

Boiling viscous liquids in natural circulation vacuum evaporators

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Abstract

Vacuum evaporative crystallizers are used by the South African sugar industry to produce more than 2 000 000 tons of sugar annually. The process fluid is very viscous and laminar conditions prevail in the apparatus. To date the equipment has been designed by rule of thumb. In order to develop a procedure for optimizing the design and operation of this equipment, research was undertaken to determine how the equations proposed for heat transfer and vapour hold-up, should be modified for boiling in laminar flow under vacuum.

It was found that the heat transfer coefficient can be estimated by the superposition method of Rohsenow and Griffith, but that different exponents and constants are required. A new equation is proposed for calculating the void fraction in the highly subcooled region, and the equation of Bowring for subcooling at the point of bubble departure must also be modified.

Nomenclature

A	flow area
B_o	constant (Equation 20)
Bl	constant (Equation 10)
C_o	flow distribution parameter
C_p	specific heat
C_{sf}	constant in Rohsenow correlation
D	tube diameter
f	friction factor
G	mass velocity
g	acceleration due to gravity
h	heat transfer
i_{fg}	latent heat of vaporisation
K	fluid consistency index
k	thermal conductivity of liquid
m	exponent in Rohsenow correlation
n	flow behaviour index
Nu	Nusselt number
Pr	Prandtl number
Q	volumetric rate of flow
Re	Reynolds number
$t(z)$	bulk liquid temperature at axial position z
u	velocity of flow
V	rising velocity of vapour
W	mass rate of flow
x	mass vapour quality
z	axial coordinate
α	void fraction
η	empirical factor for bubble departure
ρ	density
μ	viscosity
ϕ	heat flux
ϕ_c	convection heat flux
ϕ_n	nucleate boiling heat flux
σ	surface tension
ΔP_F	pressure difference due to friction
Δt_s	superheat necessary to cause nucleation
Δt_{sub}	subcooling at axial position
Δt_{subd}	subcooling at point of bubble departure

Subscripts

b at bulk conditions

d	at bubble departure
f	liquid
fo	total flow assumed liquid
g	vapour
s	at saturation
TP	two phase
w	at wall conditions

Introduction

In the sugar industry crystallization is carried out using natural circulation vacuum evaporative crystallizers known as vacuum pans. The fluid known as massecuite consists of a suspension of crystals in concentrated molasses. It is highly viscous, slightly non-Newtonian, exhibits pseudoplastic properties, and laminar conditions prevail in the apparatus.

Although research on the factors affecting the boiling characteristics of vacuum pans has been in progress for at least fifty years, improvements in the design of these crystallizers have been the result of trial and error, because no satisfactory method has yet been put forward for optimizing pan design and operation.

The equations proposed for forced convection boiling cannot be used, because they were established for turbulent flow under pressures higher than atmospheric. Boiling viscous fluid under vacuum has not been studied. Experiments were therefore designed to determine if the available equations could be adapted, and if not, what modifications would be required.

Theoretical background

The flow pattern and the liquid and tube surface temperature that occur over the length of a vertical evaporator tube are shown in Figure 1. The liquid that enters is subcooled, that is, its temperature is below the saturation temperature corresponding to the pressure at that point. In region AB heat transfer is entirely by single phase forced convection. As heat is transferred to the liquid the temperature adjacent to the heating surface increases until at point B the boiling temperature is exceeded, and vapour formation occurs, but the bulk of the liquid is still subcooled.

Between B and D the vapour bubbles grow because of

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the additional heat input, and as a result of decompression as the hydrostatic head decreases. At the same time heat penetrates towards the centre of the tube so that the liquid temperature increases. This is the region of subcooled boiling. Transition to saturated boiling occurs when the average combined enthalpy of liquid and vapour equals the saturation enthalpy of the liquid. This happens a short distance upstream of point D. At point D boiling extends across the entire sectional area of the tube, and from this point there is a gradual decrease of the fluid temperature because of the decompression effect. With viscous liquids, however, transition to saturated boiling does not take place, and subcooled boiling occurs up to the tube outlet.

