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# Gas-Liquid Two Phase Flow Phenomenon in Near Horizontal Upward and Downward Inclined Pipe Orientations

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***Abstract*—**The main purpose of this work is to experimentally investigate the effect of pipe orientation on two phase flow phenomenon. Flow pattern, void fraction and two phase pressure drop is measured in a polycarbonate pipe with an inside diameter of 12.7mm for inclination angles ranging from -20° to +20° using air-water fluid combination. The experimental data covers all flow patterns and the entire range of void fraction typically observed in two phase flow. The effect of pipe orientation on void fraction and two phase pressure drop is justified with reference to the change in flow structure and two phase flow behavior. In addition to this, the top performing void fraction and two phase pressure drop correlations available in the literature are presented and their performance is assessed against the experimental data in the present study and that available in the literature.

***Keywords*—**Flow patterns, inclined two phase flow, pressure drop, void fraction.

## I. Introduction

GAS - liquid two phase flow finds its extensive application in the field of chemical, petroleum, nuclear and refrigeration industries. For any of these industrial applications, from design point of view, the correct knowledge of flow patterns, void fraction, two phase pressure drop and two phase heat transfer coefficient is vital. In particular, the better understanding of two phase flow phenomenon in inclined systems is of utmost importance since the pipe orientation is known to substantially influence the interaction between buoyancy, gravity, inertia and surface tension forces that leads to a change in physical structure of the flow patterns and eventually alters the magnitudes of void fraction and pressure drop. The total pressure drop that occurs in the system is usually of concern for design and sizing purpose which in turn depends strongly on the correct estimation of hydrostatic two phase pressure drop that requires understanding of the flow pattern and accurate estimation of the void fraction at any given pipe orientation, pipe diameter and flow rates of individual phases. The knowledge of flow patterns is necessary since the frictional pressure drop in two phase flow is sensitive to the distribution of gas and liquid phase across the pipe cross section and along the pipe length. Both the flow patterns and void fraction are observed to change with change in pipe orientation and thus results in significantly different pressure drop even measured at similar phase mass flow rates but at different pipe orientations.

So far, majority of the research done in the field of two phase flow is for vertical and horizontal pipe orientations while little attention is paid to the two phase flow phenomenon in inclined pipe orientations. Some of the comprehensive experimental investigation of two phase flow phenomenon in inclined systems available in the literature is that of [1]-[3]. The effect of pipe orientation on the void fraction and pressure drop in non-boiling ($D$ = 25mm and 38mm) flow of air-water was analyzed by [1] and [2], and for condensing flow of R134a ($D$ = 8.5mm) by [3]. Some other studies that were focused on the inclined two phase flow phenomenon for particular flow patterns and non-circular pipe geometries include the works of [4]-[8]. References [4] and [5] studied the void fraction and two phase pressure drop in near horizontal inclined systems for slug flow using air-water and air-oil fluid combinations, respectively. Reference [6] analyzed the effect of upward pipe orientations on flow patterns, void fraction and pressure drop in a 4.5mm narrow annulus while [7] studied the void fraction in bubbly and slug flow in near vertical downward inclined pipe orientations using air-water two phase flow. Reference [8] investigated the flow pattern transitions, void fraction and pressure drop in a 50.8mm I.D. pipe inclined at +5°, -5° and 0°. They found that compared to the pressure drop, void fraction is more prominently affected by the pipe orientation. References [9] and [10] studied the effect of pipe orientation on the transition boundaries of the flow patterns and proposed a generic two phase flow pattern map applicable for upward inclined and upward and downward inclined two phase flows, respectively.

Overall, these studies reported a significant effect of pipe orientation on the physical structure of flow patterns, void fraction and two phase pressure drop and concluded that consideration of pipe orientation while analyzing two phase flow is indispensable. To further investigate and identify the two phase flow regimes and the quantitative range of two phase flow parameters that is sensitive to the pipe orientation, the present study is focused on the experimental observations of the flow patterns and measurement of void fraction and two phase pressure drop in near horizontal upward and downward pipe inclinations. The experimental measurements are carried out over a range of gas and liquid flow rates and pipe orientations in a range of $-20^o \leq \theta \leq 20^o$ that encompass all

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significant flow patterns and the entire range of void fraction of practical relevance to two phase flow applications. Although, there are several void fraction correlations available in the literature, very few of these correlations are equipped to account for the effect of change in flow patterns and pipe orientation on the void fraction. Some of the top performing correlations available in the literature are considered in this study and their performance is gauged against the experimentally measured void fraction at different pipe orientations.

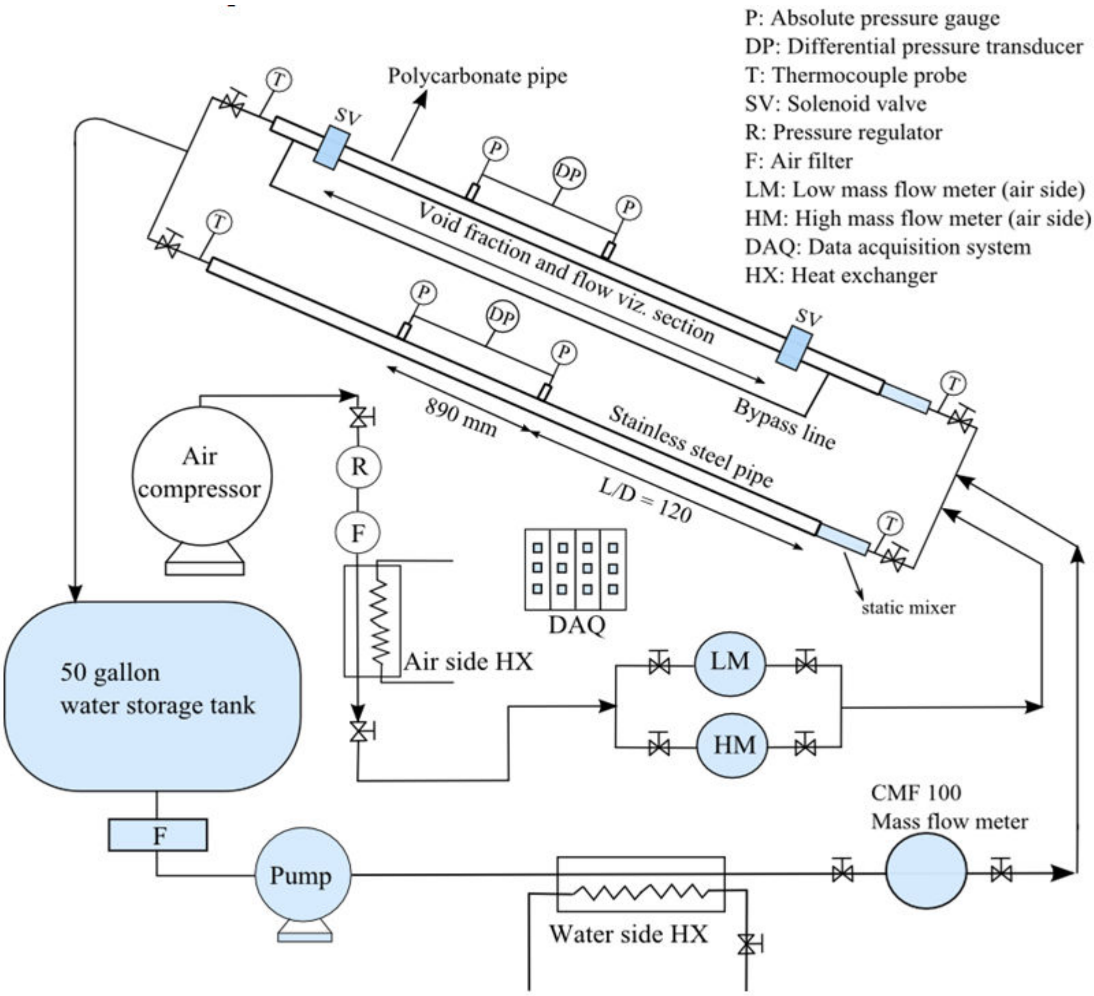



Fig. 1 Schematic of experimental setup

Additionally, a newly developed versatile correlation by [11] is presented that correctly accounts for the effect of pipe orientation on void fraction and performs better than the existing correlations for wide range of two phase flow situations. Furthermore, the performance of two phase pressure drop correlations based on the concept of homogeneous flow model is assessed against the experimental data collected in the present study.

