Shear stress in downward adiabatic annular gas-liquid flow

M V Cherdantsev, S V Isaenkov and A V Cherdantsev*  
Kutateladze Institute of Thermophysics, Novosibirsk, Russia  
*E-mail: cherdantsev@itp.nsc.ru

Abstract. Time-averaged shear stress was measured in annular gas-liquid flow. The experiments were conducted in a wide range of phases flow rates, in two pipes of different diameters, with working liquids of different viscosities. The results were compared to a number of empirical correlations. It was shown that for a fixed gas superficial velocity the shear stress shows the same linear dependence on the inlet wetting density for all pipe diameters and liquid viscosities.

1. Introduction
Annular gas-liquid flow occurs in many industrial applications in nuclear, chemical, oil-and-gas industry and thermal power engineering. It consists of liquid film travelling along duct's walls and turbulent droplet-laden gas core in the duct centre. Film surface is covered by a complex system of interfacial waves. The waves strongly affect the integral characteristics of the flow, such as heat transfer, pressure drop, droplet entrainment rate, etc. Ability to predict these characteristics is vital for safe and efficient operation of industrial plants, as well as for design of new industrial facilities.

This paper is devoted to analysis of pressure drop in annular gas-liquid flow. Pressure drop in annular flow may be orders of magnitude larger than in single-phase gas flow of the same speed [1]. Such increase occurs due to interaction of the gas stream to the interfacial waves, acting as elements of wall roughness, and due to droplet entrainment and deposition processes. Complexity of the waves and gas-waves interaction makes it difficult to develop fully physical models, able to predict the pressure drop. Thus, the existing models are mostly empirical correlations, predicting interfacial friction in annular flow based on the inlet flow rates of both phases and, frequently, mean thickness of liquid film. Several such correlations [2-5] will be used for comparison in Section 3 of the present paper.

To reduce the degree of empiricism in prediction of integral flow characteristics, it is necessary to focus on the interfacial waves and their effect on gas core and liquid film flow. E.g., in [6], a model of shear stress and entrainment in annular flow, was developed based on two-zone approach, dealing separately with large-scale disturbance waves and base film between them. In a cycle of our papers (see, e.g., [7-12]) a consecutive study of structure, formation and development of waves of different types in annular flow was conducted. In this paper we present the results of pressure drop measurements in exactly the same conditions. These measurements are expected to be useful in reaching better understanding of the flow hydrodynamics and interrelating the integral parameters of the flow to liquid film wavy structure in the future. At the present stage, we only aim at the main regularities of pressure drop behavior with inlet flow rates of the phases, pipe diameter and viscosity of working liquid.
2. Experimental details
The measurements were conducted in downwards annular gas-liquid flow, organized in the experimental rig shown in Fig. 1(a). Working liquid, initially stored in the receiving tank (1), was pumped (2) into the pressure tank (3) prior to the experiments. During the experiments, the pump was switched off so that liquid flowed into the pipe under action of gravity. Liquid flow rate was controlled by the valve (4) using the float rotameter (5). Liquid entered the distributor (6) and, being pushed through a slot inlet, appeared as a film on the inner walls of a cylindrical working section (7). Passing through the pipe, the liquid left it through a cyclone separator (8) back into the receiving tank. The gas flow was supplied from a centralized line of compressed air. Its flow rate was controlled by the valve (9) and measured with an orifice flow meter (10) equipped with U-manometer (11); the latter was calibrated using a turbine gas flow meter. Gas entered the working section through a thin-walled (0.3 mm) metal tube, passing through the liquid distributor (6) and mounted coaxially into the working section pipe (7). Scheme of the inlet of gas and liquid is shown in Fig. 1(b). Two different pipes with inner diameters $D=11.7$ mm and $D=15$ mm were used. Actual volume of the liquid distributor was about 1 liter; width of the slot is 0.5 mm and its length is 60 mm.

Figure 1. (a) Scheme of the experimental setup. The numbers are explained in the text. (b) Scheme of the inlet. (c) Scheme of pressure drop measurements.

