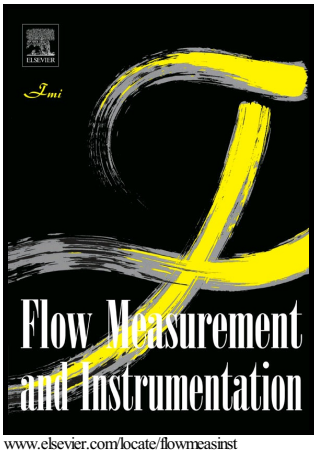

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Estimating Slug Liquid Holdup in High Viscosity Oil-Gas Two-Phase Flow

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ABSTRACT

Slug flow is one of the most critical and often encountered flow patterns in the oil and gas industry. It is characterised by intermittency which results in large fluctuations in liquid holdup and pressure gradient. A proper understanding of its parameters (such as slug holdup) is essential in the design of transport facilities (e.g. pipelines) and process equipment (slug catchers, separators etc.). In this paper, experimental investigation of slug liquid holdup (defined as the liquid volume fraction in the slug body of a slug unit) is performed. Mineral oil with viscosity, $\mu = -0.0043T^3 + 0.0389T^2 - 1.4174T + 18.141$ and air were used as test fluids. A 0.0254 m and 0.0762 m pipe internal diameters facilities with pipe lengths of 5.5 and 17 m respectively were used in the study. Electrical Capacitance Tomography was used for slug holdup measurements. Results obtained in the study shows that slug liquid holdup varied directly as the viscosity and inversely as the gas input fraction. Existing slug holdup correlations and models in literature did not sufficiently predict present experimental results. A new empirical predictive correlation for estimating slug liquid holdup was derived from present experimental databank and from data obtained in literature. The databank's liquid viscosity ranges from 0.189 – 8.0 Pa.s. Statistical analysis of the new correlation vis-à-vis existing ones showed that the present correlation gave the best performance with an average percent error, E_1; absolute average percent error, E_2 and standard deviation, E_3 of 0.001, 0.05 and 0.07 respectively, when tested on the high viscosity liquid–gas databank.

NOMENCLATURE

Symbol	Description	Unit
ρ	Density	kg/m^3
B_O	Bond Number	-
C_e	Pipeline Inclination Coefficient	-
f_s	friction factor at the pipe wall for the liquid slug body	-
Fr_m	Mixture Froude Number	-
$H_{LS}; H_s; E_s; \alpha$	Slug Liquid Holdup	-
Re_g	Gas Reynolds Number	-
Re_l	Liquid Reynolds Number	-
V_M	Mixture velocity	m/s
$V_{SL}; V_{sl}$	Liquid superficial velocity	m/s
V_T	Slug Translational Velocity	-
V_f	Film Velocity	m/s
h_a	Instantaneous liquid holdup	-
h_{th}	Liquid holdup threshold	-
l_f	Length of slug film	-
l_s	Length of liquid slug body	-
E_f	Liquid holdup of slug film	-
N_μ	Viscosity Number	-
T_{sm}	Slip velocity	N/m^2
ε_0	Electric constant	F/m
μ_M	Mixture velocity	m/s
μ_{eff}	Effective Viscosity	cP
μ_g	Gas Viscosity	cP
μ_l	Liquid Viscosity	cP
ρ_G	Gas Density	kg/m^3
ρ_{TP}	Two phase density	Kg/m^3
ρ_l	Liquid Density of water	kg/m^3
σ_{GL}	Gas-Liquid Interfacial tension	F/m
A	Inclination factor	^{o}
C	Correction factor	-
D	Pipe internal diameter potential V is placed on i.	m
Fr, N_{Fr}	Froude Number	-
T	Temperature	^{o}C
Θ	Dimensionless momentum transfer	-
θ	Angle of pipeline inclination from the horizontal	^{o}
φ	Dimensionless parameter	-

1 Introduction

Concurrent liquid-gas two-phase flow in conduits occurs frequently in several industrial processes. Under this scenario, the geometrical distribution of the two phases in the conduits can take up different patterns depending on the gas and liquid flow rates, fluid physical properties and the geometry of the conduit. One of such patterns is the slug flow. In the oil and gas industry, slug flow is ubiquitous, the understanding of its formation and hydrodynamic characteristics is essential in several applications. It is often undesirable due to its characteristic intermittency, which results in large fluctuations in liquid holdup and pressure gradient. Slug flow consists of two distinct parts; an initiating stratified liquid-gas flow with growing interfacial waves and a predominantly liquid slug that emanates from the interfacial waves' growth, which eventually bridges the cross sectional area of flow. The former is termed slug film region while the latter is the slug liquid body (slug body). Slug liquid body is relatively energetic with higher translational velocity relative to the film region. The fraction of liquid in the slug liquid body is called the slug liquid holdup. Measurement and predictions of the slug liquid holdup is essential in the design of transport facilities (e.g. pipelines) and process equipment (slug catchers, separators etc.). It is also important in the process control and structural integrity management of pipelines. Slug liquid holdup is also required as input variable in slug flow models such as the ones proposed by Dukler & Hubbard (1975), Taitel & Barnea (1990) and Zhang et al. (2003).

Researchers have conducted several studies on slug liquid holdup in liquid-gas two-phase flow. One of the earliest work was conducted by Gregory et al. (1977). The authors measured the average slug liquid holdup, H_s for an estimated 30 slugs observed for each of their test condition in a 0.0258 and 0.0512 m pipe internal diameters. Viscosity and density of the oil used in their study were 0.00675 Pa.s and 858 kg/m^3 (at 25 °C) respectively. They proposed a correlation for liquid holdup, H_{ls} as:

$$H_{ls} = \frac{1}{1 + \left(\frac{V_M}{8.66}\right)^{1.39}} \tag{1}$$

where V_M is defined as the mixture velocity ($V_{SG} + V_{SL}$) in m/s. The authors stated that the correlation was not reliable beyond a mixture velocity of 10 m/s. However, the proposed correlation did not account for the fluid physical properties (such as density, viscosity etc.) and it was also dimensionally inconsistent. Pereyra et al. (2012) evaluated the predictive

performance of Gregory et al. (1977) correlation They concluded that it was the best performing correlation in terms of simplicity and accuracy on a low liquid viscosity horizontal and near-horizontal data bank.

Malnes (1983) studied slug liquid holdup using air and light oil as test fluids. The internal diameters of pipes used in his study were 25.8 and 51.2-mm. In proposing his model, Malnes (1983) modified Gregory et al. (1977) correlation. The modified correlation accounted for fluid physical properties such as surface tension and liquid density. The modified correlation was given by:

$$H_{ls} = 1 - \frac{V_M}{\left[V_M + 83\left(\frac{g\sigma_{GL}}{\rho_L}\right)^{1/4}\right]} \tag{2}$$

Ferschneider (1983) performed experiments on a flow loop with ID pipe 146-mm and inclination angle 0 – +4°. Natural gas and condensate were used as test fluid to experimentally study void fraction in the slug body. A correlation was proposed for void fraction in the liquid slug body. Here, we present a modified version which yields the slug liquid holdup, proposed thus:

$$H_{ls} = \frac{1}{\left\{1 + \left[\frac{V_m}{\sqrt{\left(1 - \frac{\rho_G}{\rho_L}\right) gD}}\right]^2 \Big/ \left(\frac{A}{Bo^{\beta}}\right)^2\right\}^2} \tag{3}$$

where Bo, the Bond number is given by:

$$Bo = \frac{(\rho_L - \rho_G) g D^2}{\sigma} \qquad 0 < \theta \leq 90°$$

Abdul-Majeed (1999) developed a new empirical correlation based on 423 data points from available literature. The databank spanned the following range: pipe inclination, -10 – 9°, pipe internal diameter, 0.0258 – 0.171-m and liquid viscosity, 0.001 – 0.007 Pa.s. He proposed a correlation based on the mixture velocity and provided a correction factor, C for the fluid viscosity and A to account for the effect of inclination thus:

$$H_{ls} = (1.009 - CV_M)\, A \tag{4}$$

where

$$C = 0.006 + 1.3377\frac{\mu_g}{\mu_l}$$

$$A = \begin{cases} 1 & \varphi \leq 1 \quad \text{(horizontal and downward flow)} \\ 1 - \sin\varphi & \varphi > 1 \quad \text{(upward flow)} \end{cases}$$

Marcano et al. (1998) studied slug liquid holdup in a 78-mm ID pipe using air and low viscosity oil as test fluids. The authors proposed a correlation for the slug liquid holdup thus:

$$H_{ls} = 1 - \frac{1}{\left[1.001 + 0.0179V_M\left(0.0179{V_M}^2\right)^{1/4}\right]} \tag{5}$$

Gomez et al. (2000) used 283 experimental data points from a variety of internal pipe diameters ranging from 0.051 – 0.203-m, pipe inclinations varying from 0 – 90°. Liquids used in the databank were water and light oil with viscosity varying from 0.001 – 0.065 Pa.s. Freon and air were the gas in the databank. Gomez et al. (2000) developed a new dimensionless predictive correlation thus:

$$E_s = e^{-(0.45\theta + C\,Re)} \qquad 0 < \theta \leq 90° \tag{6}$$

where θ is the angle of inclination from the horizontal, $C = 2.48 \times 10^{-6}$ and the Reynolds number, Re is defined as:

$$Re = \frac{\rho_L V_M D}{\mu_L}$$

The correlation was evaluated against experimental results obtained from Nuland et al. (1997) with an absolute average error of 14.2%.

