CRANFIELD UNIVERSITY

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MULTIPHASE FLOW INSTABILITY AND ACTIVE SLUG CONTROL SOLUTIONS

SCHOOL OF WATER, ENERGY AND ENVIRONMENT
OIL AND GAS ENGINEERING CENTRE

PhD
Academic Year: 2015 - 2018

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This thesis is submitted in partial fulfilment of the requirements for the degree of PhD in Energy

ABSTRACT

Slugging as a flow assurance challenge is an upsetting condition to the oil and gas industry due to the instabilities it poses on the system. The negative repercussions associated with slug flow stems from the inlet through to the topside facilities where processing is done. Active control has been established as one of the best techniques to eradicate slug and its accompanying challenges however the controller robustness and some setbacks make improvement a necessity. In that vein, the Inferential slug controller which uses a combination of topside measurement signals to produce a single control variable which is more sensitive to slug variations hence can effectively be used to control slug, was invented. Again the robustness of this controller has been in question.

This study presents a comprehensive and systematic analysis of the Inferential slug controller design for system stability analysis and maximising throughput from unstable riser pipeline system configurations in the quest to advance this technology. The inferential slug controller's robustness was assessed by implementing this technique on several pipeline riser systems including U-shape and S-shape riser configurations. Prior to that, the flow behaviour for a wide range of flow conditions was investigated, highlighting the impact of geometry on unstable slug flow through the OLGA flow simulator (modelling) and experiments. New and unused measurement signals from the topside of either the riser/platforms were deployed in the inferential slug control technology to make the controller more sensitive and robust. A simplistic nevertheless robust procedure for designing the inferential slug controller was proposed. Unstable slug flow conditions were observed to stabilise at a relatively larger valve opening compared with that seen in open loop.

The inferential slug controller technology is further extended to deal with systems with variable time delay using a proposed modified Smith predictor model. The modified Smith predictor was recorded to improve and stabilise a pipeline riser system which has deteriorated in control performance due to time delay in the system, a resultant of large stroke time in the valve. This in practicality means an increased production through the system.

In advancing the ISC technology to be deployed on offshore fields in conjunction with other passive slug mitigation techniques, the slug mitigation potential of pseudo spiral tube (PST) was assessed when installed at the topside of the riser system. The analysis showed that the PST pipe section (spiral and wavy piece) when installed at the topside of the riser system, possesses some mitigation potential. Four different slug regions was identified for the entire pipeline system. The first region being a slug flow occurrence in the system with and without the PST whiles the second region is the region where slugging occurs in the system but disappears when coupled with the PST and the opposite describes the third region. Lastly, the fourth region is described as that region where slugging flow exist for the system coupled with the spiral pipe section and without any PST (plain) but slugging flow disappears when the system is coupled with the wavy pipe section. The wavy or spiral pipe section coupled with the S-shape riser system have slug mitigation capabilities when they are installed at the top of the riser although its effectiveness of slug mitigation depends on the flow condition. This is evident in the significant reduction in the riserbase pressure oscillation magnitude and the significant reduction in the slug envelope (region) when the system was coupled with the wavy or spiral pipe section relative to the plain system.

Keywords:

Slugging, Active control, inferential slug controller, Pseudo spiral tube, Smith predictor,

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LIST OF ABBREVIATIONS

BP	British petroleum
CAPEX	Capital expenses
DIE	Delivery, Integrity & Energy
FIR	Finite impulse response
FOPDT	First order plus dead time
GVF	Gas volume fraction
ID	Internal diameter
ISC	Inferential slug controller
OPC	OLE for process control
OLGA	Oil and Gas
PCA	Principal component analysis
PD	Proportional Derivative
PI	Proportional Integral
PID	Proportional Integral and Derivative
PST	Pseudo spiral tube
PVC	Polyvinyl chloride
SCADA	Supervisory control and data acquisition
SI	Slug Index
UFOPDT	Unstable first order plus dead time

1 INTRODUCTION

1.1 Background and motivation for study

The expanding demand for energy on the world market has hard-pressed and forced the oil and gas industry to generate some more energy to meet the demand. This has driven the industry to undertake several exploration and exploitation of new and untapped supplementary reserves at a very fast pace to increase production volumes even though the market prices are not always favourable. Maintaining a constant or steady production by means of managing the pressures has been a great challenge to flow assurance in the oil and gas industry. Flow assurance is the ability to safely and economically transport flow from the well / reservoir to the platform or the seabed to the onshore facilities through platforms or offshore facilities. The regulatory and maintenance of steady flow delivery from the system outlet are key considerations of multiphase flow technology while still minimising the process upsets.

Transportation of fluids is an inevitable process in the oil and gas industry and often done through a single pipeline system. The popularity of satellite field is on the rise since most and recently found fields are not economical to run as a standalone hence made by tying in of an existing production pipelines which results in complex pipeline configurations and long pipeline networks. In delivering fluids through these complex and long pipeline systems, the oil and gas industry is faced with the challenge of maintaining steady production due to the difficulty in managing the pressures in the system. The decline in production pressures over time is a contributing factor for the challenges faced. Most supplementary reserves which form the majority of the total number of production wells have already hit their peak season and are at a decline in terms of rate of production due to a reduced pressure. As the well matures in age, there is a reduction in flow and pressure making self-lifting almost an impossible activity and initiates several flow assurance challenges. One of such flow assurance challenges is slugging flow regime (Zheng, Brill and Taitel, 1994), a complex multiphase flow phenomenon.

Slugging flow is primarily characterized by large pressure and fluctuating flow, which often features an alternating flow of both gas and liquids in an irregular manner. Slugging can lead to major disruptions on the topside facilities and reduces the life of the production systems. Significant effort has been channelled towards the understanding of the slugging flow phenomenon and the established mitigation techniques. Active slug control mitigation technique has been established as the most prolific method for minimizing slugging flow. In the quest to optimize flow / production, the controller robustness for several slug control techniques has been questionable. Again, the complexity of multiphase flow in pipeline (complex and long flow configuration) is something worth understanding.

1.1.1 Classification of multiphase system challenges (DIE concept)

All possible issues that could potentially be encountered in a multiphase flow system have been categorised in three. These three includes enablement of delivery, system integrity and the right amount of energy (Montgomery, 2002). This is acronym as the DIE concept simply for Delivery, Integrity and Energy.

1.1.1.1 Delivery (D)

Steady flow at the outlet is achieved by ensuring a right system design that would maintain and aid fluid transport. This design however includes the receiving and handling facilities which would help in the successful management of any potential flow (gas, liquids or both) variations. The flow occurrences are predicted using some transient and steady models to achieve good operational planning and system design.

1.1.1.2 Integrity (I)

The integrity of the system is of great importance for maintaining steady flow delivery at the outlet of the system. This is observed in fluid containment and the elimination any form of impediment to the flow through the system. All system mechanical and chemical interactions are catered for under this classification.

1.1.1.3 Energy (E)

Safeguarding the system such that there is sufficient energy to move the fluid from the source to the outlet of the system. Prediction of the temperatures and pressures along the entire system to investigate the loss characteristics between each two points along the system is of great essence. This is to explore the potential of additional energy along the pipeline system.

An overlook of the DIE concept could have an impact on the flow and system as a whole. Emulsion formation, wax / hydrate formation, larger temperature changes over the riser length especially for long lengths which affect flow characteristics and the chemical properties are likely effects if the DIE concept is disregarded. Without compromising the drag and mechanical performance of the riser, insulation is sufficiently added to the pipeline riser system to help eliminate such issues.

Figure 1-1 shows a pictorial projection of the issues categorisation by Montgomery, 2002.

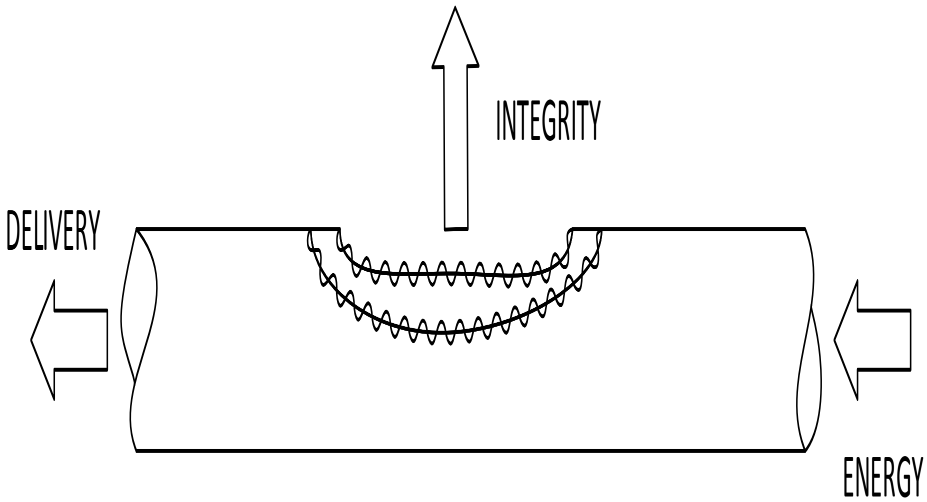


Figure 1-1 Delivery, integrity and energy concept projection

1.2 Fact finding

Since the identification of slugging flow in the oil and gas industry, vast work and effort have been routed to actively control slug flow (Sivertsen and Skogestad, 2000; Henriot et al., 2002; Kovalev et al., 2004; Ogazi et al., 2010; Vidal et al., 2013). The innovative inferential slug controller (ISC) technology was introduced to eliminate the setbacks posed by using measurement signals from the riserbase or upstream the riser to control slug (Cao, Yeung and Lao, 2010a). However, in the design of the ISC, the controller parameters were determined by trial and error which leaves the ISC technology optimization potential questionable. In the quest to advance the ISC technology, a systematic approach to determine the controller parameter would be established which in tend improves and optimize the ISC technology making it more robust in its operation. Notwithstanding other measurement signals from the topside of either the riser or platforms would be investigated as a means of improving the ISC.

The recent trend of tying in pipeline from new reserves into existing pipeline system has greatly advanced the pipeline configurations observed in the industry. Despite the advancement of the pipeline configuration in the oil and gas industry, there has been minimum attention on the effect this pipeline configuration possesses on the entire system even though the integrity of the pipeline has been looked into. Hence the renewed interest in the flow behaviour and an investigation into the initiation of flow instabilities in different pipeline riser configurations.

Wavy pipe when installed in the horizontal section prior to the riser was established to mitigate flow oscillation (Xing et al., 2013a). The wavy pipe greatly reduces the slug length hence improves system stability however, this cannot be deployed on already existing facilities since the riserbase is not readily accessible and involves huge cost in installing the pipe section at the base of the riser. To avoid this setback, the slug mitigation potential of a pseudo spiral tube (wavy pipe configuration and spiral pipe configuration) is investigated when installed at the topside of the riser pipeline system so to enable its use in conjunction with active control techniques.

Again some control techniques fail or degrade in terms of performance when deployed on real fields (Cao, Yeung and Lao, 2008). This has deterred lots of companies to try new technologies unless proven beyond reasonable doubt which makes most promising techniques shelved ideas. Interest is placed on some factors leading to the failures or poor performance of controllers and solutions to these problems.

Advancing the cause of active slug control technique (ISC technology) to be a widely accepted method has been the drive for this research. In so doing, this technology is tested to suit several flow conditions and different pipeline configurations and setups such as U-shape riser, S-shape riser or even PST configurations. The observed facts found from existing works have triggered the motivation for this research which includes: understanding the dynamics of unstable flow in riser pipelines systems, identifying the effect of pipeline riser configurations on unstable flow or slug flow, assessing the slug attenuation potential of the wavy and spiral pipe sections and advancing and optimization of active slug control techniques.

1.3 Project aim and objectives

The research is intended to advance and optimize the inferential slug control (ISC) technology while attenuating and stabilizing unstable (slug) flow at larger valve opening relative to manual choking for different riser configurations and pipeline setup.

To accomplish the quest of this research, the following specific objectives were set:

- Investigate the flow dynamics, slug envelopes (region where slug flow occurs) and stability analysis in U-shape riser pipeline configuration.
- Investigate by experiment the initiation of flow instabilities in different pipeline riser systems and the entire production operation.
- Investigate the flow dynamics, slug envelopes and stability analysis in S-shape riser pipeline configuration.
- Investigate stabilising unstable flow conditions by parameter variation technique (using choke valve) for different riser configurations.
- Advance the Inferential slug control (ISC) technology and implement it on different riser configurations (U-shape and S-shape risers) to mitigate slug.
- Investigate unstable slug flow mitigation / attenuation potential of a Pseudo Spiral Tube (wavy and spiral pipe section) when installed at the topside of the riser and assess its capacity to enhance the parameter variation technique.

1.4 Methodology

Experimental runs and process simulation approach are the methods adopted in this work. A brief framework of the methodology is defined in this section.

1.4.1 Experimental run of gas –liquid flow behaviour and control in 2 inch pipelines of different configurations

The behaviour of gas – liquid flow on 2 inch riser pipelines of different configurations (Purely vertical, U-shape and S-shape) was investigated. The multiphase flow facility in Cranfield University was deployed for this study. This study was to understand the flow dynamics in the pipeline riser systems (vertical, horizontal and the riser). The initiation of slug flow in all these flow loops was assessed so to understand the impact of flow geometry on the dynamics of flow. Slug mitigation potential of some passive and active slug mitigation techniques were assessed.

Pseudo spiral tube, PST (wavy and spiral pipe), a passive slug mitigation technique was employed for subduing slug. The PST injunction with the topside choke valve variation was investigated to establish the ability of this technique to alter the system behaviour in a riser pipeline system. The PST was implemented as a riser top flow conditioner, which is designed as an up and down flow pipe, installed upstream of the gas-liquid separator for stabilising / subduing slug flow. The wavy pipe installed at the base of the riser showed slug attenuation benefits however this cannot be deployed for already existing facilities hence the PST position. Features and description of the PST will be discussed in Chapter 5.

Active slug mitigation techniques were assessed using singular measurement signal as the control variable and a combination of multiple measurement signal as the control variable. The slug mitigation potential of some measurement signals and the Inferential Slug Control technology was tested on different flow loops. This was investigated to know the viability of using such technology on real life fields and hence aid in the optimization of the slug control technique. Further details on the flow loop descriptions, instrumentation set up, data acquisition and the operating procedure is documented in later chapters of this work.

1.4.2 Simulation of unstable gas – liquid flow behaviour and control in large diameter pipelines of different configuration

In-depth knowledge was derived for gas – liquid flow in industrial sized pipeline riser systems through simulation. OLGA, Matlab and OLGA-Matlab coupled together through OLE for process control (OPC) was used to model these industrial pipeline systems. This was to investigate and establish the use of the ISC technology in real fields. The models used were tuned to accurately represent real systems thus representing the complexity observed in multiphase flow. Other measurement signals and components required to be investigated but not available on the experimental set-up was performed through simulation. An advancement of the ISC technology was delved into using relatively larger diameter pipeline systems. Outline of the pipeline systems, configuration, and activities performed and set-up is outlined in subsequent chapters.

1.4.3 Slug mitigation technique

The ISC technology was used as the slug mitigation technique in this study. Cao, Yeung and Lao, (2010b) on the basis of the problems and setbacks encountered with using the riserbase pressure as a control variable to help stabilise slugging flow, produced a novel controller commonly known as Inferential Slug Control (ISC). This novel approach uses multiple topside measurements as the input to attain a percentage choke valve opening that stabilizes flow within the slugging regime in riser systems and minimizes the impact of slug control on overall production. The choke valve position function from the control equation is given in the form;

$$V = V_o + K(W^T Y - R)$$

where V_o is the choke valve nominal value which is predetermined and manually set to a position where the flow becomes stable or within an acceptable range, K is the control gain, W is the measurement weights, Y is the vector of measurements, $W^T Y$ is the control variable, and R is the set point of the control variable.

The mode of operation, input variables, principle, control design and corresponding set point to calculate the combination coefficients of the ISC technology will be further discussed.

1.5 Thesis outline

Chapter 2 details a literature review on multiphase flow with prominence on flow instabilities (slug flow) in pipeline riser systems and how this flow instabilities (slug flow) is subdued. The use of different slug mitigation techniques (passive and active) previously investigated is presented. The chapter ends with the aim and objectives in a technical perspective based on the literature survey.

Chapter 3 is devoted to extending the use of the ISC technology for a U-shape riser pipeline configurations. To accomplish this fact an understanding of the flow dynamics in a 2 inch U-shape pipeline riser system with much emphasis on unstable slug flow or instability in the U-shape pipeline riser is assessed. The initiation of this flow instabilities in the U-shape pipeline riser system is investigated experimentally. Understanding the flow behaviour in the U-shape riser could help in developing effective control techniques. With the quest of stabilising unstable flow behaviour using the ISC technology, some single measurement signals from the topside of the riser is used as a control variable to stabilise unstable slug flow. Numerical simulation tool, OLGA, is used to validate the experimental studies by using extra measurement from a remote platform to extend the ISC technology.

Chapter 4 is committed to extending the ISC technology for an S-shape riser pipeline configuration. In this quest, understanding the hydrodynamics for a gas - liquid flow in an S-shape pipeline riser system is of great essence. With flow instabilities / unstable slug flow as a point of interest, slug envelope is established from the different flow combination in the S-shape riser system. S-shape pipeline riser configuration dip effect on flow behaviour is also presented. Slug flow stability analysis using the ISC technology is presented. The stability analysis conducted experiments are validated using the numerical simulation tool, OLGA.

Chapter 5 is dedicated to extending the ISC to work with pseudo spiral tubes (PST – wavy pipe configuration and spiral pipe configuration). Unstable slug flow attenuation in S-shape riser pipeline configuration using a wavy and spiral pipe section installed at the topside of the riser is examined. This illustrates how the installed wavy or spiral pipe helps stabilise relative unstable flow condition. Again, a slug envelope comparison is drawn between the S-shape riser pipeline system with and without the PST pipe sections.

Chapter 6 presents an advancement to the ISC technology, a slug mitigation potential and expansion of the Inferential Slug Control (ISC), an active slug control strategy that uses measurement signals from the topside of the riser, to deal with pipeline riser systems with large stroke time.

Chapter 7 presents a general discussion on the general findings, conclusions and suggests future recommendations for further propositions.

1.6 Publications

Table 1-1 shows the publications that have resulted in the course of this research and investigation.

Table 1-1 Resulting publications from this research

Paper Title	**Conference Name**	**Paper Status**	**Action**
Stability of severe slug flow in U-shape riser	22nd International Conference on Automation & Computing (ICAC'16), University of Essex, Colchester, United Kingdom, 7-8 September 2016	Accepted	Conference Presentation
Smith predictor for slug control with a large valve stroke time	27th European Symposium on Computer Aided Process Engineering – ESCAPE 27 October 1st - 5th, 2017, Barcelona, Spain	Accepted	Conference Presentation

2 LITERATURE REVIEW

2.1 Introduction

Multiphase flow is the simultaneous flow of more than one phase in a solitary pipeline. Multiphase flow can either be two-phase, three-phase or four-phase (oil, water, gas and sand particles) depending on the flow composition or blend. The composition of these fluids determines the flow distribution of the fluid when being transported which is of great importance to various industrial processes in the power and energy generation sectors, food and beverage establishments, medicine manufacturers, aerospace, oil and gas and their likes. This is because the flow characteristics of a multiphase flow are adopted to the measure of the efficiency and the effectiveness of an industrial process involving multiphase transportation (Brennen, 2005). For the benefit of this thesis, two phase flow with special interest in gas-liquid flows would be considered because of its popularity in the oil and gas petroleum industry. A suitable starting point in understanding multiphase flows is assessing the interaction (slip effect or miscibility) between the phases involved.

2.1.1 **Slip effects in multiphase flow**

A suitable initial point for understanding multiphase flows is a phenomenological description of the mechanism of geometric distributions or flow patterns that are observed in common multiphase flows. Gaseous phases are often miscible while on the contrary, liquid phases may either be miscible or immiscible, thus forming other complex phases. Liquid-Liquid interface is influenced by interfacial tension while gas-liquid interface is influenced by surface tension. Thus, the liquid-liquid interface is affected by the cohesive force that exist among the liquid molecules while the gas-liquid interface is influenced by the adhesive force between the phases.

When fluids (liquid and gas) are transported horizontally in the oil and gas industry, gases tend to move faster than the liquid phase since it is the lighter phase. However in flow lines with some downward inclination the liquid moves comparatively faster downward since it is the heavier phase. In horizontal multiphase pipelines, the liquid holdup is greater than the liquid volumetric fraction and the reverse happens in vertical pipelines or pipelines with an inclined topography (Sætre, Johansen and Tjugum, 2012).

The purpose of this chapter is to expound concisely on the theory of two-phase and multiphase flow in general. The remaining of the chapter is branched in three main sections which is organized as follows; the first section outlines some applicable basics of two phase flow and some frequently used terminologies referred to in this work. The second section elaborates on slugging flow concept in both horizontal and vertical pipeline systems. The categories and type of slugging flow are also outlined here. The third section presents an extensive literature survey on slug flow in vertical pipes with special interest in unstable slugging flow mitigation and control. This is assessed both through experimental studies and finally the use of OLGA theory code and state of the art review on the application of OLGA in multiphase flow.

2.2 Fundamentals of two phase flows

2.2.1 Basic terminologies

Several terminologies have been given and used in multiphase flow operations for which some are outlined in this section.

Phases

In oil and gas fields (multiphase flow) just like any other industry, phases are distinct and consistent form of matter. In multiphase flow, the continuous phase also known as the primary phase is the phase with the bulk component also known as the carrier fluid. Whiles the discrete phase, also known as the secondary phase, is the phase which comes in the form of droplets or bubbles in two phase flows. The secondary phase is often carried along by the primary fluid hence would have the same velocity, whereas in other times would have a different velocity (its own velocity) to the primary phase.

Velocities

Generally in multiphase flow, two different velocity terminologies are deployed, namely superficial velocities and phase velocities.

Superficial velocities

Superficial velocities either for liquid or gas, is the velocity of phase assuming the fluid if not affected by any other phase, thus its flowing as a lone phase in the entire pipeline.

This is influenced by the flowrate of the phase through the system, the density of the phase and the pipeline diameter. Mathematically,

$$Superficial\ velocity, u_{sf} = \frac{Volumetric\ flowrate\ of\ phase, \dot{V}}{Area\ of\ the\ pipe, A} \tag{2-1}$$

Phase velocities

Phase velocity, also known as actual velocity takes into account the volume fraction of the phase. Mathematically, phase velocity is a multiplication volume fraction and the superficial velocities.

Volume fraction / void fraction

Volume fraction or void fraction is the ratio of the volume of gas phase (either the dispersed or continuous) to the total volume through the system. This is mathematically represented by

$$Volume\ fraction, \epsilon^{i} = \frac{Volume\ of\ Gas\ phase, V_g}{Total\ volume\ through\ the\ system, V_T} \tag{2-2}$$

Basically, a summation of the volume fractions of all the phases in the system must be unity. Void fraction is mostly used for gas phase whiles that of the liquid phase is termed 'hold-up'. These terms through multiphase systems have been used interchangeably. These terms are definitely the most essential parameters that relate the two phases and thus provide essential evidence of the combined two phase behaviour. For two phase systems, hold-up is given as,

$$Liquid\ Hold-up, \varepsilon = \frac{Volume\ of\ Liquid\ phase, V_l}{Total\ volume\ through\ the\ system, V_T} \tag{2-3}$$

$$Liquid\ Hold-up, \varepsilon = 1 - \alpha_g \tag{2-4}$$

Pressure gradient

The pressure gradient adopted in this study, is the rate of change of pressure with either a flow parameter, time or even the distance along the pipeline system. The pressure gradient of the system sums all the static and dynamic pressures within the system. However, some of the pressure components could be very small, hence negligible in this study.

2.2.2 Two phase flow patterns (gas-liquid flow)

Gas-liquid flows in pipes can embrace several physical distributions known as flow patterns. Flow patterns describe the geometric distribution of the phases in a particular flow composition and are characterized by the distribution of the phases. This defines the structure of the flow in the conduit in which they are flowing. The pattern of flow in multiphase flow is subject to the operating conditions of the process, the geometry of the conduit, the properties of the gas and liquid phases, the flowrate of both the gas and liquid phases and their likes. Flow patterns influence and determine the physical and thermodynamic features (mass and heat exchange) of industrial processes.

The internal geometries of flow and interfacial area determine the momentum, rate of change in energy between the internal phases and the rate of change in energy to external environment hence the importance of studying the flow patterns of systems. The challenge however is the prediction of the flow patterns for a combination of flow operating conditions and the characteristics of the phases as well as points of transition from one pattern to the other.

They have been given unique names which are now relatively standard. Flow patterns and regimes differ from flow composition due to the difference in the component velocities which sometimes causes slippage depending on the geometric configuration of the pipe, the size of the pipe, the flowrate of each phase and their likes. An appropriate starting point is a phenomenological description of the geometric distributions or flow patterns that are observed in common multi-phase flows. Below follow the various flow patterns observed in both vertical and horizontal pipelines for gas-liquid flow systems. These flow patterns can be identified by several methods, which however include the unbiased indicator procedures (void fraction fluctuations, pressure fluctuations, x-rays, gamma-rays, fluorescent light, tomography and their likes) or traditional techniques (direct visual observation or photography in transparent pipe).

2.2.2.1 Two phase flow patterns in vertical pipes

This section describes the flow patterns observed in vertical pipes and hence identifies the instabilities and factors that can lead to a transition between flow patterns. Hu et al., (2006) and Eikrem, Aamo and Foss, (2008) stated that these flow patterns could also happen in a gas-lifting production well due to a casing heading mechanism.

Different composition of gas and liquid in vertical pipes exhibit several flow patterns or regimes. These flow regimes are unique in their characteristics and features. There are several flow patterns described in literature but Figure 2-1 shows a schematic of some few types of flow patterns that occur in vertical flow lines which are widely accepted and known.

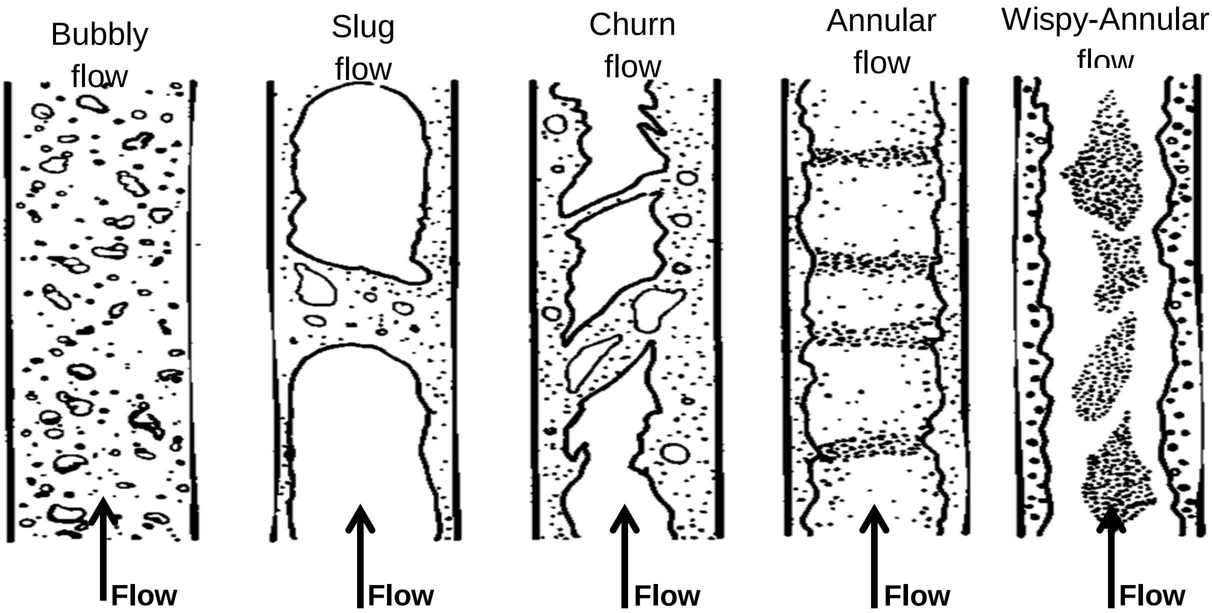


Figure 2-1 Schematic Representation of flow regimes in vertical pipes (Vince and Lahey, 1982).

Bubbly Flow: Bubbly flow is observed when the gas phase of a multiphase flow is dispersed in the liquid phase. This dispersion of the gas phase happens in the form of discrete bubbles and continuously. The bubbles are in usually spherical in nature and are relatively smaller in diameter with respect to the diameter of the pipe.

Slug flow: With an increase in the amount of gas in the flow, the bubble flow increases and begins to collide with each other. They coalesce and form larger diameter bubbles that approach the diameter of the pipeline and are separated by packets of liquid which could also contain smaller entrained bubbles. In some instances, severe slugging happens through the system. Severe slugging regime is mainly characterized by fluctuating flow, which often features an alternating flow of both gas and liquids in an irregular manner. The characteristics of severe slug flow includes large pressure and flow fluctuations, which can lead to major disruptions on the topside or processing facilities.

Churn Flow: Churn flow happens when there is instability in the flow of the fluid with an oscillatory flow occurring. This is as a result of the velocity increase of the gas flow which also causes an increase in the bubble velocity. The flow pattern is a transitional flow of a multiphase flow from slug to annular flow.

Annular Flow: Annular flow is seen when the gas phase occupies the central part of the pipeline whiles the liquid flow on the pipe walls as a film. This happens as a result of high gas velocity that causes interfacial shear on the liquid film which is dominant over the force of gravity. Additionally there are some entrained liquid in the gas core in the form of droplets.

Wispy annular flow: Wispy annular flow pattern is a type of annular flow with an increase in the concentration of liquid droplet in the gas core. This happens as a result of an increase in liquid flowrate.

2.2.2.2 Two phase flow patterns in horizontal pipes

This section describes the flow patterns observed in horizontal pipes and hence identifies the instabilities that can lead to a transition between flow patterns. Flow patterns in horizontal pipelines are very much similar to that in vertical pipelines just that the flow in horizontal pipes is influenced by gravity in that the liquids which is the more dense component stays at the base of the pipeline while the gas flows on the upper part of the pipeline. Flow patterns in horizontal pipelines are uniquely characterized and categorized as shown in Figure 2-2.

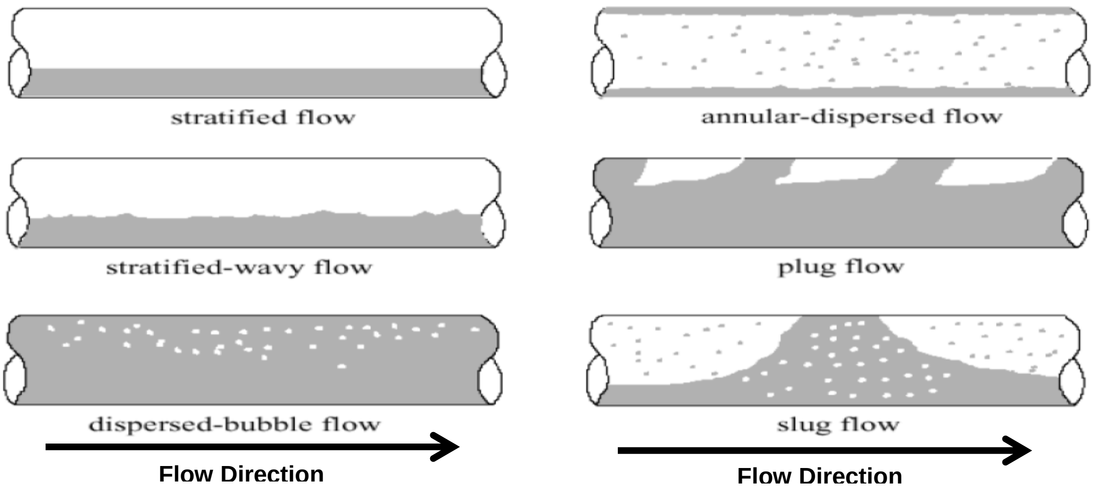


Figure 2-2 Schematic representation of flow regimes in horizontal pipes (Panton and Barajas, 1993)

Bubbly Flow: Bubbly flow pattern often occurs with high mass flow rates and it is distinguished by dispersed discrete gas bubbles in liquid with majority of the bubbles at the upper half of the horizontal tube due to buoyant force.

Plug Flow: Plug flow pattern is characterized by elongated gas bubbles separated by plugs of liquid. Due to the fact that the diameter of the plug is relatively smaller compared to the pipe diameter, the gas bubbles flow on top of a continuous liquid phase at the bottom of the pipe.

Stratified Flow: Stratified flow pattern occurs at low liquid and gas velocities or flowrates where at this point both phases even though are flow co currently flow in completely separate as two different phases with the gas flowing on top of the liquid. The phases have no effect on each other and are separated by a horizontal interface.

Wavy Flow: Wavy flow pattern is a form of stratified flow pattern but in this case the velocity of the gas is of a high velocity leading to the formation of waves on top of the interface of the gas and liquid along the flow direction. The wave amplitude is dependent on the velocities of the gas and the liquid. In wavy flow patterns the crest of the wave never gets to the top of the pipe line.

Slug Flow: Slug flow pattern happens at very high gas velocities in such a way that it increases the wavy flow region. This wavy flow region is increased to the extent that it gets to the height of the pipeline. They are often described as waves with larger amplitude.

Annular Flow: Annular flow pattern occurs at a much higher gas velocity compared to slug flow pattern. This high gas velocity forms an annular film of liquid around the pipe perimeter. The base of the pipeline has a greater share of liquid compared to that of the top of the pipe due to gravitational effects.

2.2.3 **Flow pattern identification**

Normally flow regimes are classified and identified on flow regime maps which usually are a plot of the liquid and gas superficial velocities within a range of conditions. There are many factors influencing the flow regime of any given condition and since a minority of them are considered in constructing a flow regime map, the identification of flow regimes using solely flow regime maps is practically not ideal.(Zhou, Kaasa and Aamo, 2008).There are mainly two methods used in identifying flow regimes when constructing flow regime maps namely; direct observation thus visual inspection and extraction of characteristic variable from signal fluctuation in both phases (Sun et al., 2002).

Direct Observation

Also known as the subjective descriptive type of flow regime identification since it could vary with each individual's observation. Methods such as human visual inspection or high speed cameras and their likes are deployed in determining the flow regime (Zhou, Kaasa and Aamo, 2008).

Extraction of signal fluctuation (indirect observation method)

Pressures and change or fluctuations in pressures signals are acquired and analysed using well defined methods to determine the flow regime of the composition. The spectral content of the unsteady signals from either the pressures or the volume fractions are used in instances where visual information is difficult to be obtained (Jones and Zuber, 1975). Flow parameters such as void fractions, liquid hold ups, density counts etc. can be used to determine the flow regime of a flow (Zhou, Kaasa and Aamo, 2008). This signal fluctuations could be used to yield the probabilistic density function (PDF) which enables that at any point in time the signal would be identified with a value within a set range.

2.2.4 Parameters influencing flow patterns in pipes

Several process parameters, configurations or even transport properties, determine the geometric distribution of the phases in the pipeline. Subject to the explicit system configuration and operating conditions, flow pattern may happen at numerous diverse symmetrical positions within the offshore upstream production system. For this study however flow regime maps will be assessed considering the parameters discussed below.

2.2.4.1 Number of phases

The number of phases and the properties of the phases such as the density difference, viscosity and even surface tension primarily contribute to the flow regime observed in different flow systems. Different phases dictate the superficial velocities which are the main parameters deployed to determine the flow pattern in multiphase flows.

Superficial velocity is a hypothetical velocity which is determined as if the flowing fluid was just of a particular phase. The number of phases also dictates the slippage between the different phases involved.

2.2.4.2 Flow geometry

The flow geometry of the pipeline system has a tremendous influence on the flow regime observed in each flow system. However the flow geometry is dependent on the water current gradient which dictates the riser shape (configuration), the choice of riser and the position of the riser limb. This flow geometry dictates the instabilities at the interface of gas and liquids. Figure 2-3 shows some riser configuration existing in the oil and gas industry. These have different and unique characteristics in terms of the geometry scales which relate to the volume fractions and velocity ratio amongst the phases and the associated pipe roughness. The focus of this research is not on the pipeline configurations existing in the industry hence much details would not be given. Basically, factors such as the topography of the seabed, water wave current and depth of the sea influence the flow geometry used for a production system.

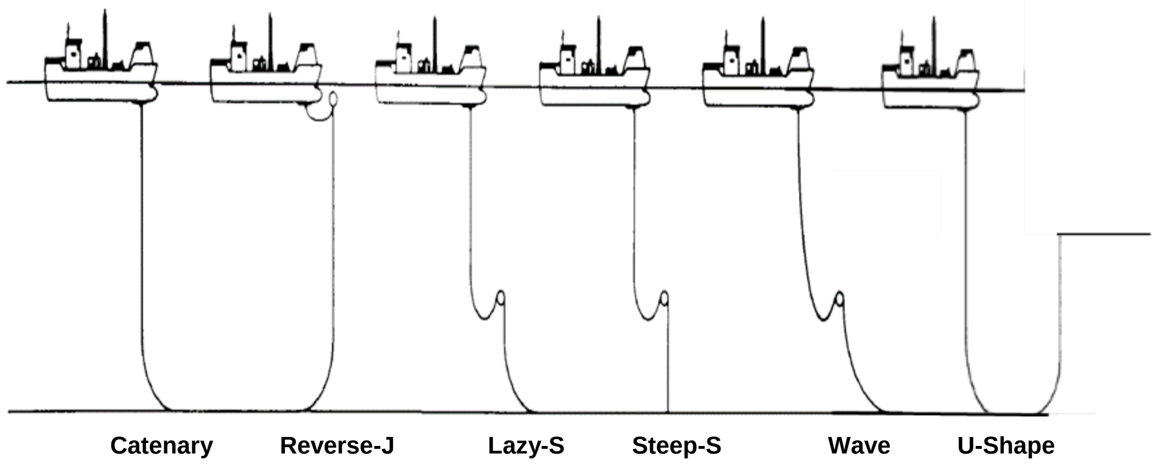


Figure 2-3 Riser configuration in the oil and gas industry

2.2.4.3 Flow conditions

Flow conditions depict the fluid concentrations and the fluid fractions which in effect result in the flow regimes observed in different flow loops. Fluid fractions determines the gas void fractions and liquid hold ups. This shows the flow and pressure fluctuations seen in a system. The conditional flow of a particular system translates to what flow regime is observed.

2.2.5 **Flow pattern detection through pipelines**

Generally flow behavior or characteristics could either be observed or predicted. Relevant dynamic theories have been used to determine the occurrence of several flow patterns by using some flow parameter or even operating conditions. These can either be empirical or not but however have their own constraints. This section addresses some non-empirical and empirical means of flow pattern detection observed in the literature.

2.2.5.1 Non empirical means of flow pattern detection

Simple criteria could be used to predict flow patterns other than any other means. For instance a criterion which is widely accepted and of great essence to the oil and gas industry is the formation of stratified flow in horizontal pipelines introduced by Taitel and Dukler, (1976) together with the condition stated in Schmidt, Brill and Beggs, (1980) reiterates that severe slugging could occur when there is stratified flow through the horizontal section of the pipeline system. Similarly, some other widely used criteria and constraint will be discussed in the following section.

Boe criterion

Boe's criterion was based on the observations and deductions made from Schmidt, Brill and Beggs, (1980). These authors stated that the riserbase gas pressure accumulation rate must be larger than in the pipeline section, for severe slugging to occur in the riser. However Boe's criterion assumed a constant liquid and gas flowrate at the inlet of the system, a gas mass balance in the pipeline and a pressure balance over the riser. Again he assumed that the flow should be stable and steady if the liquid column is stable. Mathematically, the Boe's criterion (Pedersen, Durdevic and Yang, 2015) is given as:

$$U_l^s \geq \left[\frac{P_p}{\rho_l g(1-\epsilon)\ell sin(\alpha)}\right] U_g^s \qquad \textbf{(2-5)}$$

where $\mathrm{U}_{\mathrm{g}}^{\mathrm{s}}$ and U_l^s are the superficial gas and liquid respectively at the inlet of the pipe, P_p is the riserbase pressure, α is the angle of inclination for the riser, ϵ is the liquid holdup in the riser, ℓ is the length of the horizontal pipe.

Taitel criterion

Taitel et al., (1990) on the basis of the assumptions made by Boe, stated that systems could still experience some oscillations or fluctuations even though the liquid is stable. He also concluded that for some high flowrates where the Boe's criterion predicted severe slugging behavior, the system was actually stable. Taitel's criterion is a slug criterion which outlines a correlation back pressure in the riser and holdup pressure of gas at the riserbase. Mathematically, Taitel's criterion (Pedersen, Durdevic and Yang, 2015) is given as;

$$\frac{P_s}{P_o} > \frac{\frac{\epsilon\ell + L}{\epsilon^i} - h}{\frac{P_o}{\theta g \rho_l}} \qquad \textbf{(2-6)}$$

where P_o is the atmospheric pressure, P_s is the pressure the system needs to overcome for production to proceed (back pressure) which is related to the separator pressure, ρ_l is the combined liquid density, ℓ is the length of horizontal pipeline, g is the acceleration due to gravity, L is length of pipe prior to the mixing point, ϵ is the liquid holdup in the system, ϵ^i is the void fraction of the Taylor bubble which enters the riser and its value is 0.9 for vertical riser systems, θ is referred to as the liquid holdup index which is also given by $\theta = 1 - \frac{\mathrm{U_g^s}}{U_t}$ where U_t is the Taylor bubble velocity and $\mathrm{U_g^s}$ is the gas superficial velocity

Jansen criterion

To extend Boe's criterion to include slug elimination methods, Jansen, Shoham and Taitel, (1996) expounded on the Taitel's criterion to include artificial gas lifting. Jansen assumed an injection of gas at the riser base at a constant steady rate. Mathematically, Jansen criterion (Pedersen, Durdevic and Yang, 2015) is given as;

$$\frac{P_s}{P_o} > \frac{\frac{\epsilon_{gr} L}{\epsilon^i} - h}{\frac{P_o}{\epsilon_{gr} g \rho_l}} \qquad \textbf{(2-7)}$$

But $\epsilon_{gr} = 1 - \frac{\mathrm{U_g^s}}{U_{bubble}}$ and $U_{bubble} = C_o U_s + U_d$

where U_g^s is the superficial gas velocity existing in the riser, U_s is mixture superficial velocity, U_d is the bubble drift velocity, U_{bubble} is the bubble velocity and C_o is a constant.

The use of these criteria could offer strategies for real life uses such as process design and monitoring, controller design and even system analysis and their likes (Yang, Stigkær and Løhndorf, 2013; Yang, Pedersen and Durdevic, 2014; Hua et al., 2017;). However, the use of all these criteria greatly rely upon the validity of the assumptions made in relation to the system in question. Again, they all depend solely on measurable signals or parameters which might not be necessarily available, hence an alternative approach using empirical means could be much appreciated or even using signal and data analysis.

2.2.5.2 Empirical means of flow pattern detection

Empirically flow patterns could be detected using flow regime maps which are based on the development of the co-current gas-liquid flow models to forecast two phase flow conduct call for information of flow patterns in the pipe for all possible operating conditions. For such facts flow regime maps have been established. This is basically a representation of results from interpretations of flow patterns which are plotted on a graph. The axes could be represented by the flow rates of the two phases; mass flux or even mass fraction.

Graphical representations known as flow regime maps which categorize the flow regime ranges for different phase combinations in pipelines are developed to distinguish by transitional lines and to classify the above described flow patterns. The boundaries between various flow patterns in a flow regime map occur because a regime tends to be changed as the boundary is approached and growth of this instability causes shift to another flow pattern. Studies of these flow patterns and other empirical laws (Wallis and Dobson, 1973; Weisman and Kang, 1981; Açikgöz, França and Lahey, 1992) have been deployed in the understanding of multiphase flow in general. These flow regimes identified on a flow regime map are differently characterized by the interfaces between the liquid and gas phases and the distribution of the phases to form bonds with their likes (Mandhane, Gregory and Aziz, 1974). The distribution of phases results to the difference in mass, momentum and energy transfer mechanisms and which are attributed by the densities of gas and liquid being different

(Liejin et al., 2009). The flow regime exhibited in every flow condition depends greatly on the orientation of the pipeline (vertical and horizontal pipes). Even though the flow composition could be the same, there could be a different flow pattern shown.

Based on the spectra distribution of the fluctuations in wall pressure, several works have been done on determining the classification and mapping out of flow regimes and the figures below show a flow regime map in transporting gases and liquids in both horizontal and vertical flow pipelines respectively. In spite of the numerous works done on flow regime mapping in literature, there is no generally perfect model fit to represent all flow configurations. This is because of the existence of several flow regimes maps along the same pipeline for a set of given operating conditions.

2.2.5.2.1 Horizontal pipe flow regime maps

Observed flow patterns through horizontal pipeline configurations are recorded with boundaries between the various flow regimes. Researchers such as Mandhane, Gregory and Aziz, (1974); Taitel and Dukler, (1976); Barnea, (1987), produced flow regime maps for flows through horizontal pipeline systems using both theoretical and experimental approaches. There exist some contradictions between the above mentioned maps even though some have similar or same experimental conditions. This could possibly be attributed to the means of observation (visual observation) or even not enough data points in the case of experimentally determined maps. An argument on theoretical maps not being restricted by the experimental conditions spring up but they however assume ideal behaviours hence better in terms of comparism. On the other hand an experimentally determined map closely related to or specified for an experimental setup would be better. Figure 2-4 shows a flow regime map produced through a horizontal flow pipeline (Weisman and Kang, 1981).

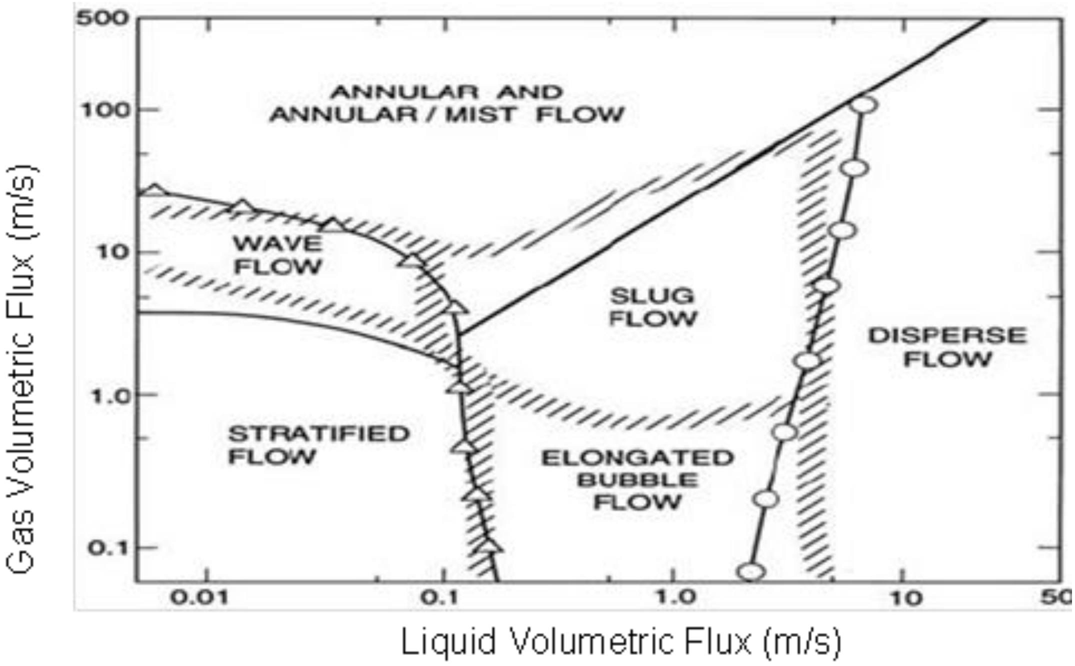


Figure 2-4 Flow regime map for horizontal flow pipelines (Weisman and Kang, 1981)

2.2.5.2.2 Vertical pipe flow regime maps

Several researchers have attempted to develop flow regime maps through vertical pipeline configuration. Theoretically, Barnea, (1987) used a unified model to develop a flow regime map through a vertical downward flow pipeline. Oshinowo and Charles, (1974) and Golan and Stenning, (1969) developed through a vertical downward pipeline, a flow regime map using experiments. Similarly, Taitel and Barnea, (1983) and Barnea, (1987) unified model was used to theoretically develop a flow regime map in an upward vertical flow pipeline.

Oshinowo and Charles, (1974) experimentally developed a flow regime map for flow through a small diameter vertical pipeline system both for upward and downward flow. The operating pressure for the experiment was 0.71 barg with a gas flow rate range of 1-104 kg/h and a 19-3540 kg/h liquid flowrate range. He suggested that the flow patterns depends on volumetric flowrate of gas and other fluid dynamic properties similar to previously developed maps.

Similarly, Hewitt and Roberts, (1969) developed a flow regime map for water/air and water/steam in an upward vertical pipeline over a range of pressure (absolute pressure – 142700 Pa – 542600 Pa) in a small diameter pipeline (31.35 mm). This maps was investigated for a gas flowrate range of 1.36 kg/h – 544.3 kg/h and a liquid flow rate between 226.8 kg/h – 8164.7 kg/h.

Of about 6000 experimental points from pipe diameters in the range of 1.27 to 16.51 cm at a given operating condition where analyzed by (Mandhane, Gregory and Aziz, 1974) to develop a flow regime map with the superficial velocities as the coordinates and that is represented by the map below.

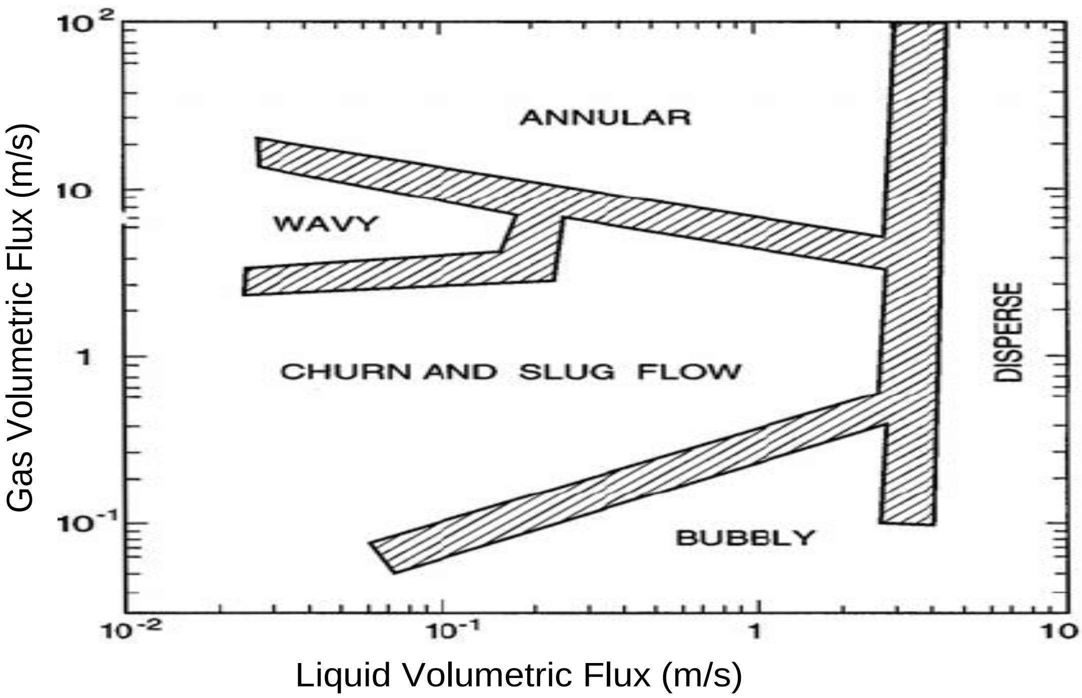


Figure 2-5 Flow pattern map for vertical flow pipelines (Mandhane, Gregory and Aziz, 1974)

One major challenge design engineers face with multiphase systems is the sensitivity of the fluid mass, momentum and energy transfer rate to the topology of the components in the flowing fluid which influences practically the geometric distribution (Abdulkadir, 2011). Thus, the interfacial area available for the phases to exchange mass, momentum or energy is strongly affected by the geometry of the system.

For the point of interest of this research being slug flow investigation and control, flow regime maps were developed in this report to estimate the slug envelope from which further investigations would be carried out. The procedure for identifying the flow pattern was by visual inspection and an analysis of unsteady pressures of the spectral content. Jones and Zuber, (1975) established that for some circumstances, the fluctuations in the volume fractions could also be used.

2.3 Multiphase slug flow

In the oil and gas industry, thus either offshore or onshore, conveyance of hydrocarbons through unitary pipeline is an inevitable and an imperative activity. Water, gas or / and even condensate/oil are conveyed through singular pipelines from the well through to the topside facilities where these fluids are being processed. The industry is faced with several challenges in undertaking this activity of which some includes wax formation, hydrate formation or even irregularity in flowrate / problematic flow regimes (slugging flow) and their likes. The irregularity in the dissimilar phases that coexist in these conduits, sometimes have significant impact on the outlet parameters and equipment of the pipeline rendering it unsafe. Knowledge on slug flow, mode of development and its characteristics is of tremendous relevance in order to design functional and safe systems while avoiding any repetitive problems related with the transportation of multiphase fluids.

With the recent trends of generating marginal gas/condensate and oil fields to existing facilities with subsea tie-backs, explorations moving and the development of field in deep water fields and producing from existing brown fields (towards the end of field life cycle), unstable and fluctuating production or slugging pipeline systems are observed more often.

Slugging is a displeasing situation that troubles production in offshore systems. This is an impairment to the system as a whole and is often faced in brown fields (matured fields). When producing hydrocarbons in multiphase flow, uneven production of the dissimilar fluid phases may occur at the outlet of the pipeline. Large variabilities in gas and/or liquid production can lead to key setbacks on the topside facilities. Inlet separator level becomes challenging to control and aggregate the threat of liquid carry-over into the gas processing systems. Likewise, the pressure surges complementing the liquid surges may impede the gas compression facilities. These production variations are initiated by the fluid composition, geometrical layout and operating conditions of the production system. Especially with pipeline riser systems, the riser may have a large effect on the amplitude of production oscillations. Hence, slug alleviation procedures may accomplish substantial value impact on the business. Understanding the slugging flow concept is paramount to its mitigation techniques therefore this would be discussed in the next section.

2.3.1 **Concept of slugging**

Slug flow regime is a flow pattern that could exist irrespective of the line configuration. This however happens as a result of different velocities existing between the phases in the multiphase flow. Slugging flow in horizontal pipeline is characterized by an elongated gas bubble itinerating at a comparatively low gas velocity to the liquid velocity at the upper section of the pipe. Conversely, the reverse happens in slug flow within vertical pipelines, in that, the elongated gas travels at a higher velocity than the liquid phase. Along a pipeline system, with an individual observing at a stationary point, there exist a track of a structure of slugs of liquid comprehending dispersed bubbles, each looking like a length of bubble, sporadic with units of disjointed flow with long bubbles. The flow still fluctuates or oscillates even with a fixed inlet flow condition (inlet gas and liquid flowrate maintained constant).

At very steady flow rates of both gas and liquid phases, slug flow is however still considered unsteady since slug flow does not mean flow intermittence. Slugging flow is mainly characterized by various system parameters which are mostly predicted with the aid of empirical correlations or mechanistic models (Hill and Wood, 1990). Void fraction, slug frequency and slug length are such common parameters used for such activities. Slug frequency amongst the slug flow characteristics is considered the most crucial due to its impact on several operational problems. Well pressure fluctuation, downstream facility flooding, pipeline corrosion and structural instabilities are mostly influenced by slug frequency. The higher the slug frequency, the much impact it has on the operation of the system. Slug frequency is defined as the number of slugs passing at a reference point over a period of time. At a much higher liquid flowrate and a relatively very low gas flow rate (Ogazi, 2011), it has been proven by experiment that along the pipeline slug frequency reduces. Conversely, the reverse happens for a very low liquid flow rate and a very high gas flow rate (Hill et al., 1996). The main concern of this study, is to deal with the slug flow development along a pipeline riser system configuration.

2.3.2 **Slug flow phenomena**

Slugging in pipelines refers to the irregularity of gas and liquid phases within a cross section of a pipeline. During the transportation of gas and liquid phases in the oil and gas industry, there exist an uneven distribution of gas and liquid thereby producing surges of gas. Lumps of the liquid in the multiphase flow follows the gas surges causing varying flows in the pipeline. The presence of slug flow can generate difficulties for both design engineers and even operators. Several factors affect the extent of slugging in flowlines and these dictates the degree of severity. Such factors include the production rates (liquids and gas rate), the topography and orientation of the flowline and the flowline pressures associated with the system.

There are some great benefits associated with the removal or reduction of slug as a flow assurance challenge. On the economic side of production, the recovery of oil would be enhanced and there is a possibility of an increase and accelerated production rate which translates to huge financial boost for oil and gas production. Again, regularity of flow either by reduction or removal of slug would ensure that flow into the separator becomes stable hence a reduction in the frequency of shutdowns thereby making the system safe to operate. Also reducing the irregularities of flow would enhance the reduction in size of new production facilities such that there would not be a need for slug catchers and over specification of production system capacity since there is a steady flow. Finally, minimising the irregularities of flow in flowlines would aid in the extension of production life of fields due to reduced pipeline pressures (Taitel et al., 1990; Hassanein and Fairhurst, 1998; Aamo et al., 2005; Eikrem, Aamo and Foss, 2006; Di Meglio et al., 2012)

2.3.2.1 Types of slug

Slug flow are grouped into several forms and are characterized by the mechanism of formation. Slug types have been given unique names in literature which have been accepted generally by the industry. Below are some definition of slug types, mode of occurrence and some distinct features about them.

Hydrodynamic slugging

The liquid build up in this type of slug are often short but appears frequently. Hydrodynamic slugs normally occur in horizontal part of flow lines and can sometimes occur in wells and risers. It happens as a results of the differences in the fluid (gas and liquid) velocities. Hydrodynamic slug flow in most cases is well handled by the inlet separator since the quantity of liquid in each slug is relatively small compared to the available space of the separator thereby posing less problems.

Riser slugging

Riser slugging occurs usually when the flowrate is low. This often happens in matured or wells approaching the end of their field lifetime thereby a drastic reduction of the flow line pressure due to a low well pressure. Riser slugging also known as severe slugging is mostly induced by the presence of riser systems and could appear at a riser part due to significant gravity influence (Di-Meglio et al., 2012; Jahanshahi, Skogestad and Grtli, 2013).

Ideally severe slugging condition in pipeline systems befall when there exist a dip in the pipeline and usually followed by a riser pipeline system. This was proposed by Schmidt, Brill and Beggs, (1980). It stated that three conditions are needed to be satisfied for severe slugging to occur, namely; an inclined (downward configuration) pipeline system, stratified flow in the horizontal section of the pipeline and the rate of hydrostatic head accumulation greater than the increase in rate of the pipeline gas pressure at the base of the riser system. This was confirmed by Pots, Bromilow and Konijn, (1987). Severe slugging regime is mainly characterized by a large fluctuating flow, which often features an alternating flow of both gas and liquids but in an irregular manner. Severe slugging occurs in a cyclic manner with slug formation; where liquid accumulates at the riserbase hence blocks the flow of through to the top till the riser gets filled with the liquid phase resulting to the highest pressure in the pipe, slug production; where slug flows out of the riser system, gas blowout; where the gas pressure exceeds the hydrostatic pressure hence gas comes out from the top and liquid fall-back; preceding the gas blowout the liquid falls back on the sides of the walls of the riser and initiates the whole process again, being its features as shown in Figure 2-6. Severe slugging can lead to major disruptions on the topside facilities and can reduce life of the production systems due to high or fluctuating pipeline pressure (Ogazi et al., 2010).

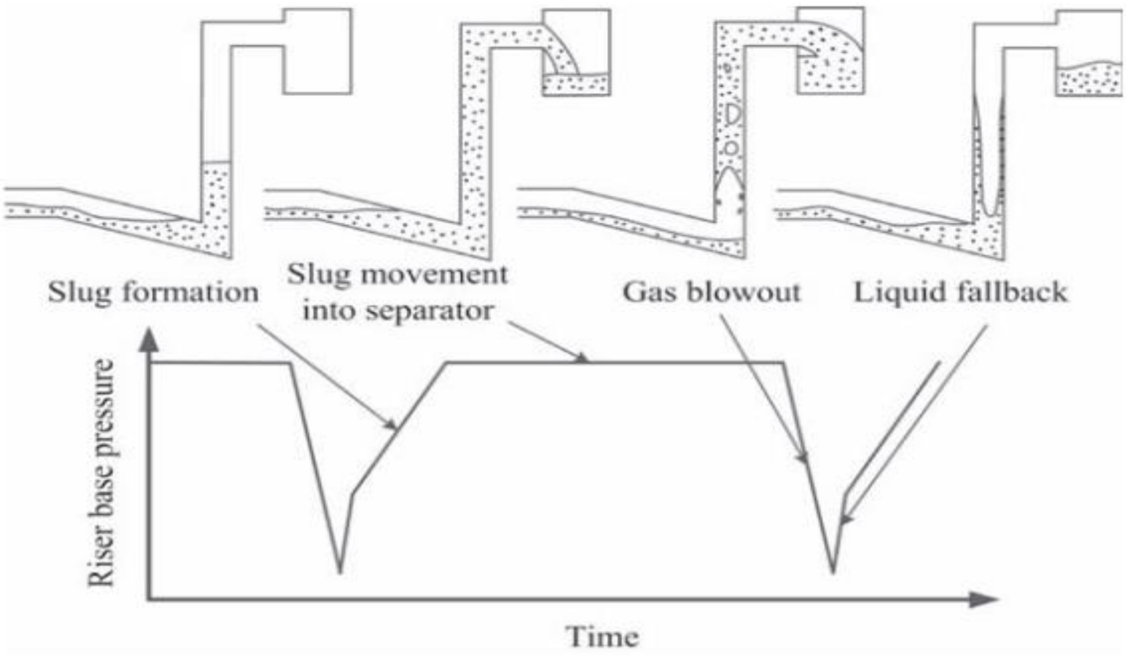


Figure 2-6 Mechanism of severe slugging (Han and Guo, 2015)

Terrain slugging

Terrain slugging is similar to riser slugging or can even be classified as a form of riser slugging however the line blockage by the liquid is as a result of the inclination in the pipe being caused by the terrain. This is as a result of irregular nature of the seabed surface. Thus a terrain slugging may perhaps happen in transport pipelines owing to the seafloor altitudes (Ogazi, 2011; Jansen, Shoham and Taitel, 1996). This however makes liquids to accumulate at the lowest point causing blockage of the gas.

Transient slugging

Transient slugging is a form of slug which is induced into the system as a result of changes in operations or an instability at the interface of the gas and liquid. Different operational changes such as sweep-out and accumulation of liquids from the dip in the pipeline elevation profile upstream of the riser, ramp up, start-up and their likes could cause transient slugging in pipeline systems (Sivertsen, Storkaas and Skogestad, 2010). Transient slugs could appear in diverse sizes and can result in

- Trips due to high or low separator level
- Reduced operating envelope to accommodate variabilities hence resulting in production delay
- Stoppage of compression train due to liquid carry-over or pressure surges

- Unstable topside processing — large periodic variations in both liquid and gas rates at the topside
- Amplified strain on equipment reliability

Pig induced slugging

Pig induced slugging is a form of operation induced slugging which happens during shut downs where pig is sent through the pipe (pigging). The liquid arising from the push out in the pipeline is what is classified as pig induced slug. This form of slug happens outside production periods and can easily be regulated.

2.3.2.2 Pressure – flow relationship

Flow in pipelines exhibit a pressure drop across due to kinetic losses as a result of friction. This is evident in that the entry pressure varies from the pressure at the exit of the conduit. Varying pressure drop is observed for flows with different characteristics. The higher the disturbance in a system the higher the pressure drop observed.

A system with slugging condition has a much higher pressure drop magnitude than a steady flow or even a stratified flow. Flows with higher slug frequencies have higher pressure drops. Physically, slug frequency could be said to be directly proportional to the pressure drop across a conduit.

2.3.3 Effects of slug

Slugging as a flow assurance challenge has undesirable effects on the oil and gas production and the industry as a whole. Due to the oscillatory nature of pressures in the system as a result of slug flow, wear on the processing equipment may result. This would in effect reduce the reliability of the process equipment and the lifetime of the production system as a whole. This causes system fittings such as valves, seals and their likes to wear quickly. The resulting unreliability of the equipment would increase the maintenance cost and the cost of production on a whole relative to an even flow of gas and liquid through the system. On a whole, slug flow increases the probability of failure of the pipeline system.

Again reduction of performance of the well could be observed due to the fluctuating and oscillating pressures of the system. This is more seen in systems with severe slugging conditions. For gas lifted wells significant drop of the gas lift efficiency is observed. Also gas may be accompanied with liquids into the compressors due to the varying gas flowrates resulting in varying separator pressure. There may be poor separation and in some cases separator flooding due to the effect of slugging on the inlet liquid level of the separator.

Furthermore, slug flow causes the system to produce at a lower operating capacity relative to the system design capacity. Also flaring of gases could be associated with the system since there is varying gas flow rate. This as per regulation is not accepted thereby causing huge financial impact on production.

2.4 Slug mitigation

Several researchers have introduced numerous methods to attenuate severe slugging flow to prevent the negative repercussions associated with it. The interest of the oil and gas industry is to subdue, eradicate, control or tame slug flow in pipeline systems. Based on theoretical, experimental works and real field studies (live field) numerous slug mitigation and control have been proposed and illustrated in the literature. Topside choking, riserbase gas injection, multiphase flow homogenizer introduction, subsea processing and separation, active control measures, gas re-injection by means of re-routing, design modification and introduction of slug catchers are all slug mitigation techniques introduced in literature. Broadly, these attempted severe slugging mitigation techniques by the numerous researchers have been categorised as either passive or active slug mitigation technique. The variation between both techniques is established as whether the mitigation procedure is introduced through an external influence or otherwise. Some established techniques have already been tested and deployed on real fields whiles others are still at the developing stage. Schmidt, Doty and Dutta-Roy (1985) established the conditions for which severe slugging occurs in pipeline riser systems as;

- Flow prior to the riser pipeline system should be operating in a stratified flow regime
- The topography of the pipeline system must have a low point where liquid could accumulate and block flow usually at the base of the riser
- The ratio between the gas compression in the horizontal and the pressure hydrostatic head increment in the riser should be less than unity. Thus, low inlet

liquid and gas flowrate so that the growth rate of the hydrostatic pressure at the riserbase is greater than that of the gas pressure in the pipeline

Hence with these established conditions, many of the mitigation techniques seek to address any of these conditions. Some of these works aimed at only taming slugging flow, whilst others considered the profitability margin of the production system on a whole. Some work from literature produced very great ideas to dealing with slug flow. They either addressed slug flow by either containing, managing or even accepting it. For the sake of this work, the various works from literature classified under both passive and active slug mitigation techniques would be outlined next to enlighten the advancements achieved over the past years in the quest to mitigate slugging flow.

2.4.1 Passive slug mitigation methods

Passive slug mitigation methods are those methods that usually do not require external interference thus, they take the form of mostly changes to system design. Techniques such as the design of slug catchers, self-gas lifting thus re-routing the gas in the pipeline to the riser, flow conditioning by means of modifying the flow regime, design of dual risers or even reduction in the pipeline diameter at a point in the system (use of Venturi) are all forms of passive mitigation techniques. These helps to suppress slugging flow hence produces a stable flow and maintains a safe operation within offshore production systems. Slug suppressions are mainly as a result of changes to the flow conditions or adjustment to the outlets of the riser system hence moving the production out of the slugging region. Passive slug mitigation techniques are mostly less flexible since they cannot be adjusted once they are installed. Positively, they are less expensive in terms of running (less resources to run) and installation (less equipment) cost and could collaboratively be used in conjunction with other slug mitigation techniques.

2.4.1.1 Flow conditioning as a passive slug mitigation method

Slugging flow mitigation by means of flow conditioning as established by many researchers has been the installation of equipment or components either at the base of the pipeline riser system or even at the top to alter the flow pattern or regime. As discussed in previous sections, one of the prerequisite for severe slugging to occur as established by. Schmidt, Doty and Dutta-Roy (1985) is to have a stratified flow prior to the base of the riser and also relatively low flowrate through the pipeline so there is no enough energy to transport the flow to the top, hence this techniques seek to tackle these factors. Flow conditioners aimed at changing the flow regime to a more operational friendly condition.

A helical pipe section was used to examine a singular (liquid) and two-phase (gas/liquid) flow behaviour through a pipeline system in Adedigba, (2007). Reinforced flexible pipe of diameter 50 mm was used to produce the helical pipe section used in this experiment by enfolding it around a 19 mm steel pipe to produce a low amplitude helical (less than 0.5 amplitude to diameter ratio). The configuration of the helical that was used for the study is shown in Figure 2-7.

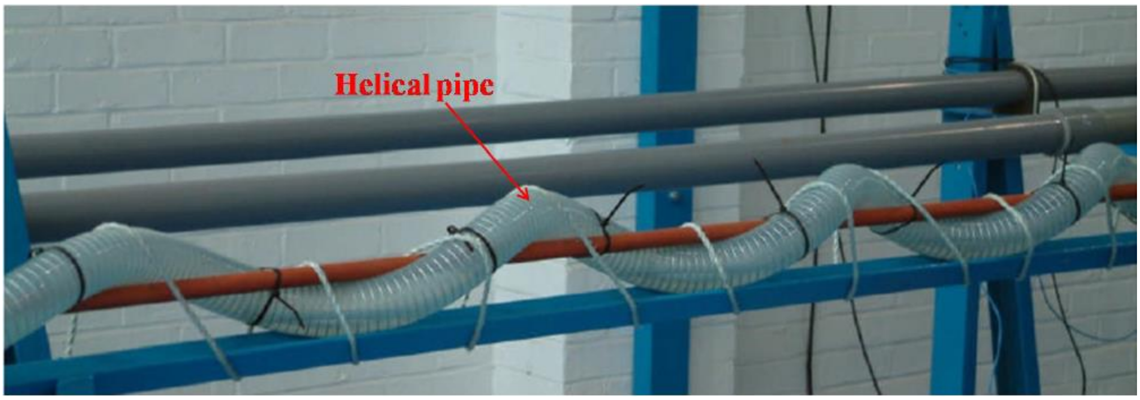


Figure 2-7 Low amplitude helical pipe (Adedigba, 2007)

It was established from the experiments that even though for some superficial liquid and gas velocities that exhibited stratified or slugging flow prior to the helical pipe, a bubbly flow existed in the helical pipe and beyond. These observations proved that the helical pipe had a slug mitigation potential when installed upstream of the riserbase. The benefits derived from the use of the helical pipe included a reduction in the slugging flow region hence a reduction in the amplitude of oscillation in pressure and flow in most flow conditions that exhibited severe slugging behaviour.

Shen and Yeung, (2008) on the basis of the benefits derived from the helical pipe introduced the Pseudo Spiral Tube (PST) a novel pipe section. Pipe fittings and standard elbows were constructed to make up the PST to defer from the Small Amplitude Helical Technology (SMAHT) which was patented from the work in Adedigba, (2007). The PST pipe section had a uniform pipe diameter throughout and was dependent on the diameter, the internal angle of elbow and the twist angle between adjacent elbows. Yeung and Cao, (2007) experimented with two PST geometries to assess its severe slugging mitigation potential. The first PST geometry was constructed from seven 90 degree elbows twisted at an angle of 180 degree forming a radius to diameter ratio of 2.2. While the second PST pipe section yielded from 26 standard 45 degree elbows twisted at an angle of 90 degree forming a radius to diameter ratio of 1.5. A pictorial view of these PST pipe sections is showed in Figure 2-8.

a) PST1

b) PST2

Figure 2-8 PST pipe sections used by Yeung and Cao, (2007)

The first PST mimics a wavy pipe section whiles that of the second one a helical pipe section. Conclusions drawn from this work showed that the wavy shaped PST had a greater slug mitigation potential relative to the helical shaped when both were installed upstream the riser base.

Furthermore (Xing, 2011; Xing et al., 2013a, 2013b, 2014; Xing and Yeung, 2010) continually investigated the PST (wavy pipe configuration) for severe slugging through both experiments and modelling. Different positions on the horizontal pipe prior to the riserbase was assessed by Xing, (2011). It was concluded that the wavy pipe performed best by accelerating the gas flow towards the riserbase and enhances mixing the two phase flow when installed with its outlet directed towards the upstream of the riserbase. Thus, the location and position of the wavy pipe in the pipeline had significant impact on the mitigating tendencies. Again, there existed a lower average differential pressure across the riser system with the wavy pipe relative to that without the wavy pipe section. Also, the wavy pipe section significantly reduced the slug length and initiated bubble penetration stage by accelerating the gas flow rate at the base. Models developed for the system with the wavy pipe showed that the slug length reduced with increasing amplitude of the wavy pipe. Similarly, the slug length reduced with increasing length of the wavy pipe. However Xing, (2011) and Xing et al., (2014) through modelling the use of the wavy pipe for slugging flow mitigation using computational fluid dynamics tool and experiments deduced that even though the device aids the penetration of the gas flow into the slug body hence reducing the fluid density, the slugging flow behaviour reappears in the system just few meters beyond the wavy pipe.

Brasjen et al., (2013), experimented the severe slugging mitigation potential of four different intrusive components. Different placement positions where assessed for all four mixing components shown in Figure 2-9 (choke, swirl, mixer and perforated liners). A 16 % reduction in the pressure fluctuation was achieved when the component was placed close to the exit of the pipeline. However, the pressure drop across the entire system increased significantly which could potentially cause a lost in production. Also, these component could potentially cause a restriction in some operational activities such as pigging or might even be lost in operation.

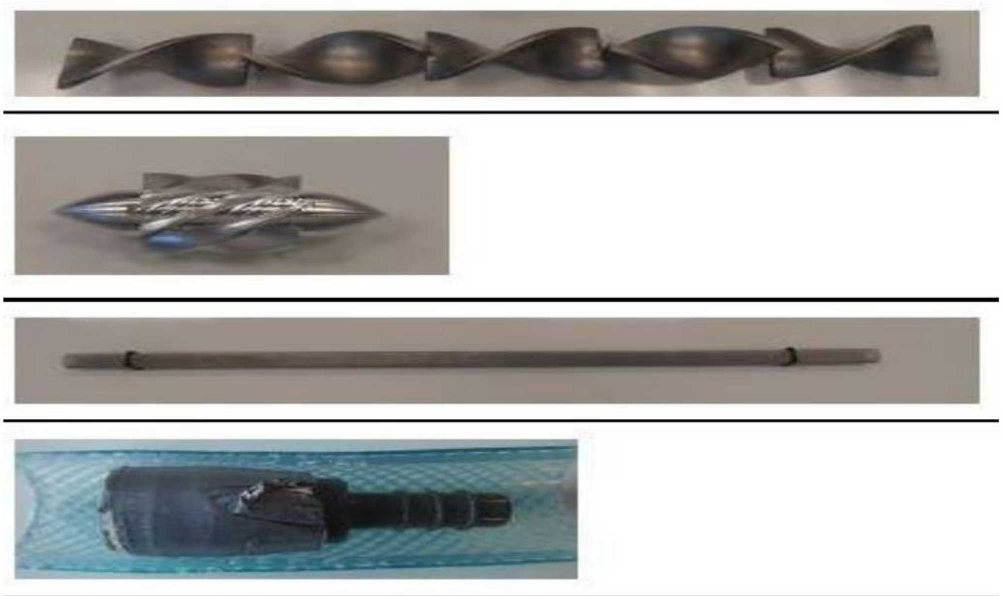


Figure 2-9 Intrusive mixing device for slug control (Brasjen et al., 2013)

A retrofit solution to solving the severe slugging condition in multiphase flows was proposed by Wyllie and Brackenridge, (1994). A small diameter pipe was inserted into the riser to create an annulus that can be deployed for gas injection. Due to the intrusive nature of the method, this technique poses a threat to some operations such as pigging since it can restrict the flow. This constraint led to a modification of the technique in Wyllie, (1995). He however replaced the small diameter pipe in the riser with a retrievable pipe. This even though solved the situation at hand, however the modified device required a work over operation for retrieving the retrievable pipe.

Some potential benefits and setbacks associated with other non-intrusive flow conditioners have been reported in Almeida and Goncalves, 1999; Schrama and Fernandes, 2005; Makogan and Brook, 2007; Makogon, Estanga and Sarica, 2011. Even though additional work is continuing to unravel the operability concerns, these methods are not freed of the constraint of potential operability concerns and additional pressure drop across the riser pipeline system. This work however assesses the slug mitigation potential of both the wavy pipe and spiral pipe (PST) configurations when installed at the top of the pipeline riser configuration.

2.4.1.2 Slug catchers as a passive slug mitigation technique

Slug catcher have been explored as a severe slugging mitigation method and has been established as one of the most popular means of mitigating slugging flow. Slug catchers are large vessels in a form of a separator (liquid / gas separator) and often installed at the topside of the processing facilities, mostly the first component of the train. Slug catchers could be categorised as finger-type, vessel-type or parking-loop with regards to its structure. The primary function of the slug catcher is to serve as a buffer volume to accommodate the changes in the upcoming gas and liquid flowrates whiles preventing unexpected overflow of the system.

Determining its size is critical to optimal operation. Focus has been placed on processing facilities capabilities of handling liquids in assessing slugging situation. In general, a typical design tactic for slug handling has been to size inlet separators to handle the largest projected liquid slug capacity. Typically, motivated by either terrain produced liquid surges in the course of production or even in transient set-ups such as blowdown, ramp up, or start up. The fundamental purpose of a slug catcher is to remove free gas from the liquid phase and supply a relatively constant flow of liquid to the rest of the processing facility. Schmidt, Z, Doty, R. D, Dutta-Roy, (1985) and Schmidt, Brill and Beggs, (1980) presented that slug catchers are often designed much larger than their projected volume due to the fear of severe slugging challenges which in effect is not economical. However, installing a slug catcher is not a real solution to counteract slug flow, but rather a way of accepting the problem. In that sense it is an expensive solution to the problem. The quality of the solution is dependent on whether the slug catcher is sized correctly which essentially has to be determined in the design stage. At that time the exact size of the slugs are not known so it has to be designed for the worst case scenario which is an unnecessarily expensive solution to the problem as well as not always appropriate with respect to space in an offshore platform. Slug catchers are expensive and are only supplied at the design stage that means dependent on a good estimate of slug size. The downside on slug catchers for slug mitigation has been on the way to estimate the slug size from which the buffer size could also be projected.

To resolve the problem of oversizing, Miyoshi, Doty and Schmidt, (1988) developed models for quantifying the critical parameters aiding the performance of a slug catcher. Even though the models achieved some degree quantification for the slug catcher in terms of sizing, the practice of oversizing is yet to be side-lined due to the current trend of production in deeper sea levels. Slug catcher used in offshore fields is less attractive due to space constraint and its inability to deal with gas surges as established by Kovalev et al., (2004) and Kovalev, Cruickshank and Purvis, (2003) which could potentially lead to gas flaring.

2.4.1.3 Flow loop design change as a passive slug mitigation potential

2.4.1.3.1 Dual riser

Kaasa, (1990), proposed a severe slugging elimination method using a second riser to connect the pipeline to the platform. This second riser due to its downward inclination acts similar to a slug catcher and equipped with pressure control valves to deal with the fluctuations in pressure. The position of the second riser is at a point on the pipeline where all the gas or majority of the gas is diverted into it due to the associated stratified nature. The economic viability of a second riser possesses a great challenge for this technique. Again a significant reduction of production could be observed since the original riser would be full of liquid thereby imposing a back pressure to the system while the second riser could also be blocked by entrained liquid that could have found its way through. Similarly, the concept of dual riser was further studied by Prickaerts, Haandrikman and Henkes, (2013), however the motivation of this work was to examine the behaviour or instabilities in the pipeline prior to the dual riser. The schematic of the study done by Prickaerts, Haandrikman and Henkes, (2013) is shown in Figure 2-10.

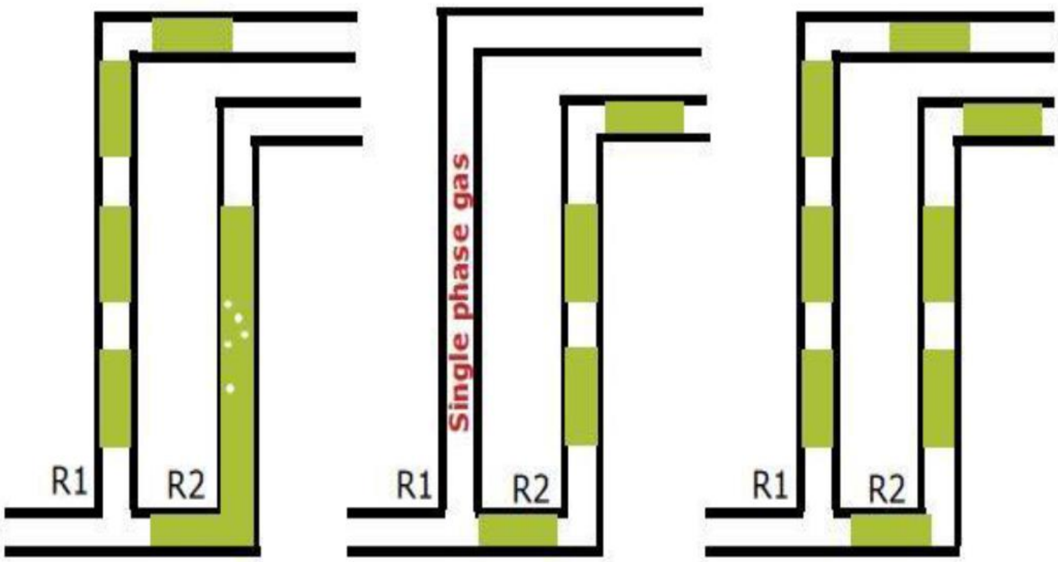


Figure 2-10 Dual riser system for slug mitigation (Prickaerts, Haandrikman and Henkes, 2013)

Song and Kouba, (2000) proposed the use of different pipelines to transport the different phases prior to the topside. From this work, the produced fluids were basically separated subsea hence serving as a slugging flow mitigation technique. On the positive hand, this method was outlined to prevent back pressure in the riser pipeline system hence maximum production capacity. However on the other hand, the cost associated with setting up the system (extra cost of an additional pipeline, separator cost and the pumps and compressors in transporting the separate phases) does not make economic sense. Also, entrained gas phase that could potentially find its way in the liquid pipeline could initiate the slugging process again. Furthermore, the liquid phase could potentially find its way into the gas phase pipeline hence blocking its path causing a build-up of pressure and introducing slugging in the system.

2.4.1.3.2 Additional vessel

McGuinness and Cooke, (1993), presented a pre-separation of fluid phases at a satellite platform before being transported to the main production platform for a Shell operated St Joseph field, Sabah, Malaysia. The case involved an observed severe slugging condition when a new satellite field was introduced on the stream. Expansion and compression of the gas was observed due to the increase in combined diameter. Separation of the liquid and gases into distinct flows bids an effective means of evading the slugging flow difficulty. Furthermore, the transportation of the partly stabilised crude straight through bypassing the low pressure (LP) separator and the atmospheric surge drum, permits an additional decrease in the production manifold pressure and therefore an additional improvement of the well output. Haandrikman et al, (1999) presented an introduction of an additional small pressurized closed vessel upstream of the first stage separator in order to cope and mitigate severe slugging flow. This technique faces however the constraint of space since the platform requires additional space to accommodate this small pressurized closed vessel. It also comes some additional cost at the design stage thereby increasing the capital expenditure (CAPEX).

