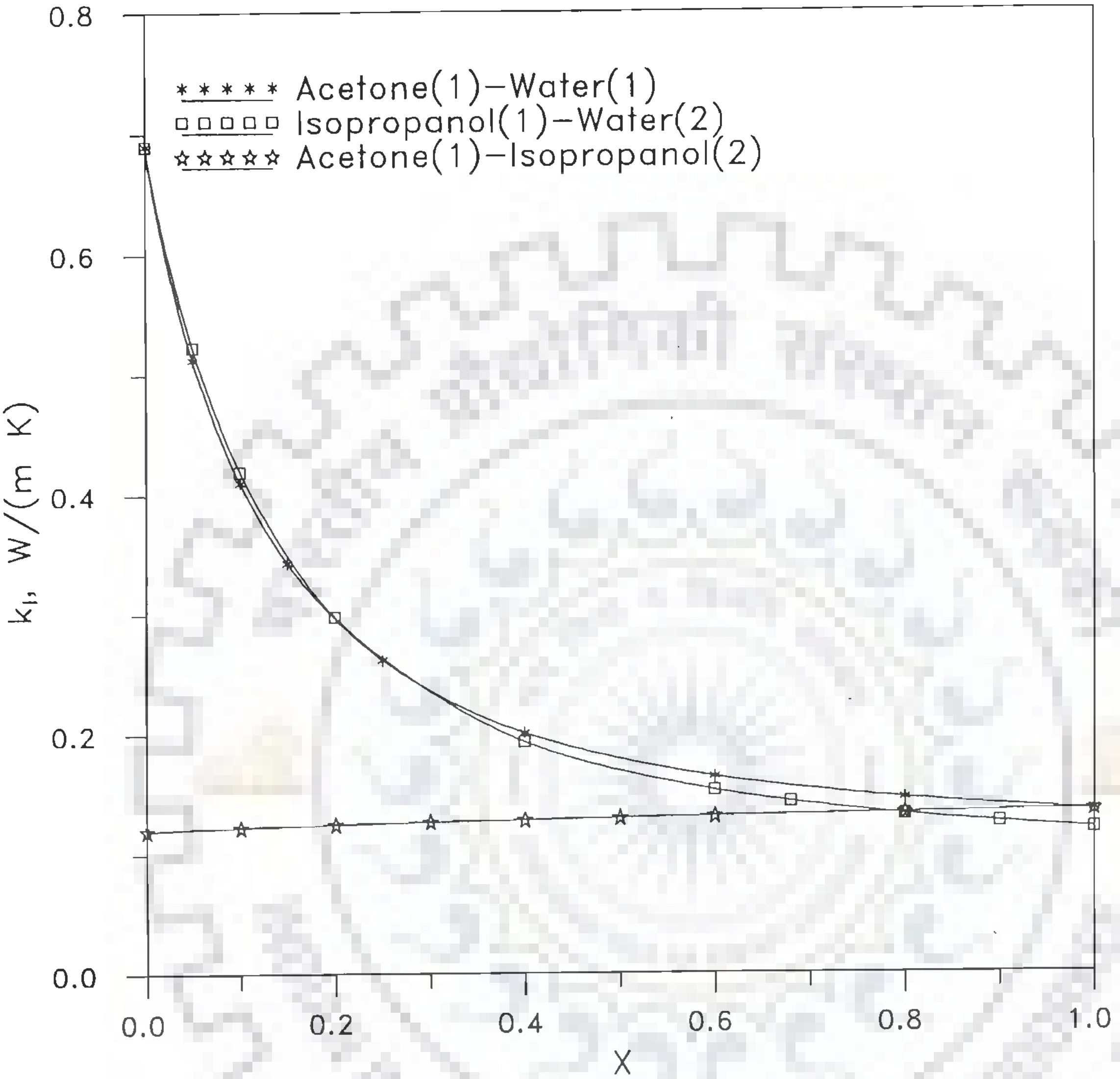


Fig. C.12 Variation of liquid thermal conductivity with mole fraction of more volatile component coefficient is the one due to Wilke-Chang(1955). This technique was used to estimate binary mass diffusivities in the present investigation. The Wilke-Chang correlation is:

$$D_{12} = 7.4 \times 10^{-8} (\Phi\, MW_2)^{0.5}\, T / (\mu_2\, V_1^{0.6}) \qquad \text{(C-58)}$$

where,

D_{12} = mutual diffusion coefficient of solute 1 at very low concentration in solute 2, cm^2/s.

MW_2 = molecular weight of solvent 2, gm/mol.

T = Temperature, K.

μ_2 = viscosity of solvent 2, cp

V_1 = molar volume of solute 1 at its normal boiling point, $cm^3/mole$.

Φ = association factor(dimensionless)

Wilke and Chang recommended that the associated factor Φ should be chosen as 2.6, if the solvent is water, 1.9 if it is methanol, 1.5 if it is ethanol, and 1.0 for other solvents. An average error of 10% was reported by Wilke and Chang for 250 solute-solvent systems tested by them.

Algorithm for Estimation of Mass Diffusion coefficient of Mixture

Calculate mass diffusion coefficient from Eq.(C-58), using the value of μ and V from steps 2 of subsection C.3.6 and step 2 of subsection C.3.3, respectively.

A complete listing of the physico-thermal properties of ternary mixtures used in this investigation are also given in Table C.9.

Table C.9 Thermodynamic and transport properties of the acetone(1)-isopropanol(2)-water(3) ternary mixture.

Composition	ρ_l, kg/m^3	ρ_v, kg/m^3	Cp, kJ/kg K	k, W/m K	σ, N/m	h_{fg}, kJ /kg	μ,N s/m^2	ΔT_{BR}, K
X_1 = 0.011 X_2 = 0.074	905.34	0.7503	3.982	0.4448	0.0542	1699.2	0.3864	17.3
X_1 = 0.020 X_2 = 0.019	911.47	0.6770	4.059	0.5475	0.0589	1886.1	0.3508	17.1
X_1 = 0.050 X_2 = 0.040	918.67	0.7693	3.860	0.4313	0.0548	1661.2	0.3864	24.1
X_1 = 0.083 X_2 = 0.036	921.87	0.8181	3.724	0.3838	0.0526	1541.1	0.3853	26.2
X_1 = 0.124 X_2 = 0.290	911.07	1.2541	3.366	0.1917	0.0336	876.3	0.4538	12..8
X_1 = 0.155 X_2 = 0.150	911.24	1.0960	3.373	0.2318	0.0395	1067.8	0.4038	20.1
X_1 = 0.212 X_2 = 0.182	907.88	1.2354	3.216	0.1993	0.03527	927.2	0.4037	17.6
X_1 = 0.266 X_2 = 0.282	902.16	1.4693	3.080	0.1652	0.02911	750.5	0.4170	12.2
X_1 = 0.294 X_2 = 0.484	898.81	1.8218	3.001	0.1389	0.0224	606.6	0.4488	6.5
X_1 = 0.389 X_2 = 0.212	892.34	1.5617	2.893	0.1595	0.0280	702.8	0.3711	13.2
X_1 = 0.524 X_2 = 0.267	877.89	1.8647	2.718	0.1419	0.0231	567.4	0.3562	6.8
X_1 = 0.699 X_2 = 0.105	857.98	1.888	2.532	0.1430	0.0234	503.1	0.2848	5.9

APPENDIX D

EXPERIMENTAL DATA

Table D.1 Experimental Data of Heat Transfer From A horizontal **Plain Tube** To Pool of Saturated **Pure Liquids**

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					Acetone					
1	10391	65.008	64.437	64.300	64.460	56.121	56.075	56.051	56.075	1227
2	18893	67.147	65.799	65.548	65.526	56.192	56.121	56.075	56.121	1913
3	27395	68.828	67.002	66.659	66.591	56.192	56.145	56.121	56.145	2464
4	35897	70.315	68.227	67.861	67.725	56.238	56.192	56.168	56.192	2910
5	44399	70.916	68.903	68.516	68.379	56.238	56.192	56.168	56.192	3420
6	52901	71.630	69.413	69.187	69.051	56.238	56.192	56.168	56.192	3883
7	57624	72.065	69.734	69.531	69.418	56.238	56.192	56.168	56.192	4119
					Isopropanol					
1	10391	90.781	89.557	89.778	89.203	82.478	82.367	82.323	82.367	1396
2	18893	92.576	91.047	91.288	90.392	82.522	82.434	82.389	82.434	2127
3	27395	94.064	92.514	92.536	91.685	82.544	82.478	82.434	82.478	2681
4	35897	94.789	93.304	93.261	92.453	82.544	82.500	82.456	82.500	3278
5	44399	95.514	93.942	93.876	93.156	82.544	82.500	82.456	82.500	3820
6	52901	96.064	94.579	94.382	93.858	82.544	82.500	82.456	82.500	4329
7	57624	96.456	94.863	94.710	94.229	82.544	82.500	82.456	82.500	4586
					Distilled Water					
1	18893	106.149	105.183	105.784	105.269	100.107	100.064	100.043	100.064	3418
2	27395	106.879	105.570	106.385	105.462	100.150	100.086	100.064	100.086	4583
3	35897	107.758	105.977	106.964	105.827	100.193	100.129	100.107	100.129	5529
4	44399	108.724	106.492	107.587	106.299	100.193	100.129	100.107	100.129	6222
5	52901	109.273	106.793	107.801	106.557	100.193	100.129	100.107	100.129	7085
6	57624	109.756	106.983	107.970	106.790	100.193	100.129	100.107	100.129	7449

Table D.2 Experimental Data of Heat Transfer From A horizontal **Plain Tube** To Pool of Saturated **Acetone(1)-Water(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					$X_1 = 0.05$					

1	10391	84.579	84.247	84.291	83.760	75.882	75.792	75.769	75.792	1236
2	18893	86.843	86.289	86.312	85.891	76.041	75.973	75.928	75.973	1825
3	27395	88.332	87.248	87.558	87.093	76.041	75.973	75.928	75.973	2366
4	35897	89.614	88.184	88.848	88.007	76.063	75.973	75.928	75.973	2831
5	44399	90.797	88.811	89.728	88.723	76.063	75.973	75.928	75.973	3281
6	52901	91.784	89.339	90.430	89.404	76.063	75.973	75.928	75.973	3711
7	57624	92.395	89.666	90.867	89.862	76.063	75.973	75.928	75.973	3916
					x_1 = 0.10					
1	18893	82.130	81.201	81.135	81.179	69.224	69.201	69.201	69.201	1548
2	27395	84.350	82.779	83.000	83.089	69.269	69.224	69.224	69.224	1947
3	35897	85.928	84.114	84.401	84.534	69.315	69.247	69.247	69.247	2319
4	44399	87.196	85.028	85.603	85.758	69.315	69.269	69.247	69.269	2671
5	52901	88.310	85.898	86.518	86.805	69.315	69.269	69.247	69.269	3004
6	57624	89.011	86.254	87.094	87.382	69.361	69.338	69.315	69.338	3184
					x_1 = 0.15					
1	18893	78.661	78.073	78.027	78.299	65.388	65.365	65.342	65.365	1465
2	27395	80.876	80.058	80.013	80.478	65.434	65.411	65.388	65.411	1833
3	35897	82.543	81.437	81.525	82.100	65.479	65.434	65.434	65.434	2181
4	44399	83.900	82.484	82.462	83.413	65.502	65.479	65.457	65.479	2525
5	52901	85.345	83.310	83.487	84.548	65.502	65.479	65.479	65.479	2831
6	57624	86.386	83.643	83.886	85.192	65.502	65.479	65.479	65.479	2987
					x_1 = 0.25					
1	18893	75.584	75.177	74.792	75.448	62.831	62.740	62.694	62.740	1512
2	27395	77.520	76.592	76.026	76.864	62.877	62.785	62.763	62.785	1964
3	35897	79.600	78.098	77.668	78.415	62.922	62.854	62.831	62.854	2304
4	44399	81.223	79.276	79.010	79.541	62.900	62.854	62.831	62.854	2627
5	52901	82.225	80.146	79.814	80.301	62.922	62.854	62.831	62.854	2979
6	57624	82.869	80.612	80.280	81.342	62.922	62.854	62.831	62.854	3130
					x_1 = 0.40					
1	10391	70.910	70.843	70.164	71.024	60.731	60.662	60.616	60.662	1032
2	18893	73.570	72.982	72.326	73.525	60.822	60.753	60.753	60.753	1532
3	27395	75.619	74.850	74.217	75.144	60.868	60.799	60.799	60.799	1937
4	35897	77.170	76.288	75.451	76.492	60.913	60.868	60.868	60.868	2320
5	44399	78.676	77.590	76.437	77.794	60.913	60.868	60.868	60.868	2651
6	52901	79.482	78.305	77.083	78.597	60.913	60.868	60.868	60.868	3025
7	57624	80.015	78.754	77.585	79.063	60.913	60.868	60.868	60.868	3206
					x_1 = 0.60					
1	10391	69.186	69.232	68.798	69.346	59.206	59.159	59.112	59.159	1041
2	18893	72.100	72.054	71.398	72.168	59.299	59.276	59.252	59.276	1493
3	27395	74.036	73.606	72.588	73.990	59.346	59.322	59.322	59.322	1926
4	35897	75.632	74.410	73.618	74.659	59.369	59.322	59.322	59.322	2355
5	44399	76.844	75.011	74.355	74.898	59.369	59.322	59.322	59.322	2785
6	52901	77.807	75.884	74.843	75.612	59.369	59.322	59.322	59.322	3167
7	57624	78.332	76.341	75.345	76.069	59.369	59.322	59.322	59.322	3353

					X_1 = 0.80					
1	10391	67.086	66.675	66.584	66.835	57.664	57.640	57.617	57.640	1135
2	18893	69.430	68.197	68.288	68.105	57.734	57.687	57.664	57.687	1747
3	27395	71.434	69.805	69.963	69.352	57.757	57.710	57.710	57.710	2206
4	35897	72.827	70.926	70.813	70.180	57.757	57.710	57.710	57.710	2666
5	44399	74.061	72.138	71.957	71.097	57.757	57.710	57.710	57.710	3043
6	52901	74.798	72.739	72.445	71.562	57.757	57.710	57.710	57.710	3489
7	57624	75.323	73.264	72.811	71.952	57.757	57.710	57.710	57.710	3690

Table D.3 Experimental Data of Heat Transfer From A horizontal **Plain Tube** To Pool of Saturated **Isopropanol(1)-Water(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					X_1 = 0.05					
1	18893	97.336	97.598	96.702	96.964	86.106	86.084	86.040	86.084	1706
2	27395	98.868	99.325	98.781	98.278	86.150	86.128	86.062	86.128	2158
3	35897	100.269	100.977	100.076	99.282	86.195	86.173	86.150	86.173	2568
4	44399	101.106	102.372	101.235	99.840	86.195	86.173	86.150	86.173	2967
5	52901	101.857	102.823	102.114	100.376	86.217	86.173	86.150	86.173	3388
6	57624	102.563	103.421	102.691	101.082	86.217	86.173	86.150	86.173	3544
					X_1 = 0.10					
1	18893	95.916	95.676	95.370	95.218	83.827	83.761	83.761	83.761	1606
2	27395	97.361	97.187	96.903	96.575	83.894	83.850	83.850	83.850	2084
3	35897	98.545	98.501	98.304	97.868	83.960	83.938	83.938	83.938	2500
4	44399	99.904	99.797	99.454	99.110	83.982	83.938	83.938	83.938	2843
5	52901	101.084	100.677	100.484	100.011	83.982	83.938	83.938	83.938	3184
6	57624	101.726	101.125	100.867	100.352	83.982	83.938	83.938	83.938	3376
					X_1 = 0.20					
1	10391	91.480	90.716	90.432	90.388	81.947	81.881	81.858	81.881	1173
2	18893	94.060	92.860	93.711	92.838	82.013	81.925	81.903	81.925	1654
3	27395	95.658	94.086	94.436	94.305	82.058	81.969	81.947	81.969	2168
4	35897	97.082	94.745	95.291	95.313	82.080	82.035	81.991	82.035	2645
5	44399	98.287	95.557	95.950	95.863	82.080	82.035	81.991	82.035	3088
6	52901	98.960	96.282	96.544	96.413	82.080	82.035	82.013	82.035	3525
7	57624	99.365	96.675	96.937	96.828	82.080	82.035	82.013	82.035	3739
					X_1 = 0.40					
1	10391	91.000	90.039	89.844	89.888	81.416	81.327	81.261	81.327	1173

2	18893	93.296	91.724	91.571	91.506	81.438	81.372	81.327	81.372	1774
3	27395	95.309	93.257	92.929	93.038	81.504	81.438	81.372	81.438	2246
4	35897	96.645	94.396	93.785	94.069	81.527	81.482	81.438	81.482	2711
5	44399	97.588	95.055	94.422	94.728	81.527	81.482	81.438	81.482	3179
6	52901	98.181	95.627	95.059	95.387	81.527	81.482	81.460	81.482	3629
7	57624	98.657	96.063	95.561	95.758	81.527	81.482	81.460	81.482	3836
						x_1 = 0.60				
1	10391	89.844	88.517	88.340	88.295	80.664	80.619	80.597	80.619	1279
2	18893	92.052	90.370	89.934	89.956	80.708	80.642	80.642	80.642	1905
3	27395	93.606	91.663	91.117	91.313	80.774	80.708	80.686	80.708	2445
4	35897	94.855	92.737	92.125	92.365	80.796	80.752	80.730	80.752	2927
5	44399	95.797	93.636	93.003	93.156	80.796	80.752	80.730	80.752	3379
6	52901	96.457	94.186	93.574	93.902	80.796	80.752	80.730	80.752	3841
7	57624	96.893	94.469	93.945	94.186	80.796	80.752	80.730	80.752	4082
						x_1 = 0.68				
1	10391	88.207	87.256	86.813	86.880	80.553	80.487	80.487	80.487	1531
2	18893	90.087	88.944	87.949	88.236	80.597	80.531	80.531	80.531	2288
3	27395	91.641	90.287	89.062	89.416	80.664	80.597	80.597	80.597	2887
4	35897	92.475	91.274	89.527	89.942	80.664	80.619	80.597	80.619	3526
5	44399	93.046	92.086	90.252	90.601	80.664	80.619	80.597	80.619	4084
6	52901	93.356	92.592	90.671	91.107	80.664	80.619	80.597	80.619	4679
7	57624	93.662	92.876	91.063	91.500	80.664	80.642	80.597	80.642	4951
						x_1 = 0.80				
1	10391	88.472	87.344	87.012	87.079	80.708	80.664	80.664	80.664	1528
2	18893	90.545	89.143	88.413	88.701	80.774	80.752	80.708	80.752	2235
3	27395	91.859	90.527	89.372	89.698	80.796	80.752	80.752	80.752	2853
4	35897	92.890	91.929	90.138	90.531	80.796	80.752	80.752	80.752	3384
5	44399	93.287	92.479	90.623	90.972	80.796	80.752	80.752	80.752	4008
6	52901	93.749	93.050	91.129	91.631	80.796	80.774	80.752	80.774	4554
7	57624	94.033	93.443	91.500	92.024	80.796	80.774	80.752	80.774	4812
						x_1 = 0.90				
1	10391	89.291	88.627	88.605	88.804	80.885	80.819	80.752	80.819	1297
2	18893	91.288	90.239	89.977	90.611	80.929	80.841	80.752	80.841	1950
3	27395	92.842	91.444	90.877	91.990	80.951	80.885	80.819	80.885	2513
4	35897	94.069	92.562	91.601	92.890	80.951	80.885	80.841	80.885	3019
5	44399	95.142	93.308	92.260	93.658	80.973	80.885	80.863	80.885	3499
6	52901	95.867	93.924	92.810	94.229	80.973	80.885	80.863	80.885	3976
7	57624	96.325	94.317	93.203	94.622	80.973	80.885	80.863	80.885	4201

