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Generalized CHF Prediction for Horizontal Tubes with Uniform Heat Flux

by

Y.L. Wong

Submitted to the Faculty of Engineering of the University of Ottawa in partial fulfilment of the requirements for the degree of

MASTER OF APPLIED SCIENCE

in Mechanical Engineering

June, 1988

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ABSTRACT

Accurate prediction of the Critical Heat Flux (CHF) or the heat flux associated with 'DNB' and 'Dryout' is of paramount importance in nuclear reactor design, reactor safety analysis and in the design of industrial boilers and heat exchangers involving two-phase flow. CHF in forced convective boiling was studied for both vertical and horizontal orientation. It is shown to be a complicated function of pressure, mass flux, hydraulic equivalent diameter and vapour quality. CHF due to phase separation and due to annular liquid film failure were analyzed. In horizontal, uniformly heated tubes, CHF was found to be lower than that for vertical upward flow under otherwise identical flow conditions. In light of the lack of a generalized prediction method for horizontal CHF, a semi-empirical correlation was developed and is capable of predicting both the upstream and downstream CHF for horizontal flow with uniform heat flux.

'Generalized Property Code' developed during the course of this study was found to be useful in predicting thermodynamic and transport properties of non-aqueous fluids and is suitable for thermalhydraulic analysis.
ACKNOWLEDGEMENT

The author wishes to express his appreciation and gratitude to Professor S.C. Cheng for his constant guidance and support. He is also greatly indebted to Dr. D.C. Groeneveld, adjunct professor, for his supervision, many invaluable suggestions and technical advice. The author would also like to thank B. Cheung who has been the major driving force for the completion of this thesis. The assistance given by Mr. C. Nguyen and L. Leung in the development of the computer code is greatly appreciated.

This study has been partially supported by the Atomic Energy of Canada Limited and the Atomic Energy Control Board.
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NOMENCLATURE

A  Flow area  
$B_o$  Bond number as defined in Eq. (2.19)  
$C_{HF}$  Critical heat flux  
$C_{HFv}$  Horizontal to vertical critical heat flux ratio  
$C_p$  Specific heat capacity at constant pressure  
$D$  Tube diameter  
$d_D$  Bubble departure diameter  
$d$  Average bubble or droplet diameter  
$F_B$  Buoyancy force acting on bubbles or droplets  
$F_D$  Drag force acting on bubbles or droplets  
$F_T$  Turbulent force acting on bubbles or droplets  
$F_r$  Froude number as defined in Eq. (5.5)  
$F_r^*$  Modified Froude number as defined in Eq. (6.4)  
$f$  Friction factor  
$G$  Mass flux  
$G_1$  Lower mass flux threshold as defined in Eq. (5.4)  
$G_2$  Upper mass flux threshold as defined in Eq. (5.2)  
$g$  Acceleration due to gravity  
$H_{fs}$  Enthalpy of vaporization  
$K_s$  Stratification factor for horizontal CHF  
$L_h$  Heated length of tube  
$M$  Molecular weight  
$P$  Pressure  
$p$  perimeter  
$P_c$  Critical pressure  

UNITS

$A$  $m^2$  
$B_o$  $kW.m^{-2}$  
$C_{HF}$  $J. kg^{-1}. C^{-1}$  
$D$  $m$  
$d_D$  $m$  
$d$  $m$  
$F_B$  $N$  
$F_D$  $N$  
$F_T$  $N$  
$F_r$  $-  
$F_r^*$  $-  
$f$  $-  
$G$  $kg.m^{-2}.s^{-1}$  
$G_1$  $kg.m^{-2}.s^{-1}$  
$G_2$  $kg.m^{-2}.s^{-1}$  
$g$  $m.s^{-2}$  
$H_{fs}$  $J.kg^{-1}$  
$K_s$  $-  
$L_h$  $m$  
$M$  $kg.mole^{-1}$  
$P$  $Pa$  
$p$  $m$  
$P_c$  $Pa$
\( Pe \) \quad \text{Peclet number, } GDC_p/\lambda_L \\
\( q \) \quad \text{Heat flux} \quad \text{[kW.m}^{-2}\text{]} \\
\( R \) \quad \text{Universal gas constant} \quad \text{[J.mole}^{-1}.K^{-1}\text{]} \\
\( Re_L \) \quad \text{Reynold number for liquid, } G D (1 - X)/\mu_f \\
\( T \) \quad \text{Temperature} \quad \text{[°C, K]} \\
\( T_1 \) \quad \text{Force balance parameter as defined in Eq. (5.18)} \\
\( T_2 \) \quad \text{Force balance parameter as defined in Eq. (5.20)} \\
\( T_3 \) \quad \text{Force balance parameter as defined in Eq. (5.27)} \\
\( T_4 \) \quad \text{Force balance parameter as defined in Eq. (5.32)} \\
\( T_5 \) \quad \text{Transit time ratio as defined in Eq. (5.40)} \\
\( T_6 \) \quad \text{Transit time ratio as defined in Eq. (5.48)} \\
\( T_c \) \quad \text{Critical temperature} \quad \text{[°C, K]} \\
\( T^+ \) \quad \text{Dimensionless parameter as defined in Eq. (5.1)} \\
\( U \) \quad \text{Phase velocity} \quad \text{[m.s}^{-1}\text{]} \\
\( u \) \quad \text{Phase velocity} \quad \text{[m.s}^{-1}\text{]} \\
\( V \) \quad \text{Molar volume} \quad \text{[m}^3.kymole}^{-1}\text{]} \\
\( V_c \) \quad \text{Critical molar volume} \quad \text{[m}^3.kymole}^{-1}\text{]} \\
\( v' \) \quad \text{Radial fluctuating velocity} \quad \text{[m.s}^{-1}\text{]} \\
\( X \) \quad \text{Vapour quality} \\
\( X_e \) \quad \text{Equilibrium quality} \\
\( X_{ed} \) \quad \text{Equilibrium quality at bubble departure} \\
\( X \) \quad \text{Lockhart-Martinelli parameter} \\
\( Y \) \quad \text{Ratio of vapour to liquid volumetric void fraction} \\
\( Z \) \quad \text{Compressibility factor, } PV/RT, \text{ Ohnesorge number} \\
\( Z_c \) \quad \text{Critical compressibility factor, } P_c V_c/RT_c \\

\textbf{Greek Symbols} \\
\( \alpha \) \quad \text{Void fraction}
\[ \beta \quad \text{Volumetric flow fraction} \]
\[ \beta_0 \quad \text{Contact angle} \]
\[ \Delta \quad \text{Change in} \]
\[ \Delta H_{in} \quad \text{Inlet subcooling} \]
\[ \lambda \quad \text{Thermal conductivity} \quad J.kg^{-1} \]
\[ \mu \quad \text{Viscosity} \quad W.m^{-1}.K^{-1} \]
\[ \theta \quad \text{Parameter relating surface tension to the normal boiling point of the substance} \]
\[ \rho \quad \text{Density} \quad kg.m^{-3} \]
\[ \sigma \quad \text{Surface tension} \quad N.m^{-1} \]
\[ \psi_H \quad \text{Merilo's CHF modeling parameter} \]
\[ \omega \quad \text{Acentric factor of the substance} \]

Subscripts

\[ c \quad \text{Critical} \]
\[ cr \quad \text{Critical} \]
\[ exp \quad \text{Experimental} \]
\[ f \quad \text{Saturated liquid} \]
\[ G \quad \text{Vapour phase} \]
\[ g \quad \text{Saturated vapour} \]
\[ hor \quad \text{Horizontal} \]
\[ i \quad \text{Interface} \]
\[ in \quad \text{Inlet} \]
\[ L \quad \text{Liquid phase} \]
\[ pred \quad \text{Predicted}^* \]
\[ r \quad \text{Reduced} \]
\[ ver \quad \text{Vertical} \]
Superscripts

- Dimensionless parameter
+ Dimensionless parameter
Chapter 1

INTRODUCTION

In most reactor and boiler systems, the energy transport involves various kinds of boiling heat transfer in two-phase flow. In two-phase heat transfer, boiling crisis is a very important phenomenon for the thermalhydraulic analysis. Figure 1 shows a typical boiling curve for two-phase flow. As can be seen from the figure, boiling crisis is one of the transition points between different heat transfer regimes.

Accurate prediction of the heat flux associated with the boiling crisis or critical heat flux (CHF) is of paramount importance in reactor design and reactor safety analysis. It is also of equal importance in devices operating under subcooled or non-boiling conditions such as in PWR. Boiling may occur under off design conditions or accident conditions in systems which are normally non-boiling and thus CHF may occur. CHF may also occur in other two-phase flow equipment such as conventional boiler systems, distillation and condensation equipment.

Over the years, hundreds of ad hoc correlations have been developed for the prediction of CHF in both vertical and horizontal tube flows. Most of them are for vertical flow and they are only applicable for a limited range of flow parameters. Recently, Groeneveld et al. (1986a) have published a generalized CHF table for vertical tube flow. This table is also applicable for non-aqueous fluids and for bundle geometries. Assessment of various published correlations undertaken by the ESDU (1986) shows this tabular method to be the best of the published methods. However, a simil-
proach for prediction of CHF in horizontal flow is not available. This is because of the
scarcity of experimental CHF data, and because the mechanisms controlling CHF in
horizontal flow are less understood than those for vertical flow.

This study is aimed at the development of a generalized semi-analytical prediction
of CHF for horizontal flow. Only the CHF for horizontal tubes with uniform heat flux
is considered in this study. Sources of experimental data are obtained from Merilo
(1977), Merilo and Ahmad (1979), Robertson (1973), Becker (1971), Leontiev et al.
(1981), Lis and Strickland (1970), Kohler and Hein (1986), Rounthwaite (1968), Ishigai
and Takagi (1970a) and Wang and Su (1980). Some of the data (those from Lis and
Strickland, Wang and Su and Ishigai and Takagi) were not used in the derivation of
the present correlation.

The mechanisms of CHF for both vertical and horizontal flow are discussed in
Chapter 2. Some CHF prediction methods available in the literature will also be
introduced. In Chapter 3, sources of experimental CHF data are outlined and the
experimental data is analysed to show the CHF trends with quality, mass velocity,
tube diameter and pressure. The experimental data was compiled and arranged in a
tabular format and is presented in Appendix A.

A generalized property code for non-aqueous fluids has also been developed for the
purpose of fluid property evaluations. The code will permit the evaluation of CHF
in horizontal flow for all transport properties for that fluid. The code predicts the
properties including specific volume, specific enthalpy, specific heat, viscosity, thermal
conductivity and surface tension. The development of this property code is presented
in Chapter 4 and the source code can be found in Appendix B.

In Chapter 5, analytical expressions have been derived considering the turbulent-
gravity force balance and bubbles or droplets transit time ratio. The dimensionless
parameters developed can be treated as measures of the degree of stratification for
two phase flow and will be used subsequently in the derivation of stratification factor
for horizontal CHF. The derivation of the stratification factor is presented in Chapter
6 with assessment technique and comparison to experimental data. The parametric
trends of the new CHF correlation have been summarized and shown to be consistent with the experimental trends.
Chapter 2

TWO-PHASE FLOW AND CRITICAL HEAT FLUX

2.1 Flow Regimes in Two-Phase Flow

In gas-liquid flow the two phases can distribute themselves in the conduit in a wide variety of ways. The phase distribution will affect the evaluation of two-phase thermalhydraulic quantities such as the void fraction, pressure drop and heat transfer in general. The phase distribution forms a number of interesting patterns called flow regimes. A particular flow regime is dependent on a number of flow parameters such as the mass flux ($G$), vapour quality ($X$), pressure ($P$) and the tube diameter ($D$). The flow regimes also vary with the tube orientation. The designation of the flow regimes is more of a subjective judgement than an objective measurement. In this thesis, only the commonly distinguishable flow regimes encountered in vertical and horizontal tubes will be discussed.

2.1.1 Flow Regimes in Vertical Flow

For vertical upward flow, there are basically five distinguishable flow regimes. They are the bubbly flow, slug flow, churn flow, wispy-annular flow and annular flow. The schematic diagram for the flow regimes encountered in vertical flow is shown in figure 2.
Bubbly Flow

In bubbly flow, the gas phase is distributed in the form of discrete bubbles completely entrained in the liquid stream. The bubbles are of approximately uniform size.

Slug Flow

In this flow regime, large bullet-shaped bubbles (slugs) are found and are separated from the tube wall by the descending film of liquid. Also, the successive vapour slugs are separated by the liquid slugs which may or may not contain smaller gas bubbles.

Churn Flow

Churn flow is very chaotic and turbulent. It is formed by the breakdown of the large vapour bubbles in slug flow (Collier, 1981). The main characteristic of the churn flow is the oscillatory or alternating direction of motion of the liquid.

Wispy-Annular Flow

Wispy-annular flow regime was first identified by Hewitt et al. (1970). In this flow regime, relatively thick liquid film flows along the tube wall together with a considerable amount of liquid entrained in the gas core. Small bubbles are found in the liquid film. This flow regime is found to occur with high mass velocities.

Annular Flow

Annular flow is characterized by continuity in the axial direction of the gas phase in the core. Similar to the wispy-annular flow, the liquid film is flowing on the tube wall with little bubble entrainment. Separated droplets are entrainment in the gas core in varying amount.
2.1.2 Flow Regimes in Horizontal Flow

For horizontal flow, the flow regimes are complicated by the asymmetry of the two-phase distribution due to the influence of gravity in the transverse direction to the flow. Figure 3 shows the generally recognised flow regimes.

Bubbly Flow

The flow pattern in this regime is similar to that in vertical flow except that the vapour phase (bubble) tends to flow along the top part of the tube. At moderate and high mass velocities, the vapour and liquid phases distribute themselves more symmetrically and hence the flow pattern resembles that in vertical flow.

Plug Flow

Plug flow is similar to the slug flow in the vertical direction except that the gas bubbles tend to travel in the upper half of the tube.

Stratified Flow

Stratified flow is also referred to as the stratified smooth flow. In this flow regime, liquid flows along the bottom of the tube. The gas-liquid interface is relatively smooth. This pattern occurs at very low liquid and vapour velocities.

Stratified Wavy (Wavy Flow)

As the vapour velocity increases, the interface of the stratified flow becomes disturbed by waves travelling in the direction of flow. The gas-liquid interface is wavy and thus the term 'wavy flow'.

Slug Flow

The waves at the interface may increase in magnitude due to further increase in the vapour velocity and finally touch the upper portion of the tube forming a complete liquid bridge. This type of flow regime is called the slug flow. The upper surface of
the tube behind the wave is wetted by a residual film which drains into the bulk of the liquid.

**Annular Flow**

Annular flow in horizontal tube is similar to that in vertical flow except that the liquid film is much thicker at the bottom of the tube than at the top. Also, the wavy structure is more pronounced at the bottom of the tube causing more liquid to be entrained into the vapour core and deposited near the top of the tube.

Each of the above mentioned flow regimes will affect the CHF mechanisms in both the vertical flow and the horizontal flow. The mechanisms of CHF will be discussed in the next section with some consideration of the influence of flow pattern.

### 2.2 CHF Mechanisms

The critical heat flux for horizontal flow may be significantly lower than that for vertical flow under otherwise identical flow conditions. Gravity, which tends to cause flow stratification in horizontal flow, is the main reason for the difference in horizontal and vertical CHF's. The mechanisms involved in each case are quite different except for flows with high mass velocities. An improved understanding of the mechanisms associated with critical heat flux will permit an improvement in current prediction methods and may lead to the development of devices and systems to avoid its occurrence in practical applications.

#### 2.2.1 CHF Mechanisms in Vertical Flow

For vertical channel flows, CHF can occur in subcooled boiling, in the low quality region and in the annular flow regime. The most commonly postulated mechanisms (Hewitt, 1982) in subcooled and low quality regions are

1. Near-wall bubble crowding and vapour blanketing
2. Local overheating following micro-layer evaporation from a nucleation site
3. Burnout associated with vapour clot or slug formation.

Figure 4 illustrates schematically the near-wall bubble crowding and vapour blanketing mechanism. A 'bubble boundary layer' builds up on the heated surface and is ultimately thick enough to prevent liquid migration to the surface, thus causing high temperature build up at the wall. The concept of boundary-layer separation for the burnout phenomenon was proposed by Kutateladze and Leontiev (1966). Tong (1972) modified the boundary-layer separation theory by pointing out that the friction factor for a channel with a bubble boundary layer is much higher than that for a smooth tube because the liquid has to flow over a rougher surface represented by the bubble boundary layer. He was able to derive a reasonably accurate semi-empirical correlation.

The second type of mechanism is shown schematically in figure 5. When a bubble grows at a nucleation site, 'micro-layer evaporation' occurs under the growing bubble, resulting in a disk of dry surface. The wall temperature at this dry patch rises rapidly and the patch can not be rewetted subsequent to bubble departure. The continually rising temperature spreads over the surface, thus giving rise to the burnout phenomenon.

Fiori and Bergles (1970) observed that CHF was often associated with the formation of vapour slugs/clots. They postulated that burnout occurs as a result of the evaporation of a thin liquid layer under the vapour slugs or clots, the vapour volumes preventing access of subcooled liquid to the surface (see figure 6). CHF in vertical annuli which occurred by the vapour-clot mechanism was studied by Rogers et al. (1982). Mechanisms in the subcooled and low-quality regions are reviewed by Tong and Hewitt (1972) and Bergles (1977).

Dryout in annular flow is most commonly associated with the drying of the liquid film by progressive entrainment, droplet deposition and evaporation. Dryout occurs as a result of the integral effect of droplet mass transfer, liquid entrainment and evaporation along the channel. Figure 7 illustrates the dryout mechanism in vertical annular flow in a tube. Other similar mechanisms of dryout in vertical annular flow were
discussed by Hewitt (1982) in more detail. A review of the CHF mechanisms on a uniformly heated surface was done by Katto (1986).

2.2.2 CHF Mechanisms in Horizontal flow

In horizontal channels, gravitational force acts on both the vapour phase and the liquid phase in the direction that is not parallel to the fluid stream, causing phase separation. In these circumstances, the mechanisms of burnout are much more complex than those for vertical flows.

For subcooled boiling, Merilo (1977) hypothesized that the migration of the lighter phase (bubbles) towards the top of the channel as a result of the transverse buoyancy force causes an uneven distribution of void. This maldistribution of void has a severe effect on heat transfer in general and CHF in particular.

The flow patterns for horizontal flow are different from those for vertical flow. Figure 8 shows schematically the transitions of flow patterns in horizontal flow. Intermittent dryout occurs in the low quality region where bubbles coalesce and form a continuous vapour cushion along the upper portion of the tube (Becker, 1971). As vaporization continues downstream, the increased stream velocity causes higher amplitude waves to form on the steam-water interface. The faster moving gas stream impinges on the waves and thus causes liquid to be entrained in the vapour core region. Some of the entrained liquid is being deposited onto the upper portion of the tube and finally, the water will again cover the full circumference of the tube and annular flow is established. The annular flow in horizontal channel is not axi-symmetric. The annular liquid film is much thinner at the tube top because of draining. Usually, the downstream dryout will occur at the top of the tube first.

Studies of CHF mechanisms in different quality regions in horizontal channels have been reviewed by Fisher et al. (1978). They have identified the CHF mechanisms in three different qualities regions:

1. Very low steam qualities
2. Low and intermediate steam qualities
3. High steam qualities.

At low steam qualities, bubbles that are produced at the wall could form ribbons of vapour along the upper surface of the channel under low flow conditions. These vapour ribbons act as barriers and inhibit the replenishment of liquid lost by draining and evaporation thus causing a premature occurrence of the CHF condition and a subsequent steep rise in the wall temperature.

The mechanisms of CHF at low and intermediate steam qualities are very different from those at very low steam qualities. For low and intermediate steam qualities, the flows in evaporating channels are characterized by alternating large splashing waves (surges) which carry the liquid to the upper surface of the channel. There is little or no droplet entrainment activities and thus no replenishment of the liquid film at the top of the channel. As a result, the liquid film at the upper surface is subjected to drainage and evaporation which will finally dryout if sufficient time is given before the next splashing wave touches the upper surface.

For high steam qualities, the flow patterns would most likely be annular: Figure 9 shows a horizontal annular flow in air and water, for typical liquid film and vapour velocities (Fisher et al., 1978). The annular liquid film at the top of the channel is thinner as a result of draining along the circumferential direction. At the bottom of the channel, larger amplitude waves give rise to a great deal of droplet entrainment into the vapour core. Little entrainment occurs at the top of the channel because here the liquid film is much smoother. When the heat flux reaches the critical value, the liquid film at the top of the channel will be completely evaporated and thus causing ‘dryout’. More details of the CHF mechanisms in horizontal annular flow will be provided in section 5.6.
2.3 Prediction of CHF in Vertical Tube Flow

2.3.1 Analytical Model

Dryout in annular flow is the most likely form to occur in practical systems, where the channel lengths are commonly several meters long. This type of dryout occurs as a result of liquid removal from the film leading to film dryout. An analytical model for dryout is possible if the rate of entrainment and deposition of droplets can be postulated. With the known local heat flux, the rate of evaporation can be calculated. The rate of change of liquid film flow rate with distance along the channel can be obtained through the mass balance equation, thus giving

\[
\frac{A}{p} \times \frac{dG_{LF}}{dz} = G_D - G_E - \frac{q}{h_{fg}}
\]  

(2.1)

where \(G_{LF}\) is the mass flux (averaged over the total cross-sectional channel area \(A\)) of that part of liquid flow which is in the film, \(z\) is the axial distance, \(p\) the channel periphery, \(G_D\) the rate of droplet deposition per unit channel surface area, \(G_E\) the rate of entrainment per unit channel surface area, \(q\) the local heat flux, and \(h_{fg}\) the latent heat. Equation (2.1) is for a steady state and it ignores the contribution to evaporation by other mechanisms such as liquid flashing as a result of the channel pressure gradient. More sophisticated expressions which take into account these factors were given by Whalley (1987).

A number of predictive methods have been proposed for the calculation of dryout in annular flow. The methods differ principally in their treatment of entrainment, deposition and evaporation processes and in the choice of boundary conditions. Perhaps, the most widely accepted model is that developed at Harwell (Whalley et al. 1974, 1975, 1978 and Hewitt and Hutchinson 1976). The Harwell model is rather general and superior to most empirical models. However, more work needs to be done in advancing understanding of entrainment and evaporation processes.
2.3.2 Empirical Correlations

Burnout mechanisms in vertical tubes are still not fully understood despite many years of intensive research. However, due to the great importance of burnout phenomena, many researchers have developed correlations for burnout. Some of them are quite successful but most of them are not. This is because most correlations were developed based on limited data sources and hence limited range of applications.

Most of the successful correlations have the general form of

\[
q_c = \frac{A' + B' \Delta h_{sub}}{C' + L_h}
\]

which was originally devised by Macbeth (1963). \( \Delta h_{sub} \) is the inlet subcooling, \( A' \), \( B' \) and \( C' \) are related to the system parameters and \( L_h \) is the tube length. Two most widely accepted correlations are Bowring's correlation (1972) and Biasi's correlation (1967).

Bowring's Correlation

Bowring's correlation for CHF in vertical up flow in round tubes gives \( q_c \) (W/m²) as follows:

\[
q_c = \frac{A' + 0.25DG\Delta h_{sub}}{C' + L_h}
\]

where \( D \) is the tube diameter (m), \( G \) is the mass flux (kg/m²s), \( \Delta h_{sub} \) is the inlet subcooling (J/kg). The tube length \( L_h \) is in meters. \( A' \) and \( C' \) are functions of \( G, D, \) pressure \( P \) (Pa), and latent heat \( h_{fg} \) (J/kg) as follows:

\[
A' = \frac{2.317(0.25h_{fg}DG)F_1}{1.0 + 0.0143F_2D^{1/2}G}
\]

\[
C' = \frac{0.077F_3DG}{1.0 + 0.347F_4(G/1356)^n}
\]

where \( n \) is given by

\[
n = 2.0 - 0.5P_R
\]
where $P_R$ is the ratio of the system pressure to a pressure of 6.895 MPa (equivalent to a pressure of 1000 psia). Thus

$$P_R = \frac{P}{(6.895 \times 10^6)}$$  \hspace{1cm} (2.4)

The functions $F_1$ to $F_4$ are also functions of $P_R$ as follows:

For $P_R < 1$:

$$F_1 = \frac{P_R^{18.647} e^{20.8(1-P_R)} + 0.917}{1.917}$$ \hspace{1cm} (2.5)

$$\frac{F_1}{F_2} = \frac{P_R^{1.316} e^{2.444(1-P_R)} + 0.309}{1.309}$$ \hspace{1cm} (2.6)

$$F_3 = \frac{P_R^{7.023} e^{16.658(1-P_R)} + 0.667}{1.667}$$ \hspace{1cm} (2.7)

$$\frac{F_4}{F_3} = P_R^{1.49}$$ \hspace{1cm} (2.8)

For $P_R > 1$:

$$F_1 = P_R^{-0.369} e^{-0.648(1-P_R)}$$ \hspace{1cm} (2.9)

$$\frac{F_1}{F_2} = P_R^{-0.448} e^{0.245(1-P_R)}$$ \hspace{1cm} (2.10)

$$F_3 = P_R^{0.219}$$ \hspace{1cm} (2.11)

$$\frac{F_4}{F_3} = P_R^{1.649}$$ \hspace{1cm} (2.12)

The ranges of parameters covered by Bowring’s correlation are as follows:

Pressure: 0.2-19.0 MPa
Tube diameter: 2-45 mm
Tube length: 0.15-3.7 m
Mass flux: 136-18,600 kg/m².s

Bowring reported that for the 3792 experimental points used, the RMS error of the correlation is 7% and the 95% confidence level is ±14%.
Biasi’s Correlation

Biasi’s correlation has the advantage of being continuous with respect to system pressure with apparently little or no loss of accuracy. His correlation gives $q_c (W/cm^2)$ as follows:

$$q_c = \frac{1.883 \times 10^3 f(P)}{D^n G^{1/6}} \left[ \frac{1}{G^{1/6}} - X \right]$$

(2.13)

for the low quality region

$$q_c = \frac{3.78 \times 10^3 h(P)}{D^n G^{0.6}} (1 - X)$$

(2.14)

for the high quality region

where $n = 0.4 \quad for \quad D \geq 1 cm$

$n = 0.6 \quad for \quad D < 1 cm$

and $f(P) = 0.7249 + 0.099 Pe^{-0.032P}$

$h(P) = -1.159 + 0.149 Pe^{-0.019P} + \frac{8.99P}{10+P^2}$

This correlation is evaluated in c.g.s units with the valid ranges as follows:

$0.3 cm < D < 3.75 cm$

$20 cm < L_h < 600 cm$

$2.7 bar < P < 140 bar$

$10 g/cm^2.s < G < 600 g/cm^2.s$

$\frac{1}{1+\rho_l/\rho_g} < X < 1$

For values of $G$ below $30 g/cm^2.s$ equation (2.14) is used whereas for other values of $G$, the critical heat flux is given as the greater of the two values obtained by equations (2.13) and (2.14). Based on the 4500 data points examined, the RMS error for this correlation was 7.26% and the 88% confidence level is ±10%.

Groeneveld and Snoek (1986) compared Biasi’s and Bowring’s correlations with the AECL tube CHF data bank (containing 10,000 CHF data points). The comparison shows Biasi’s correlation to be more promising. However, Biasi’s correlation has
significant shortcomings when compared to experimentally observed trends, e.g., increasing the flow always results in the decrease of CHF according to Biasi's correlation whereas the experimentally observed trend shows that CHF increases with an increase in mass velocity at low qualities. Biasi's correlation is not continuous with respect to quality which is another drawback and it can also predict negative CHF values. A correlations developed based on a particular set of experimental data points often result in a limited range of application and tends to lose the physical significance of the parameters involved in CHF mechanisms.

Since most empirical correlations and analytical models have a limited range of application, there exists a need for a general technique for predicting CHF in vertical tube flow.

2.3.3 Table Look-up Method

In 1977, the USSR Academy of Sciences produced a series of standard tables of critical heat flux. The CHF values are tabulated as a function of the local parameters $G$, $X$ and $P$ for an 8 mm diameter tube. For tube diameters other than 8 mm, the critical heat flux may be approximated by

$$q_c = q_{c,8mm} (8/D)^{1/2}$$  \hspace{1cm} (2.15)

for $4 \leq D \leq 16$ mm.

Similar work was continued at the University of Ottawa and Chalk River Nuclear Laboratories using a much larger data base (more than 15,000 tube CHF data). The completed table was published in 1986 (Groeneveld et al., 1986a). Groeneveld's table look-up method is valid over wide ranges of flow conditions. It is accurate, simple to use, and has the correct parametric and asymptotic trends. The CHF table for an 8 mm tube is shown in Table 1. CHF values for a tube having other than 8 mm diameter may be obtained from

$$q_c = q_{c,8mm} (8/D)^{1/3}$$  \hspace{1cm} (2.16)
for $4 \leq D \leq 16\text{mm}$.

The CHF table is also valid for bundle geometries with the appropriate correction factor. Thermal entrance effect is also accounted for in a form of length correction factor. The details of those descriptions are omitted in this work. A comparison of various CHF correlations and Groeneveld's table look-up technique against tube CHF data was made at ESDU (1986) and the results are summarized in Table 2. Bowring/Ahmad method listed in Table 2 is a combination of Bowring's correlation with Ahmad's (1973) fluid to fluid modelling parameter. Allowing for the differences in physical properties by using the Ahmad's modelling parameter, CHF for non-aqueous fluids was predicted from Bowring's correlation. The results show that the table method is superior to the other methods.

### 2.4 Prediction of CHF in Horizontal Tube Flow

#### 2.4.1 Analytical Model

Ishigai and Talagi (1970b) proposed a general theoretical analysis of burnout phenomena in horizontal annular flow. Their analysis involves the mass balance equation

$$\frac{\pi D^2}{4} \frac{dG_{LF}}{dz} = -(E_v + E_e + E_s - E_d)\pi D$$  \hspace{1cm} (2.17)

where $E_v$ is the evaporation rate due to heat addition, $E_e$ the rate of droplets entrained by evaporating steam flow, $E_s$ the rate of droplets released from the liquid film due to its instability and $E_d$ the rate of droplet deposition onto the liquid film.