Zhang et al. (2003) developed an analytical model for the prediction of slug body liquid holdup. The model was developed on the basis of a balance between the turbulent kinetic energy of the liquid phase and the surface free energy of dispersed spherical gas bubbles, it is proposed thus:

$$E_s = \frac{1}{1 + \frac{T_{sm}}{3.16[(\rho_L - \rho_G)g\sigma_L]}} \tag{7}$$

where

$$T_{sm} = \frac{1}{C_e}\left[\frac{f_s}{2}\rho_S {V_M}^2 + \frac{D}{4}\frac{\rho_L E_f (V_T - V_f)(V_M - V_f)}{L_S}\right] \quad (8)$$

$$C_e = \frac{2.5 - |\sin\theta|}{2}$$

T_{sm} is a function of the wall shear stress, slug length, translational velocity and is also affected by momentum exchange between liquid slug and liquid film in the slug unit. This model is however complex. Pereyra et al (2012) in their evaluation concluded that Zhang et al. (2003) model was computationally complicated and the most accurate in predicting slug holdup for the entire databank.

Al-Safran (2009) developed a nonlinear regression model using a databank that consisted of 410 experimental data for a wide range of fluid physical properties, operational and geometrical conditions. A mechanistic feature defined as the dimensionless momentum transfer rate between the slug body and liquid film was implemented in the model. A simplified form of the parameter is given by:

$$\Theta = \frac{H_f (V_T - V_f)(V_M - V_f)}{{V_M}^2} \quad (9)$$

and the final form of the model is given by:

$$H_{LS} = 1.05 - \frac{0.0417}{\Theta - 0.123} \quad (10)$$

The model was validated against limited data for an air-oil two phase flow system.

In recent years, following increasing interests in unconventional fossil fuel resources like heavy oil, a few studies have been conducted on slug liquid holdup in high viscosity liquid-gas flows.

Kora et al. (2011) investigated the effect of high liquid viscosity on slug liquid holdup using a test facility with 0.0508 m pipe ID. Oil viscosities tested were 0.587, 0.378, 0.257 and 0.181 Pa.s.. Kora et al. (2011) accounted for the influence of inertial and viscous forces by using non-dimensional groupings of Wallis (1969). The groupings defined as dimensionless viscosity number, N_μ (to account for viscous forces) and the dimensionless Froude number, Fr (to account for inertia and gravity forces) thus:

$$N_\mu = \frac{V_m \mu_l}{gD^2(\rho_l - \rho_g)} \tag{11}$$

$$Fr = \frac{V_m}{(gD)^{0.5}} \sqrt{\frac{\rho_l}{\rho_l - \rho_g}} \tag{12}$$

Kora *et al.* (2011) based on experimental data, proposed a new slug holdup thus:

$$E_S = \begin{cases} 1.012\exp\left(-0.085 Fr {N_\mu}^{0.2}\right) & 0.15 < Fr {N_\mu}^{0.2} < 1.5 \\ 0.9473\exp\left(-0.041 Fr {N_\mu}^{0.2}\right) & Fr {N_\mu}^{0.2} \geq 1.5 \\ 1.0 & Fr {N_\mu}^{0.2} \leq 0.15 \end{cases} \tag{13}$$

Within the experimental test matrix and fluid physical properties studied, significant effect of liquid viscosity on slug body was observed. Kora et al. (2011) conducted performance evaluation of the Gregory et al. (1977), Zhang et al. (2003) and Al-Safran (2009) correlations. The proposed correlation performed very well when tested against high viscosity data used for its development. They observed that their proposed correlation gave good predictions relative to others when mixture velocity was less than 2.0 m/s, significant discrepancies in prediction were however observed at high mixture velocities.

Xu (2012) studied slug flow in Newtonian and non-Newtonian liquids-gas flows. Tap water was used as the test Newtonian liquid while carboxymethyl cellulose (CMC) solution was used as the non-Newtonian liquid. They observed that the slug liquid holdup declined with increase in inclination angle for the Newtonian liquid and gas two-phase flow. For gas and non-Newtonian liquid two-phase flow, slug liquid holdup increased significantly at a particular mixture velocity as the liquid phase becomes more shear-thinning. This characteristic behaviour was more prominent at higher mixture velocity. For the gas-Newtonian liquid two-phase flow, the author proposed a modification to the Gregory et al. (1977) model based on his studies experimental database thus:

$$\alpha_S = \frac{1}{1 + \left(\frac{V_M}{9.514}\right)^{1.274}}; \quad 0.1\ m/s \leq V_M \leq 20\, m/s \tag{14}$$

For the gas and non-Newtonian liquid flow, they propose a new correlation thus:

$$\alpha_S = \frac{(1-\sin\theta)^{0.05}}{1+3.166\times10^{-5}{Re_L}^{1.225}}; \qquad 0° \le \theta \le 75° \tag{15}$$

where

$$Re_L = \frac{\rho_l D V_M}{\mu_{eff}} \text{ and } \mu_{eff} = D^{1-n}8^{n-1}k.V^{n-1}$$

Al-Safran et al. (2015) experimentally investigated the effects of high viscosity liquid on slug liquid holdup in horizontal pipes. They developed an empirical non-linear regression model as a function of two non-dimensional numbers defined by Wallis (1969). Data utilized for the model development ranged from liquid viscosity of 0.180 – 0.587 Pa.s. Comparative analysis of results obtained in this study was made with data obtained from Gregory et al. (1977) and Nädler & Mewes (1995) which were for liquid viscosities of 0.001 and 0.007 Pa.s respectively. The authors observed that a critical mixture velocity exists at which slug aeration process was initiated and this was a function of liquid viscosity with the critical value increasing as liquid viscosity decreased. Above the critical mixture velocity, high viscosity liquid had higher slug liquid holdup compared to the low viscosity data in literature. They attributed this to the less turbulent energy in the slug mixing zone for high viscous liquid and the thicker liquid film on the slug body. Al-Safran et al. (2015) model was proposed as:

$$E_S = 0.85 - 0.075\varphi + 0.057\sqrt{\varphi^2 + 2.27}\,; \qquad \varphi = {FrN_\mu}^{0.2} - 0.89 \tag{16}$$

The proposed model out performed 10 other models when tested against high viscosity databank used in the model development.

A summary of models and correlations for the estimation of slug liquid holdup models and correlations reviewed in this paper is shown in Table 1: Summary of Slug Liquid Holdup Models and Correlations below. From the table, most of the existing models and correlations were developed and/or validated on experimental data with liquid viscosity less than 0.6 Pa.s. In this paper, an experimental investigation of slug liquid holdup for liquid viscosity up to 8.0 Pa.s is presented. Additionally, comparative analysis of existing correlations is carried out and a modified slug holdup correlation is proposed.

Table 1: Summary of Slug Liquid Holdup Models and Correlations

Author	Pipe ID (mm)	Test Fluid		Liquid Viscosity (Pa.s)	Correlation/Model
		Liquid	Gas		
Gregory (1977)	25.8, 51.2	Light Oil	Air		$H_{ls} = \frac{1}{1 + \left(\frac{V_M}{8.66}\right)^{1.39}}$
Malnes (1983)	25.8, 51.2	Light Oil	Air		$H_{ls} = 1 - \frac{V_M}{\left[V_M + 83\left(\frac{g\sigma_{GL}}{\rho_L}\right)^{1/4}\right]}$
Ferschneider (1983)	146	Condensate	Natural Gas		$H_{ls} = \frac{1}{\left\{1 + \left[\frac{V_m}{\sqrt{\left(1 - \frac{\rho_G}{\rho_L}\right) gD}}\right]^2 \Big/ \left(\frac{A}{Bo^{\beta}}\right)^2\right\}^2}$ $Bo = \frac{(\rho_L - \rho_G) g D^2}{\sigma} \qquad 0 < \theta \leq 90°$
Marcano et al. (1998)	78	Light Oil	Air		$H_{ls} = 1 - \frac{1}{\left[1.001 + 0.0179 V_M \left(0.0179 {V_M}^2\right)^{1/4}\right]}$
Abdul-Majeed (1999)		Kerosene, Light Oil, Water	Air, Nitrogen, Freon	0.001 – 0.007	$H_{ls} = (1.009 - C V_M)\, A$
Gomez et al. (2000)	51 – 203	Light Oil, Water	Air, Freon	0.001 – 0.065	$E_s = e^{-(0.45\theta + C\, Re)} \qquad 0 < \theta \leq 90°$ $Re = \frac{\rho_L V_M D}{\mu_L}; C = 2.48 \times 10^{-6}$