Table D.4 Experimental Data of Heat Transfer From A horizontal **Plain Tube** To Pool of Saturated **Acetone(1)-Isopropanol(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					X_1 = 0.10					
1	10391	86.105	85.021	85.641	85.088	76.855	76.810	76.742	76.810	1200
2	18893	88.413	86.710	87.595	85.935	76.900	76.833	76.787	76.833	1830
3	27395	90.462	88.266	89.239	87.226	76.946	76.878	76.833	76.878	2299
4	35897	91.186	89.483	90.422	88.583	76.991	76.900	76.878	76.900	2761
5	44399	92.217	90.645	91.453	89.575	76.991	76.900	76.878	76.900	3159
6	52901	92.985	91.522	92.199	90.343	76.991	76.900	76.878	76.900	3564
					X_1 = 0.20					
1	10391	81.813	80.995	81.149	80.972	72.534	72.443	72.398	72.443	1184
2	18893	84.121	82.550	83.037	82.573	72.579	72.489	72.421	72.489	1786
3	27395	85.810	83.929	84.527	83.996	72.579	72.511	72.466	72.511	2274
4	35897	87.122	85.153	86.348	85.485	72.624	72.557	72.466	72.557	2664
5	44399	88.170	86.333	87.528	86.842	72.624	72.557	72.489	72.557	3028
6	52901	88.880	87.048	88.177	87.535	72.624	72.557	72.489	72.557	3445
					X_1 = 0.30					
1	10391	79.417	78.490	77.947	78.105	69.361	69.315	69.315	69.315	1134
2	18893	81.577	79.940	79.829	79.272	69.406	69.361	69.361	69.361	1752
3	27395	83.664	81.496	81.275	80.345	69.475	69.429	69.406	69.429	2235
4	35897	84.910	82.587	82.299	81.658	69.475	69.452	69.452	69.452	2678
5	44399	86.001	83.789	83.413	83.147	69.475	69.452	69.452	69.452	3035
6	52901	87.093	84.902	84.504	84.371	69.475	69.452	69.452	69.452	3357
					X_1 = 0.40					
1	10391	76.114	75.345	74.938	75.028	66.689	66.667	66.667	66.667	1197
2	18893	78.321	77.077	76.330	76.444	66.735	66.689	66.689	66.689	1827
3	27395	80.235	78.470	77.882	77.882	66.758	66.712	66.712	66.712	2303
4	35897	81.481	79.622	78.980	78.890	66.758	66.712	66.712	66.712	2757
5	44399	82.417	80.736	80.227	80.094	66.758	66.712	66.712	66.712	3139
6	52901	83.354	81.540	81.208	81.009	66.758	66.712	66.712	66.712	3514
					X_1 = 0.50					
1	10391	73.693	72.811	72.336	72.517	64.064	63.995	63.973	63.995	1176
2	18893	75.946	74.543	73.842	73.751	64.110	64.041	63.995	64.041	1804
3	27395	77.882	76.049	75.235	75.076	64.155	64.087	64.087	64.087	2291
4	35897	79.224	77.306	76.401	76.265	64.155	64.110	64.087	64.110	2723
5	44399	80.360	78.360	77.545	77.500	64.155	64.110	64.087	64.110	3099
6	52901	81.230	79.195	78.486	78.531	64.132	64.087	64.064	64.087	3465
					X_1 = 0.60					
1	10391	71.612	70.616	70.390	70.458	62.329	62.260	62.192	62.260	1221

2	18893	74.023	72.416	71.783	71.647	62.374	62.306	62.283	62.306	1861
3	27395	75.732	73.674	73.244	72.814	62.397	62.352	62.306	62.352	2379
4	35897	77.103	74.908	74.410	74.003	62.489	62.443	62.397	62.443	2835
5	44399	78.133	75.894	75.373	75.192	62.489	62.443	62.397	62.443	3240
6	52901	79.018	76.721	76.178	76.268	62.489	62.443	62.397	62.443	3623
					x_1 = 0.80					
1	10391	67.588	66.995	67.542	67.337	59.112	59.089	59.065	59.089	1255
2	18893	69.815	68.905	69.178	68.768	59.159	59.136	59.112	59.136	1884
3	27395	71.773	70.506	70.732	69.918	59.182	59.159	59.136	59.159	2367
4	35897	72.623	71.424	71.650	70.722	59.182	59.159	59.136	59.159	2884
5	44399	73.450	72.432	72.568	71.504	59.182	59.159	59.136	59.159	3331
6	52901	74.255	73.282	73.327	72.309	59.182	59.159	59.136	59.159	3743

Table D.5 Experimental Data of Heat Transfer From A horizontal **Plain Tube** To Pool of Saturated **Acetone(1)-Isopropanol(2)-Water(3)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					x_1 = 0.011, x_2 = 0.074					
1	10391	89.114	88.716	88.472	88.804	80.265	80.199	80.177	80.199	1213
2	18893	91.419	90.632	90.720	91.135	80.310	80.243	80.221	80.243	1762
3	27395	92.907	91.881	92.427	92.689	80.310	80.243	80.221	80.243	2242
4	35897	94.658	93.152	93.654	94.003	80.310	80.243	80.221	80.243	2637
5	44399	95.776	94.204	94.575	95.142	80.310	80.243	80.221	80.243	3027
6	52901	96.653	95.278	95.474	96.151	80.310	80.243	80.221	80.243	3384
					x_1 = 0.020, x_2 = 0.019					
1	10391	91.742	90.978	91.393	91.065	82.588	82.522	82.500	82.522	1186
2	18893	94.039	92.969	93.384	92.881	82.633	82.566	82.544	82.566	1759
3	27395	95.920	94.720	95.113	94.589	82.655	82.611	82.588	82.611	2197
4	35897	97.322	95.968	96.252	95.706	82.655	82.611	82.588	82.611	2621
5	44399	98.439	97.086	97.391	96.693	82.655	82.611	82.588	82.611	3003
6	52901	99.303	97.810	98.487	97.461	82.655	82.611	82.588	82.611	3380
					x_1 = 0.050, x_2 = 0.040					
1	10391	83.981	83.384	83.583	83.893	73.846	73.778	73.710	73.778	1046
2	18893	86.577	85.758	85.935	86.223	73.891	73.824	73.778	73.824	1537
3	27395	88.598	87.248	87.668	87.890	73.891	73.846	73.801	73.846	1956
4	35897	90.117	88.384	89.158	89.180	73.891	73.846	73.801	73.846	2337
5	44399	91.540	89.335	90.492	90.404	73.914	73.846	73.801	73.846	2676
6	52901	92.854	90.256	91.566	91.588	73.914	73.846	73.801	73.846	2986

					X_1 = 0.083, X_2 = 0.036					
1	10391	80.818	80.707	80.176	80.242	70.611	70.543	70.543	70.543	1047
2	18893	83.258	83.037	82.528	82.661	70.679	70.611	70.588	70.611	1542
3	27395	85.456	84.836	84.460	84.792	70.701	70.611	70.588	70.611	1921
4	35897	86.990	86.326	86.105	86.105	70.701	70.611	70.588	70.611	2279
5	44399	88.280	85.404	87.528	87.240	70.747	70.611	70.588	70.611	2695
6	52901	89.448	86.518	88.619	88.155	70.701	70.611	70.588	70.611	3013
					X_1 = 0.124, X_2 = 0.290					
1	10391	82.234	81.968	82.211	82.079	73.394	73.371	73.348	73.371	1187
2	18893	84.896	84.276	84.564	84.188	73.439	73.394	73.371	73.394	1705
3	27395	87.049	85.987	86.275	85.721	73.439	73.416	73.371	73.416	2132
4	35897	88.561	87.255	87.631	86.923	73.439	73.416	73.371	73.416	2531
5	44399	89.728	88.369	88.811	87.816	73.439	73.416	73.371	73.416	2908
6	52901	90.867	89.360	89.972	88.575	73.439	73.416	73.371	73.416	3249
					X_1 = 0.155, X_2 = 0.150					
1	10391	79.711	79.349	79.191	79.395	70.611	70.566	70.543	70.566	1175
2	18893	82.130	81.489	81.444	81.135	70.656	70.611	70.588	70.611	1728
3	27395	84.173	83.155	83.487	82.624	70.679	70.611	70.588	70.611	2151
4	35897	85.552	84.512	84.888	83.804	70.679	70.611	70.588	70.611	2552
5	44399	87.019	85.448	85.714	84.674	70.679	70.611	70.588	70.611	2942
6	52901	88.464	86.141	86.562	85.478	70.679	70.611	70.588	70.611	3298
					X_1 = 0.212, X_2 = 0.182					
1	10391	77.743	76.408	77.879	77.155	68.584	68.539	68.493	68.539	1187
2	18893	80.294	78.118	80.338	79.023	68.653	68.607	68.539	68.607	1743
3	27395	82.336	79.836	82.182	80.345	68.653	68.607	68.539	68.607	2179
4	35897	83.937	81.370	83.738	81.614	68.653	68.607	68.539	68.607	2553
5	44399	85.360	82.705	84.984	82.749	68.653	68.607	68.539	68.607	2893
6	52901	86.495	83.951	86.053	83.774	68.653	68.607	68.539	68.607	3213
					X_1 = 0.266, X_2 = 0.282					
1	10391	78.082	77.019	77.698	77.336	68.539	68.493	68.470	68.493	1150
2	18893	80.537	79.520	80.161	79.829	68.607	68.539	68.516	68.539	1648
3	27395	82.580	81.562	82.005	81.407	68.607	68.539	68.516	68.539	2054
4	35897	84.269	82.919	83.494	82.609	68.607	68.539	68.516	68.539	2430
5	44399	85.692	84.010	84.962	83.413	68.607	68.539	68.516	68.539	2780
6	52901	86.960	85.013	85.787	84.084	68.607	68.539	68.516	68.539	3128
					X_1 = 0.294, X_2 = 0.484					
1	10391	78.015	77.223	77.653	77.313	69.132	69.087	69.087	69.087	1229
2	18893	80.471	79.362	79.785	78.910	69.178	69.132	69.110	69.132	1800
3	27395	82.226	80.898	80.987	79.991	69.201	69.155	69.155	69.155	2310
4	35897	83.671	82.277	82.410	81.105	69.201	69.155	69.155	69.155	2720
5	44399	84.541	83.236	83.214	82.085	69.201	69.155	69.155	69.155	3148
6	52901	85.212	83.929	83.730	82.889	69.201	69.155	69.155	69.155	3581

					X_1 = 0.389, X_2 = 0.212					
1	10391	74.055	73.173	73.037	72.924	64.315	64.247	64.224	64.247	1150
2	18893	76.421	75.358	75.199	74.973	64.384	64.338	64.292	64.338	1694
3	27395	78.515	77.248	76.864	76.705	64.384	64.338	64.292	64.338	2108
4	35897	79.954	78.505	77.985	78.053	64.384	64.338	64.292	64.338	2513
5	44399	81.223	79.453	79.055	79.364	64.384	64.338	64.292	64.338	2876
6	52901	82.469	80.433	80.124	80.234	64.384	64.338	64.292	64.338	3211
					X_1 = 0.524, X_2 = 0.267					
1	10391	71.046	70.232	70.209	70.141	62.397	62.329	62.306	62.329	1288
2	18893	73.548	72.394	72.032	71.647	62.443	62.397	62.374	62.397	1889
3	27395	75.506	74.013	73.651	72.791	62.443	62.397	62.374	62.397	2364
4	35897	77.012	75.609	74.908	74.026	62.443	62.397	62.374	62.397	2764
5	44399	77.885	76.663	75.826	75.034	62.443	62.397	62.374	62.397	3183
6	52901	78.752	77.467	76.789	75.861	62.443	62.397	62.374	62.397	3571
					X_1 = 0.699, X_2 = 0.105					
1	10391	68.844	68.022	67.999	67.885	59.813	59.766	59.720	59.766	1234
2	18893	71.398	70.222	69.883	70.154	59.907	59.860	59.790	59.860	1789
3	27395	73.379	72.067	71.547	71.999	59.930	59.883	59.813	59.883	2214
4	35897	74.433	72.985	72.442	72.894	59.930	59.883	59.813	59.883	2697
5	44399	75.690	73.835	73.450	73.857	59.930	59.883	59.813	59.883	3098
6	52901	76.404	74.391	74.164	74.436	59.930	59.883	59.813	59.883	3533

Table D.6 Experimental Data of Heat Transfer From A horizontal **748 fpm Tube** To Pool of Saturated **Pure Liquids**

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					Acetone					
1	10391	60.568	59.605	60.294	59.975	56.262	56.238	56.215	56.238	2684
2	18893	61.562	60.215	61.174	60.695	56.308	56.262	56.238	56.262	4068
3	27395	62.396	60.684	61.712	61.278	56.332	56.285	56.238	56.285	5236
4	35897	63.025	61.062	62.021	61.792	56.332	56.285	56.262	56.285	6315
5	44399	63.449	61.371	62.353	62.079	56.332	56.285	56.262	56.285	7373
6	52901	63.803	61.680	62.639	62.456	56.332	56.285	56.262	56.285	8326
					Isopropanol					
1	10391	86.119	85.743	85.964	85.787	82.323	82.323	82.301	82.323	2898
2	18893	87.210	86.436	86.701	86.524	82.367	82.323	82.301	82.323	4304
3	27395	87.814	86.929	87.239	87.128	82.367	82.323	82.323	82.323	5542
4	35897	88.219	87.356	87.732	87.666	82.367	82.367	82.323	82.367	6663
5	44399	88.513	87.872	88.204	88.226	82.367	82.367	82.345	82.367	7600

6	52901	88.763	88.277	88.564	88.763	82.367	82.367	82.345	82.367	8491
					Distilled Water					
1	14170	103.450	102.313	102.914	102.828	100.000	99.913	99.869	99.913	4799
2	18893	103.849	102.776	102.991	103.162	100.129	100.086	100.043	100.086	6078
3	27395	104.365	103.121	103.464	103.765	100.172	100.129	100.107	100.129	7729
4	35897	104.774	103.272	103.809	104.152	100.172	100.129	100.107	100.129	9281
5	44399	105.055	103.295	103.960	104.390	100.172	100.129	100.107	100.129	10987
6	52901	105.357	103.382	104.305	104.606	100.172	100.129	100.107	100.129	12365

Table D.7 Experimental Data of Heat Transfer From A horizontal **748 fpm Tube** To Pool of Saturated **Acetone(1)-Water(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					X_1 = 0.05					
1	10391	81.186	80.522	80.256	80.411	75.905	75.837	75.814	75.837	2190
2	18893	82.564	81.303	81.480	81.192	75.995	75.905	75.882	75.905	3307
3	27395	83.456	81.730	82.460	81.664	76.041	75.973	75.928	75.973	4315
4	35897	84.192	82.290	83.020	82.290	76.041	75.995	75.950	75.995	5163
5	44399	84.841	82.739	83.469	82.717	76.063	75.995	75.950	75.995	5967
6	52901	85.268	83.122	83.874	83.077	76.063	75.995	75.950	75.995	6753
					X_1 = 0.10					
1	10391	75.697	74.543	73.457	75.312	69.247	69.224	69.178	69.224	1878
2	18893	76.545	75.572	76.454	76.137	69.269	69.247	69.178	69.247	2722
3	27395	77.279	76.329	77.166	76.917	69.315	69.247	69.224	69.247	3574
4	35897	78.082	76.905	78.014	77.878	69.315	69.247	69.224	69.247	4242
5	44399	78.794	77.550	78.636	78.568	69.315	69.247	69.224	69.247	4864
6	52901	79.648	78.397	77.017	79.184	69.315	69.247	69.224	69.247	5686
					X_1 = 0.15					
1	10391	71.738	70.991	71.308	71.081	65.434	65.365	65.320	65.365	1759
2	18893	72.947	71.952	72.404	72.472	65.479	65.434	65.342	65.434	2691
3	27395	73.931	72.981	73.297	73.637	65.525	65.434	65.411	65.434	3420
4	35897	74.801	73.580	74.145	74.507	65.525	65.457	65.411	65.457	4081
5	44399	75.581	73.930	74.902	75.423	65.525	65.457	65.434	65.457	4678
6	52901	76.361	74.325	75.592	76.180	65.525	65.457	65.434	65.457	5214
					X_1 = 0.25					
1	10391	68.856	68.057	68.627	68.856	62.900	62.877	62.854	62.877	1816
2	18893	70.413	69.074	69.961	70.119	62.945	62.922	62.877	62.922	2709
3	27395	71.533	69.836	70.967	71.374	62.945	62.922	62.877	62.922	3420

4	35897	72.222	70.367	72.041	72.199	62.945	62.922	62.877	62.922	4083
5	44399	72.979	70.921	72.731	72.821	62.968	62.922	62.877	62.922	4703
6	52901	73.624	71.406	73.216	73.284	62.968	62.922	62.900	62.922	5314
						X_1 = 0.40				
1	10391	66.048	65.294	65.431	65.111	60.845	60.799	60.753	60.799	2224
2	18893	67.156	66.106	66.380	65.854	60.868	60.845	60.799	60.845	3414
3	27395	68.081	66.780	67.214	66.438	60.890	60.845	60.799	60.845	4360
4	35897	69.121	67.454	67.888	66.975	60.890	60.845	60.799	60.845	5117
5	44399	69.767	67.878	68.220	67.375	60.890	60.845	60.822	60.845	5952
6	52901	70.253	68.233	68.507	67.730	60.890	60.845	60.822	60.845	6756
						X_1 = 0.60				
1	10391	64.609	63.582	64.381	63.810	59.112	59.065	59.019	59.065	2066
2	18893	65.900	64.462	65.580	64.759	59.159	59.112	59.112	59.112	3122
3	27395	66.963	65.091	66.460	65.319	59.159	59.112	59.112	59.112	4008
4	35897	67.865	65.765	66.815	65.719	59.159	59.136	59.112	59.136	4847
5	44399	68.563	66.120	67.102	65.914	59.159	59.136	59.112	59.136	5700
6	52901	69.053	66.360	67.296	66.109	59.159	59.136	59.112	59.136	6556
						X_1 = 0.80				
1	10391	62.669	62.006	62.280	61.984	57.827	57.780	57.734	57.780	2333
2	18893	63.663	62.818	63.115	62.841	57.897	57.827	57.804	57.827	3585
3	27395	64.520	63.492	63.880	63.492	57.921	57.827	57.804	57.827	4565
4	35897	65.240	63.961	64.464	63.870	57.921	57.827	57.804	57.827	5490
5	44399	65.800	64.225	64.933	64.248	57.921	57.850	57.804	57.850	6393
6	52901	66.246	64.488	65.310	64.557	57.921	57.850	57.804	57.850	7253

Table D.8 Experimental Data of Heat Transfer From A horizontal **748 fpm Tube** To Pool of Saturated **Isopropanol(1)-Water(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
						X_1 = 0.05				
1	10391	92.063	91.932	91.779	91.932	86.150	86.084	86.084	86.084	1784
2	18893	93.006	93.202	92.394	93.071	86.173	86.128	86.106	86.128	2785
3	27395	94.080	94.429	92.988	93.993	86.173	86.128	86.106	86.128	3540
4	35897	94.892	95.220	93.582	95.067	86.173	86.150	86.106	86.150	4201
5	44399	95.312	95.661	94.132	95.617	86.173	86.150	86.106	86.150	4914
6	52901	95.731	96.364	94.574	96.408	86.173	86.150	86.106	86.150	5497
						X_1 = 0.10				
1	10391	89.327	89.283	88.619	89.283	83.717	83.650	83.650	83.650	1903

2	18893	91.084	90.910	89.799	90.648	83.739	83.695	83.673	83.695	2734
3	27395	92.180	91.460	90.980	91.809	83.783	83.739	83.695	83.739	3482
4	35897	93.298	92.054	91.661	92.207	83.783	83.739	83.695	83.739	4191
5	44399	94.023	92.582	92.189	92.932	83.783	83.739	83.695	83.739	4830
6	52901	94.705	93.373	92.827	93.591	83.783	83.761	83.695	83.761	5358
					X_1 = 0.20					
1	10391	87.712	87.491	87.115	87.867	81.925	81.858	81.814	81.858	1829
2	18893	89.179	88.803	87.918	89.135	81.991	81.947	81.903	81.947	2774
3	27395	90.128	89.691	88.832	90.412	81.991	81.947	81.903	81.947	3504
4	35897	90.984	90.416	89.436	91.028	81.991	81.947	81.925	81.947	4216
5	44399	91.905	90.945	90.115	91.731	81.991	81.947	81.925	81.947	4815
6	52901	92.783	91.473	90.709	92.259	81.991	81.947	81.925	81.947	5369
					X_1 = 0.40					
1	10391	87.248	86.208	86.783	87.048	81.217	81.150	81.106	81.150	1834
2	18893	88.449	87.033	87.630	87.984	81.261	81.195	81.173	81.195	2876
3	27395	89.757	87.460	88.345	88.699	81.261	81.195	81.173	81.195	3722
4	35897	90.722	88.064	88.971	89.104	81.261	81.195	81.173	81.195	4482
5	44399	91.687	88.624	89.722	89.394	81.261	81.195	81.173	81.195	5132
6	52901	92.456	89.073	90.360	89.639	81.261	81.195	81.173	81.195	5765
					X_1 = 0.60					
1	10391	85.765	84.571	84.482	84.460	80.597	80.553	80.531	80.553	2439
2	18893	86.922	85.197	85.772	85.108	80.619	80.575	80.575	80.575	3659
3	27395	87.637	85.712	86.265	85.580	80.642	80.597	80.597	80.597	4814
4	35897	87.909	86.073	86.538	85.984	80.642	80.597	80.597	80.597	5965
5	44399	88.115	86.367	86.765	86.389	80.642	80.597	80.597	80.597	7046
6	52901	88.365	86.662	86.905	86.816	80.642	80.597	80.597	80.597	8041
					X_1 = 0.68					
1	10391	85.102	84.084	84.327	83.907	80.597	80.575	80.575	80.575	2753
2	18893	86.148	84.732	85.153	84.622	80.664	80.619	80.619	80.619	4168
3	27395	86.686	85.181	85.602	85.203	80.664	80.619	80.619	80.619	5438
4	35897	86.958	85.608	85.940	85.653	80.664	80.619	80.619	80.619	6636
5	44399	87.119	85.991	86.235	86.124	80.664	80.619	80.619	80.619	7739
6	52901	87.303	86.330	86.507	86.529	80.664	80.619	80.619	80.619	8763
					X_1 = 0.80					
1	10391	84.703	83.840	84.084	83.730	80.442	80.398	80.398	80.398	2824
2	18893	85.529	84.511	84.865	84.445	80.487	80.420	80.420	80.420	4294
3	27395	86.133	84.916	85.358	85.026	80.487	80.420	80.420	80.420	5567
4	35897	86.471	85.409	85.807	85.675	80.487	80.420	80.420	80.420	6643
5	44399	86.788	85.814	86.190	86.058	80.487	80.420	80.420	80.420	7687
6	52901	87.060	86.241	86.529	86.396	80.487	80.420	80.420	80.420	8645
					X_1 = 0.90					
1	10391	85.146	84.438	84.548	84.217	80.929	80.885	80.863	80.885	2811

