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Expansion driven Unstable Two Phase Flows in Long Risers and Wells

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Ekspansjonsdrevet ustabil tofasestrøm i lange stigerør og brønner

Expansion driven unstable two phase flows in long risers and wells

Background

There are several types of flow instabilities which can cause operational problems for pipelines and risers transporting multiphase mixtures of oil and gas. One type of instability is expansion generated flow oscillations in long risers or wells. Small amounts of trapped gas can be periodically released into the riser due to the acceleration caused by the expansion in the riser. This is a type of flow instability which has not been well studied before, both experimentally and by means of numerical simulations.

An experimental setup has been constructed in the multiphase flow laboratory at EPT, as part of a student project in 2010. Some numerical simulations have also been made with available flow simulators. The simulation results depend on the simulation methodology; two fluid models can give different results compared with front tracking models. A good set of experimental data would therefore be valuable as basis for the flow model validation and further model development.

The initial trials have demonstrated that the unstable flow can be reproduced in the laboratory setup. This provides a good starting point for a combined experimental and numerical study of this unstable flow phenomenon.

Objectives

The aim of the current study is to generate experimental data on unstable riser flows and to compare the results with numerical flow simulations.

Tasks

The work will involve the following tasks

1. Install and qualify required instrumentation on the experimental setup (impedance probes, pressure sensors, video/photo)
2. Perform experiments with varying geometry, inlet pressure and gas flow rate
3. Compare the experimental results with numerical simulations with OLGA and other available models
4. Report in a MSc thesis

” - ”

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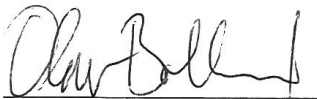
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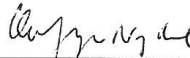
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Abstract

Flow instabilities in long wells and risers under certain flow conditions, is well known. Expansion driven flow instability (EDI) which is relatively little known, refers to occurrence of flow instabilities in long wells and risers as a result of entrapment of gas upstream of the well or riser base. EDI can also occur in gas-lift systems at low pressure and low gas injection rate.

This work was initiated to tackle flow instability problem related to deep water production operations where long wells and risers are extensively in use. The aim of this thesis is to investigate a type of flow instability known as “Expansion Driven Flow Instability (EDI) in Long Wells and Risers”. This involves experimental investigation as well numerical modelling of expansion driven flow instability in long wells and risers. Finally results of the experimental investigations are compared with numerical model data.

An experimental flow loop was setup to verify EDI at varying pipe geometry, inlet flow pressure and gas flow rate to examine the effect of EDI in long wells and risers. The laboratory experiment was conducted using air and water at atmospheric conditions, in a flowline-riser system consisting of a 32mm diameter and 9.12m long riser. The expansion driven flow cycle was captured in video recording.

Variation of inlet flow pressures was achieved by varying the height of the overflow tank. Three cases were considered, each at a different inlet pressure. Each inlet pressure of the fluid was examined against varying inclination angles of the horizontal pipe to the riser inlet. Different gas flow rates were tested at different inclination angles. It was observed that inclination angle has the greatest impact on EDI.

Experimental result of one of the cases was modelled using OLGA and the results of the experimental compared against simulation results output. Discrepancies in the two sets of results were observed in some cases. These may be attributed to simplifications and assumptions made during the simulation model build.

Both results of the experimental investigation and numerical simulation demonstrated that expansion driven flow instability can occur in laboratory setup and can probably occur in deep water natural-lift wells and risers as well as gas-lifted wells and risers under certain flow conditions and pipe inclinations.

Nomenclature

<i>Symbols</i>	<i>Units</i>	<i>Meanings</i>
α_G	[-]	Void fraction
P_{RB}	[bar]	Riser base fraction
L	[m]	Length of flowline
g	[m/s ²]	Acceleration due to gravity
ρ_g, ρ_G	[kg/m ³]	Gas density
U_{SL}	[m/s]	Superficial velocity of liquid
U_{SG}	[m/s]	Superficial velocity of gas
ρ_m	[kg/m ³]	Mixture density
ρ_l, ρ_{water}	[kg/m ³]	density of liquid/water
Δ	[-]	Change
P_{inlet}	[bar/Pa]	Inlet pressure
h	[m]	height of overflow tank/riser
M	[kg/s]	Mass flow rate of air
Q	[m ³ /s, Litre/min]	Volumetric flow rate of air

Abbreviations

<i>Acronyms</i>	<i>Meaning</i>
EOR	Enhanced Oil Recovery
EDI	Expansion Driven flow Instability
OLGA	Oil and GAs simulation tool
PVT	Pressure-Volume-Temperature
BHP	Bottom Hole Pressure
ESP	Electrical Submersible Pumping
PCP	Progressive Cavity Pumping
ESPCP	Electrical Submersible Progressive Cavity Pump
GLR	Gas-Liquid-Ratio
DPR	Gas Discharge Performance Relationship
SIPR	Shifted Inflow Performance Relationship
TPR	Tubing Performance Relationship
FPSO	Floating Production Storage and Offloading
CAPEX	Capital Expenditures
OPEX	Operational Expenditures
DAQ	Data Acquisition
IP	Impedance Probe
NTNU	Norwegian University of Science and Technology

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Chapter 1

Introduction

1.1 Background

With increasing global energy demand and decline of oil and gas in shallow waters, exploration and production of oil and gas has moved from shallow water to deep and even ultra deep water environments. This unrelenting global demand of energy has led to several developments to meet up the increasing demand. These developments include methods of optimizing production of oil and gas in extremely low pressure reservoir, deep water and ultra deep water offshore operations.

Substantial volumes of oil and gas have been delivered by deepwater basins and it still holds some resources/reserves. Several production optimization methods have been researched and adopted by companies to meet increasing energy demands. More so, there are operational challenges associated with current and future exploration and production of oil and gas in deepwater. These challenges include:

- Increasing water depth and distance from shore or existing infrastructure.
- Low pressure reservoirs requiring secondary drives or artificial lift
- Low ultimate recovery of reservoir fluids

As the development of offshore hydrocarbon reserves extends into greater water depths, long risers are required. Production risers (shown in Figure 1) wells, hereafter referred to as “vertical pipes”, used in deep and ultra deep water locations are longer than those used in shallow waters. The deeper the water is, the longer the production riser and the higher the pressure required to lift reservoir fluid to the surface. Therefore, sufficient pressure (bottomhole or reservoir) is required to transport the fluid through a flowline-riser system to the surface production platform. Also in low-pressure reservoir where long wells are used to bring reservoir fluid to the wellhead, sufficient energy is required to lift fluid to the wellhead and then manifold and onward to receiving/processing platform.

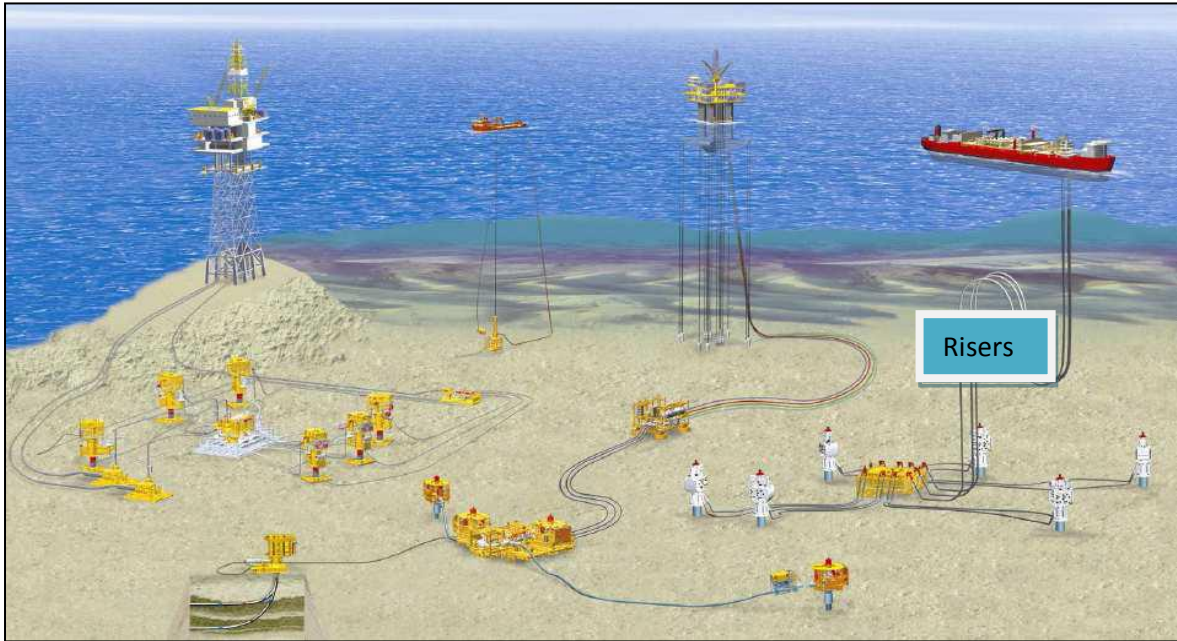


Figure 1: Offshore production system (from internet)

