Experimental Study of Two Phase Air-Water Flow in Large Diameter Vertical Pipe

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Abstract* - Recently, due to increase in production demand in nuclear and oil & gas industries, requirement to migrate towards larger pipe sizes for future developments have become essential. However, it is interesting to note that almost all the research on two phase gas-liquid flow in vertical pipe upflow is based on small diameter pipes (D ≤ 100mm) and the experimental work on the two phase gas-liquid flow in large diameter (D > 100mm) vertical pipe is scarce. Under the above circumstances, the application of modelling tools/correlations based on small diameter pipes in predicting flow behvaiour (flow pattern, void fraction, and pressure gradient) poses severe challenges in term of accuracy. The results presented in this paper is motivated by the need to introduce the research work done to the industries where the data pertaining to large diameter vertical pipe is scarce and there is a lack of understanding of two phase gas-liquid flow behaviour in large diameter (D > 100mm) vertical pipes.

The unique aspect of the results presented here is that the experimental data has been generated for a 254mm inner diameter vertical pipe that forms an excellent basis for the assessment of modelling tools/correlations. The paper presents results of (i) a systematic investigation of the flow patterns in large diameter vertical pipes and identify the transition between subsequent flow patterns, (ii) compare it directly with existing large (150mm) and small diameter data (28mm & 32mm) under same air-water superficial velocities range, (iii) exemplify that existing available empirical correlations/models/codes are significantly in error when applied to large diameter vertical pipe for predictions and (iv) lastly, assesses the predictive capability of a well known commercial multiphase flow simulator.

Keywords: *Large Diameter, Small Diameter, Air-Water, Flow Patterns, Vertical Pipe, Void Fraction, Flow regime transitions and OLGA.*

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I. Introduction

The two phase gas-liquid flow in pipes can adopt various physical configurations knows as flow patterns. These flow patterns strongly influence the key design characteristics (e.g. void fraction and pressure drop) of two phase system; hence it is important for the designer to be able to predict the flow pattern correctly. Such two phase gas-liquid flows are widely encountered in many industrial applications e.g. in nuclear and oil & gas etc. In these industries, the accurate predictions of flow patterns are extremely important for the optimum design of the systems. Thus the aim of almost all the multiphase flow research has been with the prediction of flow patterns so that the accurate theories for the flow mechanisms can be developed and the basic flow parameters (void fraction and pressure drop) in these flow patterns can be established.

Recently due to increase in production demand in above industries, requirement to migrate towards larger pipe sizes for future developments have become essential. In this context, two phase flow in the large diameter ($D > 100$mm) vertical pipe has become a subject of great interest specially in nuclear and in oil & gas industry, as for former, interest lies in the safety requirements (Khartabil and Spinks, 1995; Ohnuki and Akimoto, 2000; Shoukri *et al.*, 2000; Schlegel *et al.*, 2009) while for the later the concern is economics (Pickering *et al.*, 2001). Here it is emphasized that the pipe diameter, $D \leq 100$mm is considered as small diameter pipe while diameter, $100 < D \leq 200$mm are usually quoted as intermediate sizes and diameter, $D > 200$mm are taken as large diameter pipe. It is quite interesting to note that although vast research exist on two phase vertical pipe upflow, it is based on small diameter pipes ($D < 100$mm) and the experimental work on predicting the two phase flow behaviour in pipe diameter, $D > 100$mm is scarce. Moreover, some previous research works suggests that the flow behaviour of two phase flow in larger diameter vertical pipes is likely to be different from small diameter pipes, e.g. Slug flow, an alternate flow of liquid slugs and large bullet

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shaped smooth and elongated Taylor gas bubble does not exist for large diameter vertical pipes (Cheng *et al.*, 1998; Ohnuki and Akimoto, 2000; Shoukri *et al.*, 2000; Pickering *et al.*, 2001 and Omebere-Iyari, 2007). This notion introduces the uncertainty behind the predictive accuracy of the existing correlations and modelling tools. As most of computational codes are based on flow regime dependent constitutive equations, conditions derived from the small diameter pipes research may not be valid for large diameter vertical pipes posing severe challenges in terms of accuracy. As the experimental data obtained from conventional small diameter pipe with different flow mechanisms cannot be used for developing modelling tools for large diameter pipes, therefore new experimental data must be generated, old correlation/ constitutive equations should be validated and new relations should be modeled that will include the flow pattern (or diameter) influence on gas-liquid flow structure.

Flow pattern & its classifications

For the vertical upward flows in circular conduits, four (4) typical flow patterns **(Figure 1)** may be distinguished namely; bubbly, slug, churn and annular flows (Hewitt and Roberts, 1969; Taitel *et al.*, 1980; Mishima and Ishii, 1984). Briefly: ***(i)*** In ***Bubbly flow***, the gas is dispersed in the continuous liquid phase. Various researchers (Taitel *et al.*, 1980; McQuillan and Whalley, 1985; Barnea and Brauner, 1986) have further classified this flow as *dispersed bubbly*, where the gas phase is dispersed as small discrete bubbles in continuous liquid phase and *low liquid input bubbly* (*or non-dispersed bubbly or bubbly)* that occurs at low liquid superficial velocities only, where gas bubble of various sizes exists with some occasional coalescence in the core (Taitel *et al.*, 1980). It is to be noted that some researchers (Mishima and Ishii, 1984; Weisman and Kang, 1981) do not delineate any exact distinction between the two types and categorize both under same bubbly flow. ***(ii) Slug flow:*** As the gas superficial velocity is increased from bubbly flow, the gas bubbles begin to coalesce to form long smooth bubble with front having cap/bullet shape. This bubble is referred as Taylor bubble and is of equivalent cross section as that of the tube, being separated from the wall by a thin liquid film. The two consecutive Taylor bubbles are separated by a liquid slug that is usually aerated with gas bubbles that are shed from the tail of the leading Taylor bubble. ***(iii) Churn flow:*** As the gas superficial velocity is increased in slug flow, the Taylor bubble becomes more distorted at gas-liquid interface. The distorted bubble travels in churning motion giving rise to irregular shaped portions of gas and liquid. This flow is also called *froth flow/ churn-turbulent flow/ intermittent flow/ pulsating annular flow* (Brauner & Barnea, 1986; Weisman and Kang, 1981). Some researchers (Mao & Dukler, 1993) do not consider this flow as a separate flow and treat this flow under the slug flow. ***(iv) Annular flow:*** In this flow, the gas phase exists in the core while liquid exists as film around the periphery of the conduit and as the entrainment (drops) in the core. There are two variation of this flow (Hewitt, 1982) namely, *Wispy annular* - where entrained liquid is present as relatively large drops and liquid film contains gas bubbles and *Annular mist* in which gas occupies the centre of the core with liquid drops sizes not large and some liquid flowing along the periphery.

Review of other large diameter vertical upflow work

Although many studies have contributed to the topic of the large diameter vertical pipes (Hills, 1976; Shipley, 1984; Hills, 1992; Hirao *etal.*,1986; Ohnuki and Akimoto, 1996; Hasanein *et al.,* 1997; Cheng *et al.*, 1998; Ohnuki and Akimoto, 2000; Shoukri *et al.*,2000; Prasser *et al.*, 2002; Sun *et al.*, 2002; Oddie *et al.*, 2003; Hibiki and Ishii, 2003; Shen *et al.*, 2005; Prasser *et al.*, 2005; Omebere-Iyari *et al.*, 2007; Omebere-Iyari *et al.*, 2008; Schlegel *et al.,* 2009; Lucas *et al.*, 2010 and Shen *et al.*, 2012) majority of the work performed was restricted to pipe diameter of intermediate sizes ($D \leq 200$mm) because this was considered to be an optimum choice from cost analysis point of view. Only few studies have been conducted for very large diameter sizes ($300 < D < 500$mm) (Shipley, 1984; Ohnuki *et al.*, 1996; Hasanein *et al.,* 1997 and Yoneda *et al.*, 2002). In case of these later higher diameters, the work is been confined to very small length-to-diameter ratio (Shipley,

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1984; $(z/D)_{air\text{-}water}$ = 12.34, Ohnuki *et al.*, 1996; $(z/D)_{air\text{-}water}$ = 4.16, and Hasanein *et al.,* 1997; $(z/D)_{steam\text{-}water}$ = 7.87), hence the two phase flow is still evolving or developing, which may not depict the true two phase flow behaviour in a longer length vertical pipe. In most of the above works with exception of (Cheng *et al.*, 1998; Prasser *et al*., 2002; Omebere-Iyari *et al*., 2007; Omebere-Iyari *et al*., 2008 and Schlegel *et al.,* 2009), the major multiphase flow parameter i.e. flow patterns was studied by flow visualization only and hence can be subjective (Hills, 1976; Shoukri *et al.*,2000; Ohnuki and Akimoto, 2000; Oddie *et al.*, 2003; Shen *et al.*, 2005 and Shen *et al*., 2012). Additionally, in some of other large diameter work, flow patterns were vaguely dealt or completely ignored (Shipley, 1984; Van der Welle, 1985; Clark and Flemmer, 1986; Hirao *etal.*,1986) while in others the objective of study was determination of local flow structure i.e. local void phase and velocity distributions (Ohnuki and Akimoto, 2000; Shoukri *et al.*,2000; Prasser *et al.*, 2005; Shen *et al.*, 2005; Prasser, 2007; Prasser *et al.*, 2007; Shen *et al.*, 2010 and Schlegel *et al.,* 2010 and Lucas *et al.*, 2010), hence any comparison performed with smaller diameter pipes was limited to these parameter only. Moreover it was noticed that, some results for large diameter vertical pipes differed even for same pipe diameter, e.g. Shoukri *et al.* (2000) found that the Taitel *et al.* (1980) flow regime map was found to predict their experimental flow patterns transitions satisfactorily while other quoted that it does not (Omebere-Iyari *et al*., 2007; Omebere-Iyari *et al*., 2008). A number of work e.g. Hasanein *et al.* (1997), Yoneda *et al.* (2002), Prasser *et al*. (2007) and Omebere-Iyari *et al*. (2008) have specific nuclear application of boiling water reactors as they used steam-water as working fluid. Furthermore, in just few studies (Ohnuki and Akimoto, 2000; Shoukri *et al.*, 2000 ; Omebere-Iyari *et al*., 2007; Omebere-Iyari *et al*., 2008 and Schlegel *et al.,* 2009) the results were presented on the flow pattern maps. In the work of Omebere-Iyari *et al*. (2007), the flow patterns were deduced for nitrogen-naphtha under high pressure facility, hence it is restricted in its applicability to petroleum industry only, while in Omebere-Iyari *et al*. (2008) work, steam-water results were presented with very limited data (few data points only!) especially in the region of transition from bubbly to churn flow regime. Ohnuki and Akimoto (2000) flow pattern identification is based on visualization hence may carry some doubt while Schlegel *et al.* (2009) data can only be used for qualitative comparison as no information about operating conditions (air-water superficial velocities) are given with flow pattern identification performed by use of neural network. Overall, while the above review shows significant systematic information on the two-phase flow in large diameter vertical pipes, no previous work performed comparison with smaller diameter pipes with same working fluid under similar air-water superficial velocity range. For this reason comparison between small and large diameter is presented here in this paper so that comparison is more meaningful and straightforward.

From above brief survey it is obvious that it's a necessity to constantly increase / update the database of multiphase flow in pipes due to its empirical nature. Moreover, it is also observed that while ample work exist between 100mm to 200mm and few above 300 to 500mm currently no work is reported between 200mm and 300mm. This motivation has resulted in generating the experimental data in a large diameter (D_i = 254mm) research facility at Cranfield University, unique of its kind in UK and is the largest diameter vertical upflow setup in use in academia. The facility currently employs air-water as the working fluid because it's easier to build and operate the setup when using air-water than other working fluids that may require special handling with more complicated systems. Additionally, vast air-water two phase flow literature is available, in which case, the current use of air-water as test fluids can lead to more representative results for comparison. However, the large diameter riser facility is built with enough flexibility so that various parametric effects could be studied in future by modifying the setup as well as working fluids.

In this paper, the data generated for large diameter facility is used perform (i) a systematic investigation of the flow patterns in large diameter vertical pipes and identify the transition between

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subsequent flow patterns, (ii) compare it directly with existing large (150mm) and small diameter data (28mm & 32mm) under similar air-water superficial velocities range, (iii) exemplify that existing available empirical correlations/models/codes are significantly in error when applied to large diameter vertical pipe for predictions and (iv) lastly, present a comparison of experimental and simulated results showing the predictive capability of a well known commercial multiphase flow simulator.

II. Experimental Setup

Figure 2 shows the large diameter facility comprising of an air compressor system, water pumping system, horizontal flowline, vertical circular test section, upper plenum consisting of overhead/return tank, downcomer and a return line to sump. The overall height of the vertical circular test section is 12.2m and horizontal flowline is of 36m length, both consisting of inner diameter of 254mm schedule-40 stainless steel pipe sections. The horizontal flowline has one clear perspex section installed at approximately 2m before the base of the vertical circular test section. This transparent section helps in visual observation of the air-water flow exiting the flowline and entering in the circular test section. The vertical riser section has four (4) clear perspex sections installed at different heights for viewing the air-water flow. The water is supplied to the flowline-vertical test section from the single phase loop by a sump pump (P3). The water flow rate is regulated via a valve (VW4) and a bypass valve (VW2) and measured by Bailey Fischer & Porter XM-2000 series electromagnetic flow meter with an experimental accuracy of ±0.5% with minimum and maximum range of 5 to 200 m^3/hr. After leaving the flow meter, water flows into a piping PVC network with the different elevations from the ground and then enters into a 36m long, 254mm inner diameter horizontal schedule-40 stainless steel flowline at the ground level. The air to the facility is supplied from buffer tank to minimize the pressure pulsations from the compressor and is measured by two Fischer-Rosemount MassProbar flow meters (FT302 & FT305) with an experimental accuracy of ±1.3% with minimum and maximum ranges of 79 to 4250Sm^3/hr and 6 to 100Sm^3/hr. The flow to meters is controlled (by control valves FIC301 & FIC302) through DeltaV digital automation system. Air after metering is delivered to large diameter facility via 50.8mm pipe either to the vertical test section base (VA1) or at the inlet to horizontal flowline (VA5). The two phases i.e. water entering into the flowline and air entering either in flowline (through VA5) or in the vertical pipe base (through VA1) mixes and then flows upward into the circular test section. For the results reported here, later air entrance option i.e. air entering through vertical pipe base (through VA1) is used, while former results could be find elsewhere (Ali and Yeung, 2008a). The vertical circular test section contains three special high pressure clear perspex sections of 1m in length installed approximately at 5m, 8m and 10m heights for identifying the air-water flow patterns along the height by the high-speed video camera. A short clear perspex section is also installed at approx. 3m height, just above the air injectors and is used to observe the air injector effect. The air flow from the circular test section is vented to the atmosphere in the upper plenum while water flows into the overhead tank and then to the PVC pipe downcomer. The downcomer offers a flow path either to the sump or return to the circular test section base. In experimental data reported here only former flow path was used. The signals from the instrumentation installed at various locations in the loop were logged through dedicated LABVIEW software (Qazi & Yeung, 2006).

The whole horizontal flowline - vertical circular pipe facility is equipped with flow meters, temperature probes, pressure transducers, differential pressure cell and a water manometer. From air and water flow meters readings, air and water superficial velocities were calculated. The time-dependent pressure signal measurements were performed by flush mounted DRUCK PMP-1400 pressure transducers with the measurement range of 0-2 barg with an accuracy of ±0.15% in full scale near the entrance (base) to vertical test section and at its exit. Another pressure transducer model no. 249 from RS components with the range of 0-6 barg and accuracy ±0.25% in full scale is installed near the exit of the horizontal flowline section before the riser base to monitor the behaviour of incoming two phase flow in the riser base. Three temperature probes were used to detect inlet temperatures of air and water at their respective entrances for accurate estimates of flow. Two DRUCK PMP-4110 differential pressure cells having 0-700mbard range and nominal accuracy of 0.04% in full scale along with a water manometer were mounted in the circular test section to deduce the void fraction at the height of 5, 8 and 10m respectively.