## II. Experimental Setup

The experimental setup used for flow visualization and measurement of void fraction and pressure drop consists of a 12.7mm I.D. test section made of polycarbonate pipe and is mounted on a variable inclination frame that can be oriented from -90° to +90° including horizontal pipe orientation (see Fig. 1). The fluid combination used for generating two phase flow is distilled water and compressed air. The water stored in a 55 Gallon tank is circulated by a Bell and Gosset (series 1535, model number 3445 D10) and filtered through an Aqua-Pure AP12-T purifier before it is passed through a ITT model BCF 4063 shell and tube heat exchanger. The water then flows through Emerson (Micro Motion Elite Series model number CMF 100) Coriolis mass flow meter and then allowed to mix with air in a static mixer. The water mass flow rate is controlled by a gate valve placed after the water mass flow meter. The gas phase is the compressed air supplied by Ingersoll Rand T-30 Model 2545. Compressed air is passed through a filter and regulator circuit before it is fetched to the mass flow meter. The air mass flow rate is controlled using a Parker needle valve (Model 6A-NLL-NE-SS-V) and then allowed to mix with water in the mixer. The transparent test section made of polycarbonate material allows for flow visualization and the void fraction measurement is carried out using pneumatically operated quick closing valves. The void fraction is measured by weighing the mass of liquid trapped in the test section ($m_{meas}$) and from the total mass ($m_{tot}$) of liquid water that would accumulate in the test section of known volume. The void fraction ($\alpha$) is thus determined from (1).

$$\alpha = 1 - \frac{m_{meas}}{m_{tot}} \tag{1}$$

The two phase pressure drop between the pressure taps 0.89 m apart is measured using the Valydine DP 15 series pressure transducer (±0.25% FS accuracy) and DP 26 and DP 32 pressure diaphragms. The pressure transducer is calibrated every time the diaphragm is changed in addition to the occasional calibration to ensure repeatability. The differential pressure drop data is collected for 7-10 minutes (approx. 8000 samples) to nullify any errors in the reading due to flow fluctuation. The average value of these samples over the measured period of time is considered to be a representative of the measured values. Throughout the experiments the system temperature is maintained between 21°C and 24°C.

The total pressure drop per unit pipe length ($\Delta P/L$) in two phase flow consists of hydrostatic, accelerational and frictional components represented as shown in (2).

$$\left(\frac{\Delta P}{L}\right)_t = \left(\frac{\Delta P}{L}\right)_h + \left(\frac{\Delta P}{L}\right)_a + \left(\frac{\Delta P}{L}\right)_f \tag{2}$$

In (2), the total pressure drop is measured experimentally by measuring the pressure drop by Valydine pressure transducer and then correcting it by the pressure drop due to the amount of liquid water trapped in the connecting lines. Hydrostatic component of the two phase pressure drop is determined as a function of two phase mixture density ($\rho_M$) which in turn is calculated using the measured values of void fraction (α) and gas ($\rho_G$) and liquid ($\rho_L$) phase density in kg/m$^3$ as shown in (3). The pipe orientation ($\theta$) is measured from horizontal and '$g$' is the acceleration due to gravity with a value of 9.81 m/s$^2$.

$$\left(\frac{\Delta P}{L}\right)_h = \rho_M g \sin\theta = (\alpha\rho_G + (1-\alpha)\rho_L) g \sin\theta \tag{3}$$

The accelerational component of the two phase pressure drop is usually negligible for two phase flows (typically up to 1% of the measured pressure drop) and hence can be neglected. The frictional component of two phase pressure drop is thus calculated based on the measured total pressure drop and the hydrostatic pressure drop as shown in (4).

$$\left(\frac{\Delta P}{L}\right)_f = \left(\frac{\Delta P}{L}\right)_t - \rho_M g \sin\theta \tag{4}$$

The uncertainty in void fraction and pressure drop is carried out using the method of [12]. In a worst case scenario, the maximum uncertainty for void fraction and pressure drop measurements is found to be 14% and 16%, respectively. These high values of uncertainty correspond to very low values of void fraction and two phase pressure drop.

## III. Results and Discussion

### *A. Flow Patterns and Flow Pattern Maps*

The flow patterns in gas-liquid two phase flow are generated due to significantly different physical properties of gas and liquid phase, compressibility nature of the gas phase and their alignment with respect to each other across the pipe cross section. The knowledge of flow patterns and flow pattern maps proves instrumental in the general understanding of the physical structure and the mechanism of momentum and heat transfer in two phase flow. The flow pattern map serves as a tool to estimate the sequence of the appearance of different flow patterns with change in the gas and liquid flow rates for a given set of flow conditions. The definitions of flow patterns and their transitions are highly qualitative in nature and are mostly based on the individual's perception. The different two phase flow patterns observed in the present study are generated by varying the gas mass flow rates ($\dot{m}_G$) and keeping the liquid mass flow rates ($\dot{m}_L$) constant. The gas and liquid mass flow rates (or superficial gas ($Re_{SG}$) and liquid ($Re_{SL}$) Reynolds numbers) are varied from 0.001 to 0.2 kg/min ($140 \leq Re_{SG} \leq 19{,}000$) and 1 to 12 kg/min ($1800 \leq Re_{SL} \leq 19{,}000$) to generate different flow patterns such as bubbly, slug, intermittent, stratified and annular flow regimes (Fig. 2). Bubbly flow regime is characterized by the dispersion of small bubbles in the continuous liquid medium while the slug flow is identified as a flow structure consisting of elongated gas slugs that flow alternate to a liquid plug. The stratified flow is featured by the flow of liquid film parallel to the gas phase.

The annular flow is observed in form of liquid film flowing in contact with the pipe wall that surrounds a fast moving gas core. The definition of intermittent flow is vague since there is no particular way in which gas and liquid phases are aligned across the pipe cross section. In the present study, the intermittent flow pattern is identified based on the pulsating and chaotic nature of the two phase flow. Thus the flow structure tagged as intermittent flow in the present study consists of slug-wavy, stratified-wavy and annular-wavy flow patterns. A close observation of the bubbly flow structure as shown in Fig. 2 reveals that in downward inclinations, the gas phase is sheared by liquid phase that results into small size gas bubbles in comparison to the large elongated shaped bubbles in upward inclined flows.

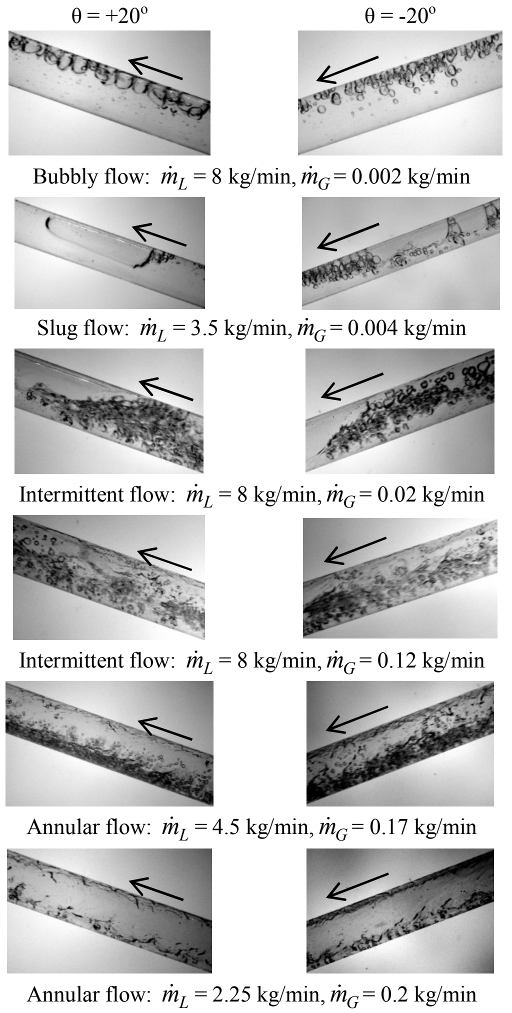



Fig. 2 Flow patterns in upward (+20°) and downward (-20°) inclined two phase flow

