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Effect of diameter on two-phase pressure drop in narrow tubes

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ABSTRACT

The effect of tube diameter on two-phase frictional pressure drop was investigated in circular tubes with inner diameters of 0.6, 1.2, 1.7, 2.6 and 3.4 mm using air and water. The gas and liquid superficial velocity ranges were 0.01–50 m/s and 0.01–3 m/s, respectively. The gas and liquid flow rates were measured and the two-phase flow pattern images were recorded using high-speed CMOS camera. Unique flow patterns were observed for smaller tube diameters. Pressure drop was measured and compared with various existing models such as homogeneous model and Lockhart–Martinelli model. It appears that the dominant effect of surface tension shrinking the flow stratification in the annular regime is important. It was found that existing models are inadequate in predicting the pressure drop for all the flow regimes visualized. Based on the analysis of present experimental frictional pressure drop data a correlation is proposed for predicting Chisholm parameter "C" in slug annular flow pattern. For all other flow regimes Chisholm's original correlation appears to be adequate except the bubbly flow regime where homogeneous model works well. The modification results in overall mean deviation of pressure drop within 25% for all tube diameters considered. This approach of flow regime based modification of liquid gas interaction parameter appears to be the key to pressure drop prediction in narrow tubes.

1. Introduction

A reliable estimate of pressure drop is very important in the design of process control systems, steam power plants, petrochemical plants, refrigeration and air-conditioning systems. Small flow channels of the order of 1–2 mm are used in compact heat exchangers. These exchangers are also used in aircraft, in air separation plants, chemical process industries, microelectronic cooling systems, research nuclear reactors, space craft, chemical processing and small sized refrigeration systems. They have very high heat duties per unit volume. Thermal system design of such equipment requires an accurate knowledge of two-phase frictional pressure drop which would improve the performance of the system. Further, the flow regimes, patterns and pressure drop in narrow tubes are different from larger diameter tubes. The effects of channel diameter, roughness of the surface, injection device and fluid properties such as surface tension and viscosity on two-phase flow are very significant. In two-phase flow, the microscale size effect starts becoming prominent at a much higher diameter, of the order of 2 mm compared to ~200 μm size in single phase gas flow. This is because the fundamental length scale in single phase flow is the mean free path which is usually in microns (for gas flow) while in two-phase flow the fundamental length scale is the bubble diameter which can be of the order of millimeters. Hence, the size effects are visible here in the millimeter range of flow passages itself.

Damianides and Westwater [1] presented flow maps for horizontal glass tubes of 1–5 mm diameter using 0.015–125.3 m/s (air) and 0.0024–5.72 m/s (water) with high speed photography and fast response pressure transducers. Bubbly, annular and intermittent flow patterns were observed.

Fourar and Bories [2] conducted experiments in artificial horizontal fractures in two experimental set ups. Size 1 × 0.5 m with gap of width 1 mm between glass plates, and size 28 × 14 cm with gap of widths 0.54 mm, 0.40 mm and 0.18 mm) between two bricks made of baked clay were used. Air and water were injected separately through two separate capillary tubes. The Lockhart–Martinelli model [3] gave a good fit for both pressure drop and liquid volume fraction against the Martinelli parameter. In Lockhart–Martinelli model [3], the Martinelli parameter, X, is a combination of the inertial and viscous forces of both phases. It is one of the most dominant parameters to correlate two-phase friction pressure gradient for mini channels.

Mishima and Hibiki [4] measured void fraction, rise velocity of slug bubbles and frictional pressure loss for air–water flows in vertical capillary tubes of inner diameters in the range 1–4 mm. The frictional pressure loss was predicted by Chisholm's equation [5] with modified Chisholm parameter C as a function of inner

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Nomenclature

Bo	Bond number, $g(\rho_l - \rho_g)D^2/\sigma$
C	constant in Chisholm correlation
D	inside diameter of the tube (m)
E	variable in Friedels correlation [14]
F	variable in Friedels Correlation [14]
H	variable in Friedels Correlation [14]
f	friction factor
Fr	Froude number, $G^2/(gD\rho_m{}^2)$
G	mass flux (kg/(m^2 s)
N	total number of data points
ΔP	pressure drop (Pa)
Re_g	superficial gas Reynolds number, GD/μ_g
Re_l	superficial liquid Reynolds number, GD/μ_l
We	Weber number, $G^2D/(\sigma\rho_m)$
x	mass fraction
X	Martinelli parameter
Greek symbols	
ϕ_l^2	two-phase friction multiplier for liquid flowing alone
μ	dynamic viscosity (Ns/m^2)
ρ	density (kg/m^3)
ρ_m	mixture density, $(x/\rho_g + (1-x)/\rho_l)^{-1}$(kg/m^3)
σ	surface tension of liquid (N/m)
Subscripts	
g	gas phase only
l	liquid phase only
m	mixture

diameter. The Chisholm's parameter C [3] was correlated with the hydraulic diameter of channel as

$$C = 21[1 - \exp(-0.319D_h)]$$

Zhang et al. [6] modified the Mishima and Hibiki [4] correlation with Laplace constant. The hydraulic diameter of channel was replaced by the non-dimensional Laplace constant based upon analysis of data by Artificial Neural Network. They mention that theoretically the Laplace constant scales the wave length of the Rayleigh–Taylor instability. When the bubbles are squeezed in the mini-channel, the formation of bubbles and the bubble movement are limited by interfacial stability. The modified correlation is of the following form

$$C = 21[1 - \exp(-0.319/Lo^*D_h)]$$

They also suggested that the modification of constant value from 0.319 to 0.674 for liquid–gas flow and 0.142 for adiabatic liquid–vapor flow better predicts the data.

Triplett et al. [7] investigated the void fraction and two-phase frictional pressure drop in microchannels. Experiments were conducted using air and water mixture in transparent circular microchannels of 1.1 and 1.45 mm inner diameter and in semi triangular microchannels with hydraulic diameters of 1.09 and 1.49 mm. Gas and liquid superficial velocities were varied between 0.02 and 80 m/s and 0.02–8 m/s, respectively. A one dimensional model was used for estimation of pressure drop using various two-phase friction models. Two-phase friction factor based on homogeneous mixture assumption provided the best agreement with experimental data. For annular flow, the homogeneous model and other widely used correlations over predicted the frictional pressure drop significantly. Significant deviations are mostly associated with slug annular and annular flow patterns and slug flow at very low liquid Reynolds number. The acceleration pressure drops were significant for tests with high liquid and gas superficial velocities. Triplett et al. [7] concluded that annular flow liquid–gas interfacial momentum transfer and wall friction in microchannels may be significantly different from similar processes in larger channels.

Lee and Lee [8] proposed a correlation for parameter C for the Lockhart–Martinelli type correlation which is of the form:

$$C = A\lambda^q\psi^r Re_{LO}^s$$

where $\psi = \frac{j\mu_L}{\sigma_{LG}}$ representing the importance of viscosity and surface tension and $\lambda = \frac{\mu_L^2}{\rho_L\sigma_{LG}D}$. The coefficient A and exponent's q, r and s are determined through the data regression process.

