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Turbulence modification in annular gas/liquid flow

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The subject of turbulence in a continuous phase and how it is modified by a second phase is a subject to which Professor Gad Hetsroni, whose 65th birthday is honoured in this Festschrift issue, has made significant contributions over many years. That it is a complex subject is well recognised. However, the problem is made more difficult by the differing features of the combination of phases that can co-exist. They each bring specific aspects.

Abstract

Measured values of turbulence intensity for annular gas/liquid flow are examined and contrasted with values from gas/solids flow. It is suggested that the larger values (relative to gas only values) found in gas/liquid flow are due to two contributions: the rough interface of the wall film and drops which are slow moving just after their creation from the wall film. A significant portion of these have Reynolds numbers >400 and, therefore, contribute to turbulence production unlike equivalent but faster moving particles in gas/solids flow. © 1999 Elsevier Science Ltd. All rights reserved.

Keywords: Gas/liquid; Drops; Turbulence; Annular flow

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1. Introduction

Early studies of turbulence modification, in pipe flows, by a second phase, include the work by Kada and Hanratty (1960) and by Boothroyd (1967) who inferred changes to turbulence from measurements of dispersion of a tracer. Subsequently, direct measurements were made using Laser Doppler techniques. For gas/solids flows there is information from Lee and Durst (1982), Maeda et al. (1980), Tsuji and Morikawa (1982), Tsuji et al. (1984) and Hosakawa et al. (1998). Zisselmar and Molerus (1979) published data for solid/liquid flows and Azzopardi and Teixeira (1994a, 1994b) for gas/liquid annular flows.

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From the data where the dispersed phase was solids, Gore and Crowe (1989) identified two types of behaviour, enhancement or suppression of turbulence. They showed that which of these occurred depended on the ratio of particle size to a turbulent length scale. If the ratio was greater than 0.1, there was enhancement. Values below this resulted in suppression. In contrast to this, the gas/liquid data of Azzopardi and Teixeira (1994a, 1994b), in the annular flow pattern, showed enhancement of the turbulence intensity even though the values of the mean drop size/ turbulent length scale ratio were below 0.1. This paper will examine the available data, consider models, present new data on drop velocities and attempt to explain this anomalous result.

In the only reported study in which the turbulence has been measured in annular gas/liquid flow in a vertical pipe, Azzopardi and Teixeira (1994b) used a Laser Doppler anemometer to measure two simultaneous components of velocity of 1 μm polystyrene tracer particles injected into the gas. The two-colour visibility technique described by Yeoman et al. (1982) was used to differentiate between these tracer particles and the water drops inherent in annular flow. This was possible, as the water drops were all much bigger than the tracer particles. In addition, in a complimentary study on the same apparatus, Azzopardi and Teixeira (1994a) measured sizes and velocity of drops using a Phase Doppler anemometer (PDA). The measured data was carefully examined to ensure that it was meaningful. For example, measurements made with the same equipment on single-phase gas flow showed profiles of mean velocity and turbulence intensity similar to the data of Laufer (1954). In addition, friction factors inferred from the extrapolation of the Reynolds stress to the wall agreed with the values from pressure drop. In two-phase flow, there was similar agreement between measured Reynolds stresses across the centre of the pipe and those inferred from pressure drops. Moreover, the exponent of the power law fit to the mean velocity profile had the same relationship with friction factor as the data from experiments from smooth and rough walled pipe carried out by Nikuradse (1932) and Nunner (1956).

Annular flow is characterised by liquid flowing partly as a film on the walls and partly as drops carried by the gas. There is constant interchange between the film and the drops. The film is very wavy and so appears to the gas as a rough wall. The main features are illustrated with data from one combination of gas and liquid flow rates. The gas superficial velocity was 30.9 m/s whilst that for the liquid was 0.016 m/s. The fraction of liquid travelling as drops was determined to be 0.14 and the mean film thickness was calculated as 0.14 mm. Fig. 1 presents the mean velocity profile for the gas together with that for the equivalent single-phase case. This shows that the velocity profile in the two-phase flow is more peaked than the gas only one. This has been observed in other annular flow studies where the mean velocity profile was measured by Pitot tube. This result is in contrast to those for gas/solids flows where the velocity profile is found to be flatter in the presence of particles. Fig. 1 also shows the measured values of drop velocity, mean and ± one standard deviation. Here, again the profile is peakier than the equivalent gas/solids case and the spread is increasing towards the wall in contrast to gas/solids data (cf. Hosakawa et al., 1998).

