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SOME RECENT RESULTS AND DEVELOPMENT IN GAS–LIQUID FLOW: A REVIEW

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1. INTRODUCTION

There are now several books and reviews available on two-phase gas–liquid flow, (Tong 1965; Wallis 1969; Hewitt & Hall-Taylor 1970; Collier 1972; Hewitt & Semeria 1972). However, the subject of two-phase flow is a very active one and new ideas, techniques and analyses are constantly cropping up which advance the state of the art. This present review aims, therefore, at supplementing the previous ones by specifically dealing with papers which have been published since the beginning of 1972. It is convenient to discuss the work under the headings of General Equation Development, Pressure Drop, Void Fraction, Studies of Individual Flow Regimes, Accelerating Flows, Diffusional and Cross-Flow Effects and Dynamic Instability.

In covering such a wide field, the objective will be to give a general indication of trends; it will obviously be impossible to deal with the details of individual papers.

2. GENERAL EQUATIONS

The commonly used equations for two-phase flow systems often contain implicit and unrecognized assumptions. For some years now, Centre d'Etude Nucléaires (CEN) in Grenoble has had a program on the development and use of more general equations for two-phase flow. By starting with these equations, even if their solution requires considerable simplification, one can see more clearly the assumptions being made and estimate the magnitude of the errors involved. During 1972, a number of further papers were published on this work. Some of these papers are reviewed in their particular context below, but a more general discussion is given by Bouré (1973). In this report, the use of the method of characteristics is explored and the physical significance of the various results obtained is discussed. Its main conclusion is that progress in the knowledge of the basic mechanisms (especially those of the mass, momentum and energy transfers between the phases) is a necessary condition for a better description of two-phase flows, even for overall phenomena such as transients. Delhaye (1973) derived the jump conditions at the interfaces, taking into account the surface tension and the material surface properties (surface mechanical and thermodynamic properties). He calculated the local entropy sources.

Beattie (1972) has published a mixing length theory for two-phase flows in which the key assumption was that the turbulence exists predominantly in the continuous phase and that the discontinuous phase makes negligible contribution to the Reynolds shear stresses. The form of the equations is thus, for bubble flow,

$$\tau = \rho_L(1 - \alpha)\overline{u'_L v'_L} \qquad [1]$$

and for drops' flow (as in the gas core of annular flow),

$$\tau = \rho_G \alpha \overline{u'_G v'_G} \qquad [2]$$

where ρ is the density, α is the void fraction, u' is the instantaneous axial velocity and v' is the instantaneous normal velocity. Subscripts L and G denote liquid and gas phases, respectively. These equations can be integrated, provided the void profiles are known. The key assumption in Beattie's analysis is that there is a linear relationship between void fraction and the local continuous phase velocity. Thus, for bubble flow,

$$\alpha = au_L + b \qquad [3]$$

and for droplet flow:

$$\alpha = cu_G + d. \qquad [4]$$

Application of the Prandtl mixing length theory (i.e. assuming mixing length equal to ky, where y is the distance from the tube wall, and constant shear stress across the channel) yields the following results for bubble flow,

$$(1 - \alpha)^{3/2} = (1 - \alpha_c)^{3/2} - \frac{u_L^* a}{\frac{2}{3}k} \ln \frac{y}{R} \qquad [5]$$

and the following result for the drops' flow,

$$\alpha^{3/2} = \alpha_c^{3/2} + \frac{u_G^* c}{\frac{2}{3}k} \ln \frac{y}{R} \qquad [6]$$

where u_L^* and u_G^* are the friction velocities for the continuous phases respectively, α_c the center line void fraction and R the radius of the tube.

Beattie showed that [3]–[6] fit measured void and velocity profiles for bubble and annular flows. For annular flow, both Harwell and CISE (Milan) have used velocity deficiency equations of the form:

$$u_G - u_{Gc} = \frac{\sqrt{\tau_i/\rho_G}}{k'} \ln \frac{y}{R} \qquad [7]$$

and found that the coefficient k' is below the classical value of 0.4 especially when the droplet concentration is high. Beattie pointed out that his equations are capable of fitting the variation of the velocity profiles over the whole range. However, there appears to be no particular justification for the basic relationships [3] & [4].

3. PRESSURE DROP

Two-phase pressure drop has long been recognized as poorly correlated and the search for better correlations has continued during the past year. However, a paper by Marinelli & Pastori (1972) must have raised some doubts about whether all this work was worthwhile. They carried out calculations of pressure drop in Boiling Water Reactors (BWR) cores using a wide variety of published correlations for the pressure drop multipliers. They found that they could predict the pressure drop to within ± 5 per cent if the single phase friction factor and single- and two-phase spacer grid loss coefficients were experimentally determined. The exception in these comparisons was the fact that when the Martinelli void fraction and friction multipliers were adopted, errors of 20 per cent to 30 per cent were found. These comparisons, however, may be somewhat misleading insofar as the BWR case covers only a relatively small range of mass velocities. Very large errors can result using existing correlations, if a wide range is considered. An example of these deviations is shown in figure 1, which illustrates some recent work carried out by the Heat Transfer and Fluid Flow Service (HTFS), Harwell. Comparisons were made between the Chawla (1972a, b) correlation and the HTFS data bank. The correlation performs badly at high mass flows as do other recognized correlations in the low mass velocity region; it would seem advisable to pay particular attention to the flow pattern.

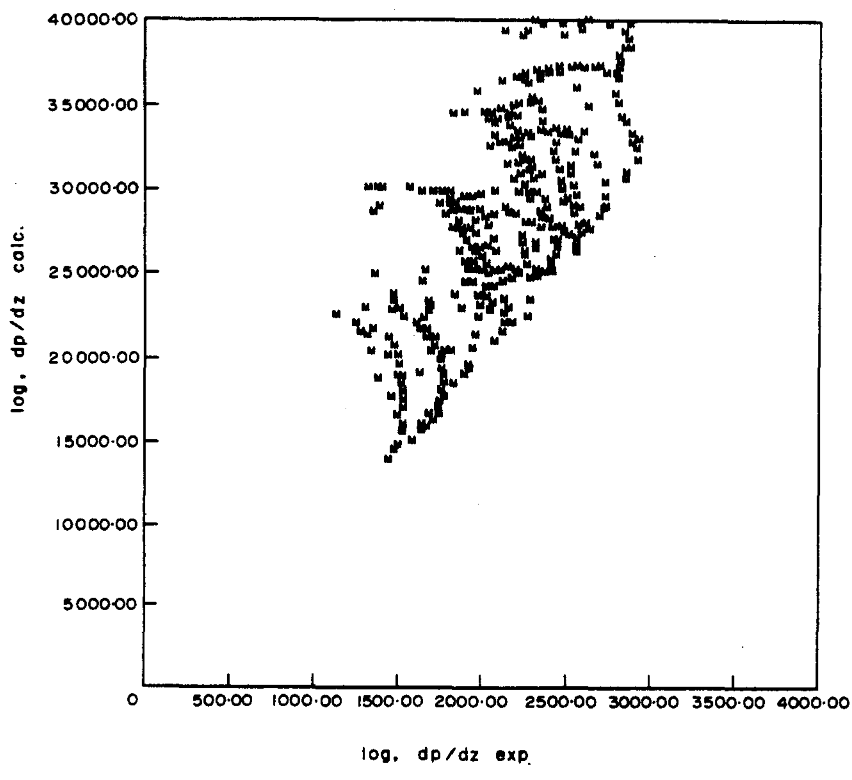

Figure 1. Comparison of Chawla correlation with data from HTFS data bank.

A number of different approaches to the problem of predicting pressure drop can be recognized. Referring to work published in the past year, one might classify these approaches as follows:

3.1. *Correlations in terms of system variables*

For some years, the team at CISE, Milan, Italy has been developing correlations for frictional pressure drop. Recently these have been extended to other geometries. The latest publication of this correlation (Lombardi & Pedrochi 1972) has the form:

$$\left(\frac{dp}{dz}\right) F = KG^n \frac{v_m^{0.86}\sigma^{0.4}}{d_0^{1.2}} \qquad [8]$$

where v_m is the mixture specific volume defined by:

$$v_m = xv_G + (1 - x)v_L \qquad [9]$$

where v_G and v_L are the specific volumes of the gas and liquid, respectively and x is the quality. In [8] σ is the surface tension, d_0 the equivalent hydraulic diameter and G the mass flux. The constant K has a value of 0.83 for round tubes and 0.213 for rod clusters and annuli. The exponent n has a value of 1.64 for round tubes and 1.6 for rod clusters and annuli. Consistent SI units should be employed in [8].

The CISE correlation is a reasonable fit to the experimental data, but tends to obscure effects of mass velocity and quality which are manifestations of changes in flow regime and which can be observed in detailed plots of the data. This is shown for some of the CISE data by Beattie (1971) as illustrated in figure 2. Beattie suggested a multiregion representation of pressure drop, including a low and high quality region in which the plot of $(\phi_{LO}^2 - 1)$ versus quality x is represented by a series of straight lines. The slopes of these straight lines are independent of mass velocity at low and high qualities, but depend on mass velocity in the medium quality region. This effect is obscured by the overall correlations and shows that two-phase pressure drop data is worthy of more detailed examination.

