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Characteristics of slug flow in a vertical narrow rectangular channel

Yang Wang, Changqi Yan *, Licheng Sun, Chaoxing Yan

Fundamental Science on Nuclear Safety and Simulation Technology Laboratory, Harbin Engineering University, Harbin, Heilongjiang 150001, China

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ABSTRACT

In the view of modeling slug flow, a detailed understanding of its hydrodynamics is of great importance. Non-intrusive flow visualization using a high speed video camera system is applied to study characteristics of slug flow in a vertical narrow rectangular channel ($3.25 \times 43\ mm^2$). The characteristics of the Taylor bubble, the liquid film and the liquid slug are studied and compared with the models available in literature. It is shown that the slug flow in the present channel is somehow different from the classical slug flow in medium size channels. The gas and liquid flow rates have significant effects on the Taylor bubble length, the thickness and velocity of liquid film at the bottom of Taylor bubble. For the continuous slug flow, the drift velocity is larger than the terminate velocity of a single Taylor bubble in stagnant liquid; the velocity of Taylor bubble could be well predicted by the Nicklin et al. correlation. The minimum stable liquid slug length is in the range from 9 to 17 hydraulic diameters in fully developed turbulent flow. Correlations for calculating the length of Taylor bubble, the thickness and velocity of liquid film at the bottom of Taylor bubble are proposed by fitting the experimental data, respectively.

1. Introduction

Slug flow is often encountered in many practical applications such as distillation columns, gas absorption units, nuclear reactors, oil–gas pipelines, and steam boilers. The complicated slug flow structure can be described as a series of slug units, each of which consists of a Taylor bubble with a liquid film around it and a portion of liquid slug behind the Taylor bubble. In view of the dominating role of the hydrodynamics of Taylor bubble in the slug flow, several experimental and theoretical works have been reported on the rise of Taylor bubbles through stationary and moving liquid in circular tubes [1–12]. The first experimental study of coalescence mechanism between two consecutive Taylor bubbles was carry out by Moissis and Griffith [5]. They found that the trailing bubble accelerates with its distance to the leading bubble decreasing and its nose sways from side to side. Campos and Guedes de Carvalho [8] did a photographic study of the flow in the wake of individual Taylor bubbles in stagnant liquid, and identified three different flow patterns in the wake of Taylor bubbles (laminar, transition and turbulent). Pinto et al. [9] investigated the Taylor bubbles rising in cocurrent flow condition and also identified three similar flow patterns in the wake of Taylor bubbles.

However, the majority of the studies are confined to slug flow in circular tubes, only a few works deal with the slug flow in narrow rectangular channels [13–19]. In spite of that, gas–liquid two-phase flow through a narrow rectangular channel has been the subject of increased research interest in the past few decades. It is encountered in many important applications including the cooling systems of various types of equipment such as high performance micro-electronics, supercomputers, high-powered lasers, medical devices, high heat-flux compact heat exchangers in spacecraft and satellites as well as research nuclear reactors with plate type fuels [20]. It can be expected that the restriction of the bubble space in the narrow channel is the cause of the difference in the two-phase flow characteristics from those in conventional channels. This may also affect heat-mass transfer characteristics during the change of phase. Up to now, there has not been a clearly definition of the size of the narrow rectangular channel in the literature. Sadatomi et al. [14] presented the flow regime map in large vertical rectangular channels and indicated that channel geometry has little influence in the noncircular channels when the channel hydraulic diameter (D_h) is larger than 10 mm. Lowry and Kawaji [21] performed an experiment in narrow vertical flow channels and found significant differences between flow in small channels to those D_h larger than 10 mm. Using these major representative research milestones, the channel classifications of narrow rectangular channel and medium size channel based on the hydraulic diameter D_h are proposed, 10 mm may be considered as the lower limit for D_h of the medium size rectangular channels.

In previous experiments on slug flow in rectangular channels, only the interface velocity of Taylor bubble is determined and no systematical information about slug flow is obtained. Nicklin et al. [4] proposed a correlation (Eq. (1)) to predict the velocity of a single Taylor bubble (V_T) in a moving liquid for circular tube based on experiments. It is generally assumed that the single

* Corresponding author. Tel./fax: +86 0451 82569655.
E-mail address: changqi_yan@163.com (C. Yan).