All the pressure sensors are installed close to clear perspex sections so that simultaneous signal acquisition and videoing can take place. The pressure transducers installed at the entrance (in the base) and near the exit of the vertical test section were used to measure the average pressure drop in the test section while the two differential pressure cells and a water manometer were used to deduce the sectional volume average void fraction at different heights along the vertical test section. The differential pressure (D/P) method determines volume average void fraction and is well known for its simplicity, low cost and ruggedness. Many previous researches (Hills, 1976; Tutu, 1984; Matsui, 1984; Anunziato and Girardi, 1984; Matsui, 1986; Ohnuki and Akimoto, 1996) have used this method for void fraction measurements in vertical two phase flows or have cross-calibrated their advanced sensors (impedance, conductivity and wire mesh sensor) against this method (Ma *et al.*, 1991; Fordham *et al.*,1999; Ohnuki and Akimoto, 2000; Shoukri *et al.*, 2000; Cheng *et al.*, 2002; Schlegel *et al.*, 2009). For the purposes of present research work, it was assumed that the differential pressure equals to hydrostatic pressure plus frictional loss and frictional loss could be represented by a correlation. To validate the above assumption, frictional loss was subtracted from measured pressure drop. The frictional loss was determined from the well known correlation of Lombardi and Carsana (1992) that was selected due to its wide parameter range and specifically developed for vertical flows. The **Figure 3** shows that the effect of frictional loss is negligible (within 4% of total) under the given water-air superficial velocities, validating the assumption that the predicted friction loss component was smaller to the extent that it does not influence the total pressure drop that was dominated by hydrostatic component (~ 96%). From deduced volume average void fraction, the probability mass function plots (hereafter referred as probability plots) were constructed to discriminate the air-water two phase flow patterns. Employing probability functions for flow pattern analysis has been documented by a numerous previous researchers and the theory could be referred from them (Jones and Zuber, 1975; Vince and Lahey, 1982; Tutu, 1984; Anunziato and Girardi, 1984; Matsui, 1986; Anunziato and Girardi, 1987; Costigan and Whalley, 1997; Cheng *et al.*, 1998; Cheng *et al.*, 2002; Omebere-Iyari *et al.*, 2007; Omebere-Iyari *et al.*, 2008; Blaney and Yeung, 2008 and Schlegel *et al.,*2009).

III. Results & Discussion

During the above adiabatic air-water flow experimental campaign, both air and water superficial velocities were varied from j_a = 0.09m/s - 2.23m/s and j_w = 0.18m/s to 1.1m/s respectively. The flow pattern identification results presented in this paper are taken at $(z/D)_{air\text{-}water} \approx 32$. The volume average void fraction calculated from this height at various air-water superficial velocities was found to be similar as that of $(z/D)_{air\text{-}water} \approx 20$ and $(z/D)_{air\text{-}water} = 39$ indicating that the flow was fully developed before it reached z/D ≈ 20. This results also agrees well with other large diameter results presented by Omebere-Iyari *et al*. (2007), Omebere-Iyari *et al*. (2008) and Schlegel *et al.* (2009) that found a fully developed flow at $(z/D)_{nitrogen\text{-}naphtha} = 15.58$, $(z/D = 7.7)_{steam\text{-}water}$ and $(z/D)_{air\text{-}water} = 16$ respectively.

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For the results reported here, water flowing path is through horizontal flowline, vertical circular test section, overhead tank, downcomer and return to sump while air entered in the vertical pipe base (through VA1) mixes with water and then flows upward into the circular test section and is vented to the atmosphere in the upper plenum. The water temperature during whole experimental campaign was monitored between 19 - 24ºC; this includes the repeatability runs performed. The existing set-up design did not allow for the experiments in annular and dispersed bubbly flow regime due to undesirable vibrations of upper plenum.

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A. Flow patterns in D = 254mm vertical pipe

A simplified classification is employed to avoid any subjectivity by considering the flow patterns in large diameter vertical pipe air-water upflow to be consisting of following: ***Dispersed bubbly flow***, ***Bubbly flow, Agitated / Clustered bubbly flow*** and ***Churn/Froth flow***. This classification is done intentionally as we planned to clear out the above delineation more clearly in ahead section. The **Figure 4** shows the sketches of the flow patterns along with the associated volume average void fraction probability plots.

Dispersed bubbly flow appeared in few experimental runs at high water and low air superficial velocities (j_a = 0.06 - 0.1m/s and j_w > 0.68m/s) only and was the consequence of low air fraction present as small discrete spherical bubbles, uniformly distributed in continuous water phase. The void fraction probability plot of this flow indicated a thin distinct single peak at low void fraction **(see Figures 4a**). This peak is different than indicated by typical bubbly flow (Jones and Zuber, 1974; Barnea *et al*., 1980; Tutu, 1984; Matsui, 1984; Anunziato and Girardi, 1987; Costigan and Whalley, 1997; Cheng *et al.*, 2002) because a zero void fraction measurement related with pure liquid is also obvious (Vince and Lahey, 1982) to clearly distinguish this flow from bubbly flow.

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Bubbly flow was obtained under low air-water superficial velocities (j_a = 0.1 - 0.3m/s and j_w < 0.6m/s) and was typically encountered when mean void fraction was greater than 0.11. This flow consisted of large population of various sizes bubbles with occasional coalescing of bubbles in the core region to form larger ones during their upward flow. The void fraction probability plot showed a single distinct peak **(Figures 4b)**. However, it can be noted that this peak is different than indicated by dispersed bubbly flow as the single peak is broader than observed for dispersed bubbly flow and is also displaced from the origin. The later observation clearly indicates that in the large diameter vertical pipe upflow condition this type of flow is associated with low liquid input only.

Agitated/ Clustered Bubbly was not observed previously in small diameter vertical pipes. This flow was obtained under the medium air superficial velocities ($0.3 < j_a \leq 1.6$m/s) and was found to be the most dominant flow throughout the large diameter vertical upflow experiments, prevailing in the same region where slug flow occurs in smaller diameter vertical upflow. The apparent distinctions of this flow from bubbly is (i) the difference in the mean void fraction values, (ii) some small bubbles flowing in the core clustering into larger ones and producing a circulatory type of rapid agitation motion in the vicinity and breaking up. This agitation was seen to increase with the increase in air superficial velocities causing smaller bubble population near the wall to move up and down in this agitation. It is emphasized here that this flow did not had any resemblance with spherical cap bubble or typical slug flow found in conventional small diameter pipes, in fact NO large smooth cap/bullet shaped Taylor bubble along with liquid slug were observed under this range of air-water superficial velocities. This observation validates the general consensus (Cheng *et al.*, 1998; Ohnuki and Akimoto, 2000; Shoukri *et al.*, 2000; Pickering *et al.*, 2001; Prasser *et al.*, 2002;

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Omebere-Iyari *et al*., 2007; Omebere-Iyari *et al*., 2008 and Schlegel *et al.*, 2009) of non existence of slug flow in large diameter vertical upflow condition.

The **Figure 4c** shows the void fraction probability plot of this flow pattern, a broad single peak at low void fraction progressively shifting toward higher void fraction with increase in air superficial velocity. This trend of distribution suggests that it possess positive skewness (i.e. right-tailed distribution). This shifting of the distribution towards higher void fraction is also accompanied by broadening of distribution as well as reduction in height. The broadening suggests that bubble size distribution was increasing by break up and reduction suggesting a further coalescence. Since more breakups of bubbles will result in more bubbles and hence more coalescence, therefore equilibrium between coalescence and breakup existed and the overall void fraction distribution shape remain uniform. This gradual shift of the distribution from low air superficial velocities to higher velocities without showing any significant changes to its shape also verifies the visual observation of the gradual and smooth transition from bubbly to agitated bubbly flow. The above statistical analysis presented further proves that **no bi-modal or twin peak** (at low and high void fractions) representing the liquid slug and Taylor bubble (as in conventional small diameter results, refer to Vince and Lahey, 1982; Tutu, 1984; Matsui, 1984; Costigan and Whalley, 1997; Cheng *et al.*, 2002) is observed under these experiments, corroborating the absence of conventional slug flow in large diameter vertical pipe upflow condition.

Churn/froth flow was observed in these experiments at the higher air superficial velocities ($j_a \geq$ 1.6m/s), when the flow gradually transformed from agitated /clustered bubbly flow. Although this flow originated from large group of bubbles clustering and agglomeration, it was unlike the previous flow (agitated bubbly flow) because of its "frothy", extremely chaotic and highly oscillatory appearance. During the flow observation it was observed that within the core region large highly distorted frothy gaseous structures of axial lengths much greater than the diameter of pipe were flowing upwards in the core section of the pipe accompanied by falling and rapidly upward moving liquid film around the periphery. The void fraction probability plot **(Figures 4d)** of this flow exhibited a broad single peak (mean value between 0.55-0.65) with negative skewness (tail extending towards the left i.e. at lower void fractions) clearly unlike the positively skewed (right-tailed distribution) agitated bubbly flow. This long thick tail towards lower void fraction indicates some liquid bridging, a typical characteristic of transitional flow while the broad peak at the higher void fraction represents the gas structures that are long and distorted in nature.

Figure 5 present the above flow patterns results on flow pattern map. In the figure, the flow patterns observed namely; dispersed bubbly (DB), bubbly or low liquid input bubbly (B), agitated / clustered bubbly (AB) and churn/froth flow (C) are shown at their respective locations. As seen in the figure, dispersed bubbly flow is observed at high liquid velocities only where coalescence seems to be suppressed by the liquid turbulence and bubbles flowed upward without any interaction with one another. Bubbly flow prevailed at low liquid input where bubbles though close in separation distance than dispersed bubbly with occasional coalescing travelled upward without any distortion or major secondary motion. The **Figure 5** also indicates that the conventional slug flow pattern has vanished from the flow pattern map. Instead, agitated/ clustered bubbly prevailed in addition to churn/froth flow. The agitated bubbly prevailed for most of condition encountered in this work. Above finding of transition from bubbly to agitated bubbly flow is in line with the visual observation of Ohnuki and Akimoto (2000) and Omebere-Iyari *et al.* (2008), however they classify this flow pattern with other names i.e. churn bubbly and churn slug by Ohnuki and Akimoto (2000 and churn turbulent by Omebere-Iyari *et al.* (2008). The figure also shows that as the air superficial velocity is increased the churn froth flow occurs. In churn flow, the flow was highly chaotic and violent with large gaseous structures travelling in the core region and liquid film travelling upward and downward along the periphery of the wall. The clear distinction of churn flow from agitated bubbly flow was performed on the basis of (void fraction) skewness where positive skewness was sufficed for agitated bubbly flow and change in skewness to negative value represented transition to churn flow. The change is skewness represented the presence of aerated liquid slugs along with large distorted gas structure with liquid film at periphery moving rapidly up and down.

B. *Existing Flow patterns maps with large diamter vertical pipe data*

The above flow patterns and their transitions observed in large diameter vertical upflow condition were compared with the theoretical predictions of two well known flow pattern maps of Taitel *et al.* (1980) and Mishima and ishii (1984) derived from small diameter vertical upflow condition. The purpose of this comparison was to determine the validity of the existing flow pattern transition models against the experimental flow patterns observed. The flow pattern maps in **Figure 6** and **Figure 7** illustrates the comparison between experimental and predicted transitions. In present work, the dispersed bubbly flow is not only occurred at lower air superficial velocities but also at slightly lower water velocity than predicted by Taitel *et al.* (1980). In fact the predictions of Taitel *et al.* (1980) occurred at a water superficial velocity of approximately one order of magnitude higher than experimentally observed transitions. It is be noted here that Costigan and Whalley (1997) in 32mm diameter vertical upflow experiments also found this model transition boundary to be higher than observed in their experiments. Present observation is also consistent with the observation of Chen *et al.* (1997) that at low air superficial velocity range this transition will also be at lower values of liquid superficial velocity unlike the trends suggested by Taitel *et al.* (1980). It is to be noted that Mishima and Ishii (1984) do not delineate any exact distinction between the above two flows (i.e. dispersed bubbly and bubbly flows), see **Figure 7**. It is interesting to note from the experimental results (**Figure 7**) that for large diameter vertical pipes, bubbly flow region became much larger compared with conventional size pipes. Both Taitel *et al.* (1980) and Mishima and Ishii (1984) models underestimate this bubbly-to-slug transition to be occurring at lower air superficial velocities. However both the above transition models are closer to actual transition at higher water velocities only. Thus while both the above flow pattern maps predict an early transition to slug flow from bubbly flow, experiments results indicate that there is NO slug flow (no bimodal peak in probability plot) instead there is gradual transition from bubbly flow to its variation agitated bubbly flow where a coalescence / clustering of bubbles and their break up process is clearly visible along with the local random liquid film movement at the wall. This deviation of the Taitel *et al.* (1980) and Mishima and Ishii (1984) bubble-to-slug transition models is due to not taking into account of diameter of the pipe. An interesting observation related to bubble-to-slug transition is that both the Taitel *et al.* (1980) and Mishima and Ishii (1984) models prediction are closer to experimental results at higher liquid velocities only and deviates at the low liquid velocities. This finding is also consistent with the experimental results of Omebere-Iyari *et al.* (2007) for pipe size of 189mm and of Omebere-Iyari *et al.* (2008) for pipe size of 194mm for which case the Taitel *et al.* (1980) bubble to slug transition model, predicted the similar trend. These trends suggest that while constant critical void fraction approach ($\alpha_{c,\ \text{taitel et al.}} = 0.25$ and $\alpha_{c,\ \text{Mishima \& Ishii}} = 0.3$) is able to predict closer actual behaviour, the approach is limited to higher water velocities only, and at lower water superficial velocity some other mechanism individually or combine with critical void fraction approach is responsible for this transition.

The experimental results and the comparison of the existing slug-to-churn flow transition models of Taitel *et al.* (1980) and Mishima and Ishii (1984) are performed next. The experimental result indicates a gradual shift from agitated bubbly to churn flow with an increase in both water and air superficial velocities near the transition region. Present results in **Figure 6** suggest that the trend

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predicted by Taitel *et al.* (1980) is in contradiction to the experimental trends. It is to be noted that for smaller pipe diameters, Taitel *et al.* (1980) transition curve terminate at slug-to-bubble transition boundary, however with increase in pipe diameter the transition curves will terminate either at: (i) slug-to-dispersed bubbly flow (i.e. reducing the churn flow region) or (ii) slug-to-annular flow thereby concluding that for very large diameter pipes the transition is from slug-to-annular flow and churn flow vanishes completely. However, for present experimental data, Taitel *et al.* (1980) slug-to-churn transition is over predicted. In comparison to above, while the general trend of current experimental boundary is consistent to Mishima and Ishii (1984) slug-to-churn boundary; it appears at significantly lower air superficial velocities (j_a = 1.6m/s) than predicted by this model (j_a = 7m/s). This means that the Mishima and Ishii (1984) model predict a higher slug-to-churn transition upon increase in diameter contrarily to seen with present data. It is to be noted that this experimental observation also corroborates the work of Ohnuki and Akimoto (2000) with 200mm vertical pipe experiments, where this transition occurred earlier than predicted by Mishima and Ishii (1984) slug-to-churn transition model. The annular transition model for both Taitel *et al.* (1980) and Mishima and Ishii (1984) could not be compared as the experiments were not performed in annular flow.

It is clear from above discussion that both Taitel *et al.* (1980) and Mishima and Ishii (1984) flow pattern maps are inadequate for predicting the flow patterns in large diameter vertical pipe upflow conditions as a whole and it is only through understanding of mechanisms involved in individual transition that can provide an appropriate model.

C. Comparsion of Flow patterns in vertical pipe

While a comprehensive and quantitative comparison with other small and large diameter vertical pipe upflow conditions would have added a greater benefit for our researchers, it is of great regret that none of the earlier large diameter (100 > D > 200) studies have conducted experiments in the similar operating conditions (e.g. working fluids: air-water, air-water superficial velocity range) and performed statistical analysis i.e. determining probability distributions functions as done in smaller diameter works (Jones and Zuber, 1975; Tutu, 1984; Matsui, 1984; Anunziato and Girardi, 1984; Matsui, 1986; Costigan and Whalley, 1997). Cheng *et al.*, 2002) to perform a systematic comparison, hence their work could not be included here. However this motivated us to perform a comparison for both cases with current experimental data separately:

(i) Current Large diameter vs. Previous Large diameter studies

While previous systematic and detailed experimental works (Ohnuki and Akimoto, 2000 and Schlegel *et al.,* 2009) could have provided a good comparison, both of these determine flow patterns by either by visualization only or do not characterizes the flow patterns according to air-water superficial velocity range. This leaves the comparison of current work with the work of Cheng *et al.* (1998) only, performed for 150mm diameter pipe for one water superficial velocity only (i.e. j_w = 0.64m/s with j_a = 0.096 to 1.113m/s). **Figure 8** presents some additional large diameter data for the comparison with the 150mm diameter work of Cheng *et al.* (1998).

