

CRANFIELD UNIVERSITY

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ACTIVE CONTROL OF HYDRODYNAMIC SLUG FLOW

Department of Offshore, Process Systems and Energy
Engineering

MSc by Research
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Supervisor: Dr. Yi Cao

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ABSTRACT

Multiphase flow is associated with concurrent flow of more than one phase (gas-liquid, liquid-solid, or gas-liquid-solid) in a conduit. The simultaneous flow of these phases in a flow line, may initiate a slug flow in the pipeline. Hydrodynamic slug flow is an alternate or irregular flow with surges of liquid slug and gas pocket. This occurs when the velocity difference between the gas flow rate and liquid flow rate is high enough resulting in an unstable hydrodynamic behaviour usually caused by the Kelvin-Helmholtz instability.

Active feedback control technology, though found effective for the control of severe slugs, has not been studied for hydrodynamic slug mitigation in the literature. This work extends active feedback control application for mitigating hydrodynamic slug problem to enhance oil production and recovery.

Active feedback Proportional-Integral (PI) control strategy based on measurement of pressure at the riser base as controlled variable with topside choking as manipulated variable was investigated through Olga simulation in this project. A control system that uses the topside choke valve to keep the pressure at the riser base at or below the average pressure in the riser slug cycle has been implemented. This has been found to prevent liquid accumulation or blockage of the flow line.

OLGA (olga is a commercial software widely tested and used in oil and gas industries) has been used to assess the capability of active feedback control strategy for hydrodynamic slug control and has been found to give useful results and most interestingly the increase in oil production and recovery. The riser slugging was suppressed and the choke valve opening was improved from 5% to 12.65% using riser base pressure as controlled variable and topside choke valve as the manipulated variable for the manual choking when compared to the automatic choking in a stabilised operation, representing an improvement of 7.65% in the valve opening. Secondly, implementing active control at open-loop condition reduced the riser base pressure from 15.3881bara to 13.4016bara.

Keywords:

Choking, multiphase, flow regime, feedback control, close-loop, open-loop, bifurcation map, OLGA

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LIST OF NOMENCLATURES

A_G =Area occupied by gas (m^2)

A_L = Area occupied by liquid (m^2)

A =Cross sectional area of the pipe (m^2)

C_D =Drag coefficient

D =Pipe inner diameter (m)

f =Darcy-Weisbach friction factor

g =Acceleration due to gravity (m/s^2)

h_L = Liquid height in the pipe(m)

L_e =Length of vertical pipe (m)

Q_G =Volumetric gas flow rate (m^3/s)

Q_L =Volumetric liquid flow rate (m^3/s)

Q_O =Volumetric oil flow rate (m^3/s)

Q_W =Volumetric water flow rate (m^3/s)

S_{GL} =Length of surface contact between gas and liquid in pipe cross-section (m)

V_G =Gas velocity (m/s)

V_{GA}^* =Gas velocity transition from stratified wavy flow to annular flow(m/s)

V_{GW}^* =Gas velocity transition from stratified flow to stratified wavy flow(m/s)

V_L =Liquid velocity (m/s)

V_{LD}^* =Liquid velocity transition from slug flow to dispersed bubble flow(m/s)

V_M =Velocity of the mixture (m/s)

V_{SG} =Superficial gas velocity (m/s)

V_{SL} = Superficial liquid velocity (m/s)

W_{eDG} = Critical droplet Weber number, between 20 or 30

α_L = Liquid volume fraction or liquid fraction

α_G = Gas volume fraction or gas fraction

\mathcal{E} = Energy dissipated per unit mass (m^2/s^3)

μ_L = Dynamic viscosity of liquid (kg/m.s)

θ = Angle of inclination of the pipe ($^\circ$), for horizontal pipe = 0°

ρ_L = Liquid density (kg/m^3)

ρ_G = Gas density (kg/m^3)

σ_{LG} = Surface tension between liquid and gas (N/m)

G_M = Production rate

B = Production index

P_{res} = Pressure of the reservoir (bara)

P_f = Flow line pressure (bara)

P_{RB} = Riser base pressure (bara)

\bar{P}_{RB} = Average riser base pressure over time T (bara)

Q_T = Total production over time T (STB/D)

P_{\max} = Maximum pressure (bara)

P_{\min} = Minimum pressure (bara)

T_P = Production period (s)

T_S = Slug period (s)

N = Number of segments

t_0 = Starting time (s)

OLGA = OilGAs

1 INTRODUCTION

1.1 Background

The ever increasing population and urbanization with its attendant high demand for energy, coupled with increase in oil prices since 1970s, has necessitated extensive research on finding new technology that can increase oil production and recovery from different fields. Today many oil wells are produced at satellite fields/hostile offshore environment where the productions from several wells are transported via manifolds in tie-in long distant pipeline from seabed to the receiving process facility. In this regard, a mixture of gas, oil, water and sometimes sand, hydrates, asphaltenes and wax are transported through distant pipelines to the platform for processing. The flow assurance challenges covers an entire spectrum of design tools, methods, equipment, knowledge and professional skills needed to ensure the safe, uninterrupted and simultaneous transport of gas, oil and water from reservoirs to the processing facility (Storkaas,2005). The cost of processing offshore is enormous in terms of Capital Expenditure (CAPEX) and Operation Expenditure (OPEX) due to technical difficulties of producing offshore, and considering the limited space available and other consideration such as harsh weather.

Slug flow that arises in multiphase (gas, oil, water) transport is a major challenge in oil exploration, production, recovery and transport. Slugging is the intermittent flow regime in which large bubbles of gas flow alternately with liquid slugs at randomly fluctuating frequency (Issa and Kempf, 2003) in pipeline. Slug causes a lot of problems due to rapid changes in gas and liquid rate entering the separators and the large variations in system pressure. Slug flow is a regular phenomenon in many engineering applications such as the transport of hydrocarbon fluids in pipelines, liquid-vapour flow in power plants and buoyancy-driven equipment (Fabre and Line', 1992). The slug can be formed in low-points in the topography of the pipeline. It can be hydrodynamic induced slugging, terrain induced slugging or operation induced slugging.

Hydrodynamic slugging, which is the main subject of this project occur in a horizontal or near horizontal pipes and can be generated by two main mechanisms (i) natural growth of hydrocarbon instability and (ii) liquid accumulation due to instantaneous imbalance between pressure and gravitational forces caused by pipe undulations (Issa and Kempf, 2003) .

For the natural growth phenomenon, small random perturbation of short wavelengths arising naturally may grow into larger and longer waves on the surface of the liquid due to the Kelvin-Helmholtz instability (Ansari, 1998). These waves may continue to grow as it transverses the length of the pipe line, picking up liquid flowing ahead of them, until they bridge the pipe cross-section, thereby forming slug. In real flow, all these events take place at different times, hence some slugs grow, while others collapse and they may travel at different speeds leading to the merging of some slugs with others (Taitel and Barnea, 1990).

In the case of liquid accumulation, slug flow may form at pipe dips due to the retardation and subsequent accumulation of liquid in the dips leading to the filling up of the cross-section with liquid. This is an extreme example of terrain induced slug flow also called “severe slugging” and occurs when a slightly inclined pipeline meets a vertical riser (Schmidt et al, 1985; Jansen et al, 1996).

Slug may arise by the combination of the mentioned mechanisms simultaneously in long hydrocarbon transport pipelines. In such cases, the slugs generated from one mechanism interact with those arising from the second leading to a complex pattern of slugs, which may overtake and combine (Issa and Kempf, 2003).

The intermittency of slug flow causes severe unsteady loading on the pipelines carrying fluid as well as on the receiving facility such as the separators. This gives rise to problems in design and therefore it is important to be able to predict the onset and subsequent development of slug flow and its control.

The purpose of this work was to investigate the capability of active feedback control strategy based on measurement of pressure or holdup transmitter at the

riser base as controlled variable with topside choking as manipulated variable with PI controller in Olga simulation to mitigate hydrodynamic slug flow.

1.2 Hydrodynamic Slugging

Hydrodynamic slug is initiated by the instability of waves on the gas /liquid interface in stratified flow. The gas /liquid interface is lifted to the top of the pipe when the velocity difference between gas phase and liquid phase is high enough. This wave growth is triggered by the Kelvin-Helmholtz instability and when the wave reaches the top of the pipe, it forms slug blocking the gas passage in the flow line see figures 1-1, 1-2, 1-3 and 1-4 respectively. At this point the liquid volume fraction (holdup) is one as the gas volume fraction tends to zero. When the slug front travels faster than the slug tail, the slug grows. Conversely, if the slug tail travels faster than the slug front, the slug decays. If the slug front and the slug tail travel at the same speed, a stable slug is obtained. When the gas velocity is high enough, gas will be entrained in the liquid as gas entrainment figure 1-1.

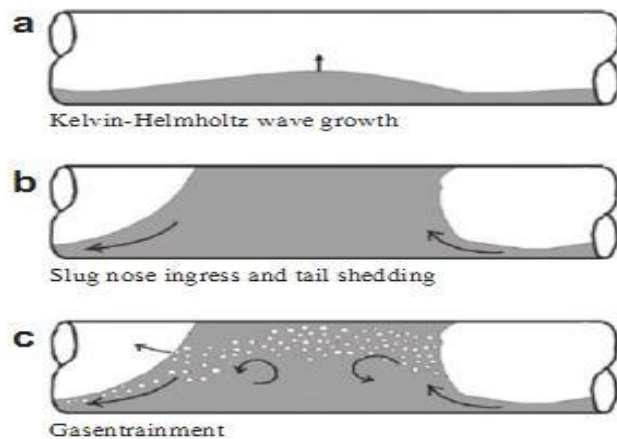


Figure 1-1 Hydrodynamic slug propagation (Varne, V. 2010)

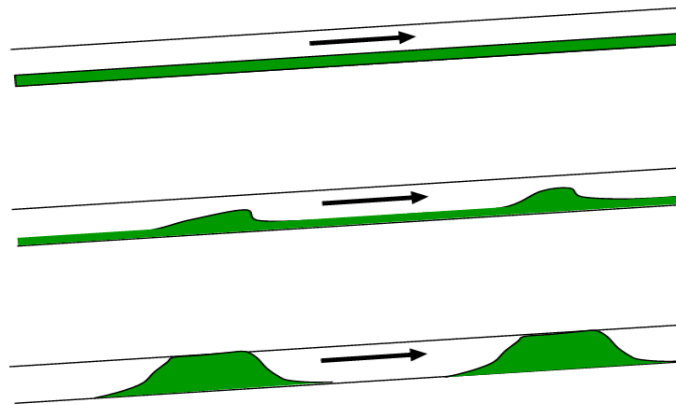


Figure 1-2 Hydrodynamic slug flow (stratified, wave instability and plugged hydrodynamic slugging, (Oram, 2013))

The holdup and surging from the horizontal flow line are transmitted to the relatively short riser and the riser may have to handle far more liquid than normal as a result of the surge from the plug of liquid. Hydrodynamic slug mitigation which is the main thrust of this project is a non-zero limit flow of liquid slug and gas pocket due to wave instability and velocity difference between the gas and liquid. Due to the dynamics of the wave instability, it is usually difficult to predict hydrodynamic slug volume.

As the multiphase fluid transverses the length of the pipeline, due to the velocity difference between the gas and the liquid and other related phenomena like wave instability, the flow regime changes from stratified, wavy and plugged hydrodynamic slugging that may block the passage of gas in the flow line see figures 1-1 (Varne, V. 2010), 1-2 (Oram, 2013) and 1-3 (Varne, V. 2010).

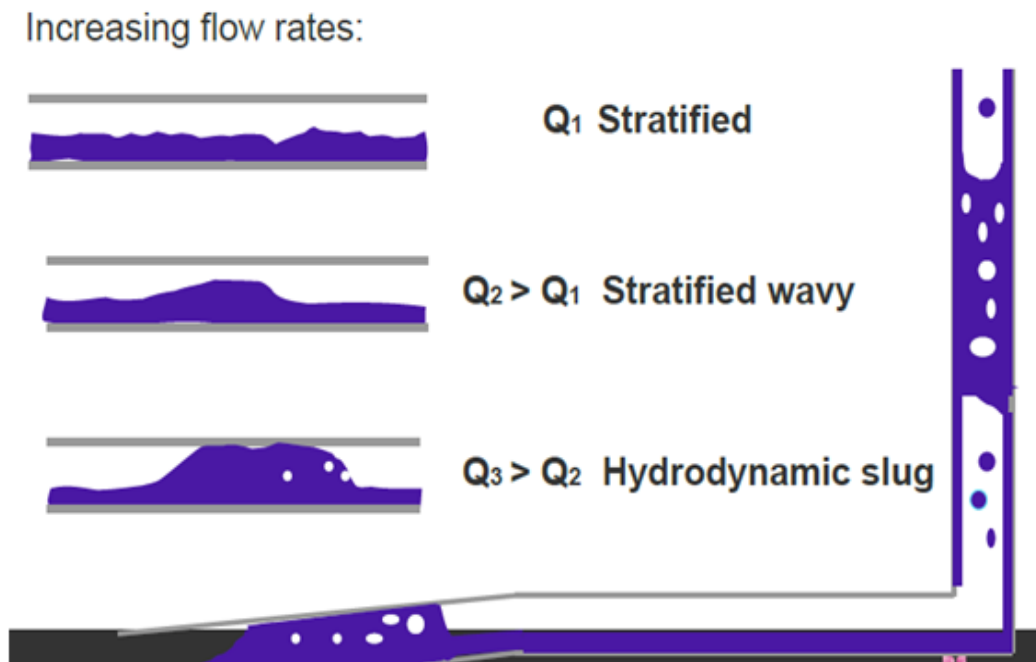


Figure 1-3 Hydrodynamic slug flow regimes (Varne, V. 2010)

The region of our interest is the unstable equilibrium that we wish to stabilise using active feedback control.

1.3 Why is Slugging a Problem?

The resulting increased topside instability caused by pressure build-up can lead to:

- Liquid overflow in the separator
- High pressure in the separator
- Poor phase separation
- Fatigue due to repeated impact
- Overload on gas compressors (Mehrdad, 2006)
- Platform trips and possible early platform abandonment
- Long term damage to the reservoir due to resulting bottom hole pressure variations, causing permanent decrease in the production of oil and gas from the reservoir (Ogazi et al, 2010)

1.4 Compare Mechanisms of Hydrodynamic and Severe Slugs.

Hydrodynamic slugging is a non-zero limit flow of liquid slug and gas pocket in a horizontal or near horizontal pipe line due to velocity difference between the gas and the liquid and wave instability in the conduit (see figure1-3). Due to the dynamics of the wave instability, it is usually difficult to predict the slug volume in hydrodynamic slugging.

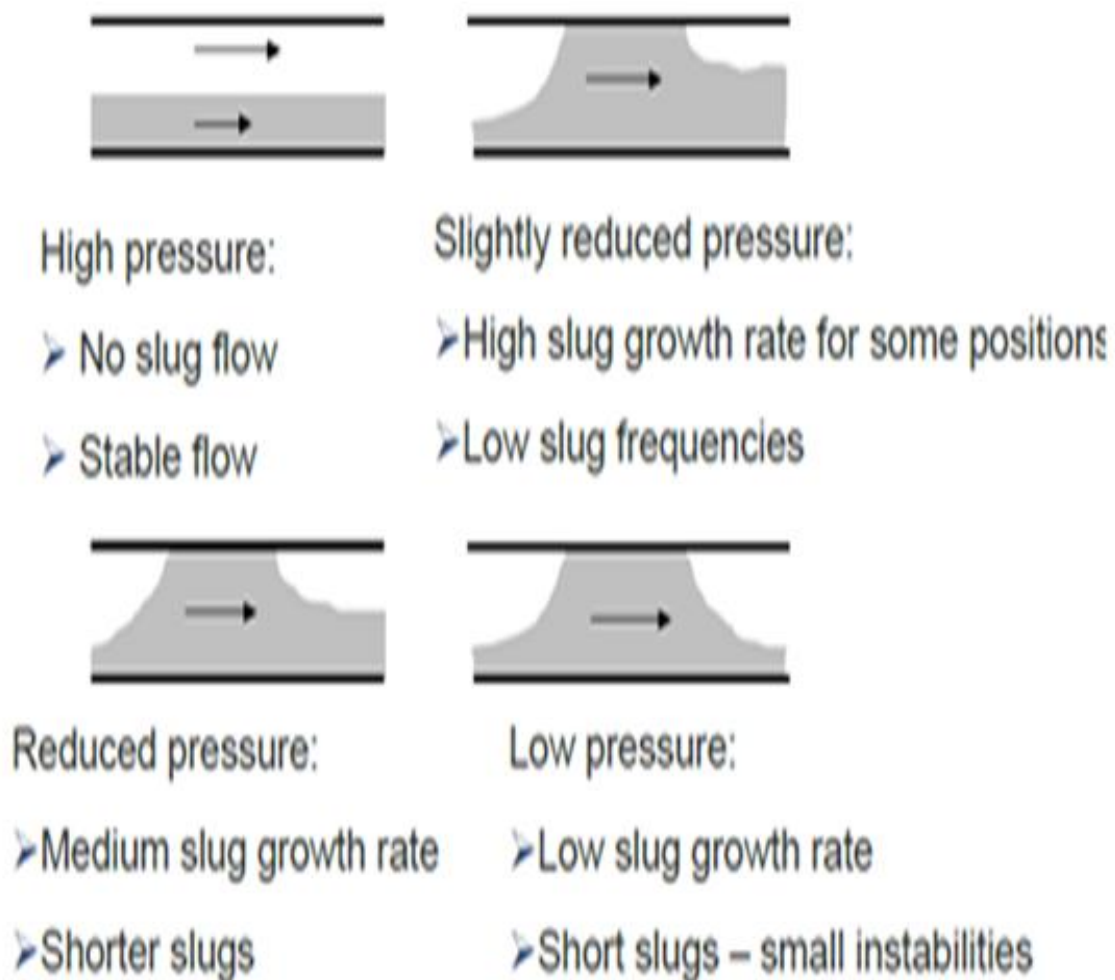


Figure 1-4 Mechanism of hydrodynamic slugs (Varne, V. 2010)

From figure 1-4, at high pressure the flow is stratified and stable. As the pressure slightly reduces due to the Bernoulli effects resulting from increased

gas velocity, a wave build-up is initiated in the flow line that can grow to fill the pipe diameter and hence block the gas flow in the pipeline.

Severe slugging or terrain induced slugging in the other hand may occur at low flow rates, when a downwards incline or horizontal pipeline is connected to a vertical riser. It is characterized by a cyclic behaviour alternating between no liquid flows at the outlet, to a high liquid delivery (surge) at the outlet. These occur when the rate of liquid flow to the riser is higher than the rate of flow up the riser and thus can cause an accumulation. The maximum slug volume in severe slugging is usually the height of the riser. This slug type is cyclic and characterized by blockage of flow at the dip or low points resulting in pressure build-up upstream the blockage until the compressed gas upstream is able to overcome the gravitational head, causing a blowout of liquid.

