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Towards Understanding the Interfacial Structures of Non-developing Slug Flow in Vertical Pipes

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Abstract **–** Non-developing slug two-phase flows in vertical pipes are widely found in various industries. These flows are of a highly complex nature largely due to the deformability of the gaseous phase resulting in unstable interfacial flow structures at different flow regimes. Such complex interfacial structures strongly control the multiphase transport phenomena including energy, mass, and momentum transfer between the phases. Therefore, a clear understanding of the behaviour characteristics of these interfacial structures is critical to the optimum design of multiphase flow systems. This study briefly provides a review on the behaviour of the gas-liquid interfacial structures for the slug flow regime in co-current upward two-phase flows. This review founds that the interfacial structures of gas-liquid interface exhibit different shapes and behaviours in non-developed compared to the fully-developed regions of slug regime. The behaviour of these structures is found to be heavily influenced by gas injector design, pipe diameter, gas and liquid phase properties, and operating flow conditions. The review also showed that the interfacial structures have been widely studied in developed region, while they have remained less understood in the non-developed region. Also, the impact of liquid and gas phases' thermo-fluid properties (density, viscosity, and surface tension) and pipe diameter on the interfacial structures in this flow regime have received the least attention.

Keywords**:** Gas-liquid; Two-phase Flows; Interfacial Structures; Slug Flow Regime; Transition Regions

1. Introduction

Upward vertical gas-liquid two-phase flow systems, one of the most complex and important forms of two-phase flow systems [1], are widely found in various industries. These systems can play the role of a transporting medium [2], where the liquid phase is circulated or transported as a result of the momentum transfer between phases in addition to the buoyancy effect. Airlift pumping, food and aquaculture systems, sewage disposal, bioreactor, etc. [2] are the most obvious examples for mass, momentum and heat transport in gas-liquid systems. In these systems, gas-liquid two-phase flows act as a high-efficiency heat/mass transfer medium [2], where thermal energy and mass species transfer between the two phases through the direct contact. Some heat transfer applications of these systems can be found in power generation industry, where tubular boilers, evaporators, condensers, thermal solar collectors, geothermal wells, etc. are used [2], [3]. Also, take the Bubble Column Reactors (BCR) widely used in many food industries as another example. These reactors are working based on direct contact mass transfer in vertical gas-liquid two-phase flows for different purposes including but not limited to oxidation, hydrogenation, chlorination, carboxylation, absorbing and desorbing Carbon Dioxide (CO_2), aerobic fermentation, and so forth [4]. According to Martin et al. [5], around one-quarter of all reactions in industries occurs in gas-liquid two-phase mediums. Note that in some applications like Bubble Column Evaporators (BCE), aquaculture oxygenation/circulation systems, and vertical gas-liquid two-phase flows, the medium can be used for simultaneous transfer of momentum, mass, and energy.

Upward vertical gas-liquid two-phase flow systems have been proven to have a highly complex nature; this complexity mainly originates from the deformability and compressibility of the gaseous phase [1], [3] and discontinuities in phases' properties [6]. In these cases, the deformable and compressible gaseous phase results in a highly complex interface between the two phases [2]. These can be in the form of infinite combinations of interfacial structures like spherical, distorted, cap, slug, and churn bubbles together with ripples, huge and disturbance waves, wisps, droplets, and liquid films have been observed at different two-phase flow regimes [7], [8]. These structures show different behaviours at different flow regimes and even different axial (non- and fully-developed regions) and radial (from the pipe centre to the walls) locations of the same flow regime [8]. Furthermore, they are heavily influenced by gas injector design conditions, pipe diameter, gas and liquid phase properties, and flow conditions [9]–[13]. These all contribute to the complex nature of such systems. On the other hand, in upward vertical gas-liquid two-phase flow systems, the two-phase flow's hydrodynamic behaviour and subsequently, multi-phase transport phenomena are strongly controlled by these interfacial structures and the interactions they have with themselves and the opposite phase. Thus, to characterize the main controlling mechanisms of energy, momentum, and mass transfer in upward vertical gas-liquid two-phase flow systems, the first step is to understand the phase distribution within these interfacial structures.