Figure 8(a) shows the present results at a constant water superficial velocity (j_w = 0.67m/s) with increasing air superficial velocity (j_a = 0.09 to 2.23m/s). As the air superficial velocity is increased at constant water superficial velocity, a number of flow patterns were observed i.e. *dispersed bubbly, bubbly, agitated / clustered bubbly and churn flow*. For same air-water superficial velocity (j_w = 0.64m/s j_a = 0.09 to 1.113m/s) range, Cheng *et al.* (1998) reports the flow as; *uniform bubbly, cap bubbly and churn flow*, refer to **Figure 8(b)**. It is to be noted that Cheng *et al.* (1998) like many previous authors (Mishima and Ishii, 1984; Weisman and Kang, 1981) do not delineate the dispersed bubbly flow from bubbly flow. Current results at very low air superficial velocity ((j_a =

0.09m/s) suggests the flow as dispersed bubbly because of its distinct, sharp uni-modal peak, mean void fraction around 0.07, lying close to origin with no visible coalescence of bubbles. However in current work for $j_a > 0.15$m/s, bubbly flow existed, this flow was unlike dispersed bubbly consisting of various shape bubbles with sporadic coalescence in the core with mean void fraction of 0.15, lying almost at same location where bubbly flow was observed by Cheng *et al.* (1998) corroborating their observation. Note that the mean void fractions defined under both cases are relatively close to each other. Cheng *et al.* (1998) observed the occasional cap bubbles at mean void fraction around 0.13 - 0.17 that increased in numbers with increase in air superficial velocity with the mean void fraction value of 0.24 - 0.4. In present work sustained large distorted type of bubble (unlike cap bubble) in the center of core was seen around the mean void fraction of 0.27 - 0.31, see **Figure 8(a)**. This is because at lower mean void fraction values (0.2 - 0.25) the formed coalescent bubble cluster while rapidly moving upward were seen to disintegrate after travelling to a short distance. It is perceived that Cheng *et al.* (1998) cap bubbly flow is similar to the agitated /clustered bubbly flow in current study. This is also obvious in **Figure 8(b)** results of (Cheng *et al.*, 1998), under similar air-water superficial velocity range (from bubbly to cap bubbly), the trend of the probability plot remained single peaked (Gaussian distribution) with positive skewness (right-hand tailed) throughout j_a = 0.09 to 1.113 m/s, which according to present analysis is the case when the flow pattern is either bubbly and/ or agitated / clustered bubbly. Interestingly this observation validates the flow pattern classification presented in current work also, as both bubbly and agitated / clustered bubbly flow indicated positive skewness. This can be verified from **Figure 8(a),** where it can be seen that for agitated bubbly flow the probability plot remained single peaked (Gaussian distribution) with positive skewness. Cheng *et al.* (1998) designated churn flow, as the air superficial velocity was increased, when mean void fraction value is still low (~ 0.34), single peak Gaussian distribution with positive skewness, refer to **Figure 8(b)**. Cheng *et al.* (1998) did not validate the basis of this designation of churn flow except from visual observation. This later observation is contrary to current experiments, as this air-water superficial velocities range exhibited agitated bubbly flow, whereas it was only after air superficial of $j_a \geq 1.6$ m/s, oscillation in overall flow with highly distorted void structure travelling in the core along with short section of aerated liquid slug were observed. This observation is supported with the probability plot in **Figure 8(a)** that indicates a shift from positive skewness to negative skewness though still the single peak. This change in skewness from positive to negative is sign of flow pattern change unlike stated in earlier work. The void fraction peak at higher void fraction with a tail extending to low void fractions explains the presence of distorted void structure (at high void fraction) and short aerated liquid slug (lower void fraction). This difference in skewness of void fraction in probability plots between the two works is clearly visible in **Figure 8**. The limited air superficial velocity range covered by Cheng *et al.* (1998) does not allow us to compare the results any further. Nevertheless, as seen from **Figure 8**, the two sets of data are generally consistent except for the discrepancy of the flow pattern regarded as churn flow by Cheng *et al.* (1998) that in fact is cap / clustered bubbly flow. This can be verified from the Schlegel *et al*. (2009) 200mm diameter data qualitatively where churn flow exhibited similar trend. Thus, this observation make us conclude that increase is diameter of pipe under vertical upflow condition offer a greater lateral flow path that provides extra degree of freedom to the air-water phases that complicates the flow structure. Under such cases the flow pattern transition from bubbly is not to conventional slug flow but to more gradual transition to agitated/clustered bubbly flow that finally turns into more violent and oscillating churn flow.

(ii) Large diameter (254mm) vs. Small diameter (28 & 32mm)

In this section we present some more supplementary large diameter data for the comparison with small diameter vertical upflow results. Such comparison of flow behvaiour in large diameter against small diameter vertical upflow under similar conditions has not been performed before. The results reported here are for constant water superficial velocity (j_w) of 0.38m/s when air superficial velocity (j_a) was increased gradually from 0.09 to 1.74/s to obtain different flow regimes. In above air-water superficial velocity range, the observed flow patterns for large diameter were *bubbly, agitated / clustered bubbly and churn flows,* see **Figure 9(a)**.

It is to be noted that the above air-water superficial velocities conditions are similar to those of Cheng *et al.* (2002) and Costigan and Whalley (1997). Cheng *et al.* (2002) data was taken for 28mm diameter pipe for water-air superficial velocity ranges of j_w = 0.35m/s and j_a = 0.0798 - 0.5m/s while Costigan and Whalley (1997) also lying in similar air-water superficial velocity range were taken in 32mm diameter vertical pipe for water superficial velocity of 0.35m/s and air superficial velocities of 0.1, 0.2, 0.5, 2.0, and 4.0 m/s. The **Figure 9(b)** and **Figure 9(c)** shows the flow patterns identification results of both these (Cheng *et al.*, 2002; Costigan and Whalley, 1997) small diameter vertical pipes data uses air-water as two phase gas-liquid test fluids, hence are more representative for making meaningful and reliable comparison.

Cheng *et al.* (2002) reported four (4) flow patterns in 28mm diameter vertical upflow, namely; *(i) discrete bubbly, (ii) clustered bubbly, (iii) cap bubbly and, (iv) slug flow,* refer to **Figure 9(b)** . For similar water superficial velocity range, the flow pattern identification results of Costigan and Whalley (1997) indicated six (6) flow patterns in 32mm diameter vertical pipe upflow, refer to **Figure 9(b)**: *(i) discrete bubbly, (ii) spherical cap bubble, (iii) slug, (iv) unstable slug, (v) churn flow and (vi) annular flow*. A satisfactory agreement exists in both small scale pipe works for the flow called as discrete bubbly; see **Figure 9(b)** and **Figure 9(c)** as similar narrow peak at low void fraction (0.08-0.14) is found in probability plots. The current results shown in **Figure 9(a)** of large diameter shows that in this region a bubbly flow (or low liquid input bubbly flow) is obtained with its corresponding void fraction probability plot, also exhibiting a thin distinct peak at lower void fractions. Though the small scale results define this as discrete bubbly where it exist as uniformly distributed bubble without bubble agglomeration or coalescence, it does not show any significant difference in mean void fraction from the bubbly flow found in large diameter. Thus it can be concluded that scale effect causes a difference in the region where existing work show bubbly (or low liquid input bubbly flow, $j_w = 0.38$m/s, $j_a = 0.09 - 0.19$m/s) flow while at the similar location ($j_w = 0.35$m/s, $j_a = 0.0798 - 0.2$m/s) discrete bubbly flow was observed by (Costigan and Whalley, 1997) and Cheng *et al.* (2002). This further indicates that although the discrete/ dispersed bubbly flow occurs for both scales of pipes, under similar gas superficial velocities, for large diameter it is found at higher liquid superficial velocities ($j_w \geq 0.68$m/s and $j_a = 0.06 - 0.11$m/s). Alternatively it can be concluded that the critical superficial liquid velocity for transition to dispersed bubbly flow increases with an increase in pipe diameter, hence all the regime transition models taking account of diameter will predict this transition more accurately. While the large diameter vertical upflow experimental runs in air superficial velocities range ($j_a = 0.09 - 0.25$m/s) reports bubbly flow with increasing agglomeration of bubbles in the core region, the small scale pipe results of Cheng *et al.* (2002) indicates an increasingly clustered bubbly flow, see **Figure 9(b)** and (Costigan and Whalley, 1997) does not report an observation for this range, see **Figure 9(c**). This delineation of clustered bubbly as a separate flow regime from cap bubbly flow complicates the description of flow patterns unnecessarily. It is assumed that the distinction of clustered bubbly ($j_w = 0.35$m/s, $j_a = 0.132$m/s) from cap bubbly in Cheng *et al.* (2002) work evolves from the fact that this work was primarily determining the flow regime transition hence much closer gas superficial velocity data points were collected whereas in (Costigan and Whalley, 1997) work the objective was flow pattern recognition hence data was much sparser in the region (i.e. $j_w = 0.35$m/s $j_a = 0.1$m/s for bubbly & 0.2m/s for cap bubbly). However, from the definition provided by former work for clustered bubbly and currently reported bubbly flow, it seems very likely that this discrepancy is due to semantic and the current bubbly flow near transition and clustered bubbly are very much similar in nature. A good agreement between both (Costigan and Whalley, 1997; Cheng *et al.*, 2002) work is found for spherical cap bubbly (or cap bubbly) flow under similar velocities (j_w = 0.356m/s, j_a = 0.154m/s and j_w = 0.35m/s, ja = 0.21m/s), refer to **Figures 9(b)** and **9(c)**. Note the mean void fraction for spherical cap bubbly (or cap bubbly flow) are almost similar in both cases (~0.2) and both indicate a single peak with a forward marching tail in probability plot, this tail becomes more and more pronounced with increase in air superficial velocity and later develops into a second peak. The spherical cap bubbly region seems to persist till $j_a \approx$ 0.5m/s after which in both works (Costigan and Whalley, 1997; Cheng *et al.*, 2002), Taylor bubble of slug flow was detected, see **Figure 9(b)** and **9(c)** where two distinct peaks in probability plots are apparent. Under the above condition, though the current large diameter results (j_w = 0.38m/s, $j_a \leq$ 0.25m/s) still indicates bubbly flow but with occasional bubble clustering / coalescence and formation of large distorted bubble in the core but there is no spherical cap bubble formation (mean void fraction around 0.21). With increasing air superficial velocity (j_a > 0.25m/s) more bubbles agglomerate to form large bubble clusters of distorted shapes in the core region along with visible secondary motion due to coalesce and breakup processes. The **Figure 9(a)** shows the corresponding void fraction probability plots, exhibiting agitated bubbly, a single peak but with a much broader distribution than seen in bubbly flow representing the wider bubble sizes due to coalesce and breakup processes. The mean void fraction during this flow remained between 0.27 – 0.41. Note that during the above increase in air superficial velocities while the void fraction probability plots remain single peaked, the skewness of the distribution was also positive. The probability plots above reports the absence of conventional slug flow in large diameter vertical pipe upflow condition as the slug flow is not visually observed neither the bimodal peak associated with it in void fraction probability plots, see **Figure 9(a)**, instead the void fraction probability plots of this (agitated bubbly) flow remained single peak, Gaussian in nature with positive skewness (right-hand tailed). In comparison to large diameter results, the slug flow was found in both small diameter works (Costigan and Whalley, 1997; Cheng *et al.*, 2002), refer to **Figure 9(b)** and **Figure 9(c)**. The slug flow exhibits a bi-modal (twin) peaks in its probability plot at high and low void fractions representing the Taylor bubble and aerated liquid slug respectively. Although (Costigan and Whalley, 1997) identified unstable slug flow in 32mm diameter experiments (j_w = 0.35m/s j_a = 2m/s) refer to **Figure 9(c)**, Cheng *et al.* (2002) did not observe the unstable slug flow due the limited test matrix of air-water superficial velocity range, refer to **Figure 9(b)**. Thus the observation of unstable slug flow pattern appears to be unique to the work of (Costigan and Whalley, 1997). Note in **Figure 9(c)**, results of (Costigan and Whalley, 1997) still shows two distinct or bi-modal peaks (as in slug flow) but the magnitude of peak at lower void fraction has reduced, this is because the liquid portion is no longer dominant (as in slug flow) and distorted highly twisting air structures formed have longer length, high travelling velocity causing a break-through the liquid slug with overall flow more chaotic than typical slug flow. According to (Costigan and Whalley, 1997) the unstable slug name conveys the description that not all of the slug structure of the flow has broken down, and some individual liquid slugs may appear occasionally. In comparison to above smaller diameter behaviour, the large diameter vertical upflow experiments upon further increase in air superficial velocity (j_a > 1.4m/s) at constant water superficial velocity (j_w = 0.35m/s) indicated transition towards churn flow. This flow was characterized by a single thick peak at higher void fractions (mean void fraction around 0.46), with distribution that is shifting to negative skewness. This negative skewness indicates shift in distribution from right to left (left-hand tail) extending towards lower void fractions, refer to **Figure 9(a)**. It is to be noted that for further increase in air superficial velocity ($j_a \leq$ 2.23m/s), experimental results for large diameter pipes indicated similar single peak with negatively skewed probability distribution. Due to limited test matrix of Cheng *et al.* (2002), further comparison could not be done but (Costigan and Whalley, 1997) results clearly validates our churn flow behaviour as they too found churn flow to possess a negatively skewed probability distribution, refer to **Figure 9(c)**. However note that in current large diameter vertical upflow work, the churn flow appears at lower

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air superficial velocities ($j_a \approx$ 1.6m/s) in comparison to (Costigan and Whalley, 1997) results ($j_a \approx$ 3.5m/s). Thus it can be concluded that while churn flow occurs for both small and large scale pipes, for large diameter pipe it is found at lower air superficial velocities i.e. $j_a \sim$ 1.6m/s then observed in small scale work of Costigan and Whalley (1997) that was found after 3m/s. Hence critical gas superficial velocity for transition to churn flow decreases with an increase in pipe diameter. The annular flow was not encountered in current work due design limitation of upper plenum of the setup but annular flow was detected by (Costigan and Whalley, 1997) at air superficial velocity ($j_a >$ 8 m/s).

Overall, there exists a reasonable agreement between small scale pipes and current large diameter results for bubbly and churn flow with the exception of their transition boundaries, the distinction arise in slug flow which seems to exist only for small diameter piping. In similar experimental conditions (i.e. under same working fluids and superficial velocity ranges) as of slug flow in small scale pipes, the large diameter pipe indicates agitated / clustered bubbly flow. This new flow pattern thus has different visual characteristics than the slug flow and is the most dominant flow in the experimental range extending over the range of that of conventional slug flow. Above behvaiour further suggests that for agitated / clustered bubbly flow, changes in the two phase flow characteristics (void fraction and pressure drop) are expected. Comparing the above flow scenario, it can be inferred that there are differences in the flow behvaiour in large and small diameter pipes under vertical upflow conditions due to the enlargement of flow path as agitated bubbly and churn/froth flow are more dominant. This means that the presence of agitated / clustered bubbly and its gradual transformation into more violent churn/froth flow is a consequence of extra degree of freedom of the two phases. This degree of freedom is small in conventional diameter pipes owing to small lateral flow path. Hence it can also be deduced that under above large diameter vertical upflow scenario, one dimensional (1D) modelling approach is not suitable as secondary flow effects exist and this may affect the hydrodynamic behaviour. However, so far, no work reports about the effect of this secondary flow in agitated bubbly flow.