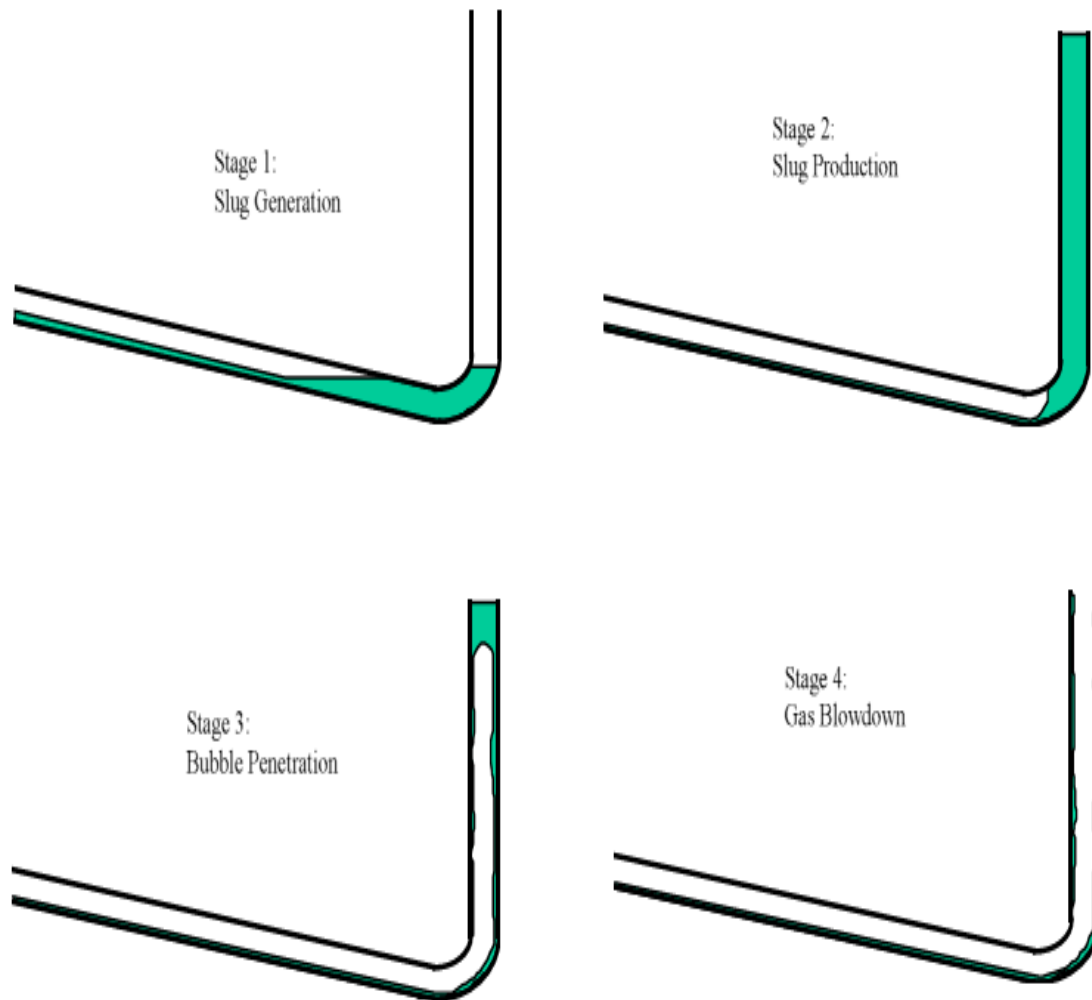


Figure 1-5 Mechanism of severe slugs or terrain induced slugs (Oram, 2013)

Under severe slug conditions, a cyclic operation is obtained. It is considered to consist of four steps (Schmidt et al, 1980; Taitel, 1986). These steps are illustrated in figure 1-5

- “(a) Liquid accumulation at the low point blocking the gas flow (slug generation)
- (b) As more gas and liquid enters into the system, the pressure will increase and the riser will be filled with liquid (slug production)

- (c) After a while the amount of gas that is blocked will be large enough to blow the liquid out as gas penetrates into the riser (bubble penetration)
- (e) After the blowout, the pressure drops and fluid falls back for a new slug cycle to start to form (slug blowout).”

1.5 Operation Induced Slugging

This type of slug could be induced by operational changes in the system, such as start-up, ramp-up, or pigging etc. During start-up, slug may be formed owing to liquid which settled at the low points in the line after shutdown. Also when there is a change in the steady condition of flow (flow rate change) in the multiphase flow line. For example, when there is a production rate drop or increase for a line operating in stratified flow, slug could be formed. Transient simulator Olga can be used to simulate such a condition.

Most of the earlier works on slug mitigation (Yocum, 1973, Schmidt et al, 1980), concentrated on the mitigation of the flow instability with little emphasis on the effect of the mitigation strategy on oil production and recovery. These limitations propelled a continued research on slug control strategies to investigate further into methods that will enhance optimal production and recovery. Recently, (Ogazi, et al, 2009), reported the effectiveness of feedback control as severe slug mitigation strategy with a robust controller, while the current work seek to extend investigation on the effectiveness of feedback control strategy to mitigate hydrodynamic slugging.

1.6 Slug Mitigation and Prevention Methods

There are a number of slug mitigation and prevention methods, which includes:

- “Increasing the flow rate
- Riser base gas injection
- Gas lift in the well
- Fixed topside choking
- Combination of gas injection and topside choking

- Slug catcher
- Active feedback control
- Modified flow line layout/riser base geometry to avoid a dip”(Yocum, 1973)

This research project utilized topside choking to control hydrodynamic slug flow problem with active feedback.

1.7 Aim.

- The aim of this research is to develop a method for hydrodynamic slug control using topside choking with active feedback control.

To achieve this aim, the following objectives were pursued.

1.8 Objectives.

- Investigate the suitability of active feedback control using topside choke for hydrodynamic slug control.
- Perform controllability analysis on the possible control variables.
- Investigate the effectiveness of this control strategy to improve oil production and recovery

1.9 Conclusion

Hydrodynamic slugs have been found to occur in a horizontal or near horizontal pipeline by two main mechanisms (i) natural growth of hydrocarbon instability due to Kelvin-Helmholtz instability (ii) liquid accumulation due to instantaneous imbalance between pressure and gravitational forces caused by pipe undulations. Slug may also arise by the combination of the two mechanisms presented simultaneously in long hydrocarbon transport pipeline. In such case, the slug generated from one mechanism interacts with those arising from the second mechanism leading to a complex pattern of slugs which may overtake and combine. The slug may grow when the slug front travels faster than the slug tail or travelling an upward inclination. It may decay when the slug tail travels faster than the slug front or travelling a downward inclination. If both the

slug front and the slug tail travel at the same speed, a stable slug may be formed.

Active feedback control technology has not been extended for the investigation of hydrodynamic slug control in the literature. This extension of the capability of active feedback control technology with topside choke valve to mitigate hydrodynamic slug flow is the main contribution of the present work.

2 LITERATURE REVIEW

2.1 Multiphase Flow

Multiphase flow is a very complex flow behavior and its description depends heavily on the flow regime detection. To be able to calculate important factors such as pressure drop and flow rates, it is critical to know the flow regime in all parts of the system.

The parameters that determine which flow regime will occur is also changing with time as the wells are getting more and more depleted at the end of their life-time. This means that the engineers must plan for different scenarios when designing the production and process system.

Slugging is a flow regime that causes a lot of problems due to rapid changes in gas and liquid rates entering the separators and large variations in system pressure. It can be hydrodynamic slugging, terrain induced slugging or operation induced slugging. Figure 2-1 show three phases water, oil and gas as they transverse a horizontal pipe cross-section.

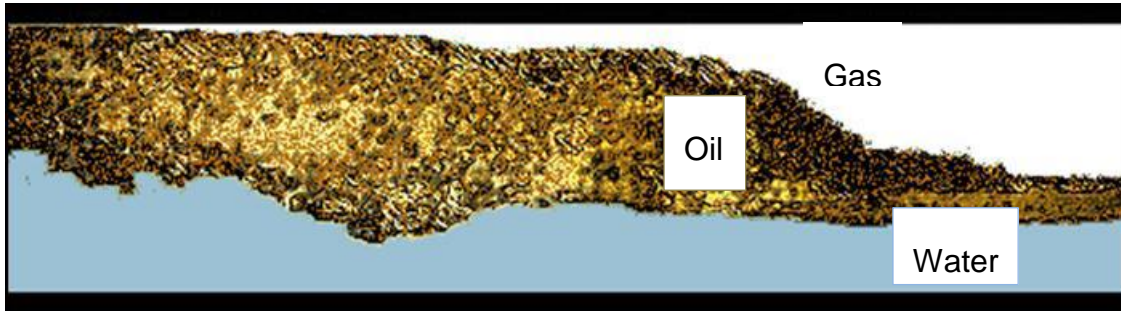


Figure 2-1 Schematic slug fronts in horizontal water-oil-gas flow line (Bratland, 2010)

The understanding of how water, oil and gas in a conduit respond to pressure changes, flow rate changes, composition, density changes, viscosity changes, and temperature changes etc, will help the operator to predict accurately the development of transient flows usually caused by slug propagation. Traditional flow pattern has been produced as a tool to predict the flow regime that will develop in the pipeline (Taitel and Dukler, 1976; Barnea, 1977).

2.2 Flow Regime Determination

Determining flow regime is critical in the analysis of multiphase flow. In cases where the flow happens to be near the border between two or even three different flow regimes, the uncertainties are generally most significant. We may also experience situations where minor changes in flow properties or inclination angle is likely to change the flow regime, and simulation may require more accurate pipe elevation profile or fluid composition data than are available. These uncertainties are investigated by simulating several times with slightly different input-data to see how the results compare. The main mechanism at work in the switching from one flow regime to another is thought to be the Bernoulli effects, which reduces the pressure if the gas velocity is increased (Bratland, 2010).

2.2.1 Flow Regime Map in Horizontal Pipe

Figure 2-2 shows the flow regimes that may develop as the multiphase fluid flows across the pipeline at varying conditions.

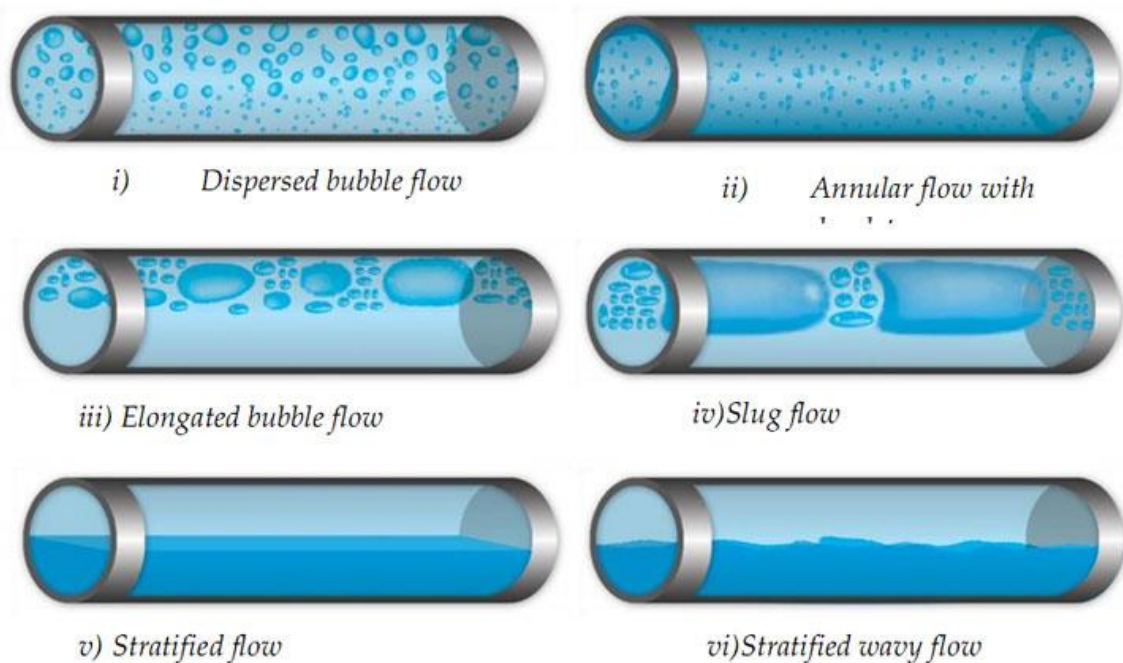


Figure 2-2 Schematics of flow regimes in horizontal pipe (Bratland, 2010).

Flow regime is a function of gas/liquid superficial velocity changes. At low gas and liquid velocity flow is stratified, increasing the liquid superficial velocity, shifts the flow regime to intermittent flow (slug region). With further increases in the liquid velocity, the flow regime becomes bubble flow.

Conversely, increasing the gas velocity will shift the flow region to the right. The flow regime become stratified wavy or annular with further increase of gas superficial velocity as in figure 2-3

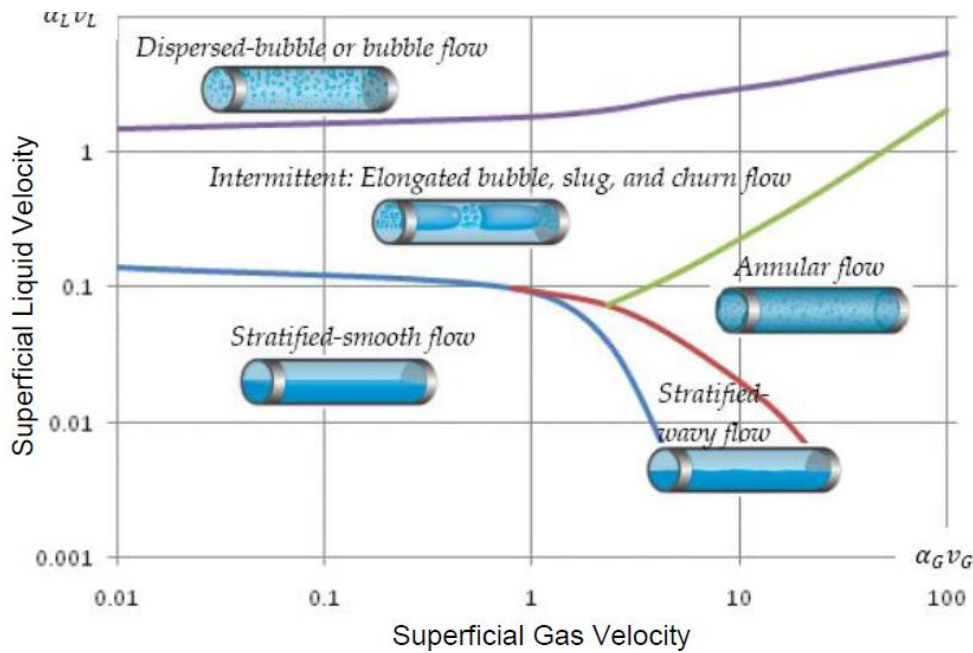


Figure 2-3 Flow regime map for horizontal pipe with gas - liquid two phase flow (Bratland, 2010).

2.3 Prediction of Flow Regime Transition in Horizontal Pipes

Mathematical model for the prediction of flow regime map was developed by (Taitel and Dukler, 1976).

2.3.1 Transition from Stratified Flow

Mathematical model (Taitel and Dukler 1976) for transition from stratified flow prediction.

$$v_{GW}^* = \left[\frac{4\mu_L(\rho_L - \rho_G)g \cos(\theta)}{v_L \rho_G \rho_L S} \right]^{0.5} \quad (2.1)$$

μ_L = Dynamic viscosity of liquid [kg/m.s]

ρ_L = Liquid density [kg/m³]

ρ_G = Gas density [kg/m³]

g = Acceleration due to gravity [m/s²]

θ = Angle of inclination of the pipe [°], for horizontal pipe= 0°

S = Sheltering coefficient=0.01

v_L = Liquid velocity [m/s]

v_G = Gas velocity [m/s]

When the gas velocity is greater than v_{GW}^* the flow regime will change from stratified flow to stratified wavy flow (blue line of figure 2-3). These flow regimes are assumed accurate within the limit of $\pm 10^\circ$ angle of inclination (Bratland, 2010)

2.3.2 Transition to Annular Flow

Bratland, (2010) reported that Bernoulli principle was applied by Taitel and Duckler to predict transition to annular flow.

$$v_{GA}^* = \left(1 - \frac{h_L}{d}\right) \left[\frac{(\rho_L - \rho_G) g \cos(\theta) A_G}{\rho_G S_{GL}}\right]^{0.5} \quad (2.2)$$

h_L = Liquid height in the pipe [m]

d = Pipe inner diameter [m]

v_{GA}^* = Gas velocity transition from stratified wavy to annular flow (m/s)

S_{GL} = Length of surface contact between gas and liquid in pipe cross-section [m]

A_G = Cross-sectional area of the gas [m²]

g = Acceleration due to gravity [m/s²]

Flow becomes annular when the gas velocity exceeds \mathcal{V}_{GA}^* (red line of figure 2-3)

Taitel et al, (1980) found that liquid height in the pipe h_L has to be less than 0.35 of internal diameter for the flow to be in annular flow, otherwise the flow would be slug flow. This is summarized in the following conditions (Bratland, 2010)

$$\text{Annular flow if } \mathcal{V}_G > \mathcal{V}_{GA}^* \quad \text{and} \quad h_L < 0.35d \quad (2.3)$$

$$\text{Slug flow if } \mathcal{V}_G > \mathcal{V}_{GA}^* \quad \text{and} \quad h_L > 0.35d \quad (2.4)$$

2.3.3 Transition to Dispersed Bubble Flow

When the liquid velocity is further increased, the flow become turbulent which leads to crushing the Taylor bubbles to small dispersed bubble (Bratland, 2010).The flow transits from slug flow to dispersed bubble flow (grey line of figure 2-3) represented by the equation 2.5 (Bratland, 2010)

$$\mathcal{V}_{LD}^* = \left[\frac{4A_G g \cos(\theta)}{S_{GL} f} \cdot \left(1 - \frac{\rho_G}{\rho_L} \right) \right]^{0.5} \quad (2.5)$$

f Darcy-Weisbach friction factor

\mathcal{V}_{LD}^* = Transition velocity from slug flow to Dispersed bubble flow. When the liquid velocity exceed \mathcal{V}_{LD}^* the flow becomes dispersed bubble flow.

2.4 Flow Regime in Vertical Pipes

It is highly dependent on the in-coming gas flow rate, as the amount of gas is gradually increased, the flow regime transit from bubble flow, slug (intermittent) flow, churn flow, and annular flow respectively in vertical pipes. For annular flow the liquid film at the wall no longer have a uniform thickness. Figure 2-4 shows the flow regime transition that may occur in vertical pipes.