2	18893	86.082	85.086	85.374	84.976	80.973	80.929	80.929	80.929	4256
3	27395	86.730	85.624	85.978	85.668	80.973	80.929	80.929	80.929	5414
4	35897	87.046	86.095	86.427	86.338	80.973	80.929	80.929	80.929	6484
5	44399	87.363	86.566	86.832	86.942	80.973	80.929	80.929	80.929	7418
6	52901	87.701	86.993	87.237	87.569	80.973	80.929	80.929	80.929	8221

Table D.9 Experimental Data of Heat Transfer From A horizontal **748 fpm Tube** To Pool of Saturated **Acetone(1)-Isopropanol(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
						X_1 = 0.10				
1	10391	82.358	81.230	81.827	81.495	76.697	76.652	76.652	76.652	2052
2	18893	83.537	82.011	82.675	82.254	76.765	76.697	76.697	76.697	3199
3	27395	84.385	82.659	83.301	82.969	76.765	76.697	76.697	76.697	4142
4	35897	85.011	83.285	83.861	83.639	76.765	76.697	76.697	76.697	4961
5	44399	85.350	83.735	84.288	84.265	76.765	76.697	76.697	76.697	5769
6	52901	85.622	84.095	84.670	84.847	76.765	76.697	76.697	76.697	6535
						X_1 = 0.20				
1	10391	78.955	77.869	78.344	78.095	72.647	72.579	72.534	72.579	1813
2	18893	80.285	78.717	79.282	78.875	72.692	72.624	72.602	72.624	2839
3	27395	81.155	79.429	79.916	79.584	72.692	72.624	72.624	72.624	3712
4	35897	81.560	79.900	80.387	80.210	72.692	72.624	72.624	72.624	4560
5	44399	81.987	80.438	80.836	80.881	72.692	72.624	72.624	72.624	5289
6	52901	82.392	80.909	81.308	81.485	72.692	72.624	72.624	72.624	5956
						X_1 = 0.30				
1	10391	75.765	74.724	75.109	74.905	69.292	69.247	69.224	69.247	1769
2	18893	77.042	75.459	76.024	75.685	69.338	69.315	69.269	69.315	2802
3	27395	77.822	76.058	76.668	76.352	69.338	69.315	69.269	69.315	3694
4	35897	78.399	76.634	77.222	76.996	69.361	69.315	69.292	69.315	4492
5	44399	78.930	77.188	77.798	77.753	69.361	69.315	69.292	69.315	5165
6	52901	79.383	77.651	78.262	78.375	69.361	69.315	69.292	69.315	5815
						X_1 = 0.40				
1	10391	73.502	72.484	72.824	72.688	66.895	66.849	66.826	66.849	1726
2	18893	74.735	73.241	73.694	73.445	66.941	66.872	66.872	66.872	2742
3	27395	75.537	73.863	74.383	74.157	66.941	66.872	66.872	66.872	3607
4	35897	76.068	74.394	74.892	74.779	66.941	66.895	66.872	66.895	4414
5	44399	76.554	74.948	75.423	75.445	66.941	66.895	66.872	66.895	5108
6	52901	77.040	75.366	75.977	76.067	66.941	66.895	66.872	66.895	5743
						X_1 = 0.50				

1	10391	70.448	69.381	69.905	69.905	64.224	64.155	64.132	64.155	1809
2	18893	71.567	70.051	70.640	70.549	64.269	64.201	64.178	64.201	2911
3	27395	72.347	70.673	71.239	71.193	64.269	64.201	64.178	64.201	3831
4	35897	72.969	71.317	71.837	71.883	64.269	64.224	64.178	64.224	4615
5	44399	73.432	71.871	72.414	72.527	64.269	64.224	64.178	64.224	5325
6	52901	73.850	72.402	73.058	73.126	64.269	64.224	64.178	64.224	5954
					X_1 = 0.60					
1	10391	68.217	67.235	67.646	67.623	62.283	62.237	62.215	62.237	1911
2	18893	69.348	67.772	68.320	68.297	62.329	62.283	62.260	62.283	3074
3	27395	70.175	68.447	68.926	68.949	62.329	62.283	62.260	62.283	4008
4	35897	70.684	69.053	69.462	69.575	62.329	62.283	62.260	62.283	4848
5	44399	71.124	69.699	69.925	70.219	62.329	62.283	62.260	62.283	5583
6	52901	71.452	70.207	70.275	70.615	62.329	62.283	62.260	62.283	6337
					X_1 = 0.80					
1	10391	64.974	64.084	64.541	64.427	59.159	59.136	59.136	59.136	1937
2	18893	66.083	64.507	65.238	65.124	59.206	59.182	59.159	59.182	3120
3	27395	66.940	64.954	65.753	65.547	59.206	59.182	59.159	59.182	4141
4	35897	67.477	65.468	66.130	65.993	59.206	59.182	59.182	59.182	5071
5	44399	67.901	65.937	66.508	66.508	59.206	59.182	59.206	59.182	5905
6	52901	68.233	66.338	66.886	67.000	59.206	59.206	59.206	59.206	6689

Table D.10 Experimental Data of Heat Transfer From A horizontal **748 fpm Tube** To Pool of Saturated **Acetone(1)-Isopropanol(2)-Water(3)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					X_1 = 0.011, X_2 = 0.074					
1	10391	85.898	84.659	85.234	84.571	80.376	80.332	80.288	80.332	2184
2	18893	87.453	86.126	86.922	86.082	80.420	80.354	80.332	80.354	3008
3	27395	88.257	86.664	87.593	86.641	80.420	80.376	80.332	80.376	3963
4	35897	89.015	87.179	88.130	87.113	80.442	80.376	80.332	80.376	4800
5	44399	89.525	87.650	88.535	87.518	80.442	80.376	80.332	80.376	5602
6	52901	89.879	88.011	88.785	87.768	80.442	80.376	80.332	80.376	6428
					X_1 = 0.050, X_2 = 0.040					
1	10391	79.407	78.457	79.814	78.502	73.937	73.869	73.846	73.869	2012
2	18893	80.728	79.554	80.462	79.531	73.982	73.937	73.891	73.937	3081
3	27395	81.708	80.403	81.531	80.557	74.005	73.937	73.914	73.937	3858
4	35897	82.489	80.896	82.135	81.007	74.005	73.937	73.914	73.937	4672
5	44399	83.027	81.345	82.673	81.434	74.005	73.937	73.914	73.937	5433

6	52901	83.476	81.750	83.122	81.816	74.005	73.937	73.914	73.937	6156
					$x_1 = 0.083,\ x_2 = 0.036$					
1	10391	76.692	75.652	76.489	75.855	70.701	70.656	70.633	70.656	1886
2	18893	77.970	76.816	77.789	76.974	70.747	70.679	70.679	70.679	2823
3	27395	78.953	77.709	78.659	78.049	70.769	70.724	70.701	70.724	3598
4	35897	79.657	78.195	79.190	78.534	70.769	70.724	70.701	70.724	4397
5	44399	80.283	78.749	79.664	79.111	70.769	70.724	70.701	70.724	5090
6	52901	80.865	79.184	80.024	79.427	70.769	70.724	70.701	70.724	5784
					$x_1 = 0.124,\ x_2 = 0.290$					
1	10391	79.294	78.367	79.113	78.434	73.507	73.439	73.416	73.439	1942
2	18893	80.595	79.395	80.506	79.486	73.552	73.462	73.416	73.462	2896
3	27395	81.774	80.203	81.597	80.248	73.552	73.484	73.462	73.484	3672
4	35897	82.776	80.984	82.378	80.984	73.552	73.484	73.462	73.484	4333
5	44399	83.602	81.588	83.137	81.544	73.552	73.507	73.462	73.507	4955
6	52901	83.874	82.126	83.675	82.104	73.552	73.507	73.462	73.507	5605
					$x_1 = 0.155,\ x_2 = 0.150$					
1	10391	76.014	75.176	75.833	75.357	70.724	70.679	70.679	70.679	2118
2	18893	77.359	76.092	76.477	76.250	70.792	70.747	70.724	70.747	3262
3	27395	78.207	76.849	77.709	77.144	70.792	70.747	70.724	70.747	4074
4	35897	78.964	77.403	78.263	77.516	70.792	70.747	70.724	70.747	4928
5	44399	79.708	77.979	78.862	78.025	70.792	70.747	70.747	70.747	5630
6	52901	80.290	78.465	79.095	78.488	70.792	70.747	70.747	70.747	6353
					$x_1 = 0.212,\ x_2 = 0.182$					
1	10391	74.023	73.163	73.683	73.389	68.470	68.425	68.402	68.425	2024
2	18893	75.255	74.214	75.029	74.576	68.516	68.470	68.470	68.470	3005
3	27395	76.533	75.039	75.628	75.447	68.516	68.470	68.470	68.470	3815
4	35897	77.516	75.797	76.679	76.272	68.516	68.470	68.470	68.470	4440
5	44399	78.364	76.441	77.482	76.984	68.516	68.470	68.447	68.470	5022
6	52901	78.962	76.995	78.148	77.651	68.516	68.470	68.447	68.470	5590
					$x_1 = 0.266,\ x_2 = 0.282$					
1	10391	74.633	73.570	74.317	73.955	68.653	68.607	68.584	68.607	1887
2	18893	76.115	74.599	75.730	75.097	68.676	68.653	68.630	68.653	2806
3	27395	77.460	75.379	76.872	75.967	68.699	68.653	68.630	68.653	3530
4	35897	78.512	76.046	77.675	76.656	68.630	68.653	68.653	68.653	4186
5	44399	79.354	76.803	78.409	77.459	68.699	68.653	68.653	68.653	4753
6	52901	79.980	77.492	79.095	78.171	68.721	68.653	68.653	68.653	5282
					$x_1 = 0.294,\ x_2 = 0.484$					
1	10391	74.430	73.525	74.136	73.253	69.269	69.224	69.201	69.224	2256
2	18893	75.730	74.418	75.391	74.237	69.315	69.247	69.224	69.247	3323
3	27395	76.691	75.130	76.329	74.949	69.315	69.269	69.224	69.269	4211
4	35897	77.358	75.774	77.018	75.525	69.315	69.269	69.224	69.269	5021

5	44399	77.889	76.283	77.527	76.034	69.315	69.269	69.224	69.269	5793
6	52901	78.307	76.655	77.877	76.452	69.315	69.269	69.224	69.269	6569
					X_1 = 0.389, X_2 = 0.212					
1	10391	70.176	68.833	69.928	69.404	64.589	64.543	64.498	64.543	2061
2	18893	71.567	70.391	71.273	70.617	64.635	64.589	64.543	64.589	2965
3	27395	72.709	71.035	72.370	71.284	64.635	64.589	64.543	64.589	3773
4	35897	73.670	71.657	73.059	72.018	64.635	64.589	64.543	64.589	4480
5	44399	74.518	72.233	73.545	72.640	64.635	64.589	64.543	64.589	5136
6	52901	75.185	72.832	74.234	73.216	64.635	64.589	64.543	64.589	5702
					X_1 = 0.524, X_2 = 0.267					
1	10391	75.448	74.498	75.131	74.294	70.204	70.136	70.113	70.136	2213
2	18893	76.861	75.527	76.612	75.481	70.249	70.181	70.136	70.181	3184
3	27395	77.732	76.058	77.370	75.944	70.249	70.181	70.158	70.181	4161
4	35897	78.534	76.702	78.014	76.498	70.249	70.204	70.158	70.204	4963
5	44399	79.177	77.165	78.703	76.961	70.249	70.204	70.158	70.204	5694
6	52901	79.825	77.809	79.117	77.538	70.249	70.204	70.158	70.204	6321

Table D.11 Experimental Data of Heat Transfer From A horizontal **1024 fpm Tube** To Pool of Saturated **Pure Liquids**

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					Acetone					
1	10391	60.111	59.184	59.605	59.278	56.262	56.238	56.238	56.238	3148
2	18893	60.991	59.827	60.215	59.850	56.285	56.262	56.262	56.262	4779
3	27395	61.620	60.387	60.684	60.250	56.285	56.262	56.262	56.262	6131
4	35897	61.998	60.742	60.948	60.559	56.238	56.215	56.215	56.215	7415
5	44399	62.147	60.891	61.188	60.754	56.238	56.215	56.215	56.215	8836
6	52901	62.456	61.086	61.429	60.927	56.238	56.215	56.215	56.215	10069
					Isopropanol					
1	10391	85.743	84.770	85.256	84.637	82.478	82.456	82.434	82.456	3927
2	18893	86.436	85.241	85.816	84.998	82.522	82.500	82.478	82.500	6050
3	27395	86.907	85.668	86.310	85.403	82.522	82.500	82.478	82.500	7670
4	35897	87.223	86.073	86.803	85.741	82.522	82.500	82.478	82.500	9065
5	44399	87.518	86.456	87.208	86.080	82.522	82.500	82.478	82.500	10289
6	52901	87.723	86.684	87.480	86.285	82.522	82.500	82.478	82.500	11644
					Distilled Water					
1	14170	102.956	101.733	102.506	101.926	100.000	99.978	99.956	99.978	6155
2	18893	103.162	101.939	102.840	102.089	100.086	100.064	100.021	100.064	7715

3	27395	103.614	102.069	103.185	102.262	100.107	100.064	100.043	100.064	10098
4	35897	103.895	102.285	103.294	102.478	100.107	100.064	100.043	100.064	12301
5	44399	104.218	102.480	103.488	102.608	100.107	100.064	100.043	100.064	14190
6	52901	104.391	102.696	103.704	102.739	100.107	100.064	100.043	100.064	15969

Table D.12 Experimental Data of Heat Transfer From A horizontal **1024 fpm Tube** To Pool of Saturated **Acetone(1)-Water(2)** Liquid Mixture

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
						x_1 = 0.05				
1	10391	80.832	79.633	80.057	79.656	75.973	75.928	75.905	75.928	2527
2	18893	81.834	80.484	81.037	80.462	76.041	75.995	75.950	75.995	3810
3	27395	82.571	81.177	82.018	81.088	76.063	76.018	75.995	76.018	4815
4	35897	82.953	81.626	82.400	81.715	76.063	76.018	75.995	76.018	5837
5	44399	83.314	81.876	82.938	82.009	76.063	76.018	75.995	76.018	6820
6	52901	83.763	82.193	83.210	82.370	76.063	76.018	75.995	76.018	7711
						x_1 = 0.10				
1	10391	73.887	73.344	73.728	73.570	69.155	69.132	69.110	69.132	2309
2	18893	74.803	74.146	74.486	74.441	69.201	69.155	69.132	69.155	3559
3	27395	75.582	74.813	75.039	75.107	69.247	69.201	69.201	69.201	4625
4	35897	76.023	75.209	75.435	75.503	69.247	69.201	69.201	69.201	5671
5	44399	76.441	75.581	75.875	75.966	69.247	69.201	69.201	69.201	6574
6	52901	76.678	75.863	76.044	76.203	69.247	69.201	69.201	69.201	7574
						x_1 = 0.15				
1	10391	70.131	69.950	69.905	70.312	65.388	65.365	65.342	65.365	2207
2	18893	71.160	70.911	70.821	71.409	65.411	65.388	65.365	65.388	3322
3	27395	71.895	71.352	71.306	72.166	65.434	65.411	65.411	65.411	4374
4	35897	72.675	72.018	71.883	73.104	65.434	65.411	65.411	65.411	5126
5	44399	73.296	72.323	72.301	73.636	65.434	65.411	65.411	65.411	5942
6	52901	73.737	72.628	72.560	73.963	65.434	65.411	65.411	65.411	6777
						x_1 = 0.25				
1	10391	67.554	66.732	67.189	66.984	62.603	62.580	62.580	62.580	2294
2	18893	68.480	67.247	67.932	67.635	62.626	62.603	62.603	62.603	3623
3	27395	69.154	67.967	68.447	68.287	62.648	62.626	62.603	62.626	4692
4	35897	69.869	68.642	68.847	68.801	62.648	62.626	62.603	62.626	5596
5	44399	70.536	68.996	69.201	69.246	62.648	62.626	62.603	62.626	6463

6	52901	70.999	69.325	69.687	69.732	62.648	62.626	62.603	62.626	7237
$x_1 = 0.40$										
1	10391	65.591	64.997	65.751	65.203	60.890	60.868	60.868	60.868	2303
2	18893	66.083	65.740	66.699	65.923	60.959	60.913	60.890	60.913	3639
3	27395	66.803	66.506	67.442	66.620	60.959	60.936	60.913	60.936	4638
4	35897	67.432	66.998	68.139	67.158	60.959	60.936	60.913	60.936	5526
5	44399	67.786	67.512	68.768	67.741	60.959	60.936	60.913	60.936	6328
6	52901	68.370	67.822	69.212	68.050	60.959	60.936	60.913	60.936	7123
$x_1 = 0.60$										
1	10391	63.536	63.125	63.217	63.194	59.206	59.182	59.182	59.182	2547
2	18893	64.622	63.914	64.188	64.256	59.252	59.229	59.206	59.229	3767
3	27395	65.387	64.451	64.680	64.565	59.252	59.229	59.229	59.229	4948
4	35897	65.833	64.852	65.034	64.989	59.252	59.229	59.229	59.229	6041
5	44399	66.576	65.047	65.480	65.389	59.252	59.229	59.229	59.229	6950
6	52901	66.863	65.219	65.721	65.584	59.252	59.229	59.229	59.229	8001
$x_1 = 0.80$										
1	10391	61.801	61.322	61.344	61.367	57.640	57.593	57.593	57.593	2697
2	18893	62.681	61.836	62.179	61.973	57.664	57.617	57.617	57.617	4163
3	27395	63.470	62.465	62.785	62.625	57.710	57.664	57.664	57.664	5308
4	35897	63.961	62.797	63.071	62.957	57.710	57.664	57.664	57.664	6502
5	44399	64.316	63.038	63.220	63.129	57.710	57.664	57.664	57.664	7721
6	52901	64.648	63.324	63.575	63.392	57.710	57.664	57.664	57.664	8730

Table D.13 Experimental Data of Heat Transfer From A horizontal **1024 fpm Tube** To Pool of Saturated **Isopropanol(1)-Water(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
$x_1 = 0.05$										
1	10391	90.447	89.526	90.032	89.703	86.062	86.018	86.018	86.018	2665
2	18893	91.150	90.233	90.670	90.517	86.084	86.062	86.018	86.062	4120
3	27395	91.722	90.718	91.329	91.198	86.128	86.106	86.084	86.106	5334
4	35897	92.229	91.093	91.683	91.879	86.128	86.106	86.106	86.106	6400
5	44399	92.691	91.447	92.080	92.102	86.128	86.106	86.106	86.106	7439
6	52901	93.045	91.822	92.434	92.499	86.128	86.106	86.106	86.106	8346
$x_1 = 0.10$										

1	10391	87.867	87.248	87.624	87.712	83.673	83.628	83.606	83.628	2612
2	18893	88.626	87.852	88.383	88.493	83.695	83.650	83.650	83.650	4040
3	27395	89.252	88.279	88.964	89.208	83.717	83.673	83.650	83.673	5220
4	35897	89.936	88.861	89.237	89.739	83.717	83.673	83.650	83.673	6227
5	44399	90.399	89.177	89.656	89.962	83.717	83.673	83.650	83.673	7254
6	52901	90.753	89.574	89.923	90.229	83.717	83.673	83.650	83.673	8213
					x_1 = 0.20					
1	10391	86.230	85.655	86.119	86.407	82.124	82.102	82.080	82.102	2597
2	18893	86.922	86.259	86.679	87.254	82.190	82.124	82.102	82.124	4069
3	27395	87.593	86.708	87.172	87.880	82.212	82.146	82.146	82.146	5293
4	35897	88.086	87.046	87.577	88.352	82.212	82.146	82.146	82.146	6407
5	44399	88.491	87.473	87.916	88.690	82.212	82.146	82.146	82.146	7424
6	52901	88.785	87.701	88.277	89.073	82.212	82.146	82.146	82.146	8402
					x_1 = 0.40					
1	10391	85.876	84.991	85.411	85.168	81.438	81.416	81.416	81.416	2637
2	18893	86.834	85.506	86.060	85.750	81.482	81.460	81.460	81.460	4133
3	27395	87.593	86.044	86.509	86.398	81.504	81.482	81.482	81.482	5321
4	35897	88.352	86.471	86.914	86.869	81.504	81.482	81.482	81.482	6338
5	44399	88.823	86.788	87.252	87.119	81.504	81.482	81.482	81.482	7390
6	52901	89.443	86.993	87.635	87.148	81.504	81.482	81.482	81.482	8374
					x_1 = 0.60					
1	10391	84.327	83.531	83.929	83.487	80.597	80.531	80.509	80.531	3172
2	18893	85.086	84.135	84.533	84.046	80.619	80.553	80.531	80.553	4862
3	27395	85.624	84.473	84.960	84.385	80.619	80.575	80.553	80.575	6401
4	35897	86.184	84.900	85.409	84.723	80.619	80.575	80.553	80.575	7600
5	44399	86.478	85.217	85.748	85.062	80.619	80.575	80.553	80.575	8800
6	52901	86.684	85.489	85.998	85.356	80.619	80.575	80.553	80.575	9979
					x_1 = 0.68					
1	10391	83.796	83.420	83.730	83.332	80.575	80.531	80.531	80.531	3432
2	18893	84.533	83.891	84.201	83.847	80.619	80.597	80.597	80.597	5374
3	27395	85.004	84.274	84.650	84.097	80.619	80.597	80.597	80.597	7018
4	35897	85.365	84.569	85.055	84.458	80.619	80.597	80.597	80.597	8429
5	44399	85.527	84.796	85.416	84.796	80.619	80.597	80.597	80.597	9799
6	52901	85.777	85.047	85.644	85.024	80.619	80.597	80.597	80.597	11090
					x_1 = 0.80					
1	10391	84.084	83.531	83.885	83.464	80.730	80.708	80.686	80.708	3426
2	18893	84.511	83.847	84.356	83.781	80.752	80.708	80.708	80.708	5549
3	27395	84.982	84.318	84.938	84.186	80.774	80.730	80.730	80.730	7088
4	35897	85.321	84.679	85.321	84.569	80.796	80.752	80.752	80.752	8529