D. Void fraction correlation - comparsion:

Prediction of average void fraction in two phase flows is highly significant as it plays a fundamental role in characterizing the distribution of the phases within the system, especially in the determination of the amount of liquid phase (holdup) retained in a system. The later aspect is indeed a crucial issue in primary heat transport systems in nuclear reactor in case of accidents and in flowline-riser system in offshore oil & gas platform. The designer needs a void fraction correlation as a closure relation to predict the two phase flow system behaviour before designing the actual system and/or simulating scenario related to that system. Thus the designing and/or reliability of any two phase flow system is dependent upon the prudent choice of the void fraction correlation used. There are considerable numbers of void fraction correlations belonging to different multiphase flow industries but there is also a considerable difference in the predictions of these correlations. The volume average void fraction data forms an excellent basis for the assessment of the predictive capability of these voids fraction correlations. Moreover, no study has so far included the assessment of void fraction correlations with respect to their applicability to large diameter vertical pipe upflow. The assessment is also important in order to determine the implications of the different flow patterns occurring in the large diameter and the conventional small diameter vertical pipe.

A large number of correlations were assessed against the experimental data of volume average void fraction (Ali and Yeung, 2008b), however here only results of selected correlations belonging to the following categories are listed: Homogenous mixture void fraction model, Separate flow model (using models based on slip ratios (j_a/j_w), based on Lockhart and Martinelli parameter (X), based on mass flux (G) and based on drift flux model), and some miscellaneous correlations (few empirical/specific flow regime correlations which do not specifically belong to either of the above categories).

Under the flow conditions of the current experiments, the total pressure gradient was dominated by hydrostatic head. This implies that in the experiments friction component was smaller to extent that it does not influence the total pressure gradient. For real conditions like in present analysis, the choice of two-phase void fraction correlation is of major significance in determining the hydrostatic pressure gradient. The results of the void fraction assessment presented in **Figures 10 to 12** indicates that many of the published correlations are not appropriate to characterize the void fraction in large diameter vertical pipes and only few have potential to perform satisfactorily.

Under current experimental conditions where flow regimes encountered were dispersed bubbly, bubbly, agitated bubbly, and churn flow regimes, the homogenous mixture model exhibits greater accuracy at very low void fraction only i.e. when the flow regime was bubbly and dispersed flow and progressively deviates with flow regime transition due to significant interphase slip between gas and liquid phase, thus the predicted values show high over prediction (52.67%), refer to **Figure 10(a)**. It is to be noted that the model is independent of the diameter of the conduit and thus can be applied easily to conditions of two phase flows (of different fluid properties) as a starting point where dispersed bubbly or dispersed droplets (or mist flows) are likely to be encountered.

In the "Kβ" void fraction correlation category (special case of homogenous mixture model) the Bankoff correlations (Neal and Kazimi, 1989) seems to improve the homogenous mixture model results by taking into account of radial non uniformity of void fraction and velocity difference (slip) between the two phases, performed satisfactorily, and should therefore be considered. From the comparison of the predicted values with experimental values, improvement in mean percentage errors (16.46%) can be seen in **Figure 10(b)**.

The void correlations based on separate flow model using slip (j_a/j_w) equation, mass flux (G) term or Lockhart-Martinelli parameter (X) showed wide variation in results. The analytical correlation of Chisholm (Chisholm, 1972) based on the velocity ratio (slip) and applicable to any fluid indicated an overall an under prediction (-28%) with largest deviation in bubbly and agitated bubbly flows **Figure 10(c)**. The void fraction correlation predictions of Lockhart and Martinelli (Butterworth, 1975) under predicted within -24% range, **Figure 10(d)**. The model gives lower mean percentage error prediction for bubbly flows and increasingly deviates for higher void fractions (agitated bubbly, unstable slug and churn flow).

Satisfactory accuracy was indicated in cases of mass flux (G) dependent void fraction correlations taking into account of diameter. Guzhov *et al.* (Garcia, 2005) correlation defines the holdup and no slip ratio dependence on Froude number which in turn based upon mixture volumetric flux and pipe diameter. The correlation shows an overall mean error of +4%, with successful application in bubbly flow and a slight over prediction for other flows, see **Figure 11(e)**. Another correlation by Premoli *et al.* (Premoli *et al.*, 1970), also known as CISE correlation is considered to be valid for wide range of data and uses a slip ratio, $s = f(x, G, \rho_l, \rho_g, \mu_l, \mu_g, \sigma, Q_l/Q_g)$ performed satisfactorily with overall mean error of performance of +10%. The correlation yields closest at intermediate void fraction than at low and higher void fraction, see **Figure 11(f).** It is to be noted that whenever approximate averages are required or for conditions where the flow type may not be known, Guzhov *et al.* (Garcia, 2005) and Premoli *et al.* (Premoli *et al.*, 1970) void fraction correlations are recommended on the basis of their simplicity and closer prediction for the experimental data analyzed.

Overall the drift flux based void fraction correlations were more successful in predicting the closer results to the experimental values with predicted values within ±5% for Chexal and Lellouche (Chexal and Lellouche, 1992), Hibiki and Ishii (Hibiki and Ishii, 2003) and Kataoka and Ishii (Kataoka and Ishii, 1987).

The Chexal and Lellouche (Chexal and Lellouche, 1992) correlation is a versatile correlation, validated against large data bank (consisting of full range of pressures, 1-145bars, mass fluxes, 0.01-5500kg/m^2-s, void fractions, 0.01-0.99 and various diameters, 0.005-0.456m) and applicable to all orientation flows, in all flow regimes with steam-water, air-water, and refrigerant two-phase flows. The void fraction predictions of this correlation for large diameter are within -5%, **Figure 11(g)** with an under prediction at low void fractions. Perhaps the most detailed and comprehensive study on drift flux modelling of large diameter pipe has been performed by Hibiki and Ishii (Hibiki and Ishii, 2003). Two inlet flow-regime (bubbly/cap bubbly at inlet or dispersed bubbly at inlet) based correlations were developed for large diameter pipe at low mixture volumetric flux. For higher mixture volumetric fluxes, the use of Ishii (Neal and Kazimi, 1989) and Ishii and Kataoka (Kataoka and Ishii, 1987) correlations are recommended. Excellent predictions with an overall mean error of 1.75% are obtained by these correlations in comparison to the other correlations used; see **Figure 11(h)**. It is to be noted that this correlation was able to predict the dispersed bubbly and bubbly flow regimes very well, a feature not exhibited by many other correlations. The Kataoka and Ishii (Kataoka and Ishii, 1987) drift flux correlation developed for the large diameter apparatus under pool conditions is a function of hydraulic diameter, density ratio and viscosity number. The modeled equations are for cap bubbly flow only and for other flow regimes (bubbly and churn flow regime), the use of Ishii drift flux model (Kataoka and Ishii, 1987) is recommended. The result of predicted values of void fraction vs. measured values shows a good agreement with an overall mean error of about 1.55%, see **Figure 12(i)**. It is to be noted that in the above correlations Hibiki and Ishii (Hibiki and Ishii, 2003) and Kataoka and Ishii (Kataoka and Ishii, 1987) are specifically develop for large diameter application. It can further be noted that with lowest mean percent error Kataoka and Ishii (Kataoka and Ishii, 1987) outperforms other empirical correlations while on the basis of standard deviation, the figures clearly shows that the Hibiki and Ishii (Hibiki and Ishii, 2003) correlation performs the best. Although the drift flux correlations are found to closely predict the experimental data, three (3) constraints are met by them (i) expressions are flow regime dependent hence are not continuous and this might give rise to numerical instabilities during computation (ii) because the models are sensitive to prediction of flow patterns, any inappropriate choice of flow pattern would increase the variance of the whole model and, (iii) many of the above correlations are iterative in nature which inhibits their frequent use in comparison to simpler correlations.

The empirical based void fraction correlation/pressure gradient methods of the oil industry showed inconsistencies excluding the Duns and Ros (Brill and Mukherjee, 1999). The Duns and Ros (Brill and Mukherjee, 1999) correlation is very popular in Oil industry and is specifically developed for gas-liquid vertical flow mixtures in wells. In current assessment some cases of bubbly flow regime were predicted correctly while in rest of the cases of agitated bubbly and churn-flows are all identified as slug flow, refer to **Figure 12(j)**. With an overall mean error +15%, the results can be considered to be satisfactory in comparison to the other methods of this class. Beggs and Brill method of pressure gradient (Brill and Mukherjee, 1999) applicable to any inclination indicated a mean percent error of 21.42%. The method calculates pressure gradient but identify the flow regimes (segregated, transition, intermittent and distributed) first and then liquid hold up for horizontal orientation which is corrected for actual pipe inclination. The method predicted the bubbly and some cases of agitated bubbly flow as transition flow while the cases of churn were identified as intermittent flow. The scatter seen in the **Figure 12(k)** is due to incorrect flow regime predictions.

OLGA (OLGA, 2007) is an extensively used multiphase simulation tool of oil and gas industries that has been specifically developed for large diameter (189mm) risers. For evaluation of void fraction in the test section, the OLGAS (steady state) model was used, the values indicated are average void fraction in the vertical riser test section. The inlet and outlet sections of the vertical riser geometry were excluded (to minimize the effects of adjoining node) from the calculation of average void fraction. The comparison of the experimental and OLGA results clearly indicates the differences, the results are around +30% in mean error, see **Figure 12(l)**. The over prediction of void fraction indicates a lower pressure drop prediction than true value which is quite an offset from designing point of view.

The results of the void fraction correlation assessment presented (in **Table I**) indicates that many of the published correlations are not appropriate to characterize the void fraction in large diameter vertical pipes and only few have potential to perform satisfactorily within the range (±30%). The important implication of this assessment is that two phase flow void fraction prediction should be based on flow pattern prediction and should include the diameter effect. The correlations taking into account of this fact are closer to experimental trends. It was noted that most of the correlations performed well in some of the flow regimes and their performance deteriorated in the other thus none of the correlation was able to predict all the flow regimes accurately. It was found that correlations that successfully predicted the agitated/ clustered bubbly and churn flows did not predicted the bubbly flow accurately and vice versa. This trend highlights the difference in the flow structure variation behind the bubbly and rest of the intermittent flows. Thus in the conditions where the prevailing flow pattern of the two phase flow is known, prior to design/simulation stage, the selection of the appropriate void fraction prediction will be closer to the true value then the value of randomly selected correlation/equation.

E. Experimental results vs. Simulations:

To further support the assertion that some of the presently used multiphase modelling tool gives questionable predictions of hydrodynamic parameters in large diameter vertical upflow conditions due to lack of experimental data, an effort was made to compare current results with a commercial software OLGA. The code is based on one-dimensional (1D), extended two-fluid model and is available as steady state processor and as the complete two-phase transient computational code with full functionality (Bendiksen *et al.*, 1991). The idea behind was to explore the predictive capability of extensively used OLGA multiphase flow simulation tool of Oil and Gas industry.

The large diameter facility was modeled as simplified horizontal flowline-vertical riser with short horizontal pipe at the end of the vertical rise (refer to inset in **Figure 2)**. Use of the short horizontal pipe at the end of vertical pipe is to avoid numerical instability during the simulations. It was further assumed that the diameter of the flowline-riser is constant with standard carbon steel properties. Three grids were implemented however, here results of one of the grid is presented; horizontal flowline of 36m divided into 40 sections (0.9m) with vertical pipe or riser modelled as 13 sections (0.9m) and short horizontal pipe at the top as 2 sections (1.5m). The PVTsim fluid property simulator was used for characterizing the air-water properties with air treated as an ideal gas. As the experiments were conducted at ambient conditions, no heat and mass transfer between the phases and the environment was assumed in simulations. A minimum runtime period of 2 hours was used for simulation. To ensure that proper convergence with appropriate time step has been achieved, the minimization of global volume error in the system was imposed. The cases discussed, uses the steady state pre-processor in the code for generating the initial values for transient simulations unless otherwise stated. OLGA version5.1 was used in this work and "Slug tracking module" was not used for intermittent cases as numerical stability problems were encountered when slug tracking option was "ON" in the simulations performed. The further details of the modelling assumptions, material specifications, boundary conditions and runtime conditions are given elsewhere (Ali and Yeung, 2010).

For initial model depicted as A1 & B1 in **Figure 13** and **Figure 14** respectively, steady state average flow rates of air and water at inlet and, constant pressure at the outlet, were imposed as the boundary conditions. A wide range of the air-water superficial velocities was covered in experiments and the flow regimes encountered ranged from stable bubble flow to highly intermittent churn/froth flow. The parameters investigated in the simulations were flow regime, flowline/riser base pressure characteristics and holdup (void fraction). However, here only results of flowline exit and near riser base pressure are presented. The methodology used was to test some stable flow cases (e.g. bubbly flow normally encountered and preferred type of flow in offshore oil-gas fields to uplift the two phase mixture) to provide confidence on the results of the code and then test some unstable flow cases (where the flow is unstable like slugging which results in offset of downstream processing facility in offshore oil-gas fields). Here, the results of only two (2) selected cases consisting the bubbly flow, and churn/froth flow in the vertical test section are presented (labeled as A & B). The later case of churn/froth flow is selected because under this condition the horizontal flowline (joining in vertical pipe base) was experiencing slugging flow. The two simulated cases in terms of superficial velocities of air and water (Case A, j_w = 0.50m/s; j_a = 0.18m/s and Case B = j_w = 0.32m/s; j_a = 2.17m/s). Before presenting the results of the simulation a brief description of the experimental flow behvaiour of the two above cases A (dispersed bubbly) & B (churn/froth flow) in the riser section are presented.

Case A: In case A, a stable plug flow was observed in the horizontal flowline which upon entering the vertical riser section via 90º bend connecting flowline and vertcial riser appeared as puredispersed bubbly flow prevailed in the riser section. This case A is of the high liquid superficial velocity with dispersed bubbly flow at the inlet to the vertcial test section, whereas the flow pattern in the flowline was plug flow. In contrast to prevailed flow patterns above, simulated flow pattern by the code for this case (A1) is stratified flow in the flowline and the bubbly flow in vertcial test section. The bubbly flow pattern in the riser is simulated by the code in this case correctly. **Figure 13** shows the simulated flowline exit and riser base pressure prediction for this case (A1). A stable profile is indicated over lying on the experimental trends of a dispersed bubbly flow. The simulated riser base and flowline pressure (case A1) is around 0.907bar in comparison to 0.912bar in the experiments (case A). The results are in good agreement with the experimental results concluding that initial model with the assumptions of steady state average flow rates of air and water at inlet and, constant pressure at the outlet adequetley defined the stable flow case.

Case B: As a second example case B was taken, this particular case is chosen to demonstrate the initial model's ability to simulate the unstable cyclic nature of the flow as observed in the experimental conditions. The case is for low superficial water velocity with highest superficial air velocity (j_w = 0.32m/s and j_a = 2.17 m/s). In fact this particular flow is a continuation of slugging flowline that had started occurring at slightly lower air velocity but for this highest air velocity it appears more periodic, refer to **Figure 14(a)** case B. According to the horizontal flow regime map (Taitel *et al.*, 1976), no slug should be present under the flow condition of the case B; however slugs were formed at these low water superficial velocities. In context of current experiments their presence (slugs) is attributed to the upstream and downstream topology of the horizontal flowline i.e. the elbow connecting horizontal flowline to the vertical riser (or pipe). It postulated that this terrain influence allows slugs to be present for the flow conditions which normally would result in stratified flow. This configuration caused the accumulation of water due to slowing down of the water and thereby initiating slug formation in stratified flow regime. The near riser base pressure response (case B) of this flow regime is indicated in **Figure 14(a)**. The figure shows the regular arrival of slug near the exit of the flowline with slugging changing to almost periodic sinusoidal type, regularly varying with minimum and maximum pressure of 0.3 to 0.6 bar and with cycle time of 10-12s. Notice two aspects first; the cyclic behaviour with slug build-up which corresponds to an increase in vertical riser section-flowline pressure due to water build-up in the vicinity; (however this water build-up is small) and second, there is no slug production period. During this partial build-up, air in the flowline is being temporarily compressed, moves in towards riser base with high velocity pushing the short accumulated water pool ahead in the vertical riser section. In this situation since the air superficial velocity is quite high compared to the water superficial velocity, consequently the water fall back is not enough to block the base completely for long and there is continuous air penetration as suggested by the constant pressure variation in the flowline. Similar pressure cycling behaviour is also observed by other researchers in smaller diameter horizontal flowline-vertical riser configuration (Schmidt *et al.*, 1980 and Fabre *et al.*, 1990). However the unique aspect of this case, in comparison to the earlier work is that liquid slug pushed up by the available air drive is dissipated completely or partially by the air, turning to churn/froth type of flow in later vertical sections. The flow appeared to be violent i.e. highly distorted bubble clusters travelling upward with high velocity. Finally, the churn frothy mixture sloshed out of the riser in the plenum with some of the liquid falling back on the upcoming flow in the riser.

