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Investigation of two-phase flow pattern, liquid holdup and pressure drop in viscous oil–gas flow

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ABSTRACT

An experimental investigation was carried out on viscous oil–gas flow characteristics in a 69 mm internal diameter pipe. Two-phase flow patterns were determined from holdup time-traces and videos of the flow field in a transparent section of the pipe, in which synthetic commercial oils (32 and 100 cP) and sulfur hexafluoride gas (SF_6) were fed at oil superficial velocities from 0.04 to 3 m/s and gas superficial velocities from 0.0075 to 3 m/s.

Introduction

Multiphase flow simulators are important tools used in the design and operation of oil and gas fields. The flow regime (flow pattern), liquid holdup and pressure drop in pipelines are of great interest for design of pipelines and equipment. The current state of the art multiphase flow simulators are mainly one-dimensional models, which require empirically based correlations for parameters like slug velocity, void fraction in slugs and interfacial friction factors.

High viscosity oils are among the most important future hydrocarbon fuels due to the increasing world energy demand and the depletion of conventional oil resources. Since most of the closure laws implemented in the commercial simulators are based on experiments using low viscosity oils, experiments using oils with higher viscosities are crucial to improve the existing closure laws. Among the earliest studies using oils with a high viscosity are the experiments performed by Kago et al. (1986), who conducted slug flow experiments in a 51.5 mm ID horizontal pipe. Water with polymers and water slurries were chosen as the liquid phases and air was chosen as the gas phase. The viscosity of the liquid phase ranged from 0.8 cP (water) to 55 cP (slurry). They proposed an empirical correlation for gas void fraction in the slug body. Andritsos and Hanratty (1987) performed experiments in 10 m long horizontal pipes with inner diameters 2.54 and 9.53 cm, using air and liquids with viscosities of 1, 4.5, 16 and 70 cP. They proposed a new correlation for the interfacial friction factor. Andritsos et al. (1989) performed experiments in a 25 m long horizontal pipe with inner diameter 9.53 cm, using water/glycerine mixtures with viscosities of 1, 20, and 100 cP. The study was focused on the stratified to slug transition at different viscosities. Nädler and Mewes (1995) investigated the effect of liquid viscosity on the phase distribution of slug flow in a horizontal pipe with an inner diameter of 59 mm. Air as the gas phase, and water and oil with viscosities in the range of 14–37 cP as the liquid phases were used. Multi-detector gamma densitometers were used to measure the liquid holdup, which was found to increase with increasing liquid viscosity. Gokcal (2005, 2008) and Kora et al. (2011) performed air–oil experiments, using a 50.8 mm ID horizontal pipe. The oil viscosity was varied by varying the temperature, and ranged from about 200 cP to 600 cP. Gokcal proposed new correlations for the Taylor bubble drift velocity and the slug frequency based on his experiments, while Kora proposed a new correlation for the void fraction in slugs. Smith et al. (2011) performed experiments in a 69 mm ID horizontal pipe in a 52 m long test section. 2 cP and 100 cP oils were used, and the gas phase was SF_6 at about 8 bara pressure. The slug flow region was found to be much smaller than in the experiments of Gokcal and Kora, most likely due to the

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higher gas density. The current commercial multiphase flow simulators were also found to significantly overpredict the pressure drop. Jeyachandra et al. (2012) recently performed more experiments in the same flow loop as Gokcal and Kora, and proposed a new correlation for the slug drift velocity as function of viscosity and pipe diameter. For further studies involving high viscosity liquids see for example the studies reported by McNeil and Stuart (2003), Schmidt et al. (2008), Valle (2000). Most of the above mentioned studies used low pressure air or nitrogen as the gas phase, which results in a gas density significantly lower than that of high pressure natural gas in most industrial cases. In general there is a lack of available experimental data using high viscosity oils, particularly in combination with a high density gas phase.

The focus of this experimental campaign was to provide new experimental data for a two-phase gas–oil pipe flow using SF_6 as gas phase, and both a medium (nominal viscosity 32 cP) and high (nominal viscosity 100 cP) viscosity oil. For each oil system, experiments were run at both 4 and 8 bara pressure, resulting in gas densities of approximately 25 and 50 kg/m^3 respectively. It is worth mentioning that the density of SF_6 at a pressure of 8 bara is equal to that of methane at 71 bara, a value highly relevant for multiphase hydrocarbon transport. The focus areas have been bubbly flow, the stratified to slug flow transition, and slug flow. The obtained experimental results are compared with the predictions from a two-phase gas–liquid point-model, which was also developed as part of the study. This model was presented in another publication by Smith et al. (2013).

