

Study of viscous oil-water-gas slug flow in horizontal pipes

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1. Introduction

Understanding liquid-liquid-gas flow is crucially important for production and transportation systems. There have been numerous investigations regarding two-phase flows in the past few years. However, a few research works have been performed, even using low viscosity oil, in three phase flows. As an example of three phase flows, the work by Acikgoz et al. (1992) can be cited, which deals with the identification of low viscosity oil/water and gas flow patterns within a horizontal pipe with an ID=19 mm. They used low oil viscosity ($\mu_o=0.116$ Pa·s at

25 °C) and density of $864 \text{ kg}\cdot\text{m}^{-3}$. The investigated operating conditions were as follows: superficial water velocity ($J_w=0.66 \text{ m}\cdot\text{s}^{-1}$), superficial oil velocity ($J_o=0.24 \text{ m}\cdot\text{s}^{-1}$), and superficial gas velocity (extended to $50 \text{ m}\cdot\text{s}^{-1}$). Many different flow patterns were recognized depending upon the fact that water and oil maintain their structures as continuous flow, defined as water-based and oil-based flow, respectively. Several attempts were made to utilize conductance sensors for measuring phase holdup and automatizing flow pattern recognition in two-phase flow, as suggested by Gao et al. (2016, 2017, 2018), but universal tools and methods are not yet available, particularly for three-phase flows.

Series of research activities have been conducted at the test facilities (WASP) at Imperial College, London. The work by Pan et al. (1995) deals with oil/water/gas flow within an ID=76.2 mm and ID=38.0 mm long horizontal pipes. Experimental tests were performed at 0.5 MPa test facilities and operating conditions was similar to Acikgoz et al. (1992). Hewitt (2005) conducted liquid-liquid-gas flow tests with $\mu_o= 0.04 \text{ Pa}\cdot\text{s}$ and high flash point oil. The pressure drop and liquid phase holdups using dual-energy gamma densitometer were measured. Two different behaviors were observed from the analysis of pressure drop measurements depending on gas flow rates. A peak at pressure drop was observed confirming the presence of inversion mechanism for higher gas flow rate. Accordingly, water continuous flow converts to oil-based continuous flow and effective viscosity becomes so high that it considerably increases pressure drop, see also the works by Hall (1992) and Odozi (2000). On the other hand, at low gas velocity, the peak of pressure drop was not observed. Keskin et al. (2007) suggested a two-step categorizing technique.

Twelve individual flow patterns were identified in a horizontal pipe, considering very low oil viscosity. Malinowsky (1975) carried out slug flow tests with $\mu_o=4-5$ mPa.s, ID=38.1 mm as well as operating pressure of 2 bar. Superficial gas, water, oil velocities were ranged, respectively, within the intervals as follows: $1.5 < J_g < 4.3$ m.s⁻¹, $0.19 < J_w < 2.08$ m.s⁻¹, and $0.26 < J_o < 1.36$ m.s⁻¹. The major observed flow pattern under investigation was slug flow, with a few tests corresponding to high gas flow rates, in which misty annular and slug was observed. He compared pressure drop of his experimental data with classical correlation of Beggs and Brill (1973), which is broadly applicable in oil industry, assuming that oil-water mixture had a linear interpolated viscosity. Finally, he concluded that linear approximation for mixture viscosity of oil-water is a poor representation due to the fact pressure drop predictions are up to 50% underpredicted. Stapelberg and Mewes (1994) measured distributed pressure gradient and slug frequency using a laser, concerning three-phase flow with $\mu_o=0.031$ Pa.s and $\rho_o=886$ kg.m⁻³. The tests were investigated in two different pipe diameters as 23.8 and 59 mm. Flow patterns were also detected and flow maps were developed.

Issa, et al (2007) studied oil-water-gas flow within a straight duct with ID=77.92 mm. Low oil viscosity ($\mu_o=45.8$ mPa.s) was considered for the tests. They examined different aspects of slug structures in terms of spatial distribution of each phase. Time-averaged liquid holdups, slug frequency were measured by using two X-ray and gamma systems, which were capable of differentiating between water and oil. It was observed that slug frequency is reduced as a result of increasing water cut at fixed superficial gas and liquid velocities. Furthermore, a peak of slug

frequency at lower water cut was detected, suggesting the inversion mechanism from oil to water continuous flow.

Multiphase flows of high viscous oil/water/gas behave differently from flows of low viscous oil/water/gas. The main reason arises from the rheological property of oil phase. In spite of the importance of high viscous oil/water/gas flows in oil industry, few studies have been conducted in horizontal pipes. The work by Bannwart (2009) can be cited, considering very viscous oil-air-water flows ($\mu_o=3.4$ Pa·s, $\rho_o=970$ kg.m⁻³ at T=20 °C) within a 28.4 mm i.d. pipe (Laboratory scale) and a 77 mm i.d. pipe (full-scale facility). The tests were performed both in a horizontal and upward vertical pipe, the influence of inclination angle on three-phase flow pattern was investigated. Nine flow patterns were identified in horizontal pipe which are Bubble gas-Bubble oil (B_g, B_o), Bubble gas-Annular oil (B_g, A_o), Bubble gas-Intermittent oil (B_g, I_o), Bubble gas-Stratified oil (B_g, S_o), Intermittent gas-Bubble oil (I_g, B_o), Intermittent gas-Annular oil (I_g, A_o), Intermittent gas-Intermittent oil (I_g, I_o), Stratified gas-Bubble oil (S_g, B_o), Stratified gas-Stratified oil (S_g, S_o). Superficial gas, water, and oil velocities varied as follows: $J_g=0.03-10$ m·s⁻¹, $J_w=0.04-0.5$ m·s⁻¹, $J_o=0.01-2.5$ m·s⁻¹. They presented pressure drop data and concluded that the existence of gas phase has a positive influence on increasing frictional pressure loss. Pressure drop was perceived to be highly dependent upon superficial velocity of phases. The pressure drop increase caused by gas injection can be damped by increase of water, promoting the lubrication mechanism and help preventing oil from adhering to the internal surface of pipe wall. Wang et al. (2013) carried out oil-water-gas experiments with much lower oil viscosity ($\mu_o=0.15-0.57$

Pa·s at 37.8-15.6 °C). The internal pipe diameter is 52.5 mm, superficial water and oil velocities varied from 0.1-1 m·s⁻¹, and gas superficial velocity ranged 1-5 m·s⁻¹. Flow patterns were monitored and images captured by means of high-speed video camera. The experimental pressure gradient was compared to the mechanistic model developed by Zhang and Sarica (2006), showing unsatisfactory agreement. Table 1 lists a summary of related studies for oil-water-gas flow in horizontal ducts. The rheological properties of gas and water are not reported because city tap water and air are used as test fluids in all cases.

Since very few experimental data was presented in the literature, the present study aims at expanding the data related to pressure drop measurement. Moreover, study of literature survey reveals that hydrodynamic behavior of high viscous oil-water-gas flow and slug characteristics in horizontal pipe is not well understood. In the current work, slug frequency was measured by signal analyses and the results were compared to the available correlations in the literature for gas-liquid flow. Geometrical dimensions of slug and bubble, which included slug body length (L_s) and bubble length (L_f) were measured by optical sensor. Statistical analysis of slug body length (PDFs) was presented. Based on measurements of slug and bubble length, a new expression for calculating slug unit length in three phase flow high viscous oil in horizontal pipe is formulated. Translational bubble velocity is measured by optical probe and video camera, results of bubble velocity ~~was~~ were compared to the correlation of Nicklin (1962) which is a drift-flux based formulation. Section 3 reports the results of experimental analyses.

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120 **Table 1.** Summary of experimental investigations on oil-water-gas flow in horizontal ducts

Author	Pipe I.D. (mm)	μ_o (mPa·s)	ρ_o (kg·m ⁻³)	Velocity range (m·s ⁻¹)
Malinowsky (1975)	38.1	4-5	855	J _o :0.26-1.36 J _w :0.19-2.08 J _g :1.5-4.3
Acikgoz et al. (1992)	19	116	864	J _o :0.26 J _w :0.66 J _g : up to 50.0
Hall (1992)	78	40-60	876.2	J _o :0-0.54 J _w :0-0.83 J _g :0.98-4.1
Stapelberg and Mewes (1994)	23.8; 59	31	858	J _o :0.062-0.244 J _w :0.16-0.18 J _g : not reported
Odozi (2000)	78	9.3-153	855-871	J _o :0.1, 0.5, 0.7 J _w :0.05-0.5 J _g :2.0-24.0
Hewitt (2005)	38; 78	40	860	J _o : 0.06-0.39 J _w : 0.21-0.54 J _g =1.6-11.7
Keskin et al. (2007)	51	13.5	858.8	J _o :0.02-1.5 J _w :0.01-1.0 J _g =0.1-7.0
Issa, et al (2007)	78	45.8	863	J _o : 0.13-0.37 J _w : 0.13-0.35 J _g =3.9-4.4
Bannwart (2009)	28.4; 77	3400	970	J _o : 0.01-2.5 J _w : 0.04-0.5

Wang et al. (2013)	52.5	150-570	884.4	$J_g=0.03-10$ $J_o:0.1-1$ $J_w:0.1-1$ $J_g=1-5$
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2. Experiments

