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Persistence of frequency in gas–liquid flows across a change in pipe diameter or orientation

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We would like to wish Geoff Hewitt a happy 80th birthday. BJA would like to say that having worked in his Division for many years at AERE Harwell and collaborated with him for the last twenty plus years, he always found that he was given space to develop his ideas. This is an attribute not often found in some of our leaders but a clear feature of Geoff.

Keywords:
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ECT

ABSTRACT

From a study of the characteristics of structures across a 67/38 mm sudden contraction, using air/silicone oil flows, it has been found that frequencies of the structures (mainly slugs) persist across the contraction. This is in contrast to the velocities and lengths which increase as they move into the smaller diameter pipe. These observations were found for both vertical and 5° upward orientations. A similar persistence of frequency has been found from four other sources in the literature: a vertical (gradual) contraction; a horizontal Venturi; and two cases of horizontal pipe, 90° bend and vertical riser combination. The latter were at two contrasting conditions: (i) at atmospheric pressure with air/water in small diameter (34 mm) pipes; (ii) at 20 bar in larger diameter pipes (189 mm) using nitrogen and naphtha.

1. Introduction

The upstream oil/gas production industry needs to calculate the flow rate/pressure drop/geometry relationships for complex piping systems as part of the design of production systems, particularly for offshore installations. In current methodologies, complex systems are divided into sections of constant diameter and orientation. The flow/pressure information is generated for this section, often by subdividing it into smaller axial steps. Where there are changes in pipe diameter, orientation, the additional pressure drop due to the fitting, i.e., the expansion, contraction or bend is then calculated before the next straight section of pipe is tackled. However, this ignores the transport of information through the system. This is clearest when the flow pattern is the same on both sides of the fitting. It has been most clearly illustrated by Saidj et al. (2014) who pointed out that, for same pipe diameter and flow rates, the frequency of slugs is found to be the same before and after the bend. This is significantly different from experiments with just straight pipes, where the gas and liquid are mixed at the entrance of the pipe. In those cases the slug frequency is much greater when the pipe is vertical than when it is horizontal.

Gas–liquid two-phase flow is probably the most common combination of phases in multiphase flow. Consequently, there has

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been much research in this area. There is strong interest from the part of the oil/gas industry engaged in extracting the oil and gas from reservoirs as well as from companies operating in oil refineries and hydrocarbon processing plant. In addition, there are also needs from the power generation industries (nuclear or fossil) and the chemical process industry (reboilers and condensers). The transport of gas–liquid flow plays a major role in their applications and clearly the ability to predict the fluid flow behaviour of these processes. Hence, research in this field of multiphase flow is very important from engineering and economical point of view to improve safety, reliability, sustainability, efficiency and significant decrease in frequency of maintenance multiphase flow equipment in for example the petroleum and chemical industry Azzopardi (2006). A persistent theme throughout the study of multiphase flow is the need to model and predict the detailed behaviour of those flows and phenomena that they manifest. There are three ways in which such models are explored: experimentally, through laboratory sized models equipped with appropriate instrumentation, theoretically using mathematical equations and models for flow and computationally using the power and size of modern computers to address the complexity of the flow, Brennen (2005). In many cases the laboratory model is of a different scale to the prototype, therefore a reliable theoretical model is essential for confident extrapolation to the scale of the prototype.

Various correlations and models for gas–liquid pressure drop across sudden contraction have been proposed by different authors. Chen et al. (2010) wrote a critical literature review carried out to compare popular pressure drop correlations. This included the homogenous pressure drop proposed by Collier and Thome (1994) and Chisholm (1983), momentum and mass transfer based correlation by Schmidt and Friedel (1997) and a mechanical energy equation model across the contraction. Their analysis showed that the homogenous model gave the best predictions when the pressure drops for steam-water mixture through a sudden contraction of different area ratios were measured. Attou and Bolle (1999) identified that most of the methods above do not take the flow pattern into consideration. They devised a pressure drop correlation which involves the balance equations deduced from macroscopic mass, momentum, entropy and energy conservation laws as well as a balance of forces exerted on the dispersed bubbles phases for bubbly flow. The model was found to be in good agreement with several experimental data of literature obtained for bubbly air–water flows through thick and thin orifice. However, it should be noted that pressure drop correlations used in thick orifices is a combination of that of sudden contraction and expansion. Correlations based on the two-phase singular pressure drop multiplier have been proposed by, e.g., Attou and Bolle (1999) and Fossa and Guglielmini (2002). These are functions of inlet void fraction of the dimensions of the fitting.

Fossa et al. (2006) studied the effect of different thickness of sharp edge orifice on an intermittent horizontal air–water gas–liquid flow by evaluating the local void fraction distribution downstream the orifice. The analysis showed that the void fraction usually reaches a maximum at a distance of about one diameter from the throat. Flow in the developing region and the developing length (downstream contraction) is also dependent on the upstream flow pattern and area ratio. Fossa and Guglielmini (2002) noted that this behaviour has been observed irrespective of the orifice thickness for higher values of liquid flow rate and even more evident when the area ratio is low.