5	44399	85.549	84.951	85.615	84.841	80.796	80.752	80.752	80.752	9920
6	52901	85.843	85.312	85.954	85.157	80.796	80.752	80.752	80.752	11014
					X_1 = 0.90					
1	10391	84.416	83.774	84.062	83.641	80.819	80.774	80.774	80.774	3260
2	18893	85.064	84.223	84.599	84.135	81.062	80.819	80.819	80.819	5210
3	27395	85.624	84.672	85.093	84.584	80.863	80.841	80.841	80.841	6606
4	35897	85.962	84.945	85.276	84.967	80.863	80.841	80.841	80.841	8082
5	44399	86.257	85.261	85.770	85.305	80.863	80.841	80.841	80.841	9246
6	52901	86.485	85.467	86.020	85.555	80.863	80.841	80.841	80.841	10506

Table D.14 Experimental Data of Heat Transfer From A horizontal **1024 fpm Tube** To Pool of Saturated **Acetone(1)-Isopropanol(2)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					X_1 = 0.10					
1	10391	80.566	79.902	80.301	79.947	76.787	76.765	76.742	76.765	3043
2	18893	81.237	80.462	80.883	80.529	76.787	76.765	76.742	76.765	4708
3	27395	81.752	80.867	81.332	80.978	76.810	76.787	76.787	76.787	6171
4	35897	82.179	81.139	81.670	81.338	76.810	76.787	76.787	76.787	7496
5	44399	82.518	81.411	81.987	81.655	76.810	76.787	76.787	76.787	8706
6	52901	82.878	81.684	82.480	81.949	76.810	76.787	76.787	76.787	9698
					X_1 = 0.20					
1	10391	76.986	76.308	76.602	76.262	72.557	72.534	72.534	72.534	2598
2	18893	77.834	76.907	77.314	76.997	72.624	72.602	72.602	72.602	4058
3	27395	78.411	77.325	77.800	77.506	72.624	72.602	72.602	72.602	5317
4	35897	78.806	77.607	78.172	77.856	72.624	72.602	72.602	72.602	6524
5	44399	79.288	78.002	78.613	78.274	72.624	72.602	72.602	72.602	7479
6	52901	79.670	78.375	78.940	78.669	72.624	72.602	72.602	72.602	8389
					X_1 = 0.30					
1	10391	73.796	73.072	73.412	73.050	69.292	69.269	69.269	69.269	2561
2	18893	74.599	73.717	74.192	73.807	69.338	69.315	69.292	69.315	3966
3	27395	75.266	74.180	74.723	74.361	69.384	69.361	69.361	69.361	5202
4	35897	75.684	74.552	75.073	74.733	69.384	69.361	69.361	69.361	6360
5	44399	76.147	74.902	75.536	75.151	69.384	69.361	69.361	69.361	7317
6	52901	76.542	75.185	75.977	75.547	69.384	69.361	69.361	69.361	8207
					X_1 = 0.40					

1	10391	71.851	71.014	71.330	71.059	66.804	66.781	66.781	66.781	2295
2	18893	72.676	71.726	72.065	71.952	66.849	66.826	66.804	66.826	3580
3	27395	73.275	72.166	72.664	72.619	66.849	66.826	66.804	66.826	4679
4	35897	73.670	72.448	73.127	73.059	66.849	66.826	66.804	66.826	5744
5	44399	74.020	72.731	73.477	73.455	66.849	66.826	66.804	66.826	6733
6	52901	74.348	73.058	73.872	73.827	66.849	66.826	66.804	66.826	7612
					x_1 = 0.50					
1	10391	68.627	67.920	68.239	67.920	64.087	64.064	64.064	64.064	2530
2	18893	69.370	68.571	68.868	68.640	64.110	64.087	64.087	64.087	3961
3	27395	69.926	68.995	69.406	69.200	64.110	64.087	64.087	64.087	5179
4	35897	70.390	69.439	69.892	69.756	64.110	64.087	64.087	64.087	6214
5	44399	70.853	69.767	70.310	70.197	64.110	64.087	64.087	64.087	7174
6	52901	71.384	70.072	70.682	70.547	64.110	64.087	64.087	64.087	8041
					x_1 = 0.60					
1	10391	66.687	65.956	66.276	65.933	62.329	62.283	62.283	62.283	2652
2	18893	67.498	66.631	67.110	66.745	62.374	62.329	62.329	62.329	4058
3	27395	68.081	67.100	67.625	67.259	62.374	62.329	62.329	62.329	5293
4	35897	68.482	67.432	68.002	67.683	62.374	62.329	62.329	62.329	6457
5	44399	68.837	67.764	68.380	67.992	62.374	62.329	62.329	62.329	7522
6	52901	69.099	68.050	68.621	68.415	62.374	62.329	62.329	62.329	8524
					x_1 = 0.80					
1	10391	63.194	62.440	62.828	62.554	59.089	59.065	59.065	59.065	2821
2	18893	63.708	62.886	63.343	63.069	59.136	59.112	59.089	59.112	4564
3	27395	64.200	63.218	63.789	63.584	59.136	59.112	59.112	59.112	5982
4	35897	64.601	63.619	64.212	63.938	59.136	59.112	59.112	59.112	7216
5	44399	64.933	63.996	64.567	64.270	59.136	59.112	59.112	59.112	8340
6	52901	65.356	64.306	65.059	64.580	59.136	59.112	59.112	59.112	9269

Table D.15 Experimental Data of Heat Transfer From A horizontal **1024 fpm Tube** To Pool of Saturated **Acetone(1)-Isopropanol(2)-Water(3)** Liquid Mixtures

Run no.	Heat flux W/m^2	Outer wall temperature, T_{wo} °C				Liquid temperature, T_s °C				Heat transfer coefficient W/(m^2 K)
		Top	Side	Bottom	Side	Top	Side	Bottom	Side	
					X_1 = 0.011, X_2 = 0.074					
1	10391	85.168	84.217	84.460	84.305	80.354	80.288	80.265	80.288	2451
2	18893	86.015	85.042	85.219	85.064	80.420	80.332	80.310	80.332	3789
3	27395	86.708	85.624	85.845	85.712	80.420	80.354	80.310	80.354	4881
4	35897	87.113	86.051	86.471	86.184	80.420	80.354	80.332	80.354	5895
5	44399	87.540	86.434	86.876	86.677	80.420	80.354	80.332	80.354	6813
6	52901	87.989	86.794	87.215	87.016	80.420	80.354	80.332	80.354	7680

					$X_1 = 0.050$, $X_2 = 0.040$					
1	10391	78.389	77.371	77.620	77.394	73.620	73.575	73.529	73.575	2523
2	18893	79.486	78.241	78.603	78.513	73.665	73.597	73.552	73.597	3699
3	27395	80.203	78.999	79.451	79.225	73.665	73.597	73.575	73.597	4674
4	35897	80.741	79.414	79.967	79.812	73.665	73.597	73.575	73.597	5631
5	44399	81.301	79.885	80.305	80.305	73.665	73.597	73.575	73.597	6491
6	52901	81.772	80.246	80.754	80.821	73.665	73.597	73.575	73.597	7257
					$X_1 = 0.083$, $X_2 = 0.036$					
1	10391	75.719	74.656	75.538	74.882	70.701	70.656	70.633	70.656	2290
2	18893	76.703	75.346	76.364	75.640	70.747	70.679	70.679	70.679	3553
3	27395	77.596	76.080	77.053	76.420	70.769	70.724	70.701	70.724	4522
4	35897	78.353	76.770	77.539	77.064	70.769	70.724	70.701	70.724	5356
5	44399	78.862	77.255	77.979	77.617	70.769	70.724	70.701	70.724	6167
6	52901	79.471	77.605	78.397	78.013	70.769	70.724	70.701	70.724	6922
					$X_1 = 0.124$, $X_2 = 0.290$					
1	10391	78.050	77.394	77.462	77.303	73.439	73.394	73.371	73.394	2502
2	18893	79.237	78.490	78.807	78.558	73.507	73.462	73.416	73.462	3557
3	27395	79.938	79.293	79.805	79.518	73.529	73.462	73.416	73.462	4439
4	35897	80.453	79.834	80.365	80.055	73.529	73.462	73.416	73.462	5350
5	44399	80.836	80.195	80.903	80.438	73.529	73.462	73.416	73.462	6231
6	52901	81.197	80.489	81.285	80.732	73.529	73.348	73.416	73.348	7039
					$X_1 = 0.155$, $X_2 = 0.150$					
1	10391	75.403	74.950	74.860	75.199	70.837	70.814	70.792	70.814	2423
2	18893	76.409	75.640	75.775	76.318	70.882	70.860	70.814	70.860	3646
3	27395	77.121	76.216	76.601	77.257	70.905	70.882	70.860	70.882	4631
4	35897	77.856	76.656	77.199	77.991	70.905	70.882	70.860	70.882	5486
5	44399	78.319	77.029	77.753	78.568	70.905	70.882	70.860	70.882	6311
6	52901	78.759	77.424	78.126	79.007	70.905	70.882	70.860	70.882	7104
					$X_1 = 0.212$, $X_2 = 0.182$					
1	10391	73.389	72.620	73.072	72.801	68.470	68.425	68.402	68.425	2289
2	18893	74.599	73.468	74.192	73.717	68.516	68.470	68.470	68.470	3428
3	27395	75.560	74.225	75.175	74.496	68.539	68.493	68.470	68.493	4304
4	35897	76.340	74.824	75.842	75.141	68.539	68.493	68.470	68.493	5101
5	44399	77.052	75.287	76.396	75.672	68.539	68.493	68.470	68.493	5840
6	52901	77.628	75.705	76.655	76.067	68.539	68.493	68.470	68.493	6600
					$X_1 = 0.266$, $X_2 = 0.282$					
1	10391	73.186	72.484	73.412	72.620	68.539	68.493	68.470	68.493	2348
2	18893	74.622	73.513	74.667	73.649	68.584	68.539	68.516	68.539	3393
3	27395	75.605	74.044	75.628	74.383	68.607	68.584	68.539	68.584	4324
4	35897	76.566	74.439	76.385	74.869	68.630	68.584	68.562	68.584	5147
5	44399	77.301	75.083	76.939	75.400	68.630	68.584	68.562	68.584	5849

6	52901	77.900	75.456	77.311	75.773	68.630	68.584	68.562	68.584	6596
					$X_1 = 0.294,\ X_2 = 0.484$					
1	10391	73.751	72.959	73.389	72.982	69.178	69.155	69.155	69.155	2529
2	18893	74.622	73.717	74.260	73.739	69.201	69.178	69.155	69.178	3851
3	27395	75.288	74.270	74.881	74.451	69.201	69.178	69.155	69.178	4941
4	35897	75.752	74.643	75.367	74.824	69.201	69.178	69.155	69.178	6015
5	44399	76.192	74.970	75.785	75.083	69.201	69.178	69.155	69.178	7014
6	52901	76.701	75.298	76.203	75.388	69.201	69.178	69.155	69.178	7873
					$X_1 = 0.389,\ X_2 = 0.212$					
1	10391	69.541	69.038	68.445	68.901	64.315	64.269	64.269	64.269	2211
2	18893	70.798	70.029	69.462	69.961	64.361	64.269	64.269	64.269	3274
3	27395	71.827	70.560	70.356	71.035	64.361	64.269	64.269	64.269	4118
4	35897	72.607	71.046	71.000	71.883	64.361	64.269	64.269	64.269	4889
5	44399	73.228	71.350	71.599	72.436	64.361	64.269	64.269	64.269	5648
6	52901	73.737	71.746	72.017	72.900	64.361	64.269	64.269	64.269	6368
					$X_1 = 0.524,\ X_2 = 0.267$					
1	10391	67.349	66.413	67.052	66.185	62.466	62.420	62.374	62.420	2400
2	18893	68.640	67.316	68.343	67.179	62.511	62.466	62.420	62.466	3496
3	27395	69.700	67.944	69.360	67.922	62.557	62.511	62.466	62.511	4404
4	35897	70.412	68.505	70.050	68.505	62.557	62.511	62.489	62.511	5240
5	44399	70.966	68.974	70.626	68.996	62.557	62.511	62.489	62.511	6021
6	52901	71.429	69.596	71.022	69.415	62.557	62.511	62.489	62.511	6740

APPENDIX E

COMPUTER PROGRAMS, SAMPLE DATA AND RESULT FILES

This appendix embodies the various computer programs developed during the present investigation. The programs are written using **FORTRANT 77** and run in **IBM/PC.486**. The details of the computer programs, sample data file and results are given below. Table E.1 lists the name of computer program, associated data file and the purpose of the program.

Table E.1 Name of the computer programs alongwith its purpose and data file

S. No.	Name of the program(s)	Name of data file(s)	Purpose
1.	AMEERH.FOR	AMEEREM.DAT AMEERH.DAT	Converts thermocouple EMF to temperature, calculates wall superheat, and heat transfer coefficient
2.	AMEERHID.FOR	AMEERST.DAT	Calculates heat transfer coefficient of pure component and ideal heat transfer coefficient, h_{id}, using Stephan-Abdelsalam corrlation.
3.	AMEERCOP.FOR* * *	AMEERCOP.DAT	Calculates heat transfer coefficient of pure component using Cooper correlation.
4.	AMEERDL.FOR*, * *	AMEERD.DAT	Calculates density of liquid mixture.
5.	AMEERVL.FOR*, * *	AMEERVA.DAT	Calculates density of vapour mixture.
6.	AMEERHS.FOR*, * *	AMEERHS.DAT	Calculates differential latent heat of vaporization of mixture
7.	AMEERCP.FOR*, * *	AMEERCP.DAT	Calculates specific heat of liquid mixture

Table E-1 Contd...

8. AMEERS.FOR*,**	AMEERS.DAT	Calculates surface tension of liquid mixture.
9. AMEERVI.FOR*,**	AMEERVI.DAT	Calculates viscosity of liquid mixture.
10. AMEERK.FOR*,**	AMEERK.DAT	Calculates thermal conductivity of liquid mixture.
11. TEJA.FOR*	TEJA.DAT	Calculates the pseoducritical properties of mixture.
12. AMSHUL.FOR**	AMSHUL.DAT	Calculates the heat transfer coefficient using Schlunder correlation.
13. AMTHO.FOR***	AMTHO.DAT	Calculate the heat transfer coefficient using Thome-Shakir correlation.
14. AMPAL.FOR***	AMPAL.DAT	Calculate the heat transfer coefficient using Palen-Small correlation.
15. AMCALUS.FOR***	AMCALUS.DAT	Calculate the heat transfer coefficient using Calus-Leonidopoulos correlation.
16. AMSTPR.FOR***	AMSTPR.DAT	Calculate the heat transfer coefficient using Stephan-Preusser correlation.

* *All the programs use subroutine TEJA.FOR. However, the actual subroutine is only given in program AMEERD.FOR to avoid duplicity and to save space.*

** *The original program was received from Bajorek (1992), which was modified to suit the computer used in the present investigation.*

*** *The programs are similar to AMEERHID. FOR and thus, have not been included due to space limitations.*

PROGRAMS

```
C*    Name of the program: AMEERH.FOR
C*    This program converts thermocouple EMF to temperature and
C*    calculates the heat transfer coefficient  from experimental
C*    data
C*    Description of variables:
C*    DO    = outside diameter of heating tube, m
C*    DTH   = pitch circle diameter of wall thermocouple, m
C*    L     = heating tube length, m
C*    Q     = power supply, W
C*    DELT  = wall superheat
C*    SUM1  = average wall temperature
C*    CSUM1 = corrected average wall temperature
C*    SUM2  = average liquid temperature
C*    HF    = heat flux, W/sq.m
C*    H     = heat transfer coefficient, W/sq.m K
C*    EC(I) EMF values for which temperature EMF linear relations

C*    are given
C*   S(I), CI(I) are the slope and intercept of the straight line
C*   line equation TE(I) = EMF(I) *S(I) + CI(I)

      DIMENSION EMF(8),HF(8),H(8),EC(15),S(15),CI(15),TE(8),HP(8)
      OPEN(UNIT=1,FILE='AMEEREM.DAT')
      OPEN(UNIT=2,FILE='AMEERH.DAT')
      OPEN(UNIT=3,FILE='AMEERH.OUT')
      READ(1,*)(EC(I),S(I),CI(I),I=1,11)
      READ(2,*)(Q(I),I=1,7)
      SUM1=0.0
      SUM2=0.0
      DO 10 J=1,7
      READ(2,*)(EMF(I),I=1,8)
      CALL CONV(8,EMF,EC,S,CI,TE)
      SUM1=(TE(1)+TE(2)+TE(3)+TE(4))/4.0
      SUM2=(TE(5)+TE(6)+TE(7)+TE(8))/4.0
C*    WALL TEMPERATURE CORRECTION, (WCT)
      DO=0.0312
      DTH=0.0246
      L=0.108
      PI=3.14159
      THK=133.5
      WCT=(Q(J)/(2.0*PI*L*THK)) ALOG(DO/DTH)
      CSUM1=SUM1-WCT
      DELT=CSUM1-SUM2
      HF=Q/(PI*DO*L)
      H=HF/DELT
      WRITE(3,21)HF,(TE(I),I=1,8),H
      WRITE(3,*)HF,(TE(I),I=1,8),H
      WRITE(3,*)HF,(EMF(I),I=1,8),H
10    CONTINUE
21    FORMAT(1X,F6.0,1X,6(F6.3,1X),2(F8.1))
      STOP
      END
```

```
      SUBROUTINE CONV(N,EMF,EC,S,CI,TE)

      DIMENSION EMF(70),TE(70),EC(15),S(15),CI(15)
      DO 30 I=1,N
      DO 40 J=1,11
      EEMF=EMF(I)
      EEC=EC(J)
      IF(EEMF)50,50,60
60    IF(EEMF-EEC)70,70,40
40    CONTINUE
50    TE(I)=0.000
      IF(EEMF.EQ.5.646)TE(55)=129.85
      GO TO 30
70    TE(I)=EMF(I)*S(J)+CI(J)
30    CONTINUE
      RETURN
      END
```

To show the results obtained from the program a sample data file have created by taking the Input data from Run NO.1 to 7 of Table D.1 Appendix D. The different data files used in the above program along with the result file are given below:

Data file name: AMEEREM.DAT

This data file contains EMF to temperature conversion data.