Production from deepwater matured fields or low pressure reservoirs usually require large volumes of in-situ gas or artificial lift gas to aid the reservoir pressure to flow the produced fluid to the surface. Artificial lift or use of large volume in-situ gas has been used immensely for enhanced oil recovery (EOR) operations. The reasons can vary, but the most important with regard to flow assurance (Jayawardene, Zabarar, & Dykhno, 2007) are

- Production enhancement
- Flow stabilization
- Flow line depressurization

Artificial lift is a common technique used to enhance production from matured field or low pressure-reservoir (reservoir pressure decline can be due to depletion or as a result of increased water cut), and can also be used in early operation of a new field to forestall any flow instability that might be initiated in the course of production operations. Gas lift method is among the most widely used type of artificial lift in which high pressure lift gas is injected into wells or risers to enhance production. The injected gas lightens the fluid column, thereby increasing the velocity of the fluid in the well/riser, unlike choking (also used to stabilize fluid flow) which reduces velocity of flowing fluid to stabilize flow and improve production. Therefore, an understanding and prediction of conditions under which gas lift operation might lead to flow instability is important for effective production and/or field development planning. Gas lift techniques can be categorized into continuous and intermittent gas lift method. Continuous gas lift is a preferred method of gas lift used in the industry due to its

numerous advantages over intermittent gas lift. But, continuous gas lift is susceptible to hydrodynamic instability due to lift gas rate variation which leads to pressure and flow rate variation of flowing fluid.

Similarly, the use of natural flowing gas in 'self-lifting' is another option used to provide energy required to lift reservoir fluid to receiving facilities. This also enhances production and depends on the volume of gas required. Self lifting also is prone to flow instability.

Long oil wells and risers with highly oscillatory flow constitute a significant problem in deep offshore production. Consequently, efforts to find a lasting solution have increased both in the academia and industry.

There are several types of flow instability phenomena related to the oil and gas industry. . They are:

1. Casing heading
2. Formation heading
3. Tubing heading
4. Severe slugging
5. Density wave instability

Some of these flow instabilities occur as a result of two-phase flow and depends on the volume of the gas resulting in self-lifting of reservoir fluid to the receiving facilities or as a result of continuous gas-lifting used to enhance or to stabilize flow, as in the case of severe slugging. As mentioned earlier, production from deep water fields requires sufficient pressure of the producing fluid for its transportation from the wells to surface facilities. For deep water, in low-pressure reservoirs or fields at their decline stage, two-phase flow is predominant as the gas boils out of the liquid at low pressure. Depending on the volume of gas, flow rate of the flowing liquid, terrain and inclination of the flowline, an instability known as expansion driven flow instability may be prevalent. This flow instability behaviour can also occur in continuous gas lift system where the gas is injected upstream the riser inlet. Unlike severe slugging in which there is stratification in downward inclined pipe with the liquid blocking the gas at the bend to the riser inlet, expansion driven flow in natural flowing field occurs in both wells and risers as a result of flowline inclination, terrain and varying flow rates of flowing gas by the liquid blockage of the gas at the lower bend which leads to gas

accumulation and expansion thereby pushing the upstream fluid backward. This results in back pressure which may ‘kill’ the reservoir. Therefore there is a need to investigate into this type of flow instability because of its effect on overall production of a field. By predicting flow conditions under which oscillations due to expansion driven flow instability might occur, such scenario may be avoided.

1.2 Objectives

Research to validate expansion driven flow instability can be classified into three main approaches, namely (1) analytical approach; (2) numerical approach; and (3) experimental approach. Expansion driven flow in long vertical pipes has not gained enough attention and has not been widely addressed. In the case of expansion driven flow instability, research has not been carried out to investigate its validity except for the work done by Kjeldby (2010), a thesis work on EDI at the Norwegian University of Science and Technology. The thesis work by Kjeldby (2010) focused on theory and numerical simulations of expansion driven flow instabilities using OLGA and SLUGGIT; developing a Matlab program capable of plotting simulation results from OLGA and SLUGGIT with equal visual frames.

The current thesis work is an extension of the work by Kjeldby (2010) whereby experimental investigations as well as numerical simulation of EDI, using the same flow parameters and geometry, are carried out and results from both methods are compared. This thesis is thus carried out to investigate /verify expansion driven flow instability experimentally and compare with results from numerical simulations.

The effect of flowline inclination, height of the overflow tank (upstream pressure) and gas flow rate on expansion driven flow instability in long risers/wells were investigated. For the experiment, different angles of inclination were considered.

The objectives for this study include:

1. Installation and qualification of required instrumentation on the experimental setup (impedance probes, pressure sensors, video /photos).
2. Performance of experiment with varying inlet pressure, geometry and gas flow rate.
3. Measurements of the riser base pressure and hold variations at varying geometry, inlet pressure and gas flow rate.
4. Comparison of the experimental results with numerical simulations using OLGA.

Chapter 2

Theory

One of the flow assurance issues in deepwater well-flowline-riser is related to production delivery (Guo). In matured fields, as pressure declines, fluid phase separation becomes increasingly important as two-phase flow is formed as a result of gas bubble formation as is in the case of under-saturated reservoirs. For saturated reservoirs at its bubble point becomes under-saturated and two-phase flow becomes the case as the gas boils off and flows above the liquid. This at low pressure of the reservoir becomes a production problem as the energy of the reservoir will be insufficient to lift the fluid to the platform in deepwater system and the facility is also prone to flow instability. Two-phase flow is a discrete entity, and as a result initiates natural fluctuation of its flow parameters. Depending on the void fraction (dependent on gas flow rate), reservoir pressure, water depth and terrain geometry, flow instability in wells and pipeline-riser systems might result and there is need to have an understanding of the behaviour of two-phase flow.

Flow instability is the generalized term for unstable flow under unwanted condition. Flow instability or unstable flow is an undesirable flow phenomenon that is likely to occur in different industrial operations if not checked. Flow instabilities occur in nuclear reactors, chemical reactors, flowlines/pipelines, wells, risers etc. In the petroleum industry, flow instability in the form of high oscillatory flow in pipeline-risers and oil wells has hampered on production, leading to production loss and even safety accidents (Aamo, Eikrem, Siahaan, & Foss, 2005). Slugging is among operational problems leading to flow instability encountered during production and its effect as flow instability phenomenon need to be analysed.

Slugging is a phenomenon of fluid flow instability that creates large forms of variation in the flow rates of oil, gas and water phases in multiphase flow (Poblano, Camocho, & Fairuzo, 2005). Slugging is a generic term that refers to different types of multiphase flow instabilities encountered during oil/gas production. There are different types of slugging (Calvert & Davis, 2010):

- Hydrodynamic slugging
- Severe slugging

- Casing/tubing heading
- Density wave slugging/Tubing heading

Hydrodynamic slugging is a phenomenon that occurs in stratified flow due to wave instability on the gas-liquid interface at certain flow rates. Hydrodynamic slugs are formed in two-phase horizontal pipes and are initiated by pressure drop in liquid-gas flow that leads to increase in flow rate of the gas phase.

