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Numerical study of phase distribution phenomena and wall effects in bubbly two-phase flow

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Abstract. The results of numerical simulation of the structure of a two-phase flow of a gas–liquid bubble mixture in a vertical ascending flow in a pipe are presented. The mathematical model is based on the use of the Eulerian description of the mass and momentum conservation for the liquid and gas phases, recorded within the framework of the theory of interacting continua. To describe the bubble-size distribution, the equations of particle-number conservation for individual groups of bubbles with different constant sizes are used for each fraction, taking the processes of breakage and coalescence into account. Comparison of the results of numerical simulation with experimental data has shown that the proposed approach enables the simulation of bubble turbulent polydisperse flows in a wide range of gas concentrations.

1. Introduction

Two-phase flow refers to any fluid flow consisting of two phases, either of the same or different substances. Gas–liquid flows are by far the most complex due mainly to the fact that they combine a deformable interface and the compressibility of the gaseous phase [1-4]. This is probably the most important form of two-phase flow, and is found widely in a whole range of industrial applications. These include nuclear reactors, chemical reactors, pipeline systems for the transport of oil and gas, automotive industry, sewerage treatment plants etc. For a vertical gas–liquid two-phase flow in a pipe, the distribution of interfacial structure can take possible regimes such as bubbly flow, slug flow, churn-turbulent flow, wispy annular flow and annular flow. Thus, in order to capture and enhance the accuracy of code predictions of the transient evolution of these flow regimes, more detailed numerical methods able to model the dynamics of fluid particle interaction are necessary. In the present study, we will be concerned with isothermal bubbly flows, i.e. dispersed gas–liquid flow with the gas as the dispersed phase and liquid as the continuous phase where only momentum is exchanged between the phases. The system studied in [5] is vertical upflow of water and air in a round pipe with inner diameter D = 38.0mm. They performed experiments specifically designed to show the effects of bubble size by using a special gas injector that allows to adjust the bubble size independently of liquid and gas mass fluxes. All three parameters, average bubble size $d_b$, liquid mass flux $< J_l >$
and gas mass flux \(< J_g >\) were varied in the investigation. Radial profiles of void-fraction \(\alpha_g\), bubble-size \(d_i\), liquid velocity \(U_l\) and axial liquid turbulence intensity were measured at an axial position \(L/D = 60\), so that fully developed flow conditions are realized. A change in the void fraction profile from wall to core peak with increasing bubble size was observed as well as turbulence suppression in the pipe center for combinations of high liquid and low gas mass flux which correspond to the smallest bubble sizes.

2. Governing equations

System the Reynolds-averaged Navier-Stokes equations to describe the dynamics of a turbulent mixture of liquid and air bubbles with regard to interfacial interaction has the form

\[
\frac{\partial (\rho_i \alpha_i)}{\partial t} + \frac{\partial (\alpha_i \rho_i u_i)}{\partial x_j} = 0, \quad \alpha_l + \alpha_g = 1, \quad k = l, g
\]

\[
\frac{\partial (\alpha_i \rho_i u_i)}{\partial t} + \frac{\partial (\alpha_i \rho_i u_i u_j)}{\partial x_j} = -\alpha_i \frac{\partial p}{\partial x_i} + \alpha_i \rho_i g_i + F_{ii} + \frac{\partial}{\partial x_j} \left[ \left( \mu_k + \mu_l \right) \left( \frac{\partial u_i}{\partial x_j} + \frac{\partial u_j}{\partial x_i} - \frac{2}{3} \frac{\partial u_k}{\partial x_k} \delta_{ij} \right) \right]
\]

Here \(x_j\) - cartesian coordinates, \(u_i\) - components of velocity vector, \(t\) - time, \(p\) - pressure, \(\rho_i, \rho_g\) - density of continuous and dispersed phases respectively, \(\alpha_i\) - is the volume fraction of liquid phase in mixture, \(\delta_{ij}\) - is the Kronecker symbol, \(\mu_k, \mu_g\) - dynamic coefficient of viscosity of water and air, \(\mu_{ii}\) - is the coefficient of turbulent viscosity, \(F_{ii}\) - the strength of interfacial interaction forces. The turbulent viscosity for the carrier liquid phase are determined using a two-parameter turbulence model, modified for two-phase media [3,6]. In the equations for the transfer of turbulent kinetic energy and its dissipation introduces additional components to the kinetic energy caused by pulsations of the bubbles. The motion of the dispersed phase is determined by the forces of interfacial interaction. The main forces considered have the following components: Archimedes force, resistance force, joined the force, the rotational force of Magnus, the wall friction force [3]. To describe the distribution of bubble sizes in two-phase flow the equation for conservation of number of particles is written, taking into account the processes of coagulation and fragmentation. To solve the equation of conservation for the number of bubbles we use an approach based on the method of fractions [6]. Range of distribution of particle sizes is divided into a number of fractions with fixed boundaries, there can also be exchange of vesicles between different fractions as a result of coagulation and fragmentation. In this method, the distribution of bubble sizes is approximated by a piecewise uniform distribution, thus, the problem of describing the spectrum of droplet size is reduced to the solution of the equations for the volume concentrations of individual fractions.