2.1 Experimental facility and operating procedure

The liquid-liquid-gas test rig is located in the Laboratory of Multiphase Thermal-Fluid Dynamics at Politecnico di Milano. The facility is sketched in Fig. 1 The rheological properties of fluids are: tap water ($\mu_w=1.026\times10^{-3}$ Pa·s, $\rho_w=999$ kg·m³), Oil (type Milpar 220, $\mu_o=0.838$ Pa·s, $\rho_o=890$ kg·m⁻³, and $\sigma_o=0.02$ N·m⁻¹ at T=20 °C), air ($\mu_a=1.98\times10^{-5}$ Pa·s, $\rho_a=1.2$ kg·m⁻³). The test section is composed of a 12 m long straight transparent duct with ID=40 mm. The pressure drop measurement was performed, considering five pressure taps. Each pressure was connected to a pressure transducer (differential type with full scale: 1 psi = 6.89 kPa), positioned 6 m from pipe inlet. Two K-type thermocouples (10 % Cr-6 % Al) are used to measure both the ambient and mixture temperature, one is located outside the pipe and the second one is positioned at almost half distance of the total pipe length. Variation of volumetric fluxes for oil, water, and air are as follows: $J_o=0.36-0.71$ m·s⁻¹, $J_w=0.44-1.32$ m·s⁻¹, and $J_g=0.22-2.10$ m·s⁻¹. Water and oil flow rates were measured by using a magnetic flow meter (accuracy $\pm 0.5\%$ of reading) and a calibrated gear metering pump, respectively. Oil and water are injected in the pipe through a coaxial injector in order to aid the onset of formation of core-annular flow

in the appropriate range of superficial velocities. Water is drawn first to the test section, oil is then supplied with the selected flow rate. Water flow rate is checked and its value is adjusted to the set point, if needed. Once the two-phase flow is properly formed, the distributed pressure drop is measured. At the final stage, air is then introduced at the desired flow rate, and once the three-phase flow is suitably stabilized, the distributed pressure drop is measured. A total number of 235 pressure drop data points were acquired. The facility is equipped with an optical probe developed and designed by Arnone (2017) (see section 2.3) and a video camera (see section 2.4 for image processing), positioned 7 and 7.5 m from the inlet, to investigate both quantitative and qualitative aspects of three-phase flow.

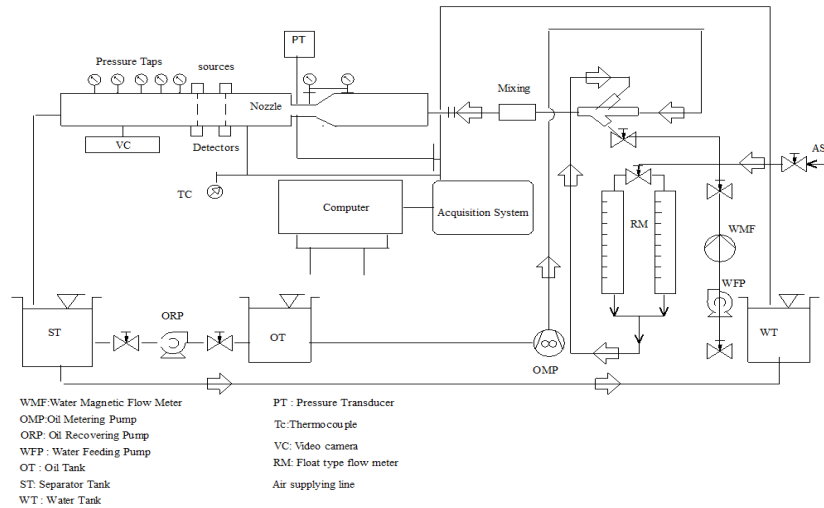


Fig 1. Experimental setup

2.2 Governing parameters

The main controlling parameters are reported in this section. Volumetric flux or superficial velocity J ($\text{m}\cdot\text{s}^{-1}$) for each phase is determined as the ratio between

individual volumetric flow rate and area of pipe cross-section, as the phase flowed alone. The total mixture volumetric flux is computed by summation of individual superficial velocity of each phase as:

$$J_t = J_o + J_w + J_g \quad (1)$$

The same definition can be applied to the total liquid superficial velocity as:

$$J_L = J_o + J_w \quad (2)$$

The ratio between gas/oil/water and total liquid volumetric fluxes is defined as:

$$\varepsilon_{Lg} = \frac{J_g}{J_L} \quad \varepsilon_{Lo} = \frac{J_o}{J_L} \quad \varepsilon_{Lw} = \frac{J_w}{J_L} \quad (3)$$

The volumetric fraction of a phase is defined as the proportion of superficial and actual velocity of single phase as follows:

$$H_o = \frac{J_o}{U_o} \quad H_w = \frac{J_w}{U_w} \quad H_g = \frac{J_g}{U_g} \quad (4)$$

The liquid holdup is then $H_L = H_o + H_w$, thus void fraction is $H_g = 1 - H_L$.

The pressure reduction factor as defined by Bannwart et al. (2004) and Poesio et al. (2009) is the ratio between liquid-liquid and three phase pressure drop, considering the same operating conditions for water and oil flow rates, that is;

$$R_{LG} = \frac{\Delta p_{oil-water}}{\Delta p_{oil-water-gas}} \quad (5)$$

Since flow pattern under investigation is slug flow, and three-phase flow mixtures are identified by elongated air bubbles, optical techniques, as introduced by Poesio

et al. (2009b), have been adopted to determine translational bubble velocity, U_t . They are illustrated in the following sections.

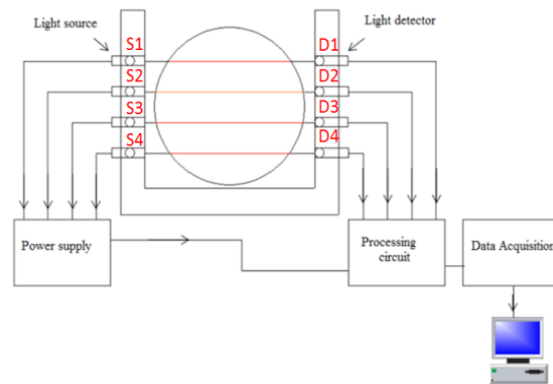
2.3 Optical probe measurements data processing

In the current study, a non-intrusive optical probe has been designed to measure the translational velocity of bubbles. During three-phase flows of oil-water-gas, different configurations of phases can be formed mainly depending on operating conditions. The phases in contact may have either smooth or wavy interfaces. The designed optical probe has to be able to detect differences between optical properties of phases in medium. Moreover, it should be capable of distinguishing between elongated bubble zone and slug liquid zone. A detailed diagram of the optical probe is depicted in Fig. 2. It comprises four points LED (Light-Emitting Diodes, which are positioned such a way that they cover the whole height of the tube with a spacing of 4 mm), four light-dependent resistance (LDR) detectors and a processing circuit.

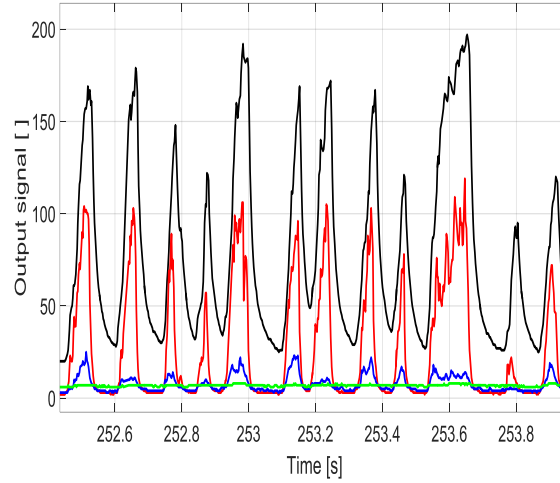
LDR sensors are placed at horizontally opposite points to capture the light generated by the LED source after its passage through the test section. The narrow light beam passes through the three-phase medium before reaching the LDR, being particularly absorbed, refracted, reflected, scattered, etc. Hence, intensity of the light captured by the detectors depends on the spatial distributions of phases. Another optical probe is positioned at an adjustable distance L

(30 cm in the present study) in order to generate time-shifted signals, which are used to determine the bubble translational velocity, as described in the following.

Electric output of the probe is obtained at the sampling frequency of 1000 Hz and saved in a PC for further processing. Because the frequency of slug flow ranged from 2 to 20 Hz, 1000 Hz is a suitable frequency for detecting it. An example of raw output signal is shown in Fig. 3 where the black and green signals denote the uppermost and lowest LEDs, respectively, mounted in a radial direction.



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Fig. 2. Schematic representation of optical probe device

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Fig. 3. Example of raw output signal for $J_o=0.71 \text{ m}\cdot\text{s}^{-1}$, $J_w=1.32 \text{ m}\cdot\text{s}^{-1}$, and $J_g=2.10 \text{ m}\cdot\text{s}^{-1}$

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The bubble translational velocity can be detected by knowing the distance (L) between two

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probes and associated time lag (τ):

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$$U_t = \frac{L}{\tau} \quad (6)$$

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In all the operating conditions, the intensity of the light captured by the two sensors below the

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pipe axis (D3 and D4 in Fig. 2) is very weak, since most of the light path covers the liquid

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phase other than gas. On the contrary, the other two sensors (D1 and D2 in Fig. 2) present a

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strong response with sharp transitions, marking the passage of air bubbles. Hence, output of

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D1 and D2 sensors has been averaged and used for data processing, in order to determine

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bubble translational velocity by means of cross-correlation technique. It is worth noting that

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distance (L) between two sensible areas (two sensors) needs to be properly selected because it

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depends on flow conditions. If bubble are too fast and short, longer distance is required. On

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the other hand, if distance is too long, it is difficult to identify the bubble because during the

passage of elongated bubbles it may happen that elongated bubbles collide each other and two sensors see different elements of bubbles. In this case, it is so difficult to obtain a time lag between two similar signals.