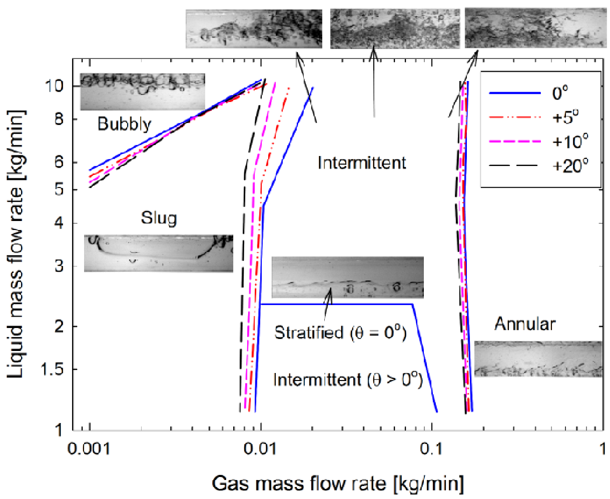



Fig. 3 Flow pattern map for upward inclined flow

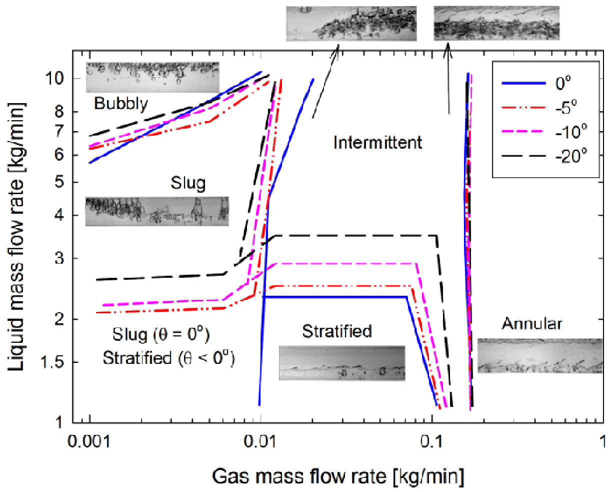


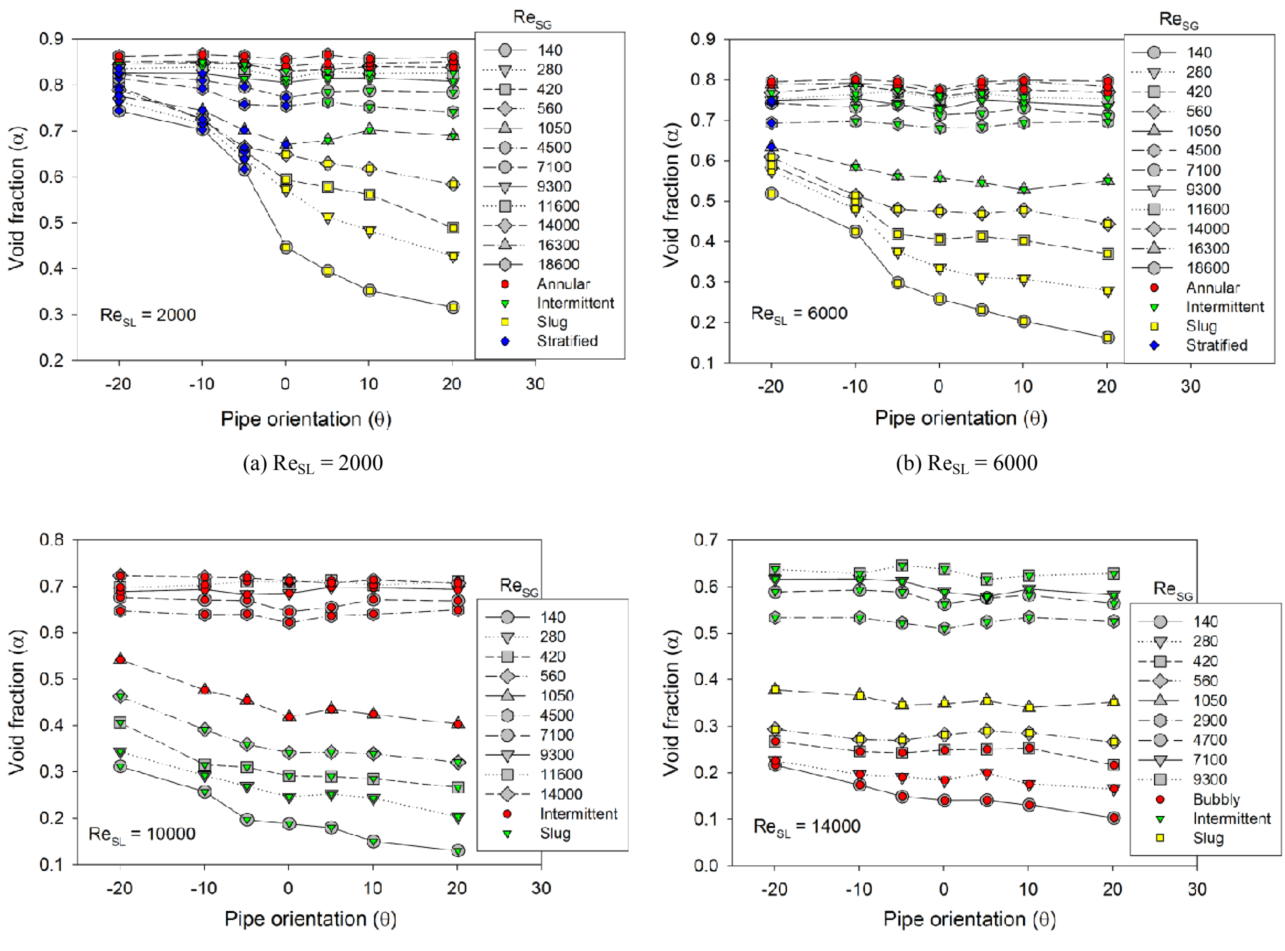

Fig. 4 Flow pattern map for downward inclined flow

In case of slug flow, significant effect of upward and downward pipe inclinations is observed in the appearance of the slug nose. In upward inclined flow, the gas slug appears to be continuous, smooth and elongated with the slug nose pointing in the same direction as that of the mean flow. Whereas, for downward inclined flow, the slug surface appears wavy with lot of smaller gas bubbles in the vicinity of the slug head and the slug nose is either flat or is pointing in a directing opposite to that of the mean flow. These dissimilarities between the bubbly and slug flow structure is found to significantly influence the void fraction to be discussed later in the next section.

It should be noted that the flow pattern maps presented in this section are exclusively developed for two phase flow of air-water through a 12.7mm I.D. polycarbonate pipe at different orientations and are not generic. However, the objective of plotting the flow pattern map is to get an understanding of the effect of pipe orientation on the transition between different flow patterns.

As illustrated in Fig. 3, for low liquid flow rates, slug flow exists at low values of gas flow rates and with increasing gas flow rates flow pattern transits to intermittent and annular flow regimes. For high liquid flow rates, bubbly, slug and intermittent flow patterns exist for low, moderate and high gas flow rates, respectively. The effect of pipe orientation on the transition boundaries with reference to the horizontal pipe orientation for different flow patterns is evident in Fig. 3. It is seen that the increase in the pipe orientation from horizontal shifts the transition between bubbly and slug flow to lower liquid flow rates and the transition between slug and intermittent flow pattern to lower gas flow rates. This early transition between the slug and intermittent flow patterns with increasing pipe orientation is possibly due to the increased buoyancy effects acting on the gas phase that aids the existence of the slug flow. It should be mentioned that the physical structure of the slug flow is significantly influenced by the phase flow rates. For low liquid and low gas flow rates, the slug flow appears in form of long and uneven length slugs moving quiescently through liquid phase at an irregular interval. With increase in the liquid and gas flow rates, the gas slug length reduces and the fast moving slugs are observed that move with a consistent frequency. There is a little effect of pipe orientation on the transition between the intermittent and annular flow, however, it is clearly seen that the increase in pipe orientation promotes early transition between intermittent and annular flow regimes. Similar observations are reported by [6], [9], [13]. It should also be noted that the stratified flow pattern exists only for horizontal flow and appears in form of intermittent flow for higher pipe inclinations ($\theta > 0^o$).

In case of downward inclined flow, the flow pattern map with reference to the horizontal pipe orientation is depicted in Fig. 4. Similar to upward inclined pipe orientations, the bubbly, slug and intermittent flow patterns appear at moderate to high liquid flow rates and for low, moderate and high gas flow rates, respectively. However, unlike upward inclined flows, stratified flow pattern exists at all downward inclined pipe orientations considered in this study. For horizontal pipe orientation, the gas phase requires higher driving potential (kinetic energy) to move downstream and carry the liquid layer parallel to it and hence the stratified flow does not occur for very low gas flow rates.