3.2. *Correlations based on dimensionless groups*

A wide variety of dimensionless groups have been used for correlating two-phase pressure drop. Another set of groups have been suggested by Kasturi & Stepanek (1972) and Stepanek & Kasturi (1972). They based their analysis on a separated flow model taking account of interactive effects by allowing the gas and liquid Reynolds numbers to affect respectively the liquid and gas friction factors. The correlating parameters used by Kasturi & Stepanek (1972) are as follows:

$$U = \left[\frac{v_L/(1-\beta)}{v_G/\beta}\right]^{0.80}\left[\frac{\rho_L(1-\beta)^2}{\rho_G\beta^2}\right]\mathrm{Re}_L^{0.70} \qquad [10]$$

$$T = \phi_G^2\,\mathrm{Re}_G \qquad [11]$$

$$S = \frac{\phi_L^2\,\mathrm{Re}_G^{0.45}}{\mathrm{Re}_L^{0.90}} \qquad [12]$$

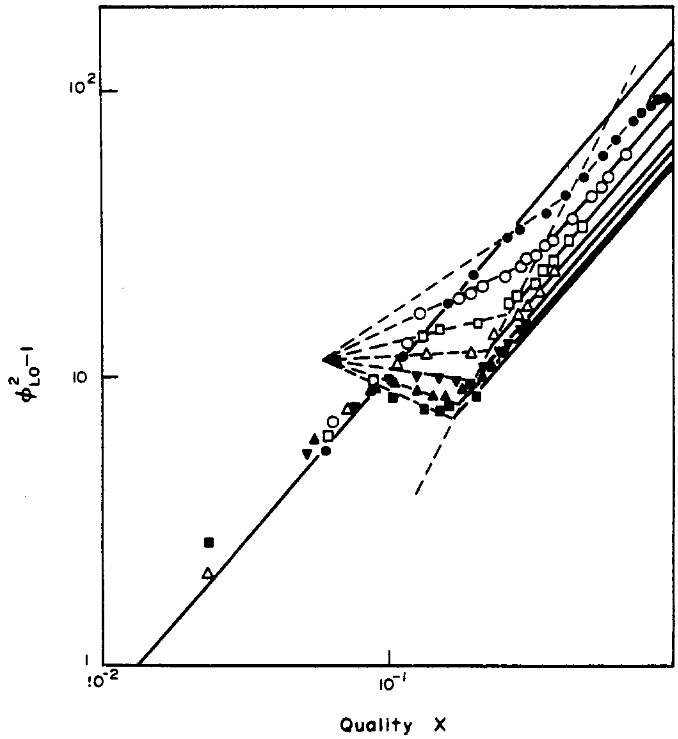

Figure 2. Pressure drop regimes revealed by plots of CISE data (Beattie, 1971).

where ν is the dynamic viscosity, β the gas to total volumetric flow ratio and Re the Reynolds number for the gas and liquid (flowing alone in the channel). The procedure for correlation was to relate the void fraction α to U via a graphical correlation and then to relate T and S to α, also graphically. It is unlikely that this correlation will work over a significant range of parameters though it represents an interesting way of introducing phase interaction into a separated flow model. Pressure drops for two-phase water–vapour and air–water flows in circular tubes have been measured by Borishansky *et al.* (1973). They correlated their data in a dimensionless number $(\Delta p_{TP} - \Delta p_L)/(\Delta p_G - \Delta p_L)$ is the parameter—rather than the usual, $\Delta p_{TP}/\Delta p_L$ or $\Delta p_{TP}/\Delta p_G$.

3.3. *Development of the Martinelli models*

The original forms of the Martinelli model are known to be inaccurate and to give a poor representation of the effects of system parameters, particularly of mass velocity. Chisholm (see for instance Chisholm & Sutherland 1969) has developed models based on the definition:

$$\phi_L^2 = 1 + \frac{C}{X} + \frac{1}{X^2} \qquad [13]$$

where X is the Martinelli parameter (square root of ratio of pressure drops of liquid and the gas flow alone). The original Martinelli curves for the various flow regimes can be fitted well by selecting a fixed value of the parameter C for each regime (for example, $C = 20$ for the turbulent-turbulent Martinelli regime). The Chisholm formulation gives the correct end point values for zero and unit quality. Also, specific values can be assigned to C corresponding to certain limiting cases. These include:

(i) Completely separated flow with no interaction between the phases which can be shown to correspond to $C = 2$.
(ii) Homogeneous flow can be represented by:

$$C = \left[\left(\frac{\rho_L}{\rho_G}\right)^{0.5} + \left(\frac{\rho_G}{\rho_L}\right)^{0.5}\right]. \qquad [14]$$

A further development consisted of applications of the model to evaporating flows (Chisholm 1972).

Johannessen (1972) examined the original Lockhart and Martinelli flow model and showed that stratified flows can be approximated by a separated flow model with no phase interaction. Although Johannessen derived a more complex equation his results can be closely represented by $C = 2$ using Chisholm's equation [13].

3.4. *System modelling*

Chawla (1972a, b) has developed a model including momentum transfer between the two phases. The basic equation employed is:

$$\left(\frac{\mathrm{d}p}{\mathrm{d}z}\right)_F = \frac{0.3164}{(Gd_0/\mu_G)^{1/4}} \cdot \frac{G^2 x^{7/4}}{2d_0\rho_G}\left(1 + \frac{\rho_G(1-x)}{x\varepsilon\rho_L}\right)^{19/8} \qquad [15]$$

where ε is a momentum exchange factor which is correlated as follows:

$$\varepsilon^{-3} = \varepsilon_1^{-3} + \varepsilon_2^{-3} \qquad [16]$$

$$\log \varepsilon_1 = 0.9592 + \log B^* \qquad [17]$$

$$\log \varepsilon_2 = [0.1675 - 0.055 \log(e/d_0)]\log B^* - 0.67 \qquad [18]$$

$$B^* = \frac{1-x}{x}\left[\frac{\mu_L^{1/6}\rho_L^{1/3}g^{1/6}}{\sqrt{G(1-x)}}\right]\left(\frac{\rho_L}{\rho_G}\right)^{-0.9}\left(\frac{\mu_L}{\mu_G}\right)^{-0.5}. \qquad [19]$$

This correlation is complex, bearing in mind the limited range as revealed by figure 1. However, much of the data shown in figure 1 may approach the condition where the two-phase velocities are equal and, as is pointed out by Chawla (1972), this case may be better approximated using the Bankoff variable density model.

An alternative phase-interaction model has been developed by Bandel & Schlunder (1972) who define the pressure gradients for the two phases in the channel as follows:

$$\left(\frac{\mathrm{d}p}{\mathrm{d}z}\right)_{TP,G} = (f_{Gw} + f_{Gi})\frac{\rho_G}{2}\bar{u}_G^2\frac{1}{\mathrm{d}_{eG}} \qquad [20]$$

$$\left(\frac{dp}{dz}\right)_{TP,L} = (f_{Lw} + f_{Li})\frac{\rho_L}{2}\bar{u}_L^2\frac{1}{d_{eL}} \qquad [21]$$

where f_{Gw} and f_{Lw} are the wall friction factors for the two phases and f_{Gi} and f_{Li} are the interfacial friction factors. Bandel & Schlunder proceed by assuming the two pressure drops as defined by [20] and [21] to be equal and by assuming that the wall friction factor is defined by the Blasius equation using a Reynolds number defined by the effective equivalent diameter of the respective phases. The result is:

$$\left(\frac{dp}{dz}\right)_{TP,G} = \left(\frac{dp}{dz}\right)_G\left(1 + \frac{F_{Gi}}{0.3164}\right)\left(\frac{d_0}{d_{eG}}\right)^{1.25}\left(\frac{1}{\alpha}\right)^{1.75} \qquad [22]$$

$$\left(\frac{dp}{dz}\right)_{TP,L} = \left(\frac{dp}{dz}\right)_L\left(1 + \frac{F_{Li}}{0.3164}\right)\left(\frac{d_0}{d_{eG}}\right)^{1.25}\left(\frac{1}{\alpha}\right)^{1.75} \qquad [23]$$

where F_{Gi} and F_{Li} are interphase friction functions which are given by Bandel & Schlunder:

$$F_{Gi} = F_{Li} = 0 \text{ for } \frac{\rho_L g}{(dp/dz)_{TP}} \geqslant \frac{50}{x} \qquad [24]$$

$$\left.\begin{aligned} F_{Gi} &= 0.6\left(\frac{1-x}{x}\right)^{0.2} \\ F_{Li} &= 0.18x^{0.3} \end{aligned}\right\} \text{ for } \frac{\rho_L g}{(dp/dz)_{TP}} \leqslant gx^{0.1}. \qquad [25]$$

Here, g is the acceleration due to gravity, d_{eG} the equivalent diameter of the gas flow and x the quality. To obtain the pressure drop, the value of (d_0/d_{eG}) is changed systematically until the pressure drops given by [22] and [23] become identical. The void fraction is obtained from conventional correlations.

Equation [24] describes (for horizontal tubes) the region of stratified flow with no phase interaction and [25] defines the annular flow region. The intermediate region between the two is covered by evaluating the pressure drop at the transition points and interpolating linearly on a log-log plot between the two extremes. Excellent agreement is obtained with pressure drop data (figure 3). For the conditions shown in figure 3, the Chawla correlation also gives good results, but the Bandel & Schlunder correlation gives much better agreement with experiments when a wider range of conditions is considered.

In predicting pressure drop in heated channels, the common practice is to use models and correlations of the type described above, but to integrate the pressure drop along the channel taking account of the variations in local quality, etc. In view of the fact that the correlation schemes outlined above do not correctly describe the physics of the processes, the departures from conditions of hydrodynamic equilibrium to which the adiabatic correlations may be presumed to approximate, are not taken into account. A typical example of this would be the case of evaporation in a tube in which dryout is occurring.

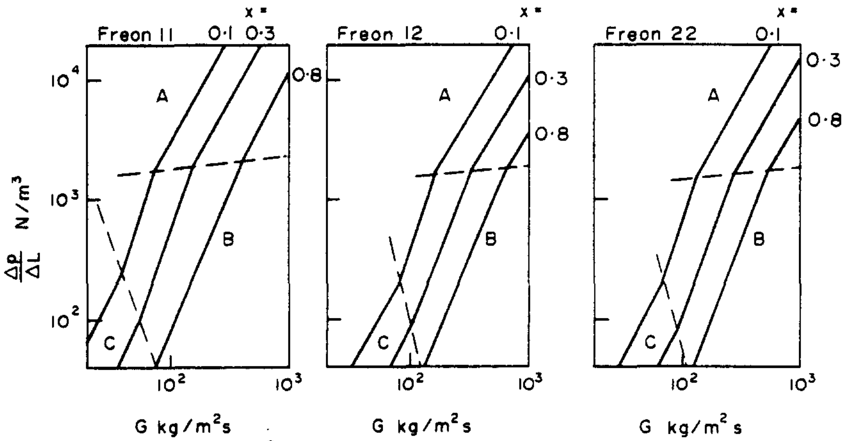

Figure 3. Data for pressure drop in refrigerant flows in horizontal tubes showing effect of flow regime (Bandel and Schlünder 1972). (A—annular flow; B—intermediate region; C—stratified flow.)

