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# HEAT TRANSFER STUDIES IN POOL BOILING OF LIQUIDS

A THESIS
*submitted in fulfilment of the*
*requirement for the award of the degree*
*of*
DOCTOR OF PHILOSOPHY
*in*
CHEMICAL ENGINEERING

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# Certificate

*Certified that the thesis entitled* "**Heat Transfer Studies in Pool Boiling of Liquids**" *which is being submitted by* **Sri PARASHU RAM SHARMA** *in fulfilment of the requirements for the award of* **DOCTOR OF PHILOSOPHY** *in Chemical Engineering of the University of Roorkee, Roorkee, is a record of candidate's own work carried out by him under my supervision and guidance. The matter embodied in this thesis has not been submitted for the award of any other degree.*

*This is further certified that the candidate has worked for a period of three years and three months from January 1974 for preparing this thesis.*

*April* 12, *1977*

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# ABSTRACT

An investigation of heat transfer in nucleate pool boiling for atmospheric and subatmospheric pressures has been carried out both analytically and experimentally.

Traditionally, the study of heat transfer in nucleate pool boiling are considered empirically, without regard for bubble dynamics. But in the present investigation, following the fact that the heat transfer in nucleate boiling is characterised by the induced turbulence due to; nucleation sites on the heating surface, bubble size and their emission frequency, a mathematical analysis has led to equation for predicting the values of absolute heat transfer coefficient, Eq. (5.22). This resultant equation relates heat transfer coefficient to the wall heat flux, system pressure and the pertinent physico-thermal properties of boiling fluids through the heating surface characteristics. But this equation is useful for calculating the absolute values of heat transfer coefficient, only if , heating surface characteristics are known as required for constant M and the values of nf for the determination of exponent a. Since the surface characteristics and the value of nf are extremely unpredictable for industrial surfaces and they differ from surface to surface, a considerable built-in difficulty is inherited in this equation. Therefore, it appears impossible to provide a

panacea for predicting the absolute values of heat transfer coefficient. However, for a given heating surface it is possible to determine the value of constant M and exponent a empirically which can be used for the calculation of absolute values of transfer coefficients. The value of exponent a has been evaluated as 0.3683, and the expression for M are represented by the set of equations, Eq. (6.3). Thus the equation for calculating the absolute values of heat transfer coefficient is as follows: while f is obtained from the respective equation, Eq. (5.15 ) or Eq. (5.19) for Jakob number less and greater than 100.

$$h = M \left[ \left\{ \left( \frac{\lambda^{2.5} \rho_v}{T_s^{1.5} \sigma} \right) \left( \frac{q}{k_\ell g} \right)^{0.3} \left( \frac{1}{C_\ell} \right)^{1.2} \right\}^{2.33} \left\{ f \right\} \right]^{0.3683} \qquad (6.4)$$

The constant M represents the combined effects of pressure and surface-liquid combination on boiling heat transfer.

A procedure has also been devised for the calculation of $h^*/h_1^*$ [cf. Eq(6.6)]. It has been found that the values of $h^*/h_1^*$ depend upon wall heat flux, system pressure and pertinent physico-thermal properties of boiling fluids. They do not depend on surface-liquid combinations. Hence this equation is useful to compare the data of different investigators obtained on differing surface-liquid combinations. It was found that this equation correlated the present data and those of Cryder and Finalborgo [5], Raben, Beaubouef and

Commerford [94] within a maximum deviation of ± 20 per cent. This equation also provides the facility for the computation of absolute values of heat transfer coefficient at subatmospheric pressures, without resort to experimentation, if the value of heat transfer coefficient at normal boiling point is available. Equation (6.6) is as follows:

$$\frac{h^*}{h_1^*} = 0.95831645(P/P_1)^{-0.3997825}\left[\left\{\left(\frac{\lambda}{\lambda_1}\right)^{2.5}\left(\frac{\rho_v}{\rho_{v1}}\right)\left(\frac{T_{s1}}{T_s}\right)^{1.5}\left(\frac{\sigma_1}{\sigma}\right)\left(\frac{k_{\ell 1}}{k_\ell}\right)^{0.3}\left(\frac{q}{q_1}\right)^{0.3}\left(\frac{C_{\ell 1}}{C_\ell}\right)^{1.2}\right\}^{2.33}\left\{\frac{f}{f_1}\right\}\right]^{0.3683} \tag{6.6}$$

Since the present study is for atmospheric and subatmospheric pressures, obviously the resultant equation, Eq.(6.6) might not correlate the boiling data for higher pressures.

A computer program was written and calculations were made to compute the heat transfer coefficients from the above resultant equations.

Apart from the analytical analysis the purpose of the investigation was also to obtain experimental data in order to verify the resultant equations from the analysis and to generate the new experimental data for subatmospheric pressures which are scanty in the literature. The experimental investigation involved the determination of heat transfer coefficient from 410 ASIS stainless steel heating surface to the boiling fluids:

distilled water, isopropanol, ethanol and methanol for the pressures ranging from 11.33 kN/m$^2$ to 98.44 kN/m$^2$ and heat flux ranging from 6870 W/m$^2$ to 41730 W/m$^2$.

The excellent consistency between the experimental data and the predicted values sufficiently proves that the present mathematical analysis based on the governing equations for nucleation sites, bubble growth, bubble size and bubble emission frequency provides an adequate procedure for heat transfer coefficient in nucleate pool boiling of fluids for the range of parameters investigated.

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# NOMENCLATURE

| | | |
|---|---|---|
| a | exponent, defined in Eq(5.20) | |
| A | heat transfer area | $m^2$ |
| C | specific heat | J/kg K |
| $c_o$ | dimensional constant, defined in Eq(5.1) | |
| $c_2$ | constant, defined in Eq(5.4a) | |
| $c_3$ | constant, defined in Eq(5.20) | |
| $C_4$ | constant, defined in Eq(6.2) | |
| $C_{sf}$ | surface-liquid combination factor | |
| d | diameter of the heating surface | m |
| D | diameter of bubble | m |
| $D_b$ | diameter of bubble at departure | m |
| f | bubble emission frequency | 1/s |
| g | acceleration due to gravity | $m/s^2$ |
| h | heat transfer coefficient | $W/m^2$ K |
| $\bar{h}$ | average heat transfer coefficient | $W/m^2$ K |
| $h^*$ | quotient at a pressure, $h/q^{0.7}$ | |
| $h^*_1$ | quotient at normal boiling point, $h/q^{0.7}$ | |
| k | thermal conductivity | W/m K |
| $\ell$ | length of the heating surface | m |
| M | dimensional constant, defined in Eq(5.21) | |
| m | exponent, defined in Eq.(5.1), equal to 2.33 | |
| n | number of active sites per unit area of the heating surface | $1/m^2$ |
| $n_4$ | exponent, defined in Eq.(6.2) | |

| | | |
|---|---|---|
| $P$ | pressure | $N/m^2$ |
| $\Delta P$ | pressure difference | $N/m^2$ |
| $q$ | heat flux | $W//m^2$ |
| $r_c$ | radius of active site | m |
| $r_r$ | radius of site for which n would be one per unit area, defined in Eq.(5.1) | |
| $T$ | temperature | K or °C |
| $\bar{T}$ | average temperature | K or °C |
| $\Delta T$ | temperature difference , $(T_w - T_\ell)$ | K or °C |
| $\overline{\Delta T}$ | average temperature difference | K or °C |
| $\Delta T_w$ | wall superheat, $(T_w - T_s)$ | K or °C |
| $\overline{\Delta T}_w$ | average wall superheat | K or °C |

Greek symbols

| | | |
|---|---|---|
| $\sigma$ | surface tension | $N/m$ |
| $\rho$ | density | $kg/m^3$ |
| $\lambda$ | latent heat of vaporization | $J/kg$ |
| $\mu$ | dynamic viscosity | $N\ s/m^2$ |
| $\nu$ | kinematic viscosity, $\mu/\rho$ | $m^2/s$ |
| $\alpha$ | thermal diffusivity, $k/C\rho$ | $m^2/s$ |
| $\theta$ | time | s |
| $\theta_d$ | departure period | s |
| $\theta_w$ | waiting period | s |
| $\delta$ | transient thermal layer thickness | m |

Dimensionless Modulii

| | | |
|---|---|---|
| $G_a$ | Gallilean number | $\frac{g\ D_b^3}{\nu^2}$ |
| $J_a$ | Jakob number | $\frac{C_\ell\ \rho_\ell\ \Delta T_w}{\rho_v\ \lambda}$ |

| | | |
|---|---|---|
| $K_P$ | Criterion for pressure term in boiling | $\frac{P}{\sqrt{g\,\sigma(\rho_\ell-\rho_v)}}$ |
| $K_t$ | Criterion for bubble break off frequency | $\frac{(\rho_v\,\lambda)^2}{C_\ell\,t_s\,\rho_\ell\sqrt{g\,\sigma(\rho_\ell-\rho_v)}}$ |
| $Nu_B$ | Nusselt number for boiling | $\frac{h\,D_b}{k_\ell}$ |
| $Pe_B$ | Peclet number for boiling | $\frac{q}{\rho_v\lambda\,\alpha}\sqrt{\frac{\sigma}{g(\rho_\ell-\rho_v)}}$ |
| $Pr$ | Prandtl number | $C_\ell\,\mu_\ell\,/\,k_\ell$ |
| $Re_B$ | Reynolds number for boiling | $\frac{q\,\rho_\ell}{\lambda\rho_v\mu_\ell}\sqrt{\frac{\sigma}{g(\rho_\ell-\rho_v)}}$ |

Subscripts

| | |
|---|---|
| i | inside |
| $\ell$ | liquid |
| o | outside |
| s | saturation |
| v | vapour |
| w | wall |
| 1 | normal boiling point |
| $\ell v$ | liquid-vapour |
| $\ell s$ | liquid-solid |
| vs | vapour-solid |
| Predt | predicted |
| Exptl | experimental |

# CHAPTER 1

## INTRODUCTION

The largest present day use of boiling of fluids is found in chemical, petro-chemical, power plant and refrigeration industries. Boiling heat transfer is advantageous in respect of enhancing heat transfer coefficient and thus reducing the weight of heat transfer equipment and consequently the cost of the equipment. This consideration has led to the use of boiling heat transfer in space vehicles also. In a nut shell, the applications of boiling heat transfer are ever-increasing with the advancement of modern technology. Keeping in view the vast number of equipment dealing with boiling heat transfer, this field of study has been the subject for active research for the last three decades in order to evolve suitable design procedures for these equipment.

A large number of empirical correlations have been recommended for the determination of boiling heat transfer coefficients. These correlations are not of general applicability for the wide range of parameters. One of the reasons of their limited applications is that these correlations have not been derived by considering the governing behaviour of bubble dynamics and surface characteristics on the boiling heat transfer which as a matter of fact, have pronounced effects on the boiling phenomenon. Obviously, these correlations have been attempted to relate heat transfer coefficient with the pertinent physico-thermal properties of the boiling fluids alone.

The effects of heating surface characteristics and bubble dynamics on heat transfer coefficient were accounted by Jakob and Linke[4] under the assumption that there exists a linear relationship between heat flux and number of active sites per unit area. They also assumed that the product of bubble emission frequency and departure diameter was a constant value. However, their correlating equation is not of general applicability. Recent studies have shown that the assumptions made by Jakob and Linke are not valid . Investigators [26], [46] [52] concluded that the relationship between heat flux and number of active sites per unit area is governed by a power law.

Until recently in all the empirical correlations and correlations due to Jakob and Linke [4] and Rohsenow [14], the bubble departure diameter was calculated by the Fritz equation [15]. Several investigations have been carried out for the determination of bubble departure diameter taking into consideration the dynamic effects of bubble [87], [88], [90], [92], [123]. It seems that its frequency and the number of active sites per unit area. A consequence of this would be that the heat transfer coefficient in nucleate pool boiling will be affected by the changes in these quantities. Therefore, an analytical model consistent with the requirements of nucleate boiling heat transfer would include the adequate determination of bubble departure diameter, its frequency and the number of active sites per unit area.

The heat transfer rate in the boiling of liquids is also influenced by the degree of superheat. In electrically heated heating surfaces, the surface temperature varies with heat flux, system pressure, physico-thermal properties of the boiling fluids and surface characteristics. Therefore, it will be of engineering interest to relate wall superheat with these parameters.

In chemical and petro-chemical industries there are many situations where boiling is carried out at atmospheric and sub-atmospheric pressures. Design data for boiling heat transfer at subatmospheric pressures are very scarce.

The considerations as above led to the present investigation with the following objectives:

Cole and Shulman [95] have conducted a careful investigation for recommending equation for bubble departure diameter at atmospheric and subatmospheric pressures. They have shown that the Fritz equation [15] is not valid for subatmospheric pressures.

Experimental investigations for bubble emission frequency using high speed photography have been made by many investigators [3], [70], [112 -116] for different boiling conditions. However, these studies do not yield expressions for direct calculations of bubble emission frequency. Besides, there is a scarcity of the bubble emission frequency relationships for subatmospheric pressure conditions. However, there exists adequate information of bubble growth rate and bubble departure diameter at subatmospheric pressures which can be used for developing analytical expressions for bubble emission frequency.

It is seen from these investigations that no single correlation for f and $D_b$ can be used for the wide range of degree of superheat and system pressures. Accordingly, care should be exercised when using them for a specific situation of boiling of fluids on heating surfaces.

It is certain that the turbulence in the superheat layer adjacent to the heating surface is markedly influenced by the bubble departure diameter,

1. To apply the appropriate governing equations of vapour bubble growth rate, bubble departure diameter, bubble emission frequency and number of active sites per unit area with a view to provide analytical equations representing the effects of wall heat flux, system pressure, physico-thermal properties of boiling fluids and surface characteristics on nucleate boiling heat transfer.

2. To experiment with a wide range of subatmospheric pressures in nucleate pool boiling of fluids of widely differing physico-thermal properties for determining the extent of validity of the present analytical model for the solution of practical problems.

3. To scrutinize the earlier analyses and empirical correlations with the help of present and existing experimental data.

4. To recommend a simplified procedure for calculating boiling heat transfer coefficients.

# CHAPTER 2

## LITERATURE REVIEW

During the last three decades a large effort has been made to researches in order to achieve basic understanding of boiling processes. An examination of investigations carried out in all these years reveals that the boiling of liquids represents the least understood form of heat transfer. No investigation has been conducted to study the boiling phenomena as a whole from the first principles. However, the various aspects of boiling heat transfer have been attempted as discrete studies. As a result of these studies, the underlying thermodynamic and hydrodynamic principles involved in the vapour bubble formation and growth and their effects on heat transfer rates are becoming known.

Keeping the aims of the present investigation in view, the literature on nucleate pool boiling heat transfer has been reviewed under the following sections:

### 2.1 EMPIRICAL CORRELATIONS

The heat transfer to the nucleate pool boiling of liquids has been the subject of active research in the past three decades. A great deal of experimental work has been done , and a large number of correlations for heat transfer coefficient have been proposed based on experimental data; however, no correlation has been proposed which possesses general applicability for wide ranges of pressure and heat flux, and for fluids of varying physico-thermal properties. In this section empirical correlations have been described with pertinent experimental variables.

The heat transfer coefficient has been related to the wall heat flux, the system pressure, the saturation temperature and the pertinent fluid properties in the following forms:

(1) In the dimensional form, nucleate boiling heat transfer coefficient may be expressed by the equation :

$$h = A\, q^{n}\, p^{m} \qquad (2.1)$$

Kutateladze and Borishanskii [1] have made an exhaustive survey of these correlations for wide ranges of heat fluxes, pressures, heating surfaces and fluids of differing properties. The survey shows that each equation is applicable only to a particular range of parameters for which it was derived. Some of the correlations have been attempted in the form:

$$h = A\, q^{n}\, ( t''/B)^{m} \qquad (2.2)$$

where $t''$ is the temperature range for which they are applicable.

(2) For many situations, the correlations for boiling heat transfer were obtained in the form of dimensionless groups, the exponents of the dimensionless groups were determined experimentally.

The general form of this type of correlationshas been expressed in the following form:

$$Nu = B\,Re^{n_1}\,Pr^{n_2}\,Ga^{n_3}\,Kp^{n_4}\,Kt^{n_5} \qquad (2.3)$$

The values of constant and exponents depend on the system conditions.

Cryder and Gilliland [2] using the methods of dimensional analysis established the correlation between variables as follows:

$$\frac{hD}{k_\ell} = 0.38\left(\frac{C_\ell\, z}{k_\ell}\right)^{0.425}\left(\frac{\Delta T S^2 D^2\, k_\ell}{z^3}\right)^{2.39}\left(\frac{z^2}{S\,D\,Y}\right)^{1.65} \qquad (2.4)$$

D = diameter of heating unit, inches

z = liquid viscosity, cp

S = specific gravity

Y = surface tension, poundals / ft

$C_\ell$ = specific heat of liquid, Btu/lb °F

The exponents in equation (2.4) were calculated from the experimental data for liquids, namely; water, carbon-tetrachloride, methanol, 1-butanol, gasoline, kerosene, glycerol and solutions of various inorganic salts.

Cryder and Finalborgo[5] have determined heat transfer coefficients for eight different liquids at boiling points both above and below atmospheric pressure experimentally. The liquids investigated by them differed in their physico-thermal properties. The heat rate varied from 439 to 2360 Btu/hr. Based on their experimental data they have concluded the following equations for heat transfer coefficient:

$$\log h = a + 2.5 \log \Delta t + bt \qquad (2.5)$$

$$\log \frac{h}{h_n} = b(t-t_n) \qquad (2.6)$$

Where $h_n$ denotes the heat transfer coefficient at the normal boiling point of a liquid, t is a temperature of boiling liquid, $t_n$ is normal boiling point of the liquid, and a and b are constants (their values depend on the nature of liquid as given in Table 2.1).

Table 2.1 Values of constants a and b in Eqs.(2.5 and 2.6)

| Liquid | a | b |
|---|---|---|
| Water | - 2.05 | 0.014 |
| Methanol | - 2.23 | 0.015 |
| Carbon tetrachloride | - 2.57 | 0.012 |
| Normal butanol | - 4.06 | 0.014 |
| 26.3% Glycerol solution | - 2.65 | 0.015 |
| Kerosene | - 5.15 | 0.012 |
| 10.1% Sodium sulphate solution | - 2.62 | 0.016 |
| 24.2 % Sodium chloride solution | - 3.61 | 0.017 |

Akin and McAdams [6] carried out an experimental investigation on a chrome-plated horizontal cylinder of diameter $1.905 \times 10^{-4}$ m immersed in liquids, namely; water, isopropanol, isobutanol and n-butanol at atmospheric pressure. They also investigated the boiling of distilled water at subatmospheric pressures ranging from 0.16 - 0.68 kg/cm$^2$.

Insinger and Bliss [7] conducted an experimental investigation with water, carbon tetrachloride, isopropanol and 40% sucrose solution at atmospheric pressure. They used a vertical heating cylinder of 6 in. length and about 1.25 in. diameter. A simplified form of the Insinger-Bliss correlation is given by:

$$\frac{h}{\sqrt{k_\ell}} \times \frac{\sqrt{\sigma}}{\rho_\ell^{0.82}} = 4\, J^{0.41}\, g^{-0.09}\, 10^{-3.2} \left( \frac{q^{0.68} C_\ell^{0.5}}{\rho_v^{0.5} \lambda^{.27}} \right) \tag{2.7}$$

Where J is mechanical equivalent of heat equal to 778 ft.lb/Btu.

This equation was found to satisfy the experimental results of Jakob and Linke [4] on water and carbon tetrachloride, of Linden and Montillon [8] and Dunn and Vincent [9] on water, and of Akin and McAdams [6] on water, isopropanol, isobutanol and n-butanol along with his own experimental data.

Bonilla and Perry [10] under-took an experimental investigation to obtain a method of interpolation between pure liquids to give boiling heat transfer coefficients for their binary mixtures. They used two heaters of effective diameter 3.58 in.and 2.62 in. respectively each consisted of 3/4 in. copper disk with copper fins 3/16 in. thick and 2 in. high welded to the bottom. The 1/4 in. spaces between fins contain flat elements wound with 1/8 in. wide nichrome ribbon , 0.283 ohm/ft, on mica cores, with asbestos insulation. The pure liquids studied were water, ethanol, n-butanol and acetone. They examined the high range Jakob and Linke equation, Eq. (2.22) as discussed later-on . The ethanol curves at pressures other than atmospheric were corrected for pressure by plotting

$$\left(\frac{v_a}{v}\right)^{1.37}\left[\frac{\sigma}{\sigma_a}\quad\frac{\rho_{\ell,a}}{\rho_\ell}\quad\frac{q}{\rho_{v,a}\lambda_a w_{b,a}}\right]$$

as abscissa. The average slope of 0.73 compares with Jakob and Linke of 0.80 (cf. equation 2.22). The modification to the equation due to Jakob and Linke as suggested by these authors consists in changing the exponent from 0.80 to 0.73, inclusion of Prandtl number raisedto the power of 1/2 , and changing the constant from 31.6 to 16.6. Thus, the proposed equation was:

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{\rho_\ell}} = 16.6\left[\frac{v_a}{v}\right]\left(\frac{\sigma}{\sigma_a}\quad\frac{\rho_{\ell,a}}{\rho_\ell}\quad\frac{q}{\rho_{v,a}\lambda\ w_{b,a}}\right)^{0.73}\left(\frac{C_\ell\mu_\ell}{k_\ell}\right)^{0.5} \qquad (2.8)$$

They have also attempted to determine the effect of pressure on heat transfer coefficient for a heat flux value of 50,000 Btu/hr ft$^2$. They found that heat transfer coefficient is directly proportional to the pressure raised to the power of 0.25. In other words it can be concluded that boiling surface does not change the ratios of the coefficients obtained at different pressures.

Cichelli and Bonilla [11] investigated the boiling of pure liquids: water, ethanol, benzene, propane, n-pentane and n-heptane; and binary mixtures: 50 mole per cent water-ethanol and 33 and 67 mole per cent propane-pentane. They observed that coefficient of heat transfer increased with pressure continuously until at critical pressure the nucleate boiling ceased to be stable. The apparatus used was an electrically heated horizontal plate. The surface was a thick copper plate with 0.002 in. of polished electroplated chromium.

Addoms [12] conducted the boiling heat transfer investigations experimentally. He used degassed distilled water on electrically heated horizontal platinum wire of diameter $6.096 \times 10^{-4}$ m. The pressure varied from 14.7 psia to 2465 psia. Addoms' data have been used extensively by other investigators for sake of comparison.

Farber and Scorah [13] also used the horizontal wire of $1.016 \times 10^{-4}$ m diameter in a pool of distilled water to carry out the investigations for nucleate boiling at atmospheric pressure.

McNelly [20] included five dimensionless groups in his correlation. The exponents and the coefficients were selected to give a reasonable fit with the experimental data from the boiling of 14 liquids. He recommended the following correlation:

$$\frac{hD}{k_\ell} = 0.255 \left[\frac{Qd}{A\lambda\mu_\ell}\right]^{0.69} \left[\frac{Pd}{\sigma}\right]^{0.31} \left[\frac{\rho_\ell}{\rho_v} - 1\right]^{0.33} \left[\frac{C_\ell\,\mu_\ell}{k_\ell}\right]^{0.69} \qquad (2.9)$$

Where,

D = diameter of heating unit, inches

d = characteristic length of heating surface, ft.

Q = heat transfer rate, Btu/hr.

$\lambda$ = heat of vaporization, Btu/lb.

P = ambient pressure on liquid, lb/sq.ft.

Averin & Kruzhilin [21] correlation is of the following form:

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{\rho_\ell - \rho_v}} = 0.082 \left[\frac{q}{\rho_v \lambda}\sqrt{\frac{\sigma}{\rho_\ell - \rho_v}}\right]^{0.7} \left[\frac{C_\ell\,\mu_\ell}{k_\ell}\right]^{-0.5} \left[\frac{J\,(\rho_v\lambda)^2}{C_\ell t_s\, \rho_\ell \sqrt{\sigma\,(\rho_\ell - \rho_v)}}\right]^{0.377} \qquad (2.10)$$

Gilmour [23] presented an expression for nucleate boiling which resembles those for convective heat transfer and condensation heat transfer. He claims that his method is in contrast to the usual methodical approach for correlating the boiling heat transfer data. In his correlation, along with three familiar dimensionless groups (St, Pr and Re) he included an additional new dimensionless group to account for pressure and surface tension effects. His expression can be represented in the following form:

$$\left[\frac{h}{CG}\right]^{a}\left[\frac{C\mu}{k}\right]^{b}\left[\frac{\rho_{\ell}\,\sigma}{P^{2}}\right]^{c}=\frac{\phi}{(DG/\mu)^{d}} \qquad (2.11)$$

The left hand side of the above expression is termed as Colburn j- function for nucleate boiling. The exponents on Stanton number is assumed to be unity and that on the Prandtl number for liquid heating is 0.6. The exponents on the pressure group and Reynold number have been computed by plotting the data of Cichelli and Bonilla [11] and Cryder and Finalborgo [5] for high pressures and subatmospheric pressures respectively. Both the data yielded the same magnitude of exponents.

The final correlation thus assumes the form:

$$\left(\frac{h}{CG}\right)\left[\frac{C\mu}{k}\right]^{0.6}\left[\frac{\rho_{\ell}\sigma}{P^{2}}\right]^{0.425}=\frac{0.001}{(DG/\mu)^{0.3}} \qquad (2.12)$$

Where $G = \frac{V \rho_\ell}{A \rho_v}$ and D is the diameter of tube.

Gilmour [23] could correlate the existing experimental data for differing liquids of various investigators with his correlation satisfactorily.

Sternling and Tichacek [28] in their attempt to generate the experimental data for binary mixtures of large relative volatility also provided the nucleate boiling heat transfer data for pure fluids: benzene, methyl chloroform, carbon tetrachloride, isopropanol, methanol and water. They measured the heat transfer coefficients at atmospheric pressure for these fluids in a pool boiler for a wide range of heat flux.

Labuntsov [36] proposed the following correlation for nucleate pool boiling. The Nusselt group in his correlation has been calculated by taking the Fritz expression for bubble diameter.

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{(\rho_\ell - \rho_v)g}} = 0.125 \left[\frac{q}{\rho_v \; \lambda \; \alpha}\sqrt{\frac{\sigma}{(\rho_\ell - \rho_v)g}}\right]^{0.65} \left[\frac{C_\ell \, \mu_\ell}{k_\ell}\right]^{-0.32} \left[\frac{(\rho_v \lambda)^2}{(t_s + 273)\rho_\ell C_\ell \sqrt{\sigma(\rho_\ell - \rho_v)g}}\right]^{0.35} \tag{2.13}$$

Hughmark [38] made a statistical analysis of nucleate pool boiling data. The exponents of several variables were determined from an analysis of the data of 23 liquids. The expression so derived is as follows:

$$q = 2.67 \times 10^{-7} \left[ \frac{(\Delta p)^{1.867} (\rho_L - \rho_v)_w^{2.27} (C_L)_w^{0.945} \, T_s^{1.618}}{(\rho_v)_w^{1.385} \, (\mu_L)_w^{1.630} \, (\lambda)_w^{1.15} \, (P/P_c)^{0.202}} \right] \tag{2.14}$$

subscript w represents that the properties evaluated at wall surface temperature.

$P$ is the ambient pressure on liquid, lb/sq.ft

$P_c$ is critical pressure, lb/sq.ft

$\Delta P$ is vapour pressure difference corresponding to $\Delta T$, lb/sq.ft

Kutateladze [39] included the pressure term in his correlation. He recommended the correlation of the following form:

$$\frac{h}{k_\ell} \sqrt{\frac{\sigma}{(\rho_L - \rho_v)\, g}} = 7.0 \times 10^{-4} \left[ \frac{q}{\rho_v \, \lambda \, \alpha} \sqrt{\frac{\sigma}{(\rho_L - \rho_v) g}} \right]^{0.7} \left[ \frac{C_\ell \mu_\ell}{k_\ell} \right]^{-0.35} \left[ \frac{P}{\sigma (\rho_L - \rho_v) g} \right]^{0.7} \tag{2.15}$$

Kichigen and Tobilevich [39] obtained the boiling heat transfer correlation by analyzing the data for a number of fluids. Their correlation is of the following empirical form:

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}} = 1.04\times 10^{-4}\left[\frac{q}{\rho_v\lambda\,\alpha}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}}\right]^{0.7}\left[\frac{P}{\sqrt{\sigma(\rho_\ell-\rho_v)g}}\right]^{0.7}\left[\frac{g}{v_\ell^2}\left(\frac{\sigma}{(\rho_\ell-\rho_v)g}\right)^{3/2}\left(1-\frac{\rho_v}{\rho_\ell}\right)\right]^{0.125} \qquad (2.16)$$

Drayer [40] extended the nucleate boiling correlations to the nucleate boiling of cryogenic fluids - hydrogen. He examined 11 existing correlations, where three appeared directly applicable for the prediction of heat transfer coefficient for liquid hydrogen over limited ranges. The Forster-Zuber [22], Forster-Grief [33] and Cryder-Gilliland [2] relations yield heat flux values in good agreement with experimental data at temperature difference in the neighbourhood of 1°F. The correlations of Hughmark [38] Gilmour [23] and McNelly [20] were fair; and those of Levy [24] , Jakob-Linke [4] , Insinger- Bliss [7], Miyauchi-Yagi [37] and Nishikawa [26] were poor . Drayer advocates that several of these correlations could be made to agree with the experimental data merely by a readjustment of some of the arbitrary constants used.

Sciance et al [41] have studied the pool boiling of saturated hydrocarbons — ethane, propane , and n-butane outside a horizontal gold- plated cylinder. They succeeded in correlating all the data of the hydrocarbons except ethane by modifying the Rohsenow equation [14] in the following form:

$$\frac{q\sqrt{\frac{\sigma}{\rho_L-\rho_v}}}{\lambda\,\mu_\ell} = C\left[\frac{C_\ell\,\Delta t}{\lambda}\left(\frac{1}{Pr}\right)^{1.18}(T_r)^{1.18}\right]^n \qquad (2.17)$$

The constant C and exponent n, are given in Table 2.2

Table 2.2 Values of Constant C and Exponent n in Eq.(2.17)

| Liquid | $C \times 10^{-5}$ | n |
|---|---|---|
| Methane | 3.25 | 2.89 |
| Propane | 5.77 | 2.60 |
| n-Butane | 2.33 | 2.84 |

Borishanskii et al [42] have studied heat transfer from a tube to water and ethanol in nucleate pool boiling. The tests were performed on three tube sizes having ID/OD equal to 6.12/6.94, 4.00/4.99 and 4.00/6.00. The majority of tests were made on first two tubes. The pressure range was 1.03 to 200 $kg_f/cm^2$ and heat flux range from $50 \times 10^3$ to approximately $1 \times 10^6$ kcal/hr $m^2$ for water. For ethanol the pressure ranged from 1 to 60 $kg_f/cm^2$ and heat flux from $22 \times 10^3$ to $700 \times 10^3$ kcal/hr $m^2$. These investigators concluded that the effect of pressure on heat transfer rate in nucleate boiling is expressed by a complex relationship and cannot be represented by a simple power law with a constant exponent on the pressure term.

Kosky and Lyon [35] have measured the nucleate pool boiling heat transfer data for pure nitrogen, oxygen, argon, methane and carbon tetrafluoride on a horizontal, flat, circular, platinum plated disk for saturation pressures ranging from 1 atm. or less to the immediate vicinity of the critical pressure. They compared all their experimental results for various fluids with several suggested nucleate boiling correlations [14],[20], [22],[23],[33], [39], [42]. These authors drew the conclusion that the Gilmour, McNelly, Kutateladze, and Borishanskii- Minchenko correlations are all superior to the Rohsenow and Forster-Zuber/Forster-Greif correlations with the following exceptions: (i) The degree of success for the Gilmour correlation holds only for nitrogen. (The order of success for the Gilmour correlation was $N_2 > Ar > O_2 > CH_4 > CF_4$ ). (ii) The Forster-Zuber/ Forster-Greif correlation was roughly as successful for the methane and carbon tetrafluoride data as any correlation tested by Kosky and Lyon but their emphasis is that those data are the poorest obtained in their work.

Frost and Li [47] attempted to evaluate the constant $C_{sf}$ and exponent r in the Rohsenow pool-boiling correlation [14] for subatmospheric pressures. The range of vacuum pressure selected by these investigators was 0.92 psia - 14.45 psia. The test-pool consisted of distilled water filled in a vacuum tank made of pyrex. The heat transfer surface was electrically heated platinum wire of 0.008 in. diameter and 3 in. length.

The authors with the method of least squares fit found that the exponent r , varied from a minimum of 0.300 to a maximum of 0.377 in comparison to the value of 0.33 obtained by Rohsenow. The value of the constant of proportionality $C_{sf}$ was obtained 0.01134 for a pressure of 14.45 psia compared to a value of 0.013 obtained by Rohsenow at atmospheric and higher pressures. However, the investigations of these authors indicate a significant effect of vacuum pressure on this constant. The values of constant $C_{sf}$ and r are listed in Table 2.3 for the vacuum pressures investigated by these investigators. An examination of Table 2.3 shows that the vacuum pressure has little influence on the exponent r while it has a significant effect on the constant $C_{sf}$ in the Rohsenow pool-boiling correlation.

Table 2.3 Values of Constant $C_{sf}$ and Exponent r from [47]

| Pressure psia | Constant $C_{sf}$ | Exponent r |
|---|---|---|
| 14.45 | 0.01134 | 0.325 |
| 4.12 | 0.00959 | 0.300 |
| 1.53 | 0.00902 | 0.377 |
| 0.92 | 0.00754 | 0.341 |

Rice and Calus [48] have carried out experimental investigations for calculating heat transfer coefficient during pool boiling. The liquids used were toluene, carbontetrachloride, methanol, n-propanol, isopropanol, water and water-isopropanol azeotrope at atmospheric pressure. The test surface was a 0.0315 cm. diameter nickle-aluminium wire 15.3 cm. long of which the 8.9 cm. midlength was used for heat transfer measurements. The wire was mounted horizontally under slight tension in a stainless steel yoke. The range of operating variables is given in Table 2.4.

The results of their investigation are correlated by the equation:

$$\frac{Nu}{K_p^{0.7}} \left[ \frac{T_s}{T_{sw}} \right]^4 = EP_e^{0.7} \qquad (2.18)$$

Where $T_s$ denotes absolute boiling point at system pressure, $T_{sw}$ absolute boiling point of water at system pressure.

Table 2.4 Range of Operating Variables in [48]

| | Atmospheric boiling point $^{o}C$ | Range of $\Delta T$ $^{o}C$ | Range of Q/A, $W/m^2$ |
|---|---|---|---|
| Toluene | 110.8 | 9.6-33.2 | 9,150-419,560 |
| Carbon tetrachloride | 76.8 | 11.9-24.7 | 35,330-307,570 |
| Methanol | 64.7 | 7.7-14.4 | 82,330-574,130 |
| N-propanol | 97.8 | 12.2-28.9 | 9,780-441,000 |
| Isopropanol | 82.5 | 6.8-15.1 | 22,710-401,900 |
| Water | 100.0 | 7.8-21.7 | 63,720-706,630 |
| Water-isopropanol azeotrope | 80.4 | 7.7-16.7 | 37,220-507,890 |

Equation (2.18) is a modification of the Borishanskii -Minchenko correlation [42]. The value of constant E is $6.30 \times 10^{-4}$ for the surface used in their work. This correlation correlated experimental data of Cichelli and Bonilla[11] when the value of constant E was taken to be $3.92 \times 10^{-4}$. The different value of E has been explained due to different surfaces used by Cichelli and Bonilla. The experimental data of Borishanskii et al for water from stainless steel surface were also correlated by equation, Eq. (2.18 ) with constant $E = 8.9 \times 10^{-4}$.

## 2.2 SEMITHEORETICAL CORRELATIONS

Amongst the many other factors, number of nucleation sites on heating surface and their characteristics play a significant role in achieving high transfer rates in boiling of liquids. The vapour bubbles, during the course of their birth, growth, collapse or break-off the heating surface, give rise to large turbulence which accounts for high heat transfer rates. It can also be noted that the nucleation sites on heat transfer surface are effective in reducing the average wall superheat required for boiling and, hence, high boiling coefficients are realised.

A survey of literature demonstrates that several investigators, basing their conclusions on theoretical considerations or experimental evidence or both, have attempted to determine exactly how heat transfer rate, heat transfer coefficient, wall superheat and the number of nucleation sites are related.

Some of the investigators [49], [14], [50] have proposed a linear relationship between the heat flux and the number of active sites on a horizontal surface. This relationship was originally propogated by Jakob[49]. However,recent investigations [26],[46],[51],[52] raise doubts on the validity of the linear relationship.

Jakob and Linke [3],[4] were perhaps the first to develop a semitheoretical model for heat transfer coefficient considering the effects of turbulence caused due to birth, growth, collapse and break-off of vapour bubbles on the heating surface. In their model the influence of number of nucleation sites is accounted by $nA_{b,1}/A$. This ratio represents the fraction of heating surface which becomes vacant due to bubble break-off . The influence of the continual displacement of liquid which develops a circulation of liquid along each vapour column is taken into account by a quantity $V_{b,1}/V_{b,2}$. Hence they related heat transfer coefficient to these ratios, bubble diameter at break-off and thermal conductivity of the fluid in the following functional form:

$$\frac{h\,D_{b,1}}{k_\ell} = \Psi\left(\frac{nA_{b,1}}{A} \cdot \frac{V_{b,2}}{V_{b,1}}\right) \qquad (2.19)$$

Using a linear relationship between the heat flux and the number of active sites on heating surface, equation, Eq.(2.19) has been reduced to the following form:

$$\frac{hD_{b,1}}{k_\ell} = \Psi\left(\frac{q}{\rho_v\,\lambda}\;\frac{1}{D_{b,1}f}\right) \qquad (2.20)$$

Later, Jakob [3] from his photographic studies concluded that $fD_{b,1}$ was nearly same for boiling water and carbon tetrachloride and approximately equals to 280 m/hr. Jakob and Linke [4] using photographic measurements reduced the Fritz equation [15].

Finally, using $D_{b,1}$ from the reduced Fritz equation and $fD_{b,1}$ equal to 280 m/hr these investigators showed that equation, Eq. (2.20) represented in the form given by Eq. (2.21)

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{\rho_\ell}} = 30\left(\frac{q}{\rho_v\,\lambda\,D_{b,1}f}\right)^{0.8} \qquad (2.21)$$

Jakob [3] in 1938 reconsidered the above model with a view to provide general applicability for pressures other than atmospheric pressure. The modified form of Eq. (2.21) was obtained and is given by Eq. (2.22).

$$\frac{h}{k_\ell}\left[\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}}\right] = 31.6\,\frac{v_{\ell,a}}{v_\ell}\left[\frac{\rho_{\ell,a}}{\rho_\ell}\,\frac{\sigma}{\sigma_a}\,\frac{q}{\rho_{v,a}\,\lambda_a D_{b,a}f}\right]^{0.8} \qquad (2.22)$$

Where the subscript 'a' denotes a physical property at the normal boiling point $v_{\ell,a}/v_\ell$ has been included empirically. In the derivation of this equation $D_{b,1}$ and f, and their product were considered to depend upon the pressure.

Jicina-Molozhin and Kutateladze [19] have scrutinized Jakob and Linke correation [4] and Jakob correlation [3] for the experimental data for nucleate pool boiling of carbon tetrachloride, 26% water-glycerine solution, 24% aqueous sodium chloride solution, water and mercury. The pressure range for these experimental data was from 0.15 to 10.0 kg/cm$^2$. These authors observed that the Jakob correlation, Eq. (2.22), does not correlate the experimental data satisfactorily. The possible reason for the reported deviation has been that the Jakob correlation does not account for the effect of pressure on $fD_b$. Jakob assumed the value of $fD_b$ a constant. These authors have related $fD_b$ with pressure and physical properties of the liquid by the following expression:

$$fD_b = C_{sf} \sqrt{\frac{g \rho_\ell \sigma}{P \rho_v}} \qquad (2.23)$$

Where $C_{sf}$ represents liquid-solid combination factor and depends upon the wetting property of the liquid. Further, they found that their experimental data and those of Cryder and Finalborgo [5] are well-represented by the following equation:

$$\frac{h^*}{h_a^*} = \frac{k}{k_a} \; \frac{D_{b,a}}{D_b} \left[\frac{\lambda_a \rho_{v,a}}{\lambda \rho_v}\right]^{0.7} \left[\frac{f_a D_{b,a}}{f D_b}\right]^{0.7} \qquad (2.24)$$

where $h^* = \frac{h}{q^{0.7}}$, and subscript a denotes a physical property at the pressure of 1 kg/cm$^2$.

In 1952 Rohsenow [14] also attempted to derive equation for nucleate boiling heat transfer theoretically. He developed a correlation of Nusselt-Reynold-Prandtl type. Rohsenow defined the bubble Reynold number and bubble Nusselt number. The quantities needed for these dimensionless groups such as bubble diameter and product of bubble diameter and frequency were inserted from Fritz [15] and Jakob [3], [16]. Nevertheless, he introduced a new dimensionless group $C_{\ell} T_x / h_{fg}$ in place of bubble Nusselt number which represents the ratio of liquid superheat enthalpy at the surface temperature to the latent enthalpy of evaporation. The basic assumptions involved in this correlation are; (i) heat transfer rate per unit heating surface area, $q$ , is proportional to the heat transfer rate to bubble per unit heating surface area while bubble remains attached to the surface $(q)_b$, (ii) the product of frequency and bubble diameter is constant as in the investigation of Jakob and Linke [4], and (iii) contact angle $\beta$ remains independent of pressure and its effect has been considered through surface-liquid combination factor $C_{sf}$; $C_{sf} = \text{constant}\ (\beta)^r$ , $r = 0.33$.

Rohsenow obtained the values of exponents over Re,b and Pr,b and value of $C_{sf}$ by applying the proposed correlation to the data of Addoms [12] which cover the pressure range of 14.7 psia to 2465 psia. His correlation thus assumes the following form:

$$\frac{C_{\ell} T_x}{h_{fg}} = C_{sf} \left[ \frac{q/A}{\mu_{\ell}\, h_{fg}} \sqrt{\frac{g_o\, \sigma}{g(\rho_{\ell} - \rho_v)}} \right]^{0.33} \left[ \frac{C_{\ell}\, \mu_{\ell}}{k_{\ell}} \right]^{1.7} \qquad (2.25)$$

Equation (2.25) may be rewritten in the form:

$$\frac{h_x}{k_\ell}\sqrt{\frac{g_o\,\sigma}{g(\rho_\ell-\rho_v)}} = \frac{1}{C_{sf}}\left[\frac{q}{\mu_\ell\lambda}\sqrt{\frac{g_o\,\sigma}{g(\rho_\ell-\rho_v)}}\right]^{0.667}\left(\frac{C_\ell\mu_\ell}{k_\ell}\right)^{-0.7} \qquad (2.26)$$

or $\quad Nu,b = \frac{1}{C_{sf}}\,(Re,b)^{0.667}\,(Pr)^{-.0.7}$

The value of $C_{sf}$ in equation (2.25) are different for each of the surface-fluid combination and are given in Table 2.5 for the data of various investigators.

Table 2.5 Values of Constant $C_{sf}$ in Eq. (2.25)

| Surface-fluid combination | $C_{sf}$ |
|---|---|
| Water-Nickle [17] | 0.006 |
| Water-Platinum [12] | 0.013 |
| Water-Copper [18] | 0.013 |
| Water-Brass [5] | 0.006 |
| $CCl_4$ - Copper [18] | 0.013 |
| Benzene - Chromium [11] | 0.010 |
| n-pentane-Chromium [11] | 0.015 |
| Ethanol-Chromium [11 ] | 0.0027 |
| Isopropanol-Copper [18] | 0.0025 |
| 35% $K_2CO_3$ - Copper [18] | 0.0054 |
| 50% $K_2CO_3$ - Copper [18] | 0.0027 |
| n-Butyl Alcohol-Copper [18] | 0.0030 |

Forster and Zuber [22] expression is a modification of the Rohsenow correlation [14]. They developed their correlation based on a study of bubble dynamics. In this case the bubble growth rate was assumed to be the dominant mechanism. The bubble diameter used was:

$$D_b = \frac{(\Delta T) C_\ell \, \rho_\ell \sqrt{\pi \alpha}}{\lambda \, \rho_v} \left[ \frac{\rho_\ell \, R_i^3}{2\sigma} \right]^{1/4}$$

Their relationship between variables in the nucleate boiling regime is as follows:

$$\frac{C_\ell \, \rho_\ell \sqrt{\pi \alpha_\ell} \; Q}{k_\ell \, \lambda \, \rho_v \, A} \left( \frac{2\sigma}{\Delta P} \right)^{0.5} \left( \frac{\rho_\ell}{g \Delta P} \right)^{0.5} = 0.0015 \left[ \frac{\rho_\ell}{\mu_\ell} \left( \frac{C_\ell \, \rho_\ell \, \Delta T \sqrt{\pi \alpha_\ell}}{\lambda \, \rho_v} \right)^2 \right]^{0.62} \left( \frac{C_\ell \, \mu_\ell}{k_\ell} \right)^{0.33} \qquad (2.27)$$

This rather formidable expression has not been extensively tested.

Levy [24] employed a simplified model of the boiling mechanism close to the heated surface. By utilising the bubble growth rate equation of Forster and Zuber [25] he deduced the following correlation:

$$q = \frac{k_\ell \, C_L \, \rho_L^2}{\sigma \, T_s (\rho_L - \rho_v)} \; \frac{1}{B_L} \; (\Delta T)^3 \qquad (2.28)$$

The coefficient $B_L$ was determined empirically employing the data of various investigators [11], [12], [18] and was found to be a function only of the product $\rho_v h_{fg}$.

To prove the validity of the recommended equation it was applied to the experimental data of several investigators [11], [12], [18] and they indicated that the proposed equation is of a generalised nature applicable to a widely differing fluids from low to very high pressures. He concluded that proposed equation is independent of heating surface-liquid combination. However, Levy noticed several deviations from the proposed equation and these were probably attributable to the evaluation of coefficient $B_L$.

Nishikawa et al [26], [32], [51] undertook a comprehensive investigation to determine the effect of surface nucleation sites on heat flux, heat transfer coefficient and wall superheat quantitatively. Based on their experimental data the following relationships have been concluded:

$$q \propto n^{1/2} \tag{2.29}$$

$$h \propto (f D_b^3 n)^{1/3} \tag{2.30}$$

$$\Delta T_w \propto q^{2/3} n^{-1/6} \tag{2.31}$$

Where n denotes the number of nucleation sites per unit area of the heating surface.

The applicability of these relationships has been reported for a pressure range of 0.4 $kg_f/cm^2$ to

1.03 $kg_f/cm^2$, a heat flux range upto 66,000 $kcal/hr\ m^2$ and a maximum population count of 8 active sites per square inch.

The Nishikawa relationship for heat transfer coefficient has been successful in correlating experimental data of Kurihara and Myers [52] for water and four organic liquids at heat fluxes upto 92,500 $kcal/hr\ m^2$ and 28 active sites per square inch. Kurihara and Myers [52] have concluded that neither the degree of surface roughness nor the nature of liquid had an effect on the Nishikawa relationship between heat transfer coefficient and active-site population.

Nishikawa and Urakawa [26], [27], [29-32] undertook an investigation of nucleate boiling of water under reduced pressure ranging from 0.4 $kg/cm^2$ to atmospheric pressure both experimentally and analytically. The mechanism of boiling heat transfer has been discussed theoretically and the following expressions in dimensionless form have been recommended for heat transfer coefficient.

For laminar flow:

$$\frac{hR}{k_\ell} = 6.35\left[\left(\frac{1}{M_s^2\, P}\;\frac{C_\ell\, \rho_\ell^2}{k_\ell\, \sigma\, \lambda\, \rho_v}\right)^{1/2} R^{3/2}\, q\right]^{2/3} \qquad (2.32)$$

For turbulent flow:

$$\frac{hR}{k_\ell} = 8.26\left[\left(\frac{1}{M_s^2\, P}\;\frac{C_\ell\, \rho_\ell^2}{k_\ell\, \sigma\, \lambda\, \rho_v}\right)^{1/2} R^{3/2}\, q\right]^{8/11} \qquad (2.33)$$

Where;

$M_s$ = 900 , $m^{-1}$
$P$ = 1,699 , kcal/h
$R$ = characteristic dimension of the heating surface.

The average line for experimental points can be represented by the following equation:

$$\frac{hR}{k_\ell} = 8.0 \left[ \left( \frac{1}{M_s^2 P} \, \frac{C_\ell \, \rho_\ell^2}{k_\ell \, \sigma \, \lambda \, \rho_v} \right)^{1/2} R^{3/2} \, q \right]^{2/3} \qquad (2.34)$$

Eq. (2.34) is valid only for clean smooth heating surface under atmospheric pressure. This equation has been further generalised by including foamability factor , $\frac{\xi}{\xi_s}$ , for the dirty or rough surface and pressure factor, $\frac{p}{p_s}$ , pressures higher and lower than atmospheric. The final general correlation assumes the following form:

$$\frac{hR}{k_\ell} = 8.0 \left[ \left( \frac{\xi}{\xi_s} \right)^{1/2} \left( \frac{p}{p_s} \right) \left( \frac{1}{M_s^2 \, P} \, \frac{C_\ell \, \rho_\ell^2}{k_\ell \sigma \, \lambda \, \rho_v} \right)^{1/2} R^{3/2} \, q \right]^{2/3} \qquad (2.35)$$

where;

$\xi$ = coefficient of foaming ability for any combination of surface and liquid.
$\xi_s$ = coefficient of foaming ability for the combination of clean smooth surface and pure liquid
$p$ = pressure
$p_s$ = atmospheric pressure

Forster and Greif [ 33 ] have analyzed the various proposed heat transfer mechanisms. They concluded that a vapour-liquid exchange mechanism can explain the rate of heat transfer in boiling. Two different correlations relating the boiling heat flux with superheat, system pressure and properties of the fluid have been prpposed as follows:

$$q = c_1 \frac{k_\ell}{2\sigma} \Delta P \; \Delta T_1 \left( \frac{\rho_\ell}{\mu_\ell} A^2 \right)^{1/5} \left( \frac{\mu_\ell c_\ell}{k_\ell} \right)^{1/3} \qquad (2.36)$$

and

$$q = 1.2 \times 10^{-3} \; \frac{\alpha \, c_\ell \, \rho_\ell \, (t_s+273)}{\lambda \, \rho_v \, \sigma^{1/2}} \left( \frac{c_\ell (t_s+273) \alpha^{1/2}}{(\lambda \, \rho_v)^2} \right)^{1/4} \left( \frac{\rho_\ell}{\mu_\ell} \right)^{5/8} \left( \frac{\mu_\ell \, c_\ell}{k_\ell} \right)^{1/3} \Delta P^2 \qquad (2.37)$$

constant $c_1$ was evaluated from the experimental data of Gunther and Kreith [34] and Cichelli and Bonilla [11] for water, the coefficient $c_1$ was found to be $0.7 \times 10^{-2}$ Data only at pressures of 50 atm and 1 atm were compared. Eq. (2.36) was also tested for other liquids including mercury. The equation predicted the heat flux correctly for high and low pressure once the coefficient $c_1$ was determined. From a measurement at atmospheric pressure, the coefficient $c_1$ varied (from water to mercury) by a factor of 2.

Eq. (2.37) was tested by comparing with experimental data for the liquids: water from 1 to 50 atm, n-butyl alcohol for 50 psia, aniline for 35 psia and mercury for 1 and 3 atm.

It is important to note that the above equations assume the following relationship:

$$\left[\frac{d(\log \Delta T_1)}{d(\log p)}\right]_{q\,=\,\text{constant}} = -\frac{1}{4}$$

The above relationship holds true for low pressures. Obviously these correlations might not correlate the boiling heat transfer data for high pressures. This has been pointed out by Leinhard in his private communication to Kosky and Lyon [35] Another point can also be noted that these authors did not use the value of $D_b$ as its determination was not known to these authors. Therefore, they have argued from physical considerations the important parameters and dimensionless groups which can be used to describe the heat transfer based on vapour-liquid exchange mechanism.

Miyauchi and Yagi [37] studied the relationship between heat transfer coefficient and the population of active nucleation sites. Their resulting expression, which has two dimensionless groups is ;

$$\frac{h}{k_\ell} = c_1 \left(\frac{q\,\rho_\ell}{\lambda\,\rho_v\,\mu_\ell}\right)^{0.74} \left(\frac{\rho_v}{\rho_o}\right)^{0.69} \left(\frac{c_\ell\,\mu_\ell}{k_\ell}\right)^{0.63} \tag{2.38}$$

Coefficient $c_1$ depends on the roughness of the heating surface and on the liquid employed. $\rho_o$ is the vapour density at atmospheric pressure in lb/cu.ft.

Gaertner and Westwater [46] carried out photographic investigation to determine the effect of population of active sites on boiling heat transfer rates. The experimental facility used by them consisted of 20% aqueous solution of nickel salt boiling at atmospheric pressure on a horizontal, flat, copper surface 2 in. in diameter. The heat flux was varied from 7,680 to 535,000 Btu/hr sq.ft and the $\Delta T$ from $17.3^oF$ to $218.8^oF$ . Based on experimental data the following relationships for heat flux and heat transfer coefficient in terms of active-site population have been proposed:

$$q = 1400\ n^{0.47} \tag{2.39}$$

$$h = 49\ n^{0.33} \tag{2.40}$$

In 1969 Mikic and Rohsenow [43] have studied boiling heat transfer analytically. Starting with the basic mechanism for a single active cavity site, these authors have related average heat flux from the heated surface to the active-site density, the average frequency for bubble departure, the bubble diameter at departure and wall superheat. They have used the results of Brown [53] for determining the number of active-sites. The number of active-sites is first related to the surface cavity distribution and then to fluid properties and the wall superheat by the following expression:

$$n = c_1 r_s^m \left(\frac{\lambda \rho_v}{2T_s \sigma}\right)^m \Delta T^m \qquad (2.41)$$

Where $r_s$ is a radius for which n would be one per unit area, and $c_1$ is dimensional constant (1/unit area) . The value for m was taken to be 2.5 for water and 3.0 for n-pentane, benzene, and ethanol.

The heat flux is related to the pertinent quantities by the following functional equation:

$$q = c_1 \frac{r_s^m}{\sqrt{\pi}\; 2^{m-1}} (k \rho C)_\ell^{1/2} \left(\frac{\lambda \rho_v}{T_s \sigma}\right)^m \sqrt{f}\; D_b^2 \, \Delta T^{m+1} \qquad (2.42)$$

Equation (2.42) was derived by assuming that the main mechanism of heat transfer in nucleate boiling is transient heat conduction to, and subsequent replacement of, superheated layer around boiling sites associated with bubble departure. The correlation incorporates effects of heat transfer surface characteristics through the quantities $c_1$ and $r_s$ and thus allows for direct forms of q vs $\Delta T$ relations.

Following the model as above the heat flux due to the boiling alone from the heated surface has been related to the wall superheat and pertinent fluid properties by the equation:

$$(q)_b = c_1 \frac{c_2 c_3^{1/2}\, r_s^m}{\sqrt{\pi}\; 2^{m-1}} (k\,\rho\, c)_\ell^{1/2} \left(\frac{\lambda\,\rho_v}{T_s\,\sigma}\right)^m \left[\frac{\sigma\, g_o\, g\,(\rho_\ell-\rho_v)}{\rho_\ell^2}\right]^{1/8} \left[\frac{\sigma\, g_o}{g(\rho_\ell-\rho_v)}\right]^{3/4} (Ja^*)^{\frac{15}{8}}\, \Delta T^{m+1} \tag{2.43}$$

Where, $c_1$ = dimensional constant ( 1/unit area)

$c_2$ = $1.5 \times 10^{-4}$ for water

= $4.65 \times 10^{-4}$ for other liquids

$c_3$ = 0.6 is chosen as an average value

$Ja^*$ = Modified Jakob number $\left[\frac{C_\ell\,\rho_\ell\,T_s}{\rho_v\,\lambda}\right]$

m = 2.5 for water

= 3.0 for liquids: n-pentane, benzene and ethyl alcohol

or normalising heat flux:

$$\frac{(q)_b \sqrt{\dfrac{\sigma}{g(\rho_\ell-\rho_v)}}}{\mu_\ell\,\lambda} = B\,(\phi\ \Delta T)^{m+1} \tag{2.44}$$

Where,

$$\phi^{(m+1)} = \frac{k_\ell^{1/2}\ \rho_\ell^{17/8}\ c_\ell^{19/8}\ \lambda^{(m-\frac{23}{8})}\ \rho_v^{(m-\frac{15}{8})}}{\mu_\ell \left[(\rho_\ell - \rho_v)\right]^{9/8}\ \sigma^{(m-\frac{11}{8})}\ T_s^{(m-\frac{15}{8})}}$$

and B is a dimensional constant which depends on boiling surface properties and gravity and is given by

$$B = \left(\frac{\gamma_s J}{2}\right)^m \frac{2}{\sqrt{\pi}} \frac{(g_o)^{11/8}}{(g)^{9/8}} c_2^{5/3}\ c_3^{1/2}\ c_1$$

In order to derive equation, Eq. (2.43) the diameter of bubble size at departure was calculated from the expression of Cole and Rohsenow [44] and $fD_b$ as recommended by Cole [45].