2.4.1.4 Self-gas lifting as a passive slug mitigation method

Unlike external gas lift which was investigated by Jansen, Shoham and Taitel, (1996) and Jansen, (1990) that required the use of large amount of compressed gas supplied externally through an injection point via separate pipeline systems with the use of a compressor, the self-gas lifting slug mitigation technique uses gas flow from the pipeline upstream the riserbase. This technique however does not require additional pipeline, external gas source and hence compressors. Thus generally, the gas flow needed to mitigate slug in the system comes from the pipeline itself, usually the pipeline upstream of the base of the riser.

Sarica and Tengesdal, 2000; Tengesdal, 2002; Tengesdal, Sarica and Thompson, 2003; Tengesdal, Thompson and Sarica, 2003, 2005 investigated self-gas lifting technique as a slug mitigation method both experimentally and through modelling. Two self-gas lifting procedures were proposed and investigated in Sarica and Tengesdal, (2000). The configurations of the proposed self-gas lifting methods are shown in Figure 2-11.

Principally, this technique works by connecting a small diameter pipe from the inclined section of the horizontal to mid-way into the riser which in effect serves as a by-pass of the riserbase. The small diameter pipe basically transfers gas to above the riserbase resulting in a reduction of pipeline pressure and of the hydrostatic head in the system.

Ideally, this technique is faced with the hurdle of dealing with entrained liquids accompanying the gas flow which could potentially block the gas path when imitated in the field hence experiencing a reoccurring slugging flow behaviour.

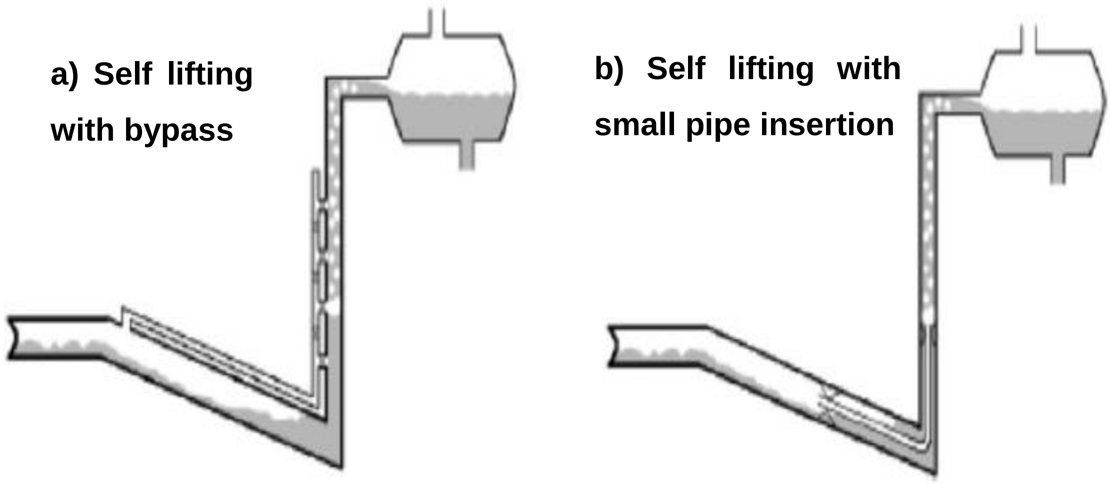


Figure 2-11 Self-gas lifting techniques (Sarica and Tengesdal, 2000)

Again Sarica and Tengesdal, 2000; Tengesdal, 2002 and Tengesdal, Sarica and Thompson, 2003 continued investigating this technique by the use of a small diameter pipe which was inserted within the pipe as shown in Figure 2-11. Conclusions drawn from this suggests that even though an improvement could be observed, the intrusive nature of the small diameter pipe could potentially hinder some operational activities such as pigging.

Generally for already existing installations where the problem of slug flow is being encountered and for compact separation units, the option of design and operational changes as a solution to deal with slugging may not be appropriate hence much better options were explored.

2.4.2 Active slug mitigation methods

Several slugging flow mitigation techniques have been categorized as active slug mitigation methods due the external interferences principally in their execution. Manual choking, external gas lifting and active control methods (automatic choking) are all forms of active slug flow mitigation techniques established in the literature. This section outlines a review of some relevant active slug mitigation methods with much emphasis on active control methods.

2.4.2.1 External gas lift as an active slug mitigation method

External gas lifting has been established as an active slugging flow mitigation method due to the use of an external gas compressor required to supply the gas for injection purposes. External gas lift was deployed to aid severe slugging mitigation by reducing the local mixture density. This technique enhances upward flow of fluid which in effect limits the liquid composition from falling back, one of the stages of severe slugging cycle. Technically, gas lift prevents the accumulation of liquid masses which could be stagnant at the base of the riser and in effect leads to severe slugging formation as established by several researchers (Yocum, 1973; Pots, Bromilow and Konijn, 1987; Hill and Wood, 1990) with the optimal point of injection and the amount of gas required as the drive.

Schmidt, Doty and Dutta-Roy, (1985) indicated that gas injection could be deployed to eliminate severe slugging flow. Large gas volume is required to enhance flow stability, which is a drawback to this technique. For this technique, a compressor is required to pressurize the gas for injection and a flow line which would transport the gas from the top to the riserbase makes this technique economically not feasible.

Pots, Bromilow and Konijn, (1987) also investigated the use of riser base gas injection as a means to suppress slug. Even though the slug cycle and severity were reduced considerably for a gas injection rate approximately half of the throughput, it was observed that it never disappeared completely even at a higher rate of three times the throughput (300 %).

Jansen, (1990) also proposed gas lift as a technique to eliminate severe slugging by combining the gas lifting technique with topside choking activities to stabilise slugging flow conditions. He further developed Quasi-equilibrium models from which it was concluded and established that significant amount of gas flow was required in order to stabilise slugging flow behaviour in the gas lifting technique. A stability criteria developed by Taitel et al., (1990) for choking and gas-lifting combination technique was also developed from this work. Even though the degree of choking and the amount of gas injected reduced significantly hence a reduction on the cost of production while increasing the production rate, the cost in compressing the gas and the amount of gas needed makes this technique not economically viable.

From the industrial point of view, Hill and Wood, (1990) described an experimental test performed on the S.E Forties field to eliminate severe slugging. The strategic adaptation for this experiment was to eliminate severe slugging in the S. E Forties field by ensuring that the flow system operates in the annular flow regime hence preventing the liquids from accumulating at the riserbase. Similarly, more gas was required to stabilise the flow completely. The high rate of gas injection could lead to an increase in pressure loss due to friction and an increase in the Joule Thompson cooling effect which is a drawback even though it is a widely used technique.

Hassanein and Fairhurst, (1998), worked on the homogenization of multiphase flow as a study to help tame slug. A significant amount of energy is consumed in the production process since more gas is introduced to the base of the riser thereby increasing significantly the gas volume fraction (GVF). Again the installation cost of a gas injection system on a whole is high thereby increasing the CAPEX. Again dealing with slug flow with external gas lift as an option that may not be a viable one for older wells since they often have a reduced lifting capacity.

2.4.2.2 Manual choking and back pressure increase

Similarly, manual choking has been established as a form active slugging flow mitigation technique due to the involvement of an operator to manually vary the valve opening during operations till the flow becomes stable. Severe slugging was observed to have been eradicated by increasing the back pressure in the pipeline (Yocum, 1973). The technicality in the increase of back pressure mitigating slug is that, it increases the pressure drop along the pipeline riser system relative to the gas velocity thereby avoiding a spontaneous blow out. This technique however, reduced the flow capacity tremendously. The imposed back pressure was observed to have reduce the capacity of flow that is reduced production even in swallow waters hence projected a worse situation in deep water production system (Tengesdal, Sarica and Thompson, 2003).

Schmidt, Brill and Beggs, (1980) recognized choking as a means to eradicate severe slugging after several experiments from which he concluded on choking as a flow stability possibility. Choking is the increase or decrease of the opening of the platform choke valve with little or no change to the flowrates and pipeline pressure. Reducing the opening of the choke valve at the platform has been the main method to increase the pressure in the line. Taitel, (1986) based on these conclusions derived a theoretical approach to elaborate on the stability capability of choking. Taitel stated that the system could achieve stability and severe slugging flow could be eliminated when the pressure at the outlet of the riser (topside pressure), approximately equal to the separator pressure, is larger than the head provided by the liquid column in the riser system pipeline configuration.

From the industrial point of view, real fields were used by Farghaly, (1987) to prove that severe slugging could be eliminated by choking on Zakum field. The effect of slug on the separator unit was minimised using the pipeline choke after it has been detected from this work. It was seen that severe slugging could be eliminated by increasing the back pressure proportionally to the increase in velocity at the choke. If the acceleration of gas into the riser is stabilised before reaching the choke, the occurrence of steady flow is surely. Slugs may return to form larger slugs at the base of the riser if the choke valve is closed much thus over choking induced slugs or even resulting in an abrupt change in operation which could hypothetically lead to the system becoming unsteady due to the multiphase flow non-linearity characteristics (Ogazi, 2011).

Even though choking has been a proven technique to eliminate severe slugging, undertaking this exercise requires extra care in order to have less increase in back pressure so as not to reduce the capacity of the production. Again, this technique is not suitable for mature wells since they mostly have reduced lifting capacity.

2.4.2.3 Active control methods

Active control based methods as an active slug mitigation technique requires a controller to influence an input element, usually the choke valve to stabilise the unstable slugging system. In the quest to optimize the topside choking technique, which has been a widely accepted method in the industry till date and eliminate the setbacks it poses on production, automation of the topside choking (Zelimir, 1977; Schmidt, Brill and Beggs, 1980; Schmidt, Doty and Dutta-Roy, 1985) as a severe slugging technique was established. The optimization of this process can be traced back to the early 1990s which has yielded the use of an automated choking (Hedne and Linga, 1990; Courbot, 1996; Henriot et al., 1999) and even a combination of choking with other mitigation techniques (Jansen and Shoham, 1991, 1994) due to the flexibility of active slugging flow mitigation technique. Active (automatic) control has been established as one of the most effective means of eliminating severe slugging flow on the field.

Automatic slugging flow control adopts an intuitive mechanism to cope or deal with slug flow. Slug handling using control methods are characterized by the use of the system (pipeline) information to regulate or adjust available degree of freedom such as choke percentage opening, pipeline pressure, separation levels and their likes to suppress and eliminate the effects of slug in the downstream units. The choice of sensors and actuators could be directed by some essential system investigation for instance input-output controllability analysis (Jahanshahi, Skogestad and Helgesen, 2012). This approach detects the slug flow and its magnitude and works by limiting its size. Again it helps to obstruct the adverse effect of the slug on the downstream systems at the production facility especially the separation and compression trains. The automatic choke mostly serves as the manipulating medium based on real time changes in the process variable (control variable) which could potentially be any process variable of good slugging representation.

Several works on active slug control has been done with different slug controllers and slug control systems (either feedback, feedforward or even cascades configurations) which have their own different functionality. Dynamic feedback control encompasses active actuation of the production choke valve. The production choke is moved in agreement with the algorithm of the dynamic feedback control. The primary function of the controllers applied in these cases aims to stabilize the system not by coping with the flow condition in the processing units downstream but by implementing a control mechanism (Havre and Dalsmo, 2001). Active slugging flow control requires minimum modification to already existing facilities which makes it advantageous in comparison to the other mitigation techniques. Hence to attain stable operating conditions, active slug control is recommended and will be discussed briefly.

Jansen and Shoham, (1991, 1994) reported that active control using the control valve at the topside of the riser as the manipulating variable and the riserbase pressure as the control variable could significantly minimise the backpressure posed on the system compared to manual choking technique while rendering the system stable from a severe slugging condition.

Hedne and Linga, (1990) through experiments assessed the performance of riser choking both manually and automatically as a means of comparison. The system was able to achieve stability at a 20 % valve opening for manual choking resulting to about 7 bar pressure drop across the choke valve. On the other hand, the active control yielded a 2.5 bar pressure drop across the choke valve when the controller was activated. A PI (Proportional-Integral) controller was however used in this study. Thus, there was a reduced pressure drop across the choke valve with the controller in action, showing the technique's supremacy over manual choking.

Courbot, (1996) proposed the implementation of an automatic slug control to prevent severe slugging (at low flow rates) in the Dunbar-Alwyn 16 inch multiphase flowline as the case study. He used a PID controller with the riserbase pressure as the control variable and the riser topside choke valve as the manipulating variable to stabilise the slugging flow. The flow through the system was kept stable by keeping the riser base pressure constant with the use of an upstream choke valve of the separator. A parallel control valve to the pipeline choke was used to control the pressure at the base of the riser. The field experience observed from the case proved successful even though there was a slight increase in the riserbase pressure. Potential reduction in production in deep waters may be posed as a constraint for this technique.

A control strategy in a transient multiphase flow simulation software that proved to have a very good performance, reduces significantly when emulated on a test rig as presented in Molyneux, Tait and Kinvig, (2000). For a particular severe slugging condition, the controller could only achieve a 50 % reduction in the pressure oscillation of the system when studied through simulation hence relatively stable. The discrepancies was however attributed to the difference in the predicted transient model and the actual system.

Similarly, Drengstig and Magndal, (2001) using a PI controller, performed severe slugging stability test with the pressure drop across the vertical riser as the control variable and the choke valve as the manipulating variable. The pressure drop across the riser was determined by the difference between the riserbase pressure of the riser and the pressure at the top of the riser. Comparing results obtained from the above study to that when using the riserbase pressure as the control variable, it was concluded that the riserbase pressure is a much desirable control variable. This was however reiterated by Storkaas, (2005) who used control theory to analyse the system and to find the most suitable control variable possible in stabilising the riser pipeline system configuration.

The use of at least a signal from the base of the riser or solely relying on measurement from the seabed, usually the riserbase pressure as seen in all the solutions discussed above, is not always reliable, much expensive and challenging to sustain. To elude using the riserbase pressure as a control variable, several other signals such as the riser outlet pressure, velocities, volumetric flow, inlet pressure, pressure drop over the choke valve and their likes have been assessed through controllability analysis using experiments and simplified models. Great understanding has been achieved for controlling severe slugging flow conditions using several control variable to choose the suitable control variable through extensive studies in literature (Storkaas, 2005; Storkaas and Skogestad, 2007; Sivertsen, Alstad and Skogestad, 2009; Sivertsen, Storkaas and Skogestad, 2010; Jahanshahi, Skogestad and Helgesen, 2012).

In the quest to sway from the use of measurement signals from the seabed, Hollenberg, de Wolf and Meiring, (1995) proposed a topside flow control technique for mitigating severe slugging flow. With the mixture velocity as the control variable and the topside choke valve as the manipulating variable, the stability performance achieved was not convincing. To improve the performance, a small separator was adopted to aid in the separation of phases and quantify the flowrates. This technique yielded a great performance in terms of stabilising the system but however there was an introduction of a significant back pressure into the system at the desirable stable condition.

Storkaas, (2005) performed a controllability analysis assessing several measurement signals. From this work, it was concluded that the riser top pressure solely was not a suitable control variable for stabilising a slugging pipeline riser system, for the reason being that the zeros of the resulting transfer function were positioned in the right half plane of the complex plane. Similarly, both pressure drop over the choke valve and the inlet pressure, solely as control variables were reported as not so suitable for stabilizing slugging flow behaviour in the riser. There was an observed time delay effect for the system using the inlet pressure as the control variable. Again, the riser outlet flowrate followed suit but was affected by low stationary gain. Further investigation (Storkaas, 2005) proved that even though the riser outlet volumetric flowrate showed poor performance when solely used, the performance improved greatly when combined in a cascade control system (used as the secondary controller) for stabilising the system. From the positive results achieved, several other cascade configuration performances were analysed by Sivertsen, Storkaas and Skogestad, (2010) using an experimental approach. The test configuration that was explored included:

- The topside pressure in the outer loop and the volumetric flowrate at the outlet in the inner loop.
- The topside pressure in the outer loop and the density at the outlet in the inner loop.
- The choke valve opening in the outer loop and the volumetric flowrate in the inner loop.
- The choke valve opening in the outer loop and the density at the outlet in the inner loop.

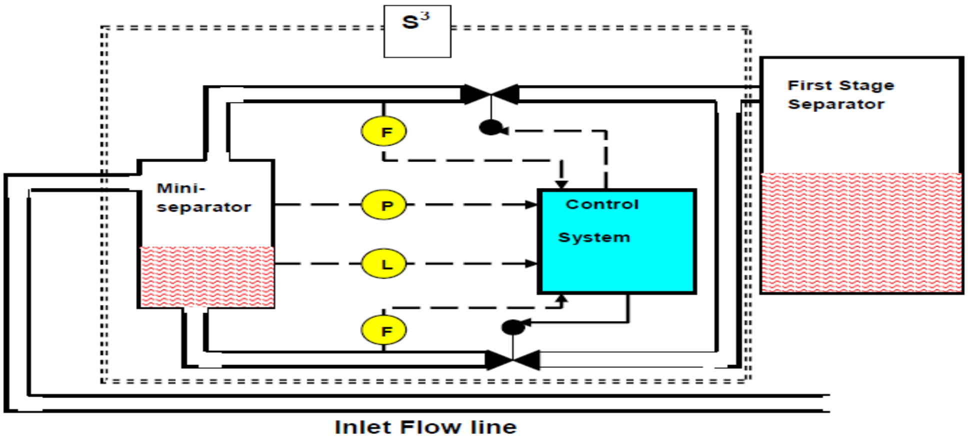


Figure 2-12 Slug suppression system (S3) control scheme (Kovalev, Cruickshank and Purvis, 2003)

The S3 slug suppression system presented in (Kovalev, Cruickshank and Purvis, 2003; Kovalev, Seelen and Haandrikman, 2004) is a system that combines the principle of a slug catcher (mini-separator) placed between the outlet of the riser and the downstream separator with active control for severe slug mitigation. It was reported to show great performance when deployed on numerous field applications. The total volumetric flowrate and liquid level in the mini separator was controlled using the gas outlet and the liquid outlet respectively but it basically functions by separating both gas and liquid phases. From the slugging flow mitigation point of view, the controlled flowrates of liquid and gas whiles keeping the liquid level and separator pressure relatively stable aided slug mitigation. A schematic diagram of the S3 slug suppression system is shown in Figure 2-12.

Kovalev et al., (2004) reported on the technology of T-junction technological liquid gas separation system, one of the best solutions technically to solve the problem of slugging in the oil and gas industry. T-junction technological liquid gas separation system, also known as vessel-less S3 was led into existence due to the expensive nature of the S3. The vessel-less S3 unlike the S3, which was a less-efficient technique, has an added control logic to make it a more-efficient method. It was an improved model of the former suggested S3 technology and a schematic diagram of the technology is shown Figure 2-13. This was accomplished by using a stratifier to decrease the volume of the vessel necessary for the device and a secondary separator in the form of T-junctions. Even though the volume was reduced significantly, the added stratifier, attached T-junctions on a downcomer which might be at an angle can add on as extra cost. Again, there was no existing scheme for sizing the entire system reported hence the T-junctions reliant on the stratifier.

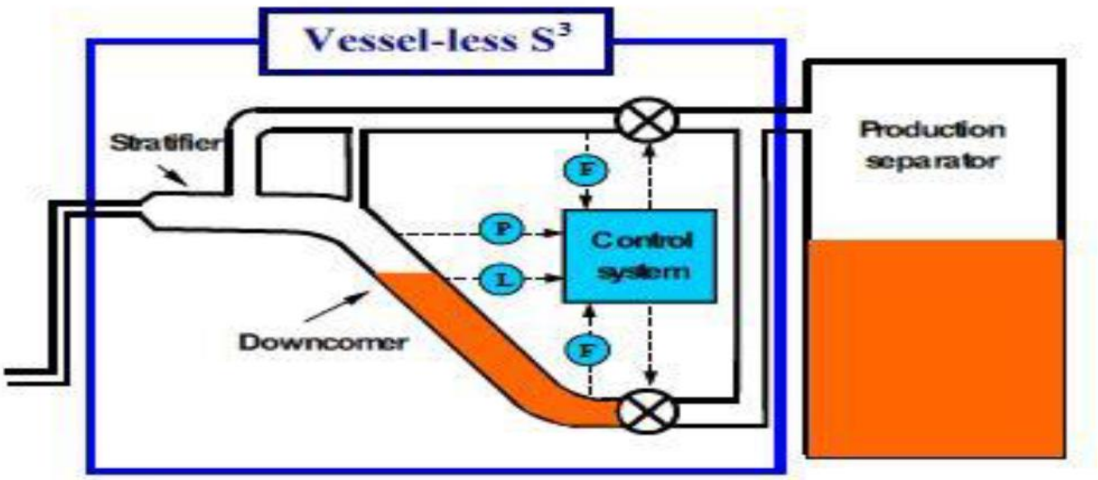


Figure 2-13 Schematic diagram of the Vessel-less S3 (Kovalev, Seelen and Haandrikman, 2004)

From the industrial point of view, Oram and Calvert, (2009) of BP developed and patented a slugging flow control system to stabilize the system at an increased production rate. This work uses a cascade slug control scheme with the primary controller (PI controller) using the riserbase pressure as the control variable and the secondary controller (PD controller) control variable as the differential pressure across the riser. The slug control scheme uses two pressure sensors for slug mitigation, one at the base of the riser and the second one placed at the topside of the riser. Using a differential pressure processor, the difference between the two pressure sensors was used as the control variable. A 10 % increase in production rate was achieved when implemented on the Vallhall production field off the coast of Norway.

ABB and BP jointly developed a slug control scheme for terrain induced slug mitigation as reported in Havre and Dalsmo, (2001). The control scheme deployed in this work is shown in Figure 2-14. The control scheme uses feedback and feedforward algorithm with both the outlet and inlet pressure being the control variables while the manipulating variable is the choke valve at the riser top. An improvement in terms of system stability was reported when this control system was deployed on the BP Hod field. Again the compressor and separator trains were observed to attain less and smoother operations while the system realized a considerable reduction in the inlet pressure which could potentially reflect in the throughput of the system. The positioning of the system, upstream the pipe close to the well, shows the tendencies of the system still experiencing hydrodynamic slugging. Again, time delay could be a major challenge due to the transmission distance (a pipe length of 13 km), hence the reason for the observed occasional slugging on the field as reported.

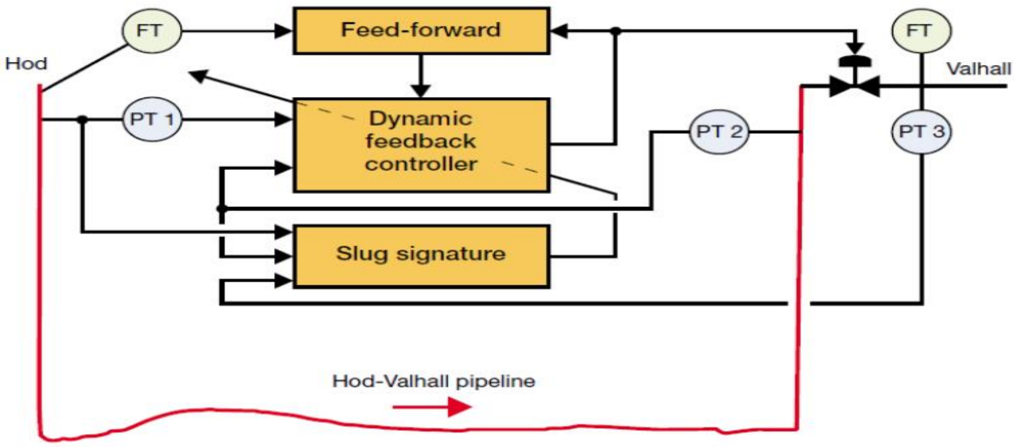


Figure 2-14 Slug control scheme with feedback and feedforward algorithm with the outlet and inlet pressure as the controller input (Havre and Dalsmo, 2001)

The pipeline riser system model developed by Storkaas, (2005) was improved and reported in Ogazi, (2011). A controllability analysis was also performed for the unstable riser pipeline configuration using the improved riser model. Deductions made from this study showed that the topside variables have slug mitigating potential, same as other variables from the seabed, hence can be used to stabilize unstable riser systems when used as controller input. From these deductions, several slug controller algorithms were developed. Ogazi, (2011) also performed slug flow stabilisation using a control valve on gas outlet of the separator instead of the choke valve at the riser top.

To avoid the seatbacks associated with the use of seabed measurement variables or a combination of downstream and upstream measurement signals and with the deductions made from Ogazi, (2011), Cao, Yeung and Lao, (2010b) developed a slug control scheme that uses only topside measurement signals. This control scheme also known as the inferential slug control (ISC) through algebraic scheming combines several topside measurements to produce a single component known as the slug index (SI) which is more sensitive to slugging flow. With the slug index as the control variable and the riser topside choke valve as the manipulating variable, this scheme showed a 10 % increase in production when deployed and tested experimentally on the 4 inch catenary riser in Cranfield University. However the design of the ISC was not outlined since a trial and error technique was used to design the controller parameters hence it is difficult to tune and potentially has weakness in robustness. Similarly Cao, (2011) reported an increased production rate of 9.6 % when tested on a 32 km long flowline leading to a riser of 100 m high at an offshore production platform located in the North Sea. Thus the pressure fluctuations reduced significantly from a magnitude of about 2 bar to 0.5 bar hence encouraging safer production.

Pedersen, Durdevic and Yang, (2014), developed a self-adjusting controller for slug control. The controller basically consisted of a supervisor controller and two other PID controllers. These controllers helped in automatically finding the optimal choke valve opening where slugging was eliminated. Thus, the operating position which yielded maximum production. The challenge faced by most of these techniques is the effective tuning of these controllers which has been addressed by several researchers (Godhavn, Fard and Fuchs, 2005; Jahanshahi et al., 2014).

Ehinmowo and Cao, (2015) proposed a new method for stabilising unstable slug flow which was observed to optimize the slug attenuation potential relative to manual choking. The reported new methodology for slug flow stability analysis from this work showed an optimization in the pressure drop across the valve compared to manual choking. Specifically, an additional 2 % valve opening corresponding to a 3 % reduction in the riserbase pressure translating to an increase in production and throughput of the system. However this method still used the riserbase pressure and the choke valve as the control variable and manipulating variable respectively hence would be still be faced with the challenge of using measurement signals from the seabed. Optimization of this process is necessary in other to have a more robust controller to yield much benefits.

From the literature, slugging in pipeline riser systems could be very challenging and understanding the mechanism of slugging could potentially aid in developing proper techniques that could aid in its mitigation. Again as established from several research, no measurement signal in the quest to develop proper and robust control system is irrelevant. The use of better control systems can minimise the slug control impact on the throughput of the system.

2.4.3 Benefits of slug mitigation

There are several benefits derived from minimizing or eliminating slug as a whole in the oil and gas industry. Recently, many reservoirs have been depleted, over the years there have been tremendous advantages derived from steady flow in production system. Controlling and mitigation of slug flow in pipelines has helped increase the safe operation of systems. Stable flow has helped with the process safety by means of smoothening the behaviour of flow during production, start up and shut in of flow systems.

Flow stability would help to increase the production of liquids. There exist for stable flow systems, less downtimes thereby continuous optimized production and faster start-up times which would in effect help reduce the price of oils translating to an increase focus on the cost of production and its efficiency. Again, production at a stable flow would enhance a more efficient use of manpower. Stable flows make wells behave in a stable manner hence the handling becomes simple thereby more energy could be directed to other duties.

2.5 Inferential slug control for flow stability

Most existing slug control techniques used the riserbase pressure which was faced with the challenge of the readily availability of the measurement at the riserbase and the significant increase in both capital and operational cost as a result of its introduction on the seabed. Cao et al., (2010) proposed a novel severe slug mitigation method using an Inferential Slug Control (ISC) to address the associated challenges with the riserbase and the problems encountered with cascade controller using topside measurements in the oil and gas industry. The input parameters applied for this controller was the measurement signals from the topside facilities only. The ISC was reported to stabilize flow for severe slugging flow condition, for a set of operating conditions in an unstable region (Cao et al., 2010). The principle behind the ISC is the combination of several topside measurements at the topside of the riser through simple algebraic scheming. Already existing facilities provide all these measurement signals since it is already deployed to control other components of the system thereby making its implementation and operation cost significantly less, rendering the control system attractive (Cao et al., 2010).

Kadulski, (2014) used proportional controllers to form a cascade configuration of the exact system which allows the control technique to be implemented in OLGA flow simulator without requiring a second software such as Simulink connection through an object linking and embedding (OLE) for process control server (OPC server). The cascade configuration and the original ISC produced the same output signal which translate to the fact that the control algorithm response was identical. Algebraic calculations were used to obtain a single control variable from several measurement signals for the reason being that, it becomes sensitive to slug flow therefore being the sole origin of alteration in the controlled variable. The control law adopted to determine the valve opening of the ISC is given by;

$$V = V_o + K(W^T Y - R) \tag{2-8}$$

where $\mathbf{V_o}$ is the choke valve nominal value which is predetermined and manually set to a position where the flow becomes stable or within an acceptable range, **K** is the control gain, $W=[k_1,k_2,\ldots,k_{n-1},k_n]T$ is the measurement weights, $Y=[y_1,y_2,\ldots,y_{n-1},y_n]$ is the vector of measurements, $\mathbf{W^T Y}$ is the control variable, **n** is the number of measurements and **R** is the set point of the control variable.

Figure 2-15 shows the layout of the ISC in cascade configuration (Kadulski, 2014). The inputs are created by signals to the proportional controllers connected in series, thereby closing lots of loops in the system. The set point for each of the controllers except the last one is created by the output of each subsequent controller.

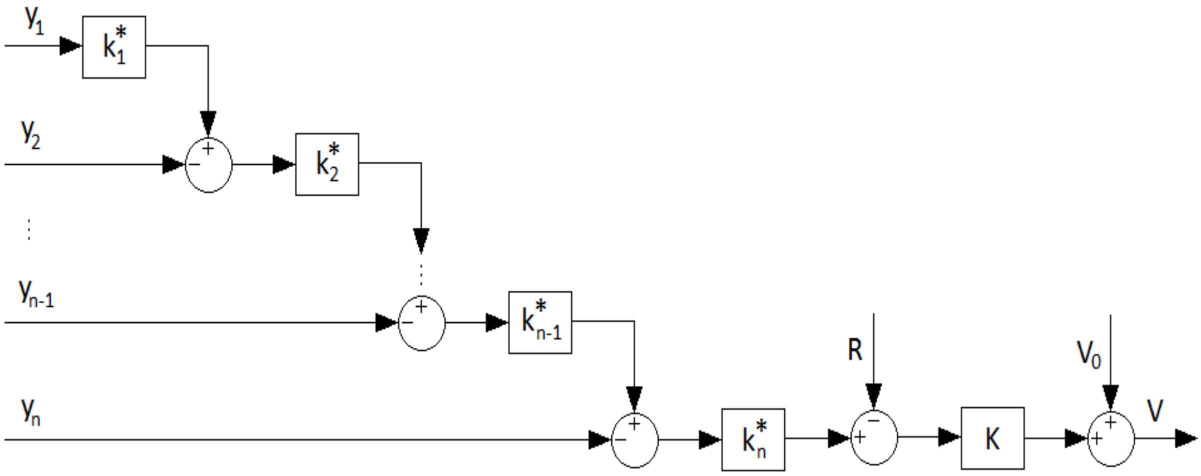


Figure 2-15 Block Diagram of the ISC System in Cascade configuration

Again, using the cascade configuration of the ISC developed by Kadulski, (2014), Kudzia, (2014) proposed the design of the ISC algorithm in a programming environment. This was to enhance the direct implementation of the ISC in tools such as the DeltaV control system. This however eliminated the need to design the controller in a different environment such as Simulink and connecting through OPC server hence reduced the capital, operational cost and reduced the probability of failure due to secondary influences such as hardware or software malfunction. Experimentally, this was tested to assess the controller robustness and performance. A stable flow was achieved for the slugging flow investigated and proved the direct implementation to be of great benefit.

Although the original ISC configuration has undergone several modifications and improvements, there is still more room to explore in terms of the controller robustness and its implementation. This work seeks to advance the ISC technology by proposing a systematic approach in designing the ISC, finding several other measurement signals which could enhance the controller and assessing the slug attenuation potential of this technique on different pipeline riser configurations.

2.6 Summary

This chapter outlined past, on-going and some key procedures related to multiphase flow in the oil and gas industry with special interest in slugging flow. Understanding the slug flow phenomena was of essence which prompted a literature survey on multiphase slug flow. The mechanism and conditions leading to slugging flow was reviewed with an attempt to also enlighten some techniques explored in the industry till date.

Some key slug elimination approaches have been looked into, deliberated on and classified on the broader sense passive slug mitigation technique (self-gas lifting, flow loop change, slug catchers and flow conditioning) and active slug mitigation techniques (external gas lift, manual choking and back pressure increase and active control methods).

The use of PST pipe as by Xing, (2011) for severe slugging flow mitigation was established to be very effective and efficient when installed at the base of the riser. However installing this pipe section at the riser base poses some difficulty on the system in terms of accessibility to the pipe section and the cost involved /associated with maintaining the pipe section fixed at the seabed. The slug mitigation technique of this method would be explored when installed at the topside.

Two key objectives of slug mitigation using active control are to eliminate the unwanted flow pattern and increase production. In the face of the advances made in slug mitigation using active control, the use of the topside measurement have shown great benefits which birth the ISC technology. Vast work has been done to develop the ISC technology for slug control but no systematic design of this controller has been established which leaves the controller robustness in question. With the controller robustness and proper design in place a further development is necessary to improve the economic aspect of the ISC technology. Thus extensive amount of work is still required and this study is committed to addressing the gaps identified. The advancement of the ISC technology on different pipeline riser configurations will be assessed next.

3 HYDRODYNAMIC SLUG FLOW IN U-SHAPE PIPELINE RISER SYSTEM

3.1 Introduction

This chapter gives a general understanding on the dynamic behaviour of the fluids in a 2 inch U-shape pipeline riser system and the impact of the downcomer on the flow behaviour with special interest in unstable flow. In the quest to stabilise unstable slug flow in the U-shape riser, an experimental study of gas-liquid flow mixture is investigated to understand the behaviour of the flow in the riser. Understanding the flow behaviour is of great essence as this aids in the design of real slug control techniques. Several research has been conducted on flow behaviours in either vertical or horizontal pipeline systems. Wallis and Dobson, (1973); Weisman and Kang, (1981); Açikgöz, França and Lahey, (1992); Hurlburt and Hanratty, (2002); Kadri et al., (2009); Krima, Cao and Lao, (2012); did some work in identifying flow regimes in horizontal pipeline systems while Baliño, Burr and Nemoto, (2010); Xing, (2011); Malekzadeh, Henkes and Mudde, (2012) studied flow patterns in vertical riser system. These flow regimes identified on a flow regime map are differently characterized by the interfaces between the liquid and gas phases and the distribution of the phases to form bonds with their likes (Mandhane, Gregory and Aziz, 1974). Several works have looked into the flow regimes and patterns in different pipeline configurations but less attention has been given to the flow behaviour in U-shape riser system which is a form of a platform to platform pipeline layout.

Experimentally, flow patterns observed from the U-shape pipeline riser configuration were used to develop a flow regime map which was then compared to that observed in the literature and similarly to a purely vertical riser with similar pipe diameter (2 inch). Thus, a slug envelope was developed for the U-shape riser to help identify in which regions slugging could occur in the system. With special interest in instabilities in the U-shape riser pipeline configuration, the initiation point of this instabilities was investigated. The procedure adopted for this study will be outlined in subsequent sections. The flow regime identification is presented first with the initiation of unstable flow (slug flow) presented later.

3.1.1 U-shape riser system configuration

The U-shape riser system is a form of a platform to platform pipeline system which can basically be characterised by two vertical pipelines (down comer and riser) and a horizontal section. The U-shape riser can be described as or similar to an extended version of an L-shape riser, purely catenary riser, or even a vertical riser configuration. The flow through the two vertical pipelines are however counter current with the horizontal pipe serving as the link between both. The downward and upward flowing vertical pipeline of the U-shape riser system is known as the downcomer and riser respectively.

The 2” U-shape flow loop used in the experimental study for this work consists of a 10 m vertical downcomer connected to a 40 m horizontal pipeline leading to a 10.5 m vertical riser. The U-shape riser system is made from pipes (stainless steel and plastic) with a uniform internal diameter of approximately 50.4 mm. The flow enters the system through the downcomer and exits through the riser into a vertical two phase separator at the top of a steel frame tower, where gas and liquid are separated. The riser is joined at the top to a purely horizontal pipe (topside) with sections of steel and PVC pipes leading to the 2 phase separator. The total length of the topside is about 3.5 m and it is equipped with a valve (choke valve), 0.3 m from the 2 phase separator where initial separation of liquid and gas takes place prior to the 3 phase horizontal separator.

Figure 3-1 shows a schematic of the three phase facility with the U-shape riser layout shown as the purple trace. The horizontal section of the U-shape riser has no inclination or declination, thus at a 0 degree angle. This serves as the channel through which the fluids are introduced into the riser and it is made from a 2 inch stainless steel schedule 20 pipe. A transparent and Perspex pipe section is fitted in the horizontal and near vertical riserbase section of the U-shape riser respectively to aid in the observation of the dynamics of the fluids through the pipe. In addition, an extra clear pipe, 1.0 meter in length, is fitted in the downcomer (about a meter from the base) to ease flow behaviour observation in the downcomer. Instrumentation positioning and data acquisition means on the purely vertical riser are shown in the appendix section A.2.1.

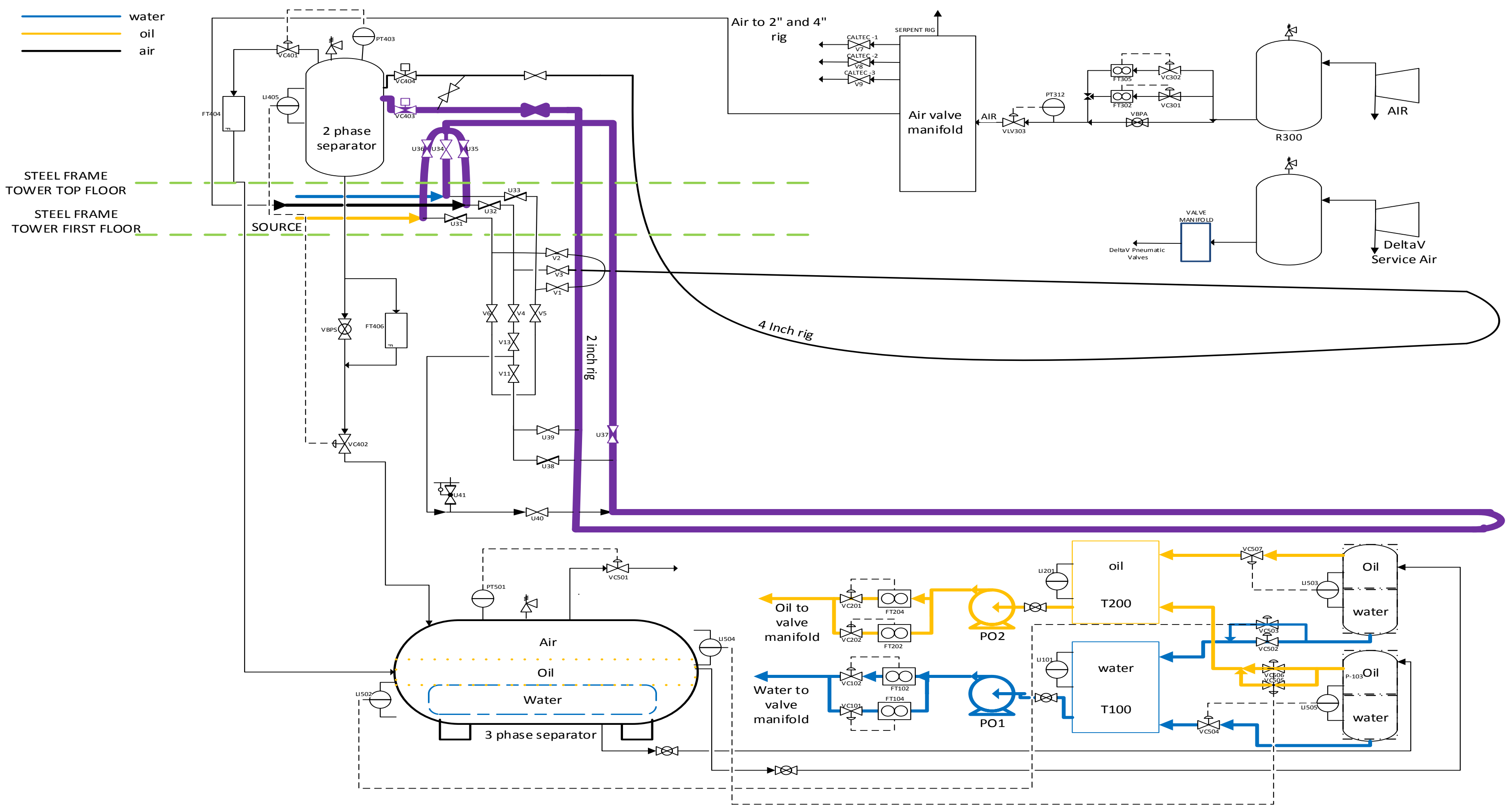


Figure 3-1 Schematic of the three phase facility, purple line - U-shape riser

3.2 Flow patterns in U-shape riser pipeline (gas - liquid flow)

Similarly to the flow in horizontal pipeline systems, the flow behavior in the U-shape riser system is influenced by gravity. The pattern of flow in multiphase flow is subject to the operating conditions of the process, the geometry of the conduit, the properties of the gas and liquid phases, the flowrate of both the gas and liquid phases and their likes. One major challenge design engineers face with multiphase systems is the sensitivity of the fluid mass, momentum and energy transfer rate to the topology of the components in the flowing fluid which influences practically the geometric distribution (Abdulkadir, 2011). Thus the interfacial area available for the phases to exchange mass, momentum or energy is strongly affected by the geometry of the system.

There are mainly two methods used in identifying flow regimes when constructing flow regime maps namely; direct observation thus visual inspection and extraction of characteristic variable from signal fluctuation in both phases (Sun et al., 2002). The procedure for identifying the flow pattern was by visual inspection and an analysis of unsteady pressures of the spectral content was used even though the fluctuations of the volume fractions as devised in some of these circumstances are established in (Jones and Zuber, 1975).

3.2.1 **Experimental procedure for flow regime identification**

Liquid flowrates within the range of 0.1 kg/s to 5 kg/s corresponding to superficial liquid velocities of 0.05 m/s to 2.47 m/s respectively were investigated against gas flowrates of 7 Sm^3/h to 150 Sm^3/h also corresponding to a superficial gas velocity of 0.18m/s to 19.87 m/s. The test matrix adopted in this study is shown in the appendix (Section A.3). Liquid (water) was pumped into the pipeline system by a 30 Hz frequency drive pump. Compressed gas flow was contacted with the pumped liquid before both transported through the downcomer of the pipeline riser system.

The flow dynamics of each condition was observed through the Perspex glass located on the vertical section of the pipeline riser system (above the riserbase). The superficial velocities of both liquid and gas were varied stepwise and each flow pattern visually observed and recorded. From the observations made, a flow regime map was developed identifying each flow pattern for the pipeline riser system.

3.2.1.1 Gas - liquid observed dynamic flow in U-shape riser system

With the point of interest in this study being the control of unstable slug flow, the identification of the slug flow region was key. Different flow points within the slug region exhibiting different characteristics were considered for further investigation. Table 3-1 shows the observed flow patterns from the different flow compositions (gas - liquid flow rates) from the test matrix.

Table 3-1 Experimentally observed flow patterns

Liquid flowrate kg/s			0.1	0.5	1	1.5	2	3	3.5	5
Liquid volumetric flowrate, m^3/s			0.0001	0.0005	0.0010	0.0015	0.0020	0.0030	0.0035	0.0050
Superficial liquid velocity per second (volumetric / area), m/s			0.05	0.25	0.49	0.74	0.99	1.48	1.73	2.47
Gas Flowrate sm3/h	Gas Flowrate nm3/h	Superficial gas velocity, m/s	OBSERVATIONS							
7.00	3.38	0.46	slug	slug	slug	slug	bubbly	bubbly	bubbly	bubbly
10.00	4.83	0.66	slug	slug	slug	slug	slug	slug	slug	bubbly
20.00	9.66	1.32	slug	slug	slug	slug	slug	slug	slug	slug
30.00	14.50	1.99	slug	slug	slug	slug	slug	slug	slug	slug
50.00	24.16	3.31	churn	slug	slug	slug	slug	slug	slug	slug
70.00	33.83	4.64	churn	slug	slug	slug	slug	slug	slug	slug
100.00	48.32	6.62	churn	slug	slug	slug	slug	slug	slug	slug
120.00	57.99	7.95	churn	slug	slug	slug	slug	slug	slug	slug
150.00	72.48	9.93	annular	slug	slug	slug	slug	slug	slug	slug
200.00	96.65	13.24	annular	slug	slug	slug	slug	slug	slug	slug
250.00	120.81	16.55	annular	annular	slug	slug	slug	slug	slug	slug
300.00	144.97	19.87	annular	annular	slug	slug	slug	slug	slug	slug

3.2.1.2 U-shape flow regime map

A total of about 100 data points were studied covering a gas superficial velocity of 0.18 m/s to 19.87 m/s and a liquid (water) superficial velocity of 0.05 m/s to 2.47 m/s on the 2 inch U-shape flow loop described earlier. Figure 3-2 shows the flow regime map obtained experimentally from the U-shape riser pipeline system. This is however compared with a flow regime map obtained from Barnea, (1987) (black line representing the transition regions).

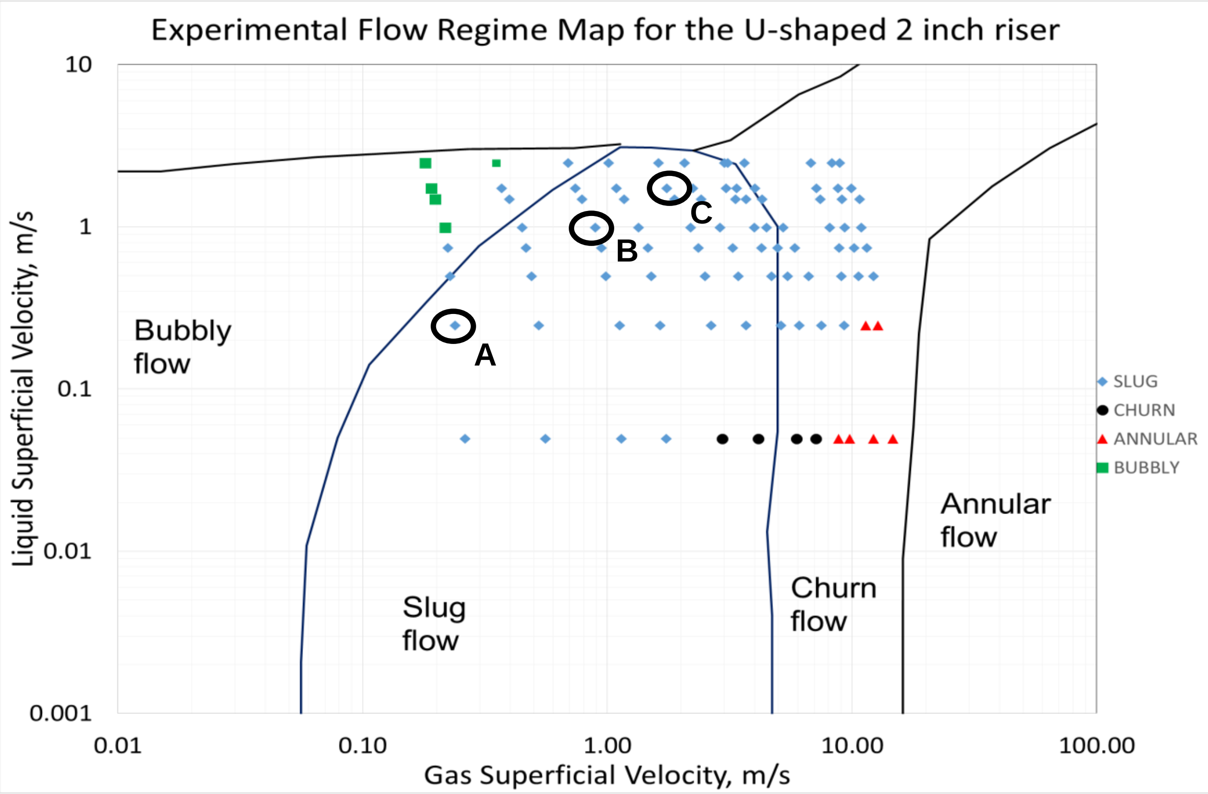


Figure 3-2 Flow regime map for the 2 inch U-shape riser loop with flow pattern boundaries from Barnea (1987), black line – Barnea boundaries.

From Figure 3-2, it could be observed that for gas superficial velocities greater than 10 m/s and high liquid superficial velocities, a slugging flow regime was still present. This indicates that the visual observations over predicts the slug flow region relative to that seen in Barnea, (1987).

Again, the slug envelope is wider and tapers towards the top compared to that observed from the literature, which could possibly mean that the slug flow from literature are mainly formed from low to medium flowrates as compared to higher flowrates. However, a considerable amount of the experimental data falls within the slug region when compared with the literature. Similar observation is made for a considerable amount of non-slugging flow conditions when compared to the literature. This could however be attributed to the difference in configuration, pipe diameter or even the mode of assessment.

3.2.1.3 Riserbase pressure trends

To understand the unstable dynamics of the flow in the U-shape riser system, three flow compositions observed to be within the unstable region are chosen to study the trend of the fluctuations or oscillations. From Figure 3-2, the three flow conditions represented by A, B and C corresponding to 0.24 m/s, 1.34 m/s and 2.23 m/s superficial gas velocities and 0.25 m/s, 0.99 m/s and 1.73 superficial liquid velocities exhibit different flow characteristics. To fully understand the dynamics and the effect the downcomer has on the flow, a slug envelope with similar flow condition was performed on a purely vertical riser system. Figure 3-3 shows flow regime map obtained through experimental from the purely vertical 2 inch riser pipeline system.

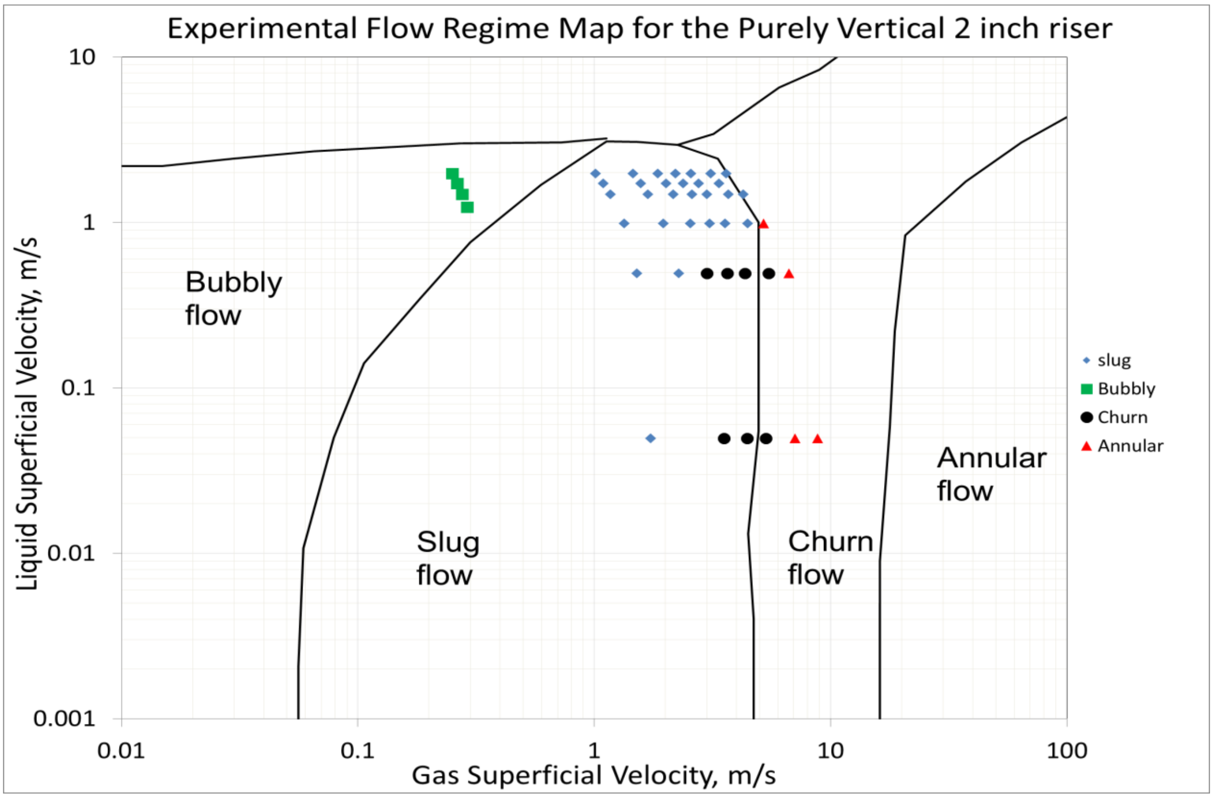
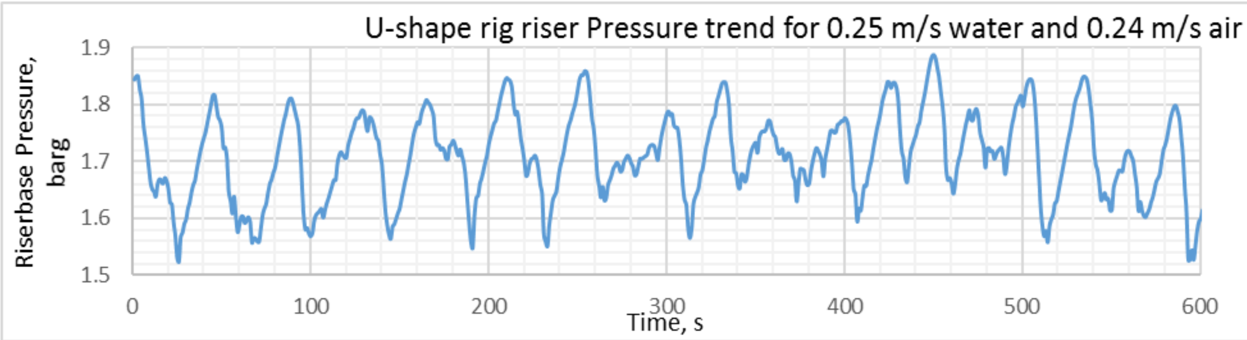
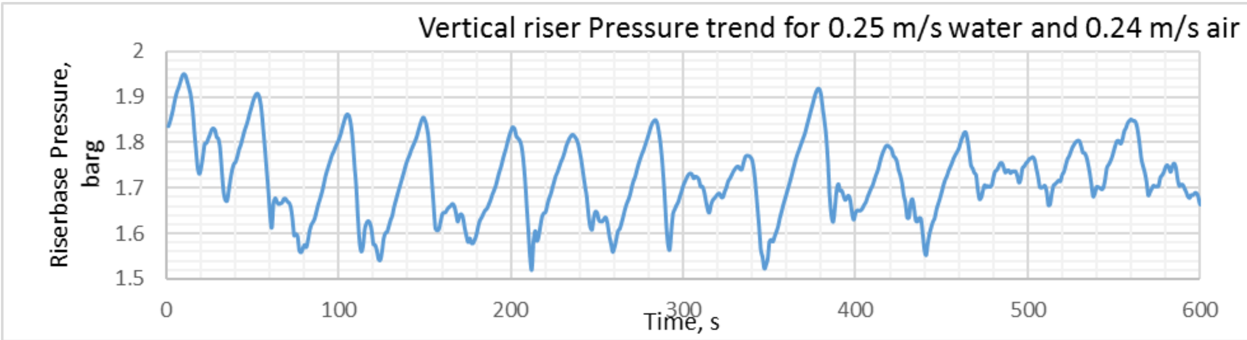


Figure 3-3 Flow regime map for a 2 inch purely vertical riser loop with flow pattern boundaries from Barnea (1987), red squares – Barnea boundaries.

Similarly, points A, B and C from Figure 3-2 fall within the unstable region in Figure 3-3. With the point of interest in this study being stabilising unstable slug flow regime, the identification of the slug region was key and different points within the slug region exhibiting different characteristics were considered for further investigation.

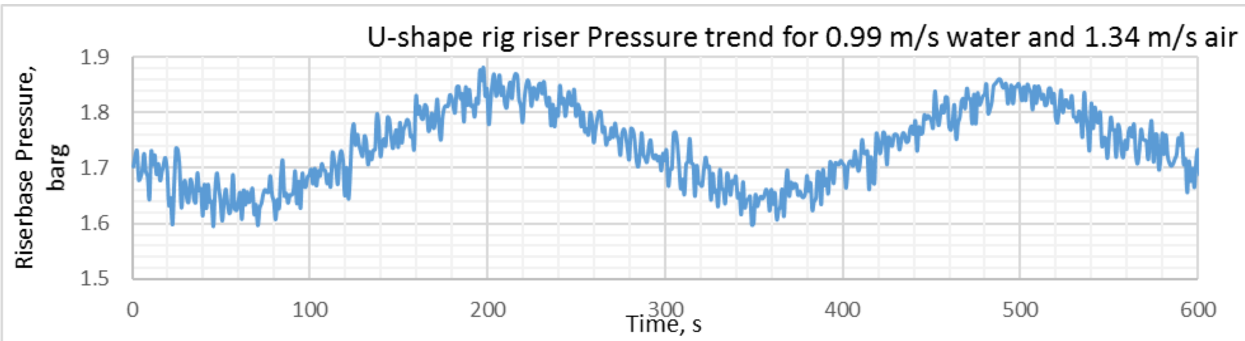


a. U-shape riser

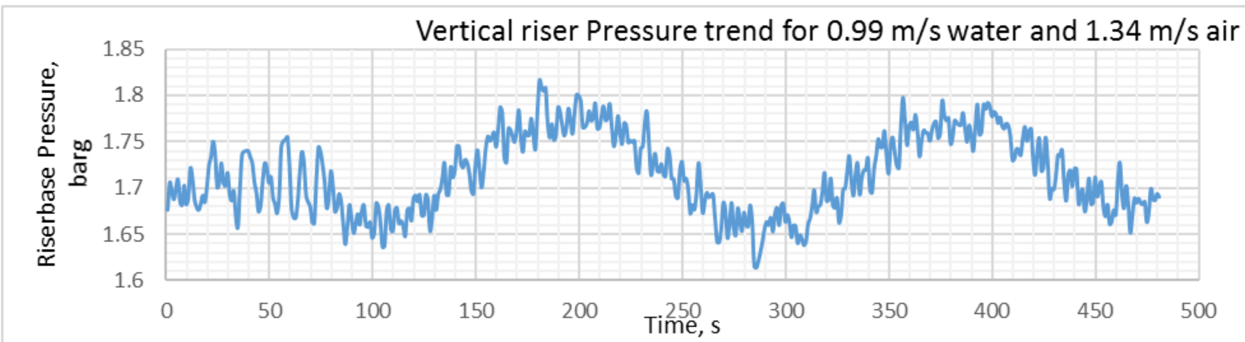


b. Vertical riser

Figure 3-4 Riserbase pressure trend for flow condition 0.25 m/s liquid and 0.24 m/s gas

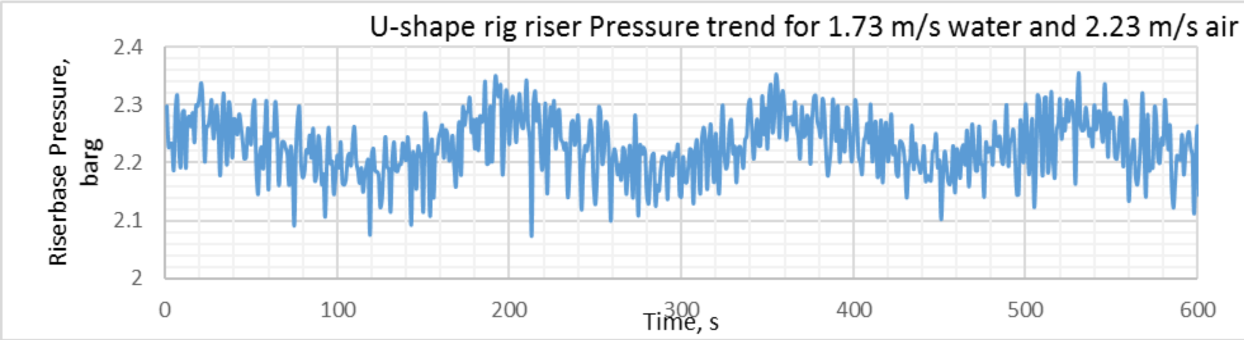


a. U-shape riser

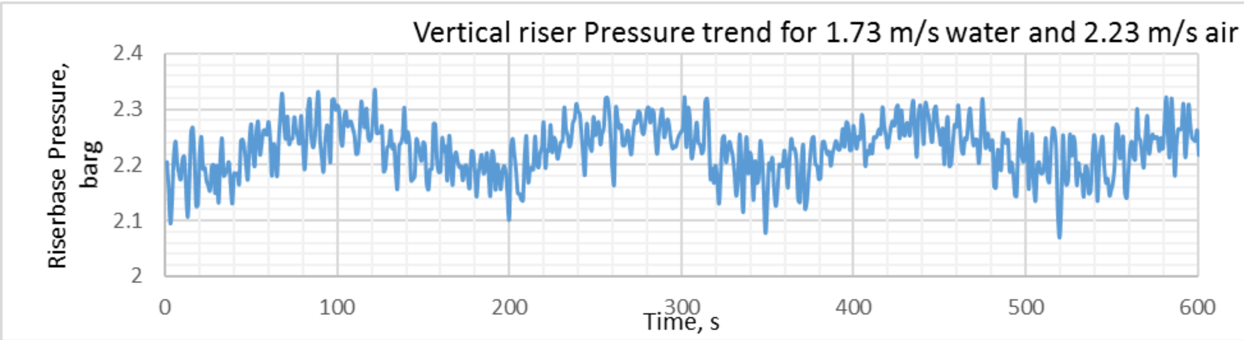


b. Vertical riser

Figure 3-5 Riserbase pressure trend for flow condition 0.99 m/s liquid and 1.34 m/s gas



a. U-shape riser



b. Vertical riser

Figure 3-6 Riserbase pressure trend for flow condition 1.73 m/s liquid and 2.23 m/s gas

Figure 3-4, Figure 3-5 and Figure 3-6 represent the riserbase pressure trend in the U-shape riser system and on the purely vertical riser for the flow conditions A, B and C from Figure 3-2.

Figure 3-4 (a) and (b), representing low gas – low liquid flowrate riserbase pressure trend on both U-shape and purely vertical risers respectively, behave similar to the characteristics of severe slugging type 2 and 3 as shown in Malekzadeh, Henkes and Mudde, (2012). Thus there exists no period where the entire riser is filled with liquid. However, the shape of fluctuation in the pressure trends represents different liquid levels translating to different volumes of produced fluids. The fluctuations in the riserbase are 0.3 - 0.4 barg in magnitude. However the magnitude of oscillation in the U-shape is slightly higher compared to that seen in the purely vertical riser. This dissimilar behaviour is traced to the geometry of the pipeline riser configuration.

Figure 3-5 (a) and (b) represent medium gas – medium liquid flowrate riserbase pressure trend on both U-shape and purely vertical risers respectively. The riserbase pressure fluctuates within a magnitude of 0.2 - 0.3 barg with low frequencies. Thus within a period of 600 seconds there were 2 complete slug cycles observed in both risers. However the pressure magnitude in the U-shape riser pipeline was slightly higher than that observed in the purely vertical riser.

Similarly, Figure 3-6 (a) and (b) represent high gas – high liquid flowrate riserbase pressure trend on both U-shape and purely vertical risers respectively. The riserbase pressure fluctuates within a magnitude of 0.2 - 0.25 barg. The frequency of oscillation is comparatively higher than that seen for medium gas – medium liquid flowrate producing about 3.5 slug cycles per 600 seconds. Again, the magnitude of oscillation in the U-shape riser is higher than that observed in the purely vertical riser. This signifies that the pipeline configuration surely has an impact on the instabilities in the system hence the investigation of the cause a necessity. A stability analysis for flow conditions A, B and C on both risers will be assessed next.

3.2.2 **Validation of model**

The experimental setup (U-shape riser system) in the laboratory deployed for this study was validated using a numerical simulation tool (OLGA) to assess and establish the potency of the results that would be derived. A U-shape riser system setup with configuration similar to that described above was model in OLGA. The pressure trend resulting from the flow condition 0.25 m/s water and 0.24 m/s air superficial velocities was assessed and investigated.

With the model equipped with slug tracking and slug tuning, the model was run. The ‘SLUGFREQCONST’ of the slug tracking was fixed at 0.005. Also the slug tuning parameters are fixed as;

DPFACT – 10

DPONSET – 1

SLUG LENGTH - $1*10^9$

UBCOEFF1 – 1

UBCOEFF2 – 1

VOIDINSLUG – 1

The corresponding pressure trend derived from the numerical simulation tool (OLGA) for the above mentioned flow condition was compared to that derived from experiment as shown in Figure 3-7.

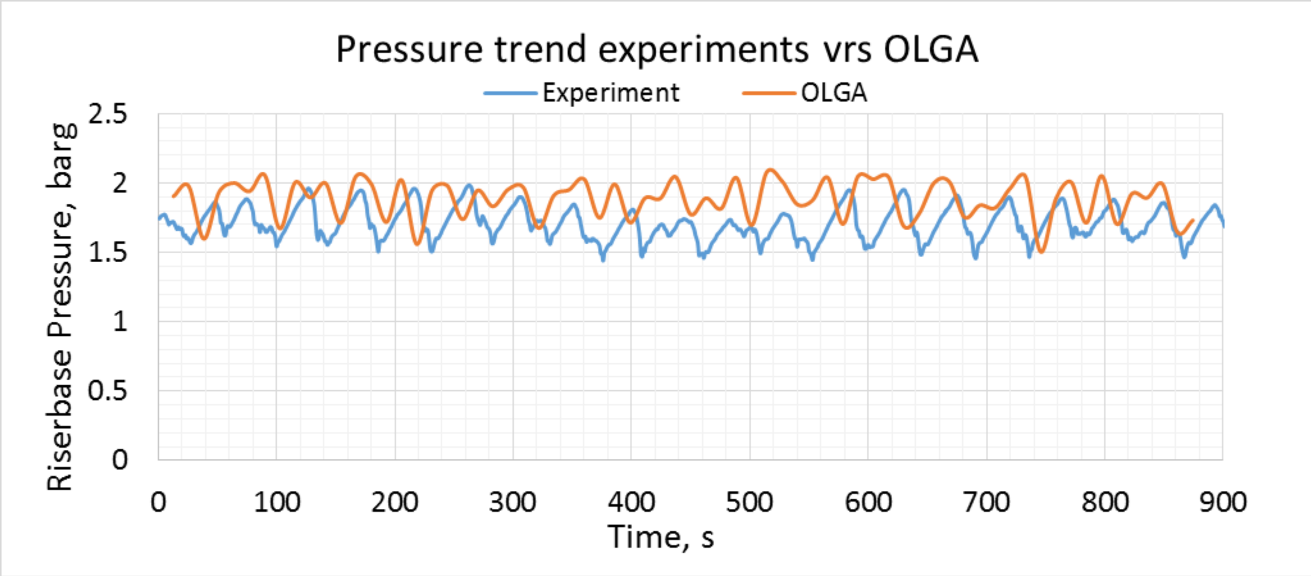


Figure 3-7 Pressure trend in 2 inch U-shape riser loop at flow conditions 0.25 m/s liquid and 0.24 m/s gas superficial velocities in both experiments and OLGA simulation

From Figure 3-7, the pressure trends for the same flow condition in both experiment and OLGA simulation has similar characteristics. However the mean riserbase pressure in OLGA is relatively higher compared to that observed from experiment. This could be attributed to the exact point at which the riserbase pressure trend was extracted and pipeline diameter as random flow patterns do often occurs in smaller diameter pipes. Table 3-2 shows the minimum, average and maximum riserbase pressure value from both experiment and OLGA, as well as the standard deviation in the riserbase pressure in both systems. Generally, both trends matches significantly. In conclusion it is possible to get similar riserbase pressure trend from experiments and numerical simulation for the same flow system and condition.

Table 3-2 Result summary for system validation of the operating condition through the U-shape in from experiments and numerical simulation (OLGA)

Riserbase Pressure	**Experiments**	**OLGA**
Minimum riserbase pressure, barg	1.440	1.504
Maximum riserbase pressure, barg	1.985	2.080
Average riserbase pressure, barg	1.710	1.884
Standard deviation	0.117	0.143

3.2.3 Stabilising unstable flow condition

The flow behaviour for different flow compositions has been studied in previous sections. From the literature it has been established that an increase in the downhole pressure of the system can help in slug mitigation. This however has been one of the common and most used method for slug elimination in the hydrocarbon industry. In this part of the study, this concept would be further explored for each of the flow conditions (low gas – low liquid flowrate (A), medium gas – medium liquid flowrate (B) and high gas – high liquid flowrate (C)) chosen from the slug envelope with the aid of a choke valve. The choke valve located at the topside of the riser was used to increase the pressure in the system. Riserbase pressure trends resulting from manual choking was used to generate bifurcation maps for the various slug conditions or different forms obtained from the flow regime map. This would aid in the understanding of the slug behaviour as well. Bifurcation maps are produced for the typical unstable slug flow conditions shown above to gain an advanced understanding of the behaviour of these slug types.

3.2.3.1 Procedure for flow stability in open loop

A bifurcation analysis (manual choking) was done to identify the stability point of the different slugging conditions chosen for further investigation work. From the flow regime map produced, three flow conditions within the slug regime were identified and investigated using the bifurcation analysis.

An unstable slug flow condition with superficial velocities of 0.24 m/s and 0.25 m/s for gas and liquid respectively, representing a low flowrates exhibiting a severe slugging condition was run. A stepwise decrease in the choke valve opening was done from a valve opening of 100 % to 10 %. Riserbase pressures for each condition were recorded. The minimum, average and maximum riserbase pressure of each condition were plotted against their corresponding percentage valve opening (bifurcation map) where the critical bifurcation point was identified.

Consequently, the same process was done for the two other chosen unstable flow conditions. Bifurcation maps were developed for these operations identifying the critical bifurcation points. Same conditions were run for the purely vertical riser system to investigate the effect of configuration on flow stability.

3.2.3.2 Low gas – low liquid flowrate bifurcation map

Using the topside riser choke valve as the manipulating variable and the riserbase pressure as the controlled variable, a bifurcation map was generated for low gas – low liquid flowrate. Figure 3-8 shows the bifurcation map obtained from the U-shape riser system for a 7 Sm^3/h gas and 0.5 kg/s liquid corresponding to a 0.24 m/s and 0.25 m/s gas and liquid superficial velocities respectively. From Figure 3-8, a bifurcation point of 19 % valve opening was observed corresponding to a riserbase pressure of 2.8 barg. The region beyond the 19 % opening (increasing valve opening) is considered as unstable whiles that to the left side of that (decreasing from 19 % valve opening) is a stable region. Again, the variation in valve opening (reducing the valve opening) causes an increase in the riserbase pressure of the system. The increased riserbase pressure causes the unstable low gas – low liquid flowrate to be relatively stable a resultant of increased pressure drop across the choke valve. This explains the bane for choking to be used to stabilise unstable slug flow condition.

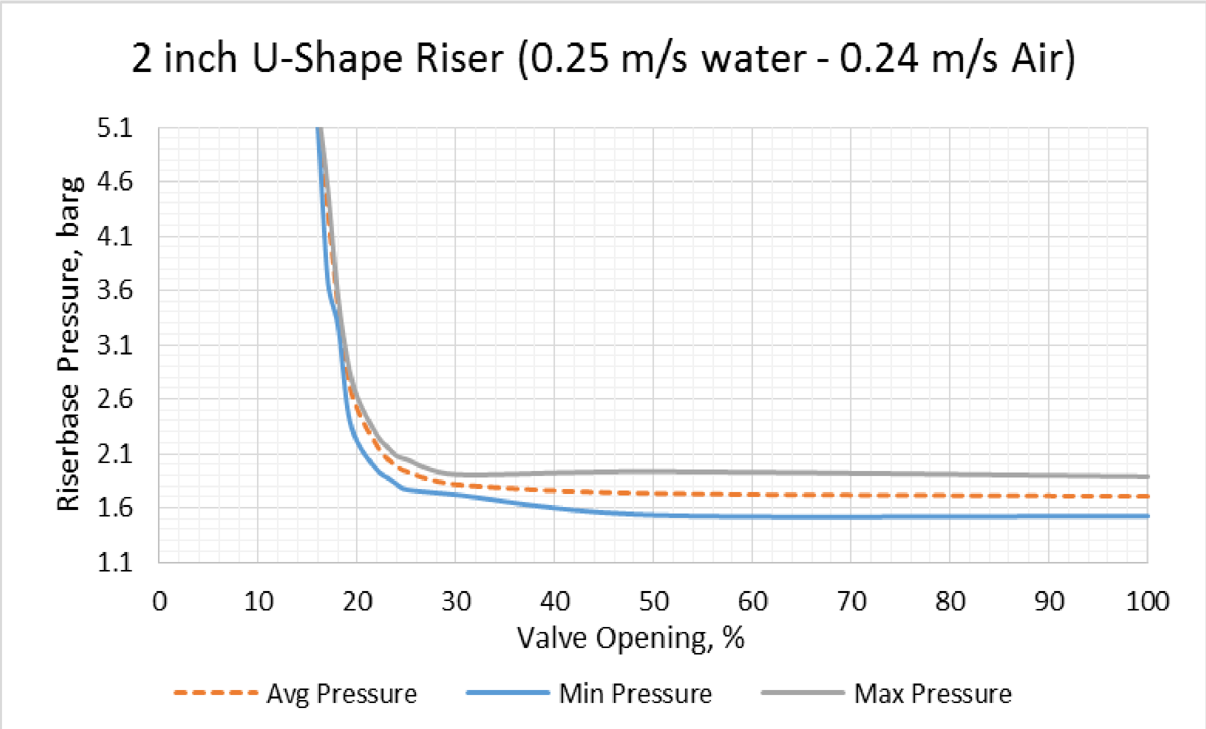


Figure 3-8 Bifurcation map for a 2 inch U-shape riser loop at flow conditions 0.25 m/s liquid and 0.24 m/s gas superficial velocities

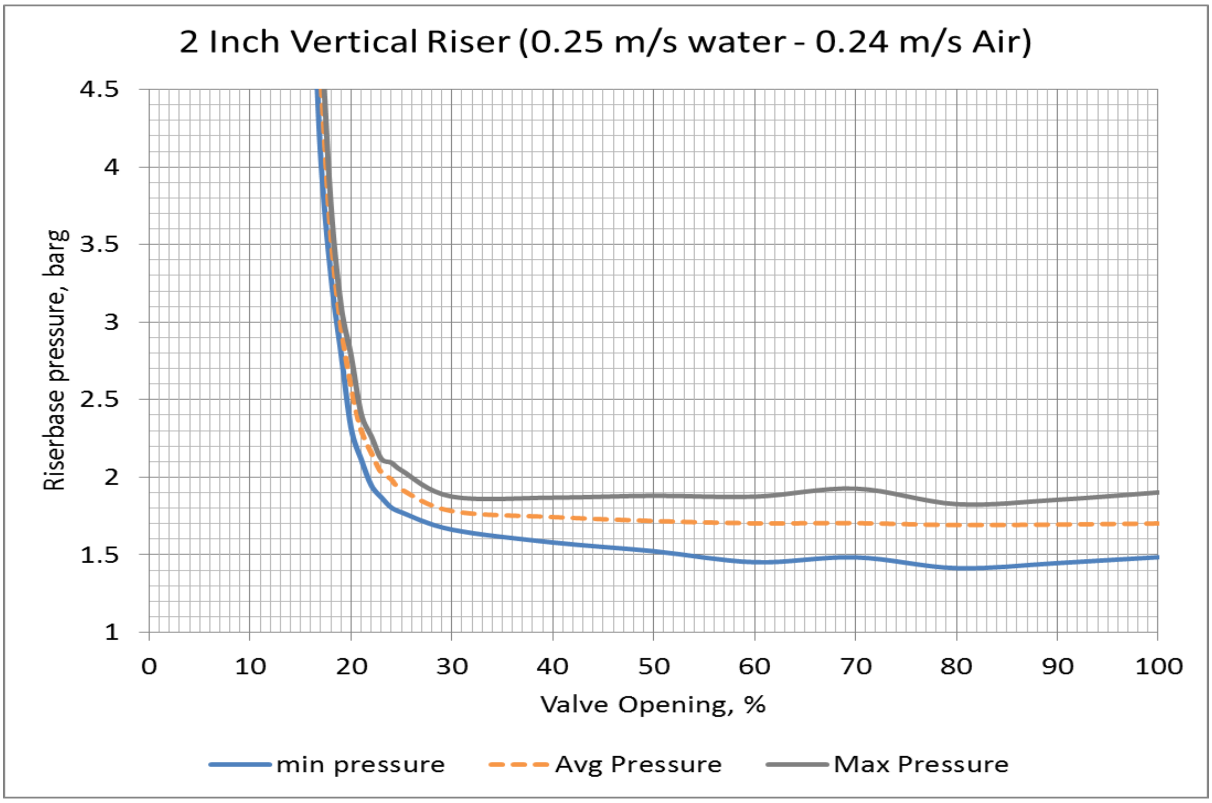


Figure 3-9 Bifurcation map for a 2 inch purely vertical riser loop at flow conditions 0.25 m/s liquid and 0.24 m/s gas superficial velocities

Similarly, Figure 3-9 represents the bifurcation map obtained from the 2 inch vertical riser system for the low gas – low liquid flowrate (7 Sm^3/h gas and 0.5 kg/s liquid). This corresponds to a 0.24 m/s and 0.25 m/s superficial velocities of gas and liquid respectively. From Figure 3-9, it was observed that the pressure oscillation magnitude reduces significantly as the choke valve opening was reduced from 100 % to 19 % opening. Valve closure beyond 19 % opening showed a relatively constant oscillation even though the magnitude increases. A bifurcation point of 19 % valve opening corresponding to a riserbase pressure of 2.7 barg was seen for the low gas – low liquid flowrate on the 2 inch purely vertical riser. The region above the 19 % valve opening (right side - increasing valve opening) is considered an unstable region while that to the left side of that is a stable region.

The bifurcation maps produced in Figure 3-8 and Figure 3-9 from the U-shape riser and the purely vertical riser respectively for low gas – low liquid flowrate yielded the same critical bifurcation point, thus, 19 % choke valve opening. However, the 2 inch U-shape riser system produces a slightly higher corresponding riserbase pressure compared to the 2 inch purely vertical riser.

3.2.3.3 Medium gas – medium liquid flowrate bifurcation map

The riserbase pressure bifurcation map for medium gas – medium liquid flowrate on the 2 inch U-shape riser flow loop is illustrated in Figure 3-10. Medium gas – medium liquid flowrate was represented by 30 Sm^3/h gas and 2 kg/s liquid which is equivalent to 1.34 m/s and 0.99 m/s gas and liquid superficial velocities respectively. From Figure 3-10, 29 % choke valve opening was seen to be the bifurcation point of the system corresponding to a 3.35 barg pressure at the base of the riser. At 29 % choke valve opening, the maximum and minimum valve openings connect and consequently the point which differentiates the stable and unstable region of the system.

Again the same flow condition was run on the 2 inch purely vertical riser and the resulting bifurcation map shown in Figure 3-11. The bifurcation point was seen to be at 31 % valve opening which corresponds to a 3.1 barg riserbase pressure. There was a decrease in the riserbase pressure from 3.35 barg at 29 % choke valve opening as seen in Figure 3-10 for the U-shape riser to 3.1 barg at 31 % choke valve opening in Figure 3-11 for the purely vertical riser. This indicates that an increase in the topside choke valve opening, eases of the pressure in the pipeline system.