Fig. 2 shows how the gas turbulence intensity for both the single and two-phase cases varies across the channel. This illustrates how the presence of the second phase both increases the values and radial gradient of this parameter. Again, this is in contrast to the gas/solids case where the turbulence intensity profile is flatter than in the gas only flow.

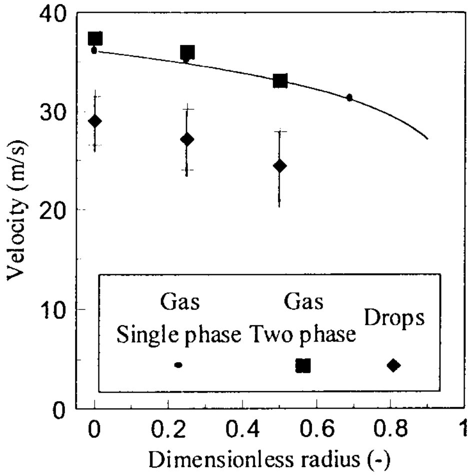

Fig. 1. Mean velocity profiles — gas superficial velocity = 30.9 m/s; liquid superficial velocity = 0.016 m/s; fraction of liquid entrained as drops = 0.14; film thickness = 0.014 mm; pressure = 150 kPa absolute; pipe diameter = 0.032 m.

Azzopardi and Teixeira (1994b) considered interface roughness in their attempt to explain the enhancement of turbulence intensity. Data from published literature indicate that, in both smooth and rough walled pipes, the radial profiles of turbulent intensities follow a standard curve when non-dimensionalised with the correct friction velocity. Azzopardi and Teixeira analysed their data in this way using interfacial friction velocities deduced from the two-phase pressure drop. These values were confirmed via interfacial shear stresses determined from measured profiles of Reynolds stress, $u'v'$. The measured turbulence intensities lie 25–96%

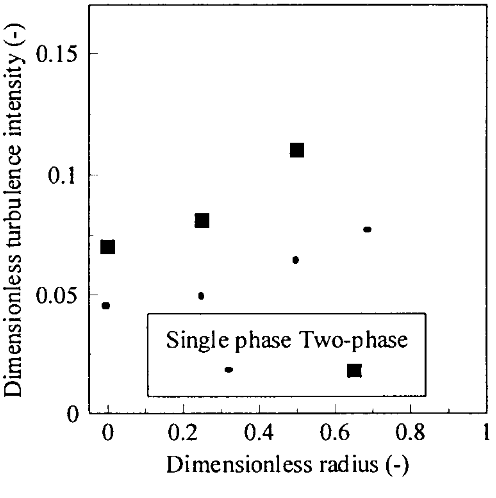

Fig. 2. Turbulence intensities — gas superficial velocity = 30.9 m/s; liquid superficial velocity = 0.016 m/s; fraction of liquid entrained as drops = 0.14; film thickness = 0.014 mm; pressure = 150 kPa absolute; pipe diameter = 0.032 m.

above the values expected from the interfacial friction velocity. Therefore, assuming that the interfacial friction velocity is an adequate measure of the interface-generated turbulence, there must be another source of turbulence that must be taken into account. This probably lies in the dispersed flow in the centre of the channel.

2. Mechanisms for turbulence modification

Several mechanisms have been advanced to explain the effect of a dispersed phase on the turbulence intensity of the continuous phase. Crowe et al. (1998) have listed a number: displacement of the flow field around a dispersed phase element; generation of wakes behind particles; dissipation of turbulence transfer of turbulence energy due to the motion of the dispersed phase; modification of velocity gradients in the carrier flow field and corresponding change in turbulence generation; introduction of additional length scales which may influence turbulence dissipation; disturbance of the flow due to particle/particle interaction.

From an order of magnitude analysis, Hetsroni (1989) deduced that particles with Reynolds numbers above 400 would augment the turbulence due to vortex shedding from the particles whilst those with lower values of Reynolds number would attenuate it. This is supported by the data of Tsuji et al. (1984) who showed that 200 μm particles have values of Reynolds number of 8–44 and diminish turbulence, and that 3000 μm particles have Reynolds numbers of 800–840 and increase turbulence. The recent data of Hosakawa et al. (1998) shows a similar result: 910 μm particles with Reynolds numbers of 350 caused an increase in turbulence. Yaun and Michalides (1992) suggested that the velocity defect in the wake of particles was responsible for the augmentation of turbulence and the work associated with the motion of the particles caused attenuation of turbulence. They provide appropriate descriptions for these terms and report good agreement with experiments. Yarin and Hetsroni (1994) developed an analysis along similar lines but employed a more detailed description of the wake. Their model correlates turbulence generated solely by particles at low concentrations.