(a) $Re_{SL} = 2000$

(b) $Re_{SL} = 6000$

(c) $Re_{SL} = 10000$

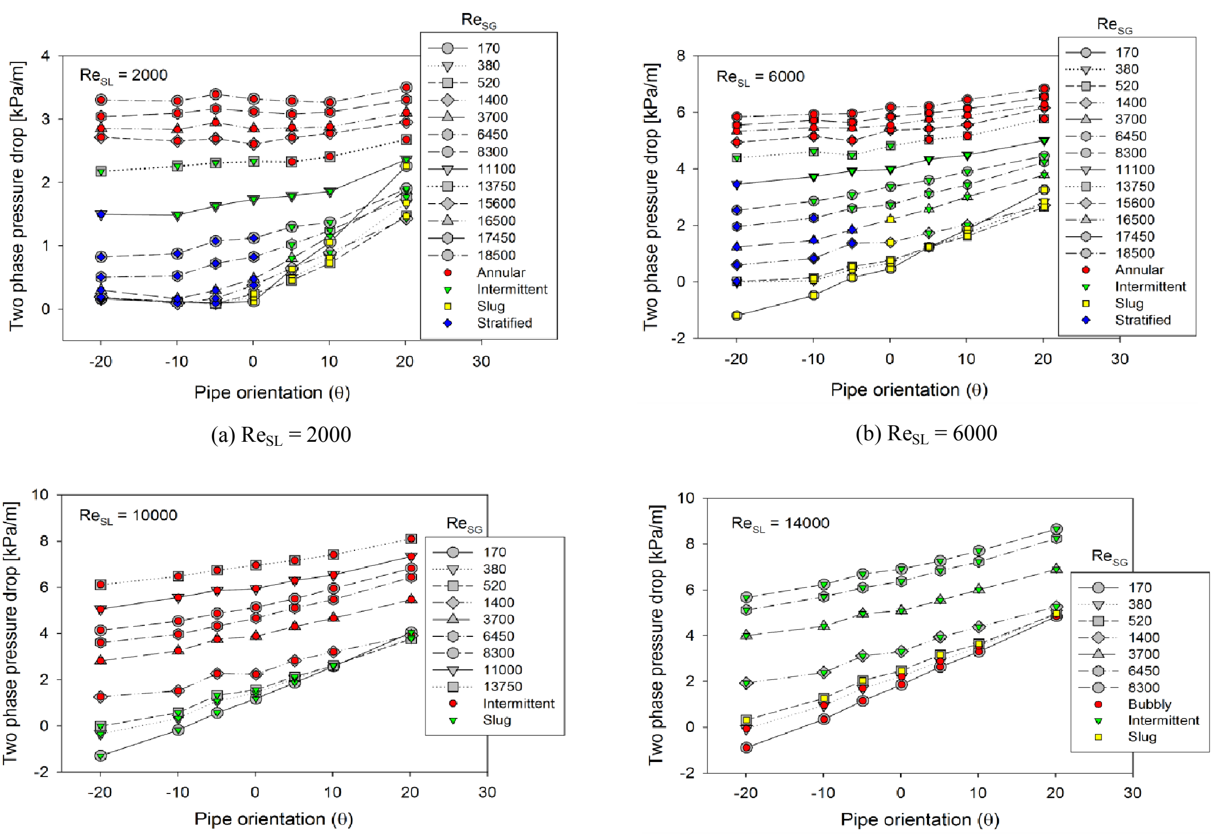
(d) $Re_{SL} = 14000$

Fig. 5 Effect of pipe orientation on void fraction for varying liquid and gas flow rates

However, the downward inclinations provide a flow potential to liquid phase in form of gravity that facilities flow stratification. It is seen that the increase in downward inclination significantly alters the transition boundary between stratified-slug, stratified-intermittent and intermittent-annular flow regimes. Increase in the downward pipe inclination with respect to horizontal is observed to delay the transition between stratified-slug and stratified-intermittent flow while expedites the transition between intermittent and annular flow regimes. Compared to the upward inclined pipe orientations, the transition boundary between intermittent and annular flow regime is negligibly affected by the increase in the downward inclinations.

### B. Void Fraction

As mentioned earlier, the correct knowledge of void fraction at various pipe orientations is crucial in calculation of two phase mixture density and hence the two phase hydrostatic pressure drop. In the present study, void fraction is measured using quick closing valves for -20$^o$ ≤ $\theta$ ≤ +20$^o$ and different flow patterns comprising a range of void fraction from 0.01 ≤ $\alpha$ ≤ 0.92. It is found that the void fraction is very sensitive to the low values of gas flow rate and increases rapidly with a small increase in the gas flow rate; whereas, for high values of gas flow rates, the void fraction is relatively insensitive and increases very gradually with increase in the gas flow rates. This relation between void fraction and gas flow rate is found to be independent of the pipe orientation. However, the effect of pipe orientation on the magnitude of the void fraction measured at fixed liquid and gas flow rates is significant. As shown in Fig. 5, the effect of pipe orientation on the void fraction is significant for the lowest liquid ($Re_{SL}$) and gas ($Re_{SG}$) superficial Reynolds numbers (or alternatively liquid and gas flow rates). For $Re_{SL}$ = 2000 and $Re_{SG}$ = 170, increase in pipe orientation in upward and downward direction with reference to horizontal causes the void fraction to decrease and increase by about 29% and 66%, respectively. At these low flow rates, a change in pipe orientation from -20$^o$ to +20$^o$ causes the void fraction to decrease from 0.743 to 0.316. Qualitatively similar results are reported by [1] and [3]. It should be noted that for low $Re_{SL}$ and $Re_{SG}$, the flow pattern in downward inclinations is stratified while in upward pipe inclinations the flow pattern is slug. Based on the flow physics and flow structure of these two flow patterns, it can be concluded that the stratification in downward inclinations result into higher residence time of the gas phase in the test section resulting into higher values of void fraction.




(a) $Re_{SL}$ = 2000
(b) $Re_{SL}$ = 6000
(c) $Re_{SL}$ = 10000
(d) $Re_{SL}$ = 14000

Fig. 6 Effect of pipe orientation on total two phase pressure drop for varying gas and liquid flow rates

Whereas, the buoyancy force acting on the slugs in upward inclinations reduces residence time of the gas phase in the test section and hence results into lower values of void fraction. At $Re_{SL}$ = 6000, even if the slug flow regime exists in both upward and downward inclinations, the significant change of the void fraction with respect to the pipe orientation can be attributed to the physical structure of the slug flow. In downward inclinations, the nose of the slug is observed to be pointing in the upstream direction indicating that the buoyancy is acting in a direction opposite of the mean flow and hence slower translational velocity of the gas slug that result into higher void fraction. In case of upward inclined flow, the nose of the slug is always pointing in the direction of flow (downstream) indicating higher slug translational velocities and hence lower values of the void fraction. With increase in $Re_{SL}$ the effect of pipe orientation on the void fraction at low $Re_{SG}$ gradually diminishes.

Nevertheless, even at $Re_{SL}$ = 14,000 and $Re_{SG}$ = 140, the change in pipe orientation from -20$^o$ to +20$^o$ causes the void fraction to decrease from 0.216 to 0.102 (≈53% decrease). It is also seen that at all liquid superficial Reynolds numbers, the void fraction at high $Re_{SG}$ corresponding to that in annular flow regime is independent of the of pipe orientation. It is interesting to see from Fig. 5 that for $Re_{SL}$ = 2000, the void fraction follows two different trends for low and high values of $Re_{SG}$ that are associated with the stratified flow regime. For $Re_{SG}$ < 1050, the void fraction in stratified flow is sensitive to the pipe orientation whereas for $Re_{SG}$ > 1050, the void fraction is relatively insensitive to the pipe orientation. These two different trends can be explained based on the flow structure of stratified flow regime at low and high gas flow rates. At low gas flow rates, the flow stratification occurs due to the gravity force acting on the liquid phase while the gas phase is subjected to the dominant buoyancy force compared to the inertia force. Thus the void fraction at these low gas flow rates, depend only on the pipe orientation that governs the thickness of the stratified liquid layer. However, in case of high gas flow rates, the stratification occurs between liquid layer and high inertia gas phase which is relatively insensitive to the buoyancy effects and shears the liquid surface and hence causes the void fraction to become insensitive to the pipe orientation. Overall, it can be concluded that the effect of pipe orientation on the void fraction is significant for low liquid and gas flow rates that typically correspond to bubbly, slug and stratified flow and is insensitive to the pipe orientation for shear driven flows such as annular flow and stratified flow with high gas flow rates. It can also be concluded that the void fraction in downward inclinations is consistently higher than that in upward inclined two phase flow.

### *C. Two Phase Pressure Drop*

As shown in (2), the total two phase pressure drop comprises of hydrostatic (function of void fraction), frictional (function of phase flow rates and flow structure) and accelerational components of two phase pressure drop. Since, the flow patterns and void fraction are both functions of pipe orientation; it is of interest to check the effect of pipe orientation on two phase pressure drop. As seen from Fig. 6, the total two phase pressure drop is very sensitive to the pipe orientation for +20o ≥ θ > 0o. At these pipe orientations, the total two phase pressure drop measured at ReSG = 170 is consistently higher than that that measured at 170 < ReSG < 1050. This tendency of higher pressure drop at low ReSG compared to higher ReSG values is more prominent in +20o pipe orientation and gradually reduces with decreasing pipe orientations. This anomaly at these pipe orientations can be explained based on the trend of void fraction illustrated in Fig. 5 and the flow reversal phenomenon to be discussed later in this section. Since for θ > 0o, the void fraction is consistently lower than that for θ < 0o, the hydrostatic component of two phase pressure drop is consistently higher for upward inclined pipe orientation that contributes to the total pressure drop. Moreover, because of the flow reversal observed at ReSG > 170, the frictional component of the two phase pressure drop reduces and hence the total pressure drop for ReSG > 170 is lower than that found at higher gas flow rates. Similar to the trend of void fraction at low ReSL and high ReSG, the two phase pressure drop is found to remain practically unaffected by the pipe orientation at ReSL = 2000 and high ReSG > 15,600 that corresponds to annular flow regime. Although, the void fraction is sensitive to low values of ReSL and ReSG for $\theta < 0^o$, the two phase pressure drop is insensitive to the change in pipe orientation. It appears that this trend is because the decrease in hydrostatic pressure drop is balanced by the increase in frictional component of two phase pressure drop with increasing downward pipe orientations. However, with increasing ReSL, the total two phase pressure drop increases with decrease in downward inclined pipe orientations.