Wang et al. [9] investigated the two-phase flow pattern and friction characteristics for an air–water system in a 3.17 mm smooth tube for a mass flux of 50–700 kg/(m^2 s). Correlations of the frictional multipliers were developed for stratified and non-stratified flow. Moriyama et al. [10] found that with decrease in tube diameter, the constant in the Chisholm correlation [5] C, decreased. They also reported that a value of $C = 0$ is more appropriate for very small tube diameters ($d < 0.1$ mm).

Wang et al. [11] compared frictional pressure drop data of refrigerants R-22, R407C and R410A in 3, 5, 7 and 9 mm diameter tubes with mass velocities ranging from 50 to 600 kg/(m^2 s). Homogeneous model and Souza and Pimenta [12] correlation gave a mean deviation of 30–35%. They modified the Chisholm correlation [5] and obtained mean deviations around 17–18%.

Chen et al. [13] developed an empirical correlation based on the homogeneous model. The empirical correlation was based on experimental two-phase pressure drop data in small diameter ($D < 10$ mm) tubes with eight refrigerant and three air–water data sets. They concluded that the Chisholm correlation [5] was not satisfactory for smaller diameter tubes. The Friedel [14] and Souza and Pimenta's [12] correlations gave fair predictions for the refrigerants but failed to predict the air–water data. Predictions by the homogeneous model gave a mean deviation of 34.7% for both refrigerant and air–water data sets.

Kawahara et al. [15] investigated experimentally the two phase characteristics in a 100 μm diameter circular tube with water and nitrogen as the two phase fluids. They observed various flow patterns. The single phase friction factor was shown to be in good agreement with the conventional laminar correlation. The two-phase friction multiplier data were over predicted by the homogeneous model, but correlated well within ±10% with the separated flow model of Lockhart–Martinelli [3].

Chung and Kawaji [16] investigated the effect of channel diameter on two-phase flow to identify the phenomena which distinguish microchannels from mini channels. Mixture of Nitrogen with water were used in circular channels of 530, 250, 100 and 50 μm diameter. A new slug flow model was proposed which predicted the two-phase frictional pressure gradient for the 100 and 50 μm channels. The two-phase pressure gradient data were compared with the predictions of the homogeneous flow model and separated flow model. Dukler et al. [17] viscosity correlation predicted the data reasonably well for 100 and 50 μm microchannels while Beattie and Whalley's [18] mixture viscosity correlation roughly predicted the two-phase pressure gradient data for 530 and 250 μm channels. For 100 and 50 μm microchannels, the two-phase friction multiplier could be predicted within ±10% by the C value correlations of Lee and Lee [19], and Mishima and Hibiki [4], while those obtained using $C = 5$ as proposed by Chisholm

[5] for conventional channels would overestimate the data. The applicable value of C decreased as the channel diameter was reduced. They mentioned that the C value for microchannels of diameter less than 50 μm would be practically zero, which corresponds to the case of a completely separated laminar flow of gas and liquid with minimal momentum coupling between the two phases.

Awad and Muzychka [20] developed expressions for obtaining bounds for two-phase frictional pressure gradient. The bound was based on the Carey correlation for turbulent–turbulent flow that uses the separate-cylinders model and the Blausius equation to represent the Fanning friction factor. The upper bound is based on separate-cylinders model for turbulent–turbulent flow that represents the Lockhart–Martinelli [3] correlation and the lower bound is based on turbulent–turbulent flow that uses the Blasius equation to represent the fanning friction factor. The mean model is based on the arithmetic mean of lower bound and upper bound.

Li and Wu [10] analyzed the experimental results of adiabatic two-phase pressure drop in micro/mini channels for both multi and single-channel configurations from collected database of 769 data points, covering 12 fluids, for a wide range of operational conditions and channel dimensions. A particular trend was observed with the Bond number (Bo) that distinguished the data in three ranges, indicating the relative importance of surface tension and they proposed a new correlation as

$$\mathrm{Bo} \leqslant 1.5, \quad C = 11.9\mathrm{Bo}^{0.45}$$

$$1.5 < \mathrm{Bo} \leqslant 11, \quad C = 109.4(\mathrm{Bo}\mathrm{Re}_l^{0.5})^{-0.56}$$

A comprehensive review of the studies of gas–liquid two-phase flow patterns and flow pattern maps at adiabatic and diabatic conditions is given in detail by Cheng et al. [22]. The studies include modeling of flow-regime transitions, specific flow patterns, stability, and interfacial shear. Recommendations for future research directions were given. The authors through their reviews concluded that compared to that in macroscale channels, the study of flow patterns in microscale channels is still in its infancy.

As Dutkowski [23] mentions the prediction of two-phase frictional pressure drop in a conventional channel itself is difficult. This is because during a two-phase flow, it is not only the volume fraction of the phases creating the two phase system that changes, but the shape of the interfacial surface (termed as flow regimes) that undergoes various levels of deformation. As observed from the above literature, there appears to be discrepancies between the various flow regimes identified and pressure drop data compared with various existing models when using smaller diameter tubes. Deviations of data with respect to flow regimes and pressure drop of small diameter tubes from the usual larger diameter tubes have been reported. The limit of diameter at which the pressure drop of smaller diameter tube does not match the conventional models (homogeneous model, separated model) is still not clear. Existing experimental works have revealed some unique phenomena in mini channels, but still there is no general theory or correlation available. Further, only limited literature is available where pressure drop models are reported on the basis of simultaneous flow regime mapping. In the present work, an attempt has been made to present a model on frictional pressure drops based on flow regimes observed with the same experimental arrangement. Comparisons have been made with conventional frictional pressure drop models, for the meso-scale of tubes between 0.6 mm and 3.4 mm diameter tubes and subsequently an appropriate correlation which shows the importance of superficial velocities and surface tension is suggested.

2. Experiments

2.1. Experimental setup

The experimental setup used in this study is designed for adiabatic co-current flow of air–water mixtures in circular horizontal tubes. A schematic diagram of the experimental setup is shown in Fig. 1.

Distilled deionised water is pumped into the test loop by an 810 lph, 0.5 HP water pump with a bypass valve from an open tank.

