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AN INVESTIGATION OF VOID FRACTION IN LIQUID SLUGS FOR HORIZONTAL AND INCLINED GAS–LIQUID PIPE FLOW

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Abstract—The void fraction in liquid slugs has been determined for air–water flow in horizontal and near-horizontal pipes by a newly-developed conductance probe technique. A semi-empirical correlation has been developed and compared with the present measurements and available data. This correlation predicts reasonably well the observed effects of diameter, inclination and physical properties.

Key Words: void fraction, slug flow, gas–liquid flow

1. INTRODUCTION

In physical models of slug flow, certain parameters are acknowledged to have a critical influence on average hold-up, pressure drop and frequency. Two of these parameters, the Dumitrescu bubble velocity (U_B) and the gas volumetric fraction in the slugs (ϵ_s) have, unfortunately, been shown to be complicated functions of the three-dimensional flow in the slug unit, see, for instance, Bendiksen (1984) and Fernandes *et al.* (1983).

Since all comprehensive slug flow models are one-, or at most two-dimensional (e.g. Dukler & Hubbard 1975; Malnes 1982; Ferschneider 1983; Fernandes *et al.* 1983), separate models for U_B or ϵ_s are required. Several attempts to develop such models for ϵ_s have been made, from pure empirical correlations, e.g. Gregory *et al.* (1978), to more sophisticated ones, incorporating two-dimensional effects, Fernandes *et al.* (1983). The lack of reliable empirical data over a reasonable range of parameters such as pipe diameter (D), inclination (ϕ) and fluid type, however, is still quite evident.

Hubbard (1965) measured ϵ_s for air–water flow in a 3.81 cm i.d. horizontal pipe. The scatter in these data, as pointed out by Gregory *et al.* (1978), is so large that they are of limited value in verifying predictive models for ϵ_s.

In their work, Gregory *et al.* (1978) used two different horizontal test sections with pipes of i.d. = 2.58 and 5.12 cm, and lengths of 575 and 340 dia, respectively. They adopted capacitance sensors developed by Gregory & Mattar (1972) to measure ϵ_s, using air–light oil as test fluids. Although there is also some scatter in their data, a weak diameter effect might be inferred.

Gregory *et al.* (1978) also proposed a correlation for ϵ_s, but without incorporating diameter or inclination effects:

$$\epsilon_s = 1 - \frac{1}{1 + \left(\frac{U_M}{8.66}\right)^{1.39}}. \qquad [1]$$

Malnes (1982) proposed another correlation, based on the same data as Gregory *et al.* (1978):

$$\epsilon_s = \frac{U_M}{C_c + U_M}, \qquad [2]$$

where U_M is the total superficial velocity, and the dimensional coefficient C_c (in m/s) was given as

$$C_c = 83\left(\frac{g\sigma}{\rho_L}\right)^{1.4}.$$

Ferschneider (1983) reports two experiments performed in the Boussens loop with i.d. = 14.6 cm, for $\phi = 0$ and $+4°$, using natural gas (90% methane) and condensate. Based on these experiments, and possibly other data from the Boussens loop, Ferschneider (1983) proposed another empirical correlation for the void fraction in the liquid slugs:

$$\epsilon_s = 1 - \frac{1}{\left\{1 + \left[\frac{U_M}{\sqrt{\left(1 - \frac{\rho_G}{\rho_L}\right) gD}}\right]^2 \Bigg/ \left(\frac{A}{Bo^\beta}\right)^2\right\}^2}, \qquad [3]$$

where

$$Bo = \frac{(\rho_L - \rho_G) g D^2}{\sigma} \quad \text{is the Bond number}$$

and the coefficients A and β were not given. There is no dependency on pipe inclination in [3], although from the data, figures 5–10, this dependency is quite clear.

A major incentive behind this study has been to obtain high-quality void fraction data to investigate, in particular, possible diameter and inclination effects. A series of experiments has been carried out with pipes of i.d. = 5.00 and 9.00 cm, test section lengths of about 350 and 200 dia, respectively, at inclinations ranging from $-3°$ to $+0.5°$, using air–water as test fluids.

The void fraction in the liquid slugs was measured by conductance probes made of ring electrodes mounted flush with the wall. A semi-empirical correlation for ϵ_s has been developed and compared with available data. This correlation predicts very well both the observed diameter and inclination effects.

2. EXPERIMENTAL SET-UP

The experimental results presented in this paper have been obtained in a new loop designed and assembled at the University of Pisa for the study of gas–liquid flow in inclined and horizontal pipes.

The apparatus includes an inclinable bench 17 m long, the slope of which can be varied continuously in the range $\pm 7\%$ by means of a motorized support, see figure 1. The test sections consisted of two transparent Plexiglas tubes of i.d. 50 and 90 mm, and maximum length equal to the length of the bench. The tubes were made of carefully flanged, 2 m long, interchangeable sections mounted on the bench by precision supports.