A method proposed by Rohsenow and Griffith [11] to estimate the boiling heat transfer coefficient in both the subcooled and saturated regions consists in adding the single phase forced convection heat flux and the nucleate boiling heat flux

$$\phi = \phi_c + \phi_n \quad (1)$$

The single phase forced convection component is given by

$$\phi_c = h_{fo}(t_w - t(z)) \quad (2)$$

where h_{fo} , the heat transfer coefficient for the total flow assumed liquid, is calculated using the Seider and Tate equation as modified by Charm and Merrill [3] for heat transfer to pseudoplastic fluids in straight tubes

$$\frac{h_f D}{K_f} = 2.0 \left[\frac{W \cdot C_p}{k_f z} \right]^{1/3} \left[\frac{K_b (3n+1)}{k_w 2(3n-1)} \right]^{0.14} \quad (3)$$

The nucleate boiling heat flux has been correlated by Rohsenow [10] using the relation

$$\frac{C_p \cdot \Delta t_s}{i_{fg}} = C_{sf} \left\{ \frac{\phi_n}{\mu_f \cdot i_{fg}} \left[\frac{\sigma}{g(\rho_f - \rho_g)} \right]^{1/2} \right\}^{0.33} (Pr)^{1+m} \quad (4)$$

where the value of the constant C_{sf} is determined experimentally, and that of the exponent m is given as 0 for water and 0.7 for other liquids. For pseudoplastic fluids this equation expressed in terms of the consistency and flow behaviour index is

$$\frac{C_p \cdot \Delta t_s}{i_{fg}} = C_{sf} \left\{ \left[\frac{\phi_n}{i_{fg} \cdot \rho_f} \right]^{2-n} \left[\frac{\sigma}{g(\rho_f - \rho_g)} \right]^{1/2} \frac{\rho_f}{K_f} \cdot 8 \cdot \left[\frac{n}{6n+2} \right]^n \right\}^{0.33} (Pr)^{1+m} \quad (5)$$

Starting at point B when boiling begins three factors contribute to the pressure drop along the tube.

- (1) The momentum effect which results from acceleration along the tube due to evaporation and expansion of the vapour phase.
- (2) The gravitational effect which results from the change in elevation.
- (3) The frictional effect due to the shear forces acting on the two phase fluid.

The acceleration and elevation losses can be calculated from a momentum equation if the mass vapour quality and the void fraction are known. The frictional component must however be determined empirically.

Griffith and Wallis [5] have proposed a method for estimating the friction loss of liquid-gas mixtures in the bubble flow regime. They related the friction loss to a modified Fanning equation

$$\Delta P_F = \frac{2f_r \rho_f u_f^2 z}{D(1-\alpha)^2} \quad (6)$$

where f_r , the Fanning friction factor, is obtained from the conventional Reynolds number-Fanning friction factor plot using a velocity defined as

$$u_f = \frac{Q_f}{A(1-\alpha)} \quad (7)$$

This method was modified by Oliver and Wright [9] for pseudoplastic non-Newtonian two phase systems. The liquid properties are considered to be dominant. The only parameter modified by the presence of gas is the liquid velocity which is obtained from the equation.

$$u_{TP} = \frac{Q_f + Q_g}{A} \quad (8)$$

The liquid friction factor is calculated using the relation $16/\text{generalized Reynolds number}$. The two phase generalized Reynolds number is calculated using this velocity and the equation of Metzner and Reed [7].

$$Re = \frac{D^n u^{2-n} \rho_f}{K_f} \cdot 8 \cdot \left[\frac{n}{6n+2} \right]^n \quad (9)$$

The amount of vapour present in the evaporator tubes must be established to permit calculation of the elevation and acceleration losses. In the case of natural circulation evaporative crystallizers, circulation results from the difference in hydrostatic head between the two phase mixture of massecuite and liquid in the tubes and single phase liquid in the downtake. Thus to calculate the circulation rate, the volumetric fraction occupied by vapour is necessary.