Kenning and Crowe (1997) proposed that the level of turbulence would be determined by the balance of the inherent turbulence production and that due to the particles, the dissipation due to the particles and viscous dissipation. They also defined a hybrid length scale, a combination of the inherent length scale and an average inter-particle spacing. This resulted in an equation for the change in turbulence intensity.

$$\frac{\sigma-\sigma_i}{\sigma_i}=\left[\frac{L_\mathrm{h}}{L_\mathrm{i}}+\frac{L_\mathrm{h}}{k_\mathrm{i}^{3/2}}\frac{f(u-v)^2}{\tau_\mathrm{p}}\frac{\tilde{\rho}_\mathrm{p}}{\tilde{\rho}_\mathrm{g}}\right]^{1/3}-1 \tag{1}$$

where σ is the turbulence intensity, k_i is the turbulence intensity for the corresponding single-phase flow. L_h is a hybrid length scale, L_i is the length scale for the corresponding single-phase flow, u is the continuous phase velocity, v is that of the particle, f is the ratio of particle drag to Stokes drag, τ_p is the particle aerodynamic response time and $\tilde{\rho}_\mathrm{p}$,$\tilde{\rho}_\mathrm{g}$ are the bulk densities of the particles and continuous phase.

Crowe and Gillandt (1998) have derived an equation for gas phase turbulence energy in a fluid particle flow. This includes terms, which: (i) account for turbulence production by gradients in the average velocity of the gas phase and work due to the fluid/particle drag, (ii) describe the redistribution of energy between the phases and (iii) identify dissipation terms dependent on local gradients and that for dense flows is related to the interparticle distance. Their paper does not carry out a comparison with data.

Hosakawa et al. (1998) showed that turbulence intensity could be correlated by the ratio of particle induced eddy viscosity to the wall induced eddy viscosity. They determined that their data and those from other workers could be correlated using cross-section-averaged values, this ratio being defined as

$$\phi = \frac{\bar{u}_\mathrm{r} d}{\bar{u}'_\mathrm{i} D} \tag{2}$$

where u_r is the average relative velocity, u'_i is the average turbulence intensity of the continuous phase, d is the particle size and D is the pipe diameter.

3. Drop velocity data

Experiments were carried out on a facility for co-current flow of water and air at the School of Chemical, Environmental and Mining Engineering, University of Nottingham. In this, filtered air is drawn from a constant pressure receiver supplied from the compressed air main. It was metered by means of an orifice plate and fed into the bottom of the vertical test section. This was constructed from stainless steel pipe of 0.038 m internal diameter. Water was pumped from a supply tank, metered by a variable area meter and introduced into the test section through a porous wall section situated 0.5 m from the start. The measuring section was positioned 4.5 m beyond this point. There was a further 0.45 m of straight test section beyond this. The air and water emerging from the test section were separated in a large vessel, the water being returned to the supply tank, the air released to atmosphere.

Simultaneous measurements of drop size and velocity were made using a PDA technique employing a Lisatek instrument manufactured by AEA Technology Ltd (Liversey, 1988). The beam of a 200 mW Argon ion laser was split into two equal intensity beams. They are focused by a 1000 mm focal length and the scattered light collected at three positions (at an offset of 40° from the forward direction). This arrangement results in a viewed probe volume of $0.65 \times 0.62 \times 0.5$ mm, and produces an instrument with a wide dynamic range and from whose signals sphericity can be determined. Filtering of the signals is carried using an opto-electronic approach, and the signals are processed to yield the frequency of the Doppler burst (and hence velocity) and phase lag between signals from two detectors (and hence size). A microcomputer calculates the velocity and size and sorts and stores the data. For the optical geometry in these experiments, the size range is 5–650 μm.

Special test sections are required to allow for the entry of the light beams and for the exit of the scattered light without their being distorted by the curved tube wall and the highly disturbed film interface. The liquid film was withdrawn through a porous wall section just before the measuring point. Small holes were drilled in the tube wall and sealed, flat windows (25 mm diameter, 3 mm thick and flat to $\lambda/2$) were provided. A gas purge helped keep the windows free of drops. The purge flow rate was kept as low as possible so as not to affect the measurements. Extensive tests have shown that these arrangements do not affect the drop flow.