The evaporation rate $E_v$ is evaluated as follows:

$$E_v = \frac{q}{h_{fs}}$$  \hspace{1cm} (2.18)

$E_e$ was assumed proportional to the average steam velocity generated by evaporation, $E_s$ assumed proportional to the growing velocity of a crest of an unstable wave and $E_d$ was assumed proportional to the radial velocity component $v_\theta$ of the turbulence and the droplets concentration in the gas core.
Their prediction agrees quite well with the experimental data from Ishigai (1970a) but tends to be inaccurate outside that particular range of tube diameters, flow rates and pressures. The calculation procedure for their method requires numerical integration and the appropriate choice of boundary conditions, which are usually not well defined. Their model is not suitable for a wide range of flow parameters.

2.4.2 Correlations

There are not too many ad hoc correlations for the predictions of CHF in horizontal tubes (Merilo, 1979, Wang and Su, 1980 and Yoshida et al., 1980). This is mainly because of the scarcity of horizontal CHF data.

Merilo’s (1979) correlation is one of the more acceptable for the prediction of CHF in horizontal tubes. He used the method of compensated distortion in deriving the horizontal CHF modelling parameter,

\[ \psi_H = Re[Z^3Bo]^{-1.05} \left( \frac{\mu_f}{\mu_g} \right)^{0.41} \]  

(2.10)

where

\[ Re = \frac{GD}{\mu_f} \]

\[ Z = \frac{\mu_f}{\sqrt{\sigma D \rho_f}} \]

\[ Bo = \frac{(\rho_f - \rho_g)gD^2}{\sigma} \]

His correlation is in the form of

\[ \frac{CHF}{GH_{f_g}} = f(L/D, X_{in}, \rho_f/\rho_g, \psi_H) \]

\[ = 575Re_L^{-0.340}(Z^3Bo)^{0.358} \left( \frac{\mu_f}{\mu_g} \right)^{-2.18} \left( \frac{L}{D} \right)^{-0.511} \left( \frac{\rho_f}{\rho_g} - 1 \right)^{1.27} \]

\[ (1 - X_{in})^{1.64} \]  

(2.20)

It correlates 462 Freon-12 and 143 water data points with RMS error of 9.12 and 9.07% respectively. The range of experimental parameters for their data base is shown in Table 3.
Atkinson and Rogers (1982) have shown that Merilo's modelling parameter can be applied to more complex geometries such as the 37-element nuclear fuel bundles. By using Merilo's modelling parameter, Atkinson and Rogers tried two forms of CHF equation for the bundle geometry:

\[
\frac{CHF}{GH_{fg}} = a_0(\rho_f/\rho_g)^{a1}(1 - X_{in})^{a2} \psi_H^{a3} \tag{2.21}
\]

and

\[
\frac{CHF}{GH_{fg}} = b_0 \psi_H^{b1}(\rho_f/\rho_g - 1)^{b2} + m_0 \psi_H^{m1}(\rho_f/\rho_g - 1)^{m2} X_{in} \tag{2.22}
\]

Equation (2.21) correlates 62 water and 67 Freon-12 data points with an overall RMS error of 4.6% while equation (2.22) correlates the same set of data with an overall RMS error of 4.4%. The applicability of Merilo's modelling parameter to bundle geometries is also partially confirmed by Ahmad et al. (1982).

Despite the applicability of Merilo's modelling parameters to the bundle geometries, his CHF correlation for horizontal tubes (equation 2.20) has some limitations. Some of the water data listed in Appendix A were used to test his correlation. As expected, the data from Merilo (1977) and Robertson (1973) agree very well with the predicted values but for data outside the range of the parameters listed in Table 3, his correlation does not predict the CHF values satisfactorily. Table 4 summarizes the results of the comparison. Most of the large errors are associated with low pressure conditions although some large errors occur also at higher pressure (e.g. Kohler and Hein's data). The ranges of \(G\), \(X\) and \(D\) for the various experimental data can be found in Table 6. Like any empirical correlation, Equation (2.20) should not be used outside the ranges of parameters specified in Table 3.

Since Equation (2.20) is empirical in nature and based on inlet conditions, it is not expected to be able to account for the possibility of having superheated steam at the exit. Table 5 shows the exit quality as calculated through the energy balance equation with the CHF value predicted from Merilo's correlation (for \(P = 1\) MPa, \(L/D = 200\)). For inlet qualities ranging from -50% to 0%, a large portion of the exit quality calculated showed that the steam is superheated at the exit in which case the CHF should
be zero. Also, the horizontal CHF predicted is much higher than the corresponding vertical CHF value. It would only predict a zero value for horizontal CHF if $X_{in} = 1$ and thus had ignored the possibility of flow stratification at low flow and negative inlet quality.

The overall RMS error for Merilo's correlation is found to be 80% when compared to the water data listed in Appendix A (data from Ishigai et al., Rounthwaite and Lis and Strickland were not included in the comparison). So, Equation (2.20) is not generally applicable over a wide range of flow parameters. Incidentally, the Merilo's scaling parameters (equation 2.19) may still be valid for correlating horizontal CHF data in tubes provided the coefficients and exponents are adjusted. For example, Merilo's scaling parameter could be used to determine the onset of flow stratification.

Both the Yoshida et al. and Wang and Su methods predict the dryout quality. Wang and Su's Correlation is valid for water while the equation of Yoshida et al. is valid for Freon-22. The correlation of Wang and Su has a simple form of

$$X_{do} = 8.81q^{-1/8}G^{-1/4}$$

Equation (2.23) is only valid for the following range of parameters:

- pressure 4.9 MPa
- heat flux 233 - 698 kW.m$^{-2}$
- mass velocity 500 - 1100 kg.m$^{-2}$.s$^{-1}$

It predicts most of their experimental data within the deviation of 10%. However, it is limited to the above range of flow conditions only. Prediction outside the specified ranges is disastrous. Up to now, there is no correlation that is accurate and generally applicable.

### 2.4.3 Groeneveld's Method

To determine the CHF for horizontal flow, Groeneveld (1982) suggested that the CHF for vertical flow be multiplied by a correction factor $K_6$, i.e.,

$$CHF_{hor} = K_6 \times CHF_{table}$$

(2.24)
where $CHF_{\text{table}}$ is the critical heat flux value obtained from the table look-up method (Groeneveld et al., 1986a). When flow is fully stratified, the CHF for horizontal flow is zero and hence $K_0 = 0$. On the other hand, as mass velocities in horizontal flow are high, effects of tube orientation on CHF become insignificant and hence $CHF_{\text{hor}}$ may be assumed equal to CHF for vertical flow ($CHF_{\text{ver}}$). The lower and higher mass flux thresholds ($G_1$ and $G_2$ respectively) can be calculated for any set of given flow conditions by using a modified flow regime map adapted from Taitel and Dukler (1976) (see figure 10). Curve $DD'$ represents the transition criterion below which stratification is noticeable and curve $A$ represents the transition criterion below which full stratification occurs. Taitel and Dukler's transition criteria will be discussed in detail in Chapter 5. The flow parameters required for the determination of the mass flux thresholds are the quality, $X$, pressure, $P$ and tube diameter $D$.

For flow conditions between the two mass flux thresholds, $K_8$ varies and depends on the flow parameters $G$, $X$, $P$ and $D$. A simplified expression for $K_8$ factor was suggested by Groeneveld et al. (1986a) which assumes a linear relationship between the $K_8$ factor and the mass flux. The expression is shown as follows,

$$K_8 = \frac{G - G_1}{G_2 - G_1}$$  \hspace{1cm} (2.25)

where

$G$ = given mass flux

$G_1$ = lower mass flux threshold where complete stratification occurs

$G_2$ = upper mass flux threshold corresponding to the onset of noticeable stratification.

This approach has the correct asymptotic trends at the two mass flux thresholds, i.e., when the flow is fully stratified ($G \leq G_1$), the CHF will be zero and when the mass velocity is high ($G \geq G_2$), the CHF for horizontal flow will be equal to that for vertical flow.
2.5 Present Approach

Taitel and Dukler's criterion for the onset of noticeable stratification is not sufficient for an accurate prediction of the CHF ratio though its generalized nature is worth considering. An investigation of other dimensionless parameters which might serve as a controlling parameter for CHF ratio was undertaken and some improvement in the CHF ratio prediction was observed. The strategy was to derive analytically a dimensionless parameter which carries the effects of turbulence, buoyancy and other interacting forces worth considering and has the correct parametric trends. The parameter is then tested against experimental data to derive a new correlation for the correction factor for horizontal CHF. The following sections outline the preparation steps necessary for the derivation of the new correlation. The approach is described in detail in Chapter 5.
Chapter 3

EXPERIMENTAL DATA

3.1 Data Sources

Table 6 presents a list of the sources of experimental data that are used for this study. The references are listed according to the group of fluids. For each source of experimental data, the ranges of pressures, mass flow rates, qualities, diameters and heat fluxes are shown to provide a better understanding of the experimental conditions.

The data bank was arranged in groups according to inlet condition (when possible), pressure ($P$), internal tube diameter ($D$) and heated length ($L_h$). They are further arranged in subgroups according to the mass flux and quality. Within each group of results, the following measured and calculated quantities are tabulated.

1) $G$ mass flux ($kg.m^{-2}.s^{-1}$)
2) $X_{cr}$ local critical quality
3) $CHF_{hor}$ critical heat flux for horizontal flow ($kW.m^{-2}$)
4) $CHF_{ver}$ critical heat flux for vertical flow ($kW.m^{-2}$)
5) CHFR critical heat flux ratio, $CHF_{hor}/CHF_{ver}$
6) I data index for $CHF_{ver}$
   1—experimental
   2—table look-up method

In the test of Rounthwaite (1968), dryout occurred twice. It first started at low
quality and ceased at a higher quality because the vapour velocity caused a change to annular flow. Later, the failure of annular flow caused dryout at the top of the tube to begin again. For these cases, two separate points are recorded (data of Routhwaite, 1968). The final data bank is listed in Appendix A.

3.2 Parametric Trends of CHF

Parametric trends of horizontal CHF must be identified and analyzed. Any CHF prediction method developed must be able to predict those trends as well as the CHF itself. If CHF is assumed to be dependent on local conditions only, then there are four basic parameters that can affect the CHF. They are the local quality, mass flux, tube diameter and pressure. The effects of each of these parameters on CHF and CHF ratio will be discussed.

3.2.1 Dryout Quality

CHF has a distinct dependency on dryout quality. At lower dryout quality, the CHF is higher. However, we are interested in the effect of dryout quality on CHF ratio ($CHF_{hor}/CHF_{ver}$).

Merilo (1977) noted the effect of quality on the relative CHF performance of vertical and horizontal geometries. He observed that for qualities greater than 40%, the vertical and horizontal CHF curves converge for increasing quality, whereas at less than 40% quality, and higher mass flux, the curves diverge for increasing quality (refer to figure 11). However, it must be noted that this trend may be partially due to the different mass velocities at the high and low qualities. Most of the other experimenters only observed the trend of CHF with quality rather than CHF ratio with quality. Merilo's observation should be considered when one tries to derive a dimensionless parameter that controls the CHF ratio.

At higher qualities, the vapour velocity will be higher, which then causes more axial drag and turbulence and a reduced effect of gravitation on the phase distribution. The
greater influence of turbulence over the buoyancy effect means that the effect of tube orientation is less significant and thus the CHF ratio increases. For subcooled or low-quality boiling (before the onset of net vapour generation), the flow pattern for horizontal flow is very similar to that for vertical flow. This is because the mechanisms governing the CHF are very similar to those for vertical flow. Some of the subcooled data obtained by Becker (1971) shows that the CHF ratio is well below the value of 1 (figure 12). This fact can be explained by the existence of void in subcooled boiling. The amount of vapour present is significant enough to cause some stratification even though the thermodynamic quality is less than zero. Therefore, the actual quality (or the actual amount of void present) is the real factor in the stratification of the flow. Also, Becker's data were obtained at low pressures for which the density difference is high and thus stratification is more likely.

Before bubble departure occurs, the bubbles remain attached to the wall and therefore the flow pattern for horizontal flow is similar to that for vertical flow. The parametric trend is then that the CHF ratio starts at 1.0 initially for zero actual quality (or onset of nucleate boiling) and stays very close to 1.0 until at a certain point when bubbles start to detach from the wall. After bubble departure, gravitational forces start to take effect and cause stratification to occur in low quality flows. From that point on, the CHF ratio decreases sharply with quality until the quality is high enough to cause some turbulence, thus suppressing the stratification effect. Thereupon the CHF ratio increases with increasing dryout quality and may eventually approach 1.0 again. Figure 13 illustrates the variation of CHF with quality for both vertical and horizontal tube flows. It is clear from this figure that the CHF values for both vertical and horizontal flow are very similar in the subcooled and low quality regions (or the CHF ratio has a value close to 1.0). As the quality increases, the CHF for horizontal flow decreases more than $CHF_{ver}$ because of void migration (CHF ratio decreases). Finally, at higher quality when annular flow occurs, void migration is suppressed because of higher $G$ and the CHF for horizontal flow approaches that for vertical flow (CHF ratio approaches 1.0). It is important that the actual quality be used in the analysis of
stratification criteria. At high qualities, the thermodynamic quality is approximately equal to the actual quality.

3.2.2 Mass Velocity

As explained earlier, higher mass velocities will prevent stratification and induce more mixing between the liquid and vapour phases. Consequently, a more symmetrical flow distribution will result. Therefore, the CHF ratio increases with increasing mass velocity. Figure 14 shows the effects of mass velocity on CHF for horizontal flow. However, linear relationship does not seem to be a correct representation of CHF ratio variation with mass velocity. Studies to include the non-linear relationship will be discussed in the later sections. All experimenters agree that as mass velocity increases, the CHF ratio approaches 1.0 which means that the CHF for horizontal flow is close to the counterpart for vertical flow. It is always safe to say that the CHF ratio increases with mass velocity.

3.2.3 Tube Diameter

As the tube diameter increases, there is more tendency for stratification to occur because the liquid has to move through a larger distance before reaching the top part of the tube. Therefore, the CHF ratio decreases with an increase in tube diameter.

Becker (1971) and Robertson (1973) observed that for a larger diameter tube, dryout occurs at lower quality range and that a higher mass flux threshold \( G_2 \) is required in order to avoid the separation of the phases. This implies that CHF ratio decreases with increase in tube diameter. Merilo (1979) also studied the effect of tube diameter on CHF for both vertical and horizontal flow. He found a similar relationship between tube diameter and CHF as those described by Becker and Robertson. Figure 15 shows the effect of diameter on horizontal CHF for typical flow conditions.
3.2.4 Pressure

Pressure has some effects on the CHF ratio variation because it affects the buoyancy force acting on the two phases. Also, the latent heat of vaporization changes with pressure thus causing a change in the amount of heat required to evaporate the liquid film (CHF). With an increase in pressure, the density difference decreases and the buoyancy force will become less pronounced. As a result, one should expect the CHF ratio to approach 1.0 at near-critical pressure. At the same time, the density ratio \(\rho_1/\rho_2\) decreases with increasing pressure and results in a lower mixture velocity. So, less turbulence occurs and stratification seems likely to happen.

Becker (1971) observed that for high heat fluxes and high mass velocities, where downstream dryout occurred, the dryout quality increases with increasing pressure. However, for low heat fluxes and low mass velocities, the dryout quality decreases with increasing pressure. For the former case where mass velocities are high, the latent heat of vaporization decreases with an increase of pressure and therefore more vapour will be generated (thus higher quality) causing a more turbulent environment which suppresses dryout. For the latter case, where heat fluxes are low, an increase in pressure decreases the vapour velocity which decreases the mixing between the phases and causing a vapour cushion to exist at a higher mass velocity and at a lower steam quality. The effect of pressure on CHF is illustrated in figure 16.

3.2.5 Other Geometric Effects

Geometric effects such as bends in the piping just upstream of the test-section inlet can affect the phase distribution to a certain degree. The effects of bends on two-phase flow are not well understood at present. Robertson (1973) and Lis and Strickland (1970) had performed experiments with upstream bends and they found that the bend will generally tend to separate the two phases very effectively. The heavier liquid phase will collect on the outside surface of the bend and the lighter gaseous phase on the inside. The stratification effect might be suppressed or enhanced depending on
the direction of the bend. Spacers and flow obstructions also have some effects on the phase distribution. The two phases will be better mixed after passing through a spacer or flow obstruction. This is due to the generation of wakes or eddies downstream of the flow obstruction which in turn causes more turbulence in the flow. Therefore, the CHF will be enhanced if the flow encounters spacers or similar type of flow obstructions.

Among other factors that might affect the CHF is the presence of a compressible volume upstream of the test section such as a two-phase inlet condition.
Chapter 4

EVALUATION OF THERMODYNAMIC AND TRANSPORT PROPERTIES AND THERMALHYDRAULIC PARAMETERS

4.1 General

In this chapter, property codes for evaluation of fluid properties will be discussed. The property codes are capable of evaluating properties of water as well as those of non-aqueous fluids. Void fraction model, determination of actual quality and other preliminary background knowledge are reviewed and examined in this chapter.

4.2 Fluid Properties Evaluation

In all thermalhydraulic analysis, particularly in two-phase flow heat transfer, fluid properties must be known. The most common fluid media used in nuclear reactors
and commercial heat exchangers are water and refrigerants. It is important to predict accurately the thermodynamic and transport properties of these fluids. Many correlations and property prediction packages have been developed for light water and very few for heavy water. Most of them are commercial packages and proprietary software. A simple and accurate water property prediction software (UOCODE) has been developed by Cheng et al. (1987) at the University of Ottawa.

For non-aqueous fluids, no general method is available in the published literature. As a part of this thesis, a study to develop a general predictive method for property evaluation for non-aqueous, non-polar fluids was undertaken and completed. A semi-analytical predictive method was developed for non-polar fluids including the refrigerant family and methane family.

4.2.1 Properties for Water

The UOCODE developed by Cheng et al. (1987) uses simple correlations for the evaluations of water properties. It was developed for micro-computer applications. The required input parameters are the pressure and temperature. Properties evaluated are the specific volume, density, enthalpy, specific heat, thermal conductivity, viscosity and surface tension and latent heat of vaporization at the given pressure.

The range of validity of the UOCODE essentially covers all thermodynamic states from the triple point to critical point of water, i.e

\[ 0.01^\circ C < T < 800^\circ C \]
\[ 0.6113kPa < P < 22.09MPa \]

The authors reported a root mean square error of less than 3% for all properties except at near critical condition. Recently, the UOCODE was updated to include the prediction of properties for heavy water and similar accuracy was reported.
4.2.2 Properties for Non-aqueous Fluids

A generalized property code was developed for the evaluation of thermodynamic and transport properties for non-aqueous fluids. This property code makes use of a four parameter equation of state developed by Adachi et al. (1983). Thermodynamic properties of the fluids were obtained through the derivatives of this equation of state while the transport properties were evaluated through the use of empirical correlations.

Equation of State

Adachi et al.'s equation of state was developed for non-polar substances and is thus valid for fluids such as Freons and light hydrocarbons. The equation of state is given by

\[ P = \frac{RT}{V - b_1} \left( \frac{a(T)}{(V - b_2)(V + b_3)} \right) \]  

(4.1)

It is in cubic form with one temperature-dependent parameter \( a(T) \) and three temperature-independent parameters \( b_1, b_2 \) and \( b_3 \). The parameters can be expressed as follows:

\[ a(T) = \frac{AR^2T^6}{P_c}[1 + \alpha (1 - T^2)]^2 \]  

(4.2)

\[ A = 0.44869 + 0.04024\omega + 0.01111\omega^2 - 0.00576\omega^3 \]  

(4.3)

\[ \alpha = 0.4070 + 1.3787\omega - 0.2933\omega^2 \]  

(4.4)

and

\[ b_i = \frac{B_iRT_c}{P_c}, i = 1, 2, 3 \]  

(4.5)

where

\[ B_1 = 0.08974 - 0.03452\omega + 0.00330\omega^2 \]  

(4.6)

\[ B_2 = 0.03686 + 0.00405\omega - 0.01073\omega^2 + 0.00157\omega^3 \]  

(4.7)

\[ B_3 = 0.15400 + 0.14122\omega - 0.00272\omega^2 - 0.00484\omega^3 \]  

(4.8)

\( R \) is the universal gas constant, \( \omega \) is the acentric factor of the substance which may be found from Reid et al. (1977) and \( T_r \) is the reduced temperature. This equation
of state is valid for gases and includes saturated liquid states. Derivations of thermodynamic properties from this equation are briefly discussed in the following sections. The equations used for the evaluation of each fluid property are presented in Appendix B together with the source code for the property code.

Specific Volume

With $T$ and $P$ as input parameters, the molar volume can be estimated using Newton-Raphson's iteration technique (with the equation of state). To obtain the specific volume, the molar volume is divided by the molecular weight of the fluid.

Enthalpy

Enthalpy is a relative quantity that has no absolute value and it is always evaluated with respect to a reference value. The relative enthalpy concept is shown in figure 17. If the enthalpy difference between two states $P_1, T_1$ and $P_2, T_2$ is desired, there are an infinite number of calculational paths, all of which give the same numerical results. Ideal gas heat capacity at zero pressure ($C_p^0$) for various gases are readily available. A convenient path of calculating the enthalpy difference is $AQCR$ in figure 17. Therefore, the enthalpy differences between two thermodynamic states using the path $AQCR$ can be expressed as

$$H_2 - H_1 = (H^0 - H_{P_1})T_1 + \int_{T_1}^{T_2} C_p^0 dT - (H^0 - H_{P_2})T_2$$

$$= (H_{P_2} - H^0)T_2 - (H_{P_1} - H^0)T_1 + \int_{T_1}^{T_2} C_p^0 dT$$  \hspace{1cm} (4.9)

where

$$(H_{P_2} - H^0)_{T_2} = \text{Enthalpy departure function (it relates}}$$

the enthalpy at $P_2, T_2$ to a reference

state $P = 0$ at the same temperature

If one chooses $P_1$ and $T_1$ to be the reference pressure and temperature at which the enthalpy is set equal to zero, then the enthalpy at any state $P$ and $T$ can be expressed
as
\[ H = (H_P - H^0)_T - (H_{P_{ref}} - H^0)_{T_{ref}} + \int_{T_{ref}}^T C^0 PdT \]  

(4.10)

From the equation of state, the departure function can be derived as
\[ \frac{(H - H^0)}{RT_c} = (Z - 1)T_r + \frac{1}{B_2 + B_3}[A(T_r) - T_r \frac{dA(T_r)}{dT_r}] ln \frac{Z - B_2 P_r/T_r}{Z + B_3 P_r/T_r} \]
or
\[ (H - H^0)_T = RT(Z - 1) + \frac{1}{b_2 + b_3}[a(T) - T \frac{da(T)}{dT}] ln \frac{V - b_2}{V + b_3} \]  

(4.11)

where
\[ A(T_r) = A[1 + \alpha(1 - T_r^{1/2})]^2 \]  

(4.12)

The ideal gas specific heat at zero pressure for most substances can be expressed in terms of a third order polynomial function of temperature
\[ C^0_P = A + BT + CT^2 + DT^3 \]  

(4.13)

The constants A, B, C and D are obtained experimentally and a list of substances with these constants is provided by Reid et al. (1977) and other engineering handbooks.

Specific Heat Capacity at Constant Pressure \((C_P)\)

Specific heat capacity at constant pressure (or specific heat for short) is defined as the rate of change of enthalpy with respect to temperature at constant pressure. Therefore
\[ C_P - C^0_P = \frac{\partial (H - H^0)}{\partial T} \bigg|_P \]  

(4.14)

Differenting equation (4.11) w.r.t \(T\) at constant \(P\) gives
\[ C_P - C^0_P = - \frac{1}{b_2 + b_3} \frac{RT d^2 a(T)}{dT^2} ln \frac{V - b_2}{V + b_3} - R \frac{\partial V}{\partial T} \bigg|_P \left[ \frac{RT}{V - b_1} - T \frac{da(T)}{dT} \frac{1}{(V - b_2)(V + b_3)} \right] \]  

(4.15)

and
\[ \frac{\partial V}{\partial T} \bigg|_P = - \frac{(\partial P/\partial T)_V}{(\partial P/\partial V)_T} \]  

(4.16)
$(\partial P/\partial T)_V$ and $(\partial P/\partial V)_T$ can be derived from the equation of state directly. Equation (4.16) is the famous Maxwell's Law.

Three other thermophysical properties (viscosity, thermal conductivity and surface tension) are also evaluated. The correlations used are discussed below.

**Viscosity**

Viscosity can be seen as a measure of the internal fluid friction which tends to oppose any dynamic change in the fluid motion. The mechanism or theory of gas viscosity has been reasonably well clarified by the application of kinetic theory of gases but not much is understood for the theory of liquid viscosity. This section outlines the viscosity estimation techniques for both non-polar gases and liquids.

**Low-Pressure Gas Viscosity**

For low-pressure gas viscosity estimation, either the kinetic theory or the law of corresponding states approach can be used. The theoretical method requires the knowledge of collision diameter and intermolecular potential function, which are not readily available, whereas the latter approach only requires the knowledge of the critical properties and molecular weight of the substance. The corresponding states approach is simpler to use and the constants can be easily found in the open literature. Both methods give approximately the same accuracy (Reid et al., 1977). Since the corresponding states approach is easier to use, it was chosen for the estimation of low pressure viscosity for non-polar gases. The exact equation used is the Thodos et al. equation (1970)

$$
\mu^0_\xi = 4.610T_r^{0.618} - 2.04e^{-0.4437r} + 1.94e^{-1.0537r} + 0.1
$$

(4.17)

where

$$
\xi = T_c^{-1/6}M^{-1/2}P_c^{-2/3}
$$

$T_c = \text{Critical temperature, } K$

$P_c = \text{Critical pressure, atm}$
\( M = \text{Molecular weight} \)

\( \mu^0 = \text{Low-pressure gas viscosity, micro-poise} \)

The expected error for this equation is normally between 1 and 3 percent for non-polar gases. It should not be used for hydrogen, helium, diatomic halogen gases and polar gases.

**Dense-Gas Viscosity**

For non-polar gases, the most accurate prediction method for dense gas viscosity yet suggested is the Jossi et al. (1962) equation

\[
[(\mu - \mu^0)\xi + 1]^{0.25} = 1.0230 + 0.23364\rho_r + 0.58533\rho_r^2 - 0.40758\rho_r^3 + 0.093324\rho_r^4
\]

(4.18)

where \( \rho_r = \text{reduced density} \)

\( \mu = \text{dense-gas viscosity, micro-poise} \)

The relation is valid for \( 0.1 \leq \rho_r \leq 3 \). The exclusion of any specific temperature effect has been criticized by Rogers and Brickwedde (quoted by Reid et al., 1977). However, most ethane data reported by other researchers show no such temperature dependency. Despite its approximate nature, this correlation is reasonably accurate and simple to use. The agreement between experimental and estimated dense gas viscosities is reasonable. The error normally does not exceed 10 percent.

**Liquid Viscosity**

For estimation of saturated liquid viscosity (which is hereby referred to as liquid viscosity), group contribution methods are usually more accurate than corresponding states methods or other correlations. However, they are usually very complex and detailed chemistry of the substance must be known before any calculation can be made. A simple correlation for estimating low temperature liquid viscosity was suggested by Van Velzen et al. (1972). They assume that \( \log \mu_L \) is linear in \( \frac{1}{T} \). Their correlation is
shown as

$$\log \mu_L = B \left( \frac{1}{T} - \frac{1}{T^0} \right)$$ (4.19)

where $\mu_L$ = liquid viscosity, cP
$T$ = temperature, K

$B$ and $T^0$ are constants for each specific substance. A list of these constants for different substances was tabulated by Reid et al. (1977).

This relation is only valid for $T_r \leq 0.75$. For saturated liquid at higher temperatures, corresponding-states relations can be used. Letsou and Stiel (1973) suggested the correlation

$$\mu_L \xi = (\mu_L \xi)^{(0)} + (\mu_L \xi)^{(1)}$$ (4.20)

where

$$(\mu_L \xi)^{(0)} = 0.015174 - 0.02135T_r + 0.0075T_r^2$$ (4.21)

$$(\mu_L \xi)^{(1)} = 0.042552 - 0.07674T_r + 0.03407T_r^2$$ (4.22)

and $\mu_L$ is in centipoise and $\xi$ is defined in the same way as before. This relation is only valid for $0.76 \leq T_r \leq 0.98$. Generally, an average error of about 3 percent was reported by the authors for $T_r$ up to 0.92. Larger errors were found as the critical point was approached. Their method represents the best available for estimating saturated-liquid viscosities at temperature of $T_r \geq 0.76$.

**Thermal Conductivity**

Estimation of thermal conductivity for liquid or gas is similar to that of viscosity. Group contribution methods are generally the most accurate but not easy to apply because of the requirement of complex chemistry of the substance. The estimation of thermal conductivity is subdivided into three distinct thermodynamic regions, namely the low density gases, dense gases and liquid regions.

**Low Density Gases Thermal conductivity**
A simple and accurate method for the prediction of thermal conductivity for low density gases is the well-known Eucken correlation for polyatomic gases

$$\frac{\lambda^0 M}{\mu^0} = C_v + 4.47$$

(4.23)

This simple Eucken relation is surprisingly good and usually more accurate than some other modified Eucken forms. Errors are usually less than 10 percent. Other correlations like the Roy-Thodos (1968,1970) and Bromley (1952) methods are very complex and they are only slightly more accurate than the simple Eucken form. Stiel and Thodos (1964) suggested a similar form to Eucken's equation and it generally predict a slightly higher value than the Eucken form. For higher density gases, simple kinetic theory would not be sufficient to account for the interactive forces between the molecules.