As the film thins, the interfacial roughness presented by the film to the gas core decreases and this can affect the pressure drop. Calculations carried out with the recent Harwell model for annular flow are illustrated in figure 4. The local interfacial shear stress (related directly to the local pressure gradient) passes through a minimum as the burn-out point is traversed. This is in accord with experimental observations made by Tarasova & Leontev (1965). However, nothing short of full modelling of the system is likely to produce this kind of answer and it is normally necessary to have recourse to integration of the adiabatic correlations. Moreover, it should be pointed out that the average pressure drops predicted for diabatic situations are often not affected significantly by the kind of variations illustrated in figure 4.

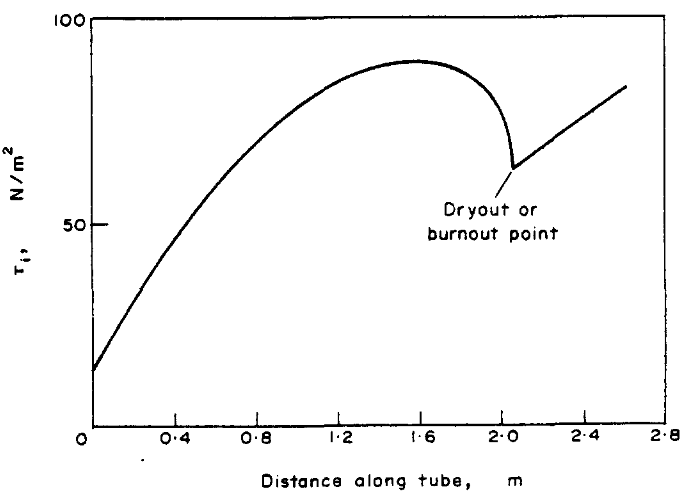

Figure 4. Interfacial shear stress (τi) as a function of distance along a heated tube in which water is being evaporated (unpublished data from AERE, Harwell).

The modelling approach has long been a subject of discussion, with particular reference to Freon modelling of critical heat flux (Bouré 1972). A further development of the modelling technique which can be used, as an alternative to general correlations, is reported by Bruce (1972) who models two-phase pressure drop between the Freon and water systems respectively. He concludes that the two-phase Euler number:

$$Eu_{TP} = (\Delta p_{TP} - \Delta p_{\text{Stat.}} - \Delta p_{LO})/(G^2/\rho_L) \tag{26}$$

is equal in the model and the prototype if the following criteria are met:

$$\begin{gathered}
\psi_p = \psi_M \\
\left(\frac{\phi}{G\lambda}\right)_p = \left(\frac{\phi}{G\lambda}\right)_M \\
\left(\frac{\Delta H_{iu}}{\lambda}\right)_p = \left(\frac{\Delta H_{in}}{\lambda}\right)_M \\
\left(\frac{L}{D}\right)_p = \left(\frac{L}{D}\right)_M \\
(\Theta)_p = (\Theta)_M \\
\left(\frac{\rho_L}{\rho_G}\right)_p = \left(\frac{\rho_L}{\rho_G}\right)_M
\end{gathered} \tag{27}$$

where the subscript p indicates the prototype and M the model. Θ is the angle of inclination of the channel and ψ is defined by:

$$\psi = \left\{\frac{GD^{0.2}}{\mu_L}\left[\frac{G}{\rho_L}\left(\frac{\rho_L^2}{\sigma g(\rho_L - \rho_G)}\right)^{0.25}\right]^{0.5}\frac{\Lambda}{1-\Lambda}\right\} \tag{28}$$

where

$$\Lambda = \frac{\rho_G}{\rho_L}\left(\frac{\mu_L}{\mu_G}\right)^{0.2}. \tag{29}$$

It is, of course, very difficult to obtain similarity in all these parameters and Bruce describes simple scaling factors which can be employed between Freon and water at various pressures.

To conclude this section, a few special cases are mentioned:

(a) *Viscous and non-Newtonian liquids.* Porzhezinskiy & Sagan (1971) reported data and presented a correlation for downflow of air-viscous fluid mixtures which is applicable to falling film evaporator systems. Studies of two-phase flow with non-Newtonian liquids are also reported by Mahalingam & Valle (1972) and by Srivastava & Narasimhamurty (1972). The increase in pressure drop observed with pseudoplastic aqueous solutions over that with pure water, depends on the flow regime (figure 5). In bubble flow and annular flow, where there is intense phase interaction, there is an increase in pressure drop. In the region of low interaction (wavy or slug flow) the difference in pressure drop is smaller. Mahalingam & Valle found that in the slug flow regime the pressure drop in two-phase flow could be actually less than that with the single flow of liquid alone. It was postulated

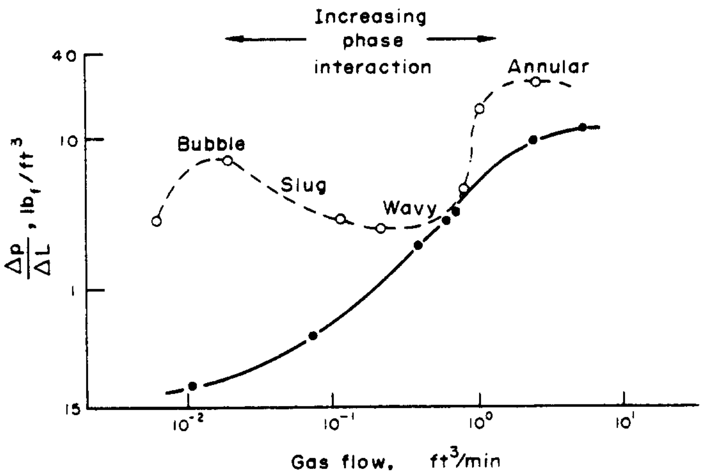

Figure 5. Pressure drop in flow of gas/pseudoplastic mixtures in horizontal tubes (Mahalingam & Valle 1972). —●— water, ——○—— water with 0.5% Methocel, −90 HG. Liquid flow, 0.0347 lb/s.

that this was due to the suppression of circulation in the liquid slugs in the non-Newtonian case.

(b) *Rod bundles.* Data for a 6 × 6 rod bundle were reported by Grillo & Mazzone (1972) who showed good agreement with the Baroczy correlation. Pressure drops across the fittings within the bundle agreed with a separated flow model, provided an arbitrary choice could be made about the relative velocity between phases.

(c) *Effect of pipe inclination.* A systematic study of the effect of pipe inclination from vertical upwards to vertical downwards flow was reported by Beggs (1972). He related the two-phase friction factor uniquely to the liquid hold up, independent of inclination of the channel. He found that the liquid hold up varied considerably with inclination and his results are discussed further in section 4 below.

(d) *Shell and tube heat exchangers.* Data for two-phase flow in a model shell and tube exchanger were reported by Grant & Murray (1972). They correlated their data in terms of the cross flow zone of the exchanger and the window zone:

Cross flow zone: $$\frac{\Delta p_{TP}}{\Delta p_{LO}} = 1 + (\Gamma^2 - 1)(x + 0.15x^{1/2} - 0.15x^{400}) \qquad [30]$$

Window zone: $$\frac{\Delta p_{TP}}{\Delta p_{LO}} = 1 + (\Gamma^2 - 1)(0.5x + 0.5x^2) \qquad [31]$$

where $$\Gamma^2 = \left(\frac{\Delta p_{GO}}{\Delta p_{LO}}\right). \qquad [32]$$

Data are also given for the flow patterns in shell and tube heat exchangers.

(e) *Coiled tubes.* A review of pressure drop and other phenomena in coiled tubes is given by Hopwood (1972) and the use of coiled tubes and other forms of steam generator for sodium reactor steam generators is given by Robin (1972).

(f) *Roughened tubes.* The effect of tube roughness on pressure drop is much less in two-phase flow than in single-phase flow, i.e. the friction multipliers used for roughened tubes are less than those for smooth tubes. The effect is illustrated in figure 6. Shires (1972) explains these effects in terms of relaxation of the flow pattern as a result of increased roughness. Obviously, if the flow is governed by the interfacial roughness, then the wall roughness has small effect and the increase in pressure drop above that of single phase flow is less than expected.

On the whole, one can state that considerable progress has been made in the field of two phase pressure drop and there is an increasing recognition of the importance of phase interaction and of the need to obtain a more fundamental understanding of the phenomena.

4. VOID FRACTION

There is still considerable interest in void fraction measurement and prediction. As was mentioned in Section 3, Stepanek & Kasturi (1972) produced a dimensionless correlation between void fraction and the group U defined by [10]. Gomezplata *et al.* (1972) presented a new correlation of void fraction:

$$U_s = [V_L/(K - \alpha) - V_G/\alpha] \qquad [33]$$

where U_s is the relative velocity, V is the superficial velocity of the liquid and gas phases and K is a parameter that incorporates the effect of the radial velocity profile. The relative velocity is given for bubble flow ($\alpha < 0.28$) by the equation:

$$U_s/U_t = 0.27 + 0.73(1 - \alpha)^{2.8}, \qquad [34]$$

where U_t is the terminal velocity of a single bubble. For slug flow conditions ($\alpha > 0.28$) the relative velocity

$$U_s = 0.35\sqrt{gD} \qquad [35]$$

was used. The parameter K was correlated by the following:

$$K = 1.0 - 0.25\exp(-0.000011\,\mathrm{Re}_{TP}) \quad \text{up flow}$$

$$K = 1.0 + 0.5\ \exp(-0.000105\,\mathrm{Re}_{TP}) \quad \text{down flow } (\alpha < 0.28)$$

$$K = 1.0 + 0.5\ \exp(-0.000015\,\mathrm{Re}_{TP}) \quad \text{down flow } (\alpha > 0.28)$$

$$\mathrm{Re}_{TP} = \frac{d_0\rho' V_t}{\mu'} \qquad [36]$$

$$\rho' = \rho_L(1 - \alpha)$$

$$\mu' = \mu_L(1 - \alpha)$$

$$V_t = V_L(1 - \alpha) + V_G\alpha.$$

These equations are solved by trial and error to give the value of α.