The validity of this equation, Eq. (2.43) was tested by using the experimental data of Addoms [12] , and Cichelli and Bonilla [11] . The proposed correlation was also found to be consistent with the low heat flux data of Gaertner and Westwater [46].

Wiebe and Judd [54] conducted an experimental investigation of temperature profiles in water boiling on a horizontal copper surface for incipient boiling conditions. They used 20,000 , 50,000 and 100,000 Btu/hr.ft$^2$ heat flux while changing the subcooling from 0 to 105 $^o$F. The temperature profiles obtained enabled the extrapolated superheated layer, $\delta$ to be evaluated as follows:

For values of (N/A) f $< 55 \times 10^3$ bubbles/in$^2$-sec.

$$\delta \propto \left[ (N/A) f \right]^{-1/2} \qquad (2.45)$$

For (N/A)f $> 55 \times 10^3$ bubbles / in$^2$- sec

$$\delta \propto \left[ (N/A) f \right]^{-1/3} \qquad (2.46)$$

They have also deduced the following respective equations for heat flux as function of active site density, N/A and frequency of bubble emission, for the above two conditions:

$$\frac{Q}{A} \propto \left[ (N/A) f \right]^{1/2} (T_w - T_\infty) \qquad (2.47)$$

and

$$\frac{Q}{A} \propto \left[ (N/ A) f \right]^{1/3} (T_w - T_\infty) \qquad (2.48)$$

Aladev [55] has studied the mechanism of heat transfer in nucleate boiling . Among many other aspects, he has attempted, the procedure for determining the heating surface temperature from which heat is transported to the boiling liquid. He has successfully correlated the experimental data for pool boiling of distilled water only for pressure range 0.09 to 200 atmospheres by the following equation:

$$\frac{\Delta T_w}{T_s} = 4.7 \times 10^{-3} \left[ \frac{10^{-6} \, q \, \lambda}{k_\ell \, T_s \, g} \right]^{0.3} \left[ \frac{\lambda}{C_\ell \, T_s} \right]^{1.2} \qquad (2.49)$$

## 2.3 SELECTED CORRELATIONS FOR POOL BOILING

Preceding sections contain a chronological literature survey of nucleate pool boiling of liquids. It shows that a large number of correlations for computing boiling heat transfer coefficients have been obtained. These correlations have been found suitable for given boiling liquids and boiling situations namely; wall heat flux, system pressure, surface -liquid combination and boiling fluids. Therefore, in order to compare the present and existing experimental data with the predictions from correlations, it is necessary that these correlations satisfy the following:

(i) The correlations are to be for boiling from the outer surface of a horizontal tube immersed in a pool of fluid.

(ii) The correlations are to be for low heat flux values.

(iii) The correlations are to be for the system at atmospheric and/or subatmospheric pressures.

(iv) The correlations are to be for the boiling liquids whose physico-thermal properties were similar to those of distilled water, isopropanol, ethanol and methanol.

(v) The correlations are to be typical to represent the different major approaches, namely; based on empirical methods, based on the Jakob Linear

Relationship between heat flux and number of active-sites, based on non-linear relationship between heat flux and number of active sites.

The selected correlations are listed in Table 2.6.

TABLE 2.6 Selected Correlations* for Pool Boiling

| Author | System Parameter | | Type of correlation | Remarks |
|---|---|---|---|---|
| | Fluids | Pressure Range | | |
| Jakob and Linke[3] | Water and carbon tetrachloride | Atmospheric and greater than atmospheric | Semi-analytical | Refer Eq.(2.22) |
| Labuntsov [36] | - | Atmospheric | Empirical | Refer Eq. (2.13) |
| Kutateladze [39] | - | 0.2-10 atm. | Empirical | Refer Eq. (2.15) |
| Kichigen and Tobilevich [39] | Many fluids | Atmospheric | Empirical | Refer Eq.(2.16) |
| Forster and Greif [33] | Water,<br>n-butyl alcohol,<br>Aniline,<br>Mercury | 1-50 atm<br>50 psia<br>35 psia<br>1-3 atm | Vapour-liquid exchange mechanism | Refer Eq.(2.37) |
| Mikic and Rohsenow [43] | Water, ethyl-alcohol, n- pentane, & benzene | Wide range of pressure (atmospheric and greater than atmospheric | Analytical | Refer Eq.(2.44) |

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* Equations transformed to as per SI units requirements

## 2.4 SURFACE CHARACTERISTICS AND BUBBLE DYNAMICS

Attempts to correlate nucleate boiling heat transfer data through fluid properties alone have failed, because the surface characteristics have a pronounced effect on the boiling phenomenon. Boiling heat transfer is characterized by the formation of small vapour bubbles at the active-sites and the bubble dynamics - growth rates, frequencies, bubble sizes at departure and interaction with the bulk liquid directly determine the amount of heat transferred. The mechanism of bubble growth rates has been extensively investigated [56-58],[25],[59-86]. The mechanism of bubble departure was investigated originally by Fritz [15] based on a static balance between the buoyant force and the surface tension force at the bubble base. Subsequently, it has been found that the Fritz equation is not of general applicability. With various ramifications the mechanism of bubble departure diameter has been reported successively by many investigators [87-100].

Recently, some attempts have been made to derive theoretical equations for predicting boiling heat transfer rates based on heating surface characteristics, bubble dynamics and boiling mechanisms.

In this section, some of the investigations pertaining to the heating surface characteristics, the bubble growth rate, the bubble diameter at departure and the bubble emission frequency have been dealt with.

### 2.4.1 Surface Characteristics

Kurihara and Myers [52] developed an equation which relates the temperature difference and surface conditions with the active site density. Their correlation failed to predict the active site density at a particular set of conditions. However, it could predict the difference in active site density between two different conditions.

Gaertner [101] undertook the study to relate the active site density to the surface temperature. His study concluded the following equation for site density:

$$(N/A) = N_o \exp(-k/T_w^3) \qquad (2.50)$$

where, k is a constant which is a function of fluid properties and surface conditions. $N_o$ is a constant.

Griffith and Wallis [102] undertook a study to investigate the role of surface conditions in nucleate boiling. For their investigation they made single cavities on different copper surfaces finished with emery paper and obtained boiling data for water, methanol and ethanol. They concluded that the mouth diameter of the cavity determines the superheat needed to initiate boiling, and its shape determines its stability once boiling has begun. On the basis of single-cavity nucleation theory, they proposed that the gross nucleation properties of a given surface for all fluids under all conditions can be characterised with a single group having the dimensions of length. This group $r_c = [\ 2\,\sigma\,T_s\ /\rho_v(T_w - T_s)\lambda\ ]$ when plotted against active site density N/A for a particular surface gave a single plot. Consequently, it was concluded that the nucleation characteristics of a surface would be known if the size distribution of active sites on the surface were known.

Brown [53] used different surfaces with various surface finishes in his studies for bubble nucleation. Brown found that a number of active sites per unit area with radii larger than $r_c$ could be correlated by the following equation:

$$N/A \;=\; c_1 \left( \frac{r_s}{r_c} \right)^m \qquad (2.51)$$

where $r_s$ is a radius for which n would be unity per unit area and $c_1$ is a dimensional constant having dimensions of (unit area)$^{-1}$ and $r_c$ is critical radius equal to $[\ 2\sigma\,T_s/\ \rho_v\,\lambda\ (T_w - T_s)\ ]$ .

Shoukri and Judd [103] have shown that Brown relationship could correlate their experimental data. The parameter group $[2\sigma\ T_s/\rho_v\ \lambda\ (T_w - T_s)\ ]$ is recommended for predicting the minimum nucleation cavity radius and therefore, the usage of this parameter group with Brown correlation is recommended as a sufficient method for describing the nucleation characteristics of a boiling surface.

### 2.4.2 Bubble Growth Rate

The process of nucleate boiling is the sum total of the processes of bubble initiation, growth and departure. Bubble growth rates have been investigated quite extensively for fluids such as water and methanol boiling from solid surfaces under conditions of atmospheric pressure or greater corresponding to Jakob numbers less than 50 [88], [104-106]. However, considerably little attention has been paid to growth rates at subatmospheric pressure conditions.

Zuber [62] studied the bubble growth rate on solid surfaces. Zuber expression for bubble growth is as follows:

$$D = \phi_c \frac{4}{\pi} J_a \sqrt{\pi \alpha t} \left[ 1 - \frac{q \sqrt{\pi \alpha t}}{2k_\ell (T_w - T_s)} \right] \qquad (2.52)$$

Hsu and Graham [105] recommended the following bubble growth equation in a non-uniform temperature field:

$$D = \frac{\pi}{\lambda \rho_v} \left\{ \frac{qt}{2\pi} + \frac{2q \delta^2}{\alpha \pi^2} \left( \frac{\pi^2}{6} \sum_{n=1}^{\infty} \frac{1}{n^2} \exp\left[ -\left(\frac{\pi n}{\delta}\right)^2 \alpha t \right] \right) \right\}$$

(2.53)

Han and Griffith [106] took into consideration the curved surface of the bubble covered by a hot liquid microlayer. Their growth equation is as follows:

$$D = \frac{\emptyset_s \emptyset_c}{\emptyset_v} \quad \frac{2k_\ell}{\lambda\ \rho_v} \left\{ \frac{2(T_w - T_s)}{\sqrt{(\pi\ \alpha)}}\, t^{1/2} - \frac{(T_w - T_o)}{\delta} \quad \frac{\delta^2}{4\alpha} \right.$$

$$\left. \left( \frac{4\alpha t}{\delta^2}\ \mathrm{erf}\, \frac{\delta}{\sqrt{4\alpha t}} + \frac{2}{\sqrt{\pi}} \quad \frac{\sqrt{4\ \alpha t}}{\delta} \exp \left[ - \frac{\delta^2}{4\alpha\ t} \right] - \mathrm{erf}\ C\, \frac{\delta}{\sqrt{4\alpha t}} \right) \right\} \tag{2.54}$$

Mikic and Rohsenow [74] developed the following equation for bubble growth in non-uniform temperature field:

$$R^+ = (\tau^+)^{1/2} \left[ 1 - \theta_1 \left\{ \left(1 + \frac{\tau_w^+}{\tau^+}\right)^{1/2} - \left(\frac{\tau_w^+}{\tau^+}\right)^{1/2} \right\} \right] \tag{2.55}$$

Cole and Shulman [66] have scrutinized the existing theories for bubble growth in a non-uniform temperature field for subatmospheric conditions. Comparison of all of the non-uniform temperature field growth equations [62] [63], [105], [106] with the experimental data indicates them to be less satisfactory than the uniform superheat expression given by $D = Ja \sqrt{\pi\ \alpha\ t}$. As a matter of fact the reasonable agreement with the magnitude of the growth data is obtained only for Jakob numbers less than 100. Above this value these authors obtain increasingly greater discrepancy between theory and experiment. Finally they recommended two separate expressions for bubble growth in an non-uniform temperature field as follows:

For $Ja < 100$

$$D = Ja \sqrt{\pi \alpha t} \qquad (2.56)$$

For $Ja > 100$

$$D = 5\, Ja^{3/4} \sqrt{\alpha t} \qquad (2.57)$$

Stewart and Cole [83] have investigated the bubble growth rates experimentally in order to determine the effect of high Jakob number conditions. In their studies Jakob number ranged from 955 to 1112. This study concludes that liquid inertia is a major factor restricting the rate of growth of vapour bubbles at low pressures.

van Stralen et al [85] have investigated experimentally the growth rate of vapour bubbles upto departure in water boiling at pressures varying from 26.7 to 2.0 kPa which correspond to Jakob number range from 108 to 2689 . During initial growth the results of Stewart and Cole [83] as regards the influence of liquid inertia are established. Experimental bubble growth is in quantitative agreement with the van Stralen et al theory [84].

### 2.4.3 Bubble Departure Diameters at Subatmospheric Pressure

Cole and Shulman [95] conducted a careful literature review for the expressions for bubble departure diameter [15], [87-94] which were developed either theoretically or empirically. They recommended that none of the correlations tested were able to satisfactorily represent subatmospheric departure diameter data. As a result of the fact that none of the correlations are very satisfactory, it is recommended that the following equation of dimensionless departure diameter with pressure be employed to obtain order of magnitude estimates of the departure diameter:

$$\frac{D_b}{\left[\frac{\sigma}{g(\rho_\ell - \rho_v)}\right]^{1/2}} = \frac{133.3}{P} \qquad (2.58)$$

where P is in $kN/m^2$.

This expression covers the widest range of experimental conditions and does not require a knowledge of heat flux level or wall superheat. In addition it also satisfies the equation recommended by Nishikawa and Urakawa [89]. Their equation is as follows:

$$D_b = 0.672\ P^{-0.575} \qquad (2.59)$$

Where $D_b$ is in inches and P is in psia.

Departure equation proposed by Cole [45] shows that the departure diameter is directly proportional to wall superheat. The equation is :

$$\frac{D_d}{\sqrt{\frac{\sigma}{g(\rho_\ell - \rho_v)}}} = 4 \times 10^{-2}\ Ja \tag{2.60}$$

Recently Cole and Rohsenow [44] have attempted to modify the previous correlations, proposed by Cole and Shulman [95] and Cole [45] for bubble departure diameter. These investigators argued that wall superheat, which appears in Jakob number, was not available for high pressure data and thus previous correlations could not be tested for the pressures greater than one atmospheric due to the lack of this information.

In this attempt Cole and Rohsenow have proposed a modified Jakob number which is free from the wall superheat term:

$$Ja^* = \frac{\rho_\ell\, c_\ell\, T_s}{\rho_v\, \lambda}$$

and they have recommended the correlations for bubble departure diameter by using this modified Jakob number. In their support to delete the wall superheat term they have mentioned the previous work [104], [107], [108], where it was found that corresponding to wall superheats ranging from 20° to 30°F, the arithmetic mean departure diameter to be approximately constant. The work of Preckshot and Denny [96], where wall superheat range was 40 to 70 °F, found an increase in departure diameter with increase in wall superheat and a decrease in this diameter with increase in active nucleation site density. Thus a surface consisting of natural nucleation sites, where both the wall superheat and bubble population density increase with heat flux level, would essentially yield a constant departure diameter due to the counteracting of each others effect. Cole and Rohsenow thus advocate that either the effect of bubble population density should be taken into account or the variation with a wall superheat term should be removed to define the bubble departure diameter. Because of the fact that no detailed experimental studies for the effect of bubble population on departure diameter is available, the Jakob number has been modified in a manner as mentioned above.

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Based on the experimental departure diameter data of various workers they have proposed two different correlations for water and other liquids.

The correlation for water is based on the experimental departure data of Semeria [91] for pressures 2 to 140 atmospheres, Tolubinskii and Ostrovsky [70] for pressures ranging from approximately 0.2 to 10 atmospheres, Cole [45] for pressures ranging from 0.066 to 0.470 atmospheres, Hatton and Hall [108] for pressures ranging from 0.12 to 1 atmosphere and Siegel and Keshock [109] for 1 atmosphere.

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The correlation is:

$$\frac{D_d}{\sqrt{\dfrac{\sigma\, g_c}{g(\rho_\ell-\rho_v)}}} = 1.5 \times 10^{-4}\, (Ja^*)^{5/4}; \text{ for } Pr < 0.2 \qquad (2.61)$$

And the correlation for other liquids is based on the experimental departure diameter data of Wanninger [110] for propane at pressures of 8.5, 11 and 14 atmosphere and iso-pentane at 1 atm; Cole [45] for toluene, n-pentane methanol, carbon tetrachloride and acetone at pressures ranging from 0.066 to 1 atm, Tolubinskii and Ostrovsky [70] for benzene, n-butyl alcohol and Freon- 12 at 1 atmosphere; McFadden and Grassmann [111] for nitrogen at 1 atmosphere; Perkins and Westwater [114] for methanol at 1 atmosphere; Preckshot and Denny [96] and Jakob [3] for carbon tetrachloride at 1 atm. The correlation is:

$$\frac{D_d}{\sqrt{\dfrac{\sigma\, g_c}{g(\rho_\ell-\rho_v)}}} = 4.65 \times 10^{-4} (Ja^*)^{5/4}; \quad \text{for } Pr < 0.2 \qquad (2.62)$$

The deviation of experimental data from either equation is approximately ± 50%. The authors point out that the deviation is due to result of neglecting the bubble population density, dynamic contact angle, active cavity size and wall superheat.

### 2.4.4 Bubble Emission Frequency at Subatmospheric Pressures

The availability of accurate and precise correlations for bubble emission frequency, f , and bubble departure diameter, $D_b$ provides a good opportunity for deriving equations for heat transfer coefficient analytically. The determination of relationship between f and $D_b$ has been a subjecst of active research. Consequently a good amount of information [3],[19],[70],[87],[88], [106], [108],[109], [111-120] exists which contains results based on one or other approaches. Recently Ivey [118] has attempted to scrutinize all the available investigations and put them in a rational form. He concludes that a single expression may not adequately correlate bubble frequency with bubble diameter for all bubble diameters in nucleate boiling. He suggests three separate regions with which the experimental data are better correlated. The regions are hydrodynamic, transition and thermodynamic.

It seems that exclusive correlations for bubble emission frequency are non-existent but for the correlations due to Han and Griffith [106], Hatton and Hall [108] and Saini [86].

Han and Griffith [106] have studied the mechanism of heat transfer in nucleate pool boiling. In addition to bubble initiation, growth and departure they recommended the calculation of bubble emission frequency by the following equation:

$$f = \frac{1}{\theta_d + \theta_w} \qquad (2.63)$$

In equation (2.63) waiting period $\theta_w$ is expressed by the following expression:

$$\theta_w = \frac{\delta^2}{\pi \alpha} = \frac{g}{4 \pi \alpha} \left[ \frac{(T_w - T_\infty) R_c}{T_w - T_{sat} \left[1 + (2\sigma / R_c \rho_v \lambda)\right]} \right]^2 \qquad (2.64)$$

and departure time $\theta_d$ can be obtained by solving following equations:

$$R_d = 0.4251 \; \phi \left[ \frac{2\sigma}{g(\rho_\ell - \rho_v)} \right]^{1/2} \qquad (2.65)$$

$$R - R_c = \frac{\phi_s \phi_c}{\phi_v} \; \frac{\alpha c \rho}{\rho_v \lambda} \left[ \frac{2\theta_w}{(\pi\alpha)^{1/2}} t^{1/2} - \frac{\theta_w - \theta_\infty}{\delta} \; \frac{\delta^2}{4\alpha} \left( \frac{4\alpha t}{\delta^2} \; \mathrm{erf} \; \frac{\delta}{(4\alpha t)^{1/2}} + \frac{2}{(\pi)^{1/2}} \; \frac{(4\alpha t)^{1/2}}{\delta} \exp\left[- \delta^2 / 4\alpha t\right] - 2 \; \mathrm{erfc} \; \frac{\delta}{(4\alpha t)^{1/2}} \right) \right] + \frac{\phi_h h_v \theta_w}{\phi_v \rho_v \lambda} t$$

(2.66)

Hatton and Hall [108] used the bubble growth equation of Plesset-Zwick [57] for departure time, $\theta_d$ and

Hsu [121] and Han - Griffith [106] expression for bubble nucleation for calculating the bubble emission frequency under the assumption of $\theta_w = 0$ . The final form of the equation is as follows :

$$f = \frac{1}{\theta_d} = \frac{3}{\pi \alpha} \left[ \frac{16 \, k_\ell \, \sigma \, T_s}{(\lambda \rho_v)^2 \, D \, D_c} \right]^2 \qquad (2.67)$$

Cole [45] has proposed the expressions for bubble frequencies and departure volumes for nucleate pool boiling of liquids at subatmospheric pressures. His expressions are the modifications over the expression which have been proposed by previous workers, who have not taken into account the effect of pressure and heat flux over the volumetric vapour flow rate ($f D_b^3$) . Recent experimental investigations [120] has determined that the volumetric vapour flow rate for the isolated bubble region in nucleate boiling is a function of both the heat flux and pressure. Thus in establishing the expressions for $fD_b^3$ or $fD_b$ these variables should be taken into account. Since the heat flux is proportional to the degree of superheat, and system pressure can be taken into account through vapour density term, the Jakob number can be included while developing the expressions for $fD_b^3$ or $fD_b$.

With the aid of his own experimental data and the data of other workers Cole proposed the following expressions:

$$fD_b^3 \propto \left[\frac{\sigma^{5/3}}{g\,\rho_\ell^{2/3}\,(\rho_\ell - \rho_v)}\right]^{3/4} \left[\frac{\rho_\ell\, C_\ell\, \Delta T}{\rho_v\, \lambda}\right]^{2} \qquad (2.68)$$

where,

$$\frac{\rho_\ell\, C_\ell\, \Delta T}{\rho_v\, \lambda} = Ja$$

and the volumetric vapour flow rate per cross section is

$$fD_b \propto \left[\frac{g(\rho_\ell - \rho_v)\sigma}{\rho_\ell^2}\right]^{1/4} \qquad (2.69)$$

which is independent of Jakob number.

Based on the bubble departure diameter expressions of Saini et al [122] Saini [123] developed three different expressions for bubble emission frequency. His expressions for different range of Jakob number are as follows:

For $Ja \leqslant 16$

$$f = \frac{1}{\dfrac{0.865}{\alpha_\ell}\left[\dfrac{k_\ell\, \Delta T}{q}\right]^2 * \left[\dfrac{\alpha_\ell}{g^2}\right]^{1/3} \left[\dfrac{6.6\, C_\ell \Delta T\, \sigma}{\alpha_\ell\, q}\right]^{2/3} / (\pi\, Ja^2)} \qquad (2.70)$$

For $16 < Ja < 100$

$$f = \frac{1}{\dfrac{0.865}{\alpha_\ell}\left[\dfrac{k_\ell \Delta T}{q}\right]^2 + \left[\dfrac{\alpha_\ell}{g^2}\right]^{1/3} \left[0.578\, Ja^{2/3} \left\{1.22 + \sqrt{1 + 2.67\,\dfrac{C_\ell \Delta T \sigma}{\alpha_\ell q\, Ja^4}}\right\}^{4/3}\right]} \qquad (2.71)$$

For Ja $\geqslant$ 100

$$f = \frac{1}{\frac{0.865}{\alpha_l}\left[\frac{k_l \, \Delta T}{q}\right]^2 + \left[\frac{\alpha_l}{g^2}\right]^{1/3} (3.4\ Ja^{1/2})} \qquad (2.72)$$

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# CHAPTER 3

## EXPERIMENTAL APPARATUS

### 3.1 DESIGN CONSIDERATIONS

As a large diameter heating surface was employed in the present investigation, it was expected that the variation in surface temperature around the circumference of the heating surface might exist. This could also be able to furnish the information about the changes in local values of heat transfer coefficient. Therefore, an important consideration was the location of surface thermocouples at different circumferential positions. Considering the bubble dynamics on the heating surface it would be required that the temperatures are measured at least at the top, at the side and at the bottom positions of the surface. This information can also be utilised for averaging the surface temperature by the method of mechanical quadrature.

Boiling fluid temperature changes around the heating surface and the heating surface is enveloped by the superheated layer of the liquid. Therefore, another important consideration was as to where measure the temperature of the boiling fluid. It was also necessary to measure liquid temperature corresponding to three above mentioned positions of wall thermocouples so that local values of heat transfer coefficient could be calculated. For calculating heat transfer coefficient, the temperature of the fluid was measured at a point outside the superheated layer. It is important at this stage to mention that the thickness of superheated liquid layer does not remain constant and it changes with the parameters namely; heat flux, pressure and properties of boiling fluids. Therefore, it was required to vary the position of the fluid thermocouples depending upon system parameters and conditions. To meet these requirements movable thermocouple probes are required.

The apparatus used in the present study was designed to conduct experiments on binary liquid mixtures as well. Therefore, design consideration was made for drawing the liquid sample and the equilibrium vapour-condensate. To keep the composition of boiling mixture at a given value it was necessary to return the vapour condensate from the condensers back to the pool of boiling fluid. The vapour condensate return line and the liquid sampling line could alter the hydrodynamic conditions in the neighbourhood of the heating surface. Therefore, special care was taken to keep these lines sufficiently away from the heating surface.

## 3.2 EXPERIMENTAL FACILITY

An experimental facility for studies in nucleate pool boiling was designed and fabricated. It is depicted schematically in Figure 3.1 and a photographic view is shown in Figure 3.2. The facility comprises mainly of

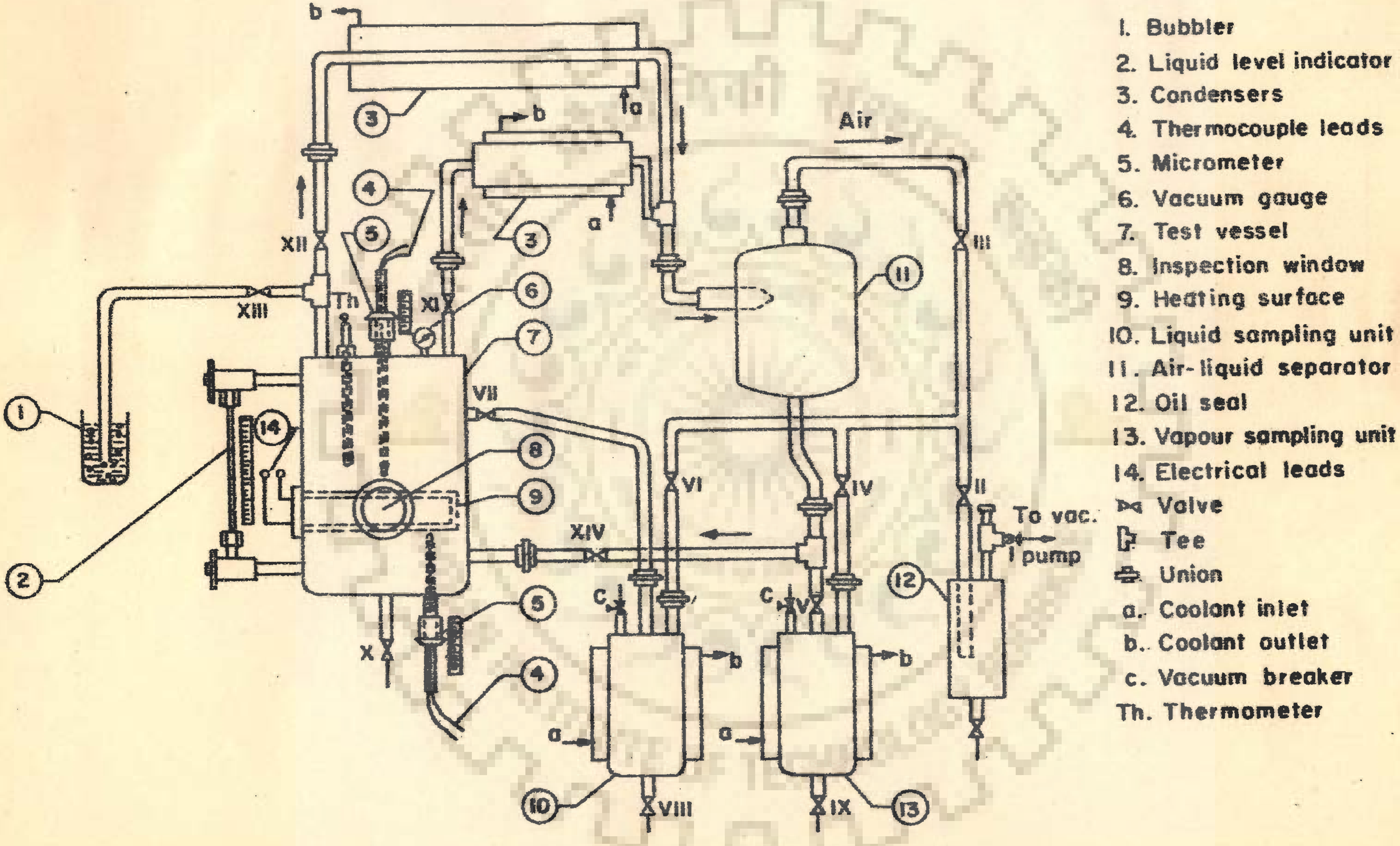


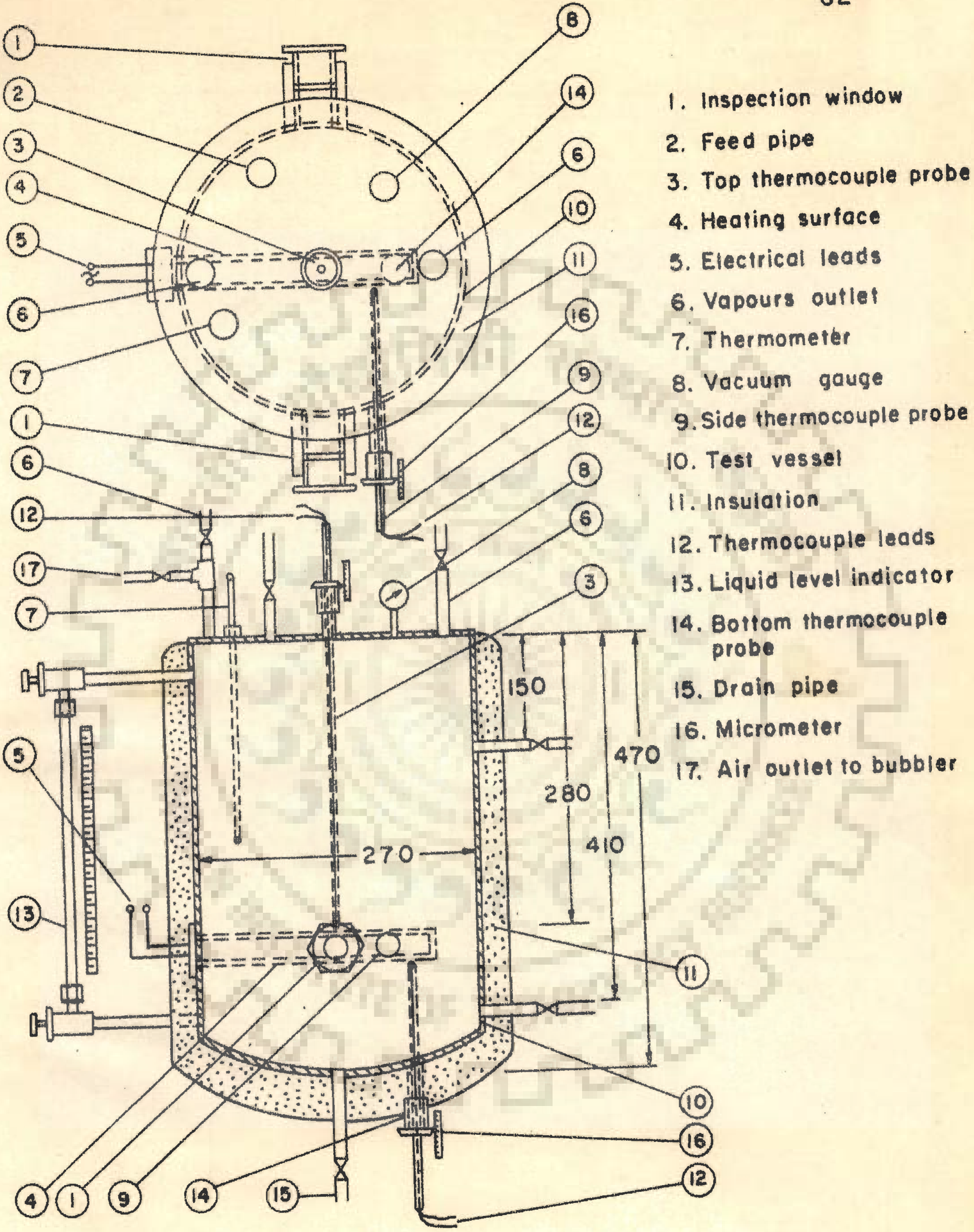
Fig 3·1 Schematic diagram of the experimental facility

Fig. 3·2 Photographic view of the experimental facility following components:

- Test vessel
- Heating surface
- Liquid thermocouple probes
- Vacuum and condensing system
- Air-liquid separator and sampling units

### 3.2.1 Test Vessel

The details of the test vessel are shown in Figure 3.3 and a photographic view in Figure 3.4. It was a cylindrical vessel made out of 304 ASIA stainless steel sheet having a welded flat top and dished bottom. The diameter of the vessel was 240 mm and height 470 mm. The vessel was filled upto a capacity of 0.017 $m^3$. Figure 3.3 shows the locations of the thermocouple probe (3), the thermometer (7) and the vacuum gauge (8) which were mounted on the vessel top (8). It also had provisions (2),(6) and (17), for charging the test fluid into the vessel, for carrying the vapours from the test vessel to the condensers, and for removing the dissolved air from the boiling liquid, respectively. A liquid level indicator (13) was attached to the left side of the vessel. The heating surface (4), over which boiling occurred, was installed horizontally within the vessel at a submergence depth of about 280 mm as photographed in Figure 3.4. The heating surface (4) was fitted to the



1. Inspection window
2. Feed pipe
3. Top thermocouple probe
4. Heating surface
5. Electrical leads
6. Vapours outlet
7. Thermometer
8. Vacuum gauge
9. Side thermocouple probe
10. Test vessel
11. Insulation
12. Thermocouple leads
13. Liquid level indicator
14. Bottom thermocouple probe
15. Drain pipe
16. Micrometer
17. Air outlet to bubbler

All dimensions in mm

Fig. 3·3 Details of test vessel

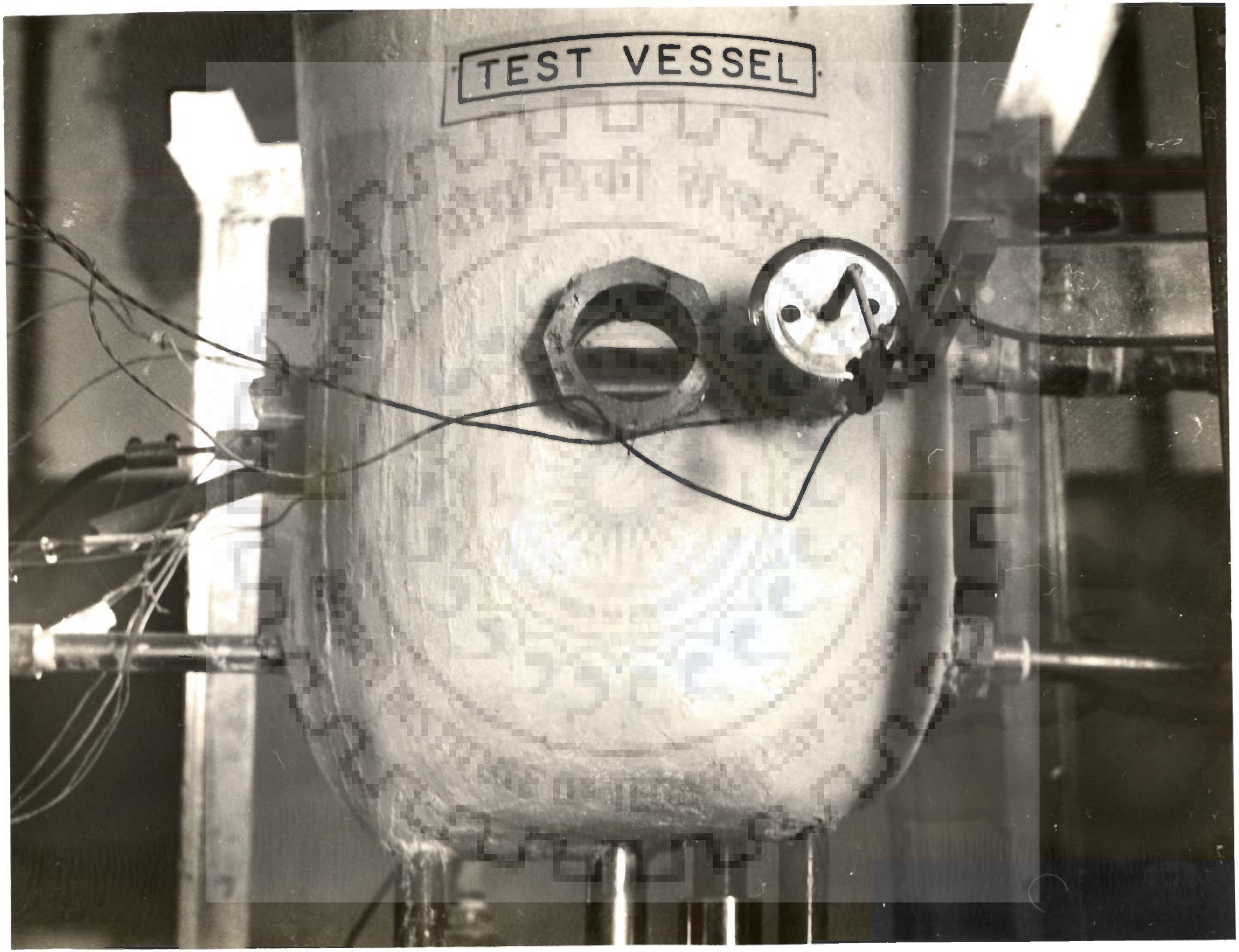


Fig. 3·4 Photographic view of the test vessel and the heating surface test vessel (10) from the left of the vessel as shown in Figure 3.3. The details of the arrangement for holding it in horizontal position is shown in Figure 3.6. Two diametrically opposite inspection windows (1) were provided at the front and rear side of the vessel for viewing the bubble dynamics. The details of the inspection window assembly is shown in Figure 3.5. Liquid thermocouple probes at the side (9) and the bottom (14) positions were attached to the vessel as depicted in Figures 3.3 and 3.4. The details of the thermocouple probes are described in Section 3.2.3. To minimize the heat losses to surroundings the vessel body except the top flange was thoroughly insulated by means of glass wool and a mixture of 85% magnesia powder and plaster of paris(11). To drain the liquid from the test vessel a drain pipe(15) with a valve was provided at the bottom of the vessel. Suitable pipe lines and control valves were also provided for liquid sampling and return of vapour condensate to the vessel as shown in Figures 3.1 and 3.3.

### 3.2.2 Heating Surface

Figure 3.4 provides a photographic view of the heating surface submerged in the pool of boiling fluid. Figure 3.6 shows the details of the heating surface, It consists of a 410 ASIS stainless steel cylinder (6) 70 mm outside diameter and 4 mm thick. The total

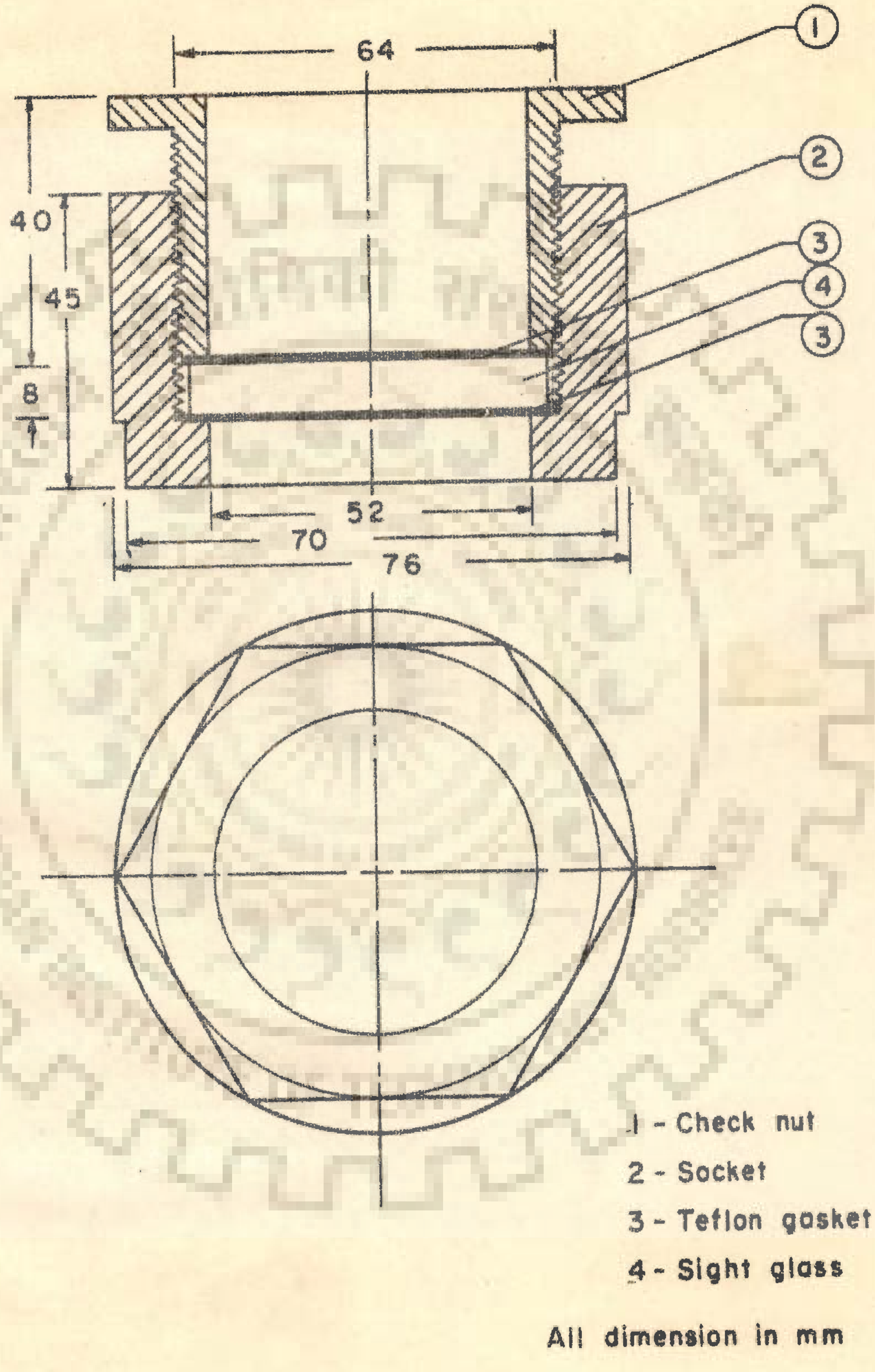
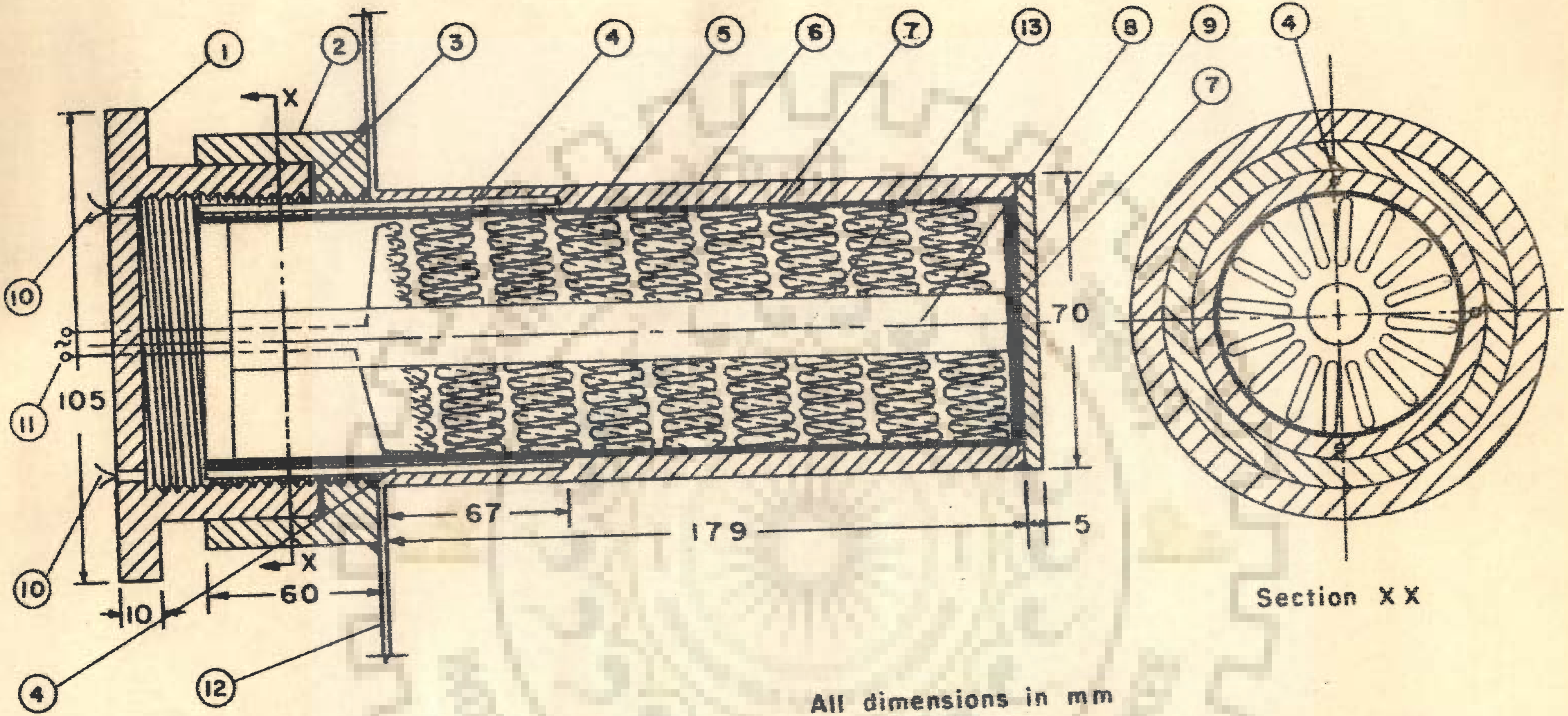


Fig. 3·5 Inspection window assembly



All dimensions in mm

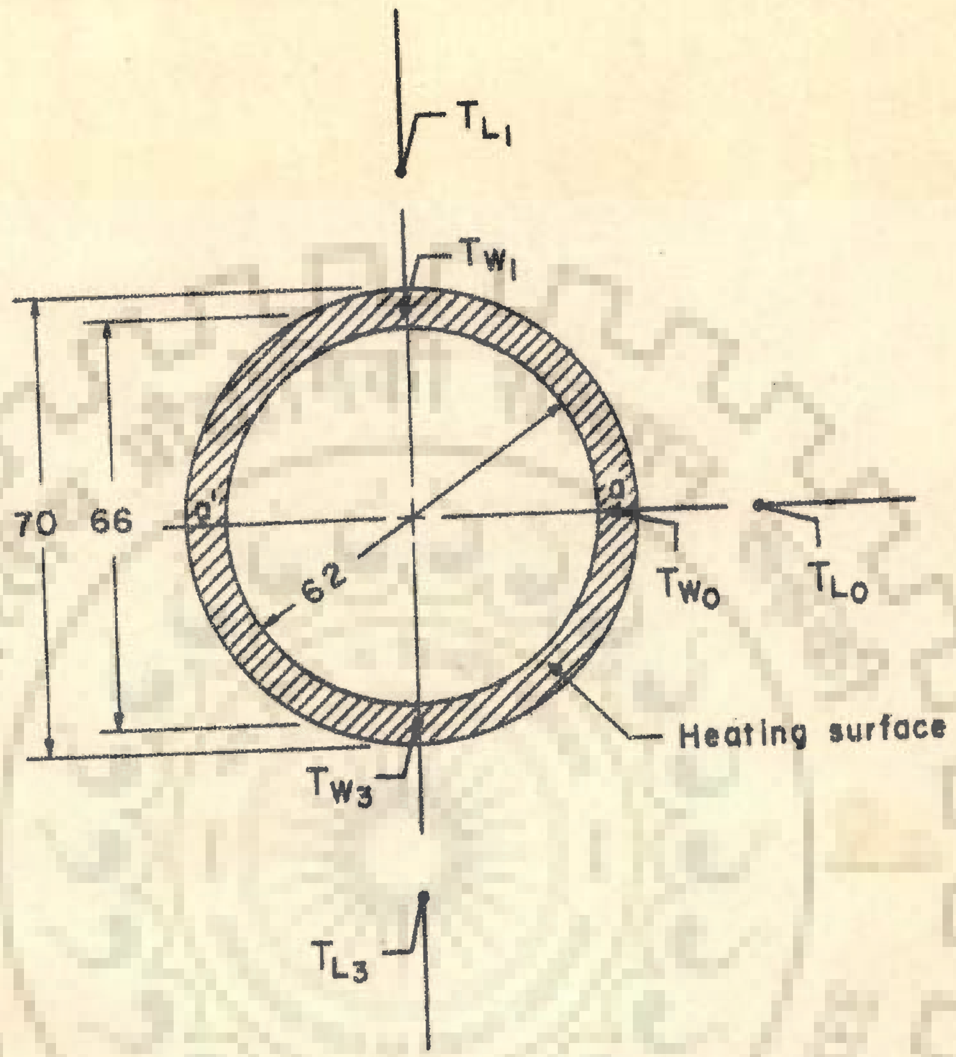
1. Check nut
2. Socket
3. Lead gasket
4. Thermocouple
5. Heating element
6. Heating surface
7. Mica sheet
8. Porcelain core
9. End plate
10. Thermocouple leads
11. Electrical leads
12. Test vessel wall
13. Asbestos rope

Fig. 3·6 Details of heating surface

length of the heating surface was 244 mm, out of which 60mm was threaded portion and 5 mm end plate (9) thickness. The outer surface of the heating surface was uniformly turned and polished by rubbing it against 0/0 emery paper. It was supported horizontally in the test vessel by means of a socket (2) welded at the left side of the vessel. A check nut (1) with lead gasket (3) was also provided to make the system leak proof.

Design considerations, as detailed in Section 3.1 dictate that three thermocouples were sufficient to know the variation of heat transfer coefficient circumferentially and to have average values of the surface temperatures. Figures 3.6 and 3.7 show the locations of these thermocouples at the side, at the top and at the bottom of the heating surface. The temperature at a' was not measured since it is equal to temperature at a, due to the symmetry of boiling conditions about the horizontal mid-plane a-a'. In Figure 3.7 the respective positions of liquid thermocouple probes are are also depicted. Three axial holes of 127 mm depth were drilled in the wall of heating surface at the above mentioned positions to place the thermocouples. Calibrated fibre-glass insulated copper constantan thermocouple wires of 24 gauge were used.

The heating surface was heated by an electric heater. A cartridge heater was designed and fabricated which provided a heat flux upto a value of 45,000 W/m$^2$. Nickle-Chrome



All dimensions in mm

T - Thermocouple

W - Wall

L - Liquid

0, 1 and 3 represent the side, top and bottom positions respectively

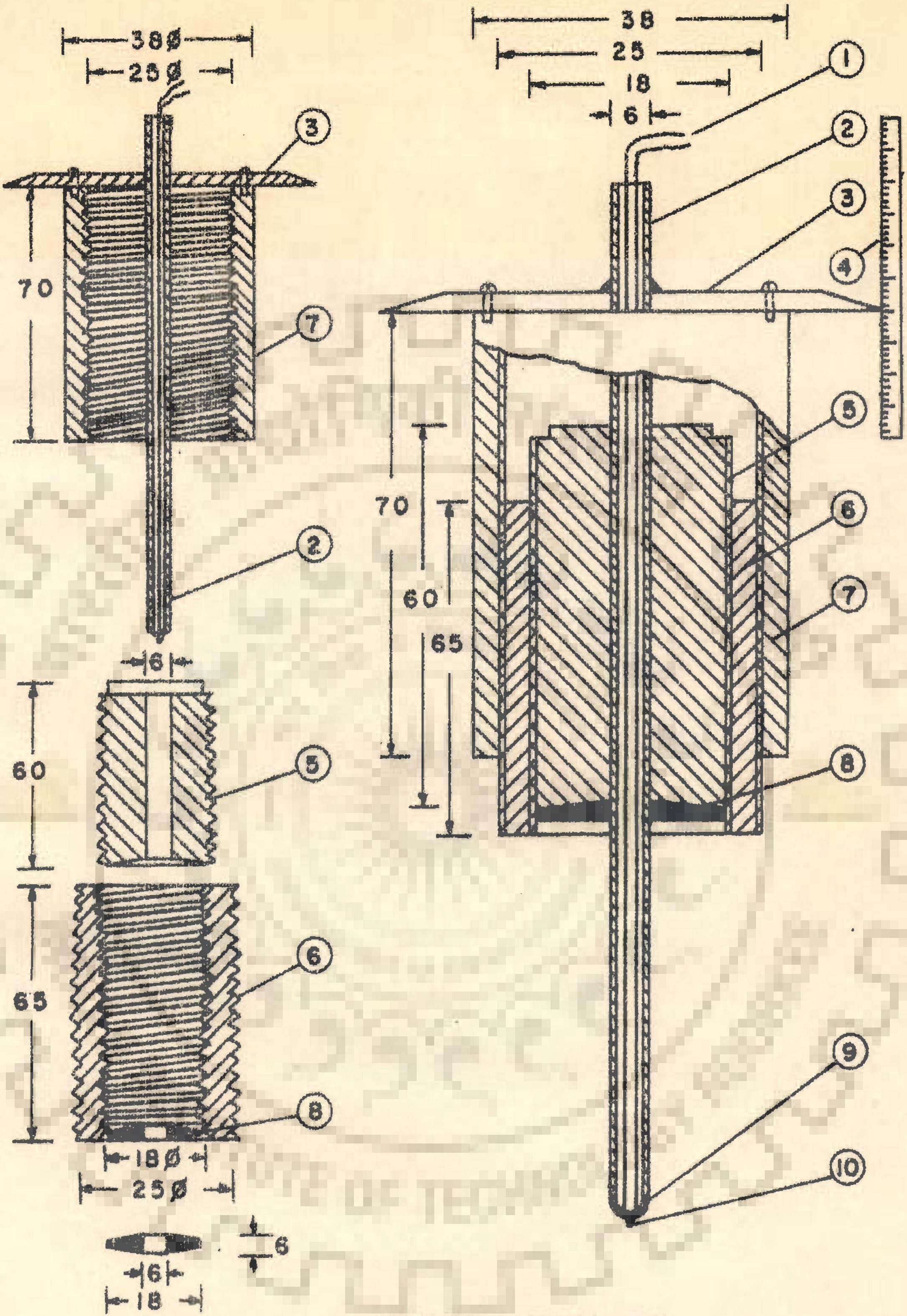
Fig. 3·7 Thermocouples layout on heating surface and in boiling liquid wire of 14 gauge having a maximum current carrying capacity of 24 amperes constituted the heating element (5). The coil (5) made of this element was wounded on a porcelain core(8) of 16 mm diameter. A fine asbestos rope (13) was provided between the two successive turns of the heating element for insulating them. The entire heating coil (5) was then thoroughly wrapped in a thin mica sheet (7) which ensured a complete electric insulation between the element and the heating surface. This cartridge heater was then placed inside the heating surface.

### 3.2.3 Liquid Thermocouple Probe

Design considerations, as detailed in Section 3.1 necessitated the measurements of fluid temperatures by moving thermocouple probes.

The details of the main components of the probe and the assembled probe are shown in Figure 3.8. The various components were the nipple (6), the check nut (5), the thermocouple carrying tube (2), the movable socket (7) with a graduated circular disc (3) and the fixed scale (4). All these components were made of stainless steel.

The nipple (6) was 65.0 mm long having 25.0 mm outside diameter and 18.0 mm inside diameter. It had full threads on its outer and inner surfaces. The outer surface was fine threaded having a pitch of 1 mm . The diameter of the vessel-end of the nipple was reduced to 6 mm, equal to



All dimensions in mm

1. Thermocouple leads
2. Thermocouple carrying tube
3. Circular scale disc
4. Vertical scale
5. Check nut
6. Nipple
7. Movable socket
8. Teflon gasket
9. Silver soldering
10. Bead

Fig. 3·8 Details of liquid thermocouple probe the outer diameter of the thermocouple carrying tube. In order to eliminate the leakage of the fluid through the vessel- end of the nipple, the check nut (5) with inside diameter of 6 mm and having threads on its outer surface matching with inner threads of the nipple (6) was fitted. The gasket (8) made of teflon was used to press against the thermocouple carrying tube (2) and thus making the system leak proof.

The thermocouple carrying tube was of 6 mm outside diameter. It had copper constantan thermocouple wires(1) of 24 gauge passing through it. The bead end (10) of the tube was silver soldered (9). The entire lengths of the thermocouple wires were sheathed with twin bored porcelain tube of 4 mm diameter having bore diameter of 0.8 mm. This ensured against any possibility of secondary junction of thermocouples inside the tube. The thermocouple wires were taken out of the other end of the tube. The thermocouple carrying tube was welded to the movable socket (7) and thus the tube traversed within the pool by means of the movable socket.

The movable socket was cylindrical in shape having fine threads of 1 mm pitch at its inner surface and having a graduated circular disc at its top. These threads matched with those on the outer surface of the nipple (6). Thus the movable socket (7) could move along the nipple and thereby thermocouple tube (2) traversed in the pool of boiling liquid. The traversed distance of the thermocouple carrying tube was measured by means of graduated movable circular scale (3) and the fixed scale (4) as shown in photograph 3.4 . The circular disk had 200 graduations on its periphery and thus a distance upto 7.5 x $10^{-3}$ mm could be read.