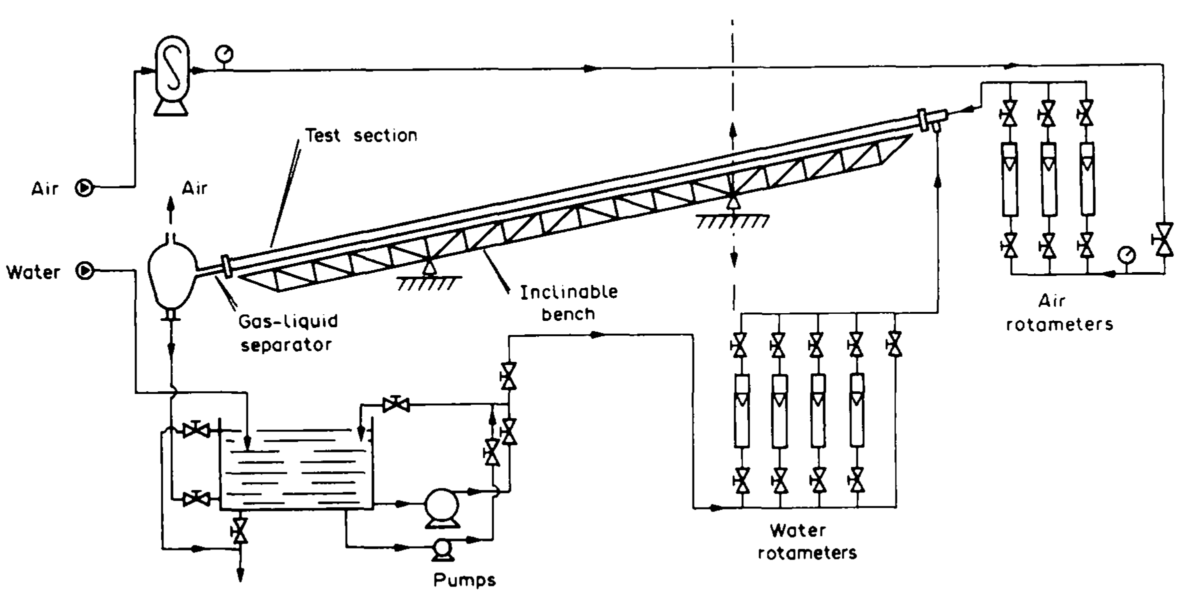

Figure 1. Schematic diagram of the test section.

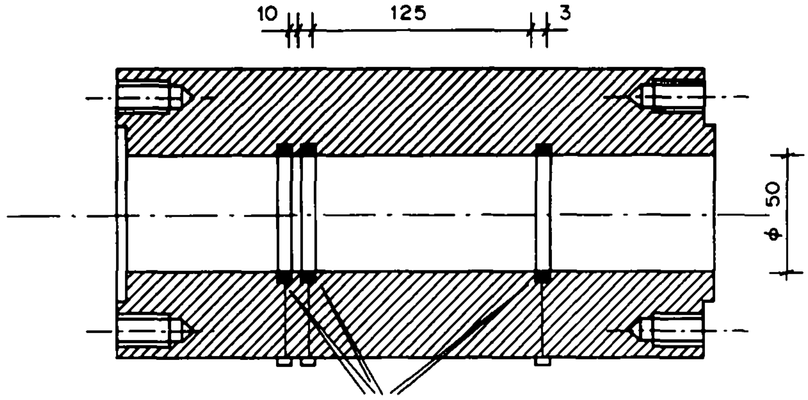

Figure 2. Schematic diagram of the conductance probe.

The liquid, water in present experiments, is circulated by two centrifugal pumps of different size. Air is supplied from a high-pressure line. Air and water are metered by two sets of rotameters. The air outlet pressure was in all cases close to atmospheric conditions.

Liquid and gas are fed to the pipe through a T-section. At the liquid entrance, stratified flow conditions are created by a thin diaphragm which separates the liquid inlet (from below) from the gas inlet (along flow direction in the pipe).

The void fraction in the liquid slugs was measured by a conductance probe made of two ring electrodes mounted flush to the pipe wall. A detailed description of the behaviour of this probe has been reported elsewhere (Andreussi *et al.* 1988), where the case of slug flow is considered in detail.

A schematic diagram of the probe used in the 50 mm line is shown in figure 2. As can be seen from this figure, the distance between the electrodes is 2.5 dia.