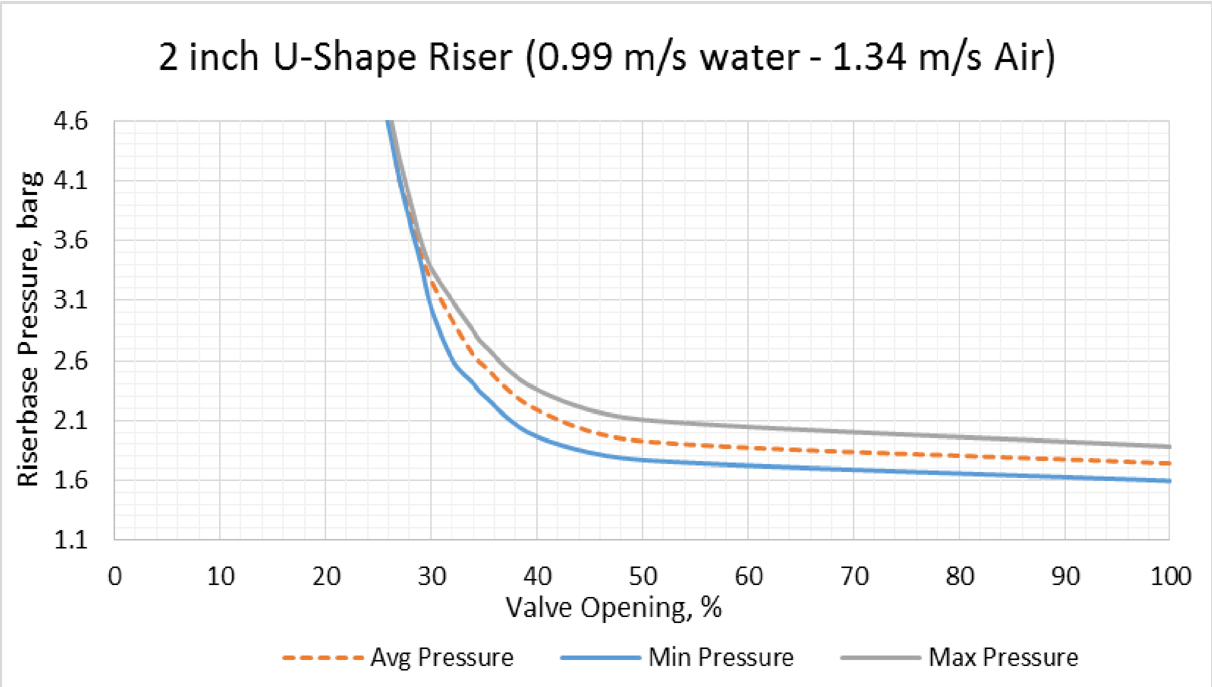


Figure 3-10 Bifurcation map for a 2 inch U-shape riser loop at flow conditions 0.99 m/s liquid and 1.34 m/s gas superficial velocities

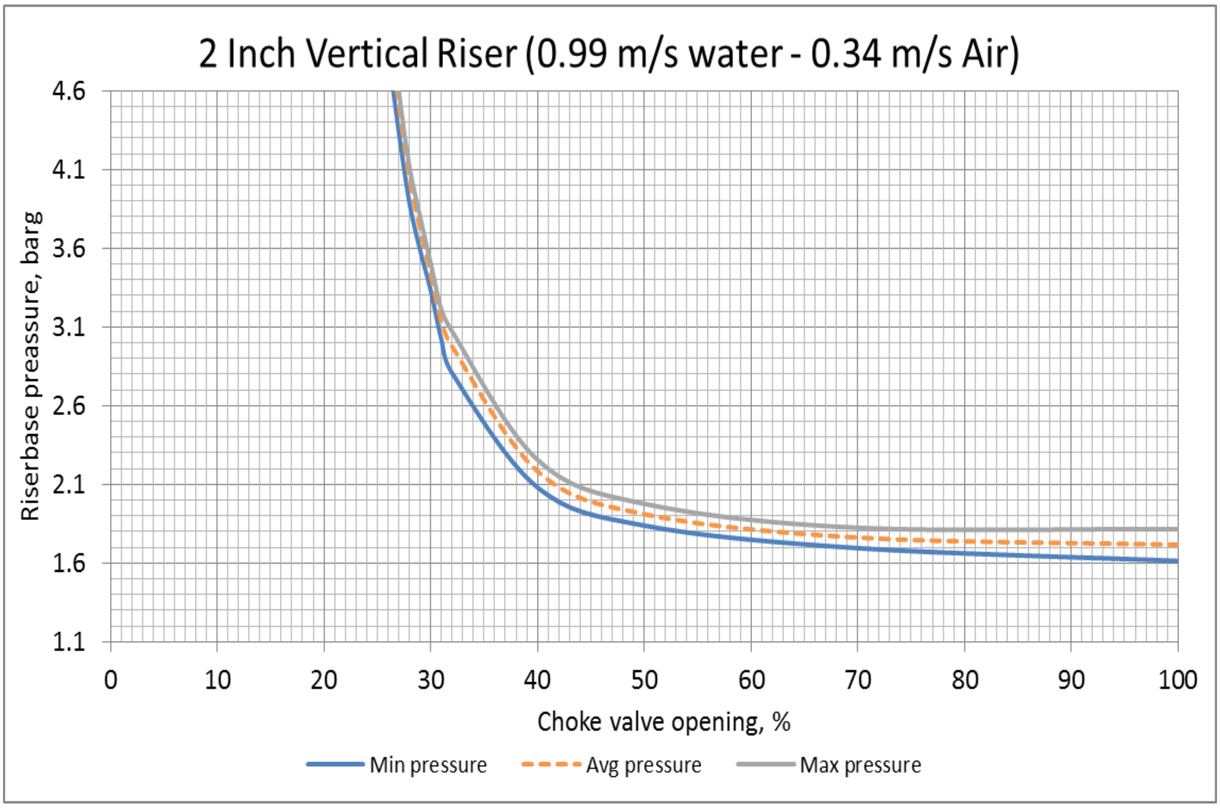


Figure 3-11 Bifurcation map for a 2 inch purely vertical riser loop at flow conditions 0.99 m/s liquid and 1.34 m/s gas superficial velocities

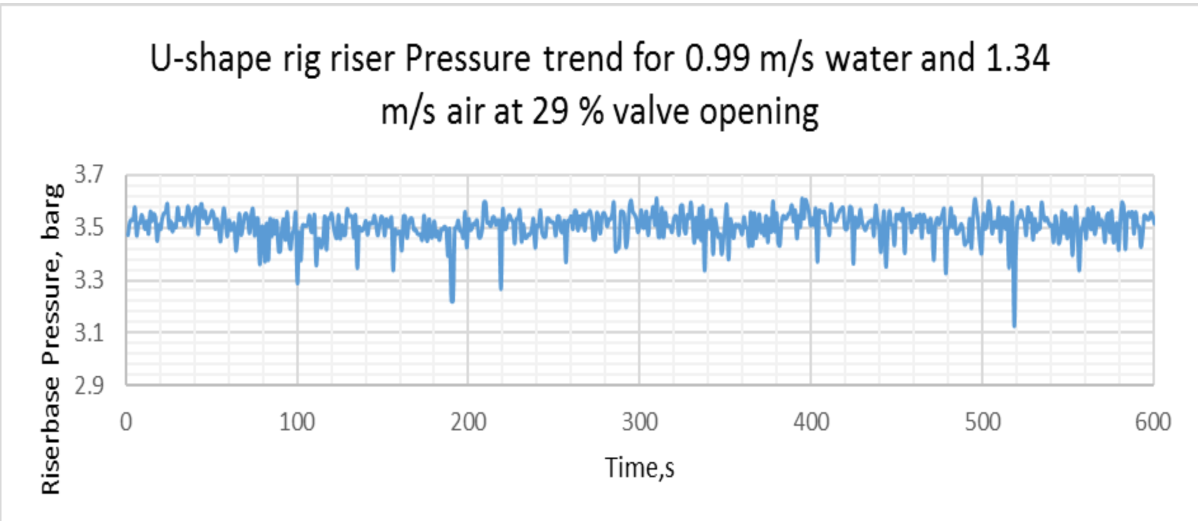


Figure 3-12 Riserbase pressure trend for a 2 inch U-shape riser loop at flow conditions 0.99 m/s liquid and 1.34 m/s gas at 29 % valve opening

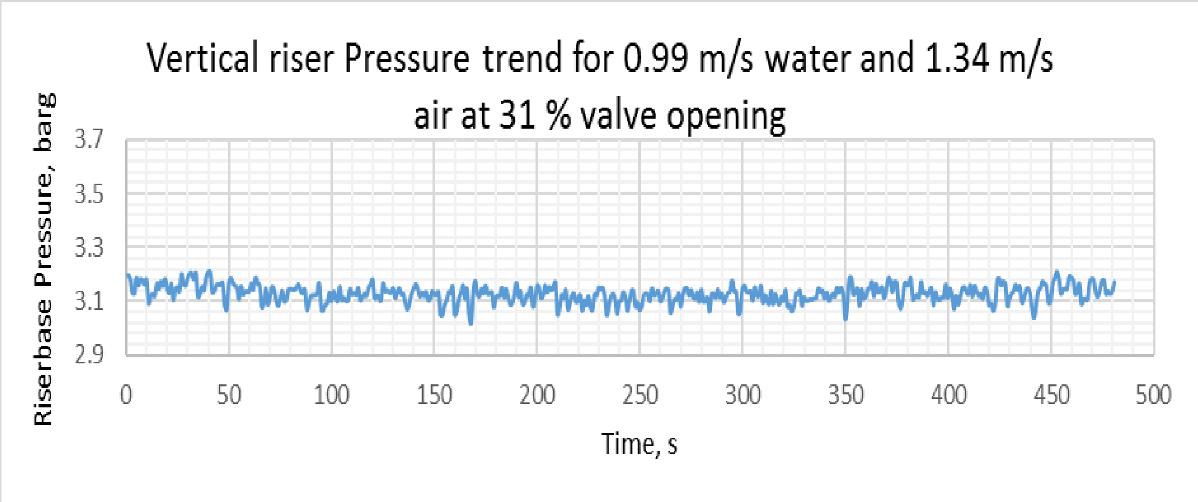


Figure 3-13 Riserbase pressure trend for a 2 inch purely vertical riser loop at flow conditions 0.99 m/s liquid and 1.34 m/s gas at 31 % valve opening

Figure 3-12 and Figure 3-13 represent the critical bifurcation points pressure trend from the base of each risers (U-shape and purely vertical riser respectively) for a moderate gas – moderate liquid flowrate condition. The pressure fluctuation seen from both are of minimum fluctuation compared to that in Figure 3-5 for the U-shape and purely vertical riser respectively when the choke valve was 100 % opened. The increased pressure drop across the choke valve as a results of the valve closure explains the relatively stable flow oscillation observed.

3.2.3.4 High gas – high liquid flowrate bifurcation map

Figure 3-14 shows the bifurcation map obtained from the U-shape riser for a high gas – high liquid flowrate represented by a 75 Sm^3/h gas flowrate and 3.5 kg/s liquid flowrate. This flowrate corresponds to a 2.23 m/s superficial gas velocity and 1.73 m/s superficial liquid velocity. From Figure 3-14 a bifurcation point of 38 % choke valve opening corresponding to a 3.7 barg riserbase pressure was observed. The choke valve operating range beyond 38 % opening was considered an unstable region thus to the right of the critical bifurcation point whiles that to the left side of the critical stable opening was considered stable. Again, after reducing the valve further into the stable region, it was observed that the system becomes unstable again which was due to the slugging induced by over choking, thus over-choking induced slugging.

Figure 3-15 shows the pressure trend in the U-shape riser system at stable condition (38 % choke valve opening). Comparatively, Figure 3-15 shows a much stable flow behaviour than observed in Figure 3-6, thus at 100 % valve opening. This is because of the attained increased pressure drop across the valve resulting from the closure of the choke valve (38 % choke valve opening).

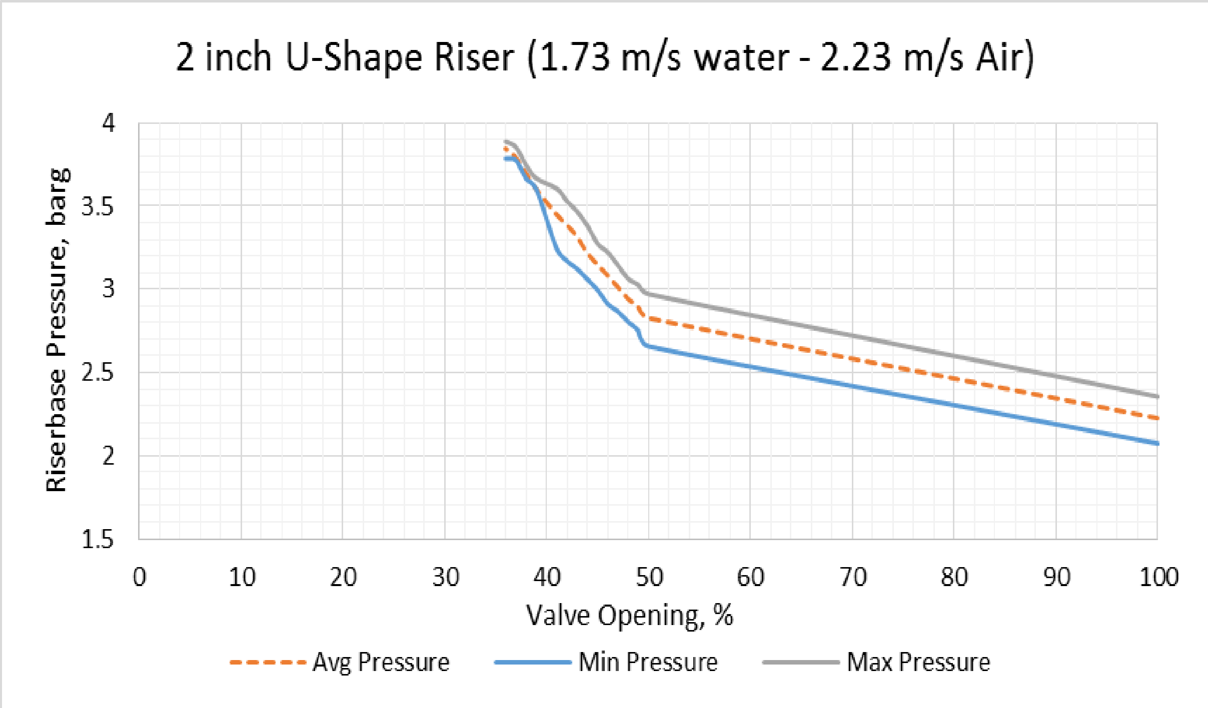


Figure 3-14 Bifurcation map for a 2 inch U-shape riser loop at flow conditions 1.73 m/s liquid and 2.23 m/s gas superficial velocities

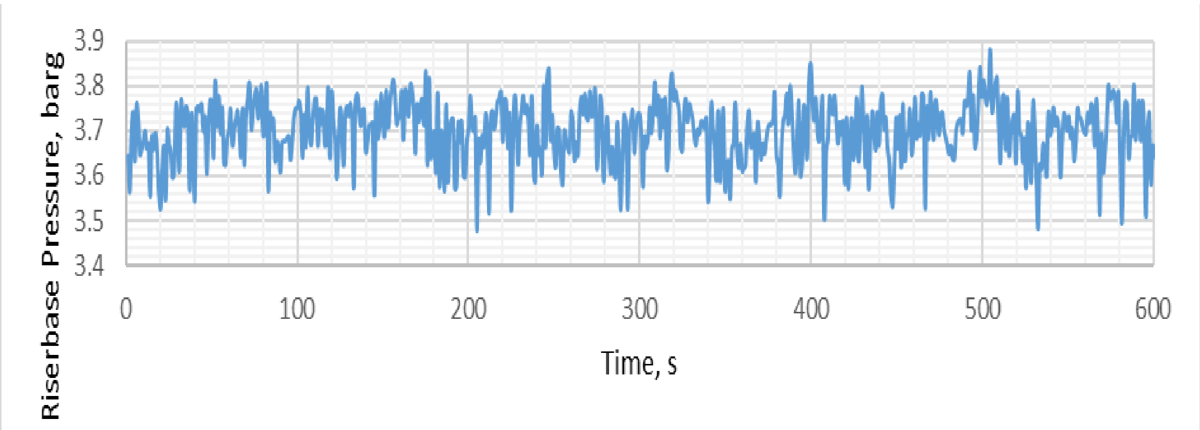


Figure 3-15 Riserbase pressure trend for a 2 inch U-shape riser loop at flow conditions 1.73 m/s liquid and 2.23 m/s gas at 38 % valve opening

Bifurcation map generated from the pure vertical riser for the same flow condition (high gas – high liquid flowrate) is shown in Figure 3-16. From Figure 3-16, 39 % choke valve opening was identified as the bifurcation point which corresponds to a riserbase pressure of 4.0 barg. For this operating condition, valve openings above 39% was seen to be an unstable region whiles valve openings 39 % and below was seen to be in the stable region. Figure 3-17 represents a relatively stable pressure trend obtained at a 39 % valve opening on the purely vertical riser for the same high gas – high liquid flowrate compared to the trend seen in Figure 3-6 for the same flow condition but at 100 % valve opening. Thus Figure 3-17 shows a stable pressure fluctuation since it was operating at a 39 % choke valve opening.

It has been presented that significant choking was required to alleviate the unstable slug flow in the pipeline riser system which unfortunately could translate to less production of fluids. It is therefore important to advance the approach to stabilising the unstable slug flow at a considerably larger valve opening. Conclusively, considering the flow conditions run on both the U-shape riser system and the purely vertical riser system, there was an obvious similarity between the stability points, however the extra riser system volume caused by the downcomer affects the stability point. Thus the valve opening had to be closed further before the system could be stabilised. This raised concerns where the actual flow instabilities in the U-shape riser system could be from and this would be investigated next. Table 3-3 presents a comparison of the outcome for the three flow conditions through both the U-shape and purely vertical riser in terms of stability analysis, thus critical bifurcation / stability valve opening and the corresponding riser base pressure.

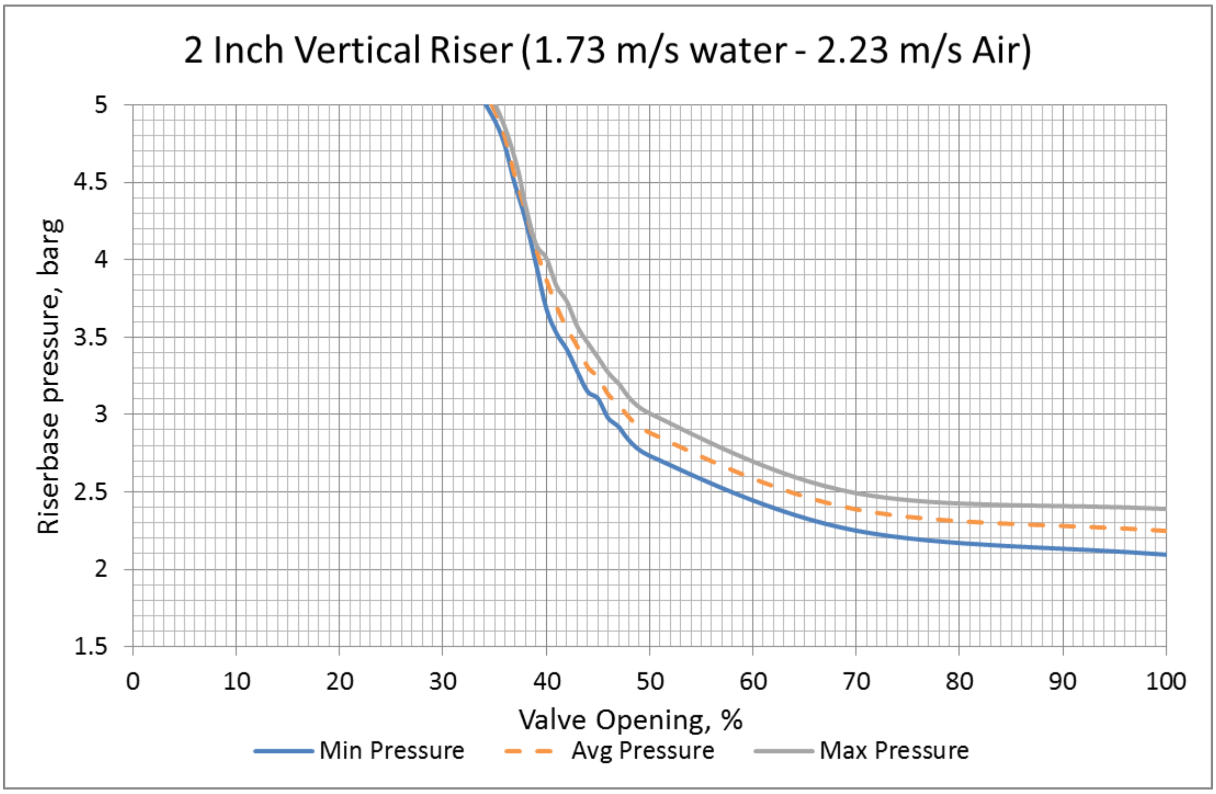


Figure 3-16 Bifurcation map for a 2 inch purely vertical riser loop at flow conditions 1.73 m/s liquid and 2.23 m/s gas superficial velocities

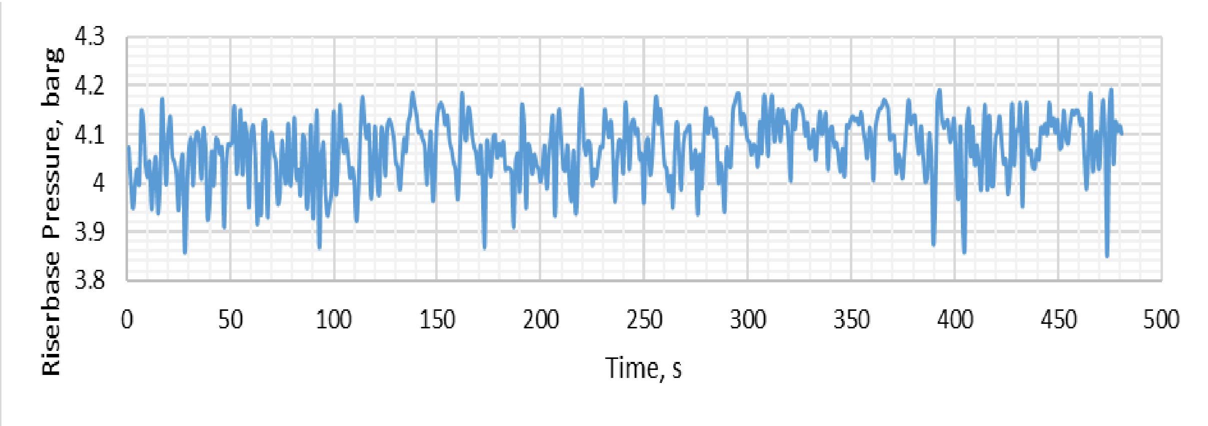


Figure 3-17 Riserbase pressure trend for a 2 inch puerly vertical riser loop at flow conditions 1.73 m/s liquid and 2.23 m/s gas at 39 % valve opening.

Table 3-3 **Result summary for stability analysis (critical bifurcation points and riserbase pressure) for operating conditions through the U-shape and vertical riser**

Flow Condition	U-Shape Riser		Purely Vertical Riser	
	Critical Bifurcation Point, %	Riserbase Pressure, barg	Critical Bifurcation Point, %	Riserbase Pressure, barg
A – 0.25 m/s Liquid and 0.24 m/s Gas	19	2.8	19	2.7
B – 0.99 m/s Liquid and 1.34 m/s Gas	29	3.35	31	3.1
C – 1.73 m/s Liquid and 2.23 m/s Gas	38	3.7	39	4.0

3.2.4 Initiation of flow instabilities in U-shape riser

Studying the dynamics of gas-liquid flow in the U-shape riser yielded a flow regime map which helped identifying the region where slug flow pattern was observed when viewed through the Perspex glass located on the vertical section of the pipeline riser system (above the riserbase). Even though the obtained flow regime map identified on the U-shape riser does not vary from that on a purely vertical riser, the stability point for some flow conditions exhibiting unstable slug flow differ in both the U-shape and the purely vertical risers hence the need to investigate this cause. To understand this concept, an investigation on the unstable slug flow region was performed to establish the initialization of slug flow in the riser.

A justification for the cause for this flow dynamics was necessary hence some further modifications were made on the U-shape riser system. These modifications included introduction of extra visual section and extra pressure transducers on the downcomer as shown in Figure 3-18 and the flow loop instrumentation shown in Table 3-4.

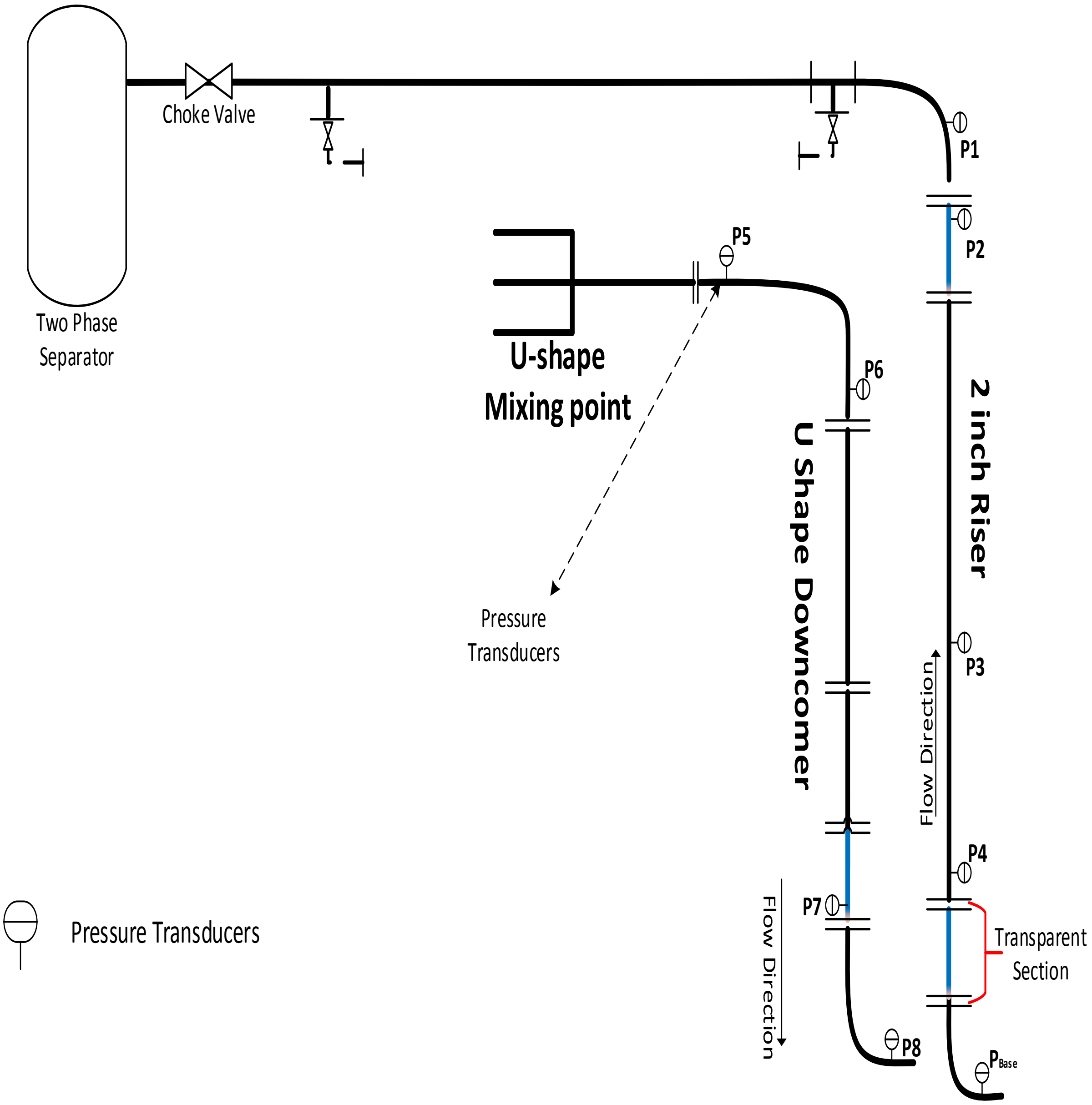


Figure 3-18 U-shape Riser system

The slug flow pattern behaviour observed in the U-shape riser was of great concern since the initialization of slug flow in the riser was not understood. The boundaries of the slug flow regime was assessed to know the initialization stage of slugging in the U-shape flow riser. An experimental run of the flow conditions exhibiting unstable slug flow in the U-shape riser was assessed. Flow pattern in the downcomer of the U-shape riser system was observed and the resulting flow regime map is shown in Figure 3-19. From Figure 3-19, four distinct and unique flow behaviours were observed in the down comer of the U-shape riser.

Table 3-4 Instrumentation list on U-shape riser

Instrumentation	Meaning
P1	Riser Top Pressure
P2	Riser Top Pressure (2)
P3	Riser Straight Pressure
P4	Riser Straight Bottom Pressure
P5	Downcomer Inlet Pressure
P6	Downcomer Top Pressure
P7	Downcomer Straight Pressure
P8	Downcomer base Pressure
P_{Base}	Riserbase Pressure

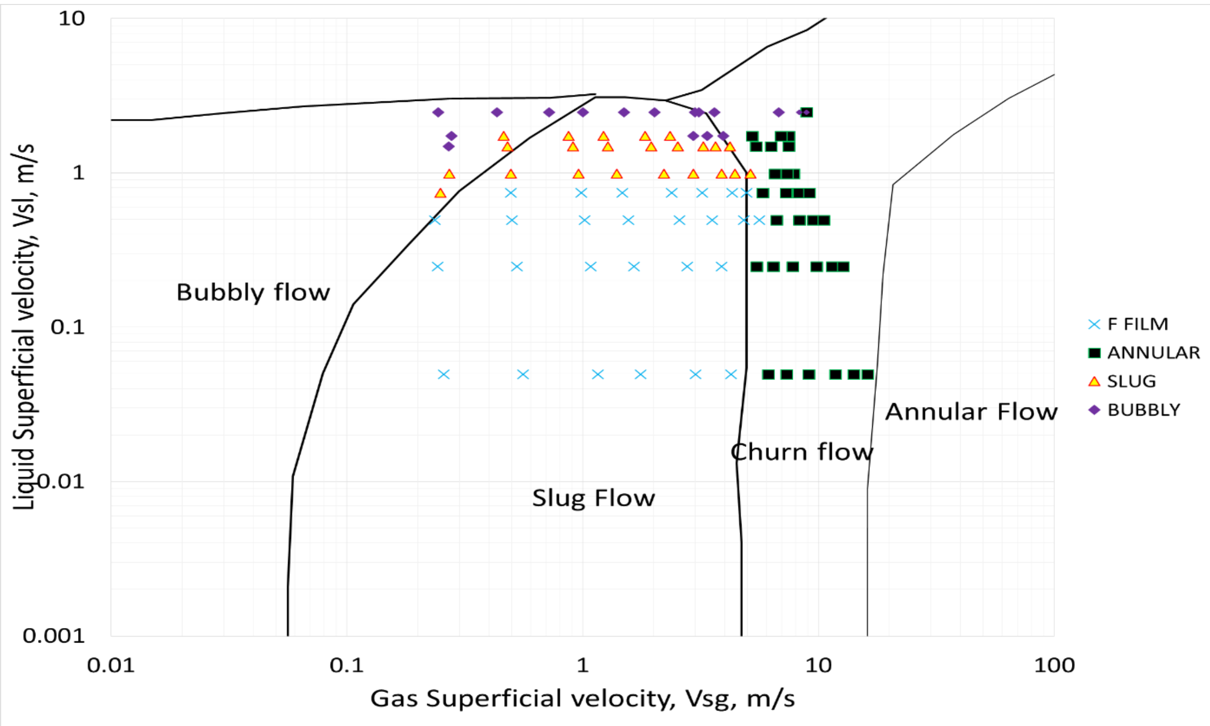


Figure 3-19 Flow regime map for a 2 inch U-shape downcomer determined experimentally

For very low gas and liquid flowrate, a free fall flow was seen in the downcomer. Increasing the liquid flowrate (mid flowrate) showed a slug flow regime in the down comer. A further increase to higher liquid flowrate showed a dispersed bubble flow.

Similarly, increasing the gas flowrate at constant liquid flow rate resulted in an annular flow regime. This was observed however for relatively medium gas flowrate. Again for high flow rate for both gas and liquid, annular flow pattern was observed, while reducing the gas flowrate at high liquid flowrate showed dispersed bubble flow.

Figure 3-20 shows a comparison of the slug behaviour regions in both the riser and the downcomer. It was observed that the slug region in the downcomer (red sectioned area) falls perfectly within the slug region of the riser (blue sectioned area). This shows that all the flow conditions exhibiting slugging flow in the downcomer, exhibits slugging flow in the riser.

However, not only flow conditions exhibiting slug flow in the downcomer translate to slug flow in the riser, since there were other flow regimes in the downcomer that also exhibited slug flow in the riser. This could means that, the initialization stage of slug flow in the riser is not necessarily from the downcomer but could be from either the flow condition itself or might also be from the horizontal section of the flow loop as established by several researchers.

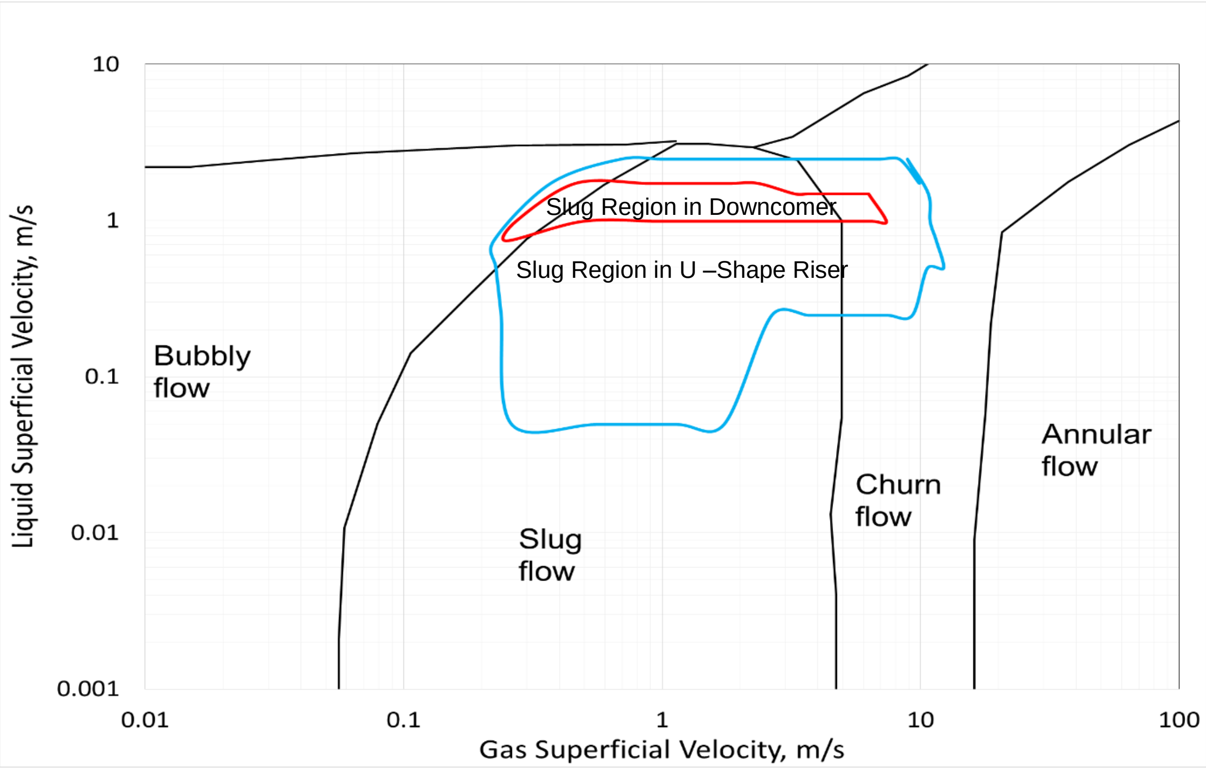


Figure 3-20 Slug region comparism for both downcomer and riser system in a 2" U-shape loop

3.2.4.1 Downcomer pressure trends

Figure 3-21 - Figure 3-24 show the downcomer pressure trends in the U-shape riser system. These figures represent the transient flow behaviour of a free fall flow, slug flow, dispersed bubble flow and an annular flow pattern in the downcomer. Different flow patterns are observed for the different flow pattern characterization.

The slug flow characteristics observed in the downcomer behave like type 2 and 3 severe slugging described in (Malekzadeh, Henkes and Mudde, 2012), however there was no time period where the entire downcomer was completely filled with liquid. However, the liquid in the downcomer build up to an extent since there is an oscillating flow observed in the system which is a resultant of the different liquid heights. The downcomer pressure fluctuates with an amplitude of about 0.1 barg which is a considerable oscillation in the 2 inch U-shape riser hence we consider to stabilise the system next using the downcomer pressures.

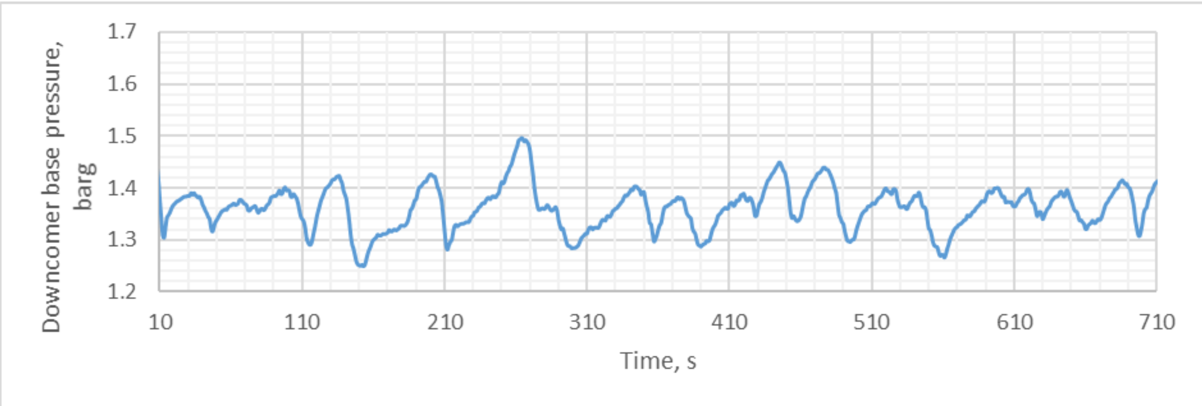


Figure 3-21 Pressure trend for a free falling flow in the 2 inch U-shape downcomer.

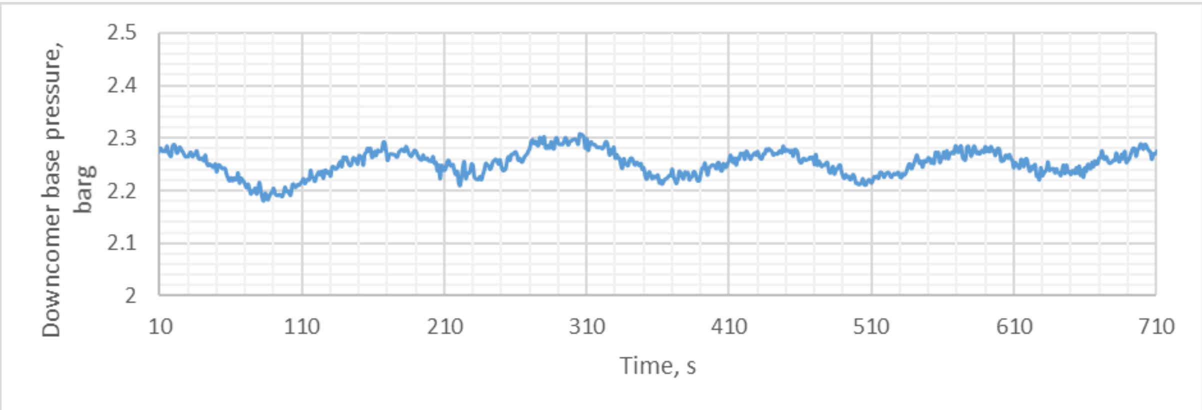


Figure 3-22 Pressure trend for a slug flow in the 2 inch U-shape downcomer.

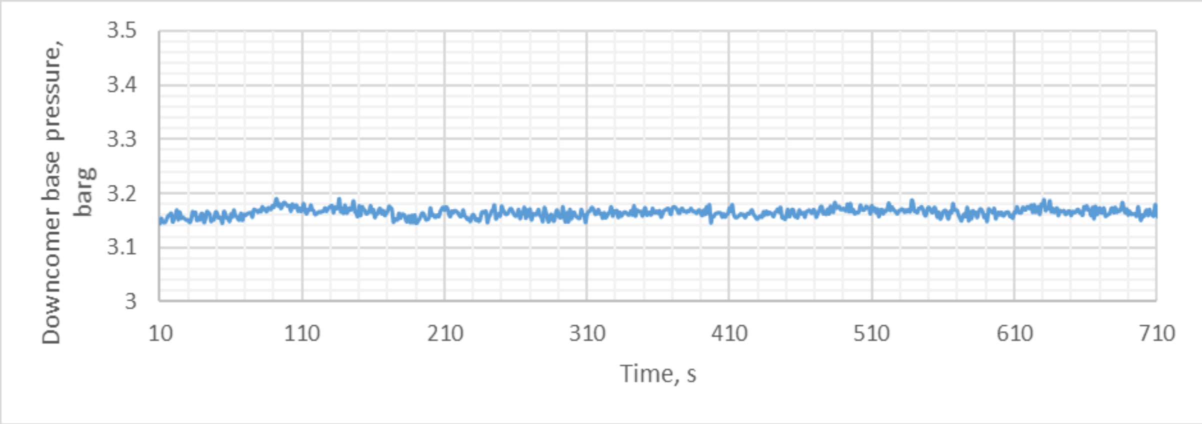


Figure 3-23 Pressure trend for a dispersed bubble flow in the 2 inch U-shape downcomer.

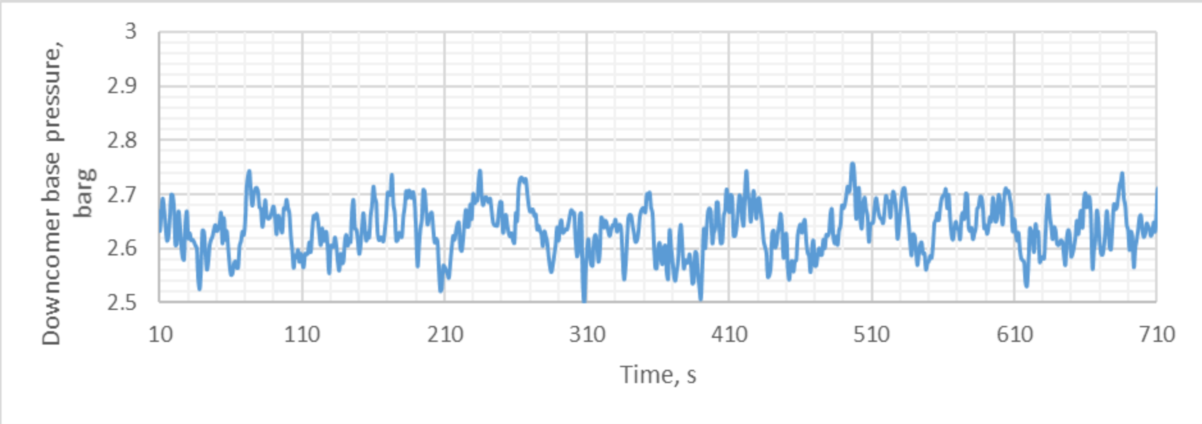


Figure 3-24 Pressure trend for an annular flow in the 2 inch U-shape downcomer.

3.2.4.2 System stability study using downcomer top pressure

To verify the slug initiation point in the U-shape riser, pressure stability test (bifurcation maps) was developed using different pressure points (downcomer top pressure, downcomer base pressure and riserbase pressure) on the U-shape riser. This study aims to establish any similarity in system response in both the riser and the downcomer of the U-shape riser with regards to the choke valve openings.

3.2.4.2.1 Low gas – low liquid flowrate bifurcation map using downcomer top pressure

From Figure 3-2 an operating condition (7 Sm^3/h gas and 0.5 kg/s of liquid) represented by point A was investigated to determine the stable and unstable operating regions in the riser and in the downcomer when varying the choke valve opening. Again, this helps in understanding the dynamic unstable slug flow behaviour in the U-shape riser system. Figure 3-25 - Figure 3-27 represent the bifurcation maps, a resultant of a parameter variation technique (using choke valve opening), obtained using the pressure at the top of the downcomer (downcomer top pressure), pressure at the base of the downcomer (downcomer base pressure) and the pressure at the base of the riser (riserbase pressure) respectively. Addressing the setbacks associated with measurement signals from the base of the riser or downcomer, the downcomer top pressure bifurcation map was of great interest as both base signals are not readily accessible especially for already existing fields.

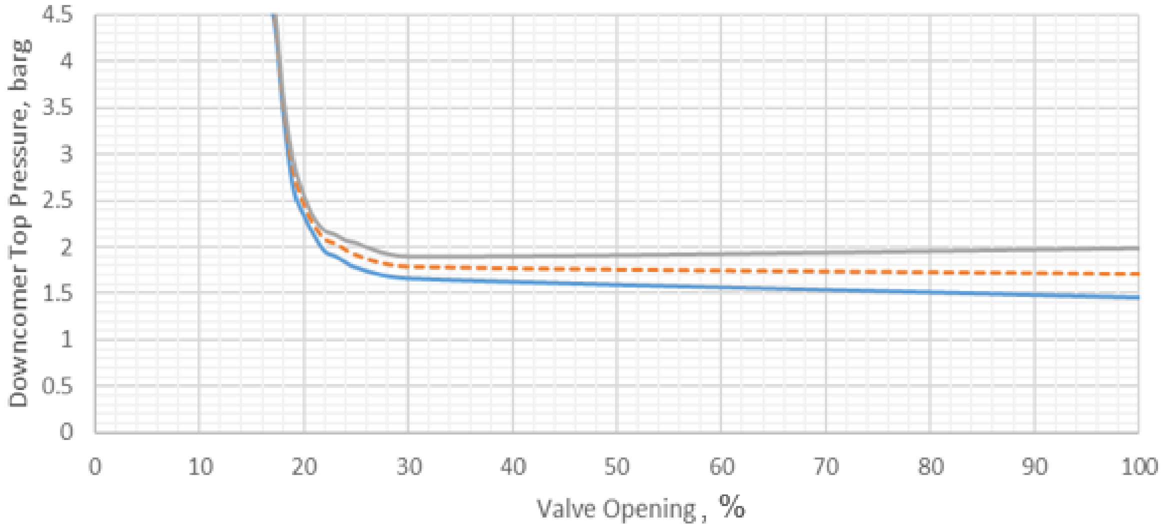


Figure 3-25 Downcomer top pressure bifurcation map (7 Sm^3/h gas and 0.5 kg/s of liquid)

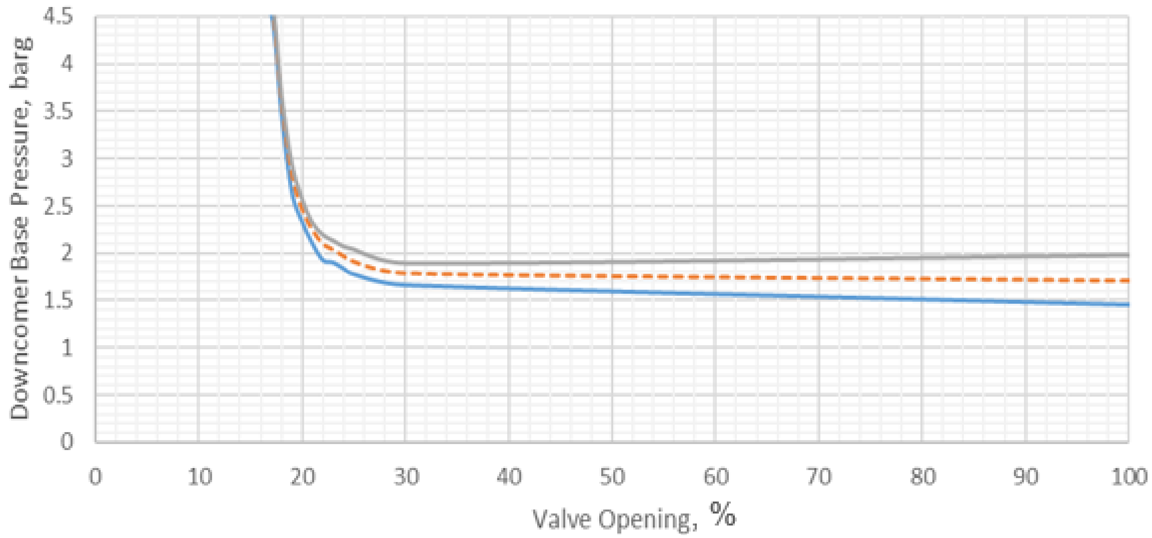


Figure 3-26 Downcomer base pressure bifurcation map (7 Sm^3/h gas and 0.5 kg/s of liquid)

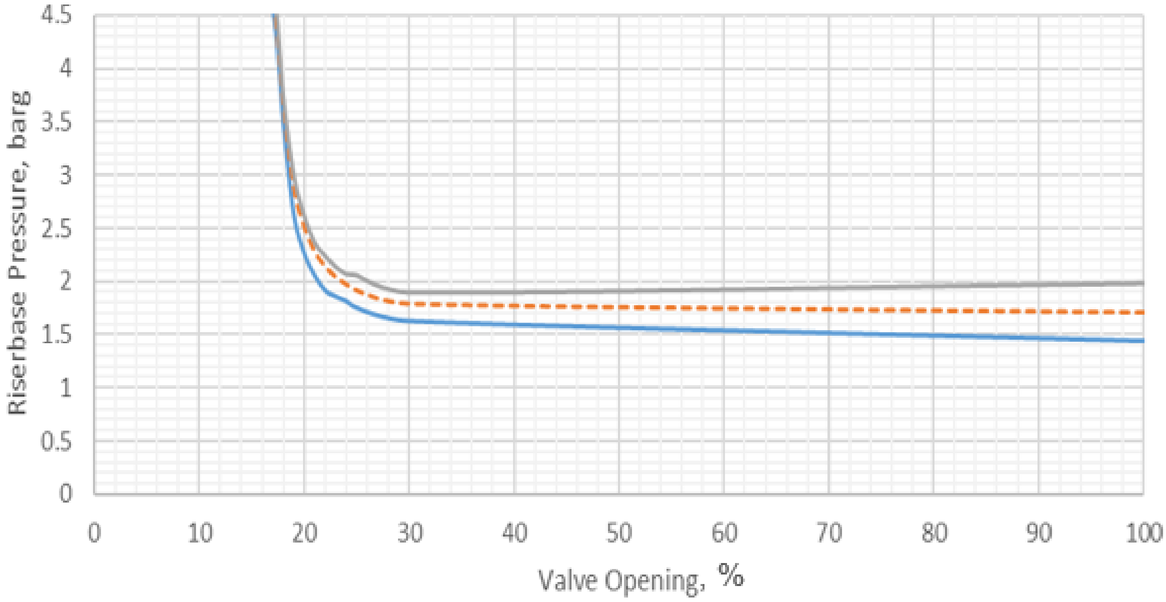


Figure 3-27 Riser base pressure bifurcation map (7 Sm^3/h gas and 0.5 kg/s of liquid)

From Figure 3-25, Figure 3-26 and Figure 3-27, 19 % choke valve opening was registered as the bifurcation point, thus, the point which transitions from the stable to unstable operation mode for the downcomer top pressure, the riserbase pressure and the downcomer base pressure bifurcation maps. This flow condition however represents a low gas – low liquid flowrate condition.

Beyond 19 % choke valve opening, thus, increasing the percentage valve opening, the system loses its stability while for valve openings below 19 % the system was stable. For the stable region, the maximum pressure curve and minimum pressure curve converge hence follow the same trend. On the contrary, for valve openings greater than 19 % valve opening, there was divergence in both the maximum and minimum pressure curves which signifies instabilities or oscillations in the system pressures (riserbase pressure, downcomer top pressure and downcomer base pressure). This explains why for all the pressure signals used the system losses its stability when the valve opening was greater than 19 %. Again, as observed from Figure 3-25, Figure 3-26 and Figure 3-27 the pressure measurement signals reduced as the valve opening reduced, signifying that reduction in the valve opening reduce the severity of the fluctuation in the pipeline system.

3.2.4.2.2 Medium gas – medium liquid flowrate bifurcation map using downcomer top pressure

Similarly, Figure 3-28, Figure 3-29 and Figure 3-30 show the bifurcation map obtained for the flow condition (30 Sm^3/h gas and 2 kg/s of liquid (Point B on Figure 3-2)) representing medium gas – medium liquid flowrate using the downcomer top pressures, downcomer base and riserbase pressure respectively.

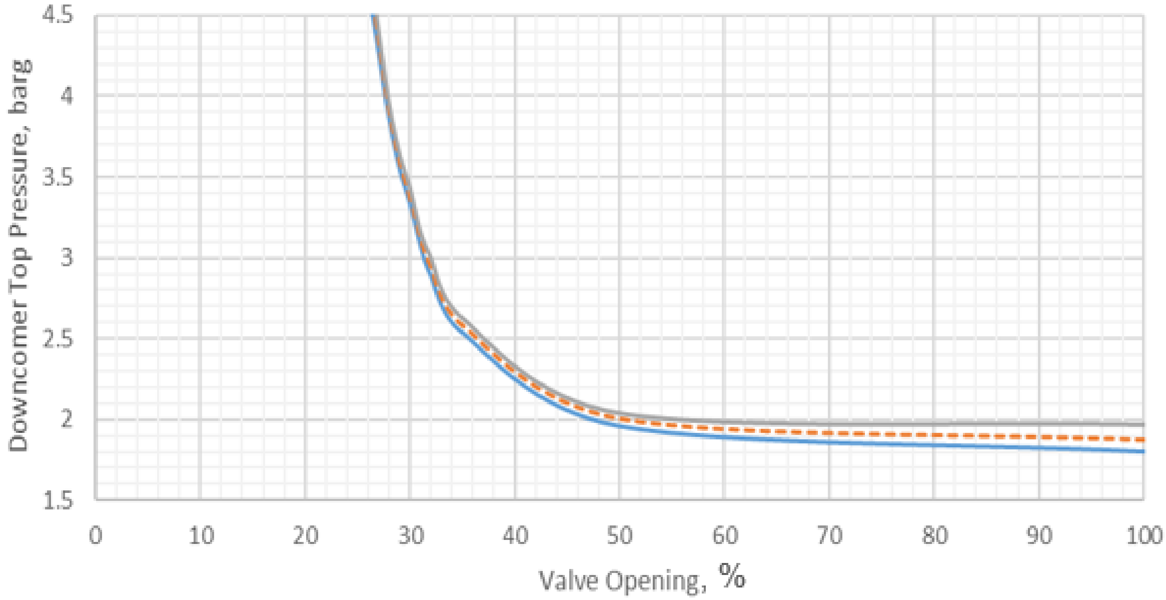


Figure 3-28 Downcomer top pressure bifurcation map (30 Sm^3/h gas and 2 kg/s of liquid)

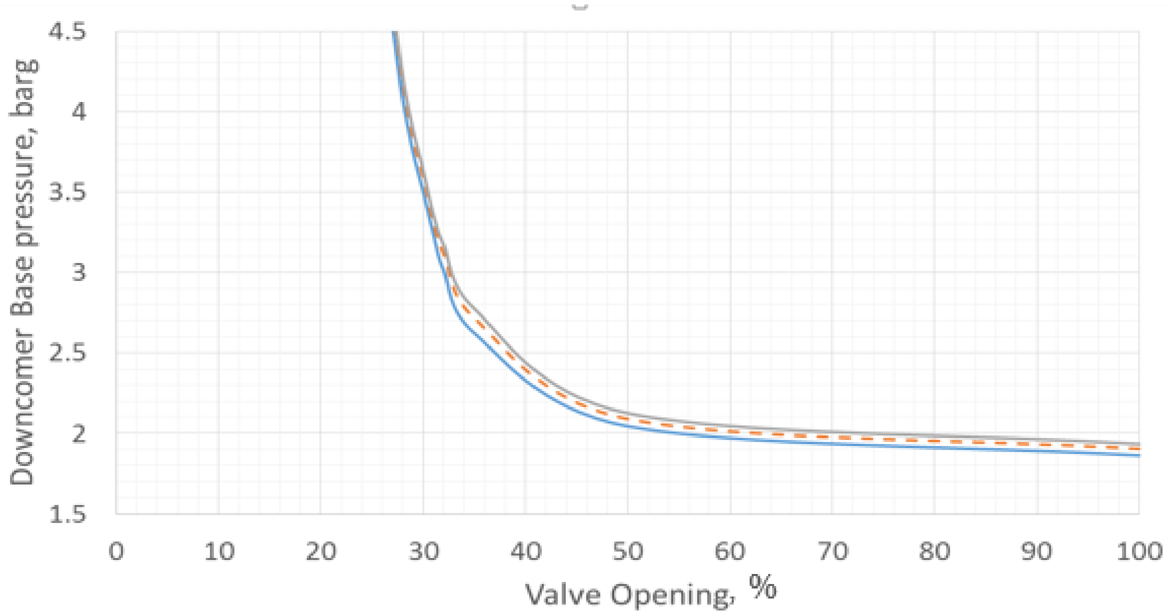


Figure 3-29 Downcomer base pressure bifurcation map (30 Sm^3/h gas and 2 kg/s of liquid)

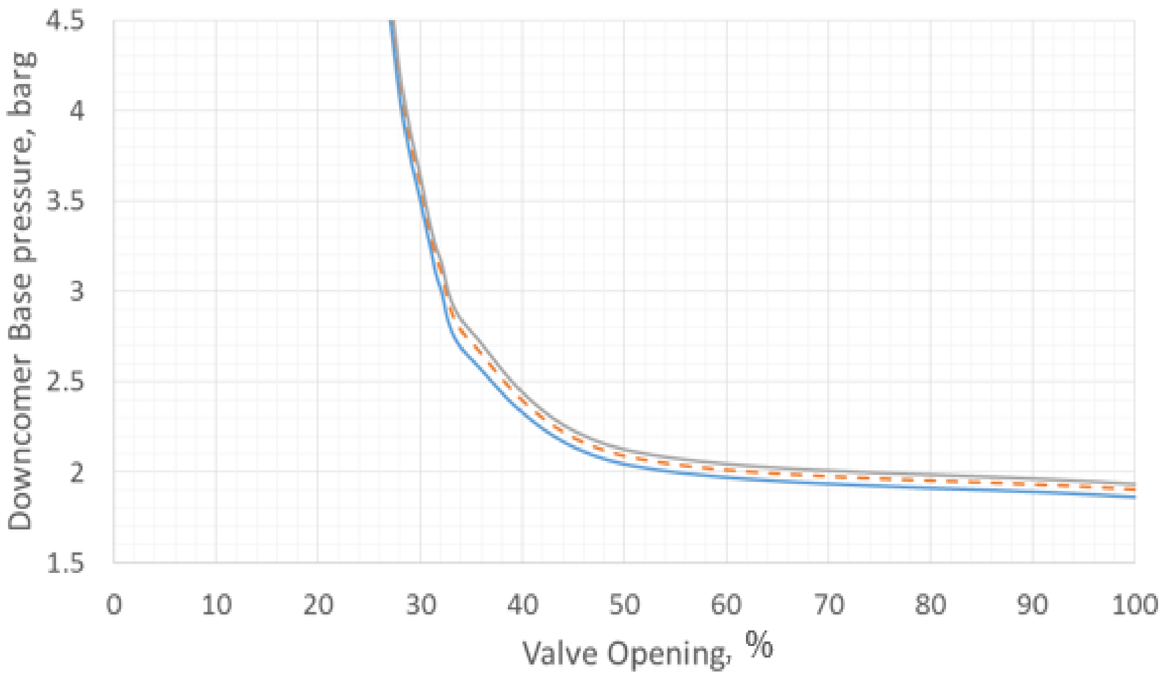


Figure 3-30 Riser base pressure bifurcation map (30 Sm³/h gas and 2 kg/s of liquid)

From Figure 3-28 and Figure 3-29, a bifurcation point of 31 % was observed for the downcomer base pressure and downcomer top pressure bifurcations. Again the riserbase pressure bifurcation map as shown in Figure 3-30 registered a bifurcation point of 29 % choke valve opening. This implies that the downcomer becomes stable at a 31 % choke valve opening however for the entire system to be stable the choke valve opening needed to be closed a further 2 % since the riserbase pressure only becomes stable at 29 % choke valve opening. The regions beyond these respective critical choke valve openings was considered an unstable region due to the oscillations in the pressures while the regions to the left side of these valve opening was considered stable. Comparatively, the pressures in the downcomer base for the different valve openings are high relative to that seen in the riserbase pressure. This could be associated with the horizontal section aiding accumulation of liquids at the base of the downcomer.

It could be deduced from this case that slugs in the riser could be formed as a result of the flow through the horizontal section of the pipeline and are not dependent on the downcomer of the U-shape riser. As shown in the medium gas – medium liquid flow condition, at 31 % valve opening, the downcomer was stable but the riserbase was not stable since it requires an extra 2 % valve closure for it to be stable as the stability point of the riserbase was at 29 % valve opening.

3.2.4.3 Validation of slug initiation in U-shape riser

From previous sections, it was observed that the flow conditions exhibiting slug flow in the downcomer also exhibited slugging in the riser, hence the validation of the slug initiation point on the U-shape riser. To confirm these findings, the same flow conditions as shown in Table 3-5 was run on a 2 inch purely vertical riser (without a downcomer) to investigate if these conditions also exhibit slugging in the purely vertical riser.

Table 3-5 Flow observations in a riser system for a 2 inch purely vertical riser

Liquid flowrate, kg/s			1.5	2	3	3.5
Liquid volumetric flowrate (m^3/s)			0.00150	0.00200	0.00301	0.00351
Superficial liquid velocity per second (m/s)			0.74	0.99	1.48	1.73
Gas Flowrate sm3/h	Gas Flowrate nm3/h	Superficial gas velocity (m/s)				
5.00	2.42	0.33	slug	slug		
10.00	4.83	0.66		slug	slug	slug
20.00	9.66	1.32		slug	slug	slug
30.00	14.50	1.99		slug	slug	slug
50.00	24.16	3.31		slug	slug	slug
70.00	33.83	4.64		slug	slug	slug
100.00	48.32	6.62		slug	slug	
120.00	57.99	7.95		slug	slug	
150.00	72.48	9.93		slug	slug	
200.00	96.65	13.24		slug	slug	
250.00	120.81	16.55		slug	slug	

Table 3-5, shows the observations made from the various flow conditions in a 2 inch purely vertical riser. Slug flow pattern was observed in the riser of the 2 inch purely vertical loop for the flow conditions that exhibited slugging flow in both the downcomer and the riser of the 2-inch U-shaped riser. This illustrates that the downcomer of the U-shape loop has minimum influence on the slugs produced in the riser of U-shape riser as represented on Figure 3-31.

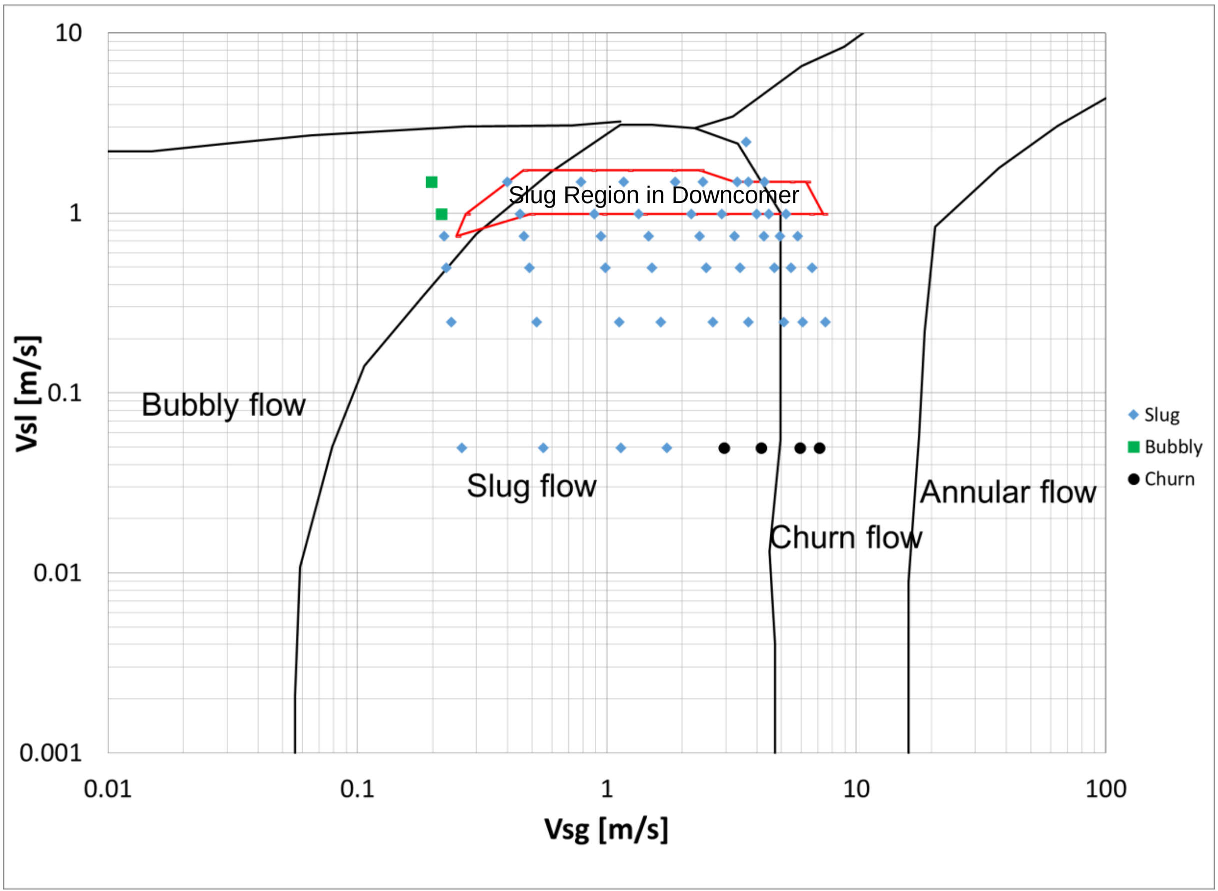


Figure 3-31 Slug region comparism for U-shape downcomer and a purely vertical riser system

From this section, it could be deduced that for flow conditions exhibiting slugging characteristics in the riser of the U-shape, its initiation may not necessarily be from the downcomer of the U-shape riser loop but from the horizontal section of the flow loop. This is shown in the study above as all the flow conditions showing a slug flow characteristics in both the riser and downcomer of the U-shape flow loop also showed slugging in the riser for a 2 inch purely vertical riser as shown in Figure 3-31.

3.3 Slug control using U-shape riser measurement signal

Active slug control, an active slug mitigation technique has been investigated in the past to understand this phenomena while improving this technology. Control based methods (Havre and Dalsmo, 2001; Storkaas, 2005; Sivertsen, Storkaas and Skogestad, 2010), active slug mitigation techniques, are known to be an effective approach to eliminate severe slugging flow in oil production riser-pipelines while maximising production at a larger valve opening. However, existing active slug control on most systems rely on information (measurements) which have to be obtained from the seabed, such as the riser base pressure. This motivated the innovation of the inferential slug controller (ISC), an active control scheme which combines several topside measurement signals to produce a single control variable which is more sensitive to slug flow through algebraic scheming. Expanding and improving the ISC has been the quest for this project.

This section discusses the use of measurement signals from the topside of pipeline riser systems to control and mitigate unstable slug flow condition in pipeline riser systems. To enable several measurement signals from the topside of a pipeline riser system to be deployed in control action for the ISC technology, the initial step was to use existing signals individually to assess its sensitivity to slug flow.

3.3.1 **Topside signal selection**

Several measurement signals exist on a pipeline riser system but the readily available signals on the 2 inch U-shape pipeline riser system include;

- Riser Outlet Pressure
- Downcomer Top Pressure
- Pressure Difference Across the Control Valve
- Riser Outlet Mass Flow Rate
- Riser Outlet Density
- Topside Two-Phase Separator Level
- Topside Two-Phase Separator Pressure
- Topside Two-Phase Separator Gas Outlet Flow Rate
- Topside Two-Phase Separator Liquid Outlet Flow Rate
- Three-Phase Separator Pressure

The most direct signals which exist in almost all risers were tested by Yeung, Cao and Lao, (2008) and was established that both the riser outlet pressure and the pressure differential across the control valve were good measurements for slug control in risers of shorter lengths. However, this was not true for higher riser systems, as the riserbase pressure signal exhibits better performance relative to that of the riser outlet pressure.

Again, even though Shell spent considerable effort in developing the S3 and the vessel less system (Kovalev, Cruickshank and Purvis, 2003; Kovalev, Seelen and Haandrikman, 2004), in the drive to establish the flowrate at the topside of the riser as a good measurement for slug control, compared to the riserbase pressure, the topside flowrate measurement signal would not be a good measure for the experiment as there is a stable reading for this signal due to a well-mixed flow of both gas and liquid. In this section however, it is attempted to compare the control performance of both the downcomer top pressure of the 2 inch U-shape riser system and the riser outlet pressure. The findings could possibly have consequences on the configuration and further expansion on the ISC technique.

3.3.2 **Slug control using riser outlet pressure**

The first step undertaken was to test the controller's ability to stabilise unstable slug flow using the riser outlet pressure at a desired flow condition. 0.5 kg/s liquid and 7 Sm^3/h gas on the 2 inch U-shape riser system from the previous sections was observed to exhibit unstable slug behaviour. From the stability analysis performed on this flow condition in open loop, the system achieved stability at a 19 % choke valve opening hence the need to stabilise the flow further in the unstable open loop region, thus at a larger valve opening using closed loop.

One such solution with proven application as an effective technique in mitigating slug /stabilise flow is the use of a PI controller. The implementation of this PI controller using the riser outlet pressure as the control variable would help stabilise the flow. The controller design and implementation will be discussed next.

3.3.2.1 Controller design

This section outlines the method of designing the PI controller, using an open loop tuning method to achieve system stability. Tuning the controller parameter is very important to achieve a stable system. For this study the open loop tuning method using the reaction curve of the entire system was used in tuning the controller. The PI controller equation is given by

$$u(t) = \left(K_p e^{(t)} + K_I \int_{t_0}^{t} e^{(t)} dt \right) \qquad \textbf{(3-1)}$$

where u is the output of the controller, t is the time, t_o is the initial time, the first term on the right of the equation representing the proportional term (P) while the second term represents the integral term (I).

To find the PI controller parameters, the choke valve opening was used to produce a step response in the riserbase pressure. A step response from a step change in the valve opening from 18 % to 19% (operating point before the critical bifurcation point) would be used to tune the controller.

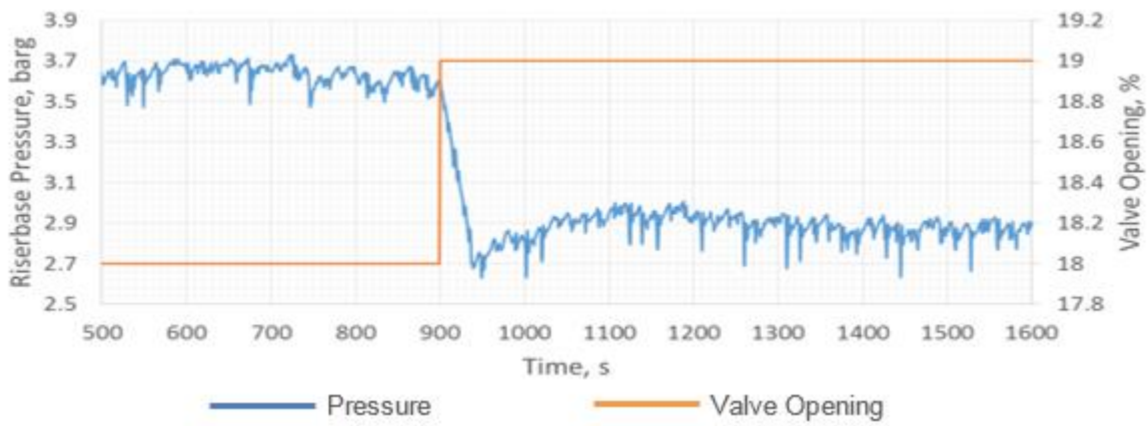


Figure 3-32 Approximate first order plus time delay of the riser system

From the step response, 18 % valve opening corresponds to 3.634 barg which then drops to 2.890 barg at a choke valve opening of 19 %. This response has a similarity to a first order system behaviour hence the PI parameters could be determined using a First Order Plus Time Delay (FOPDT) model. The time constant, τ, time delay, t_{delay}, and gain were determined from a reaction curve in order to approximate the system as a FOPDT model. Smith and Corripio, (1987) presented expressions for determining the time constant, time delay and the gain which is given by

$$K = \frac{Change\ in\ output, (\Delta c)}{change\ in\ input\ (step\ input), (\Delta m)} \qquad \textbf{(3-2)}$$

$$\tau = \frac{3}{2}(t_2 - t_1) \quad \textbf{(3-3)}$$

$$t_{delay} = t_2 - \tau \quad \textbf{(3-4)}$$

where t_1 is the time for the output to reach 0.283*Δc and t_2 is the time for the output to reach 0.632*Δc. From (3-2), (3-3) and (3-4) the gain, time constant and time delay was determined as

$\boldsymbol{K} =$ **74.377**

$\boldsymbol{\tau} =$ **27**

$\boldsymbol{t_{delay}} =$ **1**

3.3.2.1.1 Controller parameters

The Ziegler Nichols open loop tuning table was used in calculating the PI controller parameters based on the variables in the FOPDT, described above. The PI parameters obtained from the table is given by

P= 0.3267

I= 3.33

From the experimental point of view when both values were introduced into the controller block, the P value was introduced with a sign change due to the error being process set point minus the actual value measured. The Ziegler Nichols open loop tuning table is shown in the appendix C.

3.3.2.1.2 Controller implementation

The designed PI controller was tuned and implemented to assess the ability of this controller to stabilise unstable slug flow using the riser outlet pressure as the control variable. Figure 3-33 and Figure 3-34 show the riserbase pressure trend and the downcomer top pressure trend in closed loop stable condition using the riser outlet pressure signal as the control variable. From Figure 3-33 and Figure 3-34, the controller was activated after 900 seconds. Shortly after the controller was engaged the pressure (riserbase and downcomer top pressures) responses stabilises thus the pressure fluctuations / oscillations reduces significantly. Thus for both riserbase pressure and the downcomer top pressure trends, in some few seconds the flow became stable. The magnitude of the differential pressure oscillation in both the riser and the downcomer reduced drastically. Again, the average valve opening registered was 21.69 % which falls under an unstable region when the system was in open loop.

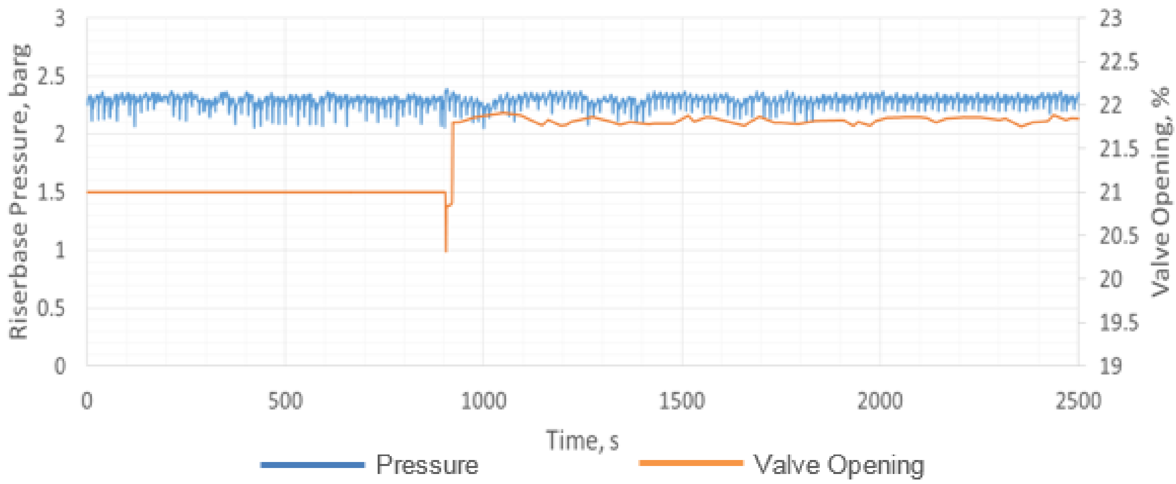


Figure 3-33 Riserbase pressure trend of the system with riser outlet pressure as the control variable

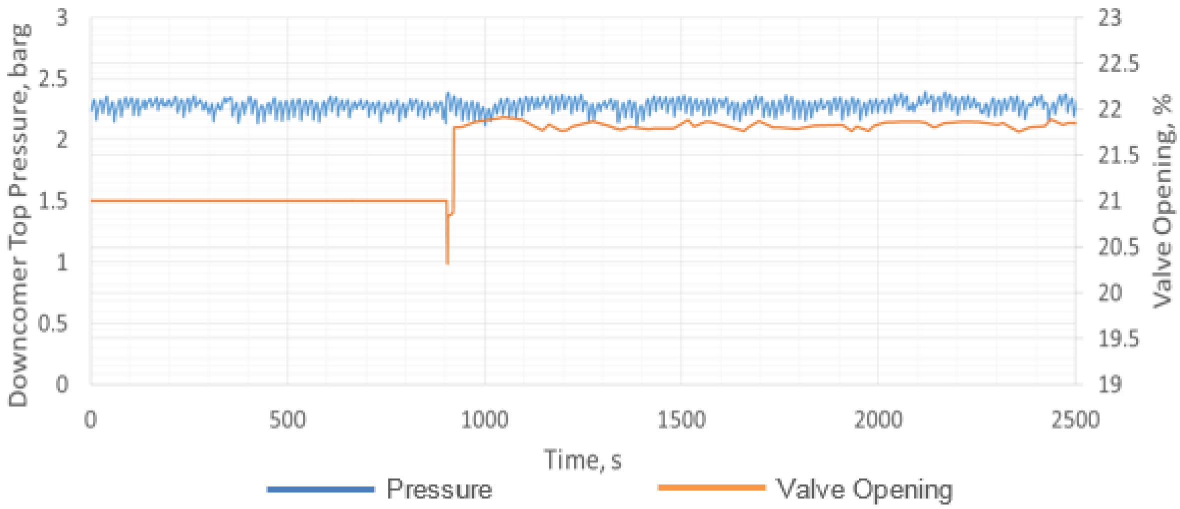


Figure 3-34 Downcomer top pressure trend of the system with riser outlet pressure as the control variable

To optimize the controller's stabilising ability, valve openings located further in the unstable region were investigated. The optimal valve opening for which the system becomes stable was observed to be 25 % corresponding to average pressures of 1.8184 bar and 1.8116 bar for both riserbase and downcomer top. Beyond 25 % valve opening, at 26 % the system becomes unstable as shown in Figure 3-35 and Figure 3-36.

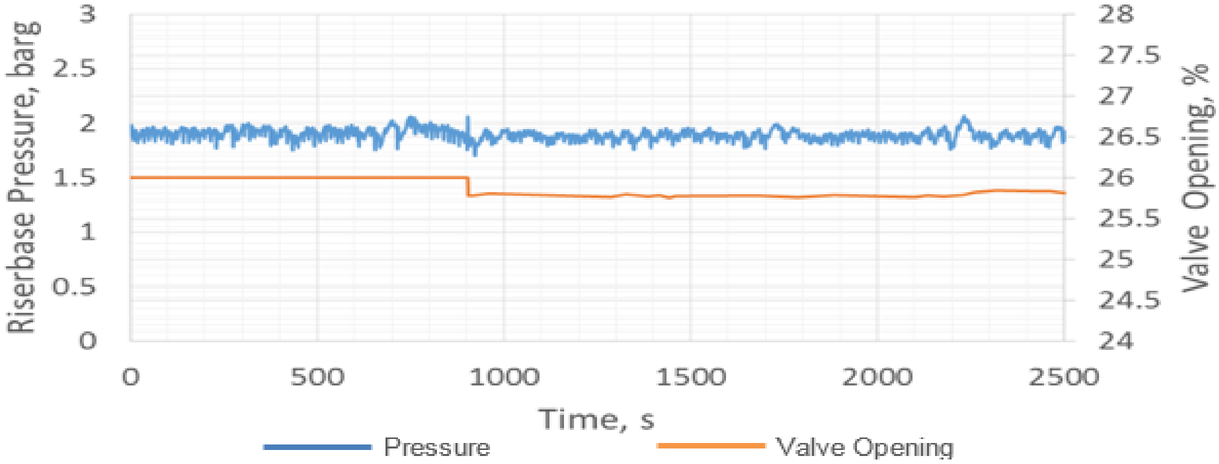


Figure 3-35 Unstable riserbase pressure trend of the system with riser outlet pressure as the control variable

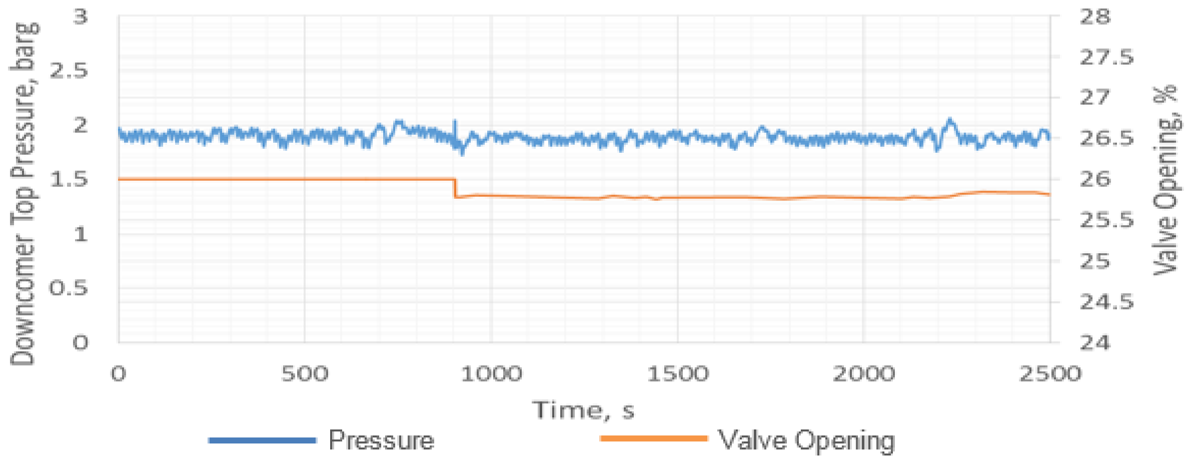


Figure 3-36 Unstable downcomer top pressure trend of the system with riser outlet pressure as the control variable

3.3.3 Slug control using downcomer top pressure

Similar to slugging flow control using the riser outlet pressure, the downcomer top pressure signal was able to stabilize the slug flow condition (0.5 kg/s and 7 Sm^3/h of liquid and gas flowrate). The controller design to stabilise the unstable slug flow using the downcomer top pressure as the control variable was done using the method adopted when the riser top pressure was used as the control variable. From (3-2), (3-3) and (3-4) the gain, time constant and time delay was determined as

$\boldsymbol{K} = \mathbf{74.353}$

$\boldsymbol{\tau} = \mathbf{85.84}$

$t_{delay} = 1.24$

The PI parameters obtained from the Ziegler Nichols open loop tuning table is given by

P= 0.838

I= 4.129

From the experimental point of view when both values were introduced into the controller block and the controller's ability to eliminate the instabilities were assessed.

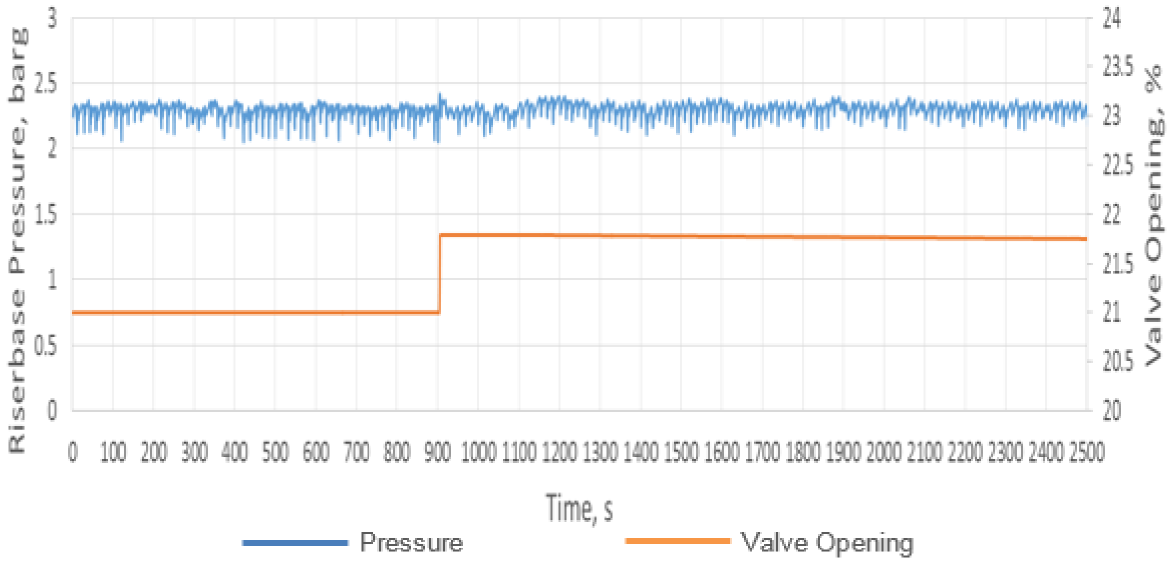


Figure 3-37 Stable riserbase pressure trend of the system with downcomer top pressure as the control variable

An attempt to stabilise the flow at 21 % choke valve opening using the controller was designed. From Figure 3-37, the pressure fluctuations reduce significantly after the controller was triggered, thus at 900 seconds. That is shortly after 900 seconds, when the controller was activated, the fluctuation in the riserbase reduces, signifying flow stability. The average corresponding valve opening for which the system becomes stable is 21.745 % which falls within the unstable region in the open loop.

Furthermore, the optimal valve opening for which the system was stable was observed to be 25 %. This corresponds to pressures of 1.8233 bar and 1.8156 bar for the riserbase and downcomer top pressures respectively. At 26 % valve opening, as shown in Figure 3-38 and Figure 3-39, both the riserbase pressure and the downcomer pressure trends are unstable. Thus there exists a continuous fluctuation of the pressure signals in both the riserbase and the downcomer top when the choke valve was made to operate at 26 % opening in closed loop.