2.3.1 Single bubble identification method

Detection of bubble translational velocity by means of cross-correlation technique is very fast but it gives only averaged value. Furthermore, structural configurations of slug flow, such as bubble length cannot be detected. The second approach is based on the so-called “single bubble identification method”. Fig. 4 shows the schematic representation of slug flow.

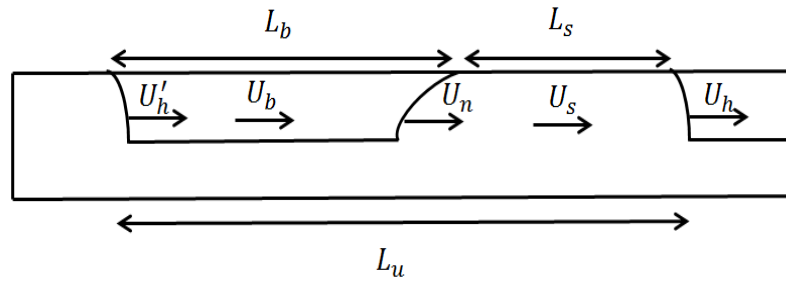


Fig. 4. Schematic representation of slug flow

As it is evident from Fig. 4, the slug flow pattern is characterized by liquid regions alternating with elongated air bubbles. Characteristic parameters of slug unit are defined as:

U_h : velocity of head of leading liquid slug

U_n : velocity of tail of leading liquid slug or head of bubble

U'_h : velocity of head of trailing liquid slug

236 L_s : liquid slug length

237 L_b : bubble length

238 L_u : total slug unit length

239 The single bubble identification method is an advantageous way to measure the head, tail
240 velocities of liquid slug and bubble, and geometrical parameter of slug unit. It is worth noting
241 that velocity of tail of liquid slug, U_n is equivalent to velocity of head of bubble. First, the raw
242 signal is normalized with power 2, in order to amplify the difference between bubble and liquid
243 passage as:

244
$$S = \left(\frac{S - \min(S)}{\max(S) - \min(S)} \right)^2 \quad (7)$$

245 An example of signal normalization is illustrated in Fig. 5. After normalization of signal, it is
246 necessary to convert the normalized signal to a binarized rectangular wave signal, by means of
247 a threshold value, suitably selected according to the operating conditions to extract some
248 information about geometrical parameters of slug flow. In the binarized signal, the value 1 and
249 0 relate to the liquid and gas phase, respectively. Fig. 6 shows typical example of binarization
250 process with a threshold value set as 0.3, where transition from 0 to 1 indicates the front of a
251 slug while transition from 1 to 0 indicates the tail of slug. The effect of threshold value on
252 capturing slug unit must be always checked. The threshold value of 0.3 seems to be a good
253 approximation because it captures all slug units.

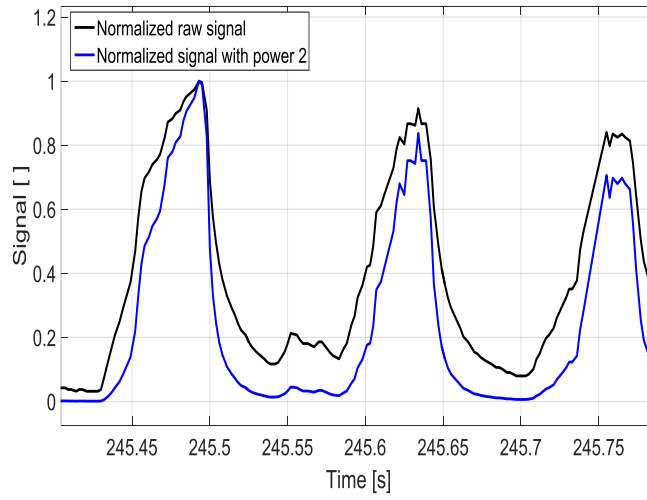


Fig. 5. Example of signal normalization

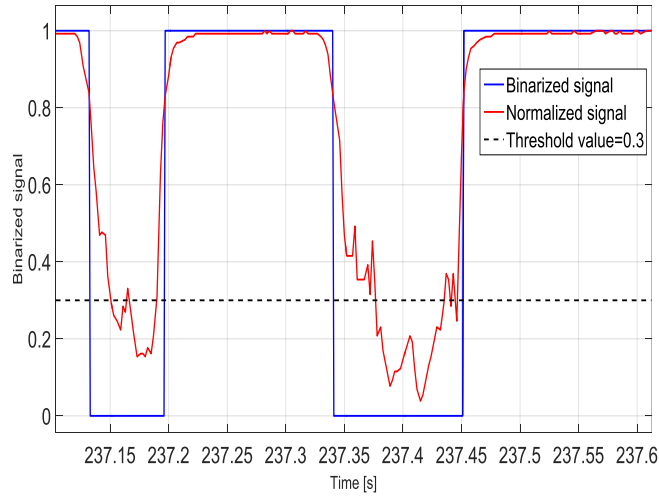


Fig. 6. Typical example of binarization process using threshold technique

After detection of rectangular signal, the algorithm is able to extract time residence of slug and bubble. To measure the bubble (U_{eb}) and liquid slug velocity (U_s), first, the characteristic times of slug unit must be suitably identified. Table 2 shows time residence of slug unit characteristic.

262 **Table 2.** Characteristic time residence of slug unit between two sensible areas

$t_{h2}-t_{h1}$	Time residence of liquid slug head between two sensible area
$t_{n2}-t_{n1}$	Time residence of liquid slug tail between two sensible area
t_s	Time passage of liquid slug from first to the second sensor
t_b	Time passage of elongated bubble from first to the second sensor

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264 Since the time residence of head and tail of slug can be detected from binarized signal, the
 265 velocity of head and tail of slug is measured by knowing the distance between two sensors as:

266
$$U_h = \frac{L}{t_{h2}-t_{h1}} \quad (8)$$

267
$$U_n = \frac{L}{t_{n2}-t_{n1}} \quad (9)$$

268 Slug (U_s) and elongated bubble velocities (U_{eb}) are measured, simply, by averaging velocity
 269 of slug head and tail:

270
$$U_s = \frac{U_h+U_n}{2} \quad U_{eb} = \frac{U_n+U'_h}{2} \quad (10)$$

271 The bubble and slug lengths would be determined by multiplication of time residence of bubble
 272 and slug between two sensible areas and respective velocities.

273
$$L_b = t_b U_{eb} \quad L_s = t_s U_s \quad (11)$$

274 The total slug unit length is the summation of bubble length (L_b) and slug body
 275 length (L_s), that is, $L_u=L_b+L_s$.

276 **2.4 Image processing**

The objective of image-processing was to measure translational velocity of slugs. Other geometrical characteristics of slugs can be directly evaluated from image processing of video cameras. To extract quantitative information regarding hydrodynamic behavior of slug flow, a series of images (250×1579 px²) were taken by video camera (NIKON D3300) at frequency 50 fps (frames per second). The visualization section is exposed by two yellow lamps. One case of the typical flow pattern is illustrated in Fig. 7. It is seen that oil is mainly opaque, whereas air and water are transparent. Hence, to distinguish air-oil and water-oil interfaces, a threshold technique is required. The image post-processing is carried out using Image Toolbox of Matlab[®]. Bubble translational velocity is computed by means of cross-correlation, using two virtual probes (probe #1 and 2 in Fig.7), positioned at the beginning and end of pipe (with known distance Δx) to ensure that even very long elongated bubbles can be captured. Therefore, translational velocity is given by:

$$U_t = \frac{|\Delta x|}{N_{frames}} \cdot pixel\ size \cdot F_s \quad (12)$$

Where N_{frames} and F_s (50 fps) are the number of frames passed between two probes and sampling frequency, respectively.