As shown by Andreussi *et al.* (1988), the conductance G_E between two ring electrodes spaced far apart (>2–3 dia) on the wall of a pipe containing a liquid with conductivity γ_L is given by

$$G_E = \gamma_L \frac{A_L}{D_E}, \qquad [4]$$

where A_L is the cross-section of the liquid phase and D_E is the distance between the electrodes.

Equation [4] can also be adopted in bubble flow with the conductivity given by the equation proposed by Maxwell (1881) or Bruggeman (1935) to describe the electric properties of a dispersion:

$$\gamma = \frac{2H_L}{3 - H_L}\gamma_L \quad \text{(Maxwell)} \qquad [5]$$

or

$$\gamma = H_L^{3/2}\gamma_L \quad \text{(Bruggeman).} \qquad [6]$$

In order to avoid any effects of the actual value of the liquid conductivity on the experimental readings, the measurements of G_E have been normalized with respect to the value of G_E relative to the pipe full of liquid. The normalized value of G_E, G_E^* can then be expressed by the following equation:

$$G_E^* = \frac{2H_L}{3 - H_L} \qquad [7]$$

or

$$G_E^* = H_L^{3/2}. \qquad [8]$$

Static and dynamic calibration tests reported by Andreussi *et al.* (1988) have shown that [8] gives a very good fit to the experimental readings. This equation has been adopted to convert conductance measurements into liquid hold-up.

The instantaneous value of the electric conductance has been determined by the electronic circuit developed by Brown *et al.* (1978). The signals from the circuit were recorded and analysed by a

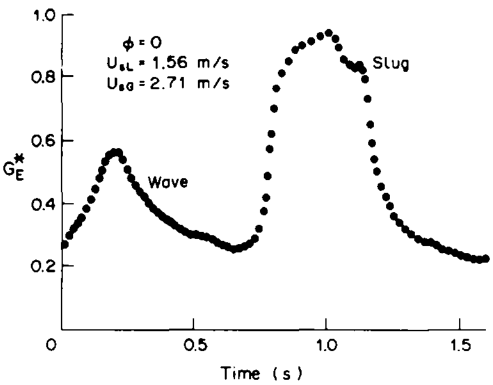

Figure 3. Typical signal in the presence of an aerated slug.

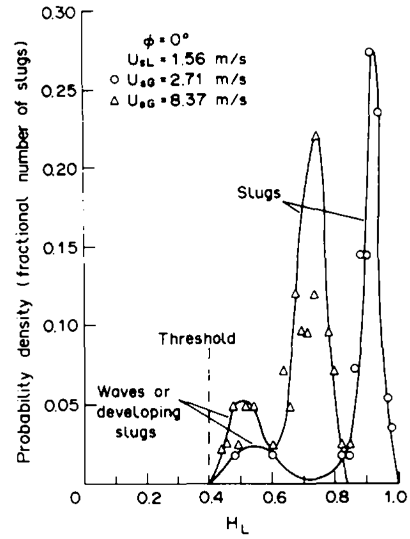

Figure 4. Probability density of slug (and wave) liquid hold-up.

personal computer. A typical signal recorded in the presence of aerated liquid slugs is shown in figure 3. From this figure it can be observed that, due to the appreciable distance between the electrodes, the arrival of a slug does not cause a sudden change in the hold-up, but the slug is preceded (and followed) by a gradual increase (decrease). The void fraction in the slug has been determined considering only the central part of the signal relative to each slug.

Close to the transition to annular flow, large disturbance waves or developing slugs can be encountered along with liquid slugs. These waves have been discriminated from slugs determining the probability density function of the hold-up in the slugs and the waves. To this purpose a threshold value for H_L, equal to 0.4 for all flow conditions, has been fixed and all perturbations above this threshold have been considered. Typical density distributions obtained by this procedure are shown in figure 4. As can be seen, these functions are, in general, bimodal and it is quite straightforward to derive from such graphs the limiting hold-up separating waves from slugs.

3. PREDICTION OF ϵ_s

The main step in the present analysis is to relate the average void fraction in the liquid slug to a balance between the net entrainment rate of small bubbles at the slug front and the net loss rate at the slug tail. It is well-known (e.g. Fernandes *et al.* 1983) that, in the range of mixture velocities encountered in slug flow, a large vortex motion is set up by the entrance of the liquid film at the slug front, as indicated in figure 5. The net entrainment rate of small bubbles at the slug front is thus assumed to consist of two terms, a pure production rate (Q_1) and a loss rate (Q_2) of small bubbles transported back into the producing Dumitrescu bubble by the vortex motion.