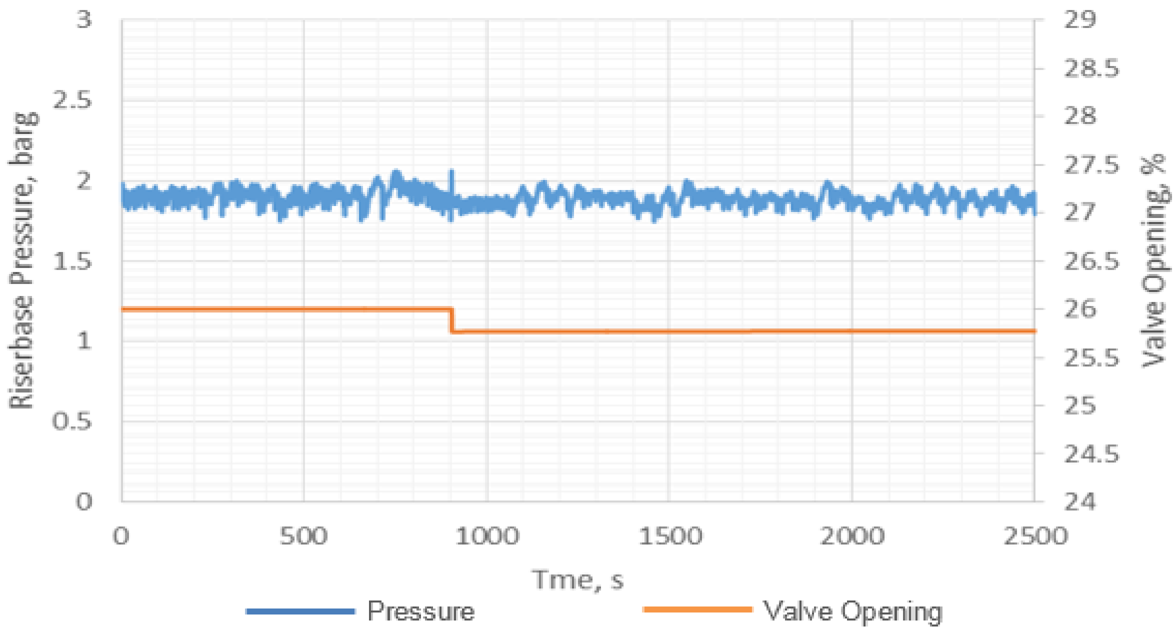


Figure 3-38 Unstable riserbase pressure trend of the system with downcomer top pressure as the control variable

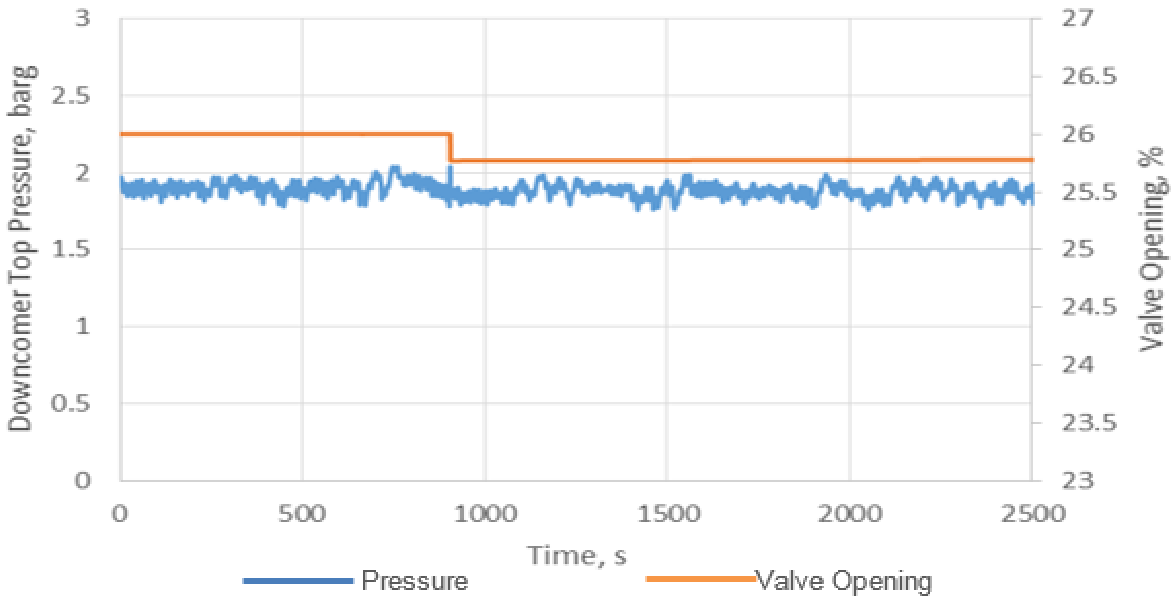


Figure 3-39 Unstable downcomer top pressure trend of the system with downcomer top pressure as the control variable

3.4 Slug control in U-shape riser using multiple topside measurements (ISC)

Slug control, a form of active slug mitigation has been on a rise since the identification of severe slugging in the oil and gas industry. In active control, the most important aspect is to know the optimum measurement signals (control variables) that can subdue the slugging flow regime. Some measurement signals' capability to control slug have been tested individually by Cao, Yeung and Lao, (2008) and it was established that both the riser outlet pressure and the pressure drop across the control valve were good measurement signals for slug control in risers of shorter lengths. Although the riserbase pressure has been recognised as the best measurement for slug control, it is not readily available in most already existing offshore configurations and the high capital and operational cost in obtaining measurement from the seabed. The setback associated with this technique was the readily availability of the measurement at the riser base. Recently, a patent (Cao, Yeung and Lao, 2010b) proposed a novel severe slug mitigation method using an Inferential Slug Controller (ISC) as a theoretically and practically sound solution, which uses topside measurement in the quest to eliminate the setbacks resulting from the use of riser base pressure. Where possible all available and relevant topside measurements are combined to form a single controlled variable, also called inferential slug index. When this slug index is maintained smoothly, the overall slugging flow is eliminated.

Although coefficients of measurement combination of the ISC can be systematically determined, the control gain of ISC is however determined by trial and error. On the other hand, a systematic approach was proposed in (Ehinmowo and Cao, 2015) to find the minimum gain for a slug controller based on outlet gas flow rate although such a measurement is generally not available in a practical system. This technique would be adopted and extended to design a controller gain of the ISC to improve the controller robustness and to make the method more realistic for actual systems.

3.4.1 Stabilising unstable slug flow behaviour

Generally, instabilities in a multiphase pipeline system are as a result of the upward flow in the system, compressibility and expansion of the gas. Due to the compressibility and expansibility nature of gases, an increment in the gas flow of a system poses two distinct effects on the riserbase pressure. That is, it either affects the system positively or negatively. Positively, it can make the system stable while on the contrary it can make the system unstable if it is dominant. Figure 3-40 shows for a specific liquid flowrate, the relationship between riserbase pressure and an increase in gas flowrate.

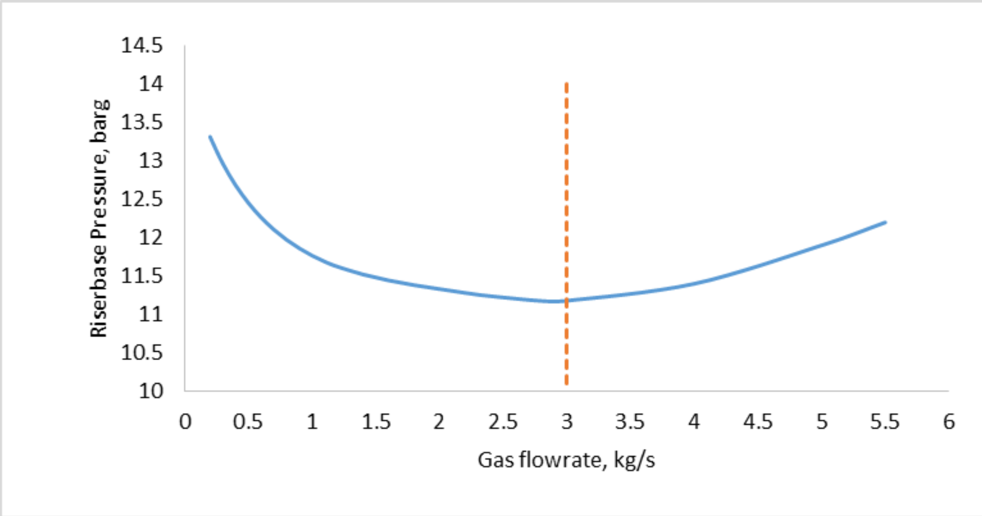


Figure 3-40 Riserbase pressure as a function of gas flow rate

At low gas flowrate corresponding to low frictional loss, an increase in the gas flowrate increases the gas –liquid ratio in the system translating to a decrease in the riserbase pressure as shown on the left side of the red line from Figure 3-40. On the other hand, at large gas flowrate there exists a dominating frictional loss, hence any additional gas flowrate would result in an increase in the frictional loss which in tend increases the riserbase pressure (right hand side of the red line in Figure 3-40) (Ehinmowo and Cao, 2015). Thus for the riser system to be stable, the riserbase pressure gradient (the riserbase response to a change in gas flowrate) must be positive. Again, when the system is unstable, the riserbase pressure slope is negative. This means that the riser system can only be stable at a considerably high gas flowrate and that explains the mitigation potential of this method. The lowest point on this curve is the minimum gas flowrate that could make the system stable for the set of boundary conditions. However better options are required to optimize the process effectively.

The riserbase pressure of a pipeline riser system considerably depends on the frictional head of the pipeline, acceleration head of the fluid, hydrostatic head or liquid head in the pipeline, the pressure drop across the choke valve and the pressure of the separator. For a pipeline riser system with flow conditions exhibiting unstable flow pattern, the system can be stabilised by choking. Choking the topside valve of the riser system increases the pressure drop across the valve which in tend increases the riserbase pressure of the system. A sufficiently large enough pressure drop across the valve can potentially make the negative slope region become positive.

The pressure drop across the choke valve of a riser pipeline system is dependent on the valve opening and the fluid flowrate through the system as shown in (3-5) assuming a linear valve characteristics.

$$\Delta P_V = a\frac{Q^2}{u^2} \tag{3-5}$$

where Q is the flowrate through the valve, u is the valve opening and 'a' is a factor dependent on the discharge coefficient, C_v and the density of the fluid, ρ, $(a = \frac{1}{\rho {C_v}^2})$.

For manual operations (choking), the valve opening is the only manipulating variable at a constant fluid flowrate. Choking the flow places a restriction on its passage through the choke valve which results to a reduction in the acceleration term during transient flow conditions, thus, there is a deceleration of flow through the choke valve. As discussed earlier, the increase in pressure of the pipeline riser system, a resultant of choking reduces the throughput of the system. Hence to increase production, a reduction in the pressure drop across the choke valve is preferred. Active control provides a solution to this effect. The controller design technique adopted in this study will be discussed next.

3.4.2 Controller design

3.4.2.1 Unstable slug flow stability

The riserbase pressure of a pipeline riser system affects its throughput while the stability is based on the riserbase gradient of the system. Therefore, the aim of designing the controller is to find a controller gain that can possibly provide a positive riserbase pressure gradient for certain flow conditions that exhibit an unstable nature in open loop. This would in effect provide a relatively lower riserbase pressure in terms of magnitude and oscillations.

For any parameter (process variable) of interest, a slight change in this parameter would drive the actual output to the set point using a feedback controller. (3-6) shows a simple feedback controller equation.

$$u = u_o + Ks \tag{3-6}$$

where u is the process output, u_o is the valve nominal value, K is the gain of the process and s is the error (set point minus controlled variable) given by Q_o-Q assuming the gas flowrate is the parameter of interest.

Using a feedback controller, the system is aimed at achieving stability at a larger valve opening relative to that observed in open loop. This however requires an additional gradient through feedback control to make up for the deficiency in gradient, a resultant of the larger valve opening.

The riserbase pressure (P_{RB}) of a pipeline riser system considerably depends on the frictional head of the pipeline (ΔP_f), acceleration head of the fluid (ΔP_a), hydrostatic head or liquid head in the pipeline (ΔP_h), the pressure drop across the choke valve (ΔP_V), and the pressure of the separator (P_s).

$$P_{RB} = \Delta P_P + \Delta P_V \tag{3-7}$$

where P_p is the riserbase pressure contribution from the frictional components (summation of static head of liquid, acceleration head, pressure drop across the valve, separator pressure and frictional head).

For a constant liquid flow rate through a multiphase flow system, a slight perturbation in the gas flowrate from (3-7) results to

$$\frac{dP_{RB}}{dQ} = \frac{d\Delta P_P}{dQ} + \frac{d\Delta P_V}{dQ} \tag{3-8}$$

From Figure 3-40, for the system to be stable, $\frac{dP_{RB}}{dQ} > 0$, thus the riserbase pressure gradient must be positive whiles $\frac{dP_{RB}}{dQ} < 0$ represents an unstable system. A negative riserbase gradient translates to an unstable flow. Thus, the right hand side of (3-8) must be greater than zero for the system to be stable. Mathematically,

$$\frac{d\Delta P_P}{dQ} + \frac{d\Delta P_V}{dQ} > 0$$

This implies that for stability to happen,

$$\left(\frac{d\Delta P_V}{dQ}\right) > -\left(\frac{d\Delta P_P}{dQ}\right) \qquad \textbf{(3-9)}$$

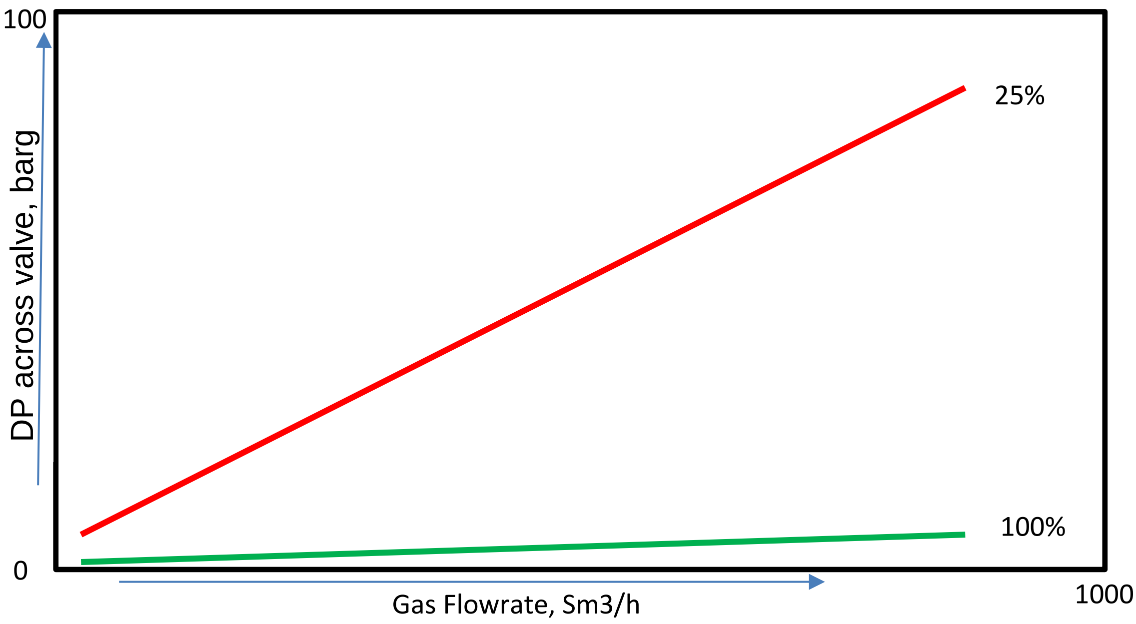


Figure 3-41 Pressure drop across valve as a function of gas flow rate

Figure 3-41 shows a pressure drop across the valve response with increasing gas flowrate for a constant liquid flowrate and valve opening. Pressure drop across the choke valve of the system was observed to increase as the gas flowrate increases for a constant valve opening. This shows that the pressure drop across the choke valve is dependent on the flow through the system which is a function of the valve opening at any particular time.

In manual control, the only manipulating variable is the valve opening and this technique has been explored by several researchers. A differentiation of (3-5) with respect to the gas flowrate yields

$$\frac{d\Delta P_v}{dQ} = a\frac{2Q}{u^2} \qquad \textbf{(3-10)}$$

Substituting (3-9) into (3-10) yields,

$$\left(a\frac{2Q}{u^2}\right) > -\left(\frac{d\Delta P_P}{dQ}\right) \tag{3-11}$$

(3-11) satisfies the condition for which an unstable flow condition would become stable under manual choking (using a choke valve) when the gas flow is perturbed. From (3-11), the left hand side of the equation (pressure drop across the valve) must be large enough which implies that the valve opening (u) must be relatively small resulting to a low throughput. This reduces the flow acceleration term thereby resulting to a reduction in flow speed and contributes greatly to the increase in the riserbase pressure in order to stabilise flow. For this reason, this technique of slug mitigation has been explored by several researchers. However a reduction in the pressure drop across the valve is always desirable as this would aid or boost production output. Regardless, without boosting production outputs while mitigating unstable slug flow, there exists the benefit of improved operations hence enabling production without huge slug catchers especially on offshore facilities. As established by several researchers, active control can help in this quest.

3.4.2.2 Active control of unstable slug flow

As discussed in previous sections, the throughput of the system is dependent on the riserbase pressure while the riserbase pressure gradient depicts the stability of the system. Under manual choking (manual control) the pressure gradient is dependent on the valve opening only since the only degree of freedom is by manipulating the choke valve as shown in (3-10). However under feedback control, the pressure drop across the valve is dependent on both the gas flowrate, Q and the valve opening, u. The valve opening, u varies with varying gas flowrate depending on the designed feedback control law. The aim of active control is to achieve a positive gradient, dP_{RB}/dQ for a given flow condition with a relative low pressure at the base of the riser. Under feedback control law in (3-6), a differentiation of (3-5) yields,

$$\frac{d\Delta P_V}{dQ} = \frac{\partial \Delta P_v}{\partial Q} + \frac{\partial \Delta P_v}{\partial u}.\frac{du}{dQ}$$

$$\frac{d\Delta P_V}{dQ} = a\frac{2Q}{u^2} + aQ^2\left(\frac{\partial}{\partial u}\left(\frac{1}{u^2}\right)\right).\frac{du}{dQ} \tag{3-12}$$

Comparatively, to maximise production under slugging conditions while stabilising flow, the second term of (3-12) must be optimised. The second term provides the extra gradient needed to accomplish a stable flow at a larger valve opening relative to manual choking.

3.4.2.3 Controller design

Operating at a larger valve opening relative to the critical point for which the system achieves stability in open loop renders a loss in gradient of the system. To achieve stability at this point, an extra gradient is required to make up for the loss resulting from the larger valve opening. From (3-6), for any parameter (process variable) of interest, a slight change in this parameter would drive the actual output to the set point using a feedback controller, $\boldsymbol{u} = u_o + K(\mathrm{Qo} - \mathrm{Q}\,)$.

But $\frac{\boldsymbol{du}}{\boldsymbol{dQ}}$ is dependent on the control variable, 'c' and the valve opening, u.

Hence $\frac{\boldsymbol{du}}{\boldsymbol{dQ}} = \frac{dc}{dQ} * \frac{du}{dc}$.

For the ISC technology, since the control variable is dependent on a combination of a number of signals from the topside of the riser, $C = w_1 y_1 + w_2 y_2 + w_3 y_3 + w_4 y_4$

$$c = W^T Y$$

$$\frac{\partial c}{\partial Q} = w_1 \frac{\partial y_1}{\partial Q} + w_2 \frac{\partial y_2}{\partial Q} + w_3 \frac{\partial y_3}{\partial Q} + w_4 \frac{\partial y_4}{\partial Q}$$

Thus

$$\frac{dc}{dQ} = W^T \frac{dY_i}{dQ}$$

$\frac{\partial c}{\partial Q}$ is estimated from the weighted deviations in measurements resulting from a perturbation in Q.

Therefore (3-12) becomes

$$\frac{d\Delta P_v}{dQ} = \frac{2aQ}{u^2} + \frac{2aQ^2}{u^3} K \left[\frac{\partial c}{\partial Q}\right]$$

$$\frac{d\Delta P_v}{dQ} = \frac{2aQ}{u^2} + \frac{2aQ^2}{u^3} K \left[W^T \frac{dY}{dQ} \right] \quad \textbf{(3-13)}$$

Therefore the stability condition for feedback control is given by

$$\frac{2aQ}{u^2} + \frac{2aQ^2}{u^3} K \left[W^T \frac{dY}{dQ} \right] > -\frac{d\Delta P_p}{dQ} \quad \textbf{(3-14)}$$

At a given valve opening, u (relatively larger valve opening compared to manual choking), there exists a minimum controller gain, K which would produce a desired $\frac{d\Delta P_v}{dQ}$ which in effect stabilises the system. Optimizing the gain value, K could be used to increase the throughput of the system on the whole.

3.4.2.4 Inferential slug controller (ISC) overview

The novel ISC proposed in (Cao, Yeung and Lao, 2010a) uses multiple topside measurements as the input to attain a percentage choke valve opening that stabilizes the flow within the slugging regime in riser systems to minimize the impact of slug on the overall production and avoid over choking. The choke valve position function from the control equation is given in the form;

$$u = u_o + K(W^T Y - R) \quad \textbf{(3-15)}$$

where u_o is the choke valve nominal value, K is the control gain which may be tuned using any available tuning technique that stabilises the flow fluctuation, W is the vector of measurement weights which is determined from samples of signals obtained over a long period of time usually more than two slug cycles when there is no controller in action, Y is the vector of measurements, W^TY is the control variable which may represent a principal component which is a linear combination of the weighted variables, and R is the set point of the control variable.

ISC creates a single controlled variable by combining several measurements to obtain a control variable which is relatively more sensitive to slug flow. ISC controls the valve openings by interpreting a combination of signals obtained from the topside through Principal component analysis (PCA) techniques.

3.4.3 **Case study**

3.4.3.1 Pipeline riser configuration

Olga simulation software was used to investigate the use of the ISC to control slug in a U-shape pipeline riser system. The U-shape riser model (B-J1) used for this study is a sub model from a satellite field courtesy Chevron Energy Technology Company. Basically the U-shape riser is divided into three sections namely the horizontal section, the down comer and the riser section which is equipped with a choke valve prior to the separator. The riser pipeline model has a diameter of 0.289 m. A schematic of the U-shape riser model is shown in Figure 3-42.

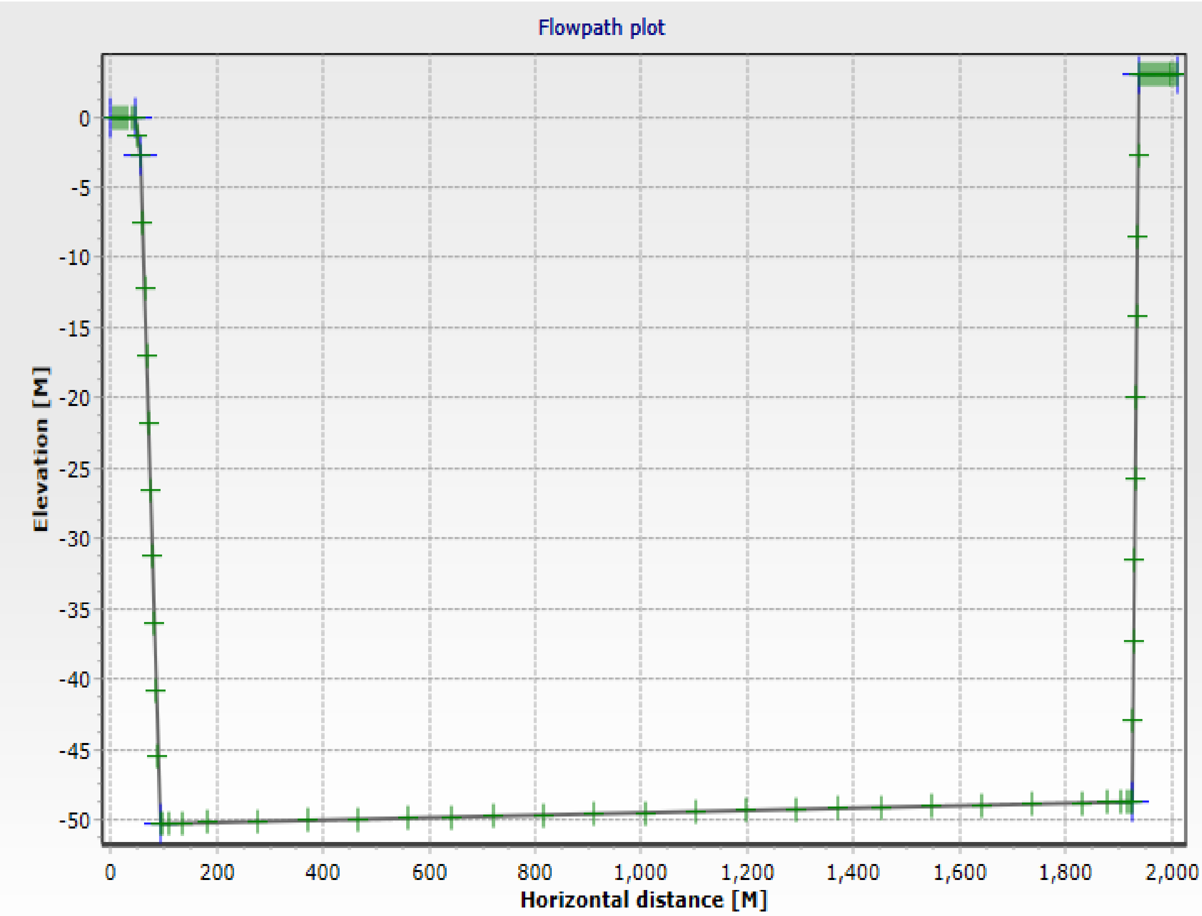


Figure 3-42 Schematic of the U-shaped pipeline system

With the point of interest being slugging flow, the operating conditions were allowed to operate in the slug region. A combination and variation of inlet conditions were investigated to develop a slug envelope. From the slug envelope, a slugging characteristic flow boundary condition as shown in Table 3-6 was used for the study.

Table 3-6 Operating conditions and parameters for the U-shape riser (B-J1)

Operating Parameters (Initial Condition)						
Source Temp. (°C)	Ambient Temp. (°C)	U value (Wat/(m*K))	Inlet Temp. (°C)	Outlet Temp. (°C)	Inlet Pressure (bar)	Outlet Pressure (bar)
44.44	23.89	69.18	47.22	47.22	12.41	12.41
Operating Conditions						
Total Mass Flow, kg/s	Gas Mass Fraction	Oil Mass Fraction	Water Mass Fraction	Inlet Temp. (°C)	Outlet Temp. (°C)	Outlet Pressure (bar)
11	0.02	0.49	0.49	44.4	23.89	10.687

3.4.3.2 Open loop system stability (Manual Choking)

Using the boundary condition shown in Table 3-6, the resulting bifurcation map is shown in Figure 3-43. From Figure 3-43, a critical bifurcation point of 5 % valve opening was achieved for the slugging flow condition above corresponding to a riserbase pressure of 16 bara. Further than the 5 % valve opening, any increase in the independent variable of the system (Valve opening) renders the system unstable. Thus beyond 5 % valve opening, a pair of complex poles crosses the imaginary axis on the S-plane which however changes the sign of the real part of the pole from negative to positive. This valve opening functions as the reference beyond which the controller to stabilize flow in an open loop unstable region would be designed, which would be discussed next.

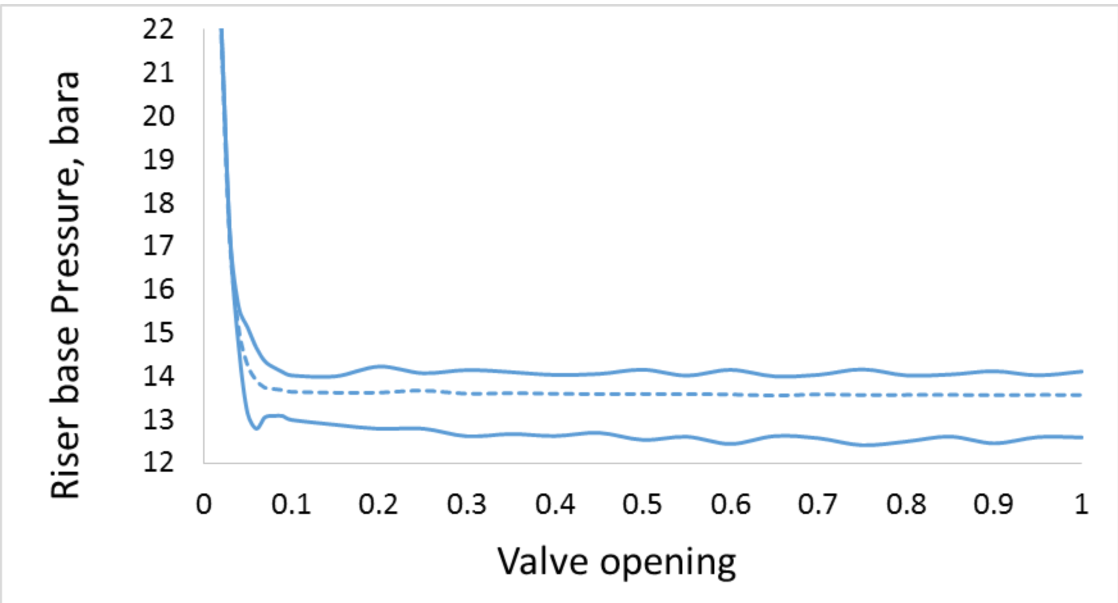


Figure 3-43 Bifurcation map for the U-shape riser

3.4.3.3 Pressure gradient and closed loop system stability

3.4.3.3.1 Pressure gradient

The stability of the pipeline riser system could be determined from the pressure gradient after a slight perturbation in the gas flowrate for constant liquid flowrate. Figure 3-44 shows the riserbase pressure response for the U-shape pipeline for the parameters in Table 3-6, with increasing gas flow flowrate at a 100 % valve opening.

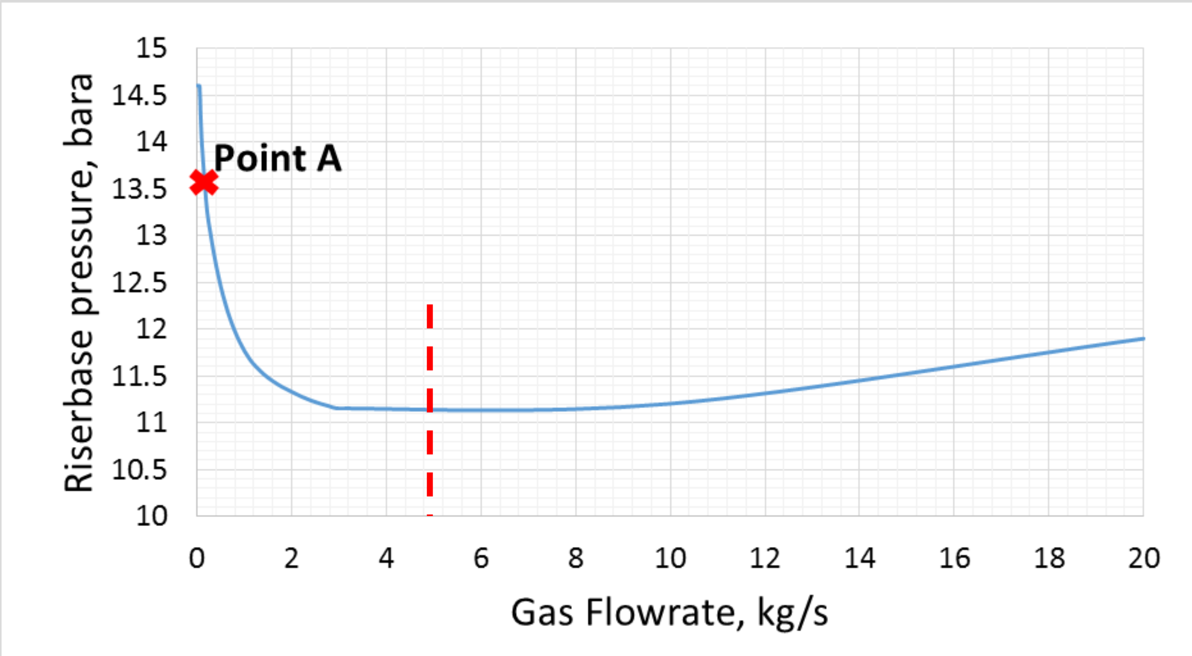


Figure 3-44 Riser base pressure response with increasing gas flowrate at 100 % valve opening

For the flow condition shown in Table 3-6 (marked as point A), the resulting pressure gradient from the riserbase pressure – gas flowrate curve shown in Figure 3-44 is -5 bar/kgs^{-1} after a 1 % perturbation in the gas flowrate. From the determined pressure gradient, it could be deduced that the system requires a minimum of 5 bar/kgs^{-1} gradient to render the system stable. Thus, a minimum of that pressure gradient needs to be supplied to the system to achieve stability.

3.4.3.3.2 Closed loop stability

In closed loop, it was purposed to stabilise the system in an open loop unstable region. Since stability was achieved at 5 % valve opening in open loop, it was aimed to stabilise the system at a valve opening of 7 %. The resulting pressure gradient for this boundary condition produces a gradient of – 2.45 bar/kgs^{-1} after a 1 % increase in the gas flowrate.

By increasing the valve opening to 7 %, the gradient supplied however was less than the gradient required to stabilise the flow at a 100 % valve opening. This infers that for the system to stabilise, an additional gradient of 2.45 bar/kgs^{-1} would be required which however comes from the control action. The controller recompenses for the forfeiture in gradient coming from the increased valve opening.

Four measurement signals from the topside of the riser were used in the ISC controller. The accuracy of the measurement signal, the signal sensitivity to noise/disturbance and the readily availability of the signals determined the signals used in this work. The riser topside pressure (PT), the downcomer top pressure (DTP), the gas mass flowrate (GG) and the liquid density (ROL) are the measurement signals deployed in this study. The OLGA model of the ISC on the U-shape riser is shown in the appendix.

From (3-13), the measurement weights, measurement coefficient and measurement signals used in calculating the gain that would provide the additional gradient to make the system stable is presented in Table 3-7. Principal component analysis was used to determine the measurement signal weight from which the ISC controller gains (cascade gain) was determined. Thus, at 7 % valve opening the resulting correlated measurement signals (riser topside pressure (PT), the downcomer top pressure (DTP), the gas mass flowrate (GG) and the liquid density (ROL)) over about 900 seconds was compressed into a 4 data cells, usually uncorrelated, that captures the essence of the original data using PCA. The uncorrelated variables is also known as the principal components. The first principal component accounts for the majority in the variability in the data as possible and the subsequent components accounts for the remaining variability in the data. The variables extracted from the larger set of data is selected based on the variable that has the highest correlations with the principal component. Basically, PCA is a factor model from which the extracted factors are based on summarizing the total variance.

From this study, the correlated data (simulation or experimental data) was loaded and normalized. The mean and standard deviation of the correlated data were computed thus mean and standard deviation of each column (correlated measurement signals). The correlated data was then standardized thus subtracting the mean from each column entry and dividing by the standard deviation. The covariance matrix of the standardized data was computed. The eigenvalues and eigenvectors of the covariance matrix was computed. A singular value decomposition of the covariance matrix is reduced to a singular row of uncorrelated variables (principal component) representing the measurement signals inputted (riser topside pressure (PT), the downcomer top pressure (DTP), the gas mass flowrate (GG) and the liquid density (ROL)). These uncorrelated variables represents the coefficient of measurement weight, $\boldsymbol{W=[k_1,k_2,\dots,k_{n-1},k_n]T,}$ with k_1, k_2, k_3 … k_n representing the first, second, third … the n^{th} column of correlated data signals. The set point for the combined variables is calculated using (3-15), with the unstable valve opening of interest known. From (3-13),

$$\frac{2aQ^2}{u^3}K\left[\frac{dc}{dQ}\right] = -2.55 \qquad \textbf{(3-16)}$$

Where $\frac{\partial C}{\partial Q}$ is therefore estimated from the weighted deviations in measurements resulting from a perturbation in Q, ‘α’ is a constant associated with valve coefficient, mixture density and the given reference liquid flow rate, Q is the gas flowrate, and ‘υ’ is the valve opening ranging from 0 to 1.

Table 3-7 Controller design parameters

Measurement Signals	**Coefficient of measurement weight**	**Deviation in Vector of measurement signal**
Riser Top Pressure (PT)	-0.703	0.0797
Downcomer Top pressure (DTP)	0.0716	0.859
Gas Mass Flowrate (GG)	0.7067	1.9521
Liquid Density (ROL)	-0.0343	0.945

The deviation in vector of measurement signals is represented by dY/dQ_g for each measurement signal used in the ISC resulting from a slight perturbation in the gas flowrate (Q_g). After a 1 % perturbation in the gas flowrate the weighted deviation in the measurement ($\frac{\partial C}{\partial Q_g}$), attained was 1.2281. From (4-1) and using Table 4-6, the minimum controller gain, K to provide the additional gradient was obtained to be

0.0361. The minimum gain, K of the ISC for any preferred pressure drop gradient at a particular valve opening could be obtained using the above expression in the quest to stabilise slug flow at an increased valve opening.

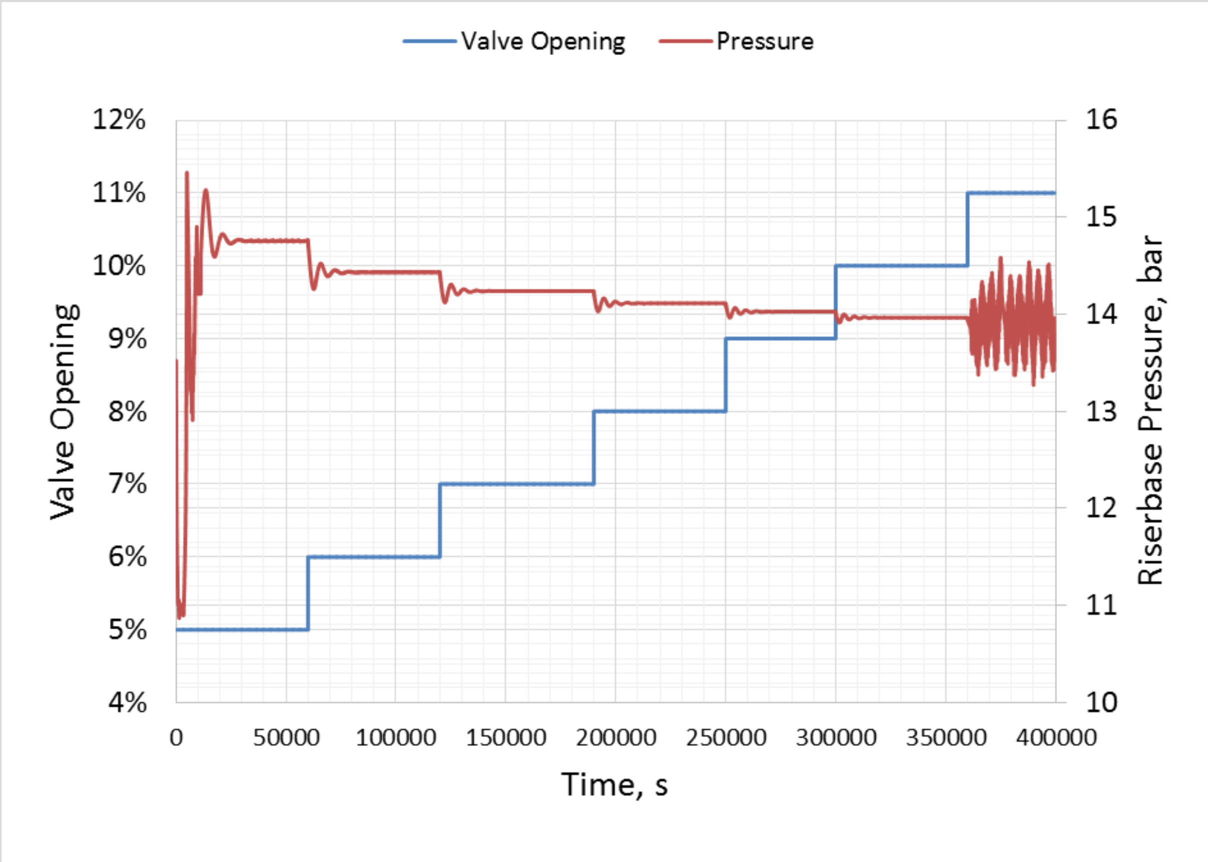


Figure 3-45 Riser base pressure response with corresponding valve opening in closed loop

Figure 3-45 represents the experimental results obtained with the ISC controller in place. From this results, the system is stabilised in an open loop unstable region. Thus, the system with the ISC in action and a stepwise increment in the choke valve of 1 %, shows that, there was an additional 5 % increment the valve opening for which the system retained its stability. Thus the maximum valve opening for which the system became stable moved from 5 % valve opening in open loop to 10 % valve opening in closed loop. Again the back pressure associated with manual choking was approximately reduced by 1 bar translating to approximately 5 % reduction in the riserbase pressure.

3.5 Summary

In this chapter, the necessity of flow loop geometry has been established. The slug envelope from both the 2-inch vertical riser and the 2 inch U-shape riser has been seen to be nearly unchanged. Most of the slug flow predicted from the experimental run on both riser match significantly with that in (Barnea, 1987). Even though in some instance there was an over prediction of certain regimes, there was a great match from both the experiments and the flow regimes from the literature.

Again, choking can be used to mitigate slug for different slug characteristic or behaviours. The choke valve needs to be closed considerably, to attain a stable flow. The degree of closure depends however on the flow characteristics. For both the U-shape riser and the purely vertical riser, there was a considerable increase in the riserbase pressure due to choking even though the pressure at the base of the riser was on the high in the U-shape compared to that of the purely vertical riser. Choking however comes with a certain degree of cost baring due to the reduced valve opening which tends to reduce the flow of the system on a whole. Therefore, there is a need to seek better ways or methods of stabilizing unstable flows in flow loops for distinct flow behaviours.

Also, it could be deduced that, for flow conditions which exhibit slugging characteristics in the riser of the U-shape, its initiation may not necessarily be from the downcomer of the U-shape riser loop but from the horizontal section of the flow loop. This is shown in the study above as all the flow conditions showing a slug flow characteristic in both the riser and downcomer of the U-shape flow loop also shows slugging in the purely vertical riser.

Furthermore, the downcomer top pressure comparatively to the riser outlet pressure is a good measurement for slug control, it cannot be relied on solely as for different flow conditions exhibiting slug flow in the riser, do not always exhibit slugging in the downcomer. Different flow regimes are observed in the downcomer which corresponds to slugging flow in the riser. Again, there could be a possibility of the downcomer length having effect on the fluctuations being observed in the downcomer. This could be an influence of the oscillations in either the downcomer base or riser base.

Finally, a systematic design of the ISC has been established and implemented on a large diameter U-shape riser system. The ISC was able to stabilize an unstable flow behaviour in the U-shape riser system at a larger valve opening relative to manual choking. Thus, system stability was achieved in an open loop unstable region hence optimizing production.

4 SLUG FLOW DYNAMICS AND CONTROL IN S-SHAPE RISER PIPELINE SYSTEMS

4.1 Introduction

The current trend of producing in deep offshore in the oil and gas industry has seen the rise of different pipeline riser system configuration. Several other riser configurations have appeared as a result of minor fields which in effect are not economical to run on a standalone processing facility or even matured fields. Tying in production pipeline systems to either already existing systems or even satellite field has been common to minimise the economic impact of the system by using common offshore processing facilities. This has led to the identification of several pipeline riser configurations; Reverse –J, Lazy-S, Steep-S, Wave, U-shape and their likes have all been discovered. The transient behaviour of fluids in some of these pipelines systems have still not been unravelled even though much effect has been channelled through modelling and experiments. Preceding studies have outlined meaningful understanding on the flow dynamics in both horizontal and vertical pipelines with the focus on unstable slug flow in pipelines (Schmidt, Brill and Beggs, 1980; Bendiksen and Espedal, 1992; Hurlburt and Hanratty, 2002; Kadri, Mudde and Oliemans, 2007; Kadri et al., 2009; Baliño, Burr and Nemoto, 2010; Xing, 2011; Malekzadeh, Henkes and Mudde, 2012). However, limited studies exist on the transient unstable slug flow in S-shape pipeline-riser systems. From Chapter 3, the 2-phase flow dynamics of a platform to platform pipeline configuration (U-shape riser) was investigated. In this chapter, another very popular riser pipeline configuration that is faced with the difficulty of flow and pressure fluctuations, the S-shape riser, will be investigated. S-shape risers have a complex geometry as will be described. An S-shape riser exhibits complex flow patterns as a result of two riser bases hence the prediction of the behavior of flow in the riser system very complicated. With special interest in unstable slug flow in the riser pipeline system, an established flow regime map was used in this study.

The chapter is structured as follows: the S-shape experimental facility used in the work is outlined next followed by section 4.2 which presents the flow patterns in the S-shape riser pipeline system from which stability analysis for different flow conditions is assessed, the S-shape riser dip effect on stability is also investigated; section 4.3

outlines the automatic control of unstable slug flow conditions in the S-shape pipeline riser system both experimentally and through simulation. The controller design and implementation with some case studies is presented. The chapter ends with a conclusion and a summary of the study outcome.

4.1.1 S-shape riser system

An S-shape riser is a form of flexible pipeline system which is basically similar to two L-shape, purely catenary, vertical riser configuration or a combination of any of these two pipeline risers joined together. The S-shape pipeline riser system could be inclined either on one or both horizontal sections. S-shape riser systems have two riser bases hence make the prediction of the hydrodynamic behaviour in the riser system complex.

For this study, the S-shape riser used is similar to two vertical risers joined together. The S-shape riser system is made up of a horizontal section, the lower limb, a down comer and an upper limb prior to the topside horizontal section. It is a uniform 2 inch (50.4 mm) internal diameter pipe. The horizontal section of the S-shape riser is 40 m in length with no inclination or declination, thus at a 0 degree angle. This serves as the channel through which the fluid is introduced into the riser. This joins the lower limb of the S-shape riser which is 5.5 m high and made from a 2 inch stainless steel schedule 20 pipe. A clear Perspex pipe is fitted within a meter length from the base of the lower limb to allow flow visualisation. The pressure at the base of the lower limb serves as the riserbase pressure (Riserbase pressure 1). The lower limb of the S-shape riser is followed by a downcomer prior to the upper limb which together form the S geometry as shown in Figure 4-2 (red line trace).

The downcomer is 1.5 m long and is declined at an angle of 45 degree serving as a declining horizontal for the upper limb. This is made from a clear PVC pipe that allows a clear view of what happens in the pipeline. The upper limb is 5.7 m high and also made from sections of clear PVC pipes aiding visualisation. The pressure at the base of the upper limb serves as the second riser base (riserbase pressure 2). The upper limb is joined at the top by a purely horizontal pipe (topside) with sections of steel and PVC pipes leading to the 2 phase separator. The total length of the topside is about 3.5 m and it’s equipped with a valve (choke valve), 0.3 m from the 2 phase separator where initial separation of liquid and gas takes place before the 3 phase horizontal separator.

S-shape riser exhibits complex flow as a result of two riser base hence the prediction of the behaviour of flow in the riser system complicated. Figure 4-1 shows a schematic of the S-shape riser used in this study.

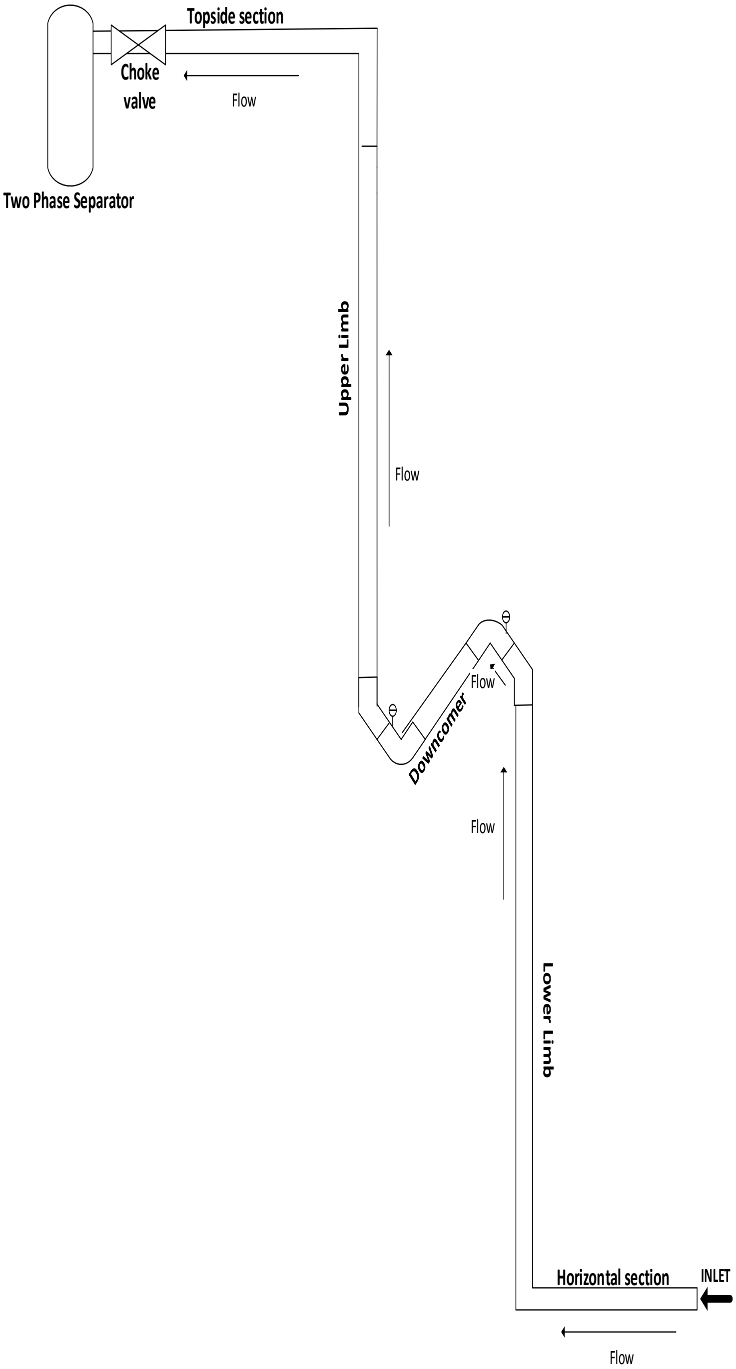


Figure 4-1 S-shape riser pipeline schematic

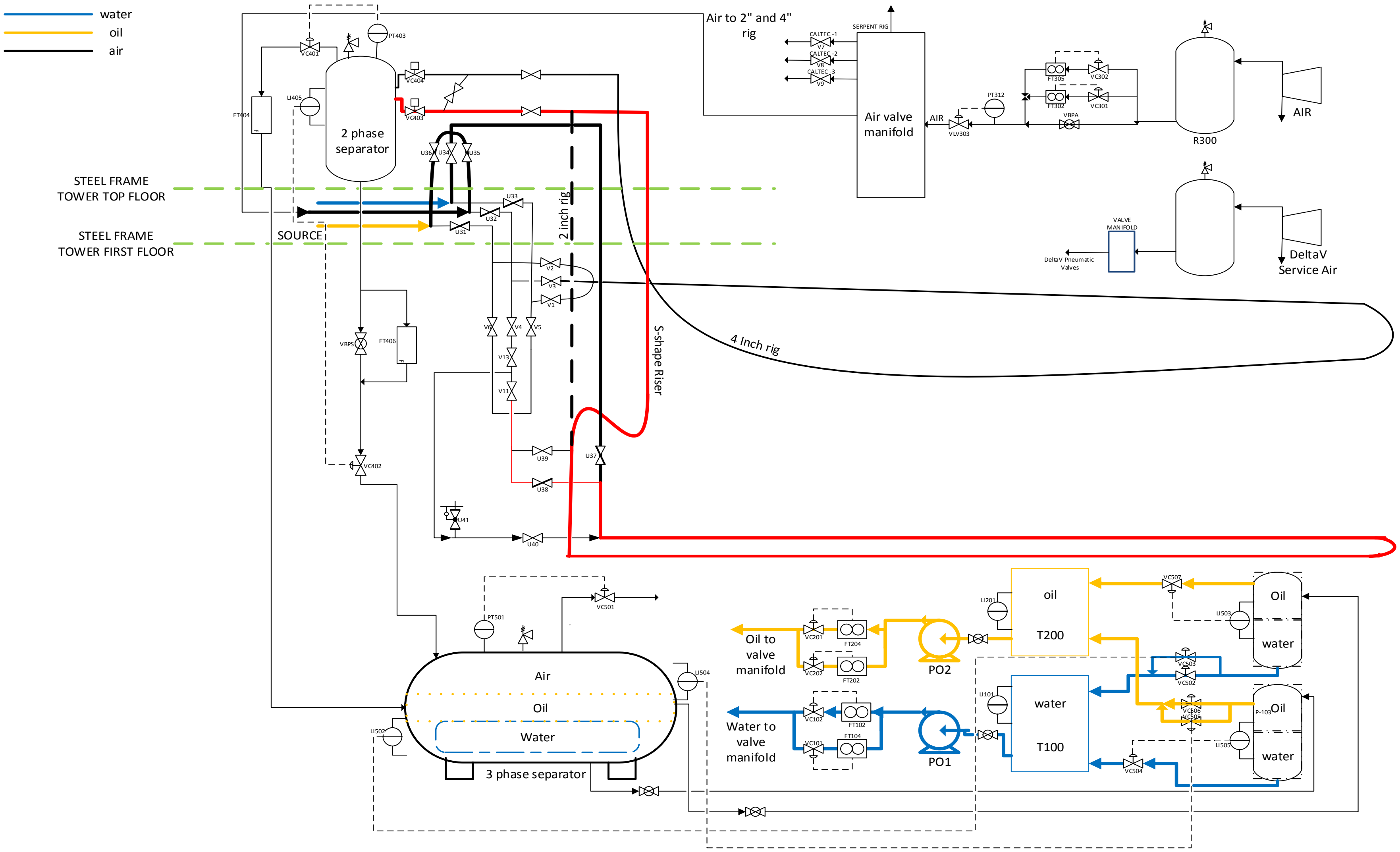


Figure 4-2 Schematic of the three phase facility (S-shape riser)

4.2 Flow patterns in S-shape riser pipeline (gas-liquid flow)

Due to complexity associated with the S-shape riser, using mechanistic numerical solution to predict flow behavior in the riser system is not always reliable hence the flow behavior would be assessed experimentally on a 2 inch S-shape riser system. The flow pattern for different flow conditions was identified using direct observation thus visual inspection and characterized using the extraction of characteristic variable from signal fluctuation in the phases involved (gas-liquid).

4.2.1 S-shape riser flow regime

4.2.1.1 Experimental procedure for flow regime identification on the S-shape riser

For the S-shape riser to be in operable mode, it was isolated from the rest of the three phase facility using the required valves. Isolators for both compressors and the pumps were turned on. The flow system was pressurised to a 1 barg pressure before the various experimental conditions were tested for the numerous cases. Liquid flowrates within the range of 0.1 kg/s to 5 kg/s corresponding to superficial liquid velocities of 0.05 m/s to 2.47 m/s respectively were investigated against gas flowrates of 7 Sm^3/h to 300 Sm^3/h also corresponding to a superficial gas velocity of 0.40m/s to 25.33 m/s. Table 4-1 shows the test matrix adopted in this study.

Table 4-1 Experimental test matrix

Component / Composition	Minimum Flowrate	Maximum Flowrate
Liquid, kg/s	0.1	5
Gas, Sm^3/h	5	300

A constant liquid flow rate was kept and the gas flow rate gradually increased till the maximum value attained. This was repeated for all the liquid flowrate of interest and experimental data recorded. The DeltaV SCADA system was used to set the various flow conditions and data acquisition. Although several other valuable information could be gathered from the 3-phase facility, the flow regime map was of primary interest.

Water flow was pumped into the system by a 30 Hz frequency drive pump into the pipeline. Compressed gas flow was contacted with the pumped liquid before both transported through the horizontal section of the pipeline riser system. The flow dynamics of each condition was observed visually through the Perspex glass and a clear PVC pipe located on the vertical section (above the riserbase) of the lower limb and the upper limb respectively. From the observations made, a flow regime map was developed identifying the flow patterns seen in the S-shape riser according to the standard flow regimes in multiphase flow. Due to the fact that the study was focused on an unstable slug flow regime, the identification of the slug envelope was paramount. Some data points within the slugging region exhibiting different characteristics were considered for further studies.

4.2.1.2 S-shape flow regime map

Research on flow pattern characterization has been beneficial and important for design optimization and two phase system operation in offshore systems. Identification of the flow patterns or regimes, thus the deformable interface of the fluids, was done using the pressure at the riserbase and visual inspection. This provides the dynamics and some information on the flow development in the system.

The interfacial structure between phases is very complicated because of the random and heterogeneous distribution of the flow structure in both spatial and temporal scales. The flow pattern, defined as the macro feature of the multiphase interface structure and its distribution, is often used to describe a multiphase flow.

A slug envelope was developed for the S-shape riser geometry and compared with one established for the U-shape riser and hence one in literature. The motivation for this is due to the limitation in the study of flow behaviour and system instabilities in S-shape riser configuration, hence the effect of riser configuration on severe slugging is yet to be understood. Again, the accuracy in understanding the characteristics and mechanism of flow in S-shape riser not only date back to pipeline design and sizing, but also the design of the processing facilities in the downstream of the pipeline and its operations.

Figure 4-3 shows the flow regime map obtained from the experimental run on the S-shape riser. From Figure 4-3, the flow behaviour data points on the S-shape riser (dotted points) was compared with a vertical flow regime map from Barnea, (1987) (black lines representing flow transition lines) and the slug flow region from the U-shape riser represented by an enclosed purple line (from chapter 3). Observation of the flow regime was done downstream of the horizontal pipe, thus a little above the riserbase of the lower limb.

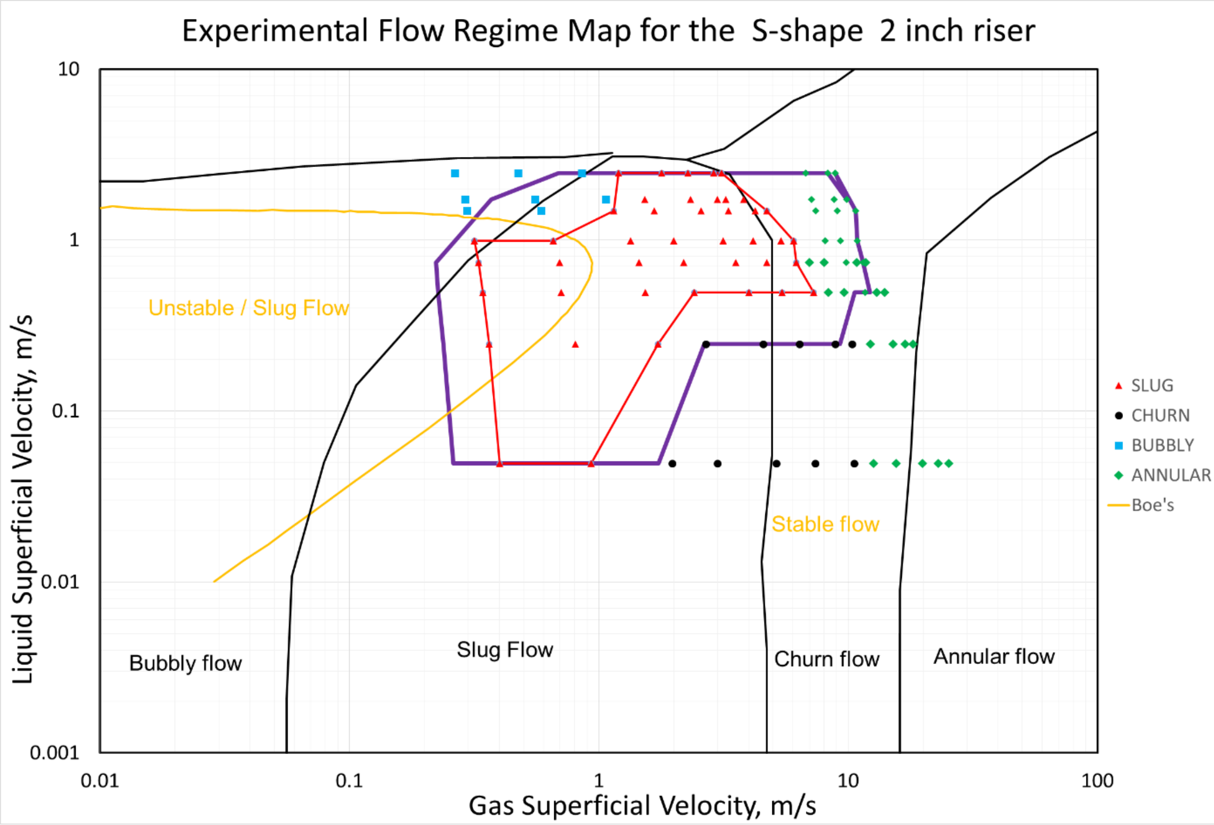


Figure 4-3 Experimental flow regime map for a 2 inch S-shape riser loop with slug, Barnea (1987) and Boe's criterion boundaries, purple-U-shape, red-S-shape

Slug flow, annular flow, dispersed bubble flow and churn flow were all observed in both pipeline riser configurations. However, some observed flow patterns on the S-shape riser system differ from that seen in the U-shape riser system. Some unstable slug flow patterns in the U-shape riser system showed annular flow (liquid surrounding tube, gas centre with or without entrained water droplets), bubbly flow (foam of bubbles in liquid or mist of droplets suspended in gas) or even churn flow (an oscillatory / chaotic flow condition with or without developed pockets of gases). At high liquid flowrate with relatively low gas flowrate, bubbly flow was observed. Increasing the gas flowrate further whiles lowering the liquid flowrate resulted to the flow regime moving through slug flow, churn flow and to annular flow.

From Figure 4-3, a considerable amount of the data points observed to be exhibiting slugging flow on the S-shape riser configuration (red boundary region) falls within the slug flow region of the flow regime map obtained from the literature (Barnea, 1987). However, beyond the slug region from the literature (black line), slug flow was still seen in the S-shape riser pipeline configuration. The slug flow region from the developed map (S-shape riser flow regime map) is wider and tapered towards the top. This could be as a result of the pipeline configuration and the mode of observation which could be subjective. Thus, the visual observation (means of detection) over predicts the slug flow region compared to that in the literature. Boe (1981), developed a criterion based on a proposal of Schmidt, Brill and Beggs, (1980), which states that the riserbase rate of the head accumulation must be greater than the gas pressure rate of increase for severe slugging to form. Comparing the slug flow region of the S-shape riser to the unstable plot region for Boe’s criterion, there are some observed slugging regions that fall within the unstable flow region (severe slugging flow), thus the orange region from Figure 4-3. This signifies that some flow conditions on the S-shape riser pipeline configuration exhibit severe slugging characteristics.

Again, in comparison with the flow regime map obtained from experiments using the U-shape riser, all the flow conditions exhibiting slug flow characteristics in the S-shape riser (red line- Figure 4-3), fall within the slug flow region on the U-shape riser (purple line - Figure 4-3). However, there are other unstable slug flow conditions on the U-shape riser which showed a relatively stable flow regime on the S-shape riser. The study of the flow regime in the S-shape pipeline riser configuration discovered that there exist different slug behaviours in the pipeline. This proves that the ‘S’ geometry does not introduce instabilities in the system but rather helps to break down the slugs in the system which is evident in the less fluctuation signal and the relative lower pressures. The pressure signal fluctuation and magnitude will be assessed next.

4.2.1.3 Riserbase pressure trend in the S-shape riser

This section seeks to investigate the dependency of flow pattern on the riser pipeline configuration. Three flow conditions exhibiting slugging characteristics (low gas – low liquid, medium gas – medium liquid and high gas – high liquid flow conditions) with trends already discussed on U-shape riser was assessed. Figure 4-4, Figure 4-5 and Figure 4-6 show the riserbase pressure trends for these flow conditions in the S-shape riser system. These riserbase pressure trends for same conditions will be compared in terms of magnitude and the magnitude of oscillations with that observed in the U-shape riser system.

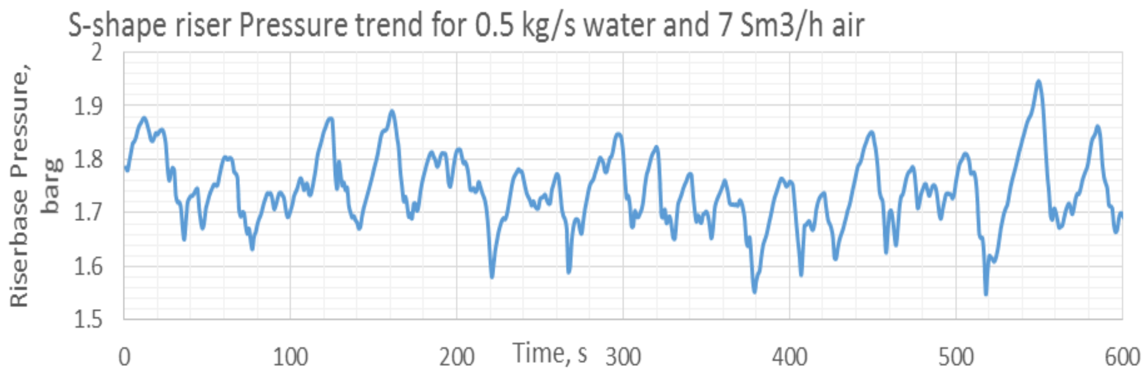


Figure 4-4 Riserbase pressure trend for the 2 inch S-shape riser loop at flow conditions 0.5 kg/s liquid and 7 Sm3/h gas.

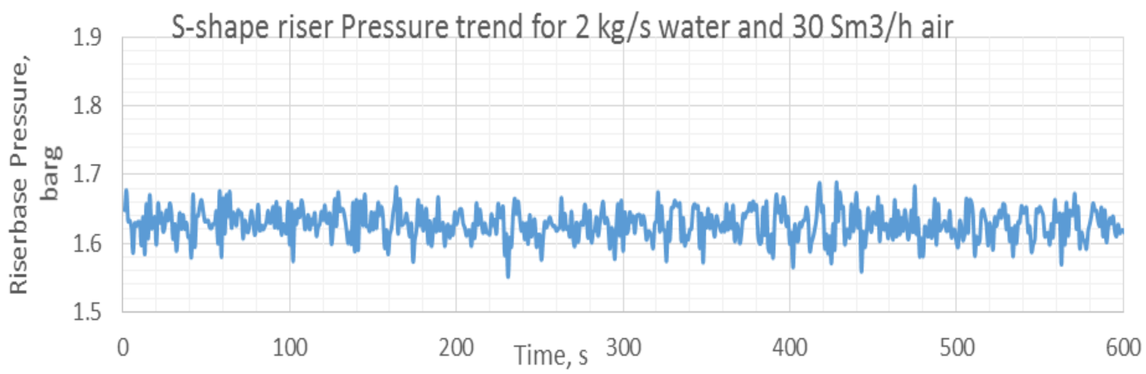


Figure 4-5 Riserbase pressure trend for the 2 inch S-shape riser loop at flow conditions 2 kg/s liquid and 30 Sm3/h gas.

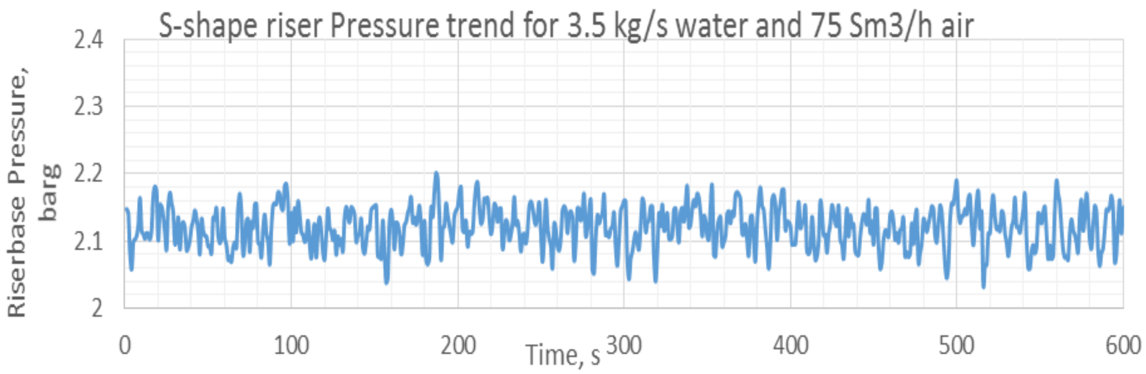


Figure 4-6 Riserbase pressure trend for the 2 inch S-shape riser loop at flow conditions 3.5 kg/s liquid and 75 Sm3/h gas.

The riserbase pressure trends for the flow conditions 0.5 kg/s liquid and 7 Sm^3/h gas, 2 kg/s liquid and 30 Sm^3/h gas and 3.5 kg/s liquid and 75 Sm^3/h gas on the S-shape riser is represented by Figure 4-4, Figure 4-5 and Figure 4-6 respectively. The pressure trends observed in the S-shape riser for low gas- low liquid flow condition, exhibiting a classic slug behaviour, comparatively have similar characteristics to that observed in the U-shape riser for the same flow conditions, however the magnitude of oscillations was relatively lower in the S-shape riser system. Increasing the flowrate through medium to high flowrate conditions, the pressure trends exhibit a transient flow behaviour which differs slightly from that observed in the U-shape riser. Again the riserbase pressure trend observed in the S-shape riser has a relatively lower pressure magnitude than that seen in the U-shape riser. This could be attributed to the length of the riser for both setups as a shorter riser length leads to lower pressures at the base of the riser due to lower liquid column in the riser. In conclusion, there are lower pressures observed in the S-shape riser pipeline configurations than in the U-shape riser for the same flow conditions. Again, the pressure fluctuations in the S-shape riser are relatively lower than those in the U-shape riser hence presumably easier to stabilise, this will be investigated next.

4.2.2 Stabilising unstable slug flow in the S-shape riser system

In this section the stability of flow in the S-shape riser would be assessed by means of investigating some flow conditions. As seen in the literature (Storkaas, 2005, Ehinmowo and Cao, 2015), increasing the pressure at the downhole could render the system stable. Parameter (choke valve) variation technique would be used to investigate the stability of the system for different flow conditions chosen from the slug envelope. This in tend increases the pressure at the base of the riser, hence observe the flow condition response to pressure. This is used to generate bifurcation diagram for the various flow conditions. This would help with the understanding of the slug flow behaviour in the S-shape riser system as well. From Figure 4-3, two flow conditions (0.5 kg/s liquid – 7 Sm^3/h gas and 1 kg/s liquid – 10 Sm^3/h gas) exhibiting classic slug flow behaviour were chosen and two other flow conditions (2 kg/s liquid – 30 Sm^3/h gas and 3.5 kg/s liquid – 75 Sm^3/h gas) exhibiting transient slugging behaviour were investigated further to assess the system stability.

4.2.2.1 Experimental procedure for flow stability by manual choking

A slug flow condition with superficial velocities of 0.36 m/s and 0.25 m/s for gas and liquid respectively, representing a low flowrate exhibiting a classic slugging behaviour condition, was run. This corresponds to 0.5 kg/s and 7 Sm3/h liquid and gas flowrate respectively. A stepwise decrease in the choke valve opening was done from a valve opening of 100 % to10 %. Riserbase pressures for each condition were recorded. The minimum, average and maximum riserbase pressures of each condition were plotted against their corresponding percentage valve opening (bifurcation map) where the critical bifurcation point was identified.

Consequently, the same process was done for superficial liquid velocity 0.49 m/s against a superficial gas velocity of 0.71 m/s corresponding to a liquid flowrate of 1 kg/s and a 10 Sm3/h flow of gas. Bifurcation maps were developed for these operations identifying the critical bifurcation points.

4.2.2.2 Classic unstable slug flow behaviour bifurcation maps

Figure 4-7 and Figure 4-8 show the riserbase pressure trend behaviour of the flow conditions exhibiting classic slug behaviour (1 kg/s liquid – 10 Sm^3/h gas and 0.5 kg/s liquid – 7 Sm^3/h gas respectively). This shows the characteristics of the observed slug flow behaviour for which the stability of these flow conditions in the S-shape riser would be assessed.

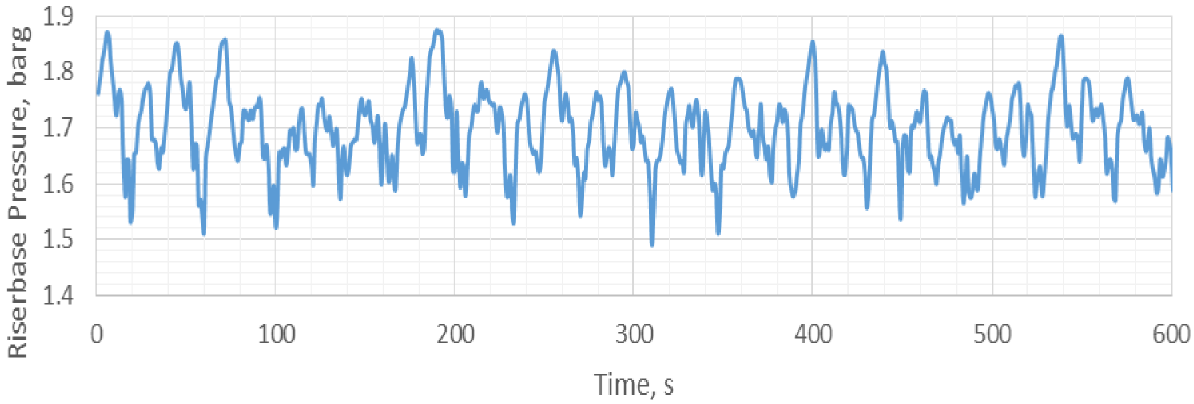


Figure 4-7 Riserbase pressure trend for a 2 inch S-shape riser loop at flow condition 1 kg/s liquid and 10 Sm^3/h gas

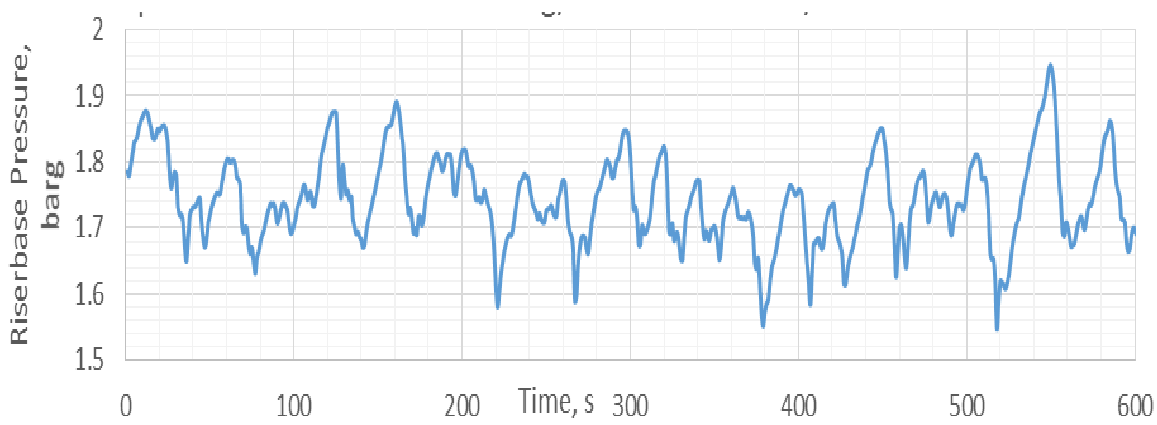


Figure 4-8 Riserbase pressure trend for a 2 inch S-shape riser loop at flow condition 0.5 kg/s liquid and 7 Sm^3/h gas

Figure 4-9 shows the bifurcation diagram obtained from the S-shape riser pipeline configuration for a liquid flowrate of 1 kg/s and a 10 Sm^3/h of gas flowrate. The pressure at the riserbase 1 (pressure at the base of the lower limb) position on the S-shape riser was used for this plot. A critical bifurcation point of 22 % valve opening corresponding to a 3.8 barg pressure was observed at the base of the lower limb. Thus the system (lower limb) was stable for valve openings equal to or less than 22 %. However for valve openings greater than 22% the system becomes unstable and this was evidence in the pressure oscillations which lead to a divergence in the curve.

Similarly, Figure 4-10 shows the bifurcation map for the same condition as above but however using riserbase 2 (pressure at the base of the upper limb). Again, 22 % valve opening corresponding to a pressure of 3.6 barg was observed to be the bifurcation point. This implies that the entire system under manual control would attain stability at a valve opening of 22 %.

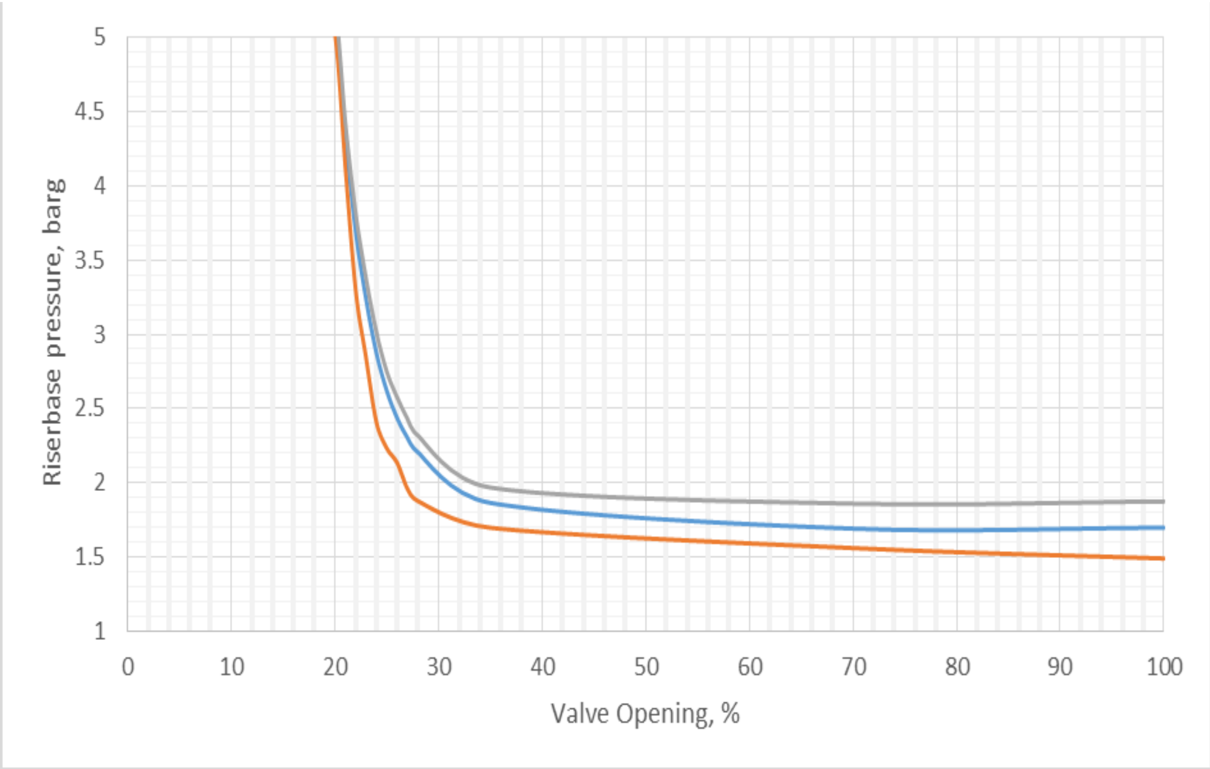


Figure 4-9 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 1 kg/s liquid and 10 Sm3/h gas flowrate at the base of the lower limb

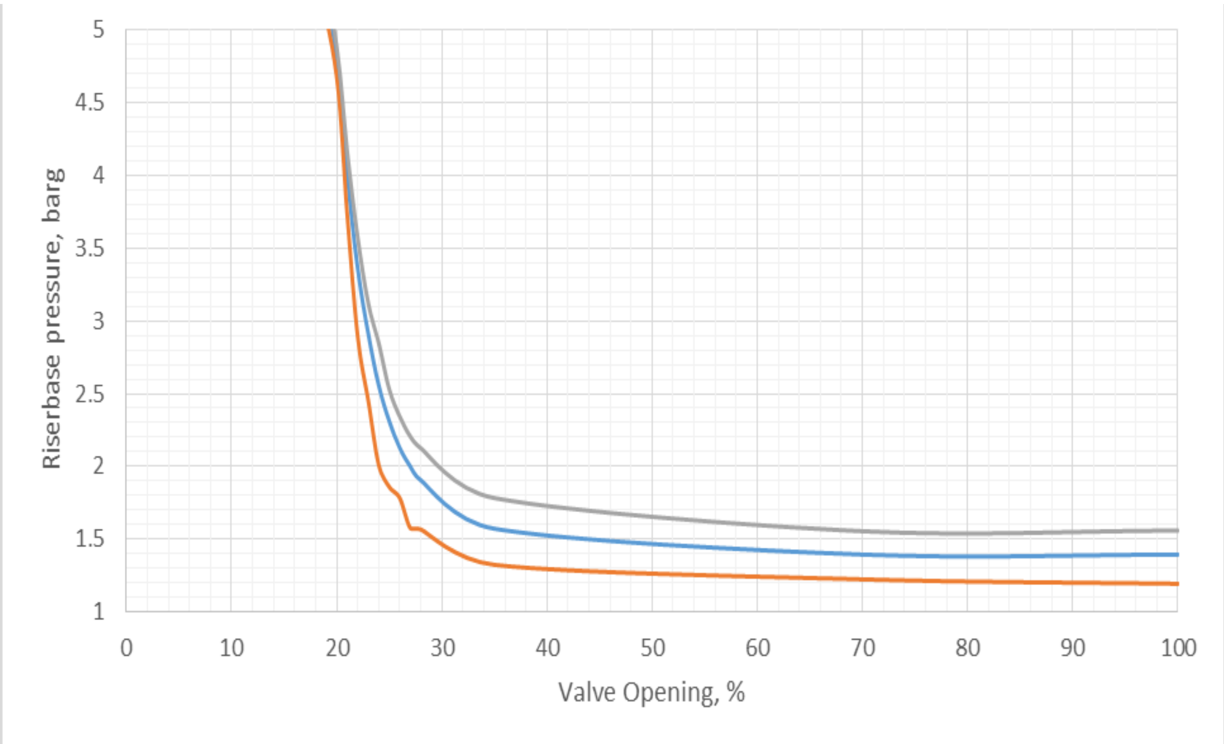


Figure 4-10 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 1 kg/s liquid and 10 Sm3/h gas flowrate at the base of the upper limb

Similarly, Figure 4-11 and Figure 4-12 represent the bifurcation map obtained from the S-shape riser pipeline configuration for a 0.5 kg/s and 7 Sm^3/h liquid and gas flowrate for the pressure at the base of the lower limb (riserbase 1) and the pressure at the base of the upper limb (riserbase 2) respectively. A bifurcation point of 19 % valve opening was observed at both the base of the lower limb and the upper limb. At 19 % valve opening, a corresponding 2.8 barg and 2.6 barg pressure was observed for the base of the lower limb and the base of the upper limb respectively. Thus, under manual choking the maximum valve opening for which the system became stable for the above flow condition was 19 %. This represents lower pressure fluctuations, a resultant of the pressure drop across the choke valve due to the closure of the valve. Again, for valve openings less than 19 %, the system was stable while the valve openings > 19 % introduces instabilities into the system (S-shape riser).