### 3.2.4 Vacuum and Condensing System

The vapours from the boiling fluids condensed externally and returned back to the pool of liquid. The condensation of the vapours was carried out in two countercurrent water-cooled condensers with the condensing vapour in the inner pipe and coolant in the annulus. These condensers(3) were located at the top of the vessel as shown in Figure 3.1 . They were pitched towards the air-liquid separator and were connected in parallel as shown in Figures 3.1 and 3.9. The condensers were designed to cause total condensation of the vapours. The separated condensate from the seperator (11) flowed through a pipe and returned to the pool of boiling fluid.

Vacuum was created by means of a vacuum pump of a "HV" series Hindustan Rotary two-stage oil immersed type with suction capacity of 125.025x$10^{-5}$ $m^3/s$. The pump was driven by a 0.37 kW motor having 1450 rpm. One of the essential features of the pump was an air Ballast which enabled the pump to attain high vacuum even when a lot of moisture and organic vapours were sucked in by the

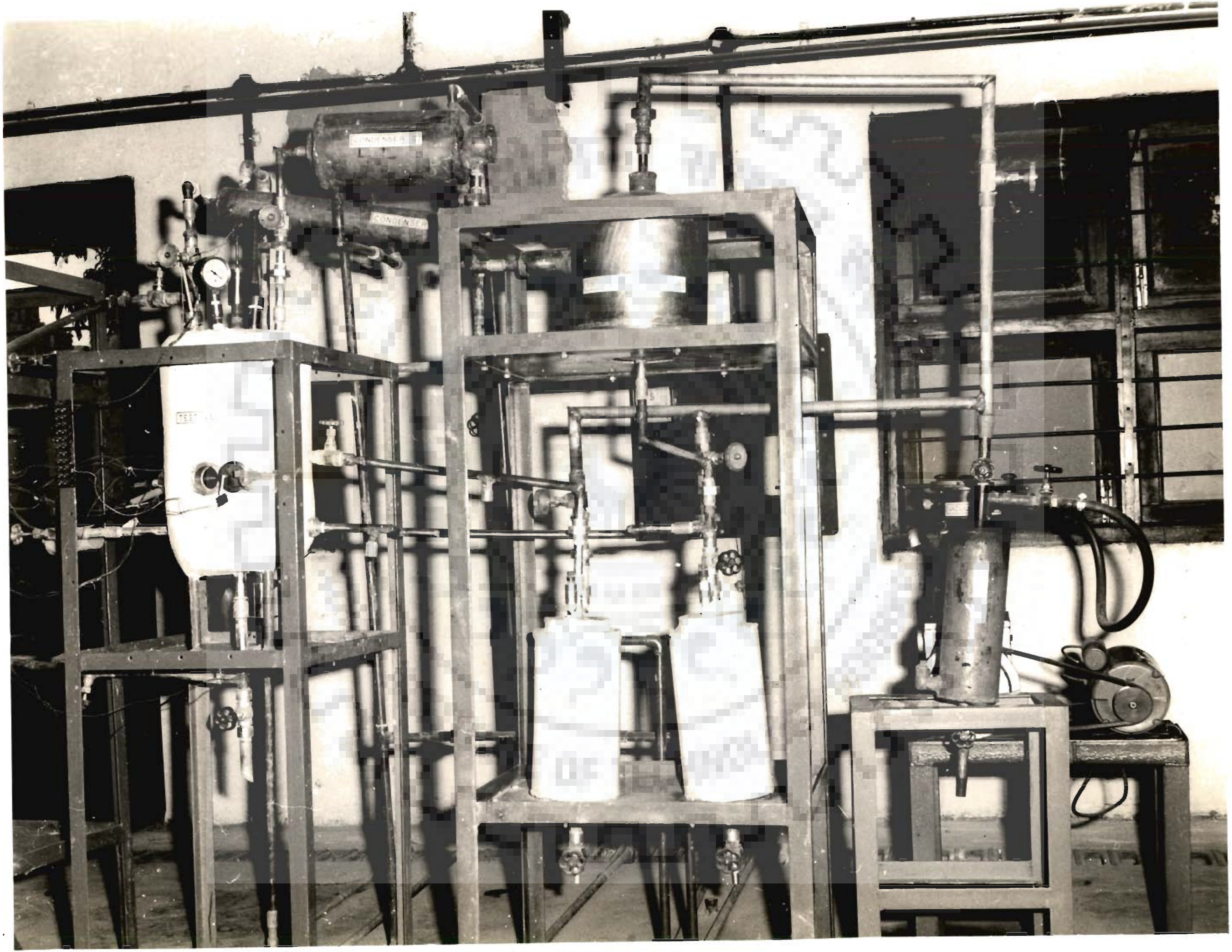


Fig. 3.9 Photographic view of the vacuum, condensing system and sampling units

pump. Drops of water particles which were released under high compression ratios, of the order of 1:700 and collected underneath the main valves were completely eliminated by the introduction of fresh atmospheric air through the air Ballast vent. The pump was protected from the organic vapours and moisture by means of an absorbent placing in suction inlet. An oil seal (12) was also provided for this purpose. To check the back flow of oil into the apparatus, valves (I) and (II) were provided at suitable locations. Pressure regulation was done with a fine needle valve (III).

### 3.2.5 Air-Liquid Separator and Sampling Units

An air-liquid separator (11) was placed between the condensers (3) and the vacuum unit as shown in Figures 3.1 and 3.9 . It was a cylindrical vessel made of stainless steel having a tangential entry for the air-liquid mixture from the condensers. The separated air passed to the vacuum pump through the pipe at the top of the separator while the condensate to the pool of boiling fluid.

Provision was made in this experimental facility for drawing out the samples of boiling fluid and the equilibrium vapours. The sampling vessels (10)& (13) had jacketted wall for the coolant to flow through the jacket in order to cool the hot samples of the fluid. A vent cock (C) provides the facility to break the vacuum in these vessels in order to take the samples from these units. Liquid sampling unit was connected to the pool of the boiling fluid with a precise control valve (vii) and the vapour sampling unit to the separator with a control valve (v). By closing the valves(iv), (v), (vi) and (vii) these units could be isolated from the rest of the apparatus. The liquid sample could be drawn by opening valve (vii) and the condensate sample by valve (v). Cool samples could be drawn out from the bottom of the vessels with the help of valves (viii) and (ix).

## 3.3 INSTRUMENTATION

The photographic view of the instrumentation used is shown in Figure 3.10. The heat to the heating surface was supplied by means of single-phase 50 c/s alternating current. Electric power input to the heating element was measured by means of calibrated precision grade ammeter and voltmeter of accuracy less than 1 per cent. The range of ammeter was from 0 to 5 amperes and that of voltmeter from 0 to 300 volts. In the present investigation the current exceeded 5 amperes. Therefore, current transformer was employed. The power supply to the element was modulated by an autotransformer. The arrangement of the electric circuit is shown in Figure 3.11. The readings of voltmeter and ammeter were noted in order to calculate the power input to the element. The power divided by the heat transfer area of the heating surface represented the heat flux.

Fig. 3·10 Photographic view of the instrumentation

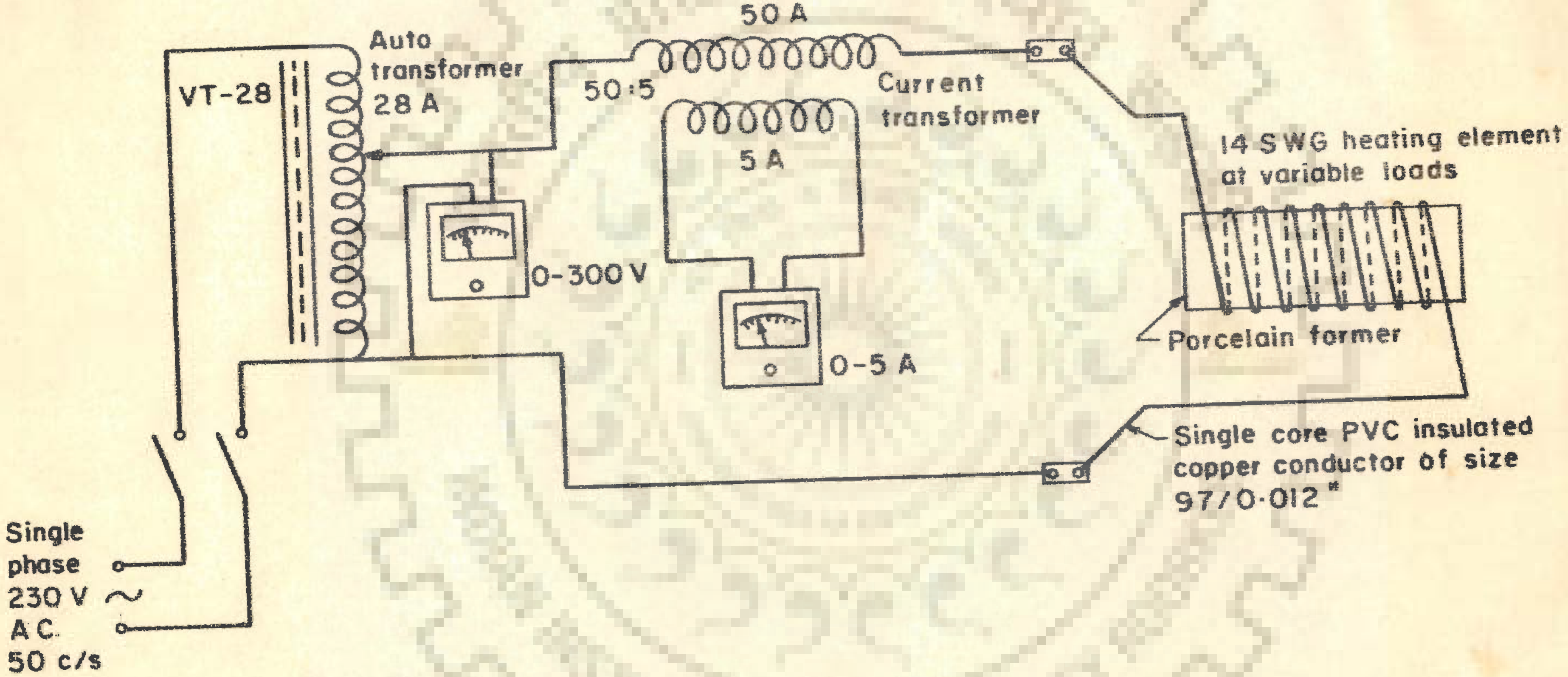


Fig. 3·11 Electric circuit for the heating element

The electromotive force of thermocouples was measured by a vernier potentiometer with a sensitive spot galvanometer. The potentiometer could read e.m.f. upto 0.001 mV with an accuracy of 0.01 per cent. A multipoint selector switch supplied by M/S Toshniwal was used to connect the thermocouples to the potentiometer. A distilled water-ice bath provided a reference temperature.

To accomplish the measurements of vacuum in the system a calibrated precision grade vacuum gauge was mounted on the top of the vessel. This pressure determines the saturation temperature of the system.

# CHAPTER 4

## EXPERIMENTAL PROCEDURE

### 4.1 TESTING OF THE EXPERIMENTAL FACILITY

A series of tests were conducted to ensure the successful operation of the various components of the experimental facility after they were fabricated and assembled. The details of these tests are:

### 4.1.1 Vacuum

As the experiments were conducted under vacuum it was necessary to establish the vacuum integrity of each of the components of the apparatus. To accomplish this, first, the individual components were tested against pressure integrity. All the exit valves were closed and compressed air was introduced to the invididual components. The compressed air was charged upto a pressure of 689.5 kN/m$^2$. All seals,connections and fittings were then checked for leaks using a soapy water solution. The components were deemed pressure tight, if, when isolated from air supply, they would maintain the 689.5 kN/m$^2$ pressure for atleast 24 hours. Finally entire experimental facility, when assembled, was tested in the similar manner for pressure integrity. After conducting the pressure tests the assembly was evacuated to 5.332 kN/m$^2$ and this was also maintained for a period of 24 hours. No change in vacuum gauge reading ensured the total assurance against any leakage.

Obviously, along with , check against any leakage, the condensers were ensured against any fluid interchange between annulus and condensing side. The sampling units were also ensured against any leakage with the jacket fluid.

### 4.1.2 Calibration of Measuring Devices

Before the thermocouples were installed, each one was checked for continuity and calibrated by means of standard mercury-in-glass thermometer of 0.1°C least count. The maximum deviation of 0.20 per cent between the readings of thermocouples and the thermometer was observed. A similar calibration was done when they were mounted on their respective positions.

The voltmeter and ammeter were calibrated against Substandard ammeter and voltmeter. The maximum deviation was less than 1 per cent.

## 4.2 DISCUSSION OF OPERATION

### 4.2.1 Initial Difficulties

Prior to commissioning of the experimental facility certain difficulties were confronted with. The major problem encountered is discussed here.

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Problem, which was immediately experienced was the unsatisfactory performance of condensers. At high values of heat flux (more than 25,000 W/m$^2$) and at low values of pressures (less than 67 kN/m$^2$) the condensers started showing unsatisfactory performance. To solve this problem the heat transfer surface for condensation was increased by installing fins over outside surface of the inner pipe and by increasing the diameter of inner pipe within physical limitations. This could mitigate the problem to quite an extent and it was possible to obtain the lowest limit of pressure as 11.33 kN/m$^2$ at a heat flux value of nearly 42,000 W/m$^2$ for the case of distilled water.

### 4.2.2 Operating Procedure

Before conducting the experiments for the boiling of a fluid the following steps were taken:

1. The experimental facility was connected to an air-compressor. The compressed air was supplied to the facility, to blow - off all the hold-up of fluid in the test vessel and connecting pipe lines. Thus, before charging the fresh fluid to the test facility it was completely free of the traces of previous fluid.

2. The test vessel was then rinsed with the fluid to be charged for the experimentation.

3. The heating surface was cleaned with distilled water, acetone and the fluid in which it was to be immersed prior to each experiment.
4. The exit valve (x) was closed (Refer Figure 3.1)
5. The liquid sampling valve (vii) was also closed.
6. The feed valve was opened and the test fluid was charged into the test vessel.
7. All the connections and fittings were checked against any leakage of the fluids.
8. Valves (xi) and (xii) were closed so that dissolved air did not circulate in the system.
9. Valve (xiii) was opened to remove the dissolved air.

After these intial operations, prior to filling of each test fluid to the test vessel, the pool of the liquid was boiled-off for six to seven hours to remove the dissolved air completely. When the bubbling ceased in bubbler(1), valve (xiii) was closed. The coolant circuits of condensers were started and valves (xi) and (xii) for vapour were opened.

Before conducting the series of experimental runs it was necessary to age and stabilize the heating surface. This was done as follows: the heating surface was submerged in the pool of fluid for a period of 24 hours followed by a boiling of 8 hours. This process was repeated till the heating surface was stabilized.

There was a consideration about the location of the liquid thermocouples. These thermocouples were to be located outside the superheated layer around the heating surface. The extent of superheated layer was found by noting the thermocouple readings at various positions away from the heating surface. The respective probes were kept at a point at which no change in fluid temperature was observed.

To begin an experiment, the equipment was adjusted to provide the chosen experimental conditions. Since the variation of heat transfer coefficieint with heat flux at a fixed pressure was the parametric variation of interest, the vacuum of the system was varied systematically. Thus a particular value of the vacuum was selected and this vacuum in the test vessel was obtained by switching -on vacuum pump and manipulating the control valve (III). After obtaining the desired value of vacuum in the system the vacuum pump was switched-off and the control valve (III) was closed. The heat flux was then adjusted by modulating the autotransformer. After the experimental conditions were adjusted the experiment was run for one to two hours until the steady state conditions reached. The steady state conditions were assumed when no variation in surface temperatures and liquid temperatures was observed with time. At steady state conditions the following informations were noted: The e.m.f. of all the surface and fluid thermocouples, the readings of voltmeter and ammeter, the readings of vacuum gauge and the barometric pressure with the help of a barometer.

The heat flux was now varied to another value and the experimental run was allowed to reach steady state conditions. All the informations, as mentioned above, were noted at steady state. Similar runs were conducted for six to seven values of heat flux for one vacuum level. Six to seven values of vacuum were employed for each fluid. Four fluids : distilled water, isopropanol, ethanol and methanol were employed for the experimentation. The range of experimental parameters is given in Table 4.1.

Table 4.1 Range of Experimental Parameters

Heating surface: 410 ASIS Stainless Steel

| Fluid | System Parameters | |
|---|---|---|
| | Heat Flux, $W/m^2$ | Pressure, $kN/m^2$ |
| 1. Distilled water | 6870 - 41,730 | 98.44 - 11.33 |
| 2. Isopropanol | 6870 - 37,709 | 97.97 - 15.33 |
| 3. Ethanol | 3307 - 33,893 | 98.18 - 19.33 |
| 4. Methanol | 6870 - 33,893 | 98.64 - 27.99 |

### 4.2.3 Reproducibility of the Surface Characteristics

Prior to the first experiment with each new experimental fluid an experiment was conducted with distilled water at atmospheric pressure to check the reproducibility of the heating surface characteristics. This experiment was always performed with maximum heat flux available in this investigation as at this heat flux all possible active sites may get activated. At the end of the experiment these results were compared with the previously conducted experiment. Within the tolerance of experimental error (like change in barometric pressure etc.), they were always equivalent.

### 4.2.4 Consistency of Experimental Data

The data produced with this experimental facility were consistent within the allowable experimental error. The data were not erratic in nature and yielded the expected trends. This trend was corroborated for those experiments also when certain experimental runs were repeated. However, experience with the facility revealed that no set of experimental conditions could be repeated lock, stock and barrel. Variations in voltage fluctuations and barometric pressure were responsible for this circumstance. Fortunately, the data were by and large consistent.

## 4.3 OPERATIONAL CONSTRAINTS

Certain physical limitations restricted upon the range of operational parameters. The range of different parameters obtained during this investigation have been listed in Table 4.1.

The parameters involved in the present investigation were heat flux, system pressure and boiling fluids. This investigation was carried out at low values of heat flux. This parameter was constrained because of the limited space available for placing the electric heater into the heating surface and due to the maximum current carrying capacity of the heating element.

The condensing unit of the facility limited the maximum attainable value of the vacuum. At higher values of vacuum, ebullition of the boiling fluid occurred. This ebullition caused an increase in the quantity of vapours which were condensed in the condensers. Physical limitations of the condensers were responsible for the insufficient condensation for this out-burst. And this involved unfavourably in reaching a steady state. Although it was possible to attain quite lower value of the pressure as of the order of 11 kN / $m^2$ in case of boiling distilled water but in case of other liquids it was restricted, relatively, to a higher value due to above mentioned difficulties.

# CHAPTER 5

## ANALYTICAL INVESTIGATIONS

### 5.1 HEAT TRANSFER COEFFICIENT IN POOL BOILING

Literature survey of Chapter 2 shows that there are two major classifications of models of heat transfer in nucleate pool boiling. They are based either on experimental evidences or theoretical considerations or both.

The models based on experimental data consider that the boiling heat transfer is analogous to the phenomenon in single-phase convective heat transfer. Correlations of this class are written in the form of dimensionless groups incorporating wall heat flux, wall superheat, system pressure and pertinent physico-thermal properties of boiling fluids. However, these correlations did not make any precise reference to the heating surface characteristics except to introduce empirical constant which accounts for the effects of specific surface-liquid combination. Some correlations have been recommended in dimensional form as well.

An important break-through in the studies of heat transfer in nucleate pool boiling was made by Jakob and Linke [3], [4], [49]. They considered the effects of heating surface characteristics on heat transfer coefficient assuming a linear relationship between the heat flux and number of active sites per unit area of the heating surface and product $fD_b$ to be a constant. Based on these assumptions an equation for heat transfer coefficient was obtained. Subsequently, in 1952 Rohsenow [14] derived a semi-theoretical equation for predicting heat fluxes during nucleate pool boiling. His analysis was based on the tacit assumptions of Jakob [3], [49]. The objection to these analyses is that they have been derived under the assumptions that heat flux changes linearly with the number of active sites and the product of bubble emission frequency and bubble departure diameter is a constant. Recent studies have revealed that bubble departure diameter depends on pressure [44], [88], [89] [91], [93], [94], [95] and frequency on pressure and heat flux [19], [32], [98] as well. Therefore, analyses due to Jakob and Linke [3], [4], [49] and Rohsenow [14] are not of general applicability for the wide ranges of pressure and heat flux.

The Jakob linear relationship between heat flux and number of active sites has been followed by the Nishikawa relationships [26], [51]. Nishikawa et al [32] have shown that the relationship between heat flux and number of active-sites per unit area is governed by a power law and not by a linear relationship as proposed by

Jakob originally. Their results were based on a maximum population count of eight active sites per square inch. The subsequent studies by Kurihara and Myers [52] and Gaertner and Westwater [46] have verified the Nishikawa relationships for twenty-eight active sites and 1,130 active sites per square inch respectively but with different exponents.

More recently Wiebe and Judd [54] have attempted the measurement of superheated layer thickness in saturated and subcooled nucleate boiling. They established a power law relationship between heat flux and product nf.

In this Chapter a new method is developed for predicting heat transfer coefficients in nucleate pool boiling at atmospheric and subatmospheric pressures from theoretical considerations.

## 5.2 ANALYSIS

Nucleate boiling is characterised by the presence of unique active sites on heating surfaces. The vapour bubbles sustain on these sites and these active sites are effective in reducing the average wall superheat required for boiling and, hence, high boiling coefficients are realised. It is interesting to speculate that the average wall superheat will depend on the number of active sites, the frequency of bubble emission and the bubble departure size. The effects of these quantities are refelected through vigorous agitation of the superheated layer adjacent to the heating surface. Recent studies by Cole and Shulman [95], Nishikawa [89] and other investigators, as mentioned previously, emphasize that bubble departure size depends on pressure of the boiling fluid. Therefore, it is apparent that the degree of agitation of superheated layer depends on number of active sites and bubble emission frequency only for a given system pressure.

From the above it can be concluded that boiling heat transfer coefficient can be considered a function of number of active sites n and bubble emission frequency f.

In the following sections relationships of n and f with wall heat flux, wall superheat, system pressure and pertinent physico-thermal properties of boiling fluids are established. Following these relationships expressions are derived which provide predictions of boiling heat transfer coefficients in terms of known quantities and constants.

### 5.2.1 Number of Active Sites

Brown [53] investigated that number of active sites per unit area n with radii larger than $r_c$ could be correlated by the following power law :

$$n = c_o \left( \frac{r_r}{r_c} \right)^m \qquad (5.1)$$

where $r_r$ is a radius for which n would be one per unit area, and $c_o$ is a dimensional constant having dimension of $(\text{unit area})^{-1}$.

According to the Laplace equation, the minimum radius of bubble is given by :

$$r_c = \frac{2\,\sigma}{\Delta P} \tag{a}$$

The difference $\Delta P$ can be expressed in terms of wall superheat as follows:

$$\Delta P = \frac{dP}{dT}\,\Delta T_w + \frac{d^2P}{dT^2}\cdot\frac{\Delta T_w^2}{2} + \ldots$$

$$\approx \frac{dP}{dT}\,\Delta T_w \tag{b}$$

Using the Clausius - Clapeyron relation and ideal gas law the following equation can be shown:

$$\frac{dP}{dT} = \frac{\lambda\,\rho_v}{T_s} \tag{c}$$

From Eqs. (a), (b) and (c) the value of $r_c$ is obtained in the following form:

$$r_c = \frac{2\,\sigma\,T_s}{\lambda\,\rho_v\,\Delta T_w} \tag{5.2}$$

When the value of $r_c$ from Eq.(5.2) is substituted in Eq.(5.1), the number of active sites as a function of wall superheat and fluid properties is obtained:

$$n = c_o\, r_r^m \left[ \frac{\lambda\, \rho_v}{2T_s\, \sigma} \right]^m \Delta T_w^m \qquad (5.3)$$

Mikic and Rohsenow [43] have shown that the value of m can be taken to be 2.5 for water implying that $q \propto \Delta T^{3.5}$ and 3.0 for organic fluids.

### 5.2.2 Wall Superheat

**Alad'ev** [55] developed the following expression which relates the wall superheat to wall heat flux and pertinent fluid properties during pool boiling of water for pressure range of 0.09 to 200 atmospheres.

$$\frac{\Delta T_w}{T_s} = 4.7 \times 10^{-3} \left[ \frac{10^{-6}\, q\, \lambda}{k_\ell\, T_s\, g} \right]^{0.3} \left[ \frac{\lambda}{C_\ell\, T_s} \right]^{1.2} . \qquad (5.4)$$

For the boiling of fluids other than water the constant of Eq.(5.4) should be replaced by some other constant $c_2$. The value of constant $c_2$ of Eq.(5.4a) is determined in Chapter 6 by least squares estimates using the experimental data in Eq.(5.4a) and is given in Table 6.1.

In order to show the extent of validity of Eq.(5.4) for the present investigation, a comparison between experimental and predicted values of $\Delta T_w$ is made in Chapter 6.

$$\frac{\Delta T_w}{T_s} = c_2 \left[\frac{10^{-6}\, q\, \lambda}{k_\ell\, T_s\, g}\right]^{0.3} \left[\frac{\lambda}{C_\ell\, T_s}\right]^{1.2} \tag{5.4a}$$

From Eqs.(5.3) and (5.4a) the following equation for number of active sites results:

$$n = c_o\, r_r^{\,m}\, c_2^m \left(\frac{\lambda\, \rho_v}{2T_s\, \sigma}\right)^m \left[\left(\frac{10^{-6}\, q\lambda}{k_\ell\, T_s\, g}\right)^{0.3} \left(\frac{\lambda}{C_\ell\, T_s}\right)^{1.2} T_s\right]^m \tag{5.5}$$

or

$$n = \frac{10^{-(1.8)m}}{2^m}\, c_o\, c_2^m\, r_r^{\,m} \left[\left(\frac{\lambda^{2.5} \rho_v}{T_s^{1.5}\, \sigma}\right)\left(\frac{q}{k_\ell\, g}\right)^{0.3} \left(\frac{1}{C_\ell}\right)^{1.2}\right]^m \tag{5.6}$$

### 5.2.3 Bubble Emission Frequency

In this section a method has been attempted which gives frequency of bubble emission as a function of Jakob number, heat flux and physical properties of boiling fluid for atmospheric and subatmospheric pressures.

The frequency of bubble emission at a single site may be written as:

$$f = \frac{1}{\theta_d + \theta_w} \tag{5.7}$$

Cole and Shulman [ 66], following the analysis for bubble growth in a uniformly superheated liquid of infinite extent have derived the following expression for bubble growth rate at moderate Jakob numbers ($J_a \leqslant 100$)

$$D(\theta) = J_a \sqrt{\pi \alpha_\ell}\ \theta^{1/2} \tag{5.8}$$

The expression for bubble departure diameter for boiling under atmospheric and subatmospheric pressures has been suggested by Cole and Shulman [95] as follows:

$$D_b = \left[\frac{133.3}{P}\right]\left[\frac{\sigma}{(\rho_\ell-\rho_v)g}\right]^{1/2} \tag{5.9}$$

Where P is in $kN/m^2$

From Eqs. (5.8) and (5.9), $\theta_d$ is calculated as follows:

$$\theta_d = \left\{ \frac{\left[133.3/P\right]^2 \left[\sigma / (\rho_\ell-\rho_v)g\right]}{\pi\ \alpha_\ell\ J_a^2} \right\} \tag{5.10}$$

Han and Griffith [106] have related the waiting period $\theta_w$ with transient thermal layer thickness, $\delta$ in the following form:

$$\delta(\theta) = \sqrt{\pi\ \alpha_\ell}\ (\theta)^{1/2} \tag{5.11}$$

where $\delta$ is given by the expression [124] :

$$\delta = 1.65\ \frac{k_\ell\ \Delta T_w}{q} \tag{5.12}$$

From Eqs.(5.11) and (5.12) the value of $\theta_w$ is obtained in the following form:

$$\theta_w = (1.65)^2 \left(\frac{k_\ell \, \Delta T_w}{q}\right)^2 \Big/ \pi \, \alpha_\ell$$

or

$$\theta_w = \frac{0.867}{\alpha_\ell} \left[\frac{k_\ell \, \Delta T_w}{q}\right]^2 \qquad (5.13)$$

Using the values of $\theta_d$ and $\theta_w$ from Eq (5.10) and Eq (5.13), respectively into Eq (5.7), an expression for frequency f is obtained as follows:

$$f = \left\{ \frac{1}{\dfrac{(133.3/P)^2 [(\sigma/(\rho_\ell - \rho_v) g]}{\pi \, \alpha_\ell \, J_a^2} + \dfrac{0.867}{\alpha_\ell} \left[\dfrac{k_\ell \, \Delta T_w}{q}\right]^2} \right\} \qquad (5.14)$$

Substituting the value of $\Delta T_w$ from Eq. (5.4a) in Eq.(5.14) the following is obtained:

$$f = \left\{ \frac{1}{\dfrac{[133.3/P]^2 [\sigma/(\rho_\ell - \rho_v) g]}{\pi \, \alpha_\ell \, J_a^2} + \dfrac{0.867}{\alpha_\ell} \left[ \dfrac{k_\ell}{q} \left\{ c_2 T_s \left(\dfrac{10^{-6} q \lambda}{k_\ell T_s g}\right)^{0.3} \left(\dfrac{\lambda}{c_\ell \, T_s}\right)^{1.2} \right\} \right]^2} \right\} \qquad (5.15)$$

For high values of Jakob number ($J_a > 100$) Cole and Shulman [66] have recommended the following empirical expression for bubble growth rate:

$$D(\theta) = 5\, J_a^{3/4} \sqrt{\alpha_\ell}\; \theta^{1/2} \tag{5.16}$$

From Eqs.(5.9) and (5.16) the value of $\theta_d$ is obtained as follows:

$$\theta_d = \frac{\left[133.3/P\right]^2 \left[\sigma / (\rho_\ell - \rho_v) g\right]}{25\, \alpha_\ell J_a^{3/2}} \tag{5.17}$$

Insertion of $\theta_w$ and $\theta_d$ from Eq. (5.13) and Eq. (5.17) in Eq. (5.7) gives an expression for f as follows:

$$f = \left\{ \frac{1}{\dfrac{\left[133.3/P\right]^2 \left[\sigma / (\rho_\ell - \rho_v) g\right]}{25\, \alpha_\ell J_a^{3/2}} + \dfrac{0.867}{\alpha_\ell} \left[ \dfrac{k_\ell \Delta T_w}{q} \right]^2} \right\} \tag{5.18}$$

Now inserting the value of $\Delta T_w$ from Eq.(5.4a) into Eq. (5.18) the following equation for bubble emission frequency results:

$$f = \left\{ \frac{1}{\dfrac{\left[133.3/P\right]^2 \left[\sigma / (\rho_\ell - \rho_v) g\right]}{25\alpha_\ell J_a^{3/2}} + \dfrac{0.867}{\alpha_\ell} \left[ \dfrac{k_\ell}{q} \left\{ c_2 T_s \left( \dfrac{10^{-6} q \lambda}{k_\ell T_s g} \right)^{0.3} \left( \dfrac{\lambda}{c_\ell T_s} \right)^{1.2} \right\} \right]^2} \right\} \tag{5.19}$$

Eq.(5.15) and Eq.(5.19) represent the expressions for f for Jakob number less and greater than 100, respectively. The value of $c_2$ is taken from Table 6.1. These expressions appear to depend upon wall heat flux also. It is interesting to mention that investigations [19], [32], [98] have also concluded that heat flux influences the values of bubble emission frequency.

Comparison of theoretical predictions from Eqs. (5.15) and (5.19) with experimental data is shown in Chapter 6 to determine the extent of validity of the proposed equations.

### 5.2.4 Heat Transfer Coefficient

It has been shown by Wiebe and Judd [54] that it is possible to evaluate heat transfer coefficient in boiling liquids from the knowledge of number of active sites per unit area n and bubble emission frequency f. In the preceding sections procedures have been described to determine these quantitites as function of known parameters except unknown quantities $c_o$ and $r_r$. Equation of Wiebe and Judd is as follows:

$$h = c_3 \, (nf)^a \qquad (5.20)$$

Where $c_3$ is a constant and, a, is 1/2 and 1/3 for values of nf less and greater than $55 \times 10^3$ Bubbles/in$^2$.sec. respectively. When nf is not known, the value of, a, can be determined by regression analysis using the experimental data. Insertion of values of n from Eq. (5.6) in Eq(5.20) yields the following equation for heat transfer coefficient:

$$h = c_3 \left[ \left\{ \frac{10^{-(1.8)m}}{2^m} c_o \, c_2^m \, r_r^m \right\} \left\{ \left( \frac{\lambda^{2.5}}{T_s^{1.5}} \frac{\rho_v}{\sigma} \right) \left( \frac{q}{k_\ell g} \right)^{0.3} \left( \frac{1}{\epsilon_\ell} \right)^{1.2} \right\}^m \times \{f\} \right]^a \quad \text{--(5.21)}$$

$$h = M \left[ \left\{ \left( \frac{\lambda^{2.5} \rho_v}{T_s^{1.5} \, \sigma} \right) \left( \frac{q}{k_\ell g} \right)^{0.3} \left( \frac{1}{\epsilon_\ell} \right)^{1.2} \right\}^m \{f\} \right]^a \quad (5.22)$$

where $f$ is given by the respective equations, Eq. (5.15) or Eq. (5.19) for Jakob number less and greater than 100

and constant $M \left[ = \left( c_3 \frac{10^{-(1.8)m}}{2^m} c_o \, c_2^m \, r_r^m \right)^a \right]$

contains quantities whose measurement is extremely difficult. In order to utilise the resultant equation, Eq. (5.22) it is necessary to determine the nature of this constant M. When M is adequately defined it becomes possible to predict heat transfer coefficients from analytical relationship (that is, from considerations of heating surface characteristics and bubble dynamics).

To obtain the value for constant M, we proceed as follows:

Eq.(5.22) provides quantitative measure of the inter-relationship of the heat transfer coefficient and of the terms: wall heat flux and pertinent physico-thermal properties of the boiling fluids. The analytical model, Eq.(5.22) does not account for the effects of pressure and surface-liquid combination explicitly. Therefore, it can be argued to treat the unknown constant M representing the combined effects of pressure and surface-liquid combination. The form of M can be intuitively expressed as follows:

$$M = \phi(P) . \psi(C_{sf}) \qquad (5.23)$$

where $\phi$ is a function of pressure only and $\psi$ is a function of surface - liquid combination factor. The present and existing experimental data will be used to show the multiplicative nature of M in Chapter 6. When one considers the bubble formation on a solid surface through which heat is transported into the fluid, the conditions for the dynamic equilibrium of a bubble attached to the heating surface are governed by the wettability of the solid by the boiling fluid among other factors. Thus the value of $\psi(C_{sf})$ , as it is , depends on the nature of heating surface and the fluid in touch with it. As a matter of fact, it is highly improbable to describe the nature of heating surface. An implication of this would be that general agreement on the absolute values of heat transfer coefficient is possible only if the same heating surface is used in various investigations. But in industrial equipment the heating surfaces used are of widely differing nature. Therefore, a rigorous generalised correlation for absolute values of heat transfer coefficient from different heating surfaces is extremely difficult. Keeping in view this inevitable difficulty and the proposed functional form of constant M it seems best to re-arrange Eq.(5.22) so as to yield the values of ratio, $h^*/h_1^*$ , where $h_1^*$ is the value at the normal boiling point. Thus the function, $\psi(C_{sf})$ is easily eliminated and the data of different investigators obtained on different heating surfaces should be correlated fairly well by the resultant equation. Therefore, the general equation for the prediction of $h^*/h_1^*$ assumes the following form where f is given by the respective equations, Eq.(5.15) or Eq.(5.19) for Jakob number less and greater than 100.

$$\frac{h^*}{h_1^*} = \frac{M}{M_1}\left[\left\{\left(\frac{\lambda}{\lambda_1}\right)^{2.5}\left(\frac{\rho_v}{\rho_{v1}}\right)\left(\frac{T_{s1}}{T_s}\right)^{1.5}\left(\frac{\sigma_1}{\sigma}\right)\left(\frac{k_{\ell 1}}{k_\ell}\right)^{0.3}\left(\frac{q}{q_1}\right)^{0.3}\left(\frac{C_{\ell 1}}{C_\ell}\right)^{1.2}\right\}^m\left\{\frac{f}{f_1}\right\}\right]^a \qquad (5.24)$$

subscript '1' denotes the values corresponding to normal boiling point.

where $M/M_1 = \phi(P)/\phi(P_1)$ which can be represented as function of $(P/P_1)$ as follows:

$$\frac{M}{M_1} = \eta(P/P_1) \qquad (5.25)$$

The analytical model, Eq. (5.24) serves to illustrate the general nature of heat transfer to boiling fluids from the heating surfaces. It is important to note that the resultant equation is independent of surface-liquid combination factor, $C_{sf}$ ; and therefore this equation should serve to compare the data on differing heating surfaces. Further, this equation, Eq. (5.24) is capable to predict heat transfer coefficients at subatmospheric pressures if the value of $h_1^*$ is adequately known.

The predicted values of wall superheat and heat transfer coefficient from the present analysis as discussed above, have been compared with the experimentally measured values of present study and existing investigations in Chapter 6.

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# CHAPTER 6

## RESULTS AND DISCUSSION

A total of 149 runs were conducted . Of these 30 were for distilled water, 45 for isopropanol, 39 for ethanol and 35 for methanol.

The data with ranges of experimental parameters are listed in Tables B-1, B-2, B-3 and B-4 of Appendix B. Columns 6 and 7 list corrected temperature difference & point values of heat transfer coefficient respectively.

In the present Chapter the experimental and the analytical results of nucleate pool boiling of fluids at atmospheric and subatmospheric pressures are discussed. Computer programs were written on an IBM 360 digital computer for calculations for all the runs reported in Appendix B. The inputs to these programs are: heat flux, heating surface temperature, liquid saturation temperature, system pressure, and physico-thermal properties of boiling fluids at their saturation temperature.

### 6.1 LIMITATIONS OF ANALYSIS

The practical limitations which have been used in the analysis of present experimental data are summarized as follows:

In the present investigation the measurement of wall temperature was done at three locations around the heating surface by means of thermocouples as discussed in Section 3.2.2. Fluid temperatures were also measured at three positions corresponding to the wall thermocouple locations. Experimental data as recorded in Appendix B show that wall and the fluid temperatures vary circumferentially. Therefore, there was a need to evaluate average wall temperature, $\bar{T}_w$ and average temperature difference, $\overline{\Delta T}$ for further analysis and interpretation of the experimental results. The values of $\overline{\Delta T}$ were required to calculate average experimental heat transfer coefficients. In order to calculate these quantities the following procedures have been employed:

Average values of $\overline{\Delta T}$ over the circumference was calculated by the method of mechanical quadrature using $\Delta T$ values at the top, at the middle and at the bottom of the heating surface. $\Delta T$ values at these locations represented the respective difference between corrected wall temperature and fluid temperature. The corrected wall temperatures were determined by subtracting temperature drop across the wall from the thermocouple readings. In some of the calculations, average values of wall temperatures and wall superheat $\overline{\Delta T}_w$ over the circumference were also required. These have also been obtained by mechanical quadrature method as detailed in Appendix A.

The fluid temperature around the heating surface was found to vary. However, the thermophysical properties of the boiling fluid were approximated to the properties of the fluid at the saturation temperature corresponding to the system pressure. This has been accepted because the variation in liquid temperature around the heating surface was small enough to cause any significant change in the pertinent physico-thermal properties of the boiling fluids.

The longitudinal conduction of heat in the heating surface has been neglected since the heating surface was a long tube of thin wall.

## 6.2 COMPARISON OF EXPERIMENTAL VALUES OF HEAT TRANSFER COEFFICIENT WITH HEAT FLUX AT ATMOSPHERIC PRESSURE

To test the extent of the validity of the widely accepted relationship; $h = bq^{0.7}$, the heat transfer coefficients of distilled water, isopropanol, ethanol and methanol are plotted in Figures 6.1 through 6.4 respectively as functions of heat flux. The experimental observations of the present study have been augmented with the data available in the literature [11], [28], [42], [98], [135]. These figures show that heat transfer coefficient — heat flux relationship ( $h \propto q^{0.7}$) is valid for all the liquids investigated. A trend similar to that illustrated on present data is evident in the data of other investigators. This ensures that the measurements and techniques used for data collection in the present investigation are reliable.

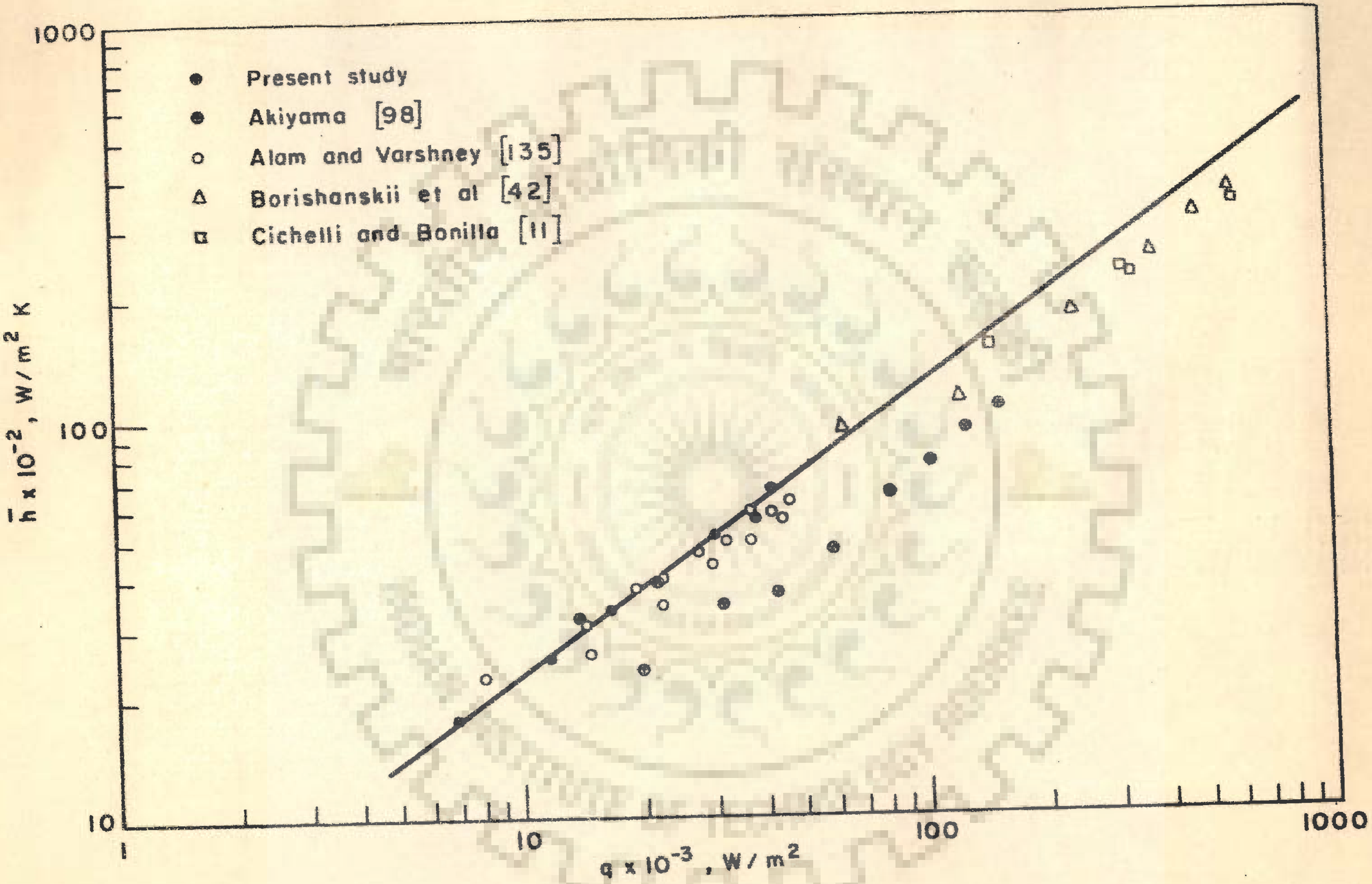

Fig. 6·1 Heat transfer coefficient – heat flux relationship for distilled water at atmospheric pressure

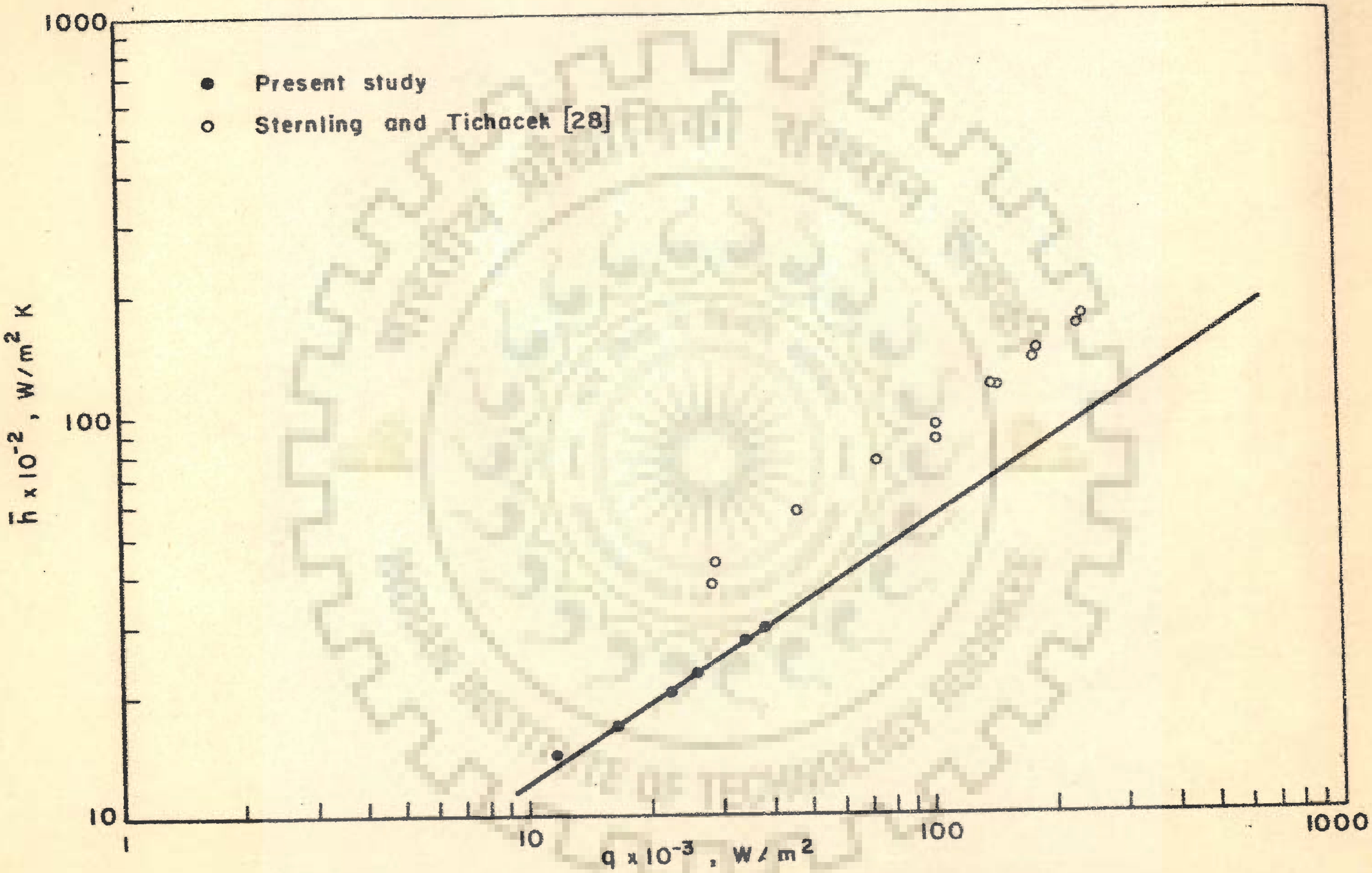

Fig. 6·2 Heat transfer coefficient - heat flux relationship for isopropanol at atmospheric pressure

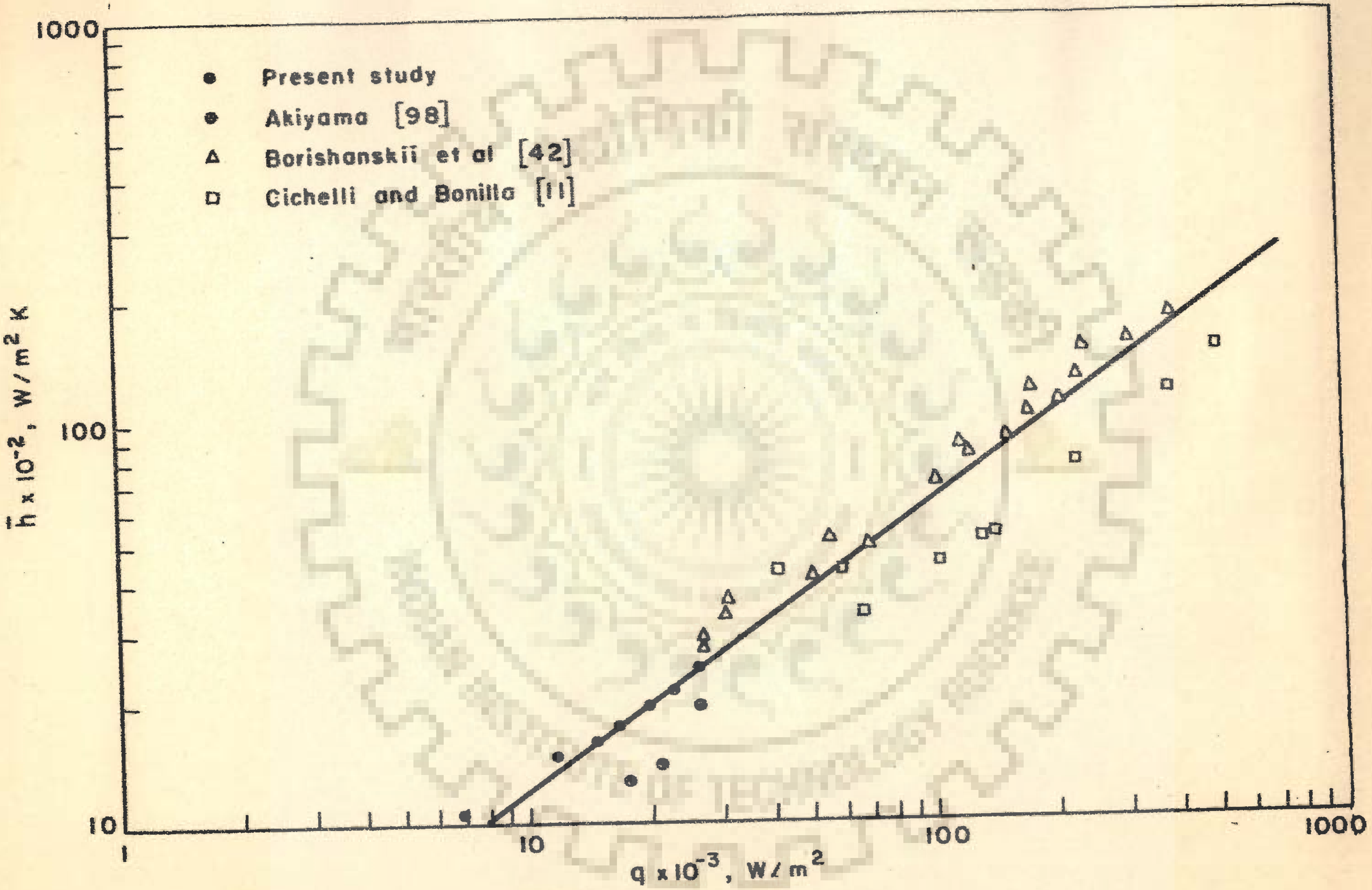

Fig. 6·3 Heat transfer coefficient-heat flux relationship for ethanol at atmospheric pressure

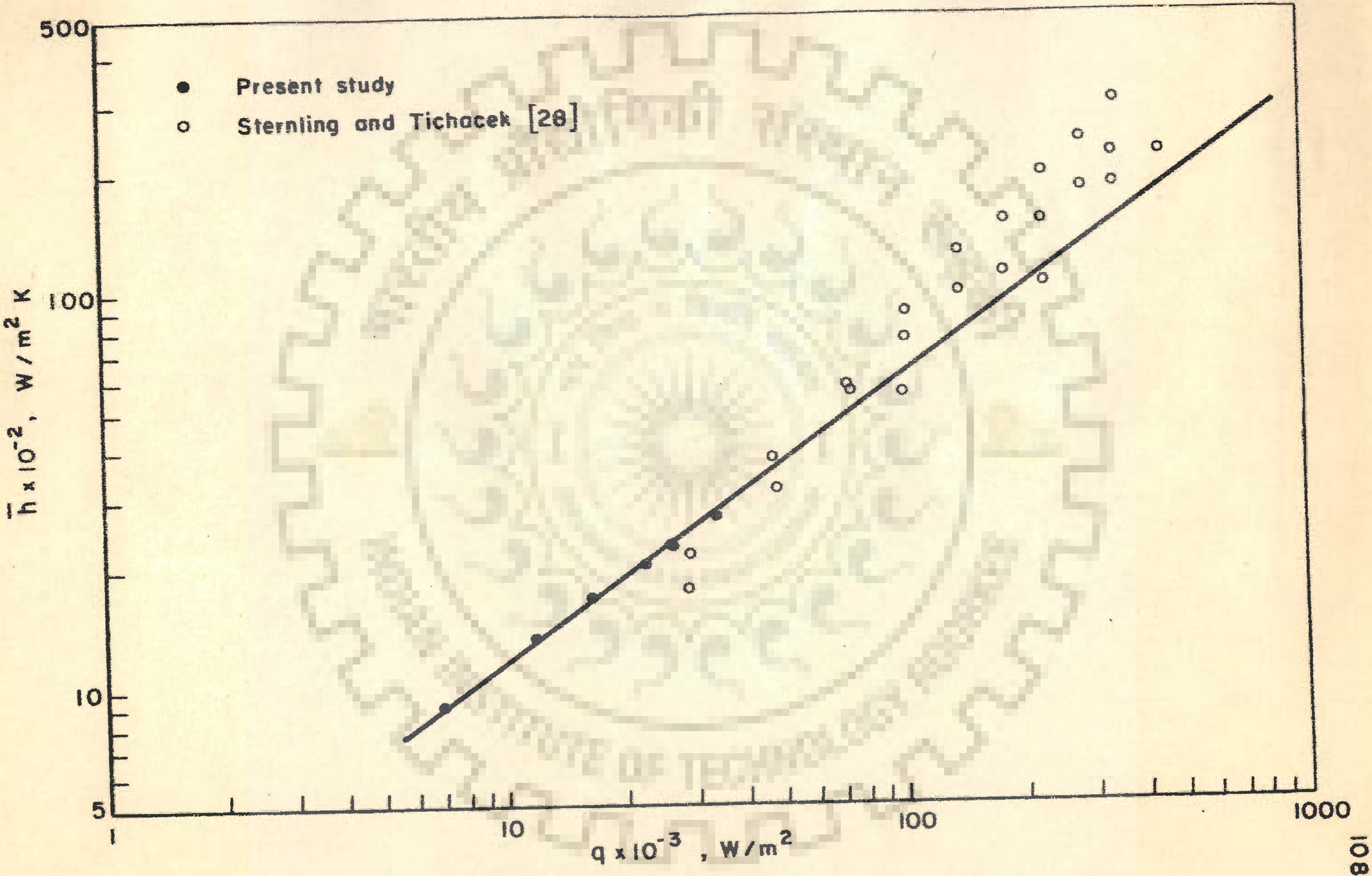

Fig. 6·4 Heat transfer coefficient – heat flux relationship for methanol at atmospheric pressure

Some of the experimental data [28], [42], [98] exhibit disagreement with the present values, but there is clearly a general correspondence between the two as a function of heat flux over a whole range of heat flux shown. It can be seen that these data form separate groupings about the lines with the slope of 0.7. This indicates that a single expression may not adequately relate heat transfer coefficient to heat flux for all heat transfer surfaces and liquids. This behaviour is not surprising but reassures the result that the characteristics of heating surface and the nature of boiling fluids in touch with it play a significant role on boiling heat transfer rates.

Whether the value of constant b is affected mainly or entirely by the surface characteristics and the nature of liquid remains to be proved.