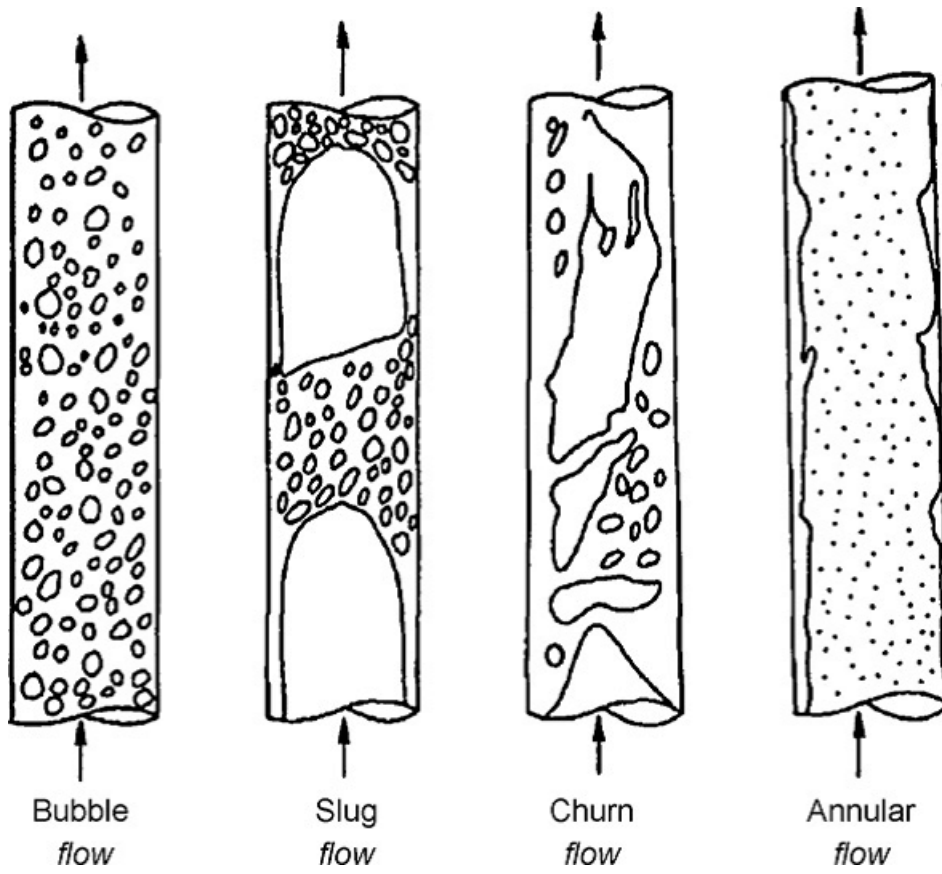


Figure 2-4 Schematic of vertical pipe flow regime (Crowe,2009)

2.5 Vertical Pipe Flow Regime Map

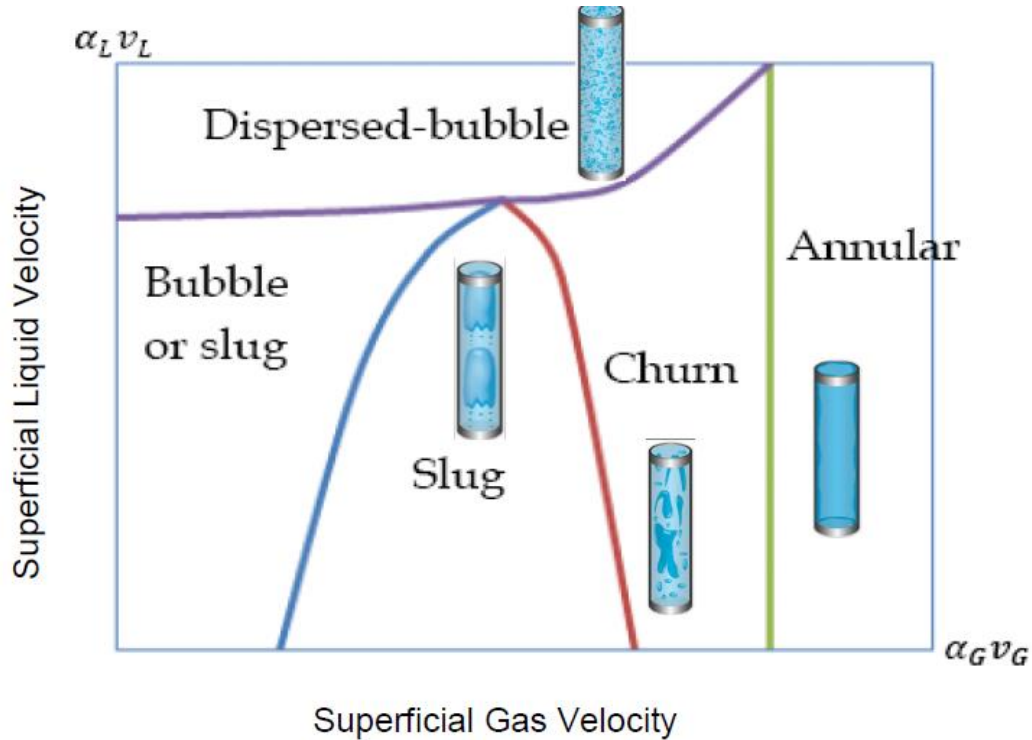


Figure 2-5 Flow regime map for vertical pipe with gas–liquid two phase flow (Bratland, 2010)

It is harder to identify visually flow regime map in vertical pipe than in horizontal pipes. The mathematical model predicted by (Taitel et al, 1980) is the most utilized for the prediction of flow regime map in vertical pipes.

2.5.1 Transition from Bubble to Slug Flow

Bubble flow does not usually exist in small diameter vertical pipes (Bratland, 2010). Transition from Bubble flow to slug was predicted by Bratland 2010 in equation 2.6 represented by blue line of figure 2-5.

$$\alpha_L v_L = 3\alpha_G v_G - 1.15 \left[\frac{(\rho_L - \rho_G)g\sigma_{LG}}{\rho_L^2} \right]^{0.25} \quad (2.6)$$

α_L = Liquid volume fraction

α_G = Gas volume fraction

v_L = Liquid velocity [m/s]

v_G = Gas velocity [m/s]

σ_{LG} = Surface tension between gas and liquid [N/m]

g = Acceleration due to gravity [m/s²]

2.5.2 Transition to Dispersed Bubble Flow

When the liquid velocity is high enough, bubble flow transition occurs, and the turbulent flow mixes the bubbles with the liquid (grey line) of figure 2-5 (Bratland, 2010)

$$\left(0.725 + 4.15 \sqrt{\frac{\alpha_G v_G}{\alpha_G v_G + \alpha_L v_L}}\right) = 2 \left[\frac{0.4 \sigma_{LG}}{(\rho_L - \rho_G)g}\right]^{0.5} \quad (2.7)$$

2.5.3 Transition from Slug to Churn Flow

Churn flow occurs when the gas flow rate increase until the slug length decays to zero. Choham, (2006) reported that flow regime at inlet of a vertical pipe is always churn flow and the flow regime changes to slug as distance into the pipe increase. Bratland, (2010) described the transition by the equation 2.8 represented by the (red line) of figure 2-5.

$$\frac{L_e}{d} = 40.6 \left(\frac{v_M}{\sqrt{gd}} + 0.22\right) \quad (2.8)$$

L_e = Length of the vertical pipe [m]

2.5.4 Transition from Churn to Annular Flow

When the gas flow rate is further increased, the flow regime changes from churn flow to annular flow as presented by the (green line) in figure 2-5 (Bratland, 2010) in equation 2.9.

$$v_G \alpha_G \approx v_G = \left[\frac{4W_{eDG}^* \sigma_{LG}(\rho_L - \rho_G)g}{3C_D \rho_G^2}\right]^{0.25} \quad (2.9)$$

W_{eDG}^* = Critical droplet Weber number, between 20 or 30

C_D = Drag coefficient C_D is obtained by iteration.

2.6 Terminology Used in Multiphase Flow Literature

This section defines some of the terminology used in the thesis as obtained from multiphase flow literature (Handbook of Multiphase Flow Metering, 2005).

2.6.1 Volume Fraction and Holdup

This is the area occupied by one phase in the cross sectional area of the pipeline (Bratland, 2010). If the area fraction is occupied by the liquid, it is termed liquid area fraction or holdup. Since volume corresponds to area if the length of that volume is infinitely small (infinitely small pipe length), this area can be termed volume fraction for the liquid or gas respectively.

$$\alpha_L = A_L/A \quad \text{and} \quad \alpha_G = A_G/A \quad (2.10)$$

α_L = Liquid volume fraction or liquid fraction

α_G = Gas volume fraction or gas fraction.

A_G = Area occupied by gas [m²]

A_L = Area occupied by liquid [m²]

A = Area of pipe cross-section [m²]

2.6.2 Superficial Velocity

The average fluid velocity in one phase is calculated by dividing the volume flow rate by the pipe cross-sectional area, as average fluid speed is difficult to calculate in multiphase flow. The assumption of single phase is made as running solely in the pipe to calculate the superficial velocity thus:

$$v_{SG} = Q_G/A \quad \text{and} \quad v_{SL} = Q_L/A \quad (2.11)$$

v_{SL} = Superficial liquid velocity [m/s]

v_{SG} = Superficial gas velocity [m/s]

Q_L = Volumetric liquid flow rate [m³/s]

Q_G = Volumetric gas flow rate [m³/s]

$$v_G = v_{SG}/\alpha_G \quad \text{and} \quad v_L = v_{SL}/\alpha_L \quad (2.12)$$

Where

v_G = Gas velocity [m/s]

v_L = Liquid velocity [m/s]

2.6.3 Water-Cut

The ratio between the volumetric flow rates of water to the total volumetric flow rate of liquid (used in oil extraction when water is produced as part of well production).

$$\text{Water-cut} = Q_w/Q_L \quad (2.13)$$

Q_W = volumetric water flow rate [m³/s]

2.6.4 Gas Oil Ratio (GOR)

Gas-Oil ratio is the ratio between produced volumetric flow rate of gas to the volumetric flow rate of oil when oil and gas are produced as part of well production.

$$\text{GOR} = Q_G/Q_o \quad (2.14)$$

Q_o = Volumetric oil flow rate [m³/s]

2.6.5 Gas Liquid Ratio (GLR)

Ratio between produced volumetric gas flow rate to the volumetric flow rate of total liquid viz (oil plus water)

$$\text{GLR} = Q_G/Q_L \quad (2.15)$$

2.7 Standard Condition

Standard conditions are internationally accepted reference measurement applied in the oil industry. The standard conditions as defined per British Standard (British Standard, 2005) at temperature 288.15k (15°C or 59F). However, imperial units referred to as field units are commonly applied by the oil industries. This imperial/field unit is applied by Olga calculation with in-built metric units. Examples of such field units are Million Standard Cubic Feet per Day (MMscf/d) for gas volumetric flow rate and Standard Barrel per Day (STB/d) for oil.

2.8 Review of slug control techniques

In order to effectively deal with the hydrodynamic slug problems, a number of publications reviewed on the earlier works were investigated to gain insight into the progress made in this area.

The earlier work on slug control reported in literature was (Yocum, 1973), which concentrated on flow stability with little emphasis on effect on production.

The publication identified several slug elimination techniques that are still referenced till today. These techniques include reduction in the pipeline diameter; splitting of the flow into multiple streams; gas injection into the riser or a combination of gas injection and choking. Yocum reported that increased back-pressure could eliminate slugging but would severely reduce the flow capacity up to 60%. Contrary to Yocum's report, Schmidt et al,(1985) noted that slugging in a pipeline riser system could be eliminated or minimized by choking at the riser top with little or no change in the flow rates and pipeline pressure.

Schmidt also indicated that elimination of slugging could be achieved by gas injection, but dismissed it as not being economically viable due to the cost of a compressor to pressurize the gas for injection and piping required to transport the gas to the base of the riser.

Hills, (1990) described riser base gas injection test performed on the S.E. Forties field to eliminate slugging. The gas injection was shown to reduce the extent of slugging. The condition for eliminating slugging using gas injection was to bring the flow regime in the riser to annular flow thus preventing liquid accumulation at the riser base.

Jansen (1990) investigated different elimination techniques such a back-pressure increase, choking, gas injection, choking and gas injection combination. He made the following observations:

“Very high back-pressures were required to eliminate severe slugging. Careful choking was needed to stabilize the flow with minimal back-pressure increase. Large amounts of gas were needed to stabilize the flow with gas injection method only” (Jansen 1990).

Choking and gas injection combination are being considered as a viable method for slug control, reducing the degree of choking and the amount of gas injection needed to stabilize the flow and yield an optimal production.

Jansen and Shoham (1994), worked together on mitigation of terrain induced slug using combination of advantages of choking and gas injection. The idea was to combine the advantages of both methods; increased choke valve opening, plus reduced gas injection rate as a viable approach to stabilized controlled start-up of a smooth flow system.

Ogazi et al, (2010) studied severe slug control with maximal choke valve opening with a robust PID controller using the Cranfield University multiphase test facility to maximize oil production.

The test facility figure 2-6 consists of a 2-inch and a 4-inch riser pipeline system that can run alternatively. The 2-inch riser is a vertical riser with upstream pipeline length of 39m inclined downward at 2° and a riser height of 11m, while the 4-inch riser is catenary with upstream pipeline length of 55m, also inclined downward at 2° and a riser height of 10.5m. Fluid for both systems is supplied from three independent single-phase sources for oil (dielectric 250), water, and air. For each riser system, the supplied fluid mixes at a mixing point into the pipeline which connects to the riser. The facility comprises of a fluid supply and metering section, test section and phase separation and measurement section respectively. The top of both risers is equipped with a topside processing facility which includes a control valve and a two-phase vertical separator that separates the fluid into liquid and gas for measuring instruments. The two-phase separator is approximately 1.2m high and 0.5m in diameter. It consists of the gas and liquid outlet control valves, pressure, flow, temperature, and level transmitters. Pressure and flow measurements are obtained at riser inlet and outlet. A schematic of this facility is shown in figure 2-6. Ogazi reported that active feedback control implemented at an open-loop unstable operating point is:

- 1) “Effective in suppressing severe slug formation and controlling severe slugging in multiphase flow pipeline, with minimal back-pressure on the riser pipeline system, and can achieve lower back-pressure than using manual choking method.
- 2) Significant reduction in back-pressure is achieved by implementing severe slug control at open-loop unstable operating point with active feedback control and oil production is increased in the system.
- 3) With the robust PID controller, the percentage increase in production increased by 7.1% more when compared to manual choking. “

Cao et al, (2011) used the Cranfield University multiphase test facility described in figure 2-6 to investigate the effectiveness of gas injection at the riser base to mitigate hydrodynamic slug. Water and air with flow rates of 0.25kg/s and 5Sm³/h respectively were used as test fluid. 125m³/h of air was injected at the bottom of the riser which stabilised the flow in the riser. Pressure differential

across the riser was used as controlled variable to control the opening of the gas injection valve and 4.19% reduction in gas injection rate was reported as achieved with active control. The present work seeks to investigate the effectiveness of active feedback control using the topside choke valve to mitigate hydrodynamic slug flow.

2.9 Control and Controllability Analysis

The primary objective of process control is to maintain a process at a desired operating conditions, safely and efficiently, while satisfying environmental and product quality requirements (Seborg et al, 2004). In feedback control system, the controller looks at the actual measured output and compares it with the desired value (set-point), and returns a corrective action when there is deviation (error) between the set-point and the measured output as may be appropriate. The three important process variables are:

- “Controlled variable (CV): These are process variables that are controlled, and the desired values of a controlled variable is referred to as its set-point
- Manipulated variable (MV): The process variables that can be adjusted in order to keep the controlled variables at or near the set-point.
- Disturbance variable (DV): These are process variables that affect the controlled variable but cannot be manipulated”.

Disturbances generally are related to changes in the operating environment of the process. The specification of CVs, MVs and DVs is a critical step in developing a control system and their selection is based on process knowledge, experience and control objective. (Seborg et al 2004)

2.10 Measurement and Actuation

Measurement devices (sensors, transmitters and actuation equipment (control valves)) are used to measure process variables and implement the calculated control action. These devices are interfaced in the control system, digital control equipment as digital computers. It is important that the controller action be

specified correctly because incorrect choice results in loss of control. The controller compares the measured value to the set-point and takes the appropriate corrective action by sending an output signal to the current -to - pressure transducer, which in turn sends a corresponding pneumatic or electric signal to the control valve (actuator).

A process control system can be categorised based on the number of input or output variables into four main types (Skogestad and Postlethwaite, 2005; Seborg et al, 2004; Ogunnaike and Ray, 1994).

- Single Input, Single Output (SISO) control system.
- Single Input, Multi Output (SIMO) control system.
- Multi Input, Single Output (MISO) control system.
- Multi Input, Multi Output (MIMO) control system.

2.11 Structure of PID Controller.

Every controller has the objective to reduce the error signal to zero (the difference between the measured value and the set-point) as represented in equation 2.16.

$$\mathbf{e}(t) = \mathbf{y}_{sp}(t) - \mathbf{y}_m(t) \quad (2.16)$$

Where $\mathbf{e}(t)$ = error signal.

$\mathbf{y}_{sp}(t)$ = Set-point.

\mathbf{y}_m = Measured value of the controlled variable.

Other performance objective will include:

- ✓ The selection of a controller to make the close-loop system stable.
- ✓ Achieve a reference-tracking objective and making the output follow the reference or set-point signal.
- ✓ If a process disturbance is present, the controller may have disturbance rejection objectives to attain.

- ✓ Some noise filtering properties may be required in the controller to attenuate any measurement noise associated with the measurement process.
- ✓ A degree of robustness in the controller design to model uncertainty may be required.

(Astron and Hagglund, 1995; Seborg et al, 2004; Ogunnaike and Ray, 1994)

Feedback controllers have been grouped into three categories in accordance with three terms PID.as represented thus;

- ❖ Proportional controller (P).
- ❖ Integral controller (I).
- ❖ Derivative controller (D).

These controllers can be paired in a manner that produces better performance in relation to the process being controlled. The most effective combinations are Seborg et al, 2004).

1. Proportional controller (P).
2. Proportional-Integral controller (PI).
3. Proportional-Derivative controller (PD).
4. Proportional-Integral-Derivative controller (PID).
5. On-Off controller.

2.12 PID Controller Equations.

2.12.1 Proportional Control.

Proportional control is denoted by the P-term in the PID controller. It is used when the control action is to be proportional to the size of the process error signal.

$$\text{Time domain} \quad u(t) = k_c e(t) + u_s \quad (2.17)$$

The gain of the Controller can be adjusted so that the change in the output of the controller can be sensitive to deviations between the set point and the controlled variable as desired (Seborg et al, 2004). The steady state value

(bias) can be adjusted using manual reset so that the output of the controller equals the steady state value when the error is zero. The transfer function of the proportional controller is given in equation 2.18.

$$\text{Laplace domain } k_c = \frac{U(s)}{E(s)} \quad (2.18)$$

Where k_c = Proportional gain.

The problem encountered in using the proportional controller is the steady state error after a sustained disturbance. The steady state error is remedied only by manual resetting. The increase in the proportional gain results in the reduction in the steady state error but this makes the system prone to oscillation.

The sign of the proportional gain can either be positive or negative to make the output of the controller to either decrease with an increase in the error (Seborg et al, 2004). When the proportional gain is negative, the process variable (riser base pressure for example) decreases when the manipulated variable (valve opening) increases. When the proportional gain is positive, the process variable (example riser base pressure) increases, when the manipulated variable (valve opening) decreases. The limitation of the proportional controller is the inability to return to the set-point after an offset (steady state error) without manual resetting. This may cause the system to oscillate (Astron and Haggund, 1995). This limitation is what the Proportional-Integral PI controller is designed to correct by taking the integral of the error from zero to time (t) and returns to zero after an offset (steady state error).

2.12.2 Proportional-Integral (PI) Controller

Proportional-Integral controller is a modification of the proportional controller with an integral mode added, it is used when it is required that the controller corrects for any steady state offset from a constant reference signal value thus (Astron and Haggund, 1995; Seborg et al, 2004; Ogunnaike and Ray,1994), combining the Proportional – Integral action gives the PI controller given as:

$$u(t) = k_c \left[e(t) + \frac{1}{\tau_I} \int_0^t e(\tau) d\tau \right] \quad (2.19)$$

The transfer function of the Proportional-Integral controller is given as :(Astron and Hagglund, 1995; Seborg et al, 2004; Ogunnaike and Ray, 1994)

$$\frac{U(s)}{E(s)} = k_c \left(1 + \frac{1}{\tau_I s} \right) = k_c \left(\frac{\tau_I s + 1}{\tau_I s} \right) \quad (2.20)$$

The integral term helps to bring the system back to the set-point by eliminating the steady state error caused by the proportional gain. When the integral time is small, “the integral action will be large this means faster elimination of the steady state error, but more oscillation. Conversely, large integral time means small integral action and slower elimination of the steady state error with less oscillation” (Seborg et al, 2004; Ogunnaike and Ray, 1994). The integral mode cannot be used as a stand-alone controller because it performs little control action until the error signal has lasted for some time.