3. Results

To verify the presented mathematical model, the experimental data of [7] were selected, where polydisperse upward bubble flow in a pipe was studied (pipe diameter \(D = 50.8\) mm; height \(H = 3060\) mm). Table presents the main parameters for different flow modes of a two-phase bubble medium in a vertical pipe (\(< J_f >\) is the reduced velocity of the liquid phase at the pipe inlet, \(< J_g >\) is the reduced velocity of the gas phase, and \(\vartheta\) is the volume flow gas concentration). The Reynolds number for the liquid phase flow varies within \(Re_l = \rho_l < J_f > D/\mu_l = (2.5 – 10.2) \times 10^4\).

The two-phase flow in a vertical pipe is assumed to be axisymmetric, so a design field consisting of a circular sector with a radius of \(r_0 = D/2 = 2.54 \times 10^{-3}\) m, a length of \(H = 3.06\) m, and a solution angle of \(5^\circ\) is selected for numerical simulation. Numerical calculations are performed for finite-volumes grids consisting of \(M_e = 32 600, 84 200,\) and \(168 600\) units. In the plane section of
ox1ox2, the number of partitions along the axis of the pipe and along its length for different grids is \( L_1 = 20 \times 200, L_2 = 40 \times 400, \) and \( L_3 = 60 \times 600 \). The numerical solution for grid \( L_2 \) differs from that for \( L_3 \) by less than 2%; therefore, all the presented numerical results are obtained using a finite-volume grid with partitioning \( L_2 \). The following parameters of the components of the two-phase water–air medium

**Table 1.** The parameters of the experimental measurements (\( d_i = 2.5 \) mm)

|      | \(<J_l>\), m/s | \(<J_g>\), m/s | \( \alpha_g \) | \( g \) |
|------|----------------|----------------|--------------|------|
| L1   | 0.986          | 0.0473         | 0.051        | 0.053|
| L2   | 0.986          | 0.321          | 0.231        | 0.279|

are set: \( \rho_l = 998.2 \text{ kg/m}^3, \rho_g = 1.2 \text{ kg/m}^3, \mu_l = 1.1 \times 10^{-3} \text{ kg/(m s)}, \) and \( \mu_g = 1.7 \times 10^{-5} \text{ kg/(m s)}. \) The particle-size distribution is presented in the form of six groups of bubbles with the minimum diameter of \( d_i^{\min} = 1 \times 10^{-5} \) m and the maximum diameter of \( d_i^{\max} = 1.2 \times 10^{-2} \) m. The calculated and experimental data on the distribution of the characteristics of the carrier and polydisperse flows in the developed turbulent flow in the section of \( x_3/D = 53.6 \) along the pipe radius are presented \((r)\) is the distance from the pipe axis). Figure 1 presents comparative data of the averaged profile of the axial component of the carrier phase velocity \( u_{3l} \) and velocity \( u_{3g} \) of the gas-liquid flow for \( <J_l> = 0.986 \text{ m/s} \) and \( <J_g> = 0.0473 \) and 0.321 m/s (modes L1 and L2). The introduction of a small concentration of bubbles \( (\alpha_g = 0.0473) \) leads to the fact that the velocity profile of the liquid phase becomes practically flat in the flow core (Figure 1a, curve 1). The effect of bubbles manifests itself in the fact that the velocity gradient increases near the wall. An increase in the velocity and concentration of bubbles \( (\alpha_g = 0.321) \) lowers the effect of the dispersed phase on the velocity profile. The gradients of the velocity

![Figure 1](image1.png)

**Figure 1.** Comparison of (curves 1 and 2) calculated results and (symbols) experimental data obtained in the L1 and L2 modes for (a) \( u_{3l} \); (b) \( u_{3g} \).

profile in the near-wall region decrease, and the profile itself becomes more parabolic, corresponding to a single-phase flow. At low flow velocities, a peak bubble concentration in the near-wall region and a uniformly low bubble concentration in the flow core at different gas concentrations are observed (Figure 2a). Similar results in the distribution diagram of \( \alpha_g \) along the pipe radius are confirmed by the experimental data of many researchers. As an example, the saddle profiles of the gas concentration along the pipe section were found experimentally in the flow of an ascending gas-liquid flow [5].
Figure 2. Comparison of (curves 1 and 2) calculated results and (symbols) experimental data obtained in the L1 and L2 modes for (a) $\alpha_g$; (b) $d_s$.