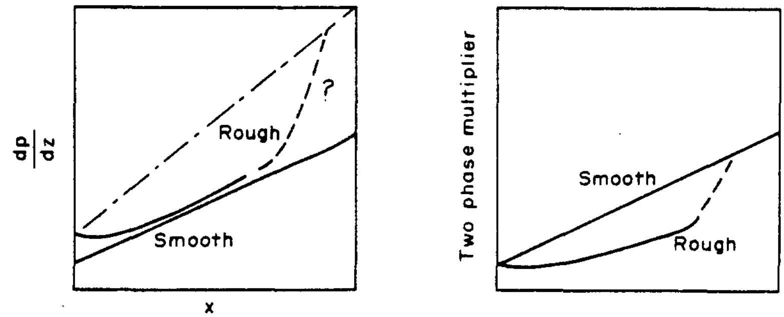

Figure 6. Effect of pipe roughness on two phase pressure drop (Shires 1972).

Measurements of void fraction are reported for a number of different situations as follows:

(a) *Rectangular channels.* Martin (1972) reported experimental data obtained for rectangular channels and found that the results were in agreement with the Bowring model at 8 MN/m^2 (80 bars).

(b) *Vertical tubes.* Bhaga & Weber (1972) reported data for void fraction in two-phase gas–liquid flow and for phase hold ups in gas–liquid–solid flow in vertical tubes. They also presented an analytical model for two and three phase flows which they show to be consistent with earlier models but, they claim, are more general.

(c) *Horizontal tubes.* Chawla (1972) discussed void fraction measurements and correlation in horizontal pipes and recommended the Bankoff method for calculation of liquid content and frictional pressure drop for conditions of high pressure with small vapour contents. For other conditions Chawla recommended his earlier correlation which involved the interaction parameter ε defined in [16]–[19].

(d) *Inclined tubes.* The data obtained by Beggs (1972) demonstrate the characteristic effect of the angle of inclination on liquid hold up (i.e. $(1 - \alpha)$) as shown in figure 7. Beggs

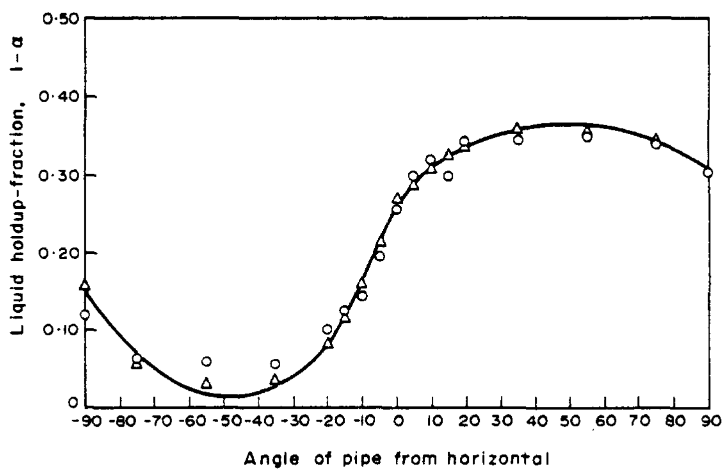

Figure 7. Effect of angle of inclination on holdup (void fraction) in two-phase flow in a tube (results of Beggs 1972).

presented empirical correlations for the effect of the angle of inclination on void fraction.

(e) *Coiled tubes.* Kasturi & Stepanek (1972) reported measurements of void fraction in coiled tubes with a variety of liquids. Systematic deviations were observed from the Lockhart–Martinelli and Dukler correlations.

(f) *In-pile measurements.* Uga *et al.* (1972) reported measurements of the steam void fraction in the Japan power demonstration reactor (JPDR). They used a resistance probe technique to obtain local and average voidage in the chimney and downcoming sections of the reactor.

(g) *Non-Newtonian liquids.* Data for void fraction with gas-pseudoplastic liquid mixtures in horizontal tubes were reported by Mahalingam & Valle (1972).

5. STUDIES OF SPECIFIC FLOW REGIMES

5.1. *Bubble flow*

Interest in two phase bubble flow continues and a number of studies of overall and local parameters have been reported in the past year. Kölbel *et al.* (1972) reported measurements of the void fraction of a bubble column by means of an X-ray absorption technique. An interesting feature was the measurement of the time variation of the gas content of the bubble column which, for low gas content, was characterized by a normal distribution. However, at high gas contents where bubble coalescence was occurring, the distribution of instantaneous measurements was skewed. The authors suggested the resolution of this skewed distribution into two normally partitioned distributions, the second corresponding to the proportion of large bubbles formed by coalescence. It was thus possible to follow the formation of the large bubbles along the tube. The results are illustrated in figure 8.

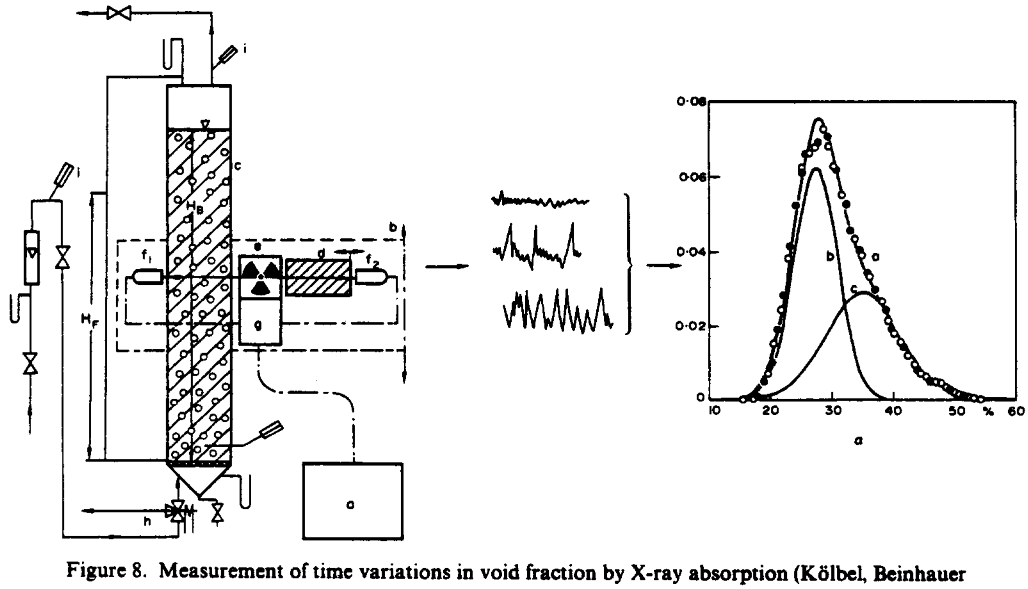

Figure 8. Measurement of time variations in void fraction by X-ray absorption (Kölbel, Beinhauer & Langemann 1972).

Another technique for measuring void fraction is to measure the electrical conductivity of the two phase mixture. Studies of electrical conductivity of mixtures of NaK and nitrogen and Hg and water were reported by Tanatugu *et al.* (1972). The electrical conductivities were close to the Maxwell theoretical value below about 20 per cent void fraction. Beyond this point the conductivity decreased sharply due to a transition from bubbly to slug flow.

Further data were obtained by using the resistance probe method for bubble size distribution and were reported by Uga (1972) and Reocreux & Flamand (1972). In-pile measurements were made in operating BWR's and the results are illustrated in figure 9. The data for penetration length distribution were converted to data for bubble diameter distribution by means of graphical differentiation. Reocreux & Flamand (1972) described the development of resistance probes for high speed flows. They used an alternating, square wave signal as the supply to the probe with a frequency between 1 kHz to 3 kHz. This prevented polarization and it was possible to process the data using a trigger unit and logical circuit with a bandwidth of about 1 MHz.

Further developments have been reported recently of optical methods for studying bubble flow. Hinata (1972) described the use of an optical fibre glass probe consisting of a bundle of fibres feeding light to a short rod with a combined bundle returning the reflected light. The probe was used for studies in mercury–air flow; in the mercury, the tip of the probe was perfectly reflecting and the light returned to a detector at the end of the returned fibres. Data was obtained for flow development and radial profiles.

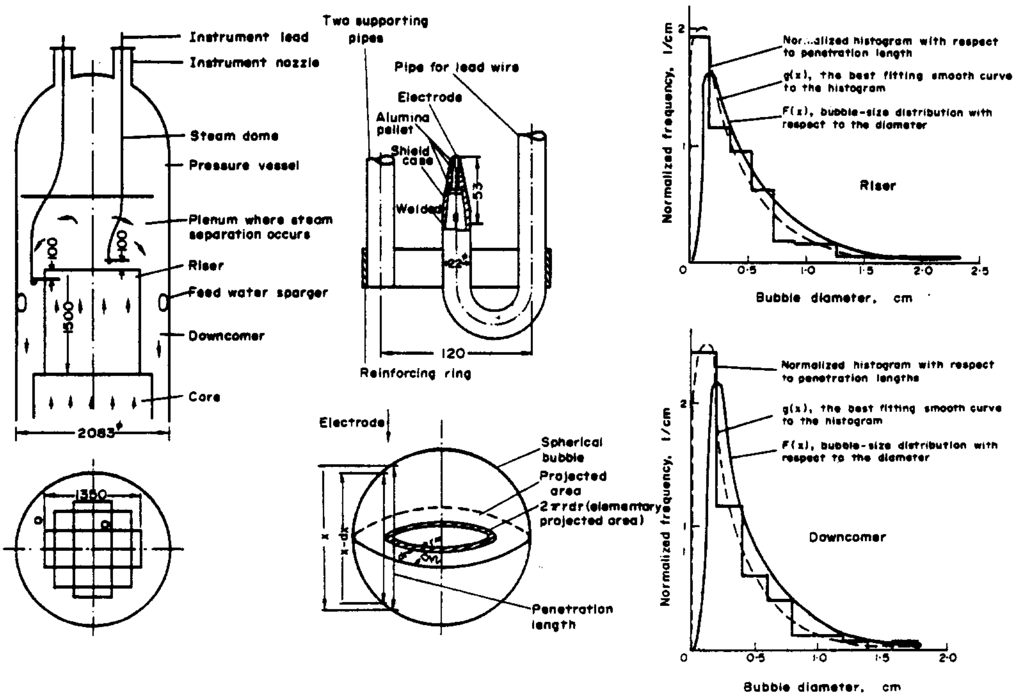

Figure 9. Bubble size distribution measurements in a BWR (Uga 1972).