### 6.3 HEAT TRANSFER COEFFICIENT – HEAT FLUX RELATIONSHIP OVER A RANGE OF SUBATMOSPHERIC PRESSURE

In this section the experimental data of the boiling fluids for subatmospheric pressure were examined by plotting them in Figures 6.5 to 6.8 . The main purpose herein is to determine the relationship between heat transfer coefficient and heat flux in boiling fluids for pressures less than atmospheric pressure. It is seen from these figures that the experimental data for subatmospheric pressures can be correlated by the same relationship as for atmospheric pressure ( $h = bq^{0.7}$) with constant b depending upon the system pressure. The constant b decreases with

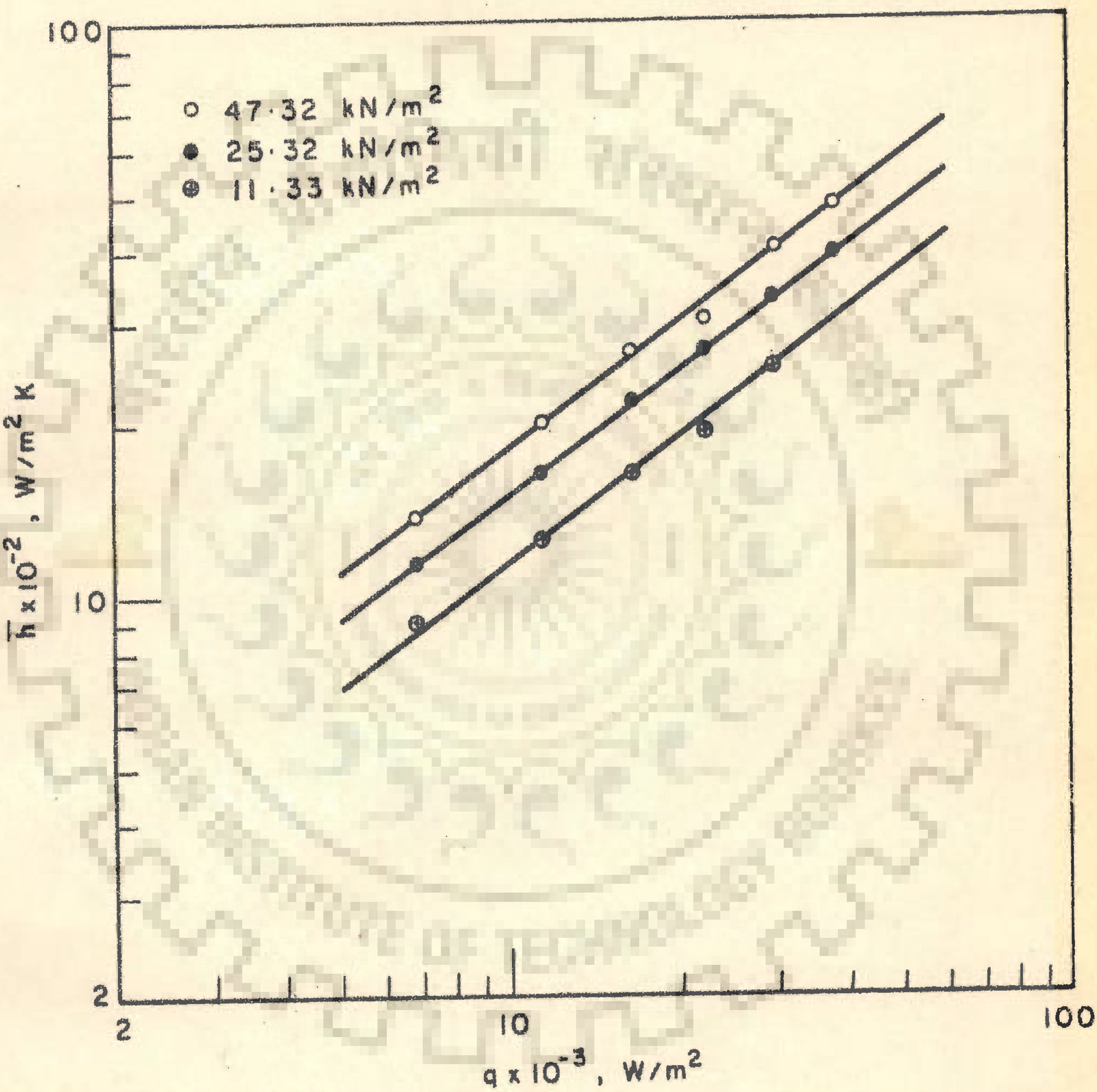

Fig. 6·5 Heat transfer coefficient–heat flux for distilled water at subatmospheric pressures

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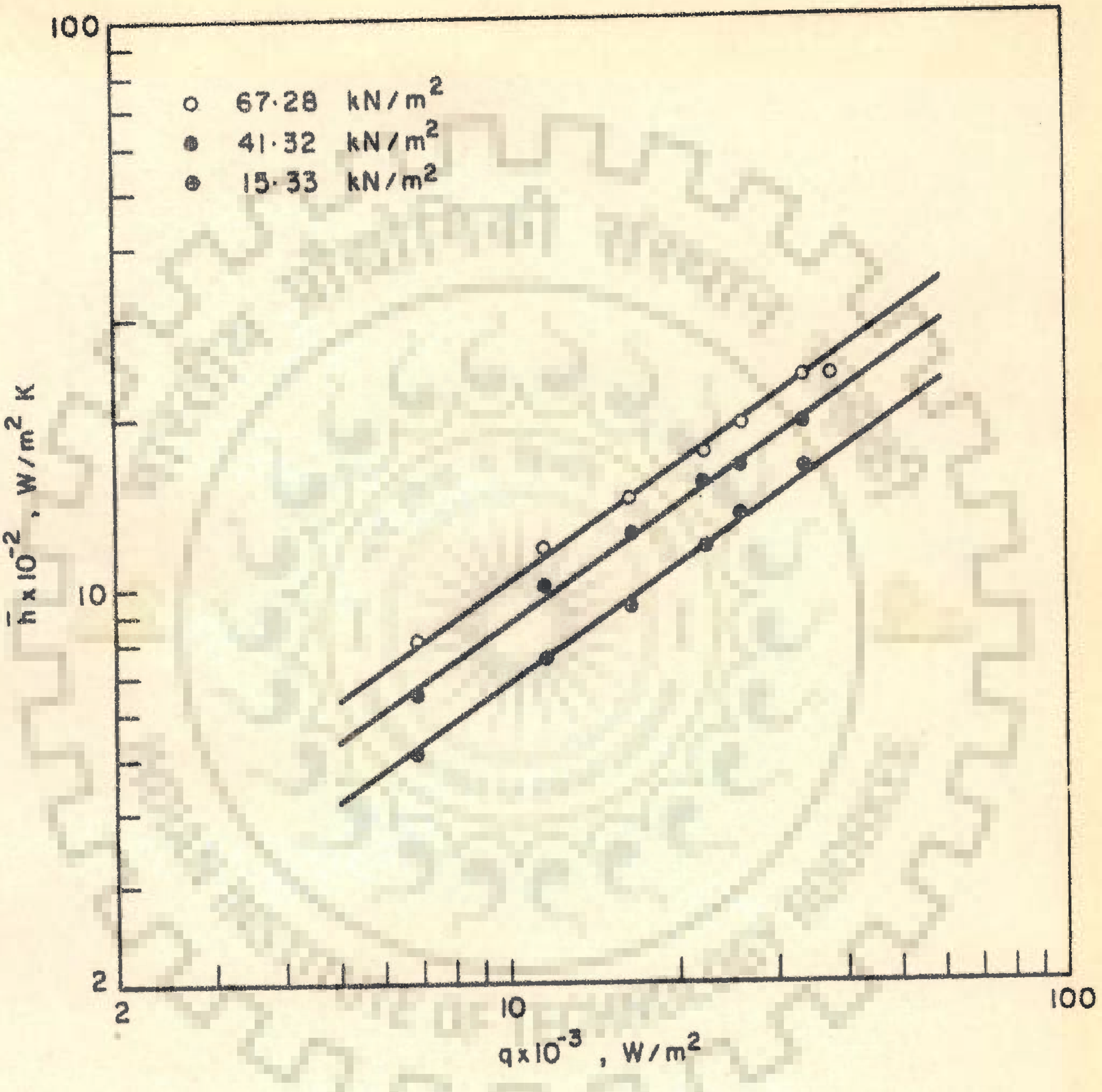

Fig. 6·6 Heat transfer coefficient-heat flux for isopropanol at subatmospheric pressures

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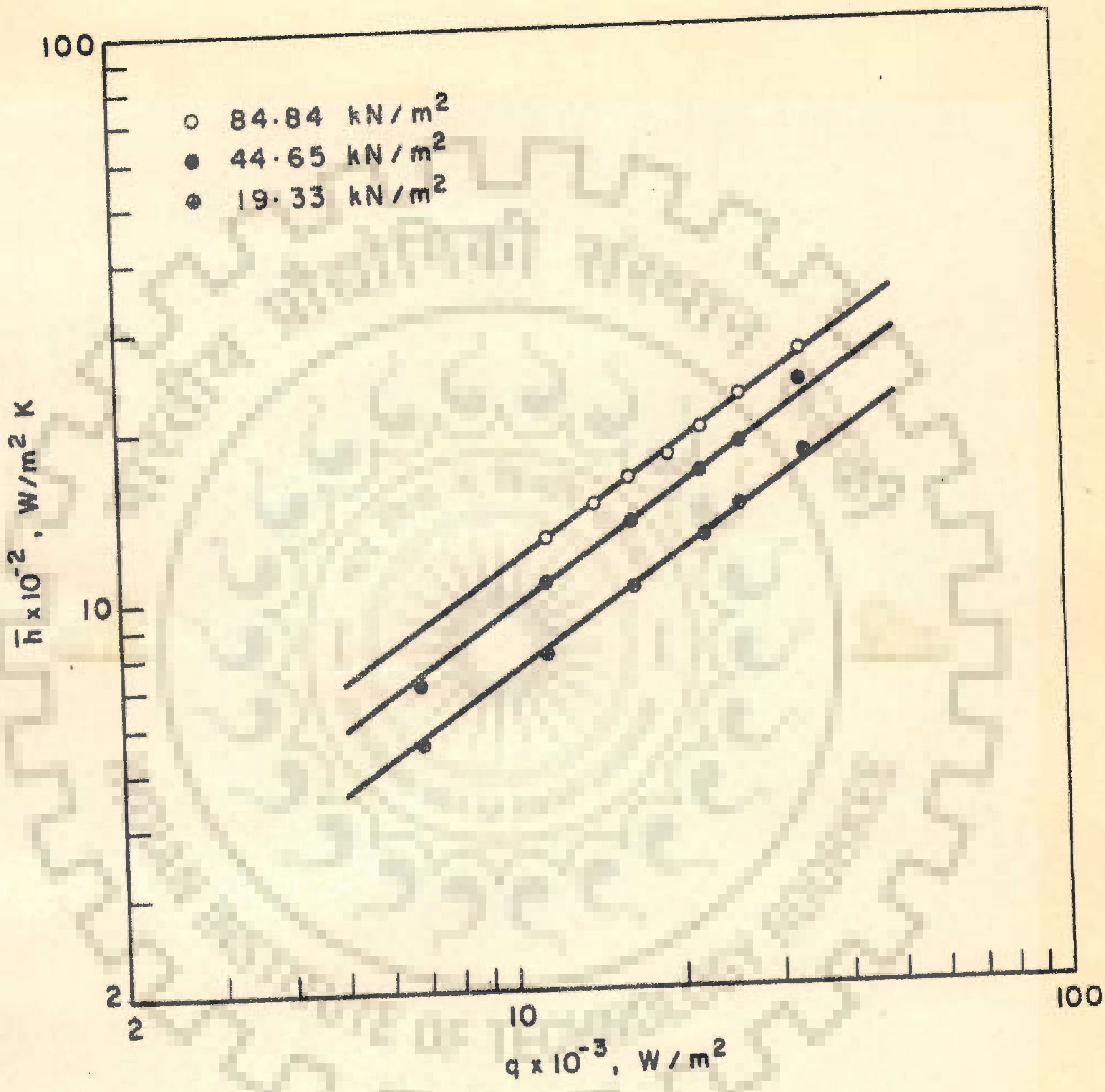

Fig. 6·7 Heat transfer coefficient–heat flux for ethanol at subatmospheric pressures

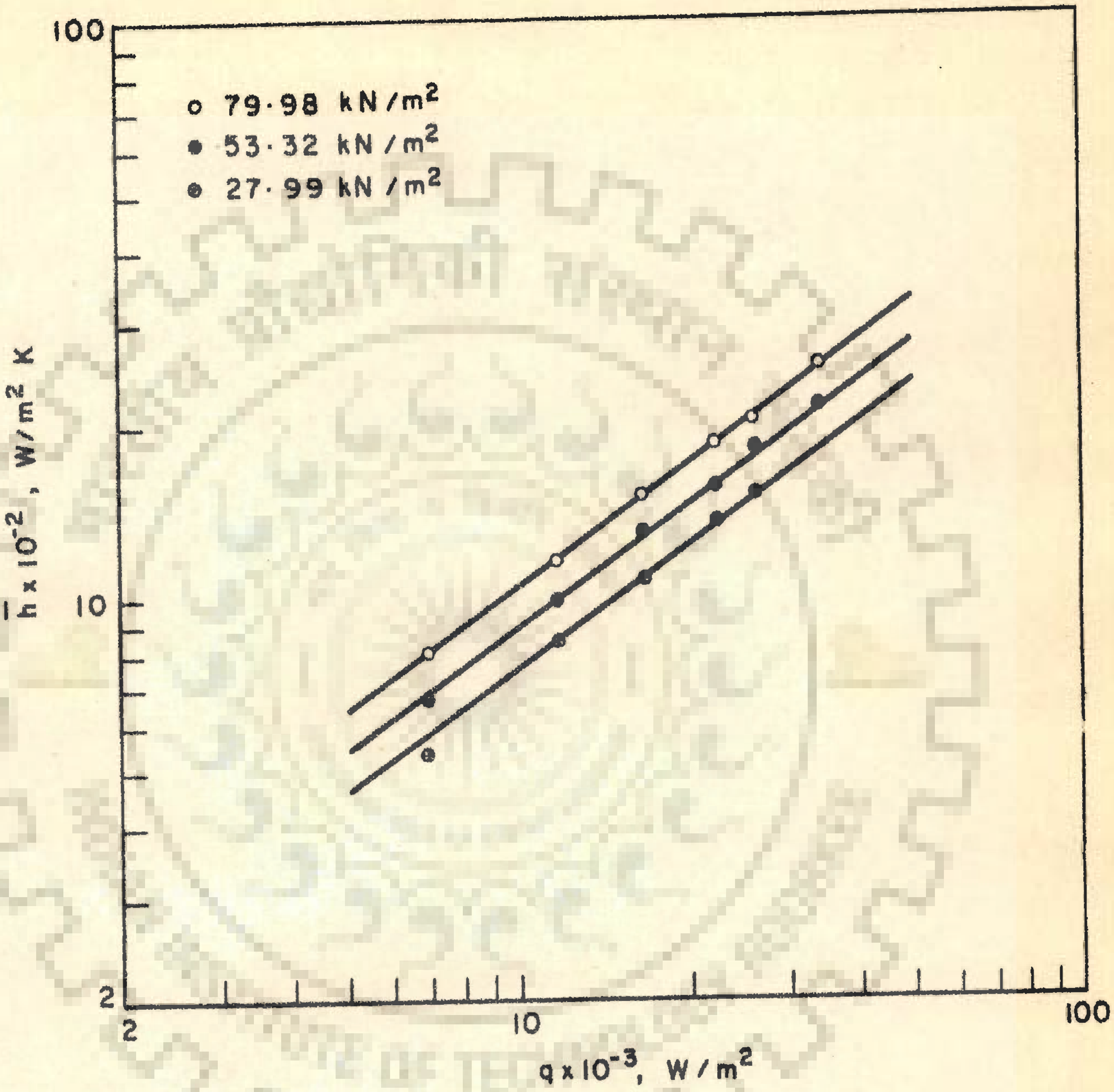

Fig. 6·8 Heat transfer coefficient – heat flux for methanol at subatmospheric pressures the decrease in pressure . The superimpos of this section and of the previous section concludes that constant b is complex which among other parameters is affected by surface characteristics, nature of boiling fluids and pressure as well. An implication of this would be that constant b cannot be incorporated in accurate theoretical analysis of the boiling heat transfer in which the proportionality constant b relating heat transfer coefficeint with heat flux is sought as an exact theoretical value.

## 6.4 COMPARISON BETWEEN EXPERIMENTAL WALL SUPERHEAT AND PREDICTED VALUES FROM ALAD'EV EQUATION

Alad'ev equation [55] relates wall superheat to wall heat flux and pertinent physico-thermal properties of water in nucleate pool boiling for pressure range from 0.09 to 200 atmospheres.

Figure 6.9 shows comparison between experimentally measured and predicted values of wall superheat from the Alad'ev equation, Eq. (5.4) as follows: The predictions are in good agreement with the experimental data for distilled water. The equation underpredicted the values for isopropanol, ethanol and methanol. The possible reason for this discrepancy among other reasons seems to be due to that this equation was derived for the experimental data of water only. Therefore, it can be concluded that the Alad'ev equation does not possess general applicability for liquids differing in their physico-thermal properties.

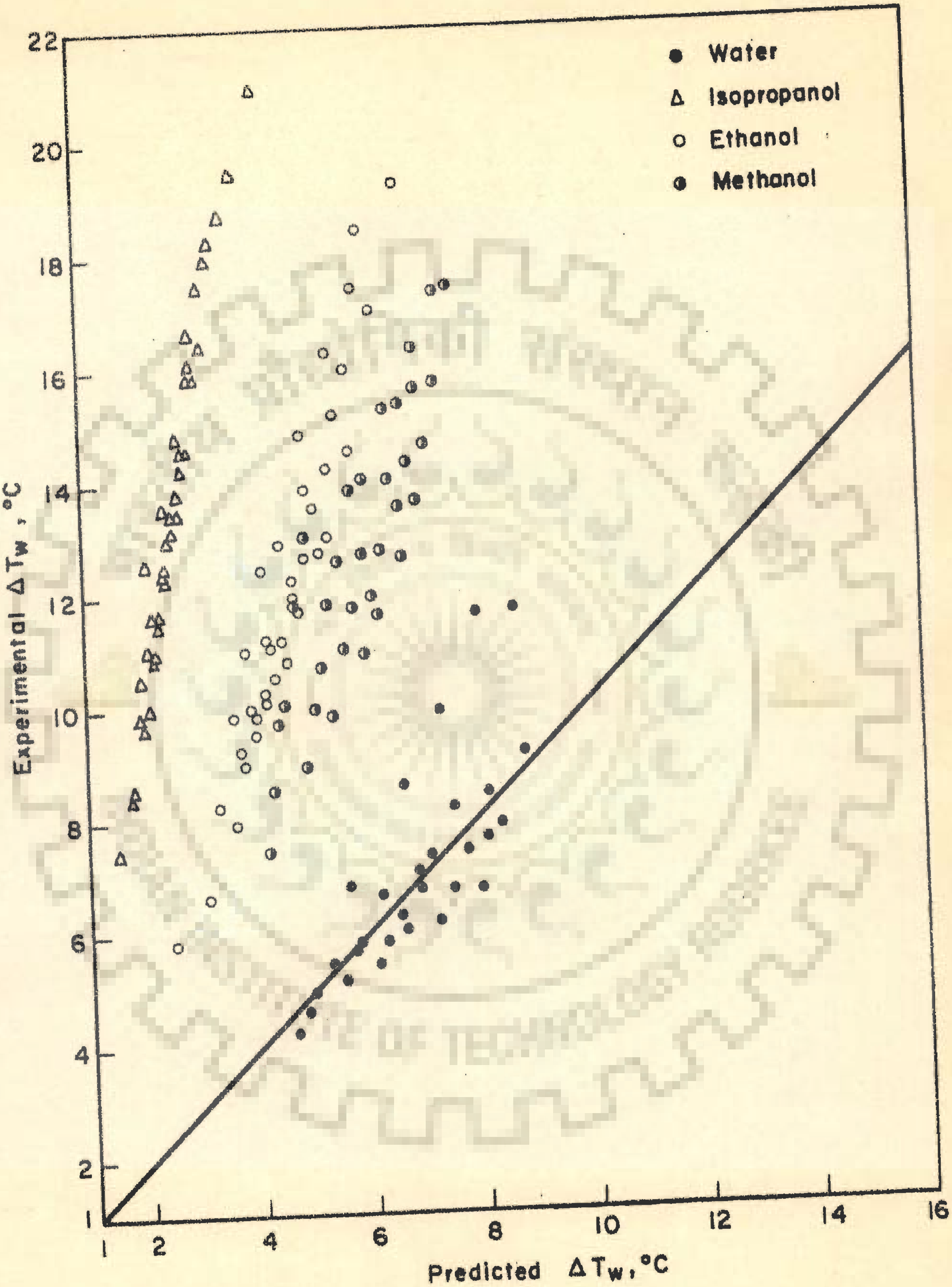

Fig. 6·9 Present experimental data compared with the Alad'ev equation, Eq(5·4)

It is intuitively obvious that the constant of the Alad'ev equation represents the surface-liquid combination effect. With this in view an attempt was made to modify the constant of the equation without altering the basic form of the equation. Computation revealed that the most optimum value of constant was found to depend upon the surface-liquid combination. For respective liquids the values of the constant are given in Table 6.1.

Table 6.1 The values of constant $c_2$ of Eq. (5.4a)

| Liquid | $c_2$ |
|---|---|
| Distilled water | $4.70 \times 10^{-3}$ |
| Isopropanol | $2.40 \times 10^{-2}$ |
| Ethanol | $1.24 \times 10^{-2}$ |
| Methanol | $9.99 \times 10^{-3}$ |

With the values of constant $c_2$ from Table 6.1 calculations were made to obtain predicted values of wall superheat. The predicted values are compared with the average experimental values for isopropanol, ethanol and methanol in Figures 6.10 and 6.11. These figures show that the comparison is excellent within a maximum of ± 15 per cent. An implication of this would be that surface-liquid combination plays a significant role on wall superheat of heating surfaces in nucleate pool boiling. Therefore, the Alad'ev equation may then be written as follows:

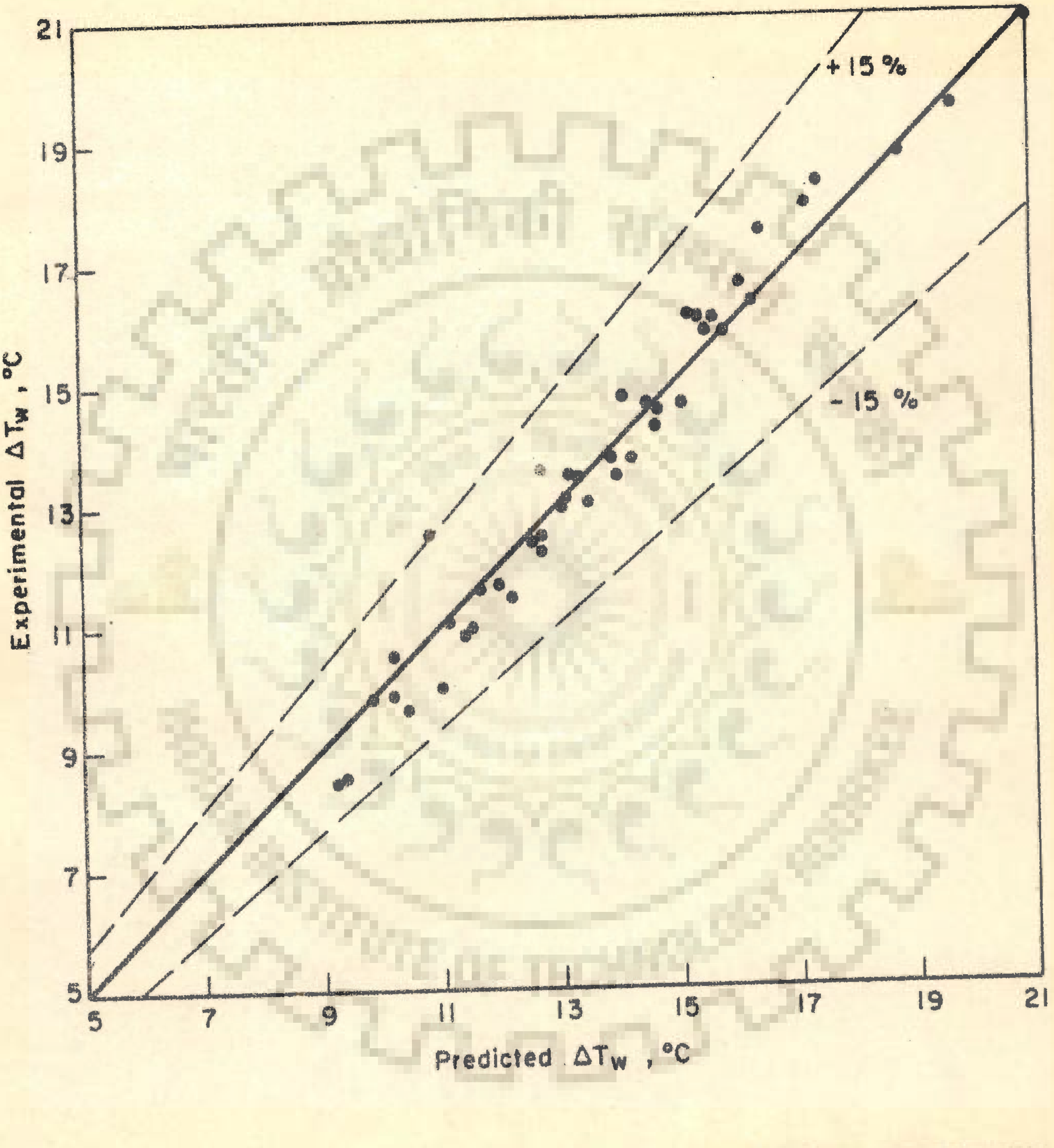

Fig. 6·10 Present experimental data of isopropanol compared with the modified Alad'ev equation, Eq(6·1)

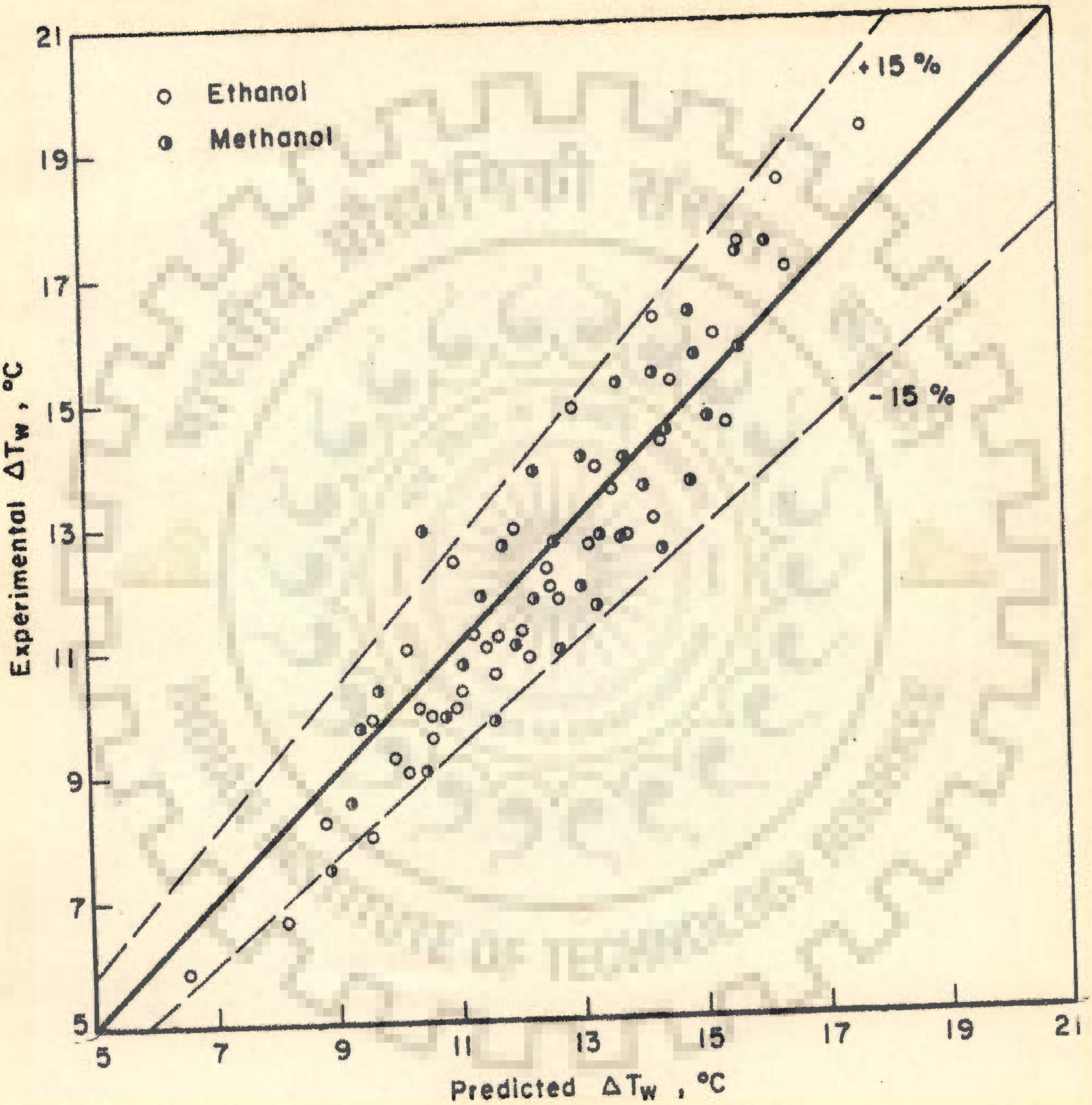

Fig. 6·11 Present experimental data compared with the modified Alad'ev equation, Eq (6·1)

$$\frac{\Delta T_w}{T_s} = c_2 \left[ \frac{10^{-6}\, q\, \lambda}{k_\ell\, T_s\, g} \right]^{0.3} \left[ \frac{\lambda}{C_\ell\, T_s} \right]^{1.2} \qquad (6.1)$$

Where constant $c_2$ is to be taken from Table 6.1 for the liquids investigated. It is important to recall that in the analytical analysis of Chapter 5 the Alad'ev equation was presumed to possess the same form [cf. (5.4a) ] as above.

## 6.5 COMPARISON BETWEEN THEORETICAL FREQUENCY FROM PROPOSED MODEL AND EXISTING EXPERIMENTAL DATA

Following the analysis of Cole and Shulman [66] for bubble growth rate and the analysis of Han and Griffith [106] for transient thermal layer thickness, $\delta$ Eqs.(5.15 & 5.19) were developed in Section 5.2.3 of Chapter 5 for the determination of bubble emission frequency, f. Cole [45] has measured experimentally the values of frequency for the given values of $\theta_w$, $\theta_d$ and Ja as parameters at atmospheric and subatmospheric pressures. Therefore, a comparison was made between the experimental values of frequency measured by Cole [45] and the calculated values from the present analysis, Eqs.(5.15 & 5.19) for almost similar values of $\theta_w$, $\theta_d$ and Ja. These values are given in Table 6.2.

Table 6.2 Comparison between predicted frequency from Eq.(5.15 or 5.19) and experimental values of Cole [45]

| Liquid | Ja | $\theta_d$, s | | $\theta_w$, s | | f, $s^{-1}$ | |
|---|---|---|---|---|---|---|---|
| | | Exptl. | Predt. | Exptl. | Predt. | Exptl. | Predt. |
| Water | 87.9 | 0.039 | 0.0477 | 0.694 | 0.3351 | 1.36 | 2.61 |
| | 191.0 | 0.034 | 0.0927 | 0.222 | 0.8060 | 3.91 | 1.11 |
| Methanol | 41.2 | 0.015 | 0.0300 | 0.087 | 0.0750 | 9.79 | 9.41 |
| | 59.6 | 0.019 | 0.0255 | 0.141 | 0.0915 | 6.22 | 8.54 |
| | 74.6 | 0.025 | 0.0358 | 0.298 | 0.3337 | 3.09 | 2.71 |

It is clear from the above table that the agreement between predicted and experimental values of frequency is fairly well and encouraging.

## 6.6 BUBBLE EMISSION FREQUENCY – HEAT FLUX RELATIONSHIP FOR ATMOSPHERIC AND SUBATMOSPHERIC PRESSURES

In Chapter 5 Eqs.(5.15 and 5.19) for f were derived from theoretical considerations. As obvious from these equations that at a given system pressure frequency depends upon heat flux. Therefore, typical plots were drawn between f and q on log log scale with pressure as parameter in Figures 6.12 through 6.15. From these figures it is seen that the variation of f with q is represented by a power law of the following form:

$$f = C_4 \, q^{n_4} \qquad (6.2)$$

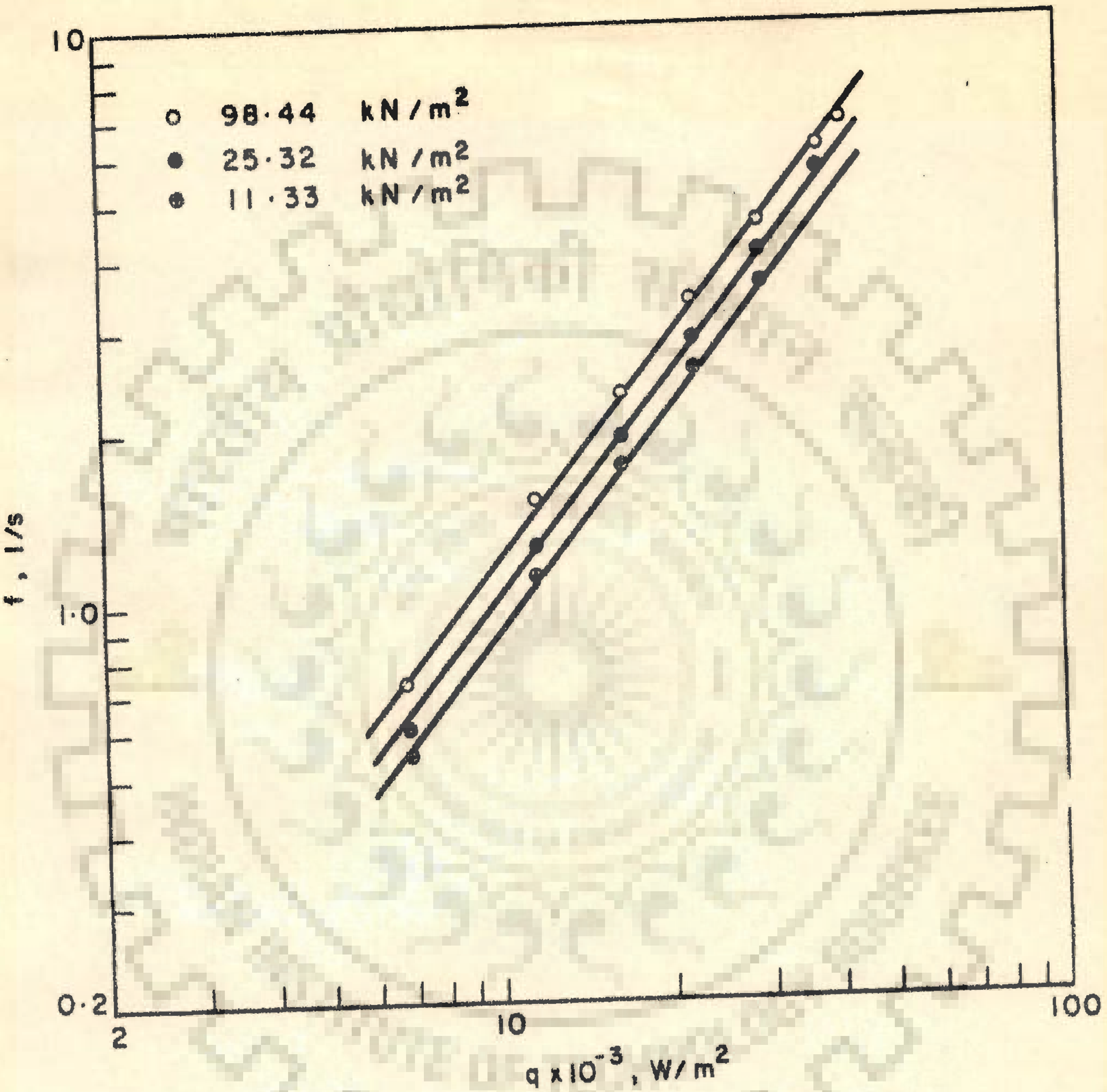

Fig. 6·12 Frequency-heat flux relationship for distilled water

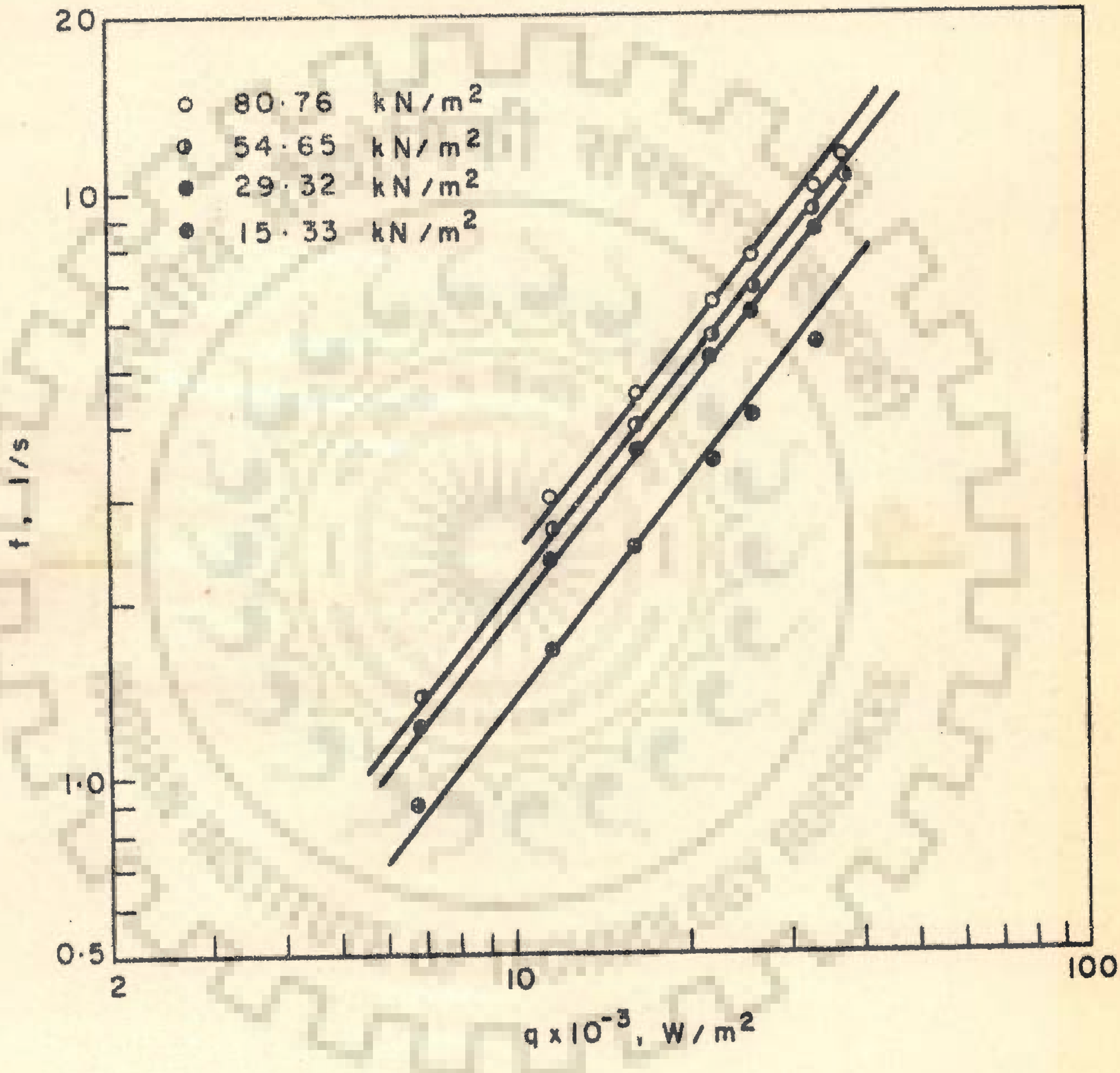

Fig. 6·13 Frequency - heat flux relationship for isopropanol

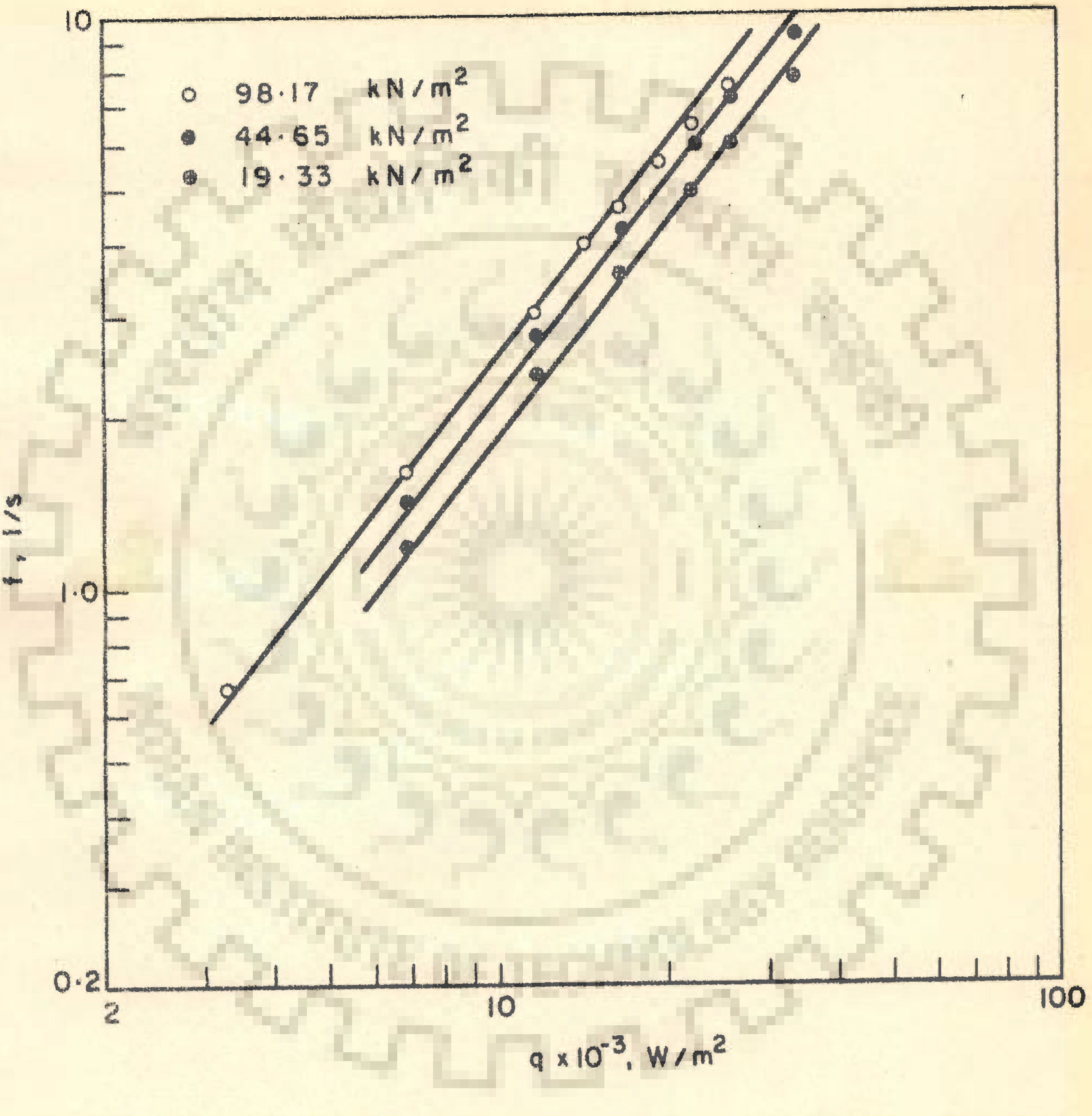

Fig. 6·14 Frequency – heat flux relationship for ethanol

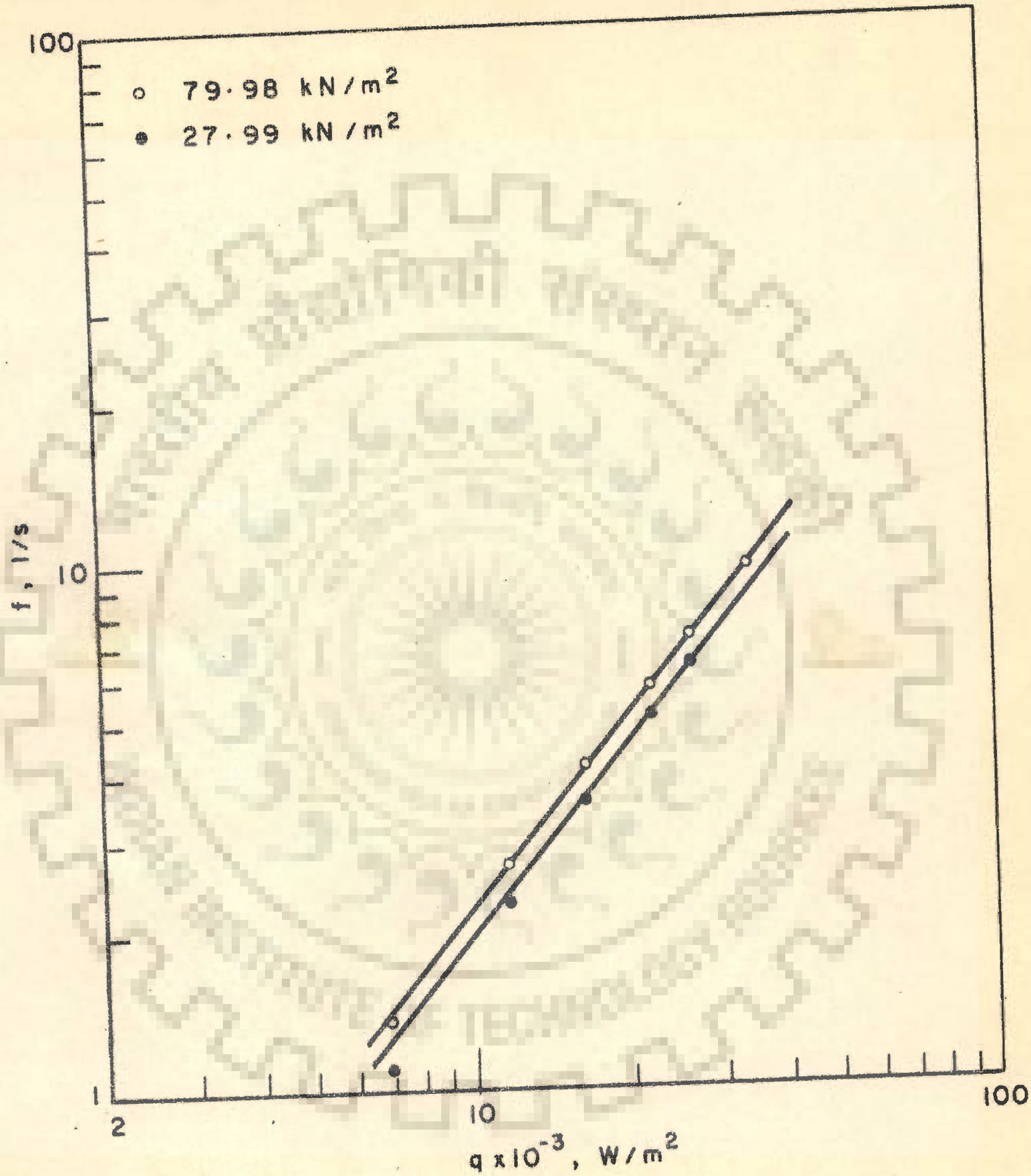

Fig. 6·15 Frequency – heat flux relationship for methanol

The values of constant $C_4$ and exponent $n_4$ were determined by the method of least squares curve fitting and are given in Table 6.3.

Table 6.3 Values of constant $C_4$ and exponent $n_4$ in Eq.(6.2)

| Distilled water | | | Isopropanol | | |
|---|---|---|---|---|---|
| Pressure kN/$m^2$ | $C_4$x$10^5$ | $n_4$ | Pressure kN/$m^2$ | $C_4$ x $10^4$ | $n_4$ |
| 98.44 | 1.506 | 1.226 | 97.97 | 0.940 | 1.120 |
| 66.65 | 0.991 | 1.265 | 80.76 | 0.685 | 1.142 |
| 47.32 | 0.864 | 1.275 | 67.28 | 0.358 | 1.199 |
| 25.33 | 0.633 | 1.299 | 54.65 | 0.360 | 1.194 |
| 11.33 | 0.578 | 1.297 | 41.32 | 0.240 | 1.233 |
| - | - | - | 29.32 | 0.255 | 1.220 |
| - | - | - | 15.33 | 0.237 | 1.186 |

| Ethanol | | | Methanol | | |
|---|---|---|---|---|---|
| Pressure kN/$m^2$ | $C_4$x$10^4$ | $n_4$ | Pressure kN/$m^2$ | $C_4$x$10^4$ | $n_4$ |
| 98.17 | 0.450 | 1.186 | 98.64 | 0.308 | 1.213 |
| 84.84 | 0.918 | 1.117 | 79.98 | 0.261 | 1.229 |
| 71.31 | 0.545 | 1.166 | 66.65 | 0.240 | 1.236 |
| 44.65 | 0.432 | 1.180 | 53.32 | 0.203 | 1.249 |
| 31.32 | 0.264 | 1.226 | 41.32 | 0.154 | 1.275 |
| 19.33 | 0.354 | 1.182 | 27.99 | 0.115 | 1.299 |

It is seen from the above table that $C_4$ is a strong function of pressure and boiling fluids while $n_4$ does not change appreciably. Therefore, the value of $n_4$ for all the liquids investigated can be averaged arithmatically and is equal to 1.221 . Thus Eq. (6.2) assumes the following form:

$$f = C_4 \, q^{1.221} \qquad (6.2a)$$

## 6.7 VARIATION OF $h^*/h_1^*$ WITH $P/P_1$

Referring to Figures 6.5 through 6.8 it is seen that the heat transfer coefficient varies with the wall heat flux, system pressure and boiling fluids , It is also a fact that h depends on heating surface as well. These figures also reveal that for a given pressure, fluid and heating surface $h^*(=h/q^{0.7})$ bears a constant value (implying $h \propto q^{0.7}$). Thus $h^*$ is a function of pressure only for a given fluid and heating surface. Therefore, a plot of values of the ratio, $h^*/h_1^*$ against $P/P_1$ would become independent of boiling fluids and heating surfaces. Such a plot would correlate the data points of different investigatioss taken on different heating surfaces. This procedure was applied to the pool boiling data of present study alongwith the data of Cryder and Finalborgo [5] for water and methanol on a brass tube, data of Akiyama et al [79] for water on stainless steel tube and of Minchenko and Firsova [136] for distilled water. In Figure 6.16, the values of $h^*/h_1^*$ versus $P/P_1$ are plotted. The figure reveals that the relation between $h^*/h_1^*$ and $P/P_1$ ; is independent of boiling fluids and the heating surfaces as envisioned above.

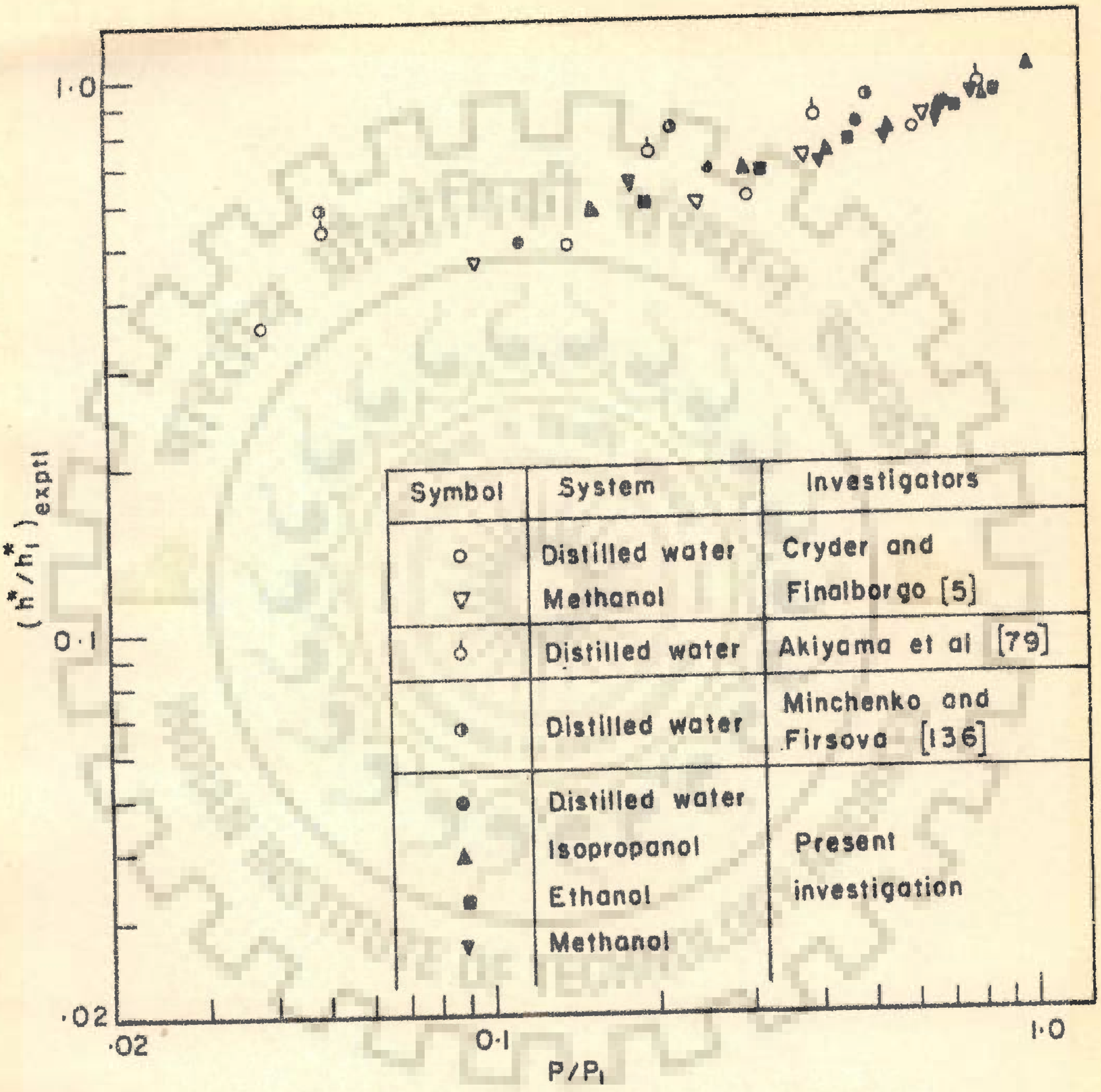

Fig. 6·16 A plot of $h^*/h_1^*$ vs $P/P_1$

## 6.8 PREDICTIONS FROM KUTATELADZE et al ANALYSIS [19]

Kutateladze et al analysis is available for predicting $h^*/h_1^*$ from the knowledge of physico-thermal properties of the boiling fluids, bubble departure diameter and bubble emission frequency. In this section the validity of the analysis is scrutinised by comparing the predicted values from it with the experimental data of present investigation and the data of Borishanskii et al [42].

Figure 6.17 shows a comparison between the experimental values of $h^*/h_1^*$ of present study with the predicted values from the above analysis [19]. It is found that the analysis is inconsistent with these experimental data.

Figure 6.18 represents comparison between the predicted values and the experimental data of Borishanskii et al for distilled water. The experimental data of Borishanskii et al corresponds to pressures greater than atmospheric pressure. It is seen from this figure that the Kutateladze et al analysis is consistent with the experimental data at pressures greater than one atmosphere. However, the analysis does not correlate the experimental data for atmospheric and subatmospheric pressures satisfactorily.