Stratified flow occurs more in downhill or horizontal flow with relatively low gas and liquid superficial velocity. With increased gas velocity, waves are formed and these waves get high enough to reach the top of the pipe thereby throttling the gas or even blocking gas flow for a moment, leading to the formation of hydrodynamic slugs or Taylor bubbles (Bratland, 2010) (shown in Figure 2). Slug formation of this sort in horizontal or near horizontal pipes and vertical pipes is called hydrodynamic slugging and the slugs formed are relatively short.

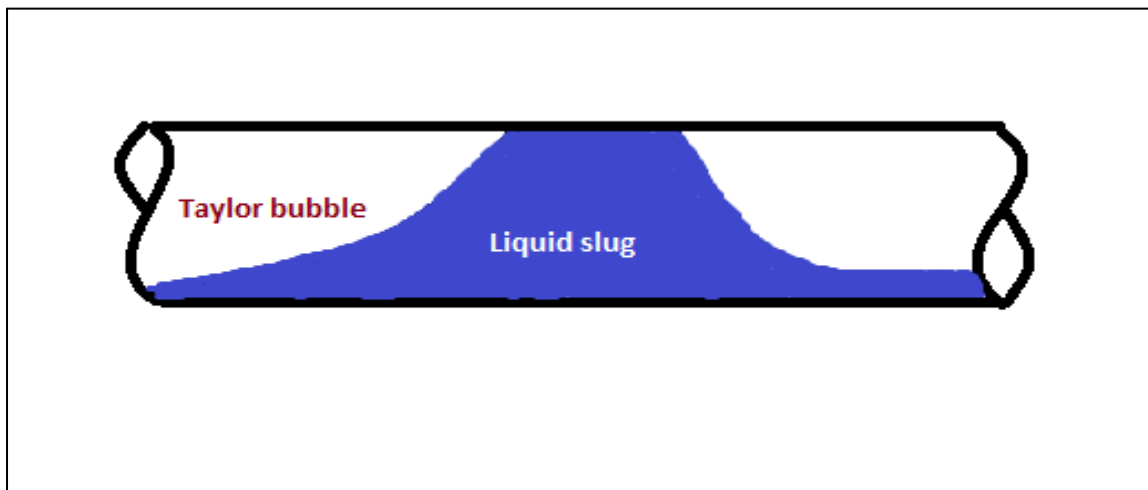


Figure 2: Schematic of hydrodynamic slugging.

Hydrodynamic slugging is characterized by a series of Taylor bubbles separated by liquid slugs. The Taylor bubbles formed are normally symmetrical in vertical pipes and can be asymmetrical or symmetrical in horizontal pipes. It does not constitute a major problem (except that it might be amplified by other factors) (Jansen, Dalsmo, Nokleberg, Havre, Kristiansen, & Lemetayer, 1999) during offshore production but the occurrence of slugs can lead to negative effect (Hanyang & Liejin, 2007).

Long risers/wells used in deepwater production fields require sufficient energy to deliver stable flow to the receiving facilities. At mature stages, reservoir pressure is reduced and this affects total production rate of the field because of low gas and liquid flow rates which are

insufficient to carry the fluid through the pipeline to the riser, resulting in intermittent flow to the receiving facilities. This intermittent flow phenomenon can be referred to as severe slugging.

Severe slugging which is a terrain dominated phenomenon is another type of flow instability that is characterized by the formation and cyclic production of long slug in a vertical pipe (riser/well) joined to a stratified flow-downward inclined pipe (pipeline/flowline) which leads to gas accumulation in the downward pipe due to liquid blockage. Terrain-induced severe slugging is a type of slugging induced by low points present in a flow line at low gas-liquid flow rate. It is dynamic and difficult to understand compared to hydrodynamic slugging. Every flow line through hilly terrain has its elevation profile, and therefore has its own slugging characteristics (Burke & Kashou).

Severe slugging can occur in two-phase flow systems in which a pipeline segment with a downward inclination angle is followed by another segment with an upward-inclination angle (Tengesdal, 2002). It has been shown that severe slugging has adverse effects on field production process. These effects are characterized by a period of no liquid and gas production followed by high liquid and gas flows, which results in high pressure and flow fluctuations making severe slugging an undesirable phenomenon (Guo, Song, & Chacko, 2005).

Conditions for severe slugging are: low gas and liquid flow rates, downward inclined pipe followed by upward inclined vertical pipe (This is a common configuration in offshore petroleum production system) and sufficient upstream compressibility.

Severe slugging can lead to reservoir flow oscillation, high production loss as a result of back pressure at the wellhead and flow instabilities in the receiving facilities (separator) which can lead to shutdown. It can be described as a type of flow instability in which there are periods of no liquid and gas production, then followed by high liquid and gas flow rate as the liquid slug is been produced. Severe slugging is a cyclic phenomenon which consists of four phases (Jansen & Shoham, 1994) and this is shown in Figure 3:

1. Slug formation,
2. Slug production,
3. Bubble penetration and blowout
4. Liquid fall back

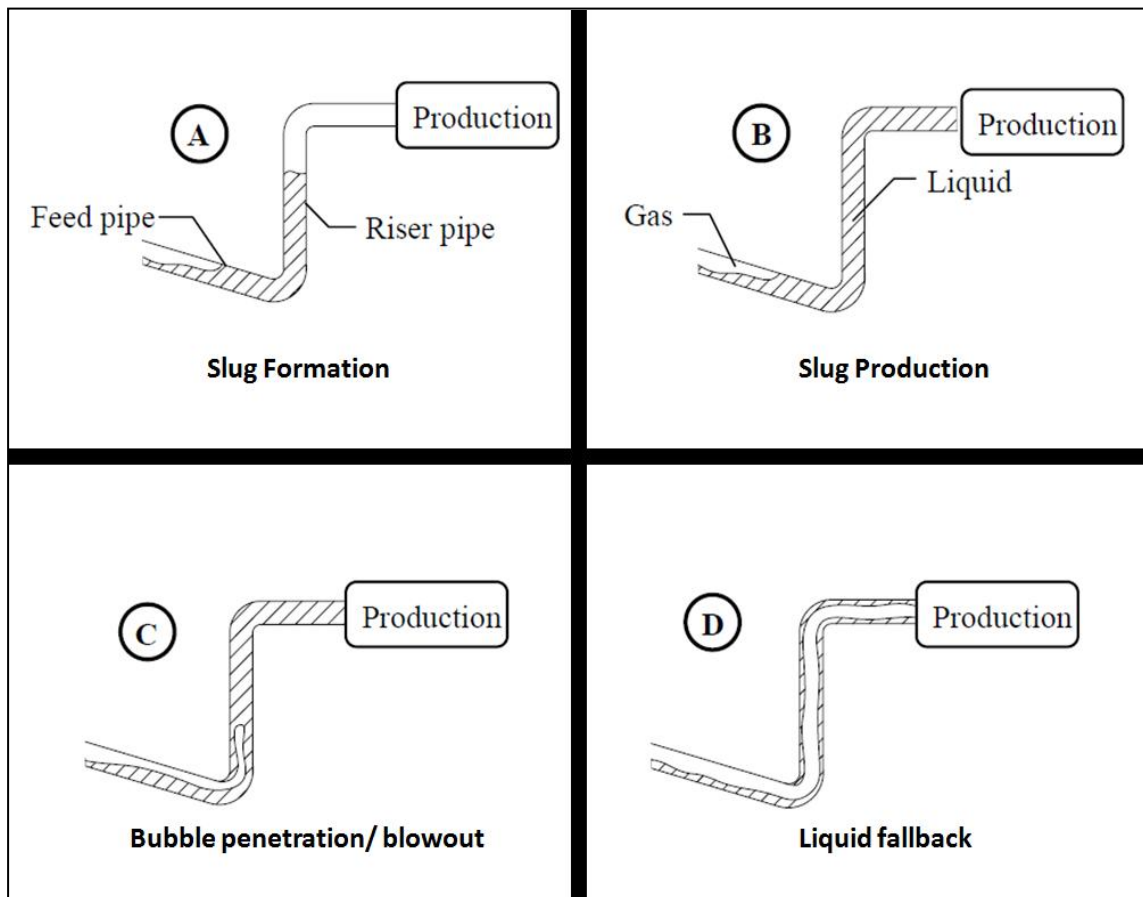


Figure 3: Stages of severe slugging (Bay, 2008)

A classification of severe slugging can be made, according to the observed flow regime, as follows (Wordsworth, Das, Loh, McNulty, Lima, & Barbuto, 1998) (Balino, Burr, & Nemoto, 2010):

- Severe slugging 1 (SS1): the liquid slug length is greater to or equal to one riser length and maximum pipeline pressure is equal to the hydrostatic head of the riser (neglecting friction pressure drop).
- Severe Slugging 2 (SS2): the liquid length is less than one riser length, with intermittent gas penetration at the bottom of the riser.
- Severe slugging 3 (SS3): there is continuous gas penetration at the bottom of the riser: visually, the flow in the riser resembles normal slug flow, but pressure, slug lengths and frequencies reveal cyclic variations of smaller periods and amplitudes compared to SS1.