For a fully-developed situation with no further accumulation of bubbles in the slug, it follows from continuity that the net loss rate at the slug tail (Q_3) equals the net production rate, or

$$Q_3 = Q_1 - Q_2. \quad [9]$$

In their analysis of void fraction in the slugs, a major assumption made by Fernandes *et al.* (1983) is that the gas bubble "production" rate into the slugs (Q_1), equals that of the gas left behind with respect to the average gas velocity in the Dumitrescu bubble. For vertical flow, Q_1 may be computed assuming rotational symmetry, and certain additional assumptions on the liquid film and gas velocity profiles, but for other inclinations the flow is three-dimensional and asymmetric, complicating the calculation.

In the present analysis, it is assumed that the bubble production rate is proportional to the film flow rate with respect to the average Dumitrescu bubble velocity, U_B, provided this flow rate exceeds a certain lower limit:

$$Q_1 = C_1 A[(1-\epsilon_B)(U_B - U_{Lf}) - U'_{Mf}], \quad [10]$$

where $(1-\epsilon_B)$ is the asymptotic liquid holdup and U_{Lf} is the average velocity of the film under the Dumitrescu bubble far from the nose. This production term incorporates three effects: the bubble production rate increases with increasing relative slug and film velocities; it also increases with film height, as it is more difficult to merge a thick film smoothly into the slug scooping it up.

The limiting velocity, U'_{Mf}, corresponds to the mixture velocity below which the film flow is continuously merged into the slug without swirl motion and there is no production of small bubbles.

It is further assumed that Q_2 is proportional to the void fraction in the liquid slug and the rise velocity of small bubbles (U_{G0}):

$$Q_2 = C_2 A \epsilon_s U_{G0}. \quad [11]$$

For surface-tension dominated flow, U_{G0} is given by Harmathy (1960) as

$$U_{G0} = 1.53\left[\frac{g\sigma}{\rho_L^2}(\rho_L - \rho_G)\right]^{1/4}. \quad [12]$$

The net bubble loss rate at the slug tail (Q_3) is given by continuity as

$$Q_3 = A\epsilon_s(U_B - U_{Gs}) + \Delta Q_f, \quad [13]$$

where ΔQ_f is a possible contribution to Q_3 from bubbles carried in the liquid film, and partly released to the following Dumitrescu bubble, see figure 5.

The fraction of bubbles in the film is, according to Fernandes *et al.* (1983), negligible for vertical flow and is assumed to be zero for all inclinations. If there were, however, a known, significant fraction of bubbles in the film, ΔQ_f may easily be computed and accounted for.

U_{Gs} is the average gas velocity in the liquid slug, approximately given by Malnes (1982) as

$$U_{Gs} = S_D(U_{Ls} + U_{G0}\sin\phi), \quad [14]$$

where U_{Ls} is the average liquid velocity in the slug, S_D is a distribution slip-ratio and U_{G0} is the rise velocity of the small bubbles in the slug. Relation [9] then yields

$$\epsilon_s(U_B - U_{Gs}) = C_1(1-\epsilon_B)(U_B - U_{Lf}) - C_1 U'_{Mf} - C_2\epsilon_s U_{G0}. \quad [15]$$

The continuity of the liquid in a system of reference following the Dumitrescu bubble is expressed as

$$(1-\epsilon_B)(U_{Lf} - U_B) = (1-\epsilon_s)(U_{Ls} - U_B) \quad [16]$$

which, inserted in [15] together with [14], yields

$$\epsilon_s = \frac{C_1(U_B - U_{Ls}) - C_1 U'_{Mf}}{(1+C_1)(U_B - U_{Ls}) + (1-S_D)U_{Ls} + C_2 U_{G0} - S_D U_{G0}\sin\phi}. \quad [17]$$

As is well-recognized, in fully-developed slug flow the Dumitrescu bubble velocity may be expressed as

$$U_B = C_0 U_{Ls} + U_0. \quad [18]$$

It has been shown experimentally by Nicklin *et al.* (1962), and theoretically for vertical flow by Collins *et al.* (1978), that $C_0 U_{Ls}$ is very close to the local velocity of liquid in front of the tip of

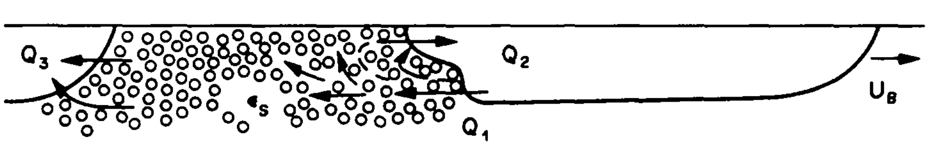

Figure 5. An illustration of the gas flow in the liquid slug.

the bubble nose, and that U_0 is the bubble rise or drift velocity in a tube with stagnant liquid at infinity. The distribution parameters C_0 and U_0 have been found empirically to be functions of inclination angle, Froude and inverse Eötvos numbers (Fr and Σ).

