Chemical Engineering Science 106 (2014) 60–75

Contents lists available at ScienceDirect

Chemical Engineering Science

journal homepage: www.elsevier.com/locate/ces

Experimental study of the hydrodynamic behaviour of slug flow in a vertical riser

M. Abdulkadir [a,b,c,*], V. Hernandez-Perez [a], I.S. Lowndes [a], B.J. Azzopardi [a], S. Dzomeku [b]

[a] Process and Environmental Engineering Research Division, Faculty of Engineering, University of Nottingham, University Park, Nottingham NG7 2RD, United Kingdom
[b] Petroleum Engineering Department, African University of Science and Technology (AUST), Abuja, Nigeria
[c] Chemical Engineering Department, Federal University of Technology, Minna, Niger State, Nigeria

HIGHLIGHTS

- Advanced instrumentation for slug flow measurement, namely ECT, under laboratory conditions, is used on a vertical riser.
- In this study, oil with viscosity five times higher than water is used, as it is more relevant to industry.
- Extraction of characteristic parameters of slug flow, such as frequency and slug length, from the measurements were carried out.

ARTICLE INFO

Article history:
Received 17 August 2013
Received in revised form
15 October 2013
Accepted 10 November 2013
Available online 22 November 2013

Keywords:
ECT
Taylor bubble
Liquid slug
Slug unit
Frequency
Void fraction

ABSTRACT

This paper presents an investigation of the hydrodynamics of slug flow in a vertical 67 mm internal diameter riser. The slug flow regime was generated using a multiphase air–silicone oil mixture over a range of gas ($0.42 < U_{SG} < 1.35$ m/s) and liquid ($0.05 < U_{SL} < 0.38$ m/s) superficial velocities. Electrical capacitance tomography (ECT) was used to determine: the velocities of the Taylor bubbles and liquid slugs, the slug frequencies, the lengths of Taylor bubbles and the liquid slugs, the void fractions within the Taylor bubbles and liquid slugs and the liquid film thicknesses. A differential pressure transducer was used to measure the pressure drops along the length of the riser. It was found that the translational velocity of a Taylor bubble (the structure velocity) was strongly dependent on the mixture superficial velocity. As the gas superficial velocity, was increased, the void fraction and the lengths of the liquid slugs and the Taylor bubbles were observed to increase. The increase in gas superficial velocity causes an increase in the frictional pressure drop within the pipe, whilst the total pressure drop (which is a sum of the hydrostatic and frictional pressure drop) along the length of the riser decreases. In addition, the frequencies of the liquid slugs were observed to increase as the liquid superficial velocity increases, but to be weakly dependant on the gas superficial velocity. The manual counting method for the determination of slug frequency was found to be in good agreement with the power spectral density (PSD) computed method.

1. Introduction

Slug flow occurs in horizontal, inclined and vertical pipes over a wide range of liquid and gas flow rates. It is the dominant flow pattern in upward inclined flow. The flow is associated with operational problems such as operation, erosion–corrosion enhancement, and structural problems occurring most especially in bends. Slug flow hydrodynamics is complex with unsteady flow behaviour characteristics. It has peculiar velocity and pressure distributions. Therefore, the predictions of the liquid hold-up, pressure drop, heat transfer, mass transfer are difficult and challenging.

The power, nuclear and chemical industries have provided a platform for strong interest in multiphase flow study. Examples of such studies are steam-water flow for power generation and flow reactors for heat and mass transfer. Additionally, applications in the petroleum industry provide another strong motivation for multiphase flow research. The transportation of multiphase involving, oil and gas in pipes is significantly reducing the cost of reservoir development. However, the main challenge confronting them is the development of multiphase technology for the transportation of oil and gas from subsea production units at large water depths to processing facilities at nearby platforms or onshore facilities. The flows in the subsea pipelines usually contain multiple phases, like oil,

* Corresponding author.
E-mail addresses: mukhau@futminna.edu.ng, mukhau@yahoo.com (M. Abdulkadir).

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http://dx.doi.org/10.1016/j.ces.2013.11.021

water and/or gas, whose composition is not known priori. The variation in the composition of fluids inside the long subsea network can lead to serious operational problems, ranging from non-continuous production or shut-down to damage equipment.

Gas with large amounts of water and or oil–water mixtures may be produced simultaneously, resulting in multiphase flow conditions in the transporting pipe system between the source and the production platform. As the fields grow older, the produced gas contains an increasing amount of water, giving rise to different mixture compositions, which affect the flow pattern and flow characteristics (Zoeteweij, 2007). For upward inclined and vertical pipe flow, slug flow can be considered as the dominant flow pattern, Hernandez-Perez (2008). This can enhance corrosion, as Kaul (1996) noted that the corrosion rate is accelerated when the flow pattern is slug flow. This flow pattern is usually characterised by an alternating flow of gas pockets and liquid slugs. Most of the gas-phase is concentrated in large bullet-shaped gas pockets, defined as Taylor bubbles. The Taylor bubbles are separated by intermediate liquid slugs, which may contain small entrained gas bubbles. A major characteristic of slug flows are their inherent unsteadiness. As this kind of flow occurs over a wide range of intermediate flow rates of gas and liquid, it is of major interest to a wide range of industrial processes that employ pipeline transport systems.