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Nomenclature

a	aspect ratio of (s/w)
C_0	distribution parameter
D	diameter (m)
D_e	eaui-periphery diameter (m)
D_h	hydraulic diameter (m)
F_{scale}	scale factor
g	gravitational acceleration (m/s^2)
j_G	gas superficial velocity (m/s)
j_G^*	dimensionless gas superficial velocity
j_L	liquid superficial velocity (m/s)
j_L^*	dimensionless liquid superficial velocity
j_{TP}	two-phase superficial velocity (m/s)
L_A	length of the wake region (m)
L_B	length of the transition region (m)
L_C	length of the fully developed region (m)
L_{min}	length of the minimum stable liquid slug (m)
L_T	Taylor bubble length (m)
R	radius of the pipe (m)
R_c	terminal radius of Taylor bubble bottom (m)
R_0	radius of Taylor bubble nose (m)
Re	Reynolds numbers of single-phase liquid
Re_{TP}	Reynolds numbers based on two-phase superficial velocity
s	height of rectangular channel (m)
V_f	liquid film velocity in the fixed frame of reference (m/s)
V_{fd}	velocity of liquid film at the bottom of Taylor bubble in the fixed frame of reference (m/s)
V_{fdr}	velocity of liquid film at the bottom of Taylor bubble in the moving frame of references (m/s)
V_{fr}	liquid film velocity in the moving frame of references (m/s)
V_0	drift velocity (m/s)
V_i	velocity (m/s)
V_{max}	maximum local velocity (m/s)
V_m	mean liquid velocity (m/s)
V_T	Taylor bubble velocity (m/s)
P	wetted perimeter (m)
w	width of rectangular channel (m)
x_1	distance from Taylor bubble nose (m)
x_2	distance from Taylor bubble bottom (m)
y	distance relative to the channel center line (m)
z	axial distance from the inlet (m)
Δx_i, Δy_i	displacements of tiny bubbles in images (pixel)
ΔN	frame interval between measured Δx_i and Δy_i (frame)

Greek letters

α	average void fraction
δ_f	liquid film thickness (mm)
δ_{fd}	thickness of liquid film at the bottom of Taylor bubble (mm)
δ_t	terminal thickness of the liquid film (mm)
η	dimensionless falling liquid film thickness
η'	modified dimensionless falling liquid film thickness
ξ	dimensionless axial distance from the Taylor bubble nose
π	circumference ratio
ρ_L	liquid density (kg/m^3)
ρ_G	gas density (kg/m^3)
$\Delta\rho$	density difference between liquid and gas (kg/m^3)
τ	time interval between two frames

Subscripts

G	gas phase
L	liquid phase
TP	two-phase

Taylor bubble velocity in a flowing liquid is a superposition of the bubble velocity in a stagnant liquid, the drift velocity (V_0), and a contribution due to the mean liquid velocity (V_m).

$$V_T = C_0 V_m + V_0 \tag{1}$$

The value of C_0 is based upon the assumption that the velocity of the Taylor bubble follows the maximum local velocity (V_{max}) in the front of the nose tip, and thus, $C_0 = V_{max}/V_m$ [4,6,7]. The value of C_0 therefore equals approximately 1.2 for fully developed turbulent flow and 2.0 for fully developed laminar flow.

This relationship has later been applied for predicting the Taylor bubble velocity in continuous slug flow in circular tube by most researchers, whereas substituting the mean liquid velocity (V_m) by the mixture velocity (j_{TP}), the sum of the liquid and gas superficial velocities j_L and j_G. Then, the relationship becomes

$$V_T = C_0 j_{TP} + V_0 \tag{2}$$

However, contrary to the proposal of Nicklin et al. [4], most investigators have taken V_0 as the intercept on the ordinate of a linear relation between V_T and j_{TP}. For the experiments of slug flow in a vertical narrow rectangular channel, researchers also have attempted to correlate the Taylor bubble velocity with the same relationship [14,16,17].

Jones and Zuber [22], Sadatomi et al. [14] and Ide et al. [16] recommended C_0 of 1.2 for slug flow in a rectangular conduit. Mishima et al. [19] reported that the value of C_0 is in the range of 1–1.2 for rectangular channels with gap sizes of 1.07 mm, 2.45 mm and 5 mm. Sowinski et al. [17] proposed that C_0 is about from 1.23 to 1.27 in narrow mini-channels. Bhusan et al. [18] reported the value of C_0 lies between 1 and 1.3 for rectangular channels 2.7 mm × 5.1 mm and 2.7 mm × 10 mm, respectively. Ishii [23] proposed the following empirical formula for C_0 in terms of density ratio for rectangular channels.

$$C_0 = 1.35 - 0.35(\rho_G/\rho_L)^{0.5} \tag{3}$$

where ρ_G and ρ_L are the densities of the gas and liquid phases, respectively.

For inertia-dominated systems, the rising velocity of Taylor bubble through a circular tube filled by stagnant liquid, the drift velocity V_0 can be expressed by Eq. (4) proposed by Dumitrescu [1], where the tube diameter D is taken as the characteristic length.

$$V_0 = 0.35\sqrt{\Delta\rho g D/\rho_L} \tag{4}$$

where $\Delta\rho$ is the density difference between the two phases, g is the gravitational acceleration.

As for the case of Taylor bubble rises through a non-circular channel, there is not a general definition on the characteristic length being extensively accepted. From time to time, several suggestions of the characteristic length were made, and several correlations for the drift velocity were also put forward.

Taking the channel width (w) and gap width (s) for being the characteristic length, Griffith [24] proposed following relationship for calculating V_0,

$$V_0 = (0.23 + 0.13s/w)\sqrt{\Delta\rho g w/\rho_L} \tag{5}$$