2.12.3 Proportional-Integral–Derivative (PID) Controller

The family of PID controller is constructed from various combinations of the proportional, integral and derivative terms as required to meet specific performance requirement. The three terms are combined together as PID to give combined total action thus:

$$\text{Time domain } u(t) = k_c \left\{ e(t) + \frac{1}{\tau_I} \int_0^t e(\tau) d\tau + \tau_D \frac{de(t)}{dt} \right\} \quad (2.21)$$

$$\text{Laplace transforms } \frac{U(s)}{E(s)} = k_c \left(1 + \frac{1}{\tau_I s} + \tau_D s \right) \quad (2.22)$$

The transfer function of the PID controller is given in its series form and parallel form as: (Astron and Hagglund, 1995)

$$\frac{U(s)}{E(s)} = k_c \left(\frac{\tau_I s + 1}{\tau_I s} \right) \left(\frac{\tau_D s + 1}{\alpha \tau_D s + 1} \right) \quad (2.23)$$

$$\frac{U(s)}{E(s)} = k_c \left(1 + \frac{1}{\tau_I s} + \tau_D s \right) \quad (2.24)$$

2.13 Controllability Analysis

Controllability analysis is an evaluation of how well a control structure was able to achieve the system's operational target (performance objective). For the riser pipeline system controllability analysis to be evaluated, defined control objective has to be specified which has practical relevance to oil and gas production. The control system that has practical relevance to oil and gas operation should achieve stable operation and optimize (increase) production. In other to achieve this operational target, a control variable that has the capability to stabilise the unstable system at large valve opening is considered of practical interest for optimal production (Ogazi et al, 2011). In this work, riser base pressure, as control variables was used to evaluate the capability of achieving the specified control objective of stable flow at large valve opening and minimised steady state error.

2.14 Control-System Structure

The controller is a parallel PID architecture connected as shown figure 2-7.

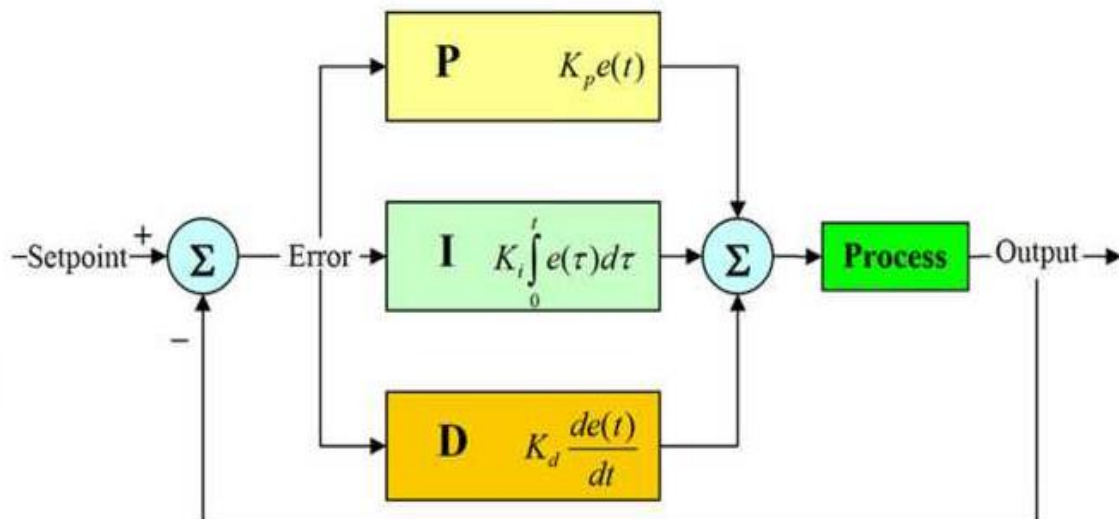


Figure 2-7 Parallel PID architecture connection (Math Works).

The green block is the system (plant) to be controlled while the other blocks are the controller. The PID system works on the error signal, which is the difference between the measured value and the desired set-point to obtain an error (e). The error signal is multiplied by the proportional gain K_P to get the proportional term (P), integrated and multiplied by the integral gain K_I to get the integral term (I) and differentiated and multiplied by the derivative gain K_D to get the derivative term (D). The error values are then summed up to give the control request that is keyed to the software request to the actuator as in figure 2-7 (Math Works).

The task is to choose controller architecture (combination) appropriate for our system and to estimate the gain values that will produce the controller response that can produce zero steady state error or minimize the steady state error as much as possible and increase the system stability as in equation 2.25.(Olga Manual).

$$u = k_c \left(e + \frac{1}{\tau_I} \int_0^t (\tau) d\tau + \tau_D \frac{de}{dt} \right) + \text{bias} \quad (2.25)$$

From the bifurcation map shown figure 4.5, the critical valve position 5% is the initial value (bias) to be implemented in equation 2.25. This is the open-loop stable valve position at which the PID is tuned.

2.15 Conclusion

Slug flow has been defined as a flow assurance challenge in multiphase transport. Slug is the intermittent flow regime in which large bubbles of gas flow alternately with liquid slugs at randomly fluctuating frequency in pipeline. Flow regime determination in the other hand is a critical issue in the analysis of multiphase flow. In cases where the flow regime happens to be near the border between two or even three different flow regimes, the uncertainties are generally most significant and minor changes in the flow properties or inclination angle is likely to change the flow regime. The main mechanism at work in switching from one flow regime to another is thought to be the Bernoulli effects, which reduces pressure when the gas velocity is increased. The flow regimes in

horizontal pipes differ from the flow regimes in vertical pipes as bubble flow does not usually exist in small diameter vertical pipes. The traditional model by Taitel and Dukler are the most applied in flow regime prediction.

The terminologies used in multiphase literatures were also presented in this chapter. Publication review of slug control techniques were discussed, showing the evolution of the different techniques that can be used to control slug flow problems and these include reduction in pipeline diameter, slug catcher, splitting of the flow into multiple streams, choke valve technology, gas injection technology, combination of gas injection and topside choke valve, active feedback control, flow line modification and layout or geometry of the flow line to avoid a dip and multivariable control. Each method has its own limitation and capabilities. In order to be able to control the riser base pressure, a control objective most relevant to oil and gas operation was defined. The controller that can achieve the systems operational target to stabilise the system at a valve opening larger than manual choking with zero steady state error or minimized steady state error was needed.

3 MODELLING THE CASE PROBLEM

An industrial scale case study of 6km flow-line and 46.2m high riser originally developed by Burke and Kashou 1996 was modelled in Olga 7.1.3 by (Hazem, 2012) and adapted for the current work for the investigation of the effectiveness of feedback control for hydrodynamic slug mitigation with pressure variation measurement used to analyse the performance of the system using pressure transmitter PT at the riser base as controlled variable. Whereas (Hazem, 2012) model investigated the use of gas injection to control hydrodynamic slugging, the current work applied topside choking with pressure transmitter PT at the riser base to investigate the effectiveness of active feedback control to mitigate hydrodynamic slugging.

The research integrates active feedback control for hydrodynamic slug control using topside choke valve to assure smooth flow and improve oil production and recovery.

3.1 Building Olga Model for the Numerical Simulation.

Olga model for the numerical simulation was built using the Burke and Kashou (1996) model as a starting point.

3.2 Introduction.

Numerical simulation is a machine thinking approach in predicting transient multiphase flow behaviour in pipeline. A number of software is available in the market to deal with numerical analysis of multiphase problems. OLGA is one of the most used and tested software in the market. Olga 7.1.3 is used in this thesis to study the effectiveness of feedback control and choking at the topside to mitigate hydrodynamic slugging.

- A case study of West African platform suffering hydrodynamic slug flow was described by (Burke and Kashou, 1996). The paper was used as starting point to build an Olga model. The aim was to obtain result similar to that of (Burke and Kashou 1996), observing how well matched is the holdup at the bottom of the riser as a validation of the model.

- Manual choking of the valve opening was investigated till stability was attained. The maximum percentage valve opening to attain stability was recorded. Stabilisation is attained when the holdup and pressure oscillation at the riser top and riser base are reduced or eliminated.

A Hopf bifurcation map of the manual choke was generated from simulation and a PI controller was designed at the critical valve position.

3.3 Simulation Start Point.

A real case problem was extensively described by Burke and Kashou of an offshore platform suffering hydrodynamic slug located at West Africa. This case problem was used as starting point to model the Olga case. The detail of the case is explained hereunder.

3.4 Pipeline Inlet Flow Rate:

Oil production 5,318 stb/d.

Gas production 5.351MMscf/d.

Water production 257stb/d.

Liquid production 5,575stb/d (oil plus water = 5,318+257).

Gas Oil Ratio (GOR) 1,006scf/stbo.

Gas Liquid Ratio (GLR) 960 scf/stbl.

Water-cut 4.61%.

Oil gravity 31.9° API.

Liquid production, GOR, percentage water-cut and oil gravity is used in the Olga model, while the rest parameters are obtained from these parameters and PVT table.

Table 3-1 shows the fluid composition as applied in the fluid PVT calculations.

Table 3-1 Burke and Kashou (1996) fluid PVT composition.

Component	Mole fraction %
C ₁	45.88
C ₂	6.64
C ₃	4.72
i-C ₄	1.2
n-C ₄	2.13
i-C ₅	1.21
n-C ₅	1.12
C ₆	2.03
C ₇	2.98
C ₈	3.62
C ₉	2.98
C ₁₀	2.67
C ₁₁	2.26
C ₁₂ ⁺	19.01
CO ₂	0.19
N ₂	0.59
Total	99.23
Sub-total C ₁ to n-C ₅	63.68%

Gas mole fraction in the fluid composition is the sum of mole fraction of C₁ till n-C₅ in Table 3-1=63.68%.

CO₂ mole fraction in gas = $0.19/63.68 \times 100=0.3\%$.

N₂ mole fraction in gas = $0.59/63.68 \times 100=0.93\%$.

3.5 Pipeline Inlet Condition.

The pipeline inlet condition stated below adapted for the investigation was initialised in the Olga model window for the numerical simulation.

Pressure in the range 20.3-21.0 bar.

Temperature 83.3°C.

3.6 Pipeline Outlet Condition.

In a similar vein the outlet condition contained below adapted for the investigation was initialised in the Olga window to specify the outlet condition for the numerical simulation.

Pressure in the range 11.3-14.8 bar.

Temperature 23.9°C.

3.7 Burke and Kashou (1996) Pipeline Profile.

Detail of (Burke and Kashou, 1996) case study platform profile inlet condition is explained in figure 3-1 as adapted for the analysis. The case problem definition, inlet and outlet condition parameters are calculated and initialised in the Olga window.

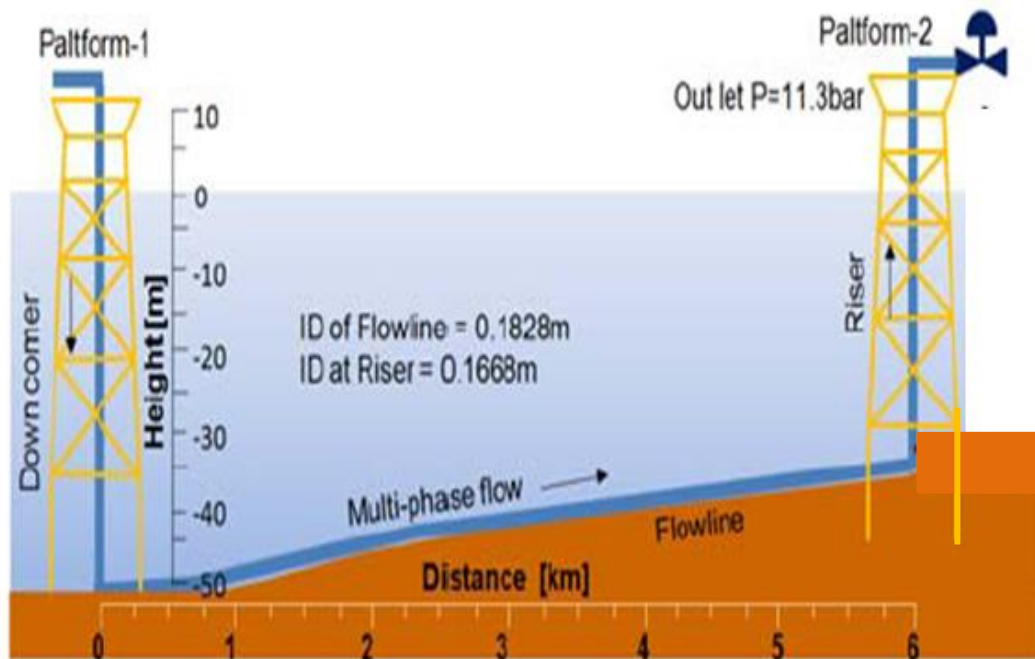


Figure 3-1 Schematic diagram of pipeline adapted from (Hazem, 2012) with choke valve at the topside used to analyse the performance of the system using topside choke valve at liquid source flow rate of 5,575stb/d, GOR 1006 and water-cut 4.61%.

The pipeline profile consists of 59.7m down-comer, 11m above the sea level, 6km flow line and 46.2m high riser.

The pipeline outlet is at 12.2m above the sea level. The surrounding condition of the sea water temperature is 22° C. It is mentioned that the pipeline is not buried and roughness is assumed to be 0.0018"(0.04572mm).

Figure 3-2 gives the detail of the pipeline components and conditions.

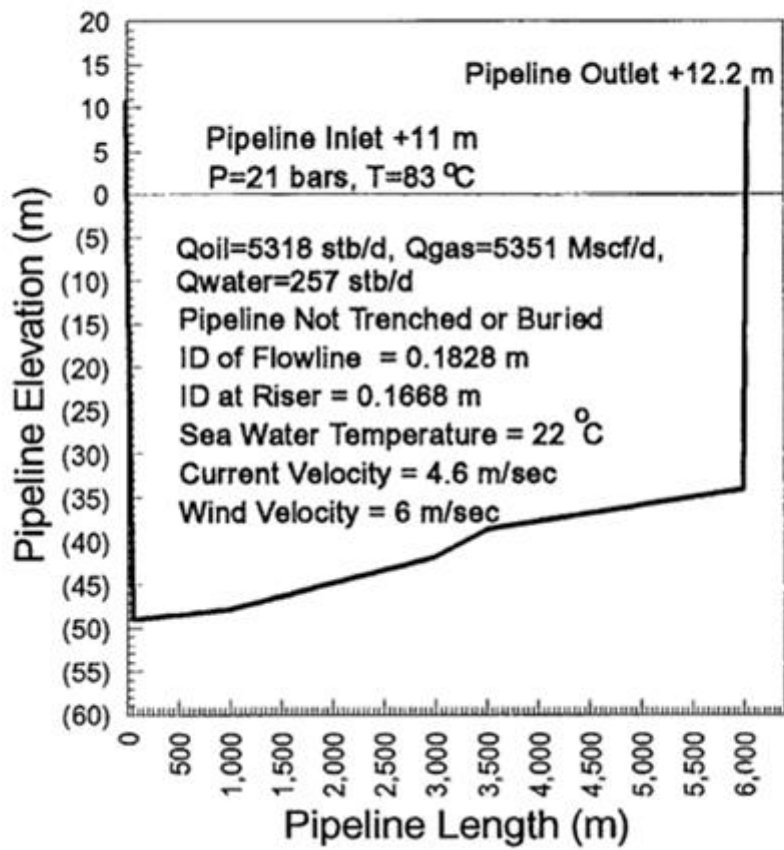


Figure 3-2 Down-comer, flow line and riser profile (Burke and Kashou 1996).

Table 3-2 gives more detail of the pipeline component ID, sections, length and elevations.

Table 3-2 Burke and Kashou (1996) Pipeline Details.

Pipe Description	Pipe ID (m)	Number of sections	Pipeline Length(m)	Pipe Elevation(m)	
Down comer	0.1668	2	2x29.85	-59.7	
Flowline-1	0.1828	10	50,90,8x100	1.2	
Flowline-2	0.1828	20	20x100	6.0	
Flowline-3	0.1828	5	5x500	3.2	
Flowline-4	0.1828	26	24x100,2x50	4.6	
Riser	0.1668	2	2x23,1	46.2	

3.8 Basic Olga Model “Texaco”.

Burke and Kashou (1996) using black oil composition was used to configure Olga model.

The pipeline setup comprise two layers of carbon steel 3.5mm thick each and an insulator two layers of poly propylene 5mm thick each. The properties of the pipe material was taken as default values assigned by Olga.

Carbon steel properties are:

Thermal capacity 470 [J/kg.°C].

Thermal conductivity 45 [W/m.K].

Density 7850 [kg/m³].

Figure 3-3 shows the wall properties and insulation.

MATERIAL : Carbon Steel		MATERIAL : Poly Propylene	
General		General	
LABEL	Carbon Steel	LABEL	Poly Propylene
CAPACITY	470 [J/kg-K]	CAPACITY	2000 [J/kg-K]
CONDUCTIVITY	45 [W/m-K]	CONDUCTIVITY	0.17 [W/m-K]
DENSITY	7850 [kg/m3]	DENSITY	750 [kg/m3]
TYPE	SOLID	TYPE	SOLID
TEMPERATURE	[C]	TEMPERATURE	[C]
Fluid		Fluid	
RADIATION	NO	RADIATION	NO
INNEREMISSIVITY	0	INNEREMISSIVITY	0
OUTEREMISSIVITY	0	OUTEREMISSIVITY	0
EXPANSION	[1/C]	EXPANSION	[1/C]
VISCOSITY	[CP]	VISCOSITY	[CP]
Phase change parameters		Phase change parameters	
PHCHMAX	[C]	PHCHMAX	[C]
PHCHMIN	[C]	PHCHMIN	[C]
CONDMULT		CONDMULT	
HCAPMULT		HCAPMULT	
FUSIONMULT		FUSIONMULT	
Elasticity		Elasticity	
EMOD	[bar]	EMOD	[bar]

Figure 3-3 Properties of carbon steel and poly propylene.

Poly propylene properties are:

Thermal capacity 2000 [J/kg.°C].

Thermal conductivity 0.17[W/m.K].

Density 750 [kg/m³].

Pipeline was named Wall-1 with the properties shown in figure 3-4.

WALL : WALL-1	
General	
LABEL	WALL-1
THICKNESS	2:3.5, 2:5 [mm]
MATERIAL	Carbon Steel, Car
ELECTRICHEAT	OFF
DISCRETIZATION	OFF
Heated walls	
POWERCONTROL	OFF
POWERMAX	[W/m]
CONTROLLERLABE	
TIME	0 [s]
POWER	[W/m]
WALLAYER	
Automatic discretization	
MAXRATIO	2 [-]
INNERDIAMETER	[m]
MAXNOLAYERS	
Elasticity	
ELASTIC	OFF
ERATIOMIN	0.1 [-]
KAPPA	[1/bar]

Figure 3-4 Pipeline wall properties.

An inlet source named oil at the first section of the pipeline was configured as closed node, implying that analysis was from the wellhead only while the pipeline outlet was configured as pressure node with pressure set at 11.3bar and temperature set at 23.9°C. Figure 3-5 shows the pipeline configuration with the nodes and the source inlet.

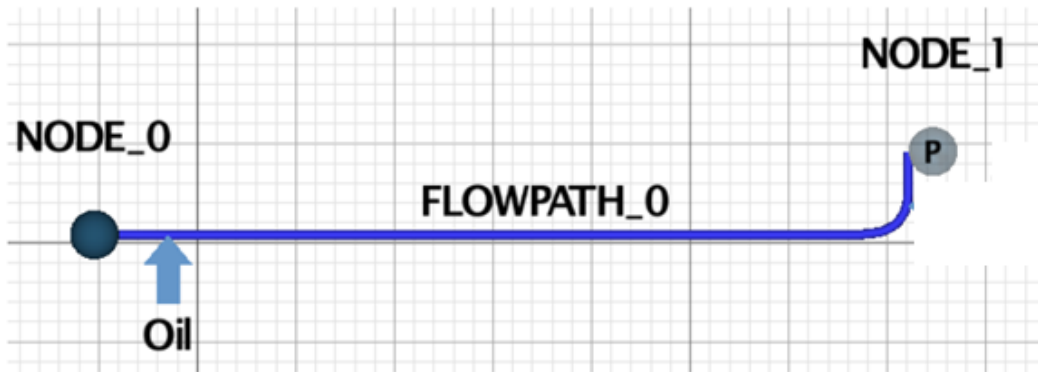


Figure 3-5 Schematic diagram of Olga model with the nodes and source inlet.

Figure 3-6 shows the node properties.