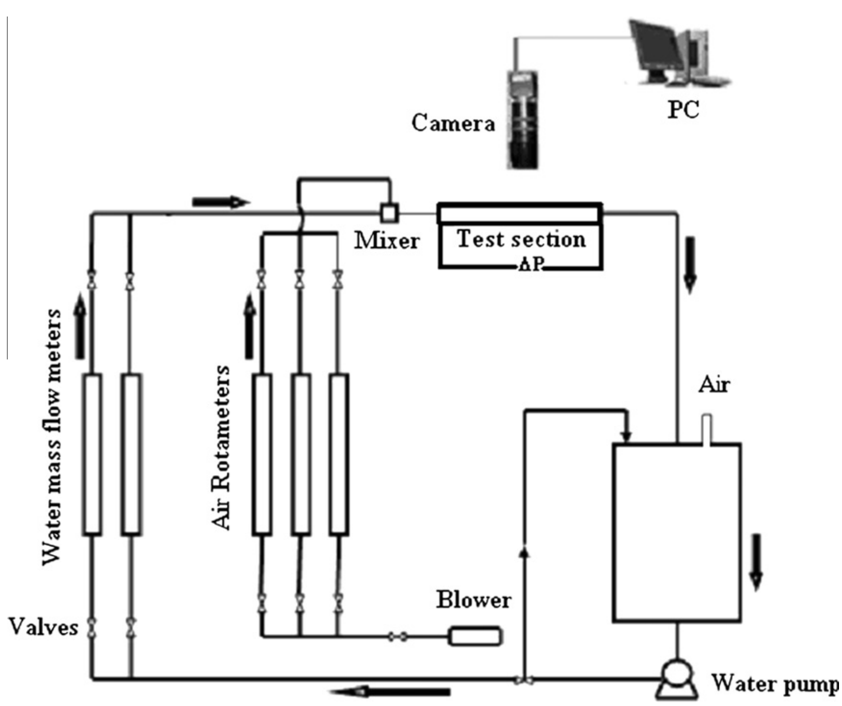

Fig. 1. Schematic diagram of experimental facility.

Atmospheric air is blown with the help of a blower. The measured temperature of water is 25 °C. Water flow rate is measured with two micromotion coriolis mass flow meters while air flow rate is measured with three Rotameters. A needle valve at the entrance for each mass flow meter and Rotameter is used for controlling the flow rate. The gas–liquid mixer installed upstream of the test section or the observation zone ensures that the two phases get thoroughly mixed before entering the test section. Details of the gas–liquid mixer and the test section are shown in Fig. 2a. Air mixes with water through four holes each of 1 mm diameter in the mixing section. The mixer section for the smallest diameter of 0.6 mm alone is different from the other tubes. For 0.6 mm tube, air is introduced in a perpendicular direction which mixes directly with water as shown in Fig. 2b. A developing length of 10 cm length is provided between the mixer and the test section (i.e., the observation zone) in order to ensure fully developed flow and to avoid mixer effects on the flow patterns and their transitions.

The test section is 10 cm long and is provided with two pressure taps at the entrance and exit that are connected to differential pressure transmitters to measure the pressure difference between test section inlet and outlet. Two transmitters are used, one for low pressure drops and the other for higher pressure drops. Rosemount (Model No. 3051s) pressure transmitter can measure pressure drops up to 10 bar while Autrol (Model No. APT3100) can measure low pressure drops of the order of millibar and up to a maximum of 1 bar. Before the experiments, the entire test loop was tested for leaks. Air pockets inside the test section and pressure taps were removed diligently before starting the experiments for each tube.

The test sections were made of silica glass with inner diameters of 0.623, 1.224, 1.732, 2.642 and 3.420 mm. The inner diameters of the capillary tubes were measured using a high speed BASLER CMOS camera (Model No. A602f supported with IEEE 1394) with a zooming lens (Navitar Zoom 18–108mmF/2.5). It has a maximum screen size resolution (pixel size) of 656 × 491 at 100 fps. This is adjustable depending upon the focusing area and frame speed. Video graphs taken using Stream Pix software was later converted into photographs and then analyzed using Windig software to determine the tube diameter. A photograph taken is shown in Fig. 3. Points were marked at 10 different places on the photograph and the diameter is determined from the average. Hereafter, the diameter of the tubes will be mentioned as 0.6, 1.2, 1.7, 2.6 and 3.4 mm only. The experimental uncertainty in air flow measurements using Rotameters are

Flow rate (lph)	Error
0–10	±0.01
11–110	±0.01
25–250	±0.1

The experimental uncertainty in water flow measurements using the micromotion Coriolis mass flow meter are

Flow rate (kg/h)	Error (%)
0–1.5 kg/h (LF series)	±1
1–2180 kg/h (CMF series)	±0.05

The experimental uncertainty in pressure drop measurements using Rosemount pressure transmitter (3051S) is ±0.025% while it is ±0.075% for Autrol Pressure Transmitter (APT 3100).

2.2. Flow visualization

The camera was mounted from the side of the test section. The test section itself is fixed on a wooden board. A floodlight of 500 W kept at a sufficient distance is used for viewing the flow patterns. Flow patterns were observed and the video graphs were recorded in a computer and subsequently analyzed frame by frame with appropriate image processing software (Stream Pix) to determine various flow regimes.

3. Pressure drop models for data reduction

Frictional pressure drop readings which were recorded in the steady state are compared with the homogeneous model, using the viscosity model of Dukler et al. [17], Chisholm modified Lockhart–Martinelli [5] model and the Friedel [14] correlation.

In the homogeneous model the flow is assumed to be one dimensional and the effective viscosity can be given by any of the following models

(a) Mc. Adams et al. [24]

$$\mu_m = \left(\frac{x}{\mu_g} + \frac{1-x}{\mu_l}\right)^{-1} \tag{1}$$

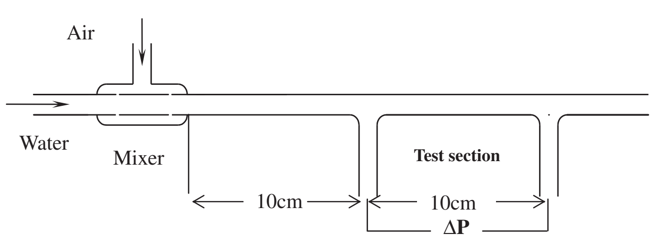

Fig. 2a. Mixer section along with test Section.

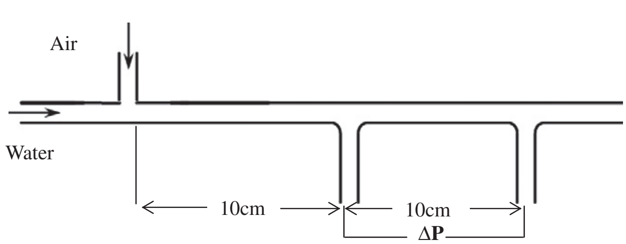

Fig. 2b. Mixer section for 0.6 mm diameter tube.

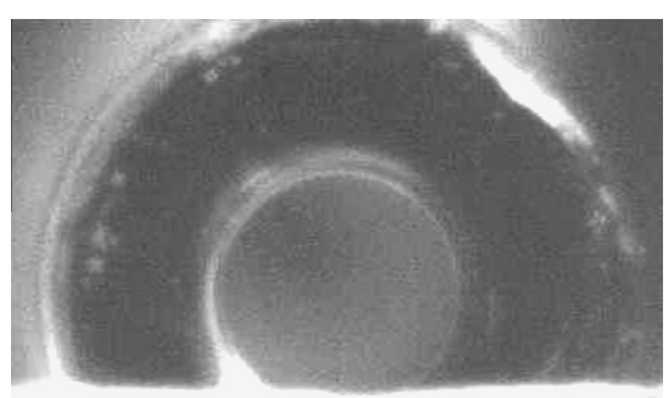

Fig. 3. Tube dia 2.6 mm.