In the following, the values of Bendiksen (1984) have been applied. A rather abrupt change in C_0 and U_0 was observed for $\mathrm{Fr} \approx 3$; where, for example, horizontally, C_0 increases from 1.05 to 1.20 and U_0 decreases from $0.54\sqrt{gD}$ to 0, if surface-tension effects can be neglected ($\Sigma \leqslant 0.05$).

Introducing [18] into [17], yields

$$\epsilon_s = \frac{U_{Ls} - U_{Mf}}{\left(\frac{1}{C_1} + 1\right)\left(U_{Ls} + \frac{U_0}{C_0 - 1}\right) + \frac{1 - S_D}{C_1(C_0 - 1)} U_{Ls} + (C_2 - S_D \sin\phi)\frac{U_{G0}}{C_1(C_0 - 1)}}. \qquad [19]$$

This may be expressed as

$$\epsilon_s = \frac{U_{Ls} - U_{Mf}}{\beta U_{Ls} + U_{M0}}, \qquad [20]$$

where

$$\beta = \left[\left(1 + \frac{1}{C_1}\right) + \frac{1 - S_D}{C_1(C_0 - 1)}\right],$$

$$U_{Mf} = \frac{(U'_{Mf} - U_0)}{(C_0 - 1)} \qquad [21]$$

and

$$U_{M0} = (C_2 - S_D \sin\phi) \cdot \frac{U_{G0}}{C_1(C_0 - 1)} + \frac{U_0}{C_0 - 1}\left(1 + \frac{1}{C_1}\right).$$

The velocity U_{Mf} gives the real lower limit, below which no bubbles may be produced. It is seen to be reduced due to the drift velocity U_0. This is to be expected, as the drift velocity is related to the thickness and velocity of the film under the Dumitrescu bubble.

The physical significance of U_{M0} may be clarified by looking at horizontal flow at high velocities ($U_0 = 0, \sin\phi = 0$) and

$$U_{M0} \to \frac{C_2}{C_1} \cdot \frac{U_{G0}}{C_0 - 1}.$$

In view of [10] and [11] U_{M0} is then proportional to the fractional loss of bubbles back into the Dumitrescu bubble (C_2) relative to the net production rate (C_1).

For steady flow, a further simplification of [20] is possible. Continuity of the total flow then yields

$$U_M = U_{SG} + U_{SL} = U_{Gs}\epsilon_s + U_{Ls}(1 - \epsilon_s) \qquad [22]$$

or, applying [14],

$$U_{Ls} = \frac{U_M - \epsilon_s S_D U_{G0} \sin\phi}{(1 - \epsilon_s) + S_D \epsilon_s}. \qquad [23]$$

This, inserted in [20], yields a second-order equation for ϵ_s.

Observing that $S_D \approx 1.0$ and $U_M \gg \epsilon_s S_D U_{G0} \sin\phi$, [23] reduces to the familiar $U_M = U_{Ls}$, and in this case

$$\epsilon_s = \frac{U_M - U_{Mf}}{\beta U_M + U_{M0}}. \qquad [24]$$

In the limit of high mixture velocities, [24] approaches a maximum value:

$$\epsilon_s \to \epsilon_{sm} = \frac{1}{\beta}.$$

In view of the definition of β in [21], the above relation states that the asymptotic void fraction in the slugs (ϵ_{sm}) only depends on the production rate, and to a smaller extent on the inclination angle and distribution slip ratio (if $S_D \neq C_0$). For $\mathrm{Fr} \leqslant 3$, the slip ratio S_D is about equal to C_0 [$S_D = C_0 \approx 1.05$; see Malnes (1982) and Bendiksen (1984)]. From [21] it then follows that $\beta = 1$.

The lack of reliable measurements of the void fraction at high mixture velocities suggests keeping $\beta = 1$ for the full range of mixture velocities investigated. This assumption can easily be removed when more data is available. The structure of the proposed correlation is acceptable from a physical standpoint, essentially in the range of low and moderate mixture velocities, which is, actually, the range of technological interest.

4. RESULTS AND DISCUSSION

The measured values of ϵ_s for different inclinations [$-3°$, $+0.3°$] are shown in figure 6 for i.d. = 5.00 cm. Data for the larger tube with i.d. = 9.00 cm were obtained for $\phi = -3.06°$ and $-0.59°$, and are presented in figure 7.