## 6.9 DETERMINATION OF CONSTANT M IN Eq(5.22)

In order to determine the value of constant M in Eq.(5.22) it was required to know the value of exponents m and a . Mikic and Rohsenow [43] recommended the value of

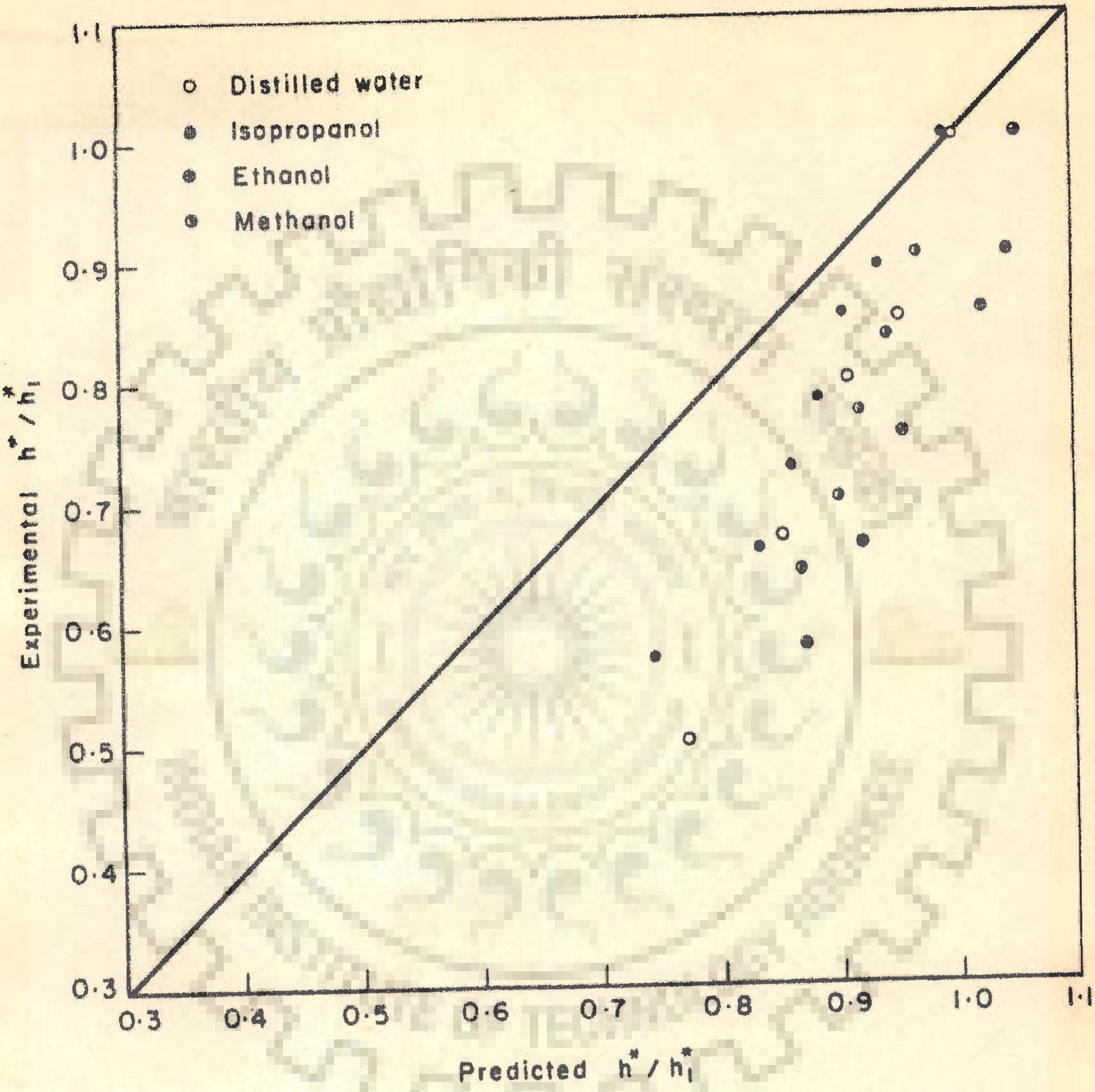

Fig. 6·17 Present experimental data compared with the Kutateladze et al analysis [19]

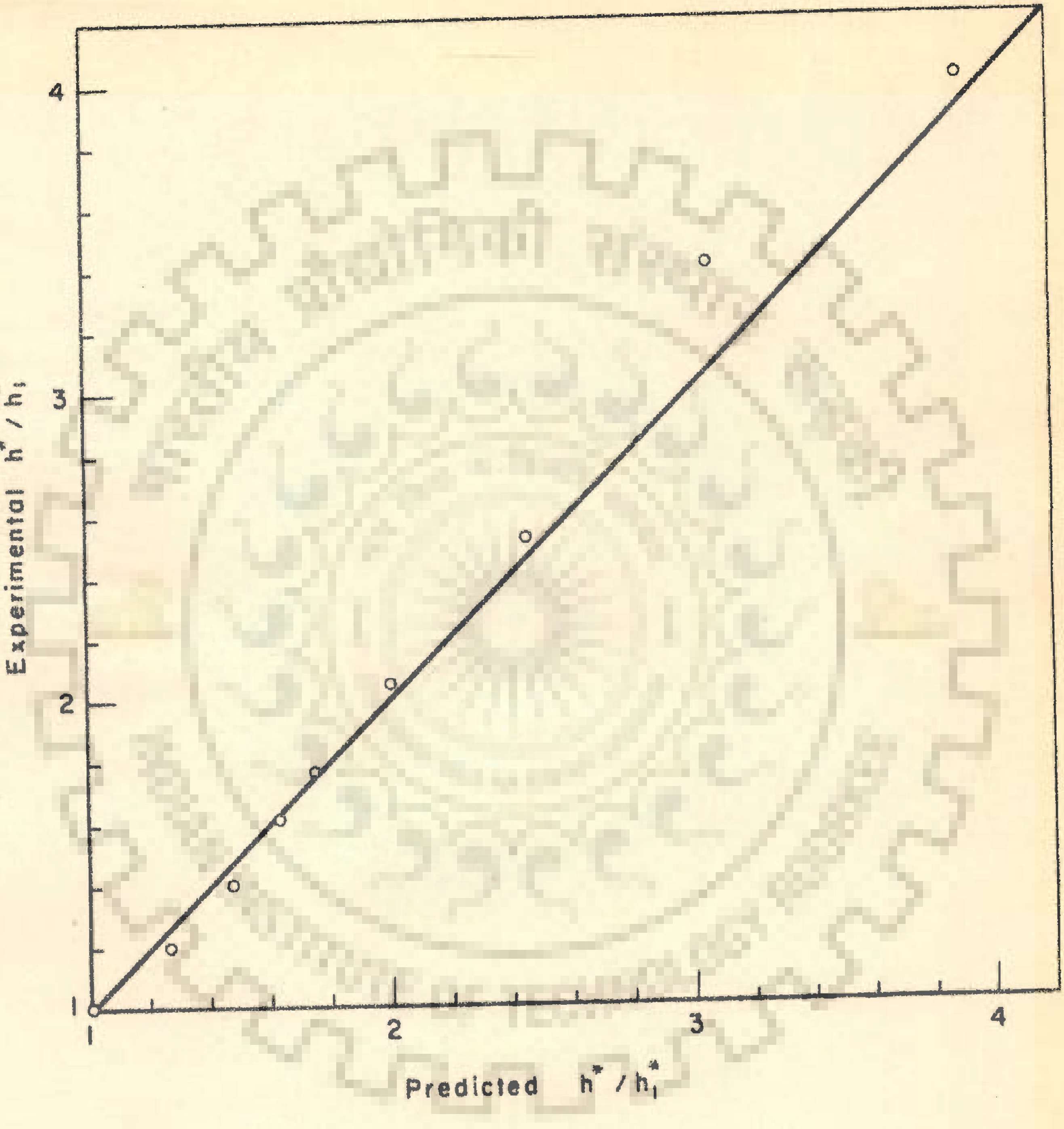

Fig. 6·18 Experimental data of Borishanskii et al [42] for distilled water compared with Kutateladze et al analysis [19] m as 2.5 for water implying that $q \propto \Delta T^{3.5}$ and 3.0 for organic liquids. However, the present data and data of many others conclude that $\bar{q}$ varies with $\Delta T$ raised to the power of 3.33. Therefore, it is thought desirable to choose the value of m as 2.33 (implying $q \propto \Delta T^{3.33}$). Exponent a was evaluated by the method of regression analysis using the entire experimental data. The values of exponent a ranged from 0.3046 to 0.3982 for all the pressures and fluids investigated. The analysis thus permitted to take an average value of 0.3683 for all the fluids employed in this investigation. It is interesting to mention here that this value is closer to 0.33 found by Wiebe and Judd [54] for the value of nf greater than $55 \times 10^3$ Bubbles/ $in^2$.sec. Hence these data seem to belong to the range of experimental parameters for which the value of nf are nearer to $55 \times 10^3$ Bubbles/ $in^2$.sec as recommended by the investigation of Wiebe and Judd which is based on careful photographic studies conducted by Judd[137].

In Chapter 5 , it is shown that M contains quantities whose measurement is extremely difficult . It is among many reasons due to the insufffcient information about the surface characteristics. Therefore, one of the suitable methods for the determination of M may be by using the experimental data in Eq.(5.22) with exponent m(=2.33) and a equal to 0.3683 as determined above. These computations revealed that the value of M changes with pressure and with surface-liquid combination as well. Table 6.4 shows the values of M for different pressures and surface-liquid combinations. Thus the argument made in Chapter 5, to treat the unknown constant M, representing the combined effects of pressure and surface-liquid combination seems to hold true.

Table 6.4 Experimental values of M for different pressures and surface-liquid combinations

Surface: Stainless steel

| Distilled Water | | Isopropanol | | Ethanol | | Methanol | |
|---|---|---|---|---|---|---|---|
| P kN/$m^2$ | Mx $10^5$ | P kN/$m^2$ | M/x$10^5$ | P kN/$m^2$ | Mx$10^5$ | P kN/$m^2$ | Mx$10^5$ |
| 98.442 | 1.14423 | 97.975 | 0.38961 | 98.175 | 0.22279 | 98.642 | 0.18169 |
| 66.650 | 1.29275 | 80.767 | 0.39247 | 84.845 | 0.22650 | 79.980 | 0.19005 |
| 47.321 | 1.55930 | 67.284 | 0.41480 | 71.315 | 0.24309 | 66.650 | 0.19864 |
| 25.327 | 2.07603 | 54.653 | 0.44132 | 44.655 | 0.29264 | 55.320 | 0.21521 |
| 11.330 | 2.92766 | 41.323 | 0.49432 | 31.325 | 0.33066 | 41.323 | 0.23885 |
| - | - | 29.326 | 0.57047 | 19.328 | 0.41347 | 27.993 | 0.29634 |
| - | - | 15.329 | 0.73346 | - | - | - | - |

Table 6.4 suggests that a mathematical relationship can be established relating constant M with system pressure P for a given surface-liquid combination. To accomplish this, regression analysis technique was employed. Three simple functions of hyperbolic, power and semilog were tried. In most of the cases these functions fitted quite well and gave coefficient of correlation varying from 0.8796 to 0.9927 for individual liquids. Tables 6.5,6.6 and 6.7 show the properties

of these respective regression lines fitted on the experimental data.

Table 6.5 Properties of hyperbolic regression lines for M and P ( $M = P/[A + BP]$ ) for different fluids

| Fluid | A | B | Coefficient of correlation |
|---|---|---|---|
| Distilled water | 8.4759386E+04 | -6.4150661E+05 | 0.9022 |
| Isopropanol | 2.7128524E+05 | -2.2458743E+06 | 0.9485 |
| Ethanol | 4.8756164E+05 | -5.1291132E+06 | 0.9583 |
| Methanol | 6.2898417E+05 | -8.3686995E+05 | 0.9736 |

Table 6.6 Properties of power regression lines for M and P ( $M = AP^B$ ) for different fluids

| Fluid | A | B | Coefficient of correlation |
|---|---|---|---|
| Distilled water | 8.6193800E-05 | -4.4409803E-01 | 0.9927 |
| Isopropanol | 1.8924030E-05 | -3.5615637E-01 | 0.9801 |
| Ethanol | 1.2634947E-05 | -3.8381813E-01 | 0.9878 |
| Methanol | 1.0130812E-05 | -3.823097E-01 | 0.9744 |

Table 6.7 Properties of semilog regression lines for M and P( $M = A + B \log_e P$ ) for different fluids

| Fluid | A | B | Coefficient of correlation |
|---|---|---|---|
| Distilled water | 4.8334136E-05 | -8.2849672E-06 | 0.9843 |
| Isopropanol | 1.2210266E-05 | -1.8919490E-06 | 0.9715 |
| Ethanol | 7.3989023E-06 | -1.1518216E-06 | 0.9810 |
| Methanol | 5.7514742E-06 | -8.8112245E-07 | 0.9601 |

An inspection of these Tables reveals that the power and semilog regression lines are fitting the data better than the hyperbolic expression. However, the power function is the best in all cases where the coefficient of correlation is of the order of 0.9927. Thus the following equations in the form of power function could represent the relation between M and P for respective fluids:

$$\left.\begin{array}{ll}\text{For Distilled water} & \\ M \approx 8.61938\text{E-}05(P)^{-4.4409803\text{E-}01} & (a) \\ \text{For Isopropanol} & \\ M \approx 1.8924030\text{E-}05(P)^{-3.5615637\text{E-}01} & (b) \\ \text{For Ethanol} & \\ M \approx 1.2634947\text{E-}05(P)^{-3.8381813\text{E-}01} & (c) \\ \text{For Methanol} & \\ M \approx 1.0130812\text{E-}05(P)^{-3.823097\text{E-}01} & (d)\end{array}\right\} (6.3)$$

Figure 6.19 shows the experimentally - determined values of M against the pressure P for different fluids investigated. The straight lines passing through the data points of respective fluids represent the equations, Eqs. (6.3a, 6.3b, 6.3c & 6.3d). It is seen that these equations well-correlate the experimental values of M.

Using the values of m and a as determined above, in Eq. (5.22) the following equation for absolute heat transfer coefficient results:

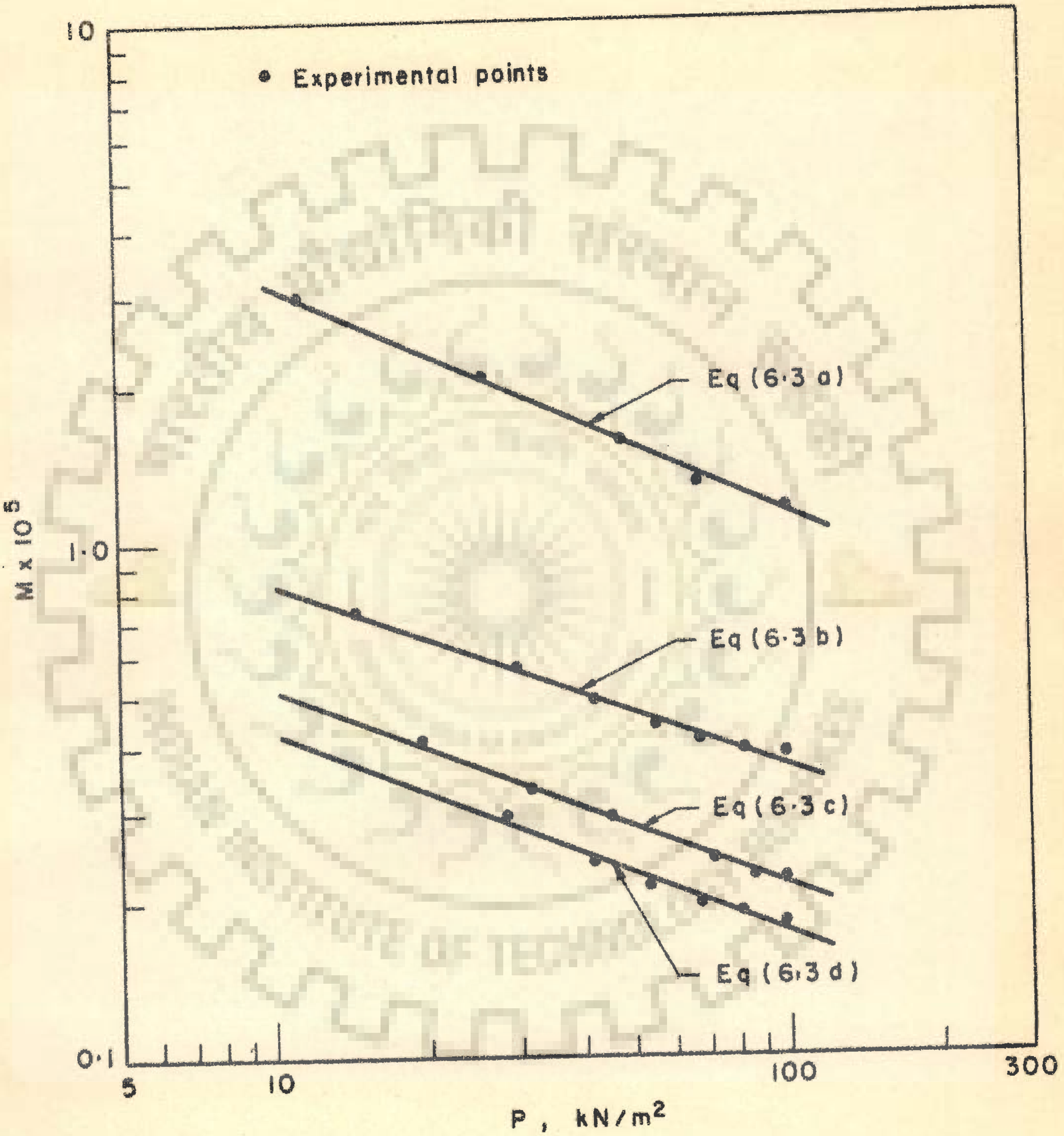

Fig. 6·19 Plots between M vs P for different fluids

$$h = M\left[\left\{\left(\frac{\lambda^{2.5}\rho_v}{T_s^{1.5}\,\sigma}\right)\left(\frac{q}{k_\ell g}\right)^{0.3}\left(\frac{1}{C_\ell}\right)^{1.2}\right\}^{2.33}\{f\}\right]^{0.3683} \qquad (6.4)$$

where M is calculated from Eq. (6.3)

It is of interest to examine the present analysis, Eq. (6.4) in conjunction with Eq. (6.2a). Using the value of f from Eq. (6.2a) into Eq. (6.4) the following equation results:

$$h = M\,C_4^{0.3683}\left[\left(\frac{\lambda^{2.5}\,\rho_v}{T_s^{1.5}\,\sigma}\right)\left(\frac{1}{k_\ell\,g}\right)^{0.3}\left(\frac{1}{C_\ell}\right)^{1.2}\right]^{2.33}\left[q^{0.7065}\right]$$

or $h$ = constant $q^{0.7065}$ for a given pressure.

Thus the widely accepted relationship that h varies with q raised to the power of 0.7 is deduced successfully.

The predicted values of absolute values of h from Eq. (6.4) are compared with the experimental ones in Figures 6.20 through 6.23 respectively for distilled water, isopropanol, ethanol and methanol. The agreement is excellent within a maximum deviation of ± 10 per cent.

## 6.10 DETERMINATION OF $M/M_1$ IN Eq.(5.25)

In order to determine the relationship between $M/M_1$ and $P/P_1$ the regression analysis, as detailed in Section 6.9, was employed considering all the data points. Fortunately, it is found that the data points for all the fluids are represented

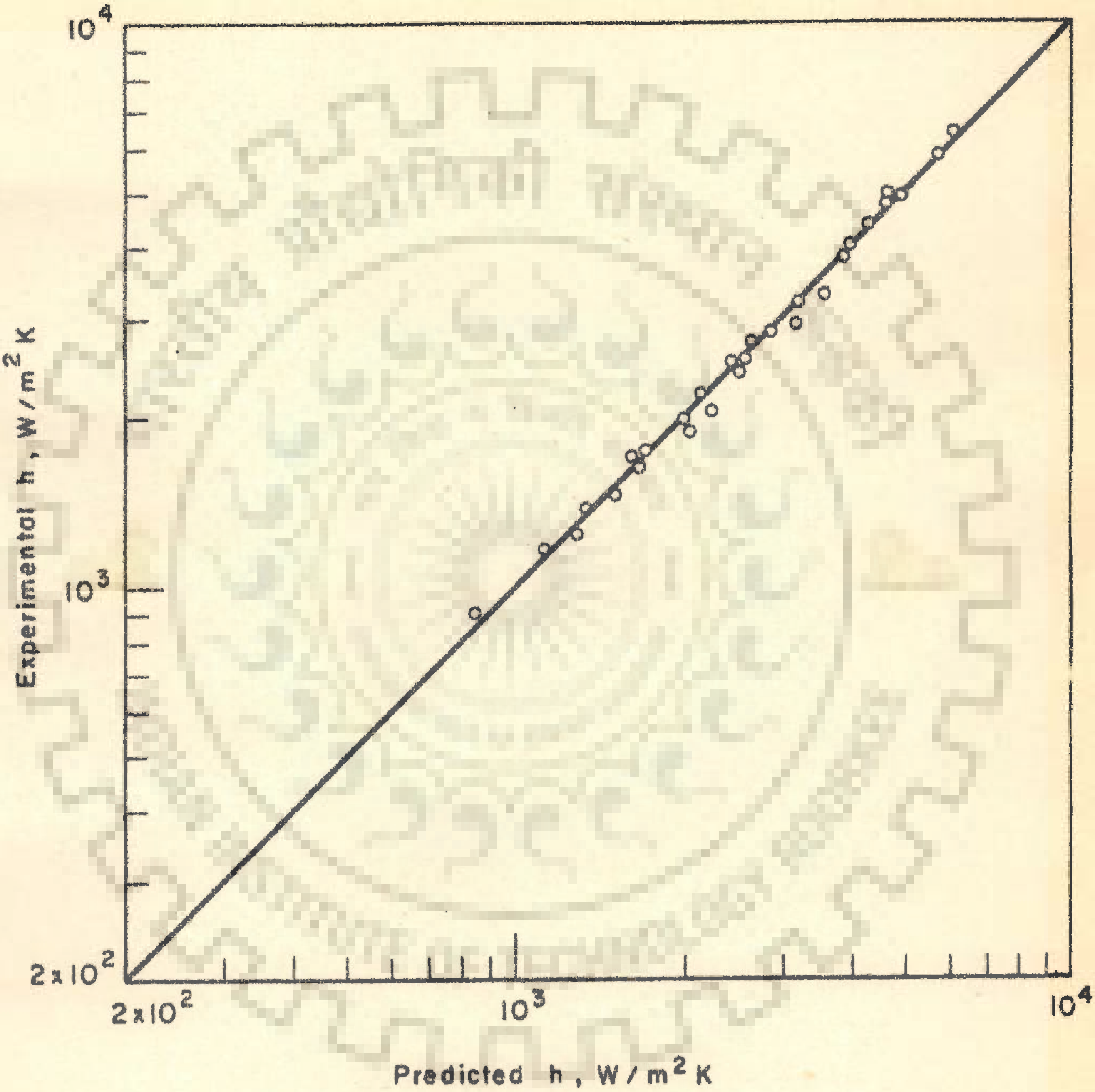

Fig. 6·20 Present experimental data for distilled water compared with the present analysis, Eqs(6·3a & 6·4)

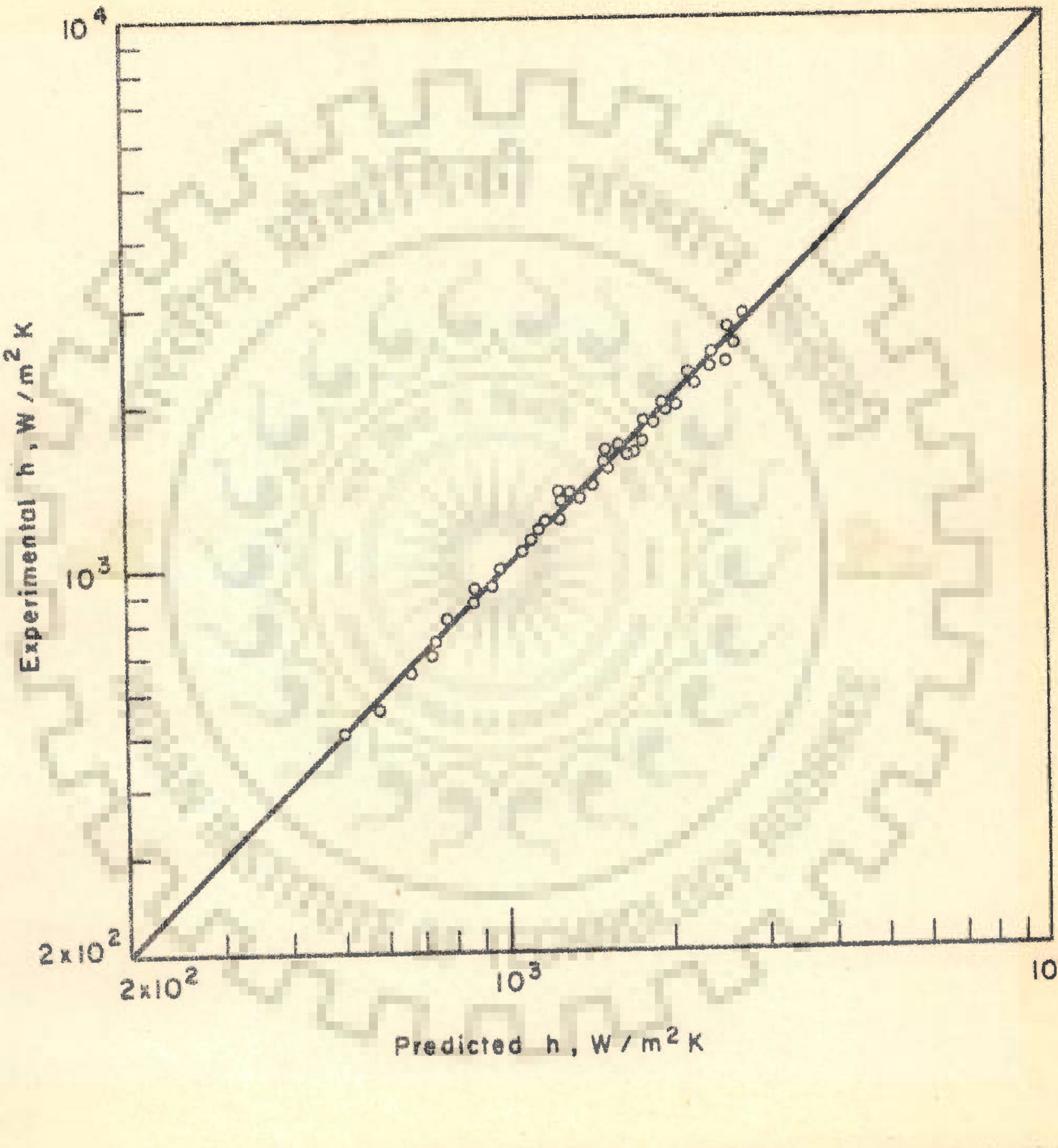

Fig. 6·21 Present experimental data for isopropanol compared with the present analysis, Eqs (6·3b & 6·4)

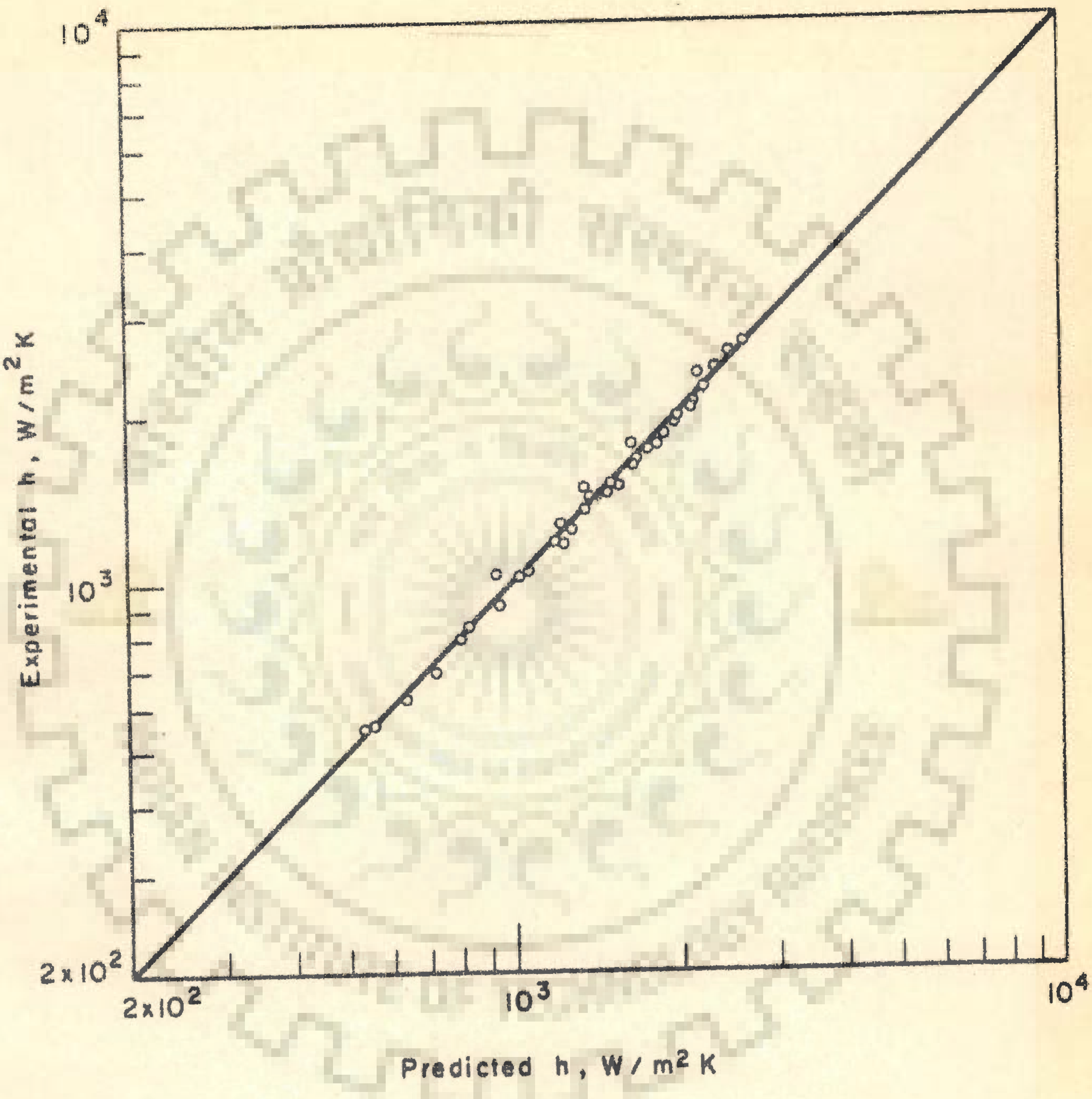

Fig. 6·22 Present experimental data for ethanol compared with the present analysis, Eqs (6·3c & 6·4)

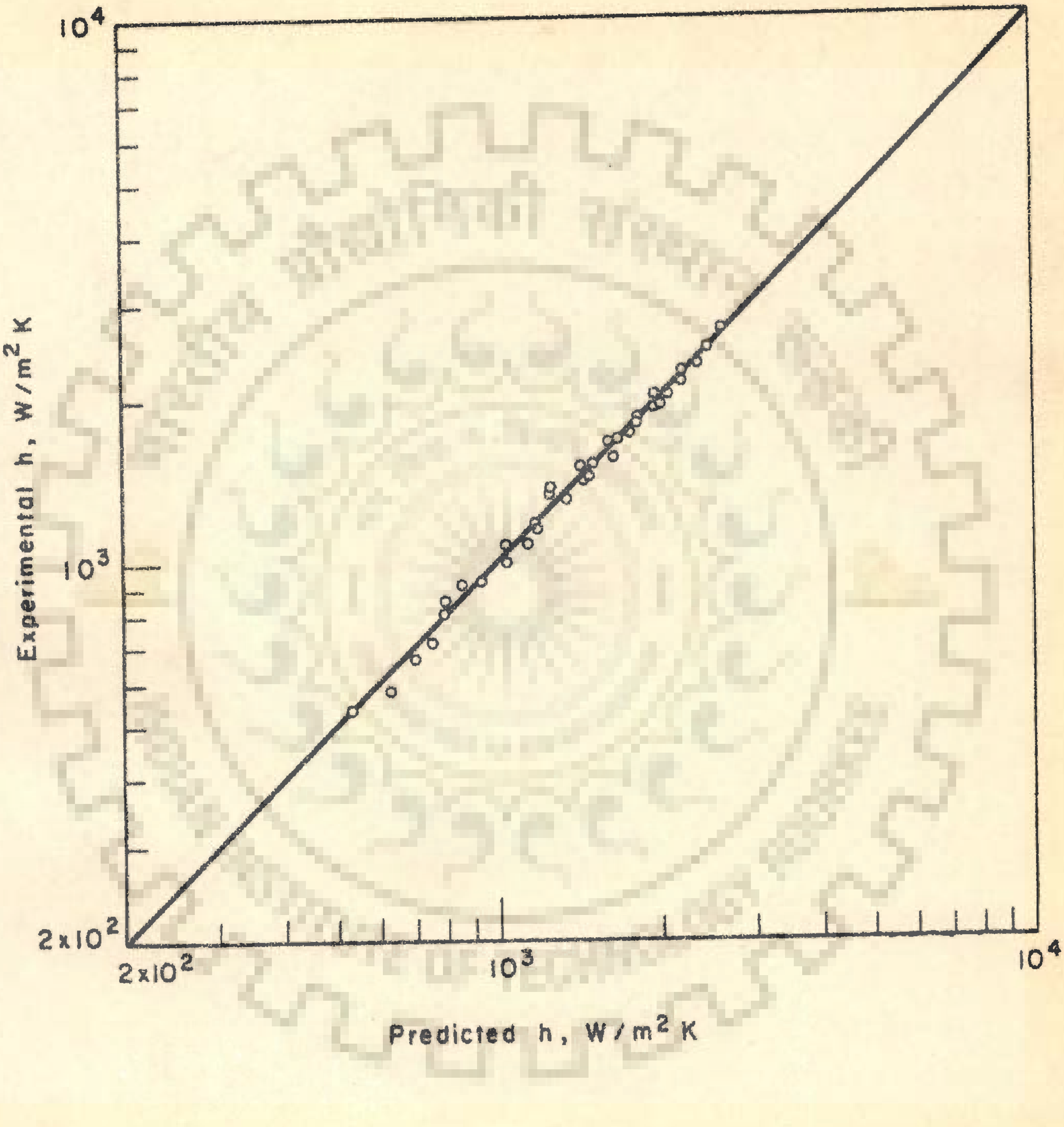

Fig. 6·23 Present experimental data for methanol compared with the present analysis, Eqs (6·3d & 6·4) by a single line. The properties of these regression lines are given in Table 6.8 . Inspection of this Table reveals that the coefficient of correlation is the best in case of power function fit. Thus the following form of power function relates all the data points:

$$\frac{M}{M_1} \approx 0.95831645\,(P/P_1)^{-0.3997825} \qquad (6.5)$$

Table 6.8 Properties of regression lines for $M/M_1$ and $P/P_1$ for all the fluids

| Regression line | A | B | Coefficient of correlation |
|---|---|---|---|
| Hyperbolic | 1.0360125E+00 | -8.9943040E-02 | 0.9185 |
| Power function | 9.5831645E-01 | -3.997825E-01 | 0.9689 |
| Semi log | 9.0278652E-01 | -5.8567074E-01 | 0.94235 |

Figure 6.24 shows a plot between the average experimental tal values of $M/M_1$ of all the fluids investigated and $P/P_1$. The solid line is the recommended equation, Eq.(6.5). From this figure it is seen that the agreement is excellent suggesting that Eq.(6.5) represents a generalised correlation for determining $M/M_1$ unlike Eq.(6.3).

It is thus seen that $M/M_1$ depends upon $P/P_1$ only as envisioned in Eq.(5.25). This seems to be possible only if M is a multiplicative function of pressure and surface-liquid combination. This is in conformity to the speculations made in Chapter 5 [cf. Eq.(5.23) ]. The fact that $M/M_1$ does depend on

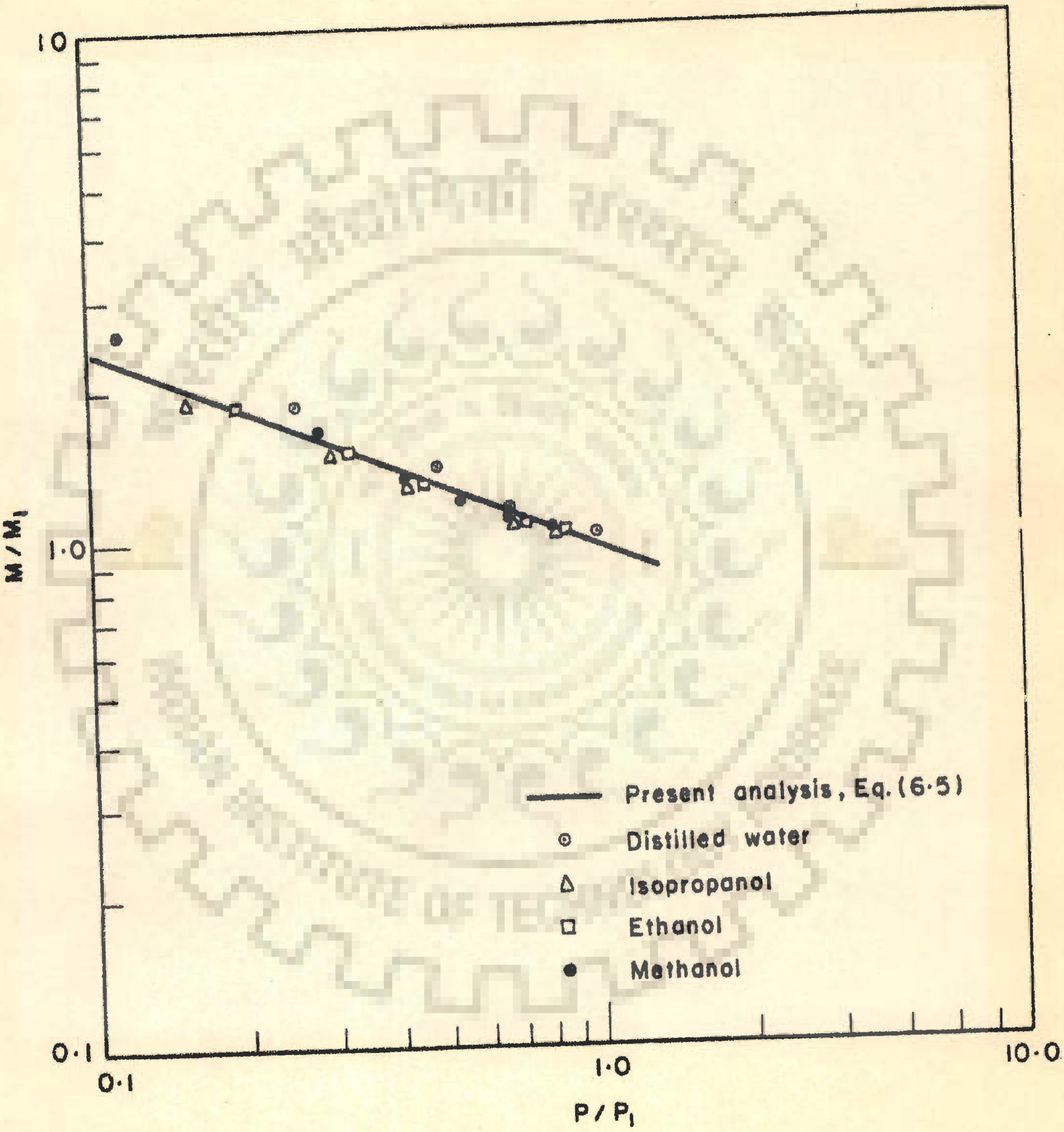

Fig. 6·24 Plot between M / $M_1$ vs P / $P_1$

pressure only and not on surface-liquid combination is scrutinized by comparing the experimental data of other investigators on different heating surfaces with the predicted values from this analysis in Section 6.11.

From the values of a (=0.3683), m(=2.33), Eq(6.5) and Eq(5.24) the following equation results for $h^*/h_1^*$ :

$$\frac{h^*}{h_1^*} = 0.95831645(P/P_1)^{-0.3997825}\left[\left\{\left(\frac{\lambda}{\lambda_1}\right)^{2.5}\left(\left(\frac{\rho_v}{\rho_{v1}}\right)\left(\frac{T_{s1}}{T_s}\right)\right)^{1.5}\left(\frac{\sigma_1}{\sigma}\right)\left(\frac{k_{\ell 1}}{k_\ell}\right)^{0.3}\left(\frac{q}{q_1}\right)^{0.3}\left(\frac{C_{\ell 1}}{C_\ell}\right)^{1.2}\right\}^{2.33} \times \left\{\frac{f}{f_1}\right\}\right]^{0.3683} \qquad (6.6)$$

Subscript '1' denotes the values corresponding to normal boiling point.

Figure 6.25 shows a comparison between the predicted values of $h^*/h_1^*$ from the proposed correlation, Eq.(6.6) and the experimental values of the present investigation. The data points show an excellent agreement with a maximum deviation of ± 10 per cent.

Further, Eq.(6.6) can be used to calculate the absolute values of heat transfer coefficient for a surface-liquid combination at subatmospheric pressures, without resort to experiment, if the value of heat transfer coefficient

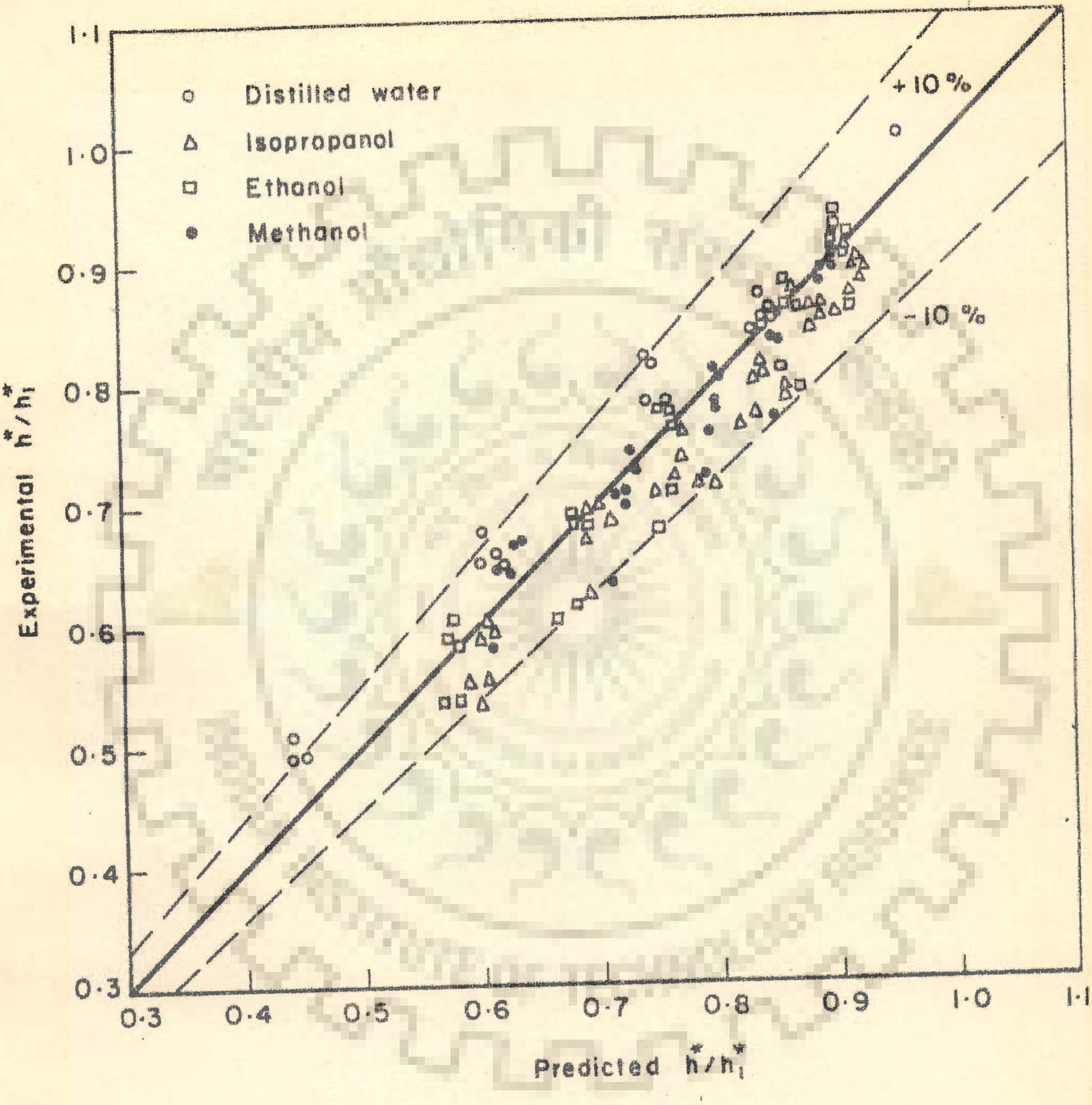

Fig. 6·25 Present experimental data compared with the present analysis, Eq (6·6) at normal boiling point is known for that surface-liquid combination.

## 6.11 COMPARISON OF PREDICTED HEAT TRANSFER COEFFICIENTS FROM PRESENT ANALYSIS WITH EXISTING DATA

To bolster the validity of the present analytical analysis, Eq.(6.6) the experimental data of Cryder and Finalborgo [5] for water, methanol and carbon tetrachloride on brass tube, and Raben et al [94] for water on the top end of a vertical cylindrical copper rod of high purity, at subatmospheric pressures were employed.

The comparison between the experimental values of $(h^*/h_1^*)$ and the predicted ones from the present analysis is made in Figure 6.26. It is seen that the analysis is consistent with the experimental data obtained on different surface-liquid combinations. It also concludes that $h^*/h_1^*$ does not depend on surface-liquid combination implying that $M/M_1$ is independent of the surface-liquid combination.

The above conclusion is in conformity with the results of Cryder and Finalborgo [5] that the ratio $h^*/h_1^*$ is relatively independent of the arrangement and composition of the heating element. Further, Bonilla and Perry [10] have also observed that the boiling surface does not change the ratios of the coefficients obtained at different pressures.

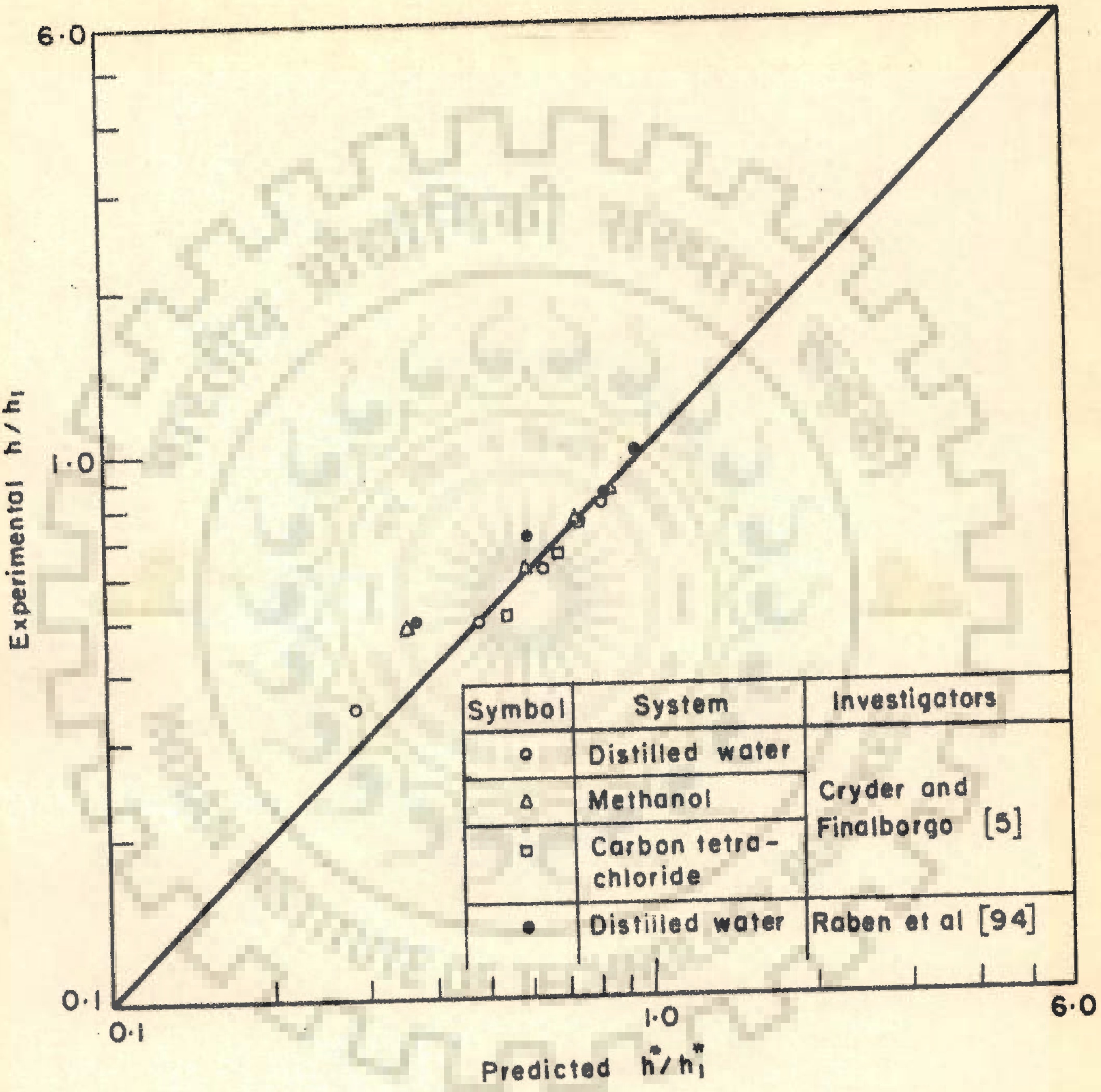

Fig. 6·26 Existing experimental data compared with the present analysis, Eq (6·6)

## 6.12 COMPARISON BETWEEN PREDICTED HEAT TRANSFER COEFFICIENTS FROM VARIOUS CORRELATIONS AND EXPERIMENTAL VALUES OF PRESENT STUDY

A comprehensive literature review in Chapter 2 shows that a large number of correlations have been proposed for heat transfer in nucleate pool boiling. Table 2.6 lists the selected heat transfer correlations for comparing the experimental data of this study with the predictions from them. The criteria of their selection have already been described. For ready reference some of the pertinent points of these correlations [4],[33],[36],[39],[43] are as follows:

Jakob and Linke [4] correlation is a semitheoretical correlation based on the assumptions that there exists a linear relationship between the heat flux and number of active sites per unit area of the heating surface and product $fD_b$ is a constant.

The correlations of Kutateladze [39] and Kichigen and Tobilevich [39] have been derived empirically by choosing pertinent groups which explain the phenomenon of nucleate pool boiling. One of the important groups included in these correlations is $K_p$ which is a criterion for pressure term in boiling.

The correlation of Labuntsov [36] is also an empirical correlation. The essential features of the correlation is that it contains criterion for bubble break-off frequency along with Peclet number for boiling.

The Forster- Greif correlation [33] is a theoretical approach for wall heat flux determination based on vapour-liquid exchange mechanism.

Mikic and Rohsenow correlation [43] assumes that the main mechanism of heat transfer in nucleate boiling is transient heat conduction to, and subsequent replacement of, the superheated layer around boiling sites associated with bubble departure.

Figures 6.27 and 6.28 represent the comparison between the present experimental data and the predicted values from the Jakob and Linke correlation [4]. Observations from these plots reveal that the Jakob -Linke correlation underpredicts the value. This observation is not surprising because this correlation was recommended for the experimental data for pressures greater than one atmosphere, while the present experimental data pertain to subatmospheric pressures. Another important reason for this disagreement is attributable, probably, to the fact that their correlation assumed the product $fD_b$ to be a constant. But the recent studies [45],[118] have shown that $fD_b$ is not a constant value and depends both upon heat flux and pressure.