Zhang et al. (2003)	$E_s = \frac{1}{1+\frac{T_{sm}}{3.16[(\rho_L-\rho_G)g\sigma_L]}}$; $T_{sm} = \frac{1}{C_e}\left[\frac{f_s}{2}\rho_S {V_M}^2 + \frac{D}{4}\frac{\rho_L E_f(V_T-V_f)(V_M-V_f)}{L_S}\right]$; $C_e = \frac{2.5-\lvert\sin\theta\rvert}{2}$				
Al-Safran (2009)					$H_{LS} = 1.05 - \frac{0.0417}{\Theta - 0.123}$ where $\Theta = \frac{H_f(V_T-V_f)(V_M-V_f)}{{V_M}^2}$
Kora *et al.* (2011)	50.4	Heavy Oil	Air	0.181 – 0.587	$E_S = \begin{cases} 1.012\exp(-0.085 Fr {N_\mu}^{0.2}) & 0.15 < Fr {N_\mu}^{0.2} < 1.5 \\ 0.9473\exp(-0.041 Fr {N_\mu}^{0.2}) & Fr {N_\mu}^{0.2} \geq 1.5 \\ 1.0 & Fr {N_\mu}^{0.2} \leq 0.15 \end{cases}$ $N_\mu = \frac{V_m \mu_l}{gD^2(\rho_l-\rho_g)}$; $Fr = \frac{V_m}{(gD)^{0.5}}\sqrt{\frac{\rho_l}{\rho_l-\rho_g}}$
Xu (2013)	25.8 - 60	Carboxymethyl cellulose (CMC)	Air	0.089 – 0.0972	For Newtonian liquid-gas flow $\alpha_S = \frac{1}{1+\left(\frac{V_M}{9.514}\right)^{1.274}}$; $0.1\ m/s \leq V_M \leq 20\,m/s$ For non-Newtonian liquid-gas flow $\alpha_S = \frac{(1-\sin\theta)^{0.05}}{1+3.166\times10^{-5}{Re_L}^{1.225}}$; $0° \leq \theta \leq 75°$ $Re_L = \frac{\rho_l D V_M}{\mu_{eff}}$; $\mu_{eff} = D^{1-n}8^{n-1}k.V^{n-1}$
Al-Safran *et al.* (2015)	50.4	Heavy Oil	Air	0.181 – 0.587	$E_S = 0.85 - 0.075\varphi + 0.057\sqrt{\varphi^2+2.27}$; $\varphi = Fr {N_\mu}^{0.2} - 0.89$

2 Experimental Setup

2.1 Test Facilities

Schematics of the two high viscosity multiphase flow facilities used in this study are shown in Figure 1 and

Figure 2. The first facility consists of 25.4-mm ID pipe with 5.5-m long test section fabricated from Perspex pipe work. The second facility is a scale up of the first facility; it consists of 76.4-mm pipe ID and 17-m long test sections also fabricated from Perspex pipe work. A flexible steel hose of the same pipe ID are used to transfer oil from the test section to the gravity driven four-phase separators in both facilities. After separation, the oil is returned to the storage tank. A detailed description of the 76.4-mm facility is given in Baba et al. (2017). The 25.4-mm ID pipe facility is described below.

An Atlas Copco® compressor with maximum discharge pressure 8-bar and maximum capacity 400-m^3/hr and an Ingersoll Rand compressor® with maximum discharge pressure of 11-bar and flow rate of 1250 m^3/hr, receives free air and compresses it before supplying to the flow loop. To prevent pulsating supply of air to the test facility, the air from the compressor is discharged into a 2.5-m^3 air tank before delivery to the test line where it is regulated. To ensure that supplied air is moisture and debris free for easy and accurate metering, dryer and filters are installed in the compressor supply lines.

Two gas flowmeters, a (Prowirl 72F15 DN15) 0.5-inch thermal mass flowmeter ranging from 0 – 20 m^3/hr and a (Prowirl 72F40 DN40) 1-inch vortex flowmeter ranging from 3 – 300 m^3/hr both manufactured by Endress+Hauser is used for air metering. Air is introduced to the main test line from a flexible steel pipe with 25.4-mm pipe ID. The flexible pipe connects to the test section through a tee-junction. Oil is introduced by a PVC pipe with 25.4-mm pipe ID, connected in series with the main test section. Both fluids mix at the tee-junction. Observation/measurement points are located about 80 pipe diameters from the last injection point in the test facility. Experimental observations showed that

this length was adequate in establishing a fully developed flow within the limits of the flow variables studied.

Oil is stored in a 0.15-m^3 tank capacity manufactured from plastic material and insulated with fibres on the periphery. A variable speed PCP with maximum capacity, 0.72-m^3/hr is used in pumping oil, Endress+Hauser's Promass 831 DN 50, a Coriolis flowmeter, with range, 0 – 180-m^3/hr is used in oil metering.

The separator is s rectangular shaped tank with viewing windows to allow for liquid levels and separation process monitoring for multiphase fluids separation. For liquid-gas flows, the residence time for complete separation of oil and air escapes to the atmosphere while oil is returned to the storage tank after a residence time of about 3 – 8 hours. The separation process is gravity driven.

2.1.1 Instrumentation and Data Acquisition System

Two GE Druck static pressure transducers, PMP 1400, with pressure range 0 – 4 barg and accuracy 0.04% over the full scale is used to obtain the static pressure in the test section, they are placed 2.17 m apart. A differential pressure transducer, Honeywell STD120, with minimum pressure drop measurement of 100 Pa and an accuracy of ±0.05% is used to measure the differential pressure. Temperature of the test fluids on the test section is measured by means of J-type thermocouples with an accuracy of ±0.1^{o}C placed at different locations.

The temperature regulator is a bath thermal circulator produced by Thermal Fisher. Copper coils submerged in the oil tank are connected to the circulator. Depending on the viscosity of oil required, either hot or cold glycol is pumped through the coils for a specific time interval until desired temperature, and thus viscosity is achieved. The circulator's temperature ranges from –5 to +50 $^{\circ}$C.

Temperature of the test fluids on the test section is measured by means of J-type thermocouples with accuracy of ±0.1°C placed at different locations.

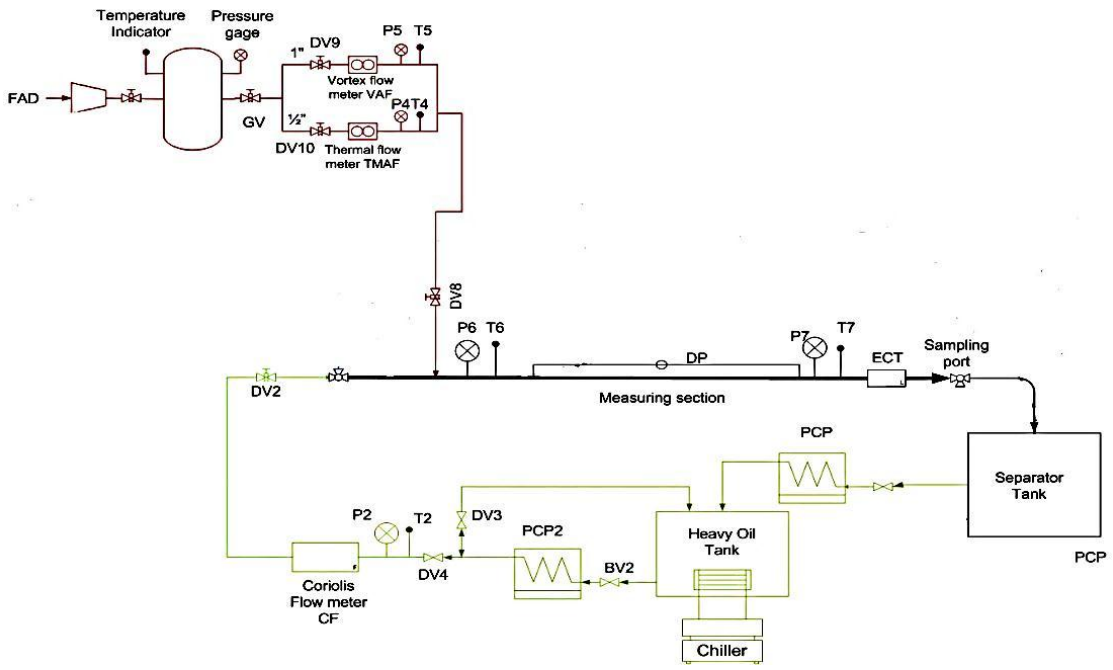


Figure 1: Schematic of the 0.0254 pipe internal diameter (ID) multiphase test facility