For a copper-constantan thermocouple and contains the different values of S(I),and CI(I) variables.

```
1.188,24.613402,0.759278,1.600,24.271800,1.165120
2.020,23.809523,1.904764,2.448,23.364486,2.803738
2.886,22.831050,4.109589,3.328,22.624434,4.705882
3.788,22.123890,6.371680,4.238,21.834060,7.467249
4.704,21.459220,9.055794,5.176,21.186440,10.33989
5.600,20.990566,11.352830
```

Data file name: AMEERH.DAT

This file contains power input, Q and EMF data as read from thermocouples,(taken from Run No. 1 to 7 of Table D.1.

```
110.0,200.0,290.0,380.0,470.0,560.0,610.0
2.680,2.655,2.656,2.649,2.282,2.280,2.279,2.280
2.784,2.725,2.713,2.714,2.285,2.282,2.280,2.282
2.868,2.788,2.770,2.773,2.285,2.283,2.282,2.283
2.944,2.852,2.830,2.836,2.287,2.285,2.284,2.285
2.981,2.892,2.869,2.875,2.287,2.285,2.284,2.285
3.023,2.925,2.909,2.915,2.287,2.285,2.284,2.285
3.048,2.945,2.931,2.936,2.287,2.285,2.284,2.285
```

Output data

Output file name: AMEERH.OUT

```
 q        Outer wall temperature, T   °C        Liquid temperature, T °C              h
                                   wo                                l

10391    65.008  64.437  64.300  64.460      56.121  56.075  56.051  56.075      1227
18893    67.147  65.799  65.548  65.526      56.192  56.121  56.075  56.121      1913
27395    68.828  67.002  66.659  66.591      56.192  56.145  56.121  56.145      2464
35897    70.315  68.227  67.861  67.725      56.238  56.192  56.168  56.192      2910
44399    70.916  68.903  68.516  68.379      56.238  56.192  56.168  56.192      3420
52901    71.630  69.413  69.187  69.051      56.238  56.192  56.168  56.192      3883
57624    72.065  69.734  69.531  69.418      56.238  56.192  56.168  56.192      4119
```

```
C     This program calculates the boiling heat transfer of liquid
C     coefficient using Stephan-Abdelsalam correlation
C     Reference: Stephan-Abdelsalam(1980)
C     SF: Surface tension; PL: Latent heat of vaporization, J/kg;
C     DL:Liquid  density, kg/m^3;  Dv:Vapour density, kg/m^3;
C     TK: Liquid thermal conductivity, W/(m K); CP:liquid specific
      heat kJ/(kg K); Y:  Contact angle, degree; TS; Liquid
      saturation temperature, K; HF: Heat flux, W/m^2 ;
      HEXP: Experimental heat transfer coefficient, W/(m^2 K)
C     HCAL: Predicted heat transfer transfer coefficient, W/(m^2 K)
C     DEV: Deviation; SD: Standard deviation;
C     ERR: Mean absolute error; N = 1 for hydrocarbons,
C     N = 2 for distilled water
C
      DIMENSION HF(10),HEXP(10)
      OPEN(UNIT=1,FILE='AMEERST.DAT')
      OPEN(UNIT=2,FILE='AMEERST.OUT')
      ST=0.0
      ERR=0.0
      READ(1,*)N,NF
      READ(1,*)(HF(I),I=1,NF)
      READ(1,*)SF,PL,DL,DV,TK,CP,Y,TS,VIS
      READ(1,*)(HEXP(I),I=1,NF)
      WRITE(2,*)
      G=9.81
      DIFF=TK/(DL*CP*1000.0)
      DIM=0.0146*Y*(2.0*SF/(G*(DL-DV)))**0.5
      Z3=CP*TS*DIM*DIM*1000.0/(DIFF*DIFF)
      Z13=(DL-DV)/DL
      Z4=(PL*DIM**2.0)/(DIFF**2.0)
      Z5=DV/DL
      DO 50 I=1,NF
      Z1=HF(I)*DIM/(TK*TS)
      IF(N.EQ.3) GO TO 60
      HCAL=0.0546*(TK/DIM)*(((Z5**0.5)*Z1)**0.674)*
     1(Z13**(-4.33))*(Z4**(0.248))
      GO TO 70
60    HCAL=0.246*1.0E+07*(TK/DIM)*(Z1**(0.673))*(Z4**(-1.58))*
     1(Z3**(1.26)*(Z13**(5.22))
70    DEV=(HCAL-HEXP(I))*100.0/HEXP(I)
      HEXP1=HEXP(I)
      HF1=HF(I)
      ST=ST+((HCAL-HEXP1)/HEXP1)**2.0
      ERR=ERR+ABS(HCAL-HEXP1)*100.0/HEXP1
      WRITE(*,*)N,HEXP1,HCAL,DEV
      WRITE(2,41)HF1,HEXP1,HCAL,DEV
41    FORMAT(3F10.1,F8.2)
50    CONTINUE
100   CONTINUE
      ST=((ST/(NF))**0.5)*100.0
      ERR=ERR/(NF)
      WRITE(2,*)
      WRITE(2,42)ST,ERR
42    FORMAT(3X,'SD =',F10.2,2X,', ERR = ',F10.2)
```

```
      WRITE(2,*)
      STOP
      END
C***************************************************************
```

INPUT DATA, PURE ACETONE, TABLE. D.1, APPENDIX D

```
1,7
10391.0,18893.0,27395.0,35897.0,44399.0,52901.0,57624.0
0.01998,506200.0,759.3,2.218,0.1347,2.279,35.0,329.3
1227.0,1913.0,2464.0,2910.0,3420.0,3883.0,4119.0
```

OUPUT DATA

```
     q          hexpl       hpred       Dev%

  10391.0      1227.0      1163.1      -5.21
  18893.0      1913.0      1736.1      -9.25
  27395.0      2464.0      2226.8      -9.63
  35897.0      2910.0      2668.9      -8.28
  44399.0      3420.0      3077.4     -10.02
  52901.0      3883.0      3460.7     -10.88
  57624.0      4119.0      3664.8     -11.03

  SD = 9.37%,    ERR = 9.18%

***************************************************************
```

LIQUID DENSITY

```
C     This program calculates specific volume of liquid mixture
C     Reference: Teja(1980)
C
      COMMON/B1/FRAC(3),TCRIT(3),VCRIT(3),ZCRIT(3),OMEGA(3),WM(3)
      COMMON/B2/ICOMP(3),PCRIT(3),ZRA(3)
      OPEN(UNIT=1,FILE='AMEERD.DAT')
      OPEN(UNIT=2,FILE='AMEERD.OUT')
      OPEN(UNIT=3,FILE='AMEERD1.OUT')
      DO 500 NO=1,4
      READ(1,*)N
      WRITE(3,*)
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(PCRIT(I),I=1,N)
      READ(1,*)(ZRA(I),I=1,N)
      READ(1,*)N1
      DO 100 K=1,N1
      KCOMP=1
      II=ICOMP(KCOMP)
      JJ=ICOMP(KCOMP+1)
      READ(1,*)TS,(FRAC(I),I=1,N)
      CALL PSEUDO(N,TCM,VCM,WMCM,ZCM)
      RU=0.08314
      TR=TBP/TCM
      CALL VLIQUID(II,TR,VL)
      VRI=(VL/VCRIT(II))
      WRITE(2,51)VRI,VL1,VL,ZRA(II),TR
51    FORMAT(5F10.4)
      VMIXL=VL
      TR=TBP/TCM
      CALL VLIQUID(JJ,TR,VL)
      VRJ=(VL/VCRIT(JJ))
      WRITE(2,52)VRJ,VL1,VL,ZRA(JJ),TR
52    FORMAT(5F10.4)
      Z2=FRAC(KCOMP)*OMEGA(II)+FRAC(KCOMP+1)*OMEGA(JJ)
      Z3=FRAC(KCOMP)+FRAC(KCOMP+1)
      OMPSU=Z2/Z3
      ZPSU=ZCRIT(JJ)+((OMPSU-OMEGA(JJ))/(OMEGA(II)-OMEGA(JJ)))*
     1(ZCRIT(II)-ZCRIT(JJ))
      VRPSU=(ZCRIT(JJ)*VRJ+((OMPSU-OMEGA(JJ))/(OMEGA(II)-OMEGA(JJ)))*
     1(ZCRIT(II)*VRI-ZCRIT(JJ)*VRJ))/ZPSU
      WRITE(2,53)OMPSU,ZPSU,VRPSU
53    FORMAT(3F12.5)
      VRPSUN=VRPSU
```

```
      DO 50 KCOMP=3,N
      II=ICOMP(KCOMP)
      ZCI=ZCRIT(II)
      OMI=OMEGA(II)
      TR=TBP/TCM
      CALL VLIQUID(II,TR,VL)
      VRI=(VL/VCRIT(II))
      WRITE(2,54)II,TR,VL1,VL,VRI
54    FORMAT(I2,4F10.4)
      Z2=0.0
      Z3=0.0
      DO 60 L=1,KCOMP
      LL=ICOMP(L)
      Z2=Z2+FRAC(L)*OMEGA(LL)
      Z3=Z3+FRAC(L)
60    CONTINUE
      OMPSUN=Z2/Z3
      ZPSUN=ZCI+((OMPSUN-OMI)/(OMPSU-OMI))*(ZPSU-ZCI)
      VRPSUN=(ZPSU*VRPSU+((OMPSUN-OMI)/(OMPSU-OMI))*
     1(ZCI*VRI-ZPSU*VRPSU))/ZPSUN
      WRITE(2,55)OMPSUN,ZPSUN,ZCI
55    FORMAT(3F10.4)
50    CONTINUE
      VMIXL=VRPSUN*VCM
      DENSL=(1.0/VMIXL)
      WRITE(2,56)VMIXL,DENSL
56    FORMAT(2F10.4)
      WRITE(3,57)FRAC(1),DENSL
57    FORMAT(2F12.4)
100   CONTINUE
500   CONTINUE
      STOP
      END

C     This subroutine calculate the mixture pseudo critical values
C
      SUBROUTINE PSEUDO (N,TCM,VCM,WMCM,ZCM)
      COMMON/B1/FRAC(3),TCRIT(3),VCRIT(3),ZCRIT(3),OMEGA(3),WM(3)
      COMMON/B2/ICOMP(3),PCRIT(3),ZRA(3)
      VCM=0.0
      TCMVCM=0.0
      WMCM=0.0
      OMCM=0.0
      DO 20 I=1,N
      DO 10 J=1,N
      II=ICOMP(I)
      JJ=ICOMP(J)
      TCI=TCRIT(II)
      VCI=VCRIT(II)
      TCJ=TCRIT(JJ)
      VCJ=VCRIT(JJ)
      VCIJ=VCI
      TCIJ=TCI*VCI
      IF(I.EQ.J)GO TO 5
```

```
      VCIJ=((VCI**(1.0/3.0)+VCJ**(1.0/3.0))**3.0)/8.0
      TCIJ=(TCI*VCI*TCJ*VCJ)**0.5
5     CONTINUE
      WRITE(2,41)VCIJ,TCIJ
41    FORMAT(2F12.6)
      Z1=FRAC(I)*FRAC(J)*VCIJ
      VCM=VCM+FRAC(I)*FRAC(J)*VCIJ
      Z2=FRAC(I)*FRAC(J)*TCIJ
      TCMVCM=TCMVCM+FRAC(I)*FRAC(J)*TCIJ
10    CONTINUE
20    CONTINUE
      TCM=TCMVCM/VCM
      WRITE(2,42)TCM,VCM,TCMVCM
42    FORMAT(3F12.6)
      DO 70 KK=1,N
      II=ICOMP(KK)
      WMCM=WMCM+FRAC(KK)*WM(II)
      ZCM=ZCM+FRAC(KK)*ZCRIT(II)
      OMCM=OMCM+FRAC(KK)*OMEGA(II)
70    CONTINUE
      WRITE(2,43)TCM,VCM,WMCM,ZCM
43    FORMAT(4F12.6)
      RETURN
      END


C     This program calculates specific volume of pure liquid
C     Reference: Spencer et al.(1972)
C
      SUBROUTINE VLIQUID(I,TR,VL)
      COMMON/B1/FRAC(3),TCRIT(3),VCRIT(3),ZCRIT(3),OMEGA(3),WM(3)
      COMMON/B2/ICOMP(3),PCRIT(3),ZRA(3)
      RU=0.08314
      V1=(1.0+(1.0-TR)**(2.0/7.0))
      VL=(RU*TCRIT(I)/WM(I)/PCRIT(I))*ZRA(I)**V1
      WRITE(2,44)I,TR,V1,VL
44    FORMAT(I2,3F10.5)
      RETURN
      END

INPUT DATA,(ACETONE-WATER)

2
1,2
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.304,0.344
58.08,18.02
47.61,221.29
0.24494,0.2338
7
349.4,0.05,0.95
342.30,0.1,0.9
```

```
338.49,0.15,0.85
335.88,0.25,0.75
333.9,0.40,0.60
332.34,0.60,0.40
330.73,0.80,0.2
```

OUTPUT DATA

```
   X1         Density, kg/m3

   0.50           999.45
   0.10           991.91
   0.15           980.87
   0.25           953.76
   0.40           912.45
   0.60           858.72
   0.80           807.75
```

****

VAPOUR DENSITY

```
C     This program calculates vapour density of mixture
C     Reference: Patel & Teja(1982)
C
      COMMON/B1/ICOMP(5),FRACX(5),FRACY(5),TCRIT(5),VCRIT(5)
      COMMON/B2/OMEGA(5),WM(5),ZCRIT(5),FRAC(5)
      COMMON/B3/PCRIT(5),F(5),PSI(5),PTEST
      OPEN(UNIT=1,FILE='AMEERVA.DAT')
      OPEN(UNIT=2,FILE='AMEERVA.OUT')
      OPEN(UNIT=3,FILE='AMEERVA1.OUT')
      DO 500 NO=1,4
      READ(1,*)N
      WRITE(3,*)
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(PCRIT(I),I=1,N)
      READ(1,*)(F(I),I=1,N)
      READ(1,*)(PSI(I),I=1,N)
      READ(1,*)N1
      DO 100 M=1,N1
      READ(1,*)TBP,(FRACX(I),I=1,N)
      READ(1,*)(FRACY(I),I=1,N)
      RU=0.08314
      PI=3.141593
      PTEST=1.01
      VMIXG=0.0
      DENSG=0.0
      VMIXL=0.0
```

```
      DENSL=0.0
      DO 5 I=1,N
      K=ICOMP(I)
5     CONTINUE
      DO 300 K=1,N
      FRAC(K)=FRACX(K)
C     IFRAC=0 user liquid mole fraction
C     IFRAC=1 user vapour mole fraction
C     IF(IFRAC.EQ.1.0) FRAC(K)=FRACY(K)
300   CONTINUE
      AM=0.0
      BM=0.0
      CM=0.0
      DO 30 I=1,N
      II=ICOMP(I)
      B=2.0-3.0*PSI(II)
      C=3.0*PSI(II)*PSI(II)
      D=-1.0*PSI(II)*PSI(II)*PSI(II)
      IFLAG=0.0
      CALL ROOT3(B,C,D,WBI,0)
      WAI=3.0*PSI(II)*PSI(II)+3.0*(1.0-2.0*PSI(II))*WBI+WBI*
     1(WBI)+1.0-3.0*PSI(II)
      WCI=1.0-3.0*PSI(II)
      AMI=WAI*RU*RU*TCRIT(II)*TCRIT(II)*(1.0+F(II)*
     1(1.0-SQRT(TBP/TCRIT(II))))**2.0/PCRIT(II)
      WRITE(2,81)IFLAG,I,WBI,WAI,WCI,AMI
81    FORMAT(2I2,3X,4F10.5)
      AMIJ=0.0
      DO 20 J=1,N
      JJ=ICOMP(J)
      IF(I.EQ.J) GO TO 10
      B=2.0-3.0*PSI(JJ)
      C=3.0*PSI(JJ)*PSI(JJ)
      D=-1.0*PSI(JJ)*PSI(JJ)*PSI(JJ)
      IFLAG=0.0
      CALL ROOT3(B,C,D,WBJ,0)
      WAJ=3.0*PSI(JJ)*PSI(JJ)+3.0*(1.0-2.0*PSI(JJ))*WBJ+WBJ*
     1(WBJ)+1.0-3.0*PSI(JJ)
      WCJ=1.0-3.0*PSI(JJ)
      AMJ=WAJ*RU*RU*TCRIT(JJ)*TCRIT(JJ)*(1.0+F(JJ)*
     1(1.0-SQRT(TBP/TCRIT(JJ))))**2.0/PCRIT(JJ)
      WRITE(2,84)IFLAG,J,WBJ,WAJ,WCJ,AMJ
84    FORMAT(2I2,3X,4F10.5)
      AMIJ=SQRT(AMI*AMJ)
      WRITE(2,85)AMIJ,II,JJ,I,J
85    FORMAT(F10.5,1X,4I2)
      GO TO 15
10    CONTINUE
      AMIJ=AMI
      WRITE(2,86)AMIJ
86    FORMAT(F10.5)
15    CONTINUE
      AM=AM+FRAC(I)*FRAC(J)*AMIJ
      WRITE(2,87)AM,I,J
```

```
87      FORMAT(F10.5,2X,2I2)
20      CONTINUE
        BM=BM+FRAC(I)*WBI*RU*TCRIT(II)/PCRIT(II)
        CM=CM+FRAC(I)*WCI*RU*TCRIT(II)/PCRIT(II)
        WRITE(2,88)I,J,BM,CM
88      FORMAT(2I2,5X,2F10.5)
30      CONTINUE
        WRITE(2,91)AM,BM,CM
91      FORMAT(3F10.5)
        B=CM-RU*TBP/PTEST
        C=AM/PTEST-BM*BM-2.0*BM*CM-RU*TBP*(BM+CM)/PTEST
        D=BM*BM*CM+RU*TBP*BM*CM/PTEST-AM*BM/PTEST
        WRITE(2,92)B,C,D
92      FORMAT(3F10.5)
        IFLAG=1
        CALL ROOT3  (B,C,D,VMIXG,1)
        CALL PSEUDO (N,TCM,VCM,WMCM,ZCM)
        VMIXG=VMIXG/WMCM
        DENSG=1.0/VMIXG
        WRITE(2,93)IFLAG,VMIXG,DENSG,WMCM
93      FORMAT(I2,4X,3F10.5)
        IFLAG=0.0
        CALL ROOT3(B,C,D,VMIXL,0)
        VMIXL=VMIXL/WMCM
        DENSL=1.0/VMIXL
        WRITE(2,94)IFLAG,VMIXL,DENSL,WMCM
94      FORMAT(I2,4X,3F10.5)
        WRITE(*,*)M,FRACX(1),VMIXL,DENSG,VMIXG
        WRITE(3,95)FRACX(1),DENSG
95      FORMAT(2F12.4)
100     CONTINUE
500     CONTINUE
        STOP
        END

C       SOLVE FOR POSITIVE ROOTS OF A CUBIC EQUATION
C
        SUBROUTINE ROOT3(B,C,D,WB,IFLAG)
        COMMON/B3/PCRT(5),F(5),PSI(5),PTEST
        COMPLEX*8   I,Y1,Y2,Y3
        I=(0.0,1.0)
        PI=3.141593
        P=(3.0*C-B*B)/3.0
        Q=(27.0*D-9.0*B*C+2.0*B**3.0)/27.0
        R=(P/3.0)**3.0+(Q/2.0)**2.0
        WRITE(2,13)B,C,D,P,Q,R
13      FORMAT(5F12.5)
        IF(R.LT.0.0) GO TO 20
        CC=-Q/2.0+SQRT(R)
        IF(CC.LT.0.) AA= -(-CC)**(1.0/3.0)
        IF(CC.GE.0.) AA= (CC)**(1.0/3.0)
        WRITE(2,14)CC,AA
14      FORMAT(5X,2F10.5)
        CC=(-Q/2.0)-SQRT(R)
```

```
      IF(CC.LT.0.) BB= -(-CC)**(1.0/3.0)
      IF(CC.GE.0.) BB= (CC)**(1.0/3.0)
      Y1=AA+BB-(B/3.0)
      Y2=-(AA+BB)/2.0+I*(SQRT(3.0)/2.0)*(AA-BB)-B/3.0
      Y3=-(AA+BB)/2.0-I*(SQRT(3.0)/2.0)*(AA-BB)-B/3.0
      WRITE(2,15)CC,BB
15    FORMAT(5X,2F10.5)
      WRITE(2,16)P,Q,R,Y1,Y2,Y3
16    FORMAT(6F10.4)
      GO TO 30
20    CONTINUE
      PHI=ACOS(SQRT((0.25*Q*Q)/((-P**3.0)/27.0)))
      Y1=2.0*SQRT(-P/3.0)*COS(PHI/3.0)
      Y2=2.0*SQRT(-P/3.0)*COS((PHI+2.0*PI)/3.0)
      Y3=2.0*SQRT(-P/3.0)*COS((PHI+4.0*PI)/3.0)
      WRITE(2,17)PHI,Y1,Y2,Y3
17    FORMAT(4F9.5)
      IF(Q.LT.0.0) GO TO 25
      Y1=-Y1
      Y2=-Y2
      Y3=-Y3
      WRITE(2,18)Y1,Y2,Y3
18    FORMAT(3F10.5)
25    CONTINUE
      Y1=Y1-B/3.0
      Y2=Y2-B/3.0
      Y3=Y3-B/3.0
      WRITE(2,19)Y1,Y2,Y3
19    FORMAT(3F10.5)
30    CONTINUE
C     DETERMINE THE MAX. POSITIVE REAL ROOT
C     DETERMINE THE MIN. POSITIVE REAL ROOT
      ROOTA=0.0
      ROOTB=0.0
      ROOTC=0.0
      IF(AIMAG(Y1).EQ.0.) ROOTA = AMAX1(REAL(Y1),0.0)
      IF(AIMAG(Y2).EQ.0.) ROOTB = AMAX1(REAL(Y2),0.0)
      IF(AIMAG(Y3).EQ.0.) ROOTC = AMAX1(REAL(Y3),0.0)
      WRITE(2,21)AIMAG(Y1),AIMAG(Y2),AIMAG(Y3)
21    FORMAT(3E10.5)
      WRITE(2,22)ROOTA,ROOTB,ROOTC
22    FORMAT(3F10.5)
      WB=1.0E+10
      IF(ROOTA.GT.0.0) WB = AMIN1(WB,ROOTA)
      IF(ROOTB.GT.0.0) WB = AMIN1(WB,ROOTB)
      IF(ROOTC.GT.0.0) WB = AMIN1(WB,ROOTC)
      IF(IFLAG.EQ.1.0) WB = AMAX1(ROOTA,ROOTB,ROOTC)
      CHECK=WB**3.0+B*WB**2.0+C*WB+D
      IF(ABS(CHECK).GT.0.5E-02) WRITE(2,23)WB,CHECK
23    FORMAT(2F10.5)
      IF(CHECK.GE.1.0E+10) WRITE(2,24)WB,CHECK
24    FORMAT(2F10.5)
      WRITE(2,26)WB,CHECK
26    FORMAT(2F10.5)
```

```
      RETURN
      END
```