Water flow rates were measured by rotameters, which were calibrated by weighing a timed efflux. They could be read within 3%. Air flow rates were measured by a standard orifice plate. The pressure drop across the orifice plate was measured by a water manometer, which could be read within 1%. Similarly, gas pressure and, hence, density could also be determined with an accuracy of better than 1%.

Drop sizes and velocities have been measured on the facility described above for a range of gas and liquid flow rates. Superficial gas velocities of 20 and 30 m/s were used. The liquid mass fluxes were between 20 and 101 kg/m^2 s. For each combination of flow rates, measurements were made over the centre 16 mm of the pipe. Each data set consisted of between 7000 and 20,000 points. An example result is given in Fig. 3, as velocity versus diameter. Also plotted is the average velocity for specific ranges of diameters. There is a small trend for this average velocity to decrease with increasing drop size. A scatter plot such as that of Fig. 3 hides much information. A better feel for the relative occurrence of the different velocities can be seen in Fig. 4.

In determining mean values, which might be calculated from the data, consideration has to be given to any biasing within the data. One particular bias concerns the size of probe volume from which drops of different diameters are measured. Because of the Gaussian intensity distribution of light within the probe volume, larger diameter drops could come from a larger probe volume and so could be over-represented. Brazier et al. (1988) suggest a method for compensating for this. When applied to the present data, only a small change is produced. Another possible bias is related to the differential residence time of drops of different velocities in the probe volume. However, as seen in Fig. 3, there are only small changes in the mean velocities of drop of different sizes. Again this effect is not considered important. A further possible effect has been identified by Tropea et al. (1996) and concerns the reception of sufficient reflectively scattered light by a PDA set up to receive refractively scattered light to alter the phase lag/diameter ratio from expected values. Tropea et al. suggest a more

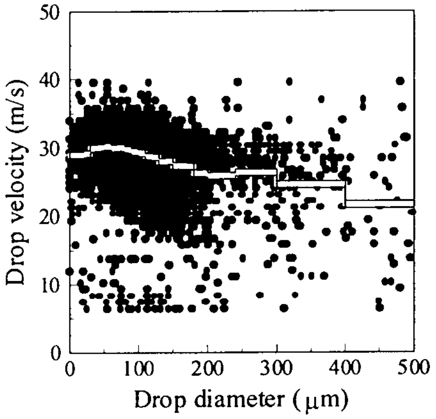

Fig. 3. Variation of drop velocity with drop diameter at centre line, gas superficial velocity = 30 m/s; liquid superficial velocity = 0.02 m/s; pressure = 150 kPa absolute; pipe diameter = 0.038 m.

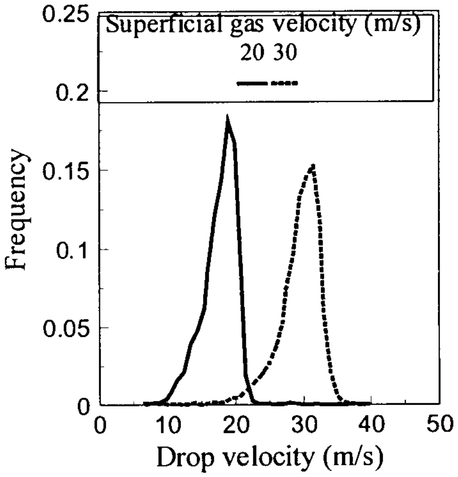

Fig. 4. Frequency distribution of drop velocities at centre line, liquid superficial velocity = 0.02 m/s; pressure = 150 kPa absolute; pipe diameter = 0.038 m.

complicated arrangement, the Dual-mode PDA, to overcome the problem. This is essentially two PDA units working together: one is a standard arrangement, the other... Tropea et al. show that the use of a dual-mode PDA eliminated a secondary peak of large drops in the volume distribution which were spurious. Here, we have taken a simpler approach. The mean velocity was calculated for the entire data set and for a subset that excluded those drops, which constituted the second (probably spurious) peak, in the volumetric drop size distribution. As the difference between these two results was small (about 1%) values from the entire data set are presented. A similar conclusion is reached for the arithmetic mean diameter. However, in the case the more commonly used Sauter mean diameter (volume/surface mean) it has been found that this parameter is much more susceptible to small changes in the number of larger drops. This aspect has been considered by Zaidi et al. (1998) who showed that, if the secondary peak is removed, the drop size distributions determined from PDA and a laser diffraction instrument were in reasonable agreement.