(b) Dukler et al. [17]

$$\mu_m = \rho_m\left(x\frac{\mu_g}{\rho_g} + (1-x)\frac{\mu_l}{\rho_l}\right) \quad (2)$$

(c) Cicchitti et al. [25]

$$\mu_m = \rho_m(x\mu_g + (1-x)\mu_l) \quad (3)$$

Kawahara [15] obtained reasonably good predictions of pressure drop with Dukler et al. [17] model for the mixture viscosity. Lockhart and Martinelli [3] analyzed two-phase flow of air and liquids such as water, kerosene, benzene and various oils in pipes with diameters varying from 0.0586 in. to 1.017 in. Their data were for larger diameter tubes and were assumed to be in annular flow regime. Later Chisholm [5] modified and gave simple expressions depending on whether the phases are in laminar or turbulent flow. The Single phase liquid multiplier is given by

$$\phi_l^2 = 1 + \frac{C}{X} + \frac{1}{X^2} \quad (4)$$

where X is the Lockhart–Martinelli parameter and is given by

$$X = \sqrt{\frac{\Delta P_l}{\Delta P_g}} \quad (5)$$

Friedel [14] developed a correlation with 25,000 data points. The smallest pipe diameter used is 4 mm. The correlation included the

(a) Bubbly flow $U_g = 0.25$m/s, $U_l = 0.92$m/s D=3.4mm

(b) Stratified smooth flow $U_g = 7.85$m/s $U_l = 0.02$m/s D=2.6mm

(c) Stratified wavy flow $U_g = 7.85$m/s $U_l = 0.04$m/s D=2.6mm

(d) Wavy annular flow $U_g = 10.5$m/s $U_l = 0.42$m/s D=2.6mm

(e) Slug annular flow $U_g = 9.18$m/s $U_l = 0.49$m/s D=1.7mm

(f) Dispersed bubbly flow $U_g = 6.15$m/s $U_l = 1.23$m/s D=1.2mm

(g) Annular flow $U_g = 30.7$m/s $U_l = 0.36$m/s D=1.2mm

(g) Slug flow $U_g = 1.96$m/s $U_l = 0.68$ m/s D=0.6mm

Fig. 4. Flow patterns (flow direction is from right to left).

gravity effect through the Froude number (Fr), and the effects of surface tension and total mass flux using the Weber number (We). The correlation is of the following form

$$\begin{aligned} \phi_l^2 &= E + \frac{3.24FH}{Fr^{0.045}We^{0.035}} \\ E &= (1-x)^2 + x^2\frac{\rho_l f_l}{\rho_g f_g} \\ F &= x^{0.78}(1-x)^{0.224} \\ H &= \left(\frac{\rho_l}{\rho_g}\right)^{0.91}\left(\frac{\mu_g}{\mu_l}\right)^{0.19}\left(1-\frac{\mu_g}{\mu_g}\right)^{0.7} \end{aligned} \tag{6}$$

Experimental pressure drop obtained for various flow regimes are then compared with homogeneous model with Dukler et al. [17] viscosity model, Chisholm correlation [5] and Friedels correlation [14] and their mean deviation is evaluated as

$$\frac{1}{N}\left(\sum_{1}^{N}\left[\frac{\Delta P_{\mathrm{mod}} - \Delta P_{\mathrm{exp}}}{\Delta P_{\mathrm{exp}}}\right]\right) X100\% \tag{7}$$

where N is the number of data points. ΔP_{mod} is the pressure drop obtained from model. ΔP_{exp} is the pressure drop obtained from experiment.

4. Results and discussion

The flow patterns obtained for various tube diameters considered are shown in Fig. 4. Bubbly, slug, slug annular and dispersed bubbly flow patterns were observed in all the tube diameters considered. Wavy annular and stratified flow were observed for the tube diameter 2.6 mm. Fig. 5a distinguishes the various flow regimes obtained for various tubes considered. To have a common nomenclature for the definition of the basic flow patterns, intermittent is used to denote slug, slug annular and wavy annular flow, while dispersed includes bubbly as well as dispersed bubbly flow. Fig. 5b shows experimental flow maps based on this nomenclature for all five tubes considered. Also a comparison is made with the flow regimes of larger diameter tubes ($D = 5$ mm) observed by Damianides and Westwater [1] to indicate the deviations for the present narrow tubes. As observed by Damianides and Westwater [1], smaller tubes require much larger gas flow to change from intermittent to annular flow. However for a tube diameter of 0.6 mm, annular flow was not observed even at high gas flow rates. Much of the difference in flow patterns is observed in intermittent regime. These aspects are examined in greater detail by Venkatesan et al. [26].

For 3.4 mm tubes, the superficial velocities varied from 0.03 m/s to 1.84 m/s for liquid and 0.15 m/s to 6.12 m/s for air. Observed flow patterns were bubbly, dispersed bubbly, slug, slug annular and wavy annular flow which is similar to that of large diameter tubes. The pressure drop obtained in various flow regimes using the 3.4 mm dia tube is shown in Fig. 6.1a. Among the three models considered, Chisholm correlation [5] predicted the overall pressure drop in the present experiments with the less mean deviation. The mean deviation with the homogeneous model along with Dukler et al. [17] viscosity model predicts bubbly and dispersed bubbly regime within 10% while Chisholm correlation [5] predicts the pressure drop with 13% and 17% mean deviation, respectively for these regimes. Hereafter, homogeneous model means homogeneous model with Dukler et al. [17] viscosity model. Slug flow was predicted by the Chisholm correlation [5] with 19% mean deviation as against 43% with the homogeneous model. However, both the models result in more mean deviation in slug annular flow pattern. Much of the difference of mean deviation in the overall pressure drop was observed in the slug annular flow regime. As visually observed, in the slug annular flow, at low liquid and medium gas velocities, the gas rises in a serpentine structure. At certain combinations of superficial liquid and gas velocities, liquid slug velocity

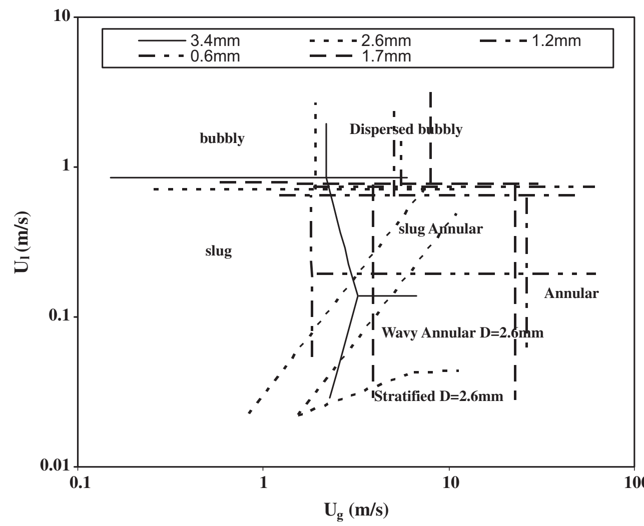

Fig. 5a. Experimental transition lines for varying diameter channels (0.6-3.4mm).