- Oscillation (OSC): there are cyclic pressure fluctuations without the spontaneous vigorous blowdown.

Severe slugging as a type of instability in flowline-riser system builds up as liquid slug at the riser base consequently trapping upstream gas. Gas trapped leads to accumulation upstream until the flowline pressure is high enough to drive the slug out of the riser. The slug length can be many times the riser length compared to normal slug flow in which the slugs are not very long, on the order of 20 pipe diameter (Pots, Bromilow, & Konijn, 1987). The severity of severe slugging increases with deep water operations when several flowline-riser system connecting underwater manifolds or satellite wells arrive at one platform separator leading to unwanted flaring and inoperability of the compressors and separators (at their design capacity) due to fluctuations in separator level and pressure. It can also lead to platform trips and shutdown. (Sarica & Tengedal, 2000) noted in their work that field development in deep water (>6000 ft) are becoming a reality and severe slugging might occur even at considerably high reservoir pressure. This attributed to the length of the deep water riser, and the expansion capacity of gas due to large hydrostatic pressure which together increases the severity of the severe slugging compared to production systems installed in shallower depth.

To eliminate severe slugging, a total gas mass flow on the order of four times inlet gas flow is needed according to (Pots, Bromilow, & Konijn, 1987), and this can be achieved by gas-lift operations. But expansion driven flow instability can also be initiated at certain flow rate of gas or as a result of gas injection valve opening, coupled with upstream flow conditions (liquid flow rate and pressure).

The stages in severe slugging include:

1. Upstream stratification leads to liquid slug formation at the foot of the riser with no liquid or gas leaving the riser outlet. This result in the hydrostatic pressure build-up in the riser.
2. Gas is trapped and accumulates at the bend as the hydrostatic pressure in the riser exceeds the gas pressure buildup in the flowline.
3. As gas accumulates, the liquid slug will grow as a result of continuous liquid supply.
4. Gas accumulates until the flowline pressure is high enough to drive the slug out of the riser followed by a gas surge triggering the start of another cycle.

The expression for the ratio of pressure build-up rates in flowline and risers as given by (Oliemans & Koninklijke, 1994) is:

$$\Pi_{SS} = \frac{p_{RB} * U_{sg}}{\alpha_G * \rho_G * U_{sl} * g * L}$$

Where;

p_{RB} = riser base pressure

U_{SG} = Superficial gas velocity

α_G = gas holdup

L = length of flowline

g = acceleration due to gravity

ρ_G = gas density

U_{SL} = superficial velocity of liquid

Severe slugging is assumed to occur for $\Pi_{SS} < 1$ and stratified flow in the flowline.

Casing heading can be likened to severe slugging due to its characterized large pressure and flow rate fluctuations that is undesirable. Casing heading is a type of heading which occurs in mature natural flowing wells completed without packers and gas-lift wells as shown in Figure 4. These wells reach a stage that the flowing fluid in the wells are at low gas and liquid superficial velocity thereby promoting large pressure and flow rate fluctuations and gravity dominance (Torre, Schmidt, Blais, Doty, & Brill, 1987) (Hu & Golan, 2003). Casing heading is characterized by periods of high pressure build-up in the casing with reduced or no production and periods of high flow rates of gas and liquid in the tubing due to intermittent gas supply from the annulus (Jahanshahi, Sakhshoor, Kharrat, & Rahnema, 2008). In casing heading space for gas accumulation is provided by the annulus between the casing and the tubing of the well.

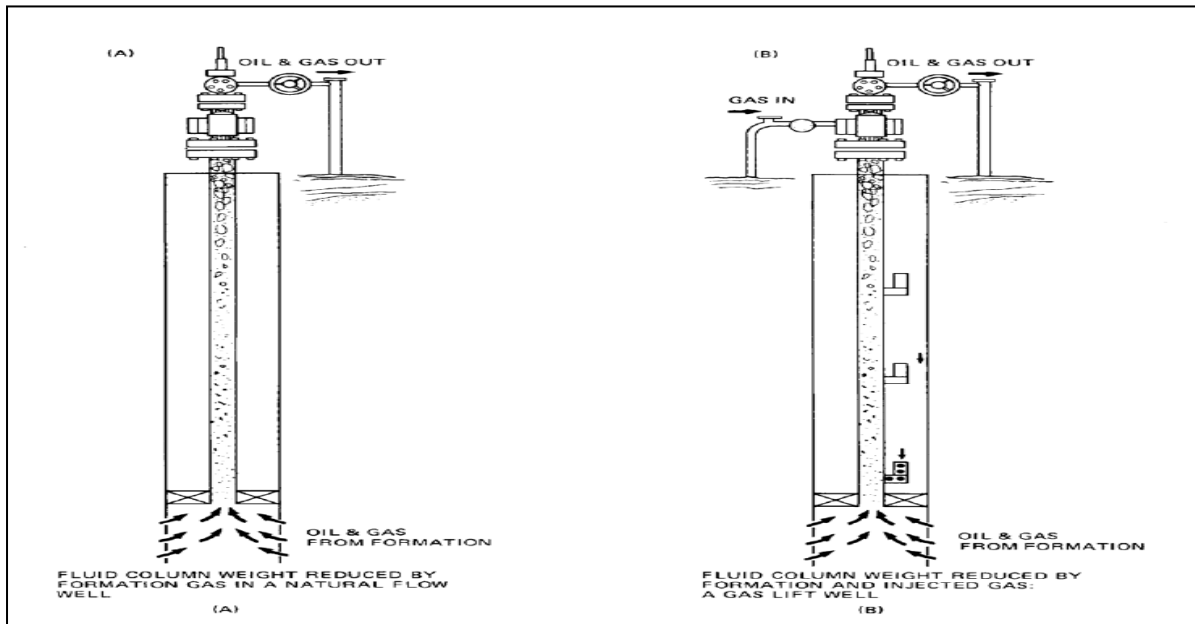


Figure 4: Natural flowing wells and gas-lifted wells (from API gas-lift vocational training series, 1993, page 3)

Casing heading in natural flowing wells can be described as follows (Torre, Schmidt, Blais, Doty, & Brill, 1987):

1. Build up of the tubing pressure at low liquid and gas superficial velocity increases backpressure and suppression of the reservoir outflow, leading to the diversion of produced liquid into the annulus. Rise in annulus liquid level compresses the stored gas in the annulus.
2. Gas volume and pressure increases in the annulus as more gas to the annulus, thereby displacing annulus liquid into the tubing. The tubing pressure increases with annulus liquid displacement. Therefore, high volume of gas in the annulus increases casing pressure.
3. Tubing pressure drops with increasing gas/liquid ratio, allowing annulus compressed gas penetration into the tubing.
4. Casing pressure drops precipitously as gas penetrates the tubing. The gas lifts liquid in the tubing to the wellhead, thereby decreasing the pressure in the tubing which leads to influx of fluid into the tubing. Another cycle begins.