INPUT DATA,(ACETONE-WATER)

```
2
1,2
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.304,0.344
58.08,18.02
47.61,221.29
0.713,0.689803
0.285,0.269
7
349.4,0.05,0.95
0.639,0.361
342.30,0.1,0.9
0.763,0.237
338.49,0.15,0.85
0.695,0.305
335.88,0.25,0.75
0.836,0.164
333.9,0.40,0.60
0.857,0.143
332.34,0.60,0.40
0.877,0.123
330.73,0.80,0.2
0.898,0.102
```

OUTPUT DATA

```
   X1          Density, kg/m3

   0.50            0.704
   0.10            0.792
   0.15            0.874
   0.25            1.029
   0.40            1.261
   0.60            1.571
   0.80            1.887
```

```
C*******************************************************************
```

LATENT HEAT OF VAPORIZATION

```
C     This  program calculates the differential latent heat of
      vaporization of mixture
C     Reference: Teja(1983), Reid et al.(1987)
C
      DIMENSION CPVAPA(5),CPVAPB(5),CPVAPC(5),CPVAPD(5)
```

```
      COMMON/B1/ICOMP(5),FRACX(50),FRACY(50),TCRIT(5),VCRIT(5)
      COMMON/B2/OMEGA(5),WM(5),ZCRIT(5),FRAC(5)
      COMMON/B3/PCRIT(5),F(5),PSI(5),PTEST
      OPEN(UNIT=1,FILE='AMEERHS.DAT')
      OPEN(UNIT=2,FILE='AMEERHS.OUT')
      OPEN(UNIT=3,FILE='AMEERHS1.OUT')
      DO 500 NO=1,4
      WRITE(3,*)
      READ(1,*)N
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(PCRIT(I),I=1,N)
      READ(1,*)(F(I),I=1,N)
      READ(1,*)(PSI(I),I=1,N)
      READ(1,*)(CPVAPA(I),I=1,N)
      READ(1,*)(CPVAPB(I),I=1,N)
      READ(1,*)(CPVAPC(I),I=1,N)
      READ(1,*)(CPVAPD(I),I=1,N)
      READ(1,*)N1
      DO 100 M=1,N1
      READ(1,*)TBP,(FRACX(I),I=1,N)
      READ(1,*)(FRACY(I),I=1,N)
      READ(1,*)VMIXL,VMIXG
      RU=0.08314
      PI=3.141593
      PTEST=1.01
      HOL=0.0
      HOV=0.0
      HVMOL=0.0
      HVKJKG=0.0
      DO 5 I=1,N
      K=ICOMP(I)
5     CONTINUE
      AM=0.0
      BM=0.0
      CM=0.0
      FPSU=0.0
      WAPSU=0.0
      DO 30 I=1,N
      II=ICOMP(I)
      B=2.0-3.0*PSI(II)
      C=3.0*PSI(II)*PSI(II)
      D=-1.0*PSI(II)*PSI(II)*PSI(II)
      CALL ROOT3(B,C,D,WBI,0)
      WAI=3.0*PSI(II)*PSI(II)+3.0*(1.0-2.0*PSI(II))*WBI+WBI*
     1(WBI)+1.0-3.0*PSI(II)
      WCI=1.0-3.0*PSI(II)
      AMI=WAI*RU*RU*TCRIT(II)*TCRIT(II)*(1.0+F(II)*
     1(1.0-SQRT(TBP/TCRIT(II))))**2.0/PCRIT(II)
      WRITE(2,82)II,B,C,D
```

```
82     FORMAT(I2,3X,3F9.4)
       WRITE(2,83)I,WBI,WAI,WCI,AMI
83     FORMAT(I2,2X,4F8.4)
       AMIJ=0.0
       DO 20 J=1,N
       JJ=ICOMP(J)
       IF(I.EQ.J) GO TO 10
       B=2.0-3.0*PSI(JJ)
       C=3.0*PSI(JJ)*PSI(JJ)
       D=-1.0*PSI(JJ)*PSI(JJ)*PSI(JJ)
       CALL ROOT3(B,C,D,WBJ,0)
       WAJ=3.0*PSI(JJ)*PSI(JJ)+3.0*(1.0-2.0*PSI(JJ))*WBJ+WBJ*
      1(WBJ)+1.0-3.0*PSI(JJ)
       WCJ=1.0-3.0*PSI(JJ)
       AMJ=WAJ*RU*RU*TCRIT(JJ)*TCRIT(JJ)*(1.0+F(JJ)*
      1(1.0-SQRT(TBP/TCRIT(JJ))))**2.0/PCRIT(JJ)
       WRITE(2,84)JJ,B,C,D
84     FORMAT(3F8.4)
       WRITE(2,85)J,WBJ,WAJ,WCJ,AMJ
85     FORMAT(I2,2X,4F8.4)
       AMIJ=SQRT(AMI*AMJ)
       WRITE(2,86)II,JJ,I,J,AMIJ
86     FORMAT(4I2,2X,F10.5)
       GO TO 15
10     CONTINUE
       AMIJ=AMI
       WRITE(2,87)I,J,AMIJ
87     FORMAT(2I2,3X,F10.5)
15     CONTINUE
       AM=AM+FRACX(I)*FRACX(J)*AMIJ
       WRITE(2,88)I,J,AM
88     FORMAT(2I2,3X,F10.5)
20     CONTINUE
       BM=BM+FRACX(I)*WBI*RU*TCRIT(II)/PCRIT(II)
       CM=CM+FRACX(I)*WCI*RU*TCRIT(II)/PCRIT(II)
       WRITE(2,89)I,J,BM,CM
89     FORMAT(2I2,5X,2F10.5)
       FPSU=FPSU+FRACX(I)*F(II)
       WAPSU=WAPSU+FRACX(I)*WAI
       WRITE(2,91)I,J,AM,BM,CM,FPSU,WAPSU
91     FORMAT(2I2,1X,5F8.4)
30     CONTINUE
       ZN=SQRT(BM*CM+0.25*(BM+CM)**2.0)
       ZM=(0.5*(BM+CM)-ZN)*(PTEST/RU/TBP)
       ZQ=(0.5*(BM+CM)+ZN)*(PTEST/RU/TBP)
       CALL PSEUDO(0,N,TCM,VCM,WMCM,ZCM)
       PCM=ZCM*RU*TCM/VCM/WMCM
       DADT=-(WAPSU*RU*RU*TCM*FPSU/PCM)*(1.0+FPSU*
      1(1.0-SQRT(TBP/TCM)))/SQRT(TBP/TCM)
       FF=100.0
       Z=PTEST*VMIXL*WMCM/(RU*TBP)
       WRITE(2,92)Z,ZM,ZQ,PCM,DADT
92     FORMAT(5F9.4)
       CH=ALOG((Z+ZM)/(Z+ZQ))
```

```
      HDEPL=FF*(RU*TBP*(Z-1.0)-(TBP*DADT-AM)*(ALOG((Z+ZM)/
     1(Z+ZQ))/(2.0*ZN)))/WMCM
      HDEPL1=RU*TBP*(Z-1.0)
      HDEPL2=TBP*DADT-AM
      HDEPL3= ALOG ((Z+ZM)/(Z+ZQ))
      HDEPL4=2.0*ZN
      WRITE(2,93)HDEPL,HDEPL1,HDEPL2,HDEPL3,HDEPL4
93    FORMAT(5F10.4)
C     CALCULATE THE IDEAL GAS ENTHALPY
      TREF=0.0
      DO 45 I=1,N
      II=ICOMP(I)
      HI=CPVAPA(II)*(TBP-TREF)+CPVAPB(II)*
     1(TBP**2.0-TREF**2.0)/2.0+
     1(CPVAPC(II))*(TBP**3-TREF**3)/3.0+CPVAPD(II)*
     1(TBP**4-TREF**4)/4.0
      HOL=HOL+FRACX(I)*HI
      HOV=HOV+FRACY(I)*HI
      WRITE(2,94)I,J,HI,HOL,HOV
94    FORMAT(2I2,2X,3F11.4)
45    CONTINUE
      HOL=HOL/WMCM
      WRITE(2,95)HOL
95    FORMAT(F12.4)
      HVMOLL=HDEPL*WMCM/1000.0
C     CALCULATE THE VAPOR PHASE ENTHALPY DEPARTURE
      AM=0.0
      BM=0.0
      CM=0.0
      FPSU=0.0
      WAPSU=0.0
      DO 80 I=1,N
      II=ICOMP(I)
      B=2.0-3.0*PSI(II)
      C=3.0*PSI(II)*PSI(II)
      D=-1.0*PSI(II)*PSI(II)*PSI(II)
      CALL ROOT3(B,C,D,WBI,0)
      WAI=3.0*PSI(II)*PSI(II)+3.0*(1.0-2.0*PSI(II))*WBI+WBI*
     1(WBI)+1.0-3.0*PSI(II)
      WCI=1.0-3.0*PSI(II)
      AMI=WAI*RU*RU*TCRIT(II)*TCRIT(II)*(1.0+F(II)*
     1(1.0-SQRT(TBP/TCRIT(II))))**2.0/PCRIT(II)
      WRITE(2,96)I,J,B,WBI,WAI,WCI,AMI
96    FORMAT(2I2,1X,5F7.4)
      AMIJ=0.0
      DO 70 J=1,N
      JJ=ICOMP(J)
      IF(I.EQ.J) GO TO 60
      B=2.0-3.0*PSI(JJ)
      C=3.0*PSI(JJ)*PSI(JJ)
      D=-1.0*PSI(JJ)*PSI(JJ)*PSI(JJ)
      CALL ROOT3(B,C,D,WBJ,0)
      WAJ=3.0*PSI(JJ)*PSI(JJ)+3.0*(1.0-2.0*PSI(JJ))*WBJ+WBJ*
     1(WBJ)+1.0-3.0*PSI(JJ)
```

```
      WCJ=1.0-3.0*PSI(JJ)
      AMJ=WAJ*RU*RU*TCRIT(JJ)*TCRIT(JJ)*(1.0+F(JJ)*
     1(1.0-SQRT(TBP/TCRIT(JJ))))**2.0/PCRIT(JJ)
      WRITE(2,97)I,J,B,WBJ,WAJ,WCJ,AMJ
97    FORMAT(2I2,1X,5F7.4)
      AMIJ=SQRT(AMI*AMJ)
      GO TO 65
60    CONTINUE
      AMIJ=AMI
65    CONTINUE
      AM=AM+FRACY(I)*FRACY(J)*AMIJ
70    CONTINUE
      BM=BM+FRACY(I)*WBI*RU*TCRIT(II)/PCRIT(II)
      CM=CM+FRACY(I)*WCI*RU*TCRIT(II)/PCRIT(II)
      FPSU=FPSU+FRACY(I)*F(II)
      WAPSU=WAPSU+FRACY(I)*WAI
      WRITE(2,98)I,J,AM,BM,CM,FPSU,WAPSU
98    FORMAT(2I2,5F8.4)
80    CONTINUE
      ZN=SQRT(BM*CM+0.25*(BM+CM)**2.0)
      ZM=(0.5*(BM+CM)-ZN)*(PTEST/RU/TBP)
      ZQ=(0.5*(BM+CM)+ZN)*(PTEST/RU/TBP)
      CALL PSEUDO(1,N,TCM,VCM,WMCM,ZCM)
      ZCM=0.23805
      PCM=ZCM*RU*TCM/VCM/WMCM
      DADT=-(WAPSU*RU*RU*TCM*FPSU/PCM)*(1.0+FPSU*(1.0-SQRT(TBP/TCM)))
     1(SQRT(TBP/TCM))
C     CONVERSION FACTOR FF TO PUT INTO KJ/KG-MOL
      FF=100.0
      Z=PTEST*VMIXG*WMCM/(RU*TBP)
      WRITE(2,111)Z,ZN,ZM,ZQ,DADT
111   FORMAT(5F10.4)
      HDEPV=FF*(RU*TBP*(Z-1.0)-(TBP*DADT-AM)*(ALOG((Z+ZM)/(Z+ZQ))/
     1(2.0*ZN)))/WMCM
      HDEPV1=RU*TBP*(Z-1.0)
      HDEPV2=TBP*DADT-AM
      HDEPV3=ALOG((Z+ZM)/(Z+ZQ))
      HDEPV4=2.0*ZN
      HVMOLV=HDEPV*WMCM/1000.0
      WRITE(2,112)HDEPV,HDEPV1,HDEPV2,HDEPV3,HDEPV4,HVMOLV
112   FORMAT(6F10.3)
      HOV=HOV/WMCM
      HVKJKG=(HDEPV-HDEPL)+(HOV-HOL)
      HVMOL=HVMOLV-HVMOLL
      WRITE(2,113)HOV,HVKJKG,HVMOL
113   FORMAT(3F11.3)
      WRITE(*,*)M,FRACX(1),HVKJKG
      WRITE(3,114)FRACX(1),HVKJKG
114   FORMAT(2F12.4)
100   CONTINUE
500   CONTINUE
      STOP
      END
```

INPUT DATA,(ACETONE-WATER)

```
2
1,2
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.309,0.344
58.08,18.02
47.61,221.29
0.713,0.689803
0.285,0.269
6.301E+00,3.224E+01
2.606E-01,1.924E-03
-1.253E-04,1.055E-05
2.038E-08,-3.596E-09
7
349.4,0.05,0.95
0.639,0.361
0.00105,1.4204
342.30,0.1,0.9
0.763,0.237
0.001085,1.263
338.49,0.15,0.85
0.695,0.305
0.00111,1.144
335.88,0.25,0.75
0.836,0.164
0.00117,0.9713
333.9,0.40,0.60
0.857,0.143
0.00123,0.793
332.34,0.60,0.40
0.877,0.123
0.00128,0.637
330.73,0.80,0.2
0.898,0.102
0.00131,0.530
```

OUTPUT DATA

```
   X1        Differential latent heat of vaporization, kJ/kg

   0.50                         1786.6
   0.10                         1547.1
   0.15                         1385.5
   0.25                         1109.5
   0.40                          872.3
   0.60                          692.1
   0.80                          584.1

C*****************************************************************
```

SURFACE TENSION

```
C     This  program calculates surface tension of mixture
C     Reference: Rice & Teja(1982)
C
      COMMON/B1/FRAC(3),TCRIT(3),VCRIT(3),ZCRIT(3),OMEGA(3),WM(3)
      COMMON/B2/ICOMP(3)
      COMMON/B3/A(3),B(3)
      OPEN(UNIT=1,FILE='AMEERS.DAT')
      OPEN(UNIT=2,FILE='AMEERS.OUT')
      OPEN(UNIT=3,FILE='AMEERS1.OUT')
      DO 500 NO=1,4
      WRITE(3,*)
      READ(1,*)N
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(A(I),I=1,N)
      READ(1,*)(B(I),I=1,N)
      READ(1,*)N1
      DO 100 K=1,N1
      KCOMP=1
      II=ICOMP(KCOMP)
      JJ=ICOMP(KCOMP+1)
      READ(1,*)TBP,(FRAC(I),I=1,N)
      CALL PSEUDO(N,TCM,VCM,WMCM,ZCM)
      E=(VCM*WMCM*1000.0)**(2.0/3.0)/TCM
      TEMP=TBP
      CALL SURFTP(II,TEMP,SURF)
      EI=(VCRIT(II)*WM(II)*1000.0)**(2.0/3.0)/TCRIT(II)
      SIEI=EI*SURF
      SPSUE=SIEI
      WRITE(2,51)E,EI,SIEI
51    FORMAT(3F12.6)
      IF(N.EQ.1) GO TO 99
      OMI=OMEGA(II)
20    CONTINUE
      Z2=0.0
      Z3=0.0
      DO 30 LL=1,KCOMP+1
      M=ICOMP(LL)
      Z2=Z2+FRAC(LL)*OMEGA(M)
      Z3=Z3+FRAC(LL)
30    CONTINUE
      OMPSU=Z2/Z3
      JJ=ICOMP(KCOMP+1)
      TEMP=TBP
      CALL SURFTP(JJ,TEMP,SURF)
      EJ=(VCRIT(JJ)*WM(JJ)*1000.0)**(2.0/3.0)/TCRIT(JJ)
      SJEJ=EJ*SURF
```

```
      WRITE(2,52)EJ,SJEJ
52    FORMAT(2F12.6)
      OMJ=OMEGA(JJ)
      SPSUE=SIEI+((OMPSU-OMI)/(OMJ-OMI))*(SJEJ-SIEI)
      OMI=OMPSU
      SIEI=SPSUE
      KCOMP=KCOMP+1
      IF(KCOMP.LT.N) GO TO 20
99    CONTINUE
      SIGMA=SPSUE/E
      WRITE(3,76)FRAC(1),SIGMA
76    FORMAT(2F12.4)
      WRITE(*,*)K,FRAC(1),SIGMA
100   CONTINUE
500   CONTINUE
      STOP
      END


C     This  program calculates surface tension of pure liquids
C     Reference: Jasper(1972)
C
      SUBROUTINE  SURFTP(I,TEMP,SURF)
      COMMON/B3/A(3),B(3)
      T=TEMP-273.15
      SURF=(A(I)-B(I)*T)/1000.0
      WRITE(2,66)SURF,T
66    FORMAT(2F10.5)
      RETURN
      END
```

INPUT DATA,(ACETONE-WATER)

```
2
1,2
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.304,0.344
58.08,18.02
26.26,75.83
0.1120,0.1477
7
349.4,0.05,0.95
342.30,0.1,0.9
338.49,0.15,0.85
335.88,0.25,0.75
333.9,0.40,0.60
332.34,0.60,0.40
330.73,0.80,0.2
```

OUTPUT DATA

```
   X1              Surface tension, N/m

   0.50                  0.05855
   0.10                  0.05469
   0.15                  0.05091
   0.25                  0.04423
   0.40                  0.03649
   0.60                  0.02911
   0.80                  0.02389
```

```
C*********************************************************************
```

THERMAL CONDUCTIVITY

```
C     This  program  calculates  liquid thermal conductivity of
C     mixture
C     Reference: Li(1976)

      DIMENSION ICOMP(3),FRAC(3),PCRIT(3),ZRA(3),VCRIT(3),OMEGA(3)
      DIMENSION ZCRIT(3)
      COMMON/B1/A(3),B(3),C(3),B1(3),ASTAR(3),BP(3),TCRIT(3),WM(3)
      OPEN(UNIT=1,FILE='AMEERK.DAT')
      OPEN(UNIT=2,FILE='AMEERK.OUT')
      OPEN(UNIT=3,FILE='AMEERK1.OUT')
      DO 500 NO=1,4
      WRITE(3,*)
      READ(1,*)N
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(A(I),I=1,N)
      READ(1,*)(B(I),I=1,N)
      READ(1,*)(C(I),I=1,N)
      READ(1,*)(B1(I),I=1,N)
      READ(1,*)(ASTAR(I),I=1,N)
      READ(1,*)(BP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(PCRIT(I),I=1,N)
      READ(1,*)(ZRA(I),I=1,N)
      READ(1,*)N1
      DO 100 L=1,N1
      READ(1,*)TBP,(FRAC(I),I=1,N)
      KCOMP=1
      II=ICOMP(KCOMP)
      JJ=ICOMP(KCOMP+1)
      CONDM=0.0
      TEMP=TBP
      IF(N.GT.1) GO TO 10
      II=ICOMP(1)
      CALL COND(II,TEMP,CONDM)
```

```
      GO TO 99
 10   CONTINUE
      VTOT=0.0
      DO 15 I=1,N
      II=ICOMP(I)
      TRI=TBP/TCRIT(II)
      WRITE(2,21)I,II,TRI
 21   FORMAT(2I2,3X,F10.5)
      RU=0.08314
      VL=(1.0+(1.0-TRI)**(2.0/7.0))
      WRITE(2,22)VL
 22   FORMAT(F9.5)
      VL=(RU*TCRIT(II)/WM(II)/PCRIT(II))*ZRA(II)**VL
      WRITE(2,23)I,TCRIT(I),VL
 23   FORMAT(I2,3X,2F10.5)
      VTOT=VTOT+FRAC(I)*VL*WM(II)/1000.0
      WRITE(2,24)VTOT,TRI,VL
 24   FORMAT(3F10.6)
 15   CONTINUE
      DO 40 I=1,N
      DO 30 J=1,N
      II=ICOMP(I)
      TRI=TBP/TCRIT(II)
      VLI=(1.0+(1.0-TRI)**(2.0/7.0))
      VLI=(RU*TCRIT(II)/WM(II)/PCRIT(II))*ZRA(II)**VLI
      PHII=(FRAC(I)*VLI*WM(II)/1000.0)/VTOT
      WRITE(2,26)I,J,II,PHII,TRI,VLI
 26   FORMAT(3I2,2X,3F10.5)
      CALL COND(II,TEMP,CONDI)
      JJ=ICOMP(J)
      TRJ=TBP/TCRIT(JJ)
      VLJ=(1.0+(1.0-TRJ)**(2.0/7.0))
      VLJ=(RU*TCRIT(JJ)/WM(JJ)/PCRIT(JJ))*ZRA(JJ)**VLJ
      PHIJ=(FRAC(J)*VLJ*WM(JJ)/1000.0)/VTOT
      WRITE(2,27)J,JJ,CONDI,TRJ,VLJ,PHIJ
 27   FORMAT(2I2,4F10.5)
      CALL COND(JJ,TEMP,CONDJ)
      CONDIJ=2.0/(1.0/CONDI+1.0/CONDJ)
      CONDM=CONDM+PHII*PHIJ*CONDIJ
      WRITE(2,28)CONDJ,CONDIJ,CONDM
 28   FORMAT(3F10.4)
 30   CONTINUE
 40   CONTINUE
      WRITE(*,*)L,FRAC(1),CONDM
      WRITE(3,29)FRAC(1),CONDM
 29   FORMAT(2F12.4)
 100  CONTINUE
 99   CONTINUE
 500  CONTINUE
      STOP
      END

C     This  program  calculates liquid thermal  conductivity of
C     pure components
```

```
C     Reference: Latini(1977)
C
      SUBROUTINE COND(I,TEMP,CONDP)
      COMMON/B1/A(3),B(3),C(3),B1(3),ASTAR(3),BP(3),TCRIT(3),WM(3)
      A1=1.2
      IF(ASTAR(I).GT.0.1) A1=0.0
      WRITE(2,32)ASTAR(I)
32    FORMAT(F10.4)
      C1=0.167
      IF(ASTAR(I).GT.0.1) C1=-0.167
      WRITE(2,33)ASTAR(I)
33    FORMAT(F10.4)
      CONDP=0.0
      CONDP=A(I)+B(I)*TEMP+C(I)*TEMP**2
      WRITE(2,61)CONDP,A(I),B(I),C(I)
61    FORMAT(4F10.5)
      IF(CONDP.GT.0.)GO TO 99
      CONDP=ASTAR(I)*BP(I)**A1/(WM(I)**B1(I)*TCRIT(I)**C1)
      WRITE(2,34)CONDP,A1,B1,C1
34    FORMAT(4F11.5)
      TR=TEMP/TCRIT(I)
      WRITE(2,35)TR,CONDP,TEMP
35    FORMAT(3F10.4)
      CONDP=CONDP*(1.0-TR)**0.38/(TR**(1.0/6.0))
      WRITE(2,36)CONDP
36    FORMAT(F10.4)
99    CONTINUE
      RETURN
      END
```

