Gas-liquid flow characteristics variation along inclined flat channel

A E Gorelikova\textsuperscript{1,2}, M A Pakhomov\textsuperscript{1}, V V Randin\textsuperscript{1,2}, V I Terekhov\textsuperscript{1} and A V Chinak\textsuperscript{1}

\textsuperscript{1}Kutateladze Institute of Thermophysics, Novosibirsk, 630090, Acad. Lavrentiev ave., 1, Russia
\textsuperscript{2}Novosibirsk State University, Novosibirsk, 630090, Pirogova str., 2, Russia

E-mail: randin@itp.nsc.ru

Abstract. The results of an experimental study of bubble diameters variation with distance from the point of gas injection in an upward bubble flow in an inclined flat channel are presented. The measurements were carried out for a superficial fluid velocity of 0.51 m/s (Re=12400) and at various values of the gas flow rate ratio. It is shown that at small values of the gas flow rate ratio ($\beta <2\%)$, the coalescence of bubbles is practically absent, therefore the angle of inclination of the channel and the distance from the point of introduction of gas into the liquid flow do not affect the average diameter of gas bubbles.

1. Introduction

Experimental studies of the upward gas-liquid flow in vertical pipes and channels are widely and in detail presented in the literature. Most experimental studies of bubble flows are devoted to flows in vertical pipes. In this case, the gas phase distribution is formed by the action of lateral forces on floating bubbles in the presence of a velocity gradient [1]. Much less attention was paid to bubbly gas-liquid flows in horizontal and inclined channels, although in this case the orientation of the channel may be very important.

The hydrodynamics of a bubbly gas-liquid flow in an inclined rectangular channel was studied in [2]. It was shown that the channel orientation affects significantly the flow hydrodynamics. In [3], a strong influence of the inclination angle on heat transfer from the wall in a bubble flow in an inclined flat channel was shown. A significant increase in the heat transfer coefficient was noted even at very low void fraction values of up to 0.01.

In paper [4], the mass transfer intensification on the wall by introducing a gas phase into the liquid flow was studied. It is shown that the absolute value of mass transfer in a two-phase flow slightly increases with increasing liquid velocity. The dispersion of the gas phase significantly affects mass transfer (with a decrease in the size of bubbles, mass transfer increases). The rapid growth of mass transfer on the wall occurs at very low void fraction (up to 0.05), after which a monotonic slow growth is observed with a further increase in void fraction. The effect of the gas phase on mass transfer increases with decreasing fluid velocity. The maximum value of mass transfer is achieved at channel inclination angles of $30\textdegree-50\textdegree$.

In [5], it was shown by direct numerical simulation that the angle of channel inclination in a rectangular inclined channel has a significant effect on the flow hydrodynamics and heat transfer. In an inclined channel, the bubbles migrate to the upper wall of the channel, increasing the mixing in the
near-wall layer, but reducing the liquid velocity compared to the single-phase flow. The heat transfer coefficient on the upper channel wall reaches a maximum at inclination angles of 30–60° from the horizontal.

The purpose of this work was to study the effect of gas bubbles and the angle of inclination on the hydrodynamics of a two-phase upward flow in an inclined flat channel.

![Experimental setup](image)

**Figure 1.** Experimental setup.
1 – main tank,
2 – centrifugal pump,
3, 4 – flow meters,
5 – valves,
6 – pre-chamber with confuser,
7–9 – setup sections,
10 – compressor,
11 – flow controller,
12 – valve,
13 – upper tank,
14–17 – temperature control system.