Dense Gases Thermal conductivity

Stiel and Thodos suggested the following correlation for the dense gas thermal conductivity estimation.

$$\begin{align*}
(\lambda - \lambda^0)\Gamma Z_c^5 &= 14.0 \times 10^{-8}(e^{0.535\rho_r} - 1) & \rho_r < 0.5 \\
(\lambda - \lambda^0)\Gamma Z_c^5 &= 13.1 \times 10^{-8}(e^{0.67\rho_r} - 1.069) & 0.5 < \rho_r < 2.0 \\
(\lambda - \lambda^0)\Gamma Z_c^5 &= 2.976 \times 10^{-8}(e^{1.155\rho_r} + 2.016) & 2.0 < \rho_r < 2.8
\end{align*}$$

(4.24)

where $\Gamma = T_e^{1/6}M^{1/2}P_e^{-2/3}$ and $Z_c$ is the critical compressibility factor of the substance.

This relation should not be used for polar substances or for hydrogen or helium. According to Reid et al. (1977), equation (4.24) appears to be the best generalized correlation.

Liquid Thermal conductivity

The thermal conductivity of liquid is quite different from that of gases. Reid et al. (1977) combined Sato's (1973) suggestion with Riedel's (1951) correlation and derived a new equation for thermal conductivity for a number of non-polar fluids in terms of
their respective normal boiling point. Their correlation has the following form

\[
\lambda_L = \frac{2.64 \times 10^{-3}}{M} \left[ \frac{3 + 20(1 - T_r)^{2/3}}{3 + 20(1 - T_{rb})^{2/3}} \right]^{2/3}
\]  
(4.25)

where \(T_{rb}\) is the normal boiling temperature. Thermal conductivity evaluated from equation (4.25) has a unit of \(cm^{-1}.s^{-1}.K^{-1}\). For \(T_r > 0.8\) the equation for high density gas thermal conductivity (eq. 4.24) can be used.

Surface Tension

Brock and Bird (1955) suggested an equation for the surface tension evaluation and was later modified by Reid et al. (1977). The modified equation relates the surface tension to the normal boiling temperature of the substance. Their relation is given in equation (4.26)

\[
\sigma = P_c^{2/3} T_c^{1/3} \theta (1 - T_r)^{11/9}
\]  
(4.26)

where \(\theta = 0.1207 \left[ \frac{1 + T_{rb} \ln P_c}{1 - T_{rb}} \right] - 0.281\)

\(P_c\) is in atm. and \(T_c\) in Kelvin. Prediction from the above equation were compared with experimental data and the authors reported an average error of less than 5%.

Detailed description of the non-aqueous fluid property code is presented in Appendix B. The property code was tested against experimental data for a number of commonly-used fluids. Generally, very good agreement was obtained between the predicted and the experimental property values.

4.3 Determination of Thermalhydraulic Parameters

A number of thermalhydraulic parameters must be known before any analysis of stratification criteria can be made. Quality and void fraction are found commonly in most of the thermalhydraulic analyses. The onset of bubble departure is important in the determination of CHF relation between horizontal and vertical flows. Bubble sizes are also needed when calculating drag forces and turbulent forces on bubbles. The following sections briefly discuss the prediction methods for the above mentioned parameters.
4.3.1 Onset of Significant Void

Onset of significant void is sometimes known as the point of net vapour generation (figure 18). It is the point where the subcooling is low enough that bubbles grow and depart into the stream. Saha and Zuber (1974) proposed a correlation based upon the combined mechanisms of hydrodynamic criteria for bubble detachment and heat transfer restraint. The equilibrium quality at bubble departure was correlated as

\[
X_{ed} = -0.0022 \frac{q_w D_h C_{PL}}{\lambda_L h_f g} \quad \text{for} \quad Pe < 70000 \quad (4.27)
\]

\[
X_{ed} = -154 \frac{q_w}{G h_f g} \quad \text{for} \quad Pe > 70000 \quad (4.28)
\]

where \( \lambda_L \) is the thermal conductivity of liquid, \( C_{PL} \) the specific heat of liquid and \( Pe \) is the Peclet number which is the product of \( Re_L \) and \( Pr_L \). A large body of data was used by Saha and Zuber in deriving this correlation. The data base used includes water with

\[
P = 0.1 - 13.8 MPa
\]

\[
G = 95 - 2760 kg/m^2.s
\]

\[
q_w = 0.28 - 1.89 MW/m^2
\]

in rectangular channels and annular and circular tubes. Data for Freon 114 were also used with

\[
P = 0.32 - 0.85 MPa
\]

\[
G = 101 - 2073 kg/m^2.s
\]

\[
q_w = 0.0063 - 5.36 MW/m^2
\]

in an annular channel.

Their correlation was chosen for the present analysis because of its simplicity and the data base used in their derivation covers a wide range of flow parameters. However, a recent study by Rogers et al. (1987) have shown that Saha-Zuber's correlation does not apply at low flow (\( G < 500 kg.m^{-2}.s^{-1} \)) and low pressure (\( P \approx 150 kPa \)) conditions.
4.3.2 Actual Quality

Usually, experimental data involving critical heat flux will contain information such as thermodynamic quality. In deriving criteria for stratification, actual quality is desired. For high quality regions, equilibrium quality \( X_e \) (or thermodynamic quality) is approximately equal to the actual quality but a similar argument can not be made in low quality or subcooled boiling regions. Estimation of actual quality is essential in low quality and subcooled boiling. Levy (1967) proposed a correlation for actual quality based on the onset of bubble departure. His equation is in the exponential form of

\[
X = X_e - X_{ed} e^{X_e/X_{ed}-1}
\]  
(4.29)

where \( X_{ed} \) is the equilibrium quality at onset of bubble departure. Saha and Zuber (1974) suggested a similar equation in the form of

\[
X = \frac{X_e - X_{ed} e^{X_e/X_{ed}-1}}{1 - X_{ed} e^{X_e/X_{ed}-1}}
\]  
(4.30)

Equations (4.27) and (4.28) can be used in the evaluation of the \( X_{ed} \). Both equations (4.29) and (4.30) were developed based on vertical tube data and have a restriction that \( X_{ed} \leq X_e \).

4.3.3 Void Fraction

In the analysis of the turbulent/buoyant force relation in two phase flow, the fraction of volume occupied by vapour in a cross section (or the void fraction, \( \alpha \)) must be known. Void fraction is related to the quality in the following fashion

\[
\frac{X}{1-X} = \frac{\alpha}{1-\alpha} \frac{S \rho_v}{\rho_i}
\]  
(4.31)

from which

\[
\alpha = \frac{X}{X + S(1-X) \frac{\rho_v}{\rho_i}}
\]  
(4.32)

where \( S \) is the ratio of the average phase velocity of the vapour to that of the liquid. This velocity ratio is usually referred to as the slip ratio.
There are a number of correlations available for the estimation of slip ratio or void fraction. Some of the more successful correlations are those of CISE (Premoli et al. 1970), Chisholm (1983), Ahmad (1970) and some proprietary correlations.

Most void fraction correlations are actually correlations of the slip ratio, which is found experimentally to depend on parameters such as \( \frac{\rho_f}{\rho_g} \), quality, mass flux and other minor variables such as tube diameter and inclination of the tube. Table 7 summarizes the dependence of slip ratio on these variables for some common correlations.

The performance of various correlations has been assessed in comparison with a wide range of data (Whalley, 1987). It was found that the most accurate generally applicable correlation is the CISE correlation. It has the following form

\[
S = 1 + E_1 \left[ \frac{Y}{1 + E_2 Y} - E_2 Y \right]^{1/2}
\]  

(4.33)

where

\[
E_1 = 1.578 Re^{-0.19} \left( \frac{\rho_f}{\rho_g} \right)^{0.22}
\]

\[
E_2 = 0.0273 We Re^{-0.51} \left( \frac{\rho_f}{\rho_g} \right)^{-0.08}
\]

\[
We = \frac{G^2 D}{\sigma \rho_f}
\]

\[
Re = \frac{GD}{\mu_f}
\]

\[
Y = \frac{\beta}{1 - \beta}
\]

\[
\beta = \frac{\rho_f X}{\rho_f X + \rho_g (1 - X)}
\]

Equation (4.33) was essentially derived for vertical flow. It would not be expected to give good results at low velocities and pressures in vertical flows. The validity of the CISE correlation under subcooled condition has not been tested. Zuber and Findlay’s (1965) equation for void fraction may be used for subcooled condition. Their equation has the form:

\[
\alpha = \frac{X \rho_f}{C_0 [X \rho_f + (1 - X) \rho_g] + \rho_g \eta \frac{\mu}{G}}
\]  

(4.34)
where

\[
\rho = \beta [1 + \left( \frac{1 - \beta}{\beta} \right)^b]
\]

\[b = \left( \frac{\rho_g}{\rho_f} \right)^{0.1}\]

\[U_{GU} = 2.9 \left[ \frac{(\rho_f - \rho_g) \sigma g}{\rho_f^2} \right]^{0.25}\]

In summary, the recommended procedure for the calculation for void fraction is to calculate the actual quality from equation (4.30) and then use equations (4.32) and (4.33) for the bulk boiling condition and equation (4.34) for the subcooled condition.

4.3.4 Bubble size

Accurate prediction of bubble size has always been a difficult task. Bubble size is closely related to the bubble departure diameter. Ever since Fritz (1935) first proposed his correlation for bubble departure diameter, i.e.

\[d_D = 0.0208 \beta_0 \sqrt{\frac{g \sigma}{\rho_f - \rho_g}}\]  \hspace{1cm} (4.35)

where \(\beta_0\) is the contact angle between the bubble and the wall surface, many other researchers have subsequently developed new correlations. Some are not very accurate while the others are too complicated. However, it must be noted that equation (4.35) is only valid for static conditions. For flow boiling, the mechanisms governing the bubble departure involves dynamic condition such as the drag force. The actual bubble departure size for flow boiling is much smaller than that for static conditions. It is not reasonable to assume that the average bubble size is the same as the bubble departure diameter.

Once the bubble is in the flowing stream, its size is determined by the balance of forces which are functions of the fluid properties and flow velocities. Average bubble size in the flowing stream can be related to the Weber number, defined as

\[We = \frac{\rho_g(u_g - u_f)^2 d}{\sigma}\]  \hspace{1cm} (4.36)
At a certain critical Weber number, the bubbles will break up, thus resulting in a maximum diameter that a bubble can attain. By assuming the average size to be proportional to the critical Weber number, equation (4.36) can be rearranged to give

\[ d = \frac{We_{crit} \sigma}{\rho_s (u_s - u_f)^2} \]

\[ = \text{Constant} \times \frac{\sigma}{\rho_s (u_s - u_f)^2} \]  \hspace{1cm} (4.37)
Chapter 5

DERIVATION OF FLOW STRATIFICATION CRITERIA

5.1 General

It is important to know the parameters controlling the onset of flow stratification. Parameters and criteria proposed by previous authors are discussed in this chapter. In addition, new dimensionless parameters are derived based on a force balance and transit time ratio. These parameters can be used as measures of the degree of stratification. They have correct parametric trends with pressure, mass velocity, quality and tube diameter. In general, attempts are made to determine the ratio of the inertia force (or turbulent force) to gravitational force.

5.2 Taitel and Dukler’s Transition Criteria

Taitel and Dukler suggested the transition criterion for flow regime transition from intermittent (slug or plug flow) to dispersed bubbly flow (curve D, figure 10). Their approach to this criterion is the balance of the radial turbulent force (the radial component of the pressure dynamic force due to the turbulence of water) with buoyancy force. Extension of this criterion to the annular flow regime (curve D’ of figure 10) will
mean that above curve $D'$, the annular flow in a horizontal tube is similar to that in a vertical tube. Therefore, the upper mass flux threshold ($G_2$) can be calculated from their criterion, i.e.

$$T^+ = \sqrt{\frac{8A_G}{\bar{P}_f \bar{D}_G (\bar{U}_L \bar{D}_L)^{0.2}}} = f_1(X) \tag{5.1}$$

where $T^+$ is

$$T^+ = \left[ \frac{|(dP/dz)_{fl}|}{g(\rho_f - \rho_g)} \right]^{0.5} \tag{5.2}$$

and can be expressed as

$$T^+ = \sqrt{\frac{0.092G_2^{1.8}(1 - X)^{1.8} \mu_f^{0.2}}{gD^{1.2} \rho_f ((\rho_f - \rho_g))}} = f_1(X) \tag{5.3}$$

where all the parameters in the first equation (with $^*$) are defined according to Taitel and Dukler (1976) and are functions of Lockhart-Martinelli parameter only. Rearranging equation (5.1) gives

$$G_2 = \left[ f_1(X)^2 \frac{gD^{1.2} \rho_f (\rho_f - \rho_g)}{0.092(1 - X)^{1.8} \mu_f^{0.2}} \right]^{0.556} \tag{5.4}$$

Similarly, the lower mass flux threshold, at which stratification occurs, can be obtained from the corresponding criterion

$$Fr^2 = \frac{G_2^2 X^2}{\rho_g (\rho_f - \rho_g) g D} = \frac{\tilde{C}_2 A_G}{\bar{U}_G \bar{D}_G / \bar{h}_L} = f_2(X) \tag{5.5}$$

which gives

$$G_1 = \left[ f_2(X) \frac{\rho_f (\rho_f - \rho_g) g D}{X^2} \right]^{0.5} \tag{5.6}$$

Equations (5.4) and (5.6) define the upper and lower mass flux thresholds respectively and can be used for estimating the $K_8$ value as defined in equation (2.25). However, comparison of $K_8$ predicted from equation (2.25) against experimental CHF ratio (figure 19) has indicated that it underpredicts the CHF ratio. Modification to equation (2.25) is necessary for a more accurate prediction of $K_8$.  

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5.3 Modified Froude Number

It is always possible to evaluate stratification effects of a two-phase flow in the horizontal tube through a dimensionless group, which has the inertia force of the steam in the numerator and the buoyancy force in the denominator. One such dimensionless group is the Froude number as defined by Taitel and Dukler in the horizontal flow regime map.

\[ Fr = \frac{GX}{\sqrt{gD\rho_s(\rho_f - \rho_g)}} \]  

(5.7)

The inertia force is approximately proportional to \( G^2 X^2 / \rho_g \) while the buoyancy force is incorporated in the \( gD(\rho_f - \rho_g) \) term.

Figure 20 shows the effect of Fr on the difference in critical steam quality between the top (\( X_{cr0} \)) and the bottom side (\( X_{cru} \)) of a horizontal tube (Kohler and Hein 1986). CHF occurs at the top first and as a result the critical steam quality at the top (\( X_{cr0} \)) is smaller. When the mass velocities are high (higher Fr value), CHF occurs both at the top and at the bottom at the same time and so it can be assumed equal to that for vertical flow. It is evident from figure 20 that for Fr higher than 15, the tube orientation has no effect on the CHF. Therefore, this serves as a criterion for determining an upper mass flux threshold value (\( G_{2,Fr} \))

\[ G_{2,Fr} = \frac{15gD\rho_s(\rho_f - \rho_g)}{X} \]  

(5.8)

and then using non-linear mass flux interpolation,

\[ K_6 = \left( \frac{G - G_1}{G_{2,Fr} - G_1} \right)^n \]  

(5.9)

the \( K_6 \) value can be estimated.
5.4 Force Balance Approach

5.4.1 General

Before deriving the criterion for the onset of noticeable flow stratification, some important interacting forces must be considered. They are the buoyancy force, turbulent force and drag force. The forces involved for each flow regime are different from another flow regime. Basically, three distinct flow regimes are involved in non-stratified flows. They are the bubbly, droplet and annular flows. This section outlines the forces involved in each flow regime and attempts to derive a force balance equation based on parameters controlling flow stratification. Radial and axial transit time ratios for bubble and droplet trajectories will also be evaluated.

5.4.2 Governing Forces in various Flow regimes

Forces in Annular Flow

In Taitel and Dukler's analysis, the buoyancy force per unit length of the gas region is

\[ F_B = g(\rho_f - \rho_g)A_G \]  \hspace{1cm} (5.10)

and radial turbulent force per unit length is established to be

\[ F_T = \frac{1}{2} \rho_f v^2 p_i \]  \hspace{1cm} (5.11)

where \( v' \) is the radial velocity fluctuation and \( p_i \) is the interface perimeter.

Forces in Bubbly Flow

The buoyancy force acting on each individual bubble can be expressed as

\[ F_B = \frac{\pi d^3 g(\rho_f - \rho_g)}{6} \]  \hspace{1cm} (5.12)

The drag force in the axial direction can be expressed as

\[ F_D = \frac{1}{2} C_D \rho_f (u_g - u_f)^2 \pi d^2 / 4 \]  \hspace{1cm} (5.13)

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Pressure gradient that exists in the surrounding fluid in the axial direction is the major reason for the difference in the velocity difference. This pressure gradient is absent in the radial direction and thus the bubbles can be assumed to follow the turbulent fluctuations of the liquid in the radial direction. According to Levich (1962), the radial turbulent force is proportional to the square of the velocity fluctuation such that for bubbles

\[ F_r = \frac{1}{2} \rho_s v'^2 \pi d^2 / 4 \]  \hspace{1cm} (5.14)

where \( v' \) is the radial velocity fluctuation as mentioned earlier and \( d \) is the diameter of the bubble.

**Forces in Droplet Flow**

The buoyancy force has a negative effect on droplets in the sense that it pulls the water droplets downward. The buoyancy force acting on an individual droplet can be expressed in a similar form to equation (5.12) except that for this case the diameter refers to the droplet diameter.

When liquid droplets are flowing in the pipe, drag forces exist between them and the moving vapour core. The drag force acting on each individual droplet is governed by

\[ F_D = \frac{1}{2} C_D \rho_s (u_s - u_f)^2 \pi d^2 / 4 \]  \hspace{1cm} (5.15)

For pre-CHF conditions, droplet flow occurs mainly in the core of the annular flow. It is not relevant to assume a radial turbulent force on droplets. Only the drag/buoyancy force ratio will be considered. However, it must be noted that droplets play an important role in annular flow and hence can affect CHF.

**Other Forces**

Some other important influencing factors involved in the stratification mechanism are the surface tension force and the viscosity effects. For small diameter tubes, liquids with contact angle < 90° (wetting liquids), have the tendency to climb the inner wall due to the capillary effect, thus forming a liquid film covering the complete inner
perimeter. The surface tension force can thus help in the prevention of stratification. An important parameter which represents the ratio of buoyant to surface tension forces is the Eotvos number. The viscosity number may be used as a measure of the viscosity effect.

The interacting forces can cause a variety of different flow patterns and it is important to relate these forces in a way that one can evaluate criteria for the onset of stratification. Some analytically and semi-analytically derived parameters are discussed in the following sections.

5.4.3 Force Balance in Annular Flow

Using an approach similar to Taitel and Dukler's (1986), the ratio of turbulent force to buoyancy force can be evaluated. Taitel and Dukler assumed the horizontal equilibrium interface perimeter, to be the boundary at which these two forces interact. In annular flow, the interface perimeter is much greater than the equilibrium interface perimeter, which is defined as the interface perimeter for smooth stratified flow (see figure 21). $p_i$ for annular flow can be estimated as

$$p_i = \pi D_i$$

but

$$\frac{D_i^2}{D^2} = \frac{A_G}{A}$$

Therefore

$$D_i = D \sqrt{\frac{A_G}{A}} = D \alpha^{1/2}$$

Substituting $D_i$ into equation (5.16) gives

$$p_i = \pi D \alpha^{1/2}$$

The root mean square value of $v'$ is approximately equal to the friction velocity (Taitel and Dukler, 1986). Thus

$$(v'^2)^{1/2} = u^* = u_f \left( \frac{f L}{2} \right)^{1/2}$$
Substituting equations (5.18) and (5.19) into (5.11) gives

\[ F_T = \frac{1}{2} \pi \rho_f u_f^2 \left( \frac{f_L}{2} \right) D \alpha^{1/2} \]  

(5.20)

Defining the parameter \( T_1 \) as

\[ T_1 = \frac{F_T}{F_B} = \frac{\frac{1}{2} \pi \rho_f u_f^2 D \alpha^{1/2} \left( \frac{f_L}{2} \right)}{g(\rho_f - \rho_g)A_G} \]

\[ = \frac{2\rho_f u_f^2 \left( \frac{f_L}{2} \right)}{gD(\rho_f - \rho_g)\alpha^{1/2}} \]  

(5.21)

Assuming \( f_L = 0.046 Re_L^{-0.2} \) for turbulent flow and substituting it into equation (5.21) yields

\[ T_1 = 0.046 Re_L^{-0.2} \left( \frac{1-X}{1-\alpha} \right)^2 \frac{G^2}{gD\rho_f(\rho_f - \rho_g)\alpha^{0.5}} \]

\[ = C_1 Re_L^{-0.2} \left( \frac{1-X}{1-\alpha} \right)^2 \frac{G^2}{gD\rho_f(\rho_f - \rho_g)\alpha^{0.5}} \]  

(5.22)

The parameter \( T_1 \) can be used as a measure of the dominance of the turbulent force over the buoyancy force. It has the correct parametric trends with \( G, X, D \) and \( P \). Generally, it increases with \( G \) and \( X \) and decreases with \( D \) and \( P \). The variation of this parameter with \( X \) and \( P \) are shown in figures 22a and 22b.

### 5.4.4 Force Balance in Bubbly Flow

The ratio of axial drag force to buoyancy force is useful in the analysis of stratification. This ratio is obtainable by dividing equation (5.13) by equation (5.12), thus giving

\[ T_2 = \frac{3}{4} \frac{C_D \rho_f (u_g - u_f)^2}{gD(\rho_f - \rho_g)} \]  

(5.23)

The bubble diameter can be expressed in terms of Weber number and fluid properties. For a sphere with \( 10^3 < Re < 2 \times 10^5 \) (Schlichting, 1968),

\[ C_D \approx 0.5 \]

where

\[ Re = \frac{\rho_f (u_g - u_f)d}{\mu_f} \]
Also, bubbles will break up at a certain critical Weber number. Assuming break-up diameter to be the average bubble diameter, by substituting equation (4.37) into equation (5.23) and knowing that \( u_g \) and \( u_f \) can be calculated by \( \frac{G X}{\rho_g \alpha} \) and \( \frac{G(1 - X)}{\rho_f (1 - \alpha)} \) respectively, \( T_2 \) can be rearranged into the form of

\[
T_2 = C_2 \left( \frac{\rho_2}{\rho_f} \right) \frac{G^4}{g \sigma \rho_f^3 (\rho_f - \rho_g)} \left[ \frac{\frac{2L}{g} (1 - \alpha) X - \alpha (1 - X)}{\alpha (1 + \alpha)} \right]^4
\]  

(5.24)

where \( C_2 \) is an arbitrary constant.

The turbulent over buoyant force ratio for bubbly flow can be obtained in a manner similar to that for annular flow. Defining \( T_3 \) as the force ratio,

\[
T_3 = \frac{F_T}{F_B} = \frac{\frac{1}{2} \rho_g u'^2 \pi d^2 / 4}{g (\rho_f - \rho_g) \pi d^2 / 6}
\]  

(5.25)

The root mean square value of \( u' \) is approximately equal to the friction velocity (Taitel and Dukler, 1986). Thus

\[
(u')^2 = u^* = u_f (f_L/2)^{1/2}
\]  

(5.26)

Substituting \( u'^2 \) into equation (5.25) yields

\[
T_3 = \frac{3 \rho_g u_f^2 (f_L/2)}{4 g d (\rho_f - \rho_g)}
\]  

(5.27)

Assuming \( f_L = 0.046 Re_L^{-0.2} \) for turbulent flow and substituting it into equation (5.27) yields

\[
T_3 = 0.0173 \left( \frac{\rho_2}{\rho_f} \right) \frac{\rho_f u_f^2 Re_L^{-0.2}}{g d (\rho_f - \rho_g)}
\]  

(5.28)

The bubble diameter may be approximated by using Weber number as described earlier, i.e., substituting equation (4.37) into equation (5.28) and rearranging yields

\[
T_3 = C_3 Re_L^{-0.2} \frac{\rho_2 u_f^2 (u_g - u_f)^2}{g (\rho_f - \rho_g)}
\]  

(5.29)

where

\[
\begin{align*}
u_g &= \frac{G X}{\rho_g \alpha} \\
u_f &= \frac{G(1 - X)}{\rho_f (1 - \alpha)}
\end{align*}
\]
Taitel and Dukler argued that when the buoyancy force is balanced by the radial turbulent force (i.e. when \( T_3 = 1.0 \)), dispersion of bubbles starts to occur, at which point the flow distribution is more symmetrical. When \( T_3 \) is much greater than 1.0; the flow distribution for horizontal flow will be very similar to that for vertical flow under otherwise identical flow conditions.

### 5.4.5 Force Balance in Droplet Flow

In horizontal flow, the buoyancy force acting upon each droplet (of diameter \( d \)) is given by equation (5.12). The ratio of the drag force (equation 5.15) to the gravity force (same as equation 5.8) can be expressed as

\[
T_4 = \frac{F_D}{F_B} = 3\frac{C_D(u_g - u_f)^2}{4g} = \frac{\rho_g}{\rho_f - \rho_g} \frac{\rho_f}{g(d\rho_f - \rho_g)} \tag{5.30}
\]

The droplet diameter can be related to Weber number, which is defined as

\[
d = \frac{W_{eL}\sigma}{\rho_f(u_g - u_f)^2} \tag{5.31}
\]

Substituting equation (5.31) into equation (5.30) and simplifying yields

\[
T_4 = \frac{3}{4} \frac{C_D}{W_{eL}} (u_g - u_f)^4 \frac{\rho_g\rho_f}{g(\rho_f - \rho_g)\sigma} \tag{5.32}
\]

The water droplets will break at a certain critical Weber number. Taking the average size of the droplets to be proportional to \( \frac{\sigma}{\rho_f(u_g - u_f)^2} \), \( W_{eL} \) can be treated as a constant. Since we are only interested in looking at the dominance of drag force over the buoyancy force, the constant terms in equation (5.32) can be combined to form an arbitrary constant and \( T_4 \) is redefined as

\[
T_4 = C_4(u_g - u_f)^4 \frac{\rho_g\rho_f}{g(\rho_f - \rho_g)\sigma} \tag{5.33}
\]

By expressing the actual velocities in terms of mass flux, quality and void fraction, equation (5.33) is rewritten as

\[
T_4 = C_4 \left( \frac{\rho_g}{\rho_f} \right) \frac{G^4}{g\sigma\rho_f^2(\rho_f - \rho_g)} \left[ \frac{\alpha_l(1 - \alpha)X - \alpha(1 - X)}{\alpha(1 - \alpha)} \right]^4 \tag{5.34}
\]
Equation (5.34) is exactly the same as equation (5.24) except for the constant $C_4$. This parameter serves as a measure of the dominance of drag force over gravitational force. A criterion for onset of stratification based on this parameter should be investigated.

The approach taken here is very simplistic in nature. A more rigorous approach was suggested by Ishii (1977).

5.5 Transit Time Approach

5.5.1 General

An alternative to the force balance approach is to consider transit time for a bubble or droplet in both the vertical and the horizontal direction. The transit time taken for a droplet or bubble to travel across the channel is compared to the transit time to travel to the exit of the tube. This transit time ratio serves as a measure of void migration and hence a measure of the void accumulation near the top of the channel. Some simplified approaches were utilized in the development of the transit time ratio for various types of flow regimes such as droplet flow, bubbly flow and annular flow.

5.5.2 Bubbly Flow

For bubbly flow, the bubbles have an average velocity of

$$u_g = \frac{G X}{\rho_g \alpha} \quad (5.35)$$

Therefore the time taken for the bubble to travel to the exit is given by

$$t_x = \frac{L_h}{u_g} \quad (5.36)$$

or

$$t_x = \frac{\rho_g \alpha}{G X} L_h \quad (5.37)$$
By neglecting drag force in the vertical direction, the force balance in the vertical direction is

\[ \frac{\pi d^3}{6 \rho_f} \frac{du_v}{dt} = \frac{\pi d^3}{6} (\rho_f - \rho_g) g, \]  

(5.38)

where \( u_v = \) bubbly velocity upward.

Solving for \( u_v \) in equation (5.38) with zero initial velocity yields

\[ u_v = \frac{(\rho_f - \rho_g) g t}{\rho_g}, \]  

(5.39)

Assume the average bubble starts to depart at the middle of the tube, the bubble rise time can be found by integrating the velocity w.r.t. time over the distance \( D/2 \), i.e.,

\[ \frac{D}{2} = \int_0^{t_v} u_v \, dt \]  

(5.40)

Substituting equation (5.39) into equation (5.40) and integrating gives

\[ \frac{D}{2} = \frac{1}{2} \frac{(\rho_f - \rho_g) g t_v^2}{\rho_g}, \]

\[ t_v^2 = \frac{D \rho_g}{g(\rho_f - \rho_g)}, \]

\[ t_v = \left[ \frac{D \rho_g}{g(\rho_f - \rho_g)} \right]^{1/2}. \]

(5.41)

Therefore, the time ratio is defined as

\[ T_3 = \frac{t_v}{t_x} = \left[ \frac{D \rho_g}{g(\rho_f - \rho_g)} \right]^{1/2} \frac{G X}{\rho_g \alpha L_h}. \]

Rearranging gives

\[ T_3 = \frac{G}{\sqrt{g D \rho_g (\rho_f - \rho_g) \alpha L_h}} \frac{X D}{\alpha L_h}. \]

(5.42)

Equation (5.42) has the correct parametric trends for heated length, inertia force and buoyancy force.

### 5.5.3 Droplet Flow

One influencing parameter for onset of stratification is the ratio of the time taken for the droplet to fall down to the bottom of the tube over the time taken for the droplet to travel to the exit.
Assume a liquid droplet is flowing with a velocity $u_f$ where

$$u_f = \frac{G(1 - X)}{\rho_f(1 - \alpha)}$$

(5.43)

Time taken for the droplet to travel to the end of the tube is given by

$$t_z = \frac{L_h}{u_f}$$

(5.44)

where $L_h =$ tube length. Substituting equation (5.43) into equation (5.44) gives

$$t_x = \frac{\rho_f(1 - \alpha)}{G(1 - X)}L_h$$

(5.45)

It is assumed that the droplets are entrained in the core of the tube and start to fall down to the bottom of the tube.