The change in void fraction that takes place in the evaporator tube is shown in Figure 1. At first the vapour formed consists of discrete bubbles attached to the tube surface, but at C they become detached to form bubbly flow. At point D boiling extends across the tube. Coalescence of the bubbles occurs fairly rapidly and transition to slug flow may take place.

The resulting void fraction is a function not only of the amount of vapour generated, but also of the rate at which it moves up the tube relative to the liquid velocity. The upward rate of flow of the vapour depends upon the vapour flow distribution that is, whether it is flowing close to the tube centerline or near the wall, and upon the bubble rise velocity.

In the highly subcooled region the bubbles grow and collapse but remain attached to the tube surface. A model

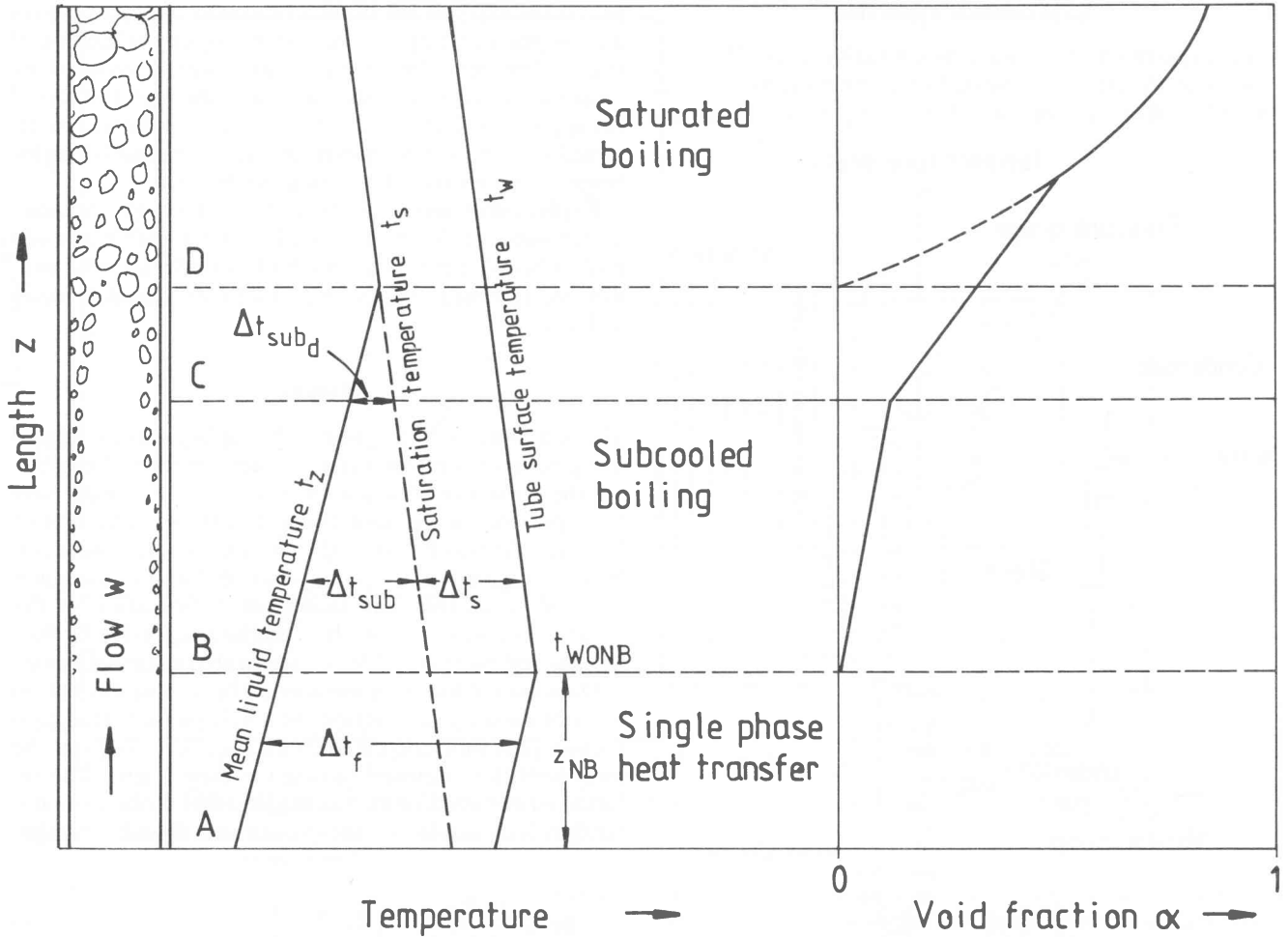


Figure 1 – Diagram of temperature & void fractions in subcooled boiling

has been proposed by Griffith et al [4] to estimate the void fraction in this region.

$$\alpha = \frac{\phi_n(\text{Pr})_f}{\text{Bl} \cdot (\text{Nu})_f \cdot h_{fo} \cdot \Delta t_{\text{sub}}} \quad (10)$$

At low pressure this relationship becomes incorrect since the diameter of the bubbles is then a function of the pressure as well as of the hydrodynamic layer thickness. Under these conditions the voidage is overestimated.

At point C the vapour bubbles start to depart from the tube surface, and there is a rapid increase in the void fraction. Bowring [2] found that the subcooling at the point of bubble departure Δt_{subd} could be approximated by the relationship