In these works, the peak distribution of the volume gas concentration in the near-wall region and the uniform flow in the core (called the “skin effect”) were obtained in the flow of a bubble mixture in the pipe at $Re = 6 \times 10^3$–$6 \times 10^4$. The difference in the effect of interphase forces on bubbles of different sizes leads to the fact that the bubbles with a diameter that is smaller than the original one are displaced into the near-wall region. Note that the considered flow mode can be characterized as a flow in that the bulk of the bubbles move in the near-wall region. The distribution of the average volume–surface diameter $d_s$ of bubbles along the pipe radius is shown in Figure 2b for two modes. At low fluid flow rates (curve 1), a uniform distribution of the mean diameter of bubbles along the channel section is observed. In this case, there are a small number of bubbles per unit volume and their effect is small; the intensity of the coagulation and fragmentation processes is insignificant, and a certain equilibrium state takes place. The number of bubbles produced by merging is approximately the same as the number of particles that have disintegrated into smaller ones as a result of collision with other particles. The increase in the velocity and concentration of the gas phase leads to an increase of $d_s$ up to 4 mm, which indicates the predominance of the intensity of coagulation of bubbles over the rate of their collapse. The motion of a gas–liquid flow with an initial inlet velocity of $<J_g>$ = 0.321 m/s is characterized by an intensification of coalescence and breakage processes; the action of transverse interfacial forces results in the redistribution of bubbles of different sizes along the channel section. Small bubbles move to the near-wall region and large bubbles accumulate in the center of the channel due to the action of interfacial forces. The flow structure at low concentrations and velocities of the gas phase, as characterized by the accumulation of bubbles at the walls, is reversed. At high velocities of the gas-liquid flow, the bulk of the bubbles move in the central region, while near the walls their concentration is negligible. As the velocity of the carrier and dispersed phases increases, the structure of the particle-size distribution changes. It is seen that the proportion of larger bubbles increases (with a size of > 5 mm) in the gas-liquid flow region. The distribution of $d_s$ along the pipe radius for both modes is shown in Figure 2b, indicating a change in the structure of a two-phase bubble flow for the L2 mode. The increase in the velocity of the liquid leads to the fact that the intensity of the collisions increases. It grows because of the higher concentration of bubbles per unit volume, causing a reduction in the distance between the bubbles to a size comparable to their diameters. Increasing the involvement of bubbles in turbulent pulsations enhances the intensity of coagulation, as evidenced by an increase in $d_s$ in the core of the flow.

The particle-size distribution in section $x/D = 53.6$ is shown in Figure 3 in the form of a fraction of particles $N_i$ of the corresponding diameter $d_i$ from the total number of bubbles in the flow region $N_s$. 
The transition from the mode L1 to mode L2 leads to an increase in $N_s = 0.62 \times 10^6$ up to $1.22 \times 10^6$. An increase in the flow rate leads to a more intensive involvement of bubbles in the turbulent pulsation movements, which are accompanied by bubble coagulation.

![Figure 3. The particle-size distribution $N_i/N_s$ for the different modes (a) L1; (b) L2.](image)

4. Conclusion

An Eulerian two-fluid model is presented to describe the process of transfer of polydisperse bubbles in turbulent gas-liquid flows with significant gas concentrations of the gas phase. A modified two-parameter model of turbulence $k-\varepsilon$ is used to simulate the turbulent flow of a liquid phase. The bubble-size distribution in a gas-liquid flow is taken into account on the basis of the partitioning of the entire particle system into separate groups (fractions) with fixed diameters, for each of which the equation of particle conservation is recorded, considering the processes of coalescence and breakage of bubbles. Based on a comparison of the results of a numerical model with the experimental data, the applicability of the proposed model for the numerical simulation of gas-liquid flows with high values of the gas concentration of the dispersed phase in a bubble upward flow in a vertical pipe is demonstrated.

References

[1] H A Jakobsen 2013 *Chemical reactor modelling. Multiphase reactive flows* (Springer)
[2] R I Nigmatulin 1987 *Dynamics of multiphase flow* (M.: Science)
[3] Ishii M, Hibiki T 2011 *Thermo-Fluid Dynamics of Two-Phase Flow* (Springer)
[4] Drew D A, Passman S L 1998 *Theory of Multicomponent Fluids* (Springer)
[5] Liu T J, Bankoff S G 1993 *Int. J. Heat Mass Transfer* 36 1049-1079
[6] Gubaidullin D A, Snigerev B A 2018 *High Temperature* 56 61-72
[7] Hibiki T, Ishii T, Xiao Z 2001 *Int. J. Heat Mass Transfer* 44 1869-1882