Landa & Tebay (1972) described a technique for measuring bubble size by measuring the size distribution of pulses of scattered light emerging from a bubble flow irradiated by a laser beam. Bubbles in the size range 0.127 mm to 1.01 mm were measured. This technique looks potentially useful, though the resolution must decrease as the bubble population increases. The concentration limit was not clear from the paper.

Matthes (1972) discussed the interpretation of data from an optical device in estimating the correlation properties of a bubble field in two-phase flow. Bell (1972) described a technique for detecting the presence of steam in cooling pipes by means of ultrasonics.

5.2. *Slug flow*

The bubble velocity, u_G in vertical slug flow is often calculated from equations of the form:

$$u_G = C_1 V_T + C_2\sqrt{gD} \qquad [37]$$

where V_T is the total superficial velocity and D the diameter of the tube. Typically, C_2 has a value of 0.35. Nobel (1972) examined various relationships for slug flow and suggested that C_1 varies with the flow conditions in the channel and is approximately equal to the reciprocal of the local void fraction at the rear end of the slug flow bubble.

Interest in horizontal slug flow continues. Greskovich & Shrier (1972) presented data for slug frequency in large diameter pipes and showed that previously published correlations of slug frequency exaggerate the diameter effect. A new graphical correlation is presented. The pressure drop in horizontal slug flow can be represented as:

$$\Delta p_S = \Delta p_{LE} + \Delta p_F \qquad [38]$$

where Δp_{LE} is the component of pressure gradient due to acceleration of the liquid as it is scooped up into the liquid slug and Δp_F is the frictional component which is generated primarily by the high velocity liquid slug head. Rosehart *et al.* (1972) postulated that Δp_{LE} would be unaffected by the addition of drag-reducing polymer for given volumetric flows of the liquid and gas. However, Δp_F would be reduced and this would give a means of separating the two terms. They reported experimental results for horizontal slug flow and these are illustrated in figure 10. Analysis of the results indicated that Δp_{LE} was in the range 2–20 times the value of Δp_F and the ratio increased with increasing superficial velocity.

5.3. *Annular flow*

Annular flow is exceedingly complex and it seems impractical to obtain fundamental models for it. Even in the relatively simple situation of falling film flow, no satisfactory general solutions have been obtained. This fact was emphasized by Webb (1972) who demonstrated that one of the classical methods (Kapitza) was unlikely to produce meaningful answers. Friedel *et al.* (1972) presented a correlation for two-phase pressure drop based on energy transfer from the gas core to the surface waves in annular flow. The expression for energy transfer is certainly very much more complicated than the simple classical one they have embodied into their analysis.

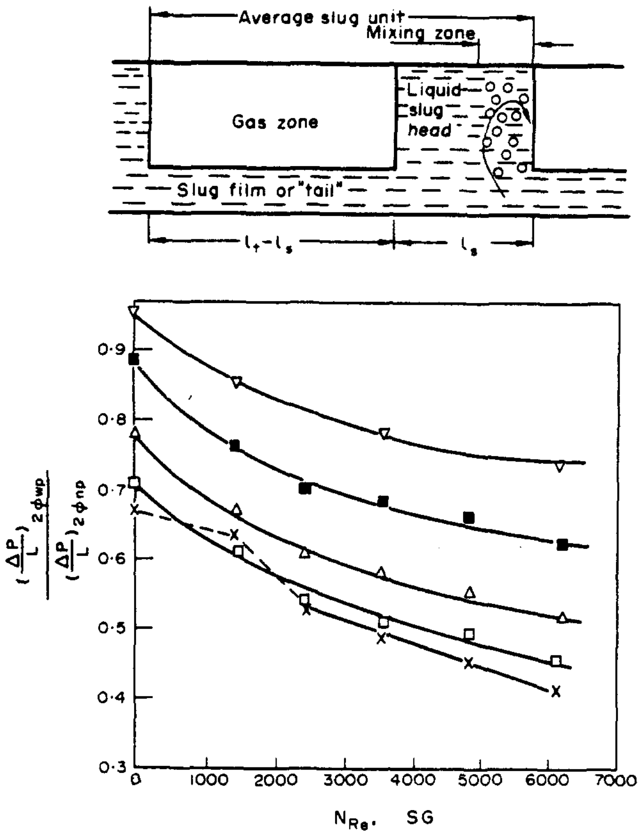

Figure 10. Two phase slug flow in horizontal tubes. Effect of drag-reducing polymer—use in estimating relative size of contributions to pressure drop. (Rosehart, Scott & Rhodes 1972).

In the absence of truly fundamental models for annular flow, it is still useful to have recourse to models which lie between fundamental ones and those which are purely empirical in nature, i.e. the dynamical model of annular flow. The dynamic model uses integral analysis coupled with localized correlations and relationships to describe the flow. This applies to both evaporating and condensing systems. Such an integral analysis for condensing annular-mist flow is given by Berry & Goss (1972), giving an encouraging prediction of the observed experimental data for condensation. Local equilibrium in entrainment is assumed. This certainly is not the case, though the deviations in condensation systems are not likely to be as significant as they are in evaporating systems. For example, if the equilibrium assumption were true in evaporating systems, burn-out would never occur!

Moeck & Stachiewicz (1972) presented an interesting new analysis for annular flow which has the following new features.

(a) A model for the liquid film was used in which the void fraction in the film was allowed to vary from the "base film" to the wave tips.

(b) The core-film interaction was broken down into the gas–film interaction and the droplet–film interaction, respectively. This differed from other analyses where the gas core was treated as homogeneous.

(c) Calculation of maximum drop size. The force on the droplets was calculated and the droplets' diameter estimated from the classical critical Weber number criterion. The results appear to be in good agreement with experimental data.

(d) Energy losses due to droplet–gas interaction (droplet drag) are shown to be significant at high entrainment and to contribute a large part of the total pressure drop.

This analysis is interesting and thought provoking. In particular, it stresses the importance of system energy loss due to the relative velocity between the droplets and the gas in the core of the annular flow.

One of the major difficulties in obtaining closed form solutions for annular flow is that of obtaining a suitable relationship for the entrainment rate. A new correlation for entrainment rate is given by Hutchinson & Whalley (1972); though this correlation is approximate, it can be built into models for annular flow with some success.

New measurements have been reported of some of the basic parameters of annular flow as follows:

(a) *Film thickness and film behaviour.* Benn (1972) described an experimental capacitance liquid film thickness monitor which is suitable for use with Freon systems. Butterworth & Pulling (1972) report a visual study of mechanisms in horizontal annular flow. A dyestuff was injected into the film and the movement of the dye step was studied as a function of flow parameters. In horizontal flow, the question is why does the liquid flow at the top of the tube? These dye injection studies strongly suggest that the surface waves, and particularly the "disturbance waves" play an important role in transferring the liquid to the top of the tube.

It is well known that local dryout phenomena can occur in rod bundles either upstream or downstream of the grid supports. An interesting set of experiments on dry patch formation in a liquid film with obstacles is reported by Shiralkar & Lahey (1972). The results are illustrated in figure 11. Dryout could occur either upstream of the obstacle or downstream, depending on its physical shape. The effect can be minimized by careful design of the obstacles and it was found that the presence of surface waves on the liquid film enhanced wetting.

(b) *Shear stress.* Instantaneous and simultaneous measurements of shear stress and liquid film thickness in horizontal film flow were reported by Zanelli & Hanratty (1973). It is shown that the shear stress and wave amplitude are in phase. However, there is evidence for laminarization of the film in the downstream zone of the wave and, in the presence of liquid entrainment, the shape of the shear stress profile is changed. Zanelli & Hanratty explained these results in terms of an approximate model. Kutateladze *et al.* (1972) have measured instantaneous shear stress in two-phase flow using the diffusion controlled electrolysis technique. The results are illustrated in figure 12. The bubble slug and annular flow regimes have characteristic spectral density curves as shown and the mean amplitude of the fluctuations in wall shear stress shows a characteristic peak in the slug flow regime.

(c) *Entrainment.* Entrainment mechanisms in annular flow were reviewed by Hewitt (1972). Yablonik & Khaimov (1972) described the measurements of the velocity of inception

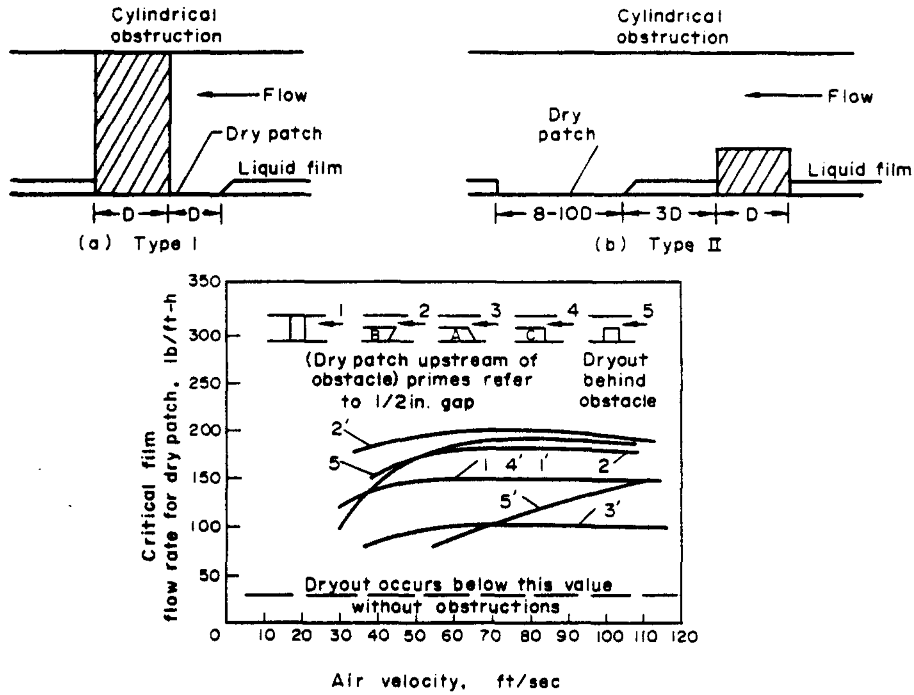

Figure 11. Effect of obstacles on breakdown of liquid films. (Shiralkar & Lahey 1972).