$$\Delta t_{\text{subd}} = \eta \cdot \phi \cdot \rho_f / G \quad (11)$$

Work by Levy [6] indicated that low pressures affect the value of the factor η . He also suggested an empirical model for the estimation of the void quality in region CD, the low subcooling region. He assumed that the true quality x' is related to thermodynamic quality x by the relationship

$$x' = x - x_d \cdot \exp(x/x_d - 1) \quad (12)$$

where x_d is the thermodynamic quality at the point of bubble departure given by

$$x_d = \frac{C_{pf}(t_f - t_s)_d}{i_{fg}} \quad (13)$$

After the point of bubble departure the void fraction is obtained from the equation of Nicklin et al. [8] who showed that the rising velocity of the vapour is made up of two components: V , its rising velocity in still liquid, plus a contribution due to the non-uniform distribution of vapour in the moving liquid.

$$\frac{Q_g}{\alpha \cdot A} = \frac{C_o(Q_g + Q_f)}{A} + V \quad (14)$$

Rouhani and Axelsson [12] found that the non-uniform flow distribution parameter C_o has a value of 1,12. Zuber and Findlay [14] gave the following expression for the rising velocity in the bubbly flow regime.

$$V = 1,53 \left[\frac{\sigma \cdot g(\rho_f - \rho_g)}{\rho_f^2} \right]^{1/4} \quad (15)$$

Experimental apparatus

The experimental vacuum evaporative crystalliser is shown in Figure 2. It consisted of a single steam jacketed Schedule 40 mild steel tube 1,3 m long and with an in-