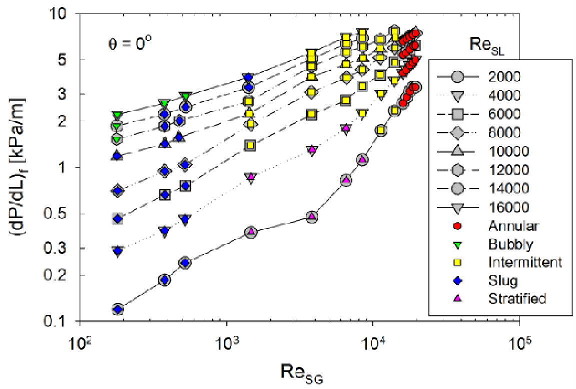
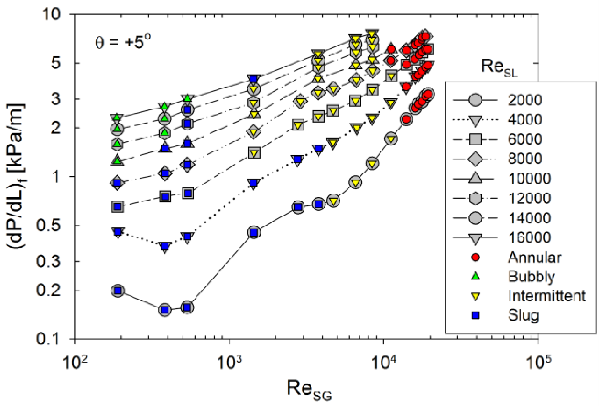
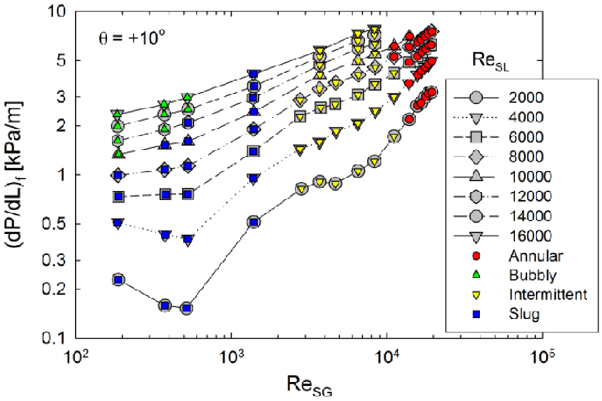


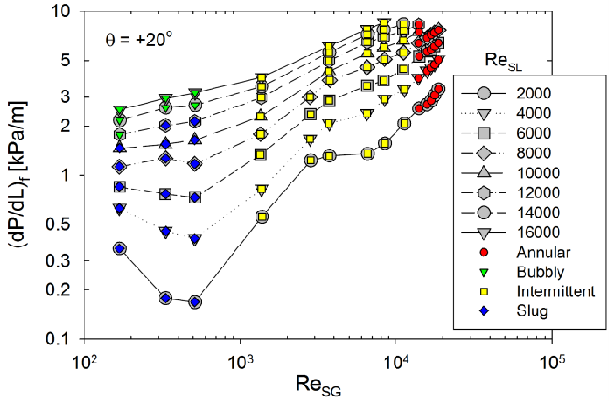

Fig. 7 Frictional component of two phase pressure drop in horizontal pipe orientation




(a) θ = +5°



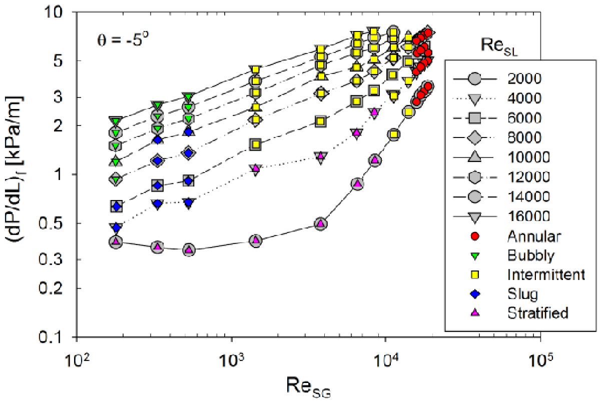

(b) θ = +10°



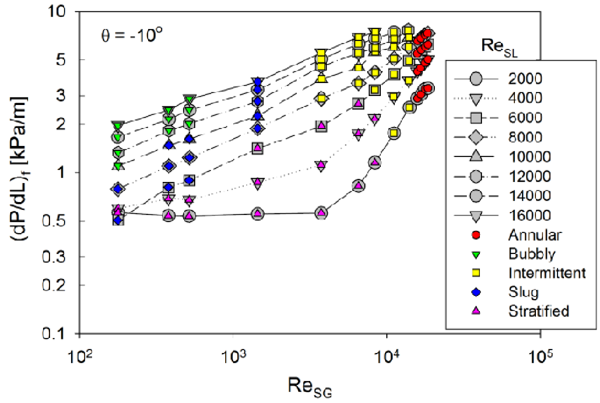

(c) θ = +20°

Fig. 8 Frictional component of two phase pressure drop in upward inclined pipe orientation

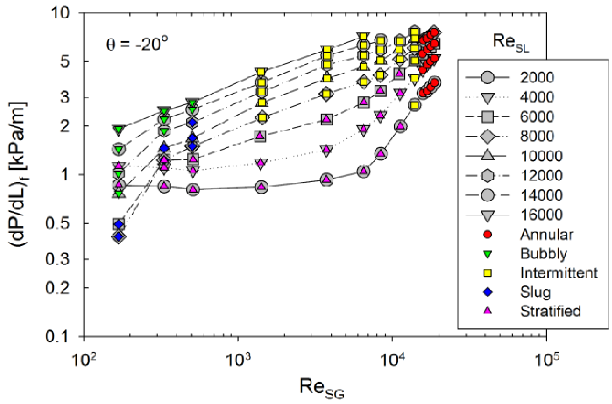
Since, most of the modeling techniques available in the literature are developed for the frictional component of the two phase flow; it is also of interest to look at the trends of frictional component of two phase pressure drop at various pipe orientations as a function of phase flow rates. Figs. 7, 8 and 9 show the variation of frictional component of two phase pressure drop for horizontal, upward inclined and downward inclined pipe orientations, respectively. In case of horizontal flow, it is seen that for all values of $Re_{SL}$, the frictional pressure drop tend to increase consistently with increase in the gas superficial Reynolds number. For a fixed $Re_{SL}$, the frictional pressure drop increases as the flow pattern transits from slug to stratified to annular and slug to intermittent to annular.




(a) θ = –5°




(b) θ = –10°




(c) θ = –20°

Fig. 9 Frictional component of two phase pressure drop in downward inclined pipe orientations

It is intuitive that the flow structure of different flow patterns dictates the magnitude of the frictional pressure drop. For separated type of flow i.e., annular flow, the frictional pressure drop is essentially due to the friction of liquid phase in contact with the pipe wall and the interfacial friction at the gas-liquid interface that results into considerable frictional pressure drop even at low $Re_{SL}$. As described earlier, the intermittent flow pattern is of chaotic and pulsating nature that significantly promotes turbulence and hence results into high values of frictional pressure drop compared to bubbly, slug and stratified flow patterns. As depicted in Fig. 8, the overall trend of the two phase frictional pressure drop with respect to change in phase flow rates is similar to that for horizontal flow except at $Re_{SL} < 8000$ and $Re_{SG} < 1000$. At these flow rates, a peculiar phenomenon of flow reversal is observed where the liquid film in contact with the pipe wall tends to flow in a direction opposite to that of the mean flow. The flow reversal phenomenon is essentially the inability of the gas phase to carry liquid phase along with it in the downstream direction and is a consequence of the balance between interfacial drag, gravitational pull acting on the liquid phase and drag exerted by either phase (gas or liquid) on the pipe wall. For high values of gas flow rates, the interfacial drag of the gas phase acting on the liquid phase and the kinetic energy of liquid phase is greater than the combined effect of the gravitational pull and the wall drag. In this case, the liquid phase moves in the same direction as that of the gas flow. With decrease in the gas flow rates, the interfacial drag exerted by the gas phase on the liquid phase decreases and further at some value of gas flow rate; the interfacial drag force is balanced by the gravitational pull acting on the liquid phase. At this moment, the wall shear stress component is negligible and the liquid phase appears to remain stationary in contact with the pipe wall. With any further decrease in the gas flow rate, the liquid phase starts to move in a direction opposite to that of the mean flow resulting into the phenomenon of flow reversal. Similar type of observations are reported by [14] and [15] for vertical upward flow. References [14] and [15] reported the appearance of flow reversal in the churn (intermittent)-annular transition regime whereas in the present study the flow reversal phenomenon is observed in slug flow regime as shown in Fig. 8. This discrepancy in the flow regime associated with flow reversal can be explained based on the experimental work of [16], who observed that the gas flow rates associated with the flow reversal phenomenon increase with increase in the pipe orientation. Thus, it is anticipated that the flow regime corresponding to flow reversal will shift from slug to intermittent-annular transition as the two phase flow orientation approaches vertical upward flow. From Fig. 8, it is evident that the phenomenon of flow reversal exacerbates with increasing pipe orientation. For $+5^o$ inclination, the pressure drop minimum is found at $Re_{SG} = 350$ that corresponds to 23% decrease in the pressure drop measured at $Re_{SG} = 170$. With increase in the pipe orientation to $+10^o$, the pressure drop minimum further shifts to $Re_{SG} = 540$ that gives 38% decrease in the pressure drop measured at $Re_{SG} = 170$. Finally, at $+20^o$, the pressure drop minimum is also seen at $Re_{SG} = 540$ that causes 54% reduction in the pressure drop compared to that measured at $Re_{SG} = 170$. It is important to mention that based on the trend of frictional pressure drop at low $Re_{SG}$, the point of pressure minimum for $+10^o$ and $+20^o$ is expected to be at $Re_{SG} > 540$. However, due to the limitation on the air side mass flow meter used in this experimental work, we could not measure data for $540 < Re_{SG} < 1100$. A comparison of the frictional pressure drop between horizontal and upward inclined flow reveals the two phase frictional pressure drop to increase consistently (except for flow reversal) with increase in the pipe orientation. In bubbly flow regime, the effect of pipe orientation on frictional pressure drop is nominal (0.3 kPa/m) however, in intermittent flow regime a 38% (1.5 kPa/m) increase in the frictional pressure drop is observed as the pipe orientation shifts from horizontal to $+20^o$.