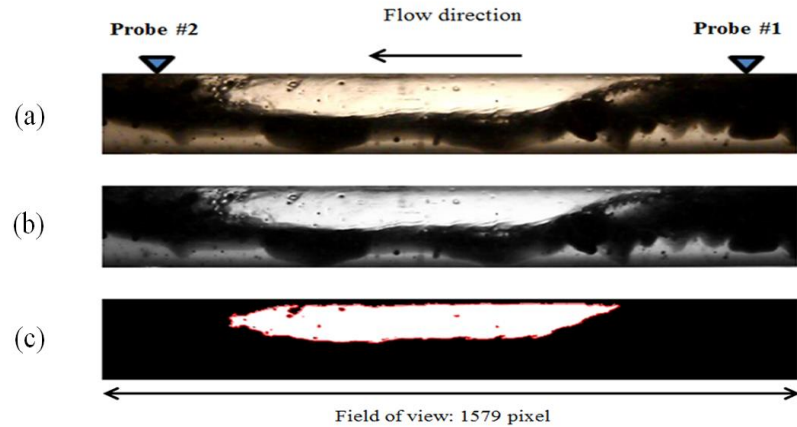


Fig. 7. Example of image post-processing to detect translational velocity, (a) original frame; (b) grey-scale image; (c) binarized frame. The triangles indicate the position of virtual probes


3. Experimental results

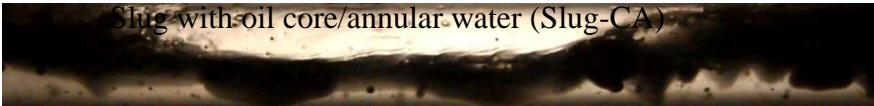

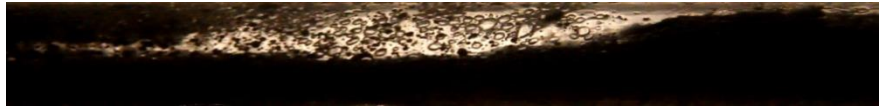
3.1 Flow pattern

Flow patterns were observed by using a digital video camera for viscous oil-water-air flow. As the slug flow was the main flow pattern, the classification was made based on the interaction between water and oil in liquid slug and film regions. Examples of monitored flow patterns are reported in Table 3 where the abbreviations are Slug-CA: Slug with oil core/annular water; Slug-FM-PGE: Slug with fully mixed oil/water and partial gas entrainment to the slug; Slug-FM-CGE: Slug with fully mixed oil/water and complete gas entrainment to the slug. The interactions between water and oil in film and slug body regions have not been widely known for such complex flow, however, it was investigated that at fixed gas velocity the film height for high viscosity flow is much greater (due to lower velocity) as compared to the lower viscosity flow (see e.g. Wang et al., 2013 and

Shmueli et al., 2015). Additionally, at lower and higher gas velocity, curvy and flat gas-liquid interface were observed, respectively (Table 3 a,b). A possible explanation for such behavior is because of decreasing the film region height with increasing J_g .

Table 3. Examples of the monitored flow patterns

Direction of flow


Flow pattern		J_o [ms ⁻¹]	J_w [ms ⁻¹]	J_g [ms ⁻¹]
(a)		0.36	0.66	0.57
(b)		0.59	0.88	1.34
(c)		0.77	1.32	1.14

The results are also shown on flow pattern map where gas and total liquid volumetric fluxes are considered as abscissa and ordinate, respectively at four different water cuts: $\epsilon_{LW}=0.48-0.64$. At fixed superficial liquid velocity, interaction between oil and water changes from core-annular flow to fully mixed flow (as shown itself by the presence of oil drops within water continuous flow) as gas velocity is increased (see, Fig. 8). Regardless of total superficial liquid velocity, this transition occurred for $J_g=0.6-0.8 \text{ m}\cdot\text{s}^{-1}$. Generally, at lower gas and liquid velocities, radial components of buoyancy forces overcome (which tend to keep the oil droplet off-center) drag forces, leading to formation of separated flow regime. Dispersion of oil into water in the film and slug body regions is highly dependent upon their input volumetric fluxes.

At higher liquid velocities (Table 3, image c), the Taylor bubble front is penetrated into the leading slug body tail and fragmented to smaller bubbles, as a result, they are transported and imbibed to the trailing slug body head. This phenomenon is affected by three main mechanisms, including shear stress caused by a shear layer between the high-velocity slug head and slow moving of leading liquid film, slug head vortex (which results from slug head circulation), and gas-carry in the proceeding liquid film (see Al-safran et al, 2015). The presence of mixing region at slug body front is not observed by visual inspection (see Table 3, a-c). Therefore, it is less likely that the second mechanism plays an important role for viscous oil-water-gas flow. Al-safran et al. (2015) evaluated the influence of high liquid viscosity on aeration of slug body for oil/gas flow and concluded that viscous forces overcome turbulent kinetic energy in the slug body mixing region, resulting in

reduction of entrainment rate. Although, interaction between the oil-water mixture and bubble in the slug head mixing region has still remained obscure for three phase flow, it seems from visual observation that the damping effect of viscous oil outweighs vortex intensity caused by water (less viscous phase).

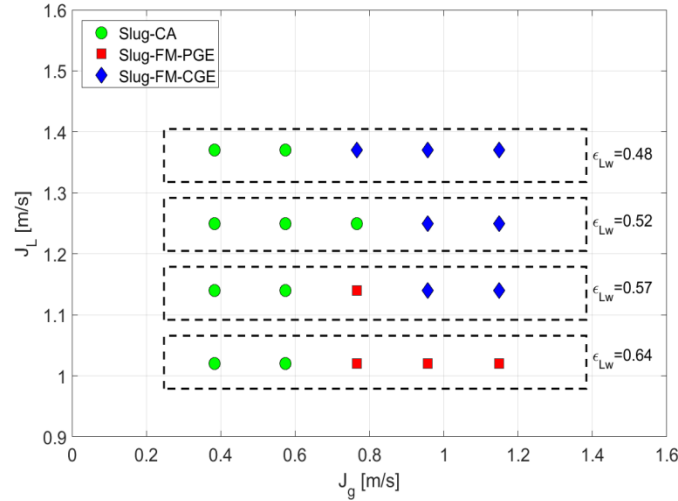


Fig. 8. Oil-water-air experimental flow map for four different Water cuts ($\epsilon_{LW}=0.48-0.64$). Slug-CA: Slug with oil core/annular water flow, Slug-FM-PGE: Slug with fully mixed oil/water flow and partial gas entrainment to the slug, Slug-FM-CGE: Slug with fully mixed oil/ water flow and complete gas entrainment to the slug

3.2 Distributed pressure drop

As stated earlier, the first objective of current research is to collect experimental data related to flows of viscous oil-water-gas. Thus, more than 200 tests were performed to measure three-phase pressure drop. Due to time constraints, the same measurement were repeated up to 5 times and the deviations resulted never higher than the measurement accuracy (i.e., ± 1.5 % of the pressure transducer full scale).

Fig. 9 depicts the results of pressure gradients plotted against ϵ_{Lg} , with superficial gas velocity as a parameter. Regular trends are observed, that is, for fixed amount of gas, increasing liquid superficial velocity (reducing gas to liquid ratio) would result in increasing of pressure gradient. At low gas values, this increase in pressure drop is more dramatic because we observed a transition from slug to plug flow regimes.

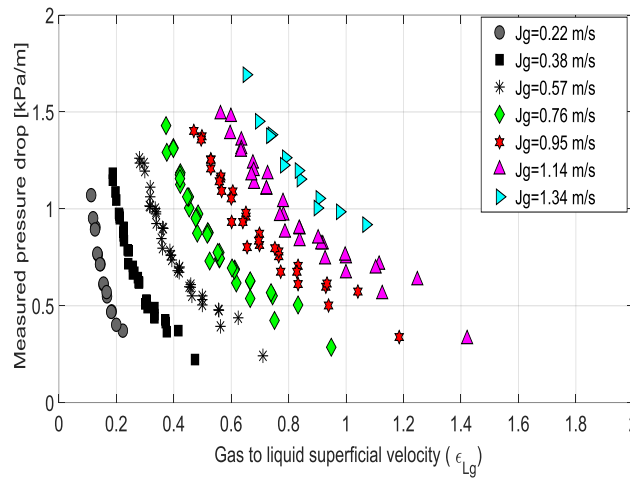


Fig. 9. Measured pressure drop ($\frac{\Delta P}{L}$) versus the ratio of gas to liquid superficial velocities (ϵ_{Lg})

Therefore, the pipe cross-sectional area has a wider contact with liquid, resulting in increasing frictional shear stresses. Analysis of pressure drop supports the assumption that liquid phases can be treated as equivalent liquid because no scattered data of pressure drop observed for fixed gas flow rate.

As investigated by other researchers (see e.g. Oliemans and Ooms, 1986, Colombo et al., 2015), oil-water core-annular flow regime is of practical interest for heavy oil transportation because of high stability and ability to reduce pressure drop. However,

as pressure is reduced the amount of dissolved gas in oil is released, so the presence of gas to liquid-liquid flow needs to be evaluated. It would be suitable to calculate pressure reduction factor for three phase flow to evaluate the influence of air addition to liquid-liquid flow, as performed by Bannwart et al. (2004). Fig. 10 depicts the results of pressure reduction factor versus the ratio of gas to liquid superficial velocity (ϵ_{LG}). Two different superficial oil velocities are investigated ($J_o=0.24 \text{ m}\cdot\text{s}^{-1}$ and $J_o=0.59 \text{ m}\cdot\text{s}^{-1}$). Pressure reduction factor has a physical meaning in a sense that if $R_{LG} < 1$ addition of gas has a positive effect and total frictional pressure drop is reduced. These results can be justified, considering the fact that gas has much lower viscosity than oil and water. Therefore, the increasing gas flow rate would result in reducing wall shear stresses and pressure gradient.