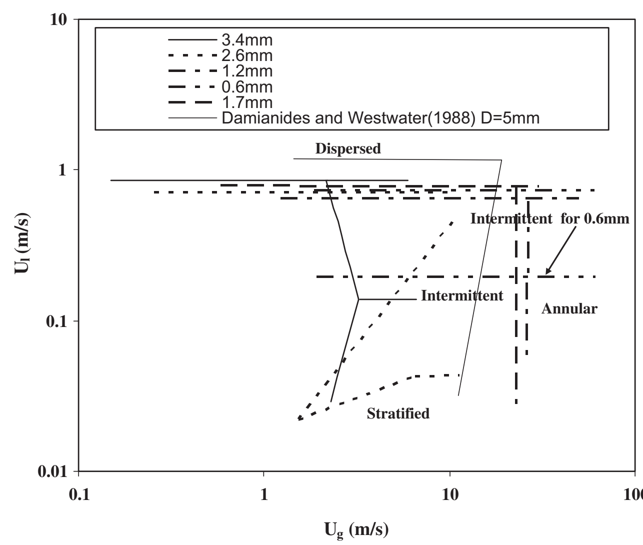

Fig. 5b. Flow map with a common nomenclature and comparison with Damianides and Westwater [1].

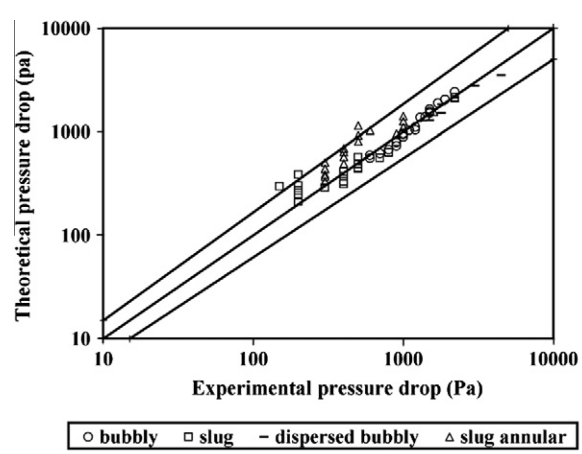

Fig. 6.1a. Experimental vs theoretical pressure drop, homogeneous model $D = 3.4$ mm.

is very low for a few moments and then suddenly accelerates after the gas slug has acquired sufficient pressure and pulls the liquid slug. There is no uniform velocity throughout the tube length during this period. Also the assumption of attainment of hydrodynamic equilibrium between the phases in the homogeneous model is less probable at high velocities. It was observed that out of 474 data points collected for all the five tubes 147 data points are of slug annular flow regime in which the mean deviation with both homogeneous and Chisholm correlation exceeded 50%. This affects the overall mean deviation and hence modifications need to be done.

In two-phase flow in macro-size channels, the capillary force is mostly negligible compared to the inertia and viscous forces. However, as the tube diameter gets smaller the capillary effect starts to play an important role in determining the behavior of two-phase flow patterns. Li and Wu [10] mentions that theoretically, there are four forces related to two-phase flow in channels: gravitational, inertia, viscous, and surface-tension forces. The comparison of the channel dimension and the nominal bubble size can be expressed in terms of the Bond number. The Bond number is a measure of the importance of body forces (almost always gravitational) compared to surface-tension forces. A high Bond number indicates that the system is relatively unaffected by surface-tension effects; a low Bond number indicates that surface tension dominates. As the channel hydraulic diameter becomes smaller, the bubbles are squeezed in the flow channel and the surface tension gradually dominates the flow. This flow pattern is not the same as that of slug flow pattern which is defined as bullet shaped. The present authors have distinguished between elongated bubble and slug flow pattern which is discussed elsewhere. Venkatesan et al. [28]. Bond number does not include the mass flux effect. The correlation suggested by Li and Wu [10] which includes Re_L for inertial effects is not included in the correlation for Bo < 1.5. The correlations of Mishima and Hibiki [4] and Zhang et al. [6] predict the Chisholm parameter C in terms of tube diameter and Laplace constant respectively. It has to be noted that even though the surface tension predominates in small diameter tubes it is the type of flow pattern that creates fluctuations in pressure drop. We have noted such fluctuations in isolated bubble and confined bubble/elongated bubble during sub-cooled boiling frictional pressure drop for mini tubes which is discussed in detail in Venkatesan et al. [27]. The type of flow pattern in turn depends on the combination of superficial liquid and superficial gas velocity in adiabatic flow. When elongated bubbles/slugs collide to have transitions to slug annular/annular flow patterns the fluctuations in frictional pressure drop will be of higher order in magnitude. So a correlation suggested for predicting pressure drop in mini channels has to be limited to the type of flow regime and it has to include the surface tension as well as inertial effects which determine the flow regime. The effect of inertia may not be said to be completely absent since majority of the liquid flow may be in laminar regime as the tube dimensions reduces while the gas phase is still in turbulent regime. The Chisholm parameter describes the level of interfacial drag or the quantitative description of interface between two phases. So our proposed correlation includes the surface tension effect by including Weber number as well as the ratio of Reynolds number of gas to liquid. The suggested correlation is of the following form.

$$\begin{aligned}&\text{Chisholm parameter } C = 4(\mathrm{We}_L)^{0.3}\left(\frac{\mathrm{Re}_G}{\mathrm{Re}_L}\right)^{0.5} \quad \text{for} \;\; \mathrm{Bo} \geqslant 1\\&\text{Chisholm parameter } C = 2(\mathrm{We}_L)^{0.5}\left(\frac{\mathrm{Re}_G}{\mathrm{Re}_L}\right)^{0.5} \quad \text{for } \mathrm{Bo} \; < 1\end{aligned} \tag{8}$$

Further, the value of parameter C becomes zero when the hydraulic diameter is as small as 0.2 mm according to Moriyama et al. [28]. The above mentioned modification in parameter C value is for slug annular regime alone. For bubbly, dispersed bubbly and annular regimes the values suggested by Chisholm [5] for C remains unchanged. With these "C" values, the mean deviation in slug annular regime is 22% as against 45% predicted by homogeneous model. Friedel correlation [14] over predicts all the data obtained in the present experiments. With "C" parameter modifications, the overall mean deviation was 18% with the Chisholm correlation [5] against 22% with the homogeneous model for all the flow regimes. All the data points are within ±50% error band line with Chisholm [5] model as shown in Fig. 6.1b.