NODE : NODE_0		NODE : NODE_1	
General		General	
LABEL	NODE_0	LABEL	NODE_1
INFO		INFO	
TYPE	CLOSED	TYPE	PRESSURE
X	0 [m]	X	0 [m]
Y	0 [m]	Y	0 [m]
Z	0 [m]	Z	0 [m]
MASSFLOW	[kg/s]	MASSFLOW	[kg/s]
VOLUME	-1 [m3]	VOLUME	-1 [m3]
Server		Server	
DPDGG	0 [Pa-s/kg]	DPDGG	0 [Pa-s/kg]
DPDGLTHL	0 [Pa-s/kg]	DPDGLTHL	0 [Pa-s/kg]
DPDGLTWT	0 [Pa-s/kg]	DPDGLTWT	0 [Pa-s/kg]
EXPOSE		EXPOSE	
Single phase		Single phase	
LINE	NO	LINE	NO
MAXPRESSUREBOC	[bara]	MAXPRESSUREBOC	[bara]
FLUIDTYPE		FLUIDTYPE	
Compositional		Compositional	
GASFRACEQ	1 [-]	GASFRACEQ	1 [-]
OILFRACEQ	1 [-]	OILFRACEQ	1 [-]
WATERFRACEQ	1 [-]	WATERFRACEQ	1 [-]
FEEDNAME		FEEDNAME	BOFEED1

Figure 3-6 Node properties.

3.9 Geometry of the Pipeline.

The geometry of the pipeline is shown in figure 3-7 with the components.

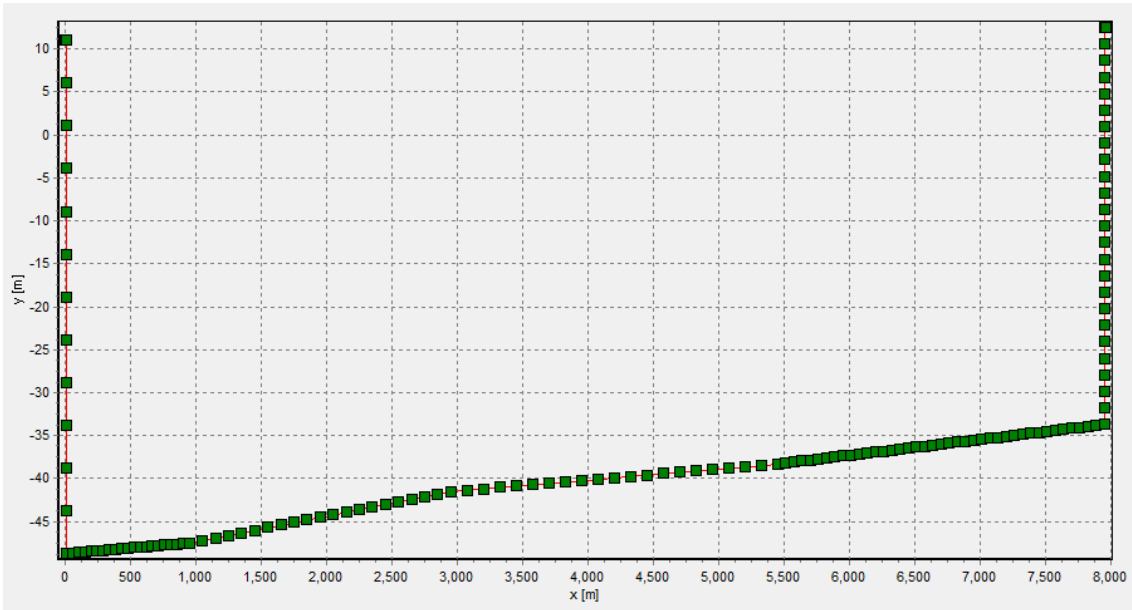


Figure 3-7 Geometry of the pipeline model Burke and Kashou 1996.

Table 3-3 Detail of pipeline geometry.

Pipe	X[m]	Y[m]	Length[m]	Elevation[m]	#Sections	Length of sections(list[m])	Diameter[m]	Roughness[m]	Wall
Start Point	0	11							
Pipe-1	10	11	10	0	4	4:2,5	0.1668	4.5672e-005	Wall-1
Pipe-2	10	-48.7	59.7	-59.7	12	12:4,975	0.1668	4.5672e-005	Wall-1
Pipe-3	949.999	-47.5	940	1.2	20	20:47	0.1828	4.5672e-005	Wall-1
Pipe-4	2949.99	-41.5	2000	6	20	20:100	0.1828	4.5672e-005	Wall-1
Pipe-5	5449.99	-38.3	2500	3.2	20	20:125	0.1828	4.5672e-005	Wall-1
Pipe-6	7949.99	-33.7	2500	4.6	40	40:62,5001	0.1828	4.5672e-005	Wall-1
Pipe-7	7949.99	12.5	46.2	46.2	24	24:1,925	0.1668	4.5672e-005	Wall-1
Pipe-8	7959.99	12.5	10	0	4	4:2,5	0.1668	4.5672e-005	Wall-1

From table 3.3 the pipe diameter in the flow line, riser and down-comer are 0.1828m, 0.1668m and 0.1668m respectively.

3.10 Fluid Composition.

Black oil compositions of three components (gas component, oil component and water component) were created as contained in the PVT fluid file. Black oil

composition was adopted when a detailed fluid property is not available from the laboratory. The following components were specified from the fluid property: gas component: specific gravity 1.732, CO₂ mole fraction 0.3%, H₂S mole fraction 0% and N₂ mole fraction 0.93%

BLACKOILCOMPONENT : Gas		BLACKOILCOMPONENT : Oil	
General		General	
LABEL	Gas	LABEL	Oil
TYPE	GAS	TYPE	OIL
GASSPECIFICGRAV	1.732	GASSPECIFICGRAV	0.64
OILSPECIFICGRAV	0.876	OILSPECIFICGRAV	0.876
APIGRAVITY	30	APIGRAVITY	31.9
WATERSPECIFICGR	1	WATERSPECIFICGR	1
H2SMOLEFRACTIO	0 [-]	H2SMOLEFRACTIO	0 [-]
CO2MOLEFRACTIC	0.3 [%]	CO2MOLEFRACTIC	0 [-]
N2MOLEFRACTION	0.93 [%]	N2MOLEFRACTION	0 [-]

BLACKOILCOMPONENT : Water		BLACKOILOPTIONS	
General		General	
LABEL	Water	GORMODEL	STANDING
TYPE	WATER	RSGO_BP-TUNING	OFF
GASSPECIFICGRAV	0.64	OILVISC-TUNING	OFF
OILSPECIFICGRAV	0.876	GOR	1006 [scf/STB]
APIGRAVITY	30	GASSPECIFICGRAV	
WATERSPECIFICGR	1	OILSPECIFICGRAV	
H2SMOLEFRACTIO	0 [-]	APIGRAVITY	
CO2MOLEFRACTIC	0 [-]	BUBBLEPRESS	[bara]
N2MOLEFRACTION	0 [-]	BUBBLETEMP	[C]
		OILVISC	[CP]
		VISCTEMP	[C]
		VISCPRESS	[bara]

Figure 3-8 Properties of black oil components.

Oil component API 31.9° gravity and water component with specific gravity 1 were initialised in the model. Black oil option was initialised STANDING so that the correlation used to calculate gas/oil ratio shall be taken as default from Olga model. Black oil feed (BOFEED-1) the well production feed which consists of

three components Oil/Gas/Water with a water- cut of 4.61% and gas oil ratio GOR of 1006scf/stb were created. The feed properties are shown figure 3-9.

Properties	
BLACKOILFEED : BOFEED1	
General	
LABEL	BOFEED1
OILCOMPONENT	Oil
GASCOMPONENT	Gas
WATERCOMPONENT	Water
WATERCUT	4.61 [%]
GOR	1006 [scf/STB]
GLR	[Sm3/Sm3]
OGR	[Sm3/Sm3]
LGR	[Sm3/Sm3]

Figure 3-9 Properties of the black oil feed.

3.11 Feed Source

Feed source are assigned to the pipeline with oil installed at the first section of pipe-1. The well feed BOFEED-1 was assigned to this source with liquid production of 5,575stb/d at a temperature of 83.3°C. Gas fraction, oil fraction and water fraction were kept as default value to take value from the fluid composition fraction figure 3.10.

Figure 3.10 shows the source properties.

SOURCE : OIL	
General	
LABEL	OIL
TIME	0 [s]
SOURCETYPE	MASS ▼
TEMPERATURE	83.3 [C]
MASSFLOW	[kg/s]
PRESSURE	[bara]
VALVEMODEL	HYDROVALVE ▼
Server	
Position	
PIPE	PIPE-1 ▼
SECTION	1 [max 4]
ABSPOSITION	[m]
POSITION	
Phase distribution	
Inhibitor tracking	
Standard conditions	
STDFLOWRATE	[Sm3/d]
PHASE	LIQUID ▼
GLR	[Sm3/Sm3]
GOR	1006 [scf/STB]
CGR	[Sm3/Sm3]
WATERCUT	0 [-]
WGR	0 [Sm3/Sm3]
MOLWEIGHT	[kg/mol]
Compositional	
FEEDNAME	BOFEED1 ...
FEEDMASSFLOW	[kg/s]
FEEDMOLEFLOW	[mol/s]
FEEDMASSFRAC	[-]
FEEDMOLEFRAC	[-]
FEEDSTDFLOW	5575 [STB/d]
FEEDVOLFRAC	[-]
GASFRACEQ	1 [-]
OILFRACEQ	1 [-]
WATERFRACEQ	1 [-]

Figure 3-10 Source properties.

3.12 Options and Integration.

Hydrodynamic slug tracking (HYDSLUG=ON) was turned on, while temperature calculation on heat transfer from inside pipe wall to the outside was applied. The rest of Olga values were kept as default, SLUGVOID=SINTEF. This correlation influenced transition from stratified flow to slug flow significantly unless slug tracking option is selected (Olga 7.1.3) figure 3.11.

OPTIONS	
General	
TEMPERATURE	WALL
STEADYSTATE	ON
SLUGVOID	SINTEF
FLASHMODEL	HYDROCARBON
WRITEPVTFILES	NO
PHASE	THREE
NOSLIP	OFF
COMPOSITIONA	BLACKOIL
WAXDEPOSITION	OFF
DRILLING	OFF
MASSEQSCHME	1STORDER
DEBUG	OFF
TRACERTRACKIN	OFF
TABLETOLERANC	OFF
HYDSLUG	ON
ELASTICWALLS	OFF
UDPLUGIN	OFF
Information	
DEBUGINFO	
Flow model	
FLOWMODEL	OLGA

INTEGRATION	
General	
ENDCONTROLLE	
ENDTIME	1000
MAXDT	0.5 [s]
MAXTIME	0 [s]
MINDT	0.001 [s]
MINTIME	0 [s]
RUNTIMESTEPAC	TRUE
STARTTIME	0 [s]
DTSTART	0.01 [s]
Server	
SIMULATIONSP	0
EXPOSE	
MAXLAGFACT	0.3

Figure 3-11 OLGA model options and integration.

3.13 Slug Tracking.

Hydrodynamic slug tracking initiated DELAYCONSTANT=150 by default as the number of pipeline diameter a slug will propagate before the next slug is initiated figure 3.12.

SLUGTRACKING	
Model	
LEVEL	OFF
HYDRODYNAMIC	ON
General	
GASENTRAINMENT	VOIDINSLUG
Limitations	
MAXNOSLUGS	
Level	
BUBBLEVOID	[-]
SLUGVOID	[-]
STARTTIME	[s]
ENDTIME	[s]
Hydrodynamic	
INITLENGTH	
INITFREQUENCY	[Hz]
DELAYCONSTANT	150
INITPOSITIONS	
INITBUBBLEVOID	1 [-]
INITSLUGVOIDS	0 [-]
INITPERIODS	[s]
INITZONELENGT	[m]
INITSTARTTIMES	[s]
INITENDTIMES	[s]

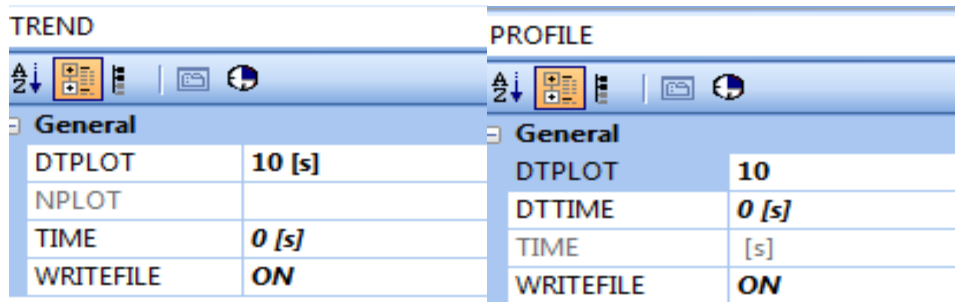
Figure 3-12 Properties of slug tracking options.

3.14 Output Options.

The output options were specified in the Olga window for the trend and profile plots.

3.14.1 Trend and Profile Properties.

The time interval between trend variable printout $DTPLOT=10[s]$ figure 3.13.



The image shows two side-by-side windows from the Olga software. The left window is titled 'TREND' and the right window is titled 'PROFILE'. Both windows have a 'General' section expanded, showing a table of properties. The 'TREND' window has a 'DTPLOT' value of '10 [s]', 'NPLOT' is empty, 'TIME' is '0 [s]', and 'WRITEFILE' is 'ON'. The 'PROFILE' window has a 'DTPLOT' value of '10', 'DTTIME' is '0 [s]', 'TIME' is '[s]', and 'WRITEFILE' is 'ON'.

TREND		PROFILE	
General		General	
DTPLOT	10 [s]	DTPLOT	10
NPLOT		DTTIME	0 [s]
TIME	0 [s]	TIME	[s]
WRITEFILE	ON	WRITEFILE	ON

Figure 3-13 Trend and profile properties.

3.15 Conclusion

An Olga model was built on the case study. The case definition statement, the inlet and outlet conditions, the fluid PVT file and the flow geometry were applied to calculate the parameters that were initialised in the Olga window to model the dynamic of the case problem in line with the field characteristics.

4 SLUG CONTROL DESIGN/TUNING.

4.1 Case Study

The industrial scale case study of 6km flow-line and 46.2m high riser was modelled in Olga 7.1.3 with pressure variation measurement used to analyse the performance of the system. The model was validated by comparing the holdup from the field case oscillation result with the holdup as calculated by Olga model to ascertain whether a tolerable matching trend result was achieved as shown in figure 4-1 for the field measurement and olga calculation respectively.

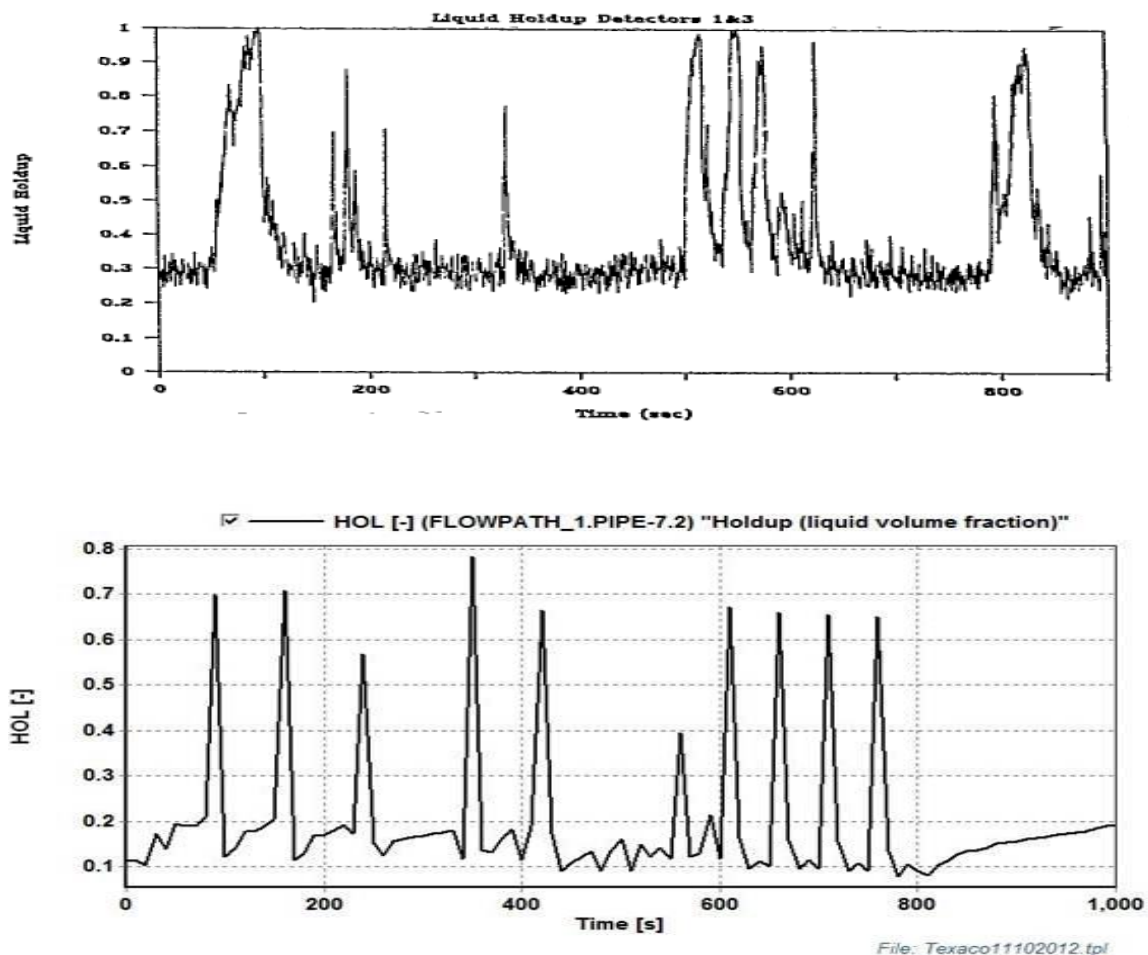


Figure 4-1 HOL field measurement with HOL as calculated by Olga model.at source liquid flow rate 5,575 STB/D, 1006scf/d GOR and 4.61% water-cut.

The results were found to match comparatively within an oscillation between 0.2 and 1.0 for the field measurement and between 0.1 and 0.8 for the Olga

calculation (both are in the range of 0.8) oscillation trend result and the model can be assumed valid and favourably matched. However, the field HOL measurement was 10% under predicted by the Olga calculation. The model is further validated by a profile plot of the flow regimes as calculated by Olga model figure 4-2. From the plot figure 4-2 the flow regime at inlet was annular (2) and as the fluid travels the length of the pipeline, the flow regime changed to slug flow (3) as can be seen in figure 4-2.