of droplet entrainment in two-phase annular flow using an electrical probe for detecting the presence of droplets. Interesting new light was thrown on the use of pitot probes in the gas core of annular flow due to a paper by Crane & Moore (1972). They presented a theoretical analysis of droplet trajectories approaching a pitot probe and showed that the differences between their calculations and an assumption of homogeneous flow was rather small

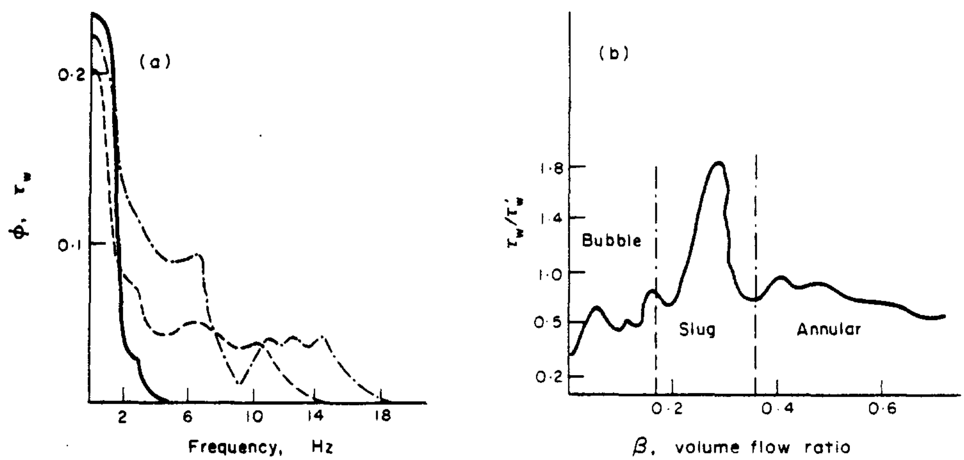

Figure 12. Measurements of wall shear stress variations in gas–liquid two-phase flow using the diffusion controlled electrolysis technique (Kutateladze *et al.* 1972).

(typically 2 per cent in extreme conditions). This confirms previous empirical findings that interpretation of pitot pressure in terms of the homogeneous model gives reasonable results for the local velocity. Determination of droplet size is important in annular flow studies and the interpretation of photographic evidence is always open to question. A critical assessment of photographic measurements for sprays is given by Clark & Dombrowski (1973).

6. ACCELERATING FLOWS

In the past year a large number of papers have been published in the area of critical two-phase flows and acoustic propagation in two-phase systems. A major review of critical flow and sound propagation phenomena was given by Hsu (1972) and a comprehensive book on supersonic two-phase flow has also been published (Saltanov 1972). Nuclear safety problems are of course the main justification of interest in this area. In addition, it should be emphasized that propagation problems arise almost inevitably when one tries to calculate numerically a non-steady flow. These propagation problems may result from the physical nature of the system or from an inadequate numerical solution of the equations. In any case they have to be studied.

Non-critical flows in orifices and nozzles were discussed at a conference on the subject held at the National Engineering Laboratory, Scotland in November, 1972.

6.1. *Non-critical converging–diverging flows*

Work on converging–diverging two-phase flows in two dimensional channels is reported by Hench & Johnston (1972) and by Wallis & Sullivan (1972). Hench & Johnston measured diffuser performance in air–water flows and their results are illustrated in figure 13. The pressure recovery factor C_p was defined as

$$C_p = \frac{p_2 - p_1 + \rho_h z}{0.5\,\rho_h V_T^2} \qquad [39]$$

where p_2 is the pressure at the outlet of the diffuser, p_1 the inlet pressure, ρ_h is the homogeneous density and z is the distance between the inlet and outlet of the diffuser. It was found that pressure recovery was close to the calculated ideal value in the bubble flow regime where there is strong interphase interaction. The pressure recovery decreases rapidly in the churn-turbulent region where the phase interaction decreases.

Flow metering systems such as orifices, nozzles and venturi meters are often used to determine two-phase flows and quality. Chisholm (1972) discussed compressibility effects in such systems. New data has been produced by Rooney (1972) for steam–water flows and by Dickson & Wood (1972) for air–oil flows, with sharp edged orifices. The use of venturis for two-phase flow metering was discussed by Harris & Shires (1972). Departures from equilibrium occur in two ways:

(a) Thompson (1972) discussed the effects of delaying vapour generation (flashing) in a liquid passing saturation temperature in the nozzle.

(b) Bakhtar (1972) discussed the effect of the formation of droplets on the delay of

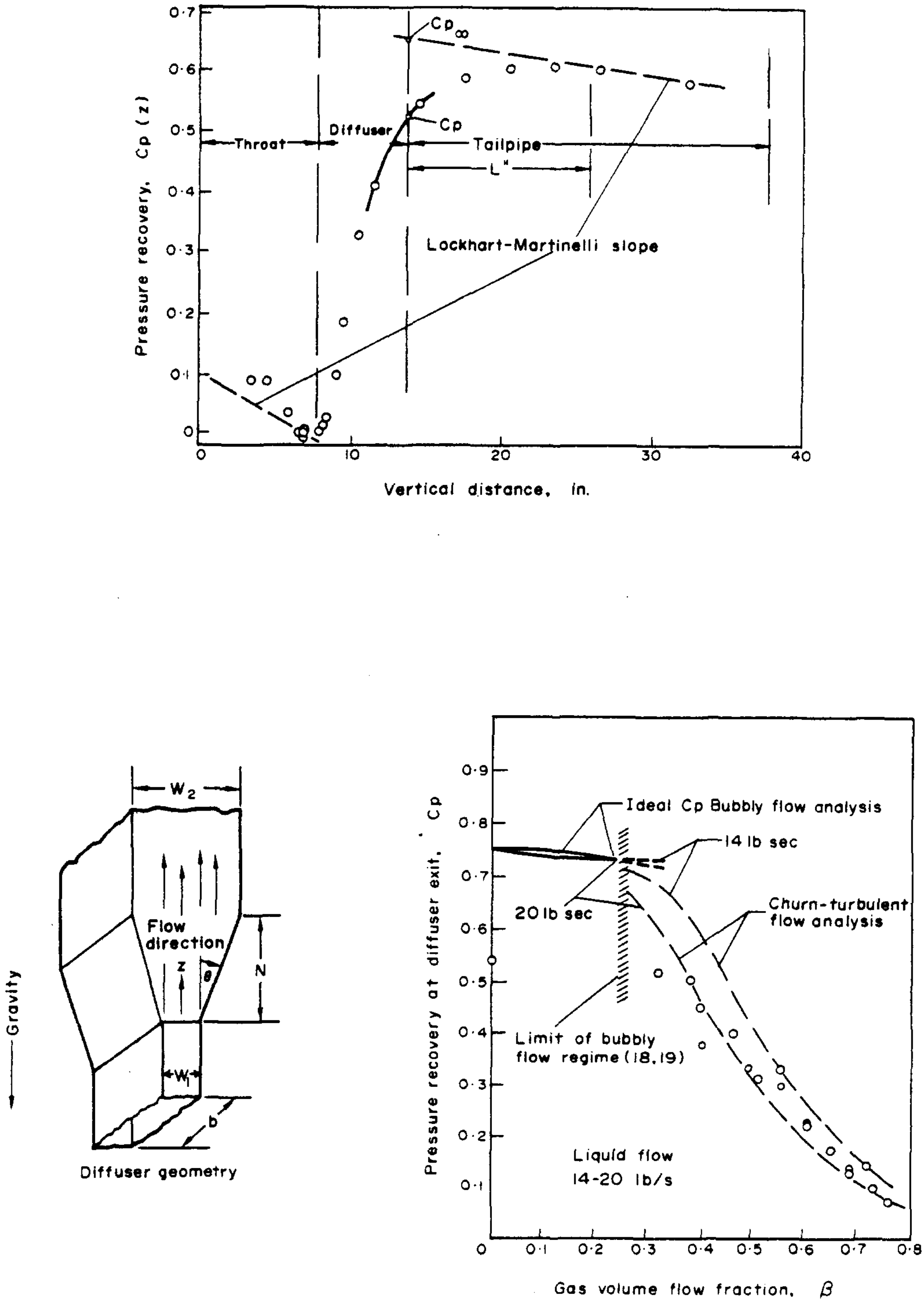

Figure 13. Performance of two dimensional diffuser (Hench & Johnston 1972).

pressure reduction in a flow of steam through orifices. This typically gives 5 per cent errors in steam orifice measurements.

Detailed local measurements of wet steam in nozzles are reported by Deich *et al.* (1972). These authors used a light scattering method to measure the drop size distribution along the lengths of the nozzle. It was found that the droplets broke up progressively as the flow proceeded through the nozzle and a reduction of mean drop size of the order of a factor of 3 was observed, in addition to a narrowing of the spectrum of droplet size distribution. The results are illustrated in figure 14.