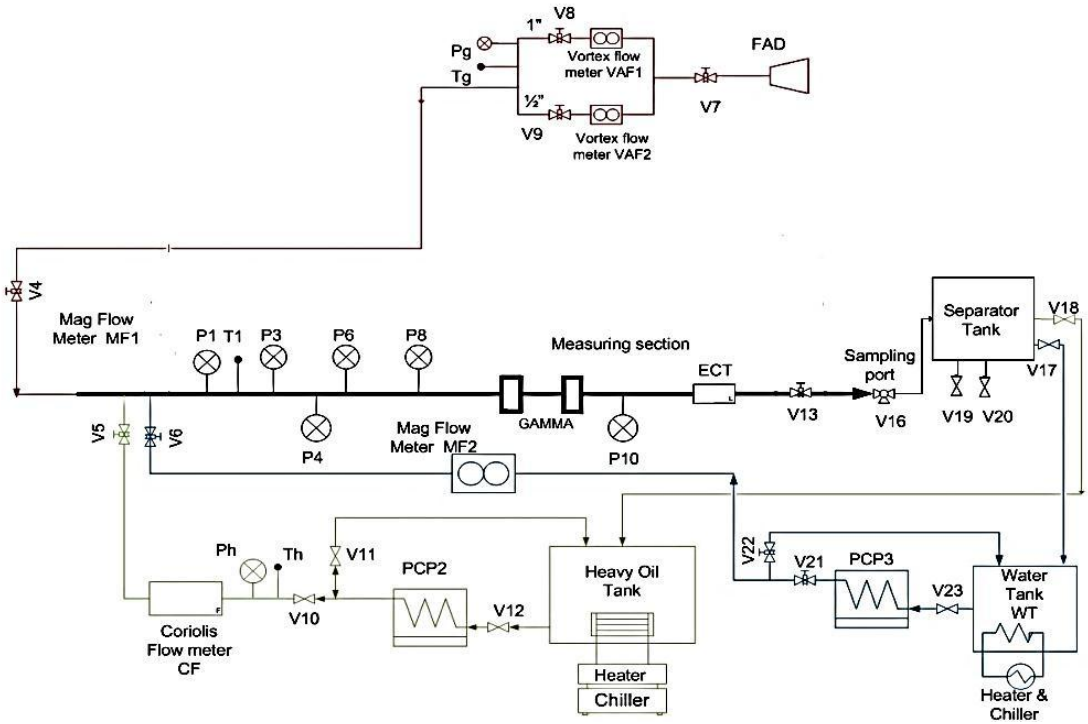


Figure 2: Schematic of the 76.4 pipe internal diameter (ID) multiphase test facility. Data acquired from the flowmeters, differential pressure transducers, pressure transducers and temperature sensors are saved to a Desktop Computer using a LabVIEW-based system (version 8.6.1). The system consists of a National Instruments (NI) USB-6210 connector board interfaces that output signals from the instrumentation using BNC coaxial cables and the desktop computer. Three

Sony camcorders, DSCH9 with 16 megapixels, high definition and 60GB HDD are used for video recordings during the test to aid visual observations.

2.2 Fluids Physical Properties and Experimental Test Matrix

Mineral oil manufactured by Total®, is used as the oil phase while compressed air is used as the gas phase in the experimental campaign.

Table **2** and Table **3** shows the fluid physical properties of the fluids used and the test matrix covered in this study. Mineral oil manufactured by Total®, is used as the oil phase while compressed air is used as the gas phase in the experimental campaign.

Table 2: Physical Properties of Fluids

Test fluids	Density (25°C, kg/m^3)	Viscosity (Pa.s)	Interfacial (liquid/air) tension (25°C, N/m)
CYL 680	≈ 918	0.90 – 5.00	
Air	1.293	0.000017	0.033

Table 3: Experimental Test Matrix

Pipe Internal Diameter (m)	Liquid Superficial Velocity (m/s)	Gas Superficial Velocity (m/s)	Liquid Viscosity (cP)
0.0754	0.05 – 0. 31	0.10 – 4.00	0.001 – 5.000
0.0254	0.03 – 0.21	0.10 – 7.00	3.300 – 8.090

3 Results and Discussion

3.1 Slug Formation Mechanism and Slug Liquid Holdup

Slug flow forms as a result of the growing sinusoidal interfacial waves between stratified liquid-gas flows in a flowline. At a favourable liquid height in the flowline, if the gas velocity is increased, the gas phase momentum increases accordingly. The increase in gas phase momentum results in a pressure decrease at the interface. Pressure decrease results in suction forces (Bernoulli Effect) acting on the liquid phase. These forces overcome the gravity and

surface tension forces on the liquid layer. This mechanism is described as the Kelvin–Helmholtz (K–H) stability criterion for transition to slug flow. At the peak of the interfacial instability, the interfacial wave grows until it becomes energetic enough to lift liquid from its layer to the top of the flowline and bridges the flowline.

Figure 1 show a typical slug flow pattern observed in this study. It can be seen that at the slug front, gas bubbles are entrained in the slug body. The slug front was observed to be highly energetic and travelled at a translational velocity higher that the mixture velocity. Due to this high velocity, the slug body accelerates the slug film resulting in gas entrainment in the slug body. Additionally, gas entrained in the liquid film on the pipe wall also contributes to gas entrainment in the slug body. Entrained gas bubbles were more in the slug front than other sections of the slug body. Al-Safran (2015) posited that entrained gas bubbles' loss occurs when the generated bubbles re-circulate from the slug front into the leading slug film tail due to their circular motion.

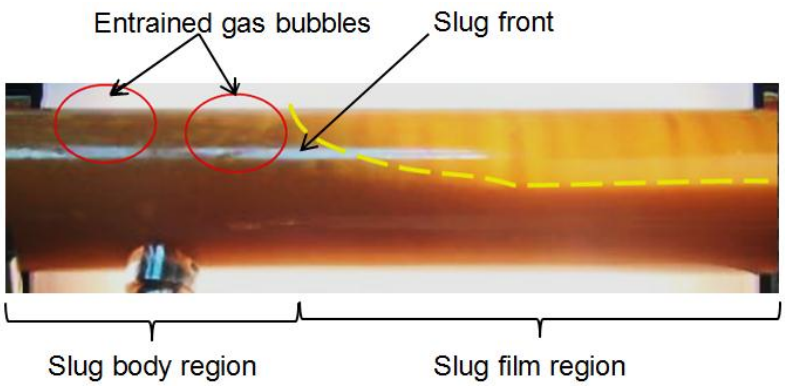


Figure 1: Slug flow pattern for high viscosity oil-gas two-phase flow obtained from video recording in the 72.4-mm pipe ID facility

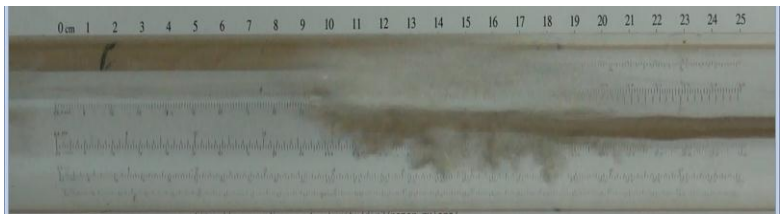


Figure 2: Slug flow pattern for high water-gas two-phase flow obtained from video recording in the 72.4-mm pipe ID facility

Figure 2 shows the slug flow pattern obtained for water-gas two-phase flow. From experimental observations, the characteristics of slug flow in the water-gas experiment are different from the oil-gas experiment. Relative to the high viscosity oil-gas flow, slug front in the water-gas experiment was more turbulent, energetic and consisted of fewer gas bubble entrainments. This implies that the slug liquid body will have fewer gas entrainments. Notably, the slug length is longer in comparison to the heavy oil-gas flow.

In this work, electrical capacitance tomography (ECT) manufactured by Industrial Tomography Systems (ITS) was used in slug holdup measurements. The sensor consists of 12 electrodes with 66 dielectric permittivity measurements. The dielectric permittivity is converted to capacitance using the Maxwell's equation. Subsequently, the linear back projection algorithm is used to process the capacitance measurements into tomograms and flow parameters such as liquid holdup, slug liquid holdup, slug frequency etc. Detailed description of the sensor is reported in Zhao et al. (2013), Archibong-Eso et al. (2014) and Baba et al. (2017).

3.2 Measurement of Slug Liquid Holdup

In Figure 3, crests in the waveforms of the time series plots are indicative of the slug (elongated liquid) body passage while troughs are indicative of the stratified film region. In determining the slug holdup, a liquid holdup threshold is defined; this also helps eliminate the classification of travelling waves as slug holdup. Several researchers have defined different threshold values with the most common ranges being between 0.7 – 0.75. Nydal (1991), Perez (2007). Recently Zhao (2013) adopted a mean value relative to the variable liquid holdups at different flow conditions; this approach is adopted in this study. The Zhao (2013) technique is particularly useful since the mean liquid holdup in the film region of the slug flow in high viscosity liquid two-phase flow may sometimes be higher than 0.7. The liquid holdup threshold h_{th} is defined as:

$$h_{th} = \frac{1}{2}[\max(h_a) + \min(h_a)] \tag{17}$$

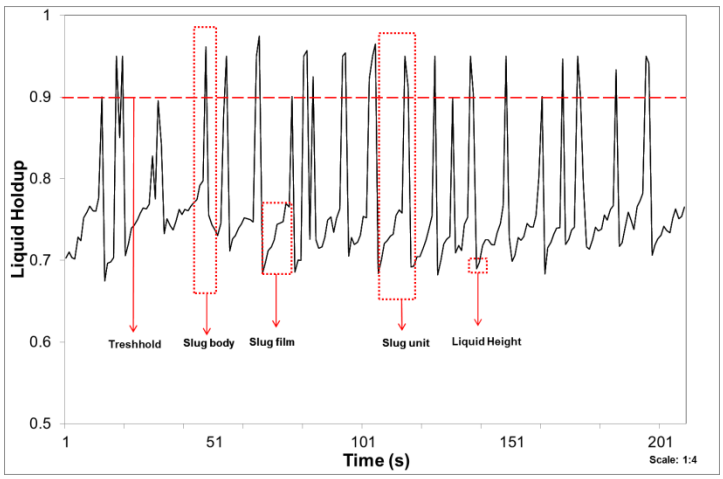
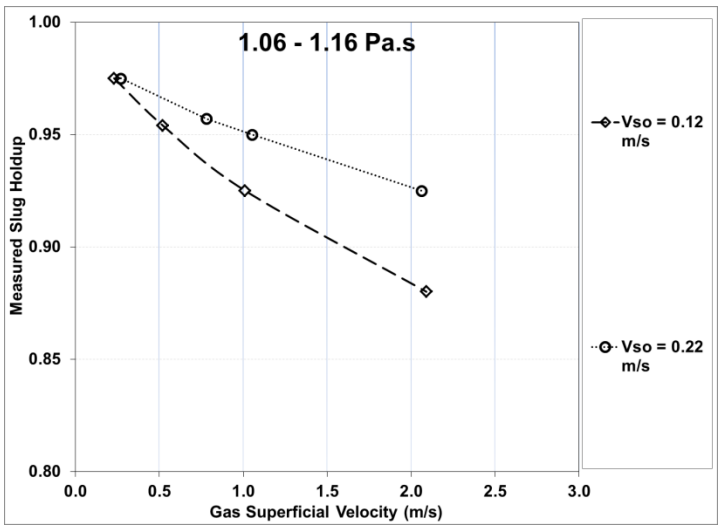


Figure 3: A typical Instantaneous liquid holdup of a time varying slug flow obtained in the study (Archibong-Eso et al. 2018)

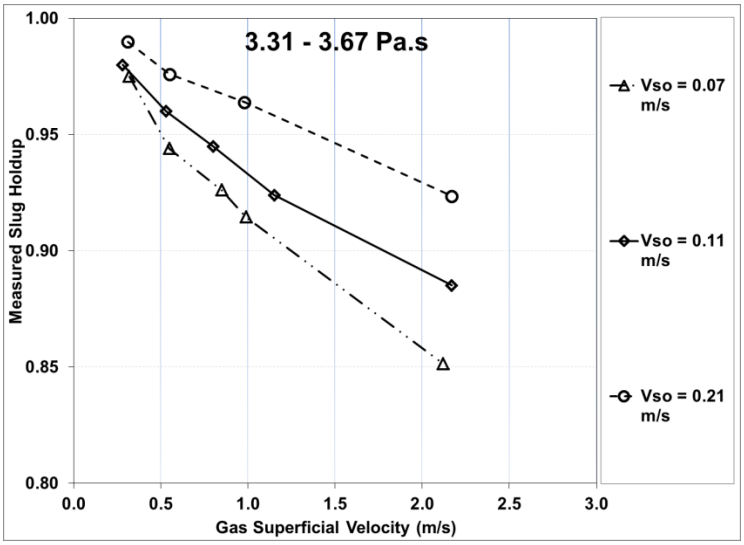
Slug holdup is computed as a mean of all the time varying instantaneous liquid holdup waveforms that exceeds this threshold.

3.3 Liquid and Gas Superficial Velocities on Effects on Slug Body Liquid Holdup

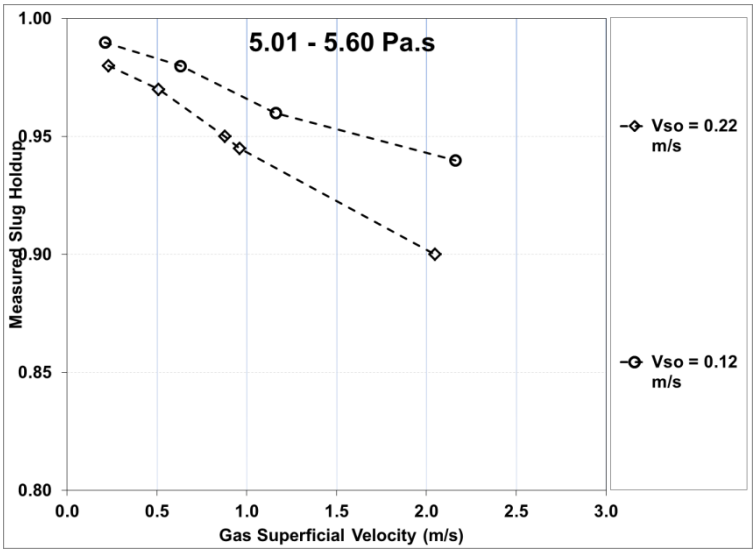
Figure 4 shows the mean slug holdup measured by ECT as a function of the gas superficial velocity. Within the experimental test matrix, slug holdup generally decreased with increase in superficial gas velocity. Relating this to observed flow patterns, it is noted that plug flow was observed close to this range of flow conditions, this lays further credence to the Barnea (1992) definition which assumed negligible gas entrainment in differentiating slug from plug flow. At relatively lower gas superficial velocities (≤ 1.0 m/s); it was observed that this decrease was not as prominent as that for higher gas superficial velocities (> 1.0 m/s). This observation may be as a result of the increased effect of liquid viscosity as higher gas superficial velocities. Slight increase in the slug holdup was also observed on increasing the superficial oil velocity in the experiments. This is expected as generally, an increase in superficial liquid velocity increases the input liquid content in the pipe thus aiding in decreasing the turbulent kinetic energy. Increased liquid superficial velocity also increases the pipe wetting on the pipe walls and by extension, the slug liquid holdup.



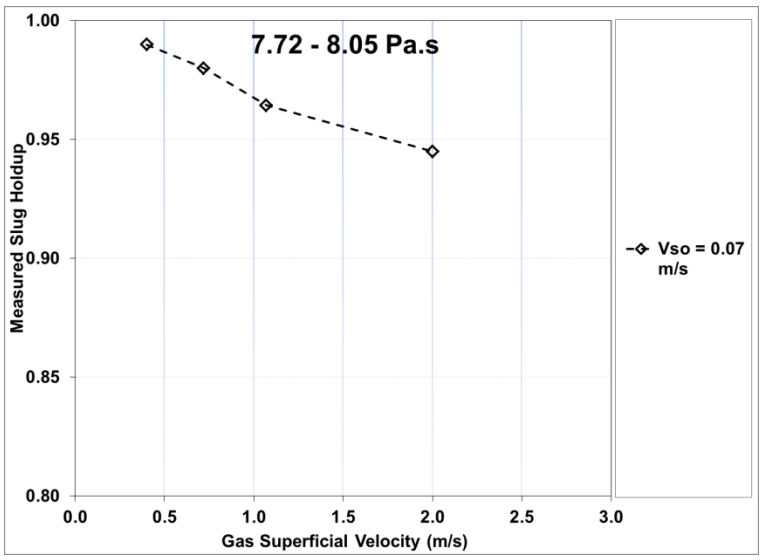
a.) Oil viscosity, 1.06 – 1.16 Pa.s



b.) Oil nominal viscosity, 3.31 – 3.67 Pa.s



c.) Oil nominal viscosity, 5.01 – 5.60 Pa.s



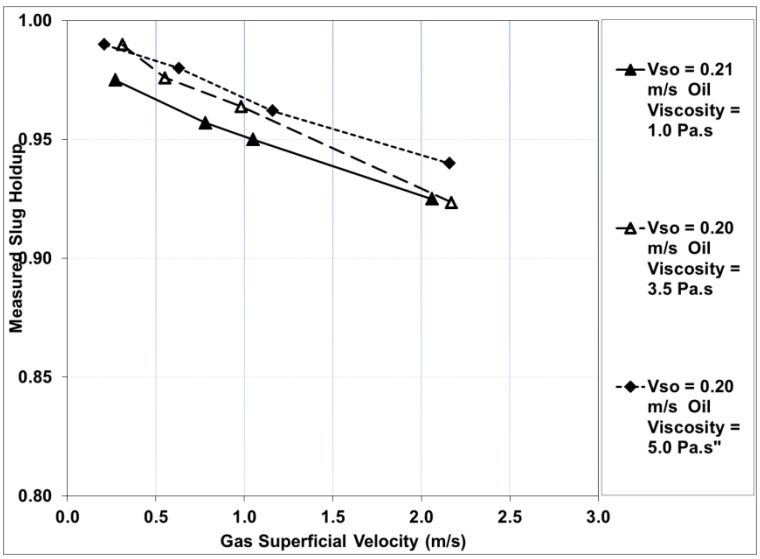
d.) Oil nominal viscosity, 7.72 – 8.05 Pa.s