OLGA'

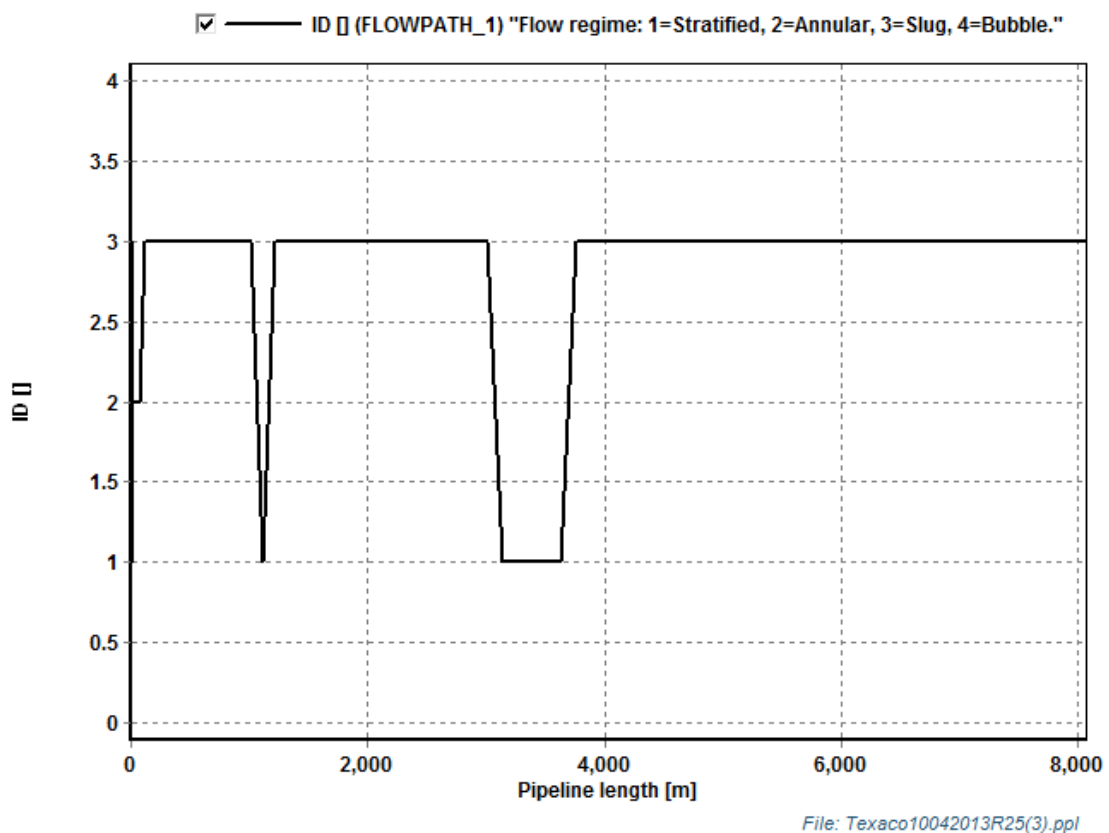


Figure 4-2 shows the flow regimes observed in the case platform indicating that flow is slug region dominated regime shown as 3.

Flow regime map for the riser using equations 2-6 to 2-9 is shown in figure 4.3. The variables values for these equations are obtained from Olga simulation at the operating point of 5,575stb/d liquid production,960scf/stb GLR and 4.61% water cut and it is marked red in the flow regime map. It can be observed from the map that the operating point marked red is within hydrodynamic slug region close to churn flow.

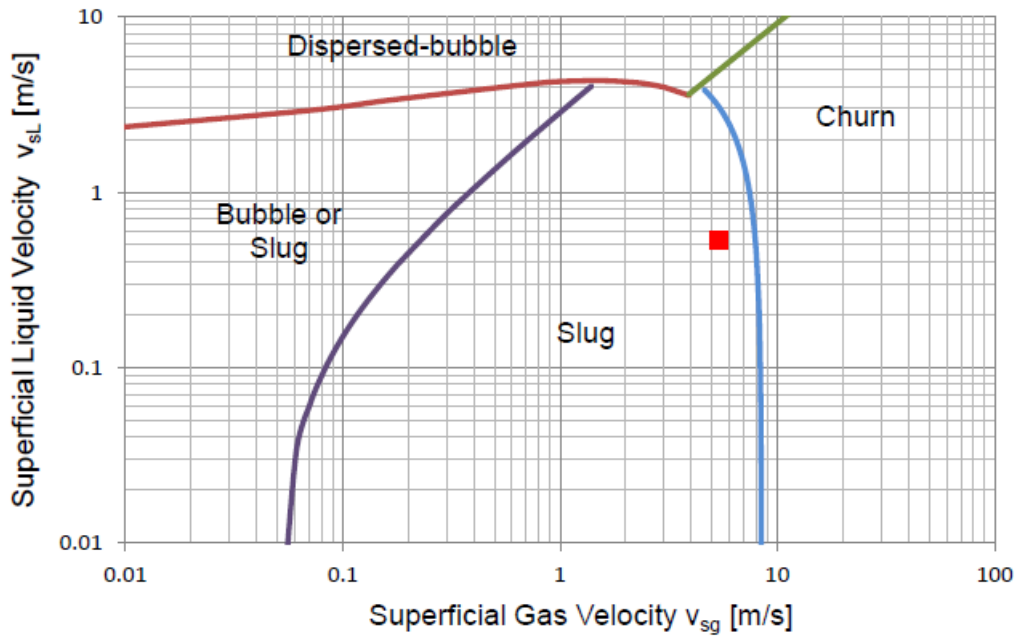


Figure 4.3 Flow regime map for the riser.' Texaco model'

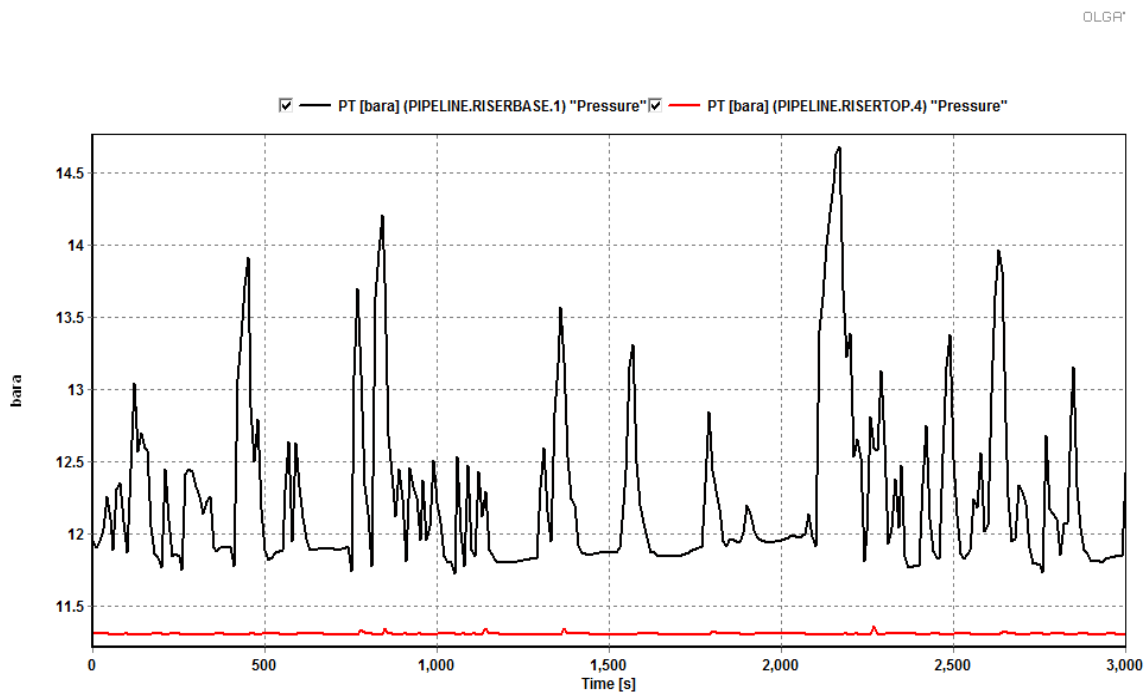


Figure 4-4 Pressure trend at the first section of the riser.

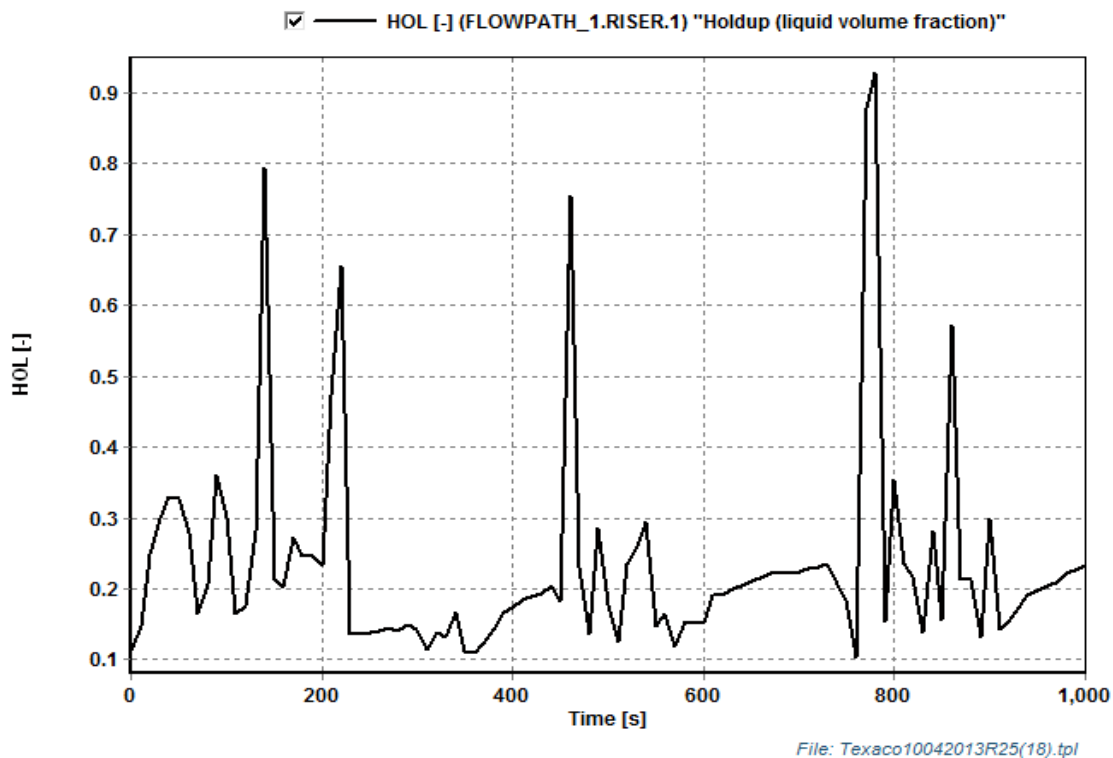


Figure 4-5 Holdup trend at the first section of the riser.

From figures 4-4 and 4-5, the system was observed to be highly unstable with pressure oscillating between 3bara and 14.8 bara figure 4-4 and holdup oscillating between 0.1 and 0.98 figure 4-5 respectively.

Through parametric study the matrix of the topside choke valve opening were [100, 90, 80, 70, 60, 50, 40, 30, 20, 18, 16, 14, 12, 10, 9, 8, 7, 6, 5, 4, 3, 2, 1].

From the plot of valve opening on the x-axis and pressure on the y- axis the bifurcation map figure 4.6 was generated.

4.2 Hopf Bifurcation Map

Hopf bifurcation occurs in a dynamic system, when the system loose stability due changes in the independent variable (Thompson and Stewart, 1986). For the riser pipeline system, Hopf bifurcation can occur if a change of the valve opening causes the system to become unstable at an operating point. Below this valve opening, the riser slugging does not exist and the flow is stable, but

pressure in the pipeline is considerably high for optimal production. This is the flow regime used when the choke valve opening is kept low as described in a bifurcation map. This implies that the point where slugging starts (onset of slugging) in open loop system (bifurcation point) is a specific parameter value where the qualitative behaviour of nonlinear differential equation system, changes from equilibrium solution to a periodic solution (Verhulst 1990). This unstable equilibrium is the operating point that was stabilized using feedback control. (See figure 4.6 Hopf bifurcation map of the industrial riser system pressure oscillation between a maximum (red line) and minimum (blue line) values shown in solid lines while the dotted (black line) represents the virtual steady state value). This bifurcation map was generated through simulation studies. The open-loop control of the industrial riser system requires the manual choke valve in order to transform the unstable flow condition in the system to stable flow condition.

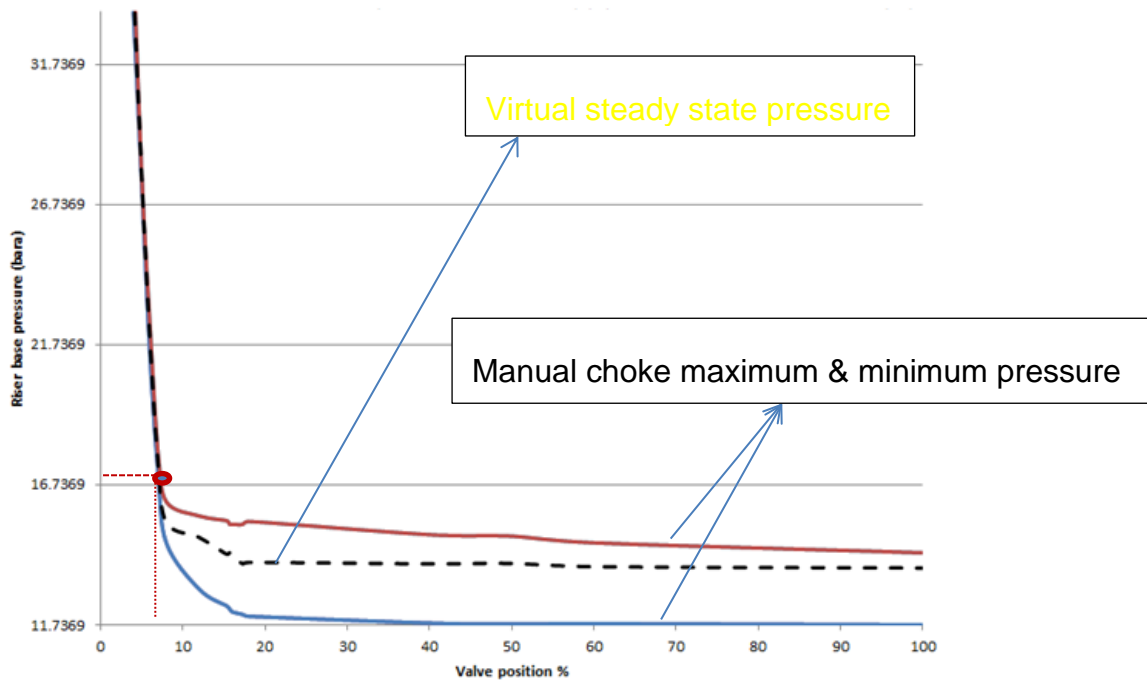


Figure 4-6 Hopf bifurcation map of the industrial riser system at liquid source flow rate 5,575std/d, GOR 1006 and 4.61% water-cut.

The bifurcation map indicates that the maximum valve opening corresponding to a stable system $u_{max} = 5\%$. For $u > u_{max}$ the system become unstable and

oscillates between a maximum and minimum pressure values. Thus u_{max} is known as the bifurcation point marked red in figure 4-6. The riser base pressure P_{RB} was calculated from the system for $5\% < u \leq 100\%$. The critical value indicated by the bifurcation map gives a minimum pressure 15.3881 bara P_{RB} and maximum u of the system to be stabilized by manual choking. The interest is to stabilise the system at unstable operating points, where the values of u are larger than this critical value such that the total pressure drop across the riser and the valve is reduced and thus the overall production is increased.

The Hopf bifurcation map shows the maximum valve opening that can stabilise the system (open -loop), a maximum manual valve opening of 5% was achieved. This valve opening is also known as the critical valve opening beyond which the system will be unstable as in figure 4.6.

At and below this valve position, slugging does not exist and the system can be operated open-loop stable without oscillation and without control. Above 5% valve opening, the system becomes unstable, with a pressure oscillation between a minimum and maximum pressure value as shown by the solid lines in figure 4.6.while the dotted line represent the virtual steady state pressure value.

4.3 Controller Design and Tuning

The controller was designed based on the critical values of the bifurcation map and subsequently tuned when the gain values have been determined.

4.3.1 Methods for Quantifying the Process Gain

The controller that has the capacity to stabilise the system at the predicted close-loop operating point to achieve the predicted optimal production is required. The Proportional, Integral and Derivative controller parameters were calculated with the control objective of a stabilised operation as well as optimized production. The method for quantifying the process gain is outlined thus:

4.3.1.1 Finding the Process Gain for Open-Loop Stable System.

The process gain values were determined from open-loop system using the process reaction curve.

4.3.1.1.1 Open Loop Tuning Rules (Process Reaction Curve).

The process reaction curve is an approximation model of the process, assuming the process behaves as a first order plus time delay system. The process reaction curve is identified by doing an open loop step test of the system and then identifying the process model parameters. The following steps were applied:

- Put the controller in manual mode
- Allow the process value (Y) to stabilise and not oscillating
- Step the output of the PI controller
- Collect data and plot the process reaction curve
- Repeat making the step in opposite direction
- $K = \text{process gain}; K = \frac{\text{Change in process value}}{\text{change in manipulated value}}$

$$K = \frac{\Delta Y}{\Delta U} \quad (4.1)$$

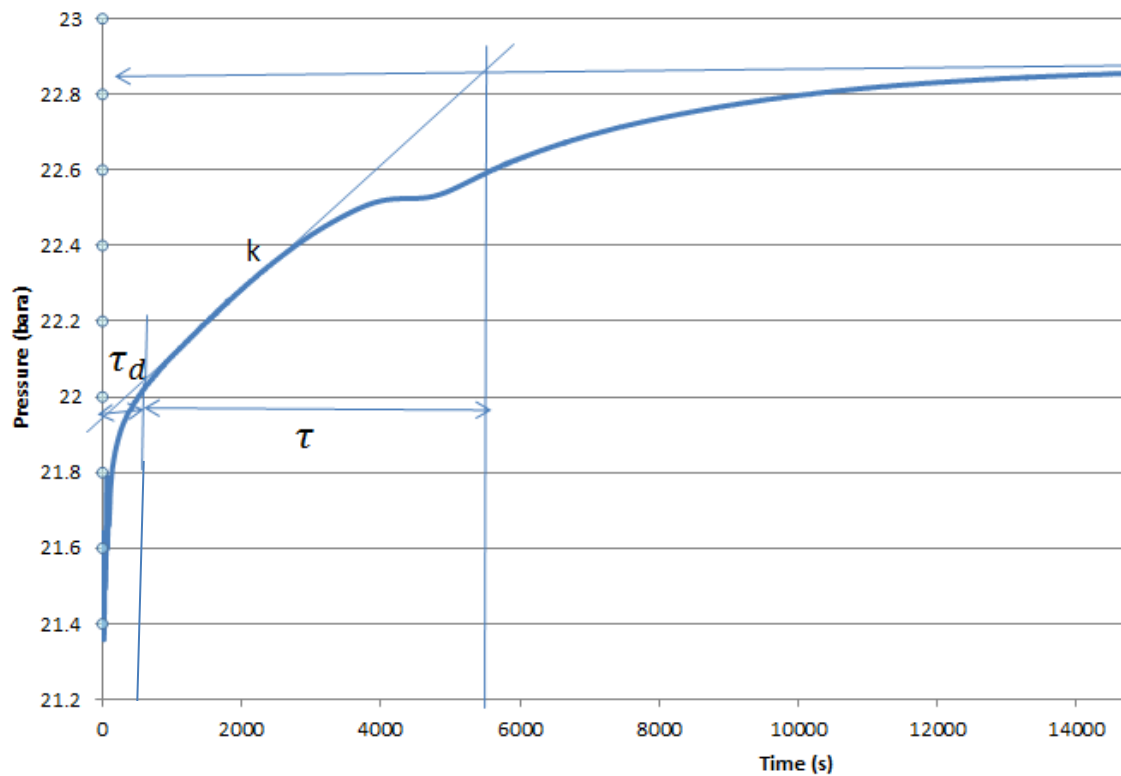


Figure 4-7 Process reaction curve $K = 75.73$, $\tau = 5700(\text{s})$, $\tau_d = 300(\text{s})$

The process parameters k was calculated from equation 4.1, while τ , τ_d as read from figure 4.7 and were then used to calculate the PI controller parameters according to the Ziegler-Nichols tuning rule as shown in table-4-1

Table 4-1 Ziegler-Nichols open-loop tuning rule.

Controller type	K_c	τ_I	τ_D
P	$\frac{1}{k} \left(\frac{\tau}{\tau_d} \right)$		
PI	$\frac{0.9}{K} \left(\frac{\tau}{\tau_d} \right)$	$3.33\tau_d$	
PID	$\frac{1.2}{K} \left(\frac{\tau}{\tau_d} \right)$	$2.0\tau_d$	$0.5\tau_d$

Recommended range of applicability $1.0 < (\tau_d/\tau) < 0.1$

4.4 Implementing Riser Base Pressure P_{RB} Control

The riser base pressure P_{RB} is the sum of the downstream pressure plus the hydrostatic pressure as a result of the weight of the riser content, friction loss and pressure due to acceleration in the riser (Storkaas, 2005). It has a very significant role in the slug control objective of stabilised flow and optimal production as in equation 4.2.

$$Q_T = \int_0^t G_M dt = B_0(P_0 - \bar{P}_{RB})T = Q_0 - Q_P \quad (4.2)$$

The target is to reduce the riser base \bar{P}_{RB} pressure and keep the pressure at the riser base at or below the average pressure in the riser slug cycle, thus preventing liquid accumulation or blockage of the flow line by manipulating the topside choke valve position to control the riser base pressure. The riser base and topside choke valve connection is shown figure 4-8

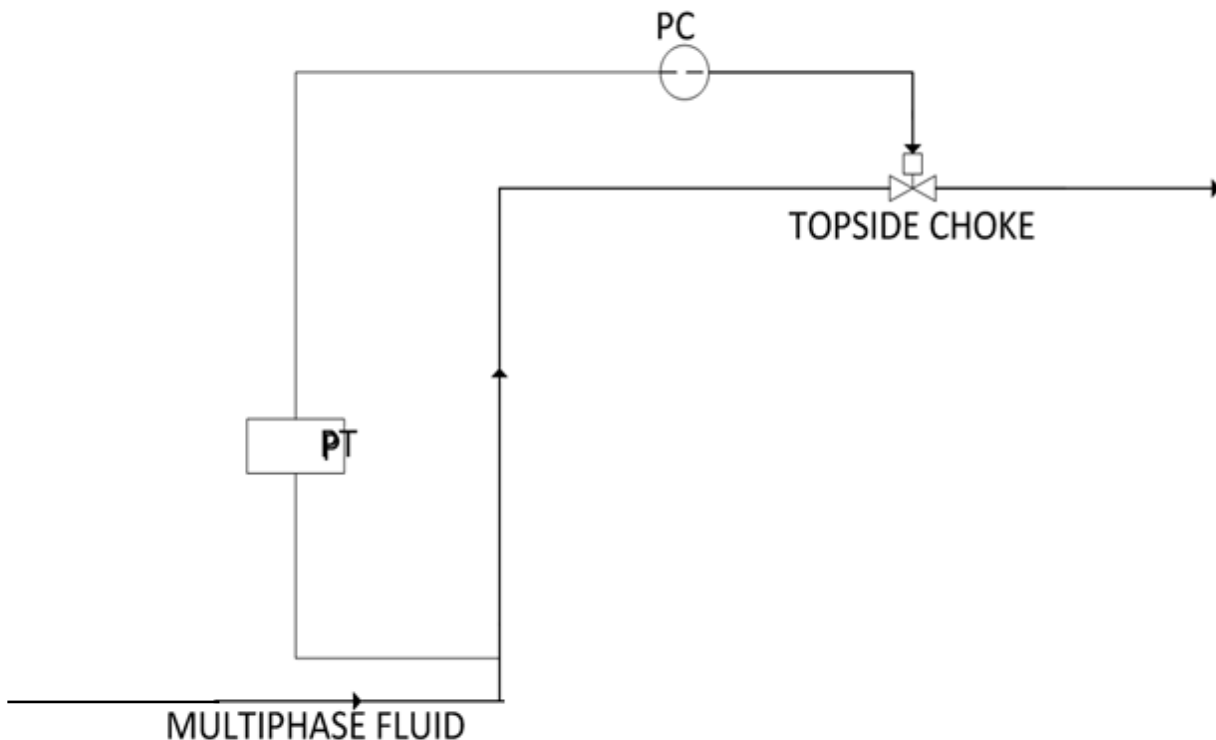


Figure 4-8 Riser base and topside choke valve connection.

The riser base measured pressure is transmitted through a pressure transmitter PT to the controller PC, which compares the measured pressure value with the desired set-point and sends an appropriate signal to the actuator (valve). The signal terminal is shown figure 4-9.

The riser base terminal is connected to the controller, whose terminal is in turn connected to the topside valve.

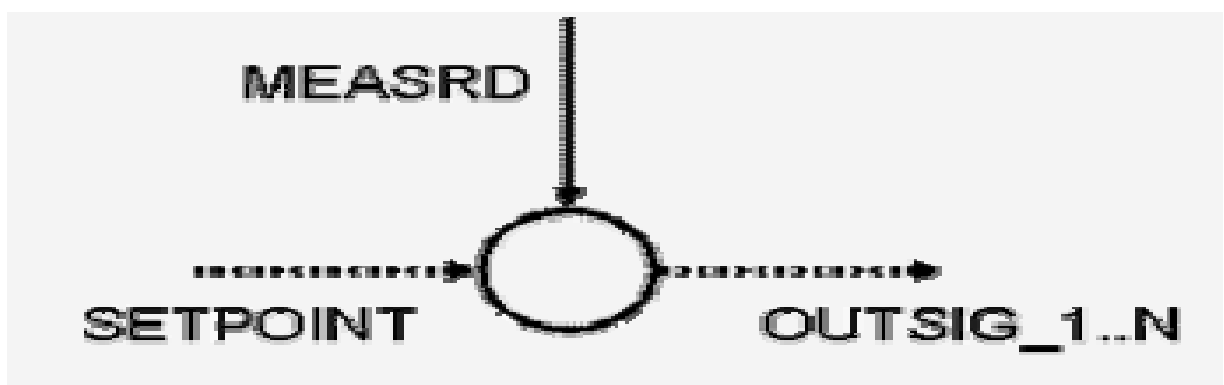


Figure 4-9 Signal terminals (OLGA Manual)

4.5 PID Tuning

Tuning is basically the process of finding the gain values (K_p K_I K_D) to meet the response time and overshoot (phase margin) specifications. The main approach to finding the gain values are: Manual tuning and Rule based tuning.

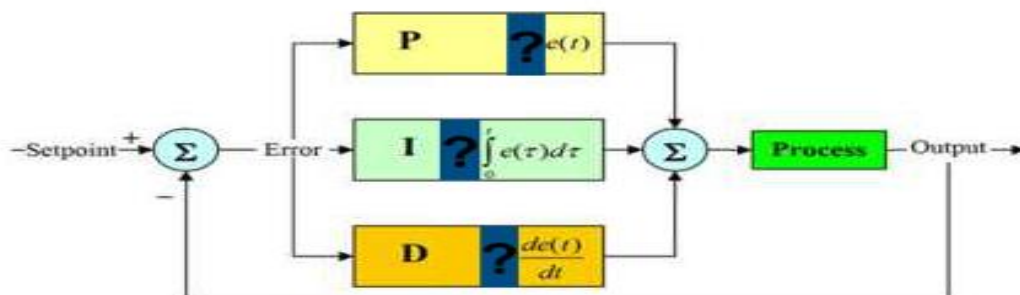


Figure 4-10 Estimating the PID gain values (Math Works, 2013).

- ❖ Manual tuning is purely based on trial and error process, time consuming, non-systematic and requires experience. It may not produce optimal design and may leads to dangerous conditions for the plant.

Table 4.2 shows the PI control parameters

Table 4-2 PI tuning parameters.

PIDCONTROLLER : PC	
General	
LABEL	PC
MAXSIGNAL	1
MINSIGNAL	0
MAXCHANGE	0.2
TIMESTEPCONTROL	ON
DELAY	0 [s]
DEFAULTINPUT	
INACTIVEMODE	
NORMRANGE	
STROKETIME	1 [s]
OPENINGTIME	10 [s]
CLOSINGTIME	10 [s]
Digital controllers	
SAMPLETIME	[s]
PID Parameters	
AMPLIFICATION	0.001
BIAS	0.05
DERIVATIVECONST	0 [s]
ERROR	0
INTEGRALCONST	100 [s]
Time varying	
TIME	0 [s]
SETPOINT	15.3881
MODE	AUTOMATIC
MANUALOUTPUT	
Server	

4.5.1 Open-Loop Tuning

The bifurcation valve position is set as the initial value at which the PI is tuned. The equivalent value of the controlled variable riser base pressure P_{RB} at the

critical point is taking as the pressure set-point. A step change from the initial valve position was applied and the proportional gain value K was calculated as in equation 4.1 and gradually increased until the system was unable to stabilise close-loop.

The process parameters $K = 75.73$, $\tau = 5700(\text{s})$, $\tau_d = 300 (\text{s})$ were then used to calculate the controller parameters from the PI controller tuning table 4-1 and the calculated values are applied to fine tune the PI for optimal performance.

- ❖ Rule based tuning algorithm: PI controller became popular due to the appearance of rule base tuning techniques such as the Ziegler- Nichols method. Rule based tuning also requires a lot of work and experience or simply cannot be applied in some open-loop unstable system. Once the gain value has been obtained in rule base gain value estimate, it cannot be fine-tuned to make the system to respond faster or to make the system more stable. Rule based tuning applies software with PI tuning algorithm and graphic user interface GUI

4.5.2 PID Tuning Algorithm and GUI

- It automatically finds the gain values to match specifications
- It provides additional tuning capability with simple slider
- It supports all types of plants including open-loop unstable plants

When more complex plant PI architectures are involved like multi-loop and multiple-input and multiple-output plants (MIMO) system,

- Existing methods such as successive loop closure requires a lot of time and experience and do not produce optimal results.
- Requires robust control toolbox
- Automatically tunes complex controllers
- The block to be tuned is only specified and the requirement

4.6 PI Implementation

- ❖ Conversion of design from continuous time to discrete time for implementation in a target micro- processor

- ❖ Scaling for fixed-point implementation. If the fixed-point data type is not scaled, the design that works well in floating point could be completely wrong on fixed-point implementation.
- ❖ Generates efficient C-code for the target micro-processor

4.7 Verify if the Design Works

- Testing before hardware prototypes are available.
- Testing corner cases
- Non-linear desktop simulation
 1. Testing different operating conditions.
 2. Testing different stages of design elaboration.
- Software-in-the loop testing
 - Use generated S-function as a way to run the generated controller code against plant model in Simulink.

4.8 Results:

The reference trajectory (controller) is the red line with time-delay necessary to approximate the model to first order system figure 4.11 plotted against the close-loop system response in black line. The gain values are adjusted until the steady state error was minimized to increase the system stability.

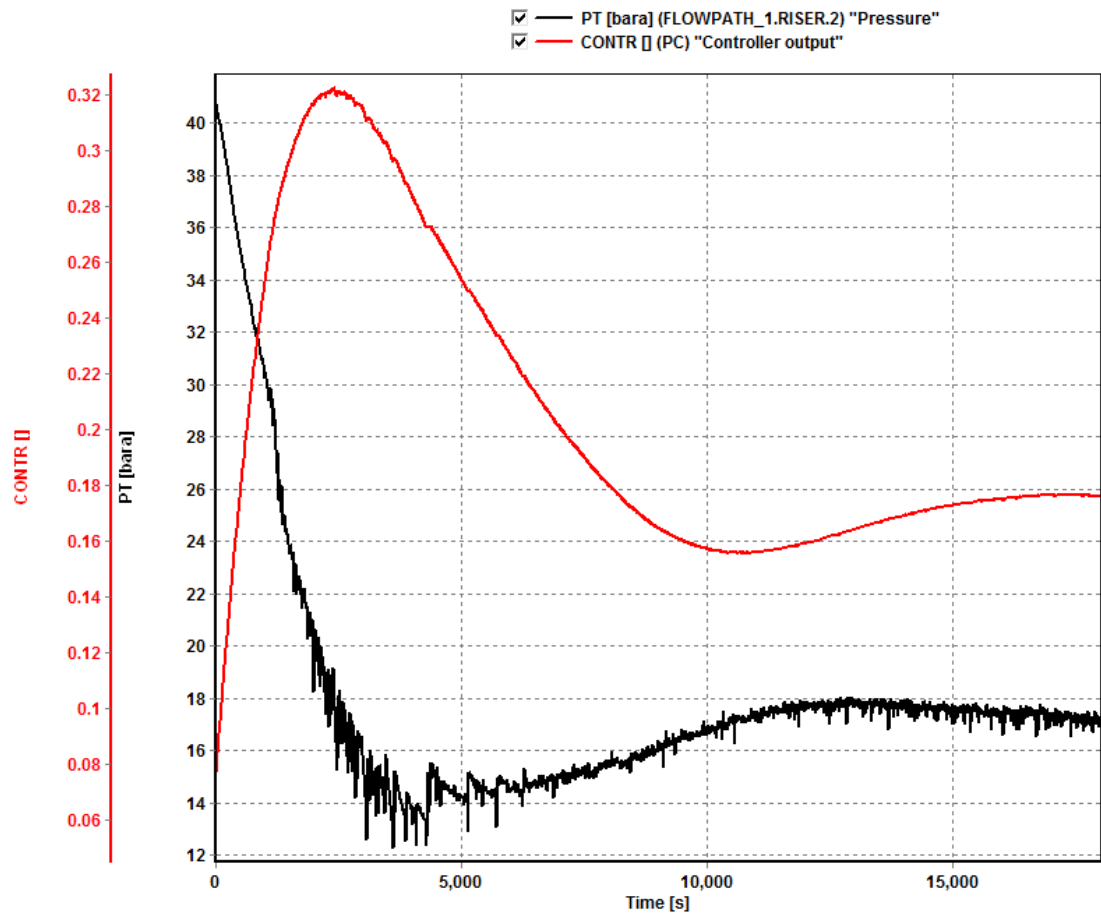


Figure 4-11 System riser base pressure response at pressure set-point of 15.3881 bara $K_c = 0.001$; $u=7.22\%$; $\tau_I = 100(s)$; $\tau_D = 0$; Measured $P_{RB} = 14.3883$ bara for 5hrs run time

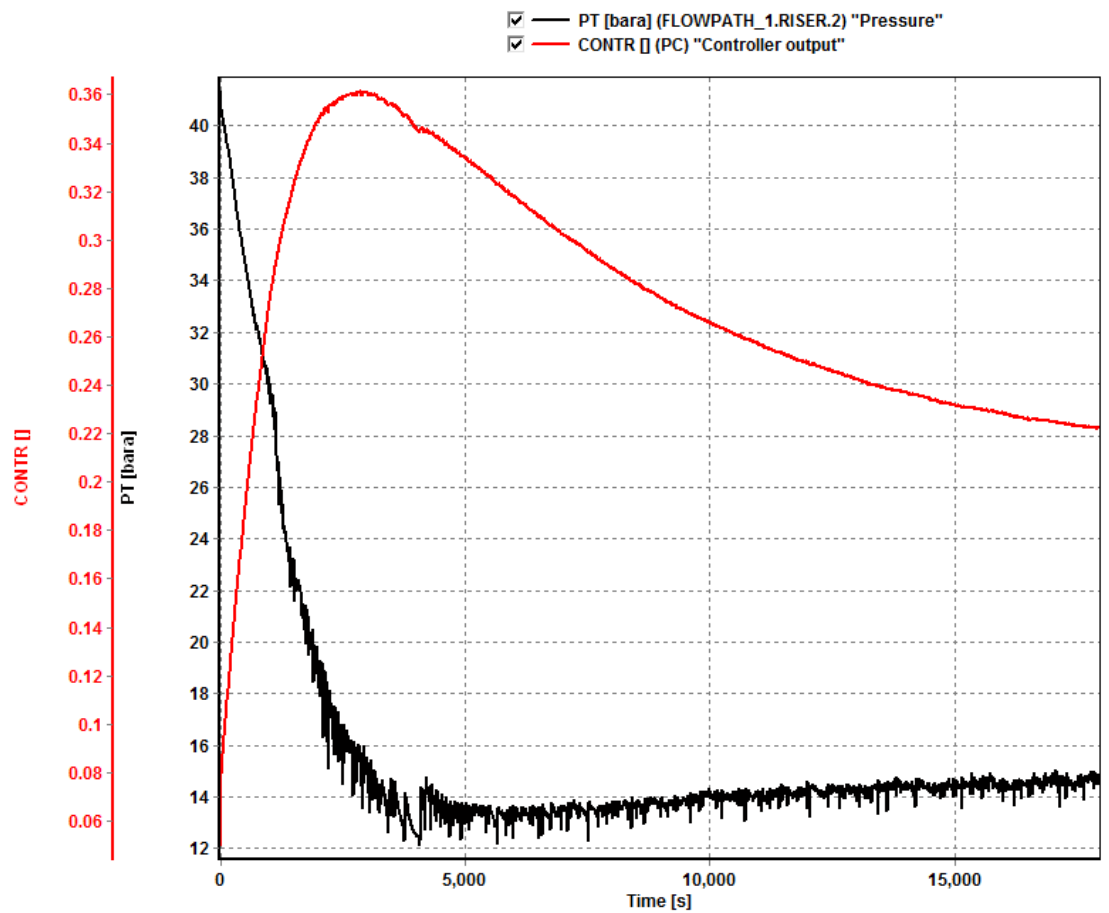


Figure 4-12 System riser base pressure response at pressure set-point of 14.3881 bara
 $K_c = 0.001$, $u=8.22\%$; $\tau_I = 100(s)$; $\tau_D = 0$; Measured $P_{RB} = 13.4016$ bara for
 5hrs run time

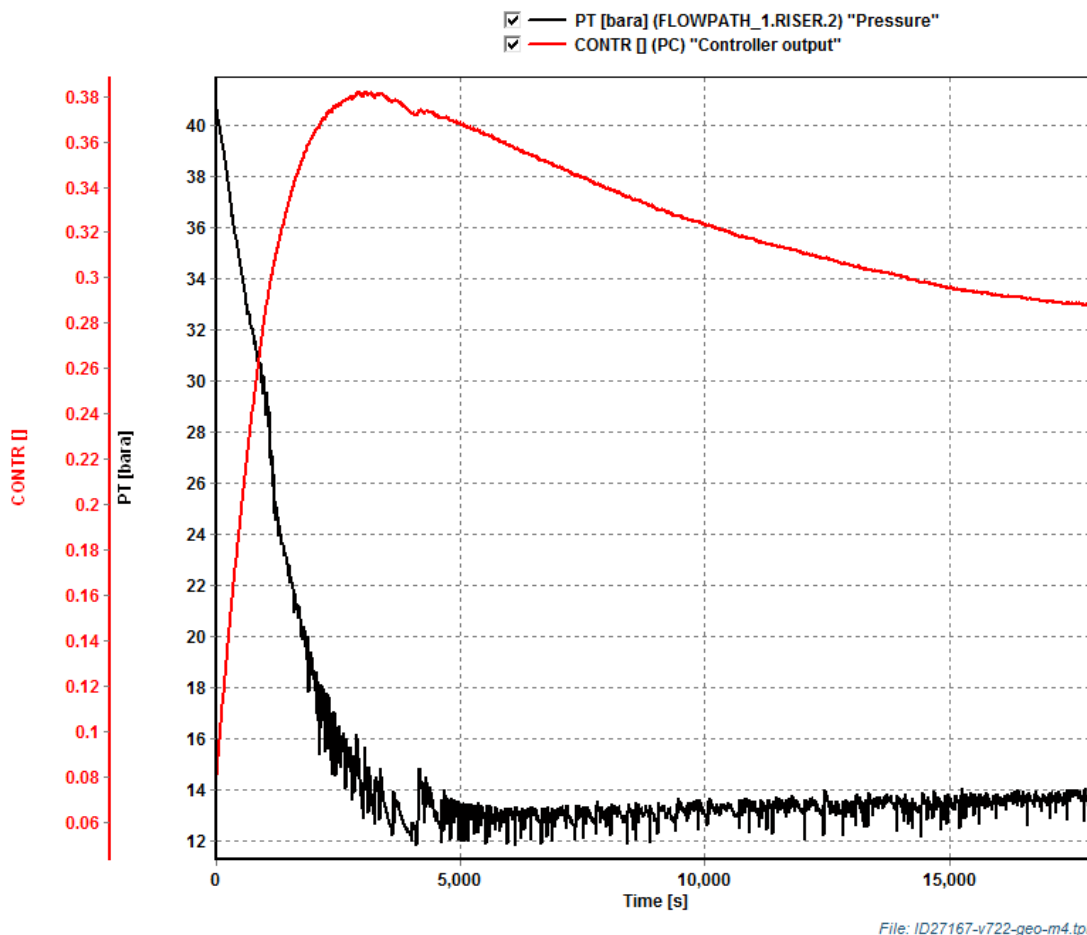


Figure 4-13 System riser base pressure response at pressure set-point of 13.3881 bara
 $K_c = 0.001$; $U=12.65\%$; $\tau_I = 100(s)$; $\tau_D = 0$; Measured $P_{RB} = 12.985$ bara for
 5hrs run time

4.9 Achievable Valve Opening to Set-Point Reduction.

The pressure set-point was initially set at 15.3881 bara

- The controller was then switched on for a 5hours simulation period. It was observed that the system was stabilised at this set-point and simulation period.
- Once the system is stabilised, the reference set-point is gradually reduced and the system is allowed to stabilise at each step reduction in the set-point.