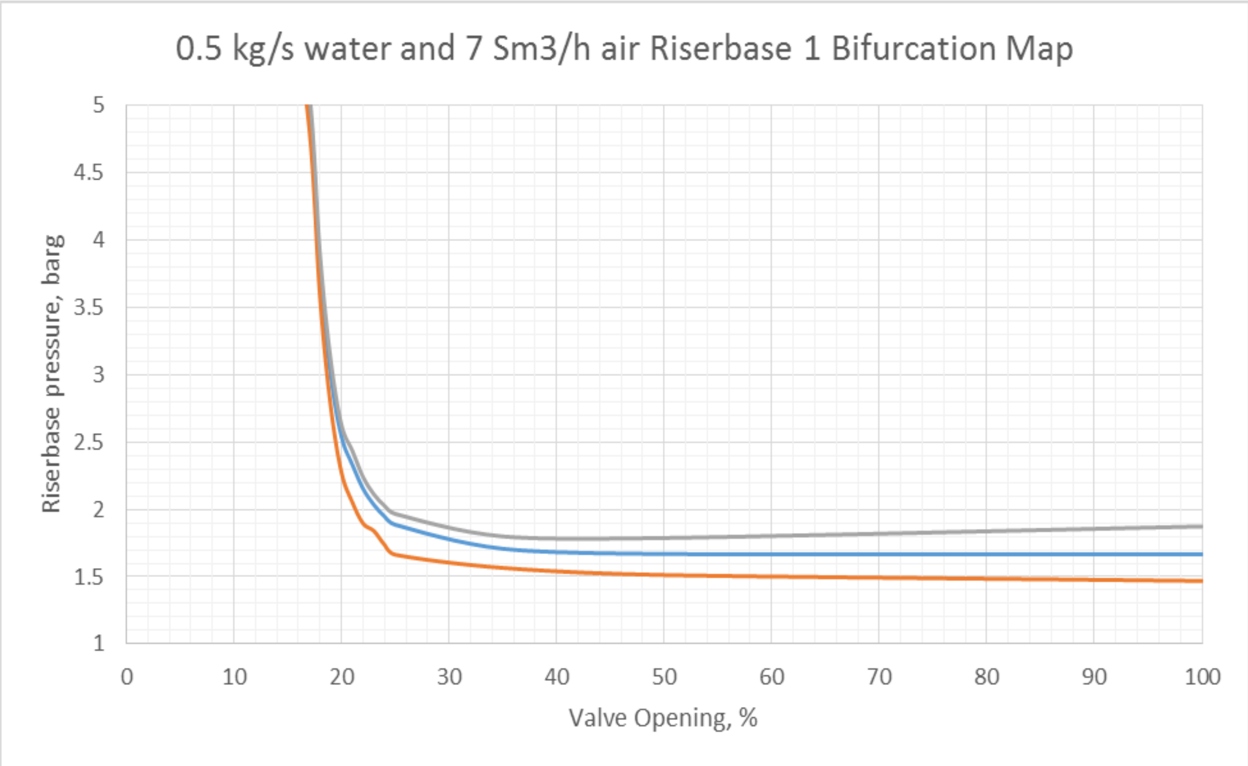


Figure 4-11 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 0.5 kg/s liquid and 7 Sm3/h gas flowrate at the base of the lower limb

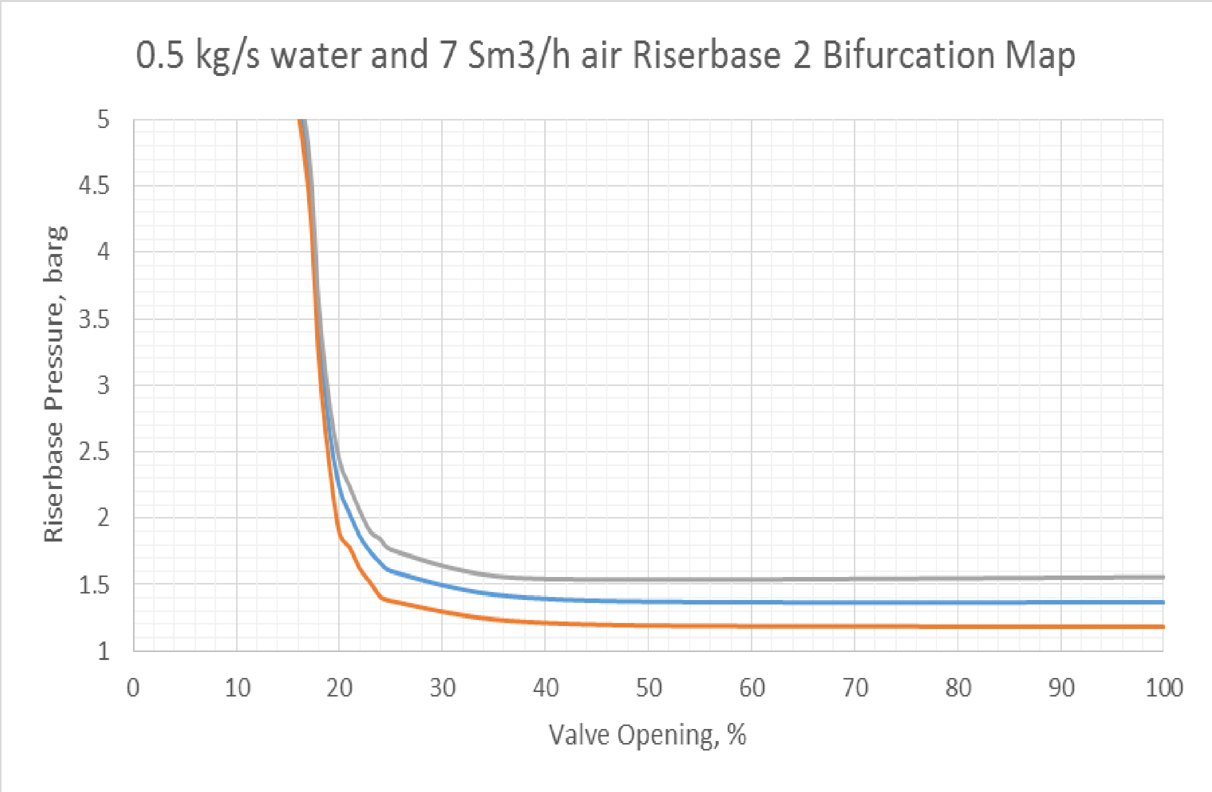


Figure 4-12 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 0.5 kg/s liquid and 7 Sm3/h gas flowrate at the base of the upper limb

In comparison the bifurcation map obtained using the riserbase for the flow condition 0.5 kg/s liquid and 7 Sm^3/h gas flowrate on the U-shape riser has similarities to that obtained from the S-shape riser pipeline configuration. A bifurcation point of 19 % valve opening corresponding to 2.8 barg riserbase pressure on the U-shape riser against 2.8 barg and 2.6 barg for the lower and upper limb respectively for the S-shape.

4.2.2.3 Transient slug flow behaviour bifurcation maps

Figure 4-13 and Figure 4-14 show the bifurcation maps obtained from the S-shape riser pipeline configuration for the flow condition of 3.5 kg/s liquid and 75 Sm^3/h gas for the pressure at the base of the lower limb (riserbase 1) and the pressure at the base of the upper limb (riserbase 2) respectively. From these figures, it could be observed that reducing the valve opening did not have any significant impact on the system stability even though there was an increase in the riserbase pressure. Thus from the bifurcation map, the bifurcation point could not be determined.

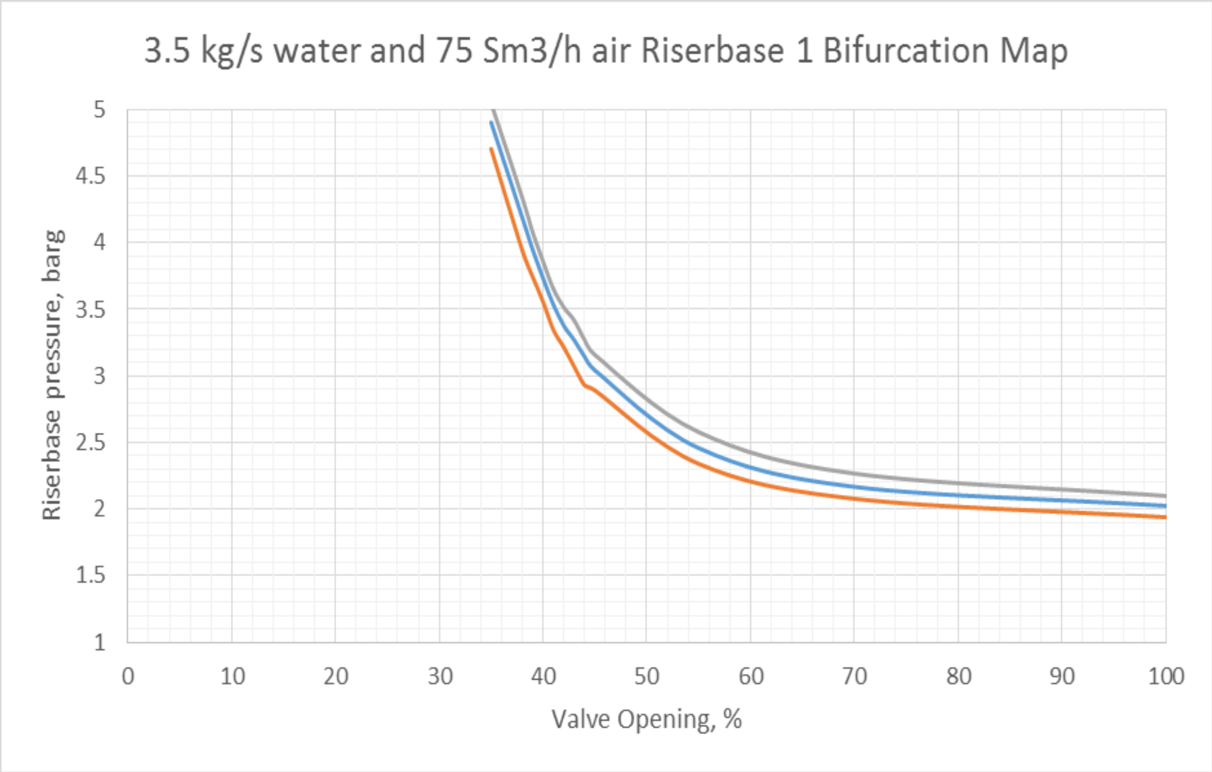


Figure 4-13 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 3.5 kg/s liquid and 75 Sm3/h gas flowrate at the base of the lower limb

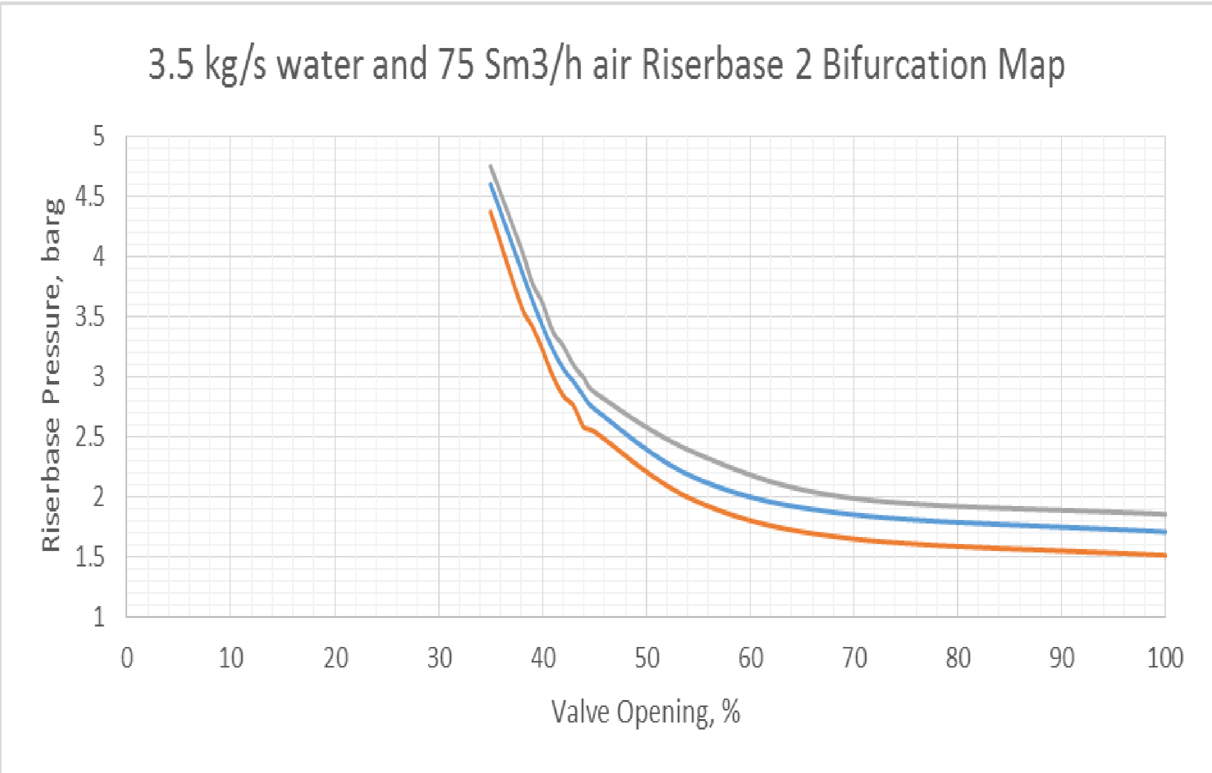


Figure 4-14 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 3.5 kg/s liquid and 75 Sm3/h gas flowrate at the base of the upper limb

Similarly for 2 kg/s liquid and 30 Sm^3/h gas flowrates on the S-shape riser, the bifurcation maps are represented by Figure 4-15 and Figure 4-16 for the pressure at the base of the lower limb and the upper limb respectively. It could be observed that the system does not stabilise with increasing the riser base pressure as established by several researchers. This could be that the stable riserbase pressure has not been reached yet since the system has an upper limit of 5 barg operating pressure for safety reasons. That is for all valve openings examined, there was a transient flow behaviour in the system. However, both riserbase behave similar to each other.

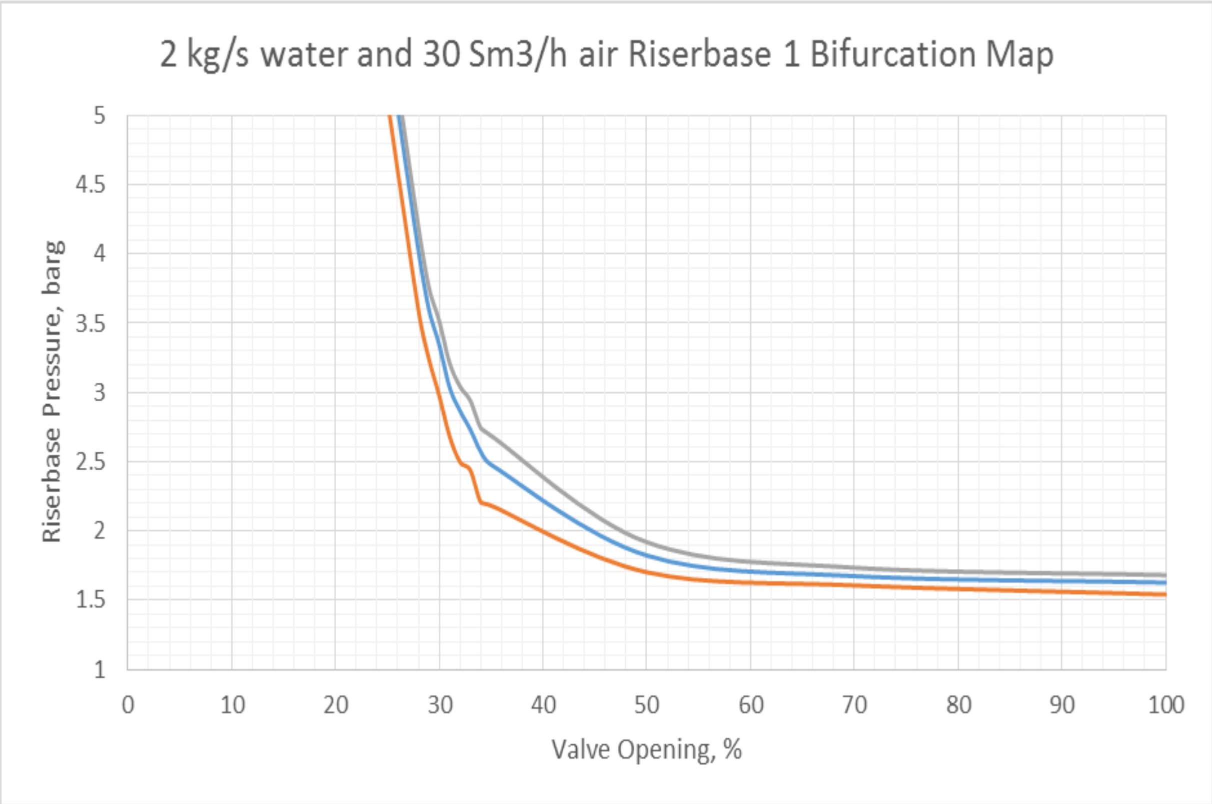


Figure 4-15 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 2 kg/s liquid and 30 Sm3/h gas flowrate at the base of the lower limb

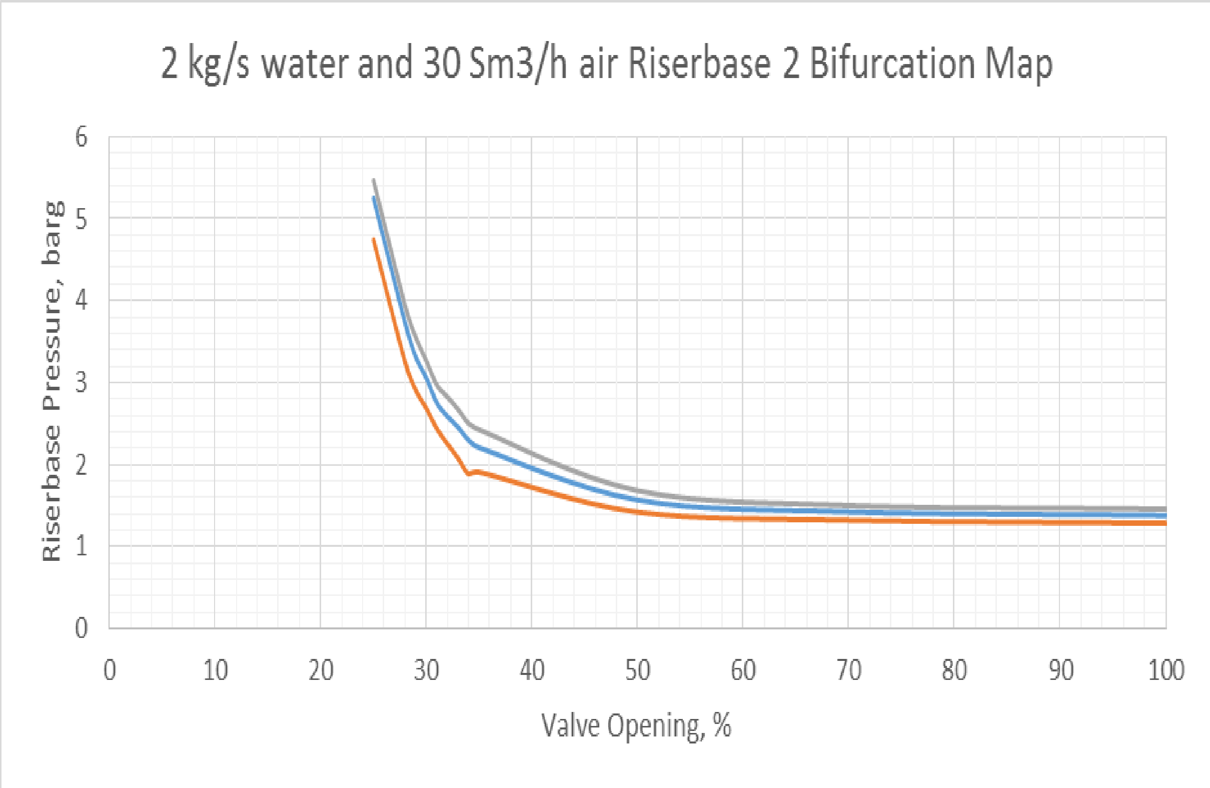


Figure 4-16 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 2 kg/s liquid and 30 Sm3/h gas flowrate at the base of the upper limb

In comparison, these flow conditions (2 kg/s liquid - 30 Sm^3/h gas and 3.5 kg/s liquid - 75 Sm^3/h gas flowrate) on the U-shape riser as shown from previous sections show a different flow behaviour to that seen in the S-shape riser in terms of stability.

For low flowrates (classical slugging conditions) on the S-shape riser, the system is responsive to a pressure increase, thus the system achieves stability for a lower choke valve opening which is similar to all other considered riser configurations. However for higher flowrates on the S-shape riser system, the system is not affected by increasing pressure by means of reducing the choke valve percentage opening. This could be a resultant of the dip and bend in the configuration which breaks the transient surge unlike that observed in other riser configurations. The S-shape riser dip effect on system stability would be assessed next.

4.2.3 S-shape riser dip effect on stability

S-shape riser is a form of flexible riser, with its features including two L-shape risers, purely catenary, vertical riser or any two of these pipeline riser configurations coupled together as described in earlier sections. Hence the S-shape riser system has two dips in its configuration. The S-shape riser system configuration is coupled mainly considering the structural integrity of the pipeline riser system, however when operating in multiphase conditions, instabilities (severe slugging) tend to occur and are difficult to stabilise. The controllability of the system could however may be compromised, thus making the system difficult to be stable or high cost of stabilising the system by just considering the structural integrity of the pipeline system in the coupling process. In this section, an assessment of the dip effect (riser length of both flowlines) on stabilising the S-shape riser system will be done. The outcome of this work would have a great deal on how flexible risers is laid out to maximize production.

OLGA simulation software was used to model an S-shape riser system for this study as this gives a considerable insight to the slug flow mechanism and makes necessary modification to the model to still suit a model of full industrial system. The use of the OLGA numerical simulation tool gives the advantage of assessing industrial systems with relatively large size pipelines in comparison to the experimental S-shape test facility. The details of the model, operating conditions, parameters and the model configuration will be presented next.

4.2.3.1 S-Shape riser – case study

The model used in this study is a replica of a single field from a satellite field from Chevron Energy Technology Company. The S-shape riser model used in this work could be sectioned into three, thus, the feed section, riser/pipeline section and the topside/processing section. The pipeline system characteristics, the flow condition and the fluid composition are considered under this study. The geometry of the S-shape riser was specially modified to attain two cases which were considered in this study. Figure 4-17 and Figure 4-18 show the flow path / pipeline configuration for the two cases under consideration. Again, the flow path geometry data for the models (Case 1 and Case 2) used excluding the well and the topside separator is presented in Table 4-2.

Table 4-2 OLGA flow path geometry data

Pipeline Section ID	Length, m	Elevation, m	Section	Diameter, m	Roughness, m	Wall Label
Flow1-1	1610	-26	16	0.279	0.0011	PRODFLOW
Riser1-1	158.704	0.302	4	0.279	0.0011	PRODRISER
Riser1-2	31.090	13.573	2	0.279	0.0011	PRODRISER
Riser1-3	48.463	39.710	2	0.279	0.0011	PRODRISER
Riser1-4	66.553	60.314	2	0.279	0.0011	PRODRISER
Riser1-5	36.747	-30.367	2	0.279	0.0011	PRODRISER
Riser1-6	42.672	-28.481	2	0.279	0.0011	PRODRISER
Riser1-7	33.364	-0.299	2	0.279	0.0011	PRODRISER
Riser1-8	33.528	20.763	2	0.279	0.0011	PRODRISER
Riser1-9	36.576	31.358	2	0.279	0.0011	PRODRISER
Riser1-10	60.960	57.516	2	0.279	0.0011	PRODRISER
Riser1-11	121.920	119.567	2	0.279	0.0011	PRODRISER
Riser1-12	109.424	108.585	2	0.279	0.0011	PRODRISER
Topside	73.152	0	4	0.356	4.56e-05	INCH14

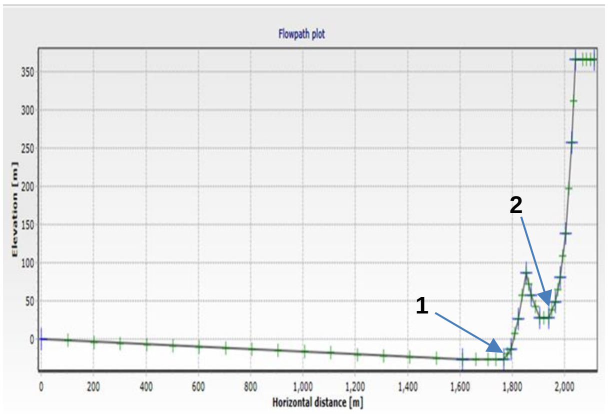


Figure 4-17 S-Shape riser configuration – Case 1

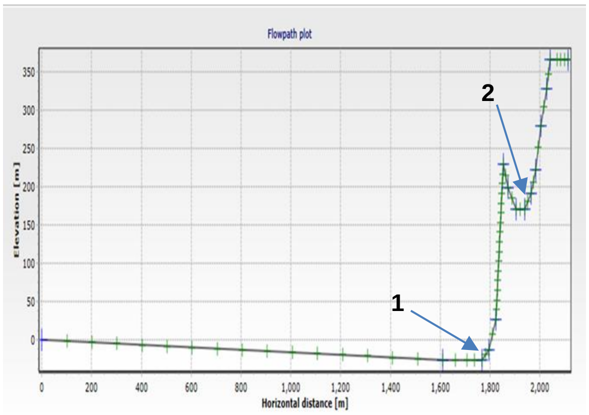


Figure 4-18 S-Shape riser configuration – Case 2

4.2.3.2 Flow path description

4.2.3.2.1 Pipeline / riser section

4.2.3.2.1.1 Case 1

The model examined is a typical S-shape riser, with two dips in configuration represented by 1 and 2 in Figure 4-17. The inlet near horizontal section is approximately 1800 m long with an equivalent riser height of approximately 365 m. The pipeline system has a uniformly distributed diameter (0.2794 m), with a choke valve of same diameter at the top of the riser. The length of the lower limb from the end of the horizontal section is 110 m while that of the upper limb (length from the ‘S’ geometry to the topside is 350 m.

Detailed pipeline length, sections and segments which defines the length, diameter, elevation and their like can be confirmed from Table 4-2 showing the profile of the pipeline with geometry shown in Figure 4-17. From Figure 4-17 the first riser base (riser base 1) is positioned 380 m below the topside while the second riser base (riser base 2) is positioned 350 m below the topside.

4.2.3.2.1.2 Case 2

Similarly, the model examined in the second case is a modified version of case one, thus a typical S-shape riser, with two dips in configuration represented by 1 and 2 in Figure 4-18. The inlet near horizontal section is approximately 1800 m long with an equivalent riser height of approximately 365 m. The pipeline system has a uniformly distributed diameter (0.2794 m), with a choke valve of same diameter at the top of the riser. The length of the lower limb from the end of the horizontal section is 260 m whiles that of the upper limb (length from the ‘S’ geometry to the topside is 210 m. Both cases have same characteristics but differ slightly in the length of both the lower and upper limbs. From Figure 4-18 the first riser base (riser base 1) is positioned 380 m below the topside while the second riser base (riser base 2) is positioned 180 m below the topside.

4.2.3.2.2 Topside / processing section

The riser outlet for both cases is connected to a horizontally oriented 2-phase gravity separator. The separator has a diameter/height of 3.6 m and length of 13.165 m. The inlet temperature and pressure of the separator is 26.67 ^{o}C and 7.93 bar. Since the outlets of the separator is controlled, the fluctuations liquid level or height could be used as a measure for the system stability.

The separator liquid outlet is controlled using an automatic PI controller while the gas outlet is controlled by a PID controller with parameters listed in Table 4-3. Valves are placed on both outlets to serve as the manipulated variable for each stream. The gas outlet of the separator is controlled using the pressure in the separator. This reflects the gas flowrate entering the separator at any point in time. Again, liquid film volume fraction serves as the controlled variable of the liquid flowrate out of the separator. The liquid film volume determines the volume of liquid in the separator.

Table 4-3 Separator controller parameters

Separator Controls and Parameters	**Liquid Outlet PID controller**	**Gas Outlet PID Controller**
Gain	2	0.00002
Bias	0.02	1
Integral Constant, s	720	75
Derivative Constant, s	0	18.75
Set point	0.597	2200000

4.2.3.2.3 Boundary conditions

The pipeline riser system inlet flowrate condition was accustomed to operate in an unstable flow pattern thus exhibits a severe slugging flow regime. This required regulating the inlet flowrate of the fluids involved while the physical properties of the system was kept constant. The system was equipped with a separator and a choke valve at the top of the riser to aid in the study. Table 4-4 shows the flow and operating conditions for which the system exhibited severe slugging. The flow and pressure fluctuation was used as a measure to determine the flow condition in the pipeline system. Flow against gravity, gas compression and expansion in the system can initiate instabilities in the system. Figure 4-19 and Figure 4-20 show the riserbase pressure response (riser base 1 and riserbase 2 respectively) observed in the S-shape riser system (Case 1) for the flow boundary condition represented in Table 4-4. Similarly, Figure 4-21 and Figure 4-22 show the riserbase pressure response (riser base 1 and riser base 2 respectively) observed in the S-shape riser system (Case 2) for the flow boundary condition from Table 4-4.

From the riser base pressure responses obtained from both cases, the system could be seen to be slugging due to the high fluctuations or oscillations in the pressures. This is evident in the separator liquid level for both cases shown in Figure 4-23 and Figure 4-24. From Figure 4-23 and Figure 4-24 the separator liquid level fluctuates within 2 to 2.5 m, which shows there are oscillations in the separator level that represents unsteady flow out of the systems.

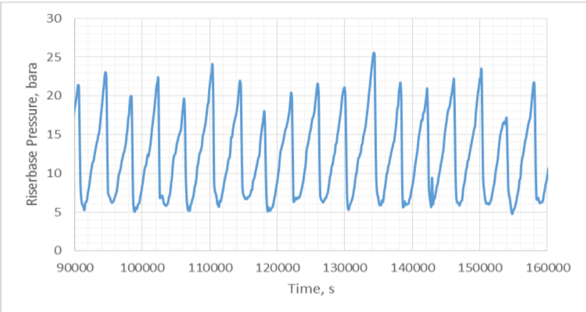


Figure 4-19 Riser base 1 pressure trend for Case 1

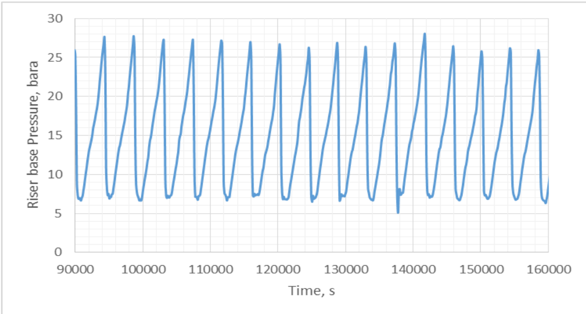


Figure 4-21 Riser base 1 pressure trend for Case 2

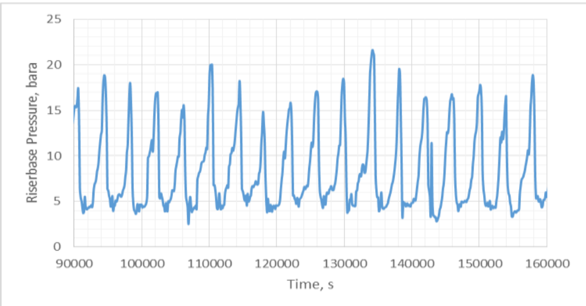


Figure 4-20 Riser base 2 pressure trend for Case 1

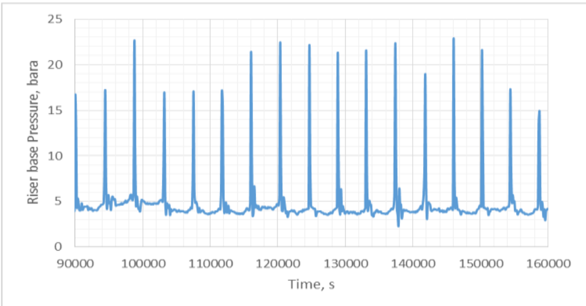


Figure 4-22 Riser base 2 pressure trend for Case 2

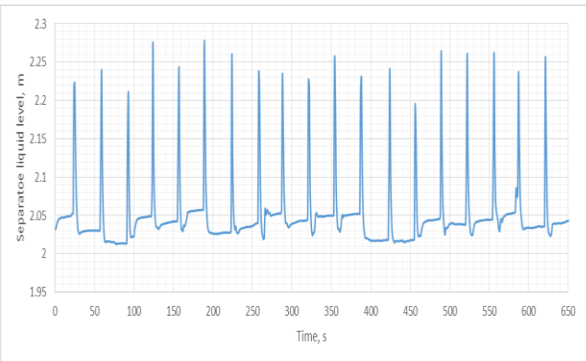


Figure 4-23 Separator liquid level for Case 1

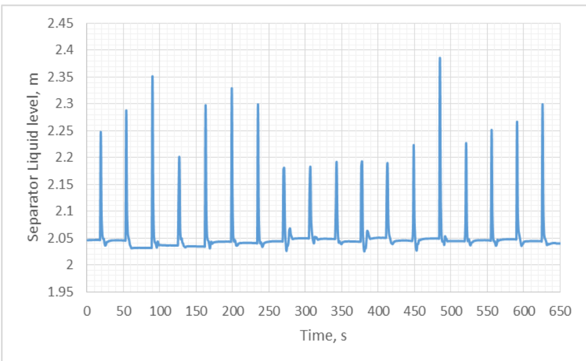


Figure 4-24 Separator liquid level for Case 2

Table 4-4 Operating conditions and parameters for the S-Shape riser that exhibits severe slugging

Operating Parameters (Initial Condition)						
Source Temp. (^{O} C)	Ambient Temp. (^{O} C)	U value (Wat/(m*K))	Inlet Temp. (^{o} C)	Outlet Temp. (^{O} C)	Inlet Pressure (bar)	Outlet Pressure (bar)
67.11	67.11	69.18	47.22	8.167	156.64	41.36
Operating Conditions						
Total Mass Flow, kg/s	Gas Mass Flow	Oil Mass Flow	Water Mass Flow	Inlet Temp. (^{o}C)	Outlet Temp. (^{o}C)	Outlet Pressure (bar)
6.9	0.9	3	3	47.22	8.17	41.36

4.2.3.3 System stability

4.2.3.3.1 Bifurcation diagram

In order to compare the difference and establish the effect of the configuration on the system stability, a parameter variation technique was adopted and the results plotted in the form of a bifurcation diagram.

4.2.3.3.1.1 Case 1

Figure 4-25 and Figure 4-26 show the bifurcation map of riser base 1 and riser base 2 respectively obtained for the boundary condition in Table 4-4 on the S-shape riser configuration in Figure 4-17. From Figure 4-25 and Figure 4-26 a bifurcation point of 9 % choke valve opening was attained corresponding to a riserbase 1 and riserbase 2 pressure of 8 barg and 7 barg respectively. Thus the S-shape riser system becomes stable when the choke valve opening is opened by 9 % of the valve diameter. For valve openings below 9 %, the system remains stable, however for valve openings greater than 9 % the system loses it stability.

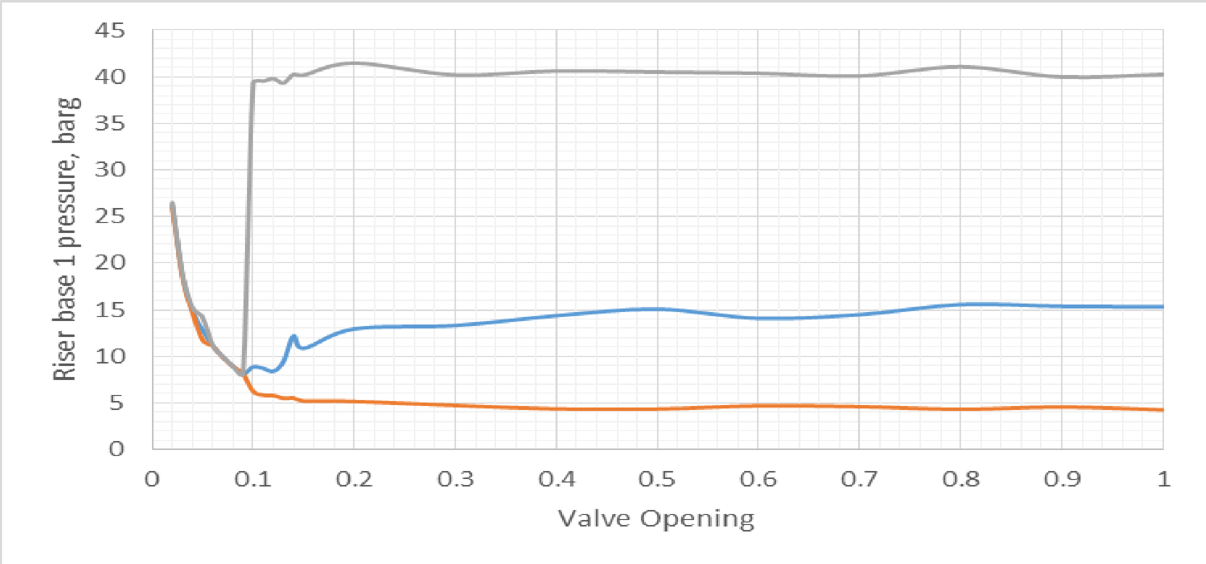


Figure 4-25 Riser base 1 pressure bifurcation for Case 1

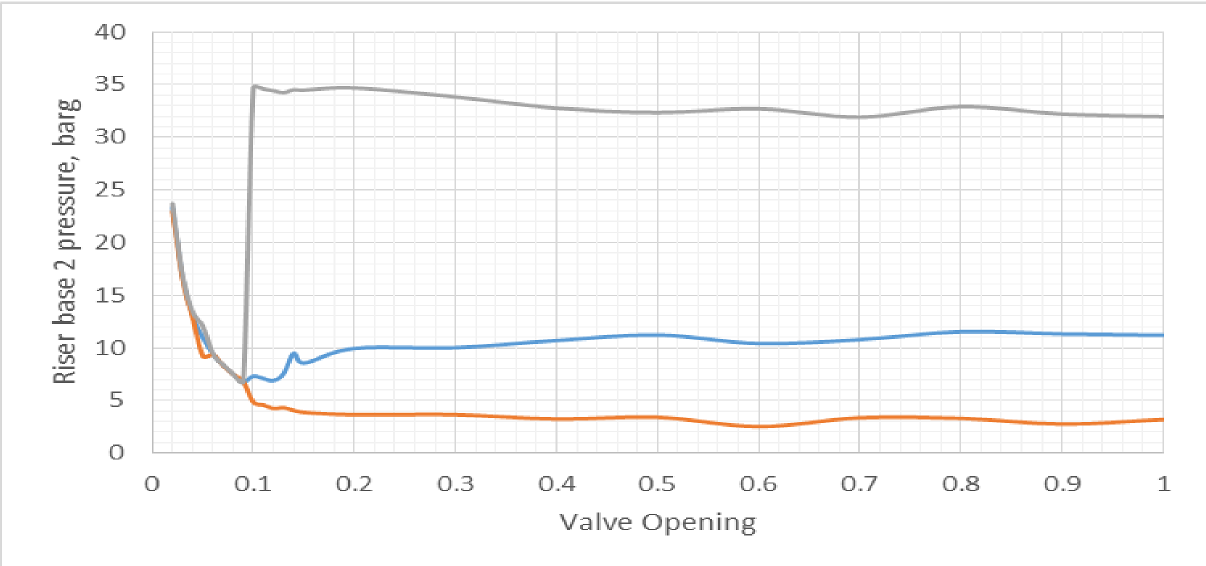


Figure 4-26 Riser base 2 pressure bifurcation for Case 1

4.2.3.3.1.2 Case 2

Figure 4-27 and Figure 4-28 show the bifurcation map of riser base 1 and riser base 2 respectively obtained for the boundary condition in Table 4-4 on the S-shape riser configuration in Figure 4-18. From Figure 4-27 and Figure 4-28 a bifurcation point of 6 % choke valve opening was attained corresponding to a riserbase 1 and riserbase 2 pressure of 7.5 barg and 6 barg respectively. Thus the S-shape riser system becomes stable when the choke valve opening is opened by 6 % of the valve diameter. For valve openings below 6 %, the system remains stable, however for valve openings greater than 6 % the system loses it stability.

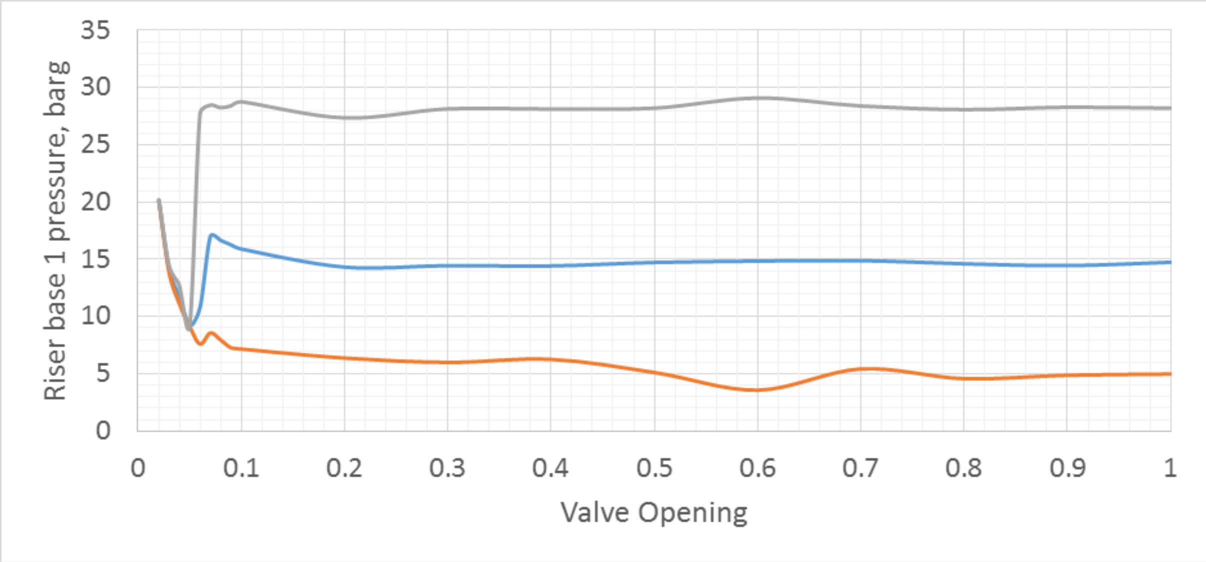


Figure 4-27 Riser base 1 pressure bifurcation for Case 2

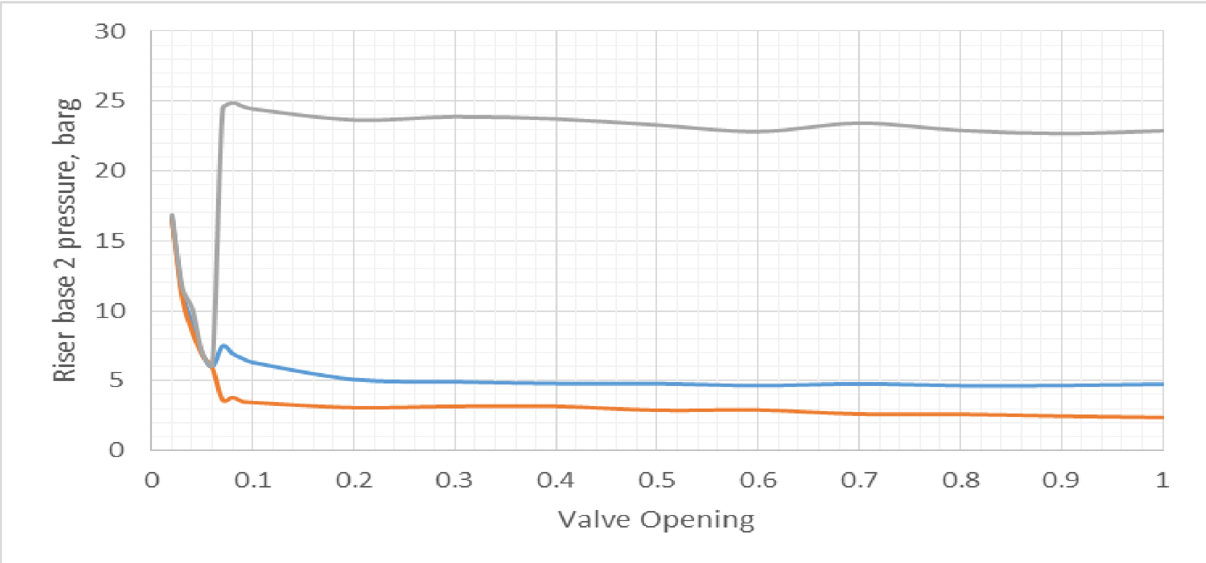


Figure 4-28 Riser base 2 pressure bifurcation for Case 2

In summary, the distance between the dips of the S-shape riser pipeline configuration plays an important role on the system stability. As observed from the cases assessed above, increasing the distance between the dips of the S-shape decreases the magnitude of the flow and pressure fluctuations in the system. However as shown in Case 2, even though the magnitude of pressure oscillations was lower, the system required a further 3 % reduction in the valve opening to render the system stable. This could be attributed to the closeness of the dip (riserbase 2) to the topside. Thus the instabilities from riserbase 2 which has an effect on riserbase 1 becomes difficult to stabilise due to the volume of the pipe before the choke valve. Again, while there is a larger valve opening related to case 1, the riserbase pressure at the stable valve opening is comparatively higher translating to higher throughput.

4.2.3.3.2 Gas injection to stabilise severe slugging

As established in Ehinmowo and Cao, (2015) for a fixed liquid flowrate an unstable riser system can considerably become stable at a high gas flowrate. This give reasons for the use of gas injection as a slug mitigation technique as established and investigated by several researchers. Instabilities (unstable slug flow) in a pipeline riser system is basically caused by the compressibility and expansion of the gas phase in the system. This system could be stable or unstable with an increment in the gas flowrate for a fixed liquid flowrate. Increment or decrease in the gas flowrate results either to an increase or decrease in the frictional loss within the system due to a change in the gas – liquid ratio.

The trade-of between the optimum points for the amount of gas to be injected to render the system stable is a crucial issue. This method will be adopted to investigate the optimal amount of gas (minimum gas flowrate) required to stabilise the S-shape riser systems in both cases (Case 1 and Case 2). This translates or reflects on the difficulty in stabilising both systems.

4.2.3.3.2.1 Case 1

Figure 4-29 and Figure 4-30 show the riserbase pressure (riserbase 1 and riserbase 2 respectively) response with increasing gas flowrate. From Figure 4-29 and Figure 4-30, the S-shape riser system in case 1 requires 5.3 kg/s gas flowrate to attain stability. Thus an additional 4.4 kg/s gas flowrate would be needed to make the system stable. The minimum point from Figure 4-29 and Figure 4-30 reflect the optimum gas flowrate (minimum gas flowrate) for which the system becomes stable. Beyond this point, the system stability still holds. This means that the system can only be stable if a considerable amount of gas flowrate was introduced into the system. The 5.3 kg/s flowrate of gas corresponds to a riserbase pressure of 5 barg and 4 barg for riserbase 1 and riserbase 2 respectively. Again, both riserbase pressure / flow characteristics become stable at the same amount gas flowrate.

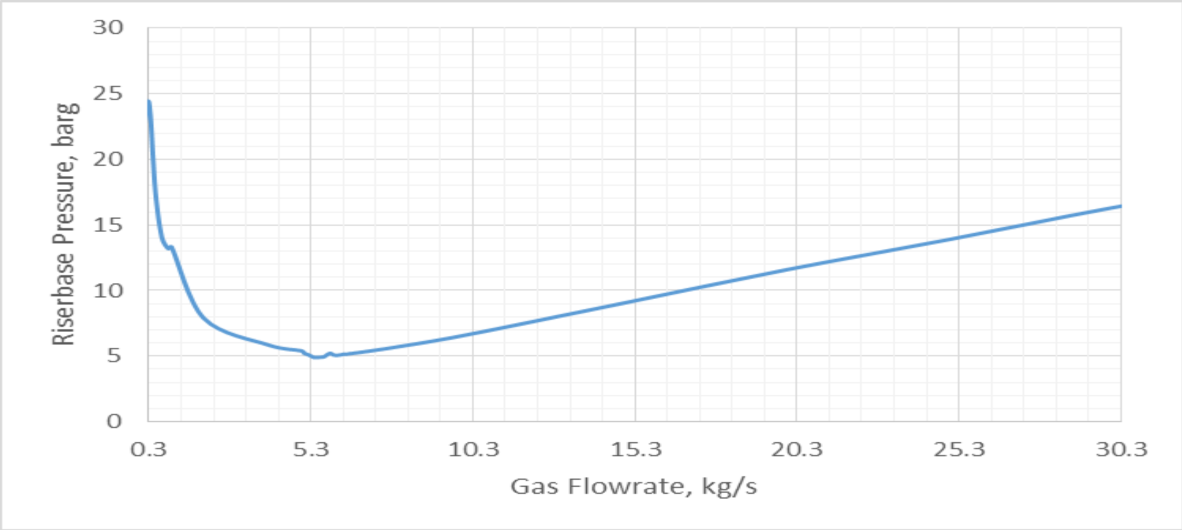


Figure 4-29 Riser base 1 pressure with varying gas flowrate for Case 1

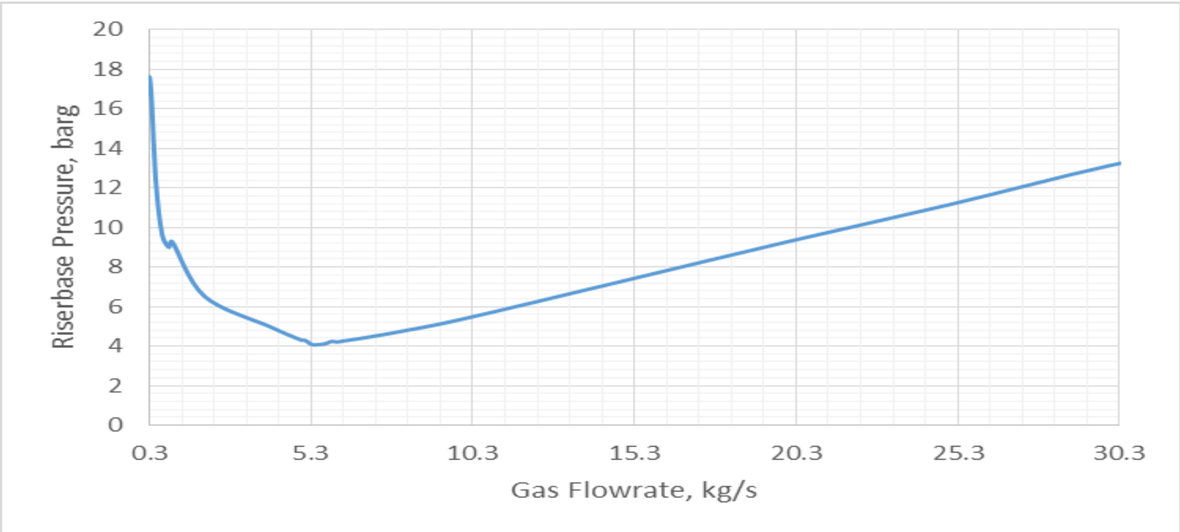


Figure 4-30 Riser base 2 pressure with varying gas flowrate for Case 1

4.2.3.3.2.2 Case 2

Similarly, for the S-shape pipeline riser system configuration in case 2, Figure 4-31 and Figure 4-32 show the riserbase pressure response with increasing gas flowrate for riserbase 1 and riserbase 2 respectively. The lowest riserbase pressure points from the plot in Figure 4-31 and Figure 4-32 reflect the minimum gas flowrate to stabilise the system. From Figure 4-31 and Figure 4-32 the system attains stability at a 6.3 kg/s gas flowrate. This corresponds to a pressure of 5 barg and 3.5 barg for riserbase 1 and riserbase 2 respectively. Thus, for the flow condition in Table 4-4, the S-shape riser configuration in case 2 can only be stable if a minimum additional gas flowrate of 5.4 kg/s was added.

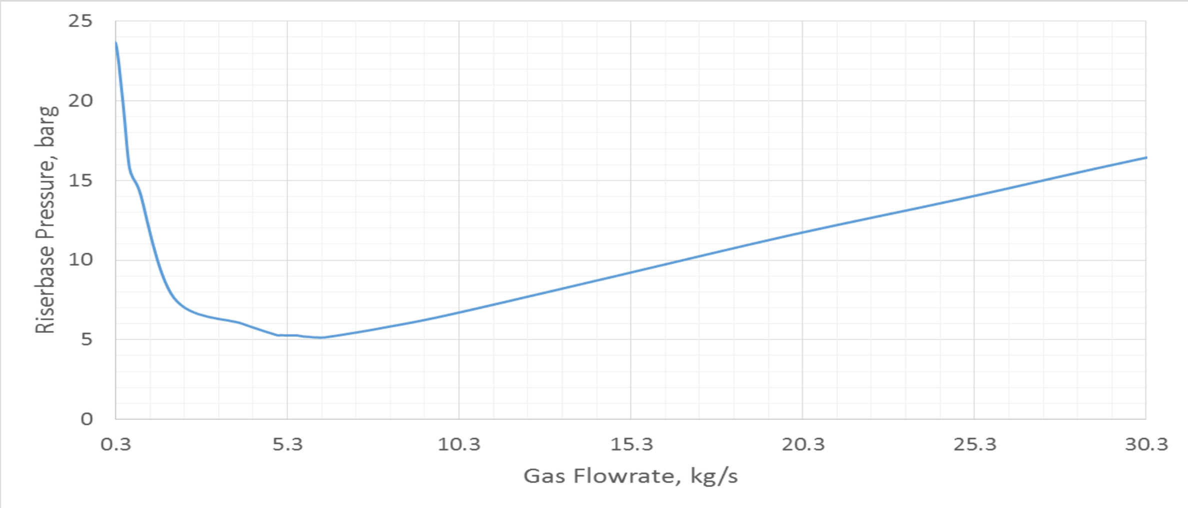


Figure 4-31 Riser base 1 pressure with varying gas flowrate for Case 2

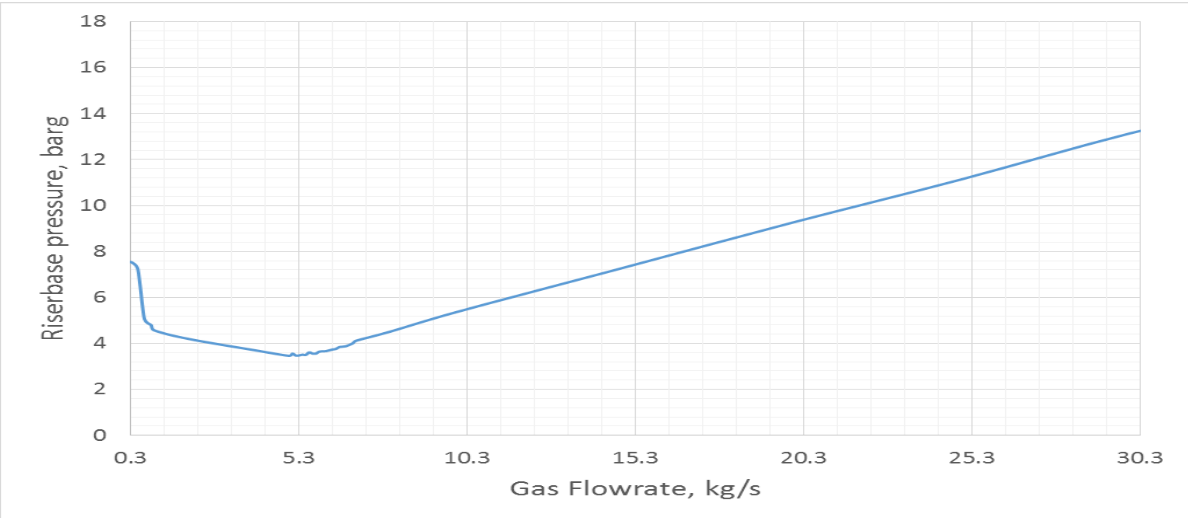


Figure 4-32 Riser base 2 pressure with varying gas flowrate for Case 2

Table 4-5 Summary of results

	CASE 1		**CASE 2**	
	RISERBASE 1	RISERBASE 2	RISERBASE 1	RISERBASE 2
Bifurcation point, %	9	9	6	6
Maximum Pressure, bara	25.58	21.62	27.98	22.88
Maximum Pressure at Bifurcation, bara	7.05	5.88	8.01	6.13
Minimum Gas Flowrate to stabilise system, kg/s	5.3	5.3	6.3	6.3
Pressure at minimum gas flowrate, barg	5	4	5	3.5
Pressure Gradient at operating condition	-0.84	-1.75	-4.4	-4

4.2.3.4 Discussion

The flow regime map obtained from the experimental run on the S-shape riser matches significantly with that in literature for hydrodynamic slugs. Regardless, there were some flow conditions that were outside the slug envelope seen in the literature even though it showed slugging behaviour on the S-shape riser pipeline. All the flow conditions exhibiting slugging flow on the S-shape riser fall within the slug envelope determined experimentally on the U-shape riser. However, there were some slug flow conditions on the U-shape riser that showed a different flow regime on the S-shape riser. This shows that the pipeline configuration has an effect on the flow patterns seen in the pipeline riser system, thus, the necessity of flow loop geometry has been established in this study. Comparatively, there exists a lower pressure fluctuation observed in the S-shape riser than that seen in the U-shape riser. The pressure magnitudes seen on the U-shape riser were higher compared with that seen in the S-shape riser. The U-shape riser has a longer continuous riser length compared to that seen in the S-shape riser resulting to different slug densities

As established by many researchers, choking can be used to mitigate slug for different slug characteristic or behaviours. The choke valve needs to be closed considerably, to attain a stable flow. The degree of closure depends however on the flow characteristics. From the stability analysis conducted on the S-shape riser, the pressures at both the base of lower and upper limb stabilise at the same valve opening. Also, for the same flow condition the system bifurcation point was same for both the U-shape and the S-shape riser. However, for the S-shape riser system some slug flow behaviours are distinct as they are unresponsive to pressure increment. Thus there are two kind of slugging flow behaviour existing in the S-shape riser system. Choking however comes with a certain degree of cost baring due to the reduced valve opening which tends to reduce the flow of the system on a whole. Therefore the need to seek better ways or methods of stabilizing flows in flow loops for distinct flow behaviours which will be addressed next.

4.3 Experimental ISC for slug control on a 2 inch S-shape riser

ISC was introduced as a theoretically and practically sound solution to stabilise unstable flows, where all available and relevant topside measurements are combined to form a single controlled variable, also called inferential slug index. When this slug index is maintained smoothly, the overall slugging flow is eliminated. The available topside measurement signals from the S-shape riser used in this work includes:

- Riser Outlet Pressure
- Pressure Drop Across the Choke Valve
- Two-Phase Separator Pressure
- Two-Phase Separator Gas Outlet Flow
- 2-Phase Separator Liquid Outlet Flow
- Three-Phase Separator Pressure

Most riser pipeline systems have readily available measurement signals from the topside and a combination of the six signals (mentioned above) for the ISC technology design was investigated experimentally to stabilise slug flow on the 2 inch S-shape riser. Flow conditions from the slug envelope developed on the S-shape riser were assessed in the quest to stabilise the unstable flow behaviour. Two different flow conditions exhibiting unique slugging characteristics will be investigated with the aim of stabilising the system using the slug control stability analysis used to design the ISC. This assesses the robustness of the ISC in a more practical approach on an S-shape riser.

This seeks to investigate how the system (flow conditions) could be stabilised at a much larger valve opening (relative to that observed in open loop) in closed loop using the ISC technology. The flow conditions and characteristic of the flow will be discussed next.

4.3.1 **0.5 kg/s liquid and 7 Sm^3/h gas flow on the S-shape riser**

As established from the previous chapter 0.5 kg/s liquid and 7 Sm^3/h of gas through the S-shape riser system exhibit a transient flow behaviour (slug behaviour). Figure 4-33 shows the pressure trend for this flow condition on the S-shape riser pipeline configuration. From Figure 4-33, the riserbase pressure oscillates between about 1.54

barg and 1.95 barg over a period of time which clearly signifies that the system is not stable.

For the system to be stable, the pressure drop across the valve must be sufficiently large that is, the valve opening must be considerably small which means low flow through the valve resulting in an increase in the riser base pressure. This leads to a reduction in the production rate as there is a reduction in the acceleration of flow.

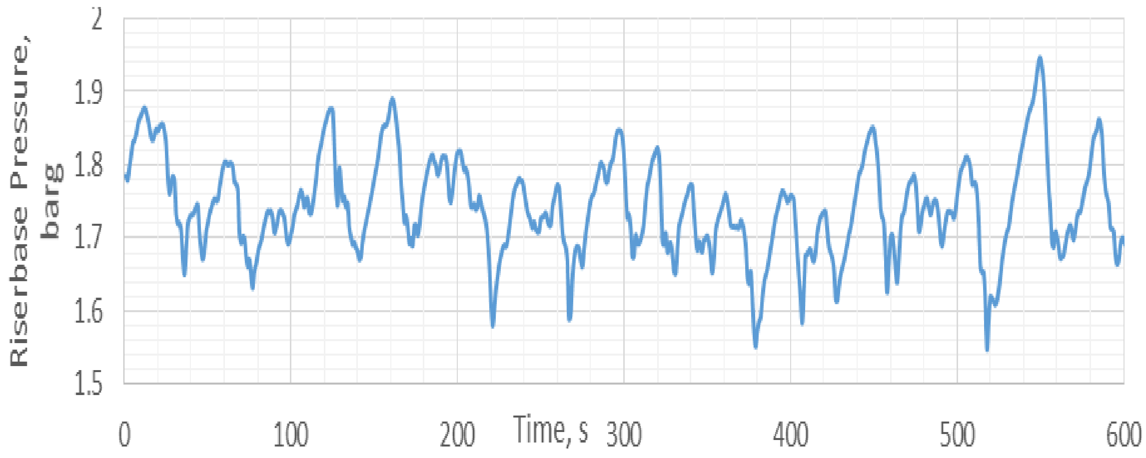


Figure 4-33 Riserbase pressure trend for the 2 inch S-shape riser loop at flow conditions 0.5 kg/s liquid and 7 Sm3/h gas

4.3.1.1 Stabilizing unstable flow condition with manual choking

A parameter variation technique was adopted to establish the stability limit in open loop to investigate the stability of the system for this flow condition at a larger valve opening in closed loop. Choking the valve at the exit of the riser has been the most common means of mitigating slugs in time past but unfortunately has a negative repercussion on production. The choke valve percentage opening is varied and the resultant pressure plotted against the respective choke valve openings. The resultant bifurcation map developed from this study is shown in Figure 4-34 and Figure 4-35.

From Figure 4-34 and Figure 4-35 the system becomes stable at a valve opening of 19 % corresponding to a pressure of 2.8 barg at riserbase 1 and 2.6 barg at riserbase 2 in open loop. For the benefit of this study, the system stability is based on physical observation and assessment of the pressure fluctuations. Riserbase 1 has a higher riserbase pressure than as observed in Riserbase 2. This could be attributed to the difference in depth from where both measurements were taken.

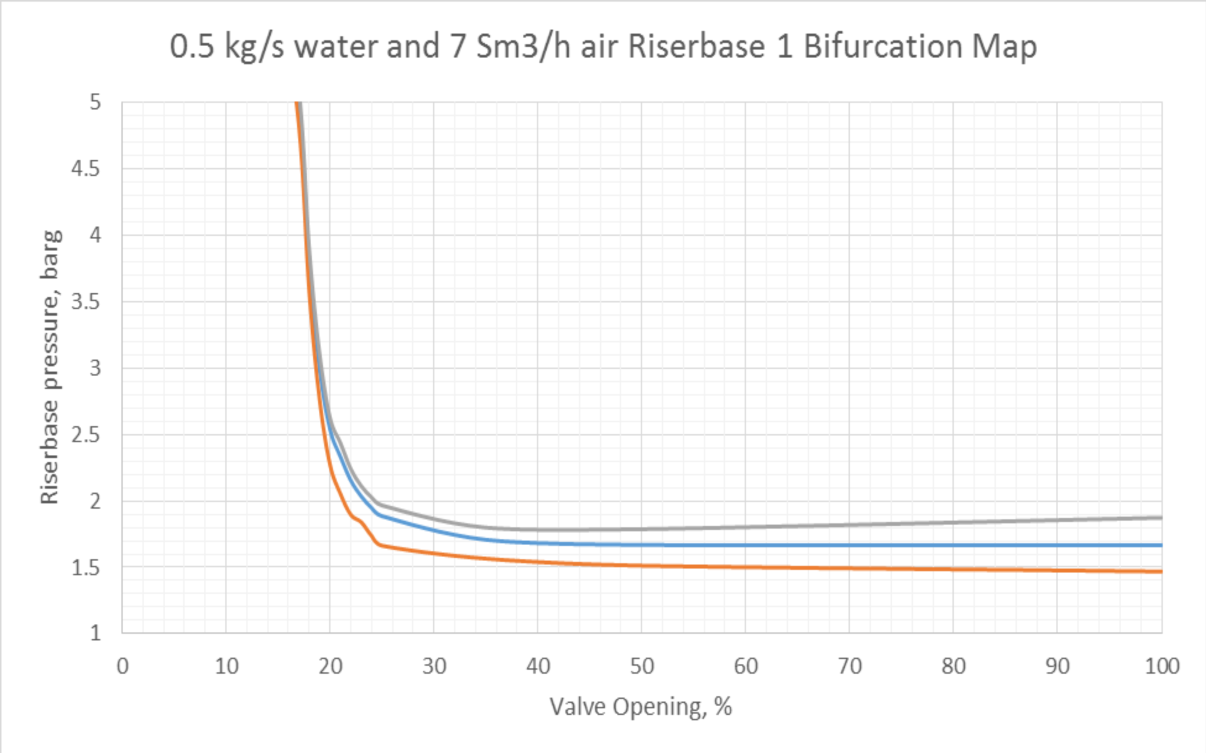


Figure 4-34 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 0.5 kg/s liquid and 7 Sm3/h gas flowrate at the base of the lower limb

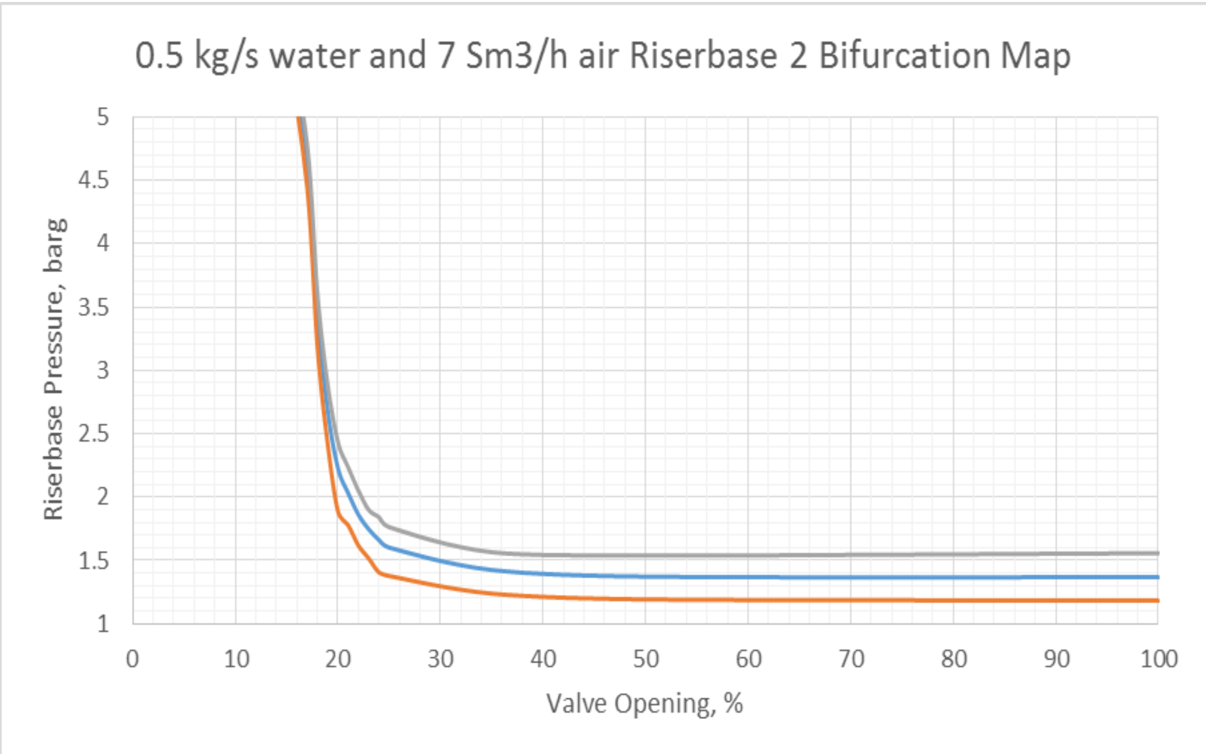


Figure 4-35 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 0.5 kg/s liquid and 7 Sm3/h gas flowrate at the base of the upper limb

The aim of applying the ISC technology on the S-shape riser system just as established by several researchers, for other active control, is to stabilise the flow beyond this valve opening (open loop stable valve opening) while maximising production.

4.3.1.2 Stabilizing unstable flow with active control (ISC)

The control variables used in the ISC technology for stabilising unstable slug flow on the S-shape riser system include; the riser outlet pressure, pressure drop across the choke valve, topside two-phase separator pressure, topside two-phase separator gas outlet flow rate, topside two-phase separator liquid outlet flow rate and three-phase separator pressure combining together to form a slug index more sensitive to slug flow.

The controller design was done using a technique established in Tandoh, Cao and Avila, (2016) as described earlier from previous sections. From Tandoh, Cao and Avila, (2016) the controller gain, K of the ISC can be obtained using (3-13)

In closed loop (for feedback control), there exists a variation in the valve opening, u with varying gas flowrate, Q. Hence the desire is to reduce the pressure drop across the valve which in tend leads to increase in production of the system. Proper control activity aims at stabilising flow beyond / at a larger valve opening, thus, an extra pressure gradient must be introduced through feedback control to compensate for the gradient loss due to increased valve opening. From (3-13), the second term provides the extra gradient which satisfies stable condition in closed loop.

Figure 4-36 shows the riserbase pressure response to a perturbation in gas flowrate since gas flowrate is the parameter of interest. Clearly the operating condition marked red ‘X’ is in the unstable region with a gradient of -0.0393 barg/kgs^{-1} for a 100 % choke valve opening after a slight perturbation in the gas flowrate. This implies that a gradient of 0.0393 is required to render the system stable.

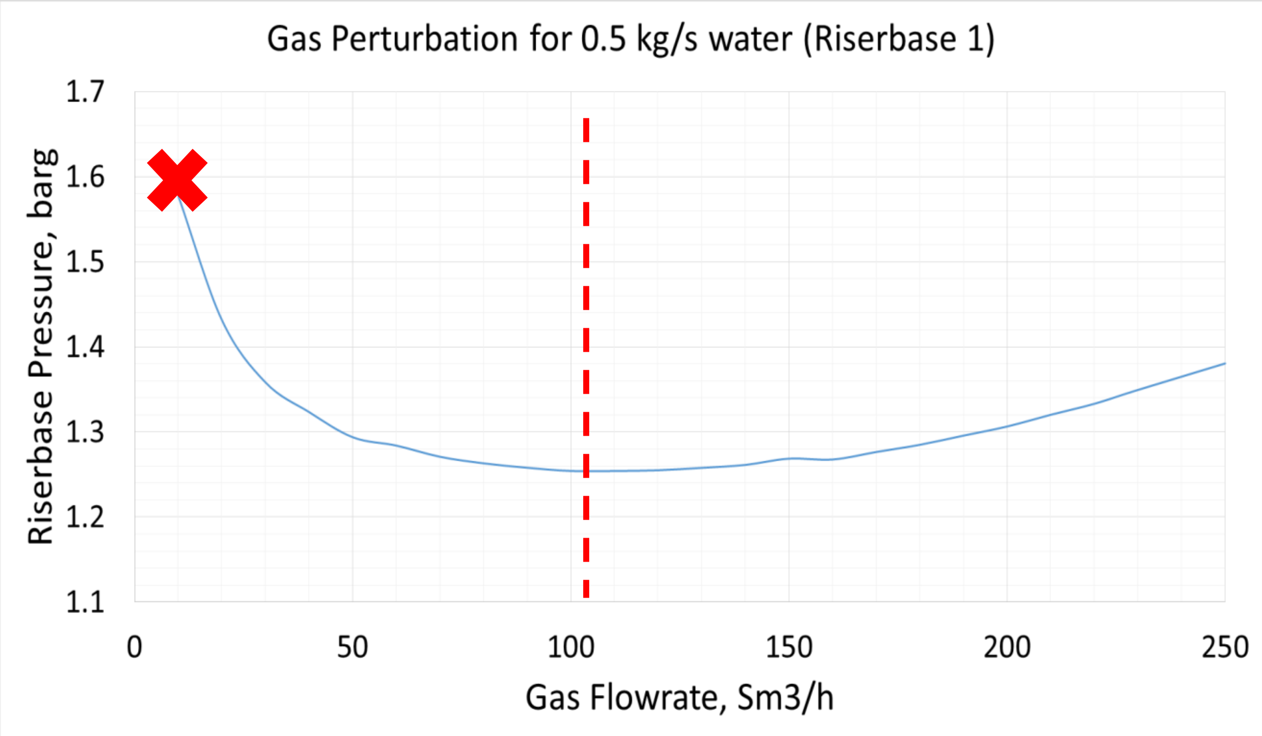


Figure 4-36 Riserbase pressure response from a perturbation in gas flowrate at a constant liquid flowrate (0.5 kg/s)

Since the bifurcation point is 19 %, we aim at stabilising the flow at 21 % valve opening which falls within the unstable flow region in open loop. At 21 %, the resultant pressure gradient is -0.008 barg/kgs^{-1} which is still in the unstable region after a slight perturbation in the gas flowrate. This means that at least an addition gradient of 0.008 barg/kgs^{-1} is needed to stabilise the flow at this valve opening. This can be achieve using the simple but robust controller (ISC). The controller gain, K, which gives this additional gradient is calculated using (3-13).

4.3.1.2.1 Controller design

Riser base pressure depicts the pipeline riser system throughput whiles the stability of the system is dictated by the pressure gradient of the system. From Figure 4-36, to the right and left of the vertical red line represents the stable and unstable flow condition respectively showing how an increase in the gas flowrate could affect the system (positively and negatively). dP/dQ>0 predicts system stability whereas dP/dQ<0 represents an unstable flow.

From Figure 4-36, the operating point marked red is on the left of the stable gas flowrate (red dotted line). This represents an unstable flow condition.

For a slight perturbation in the gas flowrate, the operating condition has a gradient of -0.0393 barg/kgs^{-1} for a fully opened valve (100 %). Again at 21 % valve opening, the point at which stability was aimed to be achieved in closed loop, a gradient of -0.008 bar/kgs^{-1} was attained at the operating point after a slight perturbation in the gas flowrate. This implies that for the system to be stable an additional 0.008 bar/kgs^{-1} gradient is required and is to be supplied by the controller.

From (3-15), the measurement signals, measurement coefficient and measurement weights used in calculating the gain that would provide the additional gradient to make the system stable is presented in Table 4-6. From (3-13),

$$\frac{2aQ^2}{u^3}K\left[\frac{dc}{dQ}\right] = -0.0312857 \qquad \textbf{(4-1)}$$

where $\frac{\partial c}{\partial Q}$ is therefore estimated from the weighted deviations in measurements resulting from a perturbation in Q, 'α' is a constant associated with valve coefficient, mixture density and the given reference liquid flow rate, Q is the gas flowrate, and 'υ' is the valve opening ranging from 0 to 1.

From (4-1) and using Table 4-6, the controller gain, K to provide the additional gradient was obtained to be 2.08. The minimum gain, K of the ISC for any preferred pressure drop gradient at a particular valve opening could be obtained using the above expression in the quest to stabilise slug flow at an increased valve opening.

Figure 4-37 represents the experimental results obtained with the controller in place. From this results, the controller was able to stabilise the flow shortly after it was activated. From time 0 seconds to 600 seconds, the system was operated in an open loop unstable region, 21 % valve open. There exist some significant riserbase pressure fluctuations with a corresponding average pressure of 2.40962 barg. After 600 seconds, the ISC controller was activated and shortly after the pressure fluctuations reduced significantly whiles registering an average riserbase pressure value of 2.35782 barg. Thus, the magnitude of the pressure and flow oscillations reduced significantly which implies the unstable pressure and flow fluctuations became relatively stable and reducing the riserbase pressure by about 2.15 %.

Table 4-6 Controller design parameters

Measurement Signals	**Measurement Weight, w**	**Measurement Coefficient, y**
Riser Outlet Pressure	-0.5050	1.58185
Pressure Drop Across the Choke Valve	-0.4747	1.01804
Topside Two-Phase Separator Pressure	-0.5060	1.000742
Topside Two-Phase Separator Gas Outlet Flow Rate	0.0811	0.8945
Topside Two-Phase Separator Liquid Outlet Flow Rate	0.0501	0.29725
Three-Phase Separator Pressure	-0.5045	0.56385

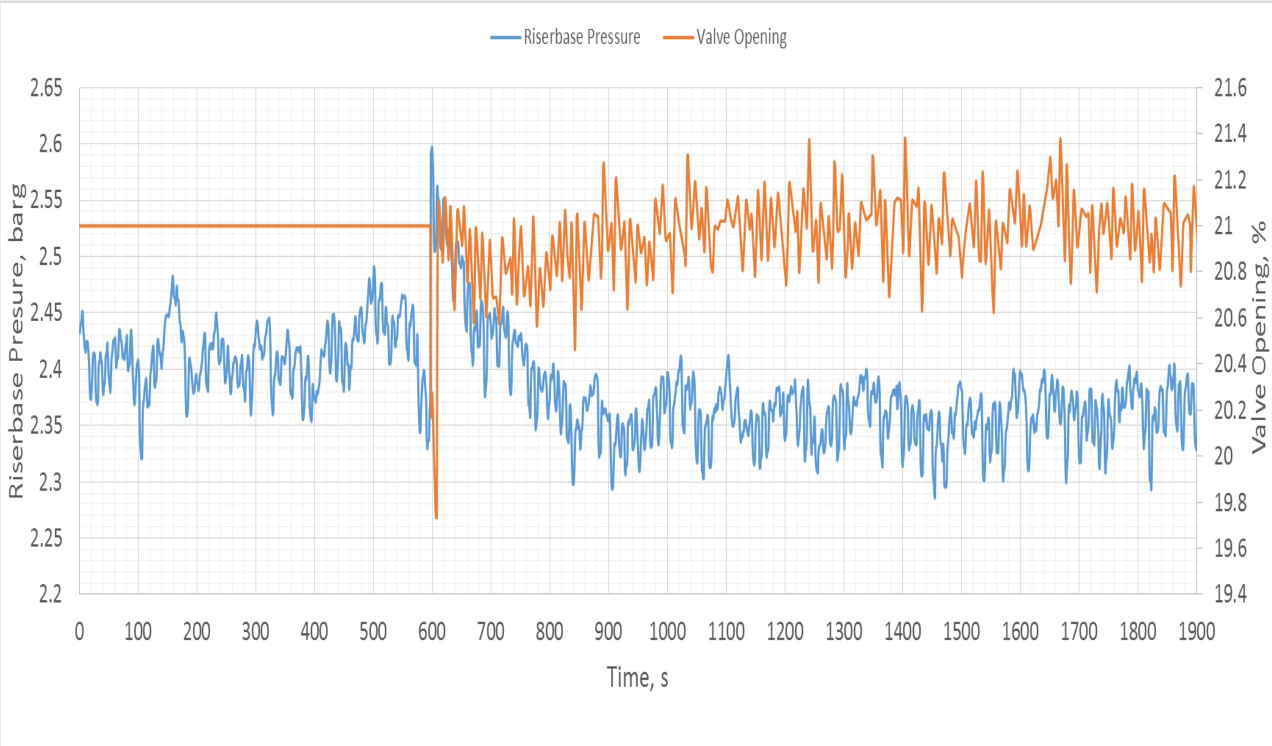


Figure 4-37 Riserbase pressure trend of the system (0.5 kg/s liquid and 7 Sm³/h gas flowrates) in closed loop (ISC)

Averagely, there was a registered percentage valve opening of 21.13 %, which in open loop falls within the unstable region. Optimization of the process was however considered as a stepwise increase of the valve opening (valve opening located further in the unstable region) was done to know the extent for which the system stability could hold. 31 % valve opening was recorded as the optimal valve opening for which the system stability could hold and the corresponding riserbase pressure at this optimal valve opening is 1.83123 barg as shown in Figure 4-38. Figure 4-38 shows the riserbase pressure trend response for the optimal stable valve opening when the controller was in operation and after the controller was turned off thus after about 1700 seconds. Beyond 31 % the system instability sets in as shown in Figure 4-39. Table 4-7 shows the valve openings and their respective resultant achievable riserbase pressure from the experiments using the ISC technology. The maximum achievable valve opening was 31 % resulting to a minimum achievable riserbase pressure of 1.83123 barg.