Two distinct effects of diameter may be observed from these measurements. First, there is a lower velocity limit, U_{Mf}, below which practically no bubbles are present in the slug, and this is shown to be dependent on pipe diameter. The asymptotic values U_{Mf}^{∞}, obtained from the available data by extrapolating the curves of ϵ_s to zero, are presented in figures 6–10. A best-fit to these data is given by

$$U_{Mf}^{\infty} = \frac{U'_{Mf}}{(C_0 - 1)} = 2.60\left[1 - 2\left(\frac{D_0}{D}\right)^2\right]\sqrt{gD}, \qquad [25]$$

where $D_0 = 2.5$ cm. There is, however, a "tail", i.e. a small fraction of bubbles present at velocities substantially below this limit (figures 6–10), which is easily explained in terms of [21]. The extrapolated limit corresponds to the first term, U_{Mf}^{∞}, whereas the "tail" is caused by the reduction due to the drift velocity, U_0, which has a positive component, even horizontally for $\mathrm{Fr} \lesssim 3$. This explains the observed tail also present in the data of Gregory *et al.* (1978) and Ferschneider (1983), see figures 8–10.

From our experiments the lower limiting velocity, U_{Mf}^{∞}, is well-represented by [25] for near-horizontal flow, as shown in figure 11. Another estimate of U_{Mf}^{∞} may be obtained from the largest stable bubble diameter, see Barnea & Brauner (1985). By equating the limiting bubble size due to turbulence with that above which bubbles attain an increasingly non-spherical shape, the following expression is derived:

$$U_{Mf}^{\infty} \approx 0.94\left\{\left[\frac{0.4\sigma}{(\rho_L - \rho_G)g}\right]^{1/2}\left(\frac{\rho_L}{\sigma}\right)^{3/5}\left(\frac{\nu_L^{0.2}}{D^{1.2}}\right)^{2/5}\right\}^{-1.12}, \qquad [26]$$

where the Blasius' friction factor has been applied.

Both expressions for U_{Mf}^{∞}, [25] and [26], have been compared with the experimental data in figure 11 and are in reasonable agreement with the data. At larger diameters, however, [26] significantly underpredicts the limiting velocity.

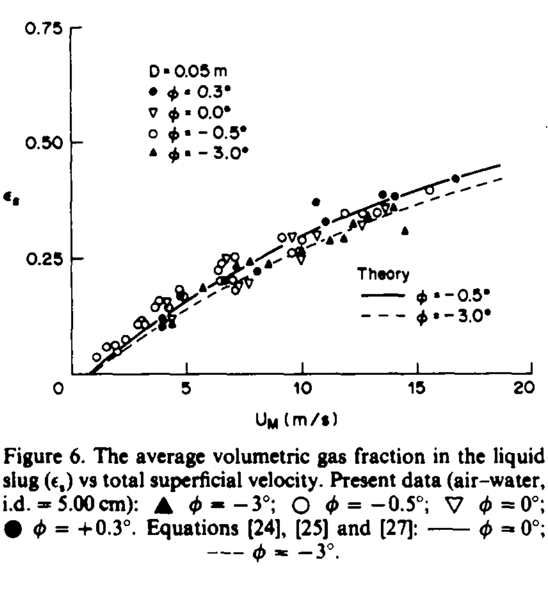

Figure 6. The average volumetric gas fraction in the liquid slug (ϵ_s) vs total superficial velocity. Present data (air–water, i.d. = 5.00 cm): ▲ $\phi = -3°$; ○ $\phi = -0.5°$; ▽ $\phi = 0°$; ● $\phi = +0.3°$. Equations [24], [25] and [27]: —— $\phi = 0°$; – – – $\phi = -3°$.

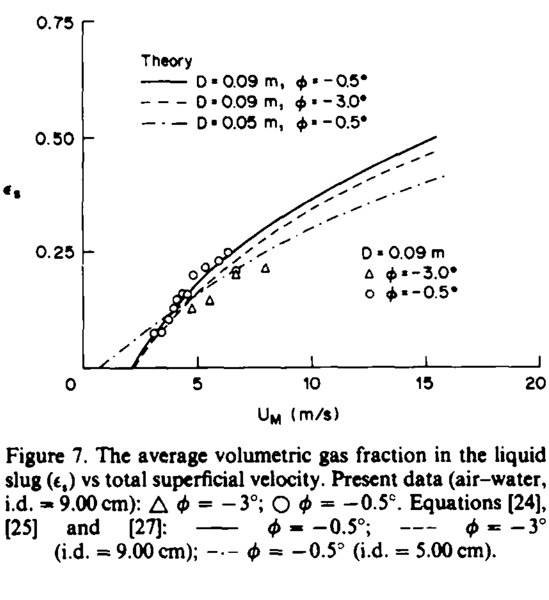

Figure 7. The average volumetric gas fraction in the liquid slug (ϵ_s) vs total superficial velocity. Present data (air–water, i.d. = 9.00 cm): △ $\phi = -3°$; ○ $\phi = -0.5°$. Equations [24], [25] and [27]: —— $\phi = -0.5°$; – – – $\phi = -3°$ (i.d. = 9.00 cm); – · – $\phi = -0.5°$ (i.d. = 5.00 cm).