- The gradual reduction in set-point yield a gradual increase in the valve opening at which the controller is stabilised.
- The reduction in set-point was continued until the system was unable to stabilise (limit of stability).
- The valve opening at which stability cannot be sustained was recorded as the achievable valve opening for the particular controller.

OLGA'

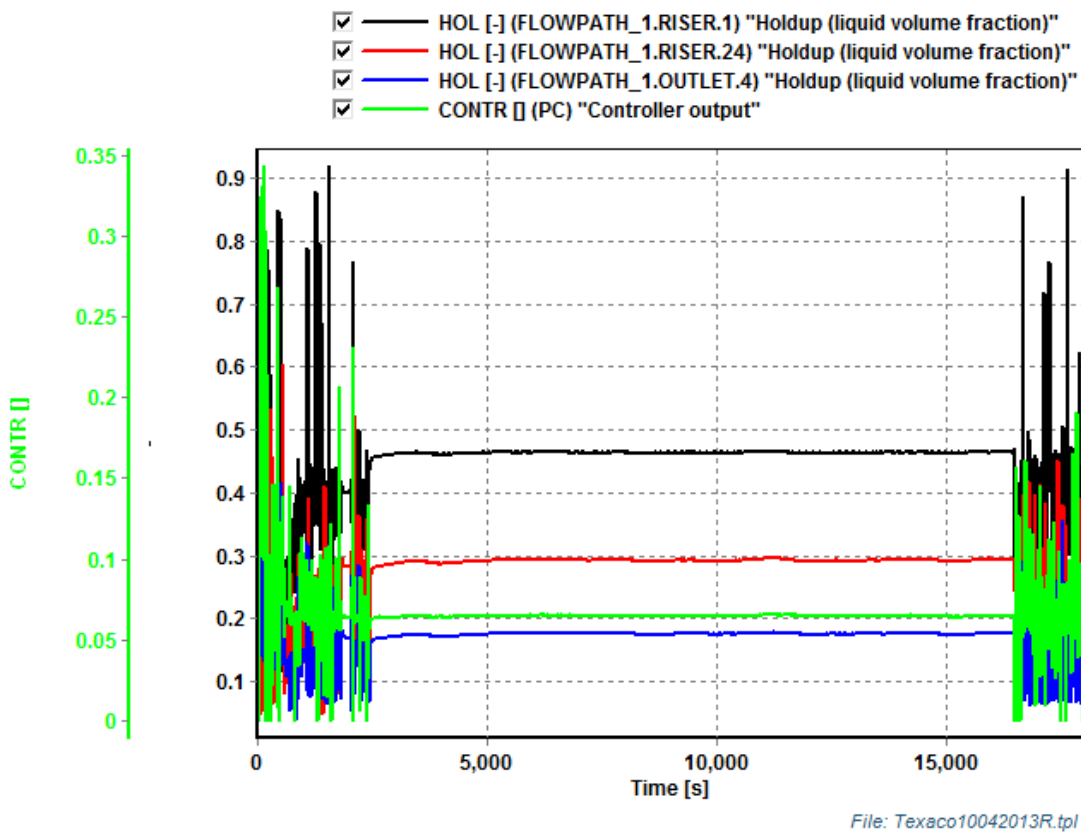


Figure 4-14 System HOL response at the riser base (0.438045); riser top (0.332417) and outlet 0.1117) at pressure set-point of 12.3881 bara $K_c = 0.001$; $u=12.65\%$; $\tau_I = 100(s)$; $\tau_D = 0$; Measured $\bar{P}_{RB} = 12.5056$ bara for 5hrs run time.

Beyond the achievable valve opening of 12.65%, the riser base slugging reappears and the system loses stability.

4.10 Loss of Stability and Continuous Oscillation.

When the stability limit of the system is exceeded the system oscillates continuously.

- As the proportional gain value K_p is increased the system stability is lost. The reduction in steady state error happens at the expense of the system stability, and the integral term was introduced to eliminate the steady state error, while the derivative term helps to increase the system stability.
- The task of estimating the gain values K_p K_I K_D from the system hardware is usually base on experience by trial and error and may not give an optimal gain values.
- Integrator windup is another challenge in tuning the PID controller. Integrator windup happens when the actuator winds fully open or fully closed and cannot get the desired set-point.
- Another issue is due to large integral value that needs to unwind which takes long time to unwind and this make the system unstable.
- Approximating derivative term (differentiating the error) introduces noise into the system, since system noise at high frequency is amplified when it is differentiated.
- Another issue encountered in PID usage is to be able to switch to the different forms P, PI and PID in ideal or parallel form, output saturation, and integrator anti-windup and bump-less transfer from one loop to the other in a multi-loop system.

4.11 Effect of Automatic Control of Topside Choke Valve Opening

The application of automatic PI feedback control on the topside choke valve, transformed the system to close-loop system and the system operate in the open-loop unstable region with increased valve opening and reduced riser base pressure represented by the green and yellow curves (figure 4-16). The

controller was designed at riser base pressure of 15.3881 bara. As the pressure set-point was gradually reduced, the valve opening increased from 5% to 12.65% a 7.65% increase in valve opening. A further increase beyond this position caused the system to lose stability and the riser base slugging reappeared.

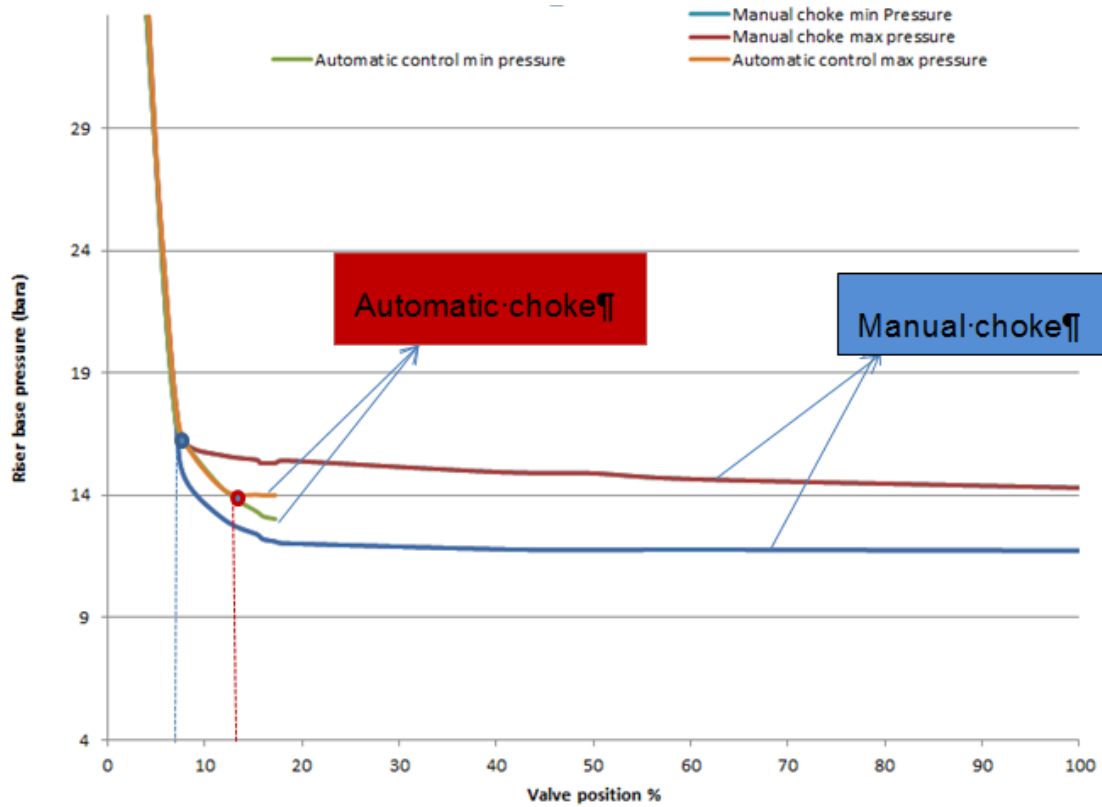


Figure 4-15 Comparing the improvement of the automatic topside choke over the manual topside choke using riser base pressure automatic control $u_c = 12.65\%$, $P_{RB} = 13.4016$ bara, $K_c = 0.001$, $\tau_I = 100$ (s) and pressure set-point 14.6675 bara. The valve opening improved by 7.65% from the manual choke.

Table 4-3 Process and controller parameters

PI	Process parameters		Controller parameters	
Valve opening	K_P	τ (s)	K_C	τ_I (s)
12.65%	75.73	5700	0.001	100

The blue and red solid curves represents the manual choke minimum and maximum pressures while the green and yellow curves represents the automatic controller minimum and maximum pressures as compared with the result of the manual control.

5 CONCLUSION / FUTURE WORK

5.1 Conclusion

A review of hydrodynamic slug control techniques, including their applications, limitations and challenges were discussed in the work. These techniques include manual choke valve technique, slug catcher, gas-injection, combination of gas-injection and choking, active feedback control of the topside choke, flow line modification/layout to avoid dips and splitting of flow into multiple streams. From the result of the investigation obtained from Olga simulation it was found thus:

- The use of manual topside choke valve alone as control strategy results in low valve opening 5%.
- The application of automatic feedback control on the topside choke valve resulted in operating the system in the open-loop unstable region.
- The application of feedback control improved the choke valve opening from 5% to 12.65%, a confirmation of Ogazi's finding that operating control at open-loop condition improves the valve opening more than manual choke.
- From the improvement on the valve opening to larger valve, feedback control is capable of improving the performance of the system at a reduced riser base pressure.
- Feedback control was able to stabilise the system and at limited valve opening of 12.65% achievable.
- There was significant reduction in back-pressure by implementing control at open-loop condition from 15.3881bara to 13.4016bara.
- Active feedback control showed interesting result in suppressing hydrodynamic slug with reduced back-pressure than manual choke
- The interesting results were the capability to operate the system in the open-loop unstable region.
- Lower back-pressure than using manual choke method thus suppressing the riser base slugging.

- The valve opening was increased from 5% to 12.65% with active control representing more than 100% increase in the valve opening, when compared with manual choke.
- This translates to an improvement in production.

FUTURE WORK:

Extensive work is still required in order to gain sufficient knowledge and understanding of hydrodynamic slugs and its control.

- It is recommended that the model be investigated on reservoir source.
- The model is recommended for validation with experimental data.
- Economic analysis to determine if the control strategy can be implemented on the reference case suffering hydrodynamic slugging is recommended.
- Another control variable should be investigated to determine which control variable can yield largest valve opening.

REFERENCES

- Ansari, M.R, (1998). *Dynamical behavior of slug initiation generated by short waves in two-phase air-water stratified flow*. ASME HTD 361. 289-295
- Åström, K. J., & Hägglund, T. (1995). *Automatic tuning of simple regulators with specifications on phase and amplitude margins*. *Automatic a*, 20(5), 645-651.
- Barnea, D. (1977), A unified model for predicting flow-pattern transitions for the whole range of pipeline inclinations. *International Journal of Multiphase Flow*, vol.13, no. 1
- Burke, N. E. and Kashou, S. F. (1996), Slug-sizing/slug-volume prediction: State of the art review and simulation", *SPE Production and Facilities*, vol. 11, no. 3, pp. 166-172.
- Bratland, O. (2010), *Pipe flow 2 Multiphase flow assurance*, drbratland.com,Norway
- British Standard (2005), *Natural gas. Standard reference conditions*, BS EN ISO 13443:2005, British standard, London, UK
- Cao, Y., Omotayo, M., Lao, L. and Yeung, H. (2011), *A study on active gas lift control for hydrodynamic slug mitigation*, 11/YC/56, Cranfield University, Bedfordshire, UK.
- Crowe,C.T.(2009), *Multiphase Flow Handbook*.
<http://www.ftp2share.com/file/26667/Multiphase%20Flow%20Handbook.rar.html>
- Fabre, J. Line, A. Peresson, L. (1992). Two fluid/two flow pattern model for transient gas liquid flow in pipes. *In 4th International conference on Multiphase Flow*, Nice, pp 269-284
- Handbook of Multiphase Flow Metering, (MPFM Handbook) (2005), Revision 2, *Norwegian Society for Oil and Gas Measurement (NFOGM)*, Oslo, Norway
- Hazem, K; Yi Cao and Liyun Lao (2012), Gas Injection for hydrodynamic slug control. *Proceedings of the 2012 IFAC Workshop on Automatic Control in Offshore Oil and Gas Production*, Norwegian University of Science and Technology, Trondheim, Norway,
May 31- June 1, 2012

- Hill, T.(1990), Gas injection at riser base solves slugging flow problems. *Oil and Gas Journal*, (USA),vol.88, no.9
- Issa, R.I and Kempf, M.H.W, 2003. Simulation of slug flow in horizontal and near horizontal pipes with two-fluid model. *International Journal of Multiphase Flow*. Pergamon
- Jansen, F.E. (1990), *Elimination of severe slugging in a pipeline-riser system*. A Master of Science thesis submitted at the Graduate School, University of Tulsa.
- Jansen, F. E. and Shoham, O. (1994), "Methods for eliminating pipeline -risers flow instabilities", *Society of Petroleum Engineers Journal*,
- Jansen, F. E., Shoham, O. and Taitel, Y. (1996), "The elimination of severe slugging—experiments and modelling", *International Journal of Multiphase Flow*, vol. 22, no. 6, pp. 1055-1072.
- Math Works, PID tuning <webinar@em.mathworks.com.
- Mehrdad,P.Fard, John-Morten Godhavn and Svein, Ivar Sagatun. (2006) Modelling of severe slug and slug control with OLGA. *SPE International symposium and exhibition on formation damage control*. Lafayette, Louisiana
- OLGA user manual version 7.1.2
- Ogazi, A I, Ogunkolade, S, Cao, Y, Lao, L and Yeung, H. (2009) Severe slugging control through open loop unstable pid tuning to increase oil production. *14th International Conference on Multiphase Production Technology*, Cannes, France
- Ogazi, A.I, Cao, Y, Lao, L and Yeung, H. (2011),Production potential of severe slugging control system.*18th International Federation of Automatic Control (IFAC) World Congress*. August, 28-September,2 2011 Milano Italy.
- Ogazi, A. I, Cao, Y, Yeung, H and Lao, L. (2010), Slug control in large valve openings to maximize oil production, *Journal of Society of Petroleum Engineers. SPE*

- Ogunnaike, B A. and Ray, W H. (1994). *Process dynamics, modelling and control*, Oxford University Press, New York.
- Oram, P (2013) Control application in industries, delivering breakthrough solution. www.bp.com
- Schmidt, Z., Brill, P. J. and Beggs, D. H. (1980), *Experimental study of severe slugging in a Two-phase-flow pipeline-riser pipe system*, vol. 20, no. 5
- Schmidt, Z., Doty, D. R. and Dutta-Roy, K. (1985), *Severe slugging in offshore pipeline riser-pipe systems*, vol. 25, no. 1.
- Seborg, D E, Edgar, T F, Mellichamp, D A, Doyle III, F J (2004), *Process dynamics and control*, Wiley
- Shoham, O. (2006), *Mechanistic Modelling of gas-liquid two-phase flow in pipes*, Society of Petroleum Engineers, New York
- Skogestad, S and Postlethwaite, I (2005), *Multivariable feedback control*. John Wiley & Sons Ltd
- Storkaas, E (2005), *Anti-slug control in pipeline riser systems*. PHD Thesis, Norwegian University of Science and Technology
- Taitel, Y. (1986), Stability of severe slugging, *International Journal of Multiphase Flow*, vol. 12, no. 2, pp. 203--217.
- Taitel, Y. and Dukler, A. E. (1976), Model for predicting flow regime transitions in horizontal and near horizontal gas-liquid flow, *AIChE Journal*, vol. 22, no. 1, pp. 47-55.
- Taitel, Y., Barnea, D. and Duckler, A. (1980), Modeling flow pattern transitions for steady upward gas-liquid flow in vertical tubes, *AICHE*, , no. 26, pp. 345.
- Taitel, Y., Barnea, D, (1990). Two-phase slug flow. *Advanced Heat Transfer* 20, 83-132
- Thompson, J.M.T. and Stewart, H.B. (1986). *Nonlinear dynamics and chaos*. John Wiley and Sons Ltd Chichester Great Britain, 1st edition
- Yocum, B. T. (1973), Offshore riser slug flow avoidance: Mathematical models for design and optimization", *SPE European Meeting*, 01/01/1973, London
- Varne Valle, (2010) *Field experience and analysis of hydrodynamic slugging*. Dynamic simulation seminar, Oct. 20 2010. www.statoil.com

Verhulst, F.A. (1990), *Nonlinear differential equations and dynamical systems*.
New York: Springer-Verlag

Appendix A Matrices of manual and automatic control

A.1 Comparing manual choke and automatic control (maximum and minimum pressures)

Manual choke			Automatic control	
Valve opening (%)	Minimum pressure (bara)	Maximum pressure (bara)	Minimum pressure (bara)	Maximum pressure (bara)
4	33.611	35.3721	33.611	35.3721
5	27.5164	27.8193	27.5164	27.8193
6	21.358	22.4285	21.358	22.4285
7	17.0904	17.8133	17.0904	17.8133
8	14.6532	16.1231	16.1231	16.1231
12	14.6532	16.1231	16.1231	16.1231
15.35	12.4081	15.4532	14.2156	14.2756
15.69	12.3081	15.3201	13.2387	14.0125
16.02	12.2031	15.321	13.1387	14.0035
17.27	12.1081	15.321	13.0287	14.002
18	12.042	15.427		
40	11.7999	14.9564		
50	11.7682	14.9015		
60	11.7815	14.665		
100	11.7369	14.3081		

A.2 Matrix of manual choke valve opening and pressure response (manual choke minimum, maximum and average pressure (bara))

Valve position (%)	Manual choke minimum pressure (bara)	Manual choke maximum pressure (bara)	Average riser base pressure (bara)
4	33.611	35.3721	34.4915
5	27.5164	27.8193	27.6678
6	21.358	22.4285	21.8932
7	17.0904	17.8133	17.4518
8	14.6532	16.1231	15.3881
9	13.6945	16.0173	14.8559
10	13.093	15.4209	14.2569
12	13.0065	15.6218	14.3141
14	13.0035	15.3976	14.2005
14.06	12.5081	15.3241	13.9161
14.56	12.608	15.324	13.966
15.08	12.5081	15.3214	13.9147
15.35	12.4081	15.4532	13.9306
15.69	12.3081	15.3201	13.8141
16	12.3074	15.3154	13.8114
16.02	12.2031	15.321	13.762
17.27	12.1081	15.321	13.7145
18	12.042	15.427	13.7345
40	11.7999	14.9564	13.7314
50	11.7682	14.9015	13.7362
60	11.7615	14.665	13.7546
70	11.7682	14.574	13.6711
80	11.7631	14.57	13.6664
90	11.7186	14.532	13.6542
100	11.7396	14.5081	13.6423