In this study, the focus is merely on the interfacial structures in both non- and fully developed regions of slug flow regime such as in airlift pumps, vertical risers, vertical evaporators, etc [14]. This flow regime is of a highly clear periodic structure, where large bubbles (so called "Taylor" bubbles) move upward with a size nearly equal to the pipe diameter and lift the liquid slugs containing small bubbles [14]. The periodic nature of this flow regime as well as the interactions between its interfacial structures have been reported to induce degradation and diminution of pipe wall and vibrations [13]. Numerous studies have investigated the dynamic behaviour of slug regime; Fernandes et al. [15], Sylvester [16], and Barnea and Ullmann [17] proposed several models for the fully-developed region of this flow regime. Faber and Line [18] performed a comprehensive review of the available existing models developed for slug flow regime. In this paper, referring to previous studies, first, slug flow development in axial direction at both non- and fully-developed regions is described; then, some explanations on bubbles behavior in radial direction at this flow regime are given. To further support the provided explanations, the slug flow structure at both non- and fully-developed regions is experimentally visualized for air-water two-phase flow in the vertical riser of an airlift pump.

2. Interfacial structures

2.1. Bubbly-to-slug transition

Bubbly flow regime is characterized by small bubbles being suspended/dispersed in a liquid continuum as discrete substances [19]. Depending on bubbles size, small bubbles disperse in a random manner and rise with different velocities [20]. With respect to the magnitude of bubble-bubble and bubble-liquid interactions, bubbly flow itself can be categorized into four different regimes [21]. These regimes are named: (a) ideally separated bubbly flow, (b) interacting bubbly flow, (c) churn-turbulent bubbly flow, and (d) clustered bubbly flow [21]. In the ideally separated bubbly flow, there are no interactions between the bubbles neither directly nor indirectly. Therefore, bubbles behaviour will be similar to single bubble behaviour in a liquid column [21]. As the bubbles number density goes up, flow regime changes to interacting bubbly flow. In this regime, bubbles start to interact with one another in both direct (due to collisions) and indirect (due to wake effects) manners [21]. A further increase in the bubbles number density leads to the emergence of churn-turbulent bubbly flow, where bubbles coalescence forms so-called "cap bubbles". In this regime, the liquid continuum contains both cap and small size bubbles. Also, the two-phase bubbly flow is extremely agitated as a result of the bubbles motion and generated local turbulence [21]. Occasionally, the large bubbles form bubble clusters behaving like single gas slugs. Having travelled a certain distance in the axial direction, these clusters sometimes coalesce that leads to gas slug formation (transition to slug flow) and sometimes separate into individual large bubbles (transition to churn flow) [21]. Hence, the clustered bubbly flow regime is indeed a transition from bubbly flow to slug or churn flow. The pipe diameter is found to be the controlling mechanism that determines to which regime the flow regime transits [22]–[24]. In addition, bubbles initial size, which is determined by gas injector design, is another factor that can affect bubbly-to-slug flow transition. At a constant gas superficial velocity, the larger the initial bubble size, the lower the liquid superficial velocity at which transition takes place [23]. Radovcich and Moissis [25] and Taitel et al. [26] attributed bubbly-to-slug flow transition to the coalescence of small bubbles. According to Jones and Zuber [27], "slug flow is simply a transitional, time periodic combination of bubbly and annular flow". Slug flow regime has been reported to have average void fraction of 0.5 (with low-peak and high-peak points around 0.2 and 0.85, respectively) [13]. The void fractions beyond which bubbly-to-slug and slug-to-churn transitions take place have been found to be 0.25 [26] and 0.52 [28], respectively.