Casing heading cycle for gas-lift wells is similar to the natural flowing well case. Casing heading occurs in gas-lift wells completed with packers in which the gas injection through the gas-lift valve in the annulus between the casing and tubing is subcritical. (Jahanshahi, Sakhshoor, Kharrat, & Rahnema, 2008) described casing head cycle as follows:

1. Pressure of lift-gas in the annulus is subcritical to the tubing pressure no gas enters the tubing. As more gas enters the annulus, annulus pressure builds up until it reaches the tubing pressure. Gas injection into the tubing begins.
2. Gas entering the tubing aerates the fluid thereby reducing fluid hydrostatic pressure and the well starts producing. This increases reservoir fluid influx into the tubing due to pressure decrease in the tubing. As more gas enters the tubing, the casing pressure drops.
3. Pressure drop in the annulus leads to less gas injection. Pressure in the tubing begins to build up leading to reservoir fluid suppression as well as gas injection blockage.

Casing and tubing flow heading causes gas lift injection fluctuation in continuous gas lift systems as a result of variations in casing and tubing pressures and this can degenerate into huge oscillations in flow parameters. Tubing and casing headings are cyclic flow instability phenomenon that occurs during gas-lift enhanced oil production of matured field.

The stages during tubing heading are:

1. Variation in tubing pressure can result to an increase in gas injection rate, decreasing production fluid density.
2. Decrease in production fluid density as a result enhances more gas to flow through the gas injection valve
3. This causes a positive feedback thereby increasing flow from the reservoir until the production fluid density in the tubing increases.
4. Increase in production fluid density results in drop in casing pressure, causing the injection flow through the gas lift valve to decrease. Also as a result of tubing fluid density increase, production pressure increases and subsequently causes a reduction of reservoir fluid entering the wellbore. Then the cycle begins.

Casing heading occurs as a result of gas injection rate fluctuation in the casing due to subcritical flow condition that arise from dynamic interaction between injection gas in the casing and multiphase fluid in the tubing(Eikrem, Aamo, & Foss, 2008).

Density wave instability is a phenomenon that describes pressure drop perturbation in wells and risers due to mixture density variation. In oil production, it is related with unstable gas-lifted wells and risers. Different types of flow instabilities in oil production systems have been recognized, and density wave instability is one of the flow instabilities that have been recently identified to affect production systems. Density wave induces large volumetric change owing to phase change leading to instability in wells and risers. In wells and risers density wave instability occurs due to change in mixture density as fluid flows in vertical pipes leading to a mixture density induced instability. And it can affect safety and the smooth operation of oil and gas production systems and could impact significantly on well productivities.

Density wave instability occurs in wells at low reservoir pressure and low gas injection rate (Hu & Golan, 2003). Observation by (Hu & Golan, 2003) indicates that unstable production due to density wave instability not only can reduce the production but also can increase flow instability in production if the gas rate is low enough.

(Dalsmo, Halvorsen, & Slupphaug, 2002) analyzed how in the past Brage field experienced casing heading and measures taken to resolve the problem which includes changing the lift gas valve, but the field still experienced slugging and it was identified as density wave instability in which the low rates in the well below the gas lift entry point lead to separation of the different phases in the well causing backflow of water and slugs to be formed in the wells. (Ambrosini, Di Marco, & Ferreri, 2000) described density wave as a phenomenon governed by lighter and heavier pocket of fluid propagating along a column. It is a type of instability that occurs in gas lifted wells/risers as a result of variation of liquid inflow from the reservoir to the wellbore or riser even at constant gas injection rate leading to variations in the liquid hold across the well/riser. Variation of the mixture density in the wells/riser invariably results in the change of hydrostatic pressure drops that is transported along the pipe as void wave. This void wave does not initially result in instability in the system because the oscillation is self-sustaining, but at certain level these small perturbation can degenerate into huge oscillations leading to an unstable flow. The instability will propagate backwards in the form of pressure, adding further to the intermittent flow from the reservoir (Calvert & Davis, 2010). The steps during one density wave cycle in a gas-lifted well (or riser) supposed the gas is

injected continuously at a constant flow rate is as follows (Jahanshahi, Sakhshoor, Kharrat, & Rahnema, 2008):

1. When there is insufficient gas injection to reduce the weight of oil column in the
2. Tubing of the well for instance, and consequently the pressure at the bottom of the well is increased and production from reservoir is decreased. Also, because of continuous injection of gas into the tubing, gas mass fraction is strictly increased and reaches its maximum value.
3. The bottomhole pressure is larger than reservoir pressure, so there is zero oil production from the reservoir and the gas mass fraction at the bottomhole is in its maximum value and a region of low density form at the bottomhole. The void region travels up along tubing as a density wave. Continuous injection of gas into the tubing reduces the weight of oil column in tubing and consequently the bottomhole pressure.
4. Decrease in bottomhole pressure results in increase of oil flow rate into the tubing and brings about the fall of the gas mass fraction in the tubing.

Density wave instability is also known to occur in other industrial systems, such as steam generators, boiling water and nuclear reactor cores (Sin'egre, Petit, Saint-Piere, & Lem'etayer). (Ambrosini, Di Marco, & Ferreri, 2000) gave a classical description of density wave instability as related to a boiling channel with an imposed constant pressure drop. The single phase and two phase region pressure drops oscillate in counter-phase as a consequence of waves of heavier or lighter fluid travelling from the inlet to the outlet of the channel.

Variation in the riser fluid density is given by:

$$\Delta\rho_m = \rho_g\alpha + \rho_l(1 - \alpha)$$

Where, ρ_m = mixture density

ρ_g = gas density

ρ_l = water density

α = void fraction = $\frac{\text{Volume or area occupied by gas}}{\text{Total volume or area}}$

Production from low pressure reservoir are prone to flow instabilities identified above and methods are adopted by the oil and gas industries to prevent or eliminate these flow instabilities that may arise at early/latter stage of production or during start-up.

Artificial lift is a technique used to obtain a higher production rate from a well by lowering the producing bottomhole pressure (BHP) on the reservoir or back-pressure on the wellhead. Artificial lift has been traditionally applied to oil wells, to boost production from the well as natural drive mechanisms decline over the well life. It has also been used to generate flow from a well in which production flow has ceased, or to increase the flow from a well in order to produce at a higher rate. Artificial lift has also been used to reduce or eliminate flow instabilities in wells and risers. Artificial lift systems have principally involved the following techniques: downhole pump in which the flowing pressure at the pump suction is lowered; gas lift system in which the density of the fluid in the tubing is lowered; plunger device that lifts liquid slugs in the wellbore using the well's own energy. To realize the maximum potential from any oil or gas field development, the most economical artificial lift method must be identified and selected (Lewis, 2007). Artificial lift usually is installed in the near vertical portion of a horizontal well and rarely in the horizontal portion, to reduce slugging and to achieve maximum drawdown (Lewis, 2007).

Types of artificial lift systems are as follows (Lewis, 2007):

- Sucker rod pumping
- Electrical submersible pumping (ESP)
- Hydraulic piston pumping
- Hydraulic jet pumping
- Plunger lift
- Progressive cavity pumping (PCP)
- Electrical submersible progressive cavity pump (ESPCP)
- Gas lift

Description of the different types of artificial lift as explained in the appendix

Amongst the different types of artificial lift, gas lift is still the preferred method of artificial lift due to its advantage over other types of artificial lift method. Choking and gas lift have been traditional methods used in the oil and gas industries to eliminate flow instabilities (but has also suffered operational disadvantage) (Jansen & Shoham, 1994).