INPUT DATA,(ACETONE-WATER)

```
2
1,2
0.0,-3.838E-01
0.0,5.254E-03
0.0,-6.369E-06
0.5,1.0
3.83E-03,4.94E-01
329.2,373.3
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.304,0.344
58.08,18.02
47.61,221.29
0.24494,0.2338
7
349.4,0.05,0.95
342.30,0.1,0.9
338.49,0.15,0.85
335.88,0.25,0.75
333.9,0.40,0.60
```

```
332.34,0.60,0.40
330.73,0.80,0.2
```

```
OUTPUT DATA

   X1       Thermal conductivity, W/(m K)

   0.50              0.5131
   0.10              0.4101
   0.15              0.3427
   0.25              0.2614
   0.40              0.2004
   0.60              0.1628
   0.80              0.1447

C******************************************************************
```

LIQUID SPECIFIC HEAT

```
C     This  program  calculates specific heat of liquid mixture
C     Reference: Teja(1983)
C
      DIMENSION PCRIT(5),ZRA(5)
      COMMON/B1/FRAC(5),TCRIT(5),VCRIT(5),ZCRIT(5),OMEGA(5),WM(5)
      COMMON/B2/ICOMP(5)
      COMMON/B3/A(5),B(5),C(5),D(5)
      OPEN(UNIT=1,FILE='AMEERCP.DAT')
      OPEN(UNIT=2,FILE='AMEERCP.OUT')
      OPEN(UNIT=3,FILE='AMEERCP1.OUT')
      DO 500 NO=1,4
      READ(1,*)N
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(PCRIT(I),I=1,N)
      READ(1,*)(ZRA(I),I=1,N)
      READ(1,*)(A(I),I=1,N)
      READ(1,*)(B(I),I=1,N)
      READ(1,*)(C(I),I=1,N)
      READ(1,*)(D(I),I=1,N)
      READ(1,*)N1
      DO 100 M=1,N1
      READ(1,*)TBP,(FRAC(I),I=1,N)
      CALL PSEUDO(N,TCM,VCM,WMCM,ZCM)
      RU=0.08314
      CP=0.0
      KCOMP=1
      II=ICOMP(KCOMP)
      CALL PSEUDO(N,TCM,VCM,WMCM,ZCM)
      R=RU*100.0/WMCM
```

```
      TEMP=TBP
      CALL CPP(II,TEMP,CPI)
      RI=RU*100.0/WM(II)
      CIRI=CPI/RI
      CPSUR=CIRI
      IF(N.EQ.1) GO TO 99
      OMI=OMEGA(II)
 20   CONTINUE
      Z2=0.0
      Z3=0.0
      DO 30 LL=1,KCOMP+1
      NN=ICOMP(LL)
      Z2=Z2+FRAC(LL)*OMEGA(NN)
      Z3=Z3+FRAC(LL)
 30   CONTINUE
      OMPSU=Z2/Z3
      JJ=ICOMP(KCOMP+1)
      TEMP=TBP
      CALL CPP(JJ,TEMP,CPJ)
      RJ=RU*100.0/WM(JJ)
      CJRJ=CPJ/RJ
      OMJ=OMEGA(JJ)
      CPSUR=CIRI+((OMPSU-OMI)/(OMJ-OMI))*(CJRJ-CIRI)
      OMI=OMPSU
      CIRI=CPSUR
      KCOMP=KCOMP+1
      IF(KCOMP.LT.N) GO TO 20
99    CONTINUE
      CP=CPSUR*R
      WRITE(3,51)FRAC(1),CP
51    FORMAT(2F12.4)
      WRITE(*,*)M,FRAC(1),CP
100   CONTINUE
500   CONTINUE
      STOP
      END


C     This program calculates specific heat of pure liquid
C     Reference: Miller et al.(1976), and Perry et al.(1984)
C
      SUBROUTINE CPP(I,T,CP)
      COMMON/B3/A(5),B(5),C(5),D(5)
      CP=A(I)+B(I)*T+C(I)*T*T+D(I)*T*T*T
      WRITE(2,52)CP
52    FORMAT(F10.5)
      RETURN
      END
```

INPUT DATA,(ACETONE-WATER)

```
2
1,2
```

```
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.304,0.344
58.08,18.02
47.61,221.29
0.24494,0.2338
3.339,2.822
-1.031E-02,1.183E-02
2.154E-05,-3.504E-05
0.0,3.60E-08
7
349.4,0.05,0.95
342.30,0.1,0.9
338.49,0.15,0.85
335.88,0.25,0.75
333.9,0.40,0.60
332.34,0.60,0.40
330.73,0.80,0.2
```

OUTPUT DATA

```
   X1          Liquid specific heat kJ/(kg K)

   0.50                    3.946
   0.10                    3.715
   0.15                    3.522
   0.25                    3.222
   0.40                    2.904
   0.60                    2.619
   0.80                    2.423
```

```
C*****************************************************************
```

VISCOSITY

```
C     This  program  calculates viscosity of liquid mixture
C     Reference: Teja & Rice(1980)
C
      COMMON/B1/FRAC(3),TCRIT(3),VCRIT(3),ZCRIT(3),OMEGA(3),WM(3)
      COMMON/B2/ICOMP(3)
      COMMON/B3/A(3),B(3),C(3),D(3)
      OPEN(UNIT=1,FILE='AMEERV.DAT')
      OPEN(UNIT=2,FILE='AMEERV.OUT')
      OPEN(UNIT=3,FILE='AMEERV1.OUT')
      DO 500 NO=1,4
      WRITE(3,*)
      READ(1,*)N
      READ(1,*)(ICOMP(I),I=1,N)
      READ(1,*)(TCRIT(I),I=1,N)
      READ(1,*)(VCRIT(I),I=1,N)
      READ(1,*)(ZCRIT(I),I=1,N)
      READ(1,*)(OMEGA(I),I=1,N)
```

```
      READ(1,*)(WM(I),I=1,N)
      READ(1,*)(A(I),I=1,N)
      READ(1,*)(B(I),I=1,N)
      READ(1,*)(C(I),I=1,N)
      READ(1,*)(D(I),I=1,N)
      READ(1,*)N1
      DO 100 L=1,N1
      KCOMP=1
      II=ICOMP(KCOMP)
      READ(1,*)TBP,(FRAC(I),I=1,N)
      CALL PSEUDO(N,TCM,VCM,WMCM,ZCM)
      E=WMCM**(-0.5)*(VCM*WMCM*1000.0)**(2.0/3.0)*(TCM**(-0.5))
      T=TBP*TCRIT(II)/TCM
      CALL VISCP(II,T,VISCO)
      EI=WM(II)**(-0.5)*(VCRIT(II)*WM(II)*1000.0)**(2.0/3.0)*
     1(TCRIT(II)**(-0.5))
      VIEI=EI*VISCO*1000.0
      VPSUE=VIEI
      WRITE(2,76)II,EI,VPSUE,VISCO
76    FORMAT(I2,3F10.6)
      IF(N.EQ.1) GO TO 99
      OMI=OMEGA(II)
20    CONTINUE
      Z2=0.0
      Z3=0.0
      DO 30 LL=1,KCOMP+1
      NN=ICOMP(LL)
      Z2=Z2+FRAC(LL)*OMEGA(NN)
      Z3=Z3+FRAC(LL)
30    CONTINUE
      OMPSU=Z2/Z3
      JJ=ICOMP(KCOMP+1)
      T=TBP*TCRIT(JJ)/TCM
      CALL VISCP(JJ,T,VISCO)
      EJ=WM(JJ)**(-0.5)*(VCRIT(JJ)*WM(JJ)*1000.0)**(2.0/3.0)*
     1(TCRIT(JJ)**(-0.5))
      VJEJ=EJ*VISCO*1000.0
      OMJ=OMEGA(JJ)
      VPSUE=EXP(LOG(VIEI)+((OMPSU-OMI)/(OMJ-OMI))*
     1(LOG(VJEJ)-LOG(VIEI)))
      WRITE(2,77)JJ,VIEI,VJEJ,VPSUE,VISCO
77    FORMAT(I2,4F10.6)
      OMI=OMPSU
      VIEI=VPSUE
      KCOMP=KCOMP+1
      IF(KCOMP.LT.N) GO TO 20
99    CONTINUE
      WRITE(2,78)E,VISMIX
78    FORMAT(2F10.6)
      VISMIX=VPSUE/(E)
      WRITE(*,*)L,FRAC(1),VISMIX
      WRITE(3,79)FRAC(1),VISMIX
79    FORMAT(2F14.5)
100   CONTINUE
```

```
500    CONTINUE
       STOP
       END


C      This   program  calculates  viscosity   of   pure  liquid
C      Reference: Yaws et al.(1976)
C
       SUBROUTINE VISCP(I,T,VISCO)
       COMMON/B3/A(3),B(3),C(3),D(3)
       VISCO=EXP(A(I)+B(I)/T+C(I)*T+D(I)*T*T)/1000.0
       WRITE(2,88)I,T,VISCO
88     FORMAT(I2,3X,2F12.6)
       RETURN
       END
```

INPUT DATA,(ACETONE-WATER)

```
2
1,2
508.15,647.3
0.003663,0.0031746
0.239,0.235
0.304,0.344
58.08,18.02
-4.033,-2.471E+01
8.456E+02,4.209E+03
0.0,4.527E-02
0.0,-3.376E-05
7
349.4,0.05,0.95
342.30,0.1,0.9
338.49,0.15,0.85
335.88,0.25,0.75
333.9,0.40,0.60
332.34,0.60,0.40
330.73,0.80,0.2
```

OUTPUT DATA

X_1	Viscosity, $N.s/m^2$
0.50	0.0003576
0.10	0.0003686
0.15	0.0003655
0.25	0.0003404
0.40	0.0003056
0.60	0.0002711
0.80	0.0002478

C**.

REFERENCES

Alam, S. S., 1972, "Nucleate Pool Boiling of Liquid Mixtures," Ph. D. Thesis, University of Roorkee, Roorkee

Alam, S. S., and Varshney, B. S., 1973a," Nucleate Pool Boiling of Liquids," Second National Heat and Mass Transfer Conference, IIT. Kanpur, Kanpur, B29-B36

Alam, S. S., and Varshney, B. S., 1973b," Pool Boiling of Liquid Mixtures," Second National Heat and Mass Transfer Conference, IIT. Kanpur, Kanpur, B37-B44

Alpay, E. H, and Balkan, F., 1989," Nucleate Pool Boiling Performance of Acetone-Ethanol and Methylene Chloride - Ethanol Binary Mixtures," Int. J. Heat Mass Transfer, Vol. 32, pp. 2403-2408

Bajorek, S. M., Lloyd, J. R., and Thome, J. R., 1989," Experimental Study of Multicomponent Pool Boiling on Smooth and Finned Surfaces," AIChE Symp. Series, Vol. 85, pp. 54-59

Bajorek, S. M., 1992, Energy System, Nuclear and Advanced Technology Division, Westinghouse Electric Corporation, Pennsylvania, Private Communication with Abdul-Ameer S. Kadhum, Chemical Engng Deptt., University of Roorkee, Roorkee, 12th Sept.

Behrens, D., and Eckermann, R., 1984," Chemistry Data Series, DECHEMA," Frankfurt, a. M., Vol.I, VLE Data Collection by Gmehling, J., Onken, U., Arlt, W., Grenzheuser, P., Weidlich, U., and Kolbe, B., 2nd ed., Elsevier, Amesterdam

Benjamin, J.E. and Westwater, J.W., 1961, " Bubble Growth in Nucleate Boiling of A Binary Mixture, "International Developments in Heat Transfer, Part II. International Heat Transfer Conference, University of Colorado, ASME, New York, pp. 212-218

Berenson, P.J., 1962," Experiments on Pool Boiling Heat Transfer," Int. J. Heat Mass Transfer, Vol. 5, pp.985-999

Bergles, A. E., 1988, " Some Perspectives on Enhanced Heat Transfer- Second Generation Heat Transfer Technology," J. of Heat Transfer, Vol. 110, pp. 1082-1096

Bonilla, C. F., and Perry, C. W., 1941," Heat Transmission to Boiling Binary Mixtures," Trans. AIChE., Vol. 37, pp. 685-787

Borishanskiy, V. M., and Minchenko, F. P., 1963," Fundamentals of Heat Transfer," Academic Press, New York, pp. 362

Bruijin, P. J., 1960," On the Asymptotic Growth Rate of Vapour Bubbles in Superheated Binary Liquid Mixtures," Physica, Vol. 6, pp. 252-264

Calus, W. F., and Leonidopoulos, D. J., 1974," Pool Boiling-Binary Liquid Mixtures," Int. J. Heat Mass Transfer, Vol. 17, pp. 249-256

Calus, W. F., and Rice, P., 1972," Pool Boiling-Binary Liquid Mixtures," Chem. Engng. Sci., Vol. 27, pp. 1687-1697

Charter, W. W., Megler, V. R., Chen, W. D., and Wang, Y. F., 1982," Atmospheric and Sub-Atmospheric Boiling of H_2O and $LiBr/H_2o$ Solutions," Int. J. of Refrigeration, Vol. 5, pp. 107-114,

Chen, j. G., Dong, L., and Zhang, H.- J., 1988," Boiling Heat Transfer on A Low-Finned Tube," Proc. Int. Symp. On Phase Change Heat Transfer, Editor: X. Mingdao, Chongging, P. R. China, pp. 129-134

Cichelli, M. T., and Bonilla, C. F., 1945," Heat Transfer to Liquids Boiling under Pressure," Trans. AIChE., Vol. 41, pp. 755-787

Cichelli, M. T., and Bonilla, C. F., 1946," Heat Transfer to Liquids Boiling under Pressure," Trans. AIChE., Vol. 42, pp. 411-412

Clark, H. B., Strenge.P. S., and Westwater, J. W., 1959," Active Sites for Nucleate Boiling," Chemical Engng. Progress Symp. Ser., Vol. 55, pp. 103-110

Cole, R., 1963," Bubble Frequencies and Departure Volumes at Subatmospheric Pressures," J. AIChE, Vol. 13, pp. 779-783

Cole, R., and Rohsenow, W. M., 1969," Correlation of Bubble Departure Diameter for Boiling of Saturated Liquids," AIChE Symp. Ser. Vol. 65, pp. 211-213

Cole, R., and Shulman, H. L., 1966," Bubble Growth Rates at High Jakob Numbers," Int. J. Heat Mass Transfer, Vol. 9, pp. 1377-1390

Cooper,M. G., 1984," Saturation Nucleate Pool Boiling - A Simple Correlation," Int. Chem. Engng. Symp. Ser. Vol. 86, pp. 785-793

Cooper, M. G., and Stone, C. R., 1981," Boiling of Binary Mixtures - Study of Individual Bubbles," Vol. 24, pp. 1937-1950

Cornwell, K., 1977," Naturally Formed Boiling Site Cavities," Letters Heat Mass Transfer, Vol. 4, pp. 63-72

Cryder, D. S. and Finalborgo, A. C., 1937," Heat Transmission from Metal Surfaces to Boiling Liquids: Effect of Temperature of the Liquid on Film Coefficient," Trans. AIChE, Vol. 33, pp. 346-362

Eddington, R. I., and Kenning, D. B. R., 1979," The Effect of Contact Angle on Bubble Nucleation," Int. J. Heat Mass Transfer, Vol. 22, pp. 1231-1236

Fisenko, V.V., Baranenko, V.I., Belov, L.A. and Korenevskiy, V.M., 1988," Effect of Dissolved Gas on Nucleate Boiling and Critical Heat Flux," Heat Transfer - Soviet Research, Vol. 20, pp. 294 -299

Florschuetz, L. W., and Rashid, A. R., 1970," Growth Rate of Free Vapour Bubbles in Binary Liquid Mixtures at Uniform Superheats," 4th International Heat Transfer Conference, Paris, Elsevier, Amsterdam, paper B7.3

Freshwater, D. C., and Pike, K. A., 1967," Vapour-Liquid Equilibrium Data for Systems of Acetone-Methanol-Isopropanol" J. of Chem. & Engng. Data, Vol. 12, pp. 179-182

Fritz, W., and Ende, W., 1935," Berechnung des Maximavolumens von Damfblasen," Phys. Z., Vol. 36, pp. 379-384

Gaertner, R. F., 1965," Photographic Study of Nucleate Pool Boiling on A Horizontal Surface," Trans. ASME, Ser. C, J. Heat Transfer, Vol. 87, pp. 17-29

Gilmour, C. H., 1958," Nucleate Boiling - A Correlation," Chem. Engng. Prog., Vol. 54, pp. 77-79

Gorenflo, D., Blein, P., Herres, G., Rott, W., Schomann, H., and Sokol, P., 1988," Heat Transfer at Pool Boiling of Mixtures with R22 and R114," Int. J. Refrig. Vol. 11, pp. 257-263

Gorenflo, D., and Fath, W., 1987," Pool Boiling Heat Transfer on Outside of Finned Tubes at High Saturation Pressure," Proc. 17th Int. Congr. Refrig., B:955-960

Grigor'ev, L. N., Khairullin, I. Kh., and Usmanov, A. G.,1968a ," An Experimental Study of the Critical Heat Flux in the Boiling of Binary Mixtures ," Int. Chem. Engng. Vol. 8, pp. 39-42

Grigor'ev, L. N., Sarkisyan, L. A., and Usmanov, A. G.,1968b ," An Experimental Study of Heat Transfer in Boiling of Three Component Mixtures ," Int. Chem. Engng. Vol. 8, pp. 76-78

Gropp, U., 1981, Verfahrenstechnik, Vol. 15, pp. 725-728

Gupta, J. M. L., 1989," Studies of Pool Boiling Heat Transfer from Heating Tube in Horizontal and Vertical Orientations," Ph. D. Thesis, University of Roorkee, Roorkee

Hahne, E., Chen Qiu-Rong, and Windisch, R., 1991," Pool Boiling Heat Transfer on Finned Tubes - An Experimental and Theoretical Study," Int. J. Heat Mass Transfer Vol. 34, pp. 2071-2079

Hahne, E., and Muller, J., 1983," Boiling on Finned Tubes and Finned Tube Bundle," Int. J. Heat Mass Transfer, Vol. 26, pp. 849-859

Happel, O., and Stephan, K., 1974," Heat Transfer from Nucleate to Film Boiling in Binary Mixtures," 5th Int. Heat Transfer Conference., Tokyo, AIChE, New York, Vol. 6, Paper B7.8,

Hatton, A. P., and Hall, I., 1966," Photographic Study of Boiling on Prepared Surfaces," Proc. 3rd Int. Heat Transfer Conference, Chicago, Vol. 4, pp. 24-37

Hinrichs, T., Henneecke, E., and Yasuda, H., 1981," The Effect of Plasma-Deposited Polymers on The Nucleate Boiling Behaviour of Copper Heat Transfer Surfaces," Int. J. Heat Mass Transfer, Vol. 24, pp. 1359-1368

Hirata, M., and Nagahama, K., 1975," Computer Aided Data Book of Vapour-Liquid Equilibria," Kodansha Limited Elsevier Scientific Publishing Company, Tokyo, Japan