Figures 6.29 and 6.30 are the plots which compare the experimental data of this study with the predictions from the Kutateladze correlation [39]. From these figures it is inferred that predicted values from the Kutateladze correlation are always lower than the experimental values. For methanol

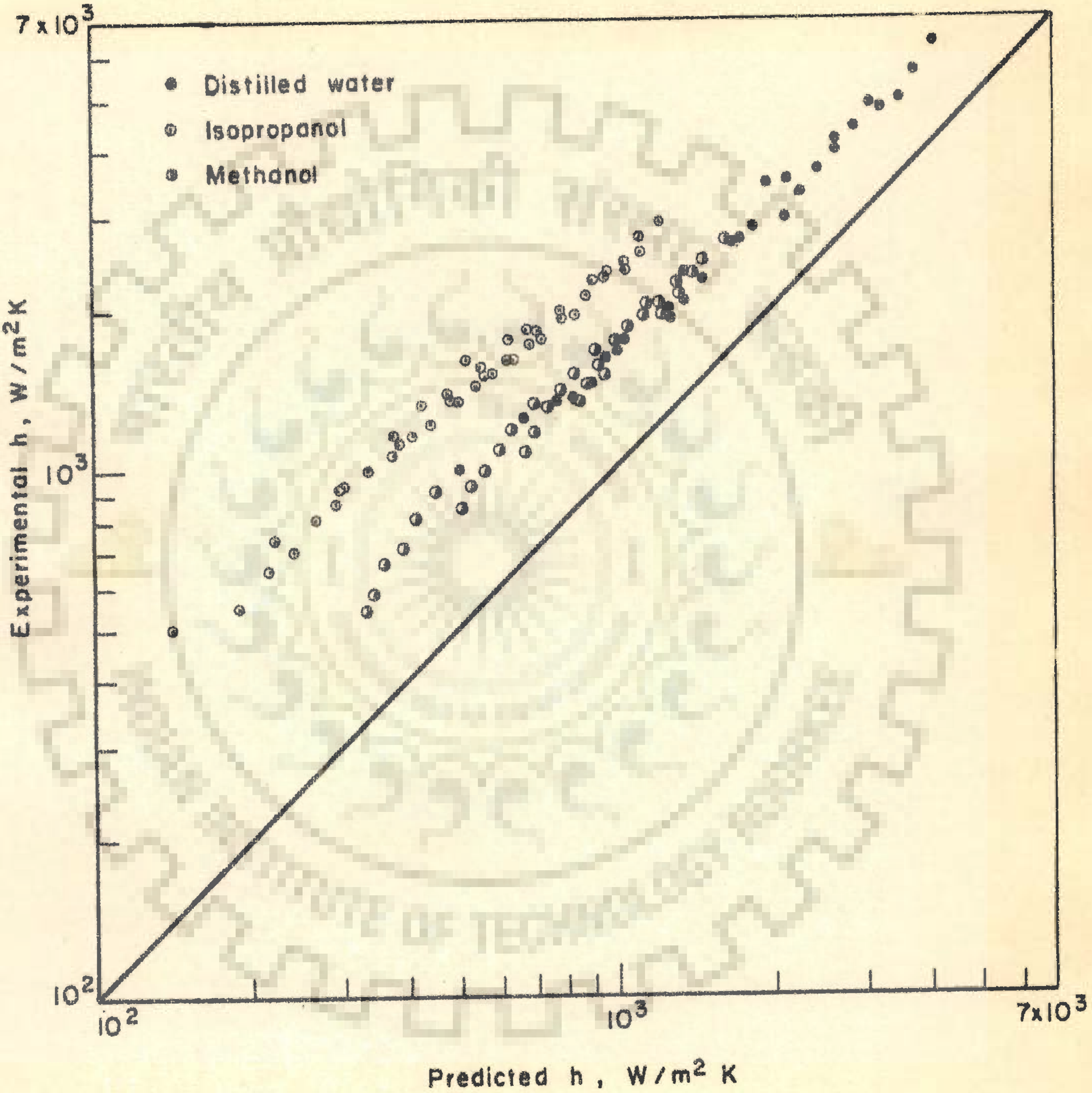

Fig. 6·27 Present experimental data compared with predictions from the Jakob & Linke correlation [4]

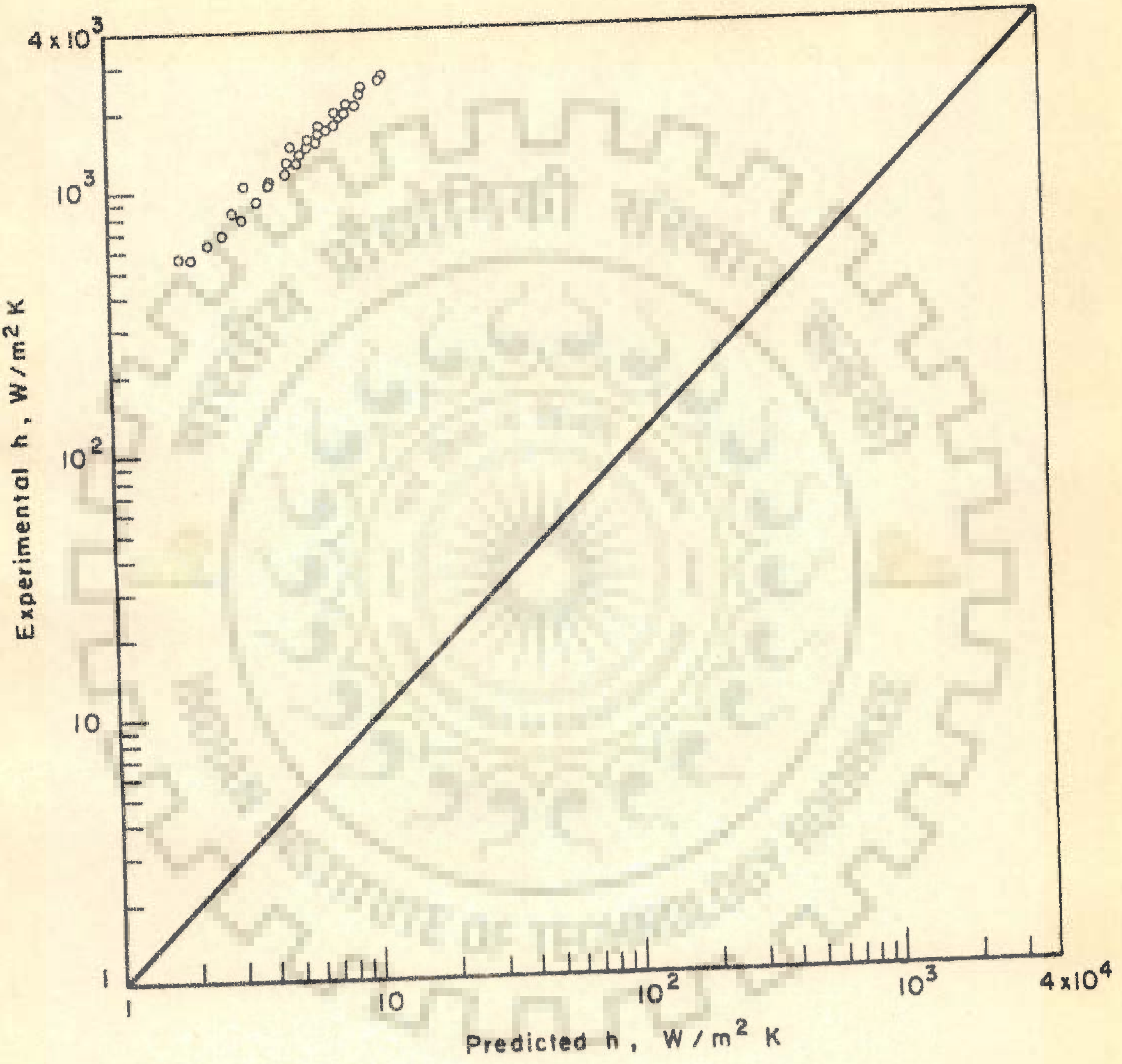

Fig. 6·28 Present experimental data for ethanol compared with predictions from the Jakob & Linke correlation [4]

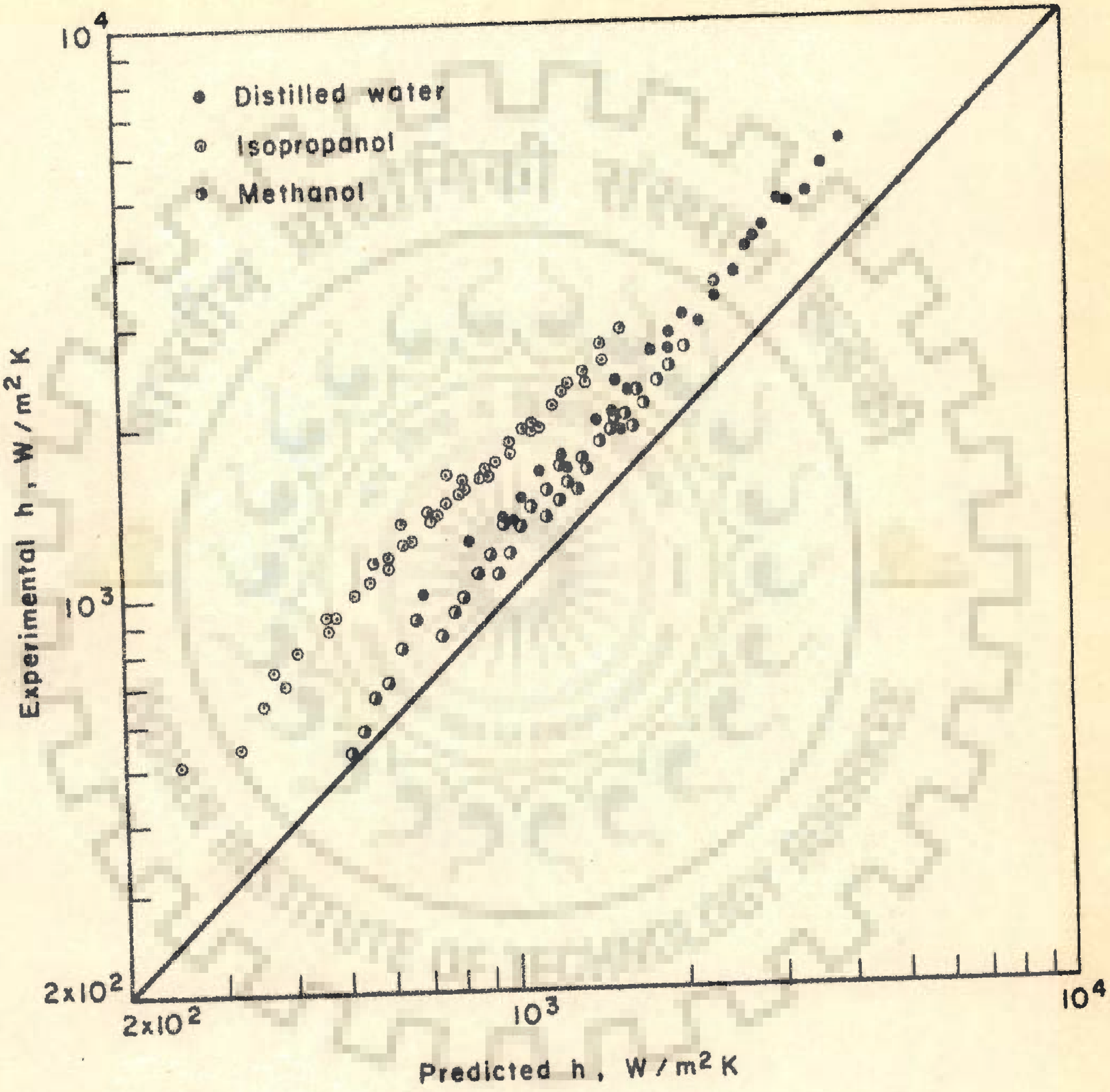

Fig. 6·29 Present experimental data compared with predictions from the Kutateladze correlation [39]

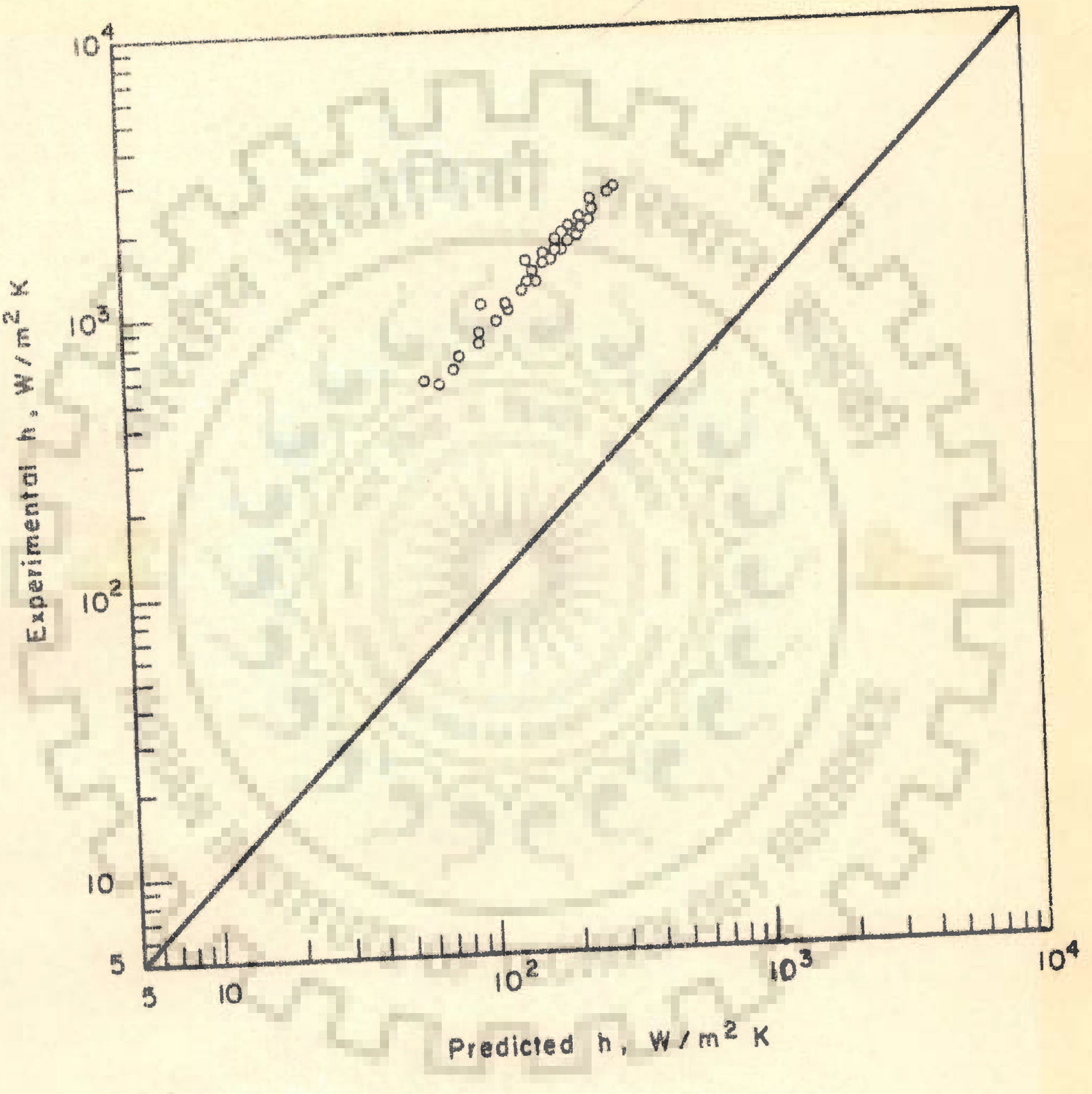

Fig. 6·30 Present experimental data for ethanol compared with predictions from the Kutateladze correlation [39] the behaviour is same as for other liquids, however, the deviation is not much. Deviation is quite appreciable for the data of ethanol.

Figures 6.31 and 6.32 compare the present experimental data with the predictions from the Kichigen and Tobilevich correlation [39]. The agreement is good for the data of methanol, fairly well for the data of distilled water and isopropanol and poor for ethanol.

Figures 6.33 and 6.34 exhibit the comparison between experimental values of heat transfer coefficient and those calculated from the Labuntsov correlation [36]. The agreement is quite good for the experimental data for isopropanol and fairly well for distilled water. The correlation overpredicts the data of methanol and underpredicts the data of ethanol.

Based on the observations as above the following points can be concluded:

1. The correlations due to Kutateladze, and Kichigen and Tobilevich appear to correlate the experimental data for methanol well.
2. The Labuntsov correlation is found to well-correlate the data for isopropanol and it correlates the data for distilled water fairly well.
3. All the correlations are poor predictors for the data of ethanol. Also, they underpredict the values of heat transfer coefficient.

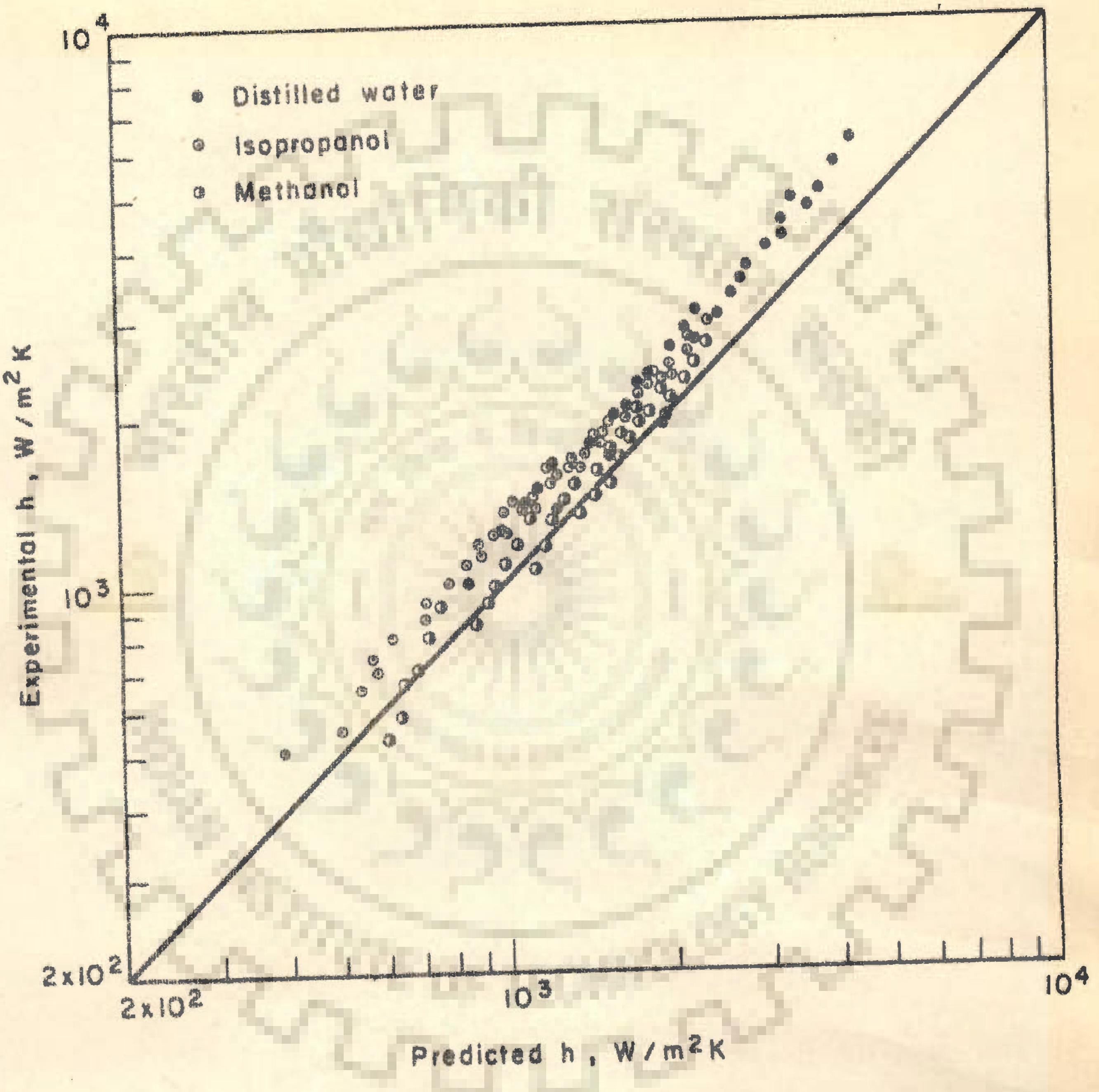

Fig. 6·31 Present experimental data compared with predictions from the Kichigen & Tobilevich correlation [39]

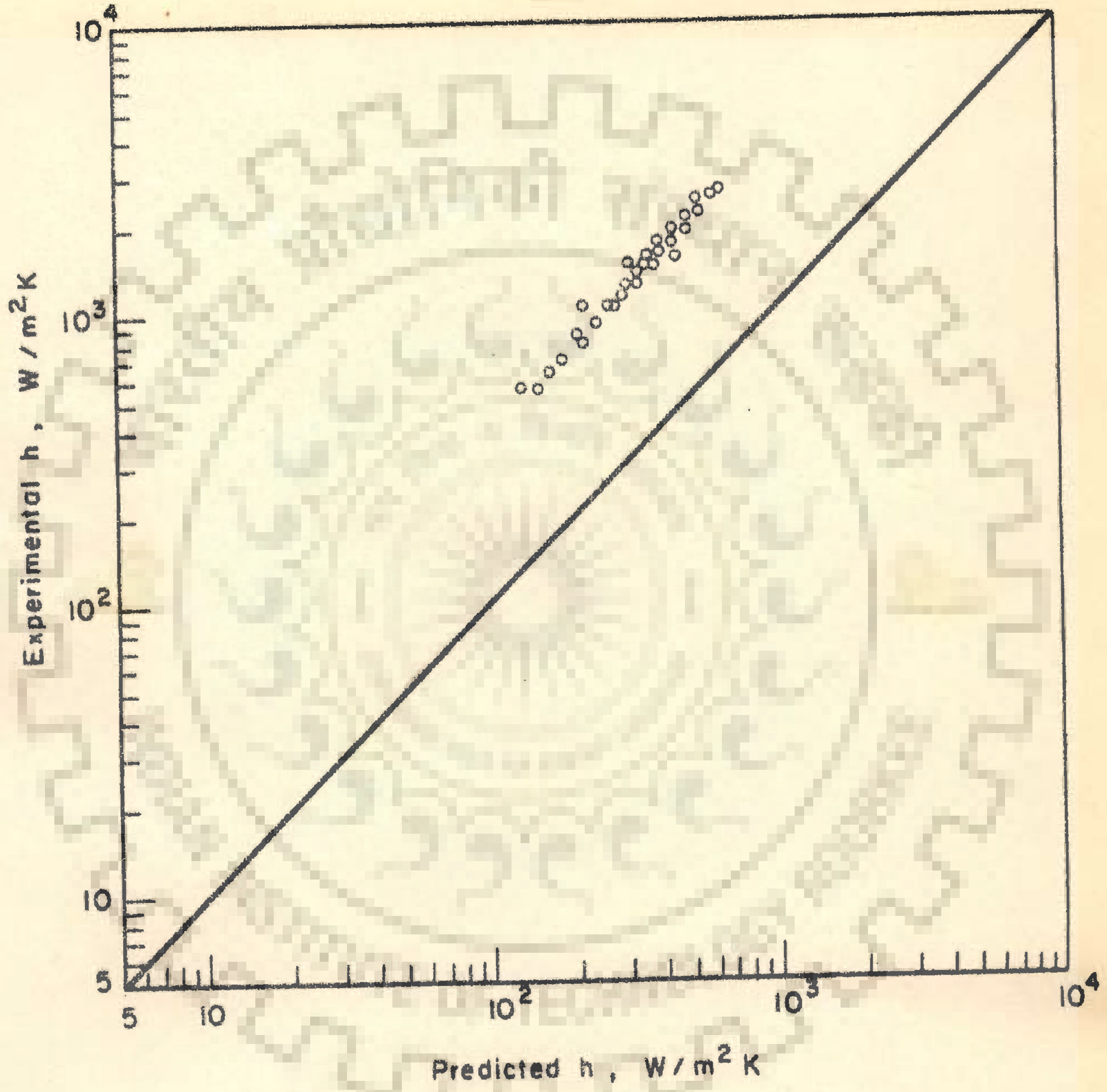

Fig. 6·32 Present experimental data for ethanol compared with predictions from the Kichigen & Tobilevich correlation [39]

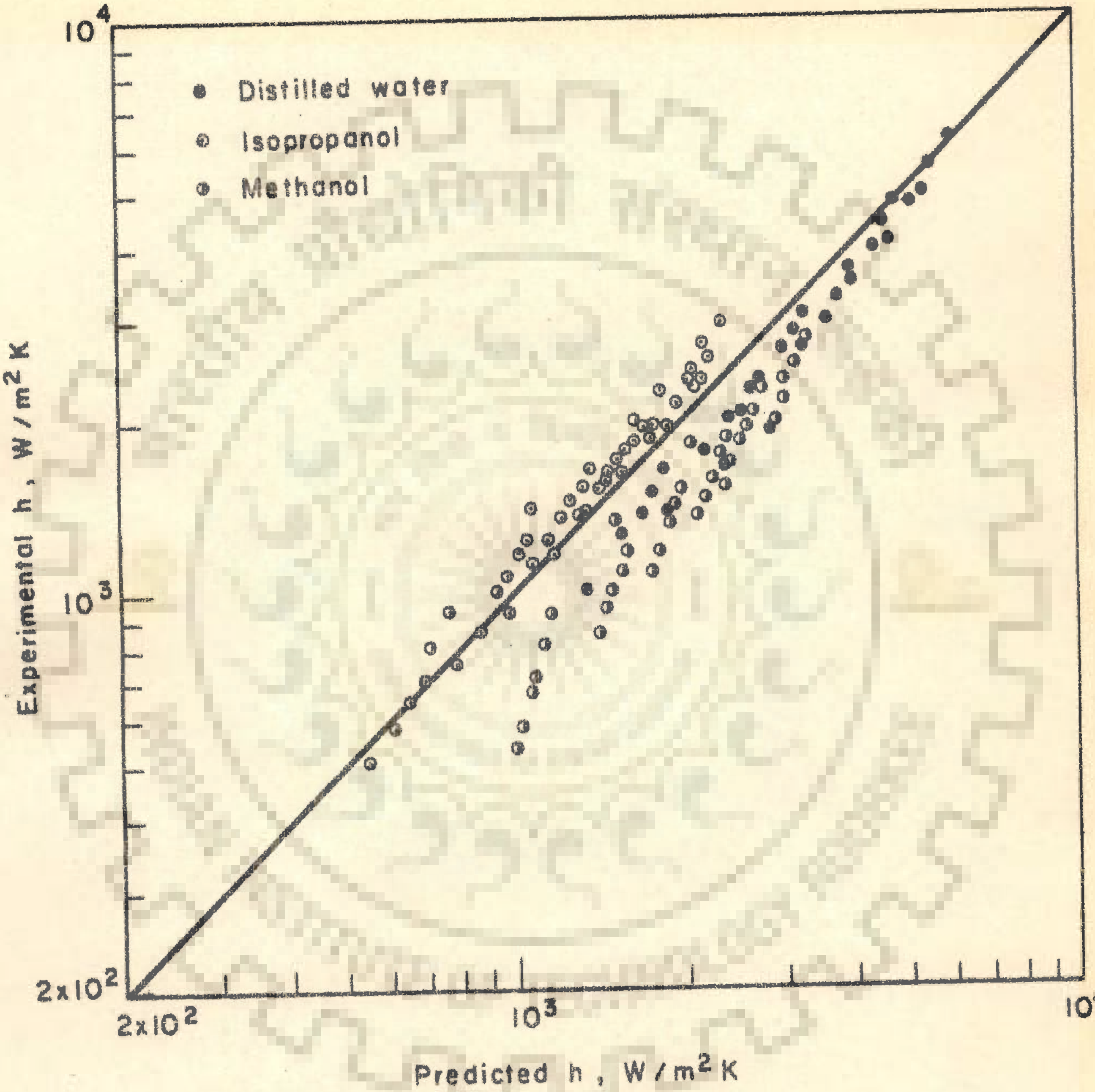

Fig. 6·33 Present experimental data compared with predictions from the Labuntsov correlation [36]

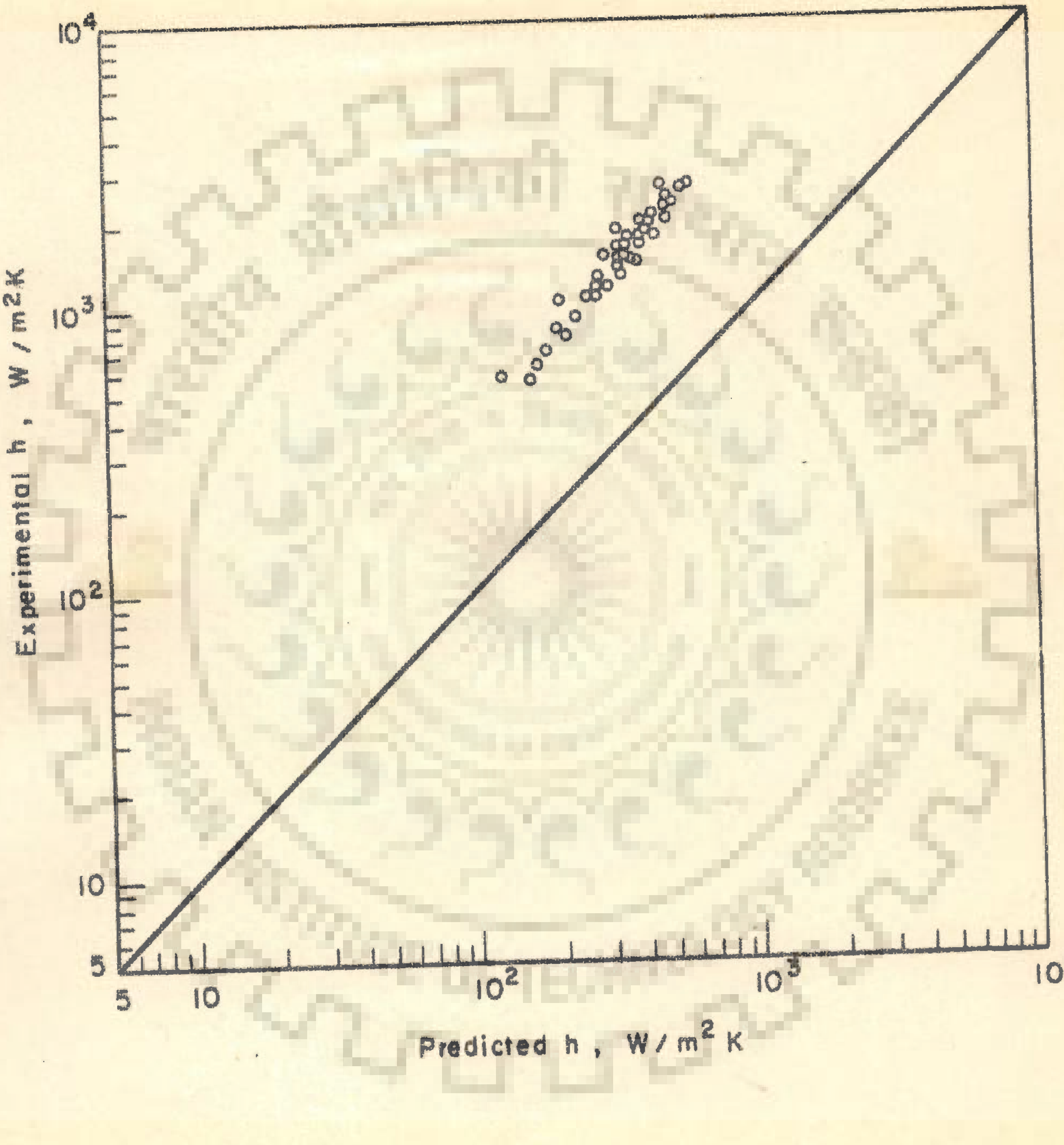

Fig. 6·34 Present experimental data for ethanol compared with predictions from the Labuntsov correlation [36]

Finally it would be concluded that none of these correlations is a good predictor for all the present experimental data at subatmospheric pressures and therefore, they do not possess general applicability for liquids differing in their physico-thermal properties.

Figures 6.35 and 6.36 represent the comparison between the present experimental values of heat flux and the predicted from the Forster and Greif correlation [33]. An inspection of these figures shows that the correlation overpredicts the value of heat flux for distilled water, isopropanol and methanol while it underpredicts the heat flux value for ethanol. These figures also reveal that their correlation seems to predict heat flux in its dependence on pressure and boiling fluid. The pressure effect is more pronounced for ethanol as shown by Figure 6.36. Thus their correlation does not possess the general applicability for the conditions employed in the present investigation.

Figures 6.37 and 6.38 represent the comparison between the present experimental data and the calculated values from the Mikic and Rohsenow correlation [43]. Abscissa $\phi \Delta T$ has been defined in Eq(2.44). It appears from these figures that this correlation satisfactorily correlates present experimental data for all the four fluids.

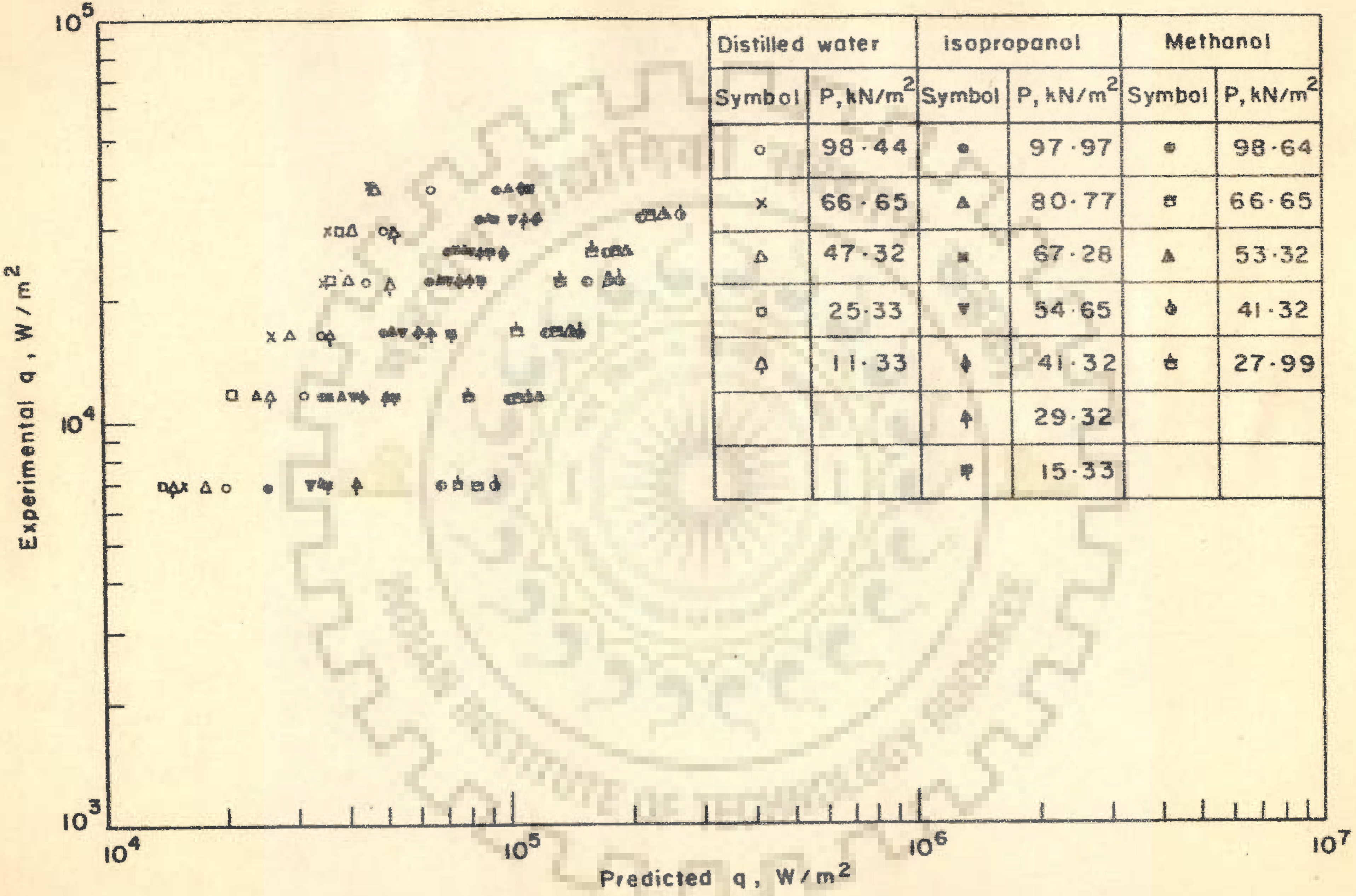

Fig. 6·35 Present experimental data compared with predictions from the Forster and Greif correlation [33]

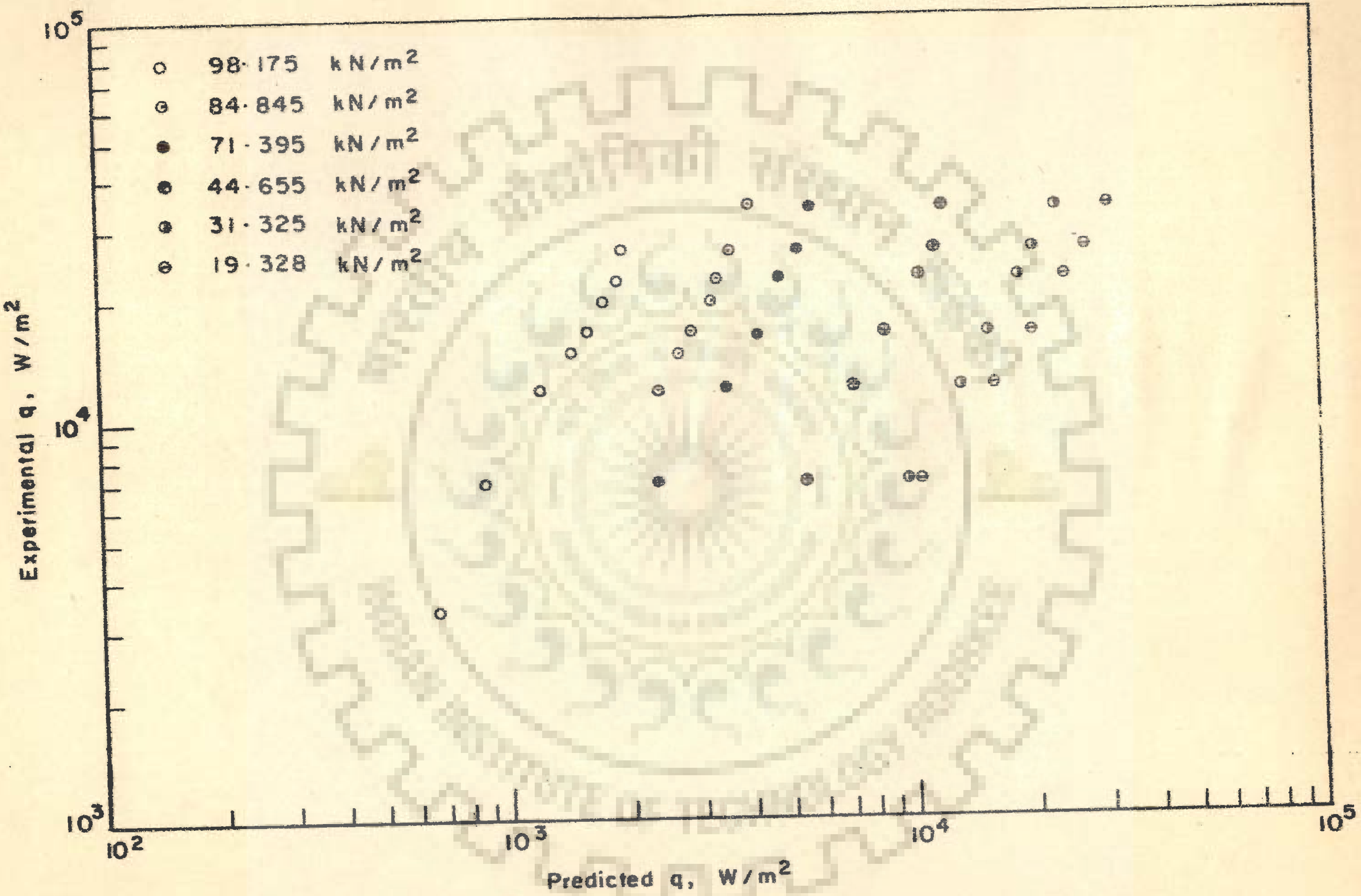

Fig. 6·36 Present experimental data for ethanol compared with predictions from the Forster and Greif correlation [33]

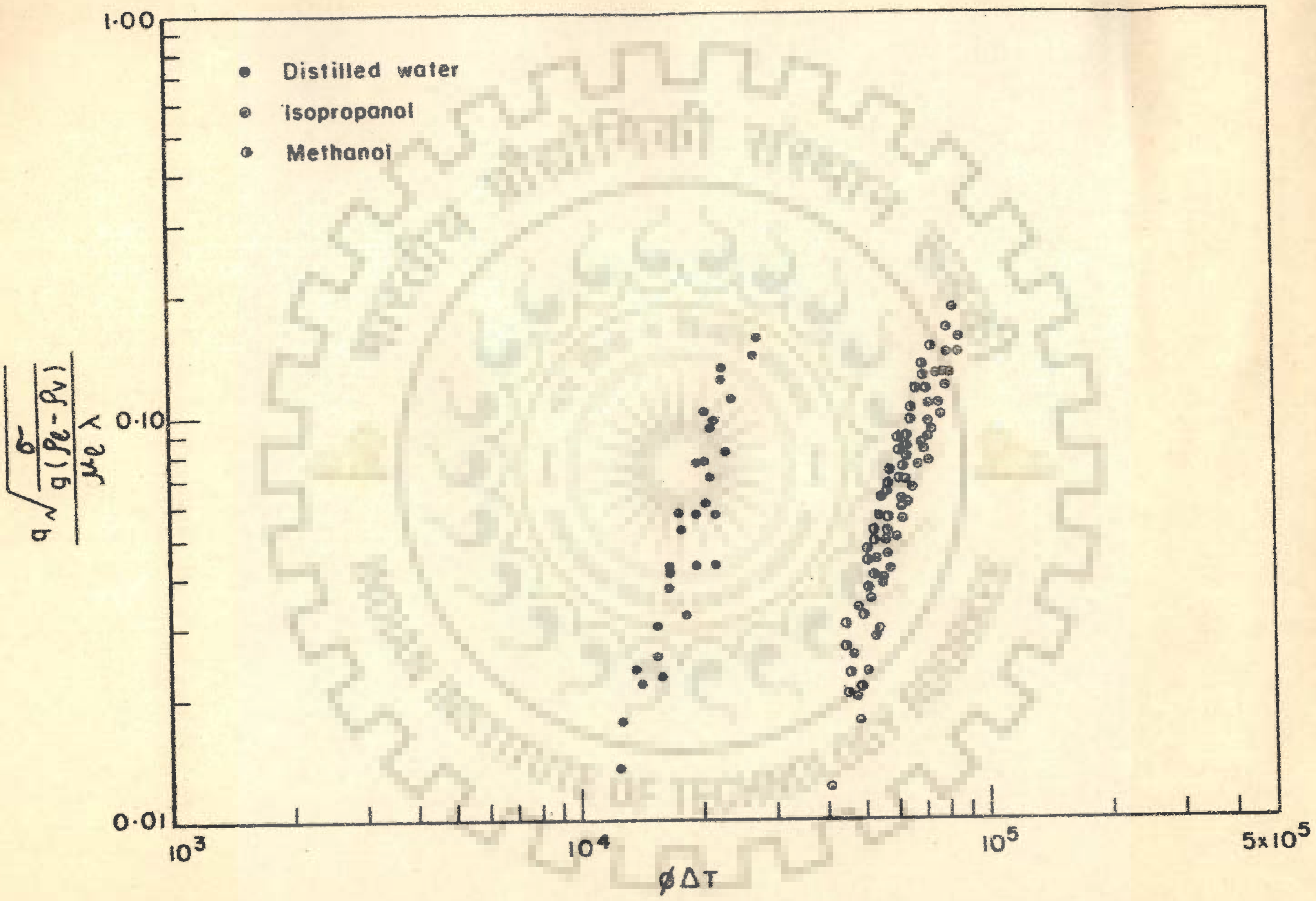

Fig. 6·37 Present experimental data compared with predictions from the Mikic and Rohsenow correlation [43]

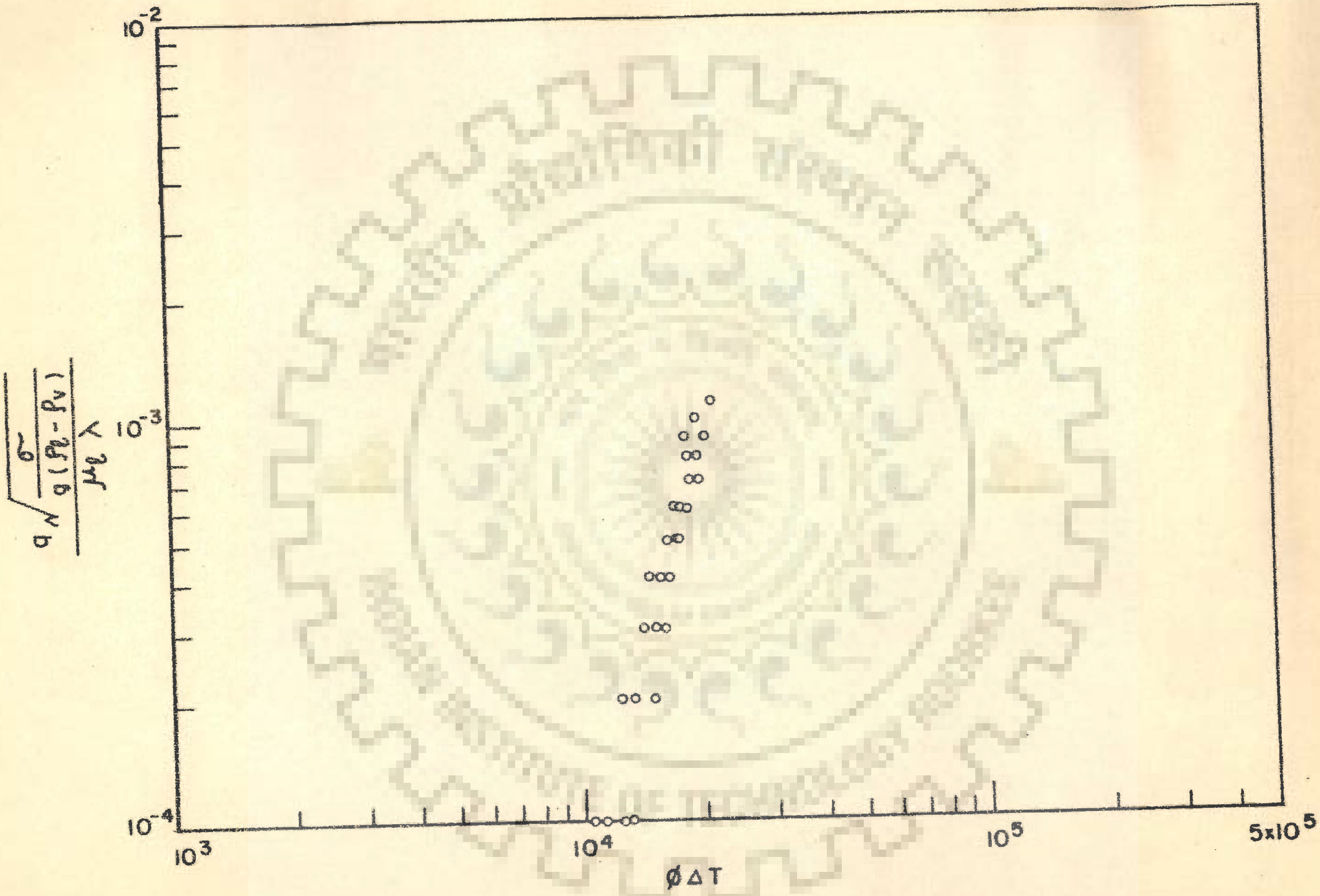

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Fig. 6·38 Present experimental data of ethanol compared with predictions from the Mikic and Rohsenow correlation [43]

# CHAPTER 7

## CONCLUSIONS AND RECOMMENDATIONS

The main conclusions drawn from the present study are as follows:

1. The present investigation has developed a new analytical model for heat transfer in nucleate pool boiling by using adequate equations for nucleation sites [53], bubble growth [66], bubble size [95] and bubble emission frequency through heating surface characteristics. The model is represented by equation, Eq. ( 5.22). This equation provides an useful procedure for the computation of absolute values of boiling heat transfer coefficient from the knowledge of wall heat flux, system pressure, and pertinent physico-thermal properties of boiling fluids through the known surface characteristics and known value of exponent a . The value of a can be determined if nf is known as recommended by Wiebe and Judd [54]. But for industrial surfaces it is highly improbable to obtain satisfactory generalised analytical equations for the surface characteristics as represented by constant M and for the values of nf . However, it is possible to have the constant M and exponent a determined empirically. For the surface characteristics used in the present investigation a procedure has been devised to determine the values of constant M as function of pressure and

the boiling fluids, and are represented by the set of equations, Eq. (6.3). The value of a is found to be independent of system parameters and has been evaluated as 0.3683. Finally, the model is represented by Eq. (6.4).

In order to obviate the unpredictable quantities built in constant M, and to provide a procedure for checking the consistency of experimental data from different heating surfaces used by different investigators, Eq.(5.22), has been transformed to Eq.(5.24). This equation incorporates the effects of wall heat flux, pertinent physico-thermal properties of boiling fluids and $M/M_1$ on $h^*/h_1^*$ . Further the values of $M/M_1$ are found to depend on system pressure only and they are independent of boiling fluids [cf. Eq.(6.5)]. The final form of Eq. (5.24) is represented by Eq. (6.6). The agreement between the experimental data of present investigation and the predicted values from the proposed analysis, Eq. (6.6) is excellent within ± 10 per cent. In addition, this equation is consistent with the experimental data of Cryder and Finalborgo [5], Raben, Beaubouef and Commerford [ 94 ] as well, which were obtained from the heating surfaces and fluids differing from each other and from that used in the present investigation. As the resultant equation, Eq. (6.6) of the present analysis is found to well-correlate the experimental data of different investigators for different conditions of heating surfaces and fluids, a considerable confidence is placed in this new analysis of heat transfer in nucleat pool boiling.

Another usefulness of this analysis is that it provides a design procedure for determining the absolute values of heat transfer coefficient at subatmospheric pressures, without resort to experimentation, from the knowledge of heat transfer coefficient at normal boiling point.

An excellent agreement between the predicted values of heat transfer coefficient and those of present investigation and of Cryder and Finalborgo [5], Raben , Beaubouef and Commerford [94] corroborates implicitly the validity of the governing equations: Eq. (5.3) for nucleation sites, Eqs (5.8 & 5.16) for bubble growth rate, and Eq. (5.9) for bubble size for atmospheric and subatmospheric pressures. Since these governing equations are for atmospheric and subatmospheric pressures, these correlation might not correlate the boiling data for higher pressure. Therefore, a care should be exercised that the resultant equation of the present analysis, Eq. (6.6) should not be used for the system pressures greater than one atmosphere.

2. As envisioned in the present investigation, a mathematical analysis consistent with the requirements of heat transfer dissipated from the heating surfaces to the boiling fluids would include the consideration of bubble dynamics on the heating surface through surface characteristics. Adequate governing equations for nucleate sites, bubble growth , bubble size and frequency should be selected in accordance with the system parameters like heat flux,

pressure and others.

3. The boiling heat transfer coefficient from the heating surface was found to vary circumferentially. It seems believable that such variation is inevitable for heating surfaces of large diameters.

4. The predictions of frequency from proposed equations, Eqs.(5.15 and 5.19) compare fairly well with the experimental values measured by Cole [45].

5. The Alad'ev equation for wall superheat was checked using experimental data . The equation was found to be consistent with the experimental data for boiling distilled water. However, the experimental data for isopropanol, ethanol and methanol could not be correlated by this equation. The constant of the Alad'ev equation was modified to fit the experimental data for the latter three liquids investigated. The modified constant was found to depend on the nature of boiling liquid to which heat was dissipated from the heating surface.

6. The average values of experimental heat transfer coefficient showed a good agreement with the existing experimental data.

7. New experimental data have been generated for subatmospheric pressures for the boiling of distilled water, isopropanol, ethanol and methanol.

As a result of the present investigation the following points emerge for future studies:

1. As indicated in the present study the proposed mathematical analysis is applicable for a limited range of parameters. In order to extend the range of applicability of the present analysis it is necessary to test it with the experimental data which are on the border lines of the tested parameters.

2. It seems that sufficient information does not exist for the prediction of wall superheat during nucleate pool boiling. Thus further research efforts are due for a generalised equation which can relate the wall superheat with the relevant parameters which are directly and easily measurable.

3. There is also a need to further the experimental programs to measure the wall temperature circumferentially at more number of points. These data would provide a more detailed thermal behaviour of the heating surfaces.

# APPENDIX A

## DETERMINATION OF AVERAGE WALL TEMPERATURE USING MECHANICAL QUADRATURE METHOD

With non-symmetrical boiling conditions around the circumference of a large diameter tube, immersed in a pool of boiling fluid, the circumferential variation in tube wall temperature might exist. In the present investigation, with tube diameter equal to 70 mm, the circumferential variation in wall temperature was observed. Numerical integration (mechanical quadrature) has been used for determining average wall temperature of heating surface. The problem of numerical integration has been solved by representing the integrand by an interpolation formula and then integrating it between the desired limits.

Let $T = f(\theta)$ denote a function which represents the values of temperature along the circumference. Let $T_0, T_1, T_2, \ldots\ldots, T_n$ be the values of the function T at equidistant values of $\theta_0$, $\theta_1$, $\theta_2$, ....., $\theta_n$ of the independent variable $\theta$.

From Newton's interpolation formula the following relationship is derived:

$$d\theta = h\,du$$

where, $u = \dfrac{\theta - \theta_0}{h}$ and h is the small interval $\Delta\theta$.

Integration of Newton's formula over n equidistant intervals of width $h (= \Delta\theta)$ leads to the following:

$$\int_{\theta_0}^{\theta_0+nh} T \, d\theta = h \int_0^n \left[ T_0 + u \, \Delta T_0 + \frac{u(u-1)}{2!} \Delta^2 T_0 + \cdots\cdots \right] du \qquad (A.1)$$

or

$$\int_{\theta_0}^{\theta_0+nh} T d\theta = h \left[ nT_0 + \frac{n^2}{2} \Delta T_0 + \left( \frac{n^3}{3} - \frac{n^2}{2} \right) \frac{\Delta^2 T_0}{2} + \cdots \right] \qquad (A.2)$$

Equation (A.2) is the general form of quadrature formula for equidistant ordinates.

In the present investigation, three thermocouples, evenly stationed, at a regular interval of $\pi/2$, were employed to record the circumferential variation in the heating surface temperature. Figure 3.7 shows locations of these thermocouples at the side, at the top and at the bottom of the heating surface. The temperature at a' was not measured since it is equal to temperature at , a, due to the symmetry of boiling conditions about the horizontal mid-plane a-a' . Therefore , with available three values two step lengths have been considered. Putting $n = 2$ , and neglecting all differences above the second (there can be no difference higher than the second, with the available three values), equation(A-2)assumes the form :

$$\int_{\theta_0}^{\theta_0+2h} T d\theta = h \left[ 2T_0 + 2 \cdot \Delta T_0 + \left( \frac{8}{3} - 2 \right) \cdot \frac{\Delta^2 T_0}{2} \right] \qquad (A.3)$$

$$\left.\begin{aligned} \text{Where}\quad \Delta T_o &= T_{w1} - T_{wo} \\ \Delta^2 T_o &= \Delta T_{w1} - \Delta T_{wo} \\ &= T_{w2} - 2T_{w1} + T_{w0} \end{aligned}\right\} \qquad (A.4)$$

Inserting the values of $\Delta T_o$ and $\Delta^2 T_o$ from (A.4) in Eq(A.3) the following is obtained:

$$\int_{\theta_o}^{\theta_o+2h} T d\theta = \frac{h}{3}\left[ T_{wo} + 4T_{w1} + T_{w2} \right] \qquad (A.5)$$

The average of the circumferential temperature is defined by the following equation:

$$\bar{T}_w = \frac{1}{2\pi} \int_0^{2\pi} T\, d\theta$$

$$= \frac{1}{2\pi}\left[ \int_0^{\pi} T\, d\theta + \int_{\pi}^{2\pi} T\, d\theta \right] \qquad (A.6)$$

From Eqs. (A.5) and (A.6) and h ( = $\pi/2$) the average value of $T_w$ is obtained as follows:

$$\bar{T}_w = \frac{1}{2\pi}\left[ \frac{\pi}{2}\ \frac{1}{3}\,(T_{wo} + 4T_{w1} + T_{w2}) + \frac{\pi}{2}\,.\ \frac{1}{3}(T_{w2} + 4T_{w3} + T_{wo}) \right]$$

$$= \frac{1}{12}\left[ 2\,T_{wo} + 2T_{w2} + 4\,T_{w1} + 4\,T_{w3} \right] \qquad (A.7)$$

By symmetry $T_{w2} = T_{wo}$ and using this in Eq. (A.7) the following equation results:

$$\bar{T}_w = \frac{1}{3} \left[ T_{wo} + T_{w1} + T_{w3} \right] \qquad (A.8)$$

Equation (A.8) shows that arithmatic mean of the thermocouple readings represents the average wall temperature.