2. Experimental setup and technique

A slightly modified experimental setup from [3] was used. The experimental setup (Fig. 1) was a two-phase circulation loop closed in liquid. The carrier fluid flow from tank 1 was fed into the test section using centrifugal pump 2 through flow meters 3 and 4. The test section is a rectangular plexiglass channel with a cross section of 10x100 mm and length of 1.7 m. Rotameters were used as liquid flow meters. The flow rate was regulated by adjusting valves 5. At the inlet of the test section, pre-chamber 6 with confuser and grid was installed to flatten the flow across the channel cross-section. The test section consisted of several sections 7–9 connected by flanges. After the test section, the liquid was supplied to the upper tank 13, where it was separated from the gas and again supplied into main tank 1. Gas (air) was supplied from compressor 10. Gas flow rate was determined using FMA5518 flow controller (OMEGA Engineering, Inc.) 11. Additional adjustment of the gas flow rate could be made by valve 12. Gas was introduced into fluid flow through forty-one capillaries with a 0.3 mm i.d. The capillaries were glued into the Plexiglas section, placed on the upper channel wall. Gas bubbles were formed at separation of gas from the ends of the capillaries, protruded by 5 mm from the upper channel wall. The gas-liquid flow, obtained by mixing gas and liquid, entered the measuring section. Measurements of the bubble diameters were made in sections that were 100, 470, and 900 mm from the gas injection point. The temperature of the test liquid was kept constant at 25°C with the help of an automatic temperature control system 14–17. The channel inclination angle \( \theta \) was counted from the vertical, so the position \( \theta=0^\circ \) corresponded to the vertical position of the channel, and \( \theta=90^\circ \) was horizontal.

The experiments were carried out at superficial liquid velocity \( \bar{u}=0.51 \text{ m/s} \) (Re=12.400). Gas flow rate ratio varied from 0.01 to 0.2. In all studied regimes, the flow remained bubbly.

To compare the results of this work with the results of [3], the same test liquid was used: a solution of potassium ferri- (0.16%) and ferrocyanide (0.21%) and sodium carbonate (2.55%) in distilled water.

The study of the gas bubble diameters was carried out using the shadow photography method (Fig. 2). Bubbles were shot by the Nikon J4 2 camera through the optical section. The shooting speed
was 120 fps at a resolution of 1280x720 pix. To ensure a uniform light field, the illumination of the flow was produced by an LED matrix 150x150 mm 1.

The resulting images were processed by software. We used the method of image processing and selection of bubbles, similar to the method described in [6]. The diameter of gas bubbles was calculated from the area of the bubble in the image as the equivalent diameter using formula \( D = \sqrt{\frac{4S}{\pi}} \). The accuracy of determining the bubble boundary was ± 1 point. For photos taken by a Nikon J4 camera on calibration frames, 1 mm = 48 points. Thus, with the size of bubbles in the diameter range of 0.3–7 mm, the relative error in determining the diameter was 0.005–0.05.

![Figure 2. Unit for gas bubble diameter studies. 1 – LED matrix, 2 – camera.](image)

The distance from the point of gas injection into the liquid flow in the experiments was 100, 470 and 900 mm. A further increase in the distance from the bubble generator to the measuring section did not significantly affect the results obtained.

### 3. Numerical model

The numerical model is based on the Eulerian (two-fluid) approach. The Eulerian approach treats the particulate phase as a continuous medium with properties analogous to those of a liquid [7]. In the two-fluid approach, both phases are considered as interacting continua. This technique involves the solution of a second set of Navier–Stokes-like equations for dispersed phase (gas bubbles) in addition to those of the carrier (fluid) phase, including: equation for continuity, two equations for momentum conservation and energy equation. The Eulerian approach is based on kinetic equations for a one-point PDF of bubble coordinates, velocity, and temperature in the turbulent Gaussian fluid flow fields [8]. Properties such as the mass of particles per unit volume are considered as a continuous property and the particle velocity is the averaged velocity over an average control volume [9]. The set of axisymmetric RANS equations is used for the fluid phase. The low-Reynolds-number elliptic blending Reynolds stress model (RSM) by [10] is used in the work. The RSM model is modified for the presence of bubbles by the model [9]. The bubble behavior in turbulent liquid and their back action on the flow is determined by drag, gravity lift, virtual mass, wall lubrication forces, turbulent transport, and turbulent diffusion [9].