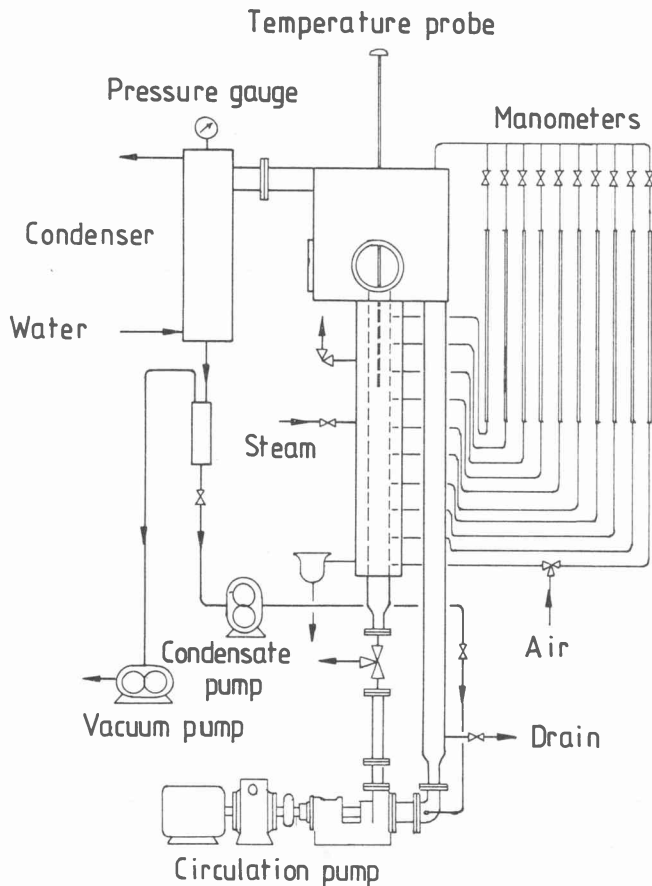


Figure 2 - Diagram of experimental pan

ternal diameter of 0,1m. These dimensions were representative of those used in some of the more recent vacuum pans. A previously calibrated positive displacement pump controlled the velocity of circulation in the tube. The condensed vapours were continuously returned to the apparatus so as to keep the concentration constant.

Bleeding air continuously through the pressure tappings

prevented entry of the boiling fluid into the manometers and clogging of the pressure tappings by crystallization of sugar. The centerline temperatures were measured by means of a probe introduced axially through the top of the apparatus. The void fraction along the tube was determined by gamma ray absorption, the source and detector being mounted on a traversing mechanism.

Experiments were carried out using syrup, molasses and massecuite. With these fluids it was possible to study both subcooled and saturated boiling under laminar conditions. The range of the operating variables is given in Table 1.

Results

The temperature of the inner tube surface was calculated using equations for film type condensation, and neglecting the resistance of any scale or oily film that may have been present on the heat transfer surfaces. The rate of heat transfer was based on the amount of condensate collected in a given time. It was assumed that the steam was dry and saturated. The saturation temperature of the boiling fluid was obtained by a method suggested by Batterham and Norgate [1] for concentrated sugar solutions.

Data for the heat flux transferred by bubble nucleation was obtained by subtracting the single phase forced convection heat flux calculated using equation (3), from the total heat flux imposed during the experiment. The residual term was correlated as suggested by Rohsenow and Griffith [11]. A plot of the generalized Prandtl number [13]

$$Pr = \frac{C_p K}{8.k} \left[\frac{u}{D} \right]^{n-1} \left[\frac{6n+2}{n} \right]^n \quad (16)$$

versus

$$\left[\frac{C_p \Delta T_s}{i_{fg}} \right] \left\{ \left[\frac{\phi_n}{i_{fg} \rho_f} \right]^{2-n} \left[\frac{\sigma}{g(\rho_f - \rho_g)} \right]^{\frac{n}{2}} \frac{\rho_f}{K_f} \cdot 8 \cdot \left[\frac{n}{6n+2} \right]^n \right\}^{-0.33} \quad (17)$$

which is shown in Figure 3 gave a value of 0,0673 for the constant C_{sf} and a value of 1,257 for exponent of the Prandtl number, so that m is equal to 0,257. It should be noted that the value of m recommended by Rohsenow

Table 1 Range of Operating Variables

Fluid	Tube inlet conditions			Flow behaviour index	Reynolds number	Prandtl number	Abs. pressure	
	Velocity (m/s)	Density (kg/m ³)	Apparent Viscosity (Pa.s)				Vapour* (kPa)	Steam (kPa)
Syrup	0,0379	1328	0,0296	1,000	357,000	147	12,5	100
	0,0792	1347	0,0791	1,000	85,400	388	27,0	122
Molasses	0,0462	1378	0,2050	1,00	60,300	695	11,5	113
	0,1210	1400	3,8900	0,932	1,820	15550	27,5	136
Massecuite	0,0462	1397	1,1300	0,980	13,200	3198	9,4	140
	0,1210	1445	12,9000	0,904	0,551	52268	26,5	173