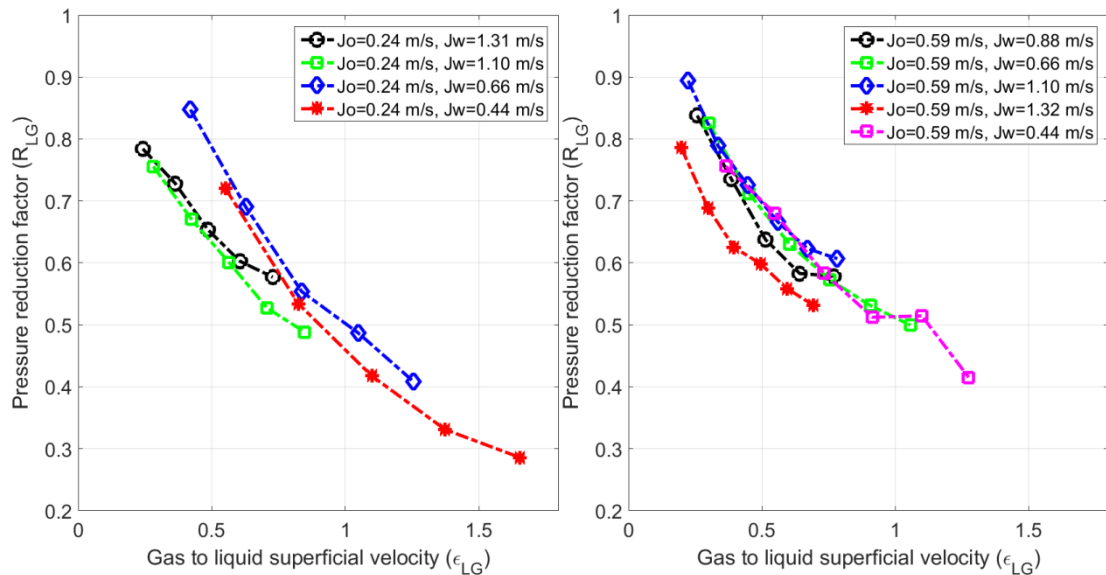


Fig. 10. Pressure reduction factor versus the ratio of gas to liquid superficial velocity for $J_o=0.24 \text{ m}\cdot\text{s}^{-1}$ and $J_o=0.59 \text{ m}\cdot\text{s}^{-1}$

Poesio et al. (2009) showed that this phenomenon occurs when flow regime of oil-water is core-annular. In conditions where transition from core-annular to stratified flow exists, no regular trend can be observed and repeatability is difficult to be obtained.

3.3 Slug body length

Because slug flow regime is a stochastic process, varied slug length can be discovered along the pipe fluctuated around its average value. So, it is widely accepted that the log-normal distribution can properly represent the slug body length as proposed by Losi et al (2016b), Fabre and Line (1992), and shown in Figs. 11-12. The results of slug body length distribution for two various superficial oil velocities ($J_o=0.48 \text{ m}\cdot\text{s}^{-1}$ and $J_o=0.71 \text{ m}\cdot\text{s}^{-1}$) are considered. On abscissa the slug body length normalized by pipe diameter and on ordinate the number of slugs captured by optical probe was plotted. In each plot, superficial gas velocity increases from left to right while superficial water velocity increases from top to bottom. Moreover, mean, median, and standard deviation values are presented. For fixed oil and water superficial velocity, as superficial gas velocity rises from $0.57 \text{ m}\cdot\text{s}^{-1}$ to $2.1 \text{ m}\cdot\text{s}^{-1}$, slug body length increases. The shape of distribution is shifted from highly right-skewed to normal like one. The number of slugs is considerably reduced due to the longer elongated bubble. As oil superficial velocity increases at constant water and gas superficial velocities, shorter slug length with higher frequency was observed. However, the shape of log-normal distribution remains unchanged. The effect of gas

superficial velocity is much more dramatic than liquid superficial velocity, probably because of pickup rate of slug by bubble (Bubble moving along stratified layer pickups liquid at their head), which finally depends on gas superficial velocity.

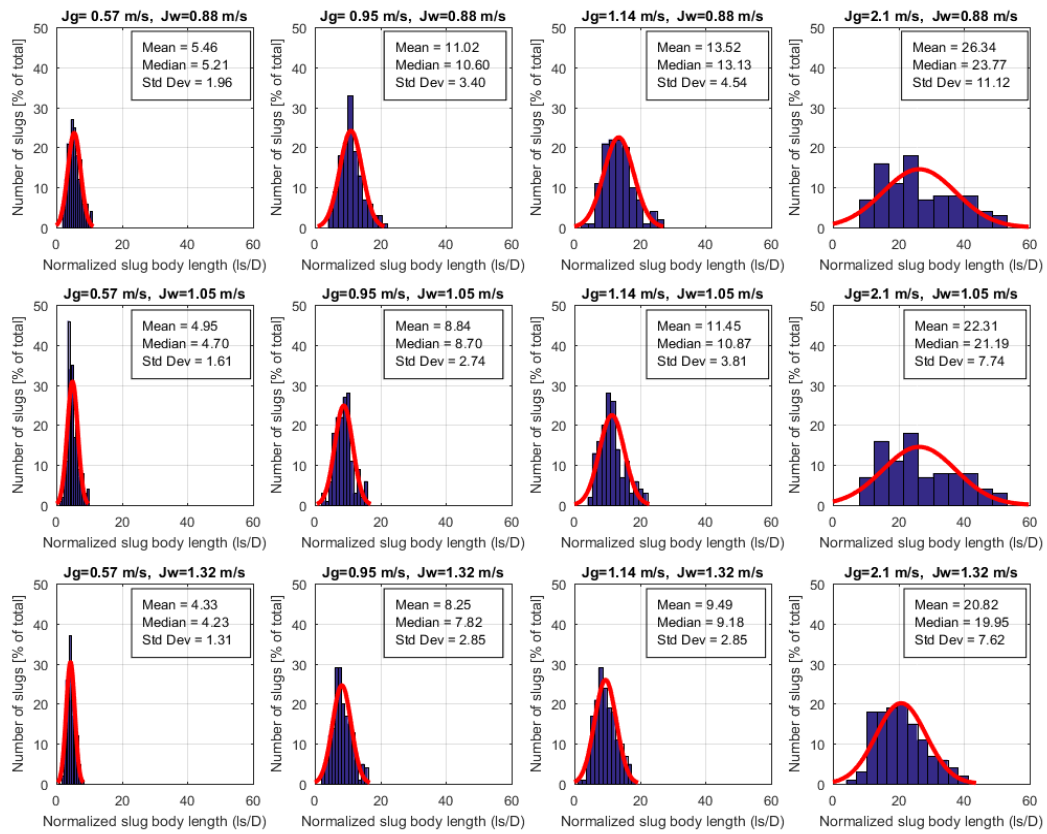


Fig. 11. Slug body length PDFs for $J_o = 0.48$ m/s

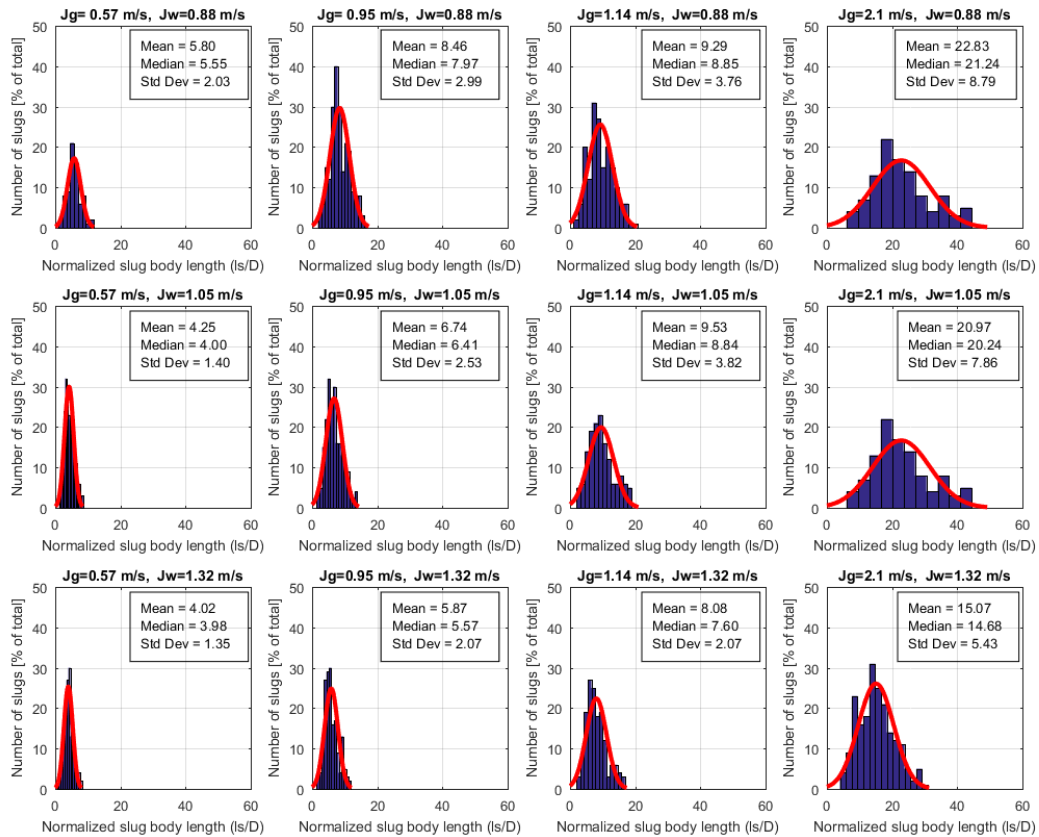


Fig. 12. Slug body length PDFs for $J_o = 0.71$ m/s

For two-phase flow of air and water, slug body length ranged between 12-30 D (Dukler and Hubbard, 1975). For two-phase flow with low viscosity of liquid, Barnea and Brauner (1985) proved that slug length is 32D. Al-Safran et al. (2011) measured liquid slug by means of laser and capacitance sensors in high viscosity oil-water. They concluded that average slug length of 10D was a reasonable approximation for high viscosity liquid-gas flow. According to the experimental measurements obtained from optical sensor, the mean slug body length ranged between 3D and 27D. Losi et al. (2016b) experimentally investigated high viscous oil-water flow within a straight duct. They proposed a correlation to compute slug

body length, considering the effect of both gas superficial velocity and pipe diameter.