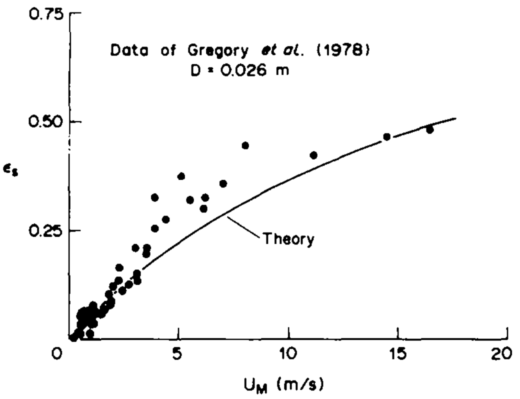

Figure 8. The average volumetric gas fraction in the liquid slug (ϵ_s) vs total superficial velocity: ● data of Gregory *et al.* (1978) (light oil–air, $\phi = 0°$, i.d. = 2.58 cm); —— [24], [25] and [27].

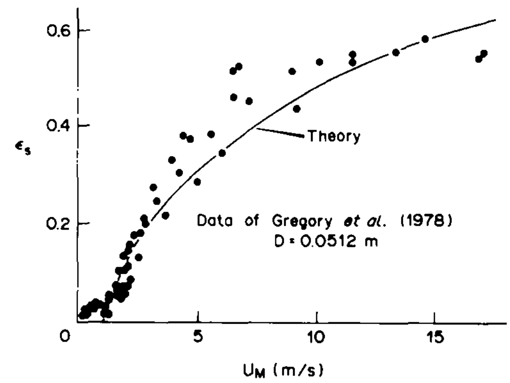

Figure 9. The average volumetric gas fraction in the liquid slug (ϵ_s) vs total superficial velocity: ● data of Gregory *et al.* (1978) (light oil–air, $\phi = 0°$, i.d. = 5.12 cm); —— [24], [25] and [27].

Secondly, a less-pronounced effect of diameter, apparent from figures 6–10, is a faster increase in ϵ_s with U_M for the larger diameter pipes. A marked difference between our data and those of Gregory *et al.* (1978) is a significantly lower void fraction at intermediate and high velocities ($U_M \lesssim 10$ m/s) for $D = 5$ cm. This is mainly thought to be a fluid properties effect (surface tension) and is explained as such through [27]. A closer investigation of the data indicates, however, that at high velocities a substantial amount of waves and pseudoslugs are present. These have been excluded in the calculation of ϵ_s in our work, which may not be the case in others, yielding higher values of ϵ_s.

Based on available data, the following empirical correlation for the remaining coefficient, U_{M0}, in [21] is proposed:

$$U_{M0} = \frac{240}{(C_0 - 1)}\sqrt{\Sigma}\,(1 - \tfrac{1}{3}\sin\phi)\left(\frac{g\sigma\Delta\rho}{\rho_L^2}\right)^{1/4} + \frac{U_0}{C_0 - 1}. \qquad [27]$$

A simplified correlation, less sensitive to the actual values of C_0, and U_0, may now be obtained by neglecting U_0 in [21] and putting $C_0 = 1.20$ for all Fr values, i.e. using [24] with U_{Mf} and U_{M0} given by [25] and [27]. This has the effect of cutting the "tail" in ϵ_s, as well as much of the angular dependency, but being identical for Fr $\gtrsim 3$, horizontally.

An effect of pipe inclination was observed, in particular for the data of Ferschneider (1983), for low mixture velocities (Fr $\lesssim 3-5$), as is to be expected from the proposed correlation, [24], [25] and [27].

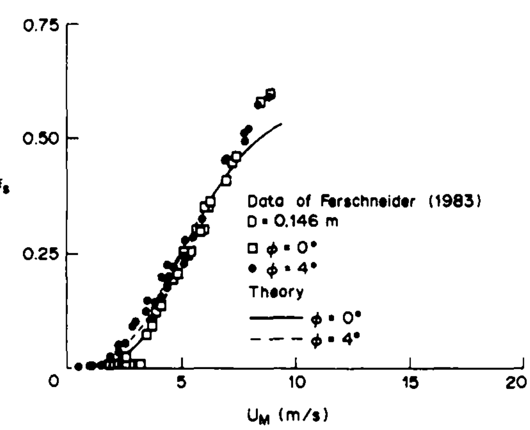

Figure 10. The average volumetric gas fraction in the liquid slug (ϵ_s) vs total superficial velocity. Data of Ferschneider (1983) (Boussens loop, gas–condensate, i.d. = 14.6 cm): □ $\phi = 0°$; ○ $\phi = +4°$. Equations [30] and [31]: —— $\phi = 0°$; – – – $\phi = +4°$.