*In vapour space

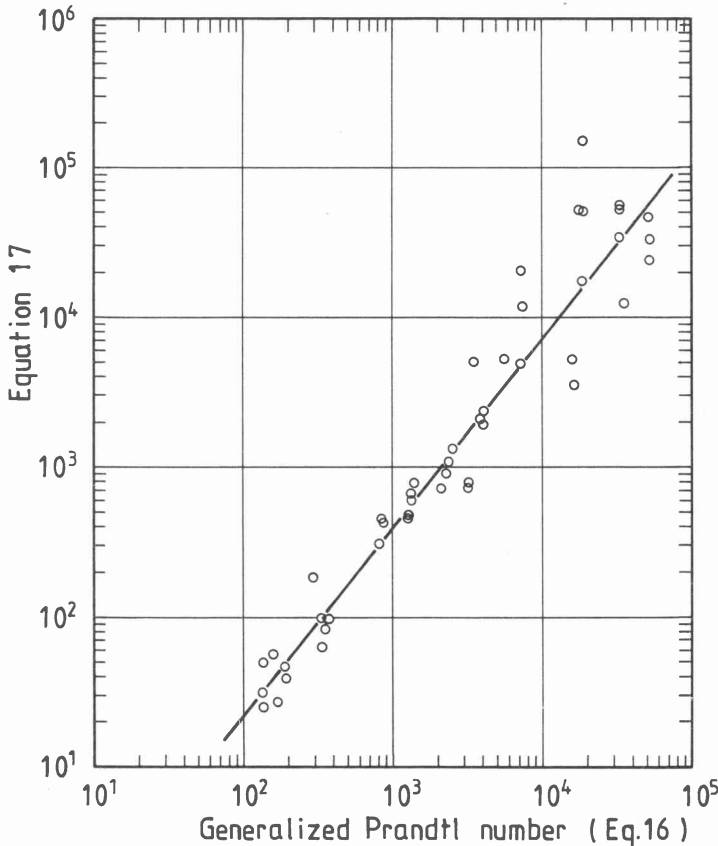


Figure 3 – Correlation of Prandtl number with Rohsenow's equation for nucleate boiling

and Griffith [11] is 0 for water and 0,7 for other liquids.

Figure 4 shows the typical void fraction profile measured by Gamma ray absorption. The low voidage associated with the highly subcooled region and the sudden increase that takes place at the point of bubble detachment can be seen. The flow distribution at the tube outlet was calculated using equation (14) in which

$$Q_f = \frac{\pi \cdot D^2 G (1 - x)}{4 \cdot \rho_f} \quad (18)$$

and

$$Q_g = \frac{\pi \cdot D^2 G \cdot x}{4 \cdot \rho_g} \quad (19)$$

The evaporation rates measured were used to calculate the mass vapour quality, x , at the tube outlet, and thus obtain the volumetric flowrate of the liquid and vapour phases. The value of V , the rising velocity of vapour, was calculated from equation (15).

The average value of C_o was 1,13 for all the runs where saturated boiling took place. This is close to the figure of 1,12 that Rouhani and Axelsson [12] found applies to most conditions.

As noted previously, the model of Griffith et al. [4] for estimation of the void fraction in the highly subcooled region is not suitable, for it gives incorrect results at low pressures. It was postulated that the void volume per unit surface area, a , would be proportional to the ratio of the boiling heat transfer coefficient h_{TP} to the single phase heat transfer coefficient h_{fo} multiplied by a proportionality constant B_o , which would be a function of the pressure