In case of downward inclined pipe orientations, for $Re_{SL} > 2000$ and $Re_{SG} > 350$, the general trend of two phase frictional pressure drop is in agreement with that of horizontal and upward inclined two phase flow. For $Re_{SL} = 2000$, the two phase frictional pressure drop is found to follow a saturated trend. The sudden increase in the frictional pressure drop from this saturating trend is found to occur at higher $Re_{SG}$ (1400, 3650 and 6200) as pipe orientation shifts from $-5^o$ to $-20^o$. This is probably because, with increase in the downward pipe orientation, the gravity assisted stratified flow exists predominantly and it requires higher gas flow rate to establish a shear driven stratified flow that causes sudden rise in the frictional pressure drop. Another explanation for this trend based on the experimental data could be given in terms of the connection between total, hydrostatic and frictional components of two phase pressure drop. As shown in Fig. 6, the total two phase pressure drop at $Re_{SL} = 2000$ and $170 \leq Re_{SG} < 3700$, remains unaffected for $-20^o$ and $-10^o$. At the same time for these values of $Re_{SL}$ and $Re_{SG}$, the hydrostatic component of two phase pressure drop (inversely proportional to the magnitude of void fraction) will decrease marginally due to very small increase in the void fraction. Now, in order to keep the total two phase pressure drop same at increasing $Re_{SG}$, the frictional pressure drop remains virtually unchanged over a range of gas superficial Reynolds numbers.

It is also seen from Fig. 9 that in downward inclinations in particular for $-10^o$ and $-20^o$ and $Re_{SL} = 6000$ and $Re_{SL} = 8000$, the frictional pressure drop at $Re_{SG} = 170$ is distinctly lower than that at $Re_{SL} < 6000$. A plausible justification for this trend can be explained by the increase in total pressure drop with increase in $Re_{SG}$ at $Re_{SL} = 6000$ (Fig. 6) and decrease in the hydrostatic component of two phase pressure drop due to increase in void fraction (Fig. 5) measured at similar phase flow rates. Thus, to compensate for the decrease in hydrostatic pressure drop and increase in the total pressure drop, the frictional component of two phase pressure drop is higher at $Re_{SG} = 350$ compared to that at $Re_{SG} = 170$.

A quick comparison between upward and downward inclinations shows that for $Re_{SL} < 10{,}000$ and $Re_{SG} < 14{,}000$, the frictional pressure drop in downward inclinations is consistently higher than that in the upward pipe inclinations. This trend is expected since the two phase flow within the range of Reynolds numbers mentioned above is not a shear driven flow and hence is sensitive to the pipe orientation.

## IV. Two Phase Flow Correlations

### A. Void Fraction Correlations

As mentioned in the introduction section, there are very few correlations available in the literature that can be used with enough accuracy to predict void fraction in inclined two phase flow. Some of these available correlations include work of [17]-[20]. Each of these correlations have their own limitations in terms of their application to a particular flow pattern, fluid combination or a pipe orientation and hence cannot be used to predict void fraction for a wide range of two phase flow conditions. Recently [11] developed a flow pattern independent drift flux model based void fraction correlation for a wide range of gas-liquid two phase flows including circular and non-circular flow geometries. Their void fraction correlation was tested against a comprehensive set of experimental data consisting of 8255 data points that comprised of air-water, refrigerants (R11, R12, R22, R134a, R114, R410A, R290 and R1234yf), air-oil, steam-water, argon-water, natural gas-water, air-kerosene, air-glycerin, argon-acetone, argon-ethanol and argon-alcohol. Their correlation is based on the concept of drift flux model whose general form is as expressed by (5) where $U_{SG}$ and $U_M$ is the gas superficial velocity (m/s) and two phase mixture velocity (m/s), respectively.

$$\alpha = \frac{U_{SG}}{C_o U_M + U_{GM}} \tag{5}$$

The distribution parameter ($C_o$) and drift velocity ($U_{GM}$) with units of m/s are the drift flux model variables. Reference [11] modeled these variables as a function of phase flow rates, fluid properties, pipe geometries, pipe diameter and void fraction as shown in (6) and (7). The variable $C_{o,1}$ required in (7) is as defined by (8) and (9) for two different flow conditions. The criteria of $0^o \geq \theta \geq -50^o$ and $Fr_{SG} \leq 0.1$ is exclusively to account for the higher values of void fraction in downward inclined pipe orientations compared to upward inclined two phase flow. This criterion is based on the physical structure of the slug and stratified flow observed in downward inclinations.

$$C_o = C_o\left(\frac{\rho_G}{\rho_L}, \frac{U_M \rho_L D_h}{\mu_L}, \frac{U_{SG}}{U_{SG}+U_{SL}}, x, f_{TP}, \theta, \alpha\right) \tag{6}$$

$$C_o = \frac{2-(\rho_G/\rho_L)^2}{1+(\mathrm{Re}_{TP}/1000)^2} + \frac{\left[\left(\sqrt{(1+(\rho_G/\rho_L)^2\cos\theta)/(1+\cos\theta)}\right)^{(1-\alpha)}\right]^{2/5} + C_{o,1}}{1+(1000/\mathrm{Re}_{TP})^2} \tag{7}$$

$$C_{o,1} = \left(C_1 - C_1\sqrt{\rho_G/\rho_L}\right)\left((2.6-\beta)^{0.15} - \sqrt{f_{TP}}\right)(1-x)^{1.5} \tag{8}$$

The constant $C_1$ in (8) is 0.2 for circular and annular pipe geometries and 0.4 for rectangular pipe geometry. The gas volumetric flow fraction is defined as $\beta = U_{SG}/(U_{SG}+U_{SL})$.

$$C_{o,1} = 0 \text{ (for } 0^o \geq \theta \geq -50^o \text{ and } Fr_{SG} \leq 0.1) \tag{9}$$

The purpose of using (9) for specific flow conditions is to make $C_o < 1$ so as to compensate for the higher values of void fraction encountered in downward inclined flow. The Froude number based on gas superficial velocity ($Fr_{SG}$) and phase densities is defined by (10).

$$Fr_{SG} = \sqrt{\frac{\rho_G}{\rho_L - \rho_G}}\frac{U_{SG}}{\sqrt{gD_h\cos\theta}} \tag{10}$$

The two phase mixture Reynolds number ($\mathrm{Re}_{TP}$) used in (7) is defined by (11) as a function of two phase mixture velocity ($U_M = U_{SL}+U_{SG}$), liquid phase density ($\rho_L$), viscosity ($\mu_L$) with units of Pa·s, and the hydraulic diameter ($D_h$).

$$\mathrm{Re}_{TP} = (U_M \rho_L D_h)/\mu_L \tag{11}$$

The two phase friction factor ($f_{TP}$) is based on the two phase Reynolds number and is calculated using equation of [21]. The drift velocity term ($U_{GM}$) required in (5) is calculated from (12) where the variables $C_2$, $C_3$ and $C_4$ are defined by (13), (14) and (15), respectively.

$$U_{GM} = (0.35\sin\theta + 0.54\cos\theta)\sqrt{\frac{gD_h\Delta\rho}{\rho_L}}\sqrt{1-\alpha}\,C_2C_3C_4 \tag{12}$$

$$C_2 = \begin{cases}\left(\dfrac{0.434}{\log_{10}(\mu_L/0.001)}\right)^{0.15} & \text{for } (\mu_L/0.001) > 10\\ 1 & \text{for } (\mu_L/0.001) \leq 10\end{cases} \tag{13}$$

$$C_3 = \begin{cases}(La/0.025)^{0.9} & \text{for La} < 0.025\\ 1 & \text{for La} \geq 0.025\end{cases} \tag{14}$$

$$\begin{aligned} &C_4 = 1\\ &C_4 = -1 \text{ for } (0 > \theta \geq -50 \text{ and } \mathrm{Fr_{SG}} \leq 0.1)\end{aligned} \tag{15}$$