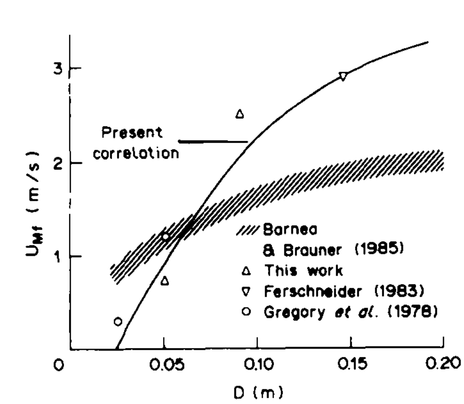

Figure 11. Extrapolated lower limiting velocity (U_{Mf}^{*}) vs pipe diameter ($\phi = 0°$): △ present data; ○ data of Gregory *et al.* (1978); ▽ data of Ferschneider (1983); —— [25]; ▨ Barnea & Brauner (1985) [26].

For down-flow a slight decrease in ϵ_s is observed for $-3°$ inclincation with respect to horizontal or up-flow (see figures 6 and 7). The reason is that the Dumitrescu bubble velocity is lower in down-flow, so that the relative slug and film velocity decreases, yielding a lower bubble production rate. For up-flow the situation is reversed, C_0 and U_0 increase (for Fr $\lesssim 3$), implying a lower U_{Mf} and a lower U_{M0}, yielding a higher ϵ_s. This is most pronounced for the large diameter pipes, e.g. the data of Ferschneider (1983), see figure 10. At higher velocities ($U_M \gtrsim 5$ m/s), there is only a weak effect of pipe inclination on ϵ_s, as is to be expected for the rather small inclinations investigated ($|\phi| \lesssim 4°$).

In our correlation for down-flow, the inclination effects apparent at $\phi = -3°$ are mainly due to a reduction in the distribution parameter, C_0, in [21] for Fr $\gtrsim 3$ (Bendiksen 1984). The predictions from the simplified correlation, applying [24], [25] and [27], are in good agreement with the data, as shown in figures 6–9.

A direct comparison of figures 6 and 7 for air–water flow, and figures 8 and 9 for air–oil flow also shows that the proposed correlation predicts the observed effects of diameter very well. This applies to the effect of fluid properties as well, which is evident from figures 6 and 9, where the predictions have been compared with data from pipes with identical diameters of 5 cm, for air–water and air–oil flow, respectively. The difference in surface tension is almost a factor of 3.

The applied coefficients in the correlation were based on available low-pressure data, so there are no density effects incorporated. For high pressures, such effects are to be expected and should be taken into account through the coefficients C_1 and C_2 in [10]. Based on the data of Ferschneider (1983), with $D = 14.6$ cm, an approximate pressure of 1.5 MPa and $\sigma = 0.028$ N/m, a rather strong dependency is observed. A preliminary extension of [24] may then be proposed:

$$\epsilon_s = \frac{U_M - U_{Mf}}{(\beta U_M + U_{M0})^m}, \quad [28]$$

where

$$m = 1 - n\frac{\rho_G}{\rho_L} \quad [29]$$

with $n = 3$.

A comparison of relation [28] with experiments relative to $\rho_G/\rho_L = 0.0188$ and $m = 0.943$ is shown in figure 10. In this figure, as the data of Ferschneider show a very low scatter and cover also the low mixture velocity region, it has been decided to represent the results obtained using the complete theory, i.e. U_{Mf} given by [21] and U_{M0} by [27]. As can be seen, the ability of the proposed correlation to also predict the observed tail in ϵ_s and the inclination effects is demonstrated.

5. CONCLUSIONS

The void fraction in liquid slugs has been determined by a newly-developed conductance probe technique. This technique allows precise and reproducible measurements of the spatial mean void fraction relative to a length of 2–3 tube diameters.

The main results of the analysis of available data (Gregory *et al.* 1978; Ferschneider 1983) and of the present observations are:

- —The effect of pipe size on void fraction in the slugs is appreciable.
- —Small variations around the horizontal position of the pipe inclination produce appreciable effects on the void fraction at low and moderate mixture velocities.
- —The void fraction in the slugs is appreciably different from zero only above a minimum value of the mixture velocity.
- —A comparison of previously published data shows that the effects of surface tension and gas density on the void fraction are strong.

A new correlation has been developed in order to fit all the available data. The proposed correlation is based on a physical approach to the slug bubble production and transport processes. The coefficients of this model have been derived empirically based on low-pressure data, and are presently not expected to describe high-pressure systems properly. The correlation, however, predicts the observed effects of pipe diameter, inclination and fluid properties on the void fraction in liquid slugs very well.

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