The most important advantages of gas lift over pumping lift method for instance which has better capacity to lower the flowing pressure of reservoir fluid are (Jahanshahi, Salahshoor, & Kharrat, 2008):

- Most pumping systems become inefficient when the GLR exceeds some high value, typically approximately 500scf/bbl (90m³/m³), because of severe gas interference. Although remedial measures are possible for conventional lift systems, gas lift systems can be directly applied to high GLR wells because the high formation GLR reduces the need for additional gas to lower the formation-flowing pressure.
- Production of solids will reduce the life of any device that is placed in the produced fluid flow stream, such as a rod pump or ESP. Gas lift systems generally are not susceptible or erosion caused by sand production and can handle a higher solids production than conventional pumping systems.
- For some applications, a higher pressure gas supply of a gas zone may be used to auto-gas-lift another zone.
- In highly deviated wells, it is difficult to use some pumping systems because of the potential for mechanical damage to deploying electric cables or rod and tubing wear for beam pumps. Gas lift systems can be deviated wells without mechanical problems. However, gas injected in the near-horizontal sections will not reduce gravity pressure effects and will in fact increase frictional losses.

Gas lift is also adaptable to reservoir changes. It is relatively simple to alter gas lift design to fit reservoir decline or an increase in water production that generally occurs at the decline age of a field. Simple modifications that can be made to gas lift installation due to reservoir change in conditions, are made at the surface without altering the downhole system (Jahanshahi, Salahshoor, & Kharrat, 2008), these changes include; replacement of gas lift valve. One major limitation of gas lift is the limit of bottomhole pressure that can be achieved.

Gas lift is inexpensive and easy to implement and requires less maintenance in comparison to other artificial lift methods. Gas lift efficiency can be enhanced by injecting bubble of smaller sizes then the bubble have a lower rise velocity in the liquid (Vazquez-Roman & Palafox-Hernandez, 2005) which gives stable flow. It used to stabilize flow in the wells, flowlines and risers.

Gas lift is a widely used type of artificial lift used to provide energy that is enough to sustain the flow of oil in the well and through the riser up to the surface with satisfactory economic return. It is a type of artificial lift method whereby external high pressure gas is injected into well/riser at some depth to augment formation gas, together lightening reservoir fluid and reducing the flowing bottomhole pressure which invariably increases the inflow of produced fluids. This allows the reservoir pressure to force the fluid to the surface. The addition of lift gas to formation gas increases the superficial gas velocity thereby enhancing a stable flow regime. It is also used to stabilize flows in wells, flowlines and risers as reservoir pressure depletes and unstable multiphase flow is initiated. The importance of gas lift was illustrated by (Ayatollahi, Narimani, & Moshfeghian, 2004) where oil production in Aghajari oil field in south of Iran are mature. Iran Aghajari Field, experienced peak production of 1 million barrels a day in 1974, and then produced a steady 850,000 barrels per day for 17 consecutive years, is now to consume 3 billion cubic feet of South Pars gas to prop up production of 187,000 barrels per day. Lift Gas are commonly injected close to the riser base (for example, Girassol field, Angola where gas is injected at the riser base for activation and flow stabilization, (Zakarian & Larrey, 2007) or vertical well to optimize production. But can also be injected at some distance upstream riser base along flowline-riser system for the purpose of optimization or stabilizing production fluid when the reservoir pressure has depleted or as a result of water depth like deep offshore operations.

Two types of gas lift methods commonly used in the industry are intermittent-flow and continuous-flow gas lift methods. As the name implies, in continuous-flow gas lift method high pressure gas is continuously injected into the produced fluid this mixes with the formation gas to lift reservoir fluid to the surface. The process for continuous-flow gas lift can be described as follows (Jahanshahi, Salahshoor, & Kharrat, 2008):

- Reduction of the fluid density and the column weight so that the pressure differential between the reservoir and the wellbore will be increased.
- Expansion of the injected gas so that it pushes liquid ahead of it, which further reduces the column weight and increases the differential between the reservoir and the wellbore.
- Displacement of liquid slugs by large bubbles of gas acting as pistons.

Whereas intermittent flow gas lift which implies intermittent injection of lift gas to producing fluid is used when continuous-flow gas lift proves to be uneconomical and this depends on factors like; reservoir pressure, tubing size, gas-to-liquid ratio and flow rate of the well. Conversion from continuous-flow gas lift to intermittent is easily achievable and commonly used by industries to achieve their optimum production rate. It is sometimes preferred to continuous gas-lift for gas-lift optimization and is dependent on well characteristics and production rate. (Ayatollahi, Narimani, & Moshfeghian, 2004) explained intermittent gas lift as a cyclic production method in which a liquid slug is allowed to build up in the tubing string of the well. The combination of surface back pressure, weight of gas column, and hydrostatic pressure of the slug reaches a specified value, leading to gas injection through the annulus casing by some type of control for a definite injection time. Under ideal conditions the liquid, in the form of a slug or piston, is propelled upwards by the energy of expanding and flowing gas beneath it (Ayatollahi, Narimani, & Moshfeghian, 2004). Production of oil slug and the gas results in pressure decrease in the tube leading to gas lift valve closure, thereby stopping gas injection. The cycle starts again with liquid build up and subsequent lift of the liquid by injected gas. Intermittent gas lift is characterized by varying pressure at the top and bottom of the liquid slug and it also poses an unsteady nature in which simulation are done to overcome these difficulties. Also intermittent gas lift is disadvantaged by the fact that intermittent liquid fall back due to friction decreases oil production rate. In all, continuous gas-lift is still the preferred option for ‘high-producers’.

Therefore, selection of the type of artificial method to use is very important for the long term profitability of any gas-lift oil well. A poor choice will lead to low production and high operating cost; therefore measures are taken to use the best suitable type of gas-lift method so as to ensure efficient and economical oil production.

(Betancourt, Dahlberg, Hovde, & Jalali, 2003) explained that production from an oil well is also possible with the help of in-situ gas as an alternative to artificial lift. Application of in-situ natural gas lift depends on standoff (with respect to initial gas-oil and water-oil contacts) and targeted production rate. Candidates for natural gas-lift are reservoirs with the oil zones between a gas-cap and a bottom aquifer and depleted oil reservoirs. The lifting requirement is largely dependent on the breakthrough of the gas-cap gas into the well as high GOR will be enough for the well to flow naturally. A downhole valve is required to control the gas rate required for “lifting” and valve position adjustment is required to maintain the desired gas flow rate.

(Xu & Golan, 1989) noted that operators of wells producing with continuous gas-lift face certain cases of difficulties in regulating and maintaining the production rate or the gas injection rate at a desired target level. As the gas injection rate fluctuates from high to low, unstable flow becomes imminent and this can lead to operational instability and eventual shutdown of production facility. To analyze well performance of a gas lift system, a riser equilibrium curve can be used using two available pressure-rate relationships at the gas injection point; the one which relates gas flow with the pressure at the injection point, called Gas Discharge Performance Relationship (DPR) is calculated from gas compressor downwards to the injection point and the other relates the liquid flow rate with the pressure at gas injection point calculated from the reservoir to the injection point and is called Shifted Inflow Performance Relationship (SIPR) (Xu & Golan, 1989) as shown in the Figure 5 below.

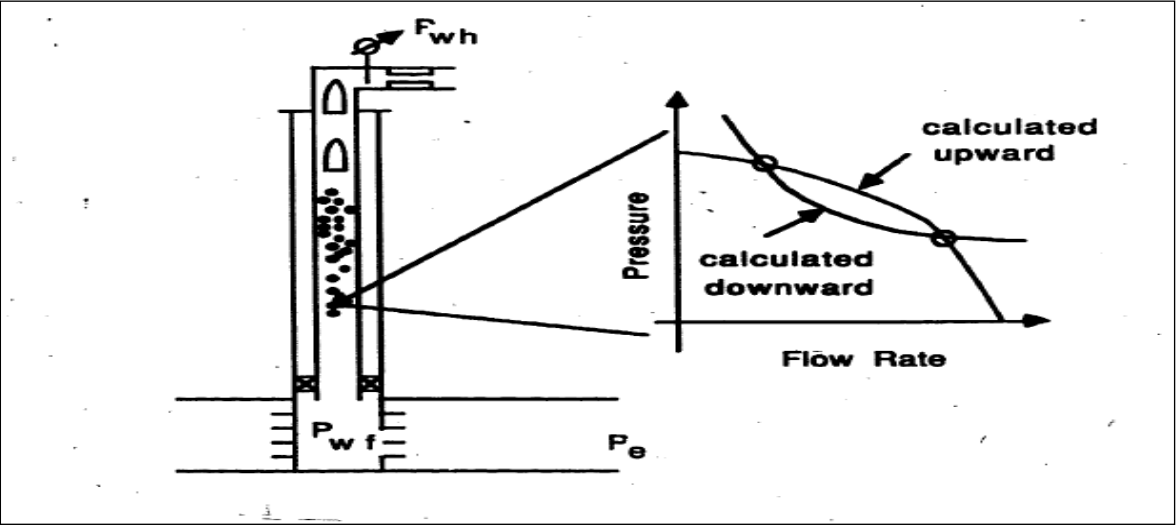


Figure 5: Determination of equilibrium conditions (Xu & Golan, 1989)

The required pressure rate is called Tubing Performance Relationship (TPR). Though, the equilibrium curve cannot guarantee that the flow is stable. Flow can only be stable when if momentarily disturbed, its new operating condition converges asymptotically to the initial ones. Dynamic flow instabilities like casing heading, density wave instability, expansion driven flow instability are detected when the equilibrium curve fail to satisfy dynamic stability criterion as represented by the Figure 6 below.