In a 2.6 mm tube, the superficial velocity ranged from 0.01 m/s to 3.4 m/s for liquid and 0.26 to 11 m/s for air. Bubbly, dispersed bubbly, slug, slug annular, wavy annular and stratified type flow patterns were observed. Pressure drop data obtained using a 2.6 mm tube is shown in Fig. 6.2a Mean deviation in bubbly regime is well predicted by the homogeneous model (5%) as against the Chisholm correlation (14%). However, the Chisholm correlation [5] predicts pressure drop in slug and dispersed bubbly flow regimes with a mean deviation of 15% and 23% as against 26% and 44%, respectively by the homogeneous model. Our proposed correlation for modification in the parameter C predicts the pressure drop for slug annular regime as 24%. Both Stratified wavy as well as stratified smooth flow patterns were observed in the 2.6 mm dia tube. Overall mean deviation was 45% with the homogeneous model while it is 20% with our proposed correlation for slug annular regime and Chisholm correlation [5]. Friedel Correlation [14] over predicts the data in all the flow regimes. 97% of data falls within ±50% error band line with Chisholm [5] model as shown in Fig. 6.2b.

With superficial velocities ranging from 0.03 m/s to 3 m/s for liquid and 0.612 to 31 m/s for air in the 1.7 mm tube, flow patterns observed were bubbly, dispersed bubbly, slug, slug annular and annular flow. Wavy annular and stratified flow patterns were not observed. Pressure drops obtained using a 1.7 mm tube are compared with other models in Fig. 6.3a. As in 3.4 mm and 2.6 mm tubes, bubbly flow regime is well predicted by the homogeneous model with 7% mean deviation. The Chisholm correlation [5] predicts slug and dispersed bubbly flow regime pressure drop with a mean deviation of 22% and 28%, respectively. Our proposed correlation for modification of parameter C predicts slug annular flow regime with 25% mean deviation. Annular flow regime is predicted by Chisholm correlation [5] with 36% mean deviation as against 40% by the homogeneous model. Overall mean deviation is less with our proposed correlation (23%) when compared to the homogenous model (46%). Friedel correlation [14] over predicts the data. 89% of data are within ±50% error band line with Chisholm [5] model which is shown in Fig. 6.3b.

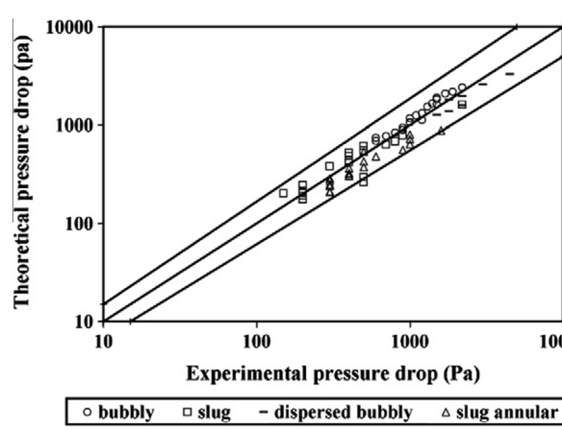

Fig. 6.1b. Experimental vs theoretical pressure drop, Chisholm model D = 3.4 mm.

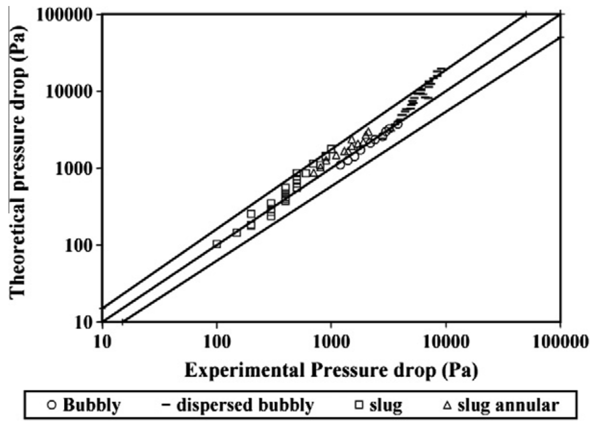

Fig. 6.2a. Experimental vs theoretical pressure drop, homogeneous model D = 2.6 mm.

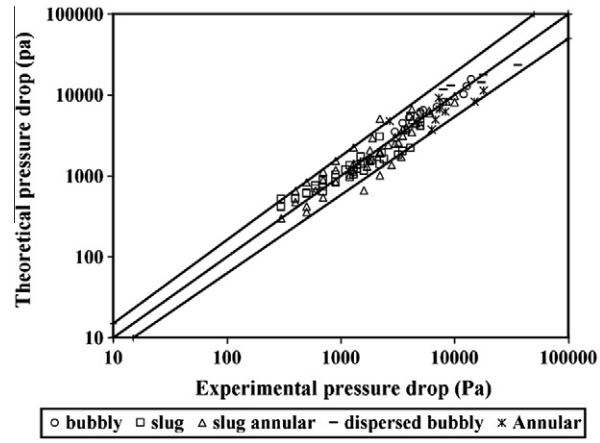

Fig. 6.3b. Experimental vs theoretical pressure drop, Chisholm model D = 1.7 mm.

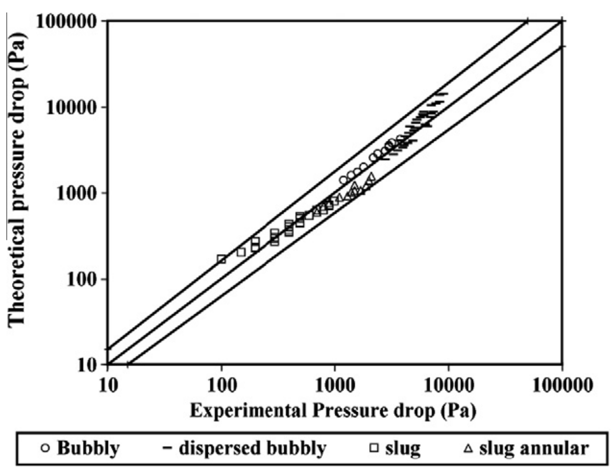

Fig. 6.2b. Experimental vs theoretical pressure drop, Chisholm model D = 2.6 mm.

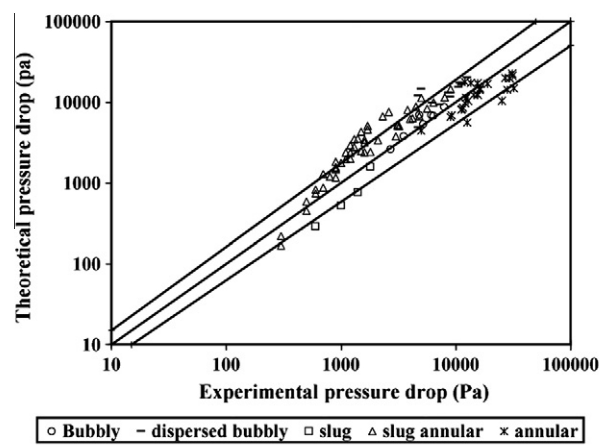

Fig. 6.4a. Experimental vs theoretical pressure drop, homogeneous model D = 1.2 mm.

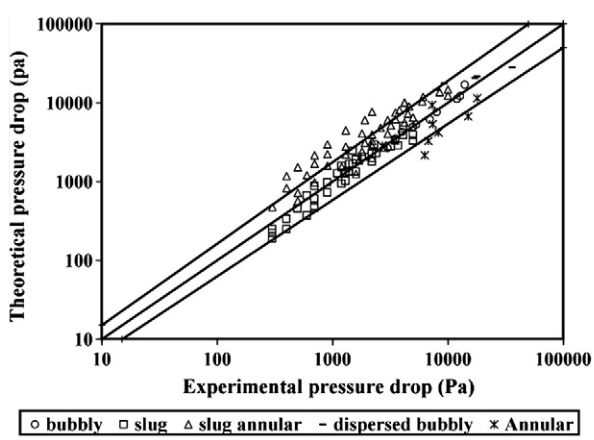

Fig. 6.3a. Experimental vs theoretical pressure drop, homogeneous model D = 1.7 mm.

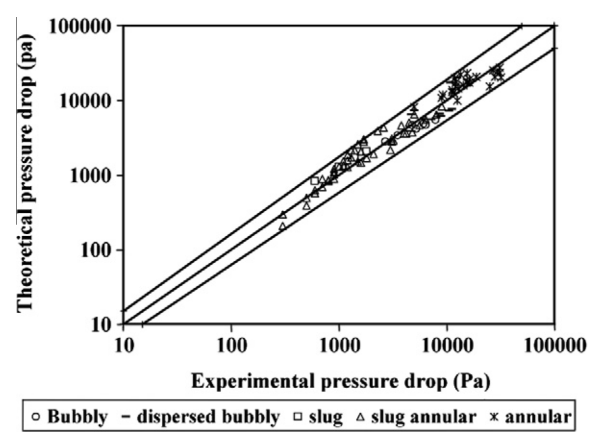

Fig. 6.4b. Experimental vs theoretical pressure drop, Chisholm model D = 1.2 mm.

The superficial velocities used in the 1.2 mm diameter tube, ranges from 0.025 to 2 m/s for liquids and 1.2 to 50 m/s for air. Flow patterns observed were bubbly, dispersed bubbly, slug, slug annular and annular type flow. As in 1.7 mm tube, wavy annular and stratified flow patterns were not observed. Pressure drop obtained using the 1.2 mm tube were compared with other models and are shown in Fig. 6.4a. The homogeneous model is better in predicting the mean deviation of the bubbly regime (13%) when compared to the Chisholm correlation [5] (15%). However slug (26%) and dispersed bubbly (34%) regimes have less mean deviation when used with the Chisholm correlation [5]. Further, slug annular flow regime with parameter "C" value modification with our correlation has a mean deviation of 25%. Annular flow regime agrees better with the Chisholm correlation [5] with 26% mean deviation. Overall mean deviation with the Chisholm correlation [5] is 25% against 67% with the homogeneous model. Similar to other tube diameters Friedel's correlation over predicts the present experimental data. 98% of data are within ±50% error band line with Chisholm [5] model as observed in Fig. 6.4b.

The superficial velocities used in a 0.6 mm diameter tube ranges from 0.16 to 2 m/s for liquids and 2 to 50 m/s for air. Four distinct flow patterns were identified. Bubbly, dispersed bubbly, slug annular and slug flow patterns were observed depending upon various combinations of superficial liquid and gas velocities. Annular flow regime was not observed in the 0.6 mm tube even at high gas velocities such as 50 m/s. According to Chung and Kawaji [16], stronger surface-tension effects in a microchannel allows the liquid film to bridge the gas core more easily than a mini channel so that the formation of annular flow would be less likely. Pressure drops obtained in a 0.6 mm tube are compared with other models in Fig. 6.5a. It can be said based on individual phase Reynolds number that almost the entire flow is in laminar regime. However, the two phase Reynolds number indicates that the two-phase flow is in the turbulent regime. This result in more mean deviation in all the flow regimes considered. Except bubbly regime, mean deviation with all three pressure drop models is more than 50% and the overall mean deviation exceeds 100%. A value of $C = 1$ over the entire regime and applying the proposed correlation for slug annular regime following results are obtained. Mean deviation in bubbly, slug, dispersed bubbly and slug annular regimes are 24%, 26%, 28% and 24%, respectively with overall mean deviation of 25% with Chisholm correlation [5]. Ninety-two percentage of data are within ±50% error band line with Chisholm [5] model as shown in Fig. 6.5b.

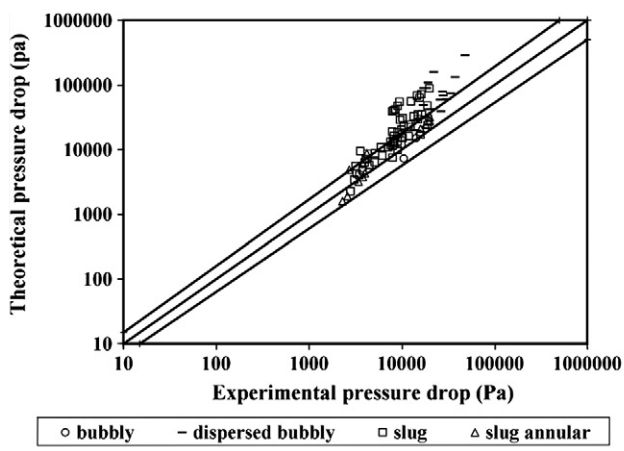

Fig. 6.5a. Experimental vs theoretical pressure drop, homogeneous model $D = 0.6$ mm.

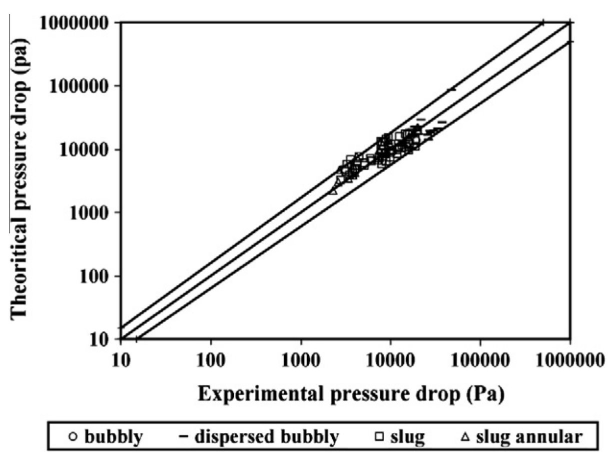

Fig. 6.5b. Experimental vs theoretical pressure drop, Chisholm model $D = 0.6$ mm.