2.2. Slug regime

2.2.1. Phase Distribution in Axial direction

Fig. 1 (adopted from Ref. [14]) and Fig. 2 schematically represent and experimentally visualize two-phase flow structure in fully-developed and quasi-developed slug flow regime, respectively. As can be seen in both figures, the flow structure is constituted of (a) large bullet-shaped gas bubbles so-called "Taylor bubbles", occupying nearly the whole cross section of the pipe, (b) a thin liquid film surrounding Taylor bubbles, (c) liquid slugs between Taylor bubbles, and (d) small gas bubbles entrained in the liquid film and slugs. According to Fig. 1, the liquid slug itself is divided into three zones; (a) swelling front zone, (b) wake zone, and (c) low-void fraction zone. Taylor bubbles have a faster motion than liquid slugs. This, in turn, leads to the transfer of a portion of liquid from the preceding liquid slug to the subsequent one in form of a falling liquid film surrounding the Taylor bubbles, as shown in Fig. 1. This liquid film itself can contain a few small bubbles. If the pipe diameter is small enough (less than 5 mm), this liquid region can be devoid of bubbles [29]. A dispersion of small bubbles is often observed in the liquid slug between Taylor bubbles [19]. These bubbles (whose population is much larger than those in the liquid film [7]) are torn off the tail of Taylor bubbles (due to the imbalance of inertia and surface tension forces [30]). Being unable to keep pace with parents, these small bubbles are swept downstream (to the liquid swelling zone) [31]. Sekoguchi [32] identified this zone which is very close to the tail of Taylor bubble; the length of this zone can be 0.15-5 pipe diameter. The small bubbles entrainment in the liquid slug has been reported to be heavily dependent upon the surrounding liquid film flow as the entrainment takes place once the film plugs into the subsequent liquid slug [33]. The small bubbles underneath the tail of Taylor bubble often re-coalesce with the tail. The reason behind this phenomenon is the wake recirculation in the wake zone which can be 1-10 pipe diameter in length [14]. The small bubbles may also get absorbed by some later bubbles or even coalesce with themselves and reach a bigger size enough to raise them with a velocity equal to that of Taylor ones [31]. Experimental observations show that the Taylor bubbles wake becomes richer in small bubbles at higher flow rates. This is because the wake is more agitated at the higher flow rates, leading to the separation of a larger number of small bubbles from the tail of the Taylor bubble [31]. Detailed evaluation of the wake zone can be found in Pinto and Campos [34], and Campos and Carvalho [35]. The wake zone is followed by another zone, named low void fraction zone, extended down to the tip of the subsequent Taylor bubble. This zone can be 1-19 pipe diameter in length [14].

The experimental observations in previous studies show that an increase in gas superficial velocity up to certain values results in longer Taylor bubbles and liquid slugs ([36]–[39]). However, it has been observed that the effect of gas superficial velocity on Taylor bubbles length is far stronger than liquid slug length [13]. According to Jaeger et al. [40], slug frequency, the number of slug units passing through a certain location per time unit, is not significantly affected by gas superficial velocity. But, liquid superficial velocity has an strong impact on this parameter. They found that the higher the liquid superficial velocity, the higher the slug frequency due to the faster flow movement. The same had been observed by Gregory and Scott [41] and Vince and Lahey [42]. One interesting point still remained not fully-answered in literature is the direct bubbly-to-churn transition in large pipe diameters ($D_p \geq 150$ mm). That is, traditional slug regime (and subsequently, Taylor bubbles) has not reported to be observed in large diameter pipes ([11], [12]).