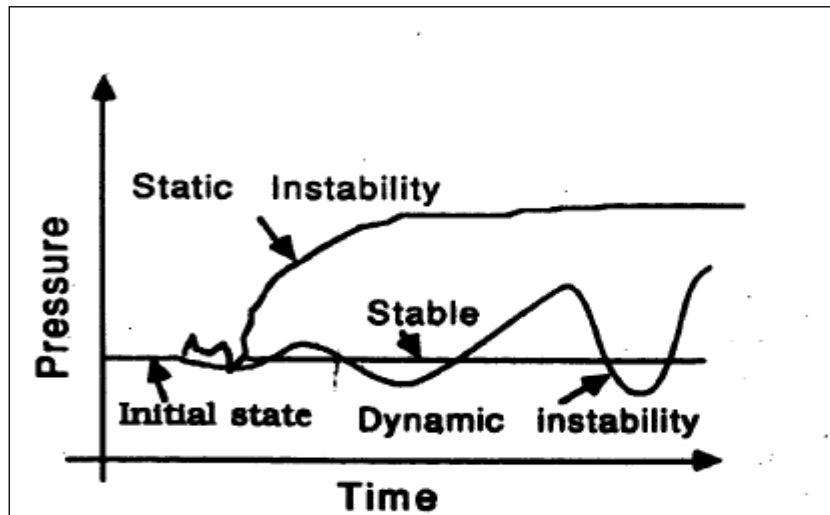


Figure 6: Illustration of stability concepts (Xu & Golan, 1989)

Gas lift and other methods of artificial lift are used by the industry to eliminate severe slugging in deep water operation. But this also comes with its own challenges. Other methods considered for the elimination of severe slugging includes; reduction of line diameter and choking. Tubing and casing heading, density wave oscillation and expansion driven flow instability are considered as instabilities related to continuous gas-lift operations thereby mimicking intermittent gas lift due to instabilities with systematic background. These phenomena can also occur in natural flowing wells. These lead to production loss. Density wave can be self-sustaining to a certain level after which it breaks down and causes flow instability in wells.

As explained earlier, gas-lift and in-situ gas-lift (gas from a natural flowing well) are used for the activation of low pressure oil wells and for long wells and risers where reservoir energy is not sufficient to lift the fluid to the surface. But oil production in gas-lifted oil wells and risers at their decline stages becomes unstable for low gas lift rates giving rise to oscillatory behaviour. (Xu & Golan, 1989) explained that flow instability is viewed by operators as an operation problem that can be solved by trial-and error adjustments of operational conditions. Control of flow instability has been discussed in many publications and in the bid to stabilize flow, operators try to adjust gas injection choke and if not successful are obliged to replace the gas injection valve with a smaller orifice sized valve.

Gas lift systems are considered to be less flexible in coping with changes well productivities and pressure variations especially in the face of water breakthrough where reservoir properties

change. Therefore, Gas lift systems sometimes operate at transient conditions and are subject to flow instabilities. (Takei, Zainal, Ramli, Matzain, Myrland, & Shariff, 2010) described flow instability condition experienced in deepwater flowline and riser of Chinguetti oil field of the Mauritania coastline. The field was developed with subsea wells, manifolds, flexible flowlines and risers tied back to a permanently moored FPSO. Analysis was carried out on the causes of the flow instability and was done by developing a production system model of the wells, flowlines and risers using OLGA. The parameters considered were; gas injection rate, location of injection point and wellhead choke opening. Lift gas was injected at the wellhead (additional gas to aid the natural flowing gas in 'self-lifting' of the reservoir fluid to the FPSO) and from the analysis stable flow was achieved which made gas-lifting a preferred option of eliminating severe slugging for the Chinguetti flowline-riser system.

As was previously stated that it was observed that to eliminate severe slugging, a total gas mass flow on the order of four times inlet gas flow is needed according to (Pots, Bromilow, & Konijn, 1987), and this can be achieved by gas-lift operations. But expansion driven flow instability can also be initiated at certain flow rate of gas or as a result of gas injection valve opening coupled with upstream flow conditions (liquid flow rate and pressure).

This implies that flow instability during production in the oil and gas industries can be caused as a result of the following reasons; geological terrain, reservoir pressure, water depth, etc. And for continuous gas lift wells/risers flow instability can be caused by; insufficient gas injection volume, gas injection rate fluctuation in the casing, incorrect port diameter size. These flow instabilities are detected via well/riser history, surface recorder of pressure fluctuations with time at the wellhead or at the platform and fluctuation of gas injection rate.

(Fairuzov, et al., 2004) expressed that operation problems that may arise as result of gas-lift instability are; compressor shutdowns caused by pressure and liquid flow rate surges, difficulties in the operation of low pressure separators and excessive gas consumption. And that this in turn may lead to high CAPEX (due to increase in compressor size) and OPEX (due to increase of gas compression cost).Operating a gas-lifted well/riser under unstable condition leads to very inefficient operation as the full potential of the gas is not properly used (Jansen, Dalsmo, Nogleberg, Havre, Kristiansen, & Lemetayer, 1999).

Expansion driven instability can be likened to density wave slugging in which there is a variation in the liquid hold-up which impacts the hydrostatic pressure and mixture density across a long riser. Any instability will be propagated backwards in the form of pressure,

adding further to the intermittent flow upstream the pipeline before the riser inlet. Expansion driven flow instabilities lead to unstable production which affect most offshore field at their decline age. Expansion driven flow instabilities in long risers is a dynamic flow instability which adds to other identified operational burden that needs be overcome for safety and stable production at the platform. It is a type of flow instability that can be initiated due to the flow rate of natural flowing gas or as a result of lift gas injection at low rate. This severity increases with the height of wells or risers at greater water depths.

Gas-lifted vertical pipe segments (wells and risers) often experience flow oscillations or instabilities that can be attributed to expansion driven flow instability which is somewhat similar to density wave instability, but further explains the lift-gas bubble behaviour and contribution in lifting the reservoir fluid up the well/riser to the receiving facility. Expansion driven flow instability also causes cyclic variation in pressure and flow rates.