In the above discussion, pressure drop data obtained for tube diameters 3.4 mm, 2.6 mm, 1.7 mm, 1.2 mm and 0.6 mm are compared with the homogeneous model with Dukler et al. viscosity model [17], Chisholm modified Lockhart–Martinelli model [5] with a newly proposed correlation for slug annular regime and Friedel's correlation [14]. Except in the bubbly regime, the Chisholm correlation [5] predicts the frictional pressure drop in other regimes with least mean deviation. A correlation for modification in the Chisholm parameter C as per (Eq. (8)) is suggested in this study, which results in better agreement with experimental data of frictional pressure drop in slug annular regime. A value of $C = 1$ for all the flow regimes and proposed correlation for slug annular regimes predicts the experimental pressure drop with less mean deviation.

5. Analysis of the correlation

Fig. 7 shows the effect of liquid Weber number (We_L) ratio on Chisholm's parameter C. Larger liquid Weber number results in increase in value of parameter C while lower value of C appears to be at lower liquid Weber number. Fig. 8 shows the effect of Re_G/Re_L ratio on parameter C. For larger tube diameters a higher

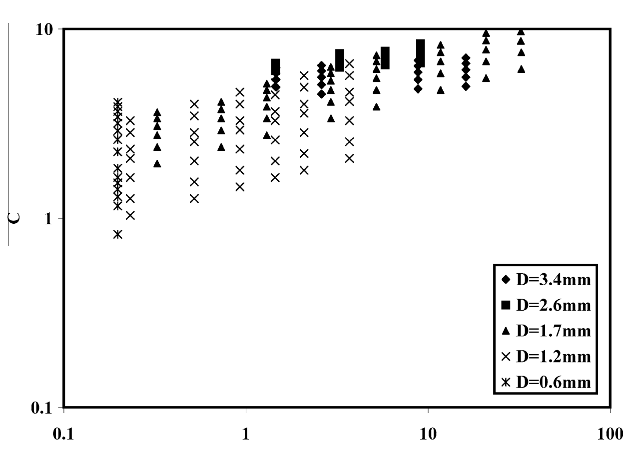

Fig. 7. Effect of liquid phase Weber number on parameter 'C'.

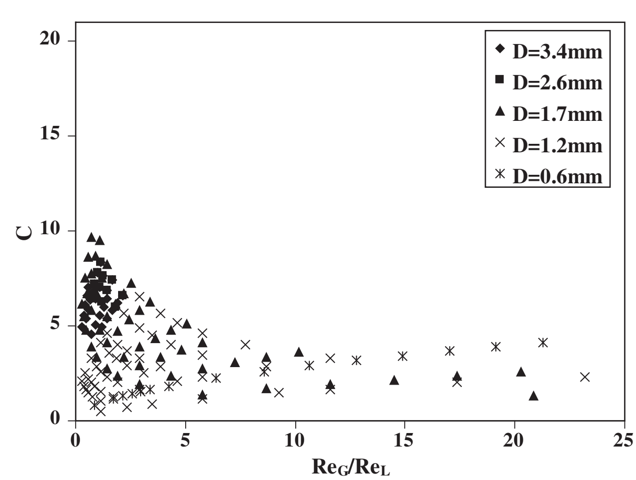

Fig. 8. Effect of Re_G/Re_L ratio on Chisholm's parameter.

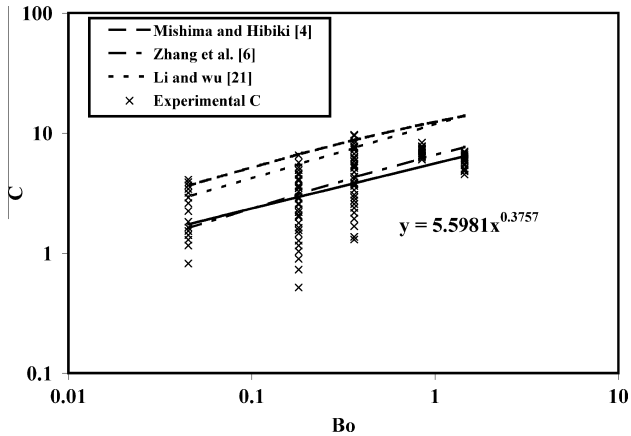

Fig. 9. Comparison of correlation's with the proposed correlation.

C value is expected which is shown and for lower tube diameters with the proposed correlation "C" values of up to 0.8 is obtained. Fig. 9 shows the parameter C value plotted against the Bond number for proposed correlation and the results are compared with the correlations of Mishima and Hibiki [4], Zhang et al. [6] and Li and Wu [21]. Due to variations in ratio of Re_G/Re_L the parameter value C varies over a range for each tube. These data values are for slug annular regime only since the mean deviation obtained in slug annular regime was much higher than all other regimes when compared with homogeneous and Chisholm model. Thus a flow regime dependent correlation which includes surface-tension effect, mass flux effect and superficial velocities is proposed for mini channels and its applicability for other fluids need to tested.

6. Conclusions

Two-phase flow patterns inside circular narrow tubes were experimentally studied using air–water mixtures. The different tube diameters used were 0.6, 1.2, 1.7, 2.6 and 3.4 mm. Superficial gas and liquid velocities ranged from 0.01 to 50 m/s and 0.01 to 3 m/s, respectively. The two-phase flow was visualized through a high-speed CMOS camera. Pressure drop was measured and compared with various models available in the literature. The major conclusions of the study can summarized as

- Bubbly flow regime is predicted well by the homogeneous model along with Dukler et al. [17] viscosity model. All other regimes deviate from this model considerably.
- Based upon flow visualization using a high speed camera, modification in Chisholm parameter C has been suggested by a correlation which includes includes surface-tension effect, mass flux effect and superficial velocities applicable for mini tubes. The correlation is applicable for slug annular regime which gives a better estimate of the frictional pressure drop when compared to the experimental data. The proposed correlation is

$$\text{The Chisholm parameter } C = 4(\mathrm{We}_L)^{0.3}\left(\frac{\mathrm{Re}_G}{\mathrm{Re}_L}\right)^{0.5} \quad \text{for Bo} > 1$$

$$\text{Chisholm parameter } \quad C = 2(\mathrm{We}_L)^{0.5}\left(\frac{\mathrm{Re}_G}{\mathrm{Re}_L}\right)^{0.5} \quad \text{for Bo} < 1$$

For bubbly, dispersed bubbly and annular flow regime, 'C' value as in the Chisholm correlation [5] predicts the frictional pressure drop well. Overall mean deviation obtained was less than 25% for 3.4, 2.6, 1.7 and 1.2 mm tubes.

- For 0.6 mm diameter tube, a value of C = 1 for all other regimes except slug annular regime where the proposed correlation is applied results in a mean deviation of 25%, which is lower than the other models.
- Friedel's correlation [14] over predicts the entire data for all tube diameters with mean deviation greater than 100%.

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