Hsu, Y. Y., and Graham, R. W., 1976, " Transport Processes in Boiling and Two Phase Systems," Hemisphere Publication Co., Washington, pp. 3-46

Hui, T. O., and Thome, J. R., 1985," A Study of Binary Mixture Boiling: Boiling Site Density and Sub-Cooled Heat Transfer ," Int. J. Heat Mass Transfer, Vol. 28, pp. 919-928

Isshiki , N., and I. Nikai, 1973," Boiling of Binary Mixtures," Heat Transfer Jap. Res. Vol. 4, pp. 56-66

Ivey, H. J., 1967," Relationship Between Bubble Frequency, Departure Diameter and Rise Velocity in Nucleate Boiling," Int. J. Heat Mass Transfer, Vol. 10, pp. 1023-1040

Jamialahmadi, M., Blochl. R., and Muller-Steinhagen, 1991," Pool Boiling Heat Transfer to Saturated Water and Refrigerant-113," The Canadian Journal of Chemical Engineering, Vol. 69, pp. 746-754

Jasper, J. J., 1972," The Surface Tension of pure Liquid Compounds," J. Phys. Chem. Ref. Data, Vol.1, pp. 841

Jensen, M. K., and Hsu, J. -T.,1988," A Parametric Study of Boiling Heat Transfer in A Horizontal Tube Bundle," J. of Heat Transfer, Vol. 110, pp. 976-981

Jincina - Molazhin, L. N., and Kutateladze, S. S., 1950," On the Problem of Effect of Pressure on the Mechanism of Vaporization in Boiling Liquid," J. Technical Physics, Vol. 20, pp. 110-116

Jungnickel, H., Wassilew, P., and Kraus, W. E., 1980," Investigations on the Heat Transfer of Boiling Binary Refrigerant Mixtures," Int. J. Refrig. Vol. 3, pp. 129-133

Kadhum, Abdul-Ameer, S., Varshney, B .S., Mohanty, B., and Govinda Rajulu, K., 1993," Heat Transfer During Pool Boiling of Hydrocarbons on Two Horizontal Tubes Placed in A Vertical Plane," Proceedings of the Eighth ISME Conference on Mechanical Engineering, IIT Delhi, Delhi, Tata McGraw-Hill Publishers, pp. 75-80,

Kartsounes, G. T., 1975," A Study of the Effect of Surface Treatment on Pool Boiling Heat Transfer in Refrigerant-12," ASHRAE Trans., Vol. 81-Part 1, pp. 320-329

Keshock, E. G., and Siegel, R., 1964, " Forces Acting on Bubbles in Nucleate Boiling Under Normal and Reduced Gravity Conditions ", NASA TN-D-99

Kirby, D. B., and Westwater, J. W., 1965," Bubble and Vapour Behaviour on A Heated Horizontal Plate During Pool Boiling Near Burnout," Chem. Engng. Prog. Symp. Ser. Vol. 61, pp. 238-248

Kojima, K., Ochi, K., and Nakazawa, Y., 1969, " Relationship Between Liquid Activity Coefficient and Composition for Ternary Systems," Int. Chem. Engng. Vol.9, pp. 342-347

Kurihara, H. M., and Myers, J. E., 1960," The Effect of Superheat and Surface Roughness on Boiling Coefficients," AIChE, J., Vol. 6, pp. 83-91

Kutateladze, S. S., 1963," Fundamentals of Heat Transfer ," Academic Press, New York, pp. 362

Kumar, 1992," Studies of Nucleate Pool Boiling Heat Transfer from Horizontal Finned Tubes to Saturated Liquids, " Ph.D. Thesis, University of Roorkee, Roorkee

Latini, G., and Pacetti, M., 1977, Therm. Conduct., Vol.15, pp. 245-248

Lee, B. I., and Kesler, M. G., 1975," A Generalized Thermodynamic Correlation Based on Three-Parameter Corresponding States," AIChE J., Vol. 21, pp. 510-527

Levy, S., 1959," Generalized Correlation of Boiling Heat Transfer," Trans. ASME, Ser. C., J. Heat Transfer, Vol. 81, pp. 37-42

Li, C. C.,1976," Thermal Conductivity of Liquid Mixtures," AIChE J., Vol. 22, pp. 927-930

McNelly, M.J.,1953," A Correlation of Rates of Heat Transfer to Nucleate Boiling of Liquids," Journal of Imperial College Chem. Engng. Soc., Vol.7, pp. 18-34

Mehrotra, R. K., 1980," Heat Transfer in Nucleate Pool Boiling of Saturated Liquids," Ph.D. Thesis, University of Roorkee, Roorkee

Miller, J. W., McGinley, J. J., and Yaws, C. L., 1976 " Thermal Conductivities of Liquids, " Chem. Engng., Vol. 83, No.23, pp. 133-136

Miller, J. W., Schorr, G. R., and Yaws, C. L., 1976 " Heat Capacity of Liquids, " Chem. Engng., Vol. 83, No. 25, 129-132

Nagel ,O., Hegner, B., and Karten, H., 1981," Criteria for the Selection and Design of Gas/Liquid Reactors," Int. Chem. Engng. Vol. 21, pp. 161-172

Nishikawa, K., and Yamagata, K., 1960," On the Correlation of Boiling Heat Transfer," Int. J. Heat Mass Transfer, Vol. 1, pp. 219

Nukiyama, S., 1934," The Maximum and Minimum Values of the Heat Transmitted from Metal to Boiling Water under Atmospheric Pressure," J. Japan Soc. Mech. Engrs., Vol. 37, pp. 367-374; Translation, Int. J. Heat Mass Transfer, Vol. 9, pp. 1419-1433(1986)

Palen, J. W., and Small,W., 1964," A New Way to Design Kettle and Internal Reboilers," Hydrocarbon Processing, Vol. 43, pp. 199-208

Palen, J. W., Taborek, J., and Yilmaz, S., 1986," Comments to the Application of Enhanced Boiling Surfaces in Tube Bundles," In Heat Exchanger Sourcebook, Ed. Palen, J. W., Washington D.C: Hemisphere, pp. 663-673

Palen, J. W., and Yang, C. C., 1983," Circulation Boiling Model for Analysis of Kettle and Internal Reboiler Performance," ASME HTD-Vol. 27, pp. 55-61

Palen, J. W., Yarden, A., and Taborek, J., 1972," Characteristics of Boiling Outside Large-Scale Horizontal Multitube Bundles ," AIChE Symp. Series, Vol. 68, pp. 50-61

Pandey, S. K., 1982," Nucleate Pool Boiling of Liquids and Their Mixtures at Sub-Atmospheric Pressures , Ph.D. Thesis, University of Roorkee, Roorkee

Pandey, S. K., Sharma, P. R., and Varshney, B. S., 1986," Heat Transfer Studies During Nucleate Pool Boiling of Binary Liquid Mixtures," Proc. 8th Int. Heat Transfer Conference, San Francisco, Vol. 4. pp. 2087-2092

Patel, P.,and Teja,A. S., 1982," A New Cubic Equation for Fluids and Fluid Mixtures", Chem. Engng. Sci., Vol. 37, pp. 463-473

Perry, R. H., Green, D. W., and Malonay, J.O., 1984," Chemical Engineering Handbook," 6th Ed., McGraw-Hill, New York

Pitzer, , K. S., Lippman, D.A., Curl, R. F., Huggins, C. M., and Peterson, D. E., 1955," The Volumetric and Thermodynamic Properties of Fluids, II. Compressibility Factor, Vapour Pressure and Entropy of Vaporization," J.. Am. Chem. Soc., Vol. 77, pp.3433

Plesset, M. S., and Zwick, S. A., 1954," The Growth of Vapour Bubbles in Superheated Liquids, " J. Appl. Phys., Vol. 25, pp. 493-500

Rabas, T. J., Taborek, J., 1986," A Survey of Turbulent Forced Convection Heat Transfer and Pressure Drop Characteristics of Low-Finned Tube Banks in Cross-Flow," ASME. HTD- Vol. 64, pp. 13-23

Reid, R. C., Prausnitz, J. M., and Poling, B.E., 1987, " The Properties of Gases and Liquids", 4th ed., McGraw-Hill Book Co.New York

Rice,P., and Teja, A. S., 1982," A Generalized Corresponding-States Method for the Prediction of Surface Tension of Pure Liquids and Liquid Mixtures," Journal of Colloid and Interface Surface, Vol. 86, pp. 158-163

Rohsenow, W. M., 1952," A Method of Correlating Heat Transfer Data for Surface Boiling of Liquids," ASME Trans., Vol. 74, pp. 969-975

Sardesai, R. G., Palen, J. W., and Thome, J. R., 1986," Nucleate Pool Boiling of Hydrocarbon Mixtures," AIChE Winter National Meeting, Miami Beach, paper No. 127a, November, pp. 2-7

Sardesai, R. G., Shock, R. W. A., and Butterworth, D., 1982," Heat and Mass Transfer Multicomponent Condensation and Boiling," Heat Transfer Engng., Vol. 3, pp. 104-114

Schlunder, E. U., 1982, " Uber den Warmeubergang bei der Blasenverdampfung von Gemischen, Verfahenstechnik, Vol. 16, pp. 692-698 and " Heat Transfer in Nucleate Boiling of Mixtures," Int. Chem. Engng., Vol. 23, pp. 589-599(1983)

Schlunder, E. U., 1983," Heat Exchanger Design Book, Vol. 5, " Physical Properties," Hemisphere Publishing Co., New York

Schlunder, E. U., 1986," Heat transfer in Nucleate Boiling of Mixtures," 8th Int. Heat Transfer Conference, San Francisco, Vol. 4, pp. 2073-2079

Schultz, R. R., and Cole, R., 1979, " Uncertainty Analysis in Boiling Nucleation," AIChE. Symposium Series. Vol.75, pp.32-39

Scriven, L. E., 1959," On the Dynamics of Phase Growth," Chem. Engng. Sci., Vol. 10, pp. 1-13

Shakir, S., 1987 ," Boiling Incipience and Heat Transfer on Smooth and Enhanced Surfaces," Ph.D. Thesis, Michigan State University, Michigan

Shakir,S., and Thome, J. R., 1986," Boiling Nucleation of Mixtures on Smooth and Enhanced Surfaces," 8th Int. Heat Transfer Conference, San Francisco, Vol. 4, pp. 2081-2086

Shakir,S., Thome, J. R., and Lioyd, J. R.,1985," Boiling of Methanol-Water Mixtures on Smooth and Enhanced Surfaces," Multiphase Flow and Heat Transfer, ASME-HTD Vol. 47,pp. 1-6.

Sharma, P. R., 1977," Heat Transfer Studies of Pool Boiling of Liquids, " Ph.D. Thesis, University of Roorkee, Roorkee

Shock, R. A. W., 1982, " Boiling in Multicomponet Fluids," in Multiphase Science and Technology, Eds., Hewitt, G. F., Delhaye, J. M., and Zuber,N., Hemisphere Publishing Corporation, New York, Vol. 1, pp. 281-386

Skinner, L. A., and Bankoff, S. G., 1964a," Dynamics of Vapour Bubbles in Spherically Symmetric Fields of General Variation," Phys. Fluids, Vol. 7, pp. 1-6

Skinner, L. A., and Bankoff, S. G., 1964b," Dynamics of Vapour Bubbles in Binary Liquids with Spherically Symmetric Initial Conditions," Phys. Fluids, Vol. 7, pp. 643-648

Spencer, C. F., and Danner, R. P., 1972, J. Chem. Engng. Data, Vol. 17, pp. 236-241

Staniszewski, B. E., 1959, " Nucleate Boiling Bubble Growth and Departure," Technical Report No. 16. Div. Sponsored Research, M.I.T., Cambridge, Mass.

Stephan, K.,1981," Heat Transfer with Natural Convection Boiling in Multicomponent Mixtures," in Heat Exchangers : Thermal-Hydraulic Fundamentals and Design, Eds. Kakas, S., Bergles, A. E., and Mayinger, F., Hemisphere Publishing Corp. Washington D.C., pp. 315-356

Stephan, K., and Abdelsalam, M., 1980," Heat Transfer Correlations for Natural Convection Boiling," Int. J. Heat Mass Transfer, Vol. 23, pp.73-87

Stephan, K., and Preusser, P., 1978," Heat Transfer in Natural Convection of Polynary Mixtures," Proceedings of 6th Int. Heat Transfer Conference, Toronto, Hemisphere Publishing Corp. Washington, DC., Vol.1, pp. 187-192

Stephan, K., and Preusser, P., 1979," Heat Transfer and Critical Heat Flux in Pool Boiling of Binary and Ternary Mixtures," Ger. Chem. Engng. Vol. 2, pp. 161-169

Tarrad, A. H., and Burnside, B. M., 1991," Instability in Pool Boiling of A Wide Boiling Mixture on A Horizontal Tube," Int. J. Heat Mass Transfer, Vol. 34, pp. 2797-2803

Teja, A. S., 1980," A Corresponding States Equation for Saturated Liquid Densities," AIChE. J. Vol. 26, pp. 337-341

Teja, A. S., 1983, " Simple Method for the Calculation of Heat Capacities of Liquid Mixtures," J. Chem. Engng. Data, Vol. 28, pp.83-85

Teja, A. S., and Rice, P., 1981, " Generalized Corresponding States Method for the Viscosities of Liquid Mixtures," Ind. Engng. Chem. Fundam., Vol. 20, pp. 77-81

Thome, J. R., 1981," Nucleate Pool Boiling of Binary Liquids: An Analytical Equation," AIChE Symp. Series, Vol. 77, pp. 238-250

Thome, J. R., 1983," Prediction of Binary Mixture Boiling Heat Transfer Coefficients Using Only Phase Equilibria Data," Int. J. Heat Mass Transfer, Vol. 26, pp. 965-974

Thome, J. R., 1989," Prediction of the Mixture Effect on Boiling in Vertical Thermosyphon Reboiler ," Heat Transfer Engng, Vol. 10, No. 2, pp. 29-38

Thome, J. R., 1990," Enhanced Boiling Heat Transfer,"Hemisphere Publication Corp., Washington D.C.

Thome, J. R., Davey, G., 1981," Bubble Growth Rates in Liquid Nitrogen, Argon and Their Mixtures," Int. J. Heat Mass Transfer, Vol. 24, pp. 89-97

Thome, J. R., and Shakir, S.,1987, " A New Correlation for Nucleate Pool Boiling of Aqueous Mixtures," AIChE Symp. Series, Vol. 83, pp. 46-51

Thome,J. R., and Shock, R. A. W.,1984, " Boiling of Multicomponent Mixtures," Advances in Heat Transfer, Eds. Hartnett, J. P., and Irivine, T. J., Academic Press, Orland, Vol. 16, pp. 59-156

Tolubinskiy, V. I., and Ostrovskiy, J. N., 1966a, " Mechanism of Vapour Formation and Rate of Heat Transfer During Boiling of Binary Solutions," Akad. Nauk, Ukr. USSR Reshul Mezhvendom, pp. 7-16

Tolubinskiy, V. I., and Ostrovskiy, J. N., 1966b, " On the Mechanism of Boiling Heat Transfer (Vapour Bubble Growth Rates in the Process of Boiling of Liquids, Solutions, and Binary Mixtures), Int. J. Heat Mass Transfer,Vol. 9, pp. 1464-1470

Tolubinskiy, V. I., and Ostrovskiy, J. N., 1969, " Mechanism of Heat Transfer in Boiling of Binary Mixtures," Heat Transfer Sov. Res., Vol. 1, No. 6, pp. 6-11

Tolubinskiy, V. I., Ostrovskiy, J. N., and Kriveshko, A.A., 1970, " Heat Transfer to Boiling Water-Glycerine Mixtures," Heat Transfer Sov. Res., Vol. 2, No. 1, pp. 22-24

Tyn, M. T., and Calus, W. F., 1975," Temperature and Concentration Dependence of Mutual Diffusion Coefficients of Some Binary Liquids Systems," J. Chem. and Engng. Data, Vol. 20. pp. 310-316

Uhlig E, and Thome J R, 1985, " Boiling of Acetone-Water Mixtures on Smooth and Enhanced Surfaces," Advances in Enhanced Heat Transfer, ASME HTD-Vol. 43, pp. 49-56

Valent, V., and Afgan, N. H., 1973," Bubble Growth Rate and Boiling Heat Transfer in Pool Boiling of Ethyl-Alcohol Mixtures," Warme-und Stoffubertragung, Vol. 6, pp.235-240

van Stralen, S. J. D., 1959," Heat Transfer to Boiling Binary Liquid Mixtures, Part 1, Br. Chem. Engng., Vol. 4, pp. 8-17

van Stralen, S. J. D., and Cole, R., 1979," Boiling Phenomena," Washington: Hemisphere Publishing Corporation, New York: McGraw-Hill Book Company, Vol. 1,pp. 219-27

van Wijk, W. R., vos, A. S., and van Stralen, S. J. D., 1956, " Heat Transfer to Boiling Binary Liquid Mixtures, " Chem. Engng. Sci., Vol. 5, pp. 68-80

Walas, S. M., 1985," Phase Equilibrium in Chemical Engineering," ButterWorth Publishers

Wall, K. W., and Park, Jr. E. L., 1978," Nucleate Boiling of n-Pentane, n-Hexane and Several Mixtures of the Two from Various Tube Arrays," Int. J. Heat Mass Transfer, Vol. 21, pp. 73-75

Webb, R. L., 1981, The Evolution of Enhanced Surface Geometries for Nucleate Boiling, " Heat Transfer Engng., Vol. 2, pp. 46-59

Webb, R. L., and Apparao, T.V.V.R., 1990," Performance of Flooded Refrigerant Evaporators with Enhanced Tubes," Heat Transfer Engineering, Vol. 11, pp. 29-43

Webb, R. L., Choi, K-D., and Apparao, T. R., 1989," A Theoretical Model for Prediction of Heat Load in Flooded Refrigerant Evaporators," ASHRAE Trans., Vol.95, pp.326-338

Webb, R. L., and Pais, C., 1992," Nucleate Pool Boiling Data for Five Refrigerants on Plain, Integral-fin and Enhanced Tube Geometries," Int. J. Heat Mass Transfer, Vol.35,pp. 1893-1904

Westwater, J. W., 1973," Development of Extended Surfaces for Use in Boiling Liquids," AIChE Symp. Series, Vol.69, pp. 1-9

Wiebe, J. R., and Judd, R. L., 1971," Superheat Layer Thickness Measurements in Saturated and Subcooled Nucleate Boiling, Trans. ASME, Ser. C, J. Heat Transfer, pp. 455-461

Wilke, C.R., and Chang, P, 1955," Correlation of Diffusion Coefficients in Dilute Solutions," AIChE J., Vol.4 , pp. 264-270

Yaws, C. L., Miller, J. W., Shah, P. N., Schorr, G. R., and Patel, D. M., 1976," Correlation Constants for Chemical Compounds," Chem. Engng. Vol. 83, pp. 153

Yilmaz, S., and Palen, J. W., 1984," Performance of Finned Tube Reboiler in Hydrocarbons Service," ASME, paper 84-HT-91

Yilmaz, S., and Westwater, J., 1981," Effect of Commercial Enhanced Surfaces on the Boiling Heat Transfer Curve," Advances in Enhanced Heat Transfer, ASME, HTD-Vol.18, pp.73-91

Zuber, N., 1959," Hydrodynamic Aspects of Boiling Heat Transfer," U. S. Atomic Energy Commission Rep. AECU-4439

Zuber, N., 1963," Nucleate Boiling, The Region of Isolated Bubbles and Similarity with Natural Convection, " Int. J. Heat Mass Transfer, Vol. 6, pp. 53-78

PUBLISHED/COMMUNICATED PAPERS

1. Kadhum, A. S., Rajulu, K. G., Varshney, B. S., and Mohanty, B., Performance of Plain and Integral-Fin Tubes During Pool Boiling of Non-Azeotropic Hydrocarbons Mixtures (Part I), Journal of Energy, Heat and Mass Transfer, 15, 239-246(1993).

2. Kadhum, A. S., Varshney, B. S., and Mohanty, B., Characterstics of Nucleate Pool Boiling of Aqueous Mixtures, ASME Winter Annual Meeting, New Orleans, USA., November,(1993).

3. Kadhum, A. S., Varshney, B. S., and Mohanty, B., A Correlation for Nucleate Pool Boiling of Binary Liquid Mixtures on Integral-Fin Tube, AIChE Winter Meeting, St. Louis, USA, (1993).

4. Kadhum, A. S., Varshney, B. S., and Mohanty, B., Boiling Heat Transfer of Acetone and Water Mixtures on Plain and Integral-fin Tubes Geometries, ASHRAE Transaction, Vol. 100, Part 1, No.-94-2-5, (1994).

5. Kadhum, A. S., Varshney, B. S., and Mohanty, B., Boiling Heat Transfer from Plain and Integral-Fin Tubes to Pool of Ternary Mixtures, 10th International Heat Transfer Conference to be held in Brighton, UK, August, (1994).(Accepted).

6. Kadhum, A. S., Varshney, B. S., and Mohanty, B., Correlations for Nucleate Pool Boiling of Binary Liquid Mixtures,Journal of Energy, Heat and Mass Transfer(To be published, 1994).