The simulation results of case B by OLGA code predicts hydrodynamic slugging in the flowline and annular flow in the riser section. Thus OLGA code also indicates that slugging in flowline appears to occur at lower air superficial velocity than indicated by Taitel *et al.* (1976) flow pattern map. Above behaviour of the code indicates that in case of the slugging in the flowline, the code is completely dissipating the created liquid slug in the base vicinity making annular flow ahead in the later sections. Although situation does corresponds to experiment slightly but slugs formed in the flowline were observed to dissipate while travelling upward in the riser in experiments. This is because the code does not classify churn/froth flow as an individual flow (OLGA only identifies bubbly, slug and annular flow under vertical upflow condition), therefore the transition from slug flow is to annular flow. The simulated flowline exit and near riser base pressure responses (case B1) from the code along with the experimental results (case B) are shown in the **Figure 14(a)**. In the figure, the pressure profile appears quite stable for both flowline and the riser base. The simulated mean pressure is around 0.208bar in comparison to mean value of 0.413bar from the experiments. It is to be noted that the flow during the experiments remained highly intermittent (0.186 – 0.714bar minimum and maximum) in comparison to simulation. This behaviour is attributed to the incorrect flow regime predictions in vertical riser, where the code predicts the case as annular flow.

OLGA preliminary model (abbreviated as A1 or B1 in **Figure 13 & Figure 14**) has been applied and the result of the two cases; one dealing with the stable bubbly flow and other related to unstable churn/froth flow were presented. Based on the results of first model it is obvious that while the code did predict the stable flow case A satisfactorily, as there is good agreement between simulated and experimental values, it was unable to predict the unstable flow case B accurately. The case B belongs to the churn/froth flow but due to incorrect flow regime (annular flow) predictions, the code predicts the case as stable flow. In fact the code, globally under predicts average riser base pressure in the unstable flow case B. Thus the average, maximum and minimum pressures in case B is also lower than actual values in experiments. This implies that the code under predicts the water inventory at the base. This later aspect motivated some modification to improve the model results for unstable flow cases. Many attempts were made in modifying the first model including the fine tuning of friction factors and varying the constant outlet boundary pressure however, these modifications did not improve the results. Thus, it was concluded that the possible option was to modify the boundary conditions, because during unstable flows, the boundaries of the system are most affected. A satisfactory alternative to this was to use the sensor time series as the boundary conditions. Since the upper plenum was open to atmosphere, therefore sensor response near the exit of the riser was used instead, while all other conditions were kept same. Admittedly, this change does bring the mean pressure in the simulation and the riser base slightly close to each other as a consequence of the head imposed (≤0.09bar). However with this change, we will still be able to examine/verify the riser base pressure trends. Thus if positive results are obtained, it at least indicates that the code is capable of capturing the dynamics of typical unstable flow phenomena in large diameter horizontal flowline-vertical riser system.

Modified Case B: The results from the extended model for the case B are presented in **Figure 14(b)**. The code still incorrectly predicts the flow regimes in flowline-riser, similar to the first model. However, the interesting change in the simulated results was the riser base pressure trend indicating the similar unstable flow behaviour with oscillations as observed in experiments (see **Figure 14a**). In comparison to the mean riser base pressure of the first model (0.208bar) and the experimental value (0.413bar) (also see **Figure 14b**), the mean riser base pressure predicted by this extended model is 0.272bar. This is the consequence of the head imposed. The predicted slugging frequency in simulation is almost similar as the experimentally observed however slugging cycle is underestimated by the code. This under prediction of the slugging cycle is accompanied by the underestimation of the slugging amplitude. Both the factors are attributed to the underestimation of the liquid holdup by the code. From the figure one can notice that the simulation indicates that there is continuous large amount of air penetration in the riser base with stratified flow in the flowline.

Both the cases (A and B) have been checked for the grid and time step independency. To ensure that a grid independent solution was achieved; three different grid sizes of 0.9m, 1.8m and 3.6m were used. Similar to above grid resolution study, a time resolution study was performed with various time steps. It was seen in simulations that the too large time step in the code resulted in an increase numerical diffusion causing large global volume error and degraded accuracy. So trials were performed with various time steps and finally the initial time step of 0.001s was manipulated from $\Delta t/10$, Δt, $10\Delta t$ in order to keep the deviation to the minimum. **Table II** summarizes the results of two cases A and B.

IV. Conclusion

An adiabatic air-water flow experiments have been performed in 254mm diameter pipe under vertical upflow condition for air and water superficial velocities of j_a = 0.09m/s - 2.23m/s and j_w = 0.18m/s to 1.1m/s respectively with mean void fraction ranged between 0.11 – 0.65. The flow patterns were also identified in above air-water superficial velocity range using visual observation and statistical analysis of sectional void fraction at $z/D \approx 32$. Four basic flow patterns were characterized namely; (i) dispersed bubbly flow, (ii) bubbly flow (at low liquid input), (iii) agitated / clustered bubbly and, (iv) churn / froth flow and an experimental flow pattern map was developed. It was found that while conventional small diameter pipe and large diameter pipe exhibit similar dispersed bubbly, bubbly flow and churn flow, the distinction arises in slug flow which seems to exist only for small diameter piping. No typical slug flow (consisting of Taylor bubble & liquid slug) is observed in current large diameter vertical upflow experiments instead agitated/ clustered bubbly prevailed in the region where a coalescence / clustering of bubbles and their break up is clearly visible along with the local random liquid film movement at the wall. Thus the conventional slug flow pattern (found is smaller diameter pipe) has vanished from the large diameter pipe under vertical upflow condition. Furthermore, dispersed bubbly flow and churn/ froth flow were observed to occur at lower water superficial velocity than in conventional small size pipe. From these results presented it can be inferred that this difference of flow behvaiour exhibited by large diameter pipe under vertical upflow condition is due to the enlargement of lateral flow path that provides extra degree of freedom to the two gas-liquid phases that complicates the flow structure and suppresses the formation of conventional bullet shaped Taylor bubble. The detailed comparison between flow patterns encountered in small diameter and large diameter vertical upflow condition also support above assertion. The comparison of the experimental results with other air-water work on large diameter vertical pipe indicates a close agreement. However, none of the flow pattern maps were able to predict the flow transitions encountered in large diameter vertical upflow condition satisfactory, putting a question mark to their applicability for these diameters.

The twelve void fraction correlations assessment indicated that void fraction correlations should be flow regime dependent. The correlations taking into account of this fact are closer to experimental trends. None of the correlations successfully predicted all the flow regimes encountered in experiments. It was found that correlations that successfully predicted the agitated / clustered bubbly and churn flow, did not predicted the bubbly flow accurately while those predicting bubbly flow showed little acceptable trend for agitated bubbly but completely deviated for churn flows. This trend highlights the difference in the flow structure variation behind these flows.

The results from numerical model clearly indicate that the effects of boundary conditions on the model are substantial. Whilst stable flows have been satisfactorily modelled with steady state average boundary conditions in OLGA, this practice was insufficient for determining the real behaviour in unstable flows in large diameter horizontal flowline-vertical riser rig. This signifies that the real behaviour of unstable flows was dominated by large transient variations at the boundary. In numerical simulation, inconsistencies were found in the prediction of flow regimes and liquid holdup, along with the under prediction of the riser base pressure. The OLGA was partially successful in qualitatively reproducing the trends and was still unable to quantitatively predict the unstable flows in large diameter horizontal flowline-vertical riser system.

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REFERENCES

Ali, S. F and Yeung, H. (2008a). "Effect of upstream conditions on the two phase flow in the large diameter vertical pipe". Presented at the 6th North American Conference on Multiphase Technology, Banff, Canada: 4-6 June 2008.

Ali, S. F. and Yeung, H. (2008b). "Performance assessment of void fraction correlations in large diameter vertical pipe upflow". In the Proceedings of 11th International Conference on Multiphase Flow in Industrial Plant, Palermo, Italy: 7-10 September 2008.

Ali, S. F. and Yeung, H. (2010). "Experimental Investigation and Numerical Simulation of Two Phase Flow in Large Diameter Horizontal Flowline-Vertical Riser". *Petroleum Science and Technology*, Volume 28, pp 1079-1095.

Annunziato, C. M. and Girardi, G. (1984). "Flow pattern identification by pressure drop and void fraction fluctuations analysis in vertical two phase flow". In the Proceedings of European Two Phase Flow Group Meeting, Rome: 19-21 June 1984.

Bankoff, S. G. (1960). "A variable density single-fluid model for two-phase flow with particular reference to steam-water flow." *Transaction of ASME, Journal of Heat Transfer*, Volume 82, pp. 265-272.

Beggs, H. D and Brill, J. P. (1973) "A Study of Two-Phase Flow in Inclined Pipes," *Journal of Petroleum Technology,* Volume 255, pp 607-17.

Bendiksen, K. H., Malnes, D., Moe, R., and Nuland, S. (1991). "The Dynamic Two-Fluid Model OLGA: Theory and Application". *SPE production engineering*, SPE Paper number 19451, pp. 171-180.

Blaney, S. and Yeung, H. (2008). "Investigation of the exploitation of a fast-sampling single gamma densitometer & pattern recognition to resolve the superficial phase velocities & liquid phase water cut of vertically upward multiphase flows". *Flow Measurement and Instrumentation*, Volume 19, pp 57-66.

Brill, J. P. and Mukherjee, H. (1999). "Multiphase Flows in Wells", *SPE Henry Doherty series: Monograph,* Volume 17. Society of Petroleum Engineers Inc, Texas.

Brauner, N. and Barnea, D. (1986). "Slug/Churn Transition Upward Gas-Liquid Flow", *Chemical Engineering Science*, Volume 41, pp 159-163.

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Butterworth D. (1975). A comparison of some void-fraction relationships for co-current gas-liquid flow, *International Journal of Multiphase Flow*, Volume 1, pp 845-850.

Chen, X. T., Cai, X. D. and Brill, J. P. (1997). “A general model for transition to dispersed bubble flow”. *Chemical Engineering Science,* Volume 52(23), pp 4373-4380.

Cheng, H., Hills, J. H. and Azzopardi, B. J. (1998) “A study to bubble-to-slug transitions in the vertical gas-liquid flow in columns of different diameter”. *International Journal of Multiphase Flow*, Volume 24(3), pp 431-452.

Cheng, H., Hills, J. H. and Azzopardi, B. J. (2002). “Effects of initial bubble size on flow pattern transitions in a 28.9 mm diameter column”. *International Journal of Multiphase Flow*, Volume 28(6), pp 1047-1062.

Chexal, B., Lellouche, G., Horowitz, J. and Healzer, J. (1991). *“*A void fraction correlation for generalized applications”. *Progress in Nuclear Energy,* Volume 27(4), pp 255-295.

Chisholm, D. (1972). "An equation for velocity ratio in two phase flow”. *National Engineering Laboratory report, Number* 535, East Kilbride, Glasgow.

Costigan, G. and Whalley P. B. (1997). “Slug flow regime identification from dynamic void fraction measurements in vertical air-water flows”. *International Journal of Multiphase Flow*, Volume 23(2), pp 263-282.

Clark, N. N. and Flemmer, R. (1986). “The effect of varying gas voidage distributions on average holdup in vertical bubble flow”, *International Journal of Multiphase Flow*, Volume 12(2), pp 299-302.

Duns, H. Jr. and Ross, N.C.J. (1963). "*Vertical Flow of Gas and Liquid Mixtures from Boreholes*," Proceedings of the Sixth World Petroleum Congress, Frankfurt, Section II, Paper 22-PD6, pp 451-465.

Fabre, J., Peresson, L., Corteville, J., Bernicot, M. and Ozon, P. (1990). Severe Slugging in Pipeline /Riser Systems, SPE Production Engineering, SPE Paper number 16846, pp 299-305.

Fordham, E. J., Lenn, C. P, Holmes, A., Simonian, S. and Ramos, R. T. (1999) “Corrections of gradiomanometer data for volume fractions in two-phase flows”, *Measurement Science and Technology*, Volume (10), pp 131-135.

Guzhov, A., Mamayev, V. and Odishariya, G. (1967). A study of transportation in gas–liquid systems. In the10th International Gas Conference, Hamburg, Germany. (Cited in: Garcia, F., Garcia, R., Joseph D.D. (2005), “Composite power law holdup correlations in horizontal pipes”, *International Journal of Multiphase Flow*, Volume 31, pp 1276-1303.)

Hasanein, H. A., Kawaji, A., Chain M. C, Yoshioka, Y. (1997) “Flow regime identification of steam-water flow in al large vertical pipe at elevated pressures”. In the Proceedings of 1997 ASME Fluids Engineering Division Summer Meeting. ASME. Vancouver, Canada: 22-26 June, 1997.

Hewitt, G. F. and Roberts, D. N. (1969). “Studies of two-phase patterns by simultaneous X-ray and flash photography”. Atomic Energy Research Establishment Report AERE-M 2159, Harwell, England. HMSO.

Hewitt, G.F. (1982). “Chapter 2: flow regimes and Chapter 10: *Measurement of Two-Phase Flow Parameters*”. In: Handbook of Multiphase Systems, Edited by Hetsroni, G. Publisher: Hemisphere/McGraw Hill, Washington.

Hibiki, T. and Ishii, M. (2003). “One dimensional drift flux model for two phase flow in large diameter pipe”. *International Journal of Heat and Mass transfer*, Volume 46(7), pp 1173-1790.

Hills, T. J. (1976). “The operation of bubble column at high throughputs: Gas hold up measurements”. *The Chemical Engineering Journal*, Volume 12, pp. 89-99.

Hills, T. J. (1993). “The behaviour of bubble column at high throughputs: Radial voidage profiles”. *The Chemical Engineering Journal*, Volume 53, pp. 115-123.

Hirao, Y., Kawanishi, A., Tsuge, A. and Kohriyama, T. (1986). “Experimental study on drift flux correlation formulas for two phase flow in large diameter tubes”. *Proceedings of 2nd International Topical Meeting on Nuclear Power Plant Thermal Hydraulic and Operations Conference*. Tokyo, Japan:. 15-17 April, 1986.

Jones, O. C. Jr. and Zuber, N. (1975). “The interrelation between void fraction fluctuations and flow patterns in two phase flow”. *International Journal of Multiphase Flow*, Volume 2 (3), pp 273-306.

Kataoka, I. and Ishii, M. (1987). “Drift flux model for large diameter pipe and new correlations for pool void fraction”. *International Journal of Heat and Mass Transfer,* Volume 30(9), pp 1927-1939.

Khartabil, H. F. and Spinks, N. J. (1995). “An Experimental Study of a Flashing-driven CANDU Moderator Cooling System”. In the proceedings of 16th Annual Conference, Canadian Nuclear Society, Saskatoon, Canada: June 4-7, 1995.

Lombardi, C., and Carsana, C. G. (1992). “Dimensionless Pressure Drop Correlation for Two-Phase Mixtures Flowing Upflow in Vertical Ducts Covering Wide Parameter Ranges”. *Heat and Technology*, Volume 10 (1-2), pp 125-141.

Lucas, D., Beyer, M., Szalinski, L. and Schutz P. (2010). “A new database on the evolution of air-water flows along a large vertical pipe”, International Journal of Thermal Science, Volume 49 (4), pp 664-674.

Ma, P. Y., Chung, N. M., Pei, B. S. (1991). “Two simplified methods to determine void fractions for two phase flows”, *Nuclear Technology*, Volume 94, pp 124-133.

Mao, Z. S. and Dukler, A. E. (1993). “The myth of churn flow”. *International Journal of Multiphase Flow*, Volume (19), pp. 377-383.