6.2. *Critical two-phase flows*

The application of the CENG's general equations to critical two-phase flows is described by Reocreux (1972a, b, 1973). The flow was described in terms of six partial differential equations and three interface equations and simplifying assumptions were made to effect the solution. Studies of low-quality critical flows were reported by Baum (1972), De Hertogh *et al.* (1972) and Lafay & Picut (1972). The first of these investigations was with gas–liquid flow and the other two with vapour–liquid flows. Baum's data for the choked flow pressure ratio are illustrated in figure 15 and it can be seen that there are considerable disagreements between the measured and theoretical pressure ratio. Baum advances an interesting explanation for these deviations: the gas bubble, as it passes the rapidly changing pressure field approaching the nozzle throat, was considered in terms of a half cycle of bubble behaviour in an oscillating pressure field. Since the behaviour of bubbles in

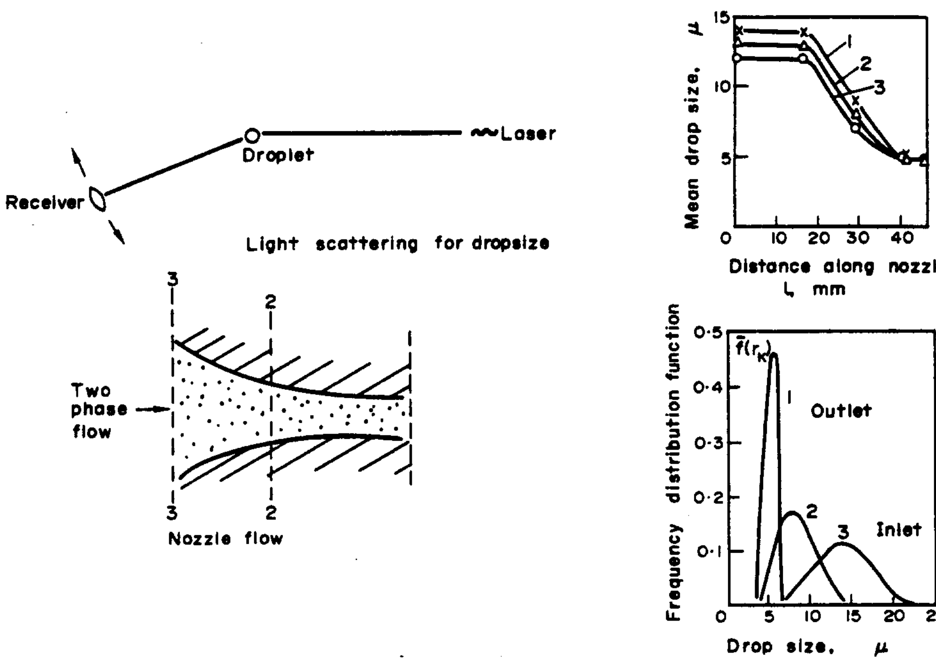

Figure 14. Light scattering method for drop size measurement in two-phase nozzle flow (Deich *et al.* 1972).

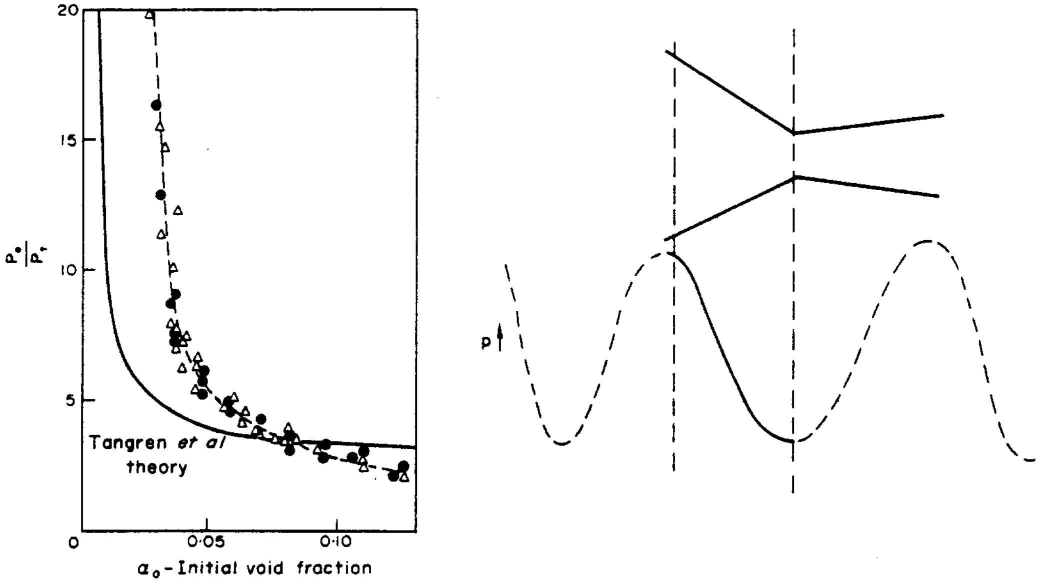

Figure 15. Pressure ratio in choked bubble flow. Deviations from theory explained in terms of bubble response. Baum (1972).

oscillating fields has been studied, Baum used the theoretical results to explain his data. Theoretically, the bubble growth lags behind the pressure field thus changing the critical flow behaviour. Baum calculated the frequencies and showed them to be consistent with this explanation.

Smith (1972) reported work carried out some years ago on an annular venturi in high quality critical flows. The pressure ratios observed fell between those for ideal isentropic flow and ideal isothermal flow. The critical flows were predicted from an analytical model, though it was necessary to introduce a "blockage factor" to account for the data. Similar blockage factors have been introduced by Wallis & Sullivan (1972) in their studies of critical flows through two dimensional nozzles.

Instabilities in critical nozzle flows of condensing steam were discussed by Yousif *et al.* (1972). In the expanding (supersonic) region, homogeneous condensation can occur giving rise to a "condensation shock". Under certain conditions, when the flow beyond the shock is reduced to a value below sonic, pressure oscillations can occur which can be important in nozzle flows and steam turbine applications.

Single phase fluids near the critical point can exhibit anomalous critical flow behaviour similar to that of two-phase flows. Hendricks *et al.* (1972) discussed the choked flow of fluid nitrogen in this region.

6.3. *Pressure pulse and sound transmission*

In two-phase mixtures, the velocity of propagation of pressure pulses is much lower than for the individual phases. Fascinating new data have become available in this area which shed new light on the phenomena.

Böckh & Chawla (1972) have reported experiments in a channel partially filled with water as illustrated in figure 16. The form of the pressure pulse translation is illustrated in the same figure and it is seen that the time for the propagation of the first change in pressure (Δt_B) is much shorter than that for the main change (Δt_Z). The first time interval corresponds to the propagation of a sound wave in the gas phase whereas the second interval corresponds to the two-phase propagation. On the basis of this data, Böckh & Chawla suggested that the energy in the particular wave is inversely proportional to the square of its velocity. The data obtained for the completely separated and stationary situation are in good agreement with those for a flowing two-phase mixture. Frankin *et al.* (1973) have, on a similar facility, studied the response of the interface to a periodic disturbance imposed on the gas-phase. For some frequency, they observed a resonance effect with standing waves on the water surface.

Data for the effect of frequency on the velocity of sound in Freon/Freon vapour mixtures was reported by Kokernak & Feldman (1972) and are shown in figure 17. A positive resonance effect is observed and this may correspond to the bubble response characteristics referred to by Baum (1972) in his study of critical flow. Kokernak & Feldman show predictions from a theoretical model; unfortunately, details of this model are not given in their paper, although presumably the model is based on some form of resonance phenomenon.

Saltanov & Kurshakov (1972) performed a theoretical study of the motion of small particles behind an oblique shock. Davydov (1972) applied the method of characteristics

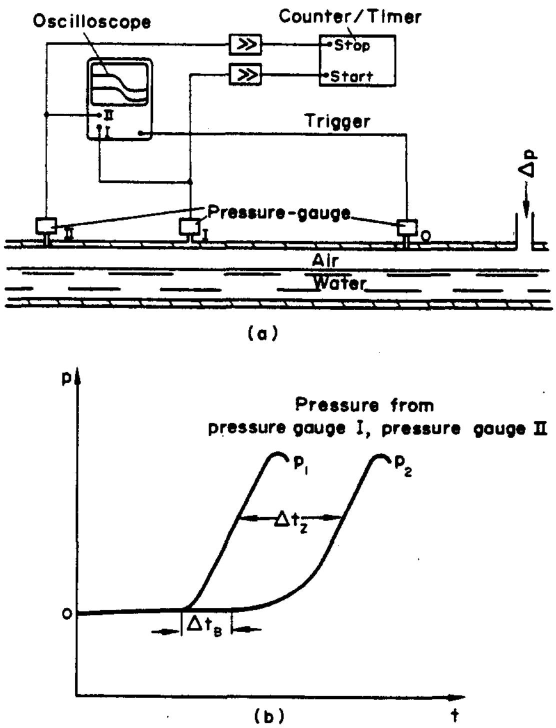

Figure 16. Pressure pulse transmission experiment of Böckh and Chawla (1972).

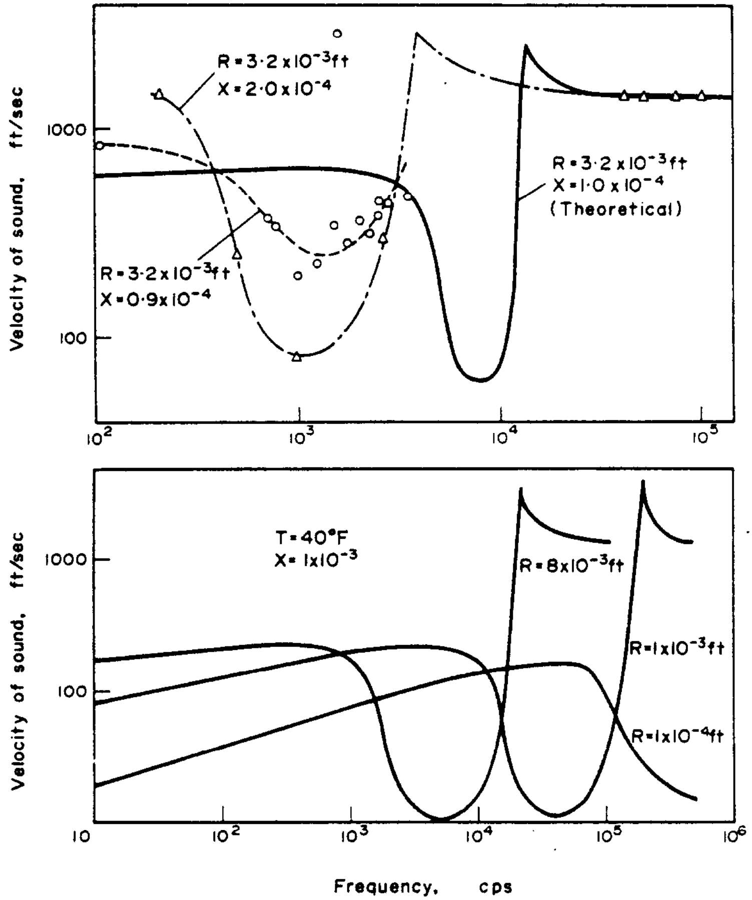

Figure 17. Effect of frequency on velocity of sound in bubbly Freon 12 vapour liquid mixtures (Kokernak & Feldman 1972).

to the problem of condensation in supersonic nozzles without equilibrium. Mukhachev & Susekov (1972) worked with suspensions and Batrak (1972) discussed the existence of bubbly supersonic flows. Using the CEǸG general equations, Bouré (1973) applied the method of characteristics to two-phase flows. In the framework of classical assumptions and neglecting the transport phenomena, the equations giving the velocities of small disturbances in two-phase flows have in practical cases two real roots and two complex ones. The real roots correspond to disturbances which Bouré called "mixed waves", to emphasize the fact that they involve void disturbances as well as pressure disturbances. These mixed waves originate from, or yield, dynamic waves in the single phase parts of the system. The complex roots possibly correspond to damped (or amplified) disturbances.