The paper is organized as follows. Following the introductory section, in which earlier investigations are summarized, the present experimental procedure is presented in Section 'Experimental procedure'. The test section, fluid systems, instrumentation, data sampling and uncertainties are presented and discussed in detail. The two-phase gas–liquid point-model is briefly reviewed in Section 'Two-fluid gas–liquid model', with details referred to Smith et al. (2013). The observed flow patterns, pressure drop, liquid holdup and other flow characteristics and experimental results, in combination with model comparisons are shown and explained before conclusions are drawn in Section 4.

Experimental procedure

Test section

A set of two-phase gas–oil experiments was carried out in SINTEF's Medium Scale Flow Loop at Tiller, Norway. The experiments were performed in a straight 51.8 m long horizontal test section, with an inner diameter of 69 mm.

The test section was constructed by 5 m steel pipe section sections connected with flanges. The pipe sections were made of stainless steel 316 L with dimensional tolerances according to EN ISO 1127/D3/T3. These tolerances imply that inner diameter was ID = 69 ± 0.5 mm. The flange connections were standard EN 1092-1 DN65 PN16 weld neck flanges (Type 11). Special precautions were taken to ensure alignment of the flanges, and 1 mm thin intra flange gaskets were used to minimize the spacing between the flanges.

Oil and gas were mixed in a full diameter *T*-junction. The gas entered through the branch connection, and this connection was oriented upwards. The test section positions are relative to the center of this *T*-junction. For low gas rates (<8 m^3/h) gas was injected through a ¾" nozzle on the side of the test section. The nozzle was located at 3.0 m.

The outlet of the test section was connected to a 4 m long hose. The inlet and outlet of this hose were at the same elevation, but the middle of the hose was lifted one pipe diameter. This was done to avoid gravitational draining of the test section at low flow rates. The nominal inner diameter of the hose was 75 mm. The hose outlet was connected to a DN 200 vertical pipe (downcomer) that connected to the inlet of the separator 4 m below. A separate DN 50 line connected the separator's gas layer to the top of the downcomer. This way, pressure fluctuations in the downcomer due to any siphoning effects were minimized.

To enable visual observation, two transparent pipe sections were added to the test section. One section was installed between 14.0 and 14.8 m and one section was installed between 34.8 and 36.8 m. The sections were made of polycarbonate pipe with OD 75 mm and ID 69 m.

The hydraulic roughness of the test section was determined to be 7 μm from single phase experiments.

Fluid system

Sulfur hexafluoride (SF_6) and synthetic oils were used as test fluids. The experiments were conducted with Nexbase 3080, which was later diluted with Exxsol D80 to obtain a lower viscosity mix fluid. Exxsol D80 oil is a dearomatized aliphatic hydrocarbon oil commonly used in flow loop experiments due to its non-flammable, nontoxic and transparent properties. Nexbase 3080 is a catalytically hydroisomerized and dewaxed base oil comprising of hydrogenated, highly isoparaffinic hydrocarbons. Nexbase 3080 was chosen since it is non-flammable, transparent, and has a high viscosity at room temperature. The physical properties of the oils are shown in Table 1.

The viscosities of the oils were measured using a Micromotion Visconic 7829 viscosimeter connected to a bypass on the oil-feed line. The viscosity values were recorded regularly together with temperature in the viscosimeter. The viscosity models based on these measurements for the 32 cP mix fluid and 100 cP Nexbase 3080 oil are shown in Eqs. (1) and (2). For Nexbase 3080:

$$\mu_o = 196 - 5T \tag{1}$$

and for the 32 cP mix:

$$\mu_o = 59.5 - 1.39T \tag{2}$$

The resulting viscosities from these models are in cP, and the parameter T is the temperature in °C.

The oil density was measured by manually recording the value shown by the coriolis flow meter. To measure the gas density, a representative gas sample was taken from the loop using a 1 L sample flask. The mass of the gas sample (weight of filled bottle – weight of empty bottle), m_s, was noted together with sample pressure p_s and temperature T_s (in Kelvin). The gas density for a certain pressure p and temperature T are calculated using extrapolation based on the ideal gas law. A density parameter β is calculated from the measurements,

$$\beta = \frac{m_s T_s}{V_s p_s} \tag{3}$$

where V_s is volume of the sample flask. Density at a pressure p and temperature T close to the pressure and temperature of the measurement are calculated using

$$\rho(p, T) = \beta \frac{p}{T} \tag{4}$$

The density parameter was measured regularly, at least once for each change in nominal pressure and temperature. The measurements show that there is no significant change in beta over the duration of the campaign and for all experiments in this campaign an average beta value of 1757 K kg/m^3/bara has been used.

The ranges of flow rates employed are given in Table 2. Based on the superficial oil velocity, the oil-phase Reynolds number $\left(Re_o = \frac{\rho_o U_{so} D}{\mu_o}\right)$ varies from 25 to 5000.