$$\frac{L_s}{D} = K \cdot \left(J_g + \frac{J_{go}^2}{J_g} \right) \quad (13)$$

Where, K is a constant which depends on liquid properties and J_{go} is the gas volumetric flux relating to the shortest slug length. The numerical values of K and J_{go} were found to be 5.3 and 0.3 from data fitting regression. Slug length data is compared with the model developed by Losi et al. (2016b) in Fig. 13, while Table 4 shows the comparison between predicted slug length by empirical correlations (proposed for gas-liquid flow) and experimental counterpart. It is observed in Fig.13 that at low gas superficial velocity, transition from slug to plug and ultimately dispersed flow regime takes place, leading to infinite length of slug body. It is evident from Fig. 13 and Table 4 that Barnea and Brauner (1985), and Al-Safran et al. (2011) models are unable to describe the behaviors of our data, whereas the approach by Losi et al. (2016b) seems to be consistent, though affected by a rather large deviation. Average relative error between experimental data and model by Losi et al. (2016b) is found to be 20.8%, while the maximum relative error is 34%. As compared to other correlations in Table 4, the lower average relative error and standard deviation of Losi et al. (2016b) model suggests the strong influence of gas superficial velocity on slug body length.

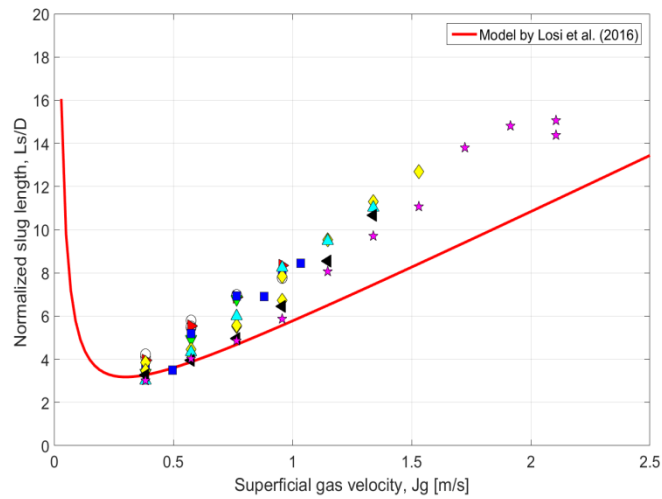


Fig. 13. Slug unit length measured by optical probe versus superficial gas velocity for varied superficial liquid velocity: \circ : $J_L = 1.36$ m/s, \triangleright : $J_L = 1.41$ m/s, ∇ : $J_L = 1.53$ m/s, \square : $J_L = 1.68$ m/s, \diamond : $J_L = 1.7$ m/s, \triangle : $J_L = 1.8$ m/s, \triangleleft : $J_L = 1.92$ m/s, \star : $J_L = 2.04$ m/s

Table 4. Comparison of slug length for different correlations of gas-liquid flow

Correlation	Avg. relative error (%)	Max. relative error (%)	St. deviation (%)
Barnea and Brauner (1985)	454.2	963.1	228.5
Al-Safran et al.(2011)	79.7	232.2	64
Losi et al. (2016b)	20.8	34	10.4

3.4 Slug frequency

The average number of slugs passing through sensors at specified time, inspected by a stationary observer is defined as slug frequency (f_s). The results of slug frequency as a function of superficial gas velocity are presented in Fig. 14. As superficial gas velocity increases, decreased slug frequency was observed, showing the presence of larger elongated bubbles and longer distances between liquid slugs. At fixed gas flow rate, it can be realized that slug frequency increases and shorter slugs form by increasing liquid superficial velocity. The agreement between the observed trends of slug frequency (Fig. 14) with the data of Issa et al. (2007) was satisfactory.

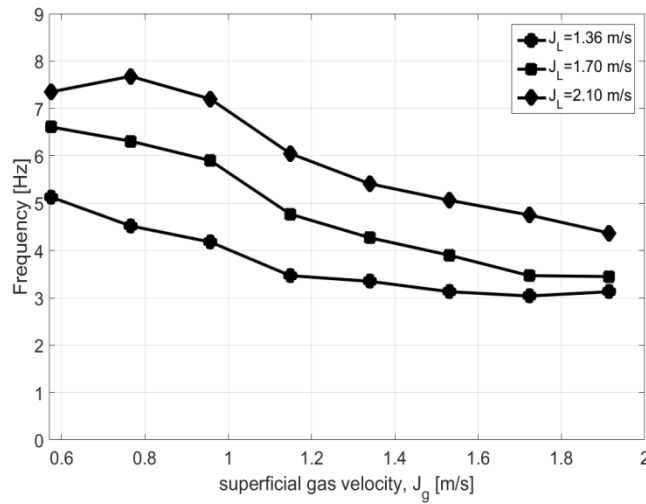


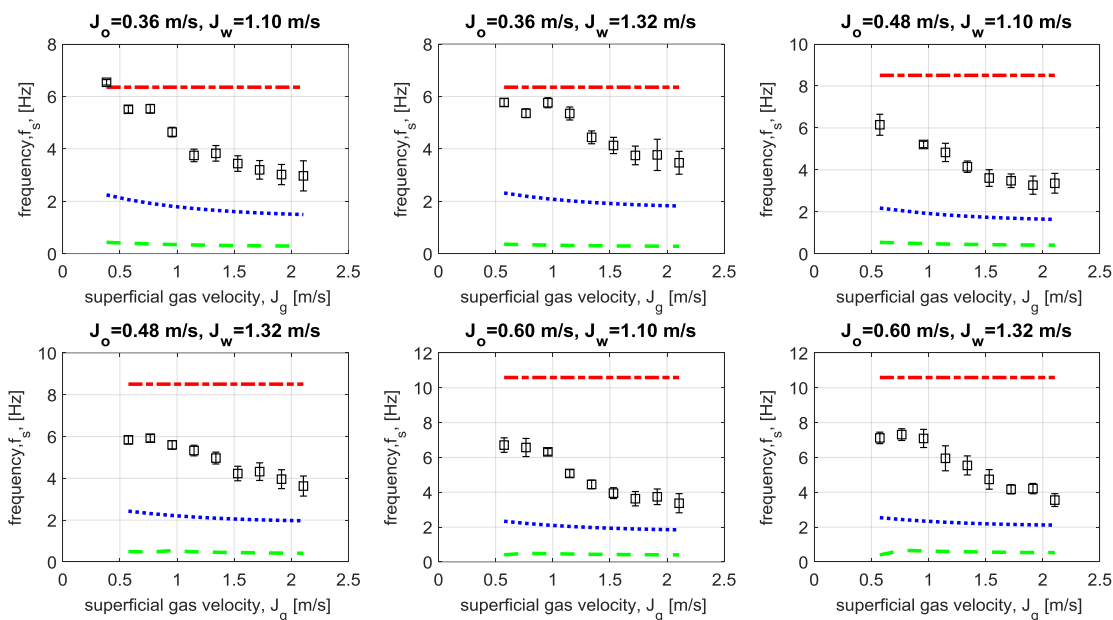
Fig. 14. Slug frequency versus superficial gas velocity for $J_L=1.36$ - 2.10 $\text{m}\cdot\text{s}^{-1}$

In order to check the possibility to broaden the application of gas-liquid flow models to compute slug frequency for gas-liquid-liquid flow, the experimental data of slug frequency for different superficial oil and water velocities (ranged within the intervals of $0.36 < J_o < 0.60$ m/s and $1.05 < J_w < 1.32$ m/s) were compared with some of

the most used correlations, listed in Table 5. A detailed description of these models was presented in the study of Hernandez-Perez et al. (2010). Among the models reported in Table 5, the correlation developed by Gokcal (2010) is based on high viscosity data and does not consider the influence of superficial gas velocity. The models developed by Gregory and Scott (1969), Greskovich and Shrier (1972) were validated for low viscosity liquid-gas flow. It is worth noting that the value of J_o was replaced by J_L for definition of $\lambda_L = J_o/J_t$ in the latter, to account for the effect of superficial water velocity. The results of comparisons between predicted values by the correlations and experimental counterpart are depicted in Fig. 15. The error bars are also presented, showing standard deviation of each slug frequency (which implies slug velocity and length) around its average value. As can be seen, all the models were not capable of describing the behavior of viscous oil-water-gas flow (see Table 6 for statistical analyses of the model's performance).