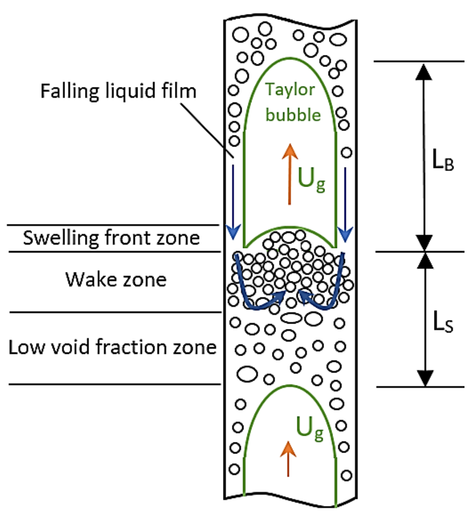
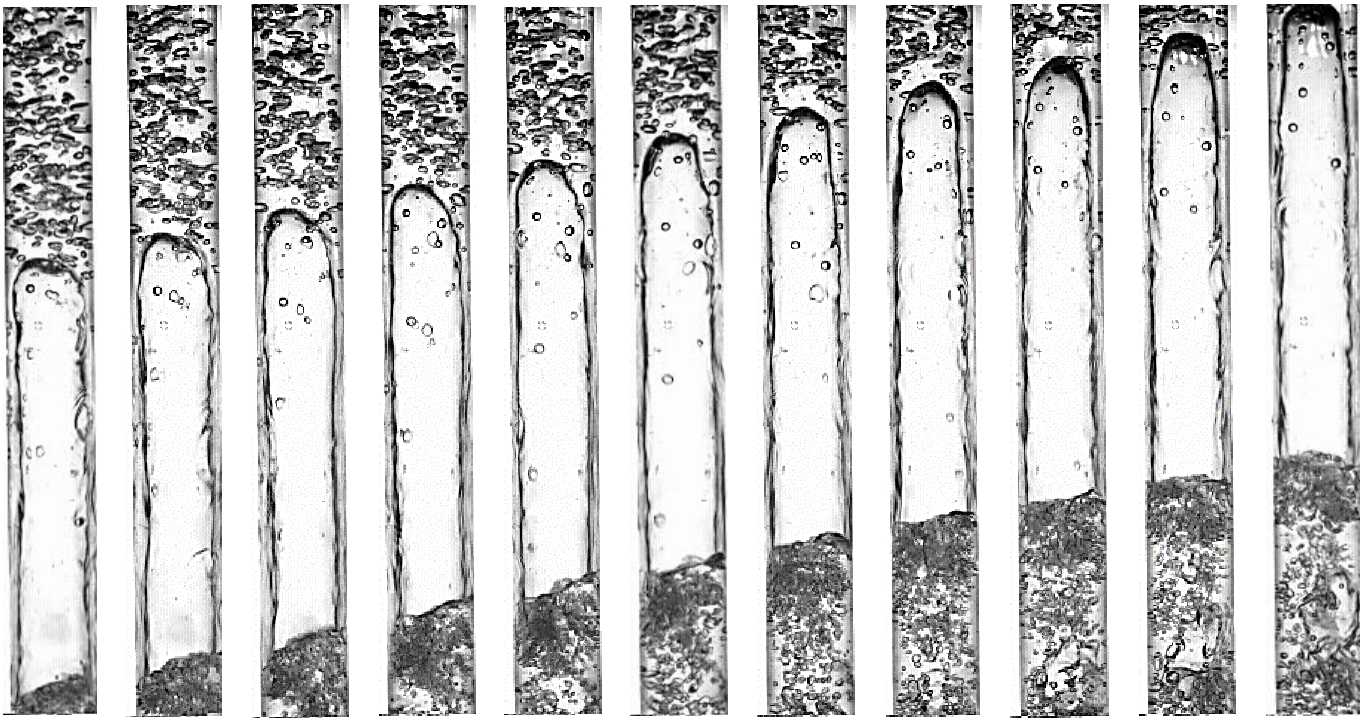


Fig. 1: Schematic slug flow regime structure (adopted from Ref. [14])



t = 0 s t = 0.02 s t = 0.04 s t = 0.06 s t = 0.08 s t = 0.1 s t = 0.12 s t = 0.14 s t = 0.16 s t = 0.18 s t = 0.20 s

Fig. 2: Taylor bubble motion in the quasi-developed region of slug flow at Z/D = 32, air flow rate of 8 Lit/min, pipe diameter of D = 1. ¼ in (The photos were taken by the authors in Multiphase Flow and Energy Lab at the University of Guelph)

Slug flow characteristics varies in axial direction until it becomes fully developed. The axial distance from the entrance needed to reach fully-developed slug flow has remained unclear so far [14]. The findings of Mi et al. [43] revealed that both Taylor bubbles and liquid slugs' lengths change along the axial direction in the developing region. This is more announced for Taylor bubbles at low liquid flow rates. Moissis and Griffith [20] characterized slug flow development in co-current upward flow by the interactions between two successive Taylor bubbles. They found that when the liquid slug length is large enough so that the Taylor bubbles are of smoothly rounded heads and rise with uniform and approximately the same velocities, the slug flow is fully-developed. That is, in the fully-developed region, there are neither direct nor indirect interactions between the two successive Taylor bubbles so that the motion and shape of each bubble is not influenced by the other one. Therefore, the two gas slugs move up with approximately the same velocities. This is because in the fully-developed region, the wake of leading bubble is smoothed out by turbulent mixing and eddies. Thus, the effect of the wake on the trailing bubble becomes negligible [31]. In this region, also, the liquid slug length is fairly regular. However, at high flow rates the lengths of both liquid slugs and Taylor bubbles are steadily less regular due the significant impacts of shear stress and turbulence [31]. On the other hand, if the liquid slug distance is smaller than a critical value, the wake of the leading Taylor bubble influences the trailing one. That is, it rises faster with a distorted nose becoming alternately eccentric on one side or another (shown in Fig. 3) and eventually agglomerates with the leading bubble [20]. To put it differently, there is a continuous sucking up of the trailing Taylor bubble in the wake flow behind leading Taylor bubbles. This process intensifies as the bubbles get closer [31]. The minimum value of the liquid slug length under/beyond which the slug flow is developing/developed was found by Pinto and Campos [44] to be four times the wake length. This distance has been also found to be insensitive to liquid and gas flow rates [20]. The rise of a Taylor bubble in the wake of the leading one contributes to a non-developed slug flow with an oscillatory nature [20]. The same observations were reported by Pinto and Campos [44]. The development of slug flow regime along the axial direction can be traced from time series void fraction data (and PDF diagrams) collected at points with different distances from the entrance [13]. Close to the entrance, the void fraction reveals a chaotic pattern and the Probability Density Functions (PDFs) fail to exhibit slug flow signature. This is mainly attributed to the impact of the gas injector [13]. However, as the flow moves up, the chaotic behaviour of void fraction is damped, the flow is recovered, and gas injector effects disappear [13]. Finally, from a certain vertical location onwards, both void fraction and PDF diagrams show similar trends at different downstream locations. That is, the flow reaches fully-developed area at this location [13]. These all represent the developing slug flow characteristics.