Examples of the radial distribution of mean drop velocity are illustrated in Fig. 5. This shows trends and scatter similar to that reported by previous workers. Fig. 6 shows the effect of liquid flow rate on the standard deviation of drop velocity (non-dimensionalised by the mean value). These drop velocity data show values and trends similar to those previously published by Tayali et al. (1990), Azzopardi and Teixeira (1994a) and Fore and Dukler (1995). Data from all sources show that the mean drop velocity at the centre-line is of about the same value as the gas superficial velocity. Given the shape of the gas velocity profile, this indicates that, in the centre of the pipe, the relative velocity between the drop and the gas is 20/30% of the gas superficial velocity. The spread of velocities is only provided by Azzopardi and Teixeira (1994a), Fore and Dukler (1995) and the data given above. Again, all show similar values of dimensional standard deviation. In the case of Fore and Dukler, this was not presented explicitly and the agreement was inferred from the half height width of frequency plots in their paper.

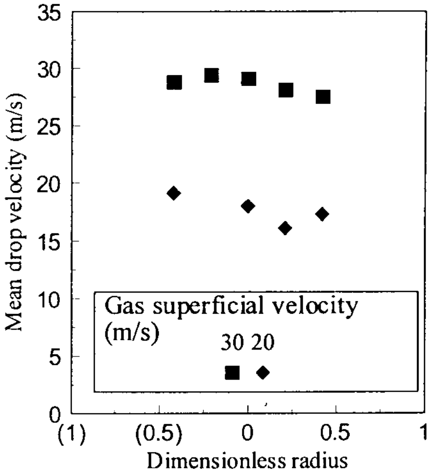

Fig. 5. Radial variation of mean drop velocity — liquid superficial velocity = 0.02 m/s; pressure = 150 kPa absolute; pipe diameter = 0.038 m.

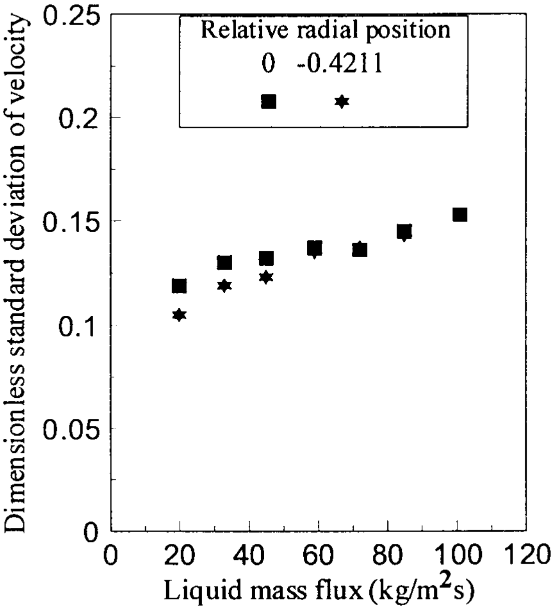

Fig. 6. Effect of liquid flow rate on dimensionless standard deviation of drop velocities — gas superficial velocity = 30 m/s; pressure = 150 kPa absolute; pipe diameter = 0.038 m.

4. Discussion

The detailed measurement from annular gas/liquid flow made by Azzopardi and Teixeira (1994a, 1994b) and presented above can be used to test the applicability of models and correlations identified above to annular flow. For the approach of Kenning and Crowe (1997), the data presented in Figs. 1 and 2 yields a relative velocity of 8.4 m/s, a characteristic time, $\tau_p = 0.037$ s, a drag ratio of 4.36 and bulk densities of 0.076 and 1.8. This results in a change in turbulence of $+0.044$. However, in these calculations a mean drop size was employed to calculate the hybrid length scale. In reality in annular flow there is not a single drop size, but a wide distribution of sizes. This would result in a shorter inter drop distance and, hence, a smaller hybrid length scale. Using this in Eq. (1) would result in a negative value and, thence, indicate damping of turbulence.