$\mu_L$ and La in (13) and (14) are the liquid phase dynamic viscosity and Laplace number ($\mathrm{La}=\sqrt{\sigma/(g\Delta\rho)}/D_h$), respectively. The σ used in Laplace number is the gas-liquid interface surface tension with units N/m. In this study, the performance of the void fraction correlation by [11] is tested against the experimental data for near horizontal upward and downward inclined pipe orientations and some of the top performing correlations available in the literature such as [18], [20], [22]-[25]. The correlations of [18], [20], [22] are the drift flux model based correlations (appropriate for buoyancy driven flows) while those of [23]-[25] are separated flow model based correlations (appropriate for shear driven flows). The void fraction experimental data at near horizontal pipe orientation consists of the data of [1]-[3] and [26] in addition to the data collected in the present study. The accuracy of these void fraction correlations is gauged based on the percentage of data points predicted within specified error bands for different ranges of void fraction. Based on the measured void fraction and associated flow patterns, the entire range of void fraction is divided into three sub ranges such that the first range of void fraction $0 \le \alpha \le 0.25$ approximate bubbly flow pattern, second range of void fraction i.e., $0.25 < \alpha \le 0.75$ corresponds to slug, intermittent and stratified flow patterns and finally the last range of void fraction i.e., $0.75 < \alpha < 1$ represent annular flow regime. For the experimental data collected in present study and that available in the literature, 144, 640 and 588 data points were categorized in each range of void fraction. Tables I and II show the performance of different void fraction correlations for these three ranges of void fraction and for $Fr_{SG} \le 0.1$, respectively. It should be noted that the performance criteria used for each range of void fraction is different and depends on the face value of the void fraction and sensitivity of void fraction to any calculated variable such as two phase mixture density. More details about this performance criteria is reported by [11] and [27]. It is evident from Table I that the correlation of [11] presented in this work performs consistently well for all three ranges of void fraction. The correlations of [24] and [28] perform slightly better than [11] for low range of void fraction. However, with increasing values of void fraction, their accuracy drops down. Similar tendency is observed for other two correlations of [18] and [22]. Table II shows the performance of different void fraction correlations for the case of low values of gas flow rates that as reported by [11] are typically associated with $Fr_{SG} \le 0.1$. It is clearly seen that the correlation of [11] gives highest accuracy by predicting 46% and 71% and 86% of data points within ±10%, ±20% and ±30% error bands, respectively. The performance of [11] for all the void fraction data considered in this study is illustrated graphically in Fig. 10.

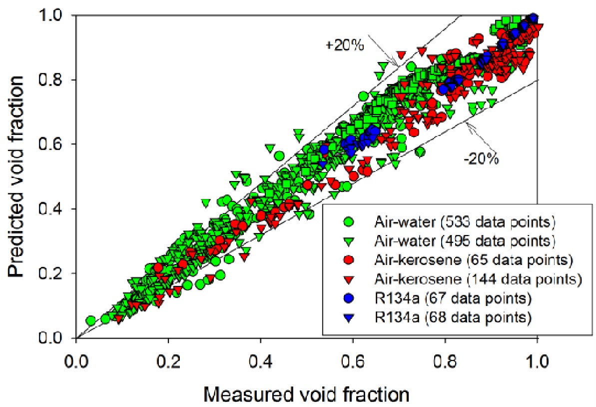



Fig. 10 Performance of [11] for near horizontal pipe orientations (●: Downward inclined, ▼: Upward inclined)

TABLE I
PERFORMANCE OF VOID FRACTION CORRELATIONS AGAINST THE EXPERIMENTAL DATA IN NEAR HORIZONTAL PIPE ORIENTATIONS

| Correlation | $(0 \le \alpha \le 0.25)^a$ (144 data points) | $(0.25 < \alpha \le 0.75)^b$ (640 data points) | $(0.75 < \alpha < 1)^c$ (588 data points) |
|---|---|---|---|
| Bhagwat and Ghajar (2014) [11] | 89 | 87 | 87 |
| Gomez et al. (2000) [18] | 91 | 84 | 41 |
| Rouhani and Axelsson (1970) [22] | 80 | 82 | 51 |
| Woldesemayat and Ghajar (2007) [20] | 30 | 79 | 86 |
| Cioncolini and Thome (2012) [25] | 0 | 64 | 88 |
| Lockhart and Martinelli (1949) [23] | 84 | 50 | 86 |
| Smith (1969) [24] | 92 | 83 | 76 |
| Permoli et al. (1970) [28] | 91 | 28 | 50 |

[a] % of data points predicted within ±30% error bands, [b] % of data points predicted within ±15% error bands, [c] % of data points predicted within ±7.5% error bands.

TABLE II
PERFORMANCE OF THE VOID FRACTION CORRELATIONS FOR LOW VALUES OF GAS FLOW RATES ($Fr_{SG} \le 0.1$)

| Correlation | % of data points within ±10% error bands | % of data points within ±20% error bands | % of data points within ±30% error bands |
|---|---|---|---|
| Bhagwat and Ghajar (2014) [11] | 46 | 71 | 86 |
| Gomez et al. (2000) [18] | 41 | 65 | 78 |
| Rouhani and Axelsson (1970) [22] | 10 | 32 | 64 |
| Woldesemayat and Ghajar (2007) [20] | 16 | 32 | 49 |
| Cioncolini and Thome (2012) [25] | 3 | 11 | 21 |
| Lockhart and Martinelli (1949) [23] | 14 | 38 | 69 |
| Smith (1969) [24] | 34 | 63 | 79 |
| Permoli et al. (1970) [28] | 12 | 40 | 72 |

Although, the overall performance of [18], [22] and [24] is comparative to that of [11] for the experimental data considered in this work, compared to all these correlations, the correlation of [11] is more accurate, versatile, robust and is applicable for a wide range of two phase flow scenarios involving wide range of system pressure, fluid properties, pipe diameters and both boiling and non-boiling two phase flows.

### *B. Pressure Drop Correlations*

Two phase flow literature reports a large number of empirical and semi-empirical two phase pressure drop models essentially based on the homogeneous flow model, separated flow model, two phase friction factor model, phenomenological model and flow pattern specific empirical models. Out of these different types of two phase flow models, the scope of the present work is to investigate the application and accuracy of the homogenous flow model based correlations to predict total two phase pressure drop in near horizontal inclined two phase flow. As mentioned earlier, the total two phase pressure drop consists of hydrostatic, frictional and accelerational components where the pressure drop due to flow acceleration could be neglected for non-boiling type of two phase flow. Using the concept of homogeneous flow model that assumes no slip condition at the gas-liquid interface and assumes thermodynamic equilibrium between the two phases, the total two phase pressure drop can be can be expressed as shown in (16). The two phase mixture density ($\rho_M$) used in (16) is based on the assumption of no slip at the gas-liquid interface and is modeled as a function of gas ($\rho_G$) and liquid ($\rho_L$) phase density and the two phase flow quality ($x$) as shown in (17).

$$\left(\frac{\Delta P}{L}\right)_t = \rho_M g \sin\theta + \frac{f_M G^2}{2\rho_M D} \tag{16}$$

$$\rho_M = \left(\frac{x}{\rho_G} + \frac{1-x}{\rho_L}\right)^{-1} \tag{17}$$

The two phase mixture friction factor ($f_M$) used in (16) is based on the two phase mixture Reynolds number ($Re_M$) defined in terms of mixture mass flux ($G$[kg/m$^2$·s]), pipe diameter ($D$) and pseudo two phase mixture dynamic viscosity ($\mu_M$ [Pa·s]) as shown in (18). Two phase literature reports several pseudo two phase dynamic viscosity models as listed in Table III to calculate two phase Reynolds number and hence the two phase friction factor to be used in (16).

$$Re_M = \frac{GD}{\mu_M} \tag{18}$$

The two phase mixture friction factor is calculated using conventional single phase friction factor correlation such as [21], [29], and [30]. In particular the correlations of [31] and [32] recommend friction factor correlations of [21] and [30], respectively. Thus, the only parameter that influences the magnitude of the predicted two phase pressure drop is the pseudo two phase mixture dynamic viscosity. The accuracy of the homogeneous flow model using the two phase mixture viscosity listed in Table III is gauged against the experimental data (655 data points for total two phase pressure drop) collected in this study. The performance of these correlations is ranked based on the number of percentage of data points predicted within ±20% and ±30% error bands for each pipe orientation. This type of performance criteria has been used in the past in the work of [33]-[35] for horizontal and vertical pipe orientations.

As reported in Table IV, for all pipe orientations considered in this study, the best performance is consistently given by [32] (Model 2) and [36]-[38]. It is evident that these correlations derived for horizontal two phase flow perform well for horizontal and upward inclined pipe orientations.