The presence of liquid slugs in the flow system gives an irregular output in terms of gas and liquid flow at the outlet of the system, or at the next processing stage. This can pose problems to the designer and operator of two-phase flow systems. Pressure drop is substantially higher in slug flow as compared to other flow regimes, and the maximum possible length of a liquid slug that might be encountered in the flow system needs to be known (Abdulkadir, 2011). In the 67 mm facility used in the current study, slugs were observed to be of about 10 pipe diameters in length, i.e., long enough for the rise velocity to be independent of the length (Griffith and Wallis, 1961). For large capacity systems in industry, these liquid slugs can even grow longer, carrying large momentum. Often, slug catching devices are used to collect the slugs, and avoid any damage to the downstream equipment. For the design of such slug catchers, it is important to know what kind of slugs to anticipate. The important question of when, and how, these slugs are formed has received much attention from research workers (Abdulkadir, 2011). However, reports on the study of the behaviour of these slugs in more industry relevant fluids are limited. For that reason, it is important to study the behaviour of slug flow in great detail for the optimal, efficient and safe design and operation of two-phase gas–liquid slug flow systems.

An analysis of the ECT data enabled the characterisation of the slug flow. This enabled the measurement of the instantaneous distribution of the flow phases over the cross-section of the pipe. The use of two such circumferential rings of sensor electrodes, 89 mm apart (also known as twin-plane sensors), enabled the determination of the translational velocity of the observed Taylor bubbles and liquid slugs. With this information, a more fundamental approach is used for improving the general knowledge on slug flow.

1.1. Background

The occurrence of slug flow in a vertical riser is a very common phenomenon under normal operating conditions within a two-phase flow facility, such as in an oil production riser. Over the past thirty years there have been a large number of research studies in this field in the peer review literature.

One of the earliest contributions to slug flow characterisation was carried out by Nicklin et al. (1962), who proposed an empirical relationship to describe the rise velocity of single Taylor air bubble in a static water column. Nicklin's empirical relationship, given by Eq. (1), describes the translational velocity of a Taylor bubble, U_N, it is the sum of the Taylor bubble rise velocity, or drift velocity, which is the velocity of a Taylor bubble in a stagnant liquid, plus the contribution of the mixture superficial velocity in the preceding slug. For the air–water system in a 26 mm bore tube considered the values of the constants C_0 and k were determined to be 1.2 and 0.35, respectively:

$$U_N = C_0 U_M + k\sqrt{gD} \tag{1}$$

where C_0 is a flow distribution coefficient, k is drift coefficient, U_M is the mixture superficial velocity and $k\sqrt{gD}$ is the drift velocity.

Moissis (1963) agreed that U_N predicted by Eq. (1) fitted his data well. Akagawa and Sakaguchi (1966) confirmed the applicability of Eq. (1) to the air–water system in a 26 mm diameter pipe and pointed out the effect of the term $k\sqrt{gD}$ is negligible except at low gas and liquid velocities. They suggested that the presence of small bubbles in the liquid slug has a slight effect on the translational velocity of a Taylor bubble and that their data indicated that C_0 is in the range of 1.25–1.35.

Brown (1965) found experimentally that the analytical solutions for the translational velocity of a Taylor bubble derived by Dumitrescu (1943) and Davies and Taylor (1950) described the behaviour of gas bubbles in low viscosity liquids well, however, they were not suitable for high viscosity liquids.

Vince and Lahey (1982) claimed that excellent correlation between the translational velocity of a Taylor bubble, U_N and the mixture superficial velocity, U_M was given by the following:

$$U_N = 1.29U_M + 0.15 \tag{2}$$

Following an analysis of experimental data in a 50 mm diameter tube, Fernandes et al. (1983) determined a slightly higher value of 1.29 for C_0. They ascribed the increase in the constant C_0 to diameter effect or to the contribution made by heading and trailing Taylor bubbles. Barnea and Shemer (1989) verified Eq. (1), using their own measurements on a 50 mm diameter tube, and used it in their calculations. A more physically based interpretation of the proposed increase in the constant was later provided by Mao and Dukler (1985). They used an aqueous electrolytic solution and air in a 50.8 mm diameter tube. They took into consideration the fact that the liquid slug in front of a Taylor bubble is aerated, and coalescence takes place between the small bubbles and the Taylor bubbles. This results in an increase of the velocity of the Taylor bubble. They derived a mathematical relationship to determine the translational velocity of a Taylor bubbles:

$$U_N = C_0 U_{LLS} + 0.35\sqrt{gD} + \Delta U_N \tag{3}$$

where, U_{LLS} is the liquid superficial velocity in liquid slug, ΔU_N is an increment of U_N as defined in Eq. (1). The value of the constant C_0 is based upon the assumption that the propagation or the front velocity of the slug, or the velocity of the interface between the gas and the liquid phases follows the maximum local velocity, $U_{\max}$ in front of the nose tip and thus $C_0 = U_{\max}/U_M$ (Nicklin et al., 1962; Bendiksen, 1984; Shemer and Barnea, 1987; Polonsky et al., 1999). The value of C_0 according to them is approximately 1.2 for fully developed turbulent flow and 2.0 for fully developed laminar flow:

$$\Delta U_N = (U_N - U_{GLS})\frac{\varepsilon_{gs}}{\varepsilon_{TB}} \tag{4}$$

$$U_{GLS} = U_0 + U_{LLS} = 1.53\left[\frac{\sigma g(\rho_L - \rho_G)}{{\rho_L}^2}\right]^{1/4}(1-\varepsilon_{gs})^{1/2} + U_{LLS} \tag{5}$$

Defining β as the ratio of the void fraction in the liquid slug and Taylor bubble:

$$\beta = \frac{\varepsilon_{gs}}{\varepsilon_{TB}} \tag{6}$$