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# APPENDIX B

## TABULATION OF EXPERIMENTAL DATA

| | |
|---|---|
| Table B-1 | Experimental Data of Heat Transfer to Boiling Distilled Water |
| Table B-2 | Experimental data of heat Transfer to Boiling Isopropanol |
| Table B-3 | Experimental Data of Heat Transfer to Boiling Ethanol |
| Table B-4 | Experimental Data of Heat Transfer to Boiling Methanol |

Nomenclature

P = System Pressure

TS = Saturation Temperature

TW = Surface Temperature

TL = Liquid Temperature

CTW = Corrected Surface Temperature

DTW = Corrected Temperature Difference

AVGTW = Average gorrected Surface Temperature

AVGDTW = Average Temperature Difference

H = Heat Transfer Coefficient

AVGH = Average Heat Transfer Coefficient

Scripts 0, 1 and 3 represent the side, top and bottom positions respectively, as shown in Figure 3.7.

```
  TABLE B-1 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING DISTILLED WATER
=====================================================================================
RUN      HEAT          TEMPERATURE          CORRECTED     CORRECTED    TRANSFER
NO.      FLUX      SURFACE       LIQUID        TW            DTW      COEFFICIENT
       W/SQ.M      DEG.C         DEG.C        DEG.C         DEG.C      W/SQ.M K
=====================================================================================

                 P=98.44 KN/SQ.N                       TS=99.00 DEG.C

  1    6870.22 TWO 103.900 TLO  99.200 CTWO 103.350 DTWO 4.150 HO 1655.17
               TW1 102.725 TL1  99.625 CTW1 102.175 DTW1 2.550 H1 2693.40
               TW3 105.575 TL3 100.050 CTW3 105.025 DTW3 4.975 H3 1380.95
            AVGTW=103.512              AVGDTW=3.892             AVGH=1765.37

  2   11755.72 TWO 105.600 TLO  99.350 CTWO 104.660 DTWO 5.310 HO 2213.80
               TW1 103.875 TL1  99.750 CTW1 102.935 DTW1 3.185 H1 3690.74
               TW3 107.375 TL3 100.400 CTW3 106.435 DTW3 6.035 H3 1947.92
            AVGTW=104.677              AVGDTW=   4.843          AVGH=2427.19

  3   16539.44 TWO 106.180 TLO  99.425 CTWO 104.000 DTWO 5.432 HO 3044.39
               TW1 104.700 TL1  99.800 CTW1 103.377 DTW1 3.577 H1 4622.85
               TW3 108.100 TL3 100.500 CTW3 106.777 DTW3 6.277 H3 2634.93
            AVGTW=104.718              AVGDTW=   5.095          AVGH=3245.99

  4   22137.40 TWO 107.050 TLO  99.625 CTWO 105.280 DTWO 5.655 HO 3914.50
               TW1 105.975 TL1  99.800 CTW1 104.205 DTW1 4.405 H1 5025.25
               TW3 109.600 TL3 100.650 CTW3 107.830 DTW3 7.180 H3 3083.20
            AVGTW=105.772              AVGDTW=   5.746          AVGH=3852.21

  5   29414.75 TWO 107.650 TLO  99.725 CTWO 105.298 DTWO 5.573 HO 5277.66
               TW1 107.450 TL1 100.150 CTW1 105.098 DTW1 4.948 H1 5944.24
               TW3 110.200 TL3 101.000 CTW3 107.848 DTW3 6.848 H3 4295.37
            AVGTW=106.081              AVGDTW=   5.789          AVGH=5081.45

  6   37709.92 TWO 109.000 TLO 100.050 CTWO 105.985 DTWO 5.935 HO 6353.52
               TW1 109.225 TL1 100.500 CTW1 106.210 DTW1 5.710 H1 6603.86
               TW3 111.950 TL3 101.250 CTW3 108.935 DTW3 7.685 H3 4906.95
            AVGTW=107.043              AVGDTW=   6.443          AVGH=5852.55

  7   41730.27 TWO 109.325 TLO 100.050 CTWO 105.988 DTWO 5.938 HO 7026.62
               TW1 109.425 TL1 100.500 CTW1 106.088 DTW1 5.588 H1 7466.66
               TW3 113.100 TL3 101.900 CTW3 109.763 DTW3 7.863 H3 5307.17
            AVGTW=107.279              AVGDTW=   6.463          AVGH=6456.79
```

```
  TABLE B-1 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING DISTILLED WATER
==============================================================================
 RUN      HEAT          TEMPERATURE              CORRECTED     CORRECTED    TRANSFER
 NO.      FLUX      SURFACE          LIQUID          TW            DTW      COEFFICIENT
         W/SQ.M      DEG.C           DEG.C         DEG.C          DEG.C     W/SQ.M K
==============================================================================

                   P=66.65 KN/SQ.M                            TS=88.50 DEG.C

   8     6870.23 TWO   93.350 TLO   88.250 CTWO   92.800 DTWO 4.550 HO 1509.69
                 TW1   91.670 TL1   88.250 CTW1   91.120 DTW1 2.870 H1 2393.17
                 TW3   96.250 TL3   89.250 CTW3   95.700 DTW3 6.450 H3 1065.03
              AVGTW=   93.207             AVGDTW=    4.624            AVGH=1485.75

   9    11752.72 TWO   95.150 TLO   88.250 CTWO   94.210 DTWO 5.960 HO 1972.37
                 TW1   92.700 TL1   88.250 CTW1   91.760 DTW1 3.510 H1 3349.03
                 TW3   97.750 TL3   89.250 CTW3   96.810 DTW3 7.560 H3 1554.95
              AVGTW=   94.260             AVGDTW=    5.676            AVGH=2070.82

  10    16539.44 TWO   95.750 TLO   88.250 CTWO   94.427 DTWO 6.177 HO 2677.26
                 TW1   93.650 TL1   88.250 CTW1   92.327 DTW1 4.077 H1 4056.02
                 TW3   98.000 TL3   89.550 CTW3   96.677 DTW3 7.127 H3 2320.43
              AVGTW=   94.477             AVGDTW=    5.794            AVGH=2854.37

  11    22137.40 TWO   96.450 TLO   88.250 CTWO   94.680 DTWO 6.430 HO 3442.71
                 TW1   94.950 TL1   88.250 CTW1   93.180 DTW1 4.930 H1 4490.14
                 TW3  100.250 TL3   89.550 CTW3   98.480 DTW3 8.930 H3 2478.93
              AVGTW=   95.446             AVGDTW=    6.763            AVGH=3273.04

  12    29414.75 TWO   96.450 TLO   88.250 CTWO   94.098 DTWO 5.848 HO 5029.51
                 TW1   96.100 TL1   88.250 CTW1   93.748 DTW1 5.498 H1 5349.66
                 TW3  100.750 TL3   89.550 CTW3   98.398 DTW3 8.848 H3 3324.29
              AVGTW=   95.415             AVGDTW=    6.731            AVGH=4369.54

  13    37709.92 TWO   97.900 TLO   88.250 CTWO   94.885 DTWO 6.635 HO 5683.24
                 TW1   97.650 TL1   88.250 CTW1   94.635 DTW1 6.385 H1 5905.76
                 TW3  102.350 TL3   89.550 CTW3   99.335 DTW3 9.785 H3 3853.74
              AVGTW=   96.285             AVGDTW=    7.601            AVGH=4960.56
```

```
   TABLE B-1 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING DISTILLED WATER
=========================================================================
 RUN     HEAT           TEMPERATURE            CORRECTED    CORRECTED    TRANSFER
 NO.     FLUX      SURFACE       LIQUID           TW           DTW      COEFFICIENT
        W/SQ.M      DEG.C        DEG.C          DEG.C        DEG.C       W/SQ.M K
=========================================================================

                   P=47.32 KN/SQ.M                         TS=80.00 DEG.C

  14     6870.22 TWO   85.400 TLO   79.900 CTWO   84.850 DTWO 4.950 HO 1387.71
                 TW1   83.550 TL1   79.900 CTW1   83.000 DTW1 3.100 H1 2215.66
                 TW3   88.400 TL3   81.050 CTW3   87.850 DTW3 6.800 H3 1010.21
              AVGTW=   85.234             AVGDTW=    4.950          AVGH=1387.71

  15    11755.72 TWO   87.250 TLO   79.900 CTWO   86.310 DTWO 6.410 HO 1833.91
                 TW1   84.500 TL1   79.900 CTW1   83.560 DTW1 3.660 H1 3211.78
                 TW3   89.550 TL3   81.050 CTW3   88.610 DTW3 7.560 H3 1554.95
              AVGTW=   86.160             AVGDTW=    5.876          AVGH=2000.34

  16    16539.44 TWO   87.900 TLO   79.900 CTWO   86.577 DTWO 6.677 HO 2476.80
                 TW1   85.550 TL1   79.900 CTW1   84.227 DTW1 4.327 H1 3821.71
                 TW3   90.500 TL3   81.450 CTW3   89.177 DTW3 7.727 H3 2140.26
              AVGTW=   86.661             AVGDTW=    6.244          AVGH=2648.67

  17    22137.40 TWO   88.700 TLO   79.900 CTWO   86.930 DTWO 7.030 HO 3148.89
                 TW1   86.900 TL1   79.900 CTW1   85.130 DTW1 5.230 H1 4232.59
                 TW3   93.150 TL3   81.450 CTW3   91.380 DTW3 9.930 H3 2229.29
              AVGTW=   87.813             AVGDTW=    7.396          AVGH=2992.80

  18    29414.76 TWO   89.200 TLO   79.900 CTWO   86.848 DTWO 6.948 HO 4233.29
                 TW1   88.000 TL1   79.900 CTW1   85.648 DTW1 5.748 H1 5117.00
                 TW3   93.250 TL3   81.450 CTW3   90.898 DTW3 9.448 H3 3113.19
              AVGTW=   87.798             AVGDTW=    7.381          AVGH=3984.78

  19    37709.92 TWO   90.000 TLO   79.900 CTWO   86.985 DTWO 7.085 HO 5322.29
                 TW1   89.400 TL1   79.900 CTW1   86.385 DTW1 6.485 H1 5814.69
                 TW3   94.575 TL3   81.450 CTW3   91.560 DTW310.110 H3 3729.86
              AVGTW=   88.310             AVGDTW=    7.893          AVGH=4777.27
```

```
  TABLE B-1 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING DISTILLED WATER
=============================================================================
RUN     HEAT          TEMPERATURE              CORRECTED     CORRECTED    TRANSFER
NO.     FLUX      SURFACE       LIQUID            TW            DTW      COEFFICIENT
       W/SQ.M     DEG.C         DEG.C           DEG.C          DEG.C      W/SQ.M K
=============================================================================

              P=25.33 KN/SQ.M                          TS=65.30 DEG.C

 20   6870.23  TWO   71.200 TL0  65.100 CTWO  70.650 DTWO   5.550 HO 1237.71
               TW1   69.025 TL1  65.100 CTW1  68.475 DTW1   3.375 H1 2035.17
               TW3   75.550 TL3  66.000 CTW3  75.000 DTW3   9.000 H3  763.36
            AVGTW=   71.375              AVGDTW=   5.975         AVGH=1149.83

 21  11755.72  TWO   72.850 TL0  65.100 CTWO  71.910 DTWO   6.810 HO 1726.20
               TW1   70.350 TL1  65.100 CTW1  69.410 DTW1   4.310 H1 2727.43
               TW3   77.030 TL3  66.000 CTW3  76.090 DTW3   10.09 H3 1165.08
            AVGTW=   72.470              AVGDTW=   7.070         AVGH=1662.80

 22  16539.44  TWO   73.800 TL0  65.100 CTWO  72.477 DTWO   7.377 HO 2241.80
               TW1   71.475 TL1  65.100 CTW1  70.152 DTW1   5.052 H1 3273.35
               TW3   78.150 TL3  66.700 CTW3  76.827 DTW3  10.127 H3 1633.20
            AVGTW=   73.152              AVGDTW=   7.512         AVGH=2199.78

 23  22137.40  TWO   74.700 TL0  65.100 CTWO  72.930 DTWO   7.830 HO 2827.17
               TW1   73.000 TL1  65.100 CTW1  71.230 DTW1   6.130 H1 3611.19
               TW3   79.125 TL3  66.700 CTW3  77.355 DTW3  10.655 H3 2077.61
            AVGTW=   73.838              AVGDTW=   8.205         AVGH=2697.96

 24  29414.76  TWO   75.500 TL0  65.100 CTWO  73.148 DTWO   8.048 HO 3654.72
               TW1   74.000 TL1  65.100 CTW1  71.648 DTW1   6.548 H1 4491.87
               TW3   81.200 TL3  66.700 CTW3  78.848 DTW3  12.148 H3 2421.36
            AVGTW=   74.548              AVGDTW=   8.915         AVGH=3299.59

 25  37709.92  TWO   76.500 TL0  65.100 CTWO  73.485 DTWO   8.385 HO 4497.16
               TW1   75.750 TL1  65.100 CTW1  72.735 DTW1   7.635 H1 4938.90
               TW3   82.675 TL3  66.700 CTW3  79.660 DTW3  12.960 H3 2909.71
            AVGTW=   75.293              AVGDTW=   9.660         AVGH=3903.72
```

```
 TABLE B-1 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING DISTILLED WATER
==============================================================================
RUN      HEAT             TEMPERATURE            CORRECTED    CORRECTED   TRANSFER
NO.      FLUX        SURFACE          LIQUID        TW           DTW     COEFFICIENT
        W/SQ.M       DEG.C            DEG.C        DEG.C        DEG.C     W/SQ.M K
==============================================================================

                    P=11.33 KN/SQ.M                          TS=48.50 DEG.C

  26    6870.23 TWO   54.153 TLO   48.300 CTWO   53.603 DTWO   5.303 HO 1295.54
                TW1   56.483 TL1   48.300 CTW1   55.933 DTW1   7.633 H1  900.07
                TW3   59.753 TL3   49.400 CTW3   59.503 DTW3   9.803 H3  700.83
             AVGTW=   56.346             AVGDTW=   7.580            AVGH=  906.26

  27   11755.73 TWO   56.337 TLO   48.300 CTWO   55.397 DTWO   7.097 HO 1656.44
                TW1   58.588 TL1   48.300 CTW1   57.649 DTW1   9.349 H1 1257.43
                TW3   62.108 TL3   49.400 CTW3   61.168 DTW3  11.768 H3  998.96
             AVGTW=   58.071             AVGDTW=   9.405            AVGH=1249.97

  28   16539.44 TWO   56.947 TLO   48.300 CTWO   55.624 DTWO   7.324 HO 2258.02
                TW1   59.448 TL1   48.300 CTW1   58.125 DTW1   9.825 H1 1683.27
                TW3   63.446 TL3   49.600 CTW3   62.123 DTW3  12.523 H3 1320.65
             AVGTW=   58.624             AVGDTW=   9.891            AVGH=1672.10

  29   22137.40 TWO   59.400 TLO   48.300 CTWO   57.630 DTWO   9.330 HO 2372.65
                TW1   61.321 TL1   48.300 CTW1   59.551 DTW1  11.251 H1 1967.55
                TW3   65.670 TL3   49.600 CTW3   63.900 DTW3  14.300 H3 1548.05
             AVGTW=   60.360             AVGDTW=  11.627            AVGH=1903.93

  30   29414.76 TWO   61.150 TLO   48.300 CTWO   58.798 DTWO  10.498 HO 2801.82
                TW1   62.751 TL1   48.300 CTW1   60.399 DTW1  12.099 H1 2431.08
                TW3   64.472 TL3   49.602 CTW3   62.120 DTW3  12.518 H3 2349.71
             AVGTW=   60.439             AVGDTW=  11.705            AVGH=2512.91
```

```
 TAELE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL
==========================================================================
RUN     HEAT           TEMPERATURE              CORRECTED    CORRECTED    TRANSFER
NO.     FLUX       SURFACE        LIQUID           TW           DTW      COEFFICIENT
      W/SQ.M       DEG.C          DEG.C          DEG.C        DEG.C       W/SQ.M K
==========================================================================

                P=97.98 KN/SQ.M                           TS=81.60 DEG.C

 31     6870.23 TWO   89.600 TL0   81.800 CTW0   89.050 DTW0   7.250 H0   947.52
                TW1   90.650 TL1   81.800 CTW1   90.100 DTW1   8.300 H1   827.66
                TW3   89.550 TL3   82.250 CTW3   89.000 DTW3   6.750 H3  1017.70
             AVGTW=   89.384                 AVGDTW=   7.434          AVGH=  924.15

 32    11755.73 TWO   91.075 TL0   81.800 CTW0   90.135 DTW0   8.335 H0  1410.37
                TW1   91.850 TL1   81.800 CTW1   90.910 DTW1   9.110 H1  1290.39
                TW3   91.400 TL3   82.700 CTW3   90.460 DTW3   7.760 H3  1514.88
             AVGTW=   90.501                 AVGDTW=   8.401          AVGH=1399.18

 33    16539.44 TWO   92.925 TL0   81.800 CTW0   91.602 DTW0   9.802 H0  1687.22
                TW1   93.675 TL1   81.800 CTW1   92.352 DTW1  10.552 H1  1567.31
                TW3   93.300 TL3   82.700 CTW3   91.977 DTW3   9.277 H3  1782.70
             AVGTW=   91.977                 AVGDTW=   9.877          AVGH=1674.41

 34    22137.40 TWO   94.225 TL0   81.800 CTW0   92.455 DTW0  10.655 H0  2077.61
                TW1   95.375 TL1   81.800 CTW1   93.605 DTW1  11.805 H1  1875.22
                TW3   95.250 TL3   82.700 CTW3   93.480 DTW3  10.780 H3  2053.52
             AVGTW=   93.180                 AVGDTW=  11.080          AVGH=1997.92

 35    26055.98 TWO   94.925 TL0   81.800 CTW0   92.841 DTW0  11.041 H0  2359.72
                TW1   95.975 TL1   81.800 CTW1   93.891 DTW1  12.091 H1  2154.82
                TW3   96.425 TL3   82.700 CTW3   94.341 DTW3  11.641 H3  2238.11
             AVGTW=   93.691                 AVGDTW=  11.591          AVGH=2247.76

 36    33893.13 TWO   96.100 TL0   81.800 CTW0   93.390 DTW0  11.590 H0  2924.24
                TW1   96.700 TL1   81.800 CTW1   93.990 DTW1  12.190 H1  2780.31
                TW3   99.050 TL3   82.700 CTW3   96.340 DTW3  13.640 H3  2484.76
             AVGTW=   94.573                 AVGDTW=  12.473          AVGH=2717.16

 37    37709.92 TWO   96.900 TL0   81.800 CTW0   93.885 DTW0  12.085 H0  3120.32
                TW1   97.250 TL1   81.800 CTW1   94.235 DTW1  12.435 H1  3032.49
                TW3  100.150 TL3   82.700 CTW3   97.135 DTW3  14.435 H3  2612.34
             AVGTW=   95.085                 AVGDTW=  12.985          AVGH=2904.05
```

```
TABLE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL
==========================================================================
RUN      HEAT          TEMPERATURE              CORRECTED    CORRECTED    TRANSFER
NO.      FLUX      SURFACE         LIQUID           TW          DTW       COEFFICIENT
        W/SQ.M      DEG.C          DEG.C          DEG.C        DEG.C      W/SQ.M K
==========================================================================

                p=80.77 KN/SQ.M                           TS=77.00 DEG.C

 38    11755.73 TWO   87.875 TL0   77.200 CTWO   86.935 DTWO   9.735 HO 1207.55
                TW1   88.575 TL1   77.200 CTW1   87.635 DTW1  10.435 H1 1126.55
                TW3   87.550 TL3   77.900 CTW3   86.610 DTW3   8.710 H3 1349.65
             AVGTW=   87.060                AVGDTW=    9.626             AVGH=1221.14

 39    16539.44 TWO   89.700 TL0   77.200 CTWO   88.377 DTWO  11.177 HO 1479.67
                TW1   90.150 TL1   77.200 CTW1   88.827 DTW1  11.627 H1 1422.41
                TW3   89.000 TL3   77.900 CTW3   87.677 DTW3   9.777 H3 1691.54
             AVGTW=   88.294                AVGDTW=   10.861             AVGH=1522.82

 40    22137.40 TWO   91.300 TL0   77.200 CTWO   89.530 DTWO  12.330 HO 1795.38
                TW1   91.850 TL1   77.200 CTW1   90.080 DTW1  12.880 H1 1718.71
                TW3   91.575 TL3   77.900 CTW3   89.805 DTW3  11.905 H3 1859.47
             AVGTW=   89.805                AVGDTW=   12.371             AVGH=1789.33

 41    26055.98 TWO   92.275 TL0   77.200 CTWO   90.191 DTWO  12.991 HO 2005.55
                TW1   92.500 TL1   77.200 CTW1   90.416 DTW1  13.216 H1 1971.41
                TW3   93.200 TL3   77.900 CTW3   91.116 DTW3  13.216 H3 1971.41
             AVGTW=   90.575                AVGDTW=   13.141             AVGH=1982.66

 42    33893.13 TWO   93.350 TL0   77.200 CTWO   90.640 DTWO  13.440 HO 2521.73
                TW1   92.500 TL1   77.200 CTW1   89.790 DTW1  12.590 H1 2691.98
                TW3   95.900 TL3   77.900 CTW3   93.190 DTW3  15.290 H3 2216.63
             AVGTW=   91.207                AVGDTW=   13.773             AVGH=2460.70

 43    37709.92 TWO   94.350 TL0   77.200 CTWO   91.335 DTWO  14.135 HO 2667.79
                TW1   93.275 TL1   77.200 CTW1   90.260 DTW1  13.060 H1 2887.37
                TW3   97.425 TL3   77.900 CTW3   94.410 DTW3  16.510 H3 2284.03
             AVGTW=   92.001                AVGDTW=   14.568             AVGH=2588.44
```

# TABLE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL

```
==========================================================================
RUN     HEAT            TEMPERATURE            CORRECTED    CORRECTED    TRANSFER
NO.     FLUX        SURFACE       LIQUID          TW           DTW       COEFFICIENT
       W/SQ.M       DEG.C         DEG.C         DEG.C         DEG.C      W/SQ.M K
==========================================================================

               P=67.28 KN/SQ.M                          TS=72.90 DEG.C

 44     6870.23 TWO   82.025 TLO   73.050 CTWO   81.475 DTWO   8.425 HO   815.38
                TW1   82.500 TL1   73.050 CTW1   81.950 DTW1   8.900 H1   771.87
                TW3   82.500 TL3   73.775 CTW3   81.950 DTW3   8.175 H3   840.32
            AVGTW=  81.792              AVGDTW=    8.500               AVGH=  808.19

 45    11755.73 TWO   84.100 TLO   73.050 CTWO   83.160 DTWO  10.110 HO  1162.76
                TW1   84.700 TL1   73.050 CTW1   83.760 DTW1  10.710 H1  1097.62
                TW3   83.800 TL3   73.775 CTW3   82.860 DTW3   9.085 H3  1293.94
            AVGTW=  83.260              AVGDTW=    9.968               AVGH=1179.28

 46    16539.44 TWO   86.025 TLO   73.050 CTWO   84.702 DTWO  11.652 HO  1419.36
                TW1   86.550 TL1   73.050 CTW1   85.227 DTW1  12.177 H1  1358.17
                TW3   85.700 TL3   73.775 CTW3   84.377 DTW3  10.602 H3  1559.92
            AVGTW=  84.769              AVGDTW=   11.477               AVGH=1441.00

 47    22137.40 TWO   88.000 TLO   73.050 CTWO   86.230 DTWO  13.180 HO  1679.59
                TW1   88.250 TL1   73.050 CTW1   86.480 DTW1  13.430 H1  1648.33
                TW3   87.900 TL3   73.775 CTW3   86.130 DTW3  12.355 H3  1791.74
            AVGTW=  86.280              AVGDTW=   12.988               AVGH=1704.38

 48    26055.98 TWO   88.675 TLO   73.050 CTWO   86.591 DTWO  13.541 HO  1924.09
                TW1   88.575 TL1   73.050 CTW1   86.491 DTW1  13.441 H1  1938.41
                TW3   89.200 TL3   73.775 CTW3   87.116 DTW3  13.341 H3  1952.94
            AVGTW=  86.733              AVGDTW=   13.441               AVGH=1938.41

 49    33893.13 TWO   90.100 TLO   73.050 CTWO   87.390 DTWO  14.340 HO  2363.47
                TW1   89.000 TL1   73.050 CTW1   86.290 DTW1  13.240 H1  2559.82
                TW3   92.850 TL3   73.950 CTW3   90.140 DTW3  16.190 H3  2093.41
            AVGTW=  87.940              AVGDTW=   14.590               AVGH=2322.97

 50    37709.92 TWO   91.100 TLO   73.050 CTWO   88.085 DTWO  15.035 HO  2508.10
                TW1   89.925 TL1   73.050 CTW1   86.910 DTW1  13.860 H1  2720.72
                TW3   96.300 TL3   73.950 CTW3   93.285 DTW3  19.335 H3  1950.32
            AVGTW=  89.426              AVGDTW=   16.076               AVGH=2345.59
```

```
 TABLE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL
==================================================================================
RUN      HEAT          TEMPERATURE              CORRECTED     CORRECTED    TRANSFER
NO.      FLUX       SURFACE         LIQUID         TW            DTW       COEFFICIENT
        W/SQ.M      DEG.C           DEG.C        DEG.C         DEG.C       W/SQ.M K
==================================================================================

                 P=54.65 KN/SQ.M                            TS=68.25 DEG.C

51     6870.22 TWO   79.500 TLO   68.500 CTWO   78.950 DTWO 10.450 HO   657.39
               TW1   78.800 TL1   68.500 CTW1   78.250 DTW1  9.750 H1   704.58
               TW3   79.400 TL3   69.650 CTW3   78.850 DTW3  9.200 H3   746.70
            AVGTW=  78.684              AVGDTW=  9.800             AVGH=  700.99

52    11755.73 TWO   80.775 TLO   68.500 CTWO   79.835 DTWO 11.335 HO  1037.10
               TW1   81.056 TL1   68.500 CTW1   80.116 DTW1 11.616 H1  1012.01
               TW3   80.450 TL3   69.650 CTW3   79.510 DTW3  9.860 H3  1192.24
            AVGTW=  79.820              AVGDTW= 10.937             AVGH=1074.84

53    16539.44 TWO   82.700 TLO   68.500 CTWO   81.377 DTWO 12.877 HO  1284.34
               TW1   82.600 TL1   68.500 CTW1   81.277 DTW1 12.777 H1  1294.39
               TW3   81.900 TL3   69.450 CTW3   80.577 DTW3 11.127 H3  1486.32
            AVGTW=  81.077              AVGDTW= 12.261             AVGH=1348.94

54    22137.40 TWO   84.900 TLO   68.500 CTWO   83.130 DTWO 14.630 HO  1513.13
               TW1   84.025 TL1   68.500 CTW1   82.255 DTW1 13.755 H1  1609.38
               TW3   84.100 TL3   69.450 CTW3   82.230 DTW3 12.880 H3  1718.71
            AVGTW=  82.571              AVGDTW= 13.755             AVGH=1609.38

55    26055.98 TWO   85.450 TLO   68.500 CTWO   83.366 DTWO 14.866 HO  1752.61
               TW1   84.350 TL1   68.500 CTW1   82.266 DTW1 13.766 H1  1892.65
               TW3   85.575 TL3   69.450 CTW3   83.491 DTW3 14.041 H3  1855.58
            AVGTW=  83.041              AVGDTW= 14.225             AVGH=1831.67

56    33893.13 TWO   87.400 TLO   68.500 CTWO   84.690 DTWO 16.190 HO  2093.41
               TW1   85.625 TL1   68.500 CTW1   82.915 DTW1 14.415 H1  2351.17
               TW3   89.350 TL3   69.550 CTW3   86.640 DTW3 16.990 H3  1994.84
            AVGTW=  84.748              AVGDTW= 15.865             AVGH=2136.29

57    37709.92 TWO   88.025 TLO   68.500 CTWO   85.010 DTWO 16.510 HO  2284.03
               TW1   85.575 TL1   68.500 CTW1   82.560 DTW1 14.060 H1  2682.02
               TW3   91.000 TL3   69.450 CTW3   87.985 DTW3 18.535 H3  2034.49
            AVGTW=  85.185              AVGDTW= 16.368             AVGH=2303.79
```

## TABLE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL

```
==================================================================================
RUN     HEAT          TEMPERATURE              CORRECTED      CORRECTED    TRANSFER
NO.     FLUX      SURFACE        LIQUID           TW            DTW       COEFFICIENT
      W/SQ.M      DEG.C          DEG.C          DEG.C          DEG.C       W/SQ.M K
==================================================================================

              P=41.32 KN/SQ.M                            TS=62.40 DEG.C

58    6870.23 TWO   74.200 TLO  62.850 CTWO   73.650 DTWO 10.800 HO   636.09
              TW1   73.725 TL1  62.850 CTW1   73.175 DTW1 10.325 H1   665.35
              TW3   74.860 TL3  63.850 CTW3   74.310 DTW3 10.460 H3   656.76
           AVGTW=   73.712             AVGDTW=  10.529               AVGH= 652.50

59   11755.73 TWO   75.575 TLO  62.850 CTWO   74.635 DTWO 11.785 HO   997.50
              TW1   75.500 TL1  62.850 CTW1   74.560 DTW1 11.710 H1  1003.89
              TW3   76.250 TL3  63.850 CTW3   75.310 DTW3 11.460 H3  1025.79
           AVGTW=   74.835             AVGDTW=  11.651               AVGH=1008.91

60   16559.44 TWO   77.850 TLO  62.850 CTWO   76.526 DTWO 13.676 HO  1210.83
              TW1   77.850 TL1  62.850 CTW1   76.526 DTW1 13.676 H1  1210.83
              TW3   78.075 TL3  63.850 CTW3   76.751 DTW3 12.901 H3  1283.56
           AVGTW=   76.601             AVGDTW=  13.417               AVGH=1234.14

61   22137.40 TWO   79.925 TLO  62.850 CTWO   78.155 DTWO 15.305 HO  1446.40
              TW1   79.350 TL1  62.850 CTW1   77.580 DTW1 14.730 H1  1502.86
              TW3   79.450 TL3  63.850 CTW3   77.680 DTW3 13.830 H3  1600.65
           AVGTW=   77.805             AVGDTW=  14.621               AVGH=1513.99

62   26055.98 TWO   81.850 TLO  62.850 CTWO   79.766 DTWO 16.916 HO  1540.23
              TW1   80.325 TL1  62.850 CTW1   78.241 DTW1 15.391 H1  1692.83
              TW3   82.150 TL3  63.850 CTW3   80.066 DTW3 16.216 H3  1606.71
           AVGTW=   79.358             AVGDTW=  16.175               AVGH=1610.85

63   33893.13 TWO   83.350 TLO  62.850 CTWO   80.640 DTWO 17.790 HO  1905.13
              TW1   80.950 TL1  62.850 CTW1   78.240 DTW1 15.390 H1  2202.22
              TW3   85.850 TL3  63.850 CTW3   83.140 DTW3 19.290 H3  1756.99
           AVGTW=   80.673             AVGDTW=  17.490               AVGH=1937.81
```

```
TABLE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL
=============================================================================
RUN     HEAT          TEMPERATURE         CORRECTED     CORRECTED    TRANSFER
NO.     FLUX      SURFACE      LIQUID        TW            DTW       COEFFICIENT
       W/SQ.M     DEG.C        DEG.C        DEG.C         DEG.C       W/SQ.M K
=============================================================================

              P=29.33 KN/SQ.M                       TS=55.20 DEG.C

64    6870.23 TWO  69.450 TLO   55.750 CTWO   68.900 DTWO 13.150 HO   522.42
              TW1  67.000 TL1   55.750 CTW1   66.450 DTW1 10.700 H1   642.03
              TW3  70.800 TL3   56.500 CTW3   70.250 DTW3 13.750 H3   499.63
           AVGTW=  68.534               AVGDTW=  12.534              AVGH=  548.12

65   11755.73 TWO  71.000 TLO   55.750 CTWO   70.060 DTWO 14.310 HO   821.49
              TW1  68.750 TL1   55.750 CTW1   67.810 DTW1 12.060 H1   974.75
              TW3  71.700 TL3   56.500 CTW3   70.760 DTW3 14.260 H3   824.37
           AVGTW=  69.543               AVGDTW=  13.543              AVGH=  868.00

66   16559.44 TWO  72.375 TLO   55.750 CTWO   71.051 DTWO 15.301 HO  1082.23
              TW1  71.150 TL1   55.750 CTW1   69.826 DTW1 14.076 H1  1176.42
              TW3  72.825 TL3   56.500 CTW3   71.501 DTW3 15.001 H3  1103.88
           AVGTW=  70.792               AVGDTW=  14.792              AVGH=1119.42

67   22137.40 TWO  74.100 TLO   55.750 CTWO   72.330 DTWO 16.580 HO  1335.17
              TW1  72.400 TL1   55.750 CTW1   70.630 DTW1 14.880 H1  1487.71
              TW3  75.000 TL3   56.500 CTW3   73.230 DTW3 16.730 H3  1323.20
           AVGTW=  72.063               AVGDTW=  16.063              AVGH=1378.11

68   26055.98 TWO  74.750 TLO   55.750 CTWO   72.666 DTWO 16.916 HO  1540.23
              TW1  73.400 TL1   55.750 CTW1   71.316 DTW1 15.566 H1  1673.80
              TW3  75.900 TL3   56.500 CTW3   73.816 DTW3 17.316 H3  1504.65
           AVGTW=  72.600               AVGDTW=  16.600              AVGH=1569.61

69   33893.13 TWO  77.200 TLO   55.750 CTWO   74.490 DTWO 18.740 HO  1808.56
              TW1  74.200 TL1   55.750 CTW1   71.490 DTW1 15.740 H1  2153.25
              TW3  79.500 TL3   56.500 CTW3   76.790 DTW3 20.290 H3  1670.40
           AVGTW=  74.257               AVGDTW=  18.257              AVGH=1856.44
```

```
 TABLE B-2 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ISOPROPANOL
=================================================================================
RUN      HEAT            TEMPERATURE            CORRECTED     CORRECTED    TRANSFER
NO.      FLUX        SURFACE        LIQUID          TW           DTW      COEFFICIENT
       W/SQ.M        DEG.C          DEG.C         DEG.C         DEG.C      W/SQ.M K
=================================================================================

                    P=15.33 KN/SQ.M                           TS=43.20 DEG.C

 70     6870.23 TWO    58.675 TLO   43.850 CTWO   58.125 DTWO 14.275 HO    481.25
                TW1    55.250 TL1   43.850 CTW1   54.700 DTW1 10.850 H1    633.16
                TW3    60.700 TL3   44.750 CTW3   60.150 DTW3 15.400 H3    446.10
             AVGTW=    57.659               AVGDTW=   13.509             AVGH=  508.56

 71    11755.73 TWO    60.950 TLO   43.850 CTWO   60.010 DTWO 16.160 HO    727.45
                TW1    57.325 TL1   43.850 CTW1   56.385 DTW1 12.535 H1    937.82
                TW3    64.400 TL3   44.750 CTW3   63.460 DTW3 18.710 H3    628.31
             AVGTW=    59.951               AVGDTW=   15.801             AVGH=  743.95

 72    16559.44 TWO    64.150 TLO   43.850 CTWO   62.826 DTWO 18.976 HO    872.64
                TW1    60.500 TL1   43.850 CTW1   59.176 DTW1 15.326 H1   1080.47
                TW3    65.550 TL3   44.750 CTW3   64.226 DTW3 19.476 H3    850.24
             AVGTW=    62.076               AVGDTW=   17.926             AVGH=  923.76

 73    22137.40 TWO    65.550 TLO   43.850 CTWO   63.780 DTWO 19.930 HO   1110.75
                TW1    61.850 TL1   43.850 CTW1   60.080 DTW1 16.230 H1   1363.96
                TW3    66.700 TL3   44.750 CTW3   64.930 DTW3 20.180 H3   1096.98
             AVGTW=    62.930               AVGDTW=   18.780             AVGH=1178.76

 74    26055.98 TWO    66.550 TLO   43.850 CTWO   64.466 DTWO 20.616 HO   1263.81
                TW1    62.550 TL1   43.850 CTW1   60.466 DTW1 16.616 H1   1568.04
                TW3    68.050 TL3   44.750 CTW3   65.966 DTW3 21.216 H3   1228.07
             AVGTW=    63.633               AVGDTW=   19.483             AVGH=1337.33

 75    33893.13 TWO    68.500 TLO   43.850 CTWO   65.790 DTWO 21.940 HO   1544.78
                TW1    64.150 TL1   43.850 CTW1   61.440 DTW1 17.590 H1   1926.80
                TW3    70.800 TL3   44.750 CTW3   68.090 DTW3 23.340 H3   1452.12
             AVGTW=    65.107               AVGDTW=   20.957             AVGH=1617.26
```

## TABLE B-3 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ETHANOL

```
===============================================================================
RUN     HEAT          TEMPERATURE               CORRECTED     CORRECTED   TRANSFER
NO.     FLUX      SURFACE        LIQUID            TW            DTW      COEFFICIENT
       W/SQ.M     DEG.C          DEG.C           DEG.C          DEG.C      W/SQ.M K
===============================================================================

                  P=98.18 KN/SQ.M                              TS=78.00 DEG.C

 76    3307.89 TWO   84.250 TLO  78.000 CTWO  83.985 DTWO   5.985 HO  552.65
               TW1   84.000 TL1  78.100 CTW1  83.735 DTW1   5.635 H1  586.97
               TW3   84.600 TL3  78.450 CTW3  84.335 DTW3   5.885 H3  562.04
            AVGTW=  84.018            AVGDTW=   5.835              AVGH= 566.85

 77    6870.23 TWO   85.450 TLO  78.250 CTWO  84.900 DTWO   6.650 HO 1033.00
               TW1   85.575 TL1  78.250 CTW1  85.025 DTW1   6.775 H1 1013.94
               TW3   85.575 TL3  78.450 CTW3  85.025 DTW3   6.575 H3 1044.78
            AVGTW=  84.984            AVGDTW=   6.667              AVGH=1030.42

 78   11755.73 TWO   87.350 TLO  78.250 CTWO  86.410 DTWO   8.160 HO 1440.62
               TW1   88.100 TL1  78.350 CTW1  87.160 DTW1   8.810 H1 1334.33
               TW3   86.550 TL3  78.700 CTW3  85.610 DTW3   6.910 H3 1701.22
            AVGTW=  86.393            AVGDTW=   7.960              AVGH=1476.81

 79   14351.15 TWO   88.250 TLO  78.250 CTWO  87.102 DTWO   8.852 HO 1621.10
               TW1   89.350 TL1  78.350 CTW1  88.202 DTW1   9.852 H1 1456.57
               TW3   88.325 TL3  78.700 CTW3  87.177 DTW3   8.477 H3 1692.81
            AVGTW=  87.494            AVGDTW=   9.061              AVGH=1583.83

 80   16539.44 TWO   89.100 TLO  78.250 CTWO  87.777 DTWO   9.527 HO 1735.92
               TW1   90.125 TL1  78.350 CTW1  88.802 DTW1  10.452 H1 1582.30
               TW3   88.675 TL3  78.700 CTW3  87.352 DTW3   8.652 H3 1911.47
            AVGTW=  87.977            AVGDTW=   9.544              AVGH=1732.89

 81   19592.88 TWO   90.000 TLO  78.250 CTWO  88.433 DTWO  10.183 HO 1923.95
               TW1   91.000 TL1  78.350 CTW1  89.433 DTW1  11.083 H1 1767.73
               TW3   89.350 TL3  78.700 CTW3  87.783 DTW3   9.083 H3 2156.94
            AVGTW=  88.550            AVGDTW=  10.116              AVGH=1936.63

 82   22137.40 TWO   91.000 TLO  78.450 CTWO  89.230 DTWO  10.780 HO 2053.52
               TW1   91.650 TL1  78.450 CTW1  89.880 DTW1  11.430 H1 1936.74
               TW3   90.200 TL3  79.000 CTW3  88.430 DTW3   9.430 H3 2347.49
            AVGTW=  89.180            AVGDTW=  10.546              AVGH=2098.95

 83   26055.98 TWO   91.400 TLO  78.450 CTWO  89.316 DTWO  10.866 HO 2397.73
               TW1   92.600 TL1  78.450 CTW1  90.516 DTW1  12.066 H1 2159.28
               TW3   90.650 TL3  79.000 CTW3  88.566 DTW3   9.566 H3 2723.54
            AVGTW=  89.466            AVGDTW=  10.833              AVGH=2405.10
```

```
TABLE B-3 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ETHANOL
=============================================================================
RUN    HEAT          TEMPERATURE            CORRECTED    CORRECTED   TRANSFER
NO.    FLUX      SURFACE        LIQUID         TW           DTW      COEFFICIENT
      W/SQ.M     DEG.C          DEG.C        DEG.C        DEG.C      W/SQ.M K
=============================================================================

              P=84.85 KN/SQ.M                        TS=73.25 DEG.C

84   11755.73 TWO   83.650 TLO  73.275 CTWO  82.710 DTWO  9.435 HO 1245.94
              TW1   84.000 TL1  73.275 CTW1  83.060 DTW1  9.785 H1 1201.38
              TW3   83.500 TL3  73.950 CTW3  82.560 DTW3  8.610 H3 1365.33
           AVGTW=   82.776           AVGDTW=   9.276           AVGH=1267.21

85   14351.15 TWO   84.400 TLO  73.275 CTWO  83.252 DTWO  9.977 HO 1438.32
              TW1   84.950 TL1  73.275 CTW1  83.802 DTW1 10.527 H1 1363.18
              TW3   84.250 TL3  73.950 CTW3  83.102 DTW3  9.152 H3 1567.97
           AVGTW=   83.386           AVGDTW=   9.886           AVGH=1451.66

86   16539.44 TWO   85.075 TLO  73.275 CTWO  83.752 DTWO 10.477 HO 1578.53
              TW1   85.575 TL1  73.275 CTW1  84.252 DTW1 10.977 H1 1506.63
              TW3   84.700 TL3  73.950 CTW3  83.377 DTW3  9.427 H3 1754.33
           AVGTW=   83.794           AVGDTW=  10.294           AVGH=1606.64

87   19592.88 TWO   86.200 TLO  73.275 CTWO  84.633 DTWO 11.358 HO 1724.93
              TW1   86.700 TL1  73.275 CTW1  85.133 DTW1 11.858 H1 1652.20
              TW3   85.750 TL3  73.950 CTW3  84.183 DTW3 10.233 H3 1914.55
           AVGTW=   84.650           AVGDTW=  11.150           AVGH=1757.16

88   22137.40 TWO   86.600 TLO  73.275 CTWO  84.830 DTWO 11.555 HO 1915.79
              TW1   86.900 TL1  73.275 CTW1  85.130 DTW1 11.855 H1 1867.31
              TW3   86.000 TL3  73.950 CTW3  84.230 DTW3 10.280 H3 2153.40
           AVGTW=   84.730           AVGDTW=  11.230           AVGH=1971.23

89   26055.98 TWO   86.900 TLO  73.275 CTWO  84.816 DTWO 11.541 HO 2257.50
              TW1   88.450 TL1  73.275 CTW1  86.366 DTW1 13.091 H1 1990.23
              TW3   86.700 TL3  73.950 CTW3  84.616 DTW3 10.666 H3 2442.68
           AVGTW=   85.266           AVGDTW=  11.766           AVGH=2214.33

90   33893.13 TWO   88.700 TLO  73.275 CTWO  85.990 DTWO 12.715 HO 2665.51
              TW1   90.100 TL1  73.275 CTW1  87.390 DTW1 14.115 H1 2401.14
              TW3   88.000 TL3  73.950 CTW3  85.290 DTW3 11.340 H3 2988.70
           AVGTW=   86.223           AVGDTW=  12.723           AVGH=2663.77
```

TABLE B-3 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ETHANOL

```
===========================================================================================
RUN      HEAT          TEMPERATURE                 CORRECTED     CORRECTED     TRANSFER
NO.      FLUX       SURFACE         LIQUID             TW           DTW       COEFFICIENT
        W/SQ.M      DEG.C           DEG.C             DEG.C        DEG.C       W/SQ.M K
===========================================================================================

                  P=71.32 KN/SQ.M                          TS=69.00 DEG.C

 91    6870.23 TWO   77.575 TLO   68.700 CTWO   77.025 DTWO   8.325 HO   825.18
               TW1   77.100 TL1   68.700 CTW1   76.550 DTW1   7.850 H1   875.10
               TW3   78.250 TL3   69.150 CTW3   77.700 DTW3   8.550 H3   803.46
            AVGTW=   77.092             AVGDTW=    8.242               AVGH=  833.52

 92   11755.73 TWO   79.800 TLO   68.700 CTWO   78.860 DTWO  10.160 HO  1157.04
               TW1   79.700 TL1   68.700 CTW1   78.760 DTW1  10.060 H1  1168.54
               TW3   80.000 TL3   69.150 CTW3   79.060 DTW3   9.910 H3  1186.23
            AVGTW=   78.893             AVGDTW=   10.043               AVGH=1170.48

 93   16539.44 TWO   81.500 TLO   68.700 CTWO   80.177 DTWO  11.477 HO  1441.00
               TW1   81.500 TL1   68.700 CTW1   80.177 DTW1  11.477 H1  1441.00
               TW3   80.700 TL3   69.150 CTW3   79.377 DTW3  10.227 H3  1617.11
            AVGTW=   79.911             AVGDTW=   11.061               AVGH=1495.28

 94   22137.40 TWO   82.800 TLO   68.700 CTWO   81.030 DTWO  12.330 HO  1795.38
               TW1   82.800 TL1   68.700 CTW1   81.030 DTW1  12.330 H1  1795.38
               TW3   82.050 TL3   69.150 CTW3   80.280 DTW3  11.130 H3  1988.94
            AVGTW=   80.780             AVGDTW=   11.930               AVGH=1855.57

 95   26055.98 TWO   83.800 TLO   68.700 CTWO   81.716 DTWO  13.016 HO  2001.70
               TW1   83.800 TL1   68.700 CTW1   81.716 DTW1  13.016 H1  2001.70
               TW3   83.000 TL3   69.150 CTW3   80.916 DTW3  11.766 H3  2214.33
            AVGTW=   81.450             AVGDTW=   12.600               AVGH=2067.89

 96   33893.13 TWO   84.900 TLO   68.700 CTWO   82.190 DTWO  13.490 HO  2512.39
               TW1   84.000 TL1   68.700 CTW1   81.290 DTW1  12.590 H1  2691.98
               TW3   84.950 TL3   69.150 CTW3   82.240 DTW3  13.090 H3  2589.16
            AVGTW=   81.907             AVGDTW=   13.057               AVGH=2595.77
```

```
  TAELE B-3 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ETHANOL
=================================================================
RUN     HEAT          TEMPERATURE             CORRECTED     CORRECTED    TRANSFER
NO.     FLUX     SURFACE          LIQUID         TW            DTW      COEFFICIENT
       W/SQ.M    DEG.C            DEG.C        DEG.C          DEG.C      W/SQ.M K
=================================================================


               P=44.65 KN/SQ.M                        TS=59.00 DEG.C

 97    6870.23 TWO    68.775 TLO   58.775 CTWO   68.225 DTWO   9.450 HO   726.95
               TW1    68.475 TL1   58.775 CTW1   67.925 DTW1   9.150 H1   750.78
               TW3    71.150 TL3   59.550 CTW3   70.600 DTW3  11.050 H3   621.70
            AVGTW= 68.917               AVGDTW=   9.884             AVGH=  695.08

 98   11755.73 TWO    70.800 TLO   58.775 CTWO   69.860 DTWO  11.085 HO  1060.49
               TW1    70.475 TL1   58.775 CTW1   69.535 DTW1  10.760 H1  1092.52
               TW3    72.350 TL3   59.550 CTW3   71.410 DTW3  11.860 H3   991.19
            AVGTW= 70.268               AVGDTW=  11.235             AVGH=1046.33

 99   16539.44 TWO    72.300 TLO   58.775 CTWO   70.977 DTWO  12.202 HO  1355.39
               TW1    72.050 TL1   58.775 CTW1   70.727 DTW1  11.952 H1  1383.73
               TW3    73.550 TL3   59.550 CTW3   72.227 DTW3  12.677 H3  1304.60
            AVGTW= 71.311               AVGDTW=  12.277             AVGH=1347.11

100   22137.40 TWO    74.500 TLO   58.775 CTWO   72.730 DTWO  13.955 HO  1586.32
               TW1    73.625 TL1   58.775 CTW1   71.855 DTW1  13.080 H1  1692.43
               TW3    75.050 TL3   59.550 CTW3   73.280 DTW3  13.730 H3  1612.31
            AVGTW= 72.621               AVGDTW=  13.588             AVGH=1629.12

101   26055.98 TWO    75.450 TLO   58.775 CTWO   73.366 DTWO  14.591 HO  1785.64
               TW1    74.500 TL1   58.775 CTW1   72.416 DTW1  13.641 H1  1909.99
               TW3    76.100 TL3   59.550 CTW3   74.016 DTW3  14.466 H3  1801.07
            AVGTW= 73.266               AVGDTW=  14.233             AVGH=1830.59

102   33893.13 TWO    76.075 TLO   58.775 CTWO   73.365 DTWO  14.590 HO  2322.97
               TW1    75.575 TL1   58.775 CTW1   72.865 DTW1  14.090 H1  2405.40
               TW3    77.100 TL3   59.550 CTW3   74.390 DTW3  14.840 H3  2283.84
            AVGTW= 73.540               AVGDTW=  14.507             AVGH=2336.32
```

## TABLE B-3 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ETHANOL

| RUN NO. | HEAT FLUX W/SQ.M | | TEMPERATURE SURFACE DEG.C | | TEMPERATURE LIQUID DEG.C | | CORRECTED TW DEG.C | | CORRECTED DTW DEG.C | | TRANSFER COEFFICIENT W/SQ.M K |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | P=31.32 KN/SQ.M | | | | | | TS=51.00 DEG.C | | |
| 103 | 6870.23 | TWO | 63.000 | TLO | 51.650 | CTWO | 62.450 | DTWO | 10.800 | HO | 636.09 |
| | | TW1 | 62.100 | TL1 | 51.650 | CTW1 | 61.550 | DTW1 | 9.900 | H1 | 693.91 |
| | | TW3 | 65.350 | TL3 | 52.350 | CTW3 | 64.800 | DTW3 | 12.450 | H3 | 551.79 |
| | | AVGTW= | 62.934 | | | AVGDTW= | 11.050 | | | AVGH= | 621.70 |
| 104 | 11755.73 | TWO | 65.200 | TLO | 51.650 | CTWO | 64.260 | DTWO | 12.610 | HO | 932.24 |
| | | TW1 | 64.800 | TL1 | 51.650 | CTW1 | 63.860 | DTW1 | 12.210 | H1 | 962.78 |
| | | TW3 | 67.275 | TL3 | 52.350 | CTW3 | 66.335 | DTW3 | 13.985 | H3 | 840.58 |
| | | AVGTW= | 64.818 | | | AVGDTW= | 12.935 | | | AVGH= | 908.82 |
| 105 | 16539.44 | TWO | 66.400 | TLO | 51.650 | CTWO | 65.077 | DTWO | 13.427 | HO | 1231.74 |
| | | TW1 | 66.000 | TL1 | 51.650 | CTW1 | 64.677 | DTW1 | 13.027 | H1 | 1269.55 |
| | | TW3 | 68.825 | TL3 | 52.350 | CTW3 | 67.502 | DTW3 | 15.152 | H3 | 1091.51 |
| | | AVGTW= | 65.752 | | | AVGDTW= | 13.869 | | | AVGH= | 1192.51 |
| 106 | 22137.40 | TWO | 68.650 | TLO | 51.650 | CTWO | 66.880 | DTWO | 15.230 | HO | 1453.52 |
| | | TW1 | 68.200 | TL1 | 51.650 | CTW1 | 66.430 | DTW1 | 14.780 | H1 | 1497.77 |
| | | TW3 | 70.000 | TL3 | 52.350 | CTW3 | 68.230 | DTW3 | 15.880 | H3 | 1394.02 |
| | | AVGTW= | 67.180 | | | AVGDTW= | 15.296 | | | AVGH= | 1447.18 |
| 107 | 26055.98 | TWO | 70.000 | TLO | 51.650 | CTWO | 67.916 | DTWO | 16.266 | HO | 1601.77 |
| | | TW1 | 68.650 | TL1 | 51.650 | CTW1 | 66.566 | DTW1 | 14.916 | H1 | 1746.74 |
| | | TW3 | 71.150 | TL3 | 52.350 | CTW3 | 69.066 | DTW3 | 16.716 | H3 | 1558.66 |
| | | AVGTW= | 67.850 | | | AVGDTW= | 15.966 | | | AVGH= | 1631.87 |
| 108 | 33893.13 | TWO | 71.700 | TLO | 51.650 | CTWO | 68.990 | DTWO | 17.340 | HO | 1954.57 |
| | | TW1 | 70.325 | TL1 | 51.650 | CTW1 | 67.615 | DTW1 | 15.965 | H1 | 2122.91 |
| | | TW3 | 72.750 | TL3 | 52.350 | CTW3 | 70.040 | DTW3 | 17.690 | H3 | 1915.90 |
| | | AVGTW= | 68.882 | | | AVGDTW= | 16.998 | | | AVGH= | 1993.86 |

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TABLE B-3 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING ETHANOL
=================================================================================
RUN      HEAT          TEMPERATURE             CORRECTED     CORRECTED    TRANSFER
NO.      FLUX       SURFACE       LIQUID          TW            DTW      COEFFICIENT
        W/SQ.M      DEG.C         DEG.C         DEG.C         DEG.C       W/SQ.M K
=================================================================================

                  P=19.33 KN/SQ.M                          TS=41.25 DEG.C

109    6870.23 TWO    53.750 TLO   40.950 CTWO   53.200 DTWO 12.250 HO   560.80
               TW1    52.800 TL1   40.950 CTW1   52.250 DTW1 11.300 H1   607.94
               TW3    56.050 TL3   41.675 CTW3   55.500 DTW3 13.825 H3   496.92
            AVGTW=    53.650               AVGDTW=   12.459             AVGH=   551.42

110   11755.73 TWO    56.550 TLO   40.950 CTWO   55.610 DTWO 14.660 HO   801.88
               TW1    55.500 TL1   40.950 CTW1   54.560 DTW1 13.610 H1   863.74
               TW3    58.875 TL3   41.675 CTW3   57.935 DTW3 16.260 H3   722.97
            AVGTW=    56.035               AVGDTW=   14.843             AVGH=   791.98

111   16539.44 TWO    58.075 TLO   40.950 CTWO   56.752 DTWO 15.802 HO  1046.62
               TW1    57.000 TL1   40.950 CTW1   55.677 DTW1 14.727 H1  1123.01
               TW3    61.175 TL3   41.675 CTW3   59.852 DTW3 18.177 H3   909.87
            AVGTW=    57.427               AVGDTW=   16.236             AVGH=1018.68

112   22137.40 TWO    59.400 TLO   40.950 CTWO   57.630 DTWO 16.680 HO  1327.16
               TW1    58.075 TL1   40.950 CTW1   56.305 DTW1 15.355 H1  1441.69
               TW3    63.700 TL3   41.675 CTW3   61.930 DTW3 20.255 H3  1092.92
            AVGTW=    58.621               AVGDTW=   17.430             AVGH=1270.06

113   26055.98 TWO    60.700 TLO   40.950 CTWO   58.616 DTWO 17.666 HO  1474.84
               TW1    59.350 TL1   40.950 CTW1   57.266 DTW1 16.316 H1  1596.87
               TW3    65.000 TL3   41.675 CTW3   62.916 DTW3 21.240 H3  1226.63
            AVGTW=    59.600               AVGDTW=   18.408             AVGH=1475.42