The numerical simulations have been carried out for polydispersed bubbly flow. In this paper, only coalescence and breakup processes have been taken. The equations for the numerical concentration and mass fraction of bubbles (the number of particles per unit volume) are written taking into account the approach [11]. A polydisperse ensemble of bubbles is modeled by a set of monodisperse groups (modes or fractions) in the framework of the \( \delta \)-approximation, where the continuous density distribution function of mass distribution of the dispersed phase is approximated by the sum of \( \delta \)-functions [11]. The basic mechanism of coalescence is associated with bubble collisions due to their entrainment into the turbulent motion of fluid [11]. It is assumed that only two bubbles collide, which
occurs most often in reality. Bubble break-up occurs due to their interaction with turbulent eddies, as shown in [12]. The rate of bubble break-up is determined by the interphase forces, which result in deformation and break-up of the bubble. The model of [12] is employed in the paper for modeling the bubble-eddy collision rate. Four δ–functions are used for all numerical simulations.

4. Results and discussion
Histograms for L=100 mm are similar for all gas flow rate ratios and inclination angles. Figures 3–5 show comparisons of histograms for distances 470 and 900 mm for different values of the volumetric gas flow rate ratio $\beta$, distances to the point where gas is introduced into the liquid flow L, and channel inclination angles $\theta$. In fig. 3-5, the ordinate is the number of bubbles of a specific diameter ($N_b$) multiplied by the bubble volume of this diameter ($V_b$) divided by the total volume of all bubbles ($Q$). The predictions have been started at $X=100$ mm, where experimental data on the mean bubble diameter were obtained.

![Figure 3](image1.png)

**Figure 3.** Bubble size distribution histograms along the inclined channel are $\beta = 1.5$ (a) and 10% (b). $\theta=30^\circ$, 1+4 – bubble distribution modes and 5, 6 – measured distributions at L = 470, 900 mm.

![Figure 4](image2.png)

**Figure 4.** Bubble size distribution histograms along the inclined channel are $\beta = 1.5$ (a) and 10% (b). $\theta=45^\circ$, 1+4 – bubble distribution modes and 5, 6 – measured distributions at L = 470, 900 mm.

At small values of the volumetric gas flow rate ratio $\beta$ (Fig. 3–5, a), a change in the distance to the observation point and channel inclination angle do not have a significant effect on the gas bubble diameters. In this case, the diameters of almost all bubbles lie in the range of 0.75–1.5 mm. This is due to the fact that with a small void fraction ($\beta=1.46\%$) the number of bubbles is small, the distance between the bubbles is large enough, and probability of their interaction is small. Therefore, coalescence of bubbles does not have a significant effect on the bubble size distribution.
Two peaks in the histograms are explained by the bubble coalescence directly near the capillaries. It is shown as well as in experiments and simulations. A similar effect was described in [13], where the generation of bubbles on a single capillary was studied. In this case, the first peak corresponds to the detachable diameter of the bubble, and the second one corresponds to the diameter of the bubble with double volume. With an increase in void fraction to $\beta=10.62\%$ (Fig. 3–5 b), the histograms obtained at distance $L=100$ mm are noticeably different from those at $L=470$ mm and $L=900$ mm. This is due to the fact that in the initial part, the bubble size lies mainly in the range of $1\div3$ mm and almost all bubbles are located along the upper channel wall in a thin layer of liquid. Therefore, an area with high concentration of bubbles is created near the upper channel wall. A large bubble concentration (a small distance between the bubbles) leads to a significant increase in the probability of bubble collisions. The interaction of bubbles leads to coalescence and increase in the average volume of bubbles as the gas-liquid mixture flows along the channel. At the same time, the number of bubbles decreases, and the distance between the bubbles increases, this leads to a decrease in the probability of interaction. Therefore, the difference between histograms for $L=470$ mm and $L=900$ mm is not so significant. In addition, an increase in the bubble diameters leads to an increase in the bubble velocity, which also accelerates the entrainment of the gas phase and reduces the concentration of bubbles.