Matsui, G. (1984) “Identification of flow regimes in vertical gas-liquid two-phase flow using differential pressure fluctuations”, *International Journal of Multiphase Flow*, Volume 10(6), pp 711-720.

Mishima, K. and Ishii, M. (1984). “Flow regime transitions criteria for upward two phase flow in vertical tubes”. *International Journal of Heat and Mass transfer*, Volume 27 (5), pp 723-737.

McQuillan, K. W. and Whalley, P. B. (1985). “Flow patterns in vertical two phase flow”. *International Journal of Multiphase Flow*, Volume 11, pp 161-175.

Neil, E. T. and Kazimi, M. S. (1989). *“Nuclear Systems I: Thermal hydraulics fundamentals”*. Hemisphere Publishing Corporation, USA.

Oddie, G., Shi, H., Durlofsky, L. J., Aziz, K., Pfeffer, B. and Holmes, J. A. (2003). “Experimental study of two and three phase flows in large diameter inclined pipes”. *International Journal of Multiphase Flow*, Volume 29(4), pp 527-720.

Ohnuki, A. and Akimoto, H. (1996). “An experimental study on developing air-water two phase flow along a large vertical pipe: effect of air injection method. *International Journal of Multiphase Flow*, Volume 22(6), pp. 1143-1154.

Ohnuki, A. and Akimoto, H. (2000). “Experimental study on transitions of flow pattern and phase distribution in upward air-water two phase flow along a large vertical pipe”. *International Journal of Multiphase Flow*, Volume 26(3), pp 367-386.

OLGA (2007). “OLGA - User Manual version 5” by SPT Group Ltd**,** Surrey, UK..

Omebere-Iyari, N. K., Azzopardi, B. J. and Ladam, Y. (2007). Two-phase flow patterns in large diameter vertical pipes at high pressures. *The AIChE Journal*, Volume 53(10), pp. 2493-2504.

Omebere-Iyari, N. K., Azzopardi, B. J., Lucas, D., Beyer, M. and Prasser, H-M. (2008) “The characteristics of gas/liquid flow in large risers at high pressures”, *International Journal of Multiphase Flow*, Volume 34(5), pp 461-476.

Prasser, H-M., Boettger, A., Beyer, M., Carl, H., Lucas, D., Schaffrath, A., Schűtz, P., Weiss, F-P., and Zschau, J. (2002). “TOPFLOW tests on the structure of the gas-liquid Interface in a large vertical pipe”. *Forschungszentrum Rossendorf Annual Report-2002*, FZR-380.

Prasser, H. M., Beyer, M., Carl, H., Gregor, S., Lucas, D., Schütz, P., and Weiss, F. P. (2005) “Evolution of the structure of a gas-liquid two-phase flow in a large vertical pipe”. Presented at: The 11th International Topical Meeting on Nuclear Reactor Thermal-Hydraulics (NURETH-11) Popes Palace Conference Centre, Avignon, France: 2-6 October, 2005.

Pickering, P. F., Hewitt, G. F., Watson, M. J. and Hale, C. P. (2001). The prediction of flows in production risers - truth and myth. In the Proceedings of Institute of International Research (IIR) Conference, Aberdeen, UK: 26-27 June, 2001.

Premoli, A., Francesco, D., and Prina, A. (1970). "An Empirical Correlation for Evaluating Two-Phase Mixture Density under Adiabatic Conditions", European Two-Phase Flow Group Meeting, Milan, Italy: 1970.

Qazi, N. and Yeung, H. (2006). “Remote access and monitoring of two phase flow rig using Web/WAP protocol”, *National Instruments: LabVIEW in the Curriculum Paper Contest*, pp 80-85.

Schlegel, J.P., Sawant, P., Paranjape, S., Ozar, B., Hibiki, T. and Ishii, M. (2009). “Void fraction and flow regime in adiabatic upward two-phase flow in large diameter vertical pipes”, *Nuclear Engineering and Design*, Volume 239(12), pp 2864-287.

Schmidt, Z., Brill, J.P. and Beggs, H.D. (1980). “Experimental study of severe slugging in a two-phase-flow pipeline - riser pipe system”. *SPE Journal*, Volume 20(5), SPE paper number 8306, pp 407-414.

Shen X, Mishima K, Nakamura H. (2005). "Two-phase phase distribution in a vertical large diameter pipe". *International Journal of Heat Mass Transfer*, Volume (48), pp 211-225.

Shen, X., Hibiki, T. and Nakamura, H. (2012). Developing structure of two-phase flow in a large diameter pipe at low liquid flow rate". *International Journal of Heat and Fluid Flow*, Volume 34, pp 70 - 84.

Shipley, D. G. (1984). "Shorter Communications: two phase flow in large diameter pipes". *Chemical Engineering Science*, Volume 39(1), pp 163-165.

Shoukri, M., Stankovic, B., Hassan, I., and Dimmick, J. (2000). "Effect of pipe diameter on flow pattern transitions and void fraction of air-water flow in vertical pipes". In the Proceedings of ICONE-8, 8th International Conference on Nuclear Engineering, Baltimore, MD, USA: 2-6 April, 2000.

Sun, X., Smith, T. R., Kim, S., Ishii, M. and Uhle, J. (2002). "Interfacial area of bubble flow in a relatively large diameter pipe". *Experimental Thermal and Fluid Science*. Volume 27, pp 97-109.

Taitel, Y., and Dukler, A.E. (1976). "A model for predicting flow regime transitions in horizontal and near horizontal gas- liquid flow", *The AIChE Journal*, Volume 22(1), pp 47-55.

Taitel, Y., Barnea, D. and Dukler, A.E. (1980). "Modelling flow pattern transitions for steady upward gas-liquid flow in vertical tubes". *The AIChE Journal*, Volume 26(3), pp 345-354.

Tutu, N. K. (1984). "Pressure drop fluctuations and bubble-slug transition in a vertical two phase air-water flow", *International Journal of Multiphase Flow*, Volume 10(2), pp 211-216.

Van der Welle. (1985). "Void fraction, bubbly velocity and bubble size in two phase flow*", International Journal of Multiphase Flow*, Volume 11(3), pp 317-345.

Vince, M. A. and Lahey Jr., R. T. (1982). "On the development of an objective flow regime indicator", *International Journal of Multiphase Flow*, Volume 8(2), pp 93-124.

Weisman, J. and Kang, S. Y. (1981). "Flow pattern transitions in vertical and upwardly inclined lines", *International Journal of Multiphase Flow*, Volume 7(3), pp 271-291.

Yoneda, K., Yasuo, A. and Okawa, T. (2002). Flow structure and bubble characteristics of steam-water two-phase flow in a large-diameter pipe. *Nuclear Engineering Design*, volume 217 (3), pp 267–281.

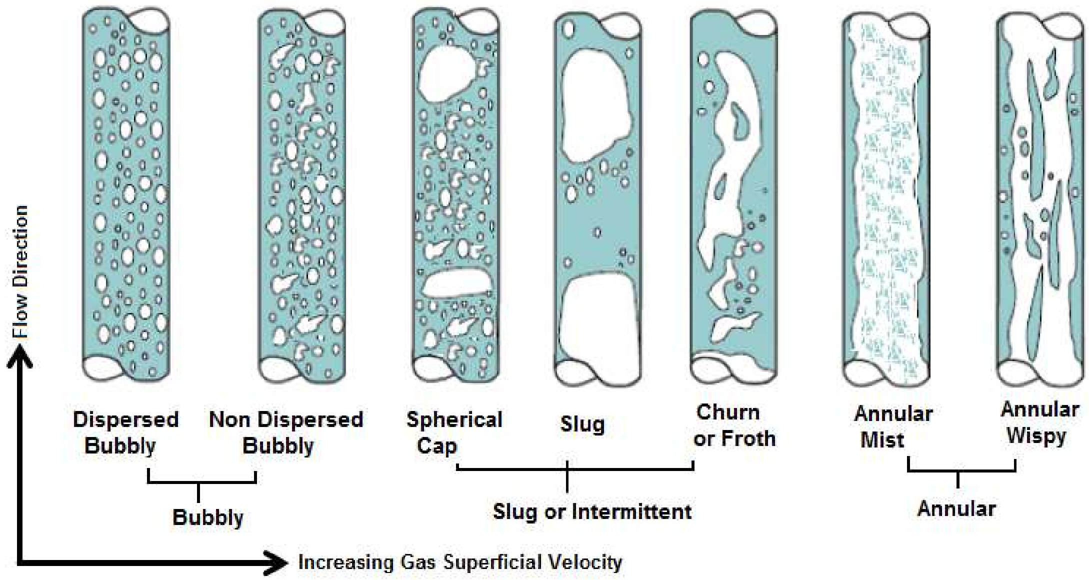

Figure 1. Commonly found flow patterns in vertical gas-liquid upflow.

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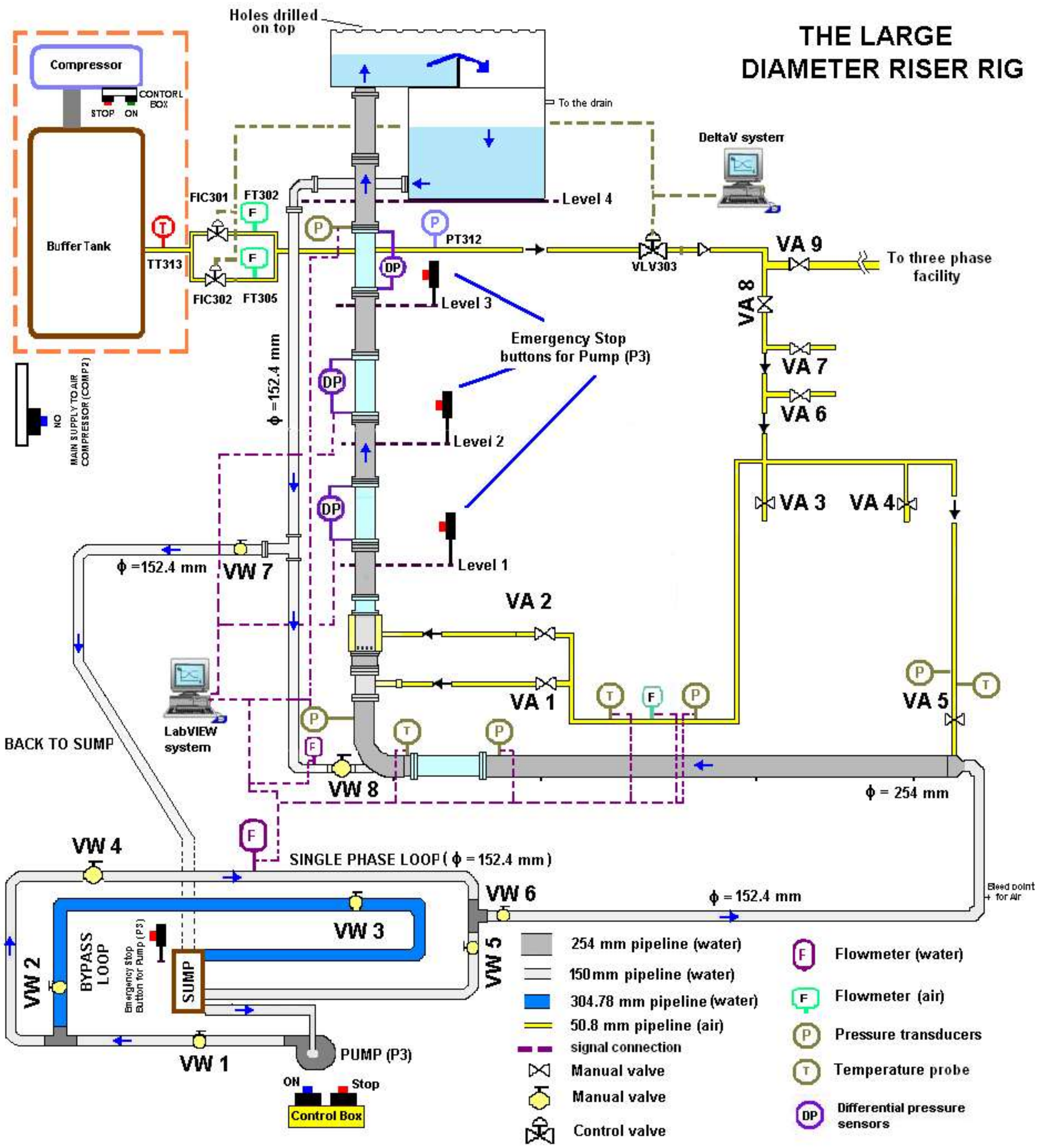

Figure 2. Schematic diagram of the large diameter facility.

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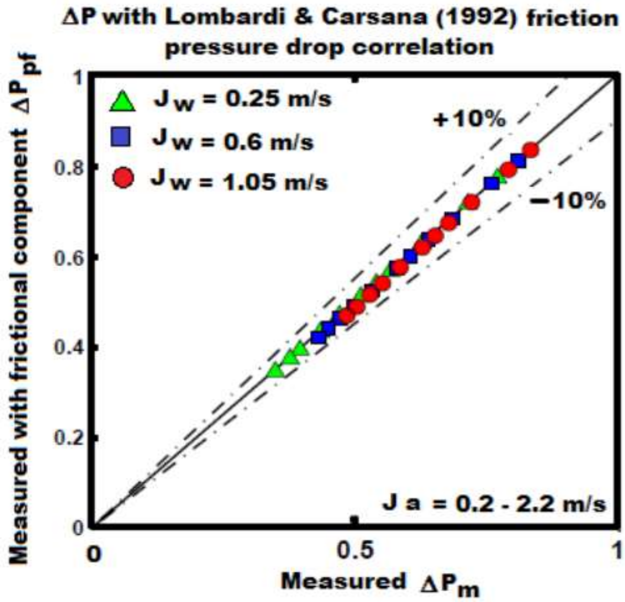

Figure 3. Effect of frictional loss in measurements with Lombardi and Carsana (1992) correlation.

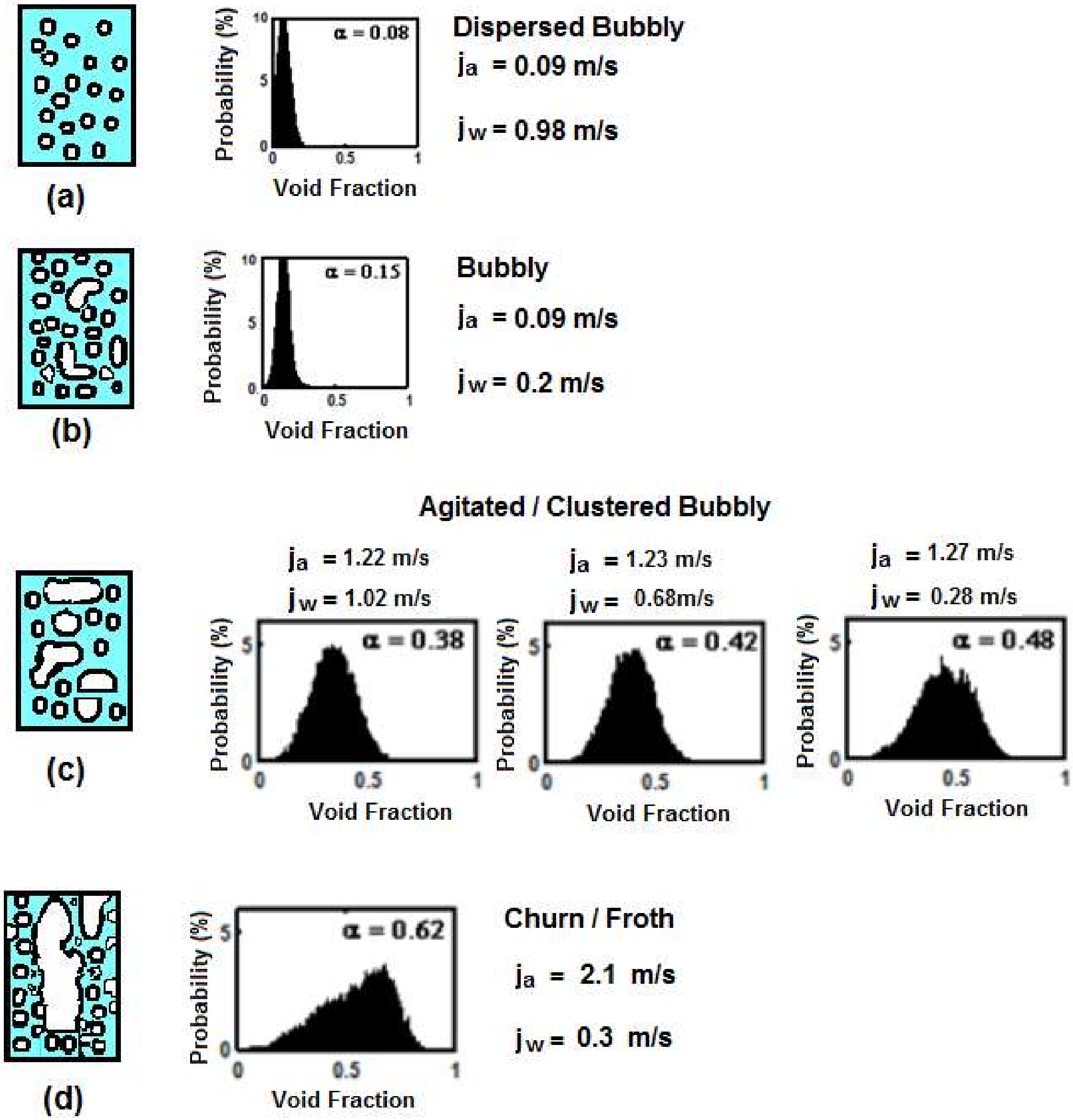

Figure 4. Sketches of observed air-water flow patterns with typical void fraction probability plots for 254mm diameter vertical pipe.