TABLE III
TWO PHASE DYNAMIC VISCOSITY MODELS

| | Correlation |
|---|---|
| Akers et al. (1959) [36] | $\mu_M = \mu_L/(1-x) + x\sqrt{\rho_L/\rho_G}$ |
| Beattie and Whalley (1982) [31] | $\mu_M = \mu_L(1-\beta)(1+2.5\beta) + \mu_G\beta$ |
| Ciccihitti et al. (1960) [39] | $\mu_M = x\mu_G + (1-x)\mu_L$ |
| Davidson et al. (1943) [40] | $\mu_M = \mu_L(1 + x(\rho_L/\rho_G) - 1)$ |
| Dukler et al. (1964) [41] | $\mu_M = \rho_M\left(x\mu_G/\rho_G + (1-x)\mu_L/\rho_L\right)$ |
| Fourar and Boris (1995) [42] | $\mu_M = (1-\beta)\mu_L + \beta\mu_G + 2\sqrt{\beta(1-\beta)\mu_L\mu_G}$ |
| Garcia et al. (2003) [43] | $\mu_M = \mu_L\rho_G / x\rho_L + (1-x)\rho_G$ |
| Lin et al. (1991) [37] | $\mu_M = (\mu_L\mu_G)/(\mu_G + x^{1.4}(\mu_L - \mu_G))$ |
| McAdams et al. (1942) [38] | $\mu_M = \left(x/\mu_G + (1-x)/\mu_L\right)^{-1}$ |
| Awad and Muzychka (2008) [32] | $\mu_M = \mu_L \dfrac{2\mu_L + \mu_G - 2(\mu_L - \mu_G)x}{2\mu_L + \mu_G + (\mu_L - \mu_G)x}$ |
| Awad and Muzychka (2008) [32] | $\mu_M = \mu_G \dfrac{2\mu_G + \mu_L - 2(\mu_G - \mu_L)x}{2\mu_G + \mu_L + (\mu_G - \mu_L)x}$ |
| Awad and Muzychka (2008) [32] | Arithmetic mean of model 1 and model 2 |
| Awad and Muzychka (2008) [32] | $\mu_M = \dfrac{1}{4}\left(\begin{array}{c}(3x-1)\mu_G + [3(1-x)-1]\mu_L + \\ \sqrt{[(3x-1)\mu_G + (3(1-x)-1)\mu_L]^2 + 8\mu_G\mu_L}\end{array}\right)$ |

TABLE IV
PERFORMANCE OF THE TWO PHASE DYNAMIC VISCOSITY MODELS AGAINST THE PRESSURE DROP DATA MEASURED IN PRESENT STUDY

| Correlation | 0° | | +5° | | +10° | | +20° | | -5° | | -10° | | -20° | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | (1) | (2) | (1) | (2) | (1) | (2) | (1) | (2) | (1) | (2) | (1) | (2) | (1) | (2) |
| [36] | 60.2 | 90.4 | 51.5 | 91.9 | 67.7 | 92 | 60.6 | 79.8 | 47.9 | 61.7 | 50.1 | 60.3 | 37.5 | 51.1 |
| [31] | 25.5 | 89.2 | 25.5 | 61.9 | 29.3 | 79.8 | 38.3 | 66.9 | 22.3 | 56.4 | 30.1 | 66.7 | 29.5 | 47.7 |
| [39] | 46.2 | 67.7 | 45.4 | 81.4 | 56.6 | 79.8 | 56.4 | 73.4 | 25.5 | 40.4 | 24.4 | 36.7 | 13.6 | 25.2 |
| [40] | 48.4 | 73.1 | 49.5 | 88.7 | 60.6 | 86.9 | 56.4 | 71.3 | 27.7 | 43.6 | 28.9 | 42.2 | 17.1 | 39.5 |
| [41] | 27.2 | 46.6 | 14 | 38.9 | 13.1 | 32.3 | 37.7 | 64.7 | 15.3 | 22.3 | 11.5 | 31.9 | 10.2 | 29.5 |
| [42] | 17.2 | 72 | 13.4 | 54.6 | 25.3 | 68.7 | 38.3 | 63.8 | 17.0 | 53.2 | 24.4 | 36.7 | 27.3 | 46.6 |
| [43] | 7.5 | 21.5 | 12.1 | 26.8 | 12.1 | 27.3 | 27.7 | 39.4 | 15.3 | 24.9 | 28.9 | 42.2 | 18.0 | 30.5 |
| [37] | 63.4 | 86.0 | 58.8 | 91.8 | 72.7 | 93.9 | 66.0 | 85.1 | 41.5 | 56.4 | 41.1 | 52.2 | 28.4 | 39.8 |
| [38] | 60.2 | 90.3 | 55.7 | 96.9 | 68.7 | 93.0 | 64.9 | 85.1 | 50.1 | 60.6 | 51.1 | 57.8 | 39.8 | 48.9 |
| [32] Model 1 | 46.2 | 67.7 | 47.4 | 81.4 | 55.6 | 78.8 | 55.3 | 73.4 | 25.5 | 40.4 | 23.3 | 36.7 | 13.6 | 25 |
| [32] Model 2 | 61.3 | 84.9 | 56.7 | 92.8 | 70.7 | 93.9 | 66.0 | 86.2 | 42.4 | 54.3 | 40.2 | 53.3 | 31.8 | 42 |
| [32] Model 3 | 49.5 | 74.2 | 49.5 | 86.6 | 61.6 | 84.8 | 61.7 | 80.9 | 29.8 | 45.7 | 30.0 | 43.3 | 18.2 | 30.7 |
| [32] Model 4 | 46.2 | 67.7 | 47.4 | 81.4 | 55.6 | 78.8 | 55.3 | 73.4 | 25.5 | 40.4 | 23.3 | 36.7 | 13.6 | 25 |

(1) % of data points predicted within ±20% error bands, (2) % of data points predicted within ±30% error bands

However, for downward pipe orientations, the accuracy of these correlations deteriorates significantly. In comparison to +5° and +10°, the accuracy of these correlations is found to drop down slightly for +20° pipe orientation. A close look at each of these pipe orientations in terms of the flow patterns reveal that the inaccuracies of these two phase dynamic viscosity models are associated with certain flow patterns. In case of upward inclined flow, the correlations listed in Table III predict the total pressure drop with an error greater than ±30% for some part of slug flow where flow reversal is observed. In addition to this, the inaccuracies of these correlations also exist in some part of intermittent flow regime typically $Re_{SL} > 12{,}000$ and $1000 \leq Re_{SG} \leq 14{,}000$. In case of downward inclined flow, most of the inaccuracies of the two phase dynamic viscosity models are associated with slug and stratified flow regimes. It is seen in the previous section that in downward pipe inclinations, the two phase flow behavior in particular for slug and stratified flow is strongly influenced by the interaction between buoyancy and inertia forces that eventually affect the void fraction, frictional pressure drop and hence the total two phase pressure drop to a considerable extent. In fact, for downward inclinations, after ignoring the data for slug and stratified flow (21 data points), the correlations of [36] and [38] predict about 78% of data points within ±30% error bands for -5° pipe orientation. Similarly for -10° and -20° pipe orientations, after ignoring slug and stratified flow data (28) and (30) data points, respectively; the accuracy of [36] improves from 60.3% and 51.1% to 82.3% and 72.1%, respectively. Likewise, a noticeable improvement in the accuracy of all other correlations is observed. Thus, it can be concluded that, the homogeneous flow model for two phase pressure drop can be used effectively only for the horizontal and upward inclined pipe orientations. However, the performance of these correlations needs to be verified against steeper pipe inclinations ($\theta > +20^{o}$) and a wide range of fluid properties and pipe diameters. In case of downward inclined two phase flow, the application of homogeneous flow model is limited to bubbly, intermittent and annular flow patterns. It is found that the homogenous flow model fails to predict the total two phase pressure drop in slug and stratified flow in downward pipe inclinations.

## V. CONCLUSIONS

The present study is focused on the experimental investigation of flow patterns, void fraction, and two phase pressure drop in near horizontal upward and downward inclined pipe orientations. Some of the distinct conclusions and salient observations done in this study are summarized below:

(1) It is found in the present study that the pipe orientation alters the transition lines between different flow patterns in particular the transition between slug-stratified, slug-intermittent and stratified-intermittent flow patterns. The transition from shear driven flow i.e., annular flow to intermittent flow is only trivially influenced by the pipe orientation.

(2) The void fraction is found to be significantly influenced by the downward pipe inclinations especially at low liquid and gas flow rates. With increasing gas flow rates, as the flow pattern approaches annular flow regime, and the effect of pipe orientation on the void fraction goes away.

(3) Similar to void fraction, total two phase pressure drop and frictional pressure drop are significantly influenced by the pipe orientation at low gas and liquid flow rates. A peculiar phenomenon of flow reversal is observed in upward inclined flow (low $Re_{SL}$ and low $Re_{SG}$ values) that noticeably affects the general trend of two phase frictional pressure drop.

(4) The performance of void fraction correlation by [11] is checked against the 1372 data points and some of the top performing correlations available in the literature. It is showed that the correlation of [11] performs consistently for the entire range of void fraction.

(5) The measured data of total two phase pressure drop at different pipe orientations is compared against the homogenous flow model (HFM). It is found that HFM cannot predict the total two phase pressure drop for slug

and stratified flow in downward inclined pipe orientations.

(6) Considering the influence of pipe orientation on the two phase flow parameters, in particular the shift in transition lines on the flow pattern maps, the tendency of higher values of void fraction in downward pipe inclinations, and flow reversal phenomenon in upward inclined two phase flow, it is of significant interest and hence is recommended to study the two phase flow phenomenon at steeper pipe inclinations.

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