Neglecting drag force upwards, the force balance in the vertical direction is

$$\frac{\pi d^3}{6} \rho_f \frac{du_v}{dt} = 6 \frac{\pi d^3}{6}(\rho_f - \rho_g)g$$

(5.46)

where $u_v =$ droplet velocity downward.

Solving for $u_v$ in equation (5.46) with zero initial velocity yields

$$u_v = \frac{(\rho_f - \rho_g)gt}{\rho_f}$$

(5.47)

Similar treatment as in the bubbly flow can be performed to determine the time taken for the droplet to reach the bottom of the tube, i.e.

$$t_v = \sqrt{\frac{D\rho_f}{g(\rho_f - \rho_g)}}$$

(5.48)

Therefore, the time ratio is defined as

$$T_0 = \frac{t_v}{t_z} \left[ \frac{D\rho_f}{g(\rho_f - \rho_g)} \right]^{1/2} \frac{G(1 - X)}{\rho_f(1 - \alpha)L_h}$$

(5.49)

Rearranging gives

$$T_0 = \frac{G}{\sqrt{gD\rho_f(\rho_f - \rho_g)\frac{1 - X}{1 - \alpha}L_h}}$$

(5.50)

The time ratio is important because the droplets formed might reach the bottom of the tube before exiting if the tube is long enough, thereby accumulating more liquid.
at the bottom. Consequently less liquid is concentrated at the top of the tube and
dryout can occur more easily.

It must be noted that the force balance and transit time ratio approaches are rather
simplistic in nature. A more rigorous approach on the relative motion between the gas
and liquid phases is given by Ishii (1977).

5.5.4 Annular Flow

Many researchers have investigated the mechanisms controlling the horizontal annular
flow configuration. Recent studies include the works of Hutchinson et al. (1974),
Laurinat et al. (1984) and T.F. Lin et al. (1985). Their works are discussed in more
details in section 5.6. The transit time approach can not be easily used for annular flow
because there is no discrete flowing particle for which the trajectory can be followed.

5.6 Annular Flow Distribution Approach

The asymmetrical distribution of the liquid film in horizontal annular flow is caused
by the gravitational force. Hutchinson et al. (1974) proposed that there is a minor
circumferential film flow from the top to the bottom of the pipe due to gravity. In order
to maintain equilibrium, the liquid film near the top of the pipe had to be replenished
by net droplet deposition, and the liquid film near the bottom had to be balanced by
net droplet entrainment (Lin et al., 1986).

In addition to the droplet exchange model discussed above, Laurinat et al. (1984)
recently proposed that there exists a secondary flow pattern in the gas core of a
horizontal annular flow. In horizontal annular flows, the asymmetrical film distribution
causes the interfacial roughness in the bottom of the pipe to be larger than that in the
top. This rougher interface at the bottom presents more resistance to the gas flow in
the axial direction and thus results in a circumferential (secondary) gas flow pattern
as shown in figure 23.

Calculations of film thickness distribution were performed at Rensselaer Poly-
technic Institute (Lin et al., 1986), based on the combination of the two mechanisms discussed. The calculated results compare favorably with experimental data. This approach can be used for estimating critical heat flux by the inclusion of heat balance equations. However, this will add a considerable amount of complexity to the problem and further investigations have to be done.
Chapter 6

DERIVATION OF ORIENTATION CORRECTION FACTOR \((K_6)\)

6.1 General

By combining two or more of the parameters developed in Chapter 5, a more complete and semi-empirical criterion for onset of stratification can be established. Several parameter groupings have been used in deriving an expression for the orientation correction factor \(K_6\). This chapter is devoted to the derivation of the \(K_6\) expression. Three different approaches were examined and the present approach gives the best result. Comparison of prediction against experimental data was made and the errors are summarized in tabular form.

6.2 Groeneveld’s Approach

Groeneveld (1982) suggested that the CHF for vertical flow be multiplied by a stratification factor \(K_6\) to obtain CHF for horizontal flow, i.e.,

\[
CHF_{\text{hor}} = K_6 \times CHF_{\text{ver}} \tag{6.1}
\]
For stratified flow, $CHF_{hor} = 0$ and $K_6 = 0$. For a mass flux greater than $G_2$ obtained from equation (5.4), the horizontal CHF is approximately equal to the vertical CHF and hence $K_6 = 1.0$. For given mass flux that is between these two threshold values, $K_6$ factors can be approximated by

$$K_6 = \frac{G - G_1}{G_2 - G_1}$$  \hspace{1cm} (6.2)

where $G_1$ is given by equation (5.6).

Comparison with data (figure 19) shows that equation (6.2) represents a lower bound to most of the data. Simple modification to include non-linear mass flux relationship can help improve the $CHF_{hor}$ prediction. A form of

$$K_6 = \left[ \frac{G - G_1}{G_2 - G_1} \right]^n$$  \hspace{1cm} (6.3)

was suggested because it has correct parametric trends. With least squares best estimate curve fitting, $n$ was found to have a value of approximately 0.62 and the root-mean-square error was found to be about 38.5% (error definition to be found in Section 6.5.1).

Figures 24 and 25 show the errors in prediction for linear and non-linear mass flux interpolation respectively. The non-linear approach gives a better prediction of CHF ratio. Both correlations give a large RMS error and this leads to the question of the reliability of curves A and DD’ in figure 10.

Work has been done by Minato et al. (1986) to derive a new model for the transition mechanism from stratified gas-liquid flow into intermittent or annular flow. Their model incorporated the contribution of liquid kinematic energy to wave growth on the interphase surface. Minato et al. compared both the Taitel and Dukler’s transition criterion and their model to their experimental data and concluded that Taitel and Dukler’s criterion provides a prediction very similar to their own in spite of its neglecting of the water kinematic energy effect. Therefore it seems that Taitel and Dukler’s transition criterion for $G_1$ is applicable.

When extrapolated to the annular flow regime, the criterion for the transition between intermittent flow and dispersed bubbly flow (curve D) may not be accurate
enough for the determination of $G_2$ due to the different mechanisms involved in the two-phase distribution. Also, the simplistic assumption represented by equation (6.3) may be another reason for the poor accuracy. Therefore, more accurate methods of estimating the $G_2$ value is required and such alternatives are discussed in the following sections.

6.3 Modified Froude Number Approach

Experimental water CHF values for horizontal flow were obtained from those listed in Appendix A and the ratio of horizontal CHF over vertical CHF was determined and plotted against $Fr$ defined in equation (5.7). Figure 26a illustrate the trend of CHF ratio versus $Fr$. It can be concluded from this figure that the CHF ratio increases with $Fr$ in general.

From an optimization technique using the least squares method, $Fr = 15$ was chosen as the criterion for the upper mass flux threshold, $G_{2,Fr}$. Prediction of $K_6$ based on this criterion and with the form of

$$K_6 = \left[\frac{G - G_{1}}{G_{2,Fr} - G_1}\right]^{0.8}$$

shows some improvement over Taitel and Dukler’s approach. The root mean square error for this equation is about 30%. From the above analysis, it is evident that the Froude number contains important parameters affecting horizontal CHF. Perhaps the heated length over diameter ratio is equally important because a long heated tube will allow a small buoyancy effect to dominate. A longer tube means a longer time available for the bubble to rise to the top of the tube before exiting.

A further modification to the Froude number is possible by introducing the heated length. One such modification is to divide $Fr$ by $(L_h/D)$ ratio, defining

$$Fr^* = Fr \frac{D}{L_h}$$

(6.4)

where $L_h$ = heated length.
A similar plot of CHF ratio vs. $Fr^*$ (figure 26b) shows a similar trend as compared to figure 26a. There is still a considerable amount of scatter in the data. No clear-cut criterion for the upper mass flux threshold can be derived from this figure. Again, from the least squares curve fitting method, $Fr^* = 0.10$ was chosen as the upper bound at which the tube orientation has no significant effect on CHF and the $G_2$ value can then be calculated as

$$G_{2,Fr^*} = 0.10 \left( \frac{L_h}{D} \right) \sqrt{gD \rho_\phi (\rho_f - \rho_\phi)} \frac{1}{X}$$  \hspace{1cm} (6.5)

With equation (6.5), the value of $K_6$ can be calculated for each given set of flow conditions by equation (6.3). For this approach, $n$ was found to be about 0.60. Therefore,

$$K_6 = \left[ \frac{G - G_1}{G_{2,Fr^*} - G_1} \right]^{0.60}$$ \hspace{1cm} (6.6)

with $G_1$ defined in equation (5.6) and $G_{2,Fr^*}$ defined in equation (6.5). The experimental CHF ratio was compared against the prediction defined in equation (6.6) (figure 27). Errors of prediction were shown in figure 28 and 29 against CHF ratio and quality respectively. This method provides a better prediction of CHF ratio as compared to the methods described in the preceding section. It is simple to use because of fewer calculation steps involved in evaluating $G_{2,Fr^*}$. The is found to RMS error is 29%.

### 6.4 Present Approach

#### 6.4.1 Summary of Parametric Groups

The parameters derived in Chapter 5 have some similar basic groups. They are summarized in Table 8. The constants $C_1$, $C_2$, $C_3$ and $C_4$ are arbitrary constants to be determined by comparison with experimental data. All the parameters $T_1$ to $T_6$ have similar parametric trends in the sense that a larger parametric value means a smaller effect of stratification. The mass flux appears in the numerator of all parameters. This simply means that all parametric values increase with mass flux. Note that parameter $T_3$ is very similar to the modified Froude number except that void fraction appears in
the denominator of the former case.

Attempts will be made to relate the CHF ratio \( \frac{CHF_{kor}}{CHF_{ver}} \) to the parameters developed in Chapter 5. Different combinations of parameters or dimensionless groups are tested against experimental data to correlate a function for the CHF ratio. The details of the assessment technique are discussed in the following section.

6.4.2 Assessment Technique

It is known that for CHF experiments, the data obtained are accurate only within the precisions of the experimental set-up. Most of the experimental set-ups permit a measurement of CHF with an accuracy of \( \pm 5\% \) (2 \( \sigma \) confidence level). Due to the wide scatter of experimental data, it is not easy to find an optimum correlation considering the number of parameters involved. An assessment technique was developed for the determination of optimum correlation of parameters.

A crude approach is to plot the CHF ratio against each parameter and combinations of parameters. Figures 30-35 show some of the typical plots. From the plots, the parameter or combinations of parameter that give the clearest trend can be chosen and correlated. One drawback of this approach is that the combinations of the parameters are user defined and are not based on any theoretical deductions. A more scientific approach is developed for this purpose.

A statistical software package called SAS (Statistical Analysis System) was used to determine the correlation coefficient of each parameter with respect to the experimental CHF data. It was found that \( T_1, T_3 \) and \( T_6 \) have a significant influence on the CHF ratio and they will be studied in more detail.

6.4.3 Derivation of CHF Correlation

From figures 30 to 35, CHF ratio seems to depend on the parameters in an exponential way. It approaches the value 1.0 as the parameter value becomes very large. Attempts were made to correlate the CHF ratio as some exponential function of the parameters.
because an exponential function give the expected behaviour of the CHF ratio. One simple function the form of

\[ CHFR = 1 - e^{-(T_i/A)^B} \]  

(6.7)

was found to correlate the CHF ratio very well where \( T_i \) is one of the parameters shown in Table 8, \( A \) and \( B \) are the empirical constants corresponding to each \( T_i \). Equation (6.7) can be rearranged to give

\[ A = \frac{T_i}{[-\ln(1 - CHFR)]^{1/B}} \]  

(6.8)

All six parameters were tested in the derivation of the correlation. All the arbitrary constants \( C_1 \) to \( C_6 \) were set equal to 1.0. For each parameter and \( B \) value, the horizontal CHF data bank was used to determine the value of \( A \) with equation (6.8). Unfortunately, due to the wide scatter of data, the value of \( A \) does not seem to be well defined. Attempts were made to determine a function for \( A \) but no suitable function was found. Figure 36 shows the values of \( A \) plotted against Martinelli's parameter for \( T_i \) with \( B = 0.5 \). No systematic trend was found for \( A \). Using the least squares fit, optimum values for \( A \) and \( B \) were obtained for each case which give the respective minimum RMS errors. Table 9 summarizes the error distributions for each parameter tested based on the constant local conditions. The comparison based constant inlet conditions is shown in Table 10. In both Tables 9 and 10, different values of \( A \) and \( B \) were tried with parameters \( T_1, T_3 \) and \( T_5 \). However, only the best correlated equations for parameters \( T_3 \) and \( T_5 \) were shown even though other constant values for \( A \) and \( B \) were investigated.

The best correlation appears to be with the use of parameter \( T_1 \) with \( B = 0.5 \) and \( A = 65 \). The use of different void correlations can affect the prediction of CHF to some extent. Taitel and Dukler’s void fraction correlation was tried without success. The CISE void correlation (see section 4.3.3) was chosen for this case because it is the most accurate generally applicable correlation (Whalley, 1987). However, the use of some proprietary correlations developed for horizontal tube flow may improve
the prediction significantly. With the CISE void correlation, the CHF ratio can be predicted by

\[ \text{CHFR} = 1 - e^{-\left(\frac{T_1}{\Delta s}\right)^{0.4}} \]  \hspace{1cm} (6.9)

where \( T_1 \) is defined in equation (5.22). Thus the CHF for horizontal flow can be predicted by

\[ \text{CHF}_{\text{hor}} = \text{CHFR} \times \text{CHF}_{\text{ver}} \]  \hspace{1cm} (6.10)

where \( \text{CHF}_{\text{ver}} \) is the CHF for vertical flow obtained from the Groeneveld look-up table. This correlation predicts the CHF ratio for water with reasonable accuracy (Tables 11 and 12) and is non-dimensional and hence may be applied to non-aqueous fluids such as refrigerants. Equations (6.9) and (6.10) were also used to predict the CHF for Freon-12 and compared to the data from Merilo and Ahmad (1979). The results are summarized in Tables 13 and 14. The parametric trends for equation (6.10) are discussed in the next section.

6.4.4 Parametric Trends of the CHF Correlation

To permit the development of a generalized correlation (applicable to a wide range of flow parameters), the correlation must have the correct asymptotic trends. Pressure, mass flux, tube diameter and quality are the four major flow parameters that affect the CHF. Parametric trends of equations (6.9) and (6.10) with respect to pressure, mass flux, tube diameter and quality are illustrated in figures 37 to 40.

Pressure has a direct influence on the density ratio, which is the main component of the buoyancy forces. The CHF ratio predicted from equation (6.9) is found to decrease with increasing pressure as would be expected from elementary considerations of the density ratio and the buoyancy forces. Increasing pressure results in the decrease of the density ratio \( \rho_f/\rho_g \) which in turn decreases the turbulent force as discussed in section 3.2.4. Figures 37a and 37b illustrate the pressure effects on the predicted CHF ratio.

The effects of mass flux on CHF ratio prediction at constant local conditions are
shown in figures 38a and 38b. Generally, the CHF ratio increases as the mass flux increases. This is in agreement with the experimentally observed trend. The quality used in the evaluation of CHFR is the actual quality. For negative thermodynamic quality, Saha and Zuber's equation presented in Chapter 4 can be used to estimate the actual quality. Figures 39a and 39b show the trends of predicted CHF ratio with quality. At constant mass flux, pressure and tube diameter, the predicted CHF ratio increases with increasing quality except at low qualities ($X_{cr} < 0.01$) when the CHF ratio decreases with increasing quality. This predicted trend conforms to that discussed in section 3.2.1. As the actual quality approaches zero, the void fraction also approaches zero resulting in a very large value of $T_1$ (due to the term $\alpha^{-0.5}$) and therefore $CHFR \rightarrow 1.0$. The influence of tube diameter on CHF ratio was demonstrated in figure 40 based on constant local conditions. CHF ratio decreases with increasing tube diameter as expected based on the discussion in Section 3.2.3.

6.5 Comparison and Error Analysis

At the present, there is no accurate and generalized prediction method for the CHF in horizontal flow. Ishigai et al. (1970b) suggested an analytical model for horizontal annular flow but their method is very complex when heating is involved. Wang and Su (1980) have correlated their experimental data to give an ad hoc equation which is not valid outside their experimental range. Their method can not be used for comparison since the wrong parametric trends of their equation often give CHF values that are thousands of times greater than the data value (except for their own data). Merilo's (1979) correlation (eq. 2.20) seems to be the most promising and is compared with the present approach (Table 12).

The present method (equation (6.9)) is compared to experimental data from sources summarized in Table 6. Two methods of comparison were made: an error analysis based on constant dryout quality condition (local parameters) and an error analysis based on system parameter (constant inlet condition approach). Since Merilo's cor-
relation is based on the constant inlet conditions, it is only compared to the present approach with constant inlet conditions (Table 12). The prediction based on constant inlet condition is more accurate than that based on constant local parameters. These two types of predictions are discussed in detail in the following sections.

6.5.1 Constant Dryout Quality Condition

A prediction based on constant dryout quality condition assumes that the CHF is a function of local parameters only. To make the comparison, CHF values are predicted directly from the equation or correlation using the local parameters given and then compared to the experimentally obtained values. The individual prediction error is defined as

\[
Error_{pred} = \frac{CHF_{pred} - CHF_{expt}}{CHF_{expt}}
\]  

(6.11)

The mean error is the average value of all the individual prediction errors. The root mean square (RMS) error is defined as

\[
RMSE = \sqrt{\frac{\sum(\text{Error}_{pred})^2}{N}}
\]

(6.12)

where \(N\) is the number of data points. The standard deviation is defined as

\[
STD = \sqrt{\frac{\sum(\text{Error}_{pred} - \text{Error}_{mean})^2}{N}}
\]

(6.13)

\(\text{Error}_{mean}\) is the arithmetic mean of the overall errors, i.e.,

\[
\text{Error}_{mean} = \frac{\sum\text{Error}_{pred}}{N}
\]

(6.14)

again, \(N\) is the number of data points.

It is clear from Tables 11 and 12 that the present method is superior to the other methods.

6.5.2 Constant Inlet Quality Condition

Predictions based on constant inlet conditions require iteration. The principle behind this is to keep the inlet parameters (such as the inlet subcooling, \(\Delta H_{in}\)) and the heated
length constant and vary the heat flux applied until CHF occurs. A prediction of CHF is made at the end of the tube corresponding to each heat flux value while keeping all the other parameters constant. At each step, the predicted CHF value is compared to the assumed heat flux value. Figure 41a shows schematically the iteration process. When the predicted CHF value is higher than the assumed heat flux, the quality at the end of the tube is increased (corresponding to an increase of assumed heat flux value) stepwise until the assumed heat flux has the same value as the predicted CHF. That final value (corresponding to the intersection of the two curves shown in figure 41a) will be the predicted CHF value used in the comparison.

Generally, the predictions based on constant inlet condition are more accurate than those based on local conditions. However, it is possible for low flow cases that predictions based on constant inlet condition give larger errors. In figure 41b, the slope of predicted CHF versus quality is positive for a range of quality. If the above iteration process is used, the final prediction will give a CHF value higher than that of the first prediction. This will give rise to a higher prediction error. Fortunately, in most industrial applications, critical heat flux occurs as a result of annular liquid film failure which corresponds to the downstream dryout for which CHF is decreasing with quality (negative slope).

Table 12 shows the error analysis for various prediction methods based on constant inlet conditions for water. Notice that predictions made from equation (6.3) improve significantly in this case when compared to the local parameters approach. On the contrary, equation (6.6) has become less accurate when constant inlet conditions are applied. Based on both the constant inlet and constant local parameters approach, it is concluded that equation (6.9) has a better overall prediction accuracy. The comparison for Freon-12 for the local parameters approach and constant inlet conditions approach are shown in Tables 13 and 14 respectively. Equation (6.3) predicts the CHF values more accurately for Freon-12.

Prediction errors for both water and Freon-12 data are also presented in figures 42 and 43. The applicability of equation (6.9) to non-aqueous fluids is partially confirmed
by the Freon data comparison even though the prediction was developed based on water data. This is because the dimensionless parameters were derived analytically based on fluid properties such as densities, viscosity and surface tension, regardless of the type of fluid involved. The errors quoted may seem very large to most people but in thermal-hydraulic analysis of two-phase flows, errors of 20% are not uncommon.

6.5.3 Error Contribution Factors

There are a number of factors which can contribute to the rather large prediction errors. They are summarised below.

1. The CHF data sources are not very consistent even with similar experimental set-ups. Some of Merilo's data and those obtained from Kohler and Hein (1986) are from very similar experimental conditions, but the CHF values quoted by both groups are quite different from each other.

2. The CHF for vertical tubes is predicted from the Table look-up method which in itself has some errors involved. Data for vertical CHF under otherwise identical conditions to horizontal tubes are scarce due to the limitations of the experimental set-up. The table look-up method is the best so far for estimating vertical CHF values for a wide range of flow conditions.

3. In the low quality and subcooled boiling region, the equilibrium quality is very different from the actual quality. Saha and Zuber's equation (Equation 4.30), used for evaluating \( X_a \) has some weaknesses, especially under high pressure and high heat flux conditions, when it could predict an actual quality of more than 50% percent for negative equilibrium quality (see figure 44).

4. In deriving the dimensionless parameters based on the forces balance, some forces were neglected such as the surface tension force (in deriving \( T_1, T_3 \) and \( T_5 \)). Also, the
viscosity effects were not taken into account in deriving most of the parameters (except for $T_1$ and $T_3$ where $\mu$ was taken into account in the Reynolds number). These forces were neglected because they are believed to be insignificant. However, part of the inaccuracy in the predictions may be attributed to the exclusion of these forces.
Chapter 7

DISCUSSION AND CONCLUSIONS

CHF mechanisms for both the vertical and horizontal flow were reviewed and some available prediction methods were examined. The analysis presented here is somewhat simplified to avoid complex mathematical modelling. A better understanding of the forces involved in the flow mechanisms is needed before an improved prediction method can be developed. Experiments to study the void migration in a horizontal flow would be helpful in establishing the onset of stratification.

The scarcity of horizontal CHF data is also an area of concern. More experimental work on horizontal CHF should be done to provide a bigger data bank and to verify the validity of the present model. The model developed here has not been tested extensively. A better model can be developed if an extensive amount of horizontal CHF data is available. Experimental trends of CHF were discussed in terms of its variation with pressure, mass flux, quality and tube diameter. It was found that the horizontal CHF generally increases with mass flux and decreases with tube diameter. The effect of quality on horizontal CHF is somewhat more complicated. If the mass flux is low, the CHF first decreases with quality and then increases with quality and decreases with quality again at high quality region (see figure 13). The pressure effect on horizontal CHF was observed by Robertson (1973) and it was found that CHF
decreases with pressure. The predicted CHF was found to have the correct trend with pressure (figures 37a and 37b).

A generalized property code for non-aqueous fluids was also developed during the course of this research. This property code helps in the determination of fluid properties necessary for the prediction of CHF. Development of this property code was discussed in Chapter 4. More detailed discussion on the subroutines and comparison with experimental data can be found in Appendix B.

A generalized semi-analytical prediction method for CHF in horizontal flow has been developed. This prediction method is only valid for uniformly heated tubes and should not be applied to tubes with variable heat flux. It was tested against most of the horizontal tube CHF data available with different flow conditions. The root mean square error is about 16%. However, the correlation over-predicts CHF more than under-predicts it. It should be used with caution if a conservative prediction is required.

Under certain flow conditions (high pressure and high mass flux), equation (6.9) could under-predict the $CHF_{\text{hor}}$ considerably. It is due to the slow approach of the exponential function towards the value of 1.0 even though the mass flux is supposed to be high enough to suppress the gravitational effect. Figure 44 shows the variation of CHF ratio with mass flux as predicted by equations (2.25) and (6.9). For mass flux greater than $G_A$, equation (2.25) predicts higher values of CHFR and thus converges towards the value of 1.0 sooner. If the maximum value of the two predictions is taken as the final predicted value, the problem of under-prediction would be solved. $CHF_{\text{hor}}$ predicted from the proposed combination were compared to the water data and the results are summarized in Table 15. The results show an improvement for the prediction of horizontal CHF (RMS error of about 15%).

The method developed is also valid for non-aqueous fluids like Freons, nitrogen, carbon dioxide and other non-polar fluids. Groeneveld (1986c) suggested a few dimensionless groups for fluid to fluid modelling and his method has been found successful in predicting CHF for vertical flow for non-aqueous fluid using Groeneveld et al.'s CHF
table. For the prediction of CHF ratio, only the flow parameters and the fluid properties need to be known. The horizontal CHF for non-aqueous fluid can be predicted by using equation (6.9) in conjunction with Groeneveld's method of predicting CHF for vertical flow.

Both equations (6.3) and (6.9) represent two of the more generally applicable prediction methods which are valid for a wide range of flow parameters and for different fluids.
FINAL REMARKS AND RECOMMENDATIONS

1. The constant inlet conditions approach should be used when computing the CHF ratio if the inlet subcooling and the heated length are known. This approach is more accurate than the constant exit conditions (local parameters) approach. An iterative process is necessary if equation (6.9) is used but it usually converges rapidly.

2. For qualities greater than 0.2, the equilibrium quality can be assumed equal to that of the actual quality. For subcooled and low quality flow boiling, the actual quality can be estimated from Saha and Zuber's equation (see section 4.3.2).

3. The mechanisms for horizontal CHF at low quality and subcooled boiling are not well understood at the present. Further experimental and analytical work for these flow conditions should be carried out before a more accurate and generalized prediction method can be developed.

4. Better predictions could result if the data were classified into different flow regimes (e.g. using the Taitel and Dukler's flow regime map) and attempts were made to correlate the data with different $T_i$ in different flow regimes.

5. Experimental and analytical work is in progress at the Atomic Energy of Canada Limited (AECL) to study the void migration in a horizontal tube. The work is aimed at developing an analytical CHF prediction for horizontal flow.
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Yoshida, S., Mori, H. and Ohno, M., (1987), 'Critical Heat Flux for Boiling Freon at High Subcritical Pressure in Horizontal Tubes', Joint ASME/JSME Conference,
Tokyo.


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Table 1. Critical Heat Flux Lookup Table (kW.m⁻²) [Groeneveld et al., 1986a]
Table 1. (Continue)

| WEIGHT | 0.25% | 0.40% | 0.50% | 0.60% | 0.70% | 0.80% | 0.90% | 1.00% | 1.10% | 1.20% | 1.30% | 1.40% | 1.50% | 1.60% | 1.70% | 1.80% | 1.90% | 2.00% |
|--------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|-------|
| 1500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 2000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 2500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 3000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 3500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 4000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 4500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 5000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 5500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 6000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 6500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 7000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 7500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 8000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 8500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 9000   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 9500   |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |
| 10000  |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |       |

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Table 2. Comparison of CHF Data in Vertical Tubes with Various Correlations for All Fluids (ESDU, 1986).