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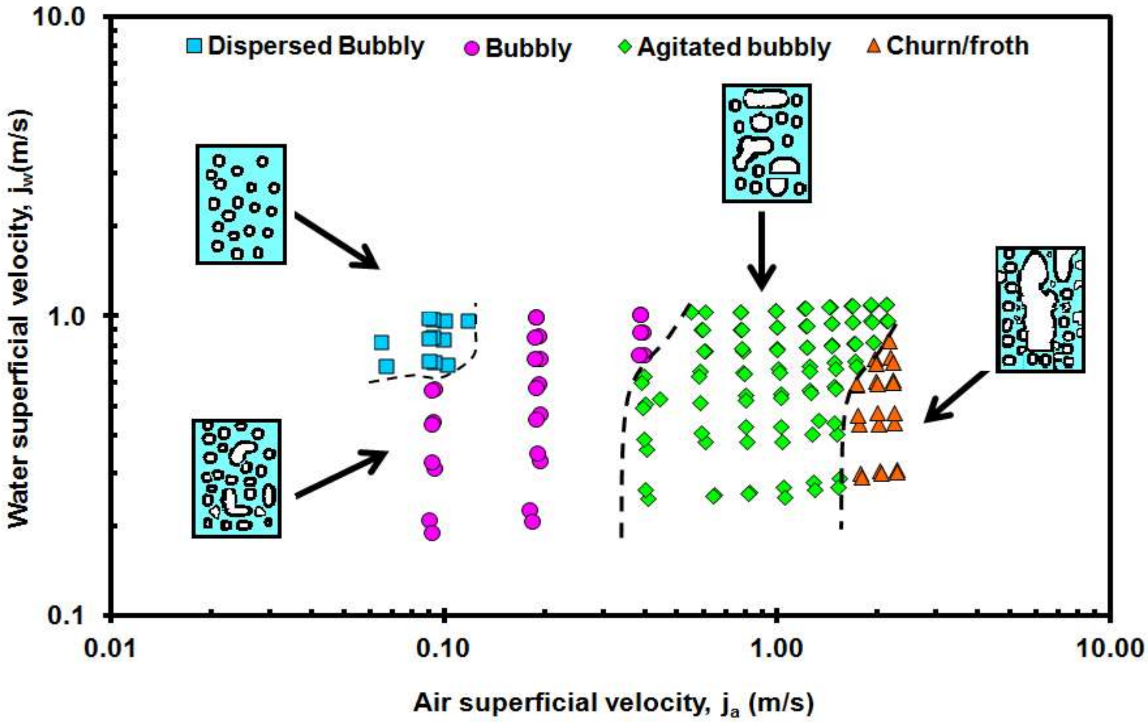

Figure 5. Experimental flow pattern map for 254 mm diameter vertical pipe.

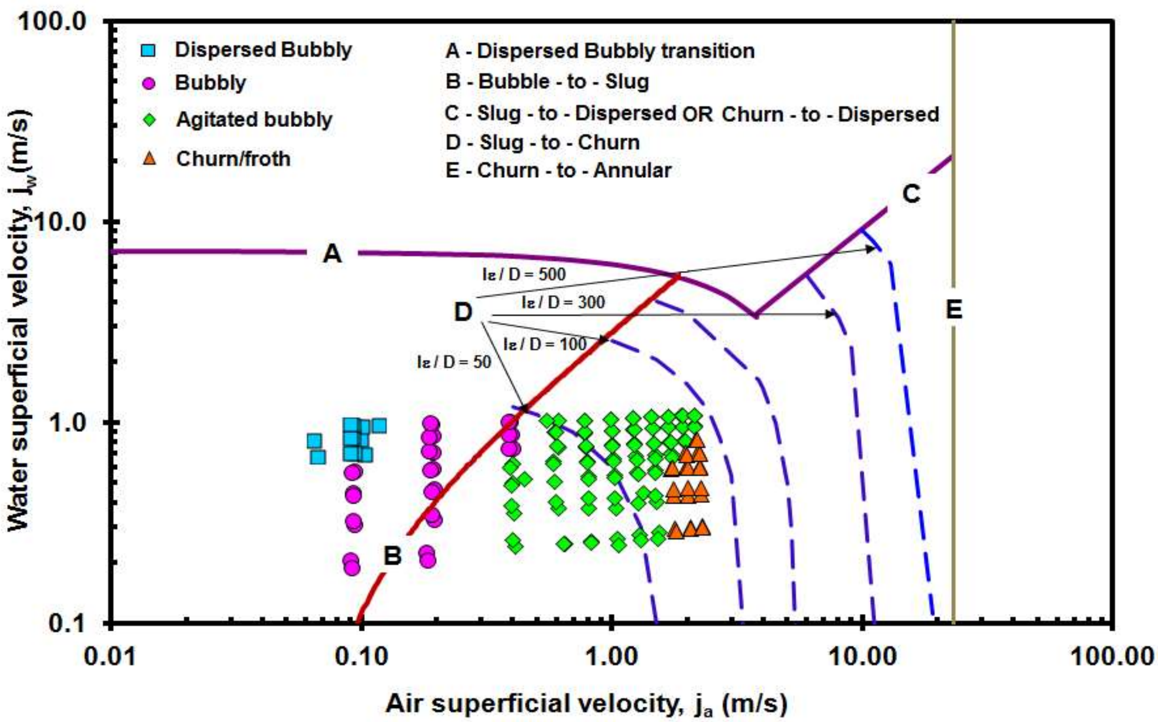

Figure 6. Flow pattern map for 254 mm diameter vertical pipe. Lines indicate the flow pattern transitions by Taitel *et al.* (1980).

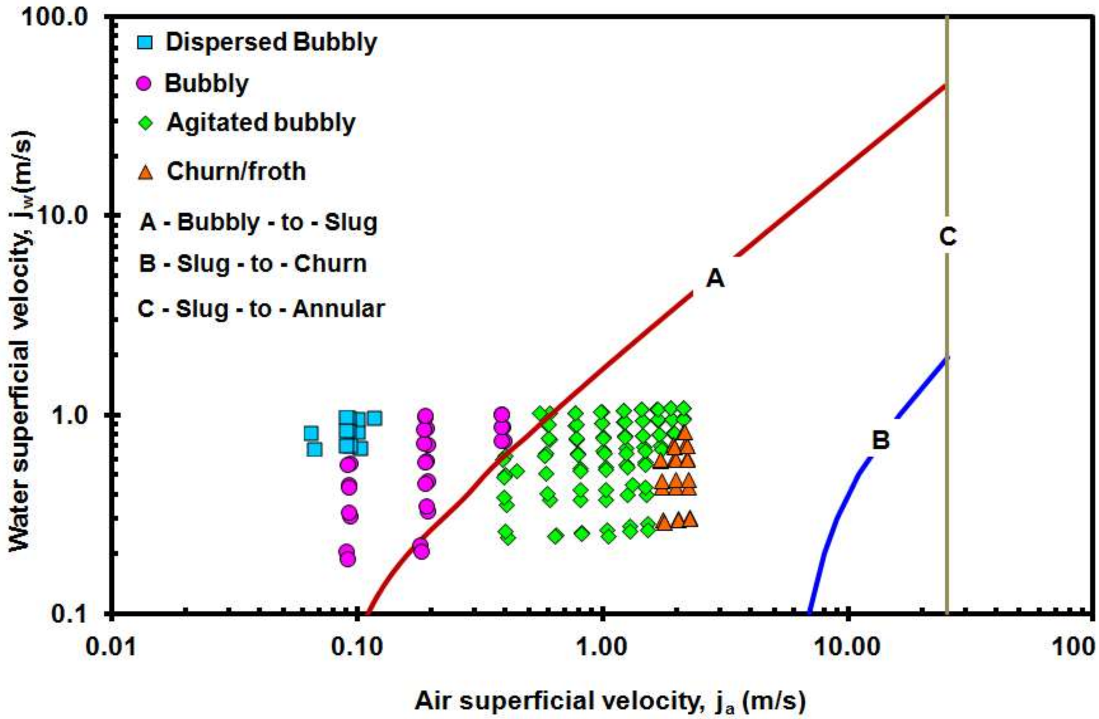

Figure 7. Flow pattern map for 254 mm diameter vertical pipe. Lines indicate the flow pattern transitions by Mishima and Ishii (1984).

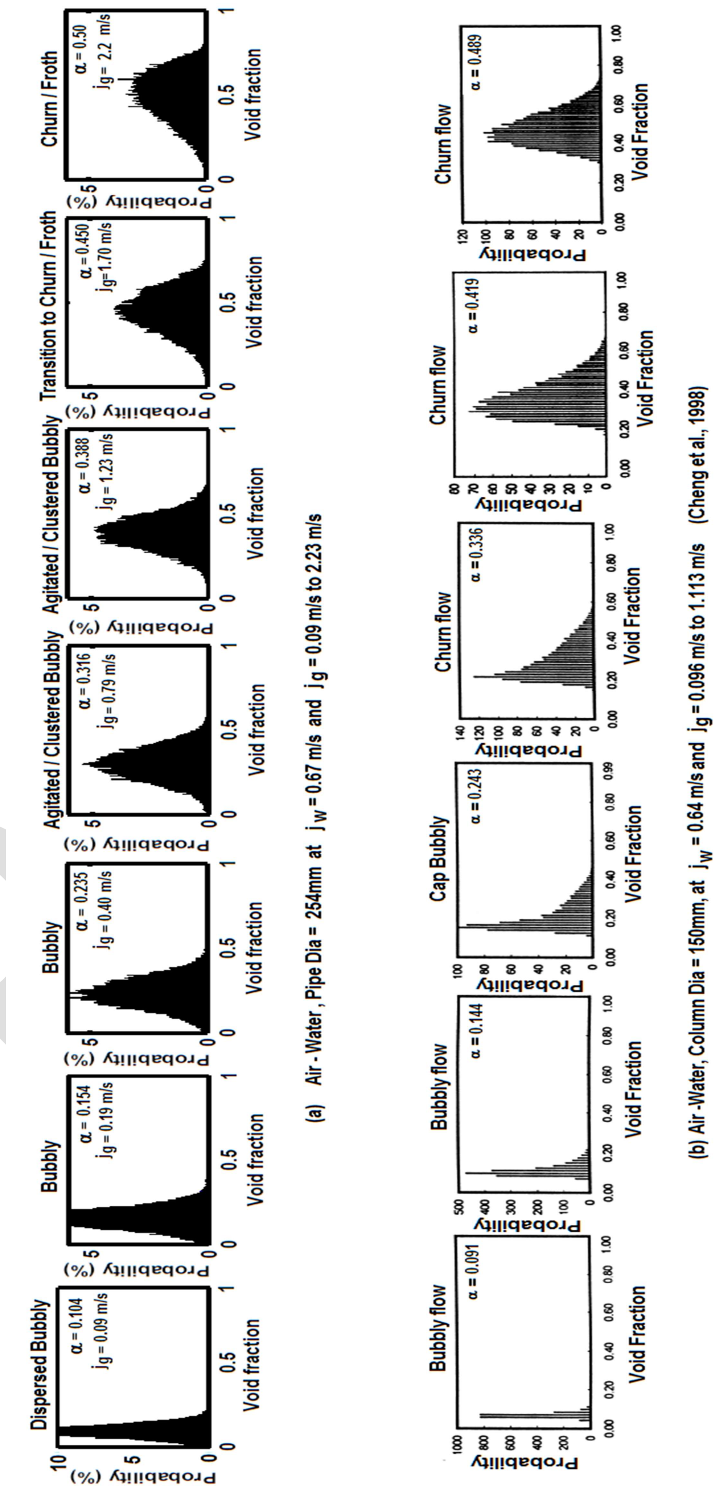

Figure 8. Air-Water flow patterns comparison between 254mm and 150mm diameter pipe under vertical upflow condition at constant water superficial velocity, $j_w = 0.67$m/s.

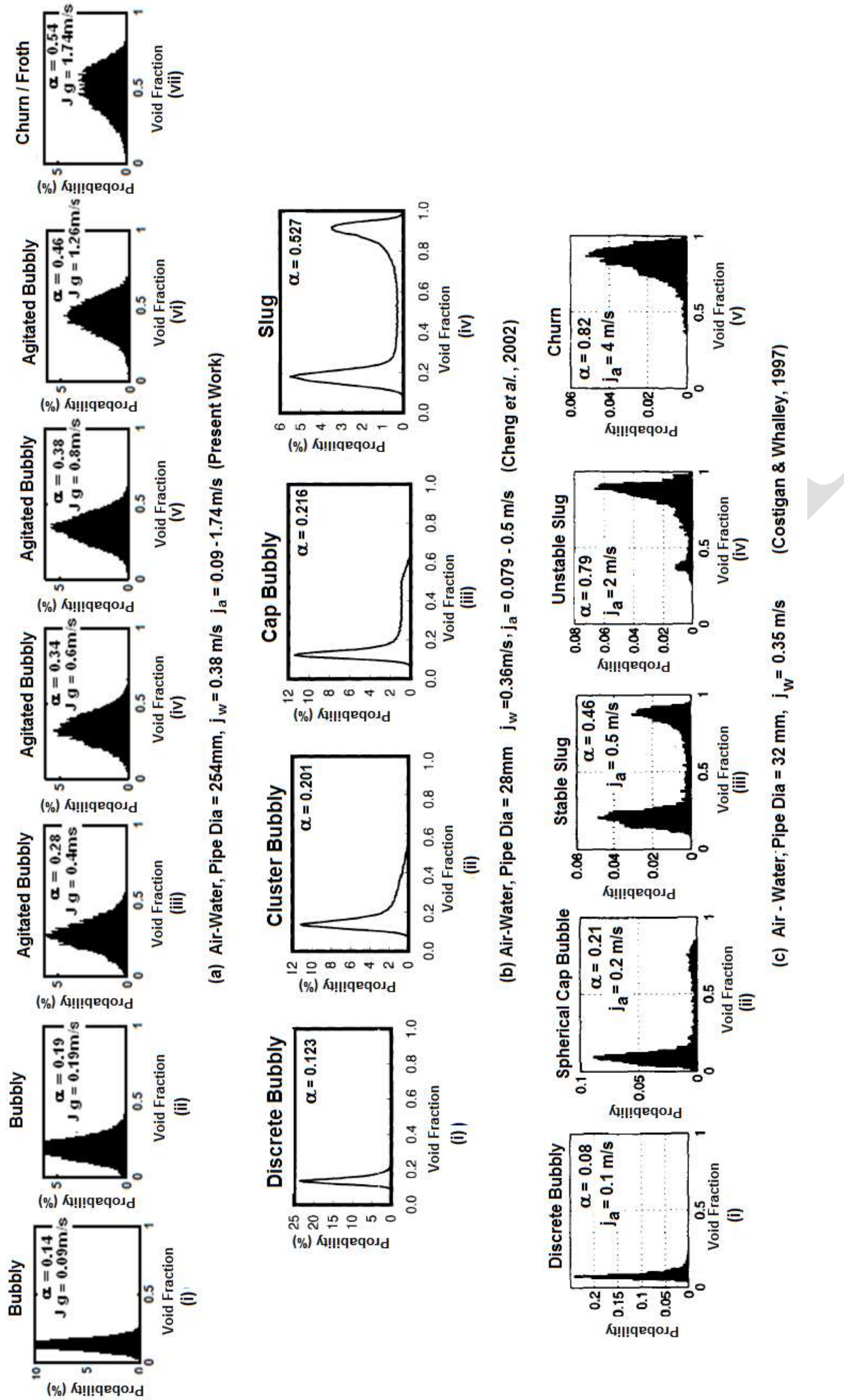

Figure 9. Air-Water flow patterns comparison between 254mm, 32mm and 28mm diameter pipes under vertical upflow condition at constant water superficial velocity, j_w = 0.38 m/s.