6.4. *Jet phenomena*

In the safety analysis of fast reactors, consideration is given to a jet of fission product gases released from a puncture in one of the fuel elements and impinging on neighbouring rods. If this jet gives rise to dewetting of the neighbouring element, a self-propagating process may occur. An experimental study of gas jets submerged in liquid and impinging onto a solid surface was reported by Bell *et al.* (1972). The system used is illustrated in figure 18. An important aspect of this problem is whether droplets are entrained from the edge of the jet and impinge on the surface to give it an effective spray cooling. Bell *et al.* measured the extent of entrainment by the ingenious technique shown in figure 18, using two separated chambers and measuring the transfer of fluid from one chamber to the next. Extensive detailed studies of the jet are reported and these include measurements of local void fraction, film thickness and droplet frequency. A theoretical analysis of this situation is given by Chawla (1972) who analysed the jet in terms of the Kelvin–Helmholtz instability. The trend predicted from the analysis appeared to be consistent with the experiments of Bell *et al.*

7. MIXING AND CROSS-FLOW IN TWO-PHASE SYSTEMS

7.1. *Residence time distribution and axial mixing*

Tracer injection techniques were used by Midoux & Charpenter (1972) for gas–liquid flow in packed columns, Chen (1972) for two-phase bubbly flow and Jagota *et al.* (1972)

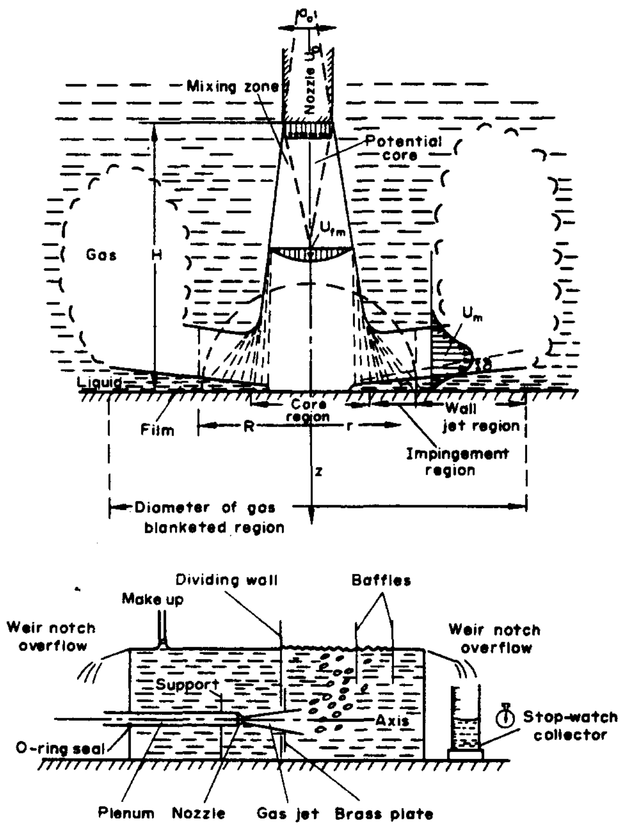

Figure 18. Experiments on two-phase jets—Bell, Boyce & Collier (1972).

for gas–liquid annular flow in vertical tubes. Midoux & Charpenter made the interesting observation that injection of a tracer at two points in a system with detection at one point, can give exactly the same information as injection at one point and detection at two points. The results obtained by Chen were for axial mixing of the liquid phase in the bubble column. A tracer was injected at the bottom of the bubble column and its progress followed by means of a conductivity cell. Chen deduced that particle dispersion increased with both gas and liquid rates.

Jagota *et al.* (1972) discussed the case of radial and axial mixing and residence time in annular flow with liquid entrainment. Their results throw interesting light on the interpretation of the experimental observations in annular flow in which pulses of radioactive tracer were injected and the transit time of the pulse was measured. Their work shows that the most important mechanism for axial dispersion is the interchange of the droplets between the gas core and the liquid film. The effects of this interchange are illustrated in figure 19. Without the interchange, the response to the injection shows two peaks (assuming that the tracer could be uniformly spread across the liquid phase) but with a large degree of interchange, there would only be one peak whose velocity of traverse would give an indication of the mean liquid velocity. Jagota *et al.* gave a detailed analysis of these interchange processes which explain not only pulse injection results but also those where there is a continuous injection.

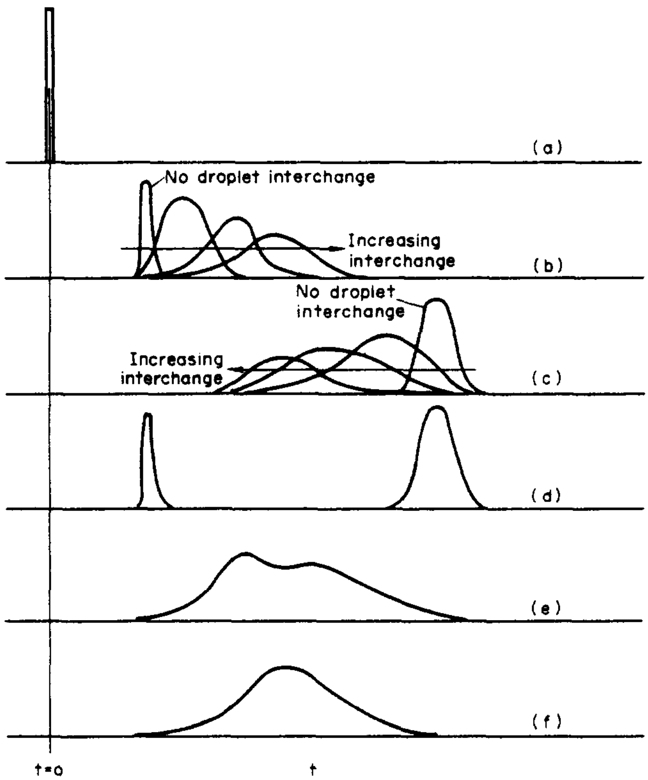

Figure 19. Movement of tracer in the liquid phase of annular flow with entrained liquid.

7.2. *Mixing effects in rod bundles*

Castellana & Casterline (1972) reported subchannel flow and enthalpy distributions in a channel typical of a BWR. Simultaneous measurements of flow and enthalpy were made at the exits of two subchannels in a 16 rod electrically heated full scale model of a typical fuel rod geometry. The authors compared their results with predictions of the COBRA II subchannel code with various values of mixing coefficient. Data for air–water flows in simulated rod bundle geometries were reported by Rudzinsky *et al.* (1972). Further developments of the theoretical representation of turbulent mixing in rod bundles was reported by Beus (1972).

Singh & St. Pierre (1973) measured the mixing rates between subchannels in a square rod array at low pressure with an air–water mixture. They found that mixing depended on the flow regime, decreasing exponentially with subchannel mass flux and quality and increasing with gap spacing.

A major difficulty in analysing flow in rod bundles is that many of the models assume a tendency towards constant void fraction in adjacent subchannels by a process of mixing. However, there is no evidence that an equilibrium situation of constant void fraction is a true one. Instead, there is a tendency for the lighter phase to migrate towards the regions of high velocity.

8. DYNAMICS AND STABILITY OF TWO-PHASE FLOW SYSTEMS

A general review on instability was given by Bergles (1972). A detailed description of the Kjeller computer code (RAMONA II) was given by Holt & Rasmussen (1972). Bouré (1972) gave a preliminary description of the GEVATRON code being developed at CENG. Dynamical models for a large once-through steam generator were described by Sanathanan *et al.* (1972).

Experimental results for low pressure instability of flow in a natural circulation loop in which boiling occurs in the riser of either *n*-pentane or water were reported by Matsui (1972). The results appear to be consistent with those observed in previous studies.

Yadigaroglu & Bergles (1972) reported experiments with boiling Freon which showed higher mode oscillations. The paper also includes theoretical developments with particular reference to the effect of heat capacity of the channel wall on the density wave oscillations.

Transient behaviour of two-phase systems as distinct from stable behaviour has been discussed by Gonzalez-Saltalo & Lahey (1972) and Hopkinson (1972). Gonzalez-Saltalo & Lahey considered the case of a heated channel where the inlet flow rate undergoes an exponential decay. This is representative of the pump rundown type of accident in a nuclear reactor. An interesting development in this paper is the use of the method of characteristics. The partial differential equations that describe the system become ordinary differential equations in a Lagrangian frame of reference and can be integrated exactly for many flow transients of interest. As the authors pointed out, the method could also be used for more complex cases where a numerical solution may otherwise be required. Hopkinson (1972) described a new model for the dynamics of steam drums with particular reference to their transient behaviour in the case of rapid pressure variations.

9. CONCLUSION

It will be seen from the above that interest in gas–liquid two-phase flow is continuing and that interesting new developments are forthcoming. It is particularly encouraging to see an increasing interest in the basic mechanisms of the phenomena. This approach is beginning to have its influence on the design methods used in engineering practice.

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