114   33893.13 TWO    63.000 TLO   40.950 CTWO   60.290 DTWO 19.340 HO  1752.45
               TW1    60.950 TL1   40.950 CTW1   58.240 DTW1 17.290 H1  1960.23
               TW3    65.500 TL3   41.675 CTW3   62.790 DTW3 21.115 H3  1605.14
            AVGTW=    60.440               AVGDTW=   19.248             AVGH=1760.80
```

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 TABLE B-4 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING METHANOL
==========================================================================
RUN      HEAT          TEMPERATURE            CORRECTED     CORRECTED    TRANSFER
NO.      FLUX      SURFACE         LIQUID         TW           DTW      COEFFICIENT
        W/SQ.M      DEG.C          DEG.C         DEG.C        DEG.C       W/SQ.M K
==========================================================================

                   P=98.64 KN/SQ.M                       TS=64.00 DEG.C

 115     6870.23 TWO    72.600 TL0    64.450 CTWO    72.050 DTWO    7.600 HO   903.89
                 TW1    71.300 TL1    64.450 CTW1    70.750 DTW1    6.300 H1  1090.38
                 TW3    74.250 TL3    65.150 CTW3    73.700 DTW3    8.550 H3   803.46
              AVGTW=    72.167               AVGDTW=    7.484              AVGH=  917.98

 116    11755.73 TWO    74.250 TL0    64.450 CTWO    73.310 DTWO    8.860 HO  1326.80
                 TW1    73.800 TL1    64.450 CTW1    72.860 DTW1    8.410 H1  1397.80
                 TW3    75.650 TL3    65.150 CTW3    74.710 DTW3    9.560 H3  1229.65
              AVGTW=    73.626               AVGDTW=    8.943              AVGH=1314.44

 117    16539.44 TWO    75.650 TL0    64.600 CTWO    74.327 DTWO    9.727 HO  1700.23
                 TW1    75.400 TL1    64.600 CTW1    74.077 DTW1    9.477 H1  1745.08
                 TW3    76.850 TL3    65.250 CTW3    75.527 DTW3   10.277 H3  1609.25
              AVGTW=    74.644               AVGDTW=    9.827              AVGH=1682.93

 118    22137.40 TWO    77.300 TL0    64.600 CTWO    75.530 DTWO   10.930 HO  2025.34
                 TW1    76.850 TL1    64.600 CTW1    75.080 DTW1   10.480 H1  2112.30
                 TW3    78.400 TL3    65.275 CTW3    76.630 DTW3   11.355 H3  1949.53
              AVGTW=    75.746               AVGDTW=   10.921              AVGH=2026.88

 119    26055.98 TWO    78.100 TL0    64.600 CTWO    76.016 DTWO   11.416 HO  2282.22
                 TW1    77.875 TL1    64.600 CTW1    75.791 DTW1   11.191 H1  2328.10
                 TW3    79.575 TL3    65.275 CTW3    77.491 DTW3   12.216 H3  2132.77
              AVGTW=    76.433               AVGDTW=   11.608              AVGH=2244.54

 120    33893.13 TWO    79.500 TL0    64.600 CTWO    76.790 DTWO   12.190 HO  2780.31
                 TW1    79.425 TL1    64.600 CTW1    76.715 DTW1   12.115 H1  2797.52
                 TW3    81.475 TL3    65.275 CTW3    78.765 DTW3   13.490 H3  2512.39
              AVGTW=    77.423               AVGDTW=   12.598              AVGH=2690.20
```

```
TABLE E-4 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING METHANOL
=====================================================================
RUN     HEAT          TEMPERATURE            CORRECTED   CORRECTED   TRANSFER
NO.     FLUX      SURFACE        LIQUID         TW          DTW     COEFFICIENT
       W/SQ.M      DEG.C         DEG.C        DEG.C       DEG.C     W/SQ.M K
=====================================================================

              p=79.98 KN/SQ.M                          TS=58.75 DEG.C

121    6870.23 TWO   67.850 TLO  58.750 CTWO  67.300 DTWO  8.550 HO  803.46
               TW1   65.650 TL1  58.750 CTW1  65.100 DTW1  6.350 H1 1081.80
               TW3   70.450 TL3  59.200 CTW3  69.900 DTW3 10.700 H3  642.03
            AVGTW= 67.434              AVGDTW=   8.534           AVGH= 805.03

122   11755.73 TWO   70.095 TLO  58.750 CTWO  69.155 DTWO 10.405 HO 1129.79
               TW1   68.300 TL1  58.750 CTW1  67.360 DTW1  8.610 H1 1365.33
               TW3   71.100 TL3  59.200 CTW3  70.160 DTW3 10.960 H3 1072.58
            AVGTW= 68.891              AVGDTW=   9.991           AVGH=1176.53

123   16539.44 TWO   71.275 TLO  58.750 CTWO  69.952 DTWO 11.202 HO 1476.37
               TW1   69.800 TL1  58.750 CTW1  68.477 DTW1  9.727 H1 1700.23
               TW3   72.600 TL3  59.200 CTW3  71.277 DTW3 12.077 H3 1369.41
            AVGTW= 69.902              AVGDTW=  11.002           AVGH=1503.21

124   22137.40 TWO   72.600 TLO  58.750 CTWO  70.830 DTWO 12.080 HO 1832.53
               TW1   71.450 TL1  58.750 CTW1  69.680 DTW1 10.930 H1 2025.34
               TW3   73.750 TL3  59.200 CTW3  71.980 DTW3 12.780 H3 1732.16
            AVGTW= 70.830              AVGDTW=  11.930           AVGH=1855.57

125   26055.98 TWO   73.775 TLO  58.750 CTWO  71.691 DTWO 12.941 HO 2013.30
               TW1   72.370 TL1  58.750 CTW1  70.286 DTW1 11.536 H1 2258.48
               TW3   75.125 TL3  59.200 CTW3  73.041 DTW3 13.841 H3 1882.39
            AVGTW= 71.673              AVGDTW=  12.773           AVGH=2039.83

126   33893.13 TWO   74.650 TLO  58.750 CTWO  71.940 DTWO 13.190 HO 2569.53
               TW1   74.100 TL1  58.750 CTW1  71.390 DTW1 12.640 H1 2681.33
               TW3   76.970 TL3  59.200 CTW3  74.260 DTW3 15.060 H3 2250.48
            AVGTW= 72.530              AVGDTW=  13.630           AVGH=2486.58
```

```
TABLE B-4 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING METHANOL
=====================================================================================
RUN      HEAT          TEMPERATURE              CORRECTED      CORRECTED    TRANSFER
NO.      FLUX        SURFACE      LIQUID           TW             DTW      COEFFICIENT
        W/SQ.M        DEG.C       DEG.C          DEG.C           DEG.C      W/SQ.M K
=====================================================================================

              P=66.65 KN/SQ.M                         TS=54.80 DEG.C

127   6870.23 TWO  65.550 TLO  54.675 CTWO  65.000 DTWO 10.325 HO  665.35
              TW1  62.600 TL1  54.675 CTW1  62.050 DTW1  7.375 H1  931.46
              TW3  67.350 TL3  55.350 CTW3  66.800 DTW3 11.450 H3  599.98
           AVGTW=  64.617              AVGDTW=  9.717             AVGH= 707.00

123  11755.73 TWO  66.700 TLO  54.675 CTWO  65.760 DTWO 11.085 HO 1060.49
              TW1  64.920 TL1  54.675 CTW1  63.980 DTW1  9.305 H1 1263.35
              TW3  68.050 TL3  55.350 CTW3  67.110 DTW3 11.760 H3  999.62
           AVGTW=  65.616              AVGDTW= 10.716             AVGH=1096.94

129  16539.44 TWO  67.850 TLO  54.675 CTWO  66.527 DTWO 11.852 HO 1395.41
              TW1  66.300 TL1  54.675 CTW1  64.977 DTW1 10.302 H1 1605.34
              TW3  69.750 TL3  55.350 CTW3  68.427 DTW3 13.077 H3 1264.70
           AVGTW=  66.644              AVGDTW= 11.744             AVGH=1408.28

130  22137.40 TWO  69.450 TLO  54.675 CTWO  67.680 DTWO 13.005 HO 1702.19
              TW1  68.100 TL1  54.675 CTW1  66.330 DTW1 11.655 H1 1899.35
              TW3  70.800 TL3  55.350 CTW3  69.030 DTW3 13.680 H3 1618.20
           AVGTW=  67.680              AVGDTW= 12.780             AVGH=1732.16

131  26055.98 TWO  70.700 TLO  54.675 CTWO  68.616 DTWO 13.941 HO 1868.89
              TW1  68.800 TL1  54.675 CTW1  66.716 DTW1 12.041 H1 2163.77
              TW3  72.150 TL3  55.350 CTW3  70.066 DTW3 14.716 H3 1770.47
           AVGTW=  68.466              AVGDTW= 13.566             AVGH=1920.55

132  33893.13 TWO  71.695 TLO  54.675 CTWO  68.985 DTWO 14.310 HO 2368.42
              TW1  70.800 TL1  54.675 CTW1  68.090 DTW1 13.415 H1 2526.43
              TW3  74.200 TL3  55.350 CTW3  71.490 DTW3 16.140 H3 2099.89
           AVGTW=  69.522              AVGDTW= 14.622             AVGH=2317.94
```

```
TABLE B-4 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING METHANOL
=================================================================================================
RUN      HEAT          TEMPERATURE               CORRECTED       CORRECTED      TRANSFER
NO.      FLUX       SURFACE       LIQUID            TW              DTW        COEFFICIENT
        W/SQ.M      DEG.C         DEG.C            DEG.C           DEG.C       W/SQ.M K
=================================================================================================

                 P=53.32 KN/SQ.M                             TS=50.00 DEG.C

133     6870.23 TWO    61.850 TL0   50.200 CTWO    61.300 DTWO 11.100 HO    618.90
                TW1    58.650 TL1   50.200 CTW1    58.100 DTW1  7.900 H1    869.57
                TW3    63.250 TL3   50.700 CTW3    62.700 DTW3 12.000 H3    572.48
             AVGTW=    60.700             AVGDTW=  10.334                AVGH= 664.81

134    11755.73 TWO    63.600 TL0   50.200 CTWO    62.660 DTWO 12.460 HO    943.46
                TW1    61.065 TL1   50.200 CTW1    60.125 DTW1  9.925 H1   1184.43
                TW3    64.695 TL3   50.700 CTW3    63.755 DTW3 13.055 H3    900.46
             AVGTW=    62.180             AVGDTW=  11.813                AVGH= 995.11

135    16539.44 TWO    64.650 TL0   50.200 CTWO    63.327 DTWO 13.127 HO   1259.88
                TW1    62.425 TL1   50.200 CTW1    61.102 DTW1 10.902 H1   1517.00
                TW3    66.125 TL3   50.700 CTW3    64.802 DTW3 14.102 H3   1172.78
             AVGTW=    63.077             AVGDTW=  12.711                AVGH=1301.18

136    22137.40 TWO    66.450 TL0   50.200 CTWO    64.680 DTWO 14.480 HO   1528.80
                TW1    64.175 TL1   50.200 CTW1    62.405 DTW1 12.205 H1   1813.76
                TW3    67.800 TL3   50.700 CTW3    66.030 DTW3 15.330 H3   1444.04
             AVGTW=    64.371             AVGDTW=  14.005                AVGH=1580.65

137    26055.98 TWO    67.050 TL0   50.200 CTWO    64.966 DTWO 14.766 HO   1764.48
                TW1    64.700 TL1   50.200 CTW1    62.616 DTW1 12.416 H1   2098.42
                TW3    68.500 TL3   50.700 CTW3    66.416 DTW3 15.716 H3   1657.83
             AVGTW=    64.666             AVGDTW=  14.300                AVGH=1822.06

138    33893.13 TWO    68.825 TL0   50.200 CTWO    66.115 DTWO 15.915 HO   2129.58
                TW1    66.750 TL1   50.200 CTW1    64.040 DTW1 13.840 H1   2448.85
                TW3    70.800 TL3   50.700 CTW3    68.090 DTW3 17.390 H3   1948.95
             AVGTW=    66.082             AVGDTW=  15.715                AVGH=2156.68
```

```
 TABLE B-4 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING METHANOL
=============================================================================
RUN      HEAT           TEMPERATURE              CORRECTED    CORRECTED   TRANSFER
NO.      FLUX       SURFACE       LIQUID            TW           DTW     COEFFICIENT
        W/SQ.M      DEG.C         DEG.C           DEG.C        DEG.C     W/SQ.M K
=============================================================================

               P=41.32 KN/SQ.M                        TS=44.00 DEG.C

 139     6870.23 TWO    57.950 TLO   44.400 CTWO    57.400 DTWO 13.000 HO   528.45
                 TW1    53.550 TL1   44.400 CTW1    53.000 DTW1  8.600 H1   798.79
                 TW3    59.325 TL3   45.000 CTW3    58.775 DTW3 13.775 H3   498.72
             AVGTW= 56.392                 AVGDTW= 11.792              AVGH= 582.60

 140    11755.73 TWO    58.245 TLO   44.400 CTWO    57.305 DTWO 12.905 HO   910.93
                 TW1    56.345 TL1   44.400 CTW1    55.405 DTW1 11.005 H1  1068.20
                 TW3    60.000 TL3   45.000 CTW3    59.060 DTW3 14.060 H3   836.10
             AVGTW= 57.256                 AVGDTW= 12.656              AVGH= 928.80

 141    16539.44 TWO    60.350 TLO   44.400 CTWO    59.027 DTWO 14.627 HO  1130.69
                 TW1    57.575 TL1   44.400 CTW1    56.252 DTW1 11.852 H1  1395.41
                 TW3    62.125 TL3   45.000 CTW3    60.802 DTW3 15.802 H3  1046.62
             AVGTW= 58.694                 AVGDTW= 14.094              AVGH=1173.47

 142    22137.40 TWO    62.100 TLO   44.400 CTWO    60.330 DTWO 15.930 HO  1389.65
                 TW1    59.400 TL1   44.400 CTW1    57.630 DTW1 13.230 H1  1673.24
                 TW3    63.650 TL3   45.000 CTW3    61.880 DTW3 16.880 H3  1311.44
             AVGTW= 59.946                 AVGDTW= 15.346              AVGH=1442.47

 143    26055.98 TWO    62.800 TLO   44.400 CTWO    60.716 DTWO 16.316 HO  1596.87
                 TW1    59.720 TL1   44.400 CTW1    57.636 DTW1 13.236 H1  1968.43
                 TW3    64.400 TL3   45.000 CTW3    62.316 DTW3 17.316 H3  1504.65
             AVGTW= 60.223                 AVGDTW= 15.623              AVGH=1667.73

 144    33893.13 TWO    64.820 TLO   44.400 CTWO    62.110 DTWO 17.710 HO  1913.74
                 TW1    62.120 TL1   44.400 CTW1    59.410 DTW1 15.010 H1  2257.97
                 TW3    67.220 TL3   45.000 CTW3    64.510 DTW3 19.510 H3  1737.18
             AVGTW= 62.010                 AVGDTW= 17.410              AVGH=1946.72
```

### TABLE B-4 EXPERIMENTAL DATA OF HEAT TRANSFER TO BOILING METHANOL

```
=================================================================================================
RUN      HEAT          TEMPERATURE             CORRECTED       CORRECTED      TRANSFER
NO.      FLUX      SURFACE        LIQUID           TW             DTW       COEFFICIENT
        W/SQ.M     DEG.C          DEG.C          DEG.C          DEG.C        W/SQ.M K
=================================================================================================

              P=27.99 KN/SQ.M                               TS=36.00 DEG.C

 145     6870.23 TWO   50.200 TLO  36.150 CTWO   49.650 DTWO 13.500 HO   508.88
                 TW1   46.600 TL1  36.150 CTW1   46.050 DTW1  9.900 H1   693.91
                 TW3   52.600 TL3  36.650 CTW3   52.050 DTW3 15.400 H3   446.10
              AVGTW=  49.250             AVGDTW=  12.934              AVGH=  531.17

 146    11755.73 TWO   51.595 TLO  36.150 CTWO   50.655 DTWO 14.505 HO   810.45
                 TW1   48.745 TL1  36.150 CTW1   47.805 DTW1 11.655 H1  1008.63
                 TW3   53.075 TL3  36.650 CTW3   52.135 DTW3 15.485 H3   759.16
              AVGTW=  50.198             AVGDTW=  13.881              AVGH=  846.84

 147    16539.44 TWO   53.275 TLO  36.150 CTWO   51.952 DTWO 15.802 HO  1046.62
                 TW1   50.375 TL1  36.150 CTW1   49.052 DTW1 12.902 H1  1281.85
                 TW3   55.150 TL3  36.650 CTW3   53.827 DTW3 17.177 H3   962.84
              AVGTW=  51.611             AVGDTW=  15.294              AVGH=1081.40

 148    22137.40 TWO   54.825 TLO  36.150 CTWO   53.055 DTWO 16.905 HO  1309.50
                 TW1   51.925 TL1  36.150 CTW1   50.155 DTW1 14.005 H1  1580.65
                 TW3   56.625 TL3  36.650 CTW3   54.855 DTW3 18.205 H3  1215.99
              AVGTW=  52.688             AVGDTW=  16.371              AVGH=1352.16

 149    26055.98 TWO   56.050 TLO  36.150 CTWO   53.966 DTWO 17.816 HO  1462.43
                 TW1   53.200 TL1  36.150 CTW1   51.116 DTW1 14.966 H1  1740.90
                 TW3   58.050 TL3  36.650 CTW3   55.966 DTW3 19.316 H3  1348.87
              AVGTW=  53.683             AVGDTW=  17.366              AVGH=1500.32
```

## C.3 ETHANOL

### Surface Tension

The following linear relationships correlate the surface tension of ethanol with saturation temperature:

For temperature range 0°C - 90°C

$$\sigma = 245.25454E-04 - 894.54545E-07*TS$$

For temperature range 90°C - 230°C

$$\sigma = 267.67141E-04 - 111.57142E-06*TS$$

where $\sigma$ has the units of N/m

### Thermal Conductivity

The relationships of the following form have been obtained for the thermal conductivity of ethanol:

For temperature range 0°C - 110°C

$$k_\ell = 170.79485E-03 - 897.17142E-07*TS$$

For temperature range 110°C - 243°C

$$k_\ell = -0.57007E+00 + 0.18558E-01*TS - 0.17283E-03*(TS**2.) + 0.70115E-06*(TS**3.) - 0.10648E-08*(TS**4.)$$

Where $k_\ell$ has the units of W/m K

### Latent Heat

The following equations have been obtained to correlate latent heat:

For temperature range 0°C - 90°C

$$\lambda = 0.10485E+07 - 0.10294E+04*TS + 0.92875E+01*(TS**2.) - 0.19893E+00*(TS**3.) + 0.89102E-03*(TS**4.)$$

# APPENDIX C

## PROPERTY ESTIMATION METHODS

Properties of pure liquids : distilled water, isopropanol, ethanol and methanol are readily available in literature [11], [125-134] as a function of temperature and pressure in tabular forms. However, these properties are either available in FPS or in CGS or in MKS system of units. Storing these vast number of data for computer calculations, a large amount of memory space is required. Therefore, these data were first converted in SI units and then suitable equations were obtained to express these properties as a function of saturation temperature, TS, in degree centigrade. Method of least squares was employed to develop these equations. These equations were the basis for a subroutine which was incorporated into the main program for the calculation of these properties. Computer programs in Fortran IV were executed on IBM 360/Model 44 computer.

### C.1 DISTILLED WATER

#### Surface Tension

The equations for surface tension of distilled water were obtained from the straight line fit of the available data. The form of correlations are:

For temperature range $0^{o}C$ - $110^{o}C$

$$\sigma = 761.30636E{-}04 - 170.67825E{-}06*TS$$

For temperature range 110°C - 310°C

$$\sigma = 826{,}90126\mathrm{E}{-}04-226.43082\mathrm{E}{-}06*\mathrm{TS}$$

where $\sigma$ has the units of N/m.

Thermal Conductivity

The following polynomial equations were obtained for correlating thermal conductivity of distilled water:

For temperature range 0°C - 110°C

$$k_{\ell} = 0{,}55081\mathrm{E}{+}00 + 0{,}26386\mathrm{E}{-}02*\mathrm{TS} - 0{,}13691\mathrm{E}{-}04*(\mathrm{TS}**2.) - 0{,}15631\mathrm{E}{-}07*(\mathrm{TS}**3.) + 0{,}20796\mathrm{E}{-}09*(\mathrm{TS}**4.)$$

For temperature range 110°C -310°C

$$k_{\ell} = 0{,}60666\mathrm{E}{+}00 + 0{,}12652\mathrm{E}{-}02*\mathrm{TS}-0{,}54635\mathrm{E}{-}05*(\mathrm{TS}**2.) + 0{,}46741\mathrm{E}{-}08*(\mathrm{TS}**3.) - 0{,}98305\mathrm{E}{-}11*(\mathrm{TS}**4.)$$

where $k_{\ell}$ has the units of W/m K

Latent Heat

The equations obtained for the latent heat are of the following form:

For temperature range 0°C - 100°C

$$\lambda = 0{,}25009\mathrm{E}{+}07 - 0{,}23655\mathrm{E}{+}04*\mathrm{TS} + 0{,}63545\mathrm{E}{+}00*(\mathrm{TS}**2.0) - 0{,}16820\mathrm{E}{-}01*(\mathrm{TS}**3.) + 0{,}28206\mathrm{E}{-}04*(\mathrm{TS}**4.)$$

For temperature range 100° - 300°C

$$\lambda = 0{,}24259\mathrm{E}{+}07 -0{,}55119\mathrm{E}{+}03*\mathrm{TS} - 0{,}15886\mathrm{E}{+}02*(\mathrm{TS}**2.) + 0{,}55052\mathrm{E}{-}01*(\mathrm{TS}**3.) - 0{,}11262\mathrm{E}{-}03*(\mathrm{TS}**4.)$$

where $\lambda$ has the units of J/kg.

## Specific Heat

Following relationships have been obtained for correlating specific heat of distilled water with saturation temperature:

For temperature range 0°C - 110°C

$$C_{\ell} = \text{0.42112E+04 - 0.21366E+01*TS + 0.35700E-01*(TS**2.)}$$
$$\text{-0.16983E-03*(TS**3.) + 0.35052E-06*(TS**4.)}$$

For temperature range 110°C - 310°C

$$C_{\ell} = \text{0.76008E+04-0.79556E+02*TS + 0.67767E+00*(TS**2.)}$$
$$\text{-0.24695E-02*(TS**3.) + 0.34187E-05*(TS**4.)}$$

Where $C_{\ell}$ has the units of J/kg K

## Liquid Density

The following polynomials relate liquid density as a function of saturation temperature for different range of temperatures:

For temperature range 0°C - 110°C

$$\rho_{\ell} = \text{0.99991E+03+042965E-01*TS - 0.70786E-02*(TS**2.)}$$
$$\text{+0.34295E-04*(TS**3.) - 0.93198E-07*(TS**4.).}$$

For temperature range 110°C - 310°C

$$\rho_{\ell} = \text{0.96214E+03+0.8393E+00*TS - 0.12197E-01*(TS**2.)}$$
$$\text{+0.39162E-04*(TS**3.) - 0.56920E-07*(TS**4.)}$$

Where $\rho_{\ell}$ has the units of kg/m$^3$

## Vapour Density

The equations obtained for the vapour density are of the following form:

For temperature range 0°C - 110°C

$$\rho_v = 0.52558E-02 + 0.17461E-03*TS + 0.21264E-04*(TS**2.) - 0.10163E-06*(TS**3.) + 0.46442E-08*(TS**4.)$$

For temperature range 110°C - 310°C

$$\rho_v = 0.28603E+02 - 0.66988E+00*TS + 0.58191E-02*(TS**2.) - 0.22209E-04*(TS**3.) + ( 0.36374E-07*(TS**4.)$$

where $\rho_v$ has the units of kg/m$^3$

## Vapour Pressure

The following polynomial was obtained for correlating vapour pressure with saturation temperature;

For temperature range 0°C - 101°C

$$P = 0.68877E+03 + 0.10886E+02*TS + 0.39057E +01*(TS**2.) -0.38904E-01(TS**3.) + 0.99335E-03*(TS**4.)$$

Where P has the units of N/m$^2$

## Viscosity

The viscosity data have been correlated by the following fifth order polynomials:

For Temperature range 0°C - 110°C

$$\mu_\ell = [179.33352E^+01 - 621.36517E-01*TS + 149.86551E-02*(TS**2.) -224.41365E-04*(TS**3.)+175.34761E-06*(TS**4.)- - 536.58698E -09*(TS**5.)] *10^{-6}$$

For temperature range 110$^{o}$C - 310$^{o}$C

$$\mu_{\ell} = [635.35320E+00 - 471.85935E-02*TS + 100.70339E-04*(TS**2.) + 278.65450E-07*(TS**3.) - 144.57533E-09*(TS**4.) + 157.95537E-12*(TS**5.)] *10^{-6}$$

Where $\mu_{\ell}$ has the units of N s/m$^2$

## C.2 ISOPROPANOL

### Surface Tension

Surface tension of isopropanol is obtained as a linear relationship with saturation temperature:

For temperature range 0$^{o}$C - 100$^{o}$C

$$\sigma = 254.45449E-04 - 830.90818E-07*TS$$

where $\sigma$ has the units of N/m

### Thermal Conductivity

The following linear relationship relates the thermal conductivity as follows:

For temperature range 0 - 85$^{o}$C

$$k_{\ell} = 153.63230E-03 - 158.16800E-06*TS$$

where $k_{\ell}$ has the units of W/m K

### Latent Heat

The values of latent heat of isopropanol between temperature range 30.2$^{o}$C to 82.6$^{o}$C has been obtained by the application of Clausius-Clapeyron equation. These values then have been converted in SI units and used in obtaining a Polynomial of fifth order which relates the latent heat of isopropanol with saturation temperature as follows:

For temperature range 30.2°C - 82.6°C

$\lambda$ = 203.03796E+04-480.80087E+02*TS + 456.30770E+00*(TS**2.)
-359.94529E-03*(TS**3.) + 243.95033E -04*(TS**4)
- 379.12940E-06*(TS**5.)

where $\lambda$ has the units of J/kg.

Specific Heat

The following linear relationship has been obtained for the specific heat for temperature range of 0°C to 85°C.

$C_\ell$ = 243.92761E+01+126.03376E-01*TS

where $C_\ell$ has the units of J/kg K

Liquid Density

Liquid density of isopropanol is available only upto a temperature range of 0°C to 30°C. A plot between liquid density and temperature is a linear relationship and it shows that the values can be extrapolated upto 90°C.

The form of the equation is

$\rho_\ell$ = 801.27856E+00-811.71419E-03*TS

where $\rho_\ell$ has the units of $kg/m^3$

Vapour Density

The vapour density of isopropanol in the temperature range 2.40°C - 82.5°C has been calculated by using ideal gas law. These values then converted into SI units and a fifth order polynomial of the following form has been obtained:

$$\rho_v = 288.90611E-04+255.00299E-05*TS - 975.27464E-08*(TS**2.) +419.90463E-08*(TS**3.) - 420.65384E-10*(TS**4.) + 386.78963E-12*(TS**5.)$$

where $\rho_v$ has the units of kg/m$^3$.

Vapour Pressure

Following is the equation for vapour pressure :

For temperature range -26.1°C - 82.5°C

$$P = 0.22520E+04-0.31228E+03*TS + 0.20468E+02*(TS**2.) -0.26690E+00*(TS**3.) + 0.29200E-02*(TS**4.)$$

where P has the units of N/m$^2$

Viscosity

The viscosity data have been correlated by the following fifth order polynomials :

For temperature range 0°C - 120°C

$$\mu_\ell = 459.10201E-05-156.16738E-06*TS + 280.63930E-08*(TS**2.) -297.86025E-10*(TS**3.) + 172.56563E-12*(TS**4.) -413.69649E-15*(TS**5.)$$

For temperature range 120°C -230°C

$$\mu_\ell = 130.98034E-05-117.39022E-07*TS - 438.09089E-11*(TS**2.) +459.53980E-12*(TS**3.) - 199.85684E-14*(TS**4.) + 261.14554E-17*(TS**5.)$$

where $\mu_\ell$ has the units of N s/m$^2$

For temperature range 90°C - 230°C

$$\lambda = 0.96103E+06 + 0.25716E+04*TS - 0.47834E+02*(TS**2.) \\ +0.24525E+00*(TS**3.) - 0.57849E-03*(TS**4.)$$

Where $\lambda$ has the units of J/kg

Specific Heat

The specific heat is represented by the following fourth order polynomial:

For temperature range 0°C - 140°C

$$C_\ell = .22663E+04 + 0.58808E+01*TS + 0.47357E-01*(TS**2.) \\ -0.29583E-04*(TS**3.) + 0.51473E-07*(TS**4.)$$

where $C_\ell$ has the units of J/kg K

Liquid Density

The following polynomials relate liquid density of ethanol as a function of saturation temperature for different range of temperatures:

For temperature range 2.8°C - 95.6°C

$$\rho_\ell = 0.82547E+03 - 0.97801E+00*TS + 0.93018E-02*(TS**2.) \\ -0.18321E-03*(TS**3.) + 0.98893E-06*(TS**4.)$$

For temperature range 95.6°C - 246.4°C

$$\rho_\ell = -0.28369E+03 +0.29204E+02*TS - 0.29719E+00*(TS**2.) \\ +0.12670E-02*(TS*3.) -0.20228E-05*(TS**4.)$$

Where $\rho_\ell$ has the units of kg/m$^3$

Vapour Density

The equations obtained for the vapour density are of the following form:

For temperature range 2.8°C - 95.6°C

$$\rho_v = 0.41022E-01 - 0.18142E-02*TS + 0.28808E-03*(TS**2.) - 0.37326E-05*(TS**3.) + 0.42259E-07*(TS**4.)$$

For temperature range 95.6°C - 246.4°C

$$\rho_v = 0.96346E+03 - 0.26052E+02*TS + 0.25854E+00*(TS**2.) - 0.11170E-02*(TS**3.) + 0.18016E-05*(TS**4.)$$

where $\rho_v$ has the units of $kg/m^3$

Vapour Pressure

The following form of the polynomial was obtained for correlating vapour pressure of ethanol with saturation temperature:

For temperature range -31.3°C - 87.4°C

$$P = 0.44420E+03 + 0.40266E+03*TS - 0.28843E+02*(TS**2.) + 0.91876E+00*(TS**3.) - 0.56114E-02*(TS**4.)$$

Where P has the units of $N/m^2$

Viscosity

The viscosity data have been correlated by the following fifth order polynomials:

For temperature range 0°C - 70°C

$\mu_\ell$ = 177.35584E-03 - 332.62798E-05*TS + 190.25797E-07*(TS**2.)
+ 420.43859E-09*(TS**3.) -921.65497E-11*(TS**4.)
+ 533.96415E-13*(TS**5.)

For temperature range 70°C - 231.67°C

$\mu_\ell$ = 123.85988E-05 - 154.34408E-07*TS + 822.04773E-10*(TS**2.
- 228.68571E-12*(TS**3.) + 361.33156E-15*(TS**4.)
- 314.72716E-18*(TS**5.)

Where $\mu_\ell$ has the units of N s/m$^2$

## C.4 METHANOL

### Surface Tension

The equations for surface tension of methanol were obtained from the straight line fit of the available data. The form of correlations are:

For temperature range 0°C - 90°C

$\sigma$ = 244.03636E-04 - 863.03030E-07*TS

For temperature range 90°C - 220°C

$\sigma$ = 274.74213E-04 - 114.92071E-06*TS

where $\sigma$ has the units of N/m

### Thermal Conductivity

The following form of linear relationship was obtained for correlating thermal conductivity .

For temperature range 0°C - 75°C

$k_\ell = 213.75939E-03 - 125.60384E-06*TS$

where $k_\ell$ has the units of W/m K.

Latent Heat

The polynomial obtained for the latent heat is of the following form:

For temperature range 0°C - 190°C

$$\lambda = 0.12087E+07 - 0.51348E+03*\ TS - 0.16398E+02*(TS**2.) \\ +0.52303E-01*(TS**3.) -0.14244E-03*(TS**4.).$$

Where $\lambda$ has the units of J/kg

Specific Heat

The following linear relationship was obtained for the specific heat of methanol:

For temperature range 0°C - 65°C

$C_\ell = 240.78574E+01 + 457.93247E-02*TS$

where $C_\ell$ has the units of J/kg K

Liquid Density

The following polynomial relates liquid density of methanol:

For temperature range 0°C - 190°C

$$\rho_\ell = 0.80955E+03-0.86131E+00*TS - 0.72658E-03*(TS**2.) \\ +0.22534E-05*(TS**3.)- 0.44294E-07*(TS**4.)$$

where $\rho_\ell$ has the units of $kg/m^3$

Vapour Density

The equations obtained for the vapour density are of the following form:

For temperature range 30°C - 100°C

$$\rho_v = 0.81702E-01 - 0.14616E-02*TS + 0.27308E-03*(TS**2.) - 0.16417E-05*(TS**3.) + 0.29594E-07*(TS**4.)$$

For temperature range 110°C - 240°C

$$\rho_v = 0.12113E + 04 - 0.32479E+02*TS + 0.32134E+00*(TS**2) - 0.13898E-02*(TS**3.) + 0.22486E-05*(TS**4.)$$

Where $\rho_v$ has the units of $kg/m^3$

Vapour Pressure

The following form of the polynomials was obtained for correlating vapour pressure:

For temperature range - 44.0°C - 64.7°C

$$P = -0.21529E+04 + 0.15060E + 04*TS - 0.84399E+02*(TS**2.) + 0.17112E+01*(TS**3.) - 0.58831E-02*(TS**4.)$$

where P has the units of $N/m^2$

Viscosity

The viscosity data have been correlated by the following fifth order polynomials.

For Temperature range 0°C - 110°C

$$\mu_\ell = 822.70646E-06 - 384.39353E-07*TS + 174.99079E-08*(TS**2.) - 373.81654E-10*(TS**3.) + 348.98035E-12*(TS**4.) - - 117.53708E -14*(TS**5.)$$

For temperature range 110°C - 220°C

$\mu_{\ell}$ = 518.99366E-06 - 341.25670E-08*TS - 262.27932E-11*(TS**2.)

+807.73165E-13*(TS**3.) -210.88401E-15*(TS**4.)

+ 984.35692E-19*(TS**5.)

Where $\mu_{\ell}$ has the units of N s/m$^2$

# APPENDIX D

## SAMPLE CALCULATIONS

Run no. 16 for distilled water is employed to demonstrate the calculational procedure. The following experimental data are taken for the run:

System pressure, P = 47.32 kN/m$^2$

Saturation temperature, $T_s$ = 80.0 °C

Heat flux, q = 16539.44 W/m$^2$

O,D, of heating surface = 70 mm

I,D, of heating surface = 62 mm

Circumferential temperatures of heating surface and corresponding temperatures of fluid are:

| | Temperatures, °C | | |
|---|---|---|---|
| | Side | Top | Bottom |
| Heating surface | 87.900 | 85.550 | 90.500 |
| Test-fluid | 79.900 | 79.900 | 81.450 |

The pertinent physico-thermal properties for distilled water at the saturation temperature (80°C) are obtained from Appendix C and are as follows:

$\sigma = 0.0625$ N/m ; $\rho_\ell = 971.8$ kg/m$^3$

$k_\ell = 0.6747$ W/m $^\circ$C ; $\rho_v = 0.2923$kg/m$^3$

$C_\ell = 4.196$ kJ/kg$^\circ$C ; $\mu_\ell = 0.3481$ mN s/m$^2$

$\lambda = 2.308$ MJ/kg ; $\alpha = 1.6546\text{x}10^{-7}$ m$^2$/s

## SURFACE TEMPERATURE CORRECTION

For the thin walled cylinder as in the present investigation, the temperature drop across the wall is calculated by the following equation:

$$T_i - T_o = \frac{q d_o}{2k} \ln \frac{d_o}{d_h} \qquad \text{(D.1)}$$

Where,

$d_o$ = outside diameter of the heating surface, m

$d_h$ = inside diameter of the heating surface + $\frac{1}{2}(d_o - d_i)$

$$T_i - T_o = \frac{16539.44 \text{ x } 70 \text{ x } 10^{-3}}{2 \text{ x } 22.15 \text{ x } 1.163} \ln \frac{70}{66}$$

$$= \underline{1.323\ ^\circ\text{C}}$$

Therefore, corrected surface temperatures are as follows:

$T_{wo} = 87.900 - 1.323 = 86.577\ ^\circ$C

$T_{w1} = 85.550 - 1.323 = 84.227\ ^\circ$C

$T_{w3} = 90.500 - 1.323 = 89.177\ ^\circ$C

Subscripts 0,1 and 3 represent the side, top and bottom positions of the wall thermocouples respectively.

The procedure for the calculation of point and average heat transfer coefficients is as follows:

## AVERAGE SURFACE TEMPERATURE

Eq. (A.7) provides the calculation for average surface temperature

$$\bar{T}_w = \frac{1}{3}\left[ T_{wo} + T_{w1} + T_{w_3} \right] \qquad (A.7)$$

$$= \frac{1}{3}\left[86.577 + 84.227 + 89.177\right]$$

$$= \underline{86.660}\ ^{o}C$$

## AVERAGE FLUID TEMPERATURE

$\bar{T}_\ell$ is calculated similar to $\bar{T}_w$ as follows:

$$\bar{T}_\ell = \frac{1}{3}\left[79.900 + 79.900 + 81.450\right]$$

$$= \underline{80.416}\ ^{o}C$$

## TEMPERATURE DIFFERENCE

Temperature difference at the side, at the top and at the bottom of heating surface are as follows:

$$\Delta T_o = 86.577 - 79.900 = 6.677\ ^{o}C$$

$$\Delta T_1 = 84.227 - 79.900 = 4.327\ ^{o}C$$

$$\Delta T_3 = 89.177 - 81.450 = 7.727\ ^{o}C$$

and average temperature difference is

$$\frac{1}{3}\left[6.677 + 4.327 + 7.727\right]$$

$$= \underline{6.243\ ^{o}C}$$

## EXPERIMENTAL HEAT TRANSFER COEFFICIENT

The point values of experimental heat transfer coefficient at the side, the top and the bottom are as follows:

$$h_o = \frac{16{,}539.44}{6.677} = \underline{2477.08} \quad W/m^2\ K$$

$$h_1 = \frac{16{,}539.44}{4.327} = \underline{3822.38} \quad W/m^2\ K$$

$$h_3 = \frac{16{,}539.44}{7.727} = \underline{2140.47} \quad W/m^2\ K$$

and average experimental heat transfer coefficient is calculated as :

$$\bar{h} = q / (\overline{\Delta T})$$

$$= \frac{16{,}539.44}{6.243} = \underline{2649.27} \quad W/m^2\ K$$

## WALL SUPERHEAT FROM ALAD'EV EQUATION

Theoretical wall superheat calculation is made using Alad'ev equation, Eq.(5.4) as follows:

$$\frac{\Delta T_w}{T_s} = 4.7 \times 10^{-3} \left[\frac{10^{-6}\ q\ \lambda}{k_\ell\ T_s\ g}\right]^{0.3} \left[\frac{\lambda}{C_\ell\ T_s}\right]^{1.2} \qquad (5.4)$$

$$\frac{\Delta T_w}{353.0} = 4.7\times10^{-3}\left[\frac{10^{-6}\times16{,}539.44\times23.08\times10^5}{0.6747\ \times\ (80+273)\ \times\ 9.81}\right]^{0.3} \times$$

$$\left[\frac{23.08\ \times\ 10^5}{4196\ \times\ (80+273)}\right]^{1.2}$$

$$\Delta T_w = \underline{6.531}\ {}^oC$$

## BUBBLE EMISSION FREQUENCY

The frequency of bubble emission is obtained from equation, Eq.(5.7)

$$f = \frac{1}{\theta_d + \theta_w} \tag{5.7}$$

$\theta_d$ from Eq.(5,10) is:

$$\theta_d = \frac{[133.3/P]^2 \; [\sigma/(\rho_\ell - \rho_v)g]}{\pi \quad \alpha_\ell \quad J_a^2} \tag{5,10}$$

$$= \frac{[133,3/47,32]^2 \quad [0,0625/(971,8-0,2923)9,81]}{\pi \times 1,6546 \times 10^{-7}(39,47)^2}$$

$$= \underline{0,06636} \quad s$$

$\theta_w$ from Eq. (5,13) is:

$$\theta_w = \frac{0,867}{\alpha_\ell} \left[ \frac{k_\ell \; \Delta T_w}{q} \right]^2 \tag{5,13}$$

$$= \frac{0,867}{1,6546 \times 10^{-7}} \left[ \frac{0,6747 \times 6,531}{16539,44} \right]^2$$

$$= \underline{0,3793} \quad s$$

Therefore,

$$f = \frac{1}{0,06636 + 0,3793}$$

$$= \underline{2,243 \; 1/s}$$

DETERMINATION OF CONSTANT M

For distilled water constant M is determined from Eq.(6.3a) as follows:

$$M = 8.61938E\text{-}05(\,P\,)^{-4.4409803E-01} \qquad (6.3a)$$

$$= 8.61938E\text{-}05(47.32)^{-4.4409803E-01}$$

$$= \underline{1.5545 \times 10^{-5}}$$

HEAT TRANSFER COEFFICIENT FROM PRESENT ANALYSIS, Eq. (6.4)

Eq. (6.4) is:

$$h = M\left[\left\{\left(\frac{\lambda^{2.5}\,\rho_v}{T_s^{\,1.5}\,\sigma}\right)\left(\frac{q}{k_\ell\,g}\right)^{0.3}\left(\frac{1}{C_\ell}\right)^{1.2}\right\}^{2.33}\{f\}\right]^{0.3683} \qquad (6.4)$$

$$h = 1.5545\text{x}10^{-5}\left[\left\{\left(\frac{(2308\text{x}10^3)^{2.5}\ \text{x}\ 0.2923}{(353.0)^{1.5}\text{x}0.0625}\right)\left(\frac{16539.44}{0.6747\text{x}9.81}\right)^{0.3}\left(\frac{1}{4196}\right)^{1.2}\right\}^{2.33}(2.243)\right]^{0.3683}$$

$$= \underline{2585.45}\ W/m^2\ K$$

$$\text{Deviation}^{*} = \frac{2649.27 - 2585.45}{2649.27} \times 100 = \underline{2.41\ \%}$$

---

[*] Deviation is calculated as percentage change with respect to the experimental value of heat transfer coefficient.

DETERMINATION OF $M/M_1$

The value of $M/M_1$ is obtained from Eq. (6.5) as follows:

$$\frac{M}{M_1} = 0.95831645 \left(\frac{P}{P_1}\right)^{-0.3997825} \qquad (6.5)$$

$$= 0.95831645 \left(\frac{47.32}{98.44}\right)^{-0.3997825}$$

$$= \underline{1.284374}$$

DETERMINATION OF $h^*/h_1^*$ FROM PRESENT ANALYSIS, Eq. (6.6)

Eq (6.6) is

$$\frac{h^*}{h_1^*} = \frac{M}{M_1}\left[\left\{\left(\frac{\lambda}{\lambda_1}\right)^{2.5}\left(\frac{\rho_v}{\rho_{v_1}}\right)\left(\frac{T_{sl}}{T_s}\right)^{1.5}\left(\frac{\sigma_1}{\sigma}\right)\left(\frac{k_{\ell 1}}{k_\ell}\right)^{0.3}\left(\frac{q}{q_1}\right)^{0.3}\left(\frac{C_{\ell 1}}{C_\ell}\right)^{1.2}\right\}^{2.33}\left\{\frac{f}{f_1}\right\}\right]^{0.3683}$$

$$= 1.284374\left[\left\{\left(\frac{2308\text{x}10^3}{2259\text{x}10^3}\right)^{2.5}\left(\frac{0.2923}{0.5784}\right)\left(\frac{372.0}{353.0}\right)^{1.5}\left(\frac{0.05923}{0.0625}\right)\left(\frac{0.6826}{0.6747}\right)^{0.3}\left(\frac{16539.44}{16539.44}\right)^{0.3}\left(\frac{4218.0}{4196.0}\right)^{1.2}\right]^{2.33} \times \left(\frac{2.243}{3.59511}\right)\right]^{0.3683}$$

$$= \underline{0.7454}$$

$$\text{Deviation} = \frac{0.8160 - 0.7454}{0.8160} \times 100$$

$$= \underline{8.65\ \%}$$

## HEAT TRANSFER PREDICTIONS FROM OTHER CORRELATIONS

The values of experimental heat transfer coefficient during saturated boiling were compared against the predicted values by a number of other correlations.

Heat Transfer Coefficient from the Correlation of Jakob and Linke [4]

The correlation is:

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{(\rho_\ell - \rho_v)g}} = 31.6\left(\frac{\mu_{\ell,a}}{\mu_\ell}\right)\left(\frac{\rho_\ell}{\rho_{\ell,a}}\right)\left[\frac{\rho_{\ell,a}}{\rho_\ell} \times \frac{\sigma}{\sigma_a} \times \frac{q}{\rho_{v,a}\lambda_a w_b}\right]^{0.8}$$

$$\frac{h}{0.6747}\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}} = 31.6\left(\frac{0.2962\times10^{-3}}{0.3481\times10^{-3}}\right)\left(\frac{971.8}{959.1}\right)$$

$$\left[\left(\frac{959.1}{971.8}\right)\left(\frac{0.0625}{0.0592}\right)\left(\frac{16539.44}{0.5784\times22.59\ \times10^5\times0.078}\right)\right]^{0.8}$$

$h = \underline{1682.19}$ W/m$^2$ K

$$\text{Deviation}^{*} = \frac{2648.67 - 1682.19}{2648.67} \times 100$$

$$= \underline{36.5\ \%}$$

---

* Deviation is calculated as percentage change with respect to the experimental value of heat transfer coefficient.

Heat Transfer Coefficient from the Correlation of Kutateladze [39]

The correlation is :

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}} = 7.0 \times 10^{-4}\left[\frac{\bar{q}}{\rho_v\lambda\ \alpha_\ell}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}}\right]^{0.7} \times$$

$$\left[\frac{P}{\sqrt{g\ \sigma\ (\rho_\ell-\rho_v)}}\right]^{0.7} \quad \left(\frac{C_\ell\ \mu_\ell}{k_\ell}\right)^{-0.35}$$

$$\frac{h}{0.6747}\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}} = 7.0\text{x}10^{-4}\left[\frac{16,539.44}{0.2923\text{x}23.08\text{x}10^{5}\text{x}1.6\text{x}10^{-7}}\right.$$

$$\left.\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}}\right]^{0.7}\left[\frac{47321.5}{\sqrt{9.81\text{x}0.0625(971.8-0.2923)}}\right]^{0.7}$$

$$\left(\frac{4196\ \text{x}\ 0.3481\text{x}10^{-3}}{0.6747}\right)^{-0.35}$$

$h = \underline{1795.12}$ W/m$^2$ K

$$\text{Deviation} = \frac{2648.67 - 1795.12}{2648.67} \text{ x } 100$$

$= \underline{32.2\ \%}$

## Heat Transfer Coefficient from the Correlation of Labuntsov [36]

The correlation is:

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}} = 0.125\left[\frac{q}{\rho_v\,\lambda\,\alpha_\ell}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}}\right]^{0.65} \times$$

$$\left[\frac{C_\ell\,\mu_\ell}{k_\ell}\right]^{-.0.32}\left[\frac{(\rho_v\,\lambda)^2}{C_\ell(t_s+273)\rho_\ell\,\sqrt{\sigma\,(\rho_\ell-\rho_v)g}}\right]^{0.35}$$

$$\frac{h}{0.6747}\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}} = 0.125\left[\frac{16539.44}{0.2923\text{x}23.08\text{x}10^5\text{x}1.6\text{x}10^{-7}}\right.$$

$$\left.\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}}\right]^{0.65}\left[\frac{4196\text{x}0.3481\text{x}10^{-3}}{0.6747}\right]^{-0.32} \times$$

$$\left[\frac{(0.2923\ \text{x}23.08\text{x}\ 10^5)^2}{4196(80+273)\ 971.8\ \sqrt{0.0625(971.8-0.2923)9.81}}\right]^{0.35}$$

$h = \underline{2995.11}$ W/m$^2$ K

$$\text{Deviation} = \frac{2648.67 - 2995.11}{2648.67} \text{ x } 100$$

$= \underline{-\ 13.1\ \%}$

Heat Transfer Coefficient from the Correlation of Kichigen and Tobilevich [39]

The correlation is :

$$\frac{h}{k_\ell}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}}$$

$$= 1.04 \times 10^{-4}\left[\frac{q}{\rho_v\lambda\,\alpha}\sqrt{\frac{\sigma}{(\rho_\ell-\rho_v)g}}\right]^{0.7}\left[\frac{P}{\sqrt{\sigma\,(\rho_\ell-\rho_v)\,g}}\right]^{0.7} X$$

$$\left[\frac{g\,\rho_\ell^2}{\mu_\ell^2}\left\{\frac{\sigma}{(\rho_\ell-\rho_v)g}\right\}^{3/2}\left(1-\frac{\rho_v}{\rho_\ell}\right)\right]^{0.125}$$

$$\frac{h}{0.6747}\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}} = 1.04\mathrm{x}10^{-4}\left[\frac{16539.44}{0.2923\mathrm{x}23.08\mathrm{x}10^{5}\mathrm{x}1.6\mathrm{x}10^{-7}}\right.$$

$$\left.\sqrt{\frac{0.0625}{(971.8-0.2923)9.81}}\right]^{0.7} \mathrm{x}\left[\frac{47321.5}{\sqrt{0.0625(971.8-0.2923)9.81}}\right]^{0.7} X$$

$$\left[9.81\left(\frac{971.8}{0.3481\mathrm{x}10^{-3}}\right)^{2}\left\{\frac{0.0625}{(971.8-0.2923)9.81}\right\}^{3/2} X\right.$$

$$\left.\left(1-\frac{0.2923}{971.8}\right)\right]^{0.125}$$

$h = \underline{2027.20}$ W/m$^2$ K

$$\text{Deviation} = \frac{2648.67 - 2027.20}{2648.67} \mathrm{x}\ 100$$

$= \underline{23.4\ \%}$

Wall Heat Flux from the Correlation of Forster and Greif [33]

The correlation is:

$$q = 1,2 \times 10^{-3} \left[ \frac{\alpha_\ell \, C_\ell \, \rho_\ell \, (t_s+273)}{\lambda \, \rho_v \, \sigma^{1/2}} \right] \left[ \frac{C_\ell \, (t_s+273) \, \alpha_\ell^{1/2}}{(\lambda \, \rho_v)^2} \right]^{1/4} \times$$

$$\left[ \frac{\rho_\ell}{\mu_\ell} \right]^{5/8} \left[ \frac{\mu_\ell \, C_\ell}{k_\ell} \right]^{1/3} (\Delta P)^2$$

$$= 1,2 \times 10^{-3} \left[ \frac{1.6\text{x}10^{-7}\text{x}4196 \text{ x}971.8(80+273)}{23,08\text{x}10^5 \text{ x } 0.2923\text{x}\sqrt{0,0625}} \right] \times$$

$$\left[ \frac{4196(80+273) \sqrt{1.6\text{x}10^{-7}}}{(23,08\text{x}10^5\text{x } 0,2923)^2} \right]^{1/4} \left[ \frac{971.8}{0,3481\text{x}10^{-3}} \right]^{5/8} \times$$

$$\left[ \frac{0,3481 \text{ x } 10^{-3} \text{ x } 4196}{0,6747} \right]^{1/3} (14346.65)^2$$

$= \underline{28906.28}$ W/m$^2$

$$\text{Deviation} = \frac{16539.44 - 28906.28}{16539.44} \times 100$$

$= - \underline{74,8}$ %

# APPENDIX E

## ANALYSIS OF ERRORS

Errors in measured heat transfer coefficient may be caused by inaccuracies in the component measurements required for the calculation of heat transfer coefficient. In order to acquire an appreciation for the accuracy of the experimental data, error analysis was performed for several experimental runs. In this Appendix a sample calculation is presented using the data of Run 16.

As mentioned above that the parameter of interest to this propagation of error calculation is the uncertainty of the experimental heat transfer coefficient. The experimental uncertainty used here, is the absolute value of maximum expected deviation from the reported experimental result.

The experimental uncertainty for the average heat transfer coefficient can be defined as:

$$E_{\bar{h}} = \left[ \sum_{i=1}^{n} \left( \frac{\partial \bar{h}}{\partial y_i} \cdot E_{yi} \right)^2 \right]^{1/2} \qquad (E.1)$$

where $y_i$ is any of n parameters of which the heat transfer coefficient is a function. Since $\bar{h}$ has been calculated from

$$\bar{h} = \frac{Q}{A(\bar{T}_{wc} - \bar{T}_{\ell})} \qquad (E.2)$$

where ,

$Q$ Power input, W

$\bar{T}_{wc}$ Average corrected wall temperature , $^{o}C$

$\bar{T}_{\ell}$ Average liquid temperature , $^{o}C$

$A$ Heat transfer area, $m^2$

Equation (E.1) can be expressed as

$$E_{\bar{h}} = \left[ \left( \frac{\partial \bar{h}}{\partial Q} E_Q \right)^2 + \left( \frac{\partial \bar{h}}{\partial A} E_A \right)^2 + \left( \frac{\partial \bar{h}}{\partial \bar{T}_{w.c}} E_{\bar{T}_{wc}} \right)^2 + \left( \frac{\partial \bar{h}}{\partial \bar{T}_{\ell}} E_{\bar{T}_{\ell}} \right)^2 \right]^{1/2} \qquad (E.3)$$

The partial derivatives existing in Eq(E.3) can be evaluated by using Eq.(E.2). Thus the uncertainty in $\bar{h}$ is

$$E_{\bar{h}} = \left[ \left( \frac{E_Q}{A(\bar{T}_{wc} - \bar{T}_{\ell})} \right)^2 + \left( - \frac{Q\, E_A}{A^2(\bar{T}_{wc} - \bar{T}_{\ell})} \right)^2 + \left( - \frac{Q\, E_{\bar{T}_{wc}}}{A(\bar{T}_{wc} - \bar{T}_{\ell})^2} \right)^2 + \left( \frac{Q\, E_{\bar{T}_{\ell}}}{A(\bar{T}_{wc} - \bar{T}_{\ell})^2} \right)^2 \right]^{1/2} \qquad (E.4)$$

Thus to evaluate Eq.(E.4) , uncertainties for the power input, for the heat transfer area, for the average corrected surface temperature and liquid temperature are required to evaluate.

EVALUATION OF UNCERTAINTY IN POWER INPUT, $E_Q$

Since,

$$Q = VI$$

Then,

$$E_Q = \left[ (VE_I)^2 + (IE_V)^2 \right]^{1/2} \qquad (E.5)$$

Where

$E_I$ is uncertainty associated with ammeter reading

$E_V$ is uncertainty associated with voltmeter reading.

For Run 16,

$I = 13$ amps $\qquad E_I = 0.05$ amps.

$V = 50$ volts $\qquad E_V = 1.0$ volt

$$Q = VI = 50 \times 13 = 650 \text{ W}$$

and $E_Q$ from Eq. (E.5);

$$E_Q = \left[ (50 \times 0.05)^2 + (13 \times 1)^2 \right]^{1/2}$$

$$= 13.238 \text{ W}$$

EVALUATION OF UNCERTINTY IN HEAT TRANSFER AREA, $E_A$

Since $A = \pi d_o \ell$

Then

$$E_A = \left[ (\pi \ell E_{do})^2 + (\pi d_o E_\ell)^2 \right]^{1/2}$$

where ,

$E_{do}$ is uncertainty associated with diameter measurement,

$E_\ell$ is uncertainty associated with length measurement

$\ell = 0.179$ m $\qquad E_\ell = 0.0005$ m

$d_o = 0.07$ m $\qquad E_{do} = 0.0001$ m

Therefore,

$$A = \pi \times 0.07 \times 0.179 = 3.93 \times 10^{-2} \ \mathrm{m}^2$$

$$E_A = \left[ (\pi \times 0.179 \times 0.0001)^2 + (\pi \times 0.07 \times 0.0005)^2 \right]^{1/2}$$

$$= 1.235 \times 10^{-4} \ \mathrm{m}^2$$

## EVALUATION OF UNCERTAINTY IN TEMPERATURES $E_{\bar{T}_{wc}}$ & $E_{\bar{T}_\ell}$

Since the average surface temperature and liquid temperature was obtained by averaging the individual temperatures

$$\bar{T}_{wc} = \frac{\sum_{i=1}^{n} T_{w_{ci}}}{n} \qquad (E.6)$$

Therefore,

$$E_{\bar{T}_{wc}} = \left[ \sum_{i=1}^{n} \left( \frac{\partial \bar{T}_{wc}}{\partial T_{wc_i}} E_{T_{wc}} \right)^2 + \left( \frac{\partial \bar{T}_{wc}}{\partial n} E_n \right)^2 \right]^{1/2} \qquad (E.7)$$

where $n = 3$ ; $\frac{\partial \bar{T}_{wc}}{\partial n} = 0$

and $\frac{\partial \bar{T}_{wc}}{\partial T_{wc_i}} = \frac{1}{n}$

Thus Eq.(E.7) gives

$$E_{\bar{T}_{wc}} = \left[ n \left( \frac{E_{T_{wc}}}{n} \right)^2 \right]^{1/2} \qquad (E.8)$$

Similarly,

$$E_{\bar{T}_\ell} = \left[ n \left( \frac{E_{T_\ell}}{n} \right)^2 \right]^{1/2} \qquad (E.9)$$

The following equation has been used to determine the corrected surface temperature $T_{wc}$

$$T_i - T_{wc} = \frac{q\,d_o}{2k} \ln \frac{d_o}{dh} \qquad (E,10)$$

where,

$T_i$ is inside surface temperature where thermocouple bead was located

$k$ is the thermal conductivity of metal

$$\text{or} \quad T_{wc} = \left(T_i - \frac{qd_o}{2k} \ln \frac{d_o}{dh}\right)$$

Therefore,

$$E_{T_{wc}} = \Bigg[ \left( E_{Ti} \right)^2 + \left( - E_q \frac{d_o}{2k} \ln \frac{d_o}{dh} \right)^2 + \left\{ \left( - \frac{q}{2k} \ln \frac{d_o}{dh} - \frac{q}{2k} \right) E_{do} \right\}^2 + \left( + \frac{qd_o}{2k^2} \ln \frac{d_o}{dh} E_k \right)^2 + \left( \frac{qd_o}{2k} \frac{1}{dh} E_{dh} \right)^2 \Bigg]^{1/2} \qquad (E,11)$$

Where $E_{Ti}$ is uncertainty associated with inside temperatuure measurement,

Now

$E_{Ti} = 0,001\ ^{o}C$ $\qquad E_k = 0,0$

$k = 25,76$ W/m K

$$q = \frac{Q}{\pi\, d_o \ell} = \pi \frac{650}{0,0393}$$

$$= 16539.44 \ \mathrm{W/m^2}$$

And

$$E_q = \left[ \left( \frac{E_Q}{\pi\, d_o\, \ell} \right)^2 + \left( - \frac{Q}{\pi\, \ell}\, \frac{1}{d_o^2}\, E_{do} \right)^2 + \left( - \frac{Q}{\pi\, d_o}\, \frac{1}{\ell^2}\, E_\ell \right)^2 \right]^{1/2} \qquad (E.12)$$

Putting the values of $E_Q$, $E_{do}$, $E_\ell$ and other quantities, we get

$$E_q = \left[ \left( \frac{13.238}{0.0393} \right)^2 + \left( - \frac{650 \times 0.0001}{0.0393 \times 0.07} \right)^2 + \left( - \frac{650 \times 0.0005}{0.0393 \times 0.179} \right)^2 \right]^{1/2}$$

$$= 343.44 \ \mathrm{W/m^2}$$

Now $d_h = \dfrac{d_i + d_o}{2}$

$$= \frac{0.062 + 0.070}{2} = 0.066 \ \mathrm{m}$$

Therefore,

$$E_{dh} = \left[ \left( \frac{1}{2} E_{di} \right)^2 + \left( \frac{1}{2} E_{do} \right)^2 \right]^{1/2}$$

And $E_{di} = E_{do}$

Therefore $E_{dh} = \left[ 2 \left( \frac{1}{2} E_{do} \right)^2 \right]^{1/2}$

$$= \left[ 2 \left( \frac{1}{2} \times 0.0001 \right)^2 \right]^{1/2}$$

$$= 7.071 \times 10^{-5} \ \mathrm{m}$$

Now putting the values of all the quantities in Eq (E.11) to determine $E_{T_{wc}}$ :

$$E_{T_{wc}} = \Bigg[ (0.001)^2 + \left( - \frac{343.44 \times 0.07}{2 \times 25.76}\, \ln \frac{0.070}{0.066} \right)^2 + \left\{ \left( - \frac{16539.44}{2 \times 25.76}\, \ln \frac{0.070}{0.066} - \frac{16539.44}{2 \times 25.76} \right) 0.0001 \right\}^2 + \left( \frac{16539.44 \times 0.070}{2 \times 25.76 \times 0.066} \times 7.071 \times 10^{-5} \right)^2 \Bigg]^{1/2}$$

$$E_{T_{wc}} = 4.9877 \times 10^{-2}$$

After knowing $E_{T_{wc}}$, $E_{\bar{T}_{wc}}$ can be calculated by using Eq (E.8).

$$E_{\bar{T}_{wc}} = \left[ 3 \left( \frac{4.9897 \times 10^{-2}}{3} \right)^2 \right]^{1/2}$$

$$= 2.8808 \times 10^{-2} \quad ; \bar{T}_{wc} = 86.66\ ^\circ C$$

And $E_{\bar{T}_\ell}$ is calculated from Eq (E.9)

Using $E_{T_\ell} = E_{T_i} = 0.001\ ^\circ C$

Therefore

$$E_{\bar{T}_\ell} = \left[ 3 \left( \frac{0.001}{3} \right)^2 \right]^{1/2}$$

$$= 5.7735 \times 10^{-4}$$

And $\bar{T}_\ell = 80.416\ ^\circ C$

Having calculated $E_{\bar{T}_{wc}}$ and $E_{\bar{T}_\ell}$, uncertainty in heat transfer coefficient $E_{\bar{h}}$ can be calculated by Eq.(E.4).

Putting the pertinent values of the various quantities in Eq.(E.4), we have

$$E_{\bar{h}} = \left[\left(\frac{13.238}{0.0393(86.66-80.416)}\right)^2 + \left(-\frac{650 \times 1.235 \times 10^{-4}}{(0.0393)^2\,(86.66-80.416)}\right)^2 + \left(\frac{-650 \times 2.88 \times 10^{-2}}{0.0393(86.66-80.416)^2}\right)^2 + \left(\frac{650 \times 5.773 \times 10^{-4}}{0.0393 \times (86.66 - 80.416)^2}\right)^2\right]^{1/2}$$

or

$$E_{\bar{h}} = 55.936$$

So $\bar{h} = 2649.27 \pm 55.936$

This uncertainty represents a typical value of the other experimental runs. Therefore, it could be said with some justification that the expected experimental uncertainties of the heat transfer coefficient for the experimental program reported here were about ± 10 per cent which is an acceptable level of error keeping in view the complexities in the boiling heat transfer process.

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