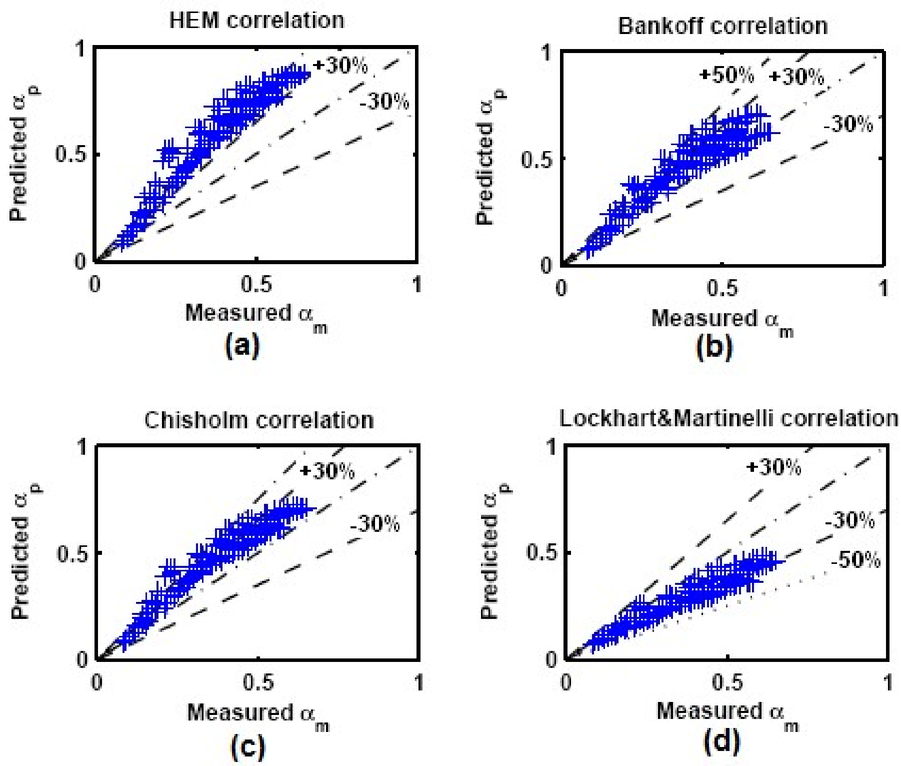

Figure 10. Comparison of the measured and predicted average void fraction for selected correlations.

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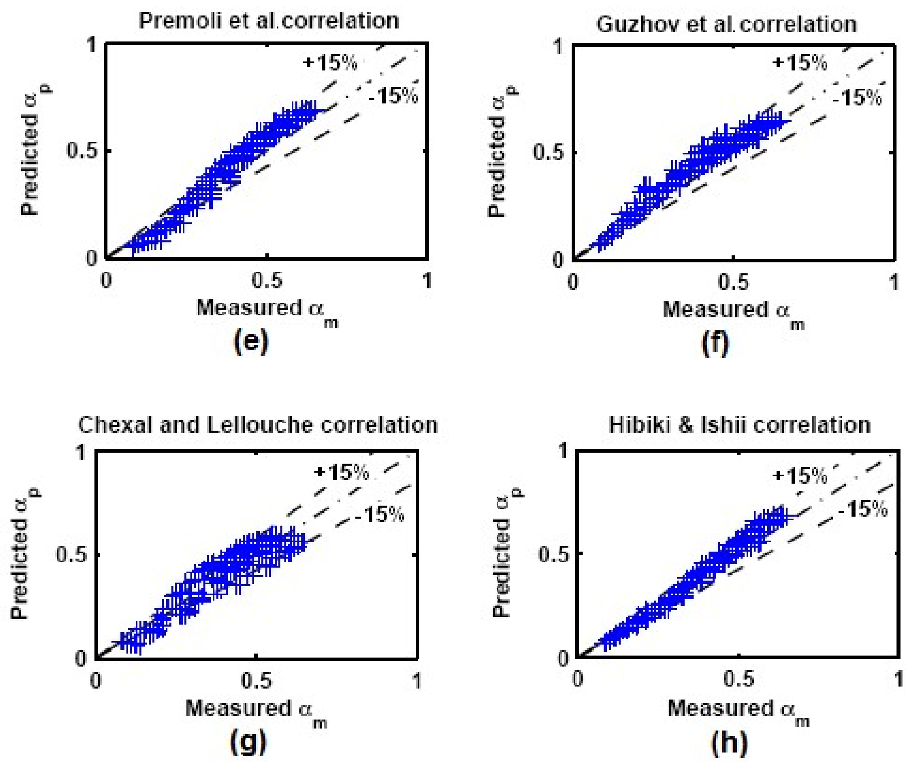

Figure 11. Comparison of the measured and predicted average void fraction for selected correlations.

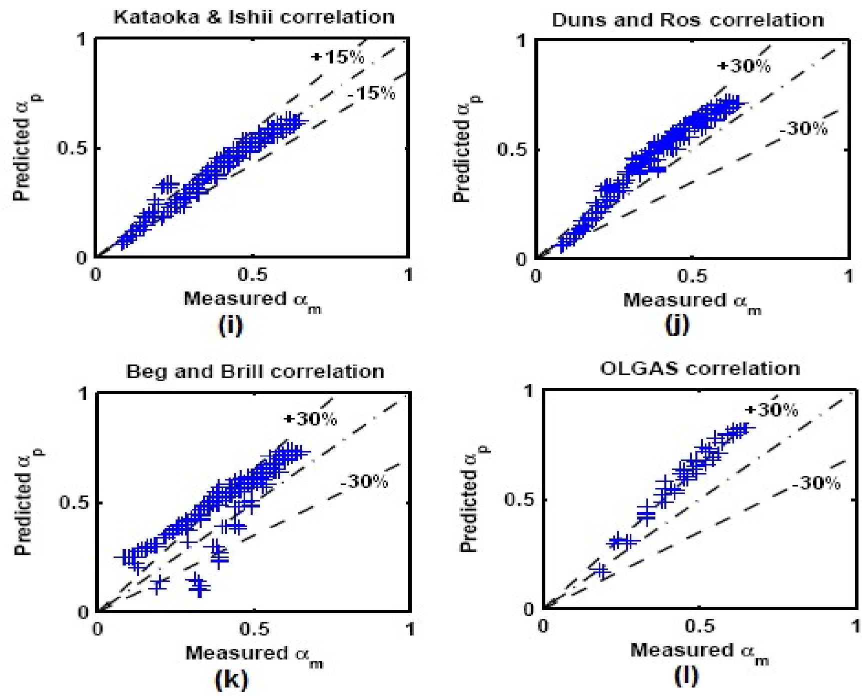

Figure 12. Comparison of the measured and predicted average void fraction for selected correlations

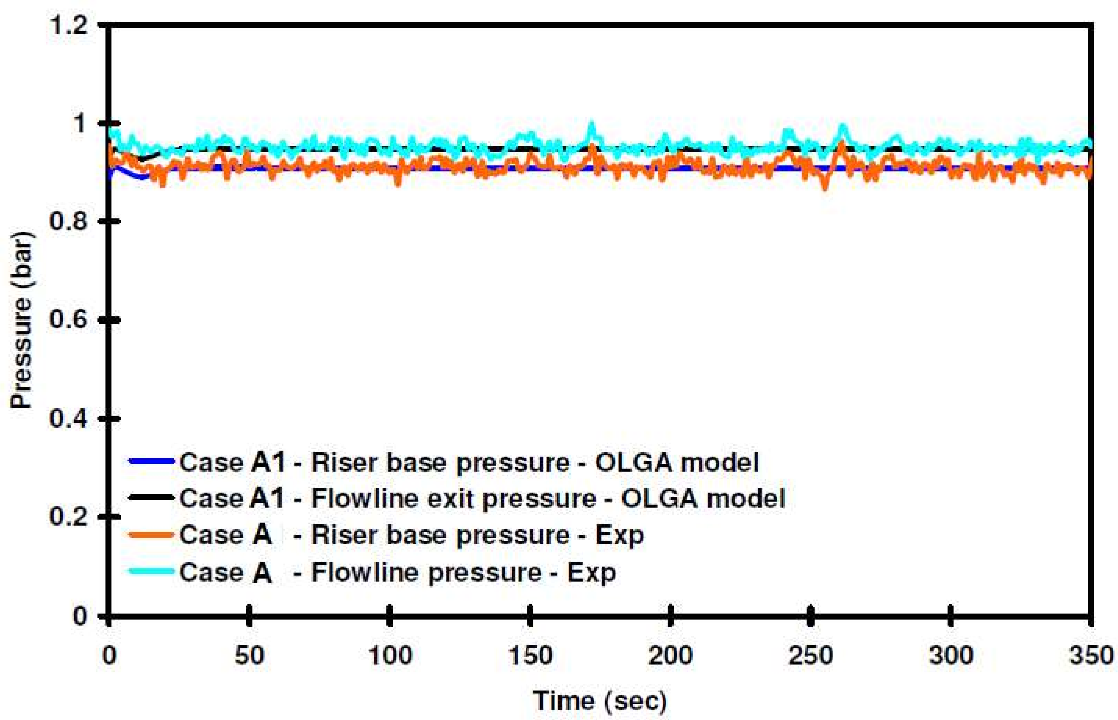

Figure 13. Comparison of experimental and simulated pressure time series results of Case A.

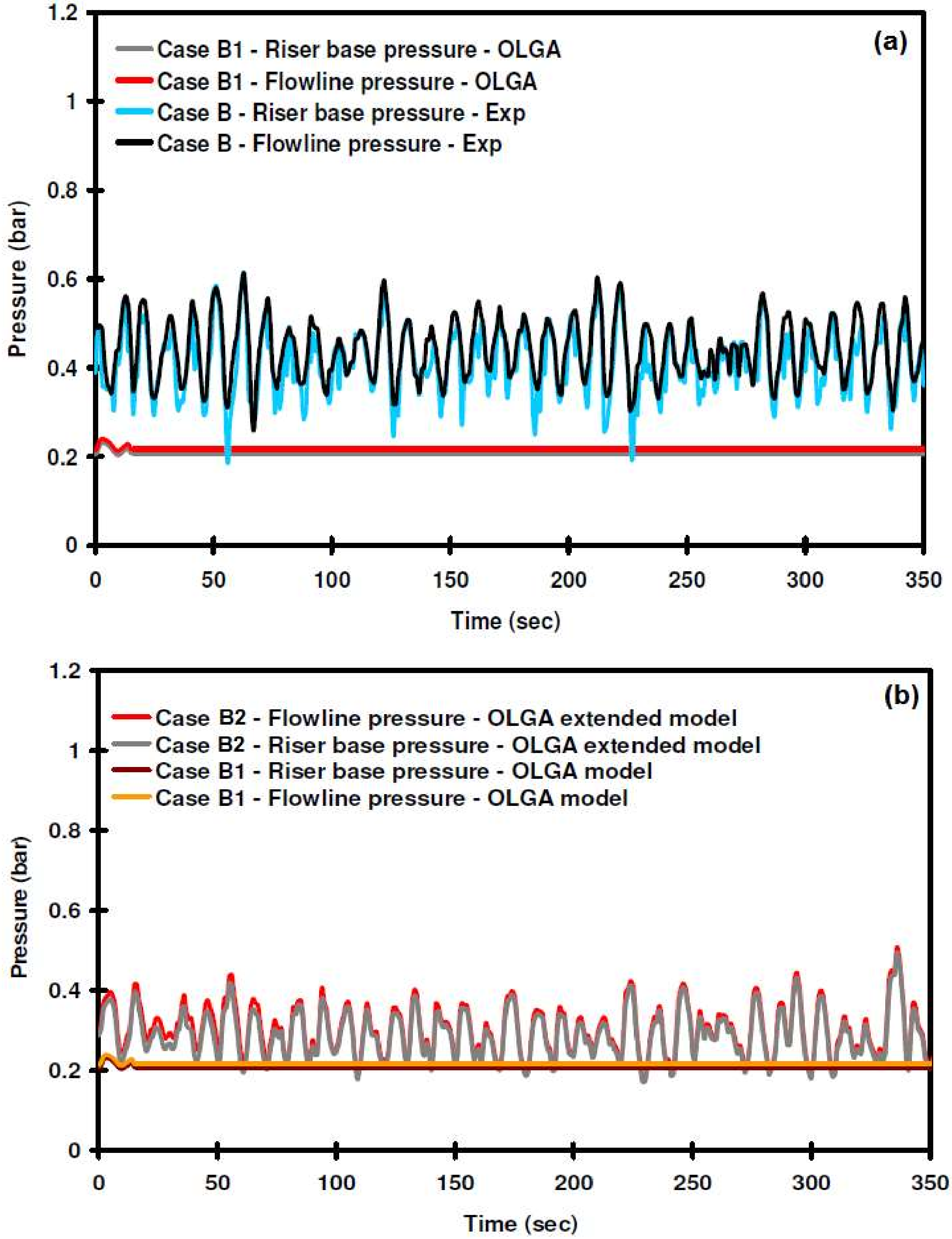

Figure 14. Comparison of experimental and simulated pressure time series results of Case B: (a) Initail model (B1) and (b) Modified model (B2).

TABLE I. THE COMPARISON OF VOID FRACTION CORRELATIONS USING LARGE DIAMETER VERTICAL UPFLOW DATA

Category	Correlation	Mean % Error	Standard Deviation (%)
Homogenous void fraction model ($\alpha=\beta$)	**HEM model**	**52.67**	**19.59**
The "$\alpha=K\beta$" forms	**Bankoff (1960)**	**16.46**	**16.52**
Based on Lockhart and Martinelli parameter (X)	**Lockhart & Martinelli (1949)**	**-24.07**	**9.66**
Based on slip ratio (S) relations.	**Chisholm (1972)**	**-27.5**	**9.89**
	Premoli *et al.* (1971)	**10.11**	**10.54**
	Guzhov *et al.* (1967)	**3.13**	**16.82**
Based on drift flux model, mostly from Nuclear industry.	**Kataoka & Ishii (1987)**	**1.55**	**10.40**
	Chexal & Lellouche (1992)	**-4.53**	**19.62**
	Hibiki & Ishii (2003)	**1.75**	**8.78**
Based on popular Oil & Gas industry.	**Duns & Ros (1963)**	**15.62**	**13.86**
	Beg & Brill (1973)	**21.42**	**50.73**
Two fluid model	**OLGA-S**	**30.26**	**11.51**

TABLE II. THE RESULT SUMMARY

Case Name	Exp Value (bar)	Steady State (bar)	Initial Model (bar)	Extended Model (bar)	[a] Δx, $\Delta x/2$, $2\Delta x$ (bar)	[b] Δt , $\Delta t/10$, $10\Delta t$ (bar)
Case A	**0.912**	**0.888**	**0.907**	**-**	**0.907, 0.852, 0.901**	**0.907, 0.909, 0.901**
Case B	**0.413**	**0.205**	**0.208**	**0.272**	**0.272, 0.269, 0.265**	**0.272, 0.273, 0.276**

[a] Grid and [b] Time step independency

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