The data of Azzopardi and Teixeira (1994a, 1994b) has been used to test the applicability of the correlation proposed by Hosakawa et al. (1998) to annular flow. From Figs. 1 and 2 values of mean relative velocity and mean inherent turbulence intensity of 8.6 and 3.23 m/s are obtained. These together with a mean drop size of 110 μm and a pipe diameter of 0.032 m result in a value of the ratio of 0.009. From the correlating graph of Hosakawa et al. (1998) this corresponds to a damping of turbulence.

Now, as the methods considered above predict damping of turbulence in contrast to the increase measured by Azzopardi and Teixeira (1994b), alternative or addition mechanisms are required to explain the change in turbulence intensity in annular gas/liquid flow. An obvious source of the extra turbulence is the rough interface presented to the gas by the wall film. However, it was pointed out above that this was not sufficient. The drop velocity data of Azzopardi and Teixeira (1994a) and the new data described above provide possible clues. If Reynolds numbers for the drops are calculated based on average values, they are typically about 100. However, as shown in Fig. 3, there are significant ranges of both drop size and velocity. If Reynolds numbers are calculated for individual drops, the distribution of probability can be obtained. Fig. 7 illustrates that there are some drops with Reynolds numbers >400, but these are only a very small proportion of the total. Similar results would be inferred from the data of Tayali et al. (1990) and Fore and Dukler (1995). Obviously, this in itself is not the explanation. The answer probably lies in the fact that drops are constantly

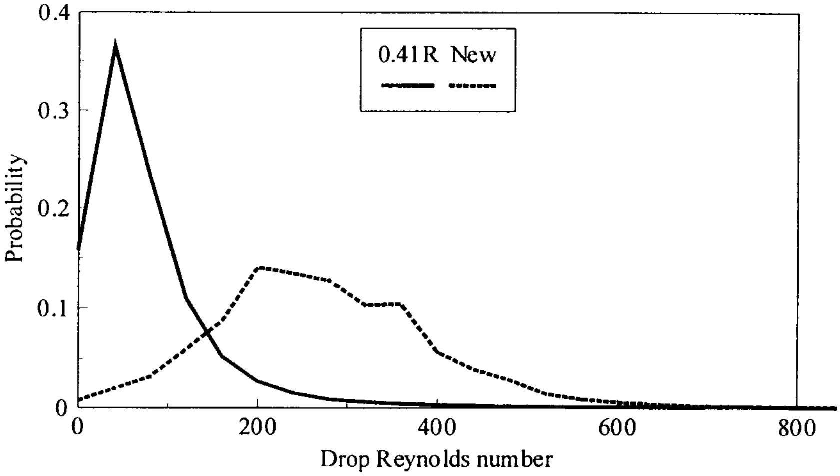

Fig. 7. Probability distribution of drop Reynolds numbers — gas superficial velocity = 30 m/s; liquid superficial velocity = 0.02 m/s; pressure = 150 kPa absolute; pipe diameter = 0.038 m.

being creased from the wall film. On acquiring independent existence, these drops would have velocities far below that of the surrounding gas. Though there are no measured values at the correct proximity to the wall film, their initial axial velocity can be approximated to that of the waves on the film from which they were created. The consequent distribution of Reynolds numbers has been estimated using the distribution of drop sizes measured nearer the pipe centre-line. They are also plotted in Fig. 7 and show that a greater proportion have Reynolds numbers >400 and are probably creating extra turbulence by vortex shedding.

The augmentation of turbulence intensity reported by Azzopardi and Teixeira (1994b) appears to disagree with the suppression of turbulence reported by Owen and Hewitt (1987). However, it is noted that Hewitt and Owen based their statement purely on the analysis of mean velocity profiles for the gas as measured by, e.g., Gill et al. (1964). These profiles they fitted to a log law equation. Using a friction velocity obtained from the two-phase pressure drop, they determined a 'two-phase von Karman constant' which they found differed from the classical single-phase value and which correlated with the ratio of gas superficial momentum to that for the gas/drop mixture. They related the change in the von Karman constant to suppression of turbulence.

The value of turbulence intensity increases with rate of entrainment. This provides further support for the idea that the increase in turbulence intensity is due to newly created drops.

5. Conclusions

From the above a simple concept for the change in turbulence intensity, for annular flow over the corresponding gas only flow, can be proposed. There are two reasons why the turbulence in augmented. Firstly, there is the presence of the liquid film, whose wavy interface acts as a rough wall to the gas. Secondly, there is the presence of newly created drops, whose low velocity relative to the gas means that they are capable of shedding vortices and so creating extra turbulence.

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