Four working liquids were used in the experiments: water at working temperature of 16°C (kinematic viscosity $v_0=1.15\times10^{-6}$ m$^2$/s) and three water-glycerol solutions: WGS1 (viscosity
$v_1=1.5 \times 10^{-6} \text{ m}^2/\text{s}$, WGS2 ($v_2=1.9 \times 10^{-6} \text{ m}^2/\text{s}$) and WGS3 ($v_3=3 \times 10^{-6} \text{ m}^2/\text{s}$). The gas was always air at nearly atmospheric pressure, superficial gas velocities, $V_G$, were within the range of 18-70 m/s. Reynolds number of liquid phase, $Re_L$, was defined as $Q_L/\pi D \nu$, where $Q_L$ is volumetric liquid flow rate at the inlet. Liquid Reynolds number varied between 140 and 500. The working temperature mentioned above was not set externally; this is an equilibrium value of temperature reached after circulation of the working liquid through the rig being sheared by high-velocity gas stream for about 1-2 hours. After this stage, the working temperature, controlled by thermometer measurements in the receiving tank, was constant for all working liquids, being equal 16$\pm$1ºC. The viscosities were measured at this temperature.

Time-averaged pressure drop in the flow was measured using a differential manometer. Manometer outlets were connected with flexible pipes to thin-walled tubes with inner diameter of 2 mm. The ends of both tubes were flush-mounted into the inner wall of the duct (see Fig. 1c). During the experiments, the manometer, the tubes and the connecting pipes were filled with working liquid. Manometer calibration was performed in a hydrostatic manner, based on a difference between heights of the liquid columns in pipes connected to the manometer's outputs. Downstream distance between the tubes, $L$, was equal 60 mm for lower gas velocities and 40 mm for high gas velocities. Average thickness of liquid film was measured using Brightness-Based Laser-Induced Fluorescence (BBLIF) technique, which is described in detail in, e.g., [11]. The measurements of both pressure drop and film thickness were taken at distances of about 40 pipe diameters below the inlet for both pipes.

3. Measurement results

The results of pressure drop measurements are presented in Fig. 2 for two different pipes. Pressure drop obviously increases with both gas and liquid flow rates. At the same time, difference in liquid viscosity does not show any noticeable effect: at the same $V_G$ and close values of $Q_L$ pressure drop shows very similar values at different viscosities. Slight variations with viscosity are observed occasionally (see, e.g. circles in Fig. 2b), but these dependencies are also characterized by different $Q_L$ and change accordingly. The ratio of liquid Reynolds numbers for the cases under comparison is different from unity (namely, it is equal to inverse viscosity ratio); nonetheless, pressure drop and shear stress remain the same.

![Figure 2. Pressure drop measurements in two pipes: (a) D=15 mm; (b) D=11.7 mm.](image-url)
Pressure drop is noticeably higher in the smaller pipe for the same combination of flow rates; average ratio of pressure drops in similar conditions is equal 1.79 with standard deviation of 0.3. This is not surprising, since larger pressure difference is required to push the gas-liquid mixture through a pipe of smaller diameter. At the same time, interfacial shear stress is much more informative parameter than the pressure drop, when it comes to describing the flow of a gas-sheared liquid film. Interfacial shear stress was calculated based on the pressure drop, $\Delta P/L$, and time-averaged film thickness, $<h>$, as follows:

$$\tau = \frac{\Delta P}{L} \left( D - 2 <h> \right) \frac{4}{D}.$$ 

Thus, at the same shear stress pressure drop will increase in smaller pipe roughly as $D^{-1}$. Besides, in a smaller pipe the wetting density, defined as $Q_L/\pi D$, will increase in the same manner at constant $Q_L$. So, assuming that shear stress is independent on pipe diameter and defined by $V_G$ and $Q_L/\pi D$, the ratio is expected to be equal to squared ratio of diameters (1.64), which is within the standard deviation range. Contributions of entrainment / deposition of droplets and gravity effect on gas phase into shear stress were neglected.