Figure 4: ECT Measured Slug Holdup in the 1 inch Horizontal Facility

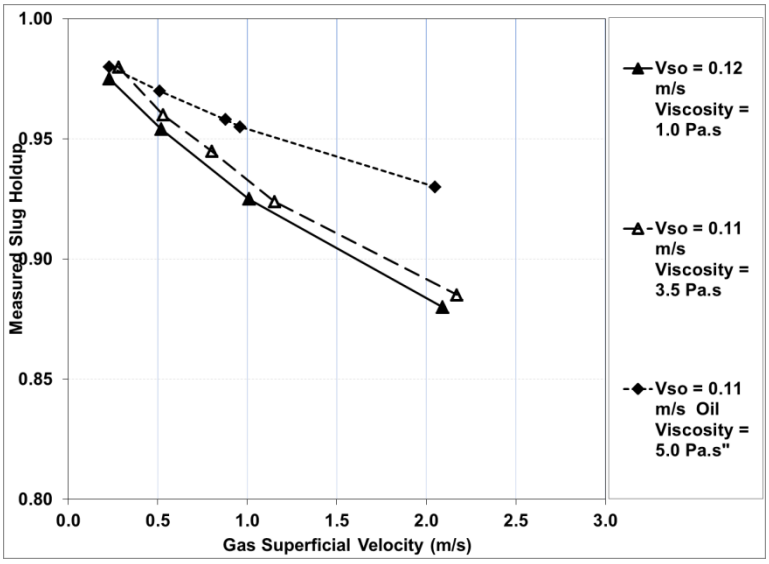
3.4 Liquid Viscosity Effects on Slug Holdup

Effect of liquid viscosity is shown in Figure 5 below by utilizing dataset of 6.9 - 8.09 Pa.s and 0.97 – 1.21 Pa.s viscosities of oil from Zhao (2014). Generally, it is observed that the slug holdup increase with increase in oil viscosities, an increase in viscosity increases the viscous shear of the oil and hence, a reduction in the gas' ability to be entrained in the oil. Increased viscosity will also help in increasing the liquid wetting on the pipe walls thus contributing to increased slug liquid holdup. Additionally, liquid viscosity reduces turbulent kinetic energy in the slug mixing region resulting in low circulation and vortex intensity and low entrainment rate at the slug front (Al-Safran, 2013). This result is consistent with results obtained by Kora *et al.* (2011).

Figures 5a and 5b show that at higher superficial velocity, effects of liquid viscosity on the slug holdup becomes less dominant. This may be explained by increased liquid content in the flow channel (as a result of increased liquid superficial velocity), which plays a larger role in the slug body liquid content compared to the increase liquid wetting on the pipe wall (as a consequence of the increased liquid viscosity).



a. Viscosity comparison in the 1 inch test facility for oil superficial velocity of 0.20 – 0.21 m/s and nominal oil viscosity of 1.0 – 5.0 Pa.s



b. Viscosity comparison in the 1 inch test facility for oil superficial velocity of 0.11 – 0.12 m/s and nominal Oil viscosity of 1.0 – 5.0 Pa.s

Figure 5: Effect of liquid viscosity on slug holdup in 1-inch horizontal pipe

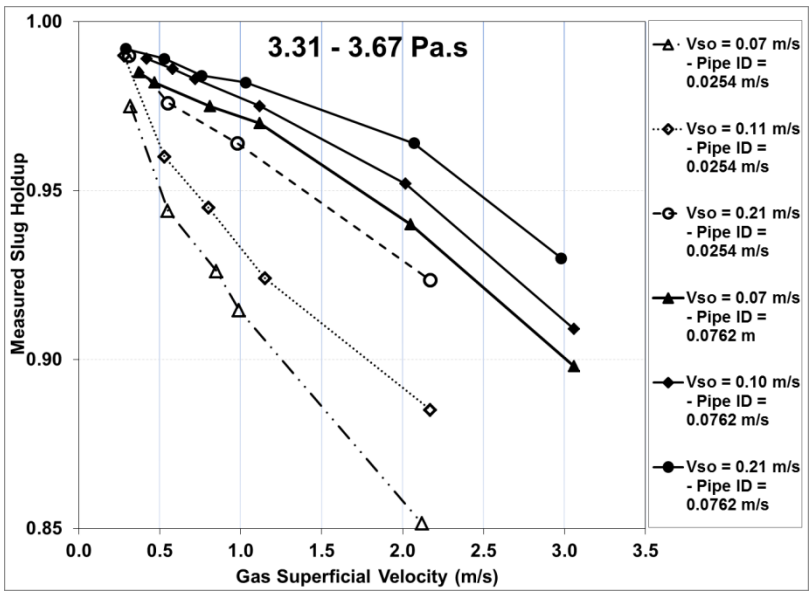


Figure 6: Effect of pipe diameter on slug holdup

3.5 Pipe Diameter Effects on Slug Frequency

Pipe diameter effect on slug body liquid holdup is shown in Figure 6 above by comparing data obtained in the 0.0254 and 0.0762 m pipe internal diameters and oil viscosity ranging from 3.31 – 3.67 Pa.s. For similar oil viscosity, it is observed that the slug liquid holdup increased with increase in pipe diameter. This is as a result of the slug aeration mechanism in high viscosity oil-gas flow. Al-Safran et al. (2013) noted that gas entrainment in the slug body is govern by the slug aeration process which involves the characteristics of bubble generation at the slug front, fragmentation in the developing mixing region, transportation to the developed slug body region and shedding at the slug back into the trailing elongated bubble. The authors related the entrained gas size to the balance between the rate of turbulent kinetic energy and gas/liquid surface tension. They stated that gas entrainments occurred when the turbulent kinetic energy is greater than the surface tension. When pipe internal diameter and thus the flow area, reduces, surface tension increases, while the turbulent kinetic energy (estimated from the mean of the turbulence normal stresses) reduces. Conversely, this result in a reduction in the gas entrained in the slug liquid holdup, for the 0.0254 compared to the 0.0762 m pipe ID (for similar oil viscosity), and thus the slug liquid holdup.

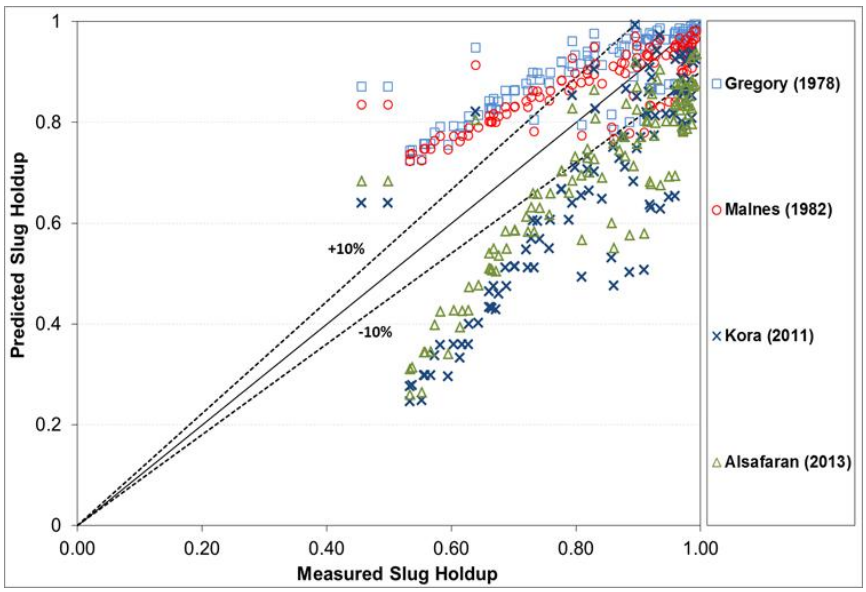


Figure 7: Predicted slug holdup as a function of the measured slug holdup

3.6 Evaluation of Slug Holdup Predictive Models

Measured slug holdups obtained in this study were compared with predictive models in literature. The Gregory *et al*. (1978), Malnes (1982), Kora *et al.* (2011) and Al-Safran *et al.* (2013) predictive models' performances were evaluated. Figure 7 shows a plot of the predicted slug holdup as function of the ECT measured slug holdup. It is observed that the models generally perform well at the highest liquid holdup but predicts poorly at relatively low slug holdup. Malnes *et al*. (1982) and Gregory *et al.* (1978) over predicted the slug holdup, both models performed similarly for the entire pipe diameters evaluated in the databank, in part; this is as result of the authors' not accounting for the effect of diameter in their models. The fact that the models were also developed for very low liquid viscosity is also a factor in its performance. The Al-Safran *et al.* (2013) and Kora *et al.* (2011) models performed relatively better when compared to the previous the aforementioned models, however, they both under predicted the slug body liquid holdup for oil viscosities at 3.0 Pa.s and higher.

3.7 Development of a New Slug Holdup Correlation

Existing models have shown promising performance for applications in high viscosity applications within predefined limits; however, an improvement is necessary to extend these limits since they not only constitute a key closure to the slug mechanistic models but are also essential in the design of some unit operations equipment.