Figure 5. Bubble size distribution histograms along the inclined channel are $\beta=1.5$ (a) and $10\%$ (b). $\theta=60^\circ$, 1–4 – bubble distribution modes and 5, 6 – measured distributions at $L=470, 900$ mm.

Figure 6 shows diagrams of the dependence of the average bubble diameters at various distances from the point of gas injection into the liquid flow. As the distance from the bubble generator increases, the average bubble diameters increase, as shown in Figure 6.

Figure 6. Mean bubble diameter for $\theta=45^\circ$. 1–distance is $L=470$ mm, 2–900 mm.
increases, the average diameter of the bubbles also increases. This is due to the coalescence of bubbles. It can be seen that with the values of the volumetric gas flow rate ratio $\beta<2\%$, the average bubble diameter grows slowly and practically does not differ for all two curves (470 mm and 900 mm). This is due to the fact that at low void fraction the interaction between bubbles is practically absent and the average bubble diameter throughout the channel does not change. As the void fraction increases, the probability of interaction between the bubbles increases. This leads to a growth in the average bubble diameter with increasing distance from the generator.

The developed model qualitatively and quantitatively predicts the distribution of the bubble diameter of the two-phase bubbly flow along the pipe radius and length in inclined bubbly flow with presence of bubbles break-up and coalescence in the range of gas volumetric flow rate ratios $\beta=1.5–10\%$.

**Conclusions**

At small values of the volumetric gas flow rate ratio ($\beta<2\%$), the coalescence of bubbles is practically absent, therefore the channel inclination angle and the distance from the point of gas injection into the liquid flow do not affect the average gas bubble diameters.

The mathematical model is based on the Eulerian approach. Bubble dynamics is described taking into account the changes in the gas bubbles density, break-up and coalescence processes. Turbulence of the carrier fluid phase is predicted using the model of Reynolds stress transport. It is shown that the developed model allows one to predict adequately the bubble diameter distributions in two-phase flow in an inclined duct in the range of gas volumetric flow rate ratios $\beta=1.5\%–10\%$.

**Acknowledgements**

The study was performed by financial support of the Ministry of Science and Higher Education by the Project No. AAAA-A18-118051690120-2 (experimental part) and Project No. AAAA-A17-117030310010-9) (mathematical model and numerical results).

**References**

[1] Zun I 1980 *Int. J. Multiphase Flow* 6 583
[2] Kashinsky O N, Chinak A V and Kaipova E V 2003 *Thermophys. and Aeromech.* 10 69
[3] Kashinsky O N, Randin V V and Chinak A V 2014 *J. Eng. Thermophys.* 23 39
[4] Kashinskii O N, Chinak A V, Smirnov B N and Uspenskii M S 1993 *J. Eng. Phys. Thermophys.* 64 422
[5] Piedra S, Lu J, Ramos E and Tryggvason G 2015 *Int. J. Heat Fluid Flow* 56 43
[6] Fu Y and Liu Y 2016 *Int. J. Multiphase Flow* 84 217
[7] Drew D A 1983 *Ann. Rev. Fluid Mech.* 15 261
[8] Zaichik L I, Skibin A P and Solov’ev S L 2004 *High Temp.* 42 101
[9] Pakhomov M A and Terekhov V I 2016 *Int. J. Heat Mass Transfer* 101 1251
[10] Lopez de Bertodano M, Lee S J, Lahey Jr R T and Drew D A 1990 *ASME J. Fluids Eng.* 112 107
[11] Mukin R V 2014 *Int. J. Multiphase Flow* 62 52
[12] Nguyen V T, Song C-H, Bae B U and Euh D J 2013 *Int. J. Multiphase Flow* 54 31
[13] Vorob’ev M A, Kashinskii O N, Lobanov P D and Chinak A V 2012 *Fluid Dynamics* 47 494