Bubble formation and expansion plays an important role in expansion driven flow instability in gas-lift systems because they are often related to strong pressure oscillations. (Mayor, Pinto, & Campos, 2008) describes bubble expansion as rise of the bubble nose region and how bubble expansion results in the upward displacement of everything ahead of the bubble by an amount proportional to the expansion of the bubble. (Mayor, Pinto, & Campos, 2008) illustrated this by considering a bubble *i*, in white in Figure 7 below, and the liquid flowing ahead of it (zone A in Figure 8). The expansion of the bubbles under bubble *i* induces a raise in the aerated liquid and gas ahead of them, proportional to the sum of the individual expansion undergone by each bubble ($\Delta h_1, \dots, \Delta h_n$), and given by:

$$\Delta Z \text{ ahead } i \text{ expansion} = \frac{S_b}{S_c} \sum_{k=i+1}^n \Delta h_k$$

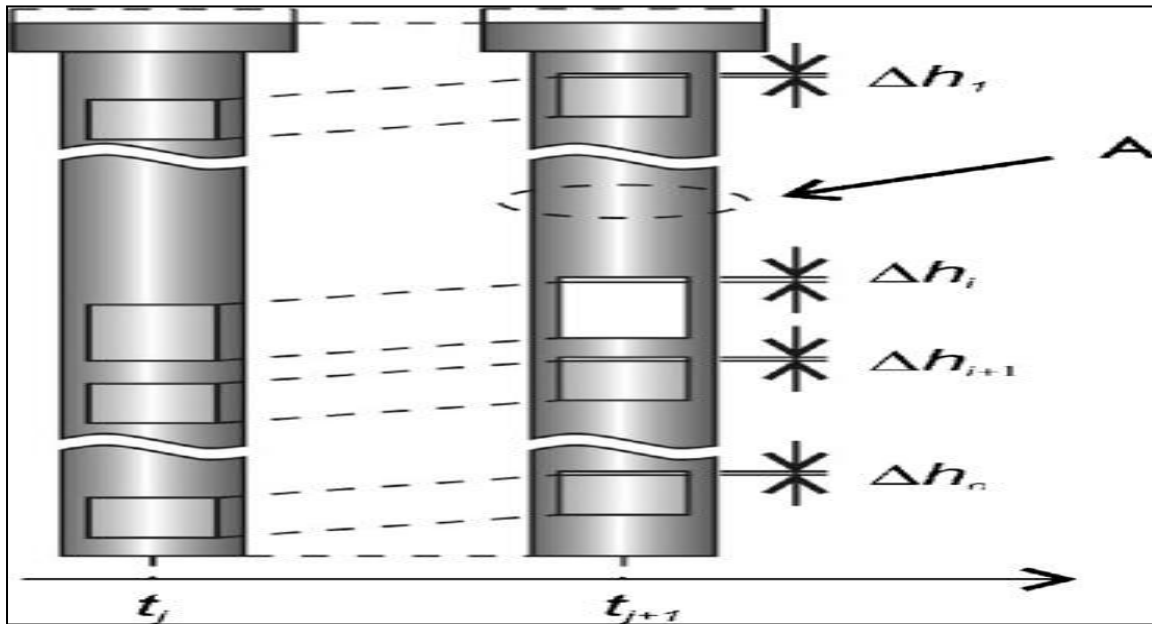


Figure 7: Two consecutive moments in the upward movement of bubbles (Mayor, Pinto, & Campos, 2008)

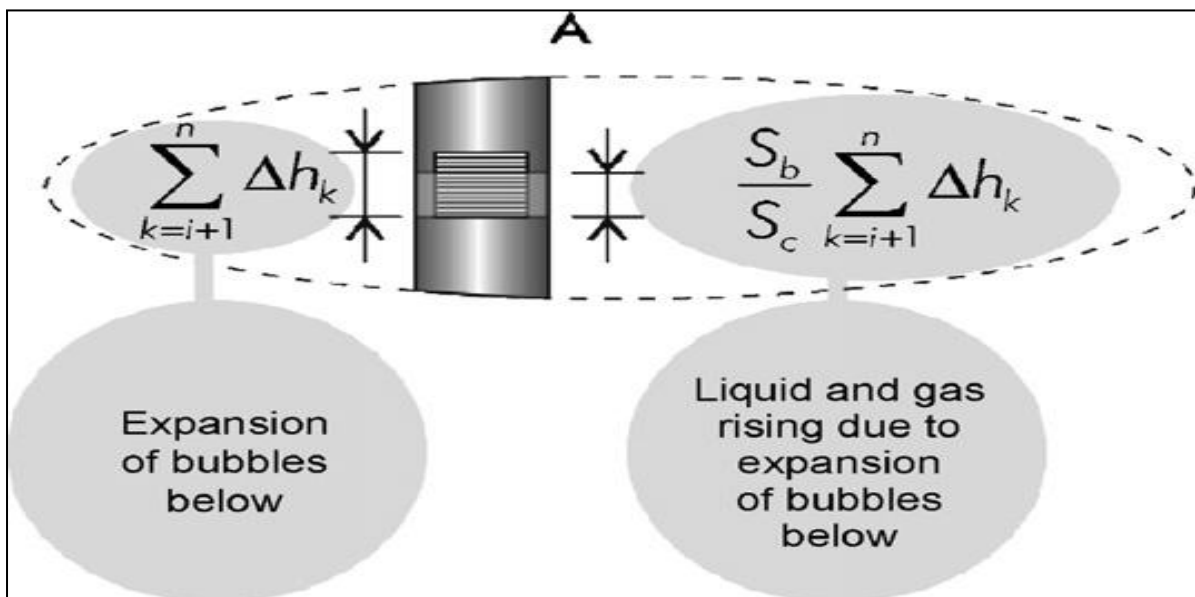


Figure 8: Liquid and gas rise ahead bubble i due to the expansion of bubbles below (Mayor, Pinto, & Campos, 2008)

This decrease in mixture density and hydrostatic pressure due to lift gas bubble behaviour results in pressure drop perturbation/oscillatory behaviour in the riser which spills over to the receiving facility at the platform. Also, the coalescence of smaller bubbles results in the formation of large bubbles thereby increasing the buoyancy and acceleration of the bubble. And as gas bubbles rises in the riser, it expands in response to the decreasing static pressure it

experiences (Issa & Kempf, 2003). This conversely reduces the residence time needed to activate the reservoir fluid and enhance stable flow to the platform.

Expansion driven flow instability can be said to be controlled by void wave propagation. (Lahey Jr., 1991) explained that void wave propagation in turn is strongly influenced by phasic slip and mass force of the gas. Other factors that promote expansion driven flow instability like other gas-lift related instabilities are; injection flow rate, injection port/valve size and tubing/riser diameter. Therefore, there is need to determine the optimal lift-gas injection point as well as the critical lift-gas flow rate to achieve a stable flow in wells and risers. Geometry of flowlines/pipelines also plays an important role in the flow behaviour. (Lahey, Jr., Park, & Drew, 1993) showed that void wave can be used to assess closure assumptions used in two fluid models where he used linear and non-linear analyses. He arrived at the conclusion that void wave is also an excellent means of testing closure assumptions on two-fluid models. Instability of void fraction wave is a hotspot and has played an important role in gas-liquid two-phase flow. And void wave velocity increases with instability/propagation.

Lisseter and Fowler (1992) analyzed models by other researchers by its applications and limitations in determining the speed of void fraction perturbations in bubbly flow citing that though their model for bubbly flow is able to simulate the propagation of voidage and can also be used to model transient phenomena like density wave oscillations. Some of the model developed for bubbly flow when compared with experiment, it was found that a small local disturbance led to large local values of the void fraction, manifested as bubbles clumping together along the pipe (Matuszkiewicz, Flamand, & Boure, 1987). And once the bubbles had clumped together, surface tension was unable (in an air-water system) to prevent the bubbles from coalescing thereby creating large Taylor bubbles which are characteristics of slug flow. This leads to the development of complex characteristic speed, invalidating and breaking down the bubbly flow equations/models since it cannot deal with phenomena such as the break up and coalescence of bubble, and so cannot track the development of the flow.

Song, No, & Chung, (1995) in their work observed that the mechanism of slug formation, critical void fraction and structural developments in bubbly flow that can lead to flow regime transition are dependent on bubble size. Though, they also observed that considering the much work that has been done on void wave and void wave propagation that there exist some level of discrepancies in previous work on critical void fraction and void fraction waves and this

difference might be attributed to two reasons: One is the inaccuracies of the measuring device, and the other is the unidentified effect of the parameter which influences the flow structures and the regime transition mechanism but not taken into account in experiments nor for physical modelling.

Pipe diameter is another important factor that determines the magnitude of flow oscillation due to flow instabilities in production systems. (Poblano, Camocho, & Fairuzo, 2005) in his work validated that the region of stable flow for a well with a tubing diameter of 31/2 inches is much larger than for the well with a tubing diameter of 7 inches. