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Vertical upward intermediate scale Taylor flow: Experiments and kinematic closure

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ABSTRACT

The vertical upward Taylor flow regime has been extensively studied at the capillary and large channel scale limits. However, flow behavior at the intermediate scale ($5 \lesssim \text{Bo} \lesssim 40$, or $6\text{ mm} \lesssim D \lesssim 17\text{ mm}$ for ambient gas–water flows) is comparatively poorly characterized. This regime is fundamentally different because classes of forces conventionally associated with either small or large Bond number flows are all relevant. In this investigation, air–water Taylor-flow experiments are conducted in 6.0, 8.0, and 9.5 mm diameter tubes. High-speed video data are collected, and automated image analysis algorithms are developed to measure flow parameters including: bubble rise velocity, liquid film thickness, void fraction, and Taylor bubble and liquid slug lengths. New correlations and flow models are developed to predict these parameters at the intermediate scale. Results from this study enable kinematic closure of intermediate scale Taylor flows.

Introduction

Background

At intermediate liquid and gas flow rates in channels, elongated relatively large diameter bubbles translate, separated by full-diameter *liquid slugs* (Taitel and Barnea, 1990). At high velocities, turbulent shear at the elongated bubble tails can entrain small bubbles in, or *aerate*, the liquid slugs (Fernandes et al., 1983). Turbulent dispersion may prevent agglomeration of these small *slug bubbles* (Barnea and Brauner, 1985). At lower velocities, bubbles are not present in the liquid slugs, and the flow pattern is sometimes referred to specifically as *elongated bubble* or *plug flow*. In small diameter (capillary) channels, the elongated bubbles tend to be symmetric with rounded *noses* and *tails*. Sometimes this configuration is specifically referred to as *bubble-train flow* (Thulasidas et al., 1995). In the vertical upward (co-flow) configuration, the flow pattern is referred to as *Taylor flow*, after the bullet shaped bubbles studied by Davies and Taylor (1950).

In vertical upward Taylor flow, elongated rising bubbles fill a major portion of the channel cross-section, and are surrounded by downward flowing (gravitationally driven) annular liquid films on the walls. The interspersed cylindrical liquid slugs rise at the bulk superficial velocity (j), which is slower than the velocity of the Taylor bubbles ($U_b > j$). This flow pattern arises in, and is employed in, many engineering applications, including:

- **Airlift and bubble pumps** – In these devices, rising Taylor bubbles lift liquid, enabling mechanically simple pumping. Relatively high pumping efficiencies can be obtained in the Taylor-flow pattern because liquid only drains downwards in the thin films around elongated bubbles (Delano, 1998). End uses include well dewatering, nuclear fuel reprocessing (de Cachard and Delhaye, 1996), and single-pressure refrigeration systems (von Platen and Munters, 1928).
- **Miniaturized heat and mass exchangers** – The Taylor-flow pattern has been targeted for monolithic catalytic reactors as a high-interfacial area density, relatively low pressure drop, operating mode (Thulasidas et al., 1995). Additionally, in microchannels, Taylor flow can occur for a large portion of the flow length during phase change processes, such as evaporation and condensation (Garimella, 2004).
- **Petrochemical processes** – Taylor flow of hydrocarbon gases and liquids often occurs in oil-well bores (Fernandes et al., 1983), and it is often necessary to employ "slug catchers" in petrochemical processes to dampen the effects of intermittent flow.

Considering these examples, Taylor flow can occur over a large span of channel diameters (μms to cms), flow rates, and fluid

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properties (water to heavy oils). Forces due to inertia (fluid acceleration, turbulence), gravity (buoyancy), friction (within the fluid medium and at the tube wall), and surface tension may thus significantly affect flow behavior. However, comprehensive investigations have thus far been available only at the asymptotic limits where some of these forces are negligible.

Extensive literature is available for Taylor flows in large scale channels. This regime has been defined using criteria based on the Bond number ($\mathrm{Bo} = (\rho_L - \rho_G)gD^2/\sigma$), a measure of relative gravitational-to-surface tension force strengths. Here, ρ_L and ρ_G represent the liquid and gas densities, respectively, g is the gravitational acceleration, D is the channel diameter, and σ is the fluid surface tension. Viana et al. (2003) identified the following criterion for large-scale flows: $\mathrm{Bo} = (\rho_L - \rho_G)gD^2/\sigma \gtrsim 40$ ($D \gtrsim 17$ mm for ambient gas–water flow). For such cases, surface tension forces are negligible, leaving flows governed by inertial, viscous, and gravitational forces. Such configurations are of particular interest for petrochemical extraction and separation, and have generally been studied in either the viscous limit (for heavy products) or the inertial (turbulent) limit for high flow rates in large scale channels. Comprehensive flow models for these configurations have been reported by Fernandes et al. (1983), Sylvester (1987) and Taitel and Barnea (1990). Detailed shadowgraphy and particle image velocimetry studies have also been conducted to measure Taylor bubble profiles and near-bubble liquid velocity fields in the large scale regime (Bugg and Saad, 2002; Nogueira et al., 2006a,b).

Many investigations have also been performed for Taylor flows at the capillary and microchannel scale. Several criteria have been proposed for defining this regime ($\mathrm{Bo} \lesssim 0.9 - 19.7$ (Khandekar et al., 2010)). This study adopts the condition of $\mathrm{Bo} \lesssim 5$, the maximum Bond number for which a Taylor bubble will rise in a stagnant liquid medium (Bendiksen, 1985). In such cases, inertial and gravitational forces are small or negligible compared to viscous and surface tension forces. Applications of interest include monolithic catalytic reactors (Thulasidas et al., 1995), microchannel heat and mass exchangers (Garimella, 2000), and fuel cells (Anderson et al., 2010; Argyropoulos et al., 1999; Hussaini and Wang, 2009). Flow models and literature reviews for Taylor flows at this scale can be found in Thulasidas et al. (1995), Garimella (2004), Liu et al. (2005) and Angeli and Gavriilidis (2008).

In contrast, vertical upward Taylor flows in the intermediate scale regime ($5 \lesssim \mathrm{Bo} \lesssim 40$) where all four classes of forces are relevant, have not yet been well characterized (Reinemann et al., 1990). At this scale, the Taylor-flow pattern is particularly attractive for gas-lift-pump and bubble-pump applications (de Cachard and Delhaye, 1996).

Analytical or *mechanistic* Taylor-flow models have been proposed in the literature in which the flow pattern is modeled as identical repeating *unit cells* of liquid slugs and elongated Taylor bubbles (Fig. 1). Appropriate closure models can then be applied for parameters including bubble rise velocity (U_b), slug or bubble length (L_s and L_b), and frictional resistances (Fernandes et al., 1983; Sylvester, 1987; Taitel and Barnea, 1990; Thulasidas et al., 1995). Many correlations have been reported for such closure models, but few are applicable in the intermediate scale regime where all four aforementioned classes of forces can be significant. Additionally, intermediate scale Taylor flows tend to span the laminar-to-turbulent transition, indicating strong sensitivity to Reynolds numbers. This investigation focuses on measurement and prediction of the Taylor bubble rise velocity, liquid film thickness, void fraction, Taylor bubble length, and liquid slug length in intermediate scale flows.

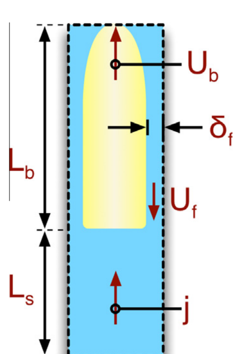

Fig. 1. Repeating unit cell model for Taylor flow.

Prior work

Taylor bubble rise velocity prediction

In large scale flows, the rise velocity of Taylor bubbles has generally been modeled in a drift flux fashion:

$$U_b = C_0 j + \Gamma\sqrt{gD} \tag{1}$$

The Γ term corresponds to the rise velocity of a bubble in a quiescent liquid medium (no net liquid flow). C_0 is referred to as the distribution parameter, and accounts for the fact that Taylor bubbles rise faster than the bulk flow because the surrounding liquid film drains downward. While the governing fluid mechanics equations are non-linear, this superposition approach for bubble velocity has proven successful in the literature. In general, these models assume that individual Taylor bubbles are long, sufficiently far apart (no drafting, Barnea and Taitel, 1993), and that liquid transport properties are dominant ($\rho_G/\rho_L \ll 1$, $\mu_G/\mu_L \ll 1$, where μ is the dynamic viscosity).

The Γ parameter is assumed independent of total flow rate (j); therefore, dimensional analysis indicates that it can be correlated in terms of the Bond number (Bo) and viscous number (N_f) (White and Beardmore, 1962).

$$\mathrm{Bo} = \frac{(\rho_L - \rho_G)gD^2}{\sigma} \qquad N_f = \sqrt{\frac{\rho_L(\rho_L - \rho_G)gD^3}{\mu_L^2}} \tag{2}$$

In the limit of negligible surface tension and viscous forces, Γ approaches approximately 0.35 (Davies and Taylor, 1950; Dumitrescu, 1943).

White and Beardmore (1962) found that surface tension effects on Γ are negligible for Bo > 70. They proposed the following correlation:

$$\Gamma = 0.345\left[1 - \exp\left(\frac{-0.01N_f}{0.345}\right)\right]\left[1 - \exp\left(\frac{3.37 - \mathrm{Bo}}{m}\right)\right] \tag{3}$$

$$m = \begin{cases} 10 & N_f > 250 \\ 69N_f^{-0.35} & 18 < N_f < 250 \\ 25 & N_f < 18 \end{cases} \tag{4}$$

Bendiksen (1985) modeled the Taylor bubble rise in the limit of negligible viscous forces and relatively small surface tension forces, and proposed the curve fit:

$$\Gamma = 0.486\sqrt{1 + \frac{20}{\mathrm{Bo}}\left(1 - \frac{6.8}{\mathrm{Bo}}\right)}\,\frac{1 - 0.96\exp(-0.0165\mathrm{Bo})}{1 - 0.52\exp(-0.0165\mathrm{Bo})} \tag{5}$$

This fit was reported to be valid for Bo > 5.7. Bendiksen (1985) also reported that $\Gamma = 0$ for Bo < 4.5; i.e. for low Bo (capillary) flow, individual elongated bubbles cannot rise without also effecting a