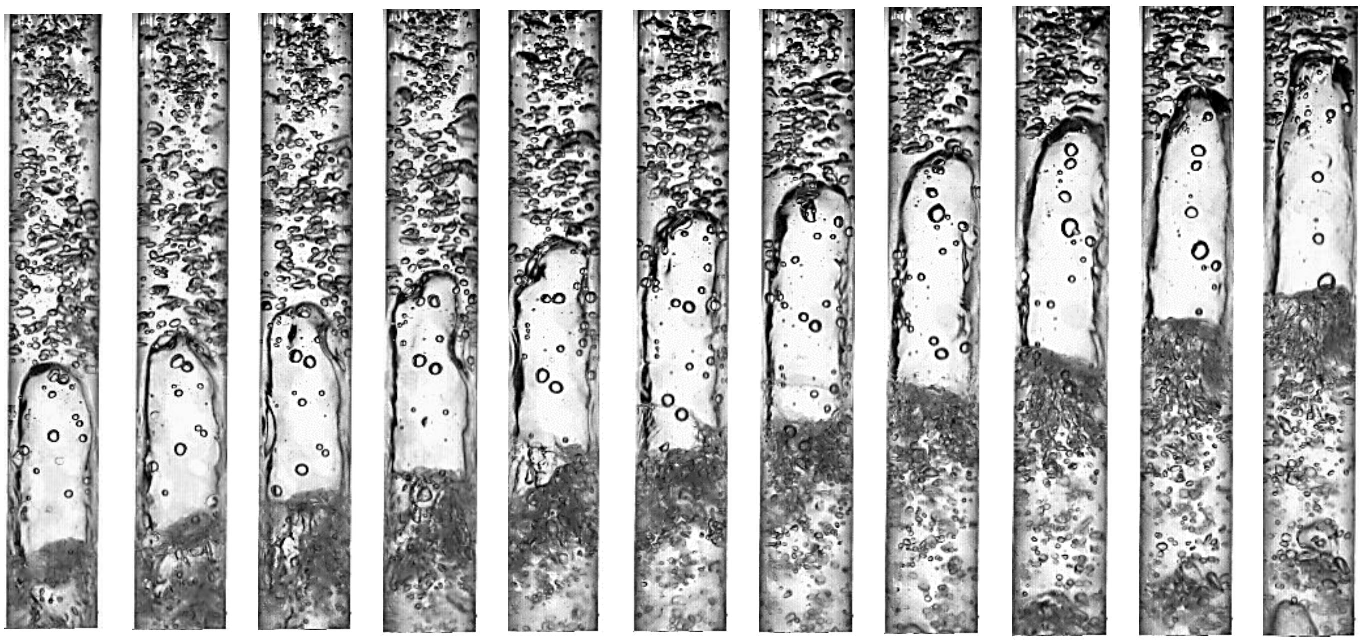


Fig. 3: Taylor bubble motion in the non-developed region of slug flow at Z/D=16, air flow rate of 8 Lit/min, pipe diameter of D=1. ¼ in (The photos were taken in Multi-phase Flow and Energy Lab at the University of Guelph)

2.2.1. Radial Phase Distribution

Local time average void fraction distribution in the radial direction for slug flow regime can be generally described by two shapes. At low gas flow rates, the distribution has a parabolic shape (with a maximum in the middle) and low values close to the walls. On the other hand, at high gas flow rates, the tip of the parabolic shape becomes somewhat flat with slightly fluctuating behaviour in the central region [7]. Moreover, the local average gas velocity distribution in radial direction is of a parabolic shape in slug flow regime. That is, gas bubbles velocities are higher in the middle of the pipe (with a peak in the centre of the pipe) due to the large concentration of gas phase close to the pipe centre [7]. The velocity of gas bubbles in both middle and wall sides increases with increasing liquid and gas superficial velocities [7]. Detailed information on the axial and radial development of the local void fraction, Interfacial Area Concentration(IAC), bubble velocity, and Sauter mean diameter of both small and large (Taylor) bubbles at bubbly-to-slug transition region and slug regime can be found in the study conducted by Wang et al. [8].

5. Conclusions

Interfacial structures behaviour of slug flow regime were briefly reviewed in the present study. It can be concluded that, Taylor and small bubbles along with liquid film and slugs were found to represent the main interfacial structures. The flow conditions, pipe diameter, and gas injector were discussed and found to strongly impact the interfacial structures at slug flow regime. Both the review and the flow visualization revealed that Taylor bubbles behaviour is considerably different in the axial direction (from non-developed to quasi- and fully-developed regions). Also, the main gaps in the literature were identified; to the best knowledge of the authors, interfacial structures have remained less understood in bubbly-to-slug transition and non-developed slug flow regions and require further investigations.

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