![Figure 3](image.png)

**Figure 3.** Comparison of interfacial friction coefficient data to the existing correlations from literature.

Fig. 3 shows a comparison of the experimental data to 4 different empirical correlations from literature. These papers provide correlations for interfacial friction, $f_i$, defined as $2\rho_g V_G^2$. Namely, in [2] $f_i$ is predicted as:

$$f_i = 0.005 \left( 1 + 300 \frac{<h>}{D} \right);$$

in [3] it is predicted as:

$$f_i = 0.316 \text{Re}_G^{0.25} (1 + 13.8 \text{We}^{0.175} \text{Re}_G^{0.7});$$

in [4] as:

$$f_i = 3.143 \times 10^{-4} + 1.158 \frac{<h>}{D};$$

and in [5] as:
It can be seen that the best prediction is given by the classical work [2], with most data within 20% error band. Correlation [4] overpredicts the current data by about 25% in general. The two other correlations show strong underprediction, especially the latest one. The discrepancy is much higher at large \( f_i \) values, i.e., at the largest liquid flow rates and the smallest gas flow rates. It can be concluded that the empirical modeling of pressure drop in annular flow is still far from complete, and further search for new approaches might yield useful results.

One flaw of the majority of the correlations regarded above is that they rely on average film thickness, which is not an input parameter of the flow, but has to be measured in order to predict pressure drop. It would be useful to construct new correlations based only on input parameters, such as inlet flow rates and physical properties of the phases and duct geometry. As it was mentioned above, at fixed gas velocity the shear stress seems to be mostly defined by the wetting density, i.e., ratio of the volumetric inlet flow rate of liquid, divided by wetted perimeter of the duct.

![Figure 4](image-url)Dependence of the shear stress on wetting density for different gas velocities. Set of flow parameters corresponding to a particular wetting density value is given in the legend on the right.

Such a representation of the data is shown in Fig. 4. For each value of superficial gas velocity, a dependence of shear stress on wetting density is shown for all liquid flow rates, all liquid viscosities and both pipe diameters. This gives about 10-20 points for each case, which all fall onto the same straight line with a relatively low scatter.

4. Discussion and conclusions
The observation illustrated by Fig. 4 is relatively unexpected on a number of reasons. Firstly, the independence of the shear stress on pipe diameter implies that the main interaction between gas core and liquid film occurs within the vicinity of the interface and is not that sensitive to what happens in the centre of the pipe. E.g., difference in pipe diameters at the same superficial gas velocity would yield different degrees of turbulence, but it does not seem to noticeably affect the shear stress. It also implies that the best parameter to describe the gas phase is its superficial velocity, rather than its Reynolds number. At the same time, it is better to characterize liquid flow by dividing its volumetric...
flow rate by the wetted perimeter ("wetting density") instead of pipe cross-section area (conventional "liquid superficial velocity").

Second arguable moment is the independence of the shear stress on the viscosity of working liquid. It implies that the liquid phase contribution into shear stress should not be described by its Reynolds number, but by dimensional wetting density. It is interesting that liquid viscosity affects the properties of disturbance waves such as speed and frequency, and these properties are better described by liquid Reynolds number than by wetting density (see [8]). This contradiction requires further investigation.

It is also interesting to compare this observation to the Wallis [2] correlation, which showed the best agreement with the present data. In this correlation the contribution of liquid phase is described by a single linear function of average film thickness. This quantity increases with wetting density, though the dependence is much weaker than linear: it would have been a square-root dependence if the velocity profile in the film was linear. At the same time, mean film thickness decreases with gas velocity, and this is the reason why the correlation reflects decrease in the interfacial friction with increasing gas velocity. It seems that the two competing effects compensate each other in terms of interrelation between the shear stress and mean film thickness, thus making the correlation work.

In future, it would be possible to construct a correlation for shear stress in form of:

\[ \tau = \frac{k(V_G)Q_L}{\pi D} + b(V_G), \]

based on dependencies shown in Fig. 4. For the present experiments, the dependencies \( k \) and \( b \) are:

\[ k(V_G) = 1875(V_G - 4.2), \quad b(V_G) = 0.014V_G^2 - 0.35V_G + 1.2, \]

though of course these coefficients are purely empirical and thus unlikely to work in different conditions. Future analysis should be aimed at providing more physical relationships for \( k \) and \( b \) coefficients by testing this approach against greater variety of flow conditions. Still, the more important goal of the future work is to interrelate the shear stress data to the wavy structure of annular flow.

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