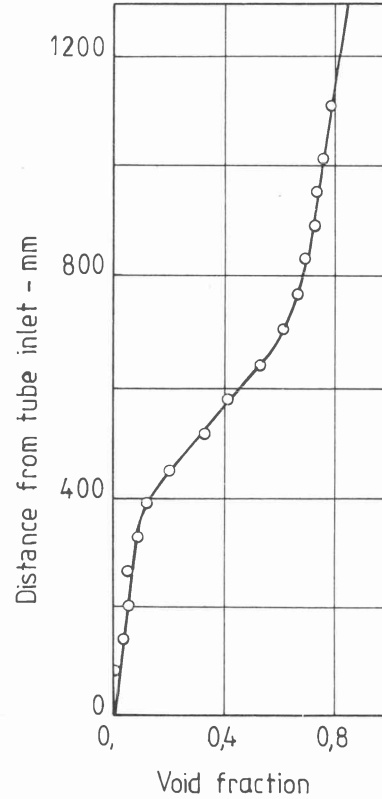


Figure 4

and of the hydrodynamic layer thickness which in turn, is related to the thickness of the thermal boundary layer (k_f/h_{fo})

$$a = B_o \cdot h_{TP} / h_{fo} \quad (20)$$

where

$$B_o = f(\text{Pr}, k_f/h_{fo}, \rho_f/\rho_g) \quad (21)$$

For a circular pipe the void volume per unit heated surface area can be related to the average local void fraction by the equation

$$a = \frac{\alpha \cdot \pi \cdot D^2}{4 \cdot \pi \cdot D} = \alpha \cdot D / 4 \quad (22)$$

which when substituted into equation (20) gives

$$\alpha = \frac{B \cdot h_{TP} \cdot k_f}{h_{fo}^2 D} \cdot (\text{Pr})^a \cdot (\rho_f/\rho_g)^b \quad (23)$$

The value of the constant B and of the exponents a and b was determined from a regression using data obtained from the void fraction profiles measured experimentally. The equation obtained was

$$\alpha = 0,00649 \frac{h_{TP} \cdot k_f}{h_{fo}^2 D} (\text{Pr})^{0,351} (\rho_f/\rho_g)^{0,414} \quad (24)$$

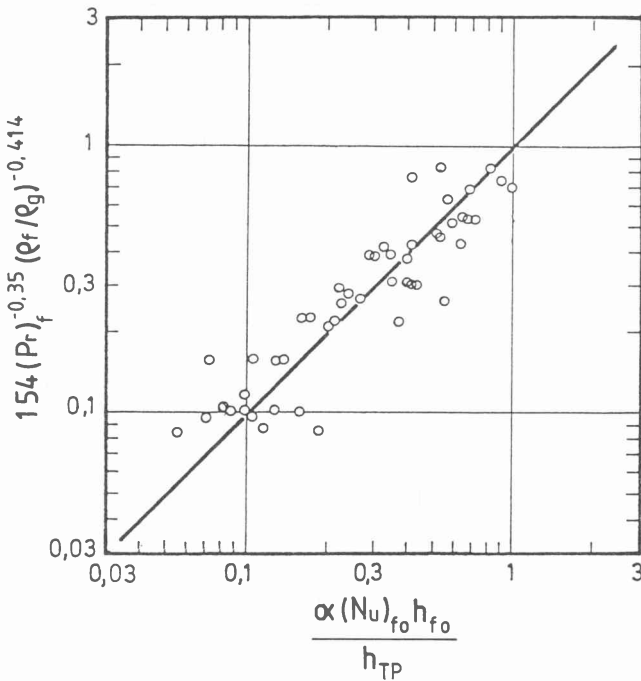


Figure 5 – Correlation for void fraction in the highly subcooled region

with a correlation coefficient of 0,91. The correlation is shown in Figure 5.

The factor η in the equation of Bowring [2] for subcooling at the point of bubble departure was found to increase with decreasing pressure in the range between 9 and 25 kPa. This is shown graphically in Figure 6. A possible explanation is that, as the void fraction increases with decreasing pressures, the vapour bubbles project further away from the heat transfer surfaces and are swept away more readily by the frictional drag.