The rise velocity of bullet shaped bubbles which occupy the greater part of the pipe cross-section was first studied analytically by Dumitrescu (1943) and Davies and Taylor (1950) who determined the rise velocity to have a value of $Fr\sqrt{(gD)}$, where D is the pipe diameter. They proposed values of Fr of 0.351 and 0.328 respectively. There have been further studies since, the most comprehensive of which is that of Viana et al. (2003) who examined the effects of liquid viscosity, surface tension and pipe diameter on the Froude number. They present an equation for Fr based of the Eötvös number ($=g\rho_l D^2/\sigma$) and a dimensionless inverse viscosity, which they term the gravity Reynolds number, ($=\sqrt{(gD^3[\rho_l-\rho_g]\rho_l)}/\eta_l$). Here ρ_l is the liquid density, ρ_g that of the gas, D is the pipe diameter, σ the surface tension and η_l the liquid viscosity. Their equation has 13 empirical constants. It has been shown, by Viana et al. (2003) to give accurate values of Fr for viscosities up to 3.8 Pa s. The above work was for bubbles rising in stagnant liquids. When there is finite gas and liquid flow rates, the Taylor bubble velocity is determined from an additive equation whose two terms are due to the bulk motion and to the drift velocity (that which would occur in stagnant liquids). This can be written as

$$u_B = C_0(u_{gs}+u_{ls}) + KFr\sqrt{gD} \tag{1}$$

With u_{gs}, u_{ls} being the gas and liquid superficial velocities respectively. Nicklin et al. (1962) reported a value of 1.2 for C_0 but noted that higher values were more appropriate as the flow rates diminished. Subsequently, this has been addressed by Collins et al. (1978), Dukler and Fabre (1994) and Guet et al. (2004). A more complicated expression was proposed for C_0 as given by Eq. (2),

$$C_0 = \frac{C_{BC}}{\left[1+\left(\frac{Re_m}{Re_c}\right)^2\right]} + \frac{C_{0,Re=\infty}}{\left[1+\left(\frac{Re_c}{Re_m}\right)^2\right]} \tag{2}$$

Values of C_{BC} = 5 or 2.27, Re_c = 4000 and $C_{0,Re=\infty}$ = 1.2. In this Re_m is defined as $\rho_l(u_{gs}+u_{ls})D/\eta_l$.

Hills (1975) and Hills and Darton (1976) noted that the velocities of larger bubbles were higher if they were travelling through a swarm of small bubbles rather than just liquid. They worked on large (square cross-section) columns and pipes where the large bubbles were of the spherical cap type. However, they also studied Taylor bubbles. In their air/water experiments they created bubbly flow of known void fraction and then released a large volume of gas. The velocity of the Taylor bubble was seen to increase as the bubbly void fraction increased. Similar results were obtained by Azzopardi et al. (2014b) who created slug flow in a 67 mm diameter, vertical pipe operated with air/water at zero liquid flow rate. The flow consisted of a series of Taylor bubbles interspersed with liquid slugs containing small bubbles. Again the velocity of the Taylor bubbles increased as the void fraction in the liquid slug increased. This data is illustrated in Fig. 1 together with data from Hills and Darton (1976) taken on a 50 mm diameter pipe (the closed symbols). Both set appear to lie on one curve. Hills and Darton proposed that the presence of small bubbles altered the shape of the large bubble making it more pointed which would increase

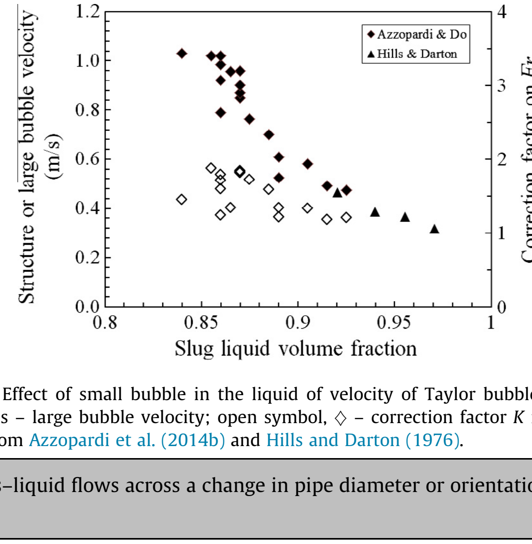

Fig. 1. Effect of small bubble in the liquid of velocity of Taylor bubbles. Closed symbols – large bubble velocity; open symbol, ◇ – correction factor K in Eq. (1). Data from Azzopardi et al. (2014b) and Hills and Darton (1976).

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