Table 5. Slug frequency correlations for gas-liquid flows from literature

Reference	Correlation	Additional information
Gregory and Scott (1969)	$f_s = 0.0226 \left[\frac{J_o}{gD} \left(\frac{19.75}{J_t} + J_t \right) \right]^{1.2}$	g:gravitational acceleration D:pipe diameter
Greskovich and Shrier (1972)	$f_s = 0.0226 \left[\lambda_L \left(\frac{2.02}{D} + Fr_m^2 \right) \right]^{1.2}$	$\lambda_L = J_o/J_t$ $Fr_m = J_t/\sqrt{gD}$
Gokcal et al. (2010)	$f_s = 2.623 \frac{1}{N_f^{0.612}} \frac{J_o}{D}$	$N_f = D^{3/2} \sqrt{\rho_o \Delta \rho g} / \mu_o$



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Fig. 15. Comparison between predicted and measured slug frequency. Experimental data is shown by square symbol while dashed green line, dotted blue line, and dash-dot red line represents the prediction by the models of Gregory-Scott (1969), Greskovich-Shrier (1972), and Gokcal et al. (2010), respectively.

Table 6 Frequency relative errors for some gas- liquid prediction correlations

Correlation	Avg. relative error (%)	Max. relative error (%)	St. deviation (%)
Gregory and Scott (1969)	-90.6	-85.0	2.1
Greskovich and Shrier (1972)	-56.2	-40.4	6.6
Gokcal et al. (2010)	87.1	214.2	52.8

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3.5 Bubble and slug unit length

In this section, the results of bubble and slug unit length, measured by optical sensors are presented. Fig. 16 illustrates elongated bubble length as a function of superficial gas velocity for varied superficial liquid velocities. Not surprisingly, increases in gas flow rates would result in longer bubbles. The results regarding slug unit measurements by optical probe versus gas superficial velocity is shown in Fig. 17. Liquid volumetric flux is considered as a parameter. In Figs. 16 and 17, different symbols denote various superficial liquid velocities. Apart from liquid superficial velocity, a regular trend is observed, suggesting considerable effect of superficial gas velocity on total slug unit length.

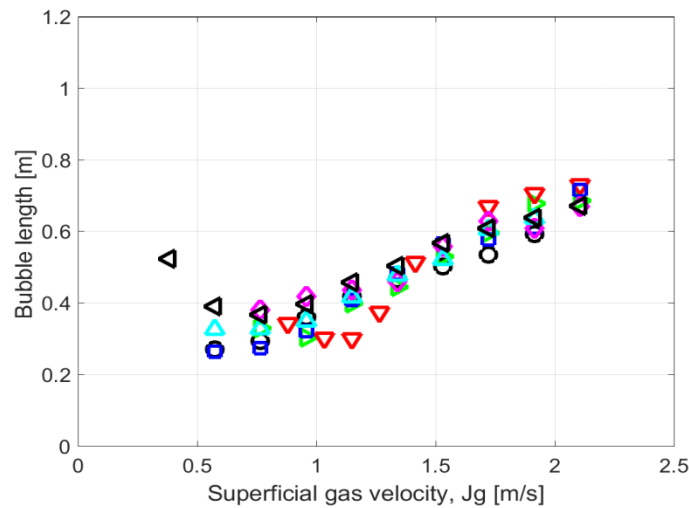
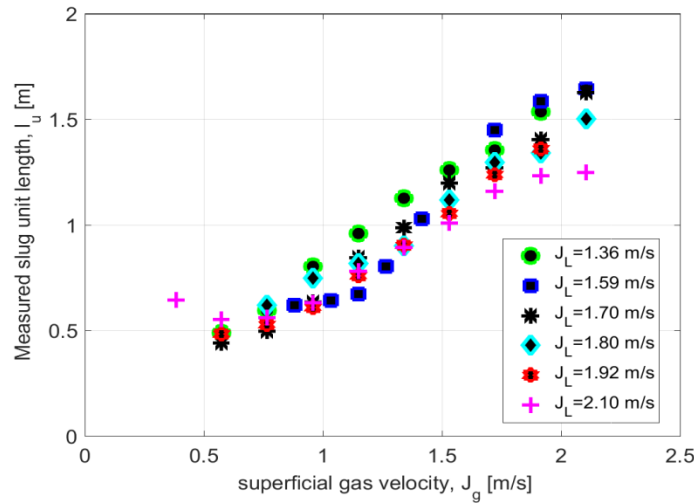


Fig. 16. Measured bubble length by optical probe versus superficial gas velocity for fixed liquid velocity; $\circ: J_L = 1.36$ m/s, $\diamond: J_L = 1.48$ m/s, $\Delta: J_L = 1.59$ m/s, $\nabla: J_L = 1.7$ m/s, $\triangleright: J_L = 1.80$ m/s, $\square: J_L = 1.92$ m/s, $\triangleleft: J_L = 2.03$ m/s

512



513

514 **Fig. 17.** Measured slug unit length by optical probe versus superficial gas
 515 velocity; \circ : $J_L = 1.36$ m/s, \square : $J_L = 1.59$ m/s, $*$: $J_L = 1.70$ m/s, \diamond : $J_L = 1.80$ m/s,
 516 \star : $J_L = 1.92$ m/s, $+$: $J_L = 2.10$ m/s

517 As can be seen from Fig. 17, it is possible to justify the behavior of slug unit length based on
 518 the physical mechanism: if the gas velocity is decreased, the structural form of slug flow regime
 519 changes into plug flow and finally followed by dispersed regime. Therefore, as a result of
 520 liquid entrainment into gas bubble, a minimum total slug length is reached. Conversely, when
 521 gas velocity is increased, the total slug unit length increases, following an exponential trend,
 522 as depicted later in Fig. 19.

523 3.6 Bubble translational velocity

524 Understanding the bubble translational velocity is crucially important because
 525 almost all mechanistic models require the information of this parameter. Bubble
 526 translational velocity has been measured by means of cross-correlation technique,

using optical sensor and video camera. It is customary to correlate bubble translational velocity to mixture superficial velocity. Fig. 18 a-b shows the results of bubble translational velocity, measured by optical probe and video camera, respectively. A linear dependency of bubble translational velocity on mixture superficial velocity is marked for both cases. Larger dispersion of data is observed for measurement of bubble translational velocity using video camera, probably due to the lower sampling frequency of video as compared to optical sensor. As first represented by Nicklin (1962), the experimental data are fitted with the regression line of $U_t = C_1 \cdot J_t + C_0$, where slope (C_1) and intercept (C_0) of line denotes the distribution parameter and drift velocity (explained by Fabre and Line (1992) as a gas bubble velocity moving through stagnant liquid), respectively. Hence, the values of C_0 becomes 1.2 and 2.0 for fully developed turbulent and laminar flow, respectively (see, e.g. Foletti et al, 2011).

As can be observed from Fig. 18 (a-b), data are well fitted to the correlation of Wallis (1969) with $C_0=1.2$ and $C_1=0$. The values of drift velocity are in agreement with other experimental investigations, Farsetti et al. (2014), Losi and Poesio (2016). The latter evaluated the influence of oil viscosity on drift velocity of a gas bubble moving in liquids for different axial positions in both horizontal and inclined pipes. They concluded that drift velocity for viscous oil-gas flow ($\mu_o=0.804 \text{ Pa}\cdot\text{s}$) is ranged between $0.0025\text{-}0.0065 \text{ m}\cdot\text{s}^{-1}$ for different axial positions in a horizontal pipe, which can be approximated equal to zero. Although the literature survey immensely suffers from the lack of experimental data and theoretical modeling to compute translation bubble velocity for viscous oil-water-air flow, the above analysis showed that

application of drift-flux model (which is originally developed for gas-liquid flow) for three phase flow of viscous oil-water-air leads to a reasonable approximation.

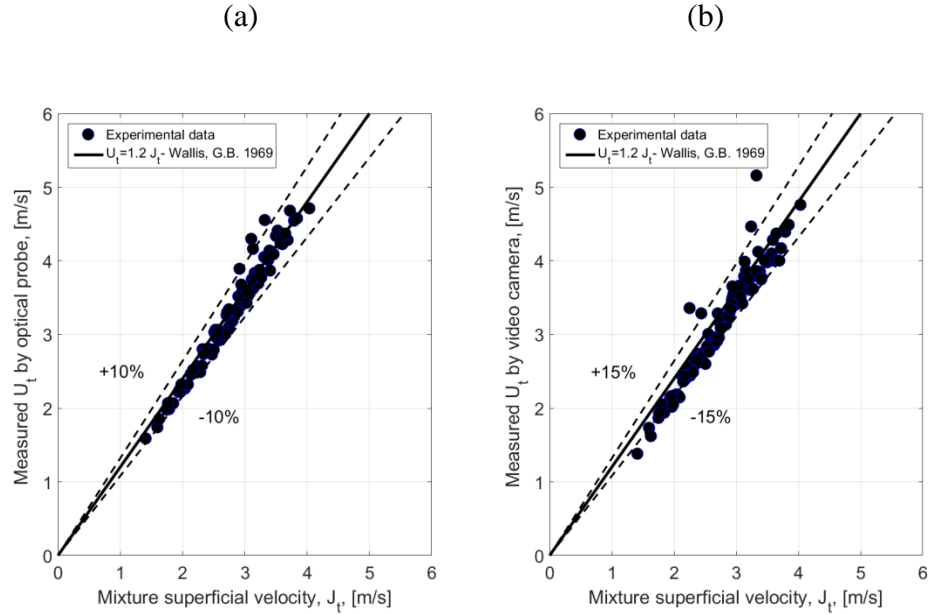


Fig. 18. Measured bubble translational velocity (U_t) versus mixture superficial velocity (J_t) by means of cross-correlation technique using (a) optical sensor, (b) video camera

3.7 Model development for slug unit length

An attempt was made to correlate the slug unit length to the input flow variables and pipe diameter. It is worth noting that as the collected data in the present study has a constant surface tension and viscosity, only inertia forces and pipe diameters are included in the modeling of total slug unit length. The slug unit length measured by optical probe is plotted against dimensionless parameter $(1+\varepsilon_{LG})$ in Fig. 19 to evaluate the influence of liquid and gas volumetric fluxes on slug unit length. At fixed liquid superficial velocity, increases gas flow rate causes increasing in slug unit length. All data collapse on a common line, so it is customary to express slug unit length normalized by pipe diameter with a power law functional form as function of operating conditions, that is:

$$\frac{l_u}{D} = C \cdot (1 + \varepsilon_{LG})^n \quad (14)$$

From experimental measurements of total unit length measured by optical sensor, coefficients C and n are found to be 7.3 and 2, respectively. Equation 14 is valid for $0.36 < J_o < 0.71 \text{ m}\cdot\text{s}^{-1}$, $0.44 < J_w < 1.32 \text{ m}\cdot\text{s}^{-1}$, and $0.22 < J_g < 2.10 \text{ m}\cdot\text{s}^{-1}$, internal diameter of 40 mm, and oil viscosity of $0.838 \text{ Pa}\cdot\text{s}$. Estimation above and beyond this range has to be done by caution as it might produce inadequate results. The model presented here has to be considered as an operative tool to compute the total slug unit length; however, it does not aim to describe the physical mechanism behind such a complex flow. Table 7 lists predicted slug unit cell from proposed model and experimental data (with respective measured bubble and slug body lengths) from current study. Those data are provided that the associated flow regimes are available.

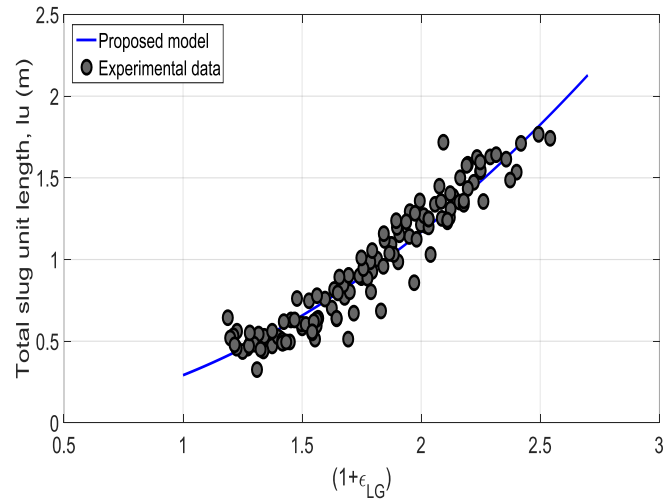


Fig. 19. Experimental total slug unit length measured by optical probe as a function of dimensionless parameter $(\frac{J_t}{J_L})$.

Table 7. Comparison of predicted slug unit cell and experimental data

J_o ($\text{m}\cdot\text{s}^{-1}$)	J_w ($\text{m}\cdot\text{s}^{-1}$)	J_g ($\text{m}\cdot\text{s}^{-1}$)	L_b (Measured) (m)	L_s (Measured) (m)	L_u (Measured) (m)	L_u (Prediction) (m)	Relative error (%)	Flow regime
0.36	1.05	0.57	0.19	0.29	0.49	0.57	17.5	Slug-CA
		0.76	0.26	0.30	0.57	0.69	21.4	Slug-FM-PGE
		0.95	0.37	0.34	0.71	0.82	15.1	Slug-FM-PGE
		1.14	0.52	0.43	0.9	0.96	0.15	Slug-FM-PGE
		1.91	0.94	0.55	1.50	1.61	7.8	Slug-FM-CGE
		2.1	1.01	0.59	1.60	1.80	1.6	Slug-FM-CGE
1.32		0.57	0.16	0.32	0.48	0.52	7.4	Slug-CA
		0.76	0.24	0.36	0.60	0.61	2.0	Slug-FM-CGE
		0.95	0.28	0.32	0.61	0.71	17.4	Slug-FM-CGE

		1.14	0.39	0.38	0.78	0.82	5.6	Slug-FM-CGE
		2.1	0.90	0.61	1.52	1.48	-2.8	Slug-FM-CGE
0.71	0.66	0.38	0.14	0.29	0.43	0.47	9.3	Slug-CA
		0.57	0.20	0.29	0.50	0.58	17.1	Slug-CA
	1.32	0.57	0.11	0.35	0.46	0.48	4.2	Slug-FM-CGE
		0.76	0.14	0.32	0.47	0.55	17.7	Slug-FM-CGE
		0.95	0.18	0.35	0.54	0.63	15.8	Slug-FM-CGE
		1.14	0.27	0.42	0.70	0.71	1.9	Slug-FM-CGE
		2.10	0.58	0.61	1.20	1.21	0.69	Slug-FM-CGE

586

587 Due to the lack of experimental data regarding total slug unit for three phase flow of viscous
588 oil-water-air in the literature survey, the proposed model is only compared with the correlation
589 developed by Cook and Behnia (2000b). They presented their model for water-air flows, taking
590 into account bubble length (L_b), superficial gas velocity, and liquid holdup in slug body region.

591
$$l_u = \frac{L_b[(1-H_{lf}) \cdot U_t - J_g]}{J_g} \quad (15)$$

592 Where H_{lf} is the mean liquid holdup in film section. They concluded that the
593 liquid holdup at bubble head ($H_{lfe} \sim \frac{U_t - J_t}{U_t}$) is different from mean value and
594 suggested that H_{lf} can be calculated as $H_{lf} \sim 1.4 H_{lfe}$. To make use of equation 15,
595 information of bubble length is required, which can be obtained from experimental
596 data. Fig. 20 illustrates the parity plot of predicted slug unit length against
597 measured slug units using 106 data points. As it is evident, the total slug unit

length predicted by proposed model is consistent with experimental data (87% of all data fall within $\pm 20\%$ relative error), with the mean absolute relative error of 10.7%. However, The model by Cook and Behnia (2000b) shows a strong under-prediction, suggesting inability of gas-liquid flow models to predict the slug unit length (the effect of the third phase (oil) is not considered for model development).

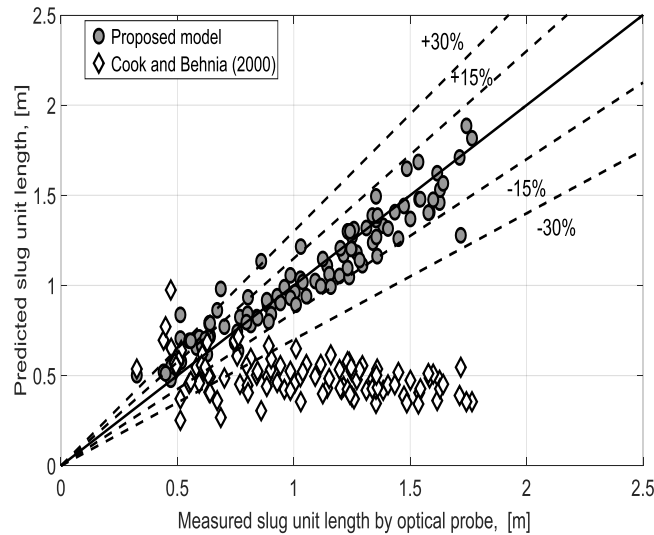


Fig. 20. Parity plot of comparison between experimental data and predicted slug unit length; dashed lines represent $\pm 15\%$ and $\pm 30\%$ deviation from bisector

4. Conclusion

In this paper, we experimentally studied and analyzed the results of three-phase flow of viscous oil-water-gas in horizontal pipe. A large number of pressure drop data set is presented. Due to limitation on operating conditions, only slug flow regime was considered with oil viscosity of 0.83 Pa.s.

Slug body, elongated bubble and total slug unit lengths are experimentally measured by optical probe. Statistical analysis of slug body length was performed,

enabling us to characterize slug flow based on probability density function (PDFs). It was found that superficial gas velocity had a profound effect on slug body and bubble length, that is, the higher gas superficial velocity, the longer slug body and bubble length was observed. A new correlation for slug unit length is developed based on experimental data obtained by optical probe for three phase flow of viscous oil-water-gas in horizontal pipe. A good agreement was observed between predicted slug unit length and measurements, considering almost all data fall into ± 20 deviation from actual values. Translation velocity of slug unit is measured by both optical probe and image processing techniques. A modified version of drift-flux model developed by Nicklin (1962) used to validate experimental data of translational bubble velocity, showing a good agreement with actual data.

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