A regression of η versus Pr indicated that the Prandtl number also has a significant influence on the subcooling at the point of bubble departure, with departure taking place at a higher subcooling as the Prandtl number increases. Again, this must result from the more intense frictional drag acting on the bubbles, since variation in the value of the Prandtl number is caused mainly by the change in viscosity.

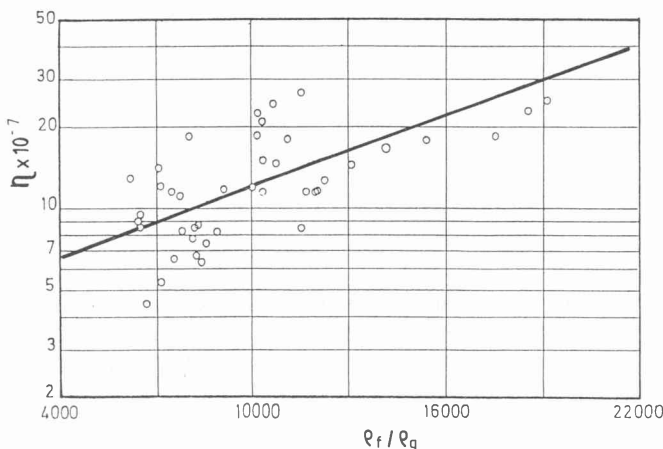


Figure 6 – Effect of pressure on factor for subcooling at bubble departure

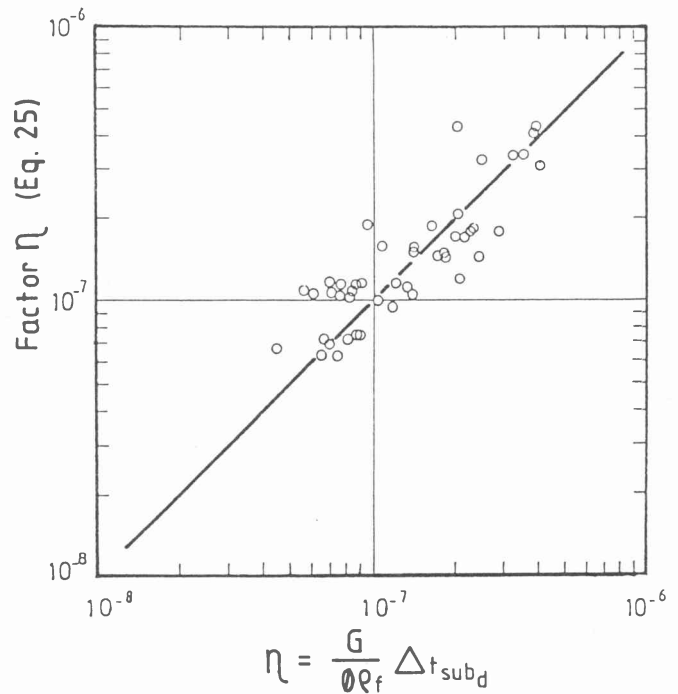


Figure 7 – Correlation of factor η for estimation of subcooling at bubble departure

It was found that under the experimental conditions the factor η can be expressed by the equation

$$\eta = 1,26 \times 10^{-8} (\text{Pr})_f^{0,254} \exp [6,73 \times 10^{-5} (\rho_f / \rho_g)] \quad (25)$$

The correlation coefficient obtained was 0,85 and the results are shown in Figure 7.

Conclusions

This research has shown that most of the equations proposed for boiling in turbulent flow under pressure require modification for use in laminar flow under vacuum. The superposition method suggested by Rohsenow and Griffith [11] works well provided that the value of the Prandtl number is raised to the power of 1,257 instead of a value of 1 for water and 1,7 for other fluids as recommended. A new equation is proposed for the estimation of the void fraction in the highly subcooled region, since that of Griffith et al. [4] is not suitable for low pressures. The equation of Bowring [2] for subcooling at the point of bubble departure must also be modified since it was found that this parameter is a function of the pressure and of the Prandtl number.

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