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*Mean Value of $\frac{q_{cr,pred} - q_{cr,expt}}{q_{cr,expt}}$
Table 3. Range of Experimental Parameters used in Deriving Eq. (2.20)

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<th>$X_m$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Merilo (1979)</td>
<td>Freon-12</td>
<td>12.6</td>
<td>193-387</td>
<td>13-20.5</td>
<td>0.7-5.4</td>
<td>-35 - 0</td>
</tr>
<tr>
<td>Merile (1977)</td>
<td>Water</td>
<td>12.6</td>
<td>193-387</td>
<td>13-20.5</td>
<td>1.0-5.7</td>
<td>-30 - 0</td>
</tr>
<tr>
<td>Robertson (1973)</td>
<td>Water</td>
<td>19.1</td>
<td>112-160</td>
<td>20.5</td>
<td>0.7-1.4</td>
<td>-29 - 7</td>
</tr>
<tr>
<td>Merilo &amp; Ahmad (1979)</td>
<td>Freon-12</td>
<td>5.3</td>
<td>193-571</td>
<td>13-20.5</td>
<td>1.6-8.1</td>
<td>-35 - 0</td>
</tr>
</tbody>
</table>
Table 4. Comparison of Merilo's (1979) Correlation with Water CHF Data

<table>
<thead>
<tr>
<th>Data Sources</th>
<th>No. of Points</th>
<th>Average Error (%)</th>
<th>RMS Error (%)</th>
<th>Pressure Range (MPa)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Robertson (1979)</td>
<td>52</td>
<td>3.65</td>
<td>8.90</td>
<td>3.4 - 6.9</td>
</tr>
<tr>
<td>Leontiev et al. (1980)</td>
<td>23</td>
<td>-7.68</td>
<td>30.19</td>
<td>6.9 - 13.6</td>
</tr>
<tr>
<td>Becker (1971)</td>
<td>82</td>
<td>67.71</td>
<td>122.19</td>
<td>1.0 - 3.0</td>
</tr>
<tr>
<td>Kohler &amp; Hein (1986)</td>
<td>69</td>
<td>55.80</td>
<td>81.36</td>
<td>10, 15, 20</td>
</tr>
<tr>
<td>Wang &amp; Su (1980)</td>
<td>19</td>
<td>-68.16</td>
<td>69.33</td>
<td>4.9</td>
</tr>
<tr>
<td>All Data</td>
<td>279</td>
<td>28.8</td>
<td>80.3</td>
<td></td>
</tr>
</tbody>
</table>
Table 5. Exit Quality Predicted by Merilo’s (1979) Correlation

\[
P = 1 \text{ MPa} \\
G = 2000 \text{ kg.m}^{-2}.\text{s}^{-1} \\
D = 0.01 \text{ m} \\
L/D = 200
\]

<table>
<thead>
<tr>
<th>(X_{\text{in}}) (%)</th>
<th>(CHF_{\text{pred}}) (kW/m(^2))</th>
<th>(X_{\text{exit}}) (%) (Calculated)</th>
</tr>
</thead>
<tbody>
<tr>
<td>-50</td>
<td>9601</td>
<td>141</td>
</tr>
<tr>
<td>-45</td>
<td>9082</td>
<td>135</td>
</tr>
<tr>
<td>-40</td>
<td>8574</td>
<td>130</td>
</tr>
<tr>
<td>-35</td>
<td>8078</td>
<td>125</td>
</tr>
<tr>
<td>-30</td>
<td>7593</td>
<td>121</td>
</tr>
<tr>
<td>-25</td>
<td>7120</td>
<td>116</td>
</tr>
<tr>
<td>-20</td>
<td>6659</td>
<td>112</td>
</tr>
<tr>
<td>-15</td>
<td>6210</td>
<td>108</td>
</tr>
<tr>
<td>-10</td>
<td>5773</td>
<td>105</td>
</tr>
<tr>
<td>-5</td>
<td>5349</td>
<td>101</td>
</tr>
<tr>
<td>0</td>
<td>4938</td>
<td>98</td>
</tr>
</tbody>
</table>


Table 6. Sources of CHF Data for Horizontal Flow

<table>
<thead>
<tr>
<th>Sources</th>
<th>Robertson*</th>
<th>Leontiev et al.</th>
<th>Becker</th>
<th>Merilo</th>
<th>Wang &amp; Su</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Fluids</strong></td>
<td>water</td>
<td>water</td>
<td>water</td>
<td>water</td>
<td>water</td>
</tr>
<tr>
<td><strong>Mass Flux</strong></td>
<td>680-</td>
<td>1000</td>
<td>160-</td>
<td>950-</td>
<td>300-</td>
</tr>
<tr>
<td>( (kg.m^{-2}.s^{-1}) )</td>
<td>1360</td>
<td>2000</td>
<td>1742</td>
<td>5700</td>
<td>1100</td>
</tr>
<tr>
<td><strong>X_{cr}</strong></td>
<td>0.16-0.55</td>
<td>0-0.55</td>
<td>0-0.71</td>
<td>0.1-0.7</td>
<td>0.65-0.98</td>
</tr>
<tr>
<td><strong>Pressure</strong></td>
<td>3.4-6.89</td>
<td>6.89-</td>
<td>1.0-3.2</td>
<td>6.89-</td>
<td>4.9</td>
</tr>
<tr>
<td>( (MPa) )</td>
<td>13.73</td>
<td></td>
<td></td>
<td>9.65</td>
<td></td>
</tr>
<tr>
<td><strong>Diameter</strong></td>
<td>12.7,</td>
<td>6.0</td>
<td>9.95,</td>
<td>12.6</td>
<td>12.0</td>
</tr>
<tr>
<td>( (mm) )</td>
<td>19.1</td>
<td></td>
<td>12.9</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Heated Length</strong></td>
<td>3.0</td>
<td>0.45</td>
<td>0.96</td>
<td>2.44,</td>
<td>2.0</td>
</tr>
<tr>
<td>( (m) )</td>
<td></td>
<td></td>
<td></td>
<td>3.66,</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>4.88</td>
<td></td>
</tr>
<tr>
<td><strong>Heat Flux</strong></td>
<td>200-</td>
<td>800-</td>
<td>750-</td>
<td>500-</td>
<td>116-</td>
</tr>
<tr>
<td>( (kW/m^2) )</td>
<td>1500</td>
<td>4450</td>
<td>4020</td>
<td>2500</td>
<td>698</td>
</tr>
<tr>
<td><strong>No. of Points</strong></td>
<td>52</td>
<td>23</td>
<td>82</td>
<td>34</td>
<td>19</td>
</tr>
</tbody>
</table>

* Some of the measurements were obtained with a sharp bend in the pipe just upstream of the test-section inlet.
Table 6. (Cont'd)

<table>
<thead>
<tr>
<th>Sources</th>
<th>Lis &amp; Strickland**</th>
<th>Kohler &amp; Hein</th>
<th>Rounthwaite</th>
<th>Ishigai et al.</th>
<th>Merilo &amp; Ahmad</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fluids</td>
<td>water</td>
<td>water</td>
<td>water</td>
<td>water</td>
<td>Freon-12</td>
</tr>
<tr>
<td>Mass Flux ($kg.m^{-2}.s^{-1}$)</td>
<td>95- 500, 2500</td>
<td>94- 214</td>
<td>95- 800</td>
<td>800 8100</td>
<td></td>
</tr>
<tr>
<td>$X_{cr}$</td>
<td>0.0-0.65</td>
<td>0.05-0.55</td>
<td>0.04-0.98</td>
<td>0.22-0.98</td>
<td>0.05-0.50</td>
</tr>
<tr>
<td>Pressure ($MPa$)</td>
<td>1.5-6.65</td>
<td>5, 10, 15, 20</td>
<td>1.48, 2.86</td>
<td>0.12-0.71</td>
<td>10.5, 15.2</td>
</tr>
<tr>
<td>Diameter ($mm$)</td>
<td>41.3</td>
<td>12.5, 24.3</td>
<td>41.3</td>
<td>5.6, 6.2, 9.95, 10.7</td>
<td>5.3</td>
</tr>
<tr>
<td>Heated Length ($m$)</td>
<td>3.0</td>
<td>6.0</td>
<td>11.6</td>
<td>0.3, 0.5, 0.75, 1.0</td>
<td>1.03-3.05</td>
</tr>
<tr>
<td>Heat Flux ($kW/m^2$)</td>
<td>63</td>
<td>150-600</td>
<td>25-66</td>
<td>300-4300</td>
<td>43-353</td>
</tr>
<tr>
<td>No. of Points</td>
<td>21</td>
<td>69</td>
<td>12</td>
<td>131</td>
<td>203</td>
</tr>
</tbody>
</table>

** Measurements were obtained with a sharp bend in the pipe just upstream of the test-section inlet.
Table 7. Dependence of Slip Ratio $S$ (Whalley, 1987)

<table>
<thead>
<tr>
<th>Correlation</th>
<th>$p_1/p_2$</th>
<th>$X$</th>
<th>$G$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Homogeneous model, $S = 1$</td>
<td>no</td>
<td>no</td>
<td>no</td>
</tr>
<tr>
<td>Zivi (1964)</td>
<td>yes</td>
<td>no</td>
<td>no</td>
</tr>
<tr>
<td>Chisholm (1972)</td>
<td>yes</td>
<td>yes</td>
<td>no</td>
</tr>
<tr>
<td>CISE (Premoli et al. 1970)</td>
<td>yes</td>
<td>yes</td>
<td>yes</td>
</tr>
<tr>
<td>Zuber et al. (1967)</td>
<td>yes</td>
<td>yes</td>
<td>yes</td>
</tr>
<tr>
<td>Bryce (1977)</td>
<td>yes</td>
<td>yes</td>
<td>yes</td>
</tr>
</tbody>
</table>

(steam-water only)
Table 8. Summary of Dimensionless Parameters

<table>
<thead>
<tr>
<th>Notation</th>
<th>Equation Form</th>
</tr>
</thead>
<tbody>
<tr>
<td>$T_1$</td>
<td>$C_1 Re_{L}^{-0.2} \left( \frac{1-X}{1-\alpha} \right)^2 \frac{G^2}{gD_{p} (\rho_f - \rho_g) \alpha^{0.5}}$ (Eq. 5.22)</td>
</tr>
<tr>
<td>$T_2$</td>
<td>$C_2 \left( \frac{\rho_f}{\rho_f} \right) \frac{G^4}{gD_{p}^2 (\rho_f - \rho_g) \alpha^{0.5}} \left[ \frac{\rho_f (1-\alpha) X - \alpha (1-X)}{\alpha (1-\alpha)} \right]^4$ (Eq. 5.24)</td>
</tr>
<tr>
<td>$T_3$</td>
<td>$C_3 Re_{L}^{0.2} \rho_g u_f^2 \frac{(u_g - u_f)^2}{g(\rho_f - \rho_g)}$ (Eq. 5.29)</td>
</tr>
<tr>
<td>$T_4$</td>
<td>$C_4 \left( \frac{\rho_f}{\rho_f} \right) \frac{G^4}{gD_{p}^2 (\rho_f - \rho_g) \alpha^{0.5}} \left[ \frac{\rho_f (1-\alpha) X - \alpha (1-X)}{\alpha (1-\alpha)} \right]^4$ (Eq. 5.34)</td>
</tr>
<tr>
<td>$T_5$</td>
<td>$\frac{G}{\sqrt{gD_{p} (\rho_f - \rho_g)}} \left( \frac{X}{\alpha} \right) \left( \frac{D}{L_h} \right)$ (Eq. 5.42)</td>
</tr>
<tr>
<td>$T_6$</td>
<td>$\frac{G}{\sqrt{gD_{p} (\rho_f - \rho_g)}} \left( \frac{1-X}{1-\alpha} \right) \left( \frac{D}{L_h} \right)$ (Eq. 5.50)</td>
</tr>
</tbody>
</table>

$T_1$ - Turbulent/buoyant force (Annular)

$T_2$ - Drag/Buoyant force (Bubbly)

$T_3$ - Turbulent/Buoyant force (Bubbly)

$T_4$ - Drag/Buoyant force (Droplet)

$T_5$ - Transit time ratio (Bubbly)

$T_6$ - Transit time ratio (Droplet)
Table 9. Comparison of Horizontal Water CHF Data with Eq. (6.7) for Various Parameters and Constants for Local Parameters Approach

<table>
<thead>
<tr>
<th>Equation</th>
<th>Mean Error</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of Data** within prediction accuracy of</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 - $e^{-\left(T_1/50^0\right)^{0.5}}$</td>
<td>0.033</td>
<td>0.281</td>
<td>0.279</td>
<td>0.2832 0.5484 0.9212</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_1/55^0\right)^{0.5}}$</td>
<td>0.038</td>
<td>0.269</td>
<td>0.269</td>
<td>0.2545 0.5771 0.9212</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_1/55^0\right)^{0.5}}$</td>
<td>0.038</td>
<td>0.269</td>
<td>0.269</td>
<td>0.2545 0.5771 0.9427</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_1/65^0\right)^{0.5}}$</td>
<td>-0.048</td>
<td>0.256</td>
<td>0.352</td>
<td>0.2581 0.5305 0.9498</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_1/70^0\right)^{0.5}}$</td>
<td>-0.069</td>
<td>0.255</td>
<td>0.245</td>
<td>0.2617 0.5233 0.9606</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_3/20^0\right)^{0.5}}$</td>
<td>-0.002</td>
<td>0.471</td>
<td>0.471</td>
<td>0.2617 0.4480 0.6810</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_3/0.06^0\right)^{0.9}}$</td>
<td>-0.065</td>
<td>0.267</td>
<td>0.259</td>
<td>0.3298 0.5448 0.9319</td>
</tr>
<tr>
<td>1 - $e^{-\left(T_3/0.065^0\right)^{0.8}}$</td>
<td>-0.066</td>
<td>0.264</td>
<td>0.255</td>
<td>0.3298 0.5914 0.9319</td>
</tr>
</tbody>
</table>

+ Mean of $\frac{CHF_{\text{pred}}-CHF_{\text{exp}}}{CHF_{\text{exp}}}$

** No. of data point = 279
Table 10. Comparison of Horizontal Water CHF Data with Eq. (6.7) for Various Parameters and Constants for Constant Inlet Conditions

<table>
<thead>
<tr>
<th>Equation</th>
<th>Mean+ Error</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of Data++ within prediction accuracy of</th>
</tr>
</thead>
<tbody>
<tr>
<td>$1 - e^{-(T_1/50)^{0.3}}$</td>
<td>0.027</td>
<td>0.169</td>
<td>-0.167</td>
<td>0.4660 0.7133 1.0000</td>
</tr>
<tr>
<td>$1 - e^{-(T_1/55)^{0.3}}$</td>
<td>0.005</td>
<td>0.163</td>
<td>0.163</td>
<td>0.4444 0.7419 1.0000</td>
</tr>
<tr>
<td>$1 - e^{-(T_1/55)^{0.3}}$</td>
<td>-0.013</td>
<td>0.159</td>
<td>0.158</td>
<td>0.4050 0.7563 1.0000</td>
</tr>
<tr>
<td>$1 - e^{-(T_1/65)^{0.3}}$</td>
<td>-0.032</td>
<td>0.158</td>
<td>0.155</td>
<td>0.4229 0.7850 1.0000</td>
</tr>
<tr>
<td>$1 - e^{-(T_1/70)^{0.3}}$</td>
<td>-0.046</td>
<td>0.159</td>
<td>0.153</td>
<td>0.4229 0.7957 1.0000</td>
</tr>
<tr>
<td>$1 - e^{-(T_1/70)^{0.3}}$</td>
<td>-0.054</td>
<td>0.421</td>
<td>0.418</td>
<td>0.4353 0.5020 0.7412</td>
</tr>
<tr>
<td>$1 - e^{-(T_3/0.06)^{0.9}}$</td>
<td>-0.067</td>
<td>0.266</td>
<td>0.258</td>
<td>0.4029 0.6439 0.9281</td>
</tr>
<tr>
<td>$1 - e^{-(T_3/0.065)^{0.8}}$</td>
<td>-0.053</td>
<td>0.243</td>
<td>0.237</td>
<td>0.4532 0.6655 0.9496</td>
</tr>
</tbody>
</table>

+ Mean of $\frac{CHF_{pred} - CHF_{exp}}{CHF_{exp}}$
++ No. of data point = 279
Table 11. Comparison of Horizontal Water CHF Data with Various Correlations for Local Parameters Approach

<table>
<thead>
<tr>
<th>Equation</th>
<th>Mean Error</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of Data** within</th>
</tr>
</thead>
<tbody>
<tr>
<td>Eq. (6.2)</td>
<td>-0.329</td>
<td>0.438</td>
<td>0.288</td>
<td>0.1756</td>
</tr>
<tr>
<td>Eq. (6.3)</td>
<td>-0.007</td>
<td>0.390</td>
<td>0.390</td>
<td>0.2043</td>
</tr>
<tr>
<td>Eq. (6.6)</td>
<td>-0.018</td>
<td>0.292</td>
<td>0.292</td>
<td>0.2545</td>
</tr>
<tr>
<td>Eq. (6.9)</td>
<td>-0.048</td>
<td>0.256</td>
<td>0.252</td>
<td>0.2581</td>
</tr>
<tr>
<td>Wang &amp; Su</td>
<td></td>
<td></td>
<td></td>
<td>Prediction is not valid due to the wrong parametric trend of the correlation</td>
</tr>
</tbody>
</table>

+ Mean of $\frac{CHFR_{pred} - CHFR_{exp}}{CHFR_{exp}}$

++ No. of data point = 279
Table 12. Comparison of Horizontal Water CHF Data with Various Correlations for Constant Inlet Conditions

<table>
<thead>
<tr>
<th>Equation</th>
<th>Mean†</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of Data++ within Prediction Accuracy of</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>10%</td>
</tr>
<tr>
<td>Eq. (6.2)</td>
<td>-0.224</td>
<td>0.304</td>
<td>0.206</td>
<td>0.2760</td>
</tr>
<tr>
<td>Eq. (6.3)</td>
<td>-0.023</td>
<td>0.162</td>
<td>0.161</td>
<td>0.3620</td>
</tr>
<tr>
<td>Eq. (6.6)</td>
<td>-0.115</td>
<td>0.424</td>
<td>0.408</td>
<td>0.4022</td>
</tr>
<tr>
<td>Eq. (6.9)</td>
<td>-0.032</td>
<td>0.158</td>
<td>0.155</td>
<td>0.4229</td>
</tr>
<tr>
<td>Merilo (1979)</td>
<td>0.288</td>
<td>0.803</td>
<td>0.620</td>
<td>0.3513</td>
</tr>
<tr>
<td>Wang &amp; Su</td>
<td>Prediction is not valid due to the wrong parametric trend of the correlation</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

† Mean of \( \frac{CHFR_{pred} - CHFR_{exp}}{CHFR_{exp}} \)

++ No. of data point = 279
Table 13. Comparison of Horizontal Freon-12 CHF Data with Various Correlations for Local Parameters Approach

<table>
<thead>
<tr>
<th>Equation</th>
<th>Mean Error</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of Data within 10%</th>
<th>Fraction of Data within 20%</th>
<th>Fraction of Data within 50%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Eq. (6.2)</td>
<td>+0.173</td>
<td>0.265</td>
<td>0.201</td>
<td>0.2020</td>
<td>0.5271</td>
<td>0.9409</td>
</tr>
<tr>
<td>Eq. (6.3)</td>
<td>+0.153</td>
<td>0.252</td>
<td>0.201</td>
<td>0.1921</td>
<td>0.4975</td>
<td>0.9606</td>
</tr>
<tr>
<td>Eq. (6.6)</td>
<td>-0.283</td>
<td>0.367</td>
<td>0.232</td>
<td>0.1773</td>
<td>0.3251</td>
<td>0.7931</td>
</tr>
<tr>
<td>Eq. (6.9)</td>
<td>+0.111</td>
<td>0.264</td>
<td>0.239</td>
<td>0.1921</td>
<td>0.4877</td>
<td>0.9704</td>
</tr>
<tr>
<td>Wang &amp; Su</td>
<td>Prediction is not valid due to the wrong parametric trend of the correlation</td>
<td></td>
<td></td>
<td></td>
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+ Mean of \( \frac{CHFR_{pred} - CHFR_{aspt}}{CHFR_{aspt}} \)

++ No. of data point = 203
Table 14. Comparison of Horizontal Freon-12 CHF Data with Various Correlations for Constant Inlet Condition Approach

<table>
<thead>
<tr>
<th>Equation</th>
<th>Mean Error</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of Data++ within Prediction Accuracy of</th>
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<td>Eq. (6.3)</td>
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<td>0.175</td>
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<td>Eq. (6.6)</td>
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<td>0.727</td>
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<tr>
<td>Eq. (6.9)</td>
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<td>0.172</td>
<td>0.149</td>
<td>0.3202</td>
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<tr>
<td>Wang &amp; Su</td>
<td>Prediction is not valid due to the wrong parametric trend of the correlation</td>
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++ Mean of $\frac{CHFR_{pred} - CHFR_{exp}}{CHFR_{exp}}$

++ No. of data point = 203
Table 15. Comparison of Horizontal Water CHF Data with The Combination Approach

| Prediction Method | Mean $^+$ | RMS Error | Standard Deviation | Fraction of Data $^{++}$ within  \
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<th></th>
</tr>
</thead>
<tbody>
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<td>10%</td>
</tr>
<tr>
<td>Constant $X_D$</td>
<td></td>
<td></td>
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</tr>
<tr>
<td>Maximum of</td>
<td></td>
<td></td>
<td></td>
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</tr>
<tr>
<td>Eq. (2.25) &amp; (6.9)</td>
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<td>0.247</td>
<td>0.247</td>
<td>0.2760</td>
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<td>Constant $X_{in}$</td>
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$^+$ Mean of $\frac{CHF_{pred} - CHF_{exp}}{CHF_{exp}}$

$^{++}$ No. of data point = 279
Figure 1. Typical Boiling Curve for Two Phase Flow.
Figure 2. Flow Regimes for Vertical Flow.
Figure 3. Flow Regimes for Horizontal Flow.
Figure 4. Near-wall Bubble Crowding and Vapor Blanketing Mechanism for Burnout Heat Flux in Subcooled and Low-quality Flows.
Figure 5. Local Overheating Following Bubble Growth from a Nucleation Center.

Figure 6. Burnout Associated with Vapor Clot or Slug Formation.
Figure 7. 'Dryout' in Vertical Tube Caused by Annular Liquid Film Failure (Collier, 1984).
Figure 8. Flow Regimes for Horizontal Flow (Becker, 1971).
Figure 9. Horizontal Annular Flow in Air and Water (Fisher, 1978).
Figure 10. Modified Generalized Flow Regime Map for Horizontal Flow.
Figure 11. Comparison of Vertical and Horizontal CHF with Water as Coolant at a Pressure of 9.65 MPa (Merilo, 1977).
Figure 12. Critical Heat Flux at Subcooled Condition (Becker, 1971).
Figure 13. Critical Heat Flux as a Function of Quality for Both Vertical and Horizontal Flow.
Figure 14. The Effect of Mass Flux on Vertical and Horizontal CHF with Freon-12 at a Pressure of 1.05 MPa (Merilo, 1977).
Figure 15. The Effect of Diameter on CHF in Vertical and Horizontal Flow on a Local Conditions Basis (Merilo and Ahmad, 1979).
Figure 16. Effect of Pressure on Horizontal CHF (Robertson, 1973).
Figure 17. Schematic Diagram Showing Possible Enthalpy Changes between Any Two States (Reid et al., 1977).
Figure 18. Onset of Bubble Detachment in Flow Boiling.
Figure 19. Comparison of $K_\theta$ Prediction against Experimental CHF Ratio (Eq. 2.21).
Figure 20. Difference of the Critical Steam Quality as a Function of the Froude Number Defined in Eq. (5.5) (Kholer and Hein, 1986).
Figure 21. Interface Perimeters for Annular Flow and Equilibrium Stratified Flow.
Figure 22a. Parametric Trend of $T_1$ with Quality.
Figure 22b. Parametric Trend of $T_1$ with Pressure.

\[ G = 1000 \text{ kg.m}^{-1}.\text{s}^{-1} \]
\[ D = 0.01 \text{ m} \]
Figure 23. Secondary Vapor Flow and Upward Liquid Flow in a Cross-sectional View of a Horizontal Annular Flow.
Figure 24. Prediction Error vs. Experimental CHF Ratio (Eq. 2.25).
Figure 25. Prediction Error vs. Experimental CHF Ratio (Eq. 6.3).
Figure 26a. Experimental CHF Ratio vs. $Fr$. 
Figure 26b. Experimental CHF ratio vs. $Fr^*$. 
Figure 27. Comparison of Experimental and Predicted CHF Ratio (Eq. 6.6).
Figure 28. Prediction Error vs. CHF Ratio (Eq. 6.6).
Figure 29. Prediction Error vs. Quality (Eq. 6.6).
Figure 30. CHF Ratio as a Function of $T_1$. 
Figure 31. CHF ratio as a Function of $T_3$. 
Figure 32. CHF Ratio as a Function of $T_5$. 
Figure 33. CHF Ratio as a Function of ($T_1 \times T_3$).
Figure 34. CHF Ratio as a Function of $(T_1 \times T^2_5)$. 

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Figure 35. CHF Ratio as a Function of \((T_3 \times T_5)\).
Figure 36. Variation of $A$ with Martinelli's Parameter.
Figure 37a. Predicted CHF Ratio as a Function of Pressure, $D = 0.01\,m$ and $X_{cr} = 0.30$. 
Figure 37b. Predicted CHF Ratio as a Function of Pressure,

\[ D = 0.01m \text{ and } G = 1500kg.m^{-2}.s^{-1}. \]
Figure 38a. Predicted CHF Ratio as a Function of Mass Flux,
$D = 0.01\,\text{m}$ and $P = 5000\,\text{kPa}$. 
Figure 38b. Predicted CHF Ratio as a Function of Mass Flux, $D = 0.01\text{m}$ and $X_{cr} = 0.30$. 
Figure 39a. Predicted CHF Ratio as a Function of Quality, \( D = 0.01m \) and \( G = 1500kg.m^{-2}.s^{-1} \).
Figure 39b. Predicted CHF Ratio as a Function of Quality,
$D = 0.01m$ and $P = 5000kPa$. 
Figure 40. Predicted CHF Ratio as a Function of Diameter, $P = 5000kPa$, $G = 1500kg.m^{-2}.s^{-1}$ and $X_{cr} = 0.30$. 


Energy balance is calculated from the following equation:

$$X = \frac{4q}{H_{fg} G} \times \frac{L}{D} - \frac{\Delta H_{in}}{H_{fg}}$$

Figure 41a. Error Based on Constant Exit and Inlet Conditions (Horizontal High Flow).
*Energy balance is calculated from the following equation

\[ X = \frac{4q}{H_{fg} G} \times \frac{L}{D} - \frac{\Delta H_{in}}{H_{fg}} \]

Figure 41b. Error Based on Constant Exit and Inlet Conditions (Horizontal Low Flow).
Figure 42. Comparison of Predicted and Experimental CHF for Water using Eq. 6.9.
Figure 43. Comparison of Predicted and Experimental CHF for Freon-12 using Eq. 6.9.
Figure 44. Actual Quality as Predicted by Saha-Zuber's Correlation (High Pressure and High Heat Flux).
Figure 45. CHF Ratio Prediction as a Function of Mass Flux.
Appendix A

CHF DATA BANK

Data for CHF in horizontal tubes were obtained from Merilo (1977), Merilo and Ahmad (1979), Robertson (1973), Becker (1971), Leontiev et al. (1981), Lis and Strickland (1970), Kohler and Hein (1986) Wang and Su (1980), Routhwaite (1968) and Ishigui et al. (1970a). Most CHF data for vertical tubes were taken from the table look-up method (Groeneveld et al., 1986a) and whenever possible, experimental data were recorded. The data index for $CHF_{ver}$ is 1 for experimental data and 2 for value taken from table look-up method. Table A-1 shows the data bank tabulated into groups according to fluids (water or Freon-12), sources of experiments and heated length. Within each group, the data are arranged into columns of mass flux ($G, kg.m^{-2}.s^{-1}$), critical quality ($X_{cr}$), pressure ($P, MPa$), tube diameter ($D, m$), $CHF_{hor} (kW.m^{-2})$, $CHF_{ver} (kW.m^{-2})$ and data index. The quantities tabulated are explained below:

1) $L_h/D$  Heated length over tube diameter ratio
2) $G$  mass flux ($kg.m^{-2}.s^{-1}$)
3) $X_{cr}$  Quality at critical heat flux
4) $P$  Pressure ($MPa$)
5) $D$  Tube diameter ($m$)
6) $CHF_{hor}$  critical heat flux for horizontal tube
flow obtained from experiment ($kW.m^{-2}$)
7) $CHF_{ver}$  critical heat flux for vertical flow ($kW.m^{-2}$)
8) $CHFR$  
   critical heat flux ratio, $CHF_{hor}/CHF_{ver}$

9) I  
   data index for $CHF_{ver}$
   1—experimental
   2—table look-up method

Some of the horizontal CHF data were much greater than the corresponding vertical CHF data. This is not likely to be correct from the consideration of physical mechanisms involved in horizontal CHF. The discrepancies could be due to experimental uncertainties or the inaccuracies in predicting the vertical CHF's for certain flow conditions. Those data for which $CHFR > 1.3$ were screened because they are not considered to be correct.
Table A-1  Data Bank for \( \text{in Horizontal Flow} \)

Fluid: Water
Source: Robertson (1973)

\[
L_i/D = 160
\]

<table>
<thead>
<tr>
<th>( G ) ((kg\cdot m^{-2}\cdot s^{-1}))</th>
<th>( X_{cr} )</th>
<th>( P ) ((MPa))</th>
<th>( D ) ((m))</th>
<th>( CHF_{hor} ) ((kW\cdot m^{-2}))</th>
<th>( CHF_{ver} ) ((kW\cdot m^{-2}))</th>
<th>( CHFR )</th>
<th>( i )</th>
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Table A-1. (Cont'd)
Fluid: Water
Source: Robertson (1973)

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<th>$P$ (MPa)</th>
<th>$D$ (m)</th>
<th>$CHF_{hor}$ (kW.m$^{-2}$)</th>
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$L_h/D = 239$

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<th>$D$ (m)</th>
<th>$CHF_{hor}$ (kW.m$^{-2}$)</th>
<th>$CHF_{ver}$ (kW.m$^{-2}$)</th>
<th>$CHF_{R}$</th>
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Table A-1 (Continue)

Fluid: Water
Source: Becker (1971)

\[ Lh/D = 96.5 \]

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Table A-1 (Continue)  
Fluid: Water  
Source: Becker (1971)

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L_{f}/D = 96.5
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Table A-1 (Continue)
Fluid: Water
Source: Becker (1971)

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Table A-1 (Continue)
Fluid: Water (Subcooled Data)
Source: Becker (1971)

\[ \frac{L_{h/D}}{D} = 74.4 \]

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<th>( X_{cr} )</th>
<th>( P ) (MPa)</th>
<th>( D ) (m)</th>
<th>( CHF_{hor} ) (kW ( \cdot m^{-2} ))</th>
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\[ \frac{L_{h/D}}{D} = 96.5 \]

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<th>( D ) (m)</th>
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<th>( CHF_{ver} ) (kW ( \cdot m^{-2} ))</th>
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Table A-1 (Continue)
Fluid: Water
Source: Merilo (1977)

\[
\frac{L_h}{D} = 194.0
\]

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<th>(D) ((m))</th>
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\[
\frac{L_h}{D} = 290
\]

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| 950. | 0.6000 | 6.8900 | 0.01260 | 777. | 1650. | 0.47091 | 2 |
| 1360. | 0.4000 | 6.8900 | 0.01260 | 970. | 2103. | 0.46125 | 2 |
| 1360. | 0.4500 | 6.8900 | 0.01260 | 870. | 1831. | 0.47515 | 2 |
| 3800. | 0.2000 | 6.8900 | 0.01260 | 2020. | 2158. | 0.93605 | 1 |
| 3800. | 0.2200 | 6.8900 | 0.01260 | 1980. | 2080. | 0.95192 | 1 |
| 3800. | 0.2400 | 6.8900 | 0.01260 | 1880. | 2000. | 0.94000 | 1 |
| 3800. | 0.2600 | 6.8900 | 0.01260 | 1770. | 1920. | 0.92187 | 1 |
| 950. | 0.4000 | 9.6500 | 0.01260 | 724. | 1198. | 0.60434 | 2 |
| 950. | 0.4500 | 9.6500 | 0.01260 | 671. | 1069. | 0.62769 | 2 |
| 3800. | 0.1200 | 9.6500 | 0.01260 | 1824. | 1838. | 0.99238 | 2 |
| 3800. | 0.1500 | 9.6500 | 0.01260 | 1618. | 1750. | 0.92457 | 1 |
| 3800. | 0.1700 | 9.6500 | 0.01260 | 1474. | 1650. | 0.89333 | 1 |
| 5700. | 0.1000 | 9.6500 | 0.01260 | 1960. | 1965. | 0.99746 | 1 |
Table A-1 (Continue)
Fluid: Water
Source: Merilo (1977)

\[ L_{h/D} = 387.0 \]

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<th>( \text{CHF}_{\text{ver}} ) ( (\text{kW} \cdot \text{m}^{-2}) )</th>
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Table A-1 (Continue)

Fluid: Water

Source: Kohler and Hein (1986)

\[ L_{h/D} = 488.0 \]

| \( G \) (kg.m\(^{-2}\).s\(^{-1}\)) | \( X_{cr} \) | \( P \) (MPa) | \( D \) (m) | \( CHF_{hor} \) (kW.m\(^{-2}\)) | \( CHF_{ver} \) (kW.m\(^{-2}\)) | \( CHFR \) | \( l \)
<table>
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<tr>
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Table A-1 (Continue)
Fluid: Water
Source: Kohler and Hein (1986)

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Table A-1 (Continue)

Fluid: Water

Source: Kohler and Hein (1986)

\[
\frac{L_h}{D} = 251.0
\]

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### Table A-1 (Continue)

**Fluid: Water**

*Source: Rounthwaite (1968)*

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$L_0/D = 280.0$
Table A-1 (Continue)

Fluid: Water
Source: Lis and Strickland (1970)

\[ \frac{L_f}{D} = 73.2 \]

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Table A.1 (Continue)

Fluid: Water


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Table A-1 (Continue)

Fluid: Water

Source: Ishigai et al. (1970a)

\[ L_{h}/D = 93.5 \]

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Fluid: Water
Source: Ishigak et al. (1970a)

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<td>0.00995</td>
<td>1454.</td>
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<td>814.</td>
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</table>

L<sub>h</sub>/D = 121.0

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<tr>
<th>G</th>
<th>X&lt;sub&gt;cr&lt;/sub&gt;</th>
<th>P</th>
<th>D</th>
<th>CHF&lt;sub&gt;hor&lt;/sub&gt;</th>
<th>CHF&lt;sub&gt;ver&lt;/sub&gt;</th>
<th>CHFR</th>
<th>l</th>
</tr>
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<tr>
<td>(kg·m&lt;sup&gt;-2&lt;/sup&gt;·s&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>(MPa)</td>
<td>(m)</td>
<td>(kW·m&lt;sup&gt;-2&lt;/sup&gt;)</td>
<td>(kW·m&lt;sup&gt;-2&lt;/sup&gt;)</td>
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Table A-1 (Continue)

Fluid: Water

Source: Ishigai et al. (1970a)

\[ L_{h}/D = 121.0 \]

<table>
<thead>
<tr>
<th>( G ) (kg m(^{-2}) s(^{-1}))</th>
<th>( X_{cr} ) (MPa)</th>
<th>( P ) (MPa)</th>
<th>( D ) (m)</th>
<th>( CHF_{hor} ) (kW m(^{-2}))</th>
<th>( CHF_{wcr} ) (kW m(^{-2}))</th>
<th>( CHF_{R} ) (kW m(^{-2}))</th>
<th>( I )</th>
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</table>

\[ L_{h}/D = 89.6 \]

| 796 | 0.333 | 0.47623 | 0.00560 | 2675 | 1865 | 1.43420 | 2 |
| 787 | 0.419 | 0.47623 | 0.00560 | 2791 | 1426 | 1.95777 | 2 |
| 779 | 0.486 | 0.47623 | 0.00560 | 2907 | 1090 | 2.66753 | 2 |
| 766 | 0.528 | 0.47623 | 0.00560 | 2849 | 961 | 2.96529 | 2 |
| 753 | 0.544 | 0.47623 | 0.00560 | 2675 | 932 | 2.87114 | 2 |
| 742 | 0.544 | 0.47623 | 0.00560 | 2500 | 963 | 2.59725 | 2 |
| 733 | 0.480 | 0.47623 | 0.00560 | 1977 | 1235 | 1.60143 | 2 |
| 596 | 0.368 | 0.47623 | 0.00560 | 2093 | 2290 | 0.91427 | 2 |
| 587 | 0.409 | 0.47623 | 0.00560 | 1977 | 2200 | 0.89888 | 2 |
| 581 | 0.512 | 0.47623 | 0.00560 | 2210 | 1684 | 1.31246 | 2 |
| 575 | 0.652 | 0.47623 | 0.00560 | 2559 | 825 | 3.10113 | 2 |
| 563 | 0.665 | 0.47623 | 0.00560 | 2384 | 752 | 3.17145 | 2 |
| 555 | 0.634 | 0.47623 | 0.00560 | 2152 | 959 | 2.24439 | 2 |
| 550 | 0.558 | 0.47623 | 0.00560 | 1686 | 1487 | 1.13376 | 2 |
| 393 | 0.744 | 0.47623 | 0.00560 | 2152 | 589 | 3.65463 | 2 |
| 390 | 0.606 | 0.47623 | 0.00560 | 2144 | 1534 | 1.13730 | 2 |
| 385 | 0.733 | 0.47623 | 0.00560 | 1949 | 637 | 3.01159 | 2 |
| 382 | 0.785 | 0.47623 | 0.00560 | 1977 | 479 | 4.12386 | 2 |
| 379 | 0.708 | 0.47623 | 0.00560 | 1744 | 723 | 2.41140 | 2 |
| 374 | 0.748 | 0.47623 | 0.00560 | 1744 | 605 | 2.88567 | 2 |
| 368 | 0.763 | 0.47623 | 0.00560 | 1686 | 585 | 2.98251 | 2 |
### Table A-1 (Continue)
**Fluid: Water**

Source: Ishigai et al. (1970a)

<table>
<thead>
<tr>
<th>$G$ (kg m$^{-2}$s$^{-1}$)</th>
<th>$X_{cr}$</th>
<th>$P$ (MPa)</th>
<th>$D$ (m)</th>
<th>CHF$_{hor}$ (kW m$^{-2}$)</th>
<th>CHF$_{ver}$ (kW m$^{-2}$)</th>
<th>CHFR</th>
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<tbody>
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<tr>
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<td>0.00560</td>
<td>3373.</td>
<td>1923.</td>
<td>1.76384</td>
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<td>1835.</td>
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$L_h/D = 89.2$

$L_h/D = 161.0$

170
Table A-1 (Continue)

**Fluid: Water**

*Source: Ishigai et al. (1970a)*

<table>
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<tr>
<th>(G) ((\text{kg}\cdot\text{m}^{-2}\cdot\text{s}^{-1}))</th>
<th>(X_{cr})</th>
<th>(P) ((\text{MPa}))</th>
<th>(D) ((\text{m}))</th>
<th>(\text{CHF}_{\text{hor}}) ((\text{kw}\cdot\text{m}^{-2}))</th>
<th>(\text{CHF}_{\text{ver}}) ((\text{kw}\cdot\text{m}^{-2}))</th>
<th>(\text{CHFR})</th>
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<tbody>
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<td>0.94048</td>
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\(L_{h/D} = 121.0\)

| 797. | 0.524 | 0.70927 | 0.00620 | 2733. | 1275. | 2.14326 | 2 |
| 767. | 0.606 | 0.70927 | 0.00620 | 2442. | 724. | 3.37353 | 2 |
| 729. | 0.671 | 0.70927 | 0.00620 | 2152. | 603. | 8.57040 | 2 |
| 597. | 0.655 | 0.70927 | 0.00620 | 2384. | 893. | 2.66994 | 2 |
| 576. | 0.712 | 0.70927 | 0.00620 | 2098. | 625. | 3.34908 | 2 |
| 547. | 0.653 | 0.70927 | 0.00620 | 1570. | 1002. | 1.56669 | 2 |
| 398. | 0.780 | 0.70927 | 0.00620 | 1803. | 641. | 2.81200 | 2 |
| 384. | 0.854 | 0.70927 | 0.00620 | 1628. | 539. | 3.02109 | 2 |
| 365. | 0.802 | 0.70927 | 0.00620 | 1279. | 605. | 2.11316 | 2 |

\(L_{h/D} = 65.4\)

| 135. | 0.643 | 0.11652 | 0.01530 | 768. | 735. | 1.04383 | 2 |
| 134. | 0.445 | 0.11652 | 0.01530 | 488. | 830. | 0.58815 | 2 |
| 134. | 0.591 | 0.11652 | 0.01530 | 605. | 795. | 0.76093 | 2 |
| 134. | 0.657 | 0.11652 | 0.01530 | 640. | 715. | 0.89443 | 2 |
| 127. | 0.613 | 0.11652 | 0.01530 | 721. | 763. | 0.94489 | 2 |
| 125. | 0.597 | 0.11652 | 0.01530 | 663. | 780. | 0.85028 | 2 |
| 124. | 0.467 | 0.11652 | 0.01530 | 477. | 809. | 0.58921 | 2 |
| 124. | 0.504 | 0.11652 | 0.01530 | 500. | 801. | 0.62432 | 2 |
| 124. | 0.649 | 0.11652 | 0.01530 | 628. | 711. | 0.88274 | 2 |
| 124. | 0.667 | 0.11652 | 0.01530 | 605. | 687. | 0.88051 | 2 |
| 96. | 0.682 | 0.11652 | 0.01530 | 570. | 617. | 0.92396 | 2 |
| 96. | 0.582 | 0.11652 | 0.01530 | 465. | 737. | 0.63146 | 2 |
| 96. | 0.517 | 0.11652 | 0.01530 | 395. | 755. | 0.52397 | 2 |
| 96. | 0.561 | 0.11652 | 0.01530 | 407. | 742. | 0.54824 | 2 |
| 96. | 0.660 | 0.11652 | 0.01530 | 488. | 646. | 0.75626 | 2 |
Table A-1 (Continue)
Fluid: Water
Source: Ishigai et al. (1970a)

<table>
<thead>
<tr>
<th>$G$ (kg.m$^{-2}$.s$^{-1}$)</th>
<th>$X_{cr}$</th>
<th>$P$ (MPa)</th>
<th>$D$ (m)</th>
<th>CHF$_{hor}$ (kW.m$^{-2}$)</th>
<th>CHF$_{ver}$ (kW.m$^{-2}$)</th>
<th>CHFR</th>
<th>$l$</th>
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<td>96.</td>
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<td>0.01530</td>
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<td>470.</td>
<td>1.16406</td>
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<tr>
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<td>0.01530</td>
<td>500.</td>
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$L_t/D = 65.4$
Table A-1  (Continue)
Fluid: Freon-12
Source: Merilo and Ahmad (1979)

\[
L_{iH/D} = 194.0
\]

<table>
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<th>G</th>
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<th>P</th>
<th>D</th>
<th>CHF_{hor}</th>
<th>CHF_{ver}</th>
<th>CHFR</th>
<th>I</th>
</tr>
</thead>
<tbody>
<tr>
<td>(kg.m(^{-2}).s(^{-1}))</td>
<td>(MPa)</td>
<td>(m)</td>
<td>(kW.m(^{-2}))</td>
<td>(kW.m(^{-2}))</td>
<td></td>
<td></td>
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</tr>
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<td>1.0500</td>
<td>0.00530</td>
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<td>167.</td>
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</tr>
<tr>
<td>1600.</td>
<td>0.3000</td>
<td>1.0500</td>
<td>0.00530</td>
<td>97.</td>
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<td>0.55232</td>
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</tr>
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<td>0.2800</td>
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<td>113.</td>
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<td>1.0500</td>
<td>0.00530</td>
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<td>180.</td>
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<td>1.0500</td>
<td>0.00530</td>
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<td>1.0500</td>
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Table A-1 (Continue)

Fluid: Freon-12

Source: Merilo and Ahmad (1979)

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\[ L_{h/D} = 290.0 \]

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Table A-1 (Continue)
Fluid: Freon-12
Source: Merilo and Ahmad (1979)

\[
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(kg.m^{-2}.s^{-1}) & (MPa) & (m) & (kW.m^{-2}) & (kW.m^{-2}) & & & \\
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1600 & 0.2800 & 1.5200 & 0.00530 & 90 & 107 & 0.83883 & 2 \\
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\end{array}
\]
### Table A.5 (Continue)

**Fluid:** Freon-12  
**Source:** Merilo and Ahmad (1979)

\[
\frac{L_H}{D} = 290.0
\]

| \(G\)  | \(X_{cr}\) | \(P\) (MPa) | \(D\) (m) | \(CHF_{tor}\) (kW m\(^{-2}\)) | \(CHF_{ver}\) (kW m\(^{-2}\)) | \(CHFR\) |  
|-------|----------|-------------|---------|-----------------|-----------------|--------|---------|  
| 6800. | 0.1980   | 1.5200      | 0.00530 | 162.            | 203.            | 0.79732| 2       |  
| 6800. | 0.1750   | 1.5200      | 0.00530 | 165.            | 206.            | 0.80108| 2       |  
| 6800. | 0.0830   | 1.5200      | 0.00530 | 241.            | 257.            | 0.93657| 2       |  
| 6800. | 0.0650   | 1.5200      | 0.00530 | 300.            | 278.            | 1.08085| 2       |  
| 8100. | 0.2650   | 1.5200      | 0.00530 | 176.            | 244.            | 0.72107| 2       |  
| 8100. | 0.2640   | 1.5200      | 0.00530 | 181.            | 244.            | 0.74128| 2       |  
| 8100. | 0.1660   | 1.5200      | 0.00530 | 189.            | 243.            | 0.77683| 2       |  
| 8100. | 0.0980   | 1.5200      | 0.00530 | 220.            | 276.            | 0.79672| 2       |  
| 8100. | 0.0770   | 1.5200      | 0.00530 | 280.            | 298.            | 0.94070| 2       |  

\[
\frac{L_H}{D} = 389.0
\]

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176
| \( G \) (kg\,m\(^{-2}\),s\(^{-1}\)) | \( X_{cr} \) | \( P \) (MPa) | \( D \) (m) | \( \text{CHF}_{\text{hor}} \) (kW\,m\(^{-2}\)) | \( \text{CHF}_{\text{ver}} \) (kW\,m\(^{-2}\)) | \( \text{CHFR} \) | \( I \) |
|---|---|---|---|---|---|---|
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| 8100. | 0.1900 | 1.0500 | 0.00530 | 172. | 203. | 0.84768 | 2 |
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| 8100. | 0.1200 | 1.0500 | 0.00530 | 183. | 230. | 0.79473 | 2 |
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**Table A-1 (Continue)**

**Fluid: Freon-12**

**Source: Merilo and Ahmad (1979)**

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L<sub>h/D</sub> = 575.0
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Source: Merilo and Ahmad (1979)

\[ L_h/D = 575.0 \]

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\[ L/h/D = 575.0 \]

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Appendix B

FLUID PROPERTIES EVALUATION

B.1 Water Properties Evaluation

The property code (UOCODE) developed by Cheng et al. (1987) was used for the evaluation of water properties. It is a convenient code to use because only pressure and temperature are required as the input parameters. The output provided by the UOCODE are the specific volume, density, enthalpy, specific heat, thermal conductivity, viscosity, surface tension and latent heat of vaporization at the given pressure.

Range of validity for the UOCODE covers a wide range of thermodynamic states from the triple point to beyond critical point of water, i.e.

\[ 0.01°C < T < 800°C \]
\[ 0.6113kPa < P < 22.09MPa \]

Comparison with data from steam table (1967 ASME steam table) were made for various properties at both saturated and superheated regions. The results are shown in figures B1 to B5. The predictions agree very well with the experimental data for most of the region except at conditions near the critical point. Undoubtedly, this property code is accurate and can be used for thermal-hydraulic analysis with good confidence.
B.2 Non-aqueous Fluid properties Evaluation

A software for non-aqueous fluid properties evaluation was developed based on a four parameter equation of state. Thermodynamic properties of the fluids were obtained through the derivatives of the equation of state while the transport properties were evaluated through the use of other empirical correlations. Adachi et al.'s equation of state (1983) is chosen because it has a simple cubic form.

\[ P = \frac{RT}{V - b_1} - \frac{a(T)}{(V - b_2)(V + b_3)} \quad (B.1) \]

It has one temperature-dependent parameter \(a(T)\) and three temperature independent parameters \(b_1, b_2\) and \(b_3\). The parameters can be expressed as follows:

\[ a(T) = \frac{AR^2T^2}{P_c}[1 + \alpha(1 - T_r^{1/2})]^2 \quad (B.2) \]

\[ A = 0.44869 + 0.0402\omega + 0.01111\omega^2 - 0.00576\omega^3 \quad (B.3) \]

\[ \alpha = 0.4070 + 1.3787\omega - 0.2933\omega^2 \quad (B.4) \]

and

\[ b_i = \frac{B_iRT_c}{P_c} \quad i = 1, 2, 3 \quad (B.5) \]

where

\[ B_1 = 0.08974 - 0.03452\omega + 0.00330\omega^2 \quad (B.6) \]

\[ B_2 = 0.03686 + 0.00405\omega - 0.01073\omega^2 + 0.00157\omega^3 \quad (B.7) \]

\[ B_3 = 0.15400 + 0.14122\omega - 0.00272\omega^2 - 0.00484\omega^3 \quad (B.8) \]

\(R\) is the universal gas constant, \(\omega\) is the acentric factor of the substance and \(T_r\) is the reduced temperature. The applicability of the proposed equation of state was tested by the authors in terms of its calculated values of several pure-component properties such as the compressibility factor \((Z)\), fugacity coefficient in the form of \(\log(f/P)\) and isothermal enthalpy departure function \((H^0 - H)_T/RT_c\). The calculated values were compared with the values reported in Pitzer’s tables (1955) along with
four other widely used equations of state. A summary of comparison is reported in Table B-1. The comparison was made at twenty-two \( T_r \) and twenty \( P_r \) conditions (0.8 < \( T_r < 4.0 \), 0.2 < \( P_r < 8.0 \), using six values of acentric factor (0.0 to 0.5 at 0.1 intervals). The average absolute deviations obtained for these three properties from Adachi et al.’s equation are the lowest.

Equation (B-1) is also valid for saturation state. Adachi et al. (1983) have compared the experimental data with the calculated values of three saturated properties (vapour pressure, liquid and vapour volumes) obtained from equation (B-1) and those obtained from the other selected equations of state. Experimental values of eighteen pure components were used in the comparison and the results are summarized in Table B-2. The overall average absolute deviation is the lowest for equation (B-1) and thus indicates that it yields satisfactory values in the overall evaluation.

**Estimation of Saturation Pressure**

At saturation, the fugacity coefficient of liquid must be equal to that of vapour in order to have thermodynamic equilibrium. Adachi et al. derived an expression for fugacity coefficient based on their equation of state:

\[
\ln(f/P) = Z - 1 - \ln Z + \ln \frac{Z}{Z - B_1 P_r/T_r} + \frac{A(T_r)}{T_r (B_2 + B_3)} \ln \frac{Z - B_2 P_r/T_r}{Z + B_3 P_r/T_r} \quad (B.9)
\]

For a given temperature, an iteration process is required in order to determine the saturation pressure; i.e., in order to have thermodynamic equilibrium, the condition that \( \ln(f/P)_V = \ln(f/P)_L \) must be satisfied. To avoid this time consuming iteration process, Lee-Kesler’s (1975) equation for vapour pressure is used. Their equation only required the critical pressure, critical temperature and the acentric factor of the substance. It is derived from the Pitzer table and has the following form:

\[
\ln P_{r,\text{sat}} = f^{(0)}(T_r) + \omega f^{(1)}(T_r) \quad (B.10)
\]

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where

\[ f^{(0)}(T_r) = 5.92714 - 6.09648T_r^{-1} - 1.28862 \ln T_r + 0.16934T_r^3 \]  (B.11)

\[ f^{(0)}(T_r) = 15.2518 - 15.6875T_r^{-1} - 13.4721 \ln T_r + 0.43577T_r^3 \]  (B.12)

Equations (B.10) to (B.12) were used to estimate the saturation pressure with good accuracy.

Derivation of Volume from Equation of State

The volume is not expressed explicitly in equation (B.1). With given temperature and pressure, an iteration process is necessary for the determination of molar volume. The Newton-Raphson method was used because the method converges rapidly. For the superheated region, only one root would be expected (figure B6). The ideal gas volume is chosen for the initial guess for this case because it is close to the actual value and thus allows earlier convergence. For the saturation region, there are three possible roots because the equation of state is in cubic form. Figure B7 shows a typical \( P-V-T \) relation at saturation. The lowest root is taken as the solution for the saturated liquid volume while the highest root corresponds to the saturated vapour volume. To obtain the lowest root, the initial guess has to be very small such as having a value of 0.001.

Derivation of Equation for Specific Enthalpy

The variation of enthalpy between two states is independent of the path chosen to pass from one state to the other. There are an infinite number of possible calculation paths, all of which give the same numerical result (see figure 15). The enthalpy difference between two states can be expressed as

\[ H = (H_P - H^0)_T - (H_{P_{ref}} - H^0)_{T_{ref}} + \int_{T_{ref}}^{T} C_P^0 dT \]  (B.13)

From the equation of state, the departure function was derived by Adachi et al. as

\[ \frac{(H - H^0)_T}{RT_c} = (Z - 1)T_r + \frac{1}{(B_2 + B_3)} [A(T_r) - T_r \frac{dA(T_r)}{dT_r}] \ln \frac{Z - B_2 P_r/T_r}{Z + B_3 P_r/T_r} \]

or
\[(H - H^0)_T = RT(Z - 1) + \frac{1}{(b_2 + b_3)} \left[ a(T) - T \frac{da(T)}{dT} \right] \ln \frac{V - b_2}{V + b_3} \quad (B.14)\]

where
\[A(T_e) = A[1 + \alpha(1 - T_e^{1/2})] \quad (B.15)\]

The ideal gas specific heat at zero pressure for most substances can be expressed in terms of a third order polynomial function of temperature
\[C_P^0 = A + BT + CT^2 + DT^3 \quad (B.16)\]

The constants A, B, C and D are obtained experimentally and a list of substances with these constants is provided by Reid et al. (1977) and other engineering handbooks.

Derivation of Equation for Specific Heat Capacity

Specific heat capacity at constant pressure is defined as the rate of change of enthalpy with temperature at constant pressure, i.e.,
\[C_P - C_P^0 = \frac{\partial}{\partial T}(H - H^0) \bigg|_P \quad (B.17)\]

Taking the partial derivative of equation (B.14) w.r.t temperature at constant pressure yields
\[C_P - C_P^0 = P \frac{\partial V}{\partial T} \bigg|_P - R + \frac{1}{(b_2 + b_3)} \left[ -T \frac{d^2 a(T)}{dT^2} \ln \frac{V - b_2}{V + b_3} \right.
\]
\[\left. + \frac{\partial V}{\partial T} \bigg|_P \left( \frac{b_2 + b_3}{(V - b_2)(V + b_3)} \right) [a(T) - T \frac{da(T)}{dT}] \right] \]
\[= -\frac{1}{(b_2 + b_3)} T \frac{d^2 a(T)}{dT^2} \ln \frac{V - b_2}{V + b_3} - R
\]
\[+ \frac{\partial V}{\partial T} \bigg|_P \left[ P + \frac{a(T)}{(V - b_2)(V + b_3)} - T \frac{da(T)}{dT} \frac{1}{(V - b_2)(V + b_3)} \right] \]
\[= -\frac{1}{(b_2 + b_3)} T \frac{d^2 a(T)}{dT^2} \ln \frac{V - b_2}{V + b_3} - R
\]
\[+ \frac{\partial V}{\partial T} \bigg|_P \left[ \frac{RT}{V - b_1} - T \frac{da(T)}{dT} \frac{1}{(V - b_2)(V + b_3)} \right] \quad (B.18)\]

where
\[\frac{\partial V}{\partial T} \bigg|_P = -\frac{(\partial P/\partial T)_T}{(\partial P/\partial V)_T} \quad (Maxwell's Law) \quad (B.19)\]
and \( \frac{da(T)}{dT} \) and \( \frac{d^2a(T)}{dT^2} \) can be obtained by taking the derivatives of equation (B.2) w.r.t. \( T \) once and twice respectively, i.e.

\[
\frac{da(T)}{dT} = -\frac{a\alpha}{T_r} \left[ \frac{1 + \alpha}{T_r^{1/2}} - \alpha \right] 
\]

(B.20)

\[
\frac{d^2a(T)}{dT^2} = \frac{a\alpha(1 + \alpha)}{2T_r^{3/2}}
\]

(B.21)

\((\partial P/\partial T)_V\) and \((\partial P/\partial V)_T\) can be derived from the equation of state directly. After taking the partial derivatives of equation (B.1) and simplifying, the final equations for \((\partial P/\partial T)_V\) and \((\partial P/\partial V)_T\) are

\[
\frac{\partial P}{\partial T} \left|_V \right. = \frac{R}{V - b_1} - \frac{da(T)}{dT} \left( \frac{1}{(V - b_2)(V + b_3)} \right)
\]

(B.22)

\[
\frac{\partial P}{\partial V} \left|_T \right. = -\frac{RT}{(V - b_1)^2} + \frac{a(T)}{(b_2 + b_3)} \left( \frac{1}{(V - b_2)^2} - \frac{1}{(V + b_3)^2} \right)
\]

(B.23)

With any \( P-V-T \) state, the specific heat capacity is calculated from equations (B.18) to (B.23).

Evaluation of Transport Properties for Non-aqueous Fluids

All the present prediction methods for transport properties are either semi-empirical or purely empirical in nature. There is no generalized and accurate analytical prediction method for viscosity, thermal conductivity, and surface tension. The equations used for transport properties prediction are discussed in a fair amount of detail in Chapter 4. This section summarizes the equations used for transport properties evaluations and their sources.

Low-Pressure Gas Viscosity

The most common form of correlation for the prediction of viscosity at low pressure is the one which makes use of corresponding states principle. This type of correlation is easy to use and only requires the knowledge of the critical properties and molecular weight of the substance. These constants are readily available in the open literature. Equation suggested by Thodos et al. (1970) was chosen for the present work:

\[
\mu^0_\xi = 4.610T_r^{0.618} - 2.04e^{-0.449T_r} + 1.94e^{-4.058T_r} + 0.1
\]

(B.24)
where
\[ \xi = T_c^{1/6} M^{-1/3} P_c^{-2/3} \]
\( T_c \) = Critical temperature, K
\( P_c \) = Critical pressure, atm
\( M \) = Molecular weight
\( \mu^0 \) = Low-pressure gas viscosity, micro-poise

The expected error for this equation is normally between 1 and 3 percent for non-polar gases. It should not be used for hydrogen, helium, diatomic halogen gases and polar gases.

**Dense-Gas Viscosity**

Jossi et al. (1962) have developed a correlation relating the dense-gas viscosity to the low density viscosity. Their equation is recommended by Reid et al. (1977) as the most accurate prediction method for non-polar dense gas viscosity. They have related the dense-gas viscosity to the low density gas viscosity in terms of the reduced density:

\[ [(\mu - \mu^0)\xi + 1]^{0.25} = 1.0230 + 0.23364\rho_r + 0.58533\rho_r^2 \]
\[ -0.40758\rho_r^3 + 0.093324\rho_r^4 \]

(B.25)

where
\( \rho_r = \) reduced density
\( \mu = \) dense-gas viscosity, micro-poise

Equation (B.25) is valid for \( 0.1 < \rho_r < 3.0 \). It can also be used for the estimation of saturated liquid viscosity as long the reduced density is within the specified limits.
Liquid Viscosity

For \( T_r < 0.75 \), the relation of Van Velzen et al. (1972) is used for the estimation of saturated liquid viscosity. They assumed that \( \log \mu_L \) is linear in \( 1/T \), i.e.

\[
\log \mu_L = B \left( \frac{1}{T} - \frac{1}{T^0} \right)
\]

where

\[ \mu_L = \text{Liquid viscosity, cP} \]
\[ T = \text{Temperature, K} \]

and \( B \) and \( T^0 \) are correlation constants for each specific substance. A list of these constants for different substances was tabulated by Reid et al. (1977).

For saturated liquids at higher temperature, corresponding-states relation is used. The correlation suggested by Letsou and Stiel (1973) is easy to use and yields accurate results up to \( T_r = 0.92 \). Their correlation is shown below.

\[
\mu_L \xi = (\mu_L \xi)^{(0)} + \omega (\mu_L \xi)^{(1)}
\]

where

\[
(\mu_L \xi)^{(0)} = 0.015174 - 0.02135 T_r + 0.0075 T_r^2
\]
\[
(\mu_L \xi)^{(1)} = 0.042552 - 0.07674 T_r + 0.0340 T_r^2
\]

The viscosity evaluated \( (\mu_L) \) is in centipoise and \( \xi \) is defined in equation (B.24).

Low Density Gas Thermal Conductivity

A simple and accurate method for the prediction of thermal conductivity for low density gases is the well-known Eucken correlation for polyatomic gases.

\[
\frac{\lambda^0 M}{\mu^0} = C^0_v + 4.47
\]

\( \lambda^0 \) is the low density gas thermal conductivity in cal/g-mole K, \( \mu^0 \) is the corresponding low density gas viscosity in poise, \( M \) is the molecular weight and \( C^0_v \) is the ideal gas specific heat at constant volume and can be obtained from

\[
C^0_v = C^0_p - R
\]

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which has the unit of cal/g-mole.K.

Dense Gas Thermal Conductivity

At higher pressure (or density), simple kinetic theory would not be sufficient to account for the interactive forces between the molecules. Additional parameter such as the density of the gas has to be taken into account. Stiel and Thodos (1964) suggested the following correlation for the dense gas thermal conductivity estimation.

\[
(\lambda - \lambda^0) \Gamma Z_c^5 = 14.0 \times 10^{-8}(e^{0.535\rho_r} - 1) \quad \rho_r < 0.5
\]

\[
(\lambda - \lambda^0) \Gamma Z_c^5 = 13.1 \times 10^{-8}(e^{0.67\rho_r} - 1.069) \quad 0.5 < \rho_r < 2.0
\]

\[
(\lambda - \lambda^0) \Gamma Z_c^5 = 2.976 \times 10^{-8}(e^{1.155\rho_r} - 2.016) \quad 2.0 < \rho_r < 2.8 \quad (B.32)
\]

where \( \Gamma = T_c^{1/3}M^{1/2}P_c^{-2/3} \) and \( Z_c \) is the critical compressibility factor of the substance. \( T_c \) is in Kelvin, \( P_c \) is in atm. and \( M \) is in gm/g-mole. Again, this relation is only good for non-polar substances.

Liquid Thermal Conductivity

Normally, liquid thermal conductivity is quite well behaved. The liquid thermal conductivity at any temperature can be estimated if the corresponding values at normal boiling temperature is known. Reid et al. (1977) suggested a correlation for non-polar fluids relating the thermal conductivity to that at normal boiling point;

\[
\lambda_L = \frac{2.64 \times 10^{-3}}{M} \left[ \frac{3 + 20(1 - T_r)^{2/3}}{3 + 20(1 - T_{rb})^{2/3}} \right] \quad (B.33)
\]

where \( T_{rb} \) is the reduced normal boiling temperature. Thermal conductivity evaluated from equation (B.33) has a unit of cal/cm.s.K. For \( T_r > 0.8 \), the equation for high density gas thermal conductivity (equation (B.32)) can be used.

Surface Tension

Reid et al. (1977) modified the equation suggested by Brock and Bird (1955) for the surface tension evaluation. The modified equation relates the surface tension to
the normal boiling temperature of the substance. Their equation is shown below.

\[ \sigma = P_c^{2/3} T_c^{1/3} \theta (1 - T_r)^{11/9} \]
\[ \theta = 0.1207 [1 + \frac{T_c \ln P_c}{(1 - T_r)}] - 0.281 \]  
(B.34)

where \( P_c \) is in atm. and \( T_c \) in Kelvins. Reid et al. (1977) reported an average error of less than 5% based on their comparison with experimental data.

**Criterion for Dense Gas**

In evaluating the transport properties for gases, different correlations are used for low and high density gases. A criterion is needed for deciding whether the gas is at low density or high density. Childs and Hanley (1968) have deduced criterion which indicates whether or not the pressure effect is significant. Their results are summarized in figure B8. An approximate equation for the above criterion is

\[ P_r = -0.125 + 0.1875 T_r \]  
(B.35)

Equation (B.35) is used to separate low density gases from high density gases.

**Accuracy of Non-Aqueous Property Code**

All the above mentioned equations were used in the development of the property code for non-aqueous fluids. This code is generally applicable to non-polar fluids with adequate accuracy. Data for Carbon dioxide, Nitrogen, and Freon-12 were used to test the accuracy of the property code. The predictions for all properties are very accurate for gases as expected. Results of the comparison for Freon-12 are shown in figures B9 to B13. For the saturated liquid properties, deviations of more than 10% are found at temperatures close to \( T_c \). More extensive numerical comparison was made and the results are summarized in Table B-3. The results indicate that the property code developed can predict the thermodynamic and transport properties of non-polar fluids satisfactorily. With the availability of new and more accurate correlations, this property code can be updated easily to produce more accurate predictions.
Programming Specifications of the Property Code

The code is written in modular form for easy updating of new equations. Flowcharts for the property code are summarized in figures B14 to B15. The source code is shown in figure B17. The code is written in Fortran-77 language and compiled with IBM Professional Fortran Compiler. It is available in micro-computer version and would require an 8087 Math Coprocessor. The required input parameters are the critical constants ($T_c$, $P_c$ and $V_c$), molecular weight ($M$), acentric factor ($\omega$) constants for ideal gas specific heat (CPVA, CPVB, CPVC and CPVD) and constants for liquid viscosity ($\bar{\eta}$ and $T^0$). A list of such constants for several commonly used fluids is shown in table B-4.
Table B-1. Comparison of Calculated Thermodynamic Properties with Pitzer's Table for a Total of 2640 Data Points at 22 $T_r$ and 20 $P_r$ Conditions

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<th>Average absolute deviation</th>
<th>$z$</th>
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<th>PR</th>
<th>HK</th>
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SRK — Soave (1972)  
PR — Peng-Robinson (1976)  
HK — Harnens and Knapp (1980)  
ICL — Ishikawa et al. (1980)  
THIS WORK — Adachi et al. (1983)
Table B-2. Comparison of Calculated and Experimental Values of Three Saturated Properties

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<td>n-Decane</td>
<td>8.79</td>
<td>19.6</td>
<td>-</td>
<td>4.90</td>
<td>9.84</td>
</tr>
<tr>
<td>Oxygen</td>
<td>1.01</td>
<td>4.44</td>
<td>3.38</td>
<td>0.63</td>
<td>7.95</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>1.30</td>
<td>6.40</td>
<td>2.77</td>
<td>1.39</td>
<td>8.12</td>
</tr>
<tr>
<td>Carbon</td>
<td>0.83</td>
<td>3.83</td>
<td>3.87</td>
<td>1.43</td>
<td>8.49</td>
</tr>
<tr>
<td>Carbon</td>
<td>0.80</td>
<td>14.5</td>
<td>1.74</td>
<td>0.57</td>
<td>4.62</td>
</tr>
<tr>
<td>Ethylene</td>
<td>1.55</td>
<td>7.49</td>
<td>2.42</td>
<td>1.14</td>
<td>6.59</td>
</tr>
<tr>
<td>Propylene</td>
<td>1.06</td>
<td>10.3</td>
<td>2.15</td>
<td>0.60</td>
<td>5.26</td>
</tr>
<tr>
<td>1-Butene</td>
<td>1.24</td>
<td>12.2</td>
<td>1.36</td>
<td>0.67</td>
<td>4.41</td>
</tr>
<tr>
<td>Acetylene</td>
<td>2.01</td>
<td>12.5</td>
<td>1.04</td>
<td>1.55</td>
<td>5.00</td>
</tr>
<tr>
<td>Overall</td>
<td>1.57</td>
<td>12.8</td>
<td>2.04</td>
<td>1.26</td>
<td>6.04</td>
</tr>
</tbody>
</table>

a Canjar and Manning (1967)
b Young (1910)
c Sage and Lacey (1950, 1955)

A $\phi P$
B $\phi V$
C $\phi V$

SRK --- Soave (1972)
PR --- Peng-Robinson (1976)
HK --- Hammens and Knapp (1980)
ICL --- Ishikawa et al. (1980)
THIS WORK --- Adachi et al. (1983)
Table B-3. Summary of Fluid Property Prediction Accuracy for Different Fluids

<table>
<thead>
<tr>
<th>Fluid Property</th>
<th>Freon-12</th>
<th></th>
<th>Nitrogen</th>
<th></th>
<th>Carbon Dioxide</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>N</td>
<td>RMS*</td>
<td>N</td>
<td>RMS*</td>
<td>N</td>
<td>RMS*</td>
</tr>
<tr>
<td>Sat. Liq. Vol.</td>
<td>24</td>
<td>0.0665</td>
<td>20</td>
<td>0.0519</td>
<td>19</td>
<td>0.0727</td>
</tr>
<tr>
<td>Sat. Vap. Vol.</td>
<td>24</td>
<td>0.0108</td>
<td>20</td>
<td>0.0160</td>
<td>19</td>
<td>0.0084</td>
</tr>
<tr>
<td>Superheated Vap. Vol.</td>
<td>35</td>
<td>0.0079</td>
<td>22</td>
<td>0.0026</td>
<td>55</td>
<td>0.0064</td>
</tr>
<tr>
<td>Sat. Liq. Enthalpy</td>
<td>24</td>
<td>0.0560</td>
<td>20</td>
<td>0.1487</td>
<td>18</td>
<td>0.0981</td>
</tr>
<tr>
<td>Sat. Vap. Enthalpy</td>
<td>24</td>
<td>0.0211</td>
<td>20</td>
<td>0.1180</td>
<td>18</td>
<td>0.0139</td>
</tr>
<tr>
<td>Hfg</td>
<td>24</td>
<td>0.0338</td>
<td>20</td>
<td>0.0444</td>
<td>18</td>
<td>0.0605</td>
</tr>
<tr>
<td>Superheated Enthalpy</td>
<td>35</td>
<td>0.0110</td>
<td>22</td>
<td>0.0166</td>
<td>55</td>
<td>0.0126</td>
</tr>
<tr>
<td>Sat. Liq. Sp. Heat</td>
<td>14</td>
<td>0.1058</td>
<td>11</td>
<td>0.1701</td>
<td>9</td>
<td>0.1291</td>
</tr>
<tr>
<td>Sat. Vap. Sp. Heat</td>
<td>14</td>
<td>0.1097</td>
<td>11</td>
<td>0.0379</td>
<td>9</td>
<td>0.1381</td>
</tr>
<tr>
<td>Superheated Sp. Heat</td>
<td>40</td>
<td>0.0254</td>
<td>22</td>
<td>0.0127</td>
<td>35</td>
<td>0.0047</td>
</tr>
<tr>
<td>Sat. Liq. Viscosity</td>
<td>14</td>
<td>0.1789</td>
<td>12</td>
<td>0.1097</td>
<td>7</td>
<td>0.0531</td>
</tr>
<tr>
<td>Sat. Vap. Viscosity</td>
<td>14</td>
<td>0.0812</td>
<td>12</td>
<td>0.0213</td>
<td>7</td>
<td>0.0202</td>
</tr>
<tr>
<td>Superheated Viscosity</td>
<td>24</td>
<td>0.0535</td>
<td>26</td>
<td>0.0081</td>
<td>23</td>
<td>0.0166</td>
</tr>
<tr>
<td>Sat. Liq. Therm. Cond.</td>
<td>15</td>
<td>0.1402</td>
<td>14</td>
<td>0.1111</td>
<td>13</td>
<td>0.0744</td>
</tr>
<tr>
<td>Sat. Vap. Therm. Cond.</td>
<td>15</td>
<td>0.0863</td>
<td>14</td>
<td>0.0957</td>
<td>13</td>
<td>0.0968</td>
</tr>
<tr>
<td>Superheated Therm. Cond.</td>
<td>24</td>
<td>0.0408</td>
<td>26</td>
<td>0.0503</td>
<td>31</td>
<td>0.0467</td>
</tr>
<tr>
<td>Overall Average</td>
<td>364</td>
<td>0.0643</td>
<td>292</td>
<td>0.0634</td>
<td>349</td>
<td>0.0533</td>
</tr>
</tbody>
</table>

* RMS error is defined as

\[
\text{RMS} = \sqrt{\frac{\text{Error}_{\text{pred}}^2}{N}}
\]
### Table B-4. List of Constant Necessary for Property Evaluation

<table>
<thead>
<tr>
<th></th>
<th>FREON-11</th>
<th>FREON-12</th>
<th>FREON-13</th>
<th>FREON-21</th>
<th>FREON-22</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>CRITICAL TEMP.</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(DEG. K)</td>
<td>471.2</td>
<td>385.0</td>
<td>302.0</td>
<td>451.6</td>
<td>369.2</td>
</tr>
<tr>
<td><strong>CRITICAL PRESS.</strong></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(PASCAL)</td>
<td>4406550</td>
<td>4122910</td>
<td>3920310</td>
<td>5166300</td>
<td>4973830</td>
</tr>
<tr>
<td><strong>CRITICAL VOLUME</strong></td>
<td>1.805E-3</td>
<td>1.795E-3</td>
<td>1.723E-3</td>
<td>1.914E-3</td>
<td>1.906E-3</td>
</tr>
<tr>
<td>(M**3/ KG)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>ACCENTRIC FACTOR</strong></td>
<td>0.188</td>
<td>0.176</td>
<td>0.180</td>
<td>0.202</td>
<td>0.215</td>
</tr>
<tr>
<td><strong>MOLECULAR WEIGHT</strong></td>
<td>137.368</td>
<td>120.914</td>
<td>104.459</td>
<td>102.923</td>
<td>86.469</td>
</tr>
</tbody>
</table>

**CONSTANTS FOR IDEAL GAS CP.**

<table>
<thead>
<tr>
<th>CPVA</th>
<th>9.789</th>
<th>7.547</th>
<th>5.449</th>
<th>5.652</th>
<th>4.132</th>
</tr>
</thead>
<tbody>
<tr>
<td>CPVD</td>
<td>9.903E-9</td>
<td>1.037E-8</td>
<td>1.065E-8</td>
<td>7.795E-9</td>
<td>7.305E-9</td>
</tr>
</tbody>
</table>

**CONSTANTS FOR LIQUID VISC.**

<table>
<thead>
<tr>
<th>BO</th>
<th>840.17</th>
<th>684.0</th>
<th>564.472</th>
<th>731.876</th>
<th>532.853</th>
</tr>
</thead>
<tbody>
<tr>
<td>TO</td>
<td>227.9</td>
<td>178.7</td>
<td>138.315</td>
<td>203.167</td>
<td>176.928</td>
</tr>
</tbody>
</table>

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Figure B1a. Comparison of Subcooled Specific Volume Predicted from UOCODE with Steam Table (ASME, 1967).

Figure B1b. Comparison of Superheated Specific Volume Predicted from UOCODE with Steam Table (ASME, 1967).
Figure B2a. Comparison of Subcooled Enthalpy Predicted from UOCODE with Steam Table (ASME, 1967).

Figure B2b. Comparison of Superheated Enthalpy Predicted from UOCODE with Steam Table (ASME, 1967).
Figure B3a. Comparison of Subcooled Specific Heat Predicted from UOCODE with Steam Table (ASME, 1967).

Figure B3b. Comparison of Superheated Specific Heat Predicted from UOCODE with Steam Table (ASME, 1967).
Figure B4a. Comparison of Subcooled Viscosity Predicted from UOCODE with Steam Table (ASME, 1967).

Figure B4b. Comparison of Superheated Viscosity Predicted from UOCODE with Steam Table (ASME, 1967).
Figure B5a. Comparison of Subcooled Thermal Conductivity Predicted from UOCODE with Steam Table (ASME, 1967).

Figure B5b. Comparison of Superheated Thermal Conductivity Predicted from UOCODE with Steam Table (ASME, 1967).
Figure B6. P-V-T Relation for a Typical Fluid at Superheated State.
Figure B7. P-V-T Relation for a Typical Fluid at Saturation State.
Figure B8. Criterion for Dense Gas (Childs and Hanley, 1968).
Fig. B9a. Comparison of Specific Volume Predicted from Non-Aqueous property Code with Freon-12 Data at Superheated State.
Fig B9b. Comparison of Specific Volume Predicted from Non-Aqueous Property Code with Freon-12 Data at Saturation State.
Fig. B10a. Comparison of Specific Enthalpy Predicted from Non-Aqueous property Code with Freon-12 Data at Superheated State.
Fig B10b. Comparison of Specific Enthalpy Predicted from Non-Aqueous Property Code with Freon-12 Data at Saturation State.
Fig B11a. Comparison of Specific Heat Capacity, Predicted from Non-Aqueous Property Code with Freon-12 Data at Superheated State.
Fig. B11b. Comparison of Specific Heat Predicted from Non-Aqueous Property Code with Freon-12 Data at Saturation State.
Fig. B12a. Comparison of Viscosity Predicted from Non-Aqueous Property Code with Freon-12 Data at Superheated State.
Fig B12b. Comparison of Viscosity Predicted from Non-Aqueous Property Code with Freon-12 Data at Saturation State.
Fig. B13a. Comparison of Thermal Conductivity Predicted from Non-Aqueous property Code with Freon-12 Data at Superheated State.
Fig B13b. Comparison of Thermal Conductivity Predicted from Non-Aqueous Property Code with Freon-12 Data at Saturation State.
Main Subroutine PROPER

Given T, P

START

CALCULATE Psat

\[ \frac{P - Psat}{Psat} \leq 0.005 \]

YES

Call SATURA

NO

P < Psat

YES

Call SUPER

NO

Call SUBCOOL

RETURN

Figure B14. Flow-Chart for Main Program of Non-aqueous Property Code.
Figure B15. Flowchart for Saturation Subroutine.
Subroutine VLIQ

START

V₀ = 0.001

.ICOUNT = 0

ICOUNT.LT.50

NO

YES

Evaluate
F(V₀) & F'(V₀)

Vnew = V₀ - \frac{F(V₀)}{F'(V₀)}

ICOUNT = ICOUNT + 1

NO

\frac{Vnew - V₀}{Vnew} \leq 0.1 %

YES

VLIQ = VNEW

RETURN

F = p - \frac{RT}{V - b₁} + \frac{a(T)}{(V - b₂)(V + b₃)}

F' = \frac{dF}{dV} |' T, p

Figure B16. Flowchart for Saturated Liquid Subroutine.
Subroutine for Non-Aqueous Fluid Property Evaluation

\( P = \) Pressure (Pa), \( T = \) Temperature (K).

**SUBROUTINE PROPER(P,T)**

*REAL MW,MW1,MW2*

*DIMENSION B(3)*

**CRIT** - External Common block in which the physical constants of the fluid are passed to this subroutine.

**TC** - Critical temperature of the substance (K).

**PC** - Critical pressure of the substance (Pa).

**VC** - Critical volume of the substance (kg.m\(^{-3}\)).

**W** - Acentric factor of the substance.

**MW** - Molecular weight of the substance (kg.kg.mole\(^{-1}\)).

**C01** - First constant for ideal gas specific heat evaluation.

**C02** - Second constant for ideal gas specific heat evaluation.

**C03** - Third constant for ideal gas specific heat evaluation.

**C04** - Fourth constant for ideal gas specific heat evaluation.

**C00** - Constant values are given such that \( C_p^0 \) has a unit of cal.gmole\(^{-1}\).0°C\(^{-1}\).

**BO** - First constant for liquid viscosity evaluation.

**TO** - Second constant for liquid viscosity evaluation.

**TREF** - Reference temperature at which the saturated liquid enthalpy is equal to zero.

**COMMON /CRIT/TC,PC,VC,W,MW,C01,C02,C03,C04,BO,TO,TREF**

Figure B1.7. Source Code for the Non-Aqueous Property Subroutine.
PROP — Common block in which all the calculated fluid properties are passed back to the calling program.

VF — Saturated liquid specific volume (m$^3$.kg$^{-1}$).
VG — Saturated vapour specific volume (m$^3$.kg$^{-1}$).
VL — Subcooled or superheated specific volume (m$^3$.kg$^{-1}$).
RHOF — Saturated liquid density (kg.m$^{-3}$).
RHOG — Saturated vapour density (kg.m$^{-3}$).
RHOL — Subcooled or superheated density (kg.m$^{-3}$).
HF — Saturated liquid enthalpy (J.kg$^{-1}$).
HG — Saturated vapour enthalpy (J.kg$^{-1}$).
HL — Subcooled or superheated enthalpy (J.kg$^{-1}$).
CPF — Saturated liquid specific heat capacity (J.kg$^{-1}$.K$^{-1}$).
CGF — Saturated vapour specific heat capacity (J.kg$^{-1}$.K$^{-1}$).
CFL — Subcooled or superheated specific heat capacity (J.kg$^{-1}$.K$^{-1}$).
UF — Saturated liquid viscosity (kg.m$^{-1}$.s$^{-1}$).
UG — Saturated vapour viscosity (kg.m$^{-1}$.s$^{-1}$).
UL — Subcooled or superheated viscosity (kg.m$^{-1}$.s$^{-1}$).
AKF — Saturated liquid thermal conductivity (W.m$^{-1}$.K$^{-1}$).
AKG — Saturated vapour thermal conductivity (W.m$^{-1}$.K$^{-1}$).
AKL — Subcooled or superheated thermal conductivity (W.m$^{-1}$.K$^{-1}$).
PRF — Saturated liquid Prandtl number.
PRG — Saturated vapour Prandtl number.
PRL — Subcooled or superheated Prandtl number.
HPF — Latent heat of vaporization.
SIGMA — Surface tension on gas-liquid interface (N.m$^{-1}$).

COMMON /PROP/VF, VG, VL, RHOF, RHOG, RHOL, HF, HG, HL, CPF, CGF, CFL,
$  \quad$ UF, UG, UL, AKF, AKG, AKL, PRF, PRG, PRL, HPF, SIGMA

CONST, PARAM — Internal Common block in which working variables are being transferred to and from the internally defined functions.

COMMON /CONST/R, B, AA, ALPHA
COMMON /PARAM/TR, PR, DEPREF, HPREF

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C******************************************************************************
C                Initializes all constant values.
C******************************************************************************

R = 8314.
B(1)=0.08974-0.03452*W+0.00330*W**2
B(2)=0.03686+0.00405*W-0.01073*W**2+0.00157*W**3
B(3)=0.15400+0.14122*W-0.00272*W**2-0.00484*W**3
AA=0.44869+0.04024*W+0.01111*W**2-0.00576*W**3
ALPHA=0.4070+1.3787*W-0.2933*W**2

C******************************************************************************
C                Determine the reference state enthalpy departure function
C                necessary for relative enthalpy evaluation.
C******************************************************************************

TR = TREF/TC
PREF=PC*PRSAT(TR,W)
VV=VWAP(TREF,PREF)
VL=VLIQ(TREF,PREF)
ZV=PREF*VV/(R*TREF)
ZL=PREF*VL/(R*TREF)
HREF=TREF*TR,ZV,ZL)
PR = PRSAT(TR,W)
DEFREF = DEPENT(TR,PR,ZV)
   TR = T/TC
   PR = F/FC

C******************************************************************************
C                When saturation state is pre-determined, either the
C                temperature or the pressure in the calling program can
C                be assigned a negative value. This will permit PROPER
C                to call the Saturation subroutine directly.
C******************************************************************************

IF (T.LT.0.001. OR. P.LT.0.) THEN
   CALL SATURA(P,T)
   RETURN
ELSE
   IF(TR.LE.1.0) THEN
      PSAT=PC*PRSAT(TR,W)
   ELSE
      CALL SUBP(P,T)
   ENDIF
   RETURN
ENDIF
C
C*************************************************************************
C If the given pressure is within 0.5% of the saturation
C pressure calculated from Lee-Kesler equation, then the
C given condition is assumed to be in saturation.
C*************************************************************************
C
IF(ABS((P-PSAT)/PSAT).LE.0.005) THEN
   CALL SATURA(P,T)
   RETURN
ELSE
C*************************************************************************
C Check for superheated or subcooled state.
C*************************************************************************
C
IF(P.LT.PSAT) THEN
   CALL SUPER(P,T)
   RETURN
ELSE
   CALL SUBCOOL(P,T)
   RETURN
ENDIF
ENDIF
END
SUBROUTINE SUPER(P,T)
REAL MW,MW1,MW2
DIMENSION B(3)
COMMON /CRIT/Tc,PC,VC,W,MW,CO1,CO2,CO3,CO4,BO,TO,TREF
COMMON /PROF/VF,VC,VL,RHOF,RHOG,RHOL,HF,HG,HL,CPF,CFG,CPL,$
   UG,UL,AKF,AKG,AKL,PRF,PRG,PRL,HFG,SIGMA
COMMON /CONST/R,B,AA,ALPHA
COMMON /PARAM/TR,PR,DEPREF,HFGREF
   V = VVAP(T,P)
   ZL = P*V/(R*T)
   VL = V/MW
   RHOL=1/VL
   VR = VL/VC

C Enthalpy is calculated relative to the reference state.
C I.e., the saturated liquid enthalpy at reference temperature
C was assumed to be zero.

DEP1 = DEPENT(TR,PR,ZL)
HL = ((CPF(T)-CPF(TREF))*4186.8+(R*T)*(DEP1-DEPREF))

CPL=CP(T,V)
UL = VISVAP(TR,PR,VR)
AKL=AVKAP(T,PR,VR)
PRL=UL+CPL/AKL
RETURN
END
Subcooled Subroutine

SUBROUTINE SUBCOOL(P, T)
REAL MW, MW1, MW2
DIMENSION B(3)
COMMON /CRIT/ TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF
COMMON /PROP/VF, VG, VL, RHOF, RHOG, RHOH, HF, HG, HL, CPF, CFG, CPL,
$       UF, UG, UL, AKF, AKG, AKL, PRF, PRG, PRL, HFG, SIGMA
COMMON /CONST/R, B, AA, ALPHA
COMMON /PARAM/TR, PR, DEPREF, HFGREF

V = VLIQ(T, P) \nZL = P*V/(R*T) \nVL = V/MW \nRHOI=1/VL \nDEP = DEPENT(TR, PR, ZL) \nHL = (((CPF(T)-CPF(TREF))*4186.8+(R*TC*(DEP-DEPREF)) \n$                                                     +MW*HFGREF
CPL = CP(T, P) \nUL = VISLIQ(T) \nAKL = AKLIQ(T, PR, VR) \nPRL = UL*CPL/AKL
RETURN
END

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Saturation Subroutine

SUBROUTINE SATURA(P,T)
REAL MW,MW1,MW2
DIMENSION B(3)
COMMON /CRIT/TC,PC,VC,W,MW,C01,C02,C03,C04,BO,TO,TREF
COMMON /PROP/VF,VG,VL,RHOF,RHOG,RHOL,HG,HL,CPF,CPG,CPL,
$        UF,UG,UL,AKF,AKG,AKL,PRF,PRG,PRL,HFG,SIGMA
COMMON /CONST/R他也,AA,ALPHA
COMMON /PARAM/TR,PR,DEPREF,HFGREF

IF(T.LT.0.001) THEN
    TSAT=TC*TRSAT(PR,W)
    T=TSAT
ELSE
    PSAT=PC*PRSAT(TR,W)
    P=PSAT
    TSAT=T
ENDIF
TR = TSAT/TC
VF1= VLIQ(TSAT,P)
VG1= VVAP(TSAT,P)
ZF = P*VF1/(R*TSAT)
ZG = P*VG1/(R*TSAT)
DEPG = DEPENT(TR,PR,ZG)
VF = VF1/MW
VG = VG1/MW
RHOF=1/VF
RHOG=1/VG
VRF=VF/VC
VRG=VG/VC
HG = ((CPO(TSAT)-CPO(TREF))*4186.8+(R*TC*(DEPG-DEPREF)))/MW
$        + HFGREF
HFG = ENTHF(TR,ZG,ZF)
IF(ABS(TSAT-TREF).LT.0.05) THEN
    HF = 0.0
ELSE
    HF = HG - HFG
ENDIF
CPF = CP(TSAT,VF1)
CPG = CP(TSAT,VG1)
UF=VISLIQ(T)
UG=VISVAP(TR,PR,VRG)
AKF = AKLIQ(T,PR,VR)
AKG = AKVAP(T,PR,VRG)
SIGMA = SIG(TR,TC,PC,W)
PRF = UP*CPG/AKG
PRG = UG*CPG/AKG
RETURN
END
Function for evaluating vapour molar volume given T and P. Newton-Raphson iteration technique is used.

FUNCTION VVAP(T, P)
REAL COUNT, MW
DIMENSION B(3), BB(3)
COMMON /CRIT/TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF
COMMON /CONST/R, BB, AA, ALPHA
TR=T/TC
DO 10 I=1, 3
   B(I)=BB(I)*R*TC/PC
10 CONTINUE
   A=AA*R**2*TC**2*((1+ALPHA*(1-SQR(T,TR)))*2)/PC

Ideal gas volume is used as the initial guess

VOLD=R*T/P
COUNT=1

If there is no convergence within 100 iteration, a message will be prompted to the user.

1 IF(COUNT.GT.100) GO TO 3
   TERM1=P*(VOLD-B(1))*(VOLD-B(2))*(VOLD+B(3)) - R*T*(VOLD-B(2)) *
   (VOLD+B(3)) + A*(VOLD-B(1))
   TERM2=P*((VOLD-B(2))*(VOLD+B(3)) + (VOLD-B(1))*(VOLD+B(3))
   +(VOLD-B(1))*(VOLD-B(2)) - R*T*((VOLD-B(2))+(VOLD+B(3)))
   +A
   VNEW=VOLD-TERM1/TERM2
   IF(ABS(VNEW-VOLD).LE.0.00010) GO TO 2
   VOLD=VNEW
   COUNT=COUNT+1
   GO TO 1
2 VVAP=VNEW
RETURN
3  VVAR=VNEW
   WRITE(*,11)
11  FORMAT(' NO CONVERGENCE WITHIN 100 ITERATIONS')
   RETURN
   END

C
C******************************************************************************
C
C Function for evaluating liquid molar volume given P and T.
C Newton-Raphson iteration technique is used.
C
C******************************************************************************
C
FUNCTION VLIQ(T,P)
REAL COUNT,MW
DIMENSION B(3),BB(3)
COMMON /CONST/R, BB, AA, ALPHA
COMMON /PROP1/TC, PC, VC, W, MW, C01, C02, C03, C04, B0, TO, TREF
   TR=T/TC
   DO 10 I=1,3
       B(I)=BB(I)*R*TC/PC
10   CONTINUE
   A=AA*R**2*TC**2*((1+ALPHA*(1-SQRT(TR)))*2)/PC

C******************************************************************************
C
C A very small value is used as the initial guess to ensure
C it will converge to the smallest root.
C
C******************************************************************************
C
   VOLD=0.001
   COUNT=1

C******************************************************************************
C
C If there is no convergence within 100 iteration, a message
C will be prompted to the user.
C
C******************************************************************************
C
1  IF(COUNT.GT.100) GO TO 3
   TERM1=P*(VOLD-B(1))*(VOLD-B(2))*(VOLD+B(3)) - R*T*(VOLD-B(2)) *
      (VOLD+B(3)) + A*(VOLD-B(1))
   TERM2=P*((VOLD-B(2))*(VOLD+B(3)) + (VOLD-B(1))*(VOLD+B(3)) +
      (VOLD-B(1))*(VOLD-B(2))) - R*T*((VOLD-B(2)) +
      (VOLD+B(3))) + A

VNEW = VOLD - TERM1 / TERM2  
IF ((ABS(VNEW - VOLD) / VOLD) .LE. 0.0001) GO TO 2  
   VOLD = VNEW  
   COUNT = COUNT + 1  
GO TO 1  
2  VLIQ = VNEW  
RETURN  
3  VLIQ = VNEW  
WRITE (*, 11)  
11 FORMAT (' NO CONVERGENCE WITHIN 100 ITERATIONS')  
RETURN  
END  

C  
C*****************************************************************************  
C  
C Function for evaluating enthalpy departure function.  
C*****************************************************************************  
C  
FUNCTION DEPENT(TR, PR, Z)  
DIMENSION B(3)  
COMMON /CONST/R,B,AA,ALPHA  
   TERM1 = (Z-1) * TR  
   TERM2 = AA * ALPHA * (1 + ALPHA * (1 - SQRT(TR))) * SQRT(TR)  
   TERM3 = AA * (1 + ALPHA * (1 - SQRT(TR))) ** 2  
   TERM4 = LOG(ABS((Z-B(2)*PR/TR)/(Z+B(3)*PR/TR)))  
   DEPENT = TERM1 + ((TERM2 + TERM3) * TERM4) / (B(2) + B(3))  
RETURN  
END  

C*****************************************************************************  
C  
C Function for evaluating Latent Heat of Vaporization using  
C Clapeyron relation.  
C*****************************************************************************  
C  
FUNCTION ENTFG(TR, ZV, ZL)  
COMMON /PROPI/TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF  
REAL MW  
R = 8314.  
PSI = 6.09648 - 1.28862*TR + 1.016082*TR**7  
  + W*(15.5875 - 13.4721*TR + 2.61462*TR**7)  
ENTFG = PSI * R * TC * (ZV - ZL) / MW  
RETURN  
END  

C  
C*****************************************************************************  
C  

Lee-Kesler equation for saturation pressure evaluation.