Table 4: High viscosity dataset used in correlation development

Dataset	Viscosity (Pa.s)	Pipe ID (m)	No. of Data
This Work	1.000 – 8.090	0.0254 – 0.0764	90
Al-Safran et al. (2009)	0.187 – 0.287	0.0580	58
Kora et al. (2011)	0.387 – 0.587	0.0580	71

Present predictive correlation utilizes a dataset with viscosity of 0.187 – 8.0 Pa.s from experiments and published literature to improve predictions, pipe internal diameter ranged from 0.0254 – 0.0762 m.

From experimental observations of the hydrodynamic behaviour of slug flow and several published works, the following functional parameters were deduced to strongly correlate slug holdup, H_{ls}:

$$H_{ls} = F(V_{so}, V_{sg}, V_m, g, D, \rho_l, \rho_g) \tag{18}$$

Dimensional analysis using the Buckingham Pi-theorem and subsequent non-dimensional groupings yielded the following dimensionless groups:

$$H_{ls} = F(N_{Fr}, Re_m, N_\mu, \frac{\rho_g}{\rho_l}) \tag{19}$$

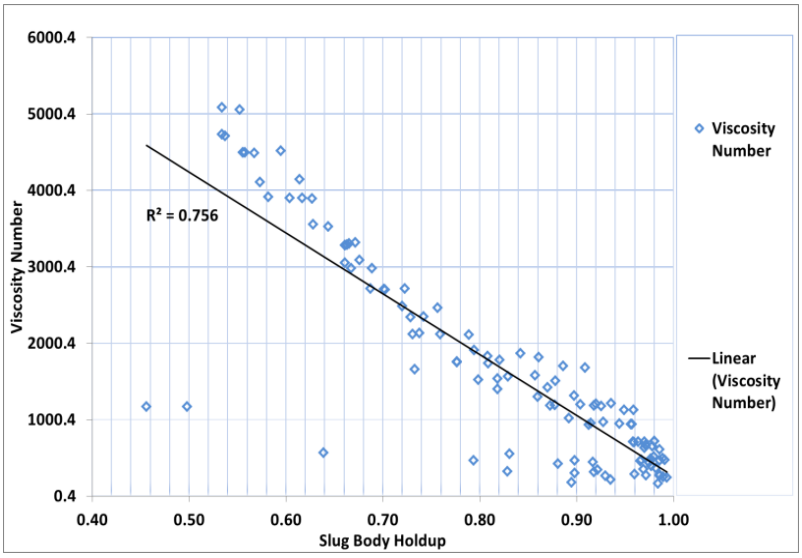
where N_{Fr}, Re_m, N_μ and ρ_g/ρ_l are the Froude number, mixture Reynolds number, viscosity number and the ratio of gas to oil densities. Partial correlation of each of the groupings was done through regression of each dimensionless group against the measured slug holdup. It was noted from preliminary analysis that N_{Fr} and N_μ showed the best and most significant correlation with the slug holdup as shown in Figure 8.

These groupings are similar to those reported and utilized by Kora *et al.* (2011) and Al-Safran *et al.* (2013). In both studies, the dimensionless numbers were defined based on Wallis (1969), this was necessary to ensure the influence of inertia and viscous force on liquid holdup is accounted for.

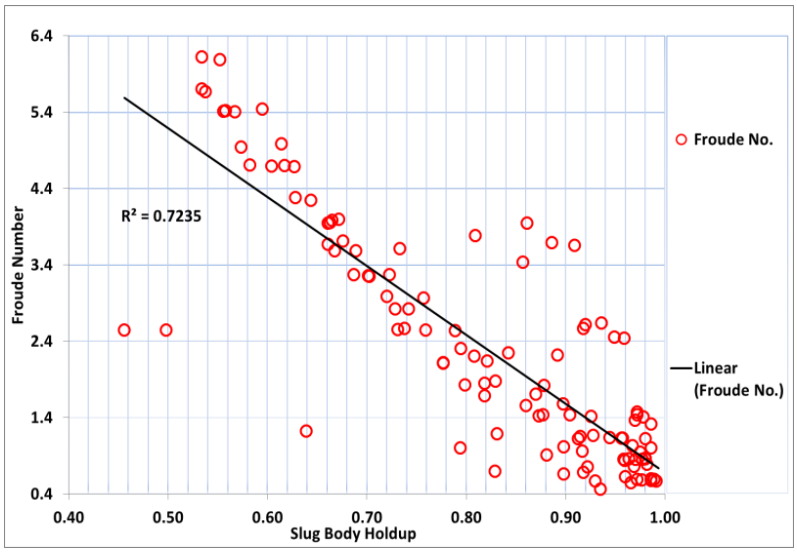
Analysis showed a strong relationship with mixture superficial velocity than both the oil and gas superficial velocities hence the decision to replace the liquid superficial velocity with the mixture velocity. The groupings are defined thus:

$$N_{Fr} = \frac{V_m}{(gD)^{0.5}} \sqrt{\frac{\rho_l}{\rho_l - \rho_g}} \tag{20}$$

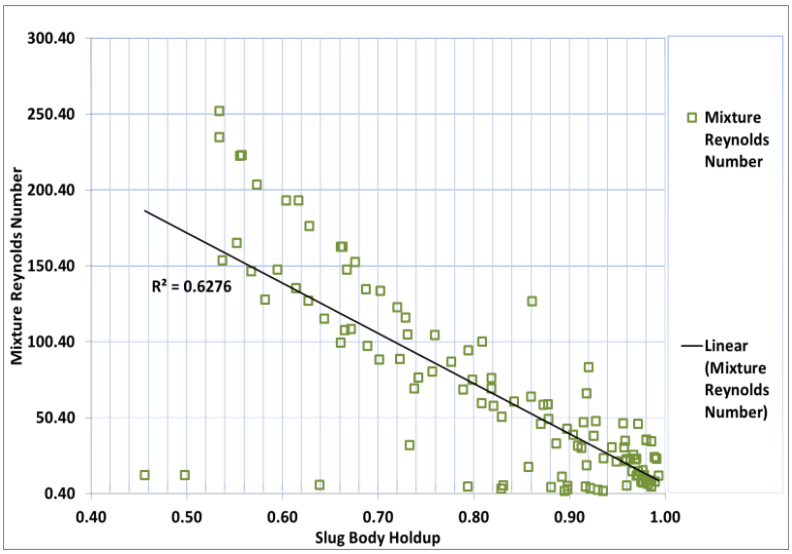
$$N_{\mu} = \frac{V_m \mu_l}{gD^2(\rho_l - \rho_g)} \tag{21}$$



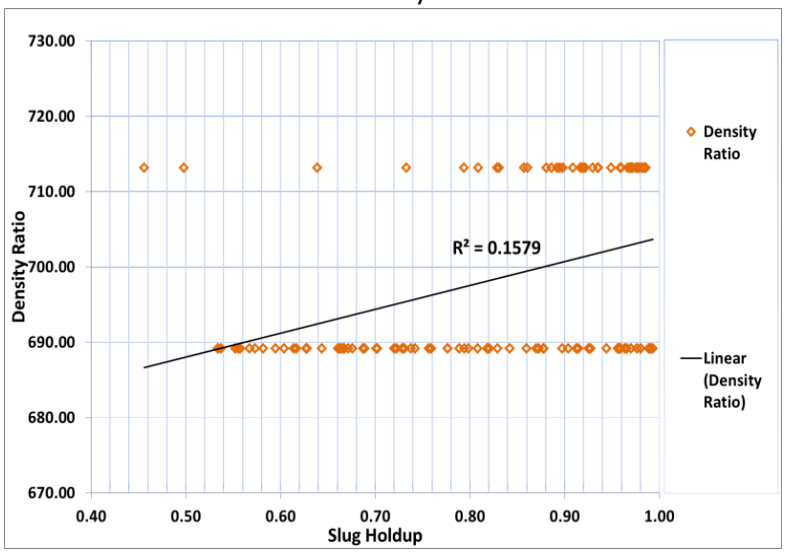
a. Viscosity Number



b. Froude Number



c. Mixture Reynolds Number



d. Density Ratio

Figure 8: Partial correlation of the dimensionless groupings

After correlation of the experimental dataset with those obtained from literature, a general non-linear relationship for the slug holdup in high viscosity oil-gas two-phase flow was proposed thus:

$$H_{ls} = 1 - 0.0336\, N_{Fr} N_{\mu}^{\;0.11} \qquad (22)$$

Table 5: Statistical Performance evaluation of proposed correlation with other correlations in literature on the high viscosity data bank

Correlation	E_1	E_2	E_3
Gregory (1978)	0.13	0.14	0.09
Malnes (1982)	-0.05	0.10	0.10
Kora *et al.* (2011)	0.14	0.16	0.11
Al-Safran *et al.* (2013)	0.13	0.15	0.09
Proposed	0.00	0.05	0.07