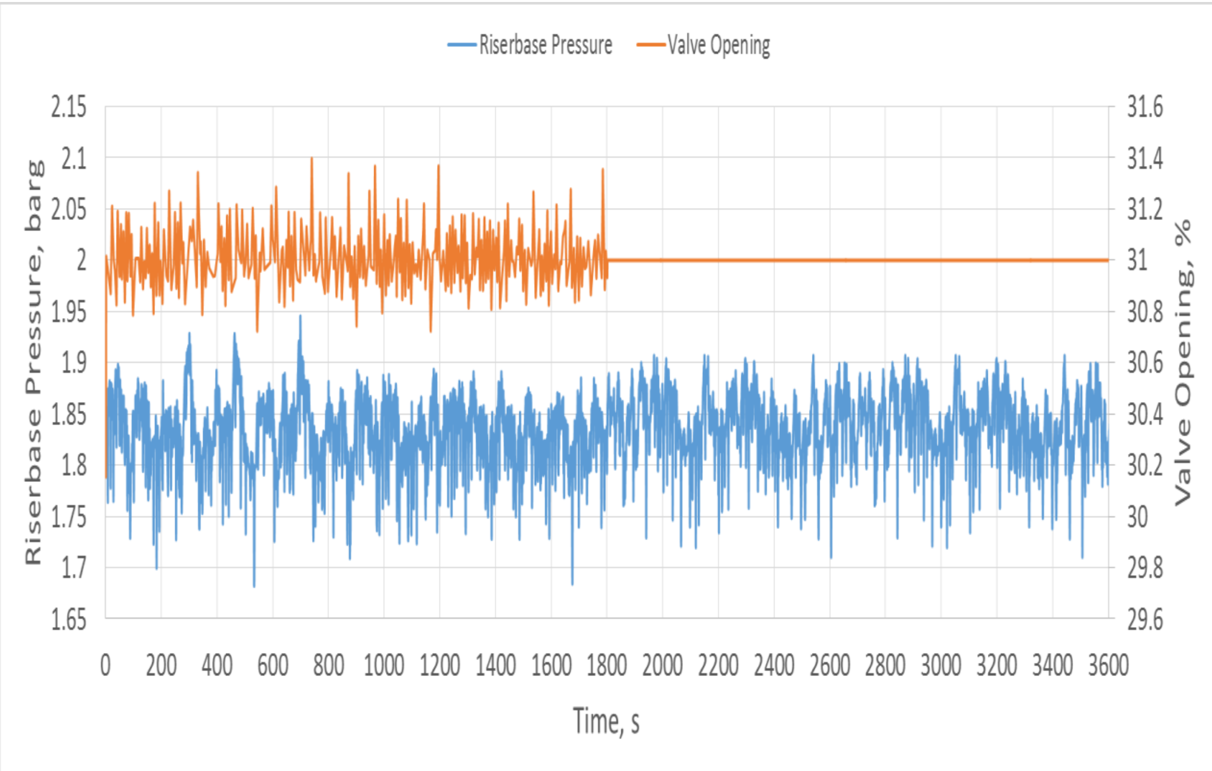


Figure 4-38 Riserbase pressure trend of the system at optimal stable valve opening without (open loop) and with (closed loop) (ISC) (0.5 kg/s liquid and 7 Sm^3/h gas flowrates (0.25 m/s and 0.54 m/s V_{sl} and V_{sg} respectively))

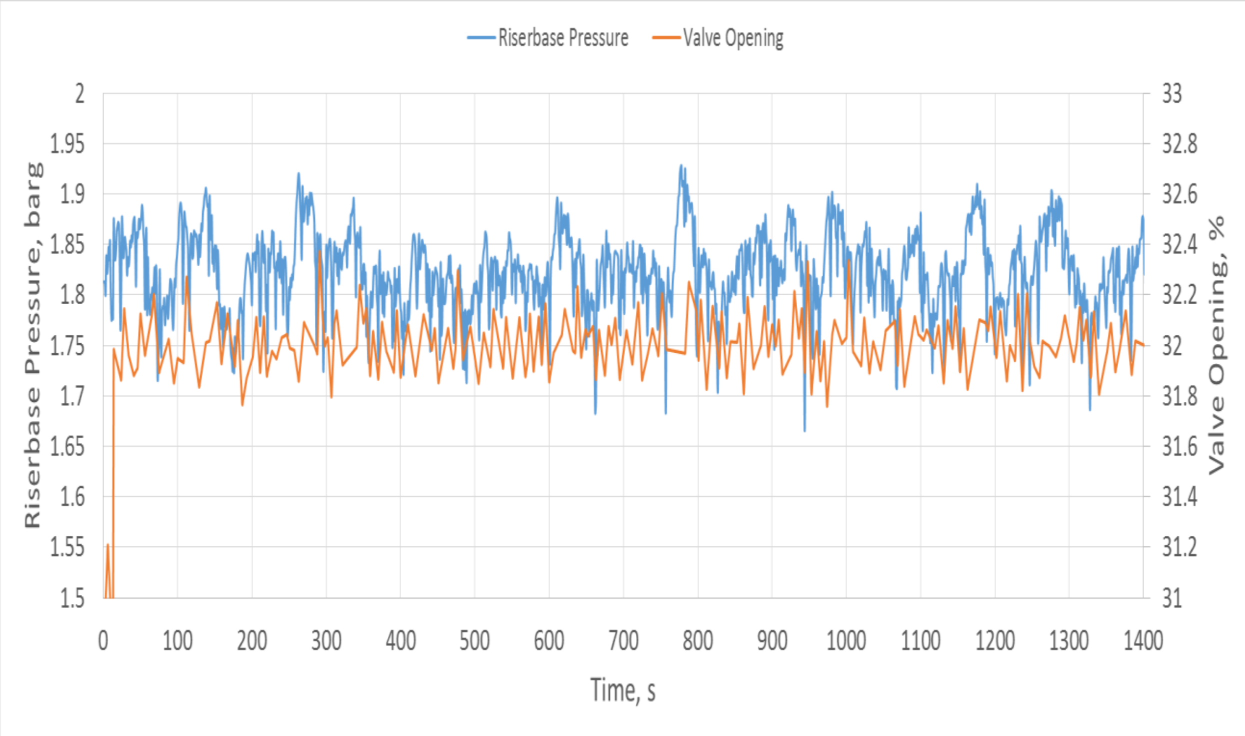


Figure 4-39 Unstable region of the Riserbase pressure trend in closed loop (ISC) (0.5 kg/s liquid and 7 Sm^3/h gas flowrates (0.25 m/s and 0.54 m/s V_{sl} and V_{sg} respectively))

Table 4-7 Resulting riserbase pressure for different valve opening.

Valve Opening, %	**Riserbase Pressure (Open loop), barg**	**Riserbase Pressure (Closed loop), barg**
19	3.0742	2.8574
20	2.6183	2.5900
21	2.4096	2.3578
22	2.2329	2.2133
23	2.1142	2.1021
24	2.0244	2.0210
25	1.9943	1.9663
30	1.8612	1.8409
31	1.8578	1.8312
32	1.8156	1.8220

4.3.2 1 kg/s liquid and 10 Sm^3/h gas flow on the S-shape riser

Similarly, 1 kg/s liquid and 10 Sm^3/h gas which represent a slugging flow condition that is responsive to pressure increase was run to test the controller's (ISC) ability to stabilise the system using the riser outlet pressure, pressure drop across the choke valve, topside two-phase separator pressure, topside two-phase separator gas outlet flow rate, topside two-phase separator liquid outlet flow rate and three-phase separator pressure.

4.3.2.1 Stabilizing unstable flow condition with manual choking

Prior to performing active control, a bifurcation map (open loop stability study) was derived for the flow condition (1 kg/s and 10 Sm^3/h of liquid and gas flowrate) as shown in Figure 4-41 and Figure 4-42. Figure 4-40 shows the pressure trend for 1 kg/s liquid and 10 Sm3/h gas flowrate on the S-shape riser pipeline configuration.

A stepwise decrease in the choke valve opening was done from a valve opening of 100 % to10 %. Riserbase pressures for each condition were recorded. The minimum, average and maximum riserbase pressures of each condition were plotted against their corresponding percentage valve opening (bifurcation map) where the critical bifurcation point was identified.

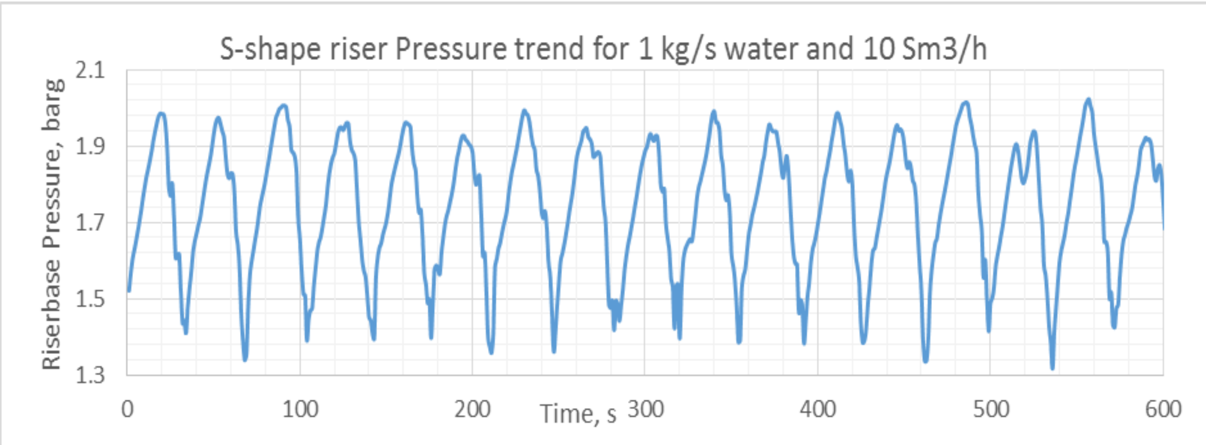


Figure 4-40 Riserbase pressure trend for the 2 inch S-shape riser loop at flow conditions 1 kg/s liquid and 10 Sm3/h gas

In open loop the system stabilises at a 22 % valve opening corresponding to a riserbase pressure valve of 3.8 barg and 3.6 barg for the lower limb and the upper limb as shown in Figure 4-41 and Figure 4-42 respectively.

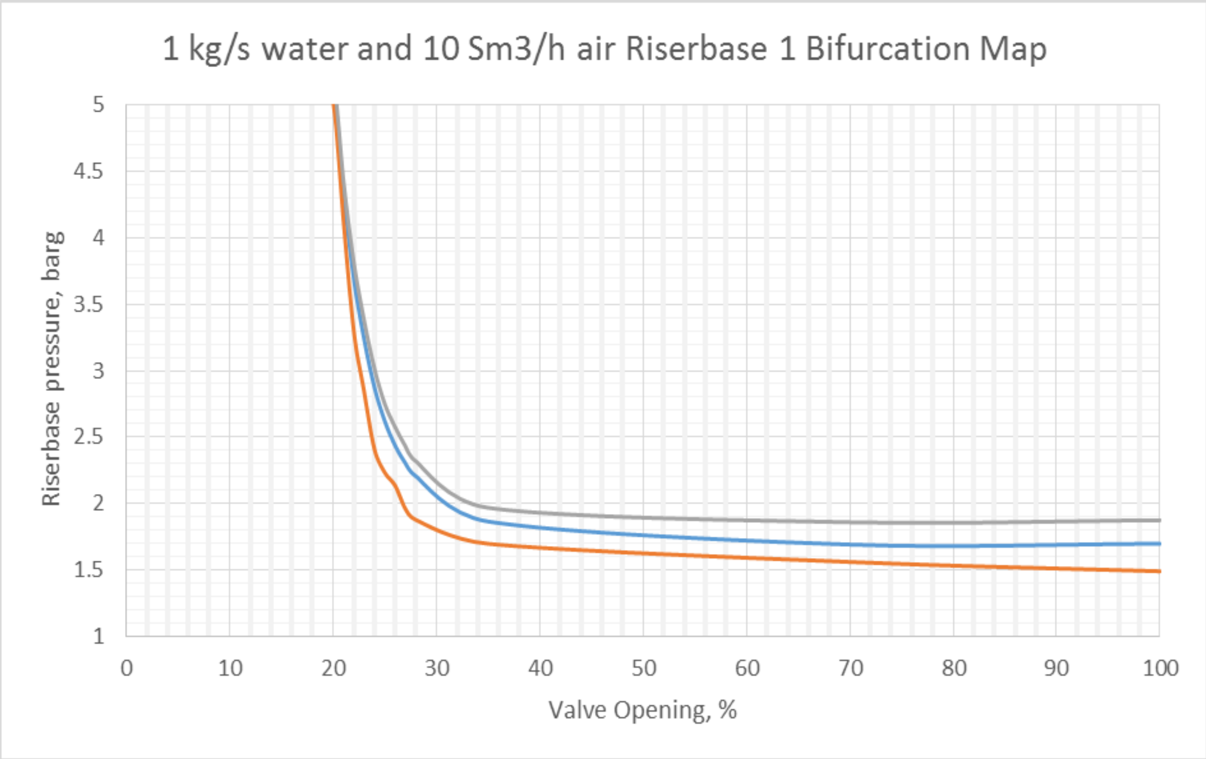


Figure 4-41 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 1 kg/s liquid and 10 Sm3/h gas flowrate at the base of the lower limb

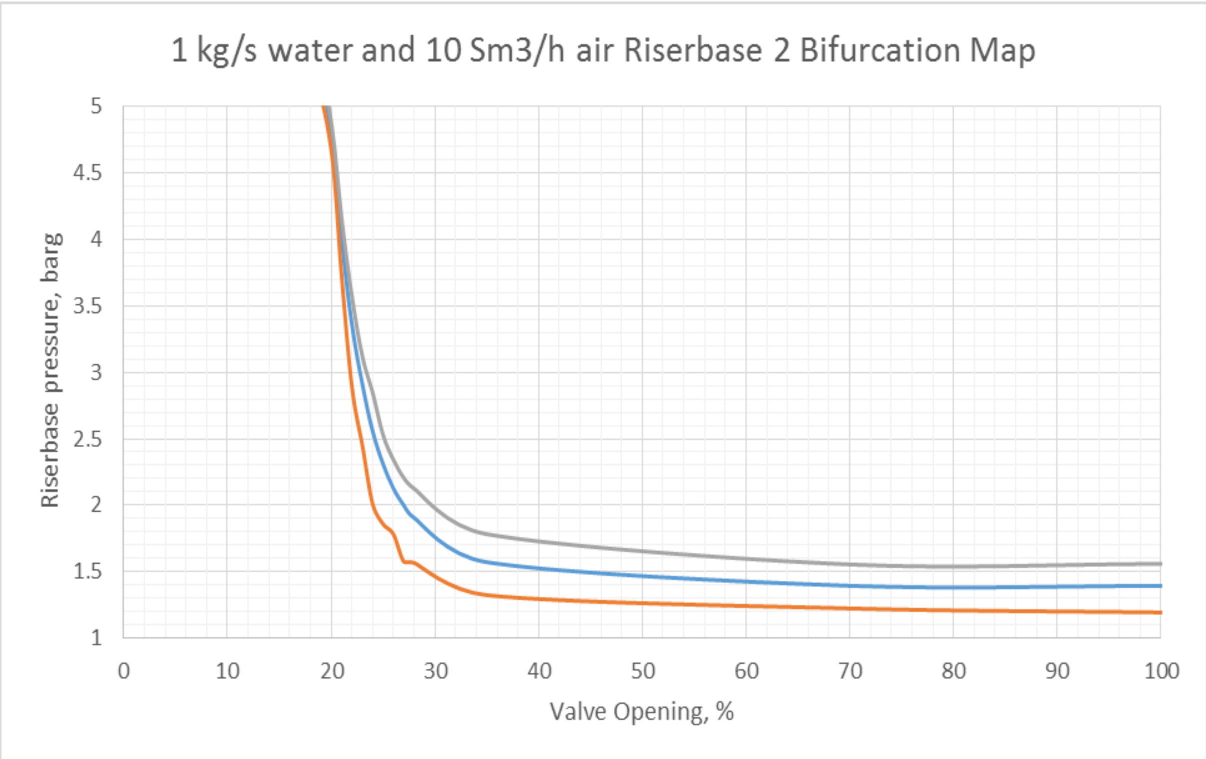


Figure 4-42 Bifurcation map for a 2 inch S-shape riser loop at flow conditions 1 kg/s liquid and 10 Sm3/h gas flowrate at the base of the upper limb

The system stability was assessed based on physical observation and analysis of the pressure fluctuations. Riserbase 1 has a higher riserbase pressure than that observed in Riserbase 2 due to the difference in depth from the point of measurement, thus, the hydrostatic head between riserbase 1 and riserbase 2. In open loop, an unstable flow condition can be stabilized (manually) when the gas flow rate is perturbed. This is given by a differentiation of the pressure with respect to the gas flowrate at a constant valve opening. For this condition to hold, the pressure drop across the valve must be sufficiently large that is, the valve opening must be considerably small which means low flow through the valve resulting in an increase in the riser base pressure. This leads to a temporary / instantaneous reduction in the production rate as there is a reduction in the acceleration of flow hence optimization of the process a necessity. For this reason active control was used to stabilise the flow beyond the bifurcation point using the ISC technology. Thus, translating to a stable flow condition at a higher valve opening which helps in optimizing the available energy. The optimized energy in the reservoir enhances produced fluids in a much steady and regular manner.

4.3.2.2 Stabilizing unstable flow with active control

The controller used in this study is the ISC controller, which combines the riser outlet pressure, pressure drop across the choke valve, topside two-phase separator pressure, topside two-phase separator gas outlet flow rate, topside two-phase separator liquid outlet flow rate and three-phase separator pressure to produce a single control variable more sensitive to slugging flow. The ISC is designed using the technique outlined in (Tandoh, Cao and Avila, 2016) as discussed in previous sections.

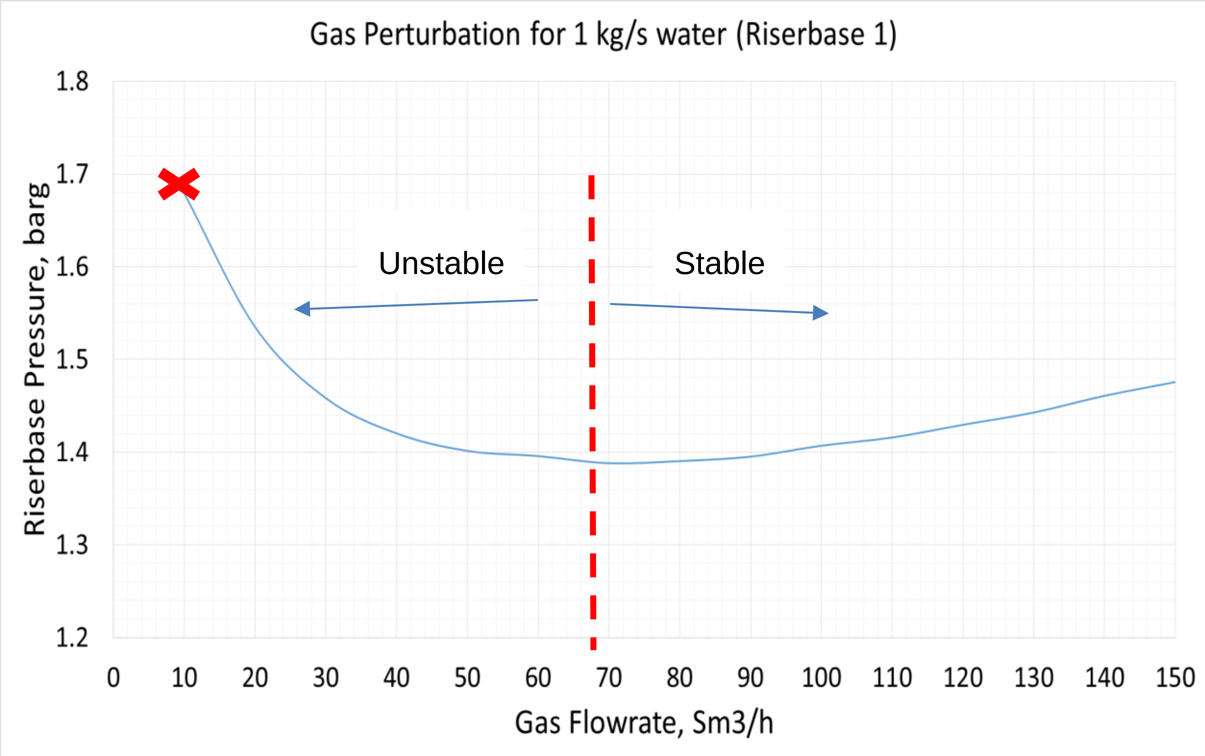


Figure 4-43 Riserbase pressure response from a perturbation in gas flowrate at a constant liquid flowrate (1 kg/s)

Figure 4-43 shows the riserbase pressure response with increasing gas flowrate. As indicated by the red mark the operating condition is in the unstable region and it has a pressure gradient of -0.0175 barg/kgs^{-1} after a slight perturbation in the gas flowrate. That means for the system to be stable it requires a minimum of 0.0175 barg/kgs^{-1}. Since in open loop the system was able to achieve stability at 22 % valve opening, we aim to stabilise the system at 24 % valve opening in closed loop. At 24 % valve opening, a pressure gradient of -0.0065 barg/kgs^{-1} was attained after a slight perturbation in the gas flowrate, representing an unstable flow. This implies that a minimum pressure gradient of an additional 0.0065 bar/kgs^{-1} would be required to stabilise the system. This could however be achieved using a controller to accommodate for the extra gradient introduced by the increased valve opening. (3-15) represents a feedback equation for the riser system using the ISC technology (the extra gradient is represented by the second term in the equation). (3-13) reduces to

$$\frac{2aQ_g^2}{u^3}K\left[\frac{dc}{dQ_g}\right] = \mathbf{-0.011} \qquad \textbf{(4-2)}$$

where $\frac{dc}{dQ_g}$ is the weighted deviations in measurements resulting from a perturbation in Q_g. The measurement weight and coefficient of measurements used in determining the weighted deviation for the flow condition are given in Table 4-8.

Table 4-8 Controller design parameters

Measurement Signals	**Measurement Weight, w**	**Measurement Coefficient, y**
Riser Outlet Pressure	-0.590	2.11013
Pressure Drop Across the Choke Valve	0.3562	1.1007
Topside Two-Phase Separator Pressure	0.2526	1.000028
Topside Two-Phase Separator Gas Outlet Flow Rate	0.3208	1.05225
Topside Two-Phase Separator Liquid Outlet Flow Rate	0.0812	1.2028
Three-Phase Separator Pressure	-0.5931	1.09456

From (4-2), K represents the gain needed to achieve the required pressure gradient that makes the system stable. Using Table 4-8, the attained controller gain, K from (4-2) was determined as 0.894. The same procedure for obtaining the controller gain seen in the previous case study was adopted to determine the attained controller gain.

From Figure 4-44 the pressure oscillations reduce relatively after the controller is activated, at about 600 seconds. The resultant average percentage valve opening observed was 24.15 %corresponding to a 2.8643 barg riser base pressure, which in open loop falls within the unstable region. Increment in the valve opening in closed loop was necessary as system optimization (controller's ability to stabilise system at valve openings located further in the unstable region) is the drive.

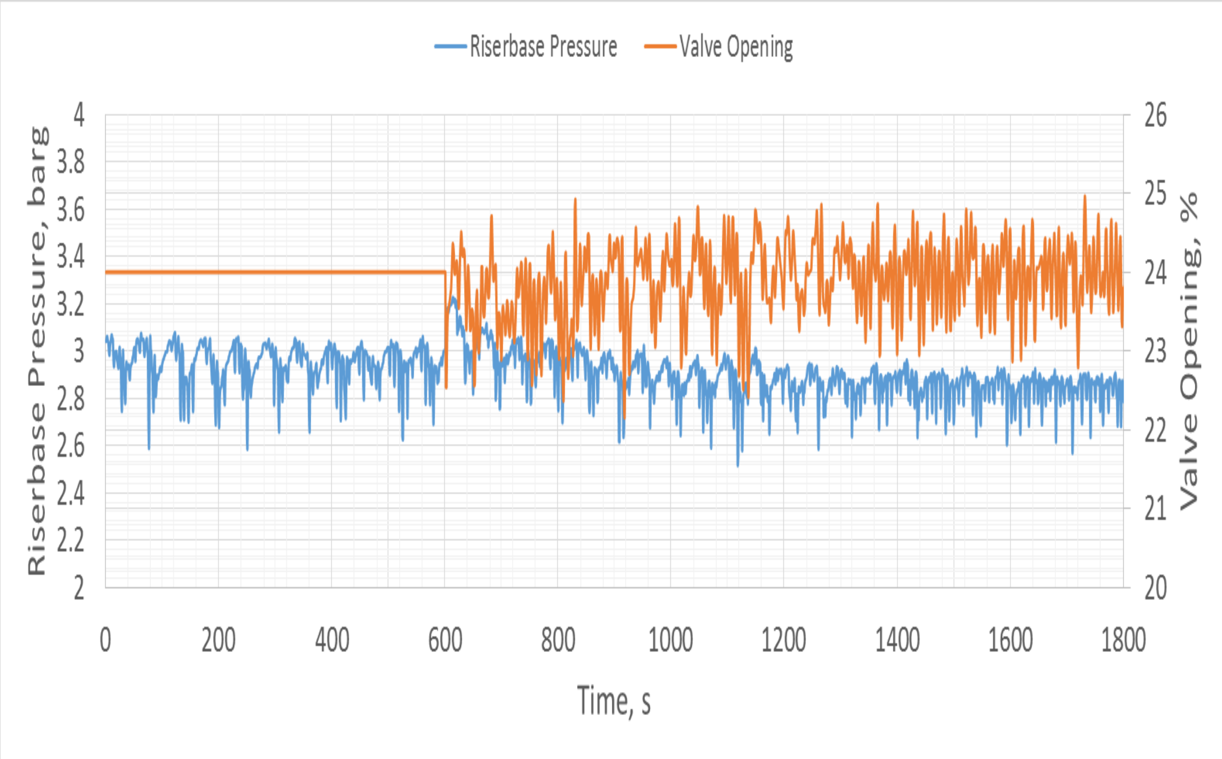


Figure 4-44 Riserbase pressure trend of the system (1 kg/s liquid and 10 Sm^3/h gas flowrates) in closed loop (ISC)

The optimal valve opening for which the system becomes stable was observed to be about 38 % as shown Figure 4-45, corresponding to a riserbase pressure of 1.89025 barg. From Figure 4-45, the system was run at an optimal valve opening of 38 % for 1200 seconds in closed loop using the ISC. This shows a relatively low pressure oscillation. Beyond 1200 seconds, the system was operated in open loop at same valve opening where the system returns to a relatively unstable condition. Beyond this valve opening (38 %) however in closed loop, the system losses its stability. Thus at 39 % valve opening, corresponding to a riserbase pressure of 1.87668 barg, the system becomes unstable as shown in Figure 4-46. Table 4-8 shows the riserbase pressure response for different choke valve openings observed in both open loop and closed loop experimental runs respectively. It can be observed that the riserbase pressures observed in closed loop are of a lower magnitude compared to that seen in open loop. Lower riserbase pressure translates to higher throughout for the same system hence the ISC technology helps increase production and relatively stabilise the fluctuations in the pipeline system.

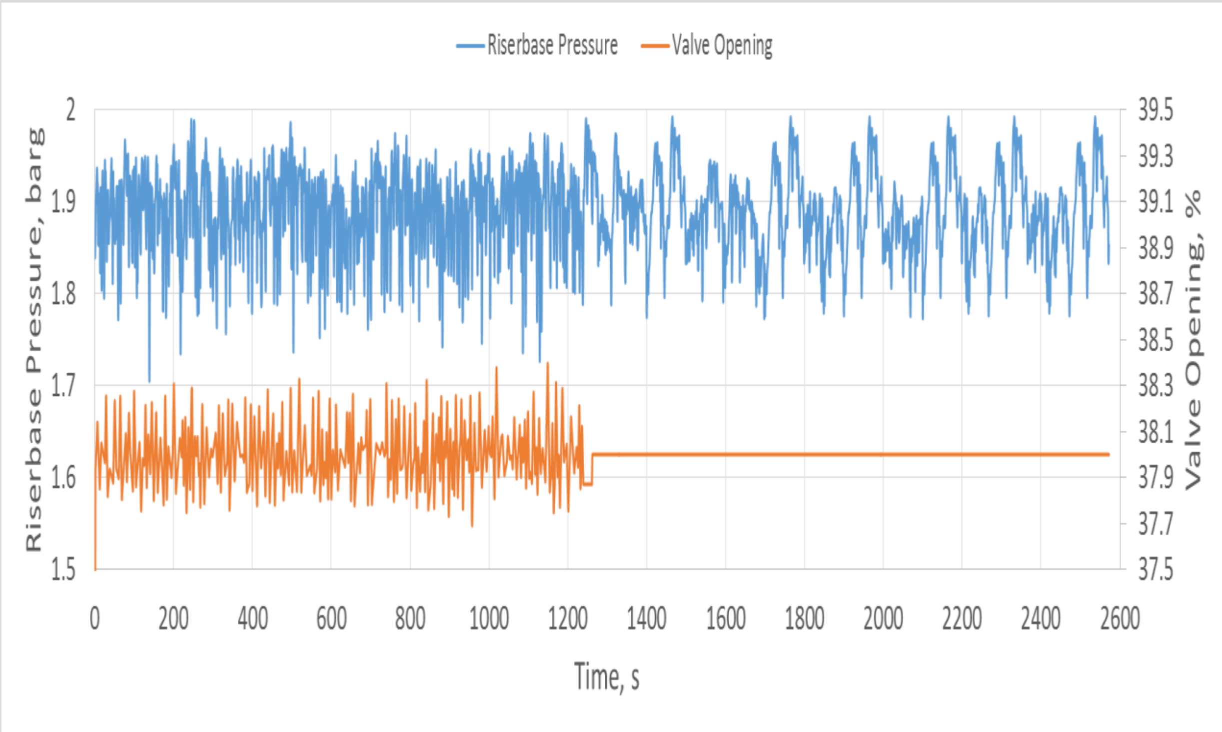


Figure 4-45 Optimal valve opening stable region of the riserbase pressure trend in closed loop (ISC) (1 kg/s liquid and 10 Sm^3/h gas flowrates (0.49 m/s and 0.71 m/s V_{sl} and V_{sg}))

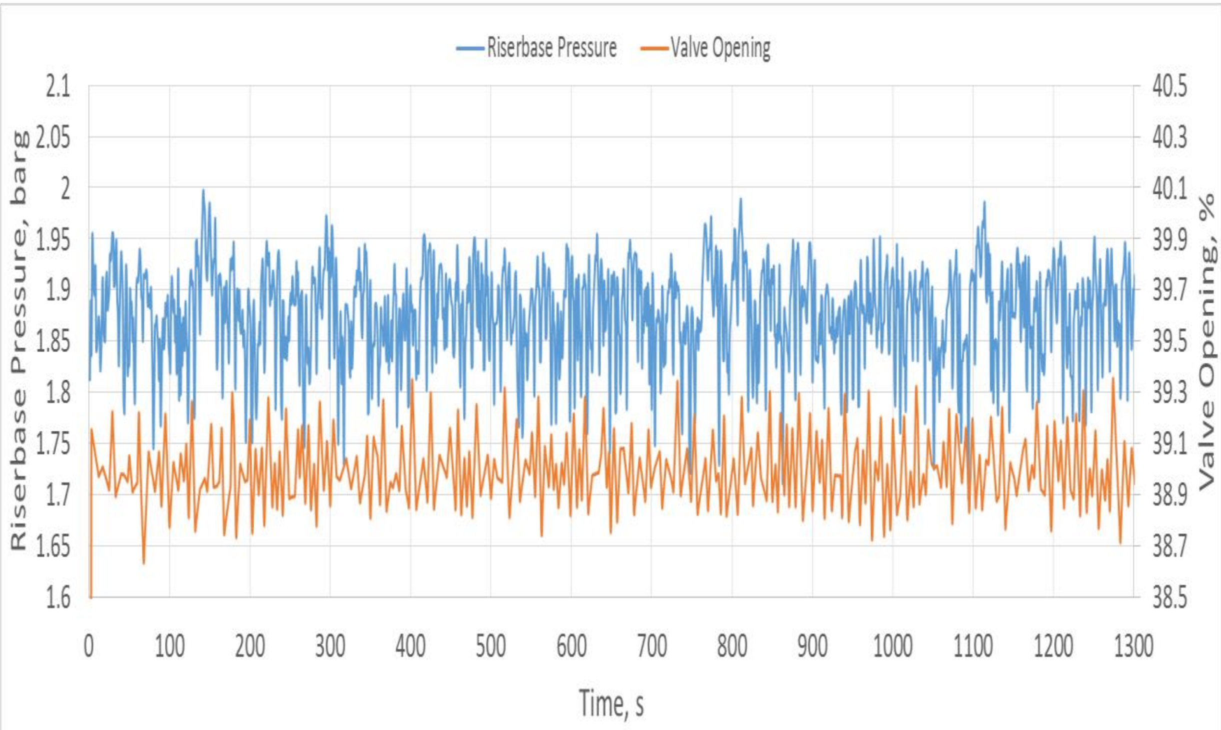


Figure 4-46 Unstable region of the riserbase pressure trend in closed loop (ISC) (1 kg/s liquid and 10 Sm^3/h gas flowrates(0.49 m/s and 0.71 m/s V_{sl} and V_{sg}))

Table 4-9 Resulting riserbase pressure for different valve opening

Valve Opening, %	Riserbase Pressure (Open loop), barg	Riserbase Pressure (Closed loop), barg
22	3.76516	3.45212
23	3.31180	3.17959
24	2.94751	2.86453
25	2.69351	2.63207
26	2.50705	2.47672
27	2.37806	2.34258
28	2.27836	2.25563
37	1.92542	1.90402
38	1.91221	1.89025
39	1.86923	1.87668

4.3.3 Summary

From the study done on the S-shape riser, it could be deduced that unstable flow in the riser could be stabilised beyond the bifurcation point. Thus the system could be rendered stable at a valve opening that exhibits unstable flow characteristics in open loop operation. The riser outlet pressure, pressure drop across the choke valve, the two-phase separator pressure and the three-phase separator pressure are a good combination of signals to be used in the ISC technology.

4.4 ISC for severe slug control on industrial S-shape riser in OLGA

An S-Shape riser is a type of pipeline configuration in the oil and gas industry through which produced fluids are transported. Topography of the seabed, wave current of the sea and water depth contribute to the configuration of risers. Systems such as the S-shape risers gather flow from several wells and the flow behaviour that co-exists in S-shape risers could be unstable which is very disturbing. Severe slug flow often happens in S-shape risers due to its configuration thus the dip in the pipeline system or even at low flowrates usually in brown fields.

For this section, an investigation was conducted on an industrial size S-shape riser with the aim of controlling or stabilising a severe slugging condition using a dynamic flow simulator, OLGA. The associated large pressure and flow fluctuations of severe slugging flow conditions have an adverse effect on the pipeline system and even production on the whole. The model used in this study is a replica of a single riser from a satellite field of Chevron Energy Technology Company. The S-shape riser model used in this work could be sectioned into three, thus, the feed section, riser/pipeline section and the topside/processing section. The pipeline system characteristics, the flow condition and fluid composition would be considered / discussed in subsequent sections of this chapter.

4.4.1 S-shape riser system

4.4.1.1 Pipeline / riser section

The model to be examined is typical S-shape riser, with two dips in configuration represented by 1 and 2 on Figure 4-47 and an inlet near horizontal section upstream the riserbase. The inlet near horizontal section is approximately 1800 m long with an equivalent riser height of approximately 365 m. The pipeline system has a uniformly distributed diameter (0.2794 m), with a choke valve of same diameter at the top of the riser.

Detailed pipeline length, sections and segments which define the length, diameter, elevation, boundary condition, pressure trends and their likes are represented in the description of Case 1 from the previous section.

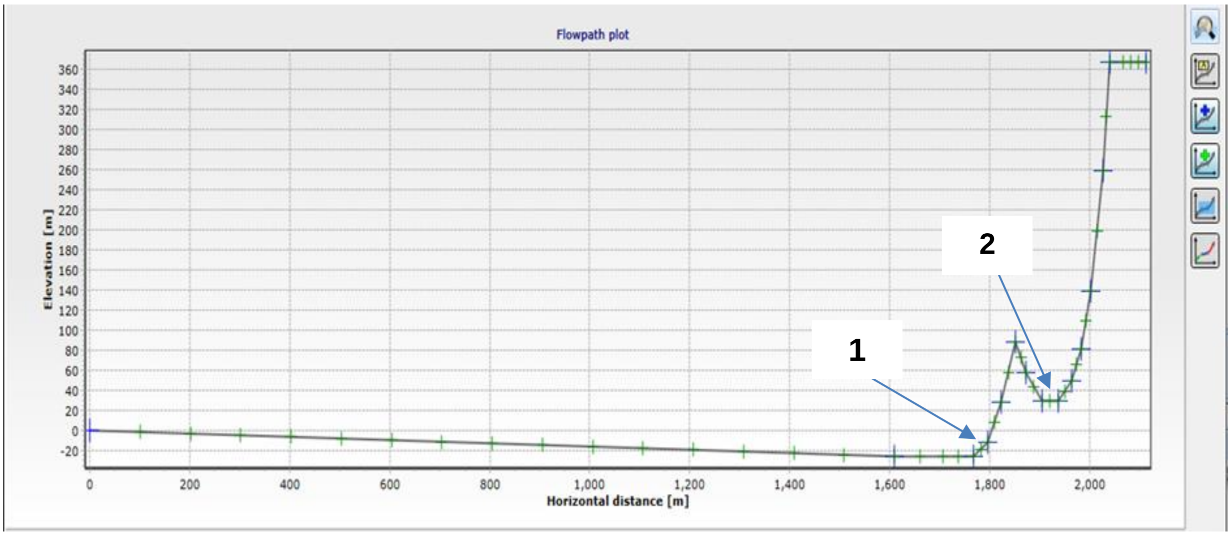


Figure 4-47 OLGA flow path configuration for the riser (S-shape)

4.4.2 Bifurcation analysis

A manual stability study was conducted on the system where the choke valve was manipulated and the resulting output (riserbase pressure) used to develop a Hopf bifurcation map. A bifurcation point was derived from the bifurcation map of the system from which we aim to stabilise the system beyond this valve opening in active control mode. The stability loss of the system is a resultant of a pair of complex poles crossing the imaginary axis on the s-plane which however changes the sign of the real part of the pole from negative to positive (Stewart and Thompson, 2001). From the bifurcation analysis, the bifurcation maps derived are shown in the previous section (Figure 4-25 and Figure 4-26 for Riser 1-2.2 and Riser 1-7.2). Riser 1-2.2 and Riser 1-7.2 represents the points marked 1 and 2 respectively from Figure 4-47.These represents the corresponding pressure trends at the first and second dips respectively of the S-shape riser for each valve manipulation. The valve was actuated to investigate the pressure response at the downhole of the system for the flow condition described earlier.

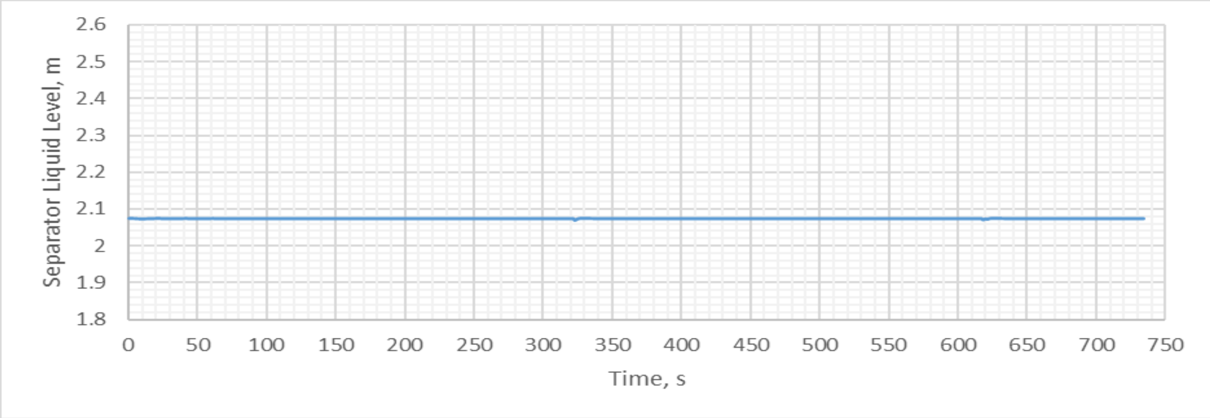


Figure 4-48 Separator liquid level at 9 % choke valve opening at the boundary condition

From Figure 4-25 and Figure 4-26, it could be observed that under manual choking, the system pressure (riserbase pressures) was stabilised at a 9 % valve opening. This forms the lowest pressure beyond which we aim to stabilise the system. At 9 % valve opening, the pressure at the riserbase was observed to be higher than at a valve opening of 100 % due to the additional static back pressure introduced by reducing the valve opening. Riserbase pressure directly relates to the rate of production of a system. From Figure 4-48, the separator liquid level at a 9 % choke valve opening at the boundary condition shows a relatively constant liquid height of about 2.08 m. This represents a steady flow in the system translating into the steady liquid level in the separator.

Instabilities in the S-shape riser system could be caused by the compression and expansion of the gas phase. Nevertheless, expansion or compression of gas in the system can greatly affect the system positively (stable) or negatively (unstable) at a constant liquid mass flowrate. Figure 4-49 shows the corresponding average riserbase pressures with varying gas flowrate while liquid mass flowrate was kept constant at a 100 % valve opening. From Figure 4-49, the region to the left and right of the broken line represents both unstable and stable operating region respectively. It could be observed that the operating point falls within the unstable region. For this system to be stable, approximately 2.1 kg/s mass flowrate of gas would be required. Thus the boundary / operating condition requires approximately 1.2 kg/s gas mass flowrate in addition to make the system stable. This explains why gas lift or injection could be used to stabilise slugging flow.

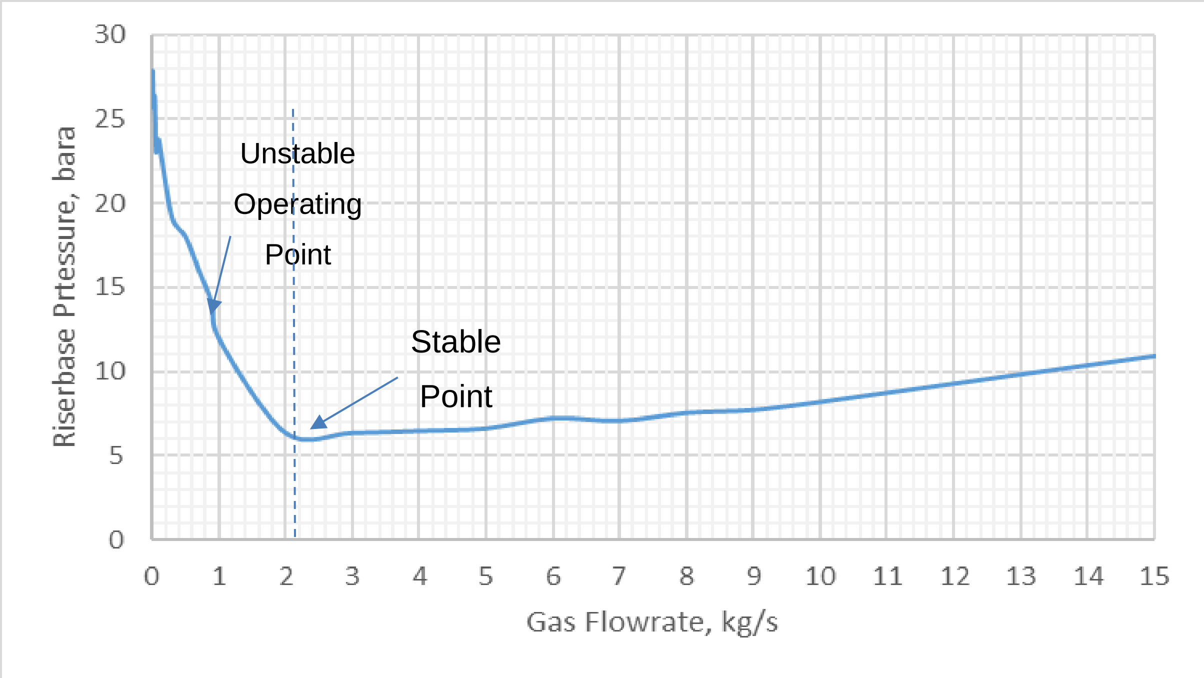


Figure 4-49 Riserbase pressure as a function of gas flow at 100 % valve opening

At a point on the curve, thus the minimum riserbase pressure point, the system becomes stable. This was when the gas flowrate was just enough to keep the liquid flowrate steady. Any further increase in the gas flowrate beyond the stable region, makes the system dominant of frictional force. The increment in the frictional force with increasing the gas flowrate makes the riserbase pressure continually increase. This translates to why there is a constant increment of the pressure drop across the choke valve with increasing gas flowrate.

4.4.3 **Stabilising slug flow using ISC controller**

4.4.3.1 Pressure gradient

Upon attaining the bifurcation point using manual choking, implementing the ISC controller and tuning the controller was required. This was aimed at stabilising flow at a larger valve opening relative to the maximum stable valve opening in open loop. From Figure 4-49, the riserbase pressure gradient after a 1 % increase in gas flowrate corresponds to -112.829 bar/kg/s, which was evident that it is in the unstable region. Hence an additional gradient of 112.829 bar/kg/s would be required to stabilise the system.

For control purpose, we aim to stabilise the system at a valve opening 11 % in active control mode. The riserbase pressure response to increasing gas flowrate for the system operating condition at 11 % valve opening is shown in Figure 4-50. After a 1 % increase in gas flowrate, the riserbase pressure gradient yielded -2.827 bar/kg/s. The negative pressure gradient shows that the system is in an unstable region and therefore requires additional pressure gradient to make the system stable. For the system to be stable at a larger valve opening relative to that in manual choking a controller was required. Necessary extension of the model was made for implementing and testing the Inferential Slug Controller (ISC).

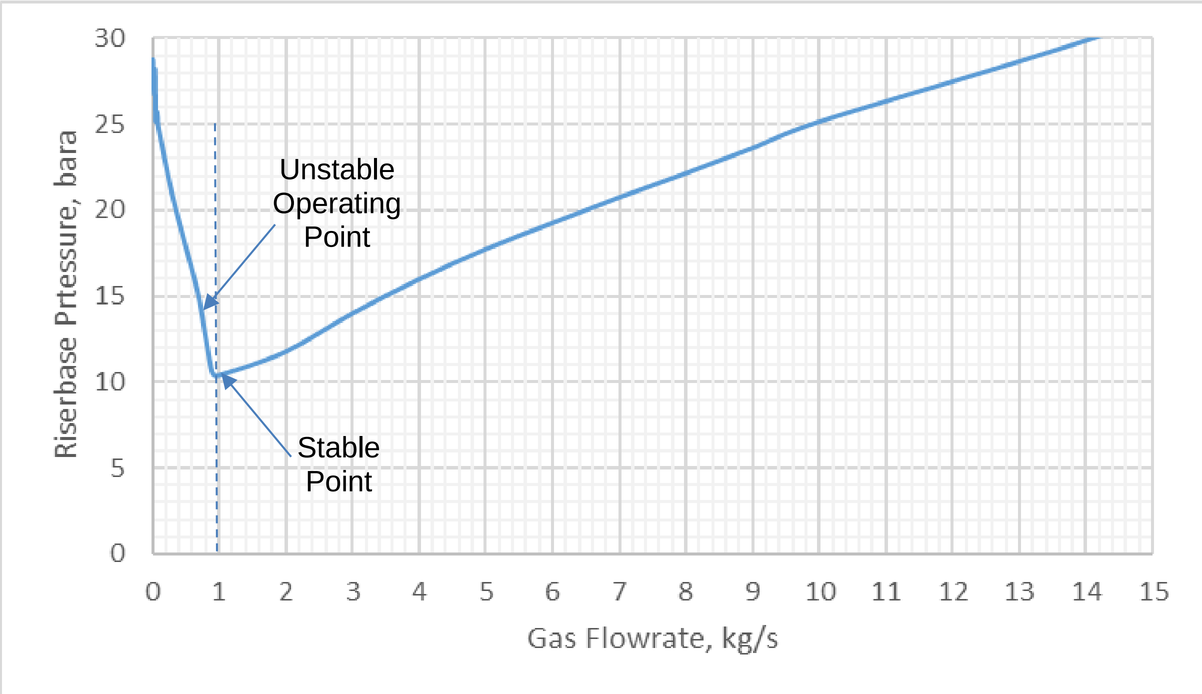


Figure 4-50 Riserbase pressure as a function of gas flow at 11 % valve opening

4.4.3.2 Implementation of ISC controller in model

For the benefit of stabilising severe slug flow on the industrial S-shape riser system, 4 measurement signals were used. The four measurement signals used in this work included; the riser top pressure (PT), the gas mass flowrate (GG), the liquid density (ROL) and the total liquid mass flowrate (GLT) due to the readily availability of these signals. Other factors that determined which signals to be deployed included the accuracy and the signal sensitivity to noise. Figure 4-51 shows the layout of the ISC on the S-shape riser in OLGA.

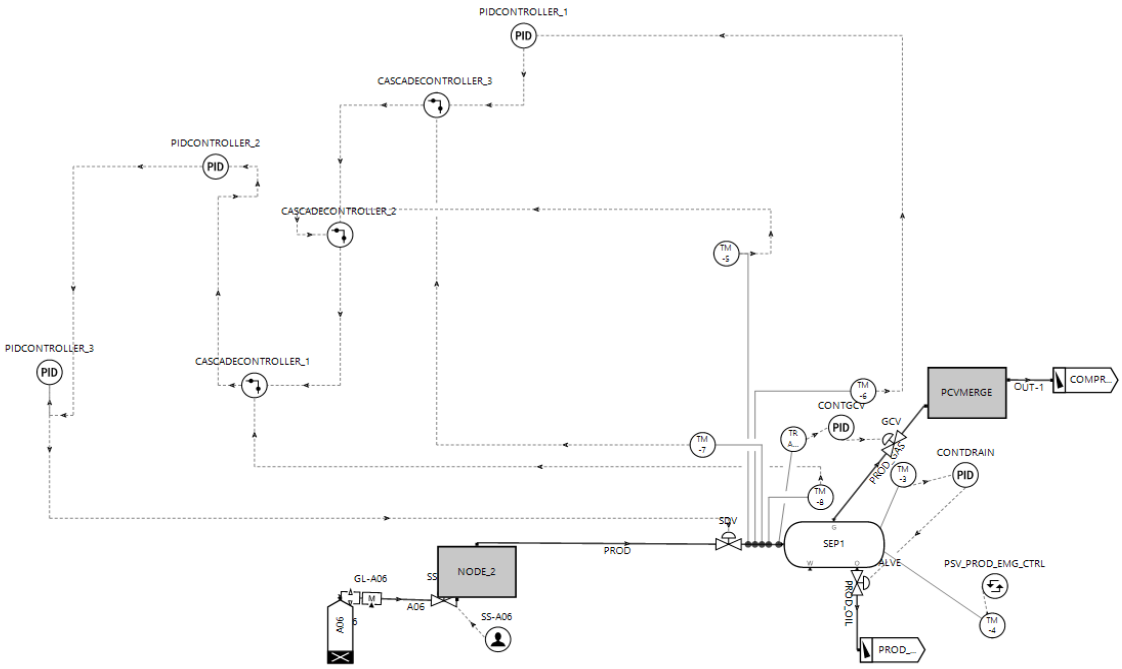


Figure 4-51 ISC layout on S-shape riser in OLGA

The ISC in OLGA was modelled in cascade configuration as shown in Figure 4-51. With the use of four measurement signals (from Figure 4-51), the valve opening was determined using the control law (3-15). Expanding (3-15) produces,

$$u = u_0 + K[(-k_4\left(\frac{k_3}{k_4}\left(\frac{k_2}{k_3}\left(-\frac{k_1}{k_2}y_1 - y_2\right)\right) - y_3\right) - y_4) - R] \qquad \textbf{(4-3)}$$

where **u_o** is the choke valve nominal value which is predetermined and manually set to a position where the flow becomes stable or within an acceptable range, **K** is the control gain, $\boldsymbol{W=[k_1,k_2,\ldots,k_{n-1},k_n]^T}$ is the vector of measurement weights, $\boldsymbol{Y=[y_1,y_2,\ldots,y_{n-1},y_n]^T}$ is the vector of measurements, **W^TY** is the control variable, **n** is the number of measurements and **R** is the set point of the control variable.

4.4.3.3 ISC-controller design

The principle behind the ISC is the combination of several topside measurements at the topside of the riser through simple algebraic scheming. Already existing facilities provide most of these measurement signals since it is already being deployed to control other components of the system thereby making it cheaper to implement (Cao, Yeung and Lao, 2010a). From the new control law, (4-3), algebraic calculation was used to obtain a single control variable from the four measurement signals for the reason being that, it becomes sensitive to slug flow therefore being the sole origin of alteration in the controlled variable.

The unidentified controller parameters was determined using the tuning technique as discussed in previous sections are presented in Tandoh, Cao and Avila, (2016).

4.4.3.4 Controller tuning and gain determination

Determining the gains of a controller is the first step in tuning a controller. This is done properly to help tame and subdue the severe slugging flow efficiently. The measurement signal weights from the four measurement signals were used to determine the gains using a systematic approach. The measurement weights and deviation in vector of measurement signal from the systematic approach are presented in Table 4-10.

Table 4-10 ISC tuning parameters

Measurement Signal	Signal Label	Measurement Weight	Deviation in Vector of Measurement Signal, dY/dQ_g	Controllers
Gas Mass Flow (GG)	k_1	0.436	1.9521	PIDCONTROLLER_1 kA - 0.7376
Liquid Density (ROL)	k_2	-0.591	0.945	CASCADECONTROLLER_3 kB – 1.3964
Topside Pressure (PT)	k_3	-0.423	0.0797	CASCADECONTROLLER_2 kC - -0.7979
Total Liquid Mass Flow (GLT)	k_4	0.531	0.859	CASCADECONTROLLER_1 kD - -0.5305

4.4.3.5 Gain implementation

The final controller parameters, thus the control gain and the valve set point for the last controller from Figure 4-51 (PIDCONTROLLER_3) were tuned using the approach illustrated in previous sections and also presented in Tandoh, Cao and Avila, (2016).

From (3-13), the estimated minimum gain of the ISC represented by K was 0.0367. The implemented controller on the system was turned on, with a 1 % stepwise increase in the choke valve, the resultant pressure trend is shown in Figure 4-52 and Figure 4-53.

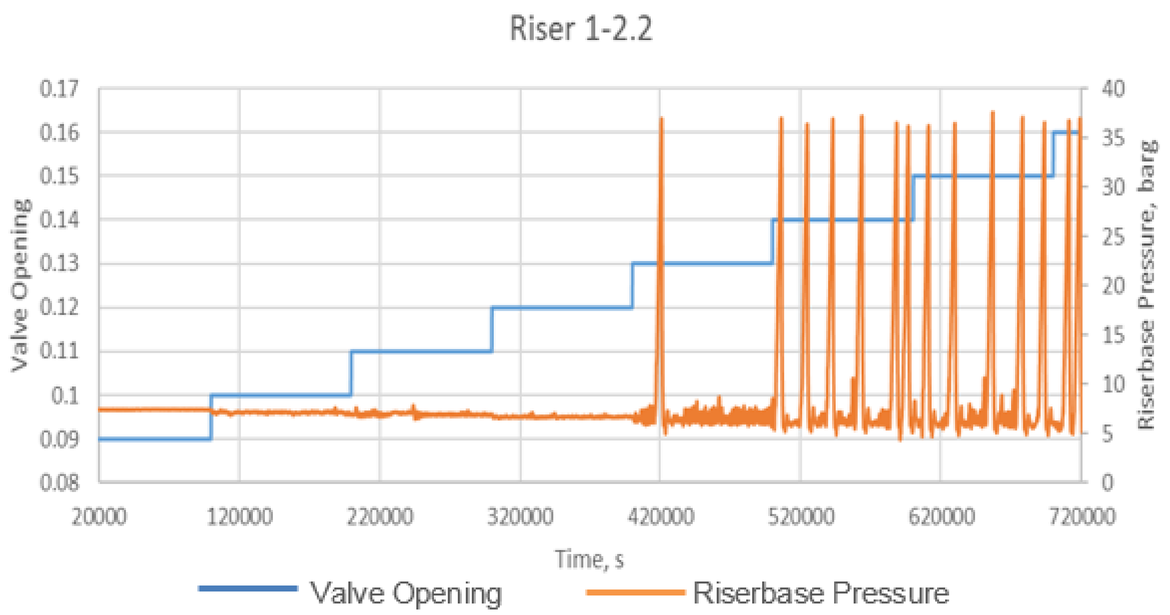


Figure 4-52 Riserbase pressure (Riser 1-2.2) response with ISC in action

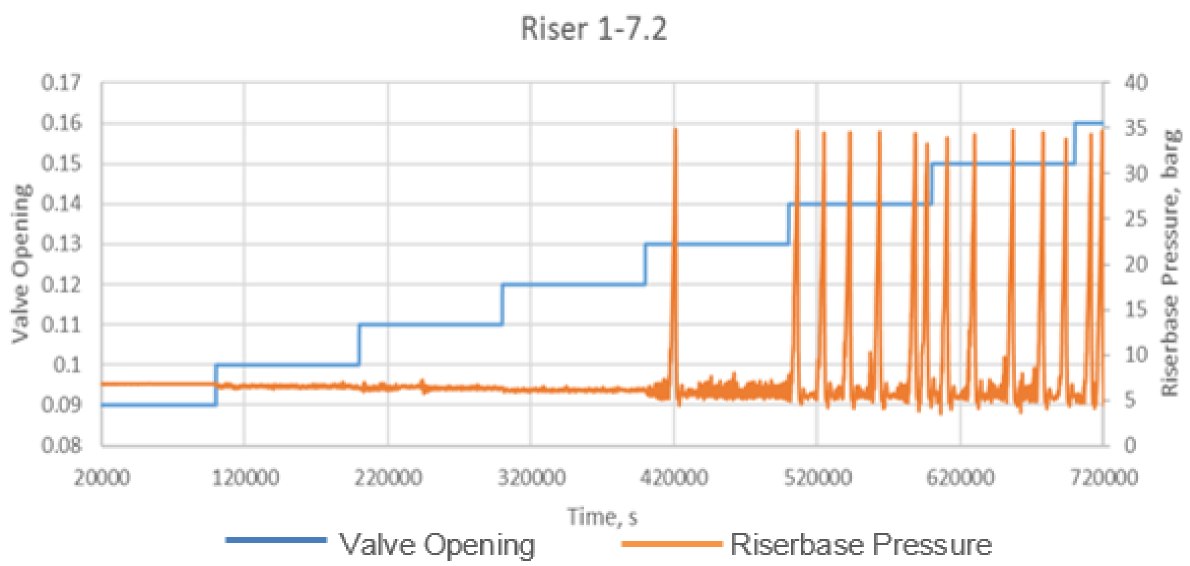


Figure 4-53 Riserbase pressure (Riser 1-7.2) response with ISC in action

From Figure 4-52 and Figure 4-53, it was observed that the ISC was able to stabilise the flow at a 12 % choke valve opening. Thus the ISC stabilised the flow at an increased 3 % in the choke valve opening.

4.4.3.6 Validation of system stability with the ISC

As a confirmation of the stability of the system (S-shape riser) equipped with the ISC at the respective investigated stable valve openings, the separator liquid level/height was assessed. Figure 4-54 shows the separator liquid level when the ISC was in action.

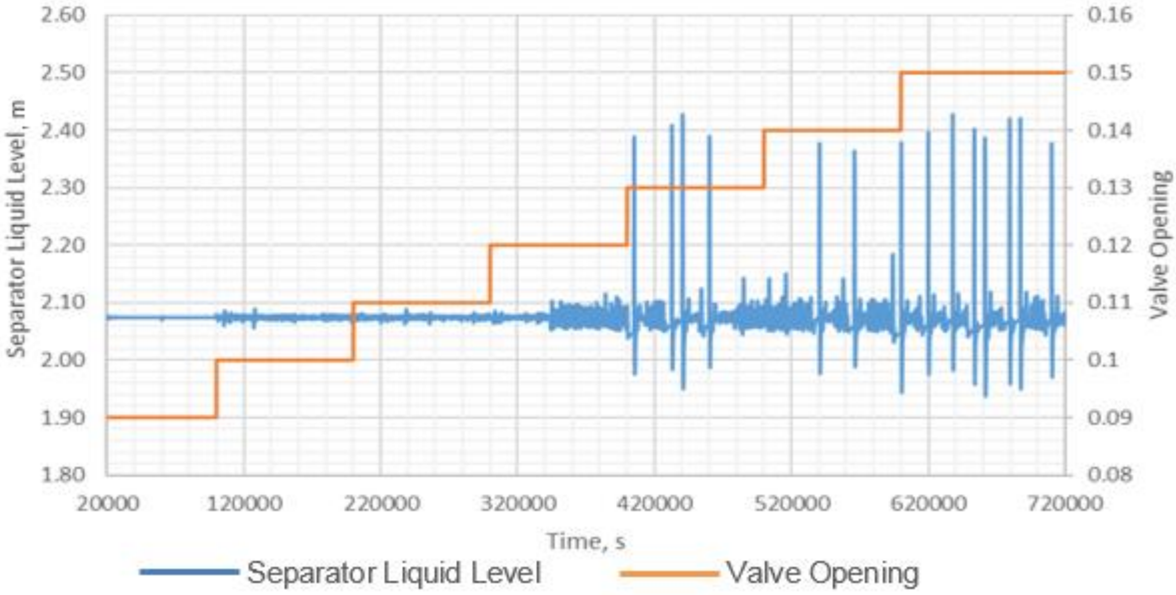


Figure 4-54 Separator liquid level for the S-shape system with ISC in action

From Figure 4-54, it was observed that the separator liquid level was relatively steady (between 2.06 m to 2.1 m) when the ISC was in action from a choke valve opening of 9 % to 12 %. For choke valve opening beyond a 12 %, the separator liquid level began to fluctuate (between 1.9 m to 2.42 m) which represents instabilities in the flow from the system.

4.5 Summary

Slugging as a flow assurance challenge has been at the centre of attraction to several research works and operators. Topside choking, the most common means of subduing slugs, has been known to affect the system negatively in the sense that it reduces the production output of the system.

From this chapter, a general understanding on the flow dynamics in the S-shape riser pipeline system was outlined. A slug envelope was defined from the established flow regime map on the S-shape riser pipeline system. A theoretical understanding of the S-shape riser dip effects on the system stability has been investigated.

The slug attenuation potential of the inferential slug controller (ISC), an active slug control technique, has been assessed through both experiments and simulations using OLGA. Comparatively, it was established that the ISC achieved stability of an unstable slug flow at a larger valve opening to choking. Thus, optimising the drop in pressure across the choke valve.

Specifically through experiments, an additional 12 and 16 percentage valve opening was achieved for the two flow conditions which translates to a reduction in the pressure at the base of the riser which in effect reflects a higher production through the system. Through simulation, an additional 3 % valve opening was achieved for system stability. This reduces the riserbase pressure which reflects in the production output of the system. The smaller additional percentage valve opening in simulation compared to experiments was because of the difference in the pipeline diameters. The best combination of signals was also assessed. The ISC robustness was a key benefactor to this achieved additional benefits however further advancement in the robustness of the controller will be explored in subsequent chapters.

5 EXPERIMENTAL STUDY ON UNSTABLE SLUG FLOW MITIGATION POTENTIAL OF PSUDO SPIRAL TUBE (PST)

5.1 Introduction

Due to the negative repercussions associated with unstable slug flow, the interest of the oil and gas industry has been to either subdue or tame slug flow in pipelines. On a broader view, slug mitigation techniques have been classified under active and passive mitigation methods. Based on theoretical, experimental works and real field studies (live field) numerous slug mitigation and control have been proposed and illustrated in the literature. Some of these works aimed at only taming slug, whilst others considered the profitability margin of the production system on a whole. Some great ideas to deal with unstable slug flow, either addressed slug flow by containing, managing or even accepting it. Regardless, some mitigation circumstance has to be established when dealing with unstable slug flow. It is patently obvious that slug flow in the oil and gas industry and its control in pipeline riser systems desire more thoughtfulness, so to avoid its effect on the system.

This chapter is directed at exploring the use of a pseudo spiral tube (PST), a passive slug mitigation technique, as a means for slug attenuation or mitigation in the 2 inch S-shape pipeline riser system configuration. In Chapter 2, a theoretical background on passive slug mitigation was provided with highlights on flow conditioners to offer a better understanding to this method. Chapter 4 has presented an understanding on the dynamic behavior of gas – liquid flow in the 2 inch S-shape riser system establishing the region where unstable flow occurs. Having established the slug envelope for a combination of several gas-liquid flow conditions, an experimental investigation of the mitigation potential of the PST on an unstable slug flow condition. The behavior of the dynamic unstable slug flow was further investigated in terms of stability analysis. This chapter also proofs the concept to demonstrate the slug mitigation potential of a Pseudo Spiral tube (PST).

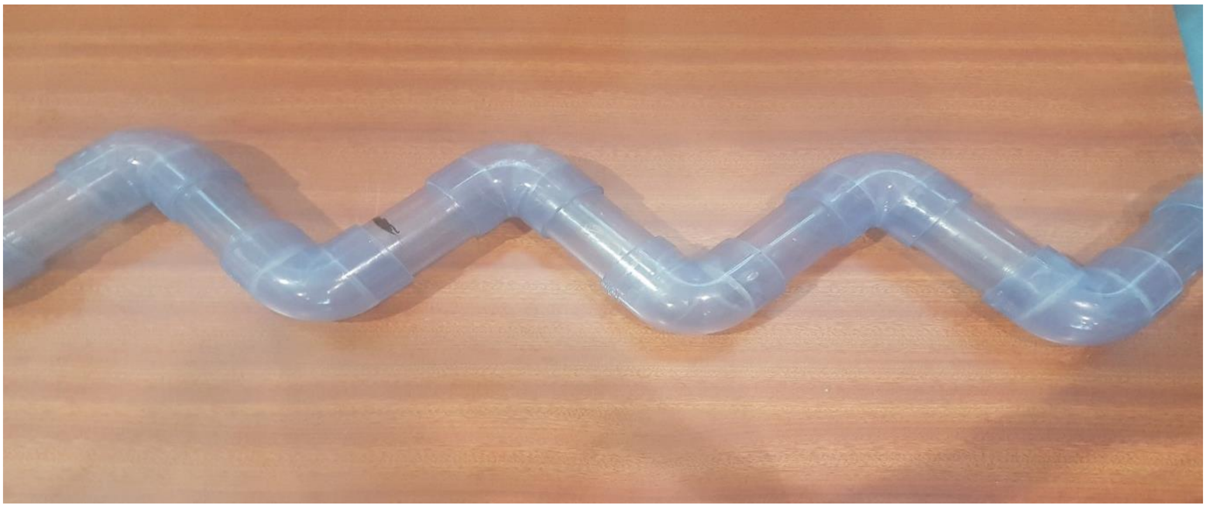
This chapter is systematized with several sections, beginning with a proof of the PST concept to mitigate unstable slug flow. A qualitative investigation of an unstable slug flow mitigation using the PST on the S-shape riser pipeline configuration is presented next. This section presents an experimentally determined flow regime map to establish the slug envelope outlining the slug mitigation potential of the PST. The slug envelope was developed when the PST was coupled with the S-shape riser and the S-shape riser without the PST using the riserbase pressure analysis and visual observation.

The third section presents an investigation of the PST in the attempt to establish the benefit of optimizing the system in terms of stability analysis. In section four, the slug mitigation potential of the PST is assessed quantitatively, quantifying the benefits of the PST using the riser outlet flow conditions and a conclusion remarks drawn in the last section.

5.2 Proof of concept

Wavy pipes as a passive slug mitigation technique was investigated by Xing et al., 2013. This demonstrated the concept that the PST has a heightening impact on the parameter variation technique and also offers a slug offsetting potential. PST pipe configurations (wavy and spiral) were installed at the upstream of the riserbase and were established to mitigate severe slugging (change the flow regime). Also Adedigba et al., 2007 studied the effects of a swirl pipe installed upstream of the riserbase on a 4 inch multiphase flow facility in Cranfield university. The hint of scheming a PST pipe was started from the belief that the gas - liquid stratified flow in the horizontal section of a riser system can be improved effectively by non-straight pipe sections. Studies on the flow features in a non-straight pipe sections have been shown. Deductions made from Adedigba et al, 2007 concluded that the swirl pipe induces mixing hence preventing stratified flow regime in a slightly downward inclined pipeline than the conventional straight horizontal pipe. This was because the gas was entrained in the liquid as the fluid mixture swirls. Again, stratified flow regime could not be re-established before the flow reaches the base of the riser hence reduced the possibility of severe slugging initiation in the system. There was also a reduction in the severity of the liquid blowdown and pressure oscillation. Finally there was an enlargement of the operating stable envelope by means of a reduction in the region of severe slugging comparatively. The same concept would be adopted for two Pseudo spiral tube (PST) pipe sections in this work but however at the topside of the riser. This helps eliminates the setbacks of installing equipment at the base of the riser which are not readily available and the cost associated with the installation. Again already existing facilities are faced with their inability of benefiting from this technique due to the cost implication.

The PST was designed such that several pipe fittings (standard elbows) are used to form a complete pipe section as shown in Figure 5-1. The schematic / setup of the PST was such that valves could be used to couple and decouple the PST pipe section to establish a connection to the 2 inch S-shape riser as shown in Figure 5-2. A detailed description of the PST and the experimental campaign will be presented next.



a) **Wavy pipe section**

b) **Spiral pipe section**

Figure 5-1 PST pipe section

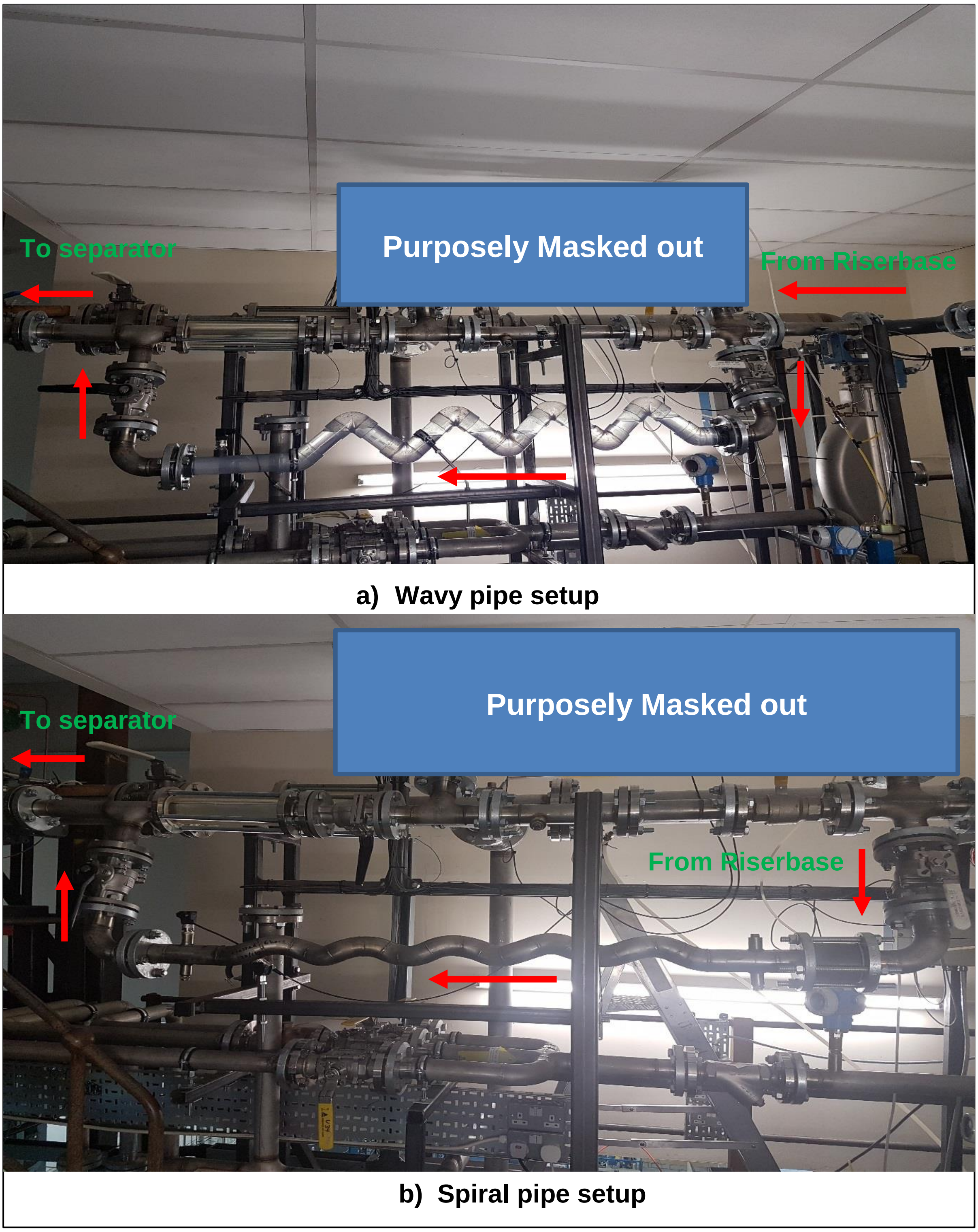
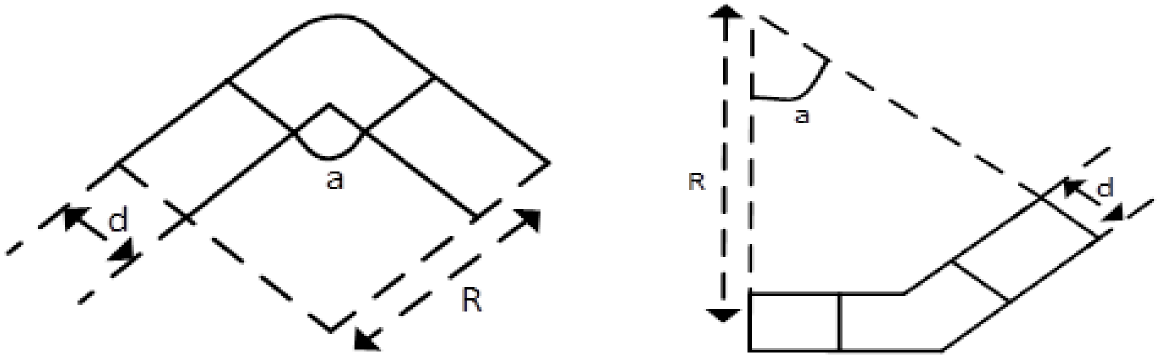


Figure 5-2 Schematic / setup of the PST on the S-shape riser

5.2.1 **PST pipe (wavy and spiral pipe) features**

For this study, the wavy and spiral pipe sections were made from standard plastic pipe bends connected in a single plane. As established in Xing et al, 2013 the least unit of the PST pipe, which gives the piping features is the piping bend which could be described by the pipe internal diameter, the bending radius and the angle of the bend as shown in Figure 5-3. Several pipeline geometries can be established from same pieces of piping bends by joining the bends in different configurations. A schematic of the piping features is represented in Figure 5-4.



a) 90 degree elbow **b) 45 degree elbow**

Figure 5-3 Piping features of the PST test piece

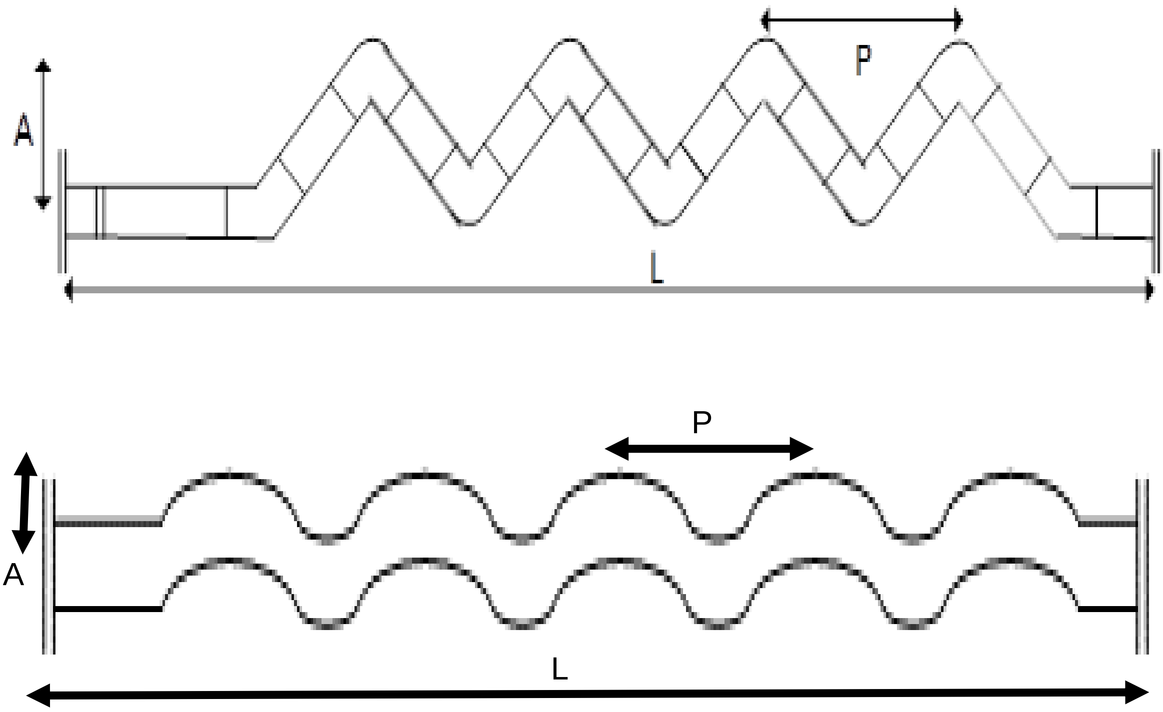


Figure 5-4 Piping features of the PST test piece

Where A is the amplitude thus the maximum space between the bend centreline and the centreline of the PST pipe, ‘P’ is the pitch; defined by the distance between consecutive dips or peaks and ‘L’ is the distance between the ends of the wavy pipe. From Xing et al, 2013, the geometrical parameter of the PST pipe is reliant on the bend angle, the ratio of the bend radius to pipe diameter and the number of bends.

The wavy pipe used for this experiment was constructed using two standard 45 degree elbow and seven short 90 degree radius bends. The 90 degree elbows are aligned adjacent to each other but joined together with a 70 mm straight transparent pipe piece. The variation in the angles of the bend (90 and 45 degree) was as a result of aligning the upstream and downstream ends of the wavy pipeline hence the 45 degree elbows at the ends. The total wavy pipe section was 1550 mm in length. The pipe fittings / elbows were made from standard clear PVC pipe. The geometrical parameter of the bends of the wavy pipe used for this experiment is presented in Table 5-1.

The spiral pipe section used for this experiment was made from 18 standard stainless steel 45 degree elbows. Each elbow was twisted clockwise at an angle of 30 degree to the adjacent one. The ends of the spiral pipe (upstream and downstream) was aligned parallel to each other. The spiral pipe piece used in this experiment was 1550 mm long. The welded spiral pipe is translucent with its geometrical parameters shown in Table 5-1.

The outlet of both PST (spiral and wavy pipes) was connected to the riser by a short horizontal pipe which is however equipped with a pressure transducer to aid the study. The slug mitigation potential of both the wavy pipe and the spiral pipe would be assessed experimentally on the 2 inch S-shape riser pipeline configuration next.

Table 5-1 PST geometrical parameters used in the experiment

	A, mm	P, mm	L, mm	d, mm	R, mm	a
45 degree bend	-	-	-	50.8	135.6	135^{0}
90 degree bend	-	-	-	50.8	115.4	90^{0}
Wavy pipe section	180	295	1550	50.8	-	-
Spiral pipe Section	36	230	1550	50.8	-	-

5.3 Experimental campaign using the PST on an S-shape riser system

Experimentally, several flow combinations were run on the 2 inch S-shape riser pipeline configuration with and without the PST pipe section. Same experimental matrix used in Chapter 4 for the flow dynamic behaviour on the S-shape riser pipelines was deployed in this study. Flow behaviour was identified and observed from the transparent section at the riser base of the S-shape riser as performed in Chapter 4. Predominantly slugging conditions are classified and defined in this study. The various observed flow behaviours will be presented next.

Furthermore representative slug flow conditions was investigated in terms of stability analysis for various mode of operations with and without choking and the results analysed. A conclusion was however drawn with a comparison between the system with the PST pipe section and without the PST pipe section.

5.3.1 Qualitative investigation of unstable slug flow mitigation of the PST on an S-shape riser

Qualitatively a wide range of flow combinations (gas - liquid flow) was run on the 2 inch S-shape riser pipeline configuration coupled with the PST pipe section (wavy and spiral pipe configuration). The resulting flow behaviour / dynamics observed by visual inspection and an analysis of the riserbase pressure was used to develop a flow regime map. The motivation for finding the flow regime map was to identify the slug envelope (region where slugging occurs) which in effect informs on how the PST pipe could reduce the slugging region. Thus the slug attenuation potential is assessed by the PST pipe's ability to reduce the unstable slug flow region in the S-shape riser. The resulting flow regime map from a combination of gas - liquid through the S-shape riser coupled with the PST pipe section (wavy and spiral pipe) will be presented next.

Table 5-2 Experimental Test Matrix

Component / Composition	Minimum Flowrate	Maximum Flowrate
Liquid, kg/s	0.1	5
Gas, Sm^3/h	5	300

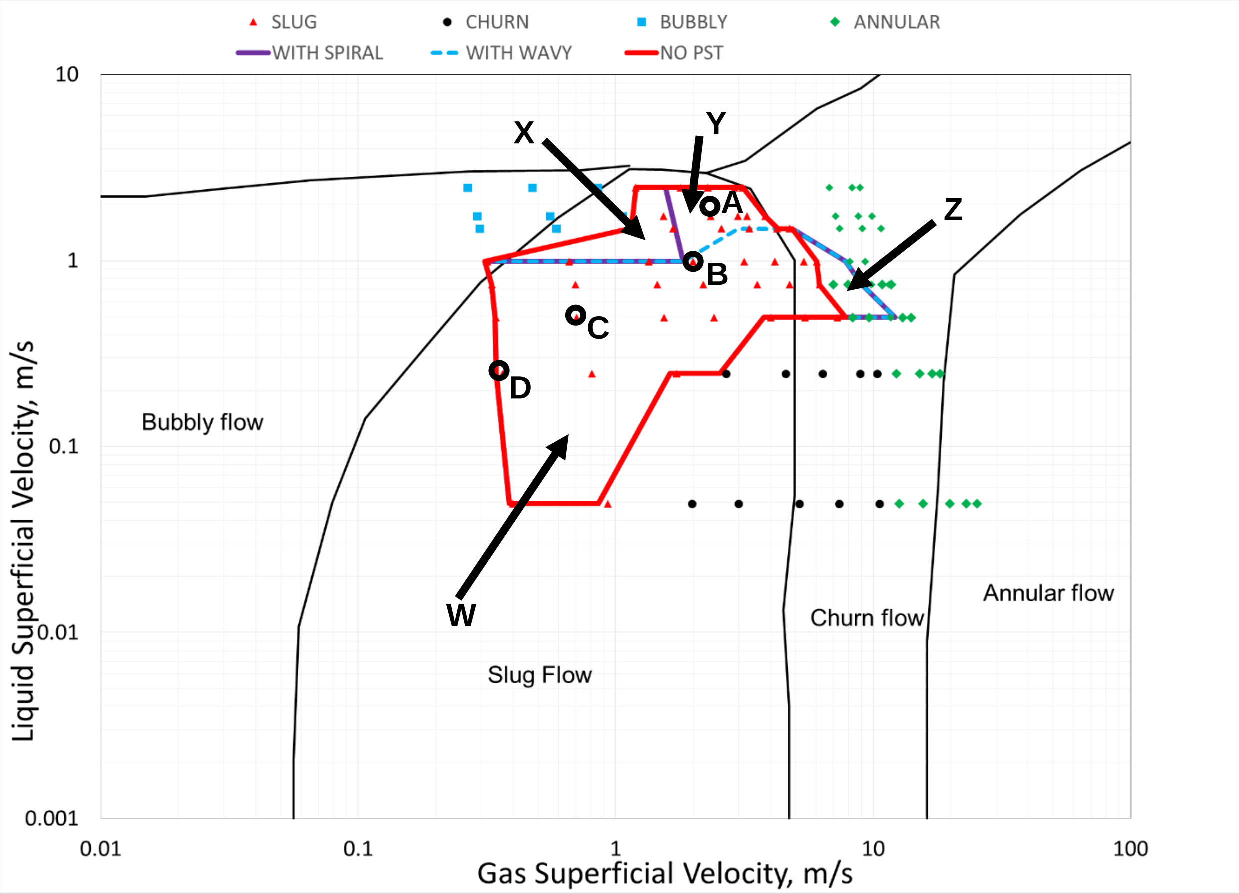


Figure 5-5 Experimental flow regime map for the S-shape riser with and without the PST pipe section

Figure 5-5 shows the flow regime map elaborating on the slug envelope obtained from the text matrix in Table 5-2 for the S-shape riser system coupled with and without the PST pipe sections. From Figure 5-5, the red markers enclosed by a red bordered line represents the unstable slug flow region observed in the plain S-shape riser (without any PST pipe section). The purple trace represents the slug envelope from the S-shape riser system coupled with a spiral pipe section while the blue dashed lines represents the slug envelope developed from the S-shape riser pipeline system coupled with the wavy pipe section. Four distinct flow behaviours were observed in the S-shape riser pipeline system. For low flowrates of both gas and liquid, slugging flow was observed, then increasing the gas flowrate with low liquid flowrate, churn flow regime was seen. A further increase in the gas flowrate puts the system into an annular flow region. In addition, for relatively high flowrate of liquid and low to medium flowrate of gas, a bubbly flow regime was detected.

At low to medium liquid flowrate and low flowrate of gas, the S-shape riser system exhibits slugging flow characteristics when coupled with or without the PST pipe sections. Increasing the gas flowrate while maintaining a relatively low liquid flowrate exhibits churn to annular flow through the S-shape riser system with and without the PST pipe sections. Annular flow regime is maintained for high liquid and high gas flowrate through the system with and without the PST pipe sections. For high liquid flowrate and relatively low gas flowrate through the system, a bubbly flow regime is observed through the system with and without the PST sections.