FUNCTION PRSAT (TR, W)
TERM = 5.92714 - 6.09648/TR - 1.28862*LOG(TR) + 0.169347*TR**6
$ + W*(15.2518 - 15.6875/TR - 13.4721*LOG(TR) + 0.43577*TR**6)
PRSAT = EXP(TERM)
RETURN
END

Determination of saturation temperature by iteration method
through the use of Lee-Kesler equation.

FUNCTION TRSAT(PR, W)
TR = 0.5
COUNT = 1.
N = 100

If there is no convergence within 100 iteration, a message
will be prompted to the user.

1 IF(COUNT.GT.N) GO TO 3
   TERM1 = LOG(PR) - 5.92714 + 6.09648/TR + 1.28862*LOG(TR) - 0.169347*
        $ TR**6 - W*(15.2518 - 15.6875/TR - 13.4721*LOG(TR) + 0.43577*TR**6)
   TERM2 = -6.09648/TR**2 + 1.28862/TR - 1.016082*TR**5 -
        $ W*(15.6875/TR**2 - 13.4721/TR + 2.61462*TR**5)
   TRNEW = TR - TERM1/TERM2
   IF (ABS(TRNEW-TR).LE.0.001) GO TO 2
   TR = TRNEW
   COUNT = COUNT + 1.
   GO TO 1
2 TRSAT = TRNEW
RETURN
3 TRSAT = TRNEW
WRITE(*,10) N
10 FORMAT(//,'T2,'NO CONVERGENCE WITHIN',14,' ITERATIONS')
RETURN
END

228
C*********************************************************************************************************C
C Integral of ideal gas specific heat capacity
C*********************************************************************************************************
C
FUNCTION CPO(T)
COMMON /PROPI/TC,PC,VC,W,MW,A,B,C,D,BO,TO,TREF
CPO = T*(A*T*(B/2+T*(C/3+D*T/4)))
RETURN
END

C*********************************************************************************************************
C Function for evaluating viscosity of gases.
C*********************************************************************************************************
C
FUNCTION VISVAP(IR,PR,VR)
REAL MW
COMMON /PROPI/TC,PC,VC,W,MW,CO1,CO2,CO3,CO4,BO,TO,TREF
PC = PC/101325.
DENSE = 0.1875*IR - 0.125
ZETA = PC**0.16667/(SQRT(MW)*PCA**0.66667)
VISG = (4.610*IR**0.618-2.04*EXP(-0.449*IR)
       + 1.94*EXP(-4.058*IR) + 0.1)/ZETA

C**********************************************************************************************************
C The low density gas viscosity is assumed first using the method of Thodos et al. If the gas is at high density, Jossi et al. equation is used.
C**********************************************************************************************************
C
IF (PR.GE.DENSE) THEN
   RHS = 1.0230 + 0.23364/VR + 0.58533/VR**2
   VISG = VISG + (RHS**4 - 1)/ZETA
ENDIF
VISVAP = VISG*1.0E-7
RETURN
END

229
Function for Evaluating liquid viscosity.

FUNCTION VISLIQ(T)
REAL MW
COMMON /PROP1/TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF
PC = PC/101300.
TR = T/TC

Van Velzen et al.'s equation is used for low temperature liquid viscosity and Letsou and Stiel's equation is used for high temperature liquid viscosity calculation.

IF(TR.GE.0.75) THEN
  TERM1 = 0.035174 - 0.02135*TR + 0.0075*TR**2
  TERM2 = 0.042552 - 0.07674*TR + 0.0340*TR**2
  ZETA = TC**0.16667/(SQRT(MW)*PCA**0.66667)
  VISLIQ = (TERM1 + W*TERM2)/ZETA)/1000.
ELSE
  VISLIQ = (EXP(BO*(1/T - 1/TO)))/1000.
ENDIF
RETURN
END

Function for evaluating specific heat capacity. Equation was derived from the equation of state suggested by Adachi et al.

FUNCTION CP(T,V)
REAL MW
DIMENSION BI(3), B(3)
COMMON /PROP1/TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF
COMMON /CONST/R, B, AA, ALPHA
TR= T/TC
DO 10 I=1,3
  BI(I) = B(I)*R*TC/PC
10 CONTINUE
A = AA*R**2*TC**2/PC
AT=A*(1+ALPHA*(1-SQRT(TR)))*2
DADT=A*ALPHA*((1+ALPHA)/SQRT(TR) - ALPHA)/TC
D2ADT2=A*ALPHA*(1+ALPHA)*TR**2*(-1.5)/(2*TC**2)
DPDUV-R/(V-BI(1)) -DADT/((V-BI(2))*(V+BI(3)))
DPDUVT=R*TR/(V-BI(1))*2 + AT*(1/(V-BI(2)))*2 -1/(V+
$ BI(3)**2)/(BI(2)+BI(3))$
$ CFG1=(-T*D2ADT2)*LOG((V-BI(2))/(V+BI(3)))/(BI(2)+BI(3))$
$ -DPDUVT*(R*I)/(V-BI(1)) -T*DADT/((V-BI(2))*(V+BI(3)))$
$ +/DPDUVT -R)+(CO1+K*(CO2+T*(CO3+T*CO4)))*4186.8$
CP=CFG1/MW
RETURN
END

C******************************************************************************
C  Function for evaluating thermal conductivity of gases.
C******************************************************************************
FUNCTION AKVAP(T,PR,VR)
REAL MW, LAMDAO
COMMON/PROPF, TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF
TR=T/TC
PCA = PC/101300.
RHOR=1/VR
ZC = PC*VC*MW/(TC*8314)
ZETA=TC**0.16667/(SQRT(MW)*PCA**0.66667)
VISO=(4.610*TR**0.618-2.04*EXP(-0.449*TR)+1.94*EXP(-4.058*TR)
$ + 0.1)/ZETA/100000.
CV = CO1+K*(CO2+T*(CO3+T*CO4)) - 1.98
LAMDAO=VISO*(1.15*CV+4.00)/MW
DENSE = 0.1875*TR - 0.125
C******************************************************************************
C  Stiel and Thodos's equations are used for both the low and
C  high density gases.
C******************************************************************************
IF(PR.LE.DENSE) THEN
  AKVAP=LAMDAO.
ELSE
  TAL = ZETA*MW
  CONSTANT=TAL*ZC**5
IF (RHOR.LT.0.5) THEN
    AKVAP = (14.0E-9*(EXP(0.535*RHOR) - 1)/CONSTANT) + LAMDAO
ELSE
    IF (RHOR.LT.2.0) THEN
        AKVAP = (13.1E-8*(EXP(0.67*RHOR) - 1.069)/CONSTANT) + LAMDAO
    ELSE
        AKVAP = (2.976E-9*(EXP(1.155*RHOR) + 2.016)/CONSTANT) + LAMDAO
    ENDIF
ENDIF
AKVAP = 418.68*AKVAP
RETURN
END

C*******************************************************************************
C Function for evaluating liquid thermal conductivity. The
C Sato-Riedel's method is employed.
C*******************************************************************************
FUNCTION AKLIQ(T, PR, VR)
COMMON /PROP1/TC, PC, VC, W, MW, CO1, CO2, CO3, CO4, BO, TO, TREF
REAL MW
TR = T/TC
IF (TR.LT.0.3) THEN
    PRB = 101300./PC
    TRB = TREF/PRB
    AKL = 2.64E-3*(3+20*(1-TR)**0.66667)/(3+20*(1-TRB)**0.66667)
    AKLIQ = 418.68*AKL/SQRT(MW)
ELSE
    AKLIQ = AKVAP(T, PR, VR)
ENDIF
RETURN
END
Function for evaluating surface tension using Brock and Brock's approach.

FUNCTION SIG(TR, TC, FC, W)
  FRE = (101300/FC)
  TER = TRSAT(FRE, W)
  THETA = 0.1207*(1+TER*LOG(FC/101300)/(1-TER)) - 0.281
  SIG = (FC/101300)**0.66667*TC**0.3333*THETA*(1-TR)**(11./9.)/1000.
RETURN
END
Appendix C

PROCEDURE FOR PREDICTING HORIZONTAL FLOW CHF

C.1 Input Parameters

The input parameters required for the prediction are the pressure in kPa, mass flux in kg.m$^{-2}$.s$^{-1}$, thermodynamic vapour quality and the tube diameter in m. With these given flow conditions, fluid properties and the parameter $T_1$ will be calculated. The CHF ratio can then be predicted based on the local parameters approach or the constant inlet condition approach.

C.2 Calculation Procedure

C.2.1 Evaluation of Fluid Properties

The saturation properties of the fluid are estimated from the UOCODE (for water) and from non-aqueous property code, PROPER (for fluids other than water). The list of constants shown in Table B-4 for the fluid of interest must be given before the properties can be calculated. These critical constants are passed to the property sub-
routine through a common block called CRIT. The estimated properties will be passed back to the calling program through a common block called PROP. An example of the calling process is shown in figure C1. The program calculates the fluid properties of freon-12 and print out the values for saturated liquid specific volume (VF), saturated vapour specific volume (VG) and the single phase specific volume (VL). For this case, the fluid is in the superheated vapour state and thus VL corresponds to the superheated vapour specific volume. Consequently, subroutine PROPER only calculate the superheated volume and assign zero values to the saturated properties. The source code for property subroutine can be found in Appendix B.

C.2.2 Evaluation of Actual Quality

When analyzing the force balance, actual void fraction and quality are needed. The method suggested by Saha and Zuber (1974) was used for actual quality evaluation (see Chapter 4). Heat flux, tube diameter and fluid properties are required for the intermediate calculation steps. Normally, the heat flux is not known if only the local parameters are given. Critical heat flux for vertical flow is used as the estimate of the heat flux. The predicted actual quality is then used to predict CHF ratio which in turn yield the value for horizontal CHF. The flowchart for actual quality evaluation subroutine is shown in figure C2. For thermodynamic qualities greater than 20%, the actual quality can be assumed equal to the thermodynamic quality.

C.2.3 Evaluation of Void Fraction

CISE (1970) equation is chosen for the slip ratio estimation and subsequently the void fraction is evaluated from the relation

\[ \alpha = \frac{X}{X + S(1 - X)L} \]

(C.1)

where \( S \) is the slip ratio calculated from equation (4.33) and \( X \) is the actual quality.
C.2.4 Evaluation of Parameter $T_1$

With the fluid properties, actual quality and void fraction, equation (5.22) is used to calculate the value of parameter $T_1$ with $C_1 = 1.0$. $Re_L$ is defined as $G(1 - X)D/\mu_f$ in equation (5.22). $T_1$ is then substituted into equation (6.9) for the evaluation of CHF ratio.

C.3 CHF Ratio Prediction Based on Local Parameter Approach

The CHF ratio ($CHFR$) is evaluated by the use of equation (6.9), i.e.;

$$CHFR = 1 - e^{-(T_1/85)^0.5} \quad (C.2)$$

Horizontal CHF values can then be obtained from

$$CHF_{hor} = CHFR \times CHF_{ver} \quad (C.3)$$

where value for $CHF_{ver}$ can be taken from Table 1.

C.4 CHF Ratio Prediction Based on Constant Inlet Condition Approach

For this method, the inlet subcooling and the heated length must be known. The exit quality is varied until the heat flux calculated from energy balance is the same as the critical heat flux predicted based on that local exit quality (see figure 41a).

C.5 User Friendly Tabular Method

Despite the correct trend of equation (6.9) with the mass flux, it would under-predict the CHF ratio considerably for certain flow conditions (high pressure and moderate
mass flux). The reason is that the exponential function would not converge to the value of 1.0 soon enough even though the mass flux value is supposed to be high enough to suppress gravity effect. A simple correction for this deficiency is to take the maximum value of $CHFR$ calculated from both equation (6.9) and equation (2.25). Figure C3 shows the prediction of CHF ratio by the two methods mentioned. For mass flux greater than $G_A$, equation (2.25) predicts a higher value of $CHFR$. If equation (2.25) is used for mass flux greater than $G_A$, the prediction will converge to the value of 1.0 sooner.

This combined approach was used to produce a CHF ratio ($CHFR$) table for water (see Table C-1). The $CHFR$ table provides users with an easy and quick way of estimating the CHF ratio. Table C-1 is only valid for an $8mm$ inner diameter tube. The CHF ratio is found to decrease with increasing tube diameter as illustrated by figure 37. Tube diameter correction factor for CHFR is approximated by

$$K_{dia} = \left(\frac{0.008}{D}\right)^{0.3}$$  \hspace{1cm} (C.4)

for $0.005m \leq D \leq 0.016m$ and tube diameter correction factor for vertical CHF is given by

$$K_1 = \left(\frac{0.008}{D}\right)^{0.33}$$  \hspace{1cm} (C.5)

Therefore, horizontal CHF can be calculated as

$$CHF_{hor} = K_{dia} \times K_1 \times CHFR_{8mm} \times CHF_{ver,8mm}$$  \hspace{1cm} (C.6)

$$CHF_{hor} = \left(\frac{0.008}{D}\right)^{0.63} \times CHFR_{8mm} \times CHF_{ver,8mm}$$  \hspace{1cm} (C.7)

where $CHFR_{8mm}$ is the $CHFR$ value taken from Table C-1 and $CHF_{ver,8mm}$ is the corresponding vertical CHF value taken from Table 1 (under the same $G$, $X$ and $P$). Table C-1 was arranged in a matrix form so that simple linear interpolation is possible. The $CHFR$ values are tabulated against thermodynamic quality so that no further calculation for actual quality is needed. Values with asterisks are extrapolated from the higher qualities region. They correspond to the conditions where the actual qualities are equal to zero. Table C-1 was used to predict CHF ratio and compared to
the data bank. The results for water and freon-12 comparison are summarized in Table C-2. It is clear that similar type of error distribution is found by using Table C-1 as compared to the step-by-step calculations outlined earlier. It is therefore worthwhile to lose some accuracy when the prediction method is so simple to apply such as the tabular approach.

C.6 Source Code for Horizontal CHF Prediction

The prediction subroutine can be divided into three levels. Level one is the main subroutine where the input parameters are read in. It is in this level that the necessary subroutines are called and the main calculations are being performed. Figure C4 shows schematically the flow of information in level one.

Level two is the main backbone of the whole program. It calls the other subroutines in level three needed for intermediate calculations. Property subroutines is being called first and subsequently subroutines such as XACT and VOID are called so that $T_1$ can be evaluated. The flowchart for CHFRAT in second Level is shown in figure C5. CHFRAT is the subroutine which calculates the CHF ratio.

Subroutines in level three are PROPER, XACT and VOID. The source code for the whole prediction program is listed in figure C6.
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Table C-1. CHF Ratio (CHF<sub>hor</sub>/CHF<sub>ver</sub>) Table for Water
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Table C-1: CHF Ratio (CHFhor/CHFver) Table for Water
Table C-2. Comparison of Prediction Accuracy Between The CHF Ratio Table and Equation (6.9)

<table>
<thead>
<tr>
<th>Prediction Method</th>
<th>Average* Error</th>
<th>RMS Error</th>
<th>Standard Deviation</th>
<th>Fraction of data** within prediction accuracy of</th>
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<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>±10%</td>
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<tr>
<td>Constant X_{PD}</td>
<td></td>
<td></td>
<td></td>
<td></td>
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<td>CHFR Table (Table C-1)</td>
<td>0.005</td>
<td>0.278</td>
<td>0.278</td>
<td>0.1885</td>
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<td>Eq. (6.9)</td>
<td>-0.044</td>
<td>0.259</td>
<td>0.255</td>
<td>0.2654</td>
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<tr>
<td>Constant X_{IN}</td>
<td></td>
<td></td>
<td></td>
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<tr>
<td>CHFR Table (Table C-1)</td>
<td>-0.022</td>
<td>0.181</td>
<td>0.181</td>
<td>0.3846</td>
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<tr>
<td>Eq. (6.9)</td>
<td>-0.029</td>
<td>0.165</td>
<td>0.165</td>
<td>0.3615</td>
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</table>

* Mean of $\frac{(\text{CHF}_\text{pred} - \text{CHF}_\text{expt})}{\text{CHF}_\text{expt}}$

** No. of data point = 260.
Example program for estimating fluid properties of Freon-12.

PROGRAM FREON12
REAL MW
COMMON /CIT/TC,PC,VC,W,MW,CO1,CO2,CO3,CO4,BO,TO,TREF
COMMON /PROP/VF,VG,VL,RHOF,RHOG,RHOL,HF,HG,HL,CPF,CFG,CPL,
$ AF,UG,UL,AKF,AKG,AKL,PRF,PRG,PRL,HFG,PSAT,TSAT,SIGMA
TO=369.2
PC=4973830.
VC=0.0019056
W=0.215
MW=86.469
CO1=4.132
CO2=3.865E-2
CO3=-2.794E-5
CO4=7.305E-9
BO=532.855
TO=156.928
TREF=233.15
P = 300000.
T = 400.
CALL PROPER(P,T)
WRITE(*,10) VF,VG,VL,HF,HG,HL,CPF,CFG,CPL,UF,UG,UL,
$ AF,AKG,AKL,SIGMA,T,P
STOP
END

VF =  0.00000E+00  VG =  0.00000E+00  VL =  0.12593E+00
HF =  0.00000E+00  HG =  0.00000E+00  HL =  0.34709E+06
CPF =  0.00000E+00  CFG =  0.00000E+00  CPL =  0.76401E+03
VISF=  0.00000E+00  VISG=  0.00000E+00  VISL=  0.17288E-04
AKF =  0.00000E+00  AKG =  0.00000E+00  AKL =  0.16752E-01
SIGMA =  0.00000E+00  T =  0.40000E+03  P =  0.30000E+06

Figure C1. Example Program listing for Evaluating Fluid Properties.
Subroutine XACT

START

Given q, X_e, D, Properties

Evaluate X_e

X_e ≤ X_ed

YES

X_e ≥ 0

YES

X_act = X_e

NO

X_act = \frac{X_e \cdot X_{ed} \cdot \exp \left( \frac{X_e}{X_{ed}} - 1 \right)}{1 - X_{ed} \cdot \exp \left( \frac{X_e}{X_{ed}} - 1 \right)}

NO

RETURN

x_{act} = 0

Figure C2. Flowchart for Actual Quality Subroutine.
Figure C3. CHF Ratio Prediction as a Function of Mass Flux.
Main Program CHFHOR

START

Input P, G, X_e, D

CALL CALCHF

CALL CHFRAT

\[ \text{CHF}_{\text{hor}} = \text{CHFR} \times \text{CHF}_{\text{ver}} \]

RETURN

Figure C4. Flowchart for Main Program of Horizontal CHF Prediction.
Subroutine CHFRAT

START

Input
P, G, X_e, D

CALL PROPER

CALL XACT

CALL VOID

Evaluate $T_1$

\[ \text{CHFR} = 1 - e^{-(T_1/6.0)^{0.5}} \]

RETURN

Figure C5. Flowchart for CHF Ratio Subroutine.
Main subroutine for horizontal CHF prediction

SUBROUTINE CHFNSRT(P, G, XE, DYM, CHFNSRT)

Call subroutine CALCHF to get the CHF value for vertical flow

CALL CALCHF(P, G, XE, DYM, CHF)

Call subroutine CHFRATIO to get the value for CHF ratio

CALL CHFRATIO(P, G, XE, DYM, CHF)

Calculating CHF value for horizontal flow

CHF = CHFNSRT*CHF
RETURN
END

Figure C6. Source code for CHF Prediction Program
Subroutine to calculate CHF ratio
Pressure in kPa
Mass flux in kg/m²s
Diameter in m.

SUBROUTINE CHFRAT(P,G,XE,DHY,CHFR)

PROP ——— Common block in which all the calculated fluid properties are passed to this calling subroutine.

COMMON/PROP/
    VE, VG, VL, RHOF, RHOG, RHOH,
    HF, HG, HL, CPF, CMG, CPL,
    VISF, VISG, VISL, AKF, AKG, AKL,
    PRF, PAG, PRL, HFG, PSAT, TSAT,
    SIGMA
PSATU = P
TSATU = 0.0

CALL PROPER(PSATU, TSATU)

If the thermodynamic quality is less than 20%, calculate the actual quality using the Saha-Zuber equation, and the vertical CHF as the estimate for heat flux. If it is greater than 20%, assume the actual quality to be the same as the thermodynamic quality.

IF (XE.LE.0.2) THEN
    CALL CALCHF (P,G,XE,DHY,CHFVER)
    QQ = CHFVER*1000.

Figure C6. (Cont'd)
CALL XACT(QQ,G,DHY,XE,X)
ELSE
  X = XE
ENDIF

Void fraction is calculated through subroutine VOID using CISE equation.

CALL VOID(G,X,DHY,ALPHA)

Value of the force balance parameter T1 is evaluated and the CHF ratio can thus be calculated

REL=G*(1-X) *DHV/VISF
TERM1=REL**(−0.2)
TERM2=(1-X)/(1-ALPHA)
TERM3=G**2/(9.81*DHV*RHOF*(RHOF-RHOC))
TT1=0.092*TERM1*TERM2**2*TERM3/SQRT(ALPHA)
CHFR=1−EXP(−(TT1/6.0)**0.5)
RETURN
END
Subroutine to calculate actual quality

P in kPa, G in kg.m\(^{-2}\).s\(^{-1}\), DHY in meter and
QQ in W.m\(^{-2}\)

SUBROUTINE XACT(QQ,G,DHY,XE,XA)
COMMON/PROP/ VF, VG, VL, RHOF, RHOG, RHL,
+ HF, HG, HL, CPF, CPF, CPL,
+ VISF, VISG, VISL, AKF, AG, AKL,
+ PRF, PRG, PRL, HFG, PSAT, TSAT,
+ SIGMA

Calculate quality at bubble departure, XED

PE = G\*DHY\*CPF/ACK
BOIL = QQ/(G\*HFG)
IF(PE.GT.70000.) THEN
  XED = -154\*BOIL
ELSE
  XED = -0.0022*(PE\*BOIL)
ENDIF

IF(XE.LE.XED) THEN
  XA = 0.0
  RETURN
ELSE
  IF(XED.GE.0.) THEN
    XA = XE
    RETURN
  ELSE
    EXPON=XE/XED
    IF(EXPON.LE.(-20.0)) EXPON=-20.0
    EE = XED*EXP(EXPON-1)
    XA = (XE - EE)/(1-EE)
  ENDIF
ENDIF
RETURN
END
C Subroutine for estimating void fraction
C
G in kg.m^-2.s^-1, DHV in m and X is the actual quality
C
SUBROUTINE VOID(G,X,DHV,ALPHA)
C
PROP —— Common block through which the calculated saturation
fluid properties are passed to this subprogram.
C
COMMON/PROP/ VF, VG, VL, RHOF, RHOG, RHOI,
+ HP, HC, HL, CPF, CPF, CPL,
+ VISF, VISG, VISL, AKF, AKG, AKL,
+ PRF, PRG, PRL, HFG, PSAT, TSAT,
+ SIGMA
C
Calculation of void fraction using CISE equation.
C
VRL = X*RHOF/(RHOG*(1-X))
RE = G*DHV/VISF
BETA=RHOF*X/(RHOF*X+RHOG*(1-X))
WE=G**2*DHV/(SIGMA*RHOF)
E1=1.578*RE**(-0.19)*(RHOF/RHOG)**0.22
E2=0.0273*WE*RE**(-0.51)*(RHOF/RHOG)**(-0.08)
YY=BETA/(1-BETA)
ARGU = YY/(1+YY*E2)-YY*E2
IF(ARGU.LE.0.) ARGU=0.0
SRATIO=1+EI*SQRT(ARGU)
IF(SRATIO.LT.1.0) SRATIO=1.0
ALPHA = 1/(1+SRATIO/VRL)
RETURN
END
SUBROUTINE CALCHF(P,G,D,X,CHF)

C**********************************************************************
C Subroutine CALCHF
C
This subroutine is used to obtain the CHF from table of discrete
values. The table was presented by Groeneveld et al.

C**********************************************************************

C**********************************************************************
C Variable declaration:
C
Character:
C
GBCM : a single character variable for channel geometry.
C
UNIT : a single character variable for unit.
C
FLUID : a single character variable for fluid type.
C
ORIENT : a single character variable for channel orientation.
C
STATE : a single character variable for state of the system.

Real:
C
X : provided quality.
C
D : provided diameter.
C
G : provided mass flux.
C
P : provided pressure.
C
INTRP3 : the interpolated value.
C
PX : discrete value of quality.
C
PG : discrete value of mass flux.
C
PP : discrete value of pressure.
C
CHFTAB : a set of 4410 discrete values of CHF arranged in three-
dimensional arrays.

Logical:
C
NFLG : Program termination flag.
C
Flag : Step termination flag.

C**********************************************************************

REAL CHFTAB(21,14,15),PX(21),PG(14),PP(15),INTRP3
LOGICAL FLAG
CALL UP TWO COMMON BLOCKS TO PASS SOME VARIABLES IN AND OUT THE SUBROUTINE.

COMMON/TABLE/CHFTAB
COMMON/ROWS/PX, PG, PP, IX, IG, IP
    H2O = P
    GH2O = G

INITIALIZE THE FLAG AND RANGE POINTER.

FLAG = .TRUE.
IX = 20
IG = 13
IP = 14

LOCATE THE DISCRETE VALUE OF QUALITY BOUNDING THE INPUT VALUE.

DO 100 I=1,20
   IF((X .LE. PX(I+1)) .AND. FLAG) THEN
      IX = I
      FLAG = .FALSE.
   END IF
100 CONTINUE

RE-INITIALIZE THE FLAG.

LOCATE THE DISCRETE VALUE OF MASS FLUX BOUNDING THE INPUT VALUE.

FLAG = .TRUE.
DO 200 I=1,13
   IF((GH2O .LE. PG(I+1)) .AND. FLAG) THEN
      IG = I
      FLAG = .FALSE.
   END IF
200 CONTINUE
C Re-initialize the flag.
C Locate the discrete value of pressure bounding the input value.
C
C*****************************************************************************

FLAG = .TRUE.
DO 300 I=1,14
   IF((PH2O .LE. PP(I+1)) .AND. FLAG) THEN
      IP = I
      FLAG = .FALSE.
   END IF
300 CONTINUE

C*****************************************************************************

C Calculate modification factor for tube diameter other than 0.008.
C
C*****************************************************************************

IF(D .GT. 0.016) THEN
   CKL = 0.7937
ELSE IF(D .LT. 0.002) THEN
   CKL = 1.5874
ELSE
   CKL = (0.008/D)**(1./3)
END IF

CHF = IN3RP3(X,GH2O,PH2O)*CKL*1000
RETURN
END
REAL FUNCTION INIRP3(X,G,P)

C******************************************************************************
C
C Function INIRP3
C
C This function is used to set up the parameters for linear
C interpolation in CHF evaluation. The parameters are the slope,
C and two boundary points.
C
C******************************************************************************

C******************************************************************************
C
C Variable declaration:
C
C
C Real:
C X : provided quality.
C G : provided mass flux.
C P : provided pressure.
C INTERP : the interpolated value.
C FX : discrete value of quality.
C FG : discrete value of mass flux.
C PP : discrete value of pressure.
C CHFTAB : a set of 4410 discrete value of CHF arranged in three-
C dimensional arrays.
C
C******************************************************************************

REAL INTERP,CHFTAB(21,14,15),FX(21),FG(14),PP(15)

C******************************************************************************
C
C Call up two common blocks to pass some variables in and out the
C subroutine.
C
C******************************************************************************

COMMON/TABLE/CHFTAB
COMMON/ROWS/FX,FG,PP,IX,IG,IP

C******************************************************************************
C
C Linear interpolate the pressure value.
C
C******************************************************************************

PARAM = (P-PP(IP))/(PP(IP+1)-PP(IP))
C1 = INTERP(CHFTAB(IX,IG,IP),CHFTAB(IX,IG,IP+1),PARAM)
C2 = INTERP(CHFTAB(IX,IG+1,IP),CHFTAB(IX,IG+1,IP+1),PARAM)
C3 = INTERP(CHFTAB(IX+1,IG,IP),CHFTAB(IX+1,IG,IP+1),PARAM)
C4 = INTERP(CHFTAB(IX+1,IG+1,IP),CHFTAB(IX+1,IG+1,IP+1),PARAM)
C Linear interpolate the mass flux value.
C
PARAM = (G-PG(IG))/(PG(IG+1)-PG(IG))
C5  = INTERP(C1,C2,PAM)
C6  = INTERP(C3,C4,PAM)

C Linear interpolate the quality value.
C
PARAM = (X-PX(IX))/(PX(IX+1)-PX(IX))
INTERP3 = INTERP(C5,C6,PAM)
RETURN
END
REAL FUNCTION INTERP(A1,A2,P)

C Function INTERP
C
This function is used to interpolate value linearly.
It requires the slope of the function and two boundary points.

C Variable declaration:
C
Real:
INTERP : the interpolated value.
P : slope of the function.
A1,A2 : the two bounds.

INTERP = A1 + (A2-A1)*P
RETURN
END