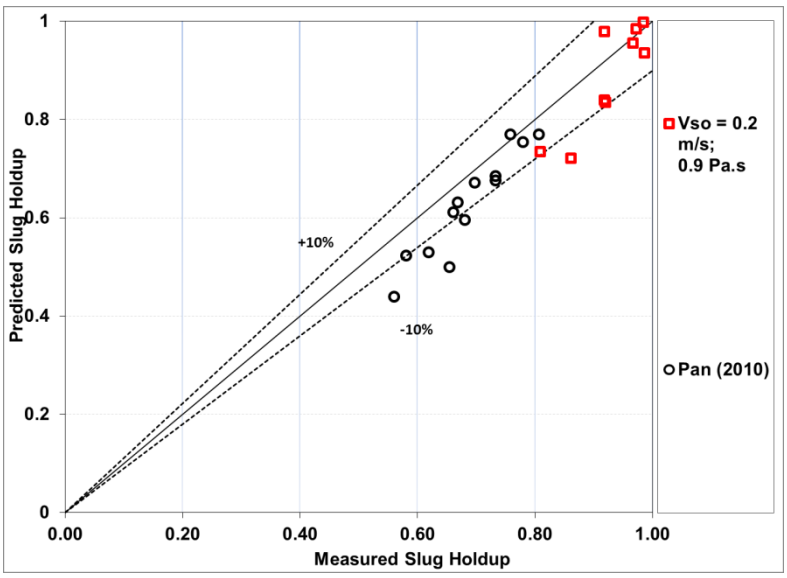


Figure 9: Validation of proposed slug frequency correlation on data obtained from 0.0762 m pipe ID facility, with oil viscosity, oil superficial velocity used are 900 cP and 0.21 m/s respectively and Pan (2010) data.

Statistical evaluation of the proposed model with other models in literature is shown in Figure 9 and Table 5. The proposed correlation out performed existing correlations with an average percent error, E_1; the absolute average percent error, E_2 and standard deviation, E_3 of 0.001, 0.05 and 0.07 respectively. In validating the model, an independent data not used in the model development was obtained from the 0.0762 m pipe ID facility. The viscosity and superficial velocity of oil were 900 cP and 0.21 m/s respectively. Results obtained from Pan (2010) at oil viscosity and superficial velocity of 0.004 pa.s and 0.4 – 0.7 m/s were also used in validating the model. Both dataset showed that the proposed correlation performed very well.

Relatively, the discrepancies in prediction is much higher in the Pan (2010) model, this is due to the oil viscosity used in that study. A need for careful consideration on the usage of the proposed correlation is advocated especially at fluid properties outside the experimental test matrix.

4 Conclusion

Slug liquid holdup (a key slug flow parameter) in horizontal high viscosity gas–liquid two-phase flow was measured using and electrical capacitance tomography (ECT) sensor. Experiments show that slug liquid holdup increased and decreased with increase in superficial liquid and gas velocities respectively. An increase in liquid viscosity had the effect of slightly increasing the slug liquid holdup. Compared to air–water experiments, the slug nose of the air–oil slug flow was less turbulent and indeed energetic due the oil viscosity inhibiting turbulence. We visualised both using high-speed imaging. Results obtained in this study were compared with correlations found in literature for slug liquid holdup estimation. It is noted that existing models did not accurately predict the slug liquid holdup for oil viscosity up to 8.0 Pa.s. This was a consequence of some of these correlations not accounting for important fluid properties like viscosity as well as the limitation of the databank used in their development. As a result, a new slug liquid holdup was proposed. The correlation covers liquid viscosity ranging from 0.189 – 8.0 Pa.s. Statistical analysis of the new correlation vis-à-vis existing ones showed that the present correlation outperformed the best performing ones surveyed in the open literature.

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References

Abdul-Majeed, G.H., (2000). Liquid slug holdup in horizontal and slightly inclined two-phase slug flow. J. Petrol. Sci. Eng., 27, 27-32

A. Archibong-Eso, Y. Baba, A. Aliyu, Y. Zhao, W. Yan, H. Yeung, 2018, On slug frequency in concurrent high viscosity liquid and gas flow, Journal of Petroleum Science and Engineering. Available at: doi.org/10.1016/j.petrol.2017.12.071.

Al-Safran, E.M., (2009). Prediction of Slug Liquid Holdup in Horizontal Pipes. *Journal of Energy Resources Technology*, 131(2), p.23001. Available at: http://energyresources.asmedigitalcollection.asme.org/article.aspx?articleid=1414905.

Al-Safran, E.M., Kora, C., Sarica, C. (2013). Prediction of Liquid Volume Fraction in Slugs in Two-Phase Horizontal Pipe Flow with High Viscosity Liquid. *International Conference on Multiphase Production Technology, Cannes, France, 2013. pp 415 - 428.*

Al-Safran, E.M., Kora, C. & Sarica, C., 2015. Prediction of slug liquid holdup in high viscosity liquid and gas two-phase flow in horizontal pipes. *Journal of Petroleum Science and Engineering*, 133, pp.566–575. Available at: http://linkinghub.elsevier.com/retrieve/pii/S0920410515300450.

Dukler, A.E. & Hubbard, M.G., 1975. Model for Gas-Liquid Slug Flow in Horizontal and Near Horizontal Tubes. *Industrial & Engineering Chemistry Fundamentals*, 14(4), pp.337–347.

Farsetti, S., Farisè, S. & Poesio, P., 2014. Experimental investigation of high viscosity oil-air intermittent flow. *Experimental Thermal and Fluid Science*, 57, pp.285–292. Available at: http://dx.doi.org/10.1016/j.expthermflusci.2013.12.004.

Gomez, L.E., Shoham, O. & Taitel, Y., 2000. Prediction of slug liquid holdup:

Horizontal to upward vertical flow. *International Journal of Multiphase Flow*, 26(3), pp.517–521.

Gregory, G.A., Nicholson, M.K. & Azlz, K., 1978. Correlation of the Liquid Volume Fraction in the Slug for Horizontal Gas-Liquid Slug Flow. *International Journal of Multiphase Flow*, 4(I), pp.33–39.

Kora, C., Sarica, C., Zhang, H-Q., Fahd, K., Al-Safran, E.M. (2011). Effects of High Oil Viscosity on Slug Liquid Holdup in Horizontal Pipes. *SPE The Canadian Unconvetional Resources Conference, 2011*, (November), pp.1–15.

Malnes, D., 1983. *Slug Flow in Vertical, Horizontal and Inclined Pipes. Report IFE/KR/E-83/ 002*, Kjeller, Norway.

Nädler, M. & Mewes, D., 1995. Effects of the liquid viscosity on the phase distributions in horizontal gas-liquid slug flow. *International Journal of Multiphase Flow*, 21(2), pp.253–266.

Nuland, S. et al., 1997. Gas Fractions in Slugs in Dense-Gas Two Phase Flow from Horizontal to 60 degrees of Inclination. In *ASME Fluids Engineering Divison Summer Meeting*.

Pan, J., (2010). Gas Entrainment in Two-Phase Gas-Liquid Slug Flow. PhD Thesis, Imperial College, London, UK

Taitel, Y. & Barnea, D., 1990. A consistent approach for calculating pressure drop in inclined slug flow. *Chemical Engineering Science*, 45(5), pp.1199–1206.

Wallis, G.B., 1969. *One Dimensional Two-Phase Flow*, New York: McGraw-Hill.

Zhang, H. et al., 2003. A unified mechanistic model for slug liquid holdup and transition between slug and dispersed bubble flows. *International Journal of Multiphase Flow*, 29(1), pp.97–107.

Zhang, H.-Q. et al., 2003. Unified Model for Gas-Liquid Pipe Flow via Slug Dynamics—Part 2: Model Validation. *Journal of Energy Resources Technology*,

125(4), p.274. Available at: http://energyresources.asmedigitalcollection.asme.org/article.aspx?articleid=1414276.

Appendix

Four statistical parameters were used to analyse the errors in this work, they include; Average Percent Error, *APE*, which indicates the magnitude of the average error; the Absolute Average Percent Error, *AAPE*, which measures the *APE* in absolute terms thus eliminating the masking effects in error predictions. The standard deviation, *STDAE* and indicates the degree of dispersion of the errors around their average value. $\in_i$ is defined as the average error for the test point i and N is the total number of test conditions in which the values $\in_i$ were obtained.

They are defined mathematically as:

$$\in_i = \frac{\in_{predicted} - \in_{measured}}{\in_{measured}} \quad \text{(A-1)}$$

$$APE = \frac{1}{N}\sum_{i=1}^{N} \in_i \quad \text{(A-2)}$$

$$AAPE = \frac{1}{N}\sum_{i=1}^{N} |\in_i| \quad \text{(A-3)}$$

$$STDAE = \frac{1}{N}\sum_{i=1}^{N} (\in_i - \overline{\in})^{0.5} \quad \text{(A-4)}$$

RESEARCH HIGHLIGHTS

- Experimental investigation of slug liquid holdup in high viscosity oil-gas two-phase flow in horizontal pipes is presented
- Existing slug liquid holdup correlations are compared with the present experimental dataset
- A new correlation for estimating slug liquid holdup is proposed
- The proposed correlation is based on a databank with a wide range of liquid viscosity (0.189 – 8.090 Pa.s)