From Figure 5-5, the entire slug envelope of the S-shape riser system coupled with and without the PST sections is categorized into four (Section W, Section X, Section Y and Section Z) as shown in Figure 5-6. Section W describes flow conditions that exhibited slug flow characteristics through the S-shape riser coupled with and without the PST pipe section. Section X particularizes on the flow conditions that exhibited slugging flow behaviour through the S-shape riser system, however the slugging flow disappeared when either the spiral pipe section or the wavy pipe section was coupled to the system. Section Y elaborates on the flow conditions that exhibited slugging characteristics through the S-shape riser system but the slug flow disappeared when the system was coupled with the wavy pipe sections whiles Section Z describes flow conditions which did not exhibit slug characteristics through the S-shape riser system but however commenced to slug when the PST section (wavy or spiral pipe section) was coupled to the system.

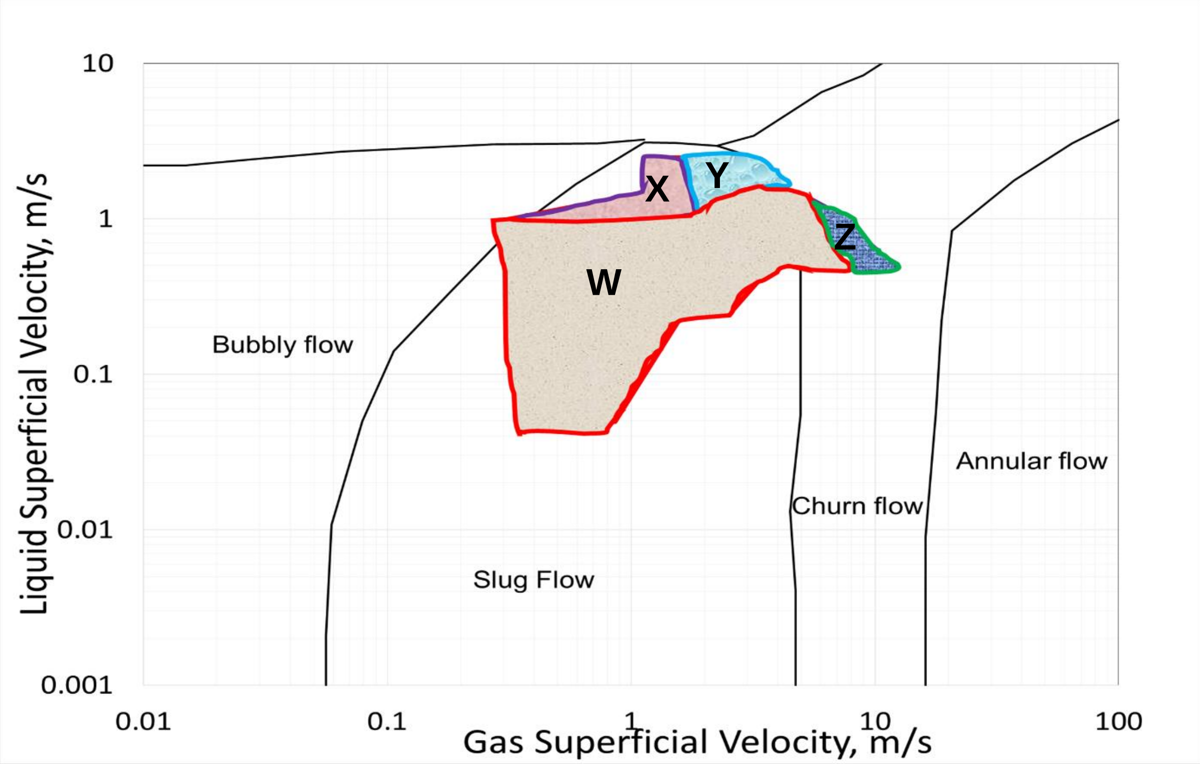


Figure 5-6 Experimental slug envelope for the S-shape riser with and without the PST pipe sections

From Figure 5-6, Section W represents a slug flow characteristic boundary conditions through the S-shape riser pipeline system. Thus, at low to medium liquid flowrate and low to medium gas flowrate, slugging flow was observed through the S-shape riser system. Similar flow behaviour was observed when the S-shape riser was coupled with the wavy pipe section as well as the spiral pipe section. This implies that for this set of flow conditions, the slugging flow characteristics was unchanged when the system was coupled with the PST pipe sections.

Again from Figure 5-6, Section X from the first glance shows the slug mitigation potential of the PST when installed at the topside of the S-shape pipeline riser system. From this region however, some flow conditions which exhibited slug flow characteristics through the S-shape riser become relatively stable when either the spiral or wavy pipe section was coupled to the system. Thus the spiral or wavy pipe section was able to tame the slugging flow characteristics hence the disappearance of the slugs. This action was observed for low gas flowrate and relatively higher liquid flowrate prior to the flow regime transitioning to bubbly flow.

Furthermore, section Y representing medium gas flowrate and relatively higher liquid flowrate shows slugging flow behaviour through the plain S-shape riser pipeline system. The slugging flow characteristics becomes resolute when the system was coupled with the spiral pipeline section but disappears when the spiral pipe section was replaced with a wavy pipe section. Thus, for this region (Y) from Figure 5-6, the wavy pipe section has the capability to mitigate or tame slug flow but the inverse occurs for the spiral pipe section.

Also, the slug flow region from the S-shape riser system marked Section Z has unique characteristics. From this region represented by relatively medium to high gas flowrate and medium liquid flowrate, annular flow regime was observed in the S-shape riser without the PST. However, instabilities sets in when the system was coupled with the spiral or wavy pipe sections. Thus, when the S-shape riser system is coupled with either the wavy or spiral pipe section, the flow exhibits a slugging flow behaviour. This implies that, there exist some non-slugging flow conditions on the S-shape riser which exhibits slugging flow when the S-shape riser is coupled with the wavy or spiral pipe section. This means that the wavy or spiral pipe section could potentially introduce instabilities into the system. The S-shape riser system coupled with the wavy pipe section and that coupled with the spiral pipe section has similar flow characteristics for these set of flow conditions. It could be deduced that the spiral or wavy pipe section is less effective for flow conditions with relatively high gas flowrate. However this discrepancies could be subject to the means of assessment (subjective) hence further investigations needed.

From Figure 5-6, for relatively high liquid flowrate and low to medium gas flowrate, categorized by Section X and Section Y, a slug flow regime was observed in the S-shape riser system. However from Figure 5-5, there exist some unstable slugging flow conditions on the S-shape riser that do not follow suite for the system coupled with the PST (wavy or spiral pipe) section. Thus the system showed significant improvement when coupled with the wavy or spiral pipe section. Generally, this shows that the wavy pipe or spiral pipe has some effect on the flow behaviour when tied with the system. Thus some unstable slug flow region could be rendered stable when the wavy or spiral pipe is introduced into the system. Comparatively, the system coupled with the wavy pipe section has greater benefits than that coupled with the spiral pipe section. This is evident in the shrink of the slug envelope (blue trace from Figure 5-5) associated with the S-shape riser system coupled with the wavy pipe section. Conclusively, the PST section's slug mitigation potential is more effective for systems with higher liquid flowrate, however the wavy pipe has greater benefits relative than the spiral pipe section. The extent to which the wavy pipe or spiral pipe could improve the flow behaviour will be assessed next using some flow conditions exhibiting unstable slug flow in the system coupled with the PST (wavy pipe and spiral pipe).

5.3.2 Unstable slug flow stability response to the PST (wavy pipe section and spiral pipe section)

From Figure 5-5, several flow conditions within the various slug envelopes marked **A, B, C** and **D** were studied in terms of stability analysis. The boundary and parameter condition of points **A, B, C** and **D** is shown in Table 5-3.

Table 5-3 Slugging flow conditions / points for stability analysis

Point Label → / **Boundary Condition** ↓	**A**	**B**	**C**	**D**
Liquid mass flowrate, kg/s	3.00	2.00	1.00	0.50
Superficial velocity of liquid, m/s	1.48	0.99	0.71	0.25
Gas volumetric flowrate, Sm^3/h	50.00	40.00	10.00	7.00
Superficial velocity of gas, m/s	2.59	2.55	0.49	0.36

The capability of the wavy pipe or spiral pipe section to enhance choking thus improving system stability was assessed. A parameter variation technique (varying the topside valve) was performed for these flow conditions to identify the critical bifurcation point (valve opening) beyond which the system could potentially lose its stability. Prior to this assessment, the riserbase pressure trends which many researchers have used in describing the flow behaviour was analysed to provide an informed behavioural outlook of the system with and without the PST section.

5.3.2.1 Pressure trends with and without PST section

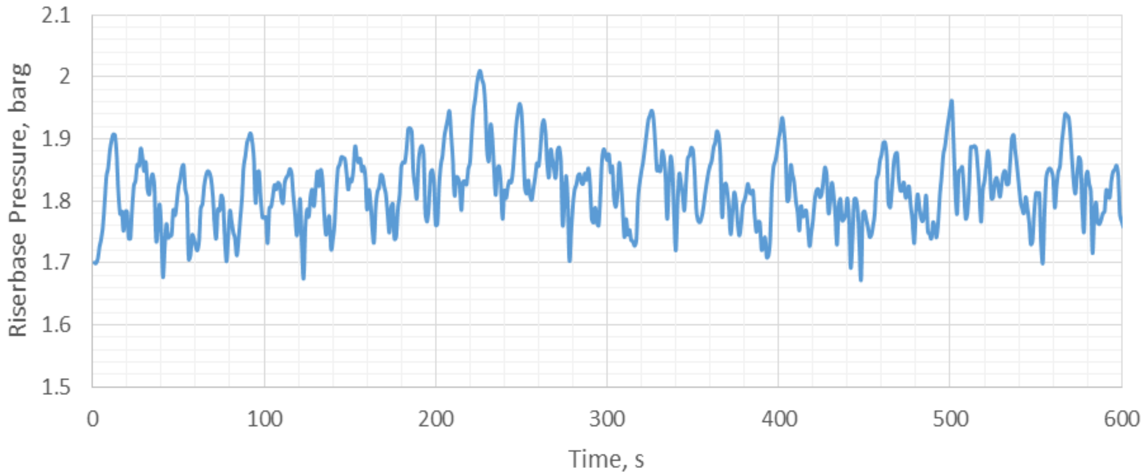
This section seeks to examine the degree of the flow / pressure fluctuations in the S-shape riser system with and without the PST pipe. Significantly this gives a clearer understanding of the impact the PST has on the system flow dynamics on a whole examining the riserbase pressure. From the slug flow region produced (Figure 5-5), two flow conditions (label **C** and label **D**) from Table 5-3 were examined.

Figure 5-7 shows the pressure trend for the flow condition labelled **C** with and without the PST. The oscillating and fluctuating pressure observed shows that the system clearly has a slugging flow characteristic.

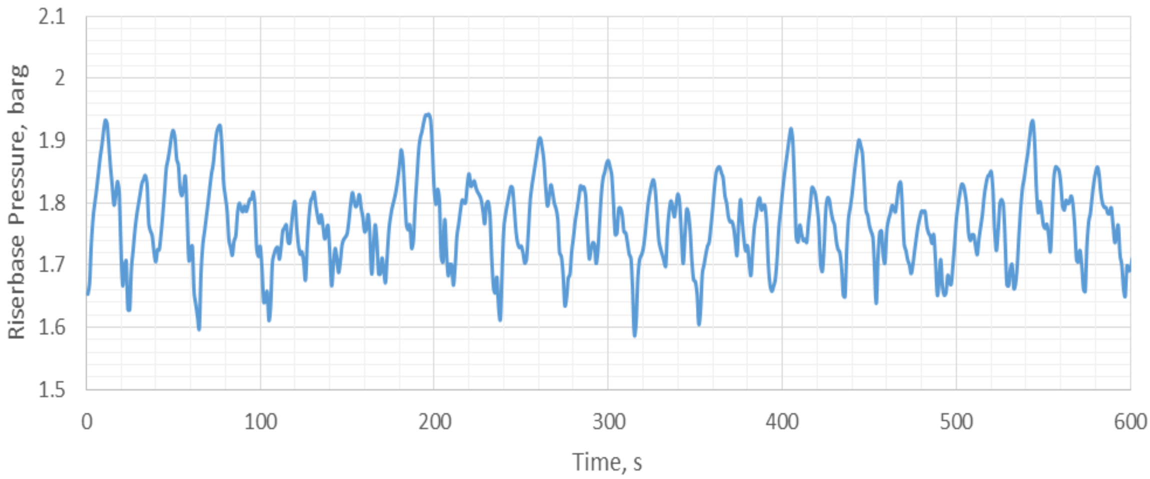
From Figure 5-7 (a) it can be observed that with the wavy pipe coupled, the system experiences relatively lower riserbase oscillation magnitude than the system without any PST section (Figure 5-7 (b)). Again, the unstable slug length / slug cycle shortens comparatively on the system with the wavy pipe but however with a higher riserbase pressure magnitude. This is shown by the standard deviation and mean riserbase pressure shown in Table 5-4 which represents a reduced slug oscillation magnitude and increased pressure loss through the riser respectively.

Similarly, Figure 5-7 (c) shows the pressure trend obtained for the flow condition through the system coupled with the spiral pipe section. The system coupled with the spiral pipe section shows a relatively shorter slug length compared to the system without the PST section. There is also a slight difference in the riserbase pressure magnitude (Figure 5-7 (b) and Figure 5-7 (c)) in the systems. However comparing the system with and without the spiral pipe, the riserbase pressure trends shows not much significant difference. However there is a reduced slug oscillation magnitude when the system is coupled with the spiral pipe section but at an increased pressure drop cost.

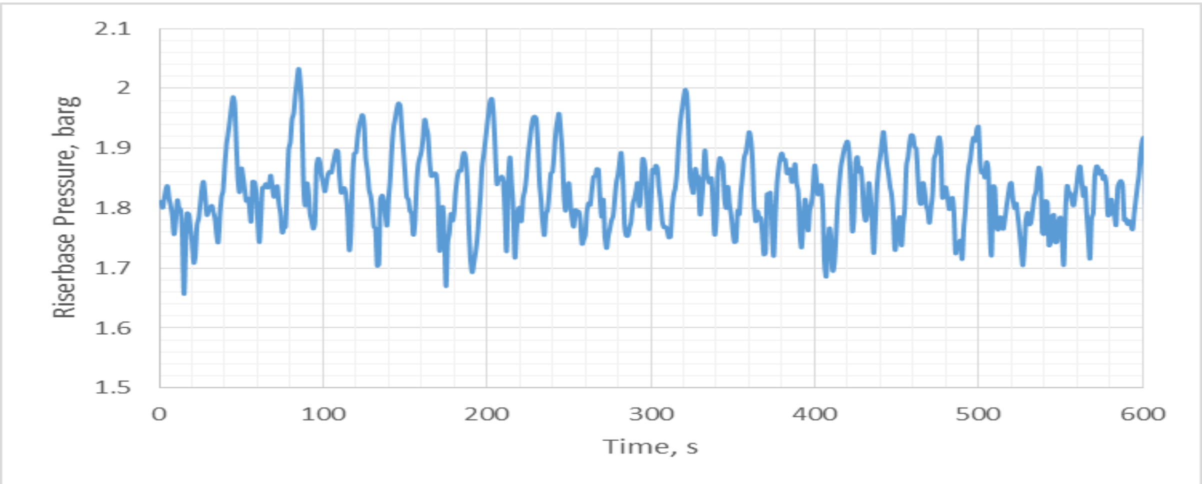
From Table 5-4, the wavy pipe section shows an improvement in the system behaviour in terms of the slug oscillation magnitude over the spiral pipe section. Again, there was an observed reduced pressure loss in the system for the system coupled with the wavy relative to the system coupled with the spiral pipe section.



a. **With wavy pipe**



b. **Without PST**



c. **With spiral pipe**

Figure 5-7 Riserbase pressure trend for 1 kg/s liquid and 10 Sm^3/h gas on the S-shape riser

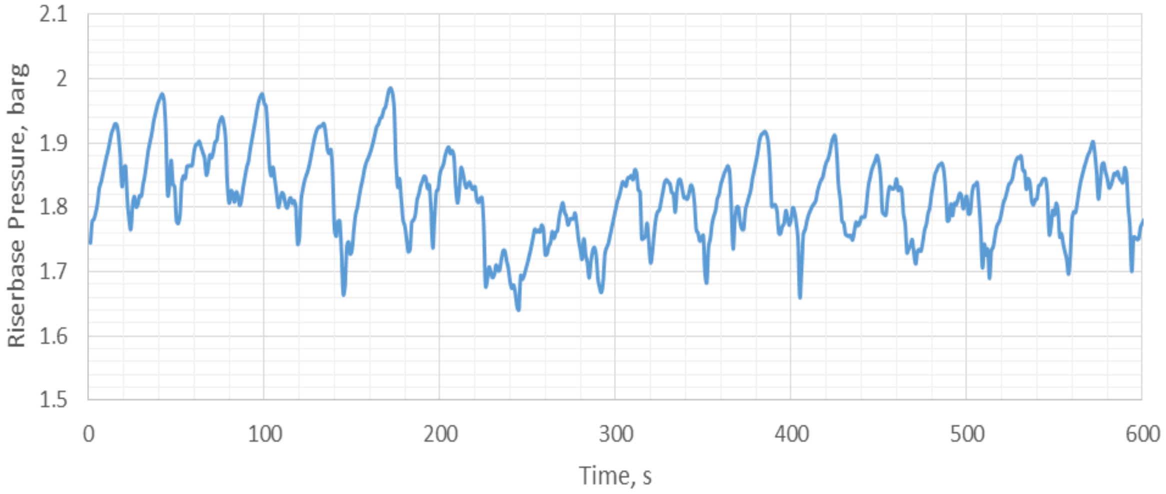
Similar to label **C**, label **D** from Table 5-3 was assessed. The resultant pressure trend is shown in the Figure 5-8. The pressure trend observed in Figure 5-8 have similar behaviour and characteristics when examined with and without the PST compared to the flow condition labelled **C**.

The riserbase pressure trend seen in Figure 5-8 represents a flow with slugging characteristics. Comparatively, the system coupled a spiral pipe section (Figure 5-8 (c)) and without the spiral pipe section (Figure 5-8 (b)) shows very similar trend except that the system coupled with the spiral pipe has slightly higher slug magnitude which is evident in the standard deviation from Table 5-4. Similarly, the system coupled with the wavy pipe (Figure 5-8 (a)) has a lower slug oscillation magnitude relative to that without any PST (Figure 5-8 (b)).

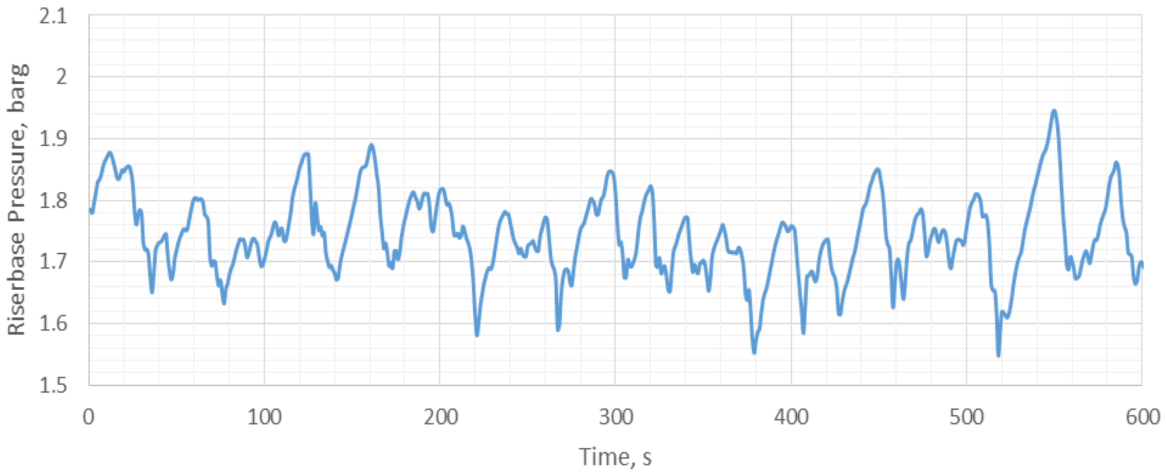
Conclusively, the PST sections (wavy and spiral) could help reduced the slug oscillation magnitude when coupled with the system at the topside of the pipeline riser system but at an increased pressure loss as elaborated in Table 5-4. The stability analysis of these two flow conditions using parameter variation technique (topside valve manual choking) would be assessed next to investigate the PST's ability to aid this method.

Table 5-4 Pressure trend result summary with and without the PST

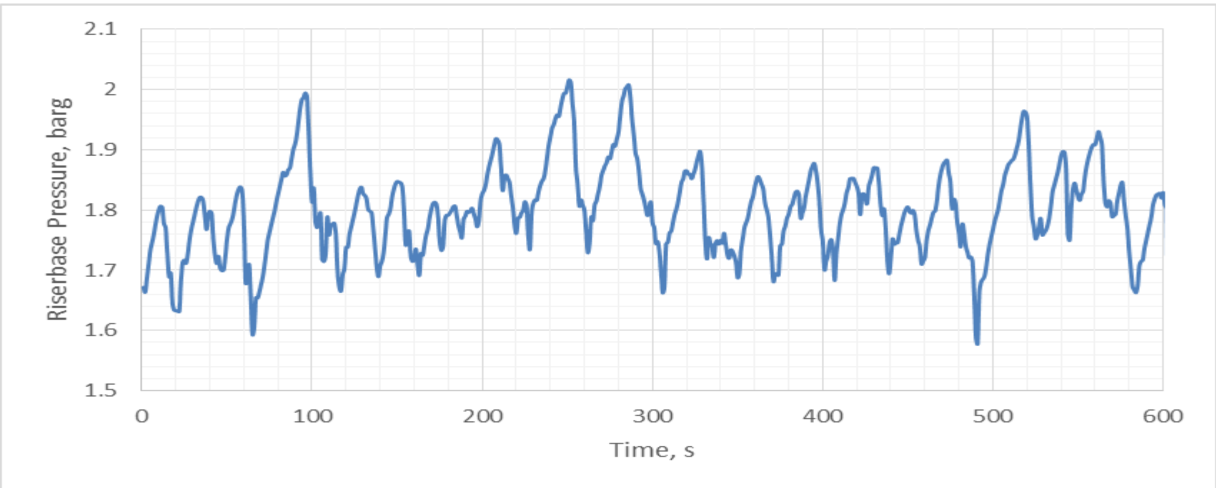
Riser configuration	Mean riserbase pressure	Standard deviation
1 kg/s Liquid and 10 Sm^3/h Gas		
Spiral pipe section	1.8281	0.0626
No PST	1.6984	0.0715
Wavy pipe section	1.7513	0.0593
0.5 kg/s Liquid and 7 Sm^3/h Gas		
Spiral pipe section	1.8006	0.0665
No PST	1.6670	0.0681
Wavy pipe section	1.7456	0.0670



a. **With wavy pipe**



b. **Without PST**



c. **With spiral pipe**

Figure 5-8 Riserbase pressure trend for 0.5 kg/s liquid and 7 Sm^3/h gas on the S-shape riser

5.3.3 Parameter variation technique using mass driven source with the PST (wavy and spiral pipe section)

Using a fixed mass source, the four boundary conditions from Table 5-3 with labels **A, B, C** and **D** were assessed using a parameter variation technique. For the benefit of this study, the S-shape riser coupled with the wavy pipe section is referred to Case 1 while the S-shape riser configuration without any PST is represented as Case 2 and the S-shape riser coupled with the spiral pipe section is Case 3. The stability analysis of these points (**A, B, C** and **D**) will be discussed next.

5.3.3.1 Label A - 3 kg/s liquid and 50 Sm^3/h gas flow

3 kg/s liquid and 50 Sm^3/h gas flowrate corresponding to superficial liquid and gas velocities of 1.48 m/s and 2.59 m/s respectively (label **A** from Table 5-3) through the S-shape riser falls within the section **Y** from Figure 5-6. This implies that there exist a relatively stable flow for the system coupled with the wavy pipe section but unstable slug flow pattern observed for the system coupled with the spiral pipe section and the system without any PST section. Thus for this flow condition, a parameter variation technique would not be necessary for the system coupled with the wavy pipe section since it already falls out of the slug envelope for the system coupled with a wavy pipe section.

Figure 5-9 and Figure 5-10 show the resulting bifurcation maps from the system without any PST section and the system coupled with the spiral pipe section respectively for 3 kg/s liquid and 50 Sm^3/h gas flow using riserbase pressure 1 (base of the riser) of the S-shape riser system. The bifurcation maps are essentially used to determine the critical bifurcation point.

From Figure 5-9 and Figure 5-10 a critical bifurcation point, which reflects the maximum valve opening beyond which the system loses its stability, of 35 % and 36 % valve opening was obtained respectively.

The critical valve opening of 35 % corresponding to a riserbase pressure of 3.8 barg was observed for the S-shape riser system without any PST pipe section as shown in Figure 5-9. Thus, the system achieves stability at a 35 % choke valve opening and beyond this valve opening, there was an initiation of instabilities in the system. This implies that for valve openings greater than 35 %, the S-shape riser system exhibits an unstable flow behaviour.

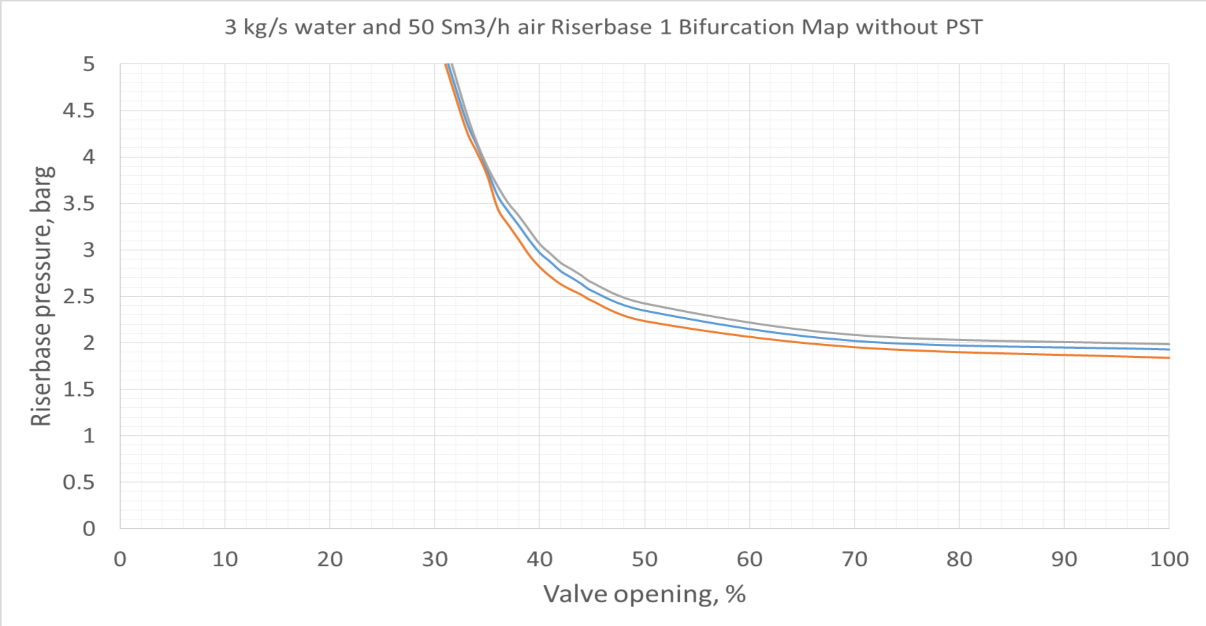


Figure 5-9 Bifurcation map 3 kg/s liquid and 50 Sm^3/h gas on the S-shape riser without any PST pipe section (Case 2) using riserbase pressure 1

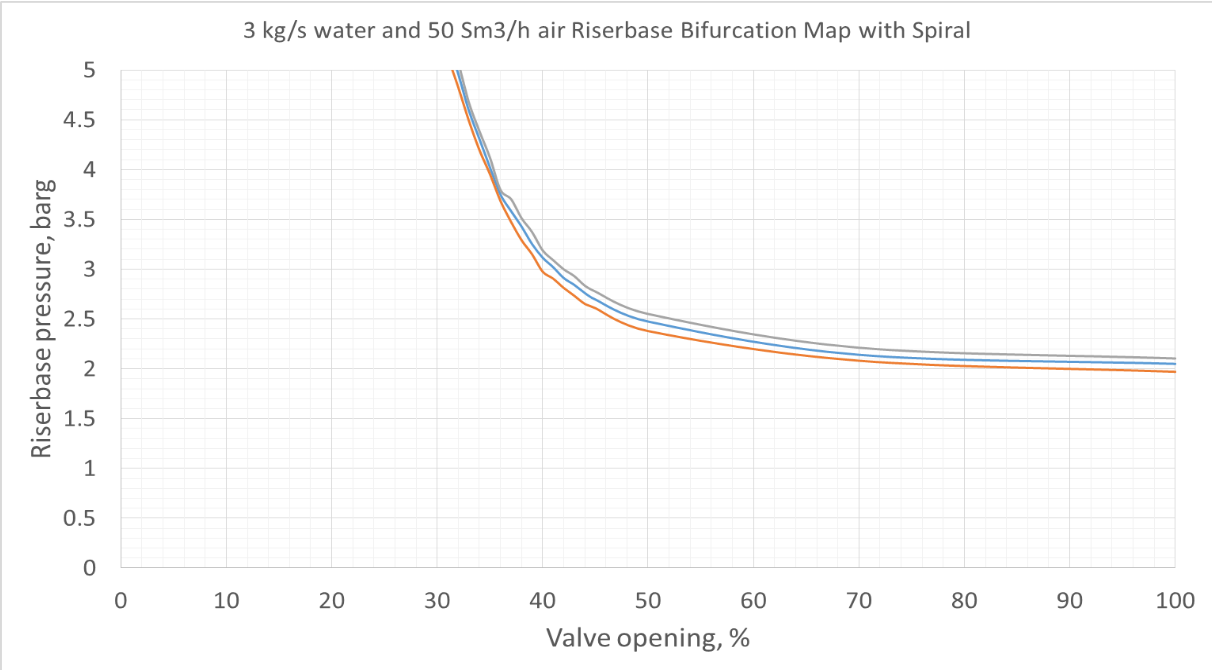


Figure 5-10 Bifurcation map 3 kg/s liquid and 50 Sm^3/h gas on the S-shape riser with the spiral pipe section (Case 3) using riserbase pressure 1

Similarly, for the S-shape riser system coupled with the spiral pipe section (Case 3), a critical bifurcation point of 36 % valve opening corresponding to a riserbase pressure of 3.7 barg was observed as shown in Figure 5-10. For valve openings equal to or less than 36 %, the system behaves in a stable manner. However, for valve openings greater than 36 %, flow instabilities sets in. Thus, the system losses its stability for valve openings greater than 36 %.

Comparatively from Figure 5-9 and Figure 5-10, it could be deduced that when the S-shape riser was coupled with the spiral pipe section, the stable valve opening increased by 1 % resulting to a reduced riserbase pressure. The increased stable valve opening translates to an increase in the system throughput hence increase in oil production. Again, there is a 2.6 % reduction in the system riserbase pressure which also translates to safer production. Conclusively, the spiral pipe section could potentially aid manual choking for some unstable slugging flow conditions when coupled at the topside of the riser pipeline system.

5.3.3.2 Label B - 2 kg/s Liquid and 40 Sm^3/h gas flow

The stability of 2 kg/s liquid and 40 Sm^3/h gas flow condition (label **B** from Table 5-3) through the S-shape riser pipeline with different configurations (the system coupled with the wavy pipe section, the system without any PST section and the system coupled with the spiral pipe section) was assessed. Flow label **B** falls within the slugging flow region **W** from Figure 5-6. This flow region however represents flow conditions that exhibited unstable slug flow in the S-shape riser pipeline system for all the system configurations considered (the system coupled with the wavy pipe section, the system without any PST section and the system coupled with the spiral pipe section). Thus, regardless of the pipeline configuration considered, the system maintained an unstable flow behaviour.

Figure 5-11, Figure 5-12 and Figure 5-13 shows the resulting bifurcation maps of the above mentioned flowrate for the S-shape pipeline configuration coupled with the wavy pipe section, the S-shape pipeline configuration without any PST and the S-shape pipeline configuration coupled with the spiral pipe section respectively.

Figure 5-11 represents the derived riserbase bifurcation map from the S-shape riser coupled with the wavy pipe section (Case 1). From Figure 5-11, a critical bifurcation point of 33 %valve opening was attained corresponding to 3.1 barg pressure at the base of the riser. Thus, for valve opening equal to or less than 33 % on this system, there was a stable flow behaviour observed whiles valve openings greater than 33 % showed an unstable flow behaviour which is evident in the pressure fluctuation or oscillation.

Also Figure 5-12 shows the riserbase bifurcation map from the S-shape riser system without any PST (Case 2) for the inlet flow condition 2 kg/s liquid and 40 Sm^3/h gas. 29 % valve opening corresponding to a riserbase pressure of 3.75 barg was attained as the critical bifurcation point. This implies that for valve openings greater than 29 % on the system configuration in Case 2, there exist an observed pressure fluctuation hence flow becomes unstable. On the other hand the system maintains its stable for valve openings less than or equal to 29 %.

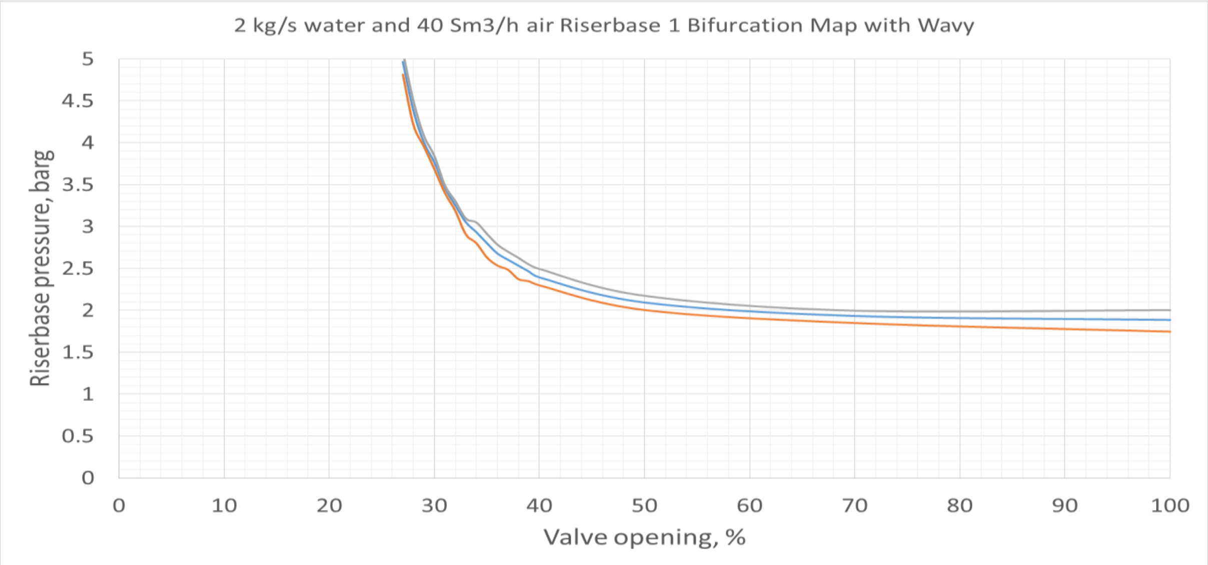


Figure 5-11 Bifurcation map 2 kg/s liquid and 40 Sm^3/h gas on the S-shape riser with the wavy pipe section (Case 1) using riserbase pressure 1

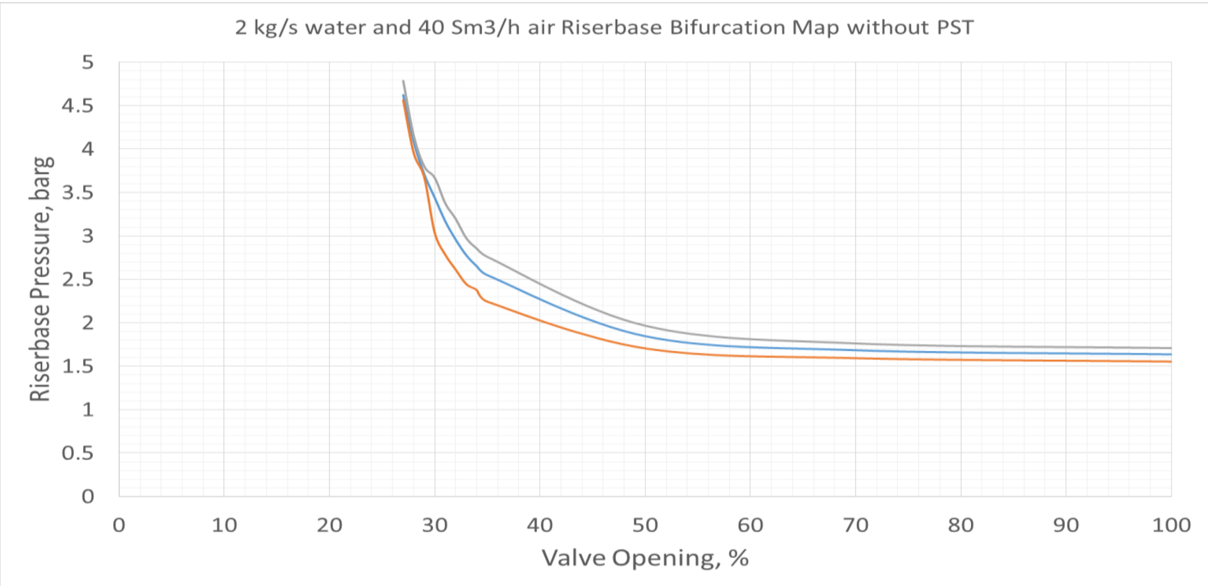


Figure 5-12 Bifurcation map 2 kg/s liquid and 40 Sm^3/h gas on the S-shape riser without any PST pipe section (Case 2) using riserbase pressure 1

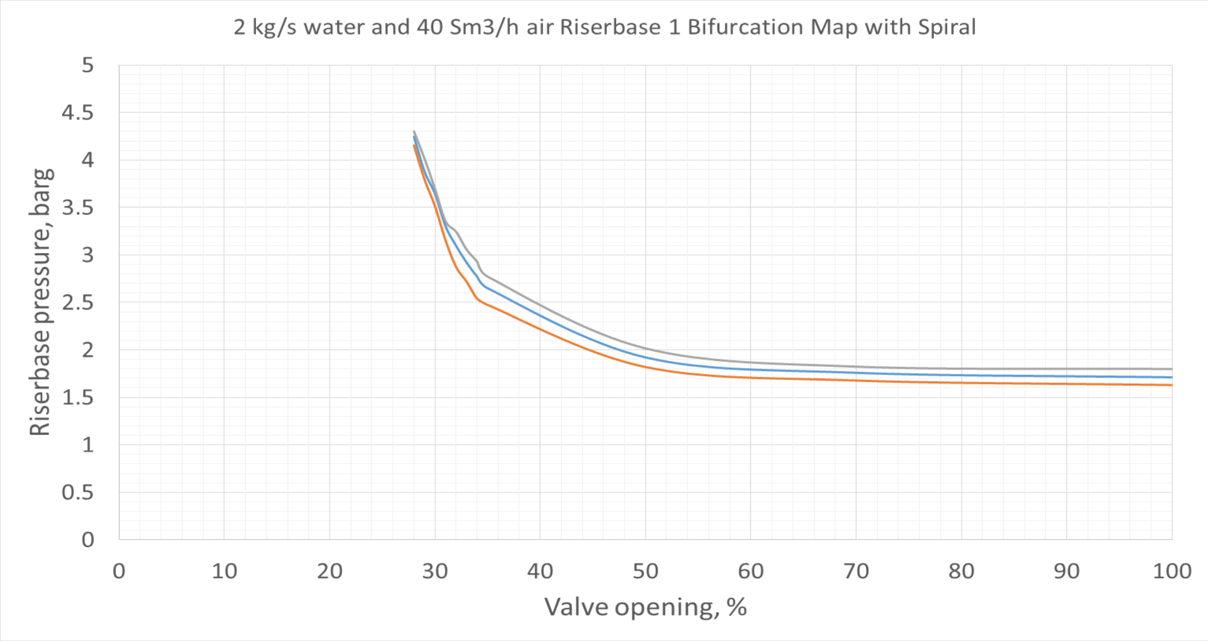


Figure 5-13 Bifurcation map 2 kg/s liquid and 40 Sm^3/h gas on the S-shape riser with the spiral pipe section (Case 3) using riserbase pressure 1

Again Figure 5-13 symbolises the riserbase bifurcation map from the S-shape riser system coupled with a spiral pipe section (Case 3) for the inlet flow condition 2 kg/s liquid and 40 Sm^3/h gas. From Figure 5-13, a critical bifurcation point of 31 % choke valve opening corresponding to a 3.3 barg riserbase pressure was achieved. Similarly, for the configuration in Case 3, the system exhibits a stable flow behaviour for valve openings equal to or less than 31 % which is evident in the relatively steady pressure trend. However, the system losses its stability when the choke valve opening exceeds 31 %.

Comparatively, for the same inlet flow condition on the different system configurations (the S-shape pipeline configuration coupled with the wavy pipe section, the S-shape pipeline configuration without any PST and the S-shape pipeline configuration coupled with the spiral pipe section), different critical bifurcation points corresponding to different riserbase pressures was achieved. Thus, a critical bifurcation point of 33 %, 29 % and 31 % was attained for the S-shape pipeline configuration coupled with the wavy pipe section, the S-shape pipeline configuration without any PST and the S-shape pipeline configuration coupled with the spiral pipe section respectively. From Figure 5-11, Figure 5-12 and Figure 5-13, it is observed that when the S-shape riser system was coupled with any of the PST pipe sections, there was an increase in the critical stable valve opening. Thus, the system coupled with the wavy pipe section (Case 1) achieves stability compared to the S-shape pipeline configuration without any PST at an additional 4 % valve opening and a lower riserbase pressure. Similarly, the system coupled with the spiral pipe section (Case 3) likened to the S-shape pipeline configuration without any PST achieves stability at a 2 % increase choke valve opening and again with lower riserbase pressure. The increased percentage stable valve opening obtained from the other system configurations (the S-shape pipeline configuration coupled with the wavy pipe section and the S-shape pipeline configuration coupled with the spiral pipe section) translates to an increase in oil production as a result of installing the PST at the topside of the riser. However, for the flow condition assessed the wavy pipe section showed greater benefit comparatively to that achieved with the spiral pipe section.

Similarly, label **C** and label **D** from Table 5-3 which falls within the slugging flow region **W** from Figure 5-6.were investigated. The results from the stability analysis of label **C** and label **D** are shown in Table 5-5.

In reality however, there is no such instance as constant mass source since all real life feed sources are driven by pressure. To examine and extend the benefit of the wavy and spiral pipe section in real life, a pressure driven is deployed to imitate real life scenarios which would be discussed next.

Table 5-5 Result summary for mass driven source stability analysis with and without PST

BOUNDARY LABEL	CONFIGURATION	CRITICAL BIFURCATION POINT, %	RISERBASE PRESSURE, barg
A	With Wavy	100	2.23
	Without PST (Plain)	35	3.8
	With Spiral	36	3.7
B	With Wavy	33	3.1
	Without PST (Plain)	29	3.75
	With Spiral	31	3.3
C	With Wavy	23	3.7
	Without PST (Plain)	22	3.8
	With Spiral	21	3.9
D	With Wavy	20	2.7
	Without PST (Plain)	19	2.8
	With Spiral	19	3.0

5.3.4 Parameter variation technique using pressure driven source with the PST (wavy and spiral pipe sections)

The S-shape riser system with and without the PST was run with a pressure driven inlet to investigate the benefit that could be derived. To enhance experiments several researchers have switched to pressure controlled systems due to their unique performance. Pressure driven sources provides the benefit of possibly controlling fluids in dead –end channels, high stability and even pulseless flow.

For the three phase facility in the Cranfield University to run as a pressure driven source (inlet), the inlet valve positions were fixed manually to an opening depending on the required flowrate. The water pump was set to a fixed frequency drive whiles the gas flow into the system was driven by the gas receiver pressure. Two pressure sources were investigated to assess the benefit of the PST (wavy and spiral pipe section) on an S-shape riser system. A riserbase bifurcation map was produced at a fixed pump drive frequency of 40 Hz. The liquid and gas inlet valves were manually fixed at an opening of 22.6 % and 6 % respectively to analyse the benefit derived when the PST was coupled to the S-shape riser.

Figure 5-14, Figure 5-15 and Figure 5-16 are the riserbase pressure bifurcation maps for the S-shape riser system coupled with the wavy pipe section (Case 1), the S-shape riser system without any PST section (Case 2) and the S-shape riser system coupled with the spiral pipe section (Case 3) respectively equipped with a pressure driven source. With the water pump drive fixed at a frequency of 40 Hz, the choke valve opening was varied and the riserbase pressure recorded to produce the pressure bifurcation map.

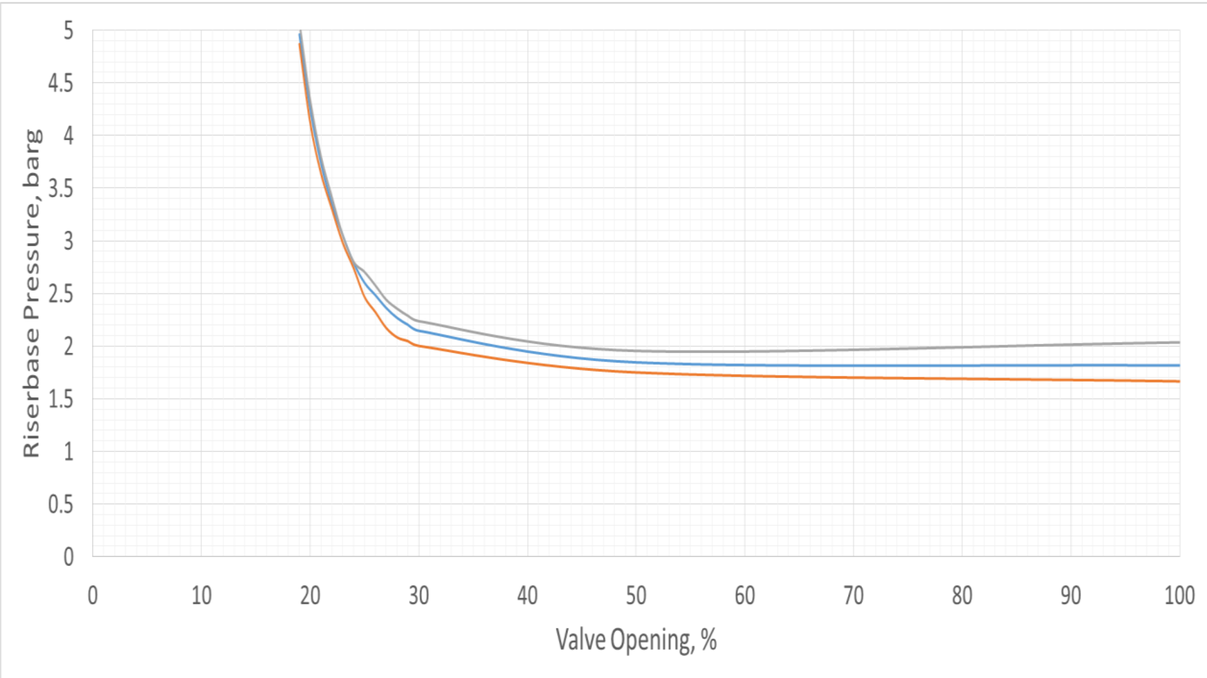


Figure 5-14 Pressure driven bifurcation map from the S-shape riser with the wavy pipe section (Case 1) using riserbase pressure 1 (40 Hz pump drive)

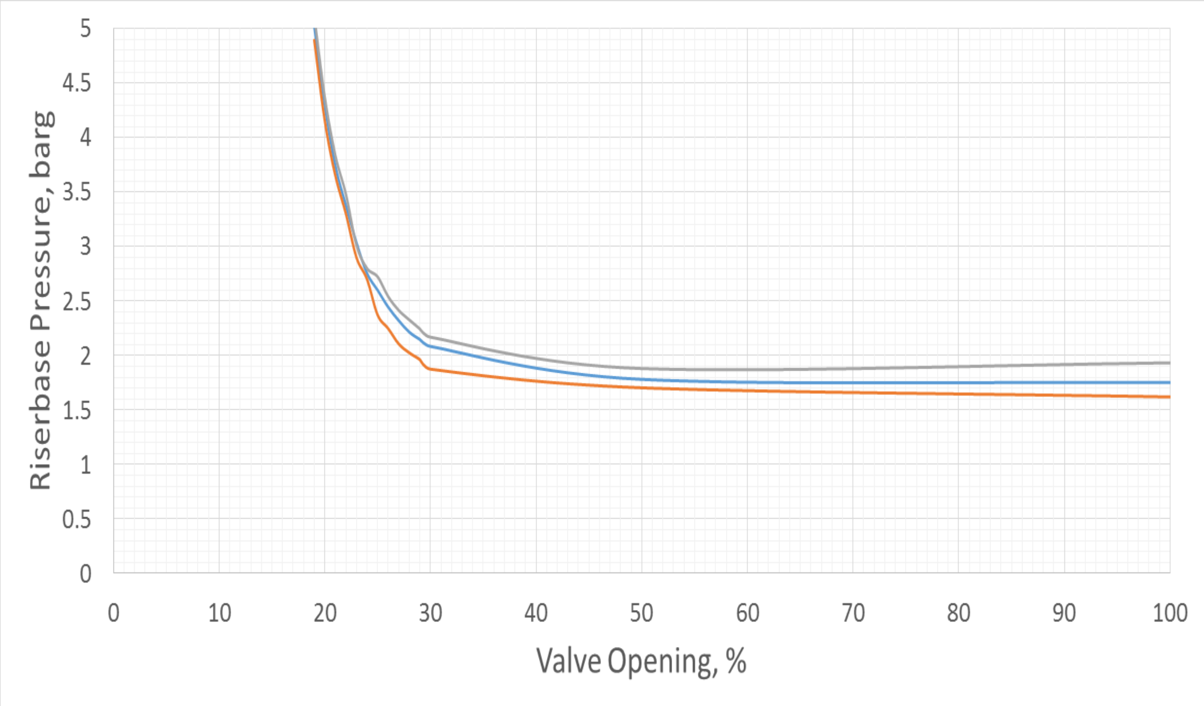


Figure 5-15 Pressure driven bifurcation map from the S-shape riser without the PST section (Case 2) using riserbase pressure 1 (40 Hz pump drive)

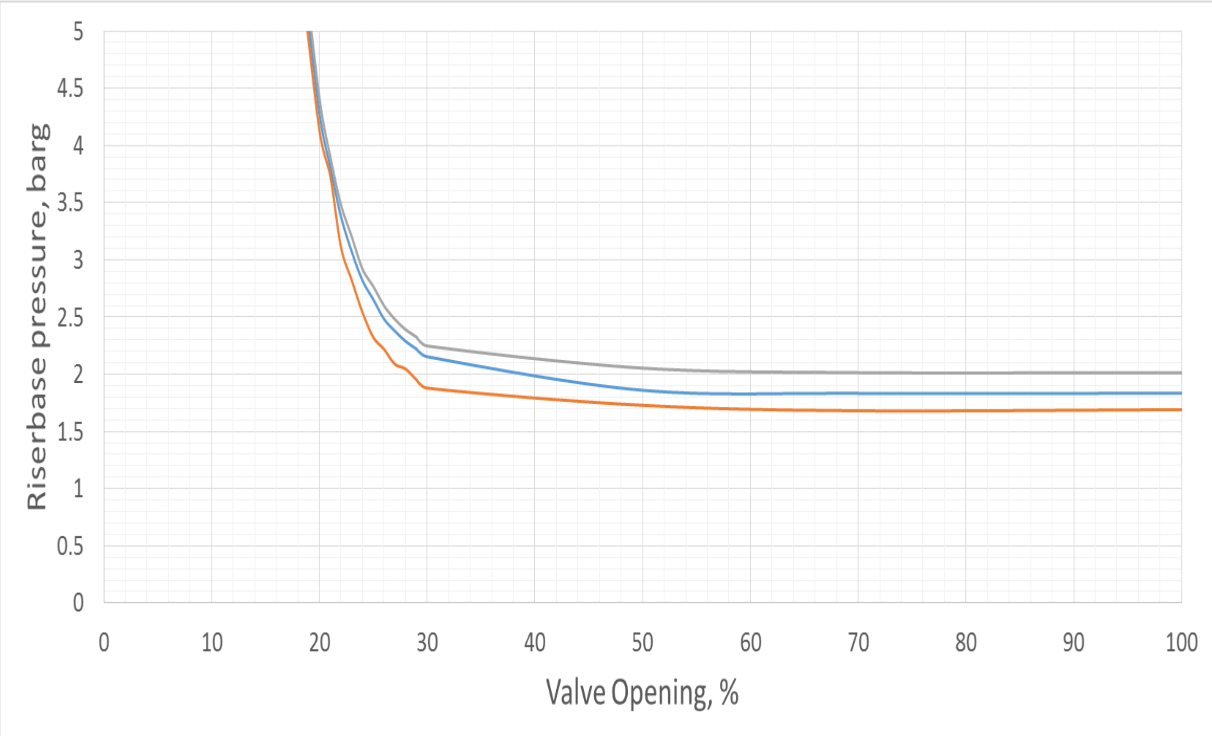


Figure 5-16 Pressure driven bifurcation map from the S-shape riser with the spiral pipe section (Case 3) using riserbase pressure 1 (40 Hz pump drive)

From Figure 5-14, the S-shape pipeline configuration coupled with the wavy pipe section (Case1) was observed to have its critical bifurcation point at a valve opening of 24 % corresponding to a riserbase pressure of 2.7 barg. At the same riserbase pressure on the S-shape pipeline configuration without any PST (Case 2), a corresponding valve opening of 24 % was obtained. For this valve opening however, there exist an unstable flow fluctuation in the system since the critical bifurcation point for the S-shape pipeline configuration without any PST (Case 2) was 23 % in open loop. Thus, for the system configuration in Case 2 to be stable, the valve opening has to be reduced a further 1 % opening to render the system stable which in effect increase the riserbase pressure.

From Figure 5-15, the S-shape pipeline configuration without any PST has a bifurcation point of 22 % valve opening corresponding to a riserbase pressure of 3.1 barg. Again, the bifurcation map produced for the S-shape pipeline configuration coupled with the spiral pipe section shown in Figure 5-16, has a critical bifurcation point of 22 % valve opening corresponding to a riserbase pressure of 3.5 barg. Comparatively, the critical stable bifurcation point for the S-shape pipeline configuration without any PST reduces when the spiral pipe section was coupled with the S-shape riser system (Case 3).

Transposing the critical stable riserbase pressure from the S-shape pipeline configuration without any PST unto the bifurcation map obtained for the S-shape pipeline configuration coupled with the spiral pipe section (Case 3), falls within an unstable region. Thus, the system coupled with the spiral pipe section observes fluctuations in the riserbase pressure at 3.1 barg which corresponds to a valve opening of 24 %.

Consequently, a much lower flowrate, thus, running the pump at a fixed frequency of 30 Hz with a fixed valve positions of 24 % and 5.9 % for both the water and gas inlet line respectively was investigated. A stability analysis performed on the system with this flow condition is summarized and presented in Table 5-6.

Table 5-6 **Result summary for pressure driven source stability analysis with and without PST**

PUMP DRIVE	CONFIGURATION	CRITICAL BIFURCATION POINT, %	RISERBASE PRESSURE, barg
40 Hz	With Wavy	24	2.7
	Without PST (Plain)	23	3.1
	With Spiral	22	3.5
30 Hz	With Wavy	23	2.9
	Without PST (Plain)	22	3.1
	With Spiral	20	3.5

5.3.5 PST effect on riser outlet conditions (topside equipment) for unstable slug flow

Quantitatively, several researchers have proposed the use of some dimensionless components to describe the benefits derived from some passive slug mitigation techniques. This illustrates the gains achieved in terms of pressure or even productivity, however these index does not give a true and elaborate representation of the actual system. In this section, the true behaviour of the system would be assessed at the outlet of the system with and without the PST section. Again, the liquid interface level of the separator at the outlet of the riser and the riserbase pressure trend were deployed for this study.

Figure 5-17 shows the riserbase pressure and separator liquid level trend for the flow condition 1 kg/s liquid and 10 Sm^3/h gas through the S-shape riser coupled with and without the wavy pipe section. From Figure 5-17, the system operates without the wavy pipe from 0 seconds to about 2900 seconds. Beyond 2900 seconds, the system was allowed to operate with the wavy pipe coupled.

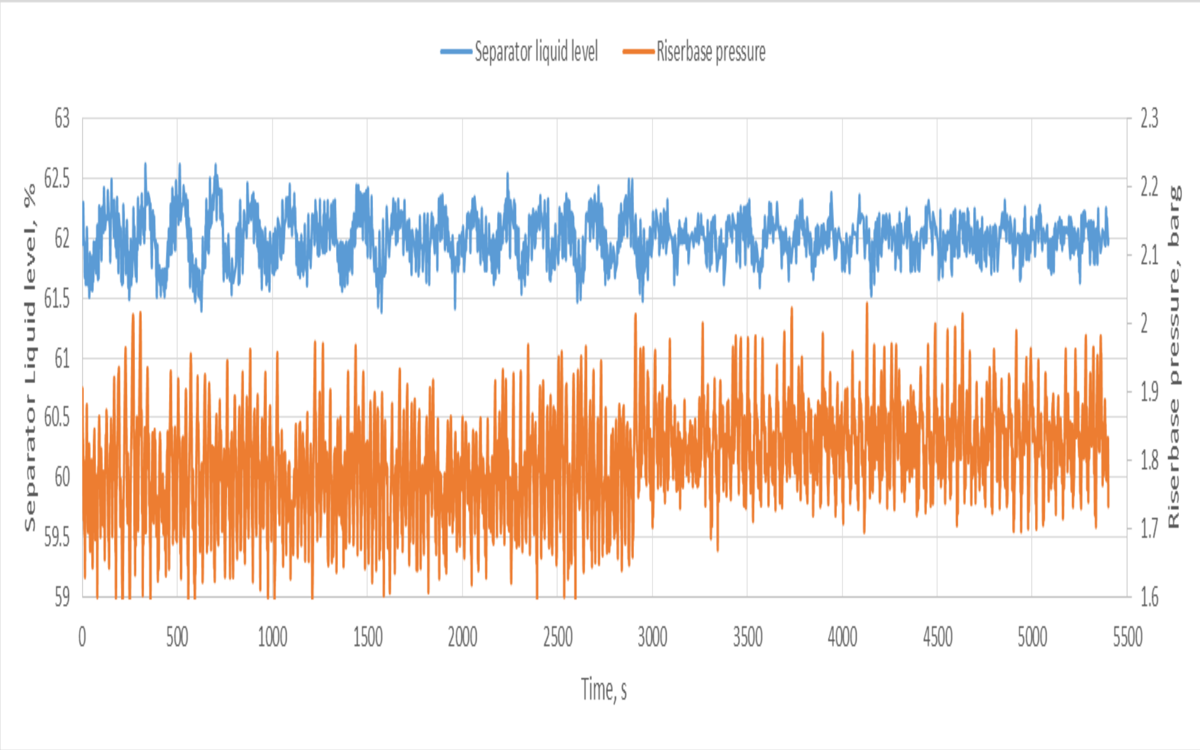


Figure 5-17 Separator liquid interface level and riserbase pressure trend for the S-shape riser with and without the wavy pipe section

From Figure 5-17, the blue trend represents the separator liquid interface level while the orange trend represents the riserbase pressure trend for the S-shape riser pipe configuration coupled with and without the wavy pipe section at a fully opened valve (100 % valve opening). It could be observed that, there is a relatively large pressure fluctuation in the system which translates and reflects in the liquid interface level in the separator before 2900 seconds. After 2900 seconds, when the wavy pipe was coupled to the system, a reduction was observed in the both the riserbase pressure and the fluctuation in the liquid interface of the separator. Thus, the liquid level interface smoothens out relatively and this could be attributed to the wavy pipe section installed at the topside of the riser pipeline system.

Similar to the quantitative analysis perform for the S-shape riser system coupled with the wavy pipe, the S-shape riser pipeline system coupled with the spiral pipe section was assessed quantitatively. An investigation was performed to assess the outlet conditions and behaviour of the flow through the S-shape riser system coupled with the spiral pipe section.

Figure 5-18 shows the riserbase pressure and separator liquid level trend for the flow condition 1 kg/s liquid and 10 Sm^3/h gas through the S-shape riser system coupled with and without the spiral pipe section at 100 % choke valve opening. From Figure 5-18, before the spiral pipe section was coupled to the system, thus before 2900 seconds, the system operated solely as the S-shape riser without any additional PST section. There exist large oscillations in the riserbase pressure (with magnitude 1.6984 barg) and hence a fluctuation in the flowrate of liquids out of the system which is observed in the large fluctuation in the separator liquid interface level. Beyond 2900 seconds from Figure 5-18, the spiral pipe section was coupled with the S-shape riser pipeline system. From 2900 second to about 4600 seconds, there was still a large fluctuation in both the riserbase pressure and the separator liquid interface level. However beyond 4600 seconds, there is a slight reduction in the magnitude of separator liquid interface level and a relatively lower riserbase pressure oscillation but of higher magnitude (1.8281 barg).

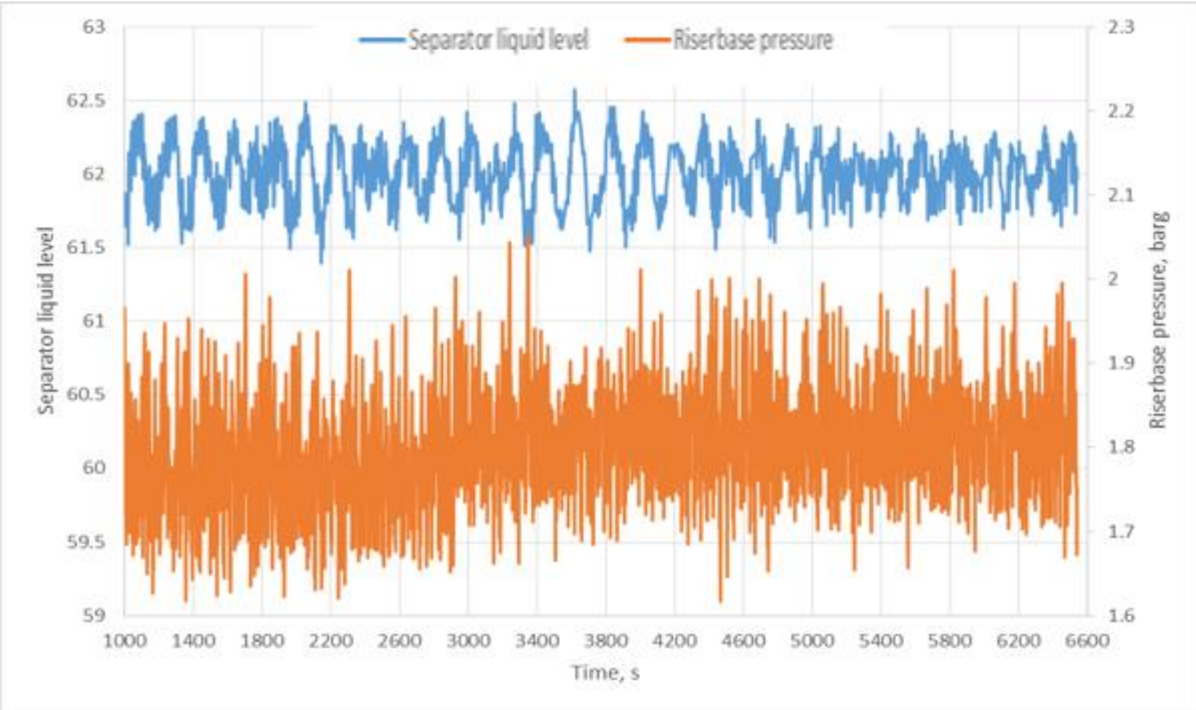


Figure 5-18 Separator liquid interface level and riserbase pressure trend for the S-shape riser with and without the spiral pipe section

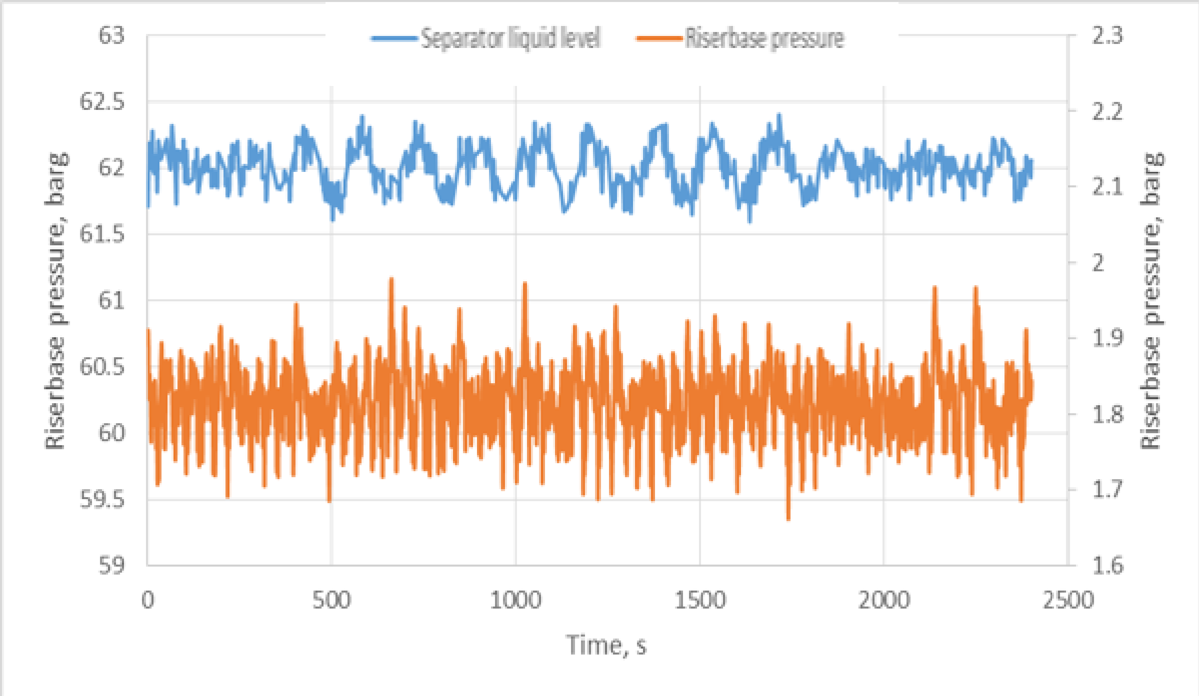


Figure 5-19 Separator liquid interface level and riserbase pressure trend for the S-shape riser without the spiral pipe section at 50 % choke valve opening

For the S-shape riser without the spiral pipe coupled to operate at this same riserbase magnitude (1.8281 barg), it requires a further reduction in the valve opening. To attain 1.8281 barg, the choke valve was closed 50 % of its original diameter. Figure 5-19 shows the separator liquid interface level and the riserbase pressure trend for the S-shape riser without the spiral pipe at a 50 % choke valve opening. This implies that the production through the choke was reduced significantly compared to the S-shape riser system coupled with the spiral pipe section. It could be deduced that the spiral pipe section has some positive impact on the system.

The benefit derived from the used of the PST (wavy and spiral pipe section) as a slug mitigation technique solely and in conjunction with other techniques will be discussed next.

5.4 Discussion

Flow regime alteration and physical characteristics parameter of the flow comportment have been assessed to investigate the PST pipe's performance ability to mitigate slug flow through a riser pipeline system. Conclusively, the PST pipe has been observed to significantly reduce the slug length in the riser pipeline system. The PST installed at the topside of the riser works by accelerating the gas phase which in effect aids bubble penetration hence mixing the gas and liquid flow through the system. Thus, the PST imparts swirl to the liquid phase causing gas entrainment at every peak and trough of the PST. Again the pipe length after the PST but before the separator does not enable re-separation of the phases hence a relatively mixed flow into the separator. It has been established through experimentations that the PST pipe section has an encouraging consequence on decreasing the slugging region and the rigorousness of the slug induced difficulties with regards to the pressure variations in the conduit and the liquid throughput of the riser.

The configuration of the PST pipe section is such that the wavy pipe has a deeper sump depth than that seen with the spiral pipe. The series of sumps aids in collecting liquid which potentially blocks the passage of gas until the pressure is built up. There is a likely possibility of blocking the entire sump in the spiral pipe section hence preventing the entirety of the gas passage in the spiral till there is sufficient build-up of pressure to cause a potential mini blow out. This action causes a slight back pressure increase hence the higher riserbase pressure for the spiral pipe section compared to the wavy pipe section. From the configuration of the PST, the troughs through liquid accumulation, traps the gas flowrate upstream the dip which leads to a build-up of pressure. A resulting mini blowout is observed which aids in the redistribution the two phase flow in the system and an aerated flow upstream the separator.

In effect the geometry of the PST pipe and the operating conditions plays a part on the flow behaviour in the pipeline riser system. For this study however, only one geometry each of the wavy and spiral pipe section was tested. This is due to the space constraint of the experimental setup. Thus, the location of the PST relative to the separator could not be manipulated and can potentially have an effect on the performance of the PST. Again, the approximate pipe length between the PST and the separator was fixed hence the effect of approximate length on the performance could not be investigated.

The wavy pipe section, showed greater benefit in stabilising unstable slug flow through the S-shape riser relative to that observed when the system is coupled with a spiral pipe section. This is evidently seen in the smaller unstable slug envelope achieved for the system configuration coupled with the wavy pipe section from the flow regime map in Figure 5-5. The wavy pipe section works by transforming the flow regime in the pipeline upstream the separator. There is a minute likelihood for the flow to observe slugging in the wavy pipe section than in a horizontal or vertical traditional pipe. The wavy pipe section has a larger amplitude to pipe diameter ratio which results to an extremely aerated flow before and after each pulse of the wavy pipe hence its greater benefit. Consequently the spiral pipe section also helps reduce the unstable slug flow region from the system without the PST but the degree of slug region reduction is smaller compared to that seen when the wavy pipe was coupled.

Also the PST pipe section through experiments has been observed to aid and enhance manual choking technique (parameter variation technique) as shown in Table 5-5 and Table 5-6 for a mass driven source and a pressure driven source respectively. The wavy pipe section derived greater benefits in aiding the manual choking technique. There was an observed increase in the stable valve opening for all the unstable slug flow conditions assessed through the system coupled with the wavy pipe section. The spiral pipe section for low flowrate through the system showed not much improvement compared to the system without any PST pipe section. However, for relatively higher flowrate through the system, the spiral pipe section showed its capability in aiding the parameter variation technique. It could be deduced that the spiral pipe section affects the flow behaviour or characteristics optimally but however not for very low flowrates.

In conclusion, even though one configuration each of both wavy pipe and spiral pipe section were deployed in this study, it could be deduced that both PST pipe sections have some slug mitigation potential. Thus, the PST pipe sections when installed at the topside of the riser has the capability of mitigating slug although its effectiveness or optimal slug mitigation is dependent on the flow condition. This study is however preliminary and would require a more comprehensive study to reveal the effective flow conditions corresponding to different design in the PST geometry.

5.5 Summary

The slug mitigation potential of the PST (wavy pipe configuration and spiral pipe configuration) when installed at the topside of the S-shape riser system has been investigated and the outcome will be outlined in this section. For the simplicity of this discussion the S-shape riser coupled with the wavy pipe section is noted as case 1, the S-shape riser system without any PST attached is noted as case 2 and the S-shape riser system coupled with the spiral pipe section would be noted as case 3.

The flow behaviour and pattern for the system was observed by investigating the riserbase pressure trends for slugging flow conditions through the plain S-shape riser system configuration (without PST), the S-shape riser system coupled with the wavy pipe section and the S-shape riser system coupled with the spiral pipe section. Four unique flow regimes were observed through the system. From the flow behaviour through the system slug flow, churn flow, annular flow and bubbly flow regime patterns were identified. With the focus of the study on slug flow, a slug flow envelope was developed for the plain S-shape riser system configuration (without PST), the S-shape riser system coupled with the wavy pipe section and the S-shape riser system coupled with the spiral pipe section. Evidently, slug flow patterns were identified using an analysis of the riserbase pressure and visual observation.

Some flow conditions exhibited slugging flow behaviour through the plain S-shape riser system configuration (without PST) and was evidently seen in the riserbase pressure fluctuations as shown from previous sections. For this same flow conditions, there existed a reduction in the riserbase pressure oscillation magnitude for the flow through the S-shape riser configuration coupled with either the wavy pipe section or the spiral pipe section but however the pressure variation in the S-shape riser coupled with the wavy pipe section was lower than that observed in the S-shape riser coupled with the spiral pipe section. Again, there was some observed back pressure imposed on the system for the S-shape riser coupled with the wavy pipe section and the S-shape riser coupled with the spiral pipe section which intend increased the riserbase pressure in both cases. In effect the spiral and wavy pipe from this analysis has some slug flow minimising effect but the S-shape riser coupled with the wavy pipe section has the maximum slugging minimising effect.

Parameter variation technique was performed for some slugging flow condition for the plain S-shape riser system configuration, the S-shape riser system coupled with the wavy pipe section and the S-shape riser system coupled with the spiral pipe section where critical bifurcation points were obtained. For the inlet of the system configured to run as a mass source, the S-shape riser system coupled with the wavy pipe section showed a larger critical bifurcation point in comparism to the S-shape riser system coupled with the spiral pipe section and the plain S-shape riser configuration (without PST). Thus, the system becomes stable at a larger valve opening in the S-shape riser system coupled with the wavy pipe section which corresponds to a lower riserbase pressure relative to the S-shape riser system coupled with the spiral pipe section (Case 3) and the plain S-shape riser configuration (without PST) (Case 2). Transposing this riserbase pressure unto case 2 and case 3, proves that there exist some instabilities in these respective cases. The S-shape riser system coupled with the spiral pipe section showed much deteriorating system behaviour compared to the plain S-shape riser configuration (without PST) for low flowrate through the system. Similar observations were made when the inlet source was fixed as a pressure driven source. It could be deduced and conclusively observed that the S-shape riser system coupled with the wavy pipe section could potentially aid parameter variation technique (manual choking) and in effect help improve the throughput of the system whiles stabilising the system at a larger valve opening. Even though the spiral pipe section possesses the same potential of aiding manual choking, it could not be conclusive for all flowrate through the system.

Quantitatively, Table 5-7 shows the outcome of the separator liquid interface level and the riserbase pressure for the plain S-shape riser configuration (without PST), the S-shape riser system coupled with the spiral pipe section and the S-shape riser system coupled with the wavy pipe section. Even though there was an increase in the riserbase pressure for both the S-shape riser system coupled with the wavy pipe section and the S-shape riser system coupled with the spiral pipe section, the magnitude of oscillation reduced significantly especially in the configuration with the wavy pipe section. The wavy pipe section had a much greater potential in smoothening out the flow.

Table 5-7 Quantitative experimental outcome

	Case 3	**Case 2**	**Case 1**
Riserbase pressure oscillation magnitude	0.356	0.3903	0.345
Mean riserbase pressure	1.8281	1.6984	1.7513

From the analysis and study conducted, it could be deduced that the wavy pipe section has a greater mitigation and optimizing potential relative to the spiral pipe section when coupled to the system. Thus, the wavy pipe section when coupled to the S-shape riser system could optimize the manual choking technique as it aids stabilising the system at a larger valve opening in open loop. Similarly there was a relatively steady flow out of the system when coupled with the wavy pipe section. This could be attributed to the rigorous mixing effect imposed on the system by the wavy amplitude of the pipe piece.

In conclusion the S-shape riser system coupled with the wavy pipe section exhibited the best outcome on all the test conditions assessed, whilst the S-shape riser system coupled with the spiral pipe section has a slighter edge over the plain S-shape riser configuration (without PST) for relatively high inlet flowrates. A comprehensive study will be required to reveal the effective flow conditions corresponding to different PST geometry design. Moreover, a further study to integrate PST configuration with ISC would also be required to establish how the PST could potentially enhance the ISC.

6 SMITH PREDICTOR FOR SLUG CONTROL WITH LARGE VALVE STROKE TIME

6.1 Introduction

Active control as a slug mitigation technique, has been established by several researchers (Yocum, 1973; Taitel, 1986; Sivertsen and Skogestad, 2000; Havre and Dalsmo, 2001; Storkaas, 2005; Cao, Yeung and Lao, 2010a; Sivertsen, Storkaas and Skogestad, 2010; Cao, 2011; Vidal et al., 2013;) as one of the most effective and optimal means of mitigating slug. This is because it stabilises an unstable slug flow at a valve opening larger than the critical manual choking point (open loop stability point). Due to the potential of active control, several other researchers have delved into it. Many outcomes from pilot scale experiments (usually 2-4 inch valves), that shows promising and improved benefits, fail to replicate when emulated on real offshore facilities (over 8 inch valve diameters). This is because of the difference in the valve stroke time (time taken for a valve to move from fully open to fully close or vice versa). Comparatively, larger diameter valves have larger stroke time than smaller diameter valves as they are designed as such. Chapter 3 and 4 addressed the flow dynamics in different riser configurations. The stabilisation of unstable slug flow conditions in these risers was attempted using the inferential slug controller.

This chapter investigates the use of Smith predictor, a well know time delay estimator to control unstable slug flow under a varying time delay resulting from large valve stroke time. The stability of the system with large stroke time is of key interest. The remaining of the chapter is organised as: Section 6.2 presents a general overview of time delay in processes and the means by which the valve stroke time introduces delay in the system; Section 6.3 gives a description of the traditional Smith predictor and the model; Section 6.4 describes and outlines a modification of the Smith predictor model to stabilise both stable and unstable systems with large valve stroke time; Section 6.5 presents a proof of the concept of extending the Smith predictor for unstable systems with varying time delay using measureable delayed input signal. Pipeline configuration, controller implementation, controller tuning and system identification with some results are presented. The chapter finally ends in Section 6.6 with the concluding remarks.

6.2 Background

6.2.1 Time delay in process systems

Time delay has been existing in real processes and these (time-delay) systems have been deployed in modelling large class of engineering systems where transmission of materials and information is required. The major sources of sluggishness in process control systems, nuclear systems, communication systems and their likes is the time delay phenomena. Time delay in system could render the system less optimal in terms of process output or even unstable. In system control, time delay makes analysing the system and the design of controllers much difficult. This has prompted an investigation in these past few decades in the quest to stabilise systems with time delays.

Smith predictor, the very first potent control technique to deal with systems with time delay was proposed by O J M Smith in the 1950s (Warwick and Rees, 1988). However, that was only for stable systems and an extension to deal with unstable systems was presented in the 1970s (Andrzej and Andrzej, 1979). Due the challenges involved with time-delay unstable systems, some researchers (Majhi and Atherton, 1998; Shamsuzzoha, Jeon and Lee, 2007; Lee, 2008; Molnar and Insperger, 2016; Ajmeri and Ali, 2017; Sanz, García and Albertos, 2017) have delved into investigating the stabilization of such systems. Double loop control scheme was used in (Park, Sung and Lee, 1998; Wang and Cai, 2002) for performance enhancement on unstable systems with time delay to address the issue of large settling time and excessive overshoots associated with previous methods. A modification of the internal model control (IMC) method was explored by Tan, Marquez and Chen, (2003) for two degree of freedom to control unstable processes with time delay to defeat the notion that IMC structure was not suitable for unstable systems due to internal instabilities. Based on the Smith predictor concept, Zhang et al., (2004) proposed a two degree of freedom control method which showed a smooth set point response for a first order unstable process with time delay without any unnecessary overshoot. To avoid the use of complex controllers, which is not helpful for industrial processes, control schemes based on a new modified Smith predictor covering for stable/unstable or minimum/non-minimum phase processes was presented in Albertos, Garcı and Ha, (2006), thus, combining the outputs for both a finite impulse response (FIR) and a stable filter for the process input and output respectively in systems with time delay. A control scheme was presented by Liu, Zhang and Gu, (2005) to stabilize unstable model with time delay, where three controllers were used to reject disturbance, stabilize the plant and to track the reference. Numerous robust control difficulties have been unravelled, but typically for systems with a single delay. In recent times, many researchers paid attention to the interval time-varying delay, where the delay fluctuates in a range for which the lower bound is not constrained to be zero.

6.2.2 Stroke time in actuator system

In the oil and gas field several valves used in operations including those for control purposes have large stroke times. This is due to the fact that these valves in the field have large diameters relative to those used in experiments. Hence they are designed as such, with large stroke time, to avoid sudden shut of valves. The consequences resulting from a sudden shut of larger diameter valves is catastrophic and would have a greater negative impact compared to smaller diameter valves.

Stroke time of the actuation system introduces some time delay in the system but for closed loop systems, the delay time is variable especially for control system in the field (oil and gas). The dependency of the time delay on the percentage opening at each sampling point introduces the variable time delay in the system. The variable time delay makes an estimation of the time delay highly uncertain together with the unstable nature of slugging flow resulting to challenges with slug control for large diameter production systems. The difficulty and challenge to control unstable systems with time delays are well recognized by many researchers (Huang and Chen, 1997). The challenge comes from the conflicting requirements where stabilizing control requires larger gain and quick response whilst the largest control gain and bandwidth have to be limited due to existing time delays.

6.3 Smith predictor

The Smith predictor is a model-based predictive controller that is effective for processes with large time delays. It has an inner loop with a core controller that can easily be designed devoid of the dead time. The effects of load disturbance and modelling error as a result of the delay in the system is enhanced through an outer loop.

The Smith predictor, has been widely used in controlling systems with delay times (Majhi and Atherton, 2000a, 2000b; Abe and Yamanaka, 2003; Kaya, 2003, 2004; Matušů and Prokop, 2011; Paulsen, 2012). Due to the configuration, the traditional Smith predictor treats a time delay as a measurement delay. Furthermore, the traditional Smith predictor requires an estimation of the time delay value at the design stage. The model of the traditional Smith predictor will be discussed in subsequent sections.

6.3.1 Smith predictor model

Figure 6-1 shows the block diagram of a traditional Smith predictor model, which consists of a predictive loop and a controller. The predictive loop is made up of basically a system model without delay and an estimated time delay model. Generally, the principle behind the traditional Smith predictor is that, the predicted output is compared with the plant output to produce an error value. If there exists a zero error in the process, then the signal from the plant model without delay can be adopted for control purpose. As such the controller can be designed without considering the time delay in the system and this in effect can improved control performance tremendously.

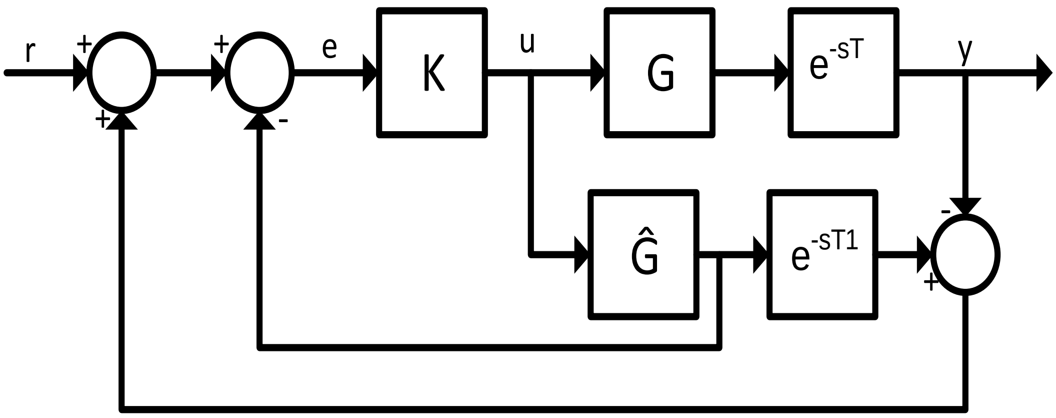


Figure 6-1 Traditional Smith predictor block diagram model

The closed loop transfer function of the traditional Smith predictor model without any model error is given in (6-1).

$$\frac{Y}{R} = \frac{K_{(s)}G_{(s)}e^{-sT}}{1 + K_{(s)}G_{(s)}} \qquad \textbf{(6-1)}$$

In the control of severe slugging flow in closed loop actuators are being used. On the fields these actuator systems have associated large stroke time which introduces a variable time delay to the control system and may destabilize the closed –loop system. The obstacle is not merely from the large time delay even though it is a recognized factor of destabilization, but also from the unreliable nature of the delay which is reliant on control increment at each sampling point rendering common techniques to handle the time delay null and void.

This requirement makes traditional Smith predictor not suitable for the variable time delay introduced by a valve stroke time. Fortunately for our system, the actual valve position, is available as a measurement. This is adopted to modify the Smith predictor model to control severe slugging conditions. The modified Smith predictor was further extended to deals with unstable systems with varying input time delay using a measurable delayed input signal. The extension and modification of the traditional Smith predictor model to deal with systems with variable time delay would be discussed next.

6.4 Modified Smith predictor model

6.4.1 **Modified Smith predictor model for stable system**

For actuation systems with large stroke time, when the Smith predictor is implemented, the time delay at the input can unvaryingly be moved to the output for a single variable system. However, at the design stage, the realistic valve opening for time delayed system is unknown at every sample point. Thus, the actual value is dependent on the input increment. The traditional Smith predictor is unable to deal with time delays introduced by valve stroke time due to the varying time delay in the system. This makes a modification to the traditional Smith predictor a necessity. In most industrial processes, the actual valve position could be measured and used to reconstruct a new Smith predictor, rendering an estimation of the varying time delay irrelevant as shown in Figure 6-2. For zero model error, the closed loop transfer function of the modified Smith predictor is the same as in (6-1).

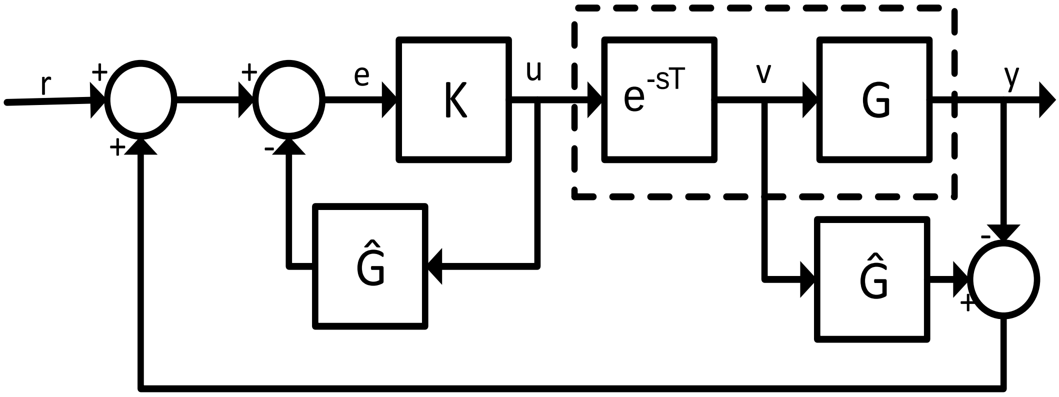


Figure 6-2 Modified Smith predictor block diagram model

6.4.2 Modified Smith predictor model for unstable system

In dealing with unstable time delay systems, Liu et. al (Liu, Zhang and Gu, 2005) proposed a scheme which decouples both the response to load disturbance and set point response by using an open loop technique for set point tracking. Liu's control scheme shown in Figure 6-3 has two degrees of freedom and is capable of dealing with unstable processes with time delays. This scheme has shown great potential in terms of load disturbance rejection and set point tracking for unstable delayed systems.

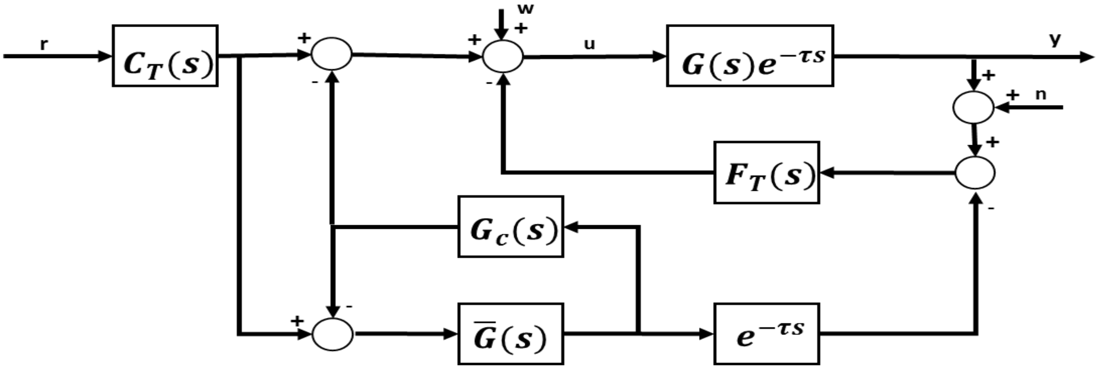


Figure 6-3 Liu's Smith predictor control scheme

Firstly, a normal controller (P or PD) is used to stabilize the set point response, then a derived set point tracking controller is obtained in terms of an integral-squared-error (ISE) specification. Again, a controller for rejecting load disturbance is derived from a sensitivity function for the closed loop system. From Figure 6-3, C_T is used for set point tracking, G_C is deployed for set point response stability whilst F_T is used for load disturbance rejection.

For our system however, due to the varying time delay, Liu's scheme is modified as shown in Figure 6-4 and used to stabilize the destabilized system as a result of the large stroke time in the valve. From Figure 6-4, the measured valve position feeds through an estimated plant model to produce a signal compared with the plant output if there are no load disturbance in the system. A second signal is produced by linking the estimated plant model to an auxiliary controller which produces a stabilized set point response and compared with the set point tracking controller output. Time delay caused by the stroke time of a valve for unstable systems, can easily be dealt with using the modified Liu's scheme. The viability of the new Smith predictor would be evaluated based on its ability to improve the control performance of a destabilized system. A case study of a stabilized unstable flow which is destabilized by a large valve stroke time is presented next to illustrate the viability of the modified Smith predictor model.

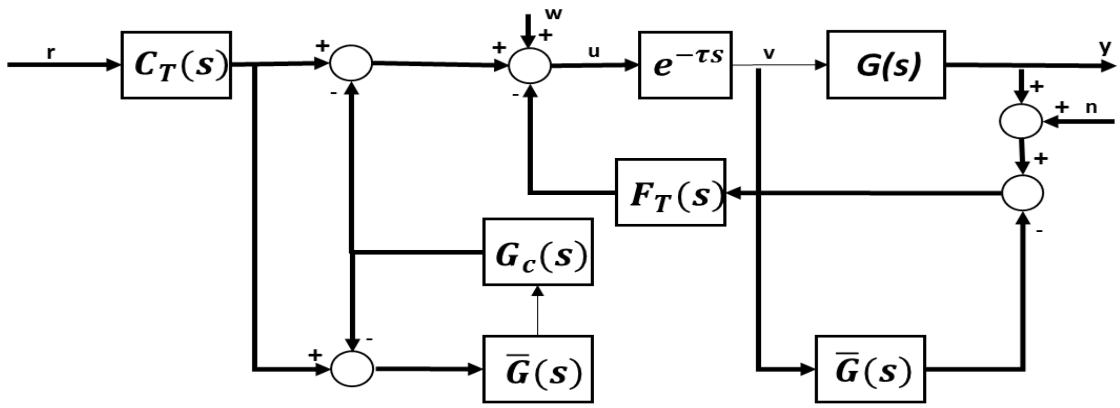


Figure 6-4 Modified Liu's scheme for variable delay

6.5 Case study

6.5.1 Stable systems

6.5.1.1 Pipeline configuration

The pipeline system used in this study is a vertical riser adopted from Ogazi, 2011. The model is an 8-inch nominal vertical riser pipeline system that has a 5000 m horizontal section and a riser height of 120 m. The pipeline system is equipped with a choke valve (8-inch) at the outlet of the riser. The schematic of the pipeline and its features is shown in Figure 6-5.

To fulfil the aim of this study, the boundary condition of the system was allowed to operate in a slug region. The pressure trend from the boundary condition in Table 6-1 shows a continuous pressure fluctuations, which represents a slug flow regime.

Table 6-1 Operating conditions and parameters for the vertical riser system

Operating Parameters (Initial Condition)						
Source Temp. (°C)	Ambient Temp. (°C)	Overall heat transfer coefficient (W/(m²*K))	Reservoir Temp. (°C)	Outlet Temp. (°C)	Reservoir Pressure (bar)	Outlet Pressure (bar)
44.44	4	5	70	80	69	30
Total Mass Flow, kg/s		**Gas Mass Flow, kg/s**		**Oil Mass Flow, kg/s**	**Water Mass Flow, kg/s**	
20.53		0.525		10	10	

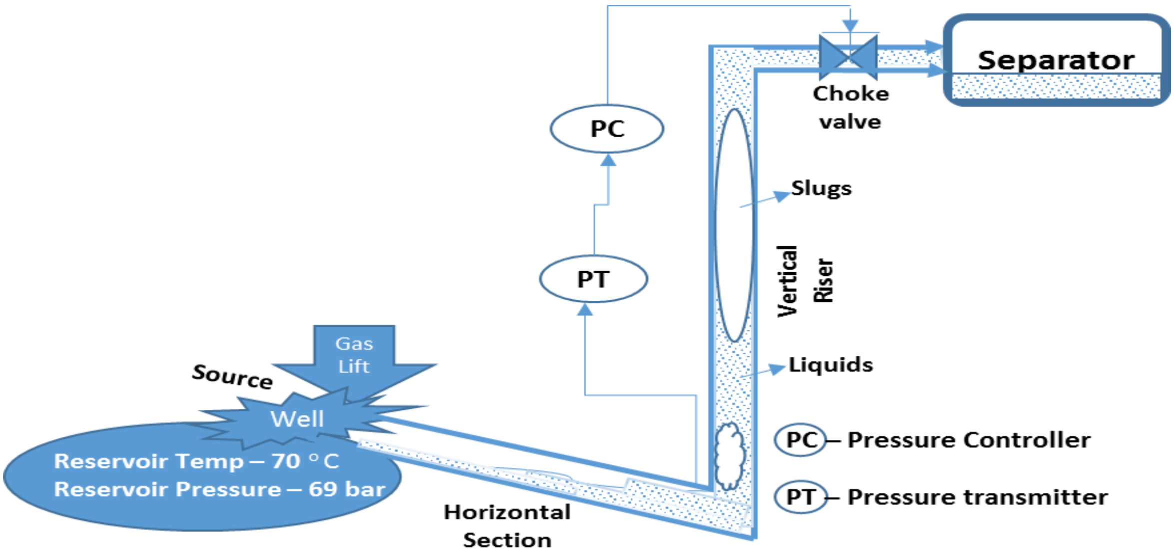


Figure 6-5 Schematic diagram of the riser model

6.5.1.2 Open loop stability

Figure 6-6 shows a system stability plot (bifurcation analysis plot) in open loop mode using the topside choke valve as the manipulated variable. Bifurcation study was performed to determine the system's critical bifurcation point, which translates to the maximum valve opening at which the system achieves stability in open loop, thus one of the first steps in controller design. With the same flowrate, the valve opening was varied and the resultant pressures plotted against their respective valve openings.

From Figure 6-6, the system achieved stability (bifurcation point) at a valve opening of 10 %, with a corresponding P_{RB} of 42.20 bara. This signifies the largest valve open at which the system was stable without any active control (open loop). For valve opening, u, less than 10 %, slugging does not exist hence there is no or minimal pressure oscillation / fluctuation. For u >10 %, the system becomes unstable again. Thus, the pipeline riser system experiences instability due to large pressure and flow oscillations observed at the base of the pipeline riser system configuration.

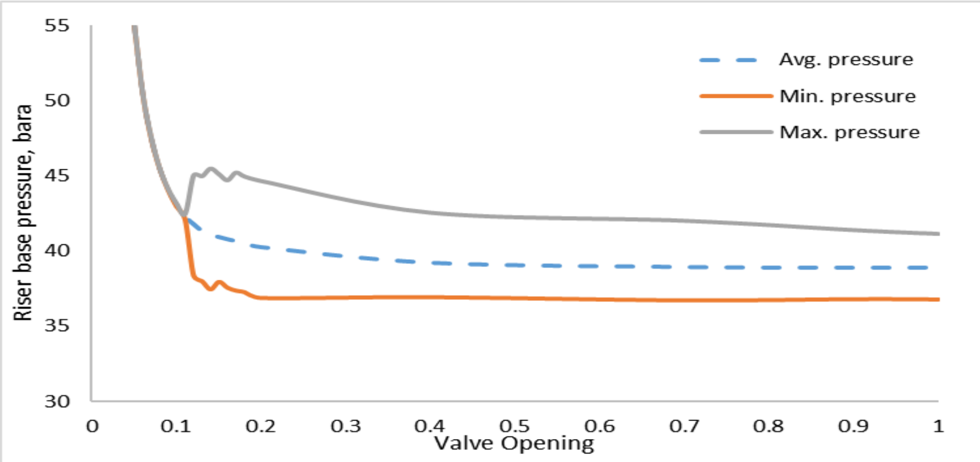


Figure 6-6 Riserbase pressure against valve opening (bifurcation map)

6.5.1.3 Controller design

Having established the stability point in open loop, the next step was to control the system at a larger valve relative to that in open loop. This would serve as a baseline to establish the effect of the stroke time on the system.

With the help of an active control, using the riser base pressure as the controlled variable, the system can be stabilized at larger valve opening. A PID controller was designed to control the riser base pressure at a given set point by regulating the choke valve at the riser top. The controller parameters were designed based a linear model corresponding to the valve opening at 11 % (Ogazi, 2011). The calculated PID controller design parameters is given by$K_C = -15$; $\tau_I = 500$; $\tau_D = 0.005$.

The estimated linear model of the system, G at 11 % valve opening (Ogazi, 2011) is given by ;

$$G = \frac{-0.258s - 0.0004248}{s^3 + 7.994s^2 - 0.002577s + 1.206e^{-005}} \quad \textbf{(6-2)}$$

Having established the system model and the controller parameters, the Simulink block was updated and simulations run. With the PID controller in action, and a 1 % stepwise increase in the choke valve opening, Figure 6-7 shows the resultant riser base pressure response. In closed loop, the system was able to achieve stability at a 13 % valve opening. For regions beyond 13 % valve opening, the system's stability could not be retained which resulted in large pressure undulations.

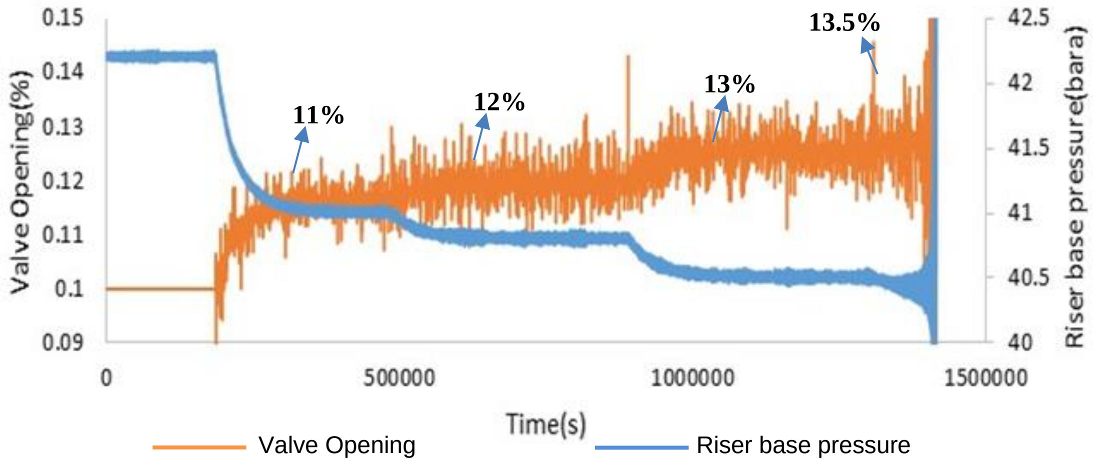


Figure 6-7 Simulation result for riserbase pressure slug control with zero stroke time

6.5.1.4 Impact of stroke time

The effect of stroke time was assessed next after establishing 13 % choke valve opening as the point for which the system with zero stroke time stabilizes under a simple PID controller. A valve stroke time of 250 seconds was introduced into the OLGA model to show its impact on control performance. Figure 6-8 shows the riserbase response in closed loop using the manipulating variable (choke valve) with a large stroke time.

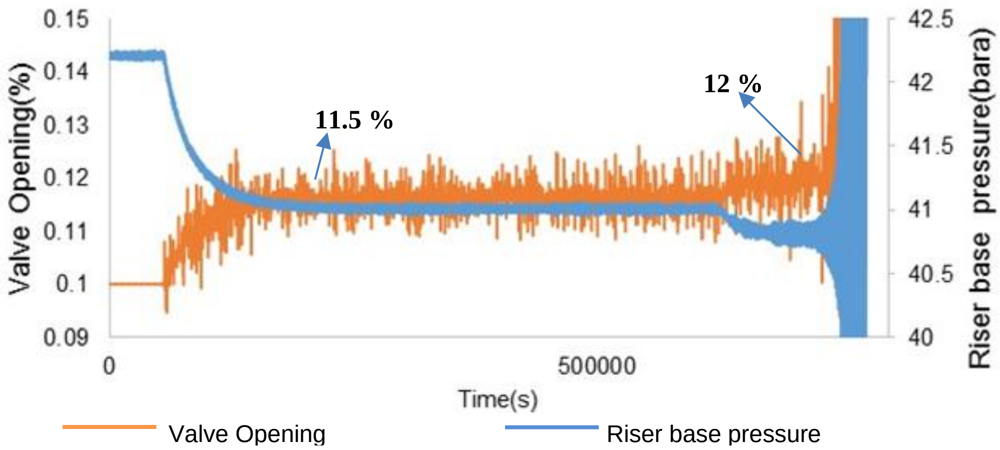


Figure 6-8 Simulation result for riserbase pressure slug control with the time delay effect

From Figure 6-8, the stroke time (time delay) definitely has an effect on the riserbase response hence the control performance. There are large pressure oscillation observed even at a choke valve opening of 12 %, thus the system under this condition become stable only at a 11.5 % valve opening. This valve opening as observed is relatively lower compared to when there was no or little valve stroke time. The large pressure oscillation at a much lower valve opening was because of the instabilities introduced by the valve stroke time, which causes a delay in the valve response. Thus evidently the valve stroke time has an adverse effect on the control performance of the system.

To improve the system response with the delay time a resultant of the large valve stroke time, the new Smith predictor control was adopted for slug control. Figure 6-9 shows the riser base pressure response using the modified Smith predictor control scheme. From Figure 6-9, the system achieved stability at a choke valve opening of 12.5 %. Comparatively, there was an increased valve opening achieved by introducing the modified Smith predictor to control unstable slugging flow the pipeline riser system configuration.

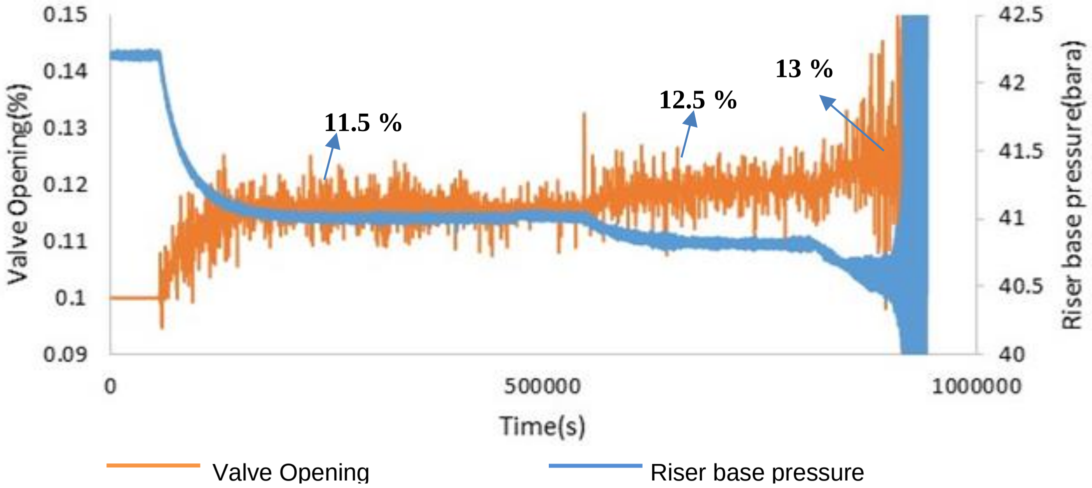


Figure 6-9 Slug control with time delay effect using the modified Smith predictor control scheme

6.5.1.5 Discussion

As shown in Figure 6-7, the controller stabilized flow from the pressure set-point change of 42.2 bara to 40.5 bara, while increasing the choke valve opening from 10

% to approximately 13 %. This translated to the minimum pressure achievable without any time delay effect.

For an inputted time delay (stroke time), thus a system with delay effect, the simulation result shown in Figure 6-8 proved that the control performance degraded when using a simple PID controller. The achieved stable pressure was changed from 40.5 bara to 41 bara for the system with the delay effect, which corresponded to a valve opening of approximately 11.5 %, thus lower than the 13 % valve opening achieved without the valve stroke time. In effect, a 2.84 % reduction in riser base pressure was observed for the system with valve stroke time compared to a 4.028 % reduction in riser base pressure for the system without valve stroke time.

Again, Figure 6-9 representing slug control scheme using the modified Smith predictor, shows system stability at a riser base pressure of 40.8 bara corresponding to 12.5 % valve opening. Comparatively, an improvement on a reduced riser base pressure from 41 bara without the modified Smith predictor to 40.8 bara with the modified Smith predictor for the system with delay effect. This represented a 3.34 % reduction in the riser base pressure with the modified Smith predictor compared to a 2.84 % reduction in the riser base pressure, which translates to an increase in oil production benefit in using the modified Smith predictor for the system with time delay in the actuation system.

The novel modified Smith predictor, which renders time delay estimation unnecessary, can deal with systems with varying time delay caused by valve stoke time. For the specific case study, the proposed modified Smith predictor is applied to slug control problem and achieves its task as an increased benefit from 2.84 % reduction in P_{RB} without the modified Smith predictor to 3.34 % decrease in P_{RB} obtained with the modified Smith predictor for the pipeline system with the delay effect. Reduction in riser base pressure for a riser system in the oil and gas industry increases the pressure drop across the riser, which in effect translates to increased oil production rate of the system. With the help of a more robust controller, greater benefit could be achieved using the modified Smith predictor for systems with large time delay.

6.5.2 Unstable system

6.5.2.1 Riser pipeline configuration

The pipeline configuration adopted for this study is a U-shape pipeline riser system which is a form of a platform to platform pipeline system. A 0.289 m uniform diameter U-shape pipeline riser system was used for this study. The features of the U-shape riser was presented in Chapter three. To satisfy the vision of this study, a slug flow regime needed to be triggered in the pipeline riser system. From a slug envelope developed, Table 6-2 shows an operating condition for which the system experienced slugging flow behaviour. The stability of this flow condition would be assessed next.

Table 6-2 Operating conditions and parameters for the U-shape riser

Operating Parameters (Initial Condition)						
Source Temp. (^{O} C)	Ambient Temp. (^{O} C)	U value (Wat/(m* K))	Inlet Temp. (^{o} C)	Outlet Temp. (^{O} C)	Inlet Pressure (bar)	Outlet Pressure (bar)
44.44	23.89	69.18	47.22	47.22	12.41	12.41
Operating Conditions						
Total Mass Flow, kg/s	Gas Mass Flow, kg/s	Oil Mass Flow, kg/s	Water Mass Flow, kg/s	Inlet Temp. (^{o}C)	Outlet Temp. (^{o}C)	Outlet Pressure (bar)
10.22	0.22	5	5	44.4	23.89	10.687

6.5.2.2 System stability (manual choking)

The stability of the unstable slug flow condition was assessed manually to achieve the critical bifurcation point. This translates to the maximum valve opening at which the flow in the pipeline system becomes stable in open loop. This is usually a performed activity prior to designing a controller, thus, a reference point for the controller to stabilize flow in an open loop unstable region.

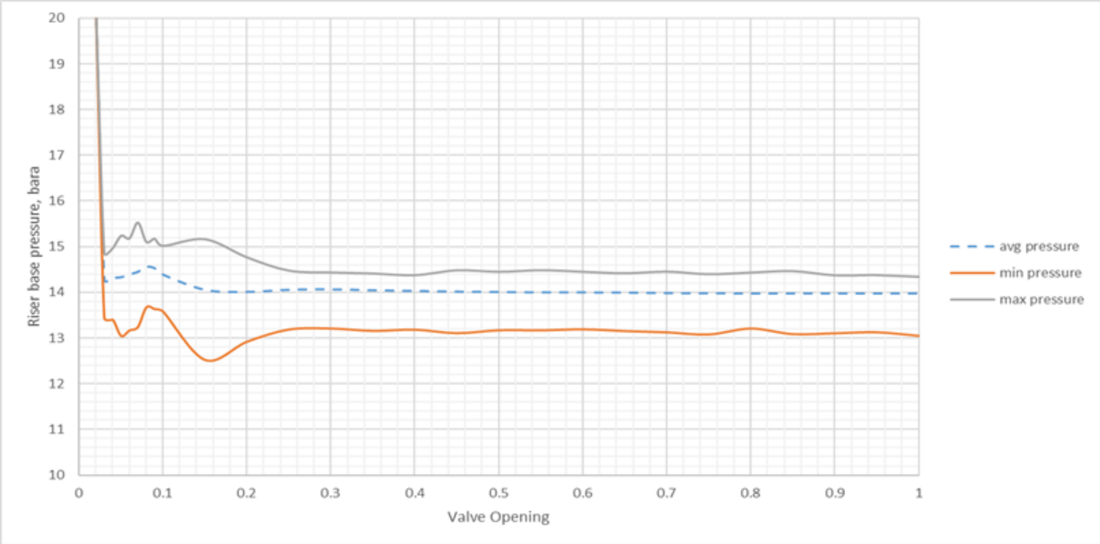


Figure 6-10 Bifurcation map for the U-shape riser

Figure 6-10 shows the bifurcation map obtained for the boundary condition shown in Table 6-2. The system stability was achieved at a valve opening of 3 %, corresponding to a pressure of 14.8 bara. From the result of the bifurcation map, stabilizing the system at the open-loop unstable region where u >3 % will be aimed in order to obtain a desired stable non-oscillatory flow regime. The controller to aid this study is discussed next.

6.5.2.3 Controller design

Having established the stability point in open loop, the next step was to control the system at a larger valve relative to that in open loop. This would serve as a baseline to establish the effect of the stroke time on the system. Riser base pressure dictates the throughput of a pipeline system while the pressure gradient dictates the stability of the system. Stability is determined by dP/dQ>0 whereas dP/dQ<0 is an unstable flow.

ISC was designed to stabilize the system at a 5 % valve opening. The controller parameters were designed base on a methodology illustrated in the previous chapters. The minimum gain, K of the ISC for any preferred pressure drop gradient at a particular valve opening could be obtained using (3-13) in the quest to stabilize slug flow at an increased valve opening.

The set of coefficients of the ISC measurement weight for gas mass flowrate (GG), liquid density (ROL) the top pressure (PT), and the total liquid mass flowrate (GLT) corresponds to (0.6385, -0.2670, 0.6449, 0.3242) respectively. From (3-13), the deviation in the vector of measurement signals represented by dY/dQg for each measurement signal GG, ROL, PT and GLT resulting from a perturbation in the Qg was obtained from simulation as -0.4360, 0.5911, 0.4233 and -0.5305 respectively. The estimated minimum gain of the ISC represented by K in (3-13) is 0.0004.

Having implemented and engaging the ISC in action with a 1 % stepwise increase in the choke valve, the resultant riserbase pressure trend is shown in Figure 6-11. It shows that the ISC was able to stabilize the flow at an increased choke valve opening, thus from a 3 % to approximately 12 % choke valve opening. Backpressure was lowered by approximately 1 bar at an increased valve opening.

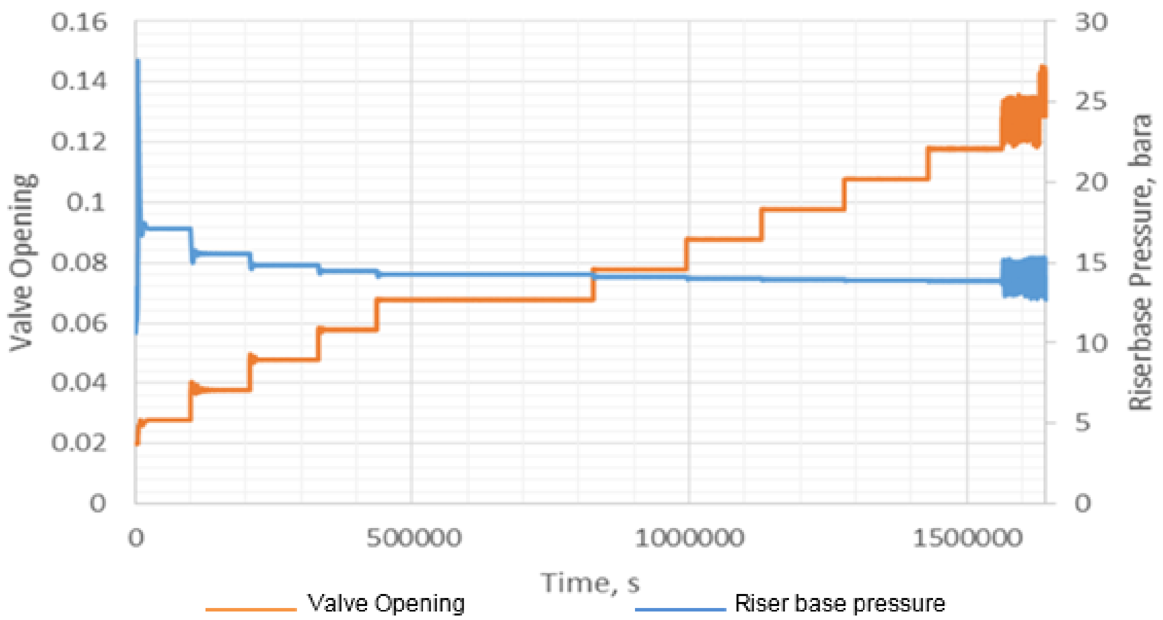


Figure 6-11 Simulation result for riser base pressure slug control with zero stroke time

6.5.2.4 Stroke time implementation

Having used the ISC to stabilize the system at an open loop unstable flow region for a zero stroke time effect, the significance of an increased stroke time on the control performance would be assessed next. For stroke time < 200 s applied to the system, the stability of the system was kept intact. However for a stroke time > 200 s the control performance begin to deteriorate.

Figure 6-12 shows the riserbase pressure response of the system in closed loop with a stroke time of 250 seconds. It could be observed that at a valve opening of 3 % valve which represented stable flow in open loop, there exist a continuous oscillation in the riser base pressure trend, representing instability or unstable flow. This signifies that truly stroke time (time delay) in actuation systems can degrade or adversely affect the performance of a controller as shown in Figure 6-12.

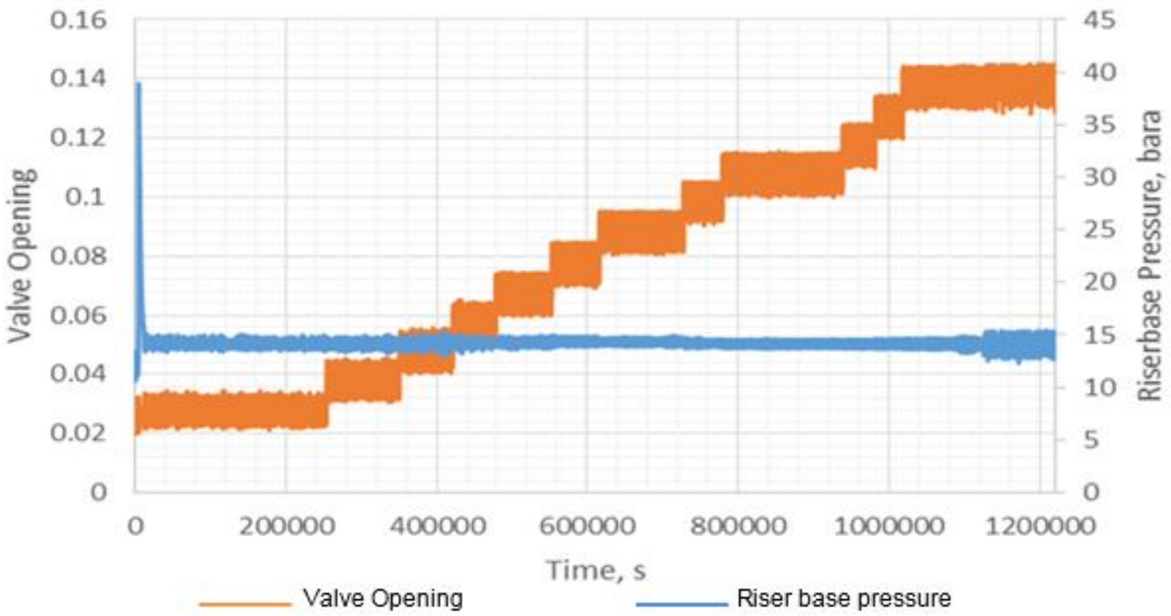


Figure 6-12 Simulation result for riser base pressure slug control with the time delay effect

6.5.2.5 Modified Smith predictor implemented with ISC for slug control with large valve stroke time

To improve the response of the system with large time delay, the modified Smith predictor was implemented with the ISC to control the instability induced by the stroke time in the choke valve. The model of the system without a delay time needs to be obtained. Due to the complexity of pipeline riser systems (U-shape riser), a mechanistic model might not be suitable to obtain the system model. Hence, a relay feedback shape factor would be adopted to identify the process model of the pipeline riser system (linear model of the system).

6.5.2.5.1 System identification

Accuracy of model prediction for systems is of great importance since the ISC is dependent on several measurement signal weight. The complexity and unsteady nature of the process in open loop makes a prediction of the model very complicated. The predicted model can however be represented by an unstable system model with time delay which could be stabilized using a constant controller. However, not all unstable process model can be stabilized using a constant controller. Effective control strategies for unstable systems have been looked into by several researchers (Kaya, 2003, 2004; Albertos, Garcı and Ha, 2006).

The process model of the pipeline riser system was identified using relay feedback shape factor which would be discussed and presented in this section. The complexity of pipeline riser system however makes it appropriate to approximate the model of the unstable open loop riser system using relay based system identification approach. Figure 6-13 shows a block diagram of a basic relay auto tuned controller structure which was used for system identification.

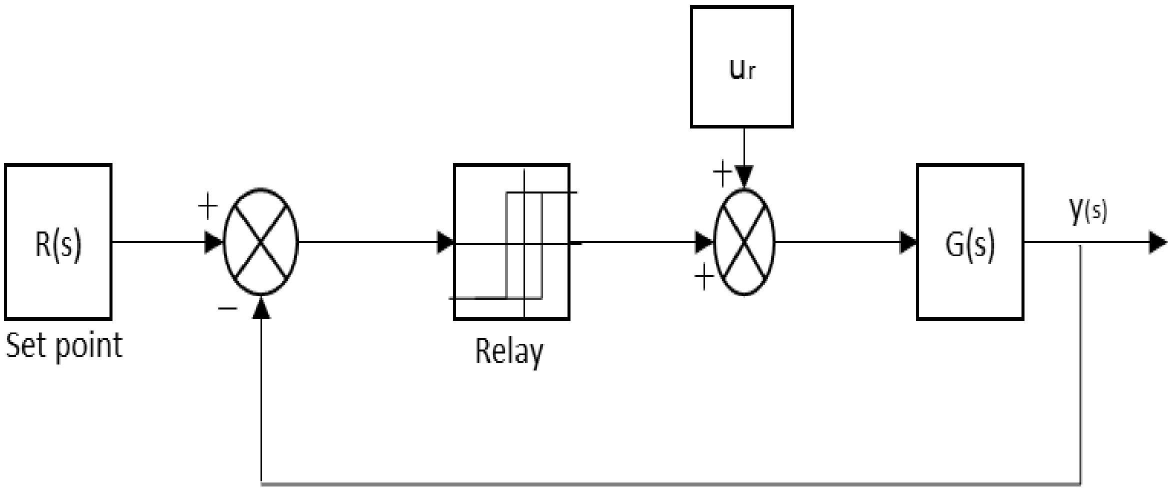


Figure 6-13 Basic relay auto tuned controller structure block diagram

The relay feedback shape factor would be adopted to help identify the pipeline riser model of the U-shape riser. To accomplish the identification procedure, the system was first connected to a relay. The reference valve opening (unstable open loop valve opening) was defined with the corresponding unstable equilibrium controlled variable. The on and off points of the relay was configured and defined. The on and off points were defined based on the output variable set point (reference valve opening) as it was set around it.

Beyond this, the relay feedback response was obtained once the system was run and the corresponding process parameters derived. Figure 6-14 shows the relay responses obtained from running the relay system identification set up of the U-shape riser.

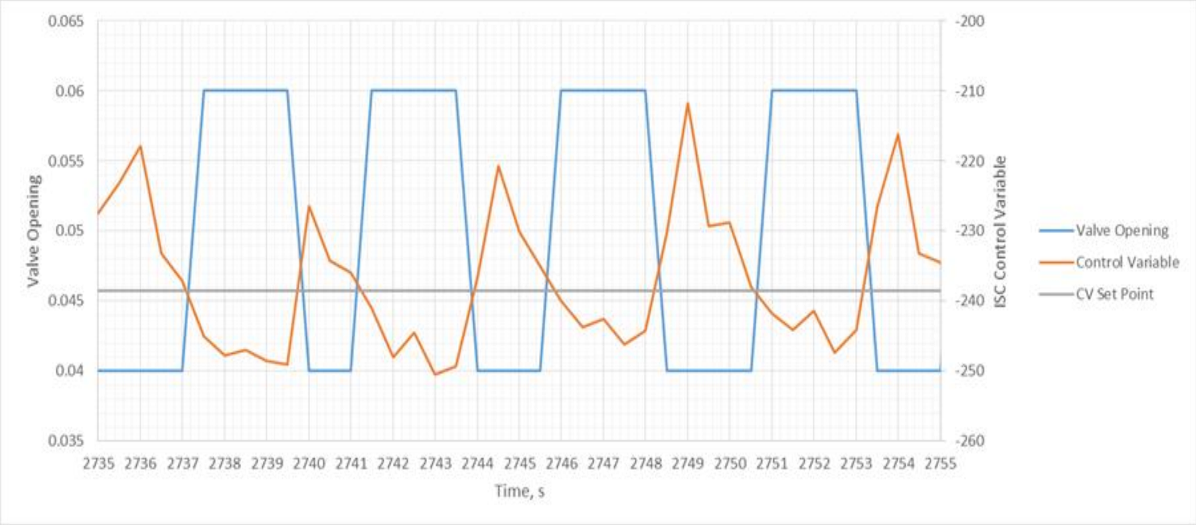


Figure 6-14 Basic relay auto tuned controller structure block diagram

The relay response as shown in Figure 6-13, having configured the relay as in Figure 6-12, shows that the system input (control variable) is increased whiles the output (valve opening) of the system decreases. As the output goes below the switch off point set in the relay, the input decreases such that the output increase again. This results to a relay feedback response which is cyclic in characteristics. The shape of the relay feedback response determines the process type, which can vary depending on the ratio of time constant to dead time and could be used to approximate the process. The period of the response oscillations was also determined from the relay feedback response.

Thyagarajan and Yu, (2002) stated that a relay feedback response with sharp edges at the peak amplitude can be approximated as a first order plus time dead time (FOPDT) but should however have sustained oscillations. As such the relay response should reach stationary oscillation in the first cycle. They stated that the shape information from the relay feedback test could be used to identify the model correctly. The transfer function of an unstable FODPT is given by (6-3).

$$G_{(s)} = \frac{K_p e^{-Ds}}{\tau s - 1} \qquad \textbf{(6-3)}$$

where τ is the time constant, Kp is the process gain, D is the dead time. However, for unstable processes, τ and Kp are given by (6-4) and (6-5) (Thyagarajan and Yu, 2003).

$$\tau = \frac{P_u/2}{\ln[1/(2e^{\frac{-D}{\tau}} - 1)]} \quad \textbf{(6-4)}$$

$$K_p = \frac{a}{h}(e^{\frac{D}{\tau}} - 1) \quad \textbf{(6-5)}$$

where Pu is period of oscillation, ‘a’ is the peak amplitude and ‘h’ is the increment in the input (relay magnitude). D, ‘a’ and Pu are all obtained from the relay shape feedback response. Also, for the limit cycle to exist, D/τ is required to be < In 2 else the ultimate period becomes infinite. Computing (6-3), (6-4) and (6-5), the transfer function of the U-shape riser pipeline system yields;

$$G_{(s)} = \frac{2581.87e^{-0.8s}}{1.16s - 1} \quad \textbf{(6-6)}$$

To stabilize the unstable time delayed system and improve the control performance, the modified Liu’s scheme (Figure 6-4) was adopted. Simplifying Figure 6-4 yields,

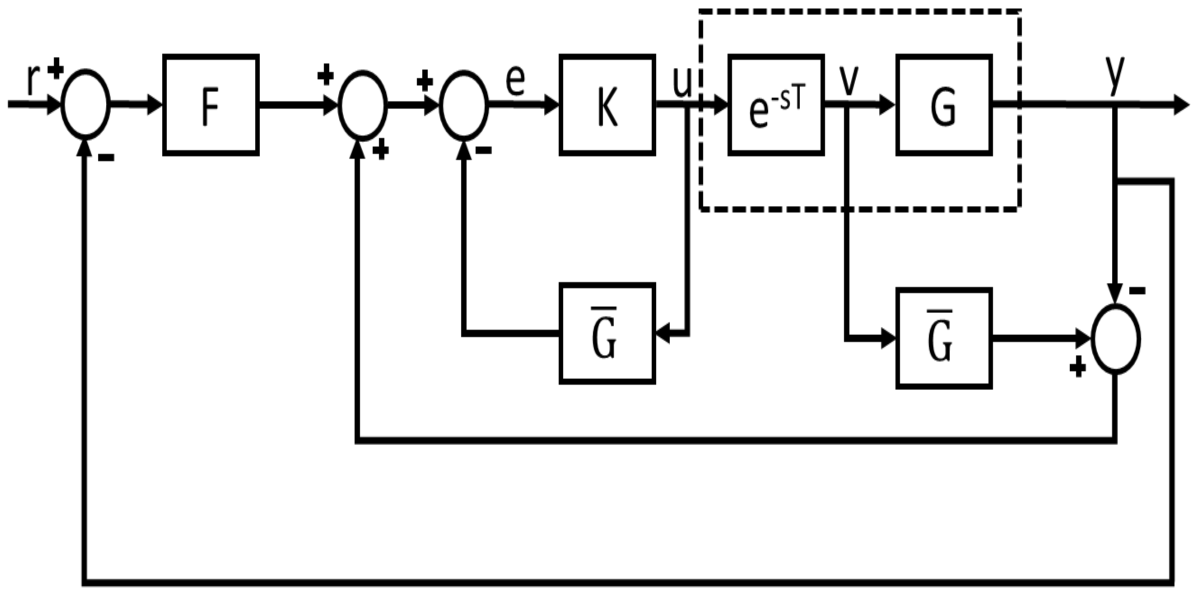


Figure 6-15 Simplified modified Liu’s scheme for variable delay

The scheme shows superiority at rejecting load disturbances which also renders the unstable model (overall model) stable in closed loop. For this case, two controllers were designed to stabilize the closed loop system whilst rejecting any load disturbance, F and to stabilize the plant model in open loop, K. Again, the ISC acts as a disturbance rejection in the model. This was of much importance for stabilizing unstable system with large stroke time. For this process, the two controllers were taken as 0.00039>K>0.00076 and

$$F = \frac{-0.4849s^2+0.81213s+1}{0.0000004s^4+0.000121s^3+0.0123s^2+0.43s+1}.$$

However the system convergence was not achievable. The stability range or gain for several process model tends not to be effective or failed even though due processes are followed in determining such controller parameters, thus no available gain to stabilize the process model. This could be associated with the differences in the process time delay and the time constant of the process model. The difficulties and challenges in controlling unstable systems with time delays are well recognized by many researchers (Huang and Chen, 1997). The challenge comes from the conflicting requirements where stabilizing control requires larger gain and quick response whilst the largest control gain and bandwidth have to be limited due to existing time delays. The approximation of the process is good for design but for simulation, it is better to use the true time delay. The difficulty in controlling the system could be due to the model having a relatively large time delay. Thus, the control difficulty depends on the relative size between the unstable time delay and the time delay hence needs to be investigated. A controllability analysis of the derived unstable first order system plus time delay would be assessed next.

6.5.2.6 Controllability analysis of an unstable first order system plus time delay (UFOPDT)

This section addresses the controllability analysis of an unstable first order system plus time delay (UFOPDT). Assuming a simple unstable transfer function given by

$$G_{(s)} = \frac{e^{-Ds}}{\tau s - 1}$$

and controller gain, K, the closed loop transfer function can be obtained as,

$$G_{p(s)} = \frac{Ke^{-Ds}}{\tau s - 1} \qquad \textbf{(6-7)}$$

Then using the Routh table, we can estimate and conclude on what the range of K is to stabilise G, and the conditions under which no K can stabilise G.

6.5.2.6.1 Unstable first order plus time delay system

The poles and zeros of a system transfer function plays a significant role in the stability of the system. Generally, physically realizable control systems must have a number of poles greater or equal to the number of zeroes. Systems satisfying this condition are referred to as proper systems. Thus,

$$G_{(s)} = \frac{N_{(s)}}{D_{(s)}}$$

where $D_{(s)}$ and $N_{(s)}$ are both polynomials with the polynomial order of $D_{(s)}$ greater than or equal to that of $N_{(s)}$. As 's' approaches a zero, the numerator of the transfer function and the transfer function in total, approaches the value 0. Also as 's' approaches a pole, the denominator of the transfer function approaches zero and the transfer function approaches infinity.

Stability of a system relates to its response to a disturbance or its input. Systems that maintains a constant state unless affected by an external action and returns to a constant state when the external action is removed is considered stable. The degree or extent of system stability, the performance of a system in steady state and system transient response makes stability analysis a necessity. This section outlines a stability analysis of an unstable first order plus time delay system. For an unstable first order plus time delay system given by;

$$G_{p(s)} = \frac{Ke^{-Ds}}{\tau s - 1} \qquad \textbf{(6-8)}$$

Using Pade approximations, the time delay could be expressed as;

$$e^{-Ds} \cong \frac{1 - \theta_1 s + \theta_2 s^2 + \cdots + \theta_n s^n}{1 + \theta_1 s + \theta_2 s^2 + \cdots + \theta_n s^n} \approx \frac{1 - \theta_1 s}{1 + \theta_1 s}$$

But $\theta_1 = D/2$. Hence **(6-8)** transforms to become,

$$G_{p(s)} = \frac{K}{\tau s - 1} \frac{(1 - \theta_1 s)}{(1 + \theta_1 s)} \qquad \textbf{(6-9)}$$

From **(6-9)**,

$$G_{p(s)} = \frac{K(1 - \theta_1 s)}{(\tau\theta s^2 + \tau s - \theta_1 s - 1)} \qquad \textbf{(6-10)}$$

From the characteristic equation $(\tau\theta s^2 + (\tau - \theta_1 - K\theta_1)s + (K - 1))$ and using the Routh table,

S²	$\boldsymbol{\tau\theta}$	$\boldsymbol{(K-1)}$
S¹	$\tau - \theta_1 - K\theta_1$	0
S⁰	$(K-1)$	

$$(K-1) > 0 \gg K > 1$$

$$\tau - \theta_1 - K\theta_1 > 0 \gg K < \frac{\tau - \theta_1}{\theta_1}$$

This implies that, the UFOPDT can only be stabilised using a basic controller gain values given as,

$$1 < K < \frac{\tau - \theta_1}{\theta_1} \qquad \textbf{(6-11)}$$

From **(6-11)**,

$$\theta_1 < \tau - \theta_1$$

$$2\theta_1 < \tau \qquad \textbf{(6-12)}$$

But $\theta_1 = D/2$

From **(6-12)**,

$$D < \tau \qquad \textbf{(6-13)}$$

From (6-13), it could be deduced that for an unstable first order system plus time delay to be stabilizable, the process time delay must be less than the time constant of the process. Hence this explains why the unstable first order system plus time delay found from (6-6) could not be stabilized. Thus, at any sampling point the time delay resulting from the stroke time of the valve makes the time delay greater than 1.16, which is the time constant.

6.5.2.7 Effect of zeros or poles on UFOPDT

As outlined above, the locations of the poles and the values of the real an imaginary parts of the pole determines the response of the system. Addition of poles to any system transfer function has the effect of pulling the root locus to the right making the system less stable while an addition of zeros to the system transfer function has the effect of pulling the root locus to the left, making the system more stable. Hence the addition of zeros (stable or unstable) to the system transfer function (UFOPDT) is much desirable and would be investigated next since the addition of poles makes the system less stable.

6.5.2.7.1 Effect of stable zeros on UFOPDT

Generally, an addition of zeros to the system, shifts the root locus towards the real axis, thus left of the s-plane. Similarly, the breakaway points shift to the left while the gain margin is increased. Thus, the system becomes less oscillatory, translating to a reduction in the settling time hence an increase in the stability of the system. For an UFOPDT with additional stable zero is given by;

$$G_{p(S)} = \frac{1}{(\tau s - 1)} \frac{(-\theta s + 1)}{(\theta s + 1)} (fs + 1)$$

$$G_{p(S)} = \frac{-\theta f s^2 + fs - \theta s + 1}{\tau \theta s^2 - \theta s + \tau s - 1}$$

Using Routh criterion, the characteristic equation of the system transfer function is given by;

$$(\tau\theta - k\theta f)s^2 + (\tau - \theta - \theta k + fk)s + (k - 1)$$

From Routh table,

$\boldsymbol{s^2}$	$\boldsymbol{\theta\tau - k\theta f}$	$\boldsymbol{k - 1}$
$\boldsymbol{s^1}$	$\tau - \theta - \theta k + fk$	0
$\boldsymbol{s^0}$	$k - 1$	

To find the gain range that makes the system stable,

$$\tau - \theta + \theta k + fk > 0 \qquad\qquad k - 1 > 0$$

$$k < \frac{-\theta + \tau}{\theta - f} \qquad\qquad k > 1$$

$$1 < k < \frac{-\theta + \tau}{\theta - f}$$

$$1 < \frac{-\theta + \tau}{\theta - f}$$

$$\tau > 2\theta - f$$

But *θ=D/2*

Hence

$$\tau > D - f$$

Deductions made from this analysis shows that for an UFOPDT with additional stable zero to be stabilizable, the time delay must be less than the system time constant plus the negative inverse of the root of the additional stable zero.

6.5.2.7.2 Effect of unstable zeros on UFOPDT

Similar to the effect of the stable zeros on UFOPDT, an UFOPDT with additional unstable zero is given by;

$$G_{p(S)} = \frac{1}{(\tau s - 1)} \frac{(-\theta s + 1)}{(\theta s + 1)} (fs - 1)$$

$$G_{p(S)} = \frac{-\theta f s^2 + fs + \theta s - 1}{\tau \theta s^2 - \theta s + \tau s - 1}$$

Using Routh criterion, the characteristic equation of the system transfer function is given by;

$$(\tau\theta - k\theta f)s^2 + (\tau - \theta + \theta k + fk)s + (-k - 1)$$

From Routh table,

$\boldsymbol{s^2}$	$\boldsymbol{\theta\tau - k\theta f}$	$\boldsymbol{-k - 1}$
$\boldsymbol{s^1}$	$\tau - \theta + \theta k + fk$	0
$\boldsymbol{s^0}$	$k - 1$	

To find the gain range that makes the system stable,

$$\tau - \theta + \theta k + fk > 0 \qquad\qquad -k - 1 > 0$$

$$k > \frac{\theta - \tau}{\theta + f} \qquad\qquad k < -1$$

$$\frac{\theta - \tau}{\theta + f} < k < -1$$

$$\frac{\theta - \tau}{\theta + f} < -1$$

$$\tau > 2\theta + f$$

But $\theta = D/2$

Hence

$$\tau > D + f$$

Similarly, it could be deduced that for an UFOPDT with additional unstable zero to be stabilizable, the time delay must be less than the system time constant minus the negative inverse of the root of the additional unstable zero.

6.6 Summary

Time delay through valve stroke time effect on pipeline riser configuration has been investigated. It was established that the time delay effect on processes reduces the performance especially in the control of slug flow. An established and known technique initiated instituted for measuring time delay in systems, was used to improve this performance. Thus, the Smith predictor model was deployed to help improve the control performance and in effect increase oil production in the hydrocarbon industry.

Most Smith predictor models requires an estimation of the time delay but the novel modified Smith predictor proposed does not require this estimation to be made. This is because the slug control system has a varying time delay, a resultant of the source of the delay. Thus, the novel modified Smith predictor renders the estimation of time delay irrelevant for input time delay and could be used to deal with processes with varying time especially caused by the valve stroke time. For specific case studies, the modified Smith predictor when applied to a slug control problem reduce the riserbase pressure significantly which in effect increased the production of the system. More robust controller was investigated to enhance the benefits derived when using the modified Smith predictor for systems with large time delay.

The modified Smith predictor was subjected to investigation for stable and unstable systems. The importance of proper estimation of models used in control was analysed. From the controllability analysis, it was established that for an unstable first order system with time delay to be stable, the time delay in the system must be less than the time constant of that same system. Thus, from a controllability analysis on unstable system proved that for an unstable system to be controllable, the time delay must be relatively smaller than the time constant

7 CONCLUSION AND FURTHER WORK

7.1 Introduction

This work has carried out an extensive study on advancing the inferential slug control technique while regularising unstable slug flow conditions and the use of a pseudo spiral tube, a passive slug mitigation technique. The conclusions drawn from the study outlined in this work are summarised and highlighted in this chapter.

7.2 Conclusion

In Chapter 2, a comprehensive review of multiphase flow instabilities with special interest in slug flow and its control techniques / application, limitations and challenges was highlighted. On the broader sense both passive slug control and active slug control techniques were addressed. Several slug control methods such as riserbase gas injection, flow conditioning, topside choking, gas re-injection and automatic control using the riser topside valve as the manipulating variable have been discussed. The use of several active slug control techniques, one of the most promising active slug mitigation techniques, and its implication on production were also studied. In spite of the advancement in the active slug control methods, which gave birth to the inferential slug control technique, it appears controller robustness is still far off. Some advancement and improvement done on the ISC technology have been discussed. The deficiency of appropriate information on the performance of current slug control system (ISC technology) is recognised as an obstacle to determining the path for further improvement. Extensive volume of work is still necessary to gain satisfactory awareness and understanding of unstable slug control using the ISC technology. In conclusion, this research fills the gap of understanding the dynamics of unstable flow in different pipeline riser systems, identifying the geometrical effect of pipeline riser configuration or layout on unstable slug flow and advancing and optimizing the inferential slug control technology to be deployed on several offshore riser facilities and configurations.

The extension of the ISC technology to be used on a U-shape riser pipeline configuration was reported in Chapter 3. Major findings including a study of the flow dynamics in a 2 inch U-shape riser system and the initiation of flow instabilities in the U-shape pipeline riser system outlining the impact of the riser geometry has been reported. The awareness of these flow dynamics and unstable regions could help with the choice of riser topside measurement signals used in the ISC and the choice of suitable and effective control strategy. The use of extra measurement signals from remote platform as control variables to stabilise unstable slug flow has been described. The downcomer top pressure has been identified as a good measurement signal for slug control, however cannot be relied upon exclusively for all flow conditions. This is because slugging flow does not always exist in both the downcomer and the riser system. Thus, there are some flow conditions which might exhibit slugging in the riser but the slugging characteristics might be existing in the downcomer. Hence could be deployed together with other measurement signals in the inferential slug control technology. The contribution of these singular measurement signals to the systematic design while improving the controller robustness of the ISC in the quest to advance the ISC technology has been explained. The knowledge on the systematic design of the ISC technology could be very valuable in the further advancement and deployment of this scheme.

The extension of the ISC technology to be used on an S-shape riser pipeline configuration was presented in Chapter 4. The dynamics of flow through the S-shape riser, slug envelope and slug flow stability criteria in the S-shape riser configuration was reported. The S-shape riser pipeline configuration dip effect on flow instabilities and behaviour was presented outlining the impact of geometry on unstable flow. The ISC technology was advanced, optimized and implemented on both small and large diameter S-shape riser pipeline configuration through experiments and numerical tools. Specific case studies showed increased choke valve opening in closed loop translating to an increased system throughput.

In Chapter 5, the unstable slug mitigation potential of the pseudo spiral tube (wavy and spiral configuration), a passive slug mitigation technique, was reported. The pseudo spiral tube (PST) through experiment was proven to have some slug attenuation prospective when implemented at the topside of the S-shape riser pipeline system although the effectiveness of the slug mitigation depends on the flow conditions. Even though the study was a preliminary investigation, the wavy pipe section coupled with the riser system was proven to have slug attenuation benefits for all the flow conditions (low and high flowrates) assessed and the ability to enhance the parameter variation technique. On the other hand the spiral pipe section coupled with the pipeline riser system was shown to only be effective for slug mitigation for relatively high flowrates whilst it’s potential deteriorate for low flowrates. The knowledge of the attenuation potential possessed by the PST in mitigating slug could be valuable in the advancement and extension of the ISC technology. Extensive work is still required to gain sufficient idea and in-depth knowledge on the mitigation potential of the PST.

Chapter 6 presented the advancement and extension of the ISC technology, an active slug control strategy that uses measurement signals from the topside of the riser, to deal with systems with time delay as a result of large valve stroke time. The Smith predictor model was specially modified not just to deal with time delay but also to deal with variable time delay in both stable and unstable systems. Specific case studies showed that the modified Smith predictor model proved to correct the degraded controller performance imposed on the system by large valve stroke time hence a regularised and optimized flow through the riser pipeline system.

7.3 Contribution of this research work

This thesis has added to knowledge in the following amongst others

- New measurement signal as control variable for slug control in pipeline riser systems has been revealed
- Systematic design, advancement, optimization and implementation of the ISC technology on different riser configurations has been established rendering the controller more robust in its operation
- The impact of pipeline geometry on flow instabilities in different pipeline riser configurations has been outlined
- The slug mitigation potential of the PST installed at the topside of the pipeline riser system has been established

7.4 Future Work

This work has presented an inclusive investigation of the systematic approach to stabilising unstable slug flow using the ISC technology whiles maximising the throughput in several riser pipeline riser configuration systems and enhancing flow stability using the PST installed at the topside of a pipeline riser system. However to achieve or further improve to enhance the outcome of this research, some further works would be a necessity.

The ISC technology has been advanced and implemented on several pipeline riser configurations using topside measurement signals combined linearly but could be explored using a non-linear combination of these signals to advance the technology further. This however would be pertinent to enlightening the concept and its indulgent and advancing the method for a much more complicated industrial systems.

More work could be done to develop mechanistic model for the S-shape and U-shape riser pipeline system to ease the investigation of the controllability of different and relevant measurement signals on the different riser configuration. This will aid further improvement of the stability analysis technique for robust slug controller investigation and design.

The S-shape riser pipeline configuration coupled with the wavy pipe section showed over the S-shape riser pipeline configuration coupled with the spiral pipe section to be the best in terms of slug mitigation potential for all the flow conditions tested whilst the system coupled with the spiral pipe section showed superiority over the plain system (without PST) when the was a large flowrate through the system. However, only one configuration each of the wavy and spiral pipe section was used hence a comprehensive study would be required to reveal the effective flow conditions corresponding to different PST (wavy and spiral) geometry design.

The PST (wavy and spiral pipe section) can also be integrated with the ISC to further advance the technology. The controllability of relevant signals that could be available from the PST section could be investigated and implemented in the ISC to improve the controller robustness.

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APPENDICES

Appendix A Experimental Facility, Operating conditions and Procedure

The overview of the multiphase facility, means of retrieving and analysing data and operating conditions is documented in this section of the appendix.

A.1 Multiphase flow facility

The experimental campaign from this thesis uses the multiphase (three-phase (air-water-oil)) flow facility housed in the Cranfield University Process systems engineering Laboratory, together with other experimental facilities. The facility is basically classified in four main sections as shown in Figure_ A-1.

Fluids (air – water mixture) through any of the flow loops is discharged into a horizontal three-phase separator where they are separated. The liquid level and the separator pressure are automatically controlled by level and pressure controllers respectively. The separated air is discharged into the atmosphere and the water is pumped back to the water tank via a coalescer.

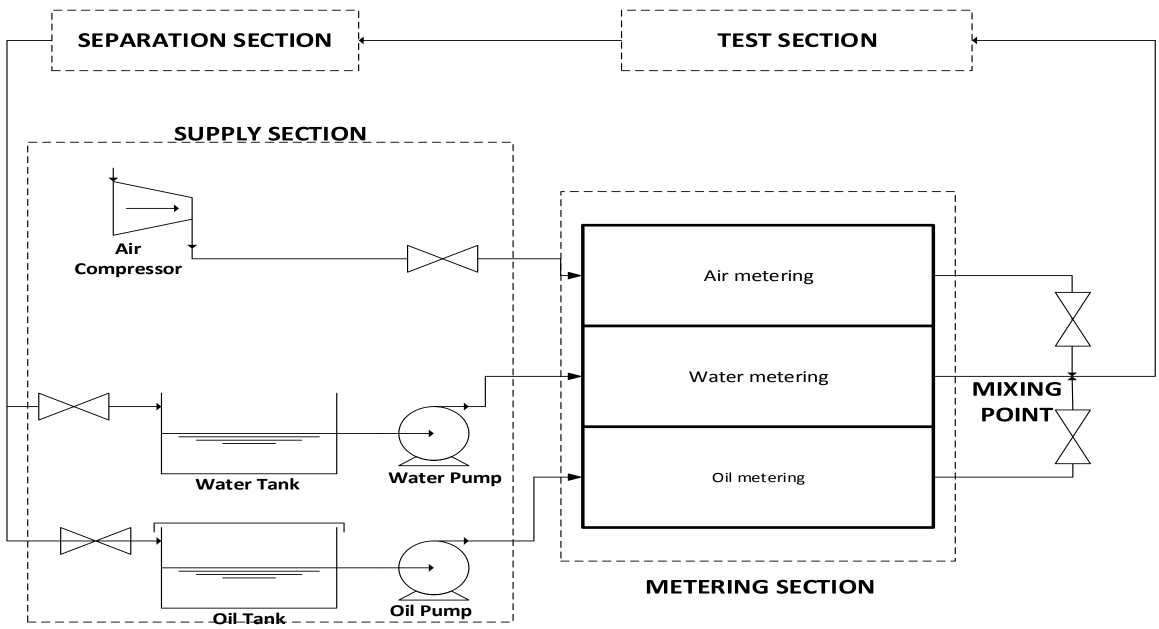


Figure_ A-1 Block flow and simple process diagram of the flow loop

A.1.1 Separation Area

The separation area deals with the separation of the different phases in the process. Separation of phases is mainly done by the first stage separator (two phase separator) and the three phase separator. The fluids are scrubbed in their individual coalescers before conveyed back into their storage tanks and the air is released into the atmosphere.

First Stage Separator

The two phase separator is located at the top of the 2 inch riser which is used mainly to separate gases and liquids. It is the first stage where separation begins and the separated fluids are transported through different pipelines into the three phase separator. It is also used to check the pressure level of the multi-phase separator and in order not to over fill this separator; a pressure controller is used to check the pressures in this separator which is displayed on the DeltaV system.

Three Phase Separator

The separator is made of an 11.12 m3 horizontal multiple phase separator (three phase) which deploys a gravity mechanism for the separation of phases. The three phase separator is regulated for better operation by the operating pressure, oil/water interface and gas/liquid interface which are controlled by a pressure gauge, and two level controllers respectively. Liquids after the separation process are sent to their respective tanks through a coalescer whiles air is exhausted into the atmosphere. Control valves are deployed to split the returning liquids thereby retaining some liquids to help maintain the liquid/liquid and gas/liquid interfaces level which is displayed in the Delta V system.

A.1.2 Test Area

The test area is mainly made of a 2 inch flow loop system. Most of the pipe making up the 2 inch loop is made of stainless steel and three Perspex glass (transparent acrylic pipe) inter-juncture. The loop is of a 20 barg rating but is limited to the maximum pressure of the air compressor which is at 7.5 barg. The liquid supply is channelled at bottom of the riser serving as the well while the air supply could either be supplied into the system by a junction that either mix with the liquid before the horizontal section of the system or at the base of the riser mainly for gas injection experiments. The operations of the test area are controlled and manipulated using the Delta V system.

A.1.3 Flow Metering Area

The flow rates of the liquid and the gas are controlled with different flow meters. The liquid flows are measured using two types of flow meters mainly a magnetic flow meter and a Coriolis flow meter. The Coriolis flow meters are used to measure smaller flow rates whiles the electromagnetic flow meters are used to measure large flow rates. The air flow rate is also measured with the help of two Rosemount mass Probar flow meters with one measuring for low flow rates and the other for large flowrates. The flowrate of the fluids are controlled with the help of control valves whose percentage openings corresponds to the flow rate supplied into the system.

A.1.4 Supply system

The supply system of the rig is mainly three but two are of interest to this project since this project utilizes only the water and air supply systems. This section is also known as the valve manifold area.

Water supply system

The water supply system is basically made of the water tank and a centrifugal pump. A 12 m3 capacity tank open to atmospheric pressure serves as the source and where it is return for the rig facility. It is located within the facility and is fitted with baffles to prevent sand from the returning liquid from entering the centrifugal pump and also enables it to settle beneath the tank. The centrifugal pump has a capacity of 40 m3/h fitted with a fixed speed drive at 30 Hz.

Air Supply System

A screw compressor is used in the supply of compressed air into the test rig. It has a capacity of 400 Sm3/h with a pressure of 10 barg as its maximum. There are two of them which are connected in parallel and both can be operated to produce any maximum flow required by the system. An air cooler and filter is deployed to enhance good quality air and at the right temperature by stripping all debris and condensates contained in the air before entering the flow meters.

A.2 Flow Data Acquisition System

Output flow data from the 2 inch experimental set up are obtained from a Delta-V Supervisory, Control and Data Acquisition (SCADA) system. The Delta-V system remotely controls the process operations and the overview of the process could be also monitored from this system. The output values from the process is recorded at a frequency of 1 Hz. Fieldbus and PROFIBUS serves as an interfacial connection between the physical test facility and the Delta-V system.

A.2.1 Instrumentation

There is several instrumentation available in the PSE laboratory in Cranfield University but with the focus of this experiment in mind (air-water test), the main ones specific to this experiment includes flow meters, pressure transducers and temperature transducers for which their details follows;

Flow meters

The 2 inch rig has two flow meters, an electromagnetic flow meter and a Coriolis flowmeter. An ABB K280/0 AS model electromagnetic flow meter located downstream the centrifugal pump to measure the water flowrate from the water supply system in higher flowrates (1 kg/s upwards) whiles a Foxboro Coriolis flow meter is used to measure water flowrates below 1 kg/s of water flowrate from the water supply source.

The electromagnetic has an accuracy of +/- 0.5 % within a range of 0-2 m3/h at a pressure of 9 bar. An Atlas Copco compressors used in the supply of compressed air into the test rig has a capability of 400 SCM/h at a pressure up to 10 barg. Two Rosemount Mass Probar flow meters of 0.5inch and 1 inch were used to measure airflow rates up to 120 Sm3/h and 4250 Sm3/h respectively.

Temperature Transducers

T – type thermocouples temperature transducers with a rangeability of –200 0C to 400 0C is installed on the flow line to determine the temperature of flow at a response frequency of 1 Hz. This thermocouples have an uncertainty of ±1% of full scale .The temperatures of oil, water and air are measured upon entry into the facility. The air temperature is kept at its required temperature with the help of an air cooler.

Pressure Transducers

To prevent interference of flow, flush mounted pressure transducers are mounted on both the air line, riser base and on the vertical section of the riser. The pressure transducer were mounted to measure the pressure on the flow line. The safe measuring range of these pressure transducers is from 0 barg to 6 barg with an uncertainty ±0.25% of full scale. Even though pressures beneath the seabed are not readily available, the riserbase pressure is of great importance to this research work.

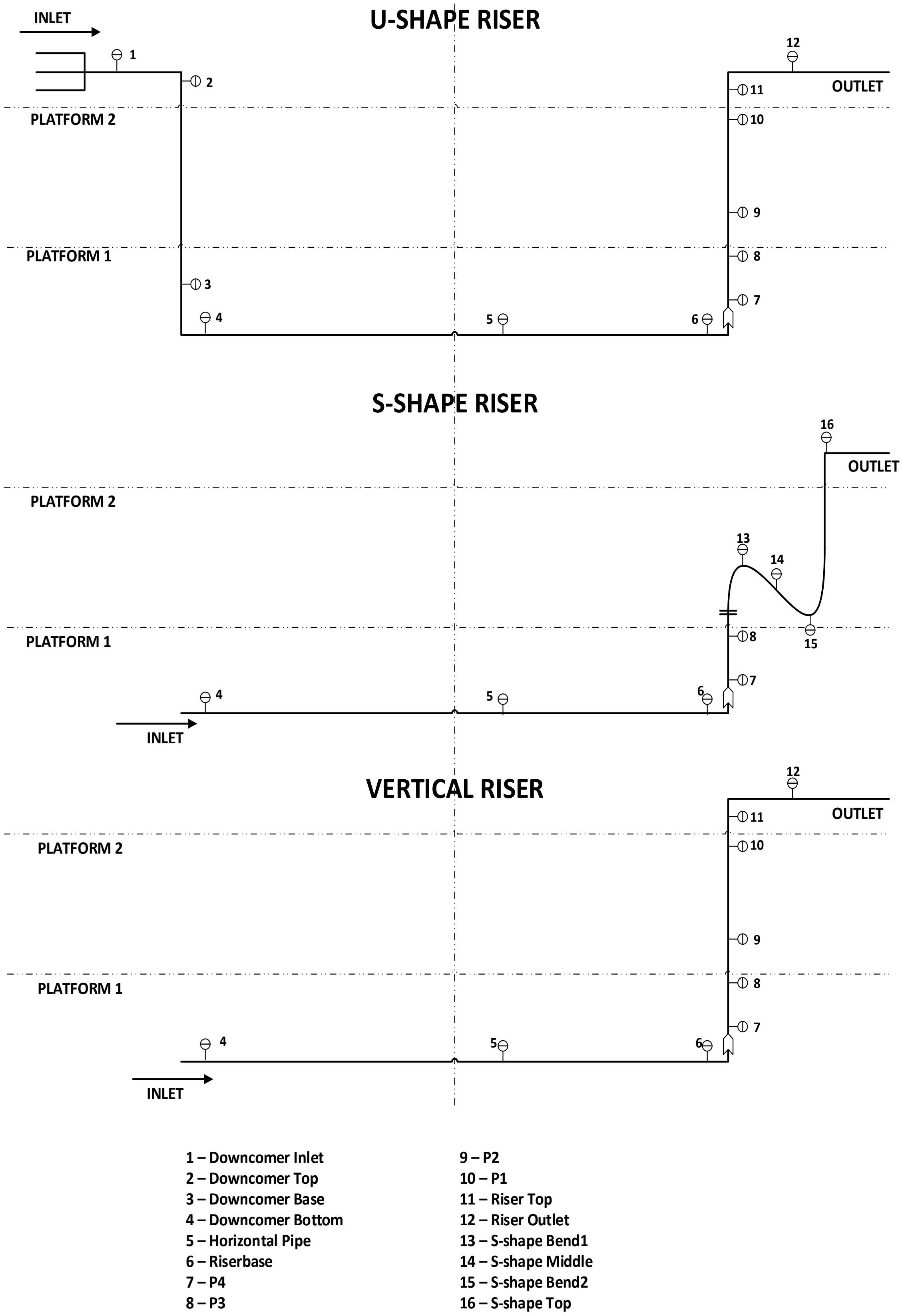


Figure_ A-2 Pressure tapings and signal acquisition from the 2" flow loops

A.3 U-shape experimental test matrix

Liquid flowrate kg/s					0.1	0.5	1	1.5	2	3	3.5	5
Liquid volumetric flowrate, m^3/s					0.00010	0.00050	0.00100	0.00150	0.00200	0.00301	0.00351	0.00501
Superficial liquid velocity, m/s					0.05	0.25	0.49	0.74	0.99	1.48	1.73	2.47
Area of Pipe, m^2	Gas Flowrate sm3/h	Gas Flowrate nm3/h	Superficial gas velocity, m/h	Superficial gas velocity, m/s								
0.0020271	5.00	2.42	1191.93	0.33	0.40	0.36	0.34	0.33	0.32	0.30	0.29	0.27
0.0020271	10.00	4.83	2383.86	0.66	0.93	0.81	0.71	0.70	0.66	0.59	0.55	0.47
0.0020271	20.00	9.66	4767.72	1.32	1.97	1.73	1.54	1.45	1.34	1.15	1.07	0.85
0.0020271	30.00	14.50	7151.58	1.99	3.00	2.68	2.41	2.19	2.00	1.67	1.53	1.20
0.0020271	50.00	24.16	11919.30	3.31	5.17	4.56	4.00	3.54	3.16	2.57	2.33	1.78
0.0020271	70.00	33.83	16687.02	4.64	7.38	6.37	5.43	4.73	4.16	3.31	2.99	2.28
0.0020271	100.00	48.32	23838.60	6.62	10.57	8.87	7.26	6.19	5.38	4.24	3.81	2.89
0.0020271	120.00	57.99	28606.32	7.95	12.64	10.35	8.34	6.99	6.04	4.73	4.03	3.40
0.0020271	150.00	72.48	35757.90	9.93	15.59	12.29	9.65	8.00	6.82	5.32	4.31	3.95
0.0020271	200.00	96.65	47677.20	13.24	19.88	15.13	11.73	9.83	7.82	6.24	5.74	4.57
0.0020271	250.00	120.81	59596.49	16.55	22.98	16.93	13.02	10.84	8.60	7.73	6.61	5.62
0.0020271	300.00	144.97	71515.79	19.87	25.33	18.21	14.03	11.74	9.52	8.21	7.31	6.73

Appendix B Large diameter riser pipeline system in OLGA

B.1 Pipeline configuration and geometry

The industrial U-shape riser model adopted in this thesis is made up of a 48 m downcomer, 1832 m horizontal section and a riser length of 51 m. The entire U-shape riser system (downcomer, horizontal pipe and riser) has a uniform diameter of 0.289 m. The geometry of the U-shape riser pipeline system is shown in Figure_B-1.

Before the numerical simulation of the model, some activities (pre-processing) are carried out. Some of such pre-processing activities include the definition of the fluid file, pipeline discretization and their likes. The fluid file, which defines the composition of the fluid, was generated using PVTsim. Details of the fluid file could not be presented in this thesis due to the sensitivity of the data. However, the pipeline geometry data showing the materials, heat transfer coefficient, roughness and the sectioning of the U-shape riser pipeline system is specified in Figure_B-2.

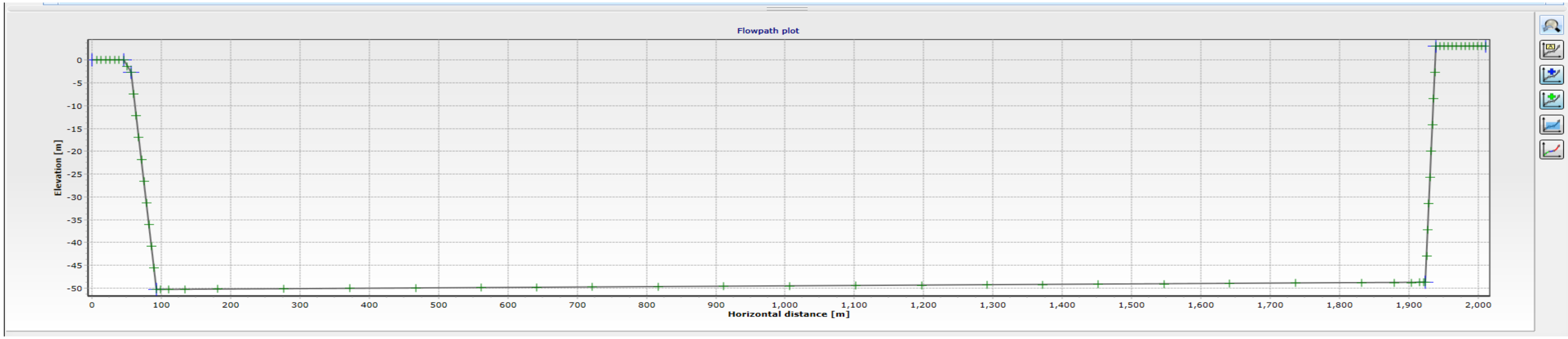
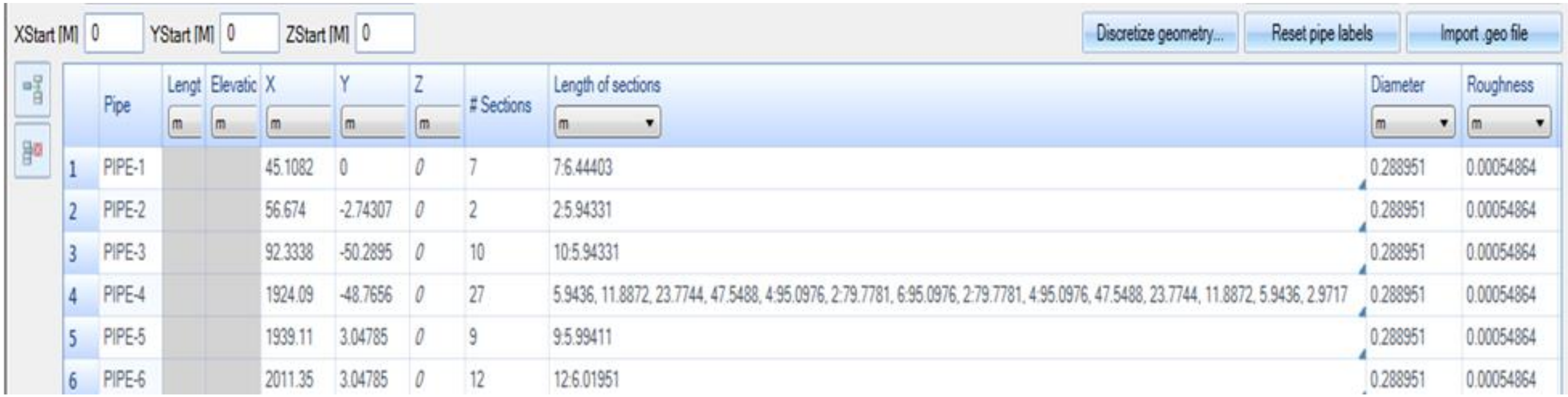


Figure_ B-1 Large diameter U-shape pipeline configuration in OLGA



XStart [M] 0 YStart [M] 0 ZStart [M] 0 | Discretize geometry... | Reset pipe labels | Import .geo file

	Pipe	Lengt (m)	Elevatic (m)	X (m)	Y (m)	Z (m)	# Sections	Length of sections (m)	Diameter (m)	Roughness (m)
1	PIPE-1			45.1082	0	0	7	7:6.44403	0.288951	0.00054864
2	PIPE-2			56.674	-2.74307	0	2	2:5.94331	0.288951	0.00054864
3	PIPE-3			92.3338	-50.2895	0	10	10:5.94331	0.288951	0.00054864
4	PIPE-4			1924.09	-48.7656	0	27	5.9436, 11.8872, 23.7744, 47.5488, 4:95.0976, 2:79.7781, 6:95.0976, 2:79.7781, 4:95.0976, 47.5488, 23.7744, 11.8872, 5.9436, 2.9717	0.288951	0.00054864
5	PIPE-5			1939.11	3.04785	0	9	9:5.99411	0.288951	0.00054864
6	PIPE-6			2011.35	3.04785	0	12	12:6.01951	0.288951	0.00054864

Figure_ B-2 U-shape pipeline properties and geometry data

B.2 Control model

Figure_B-3 shows the structural configuration of the ISC model in OLGA. This model was used to stabilise the U-shape riser system at a relatively higher valve opening in closed loop in comparison to the system in open loop.

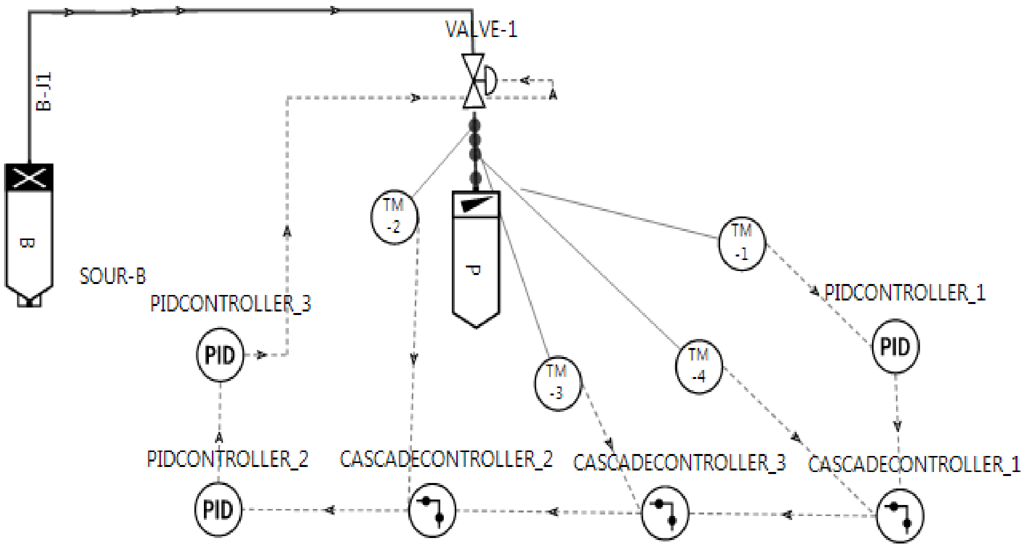


Figure_ B-3 ISC on the U-shape riser model configuration in OLGA

Appendix C Controller design

In controller design, the controller parameters could be determined using the Ziegler Nichols tuning table. Depending on the type of system in question, either the open-loop tuning method or closed-loop tuning method could be adopted. From this thesis however the open loop tuning table (Table_ C-1) was used to determine the PI parameters for controlling slug flow in the U-shape riser using either the downcomer top or riser top pressure as the control variable.

Table_ C-1 Ziegler Nichols open loop PID tuning table

Controller	$\boldsymbol{K_c}$	$\boldsymbol{\tau_I}$	$\boldsymbol{\tau_D}$
P	$\frac{\tau}{\tau_d K_P}$		
PI	$\frac{0.9\tau}{\tau_d K_P}$	$3.3\tau_d$	
PID	$\frac{1.2\tau}{\tau_d K_P}$	$2\tau_d$	$0.5\tau_d$

Table_ C-2 Ziegler Nichols closed loop PID tuning table

Controller	$\boldsymbol{K_c}$	$\boldsymbol{\tau_I}$	$\boldsymbol{\tau_D}$
P	$0.5K_u$		
PI	$0.45K_u$	$\frac{P_u}{1.2}$	
PID	$0.6K_u$	$\frac{P_u}{2}$	$\frac{P_u}{8}$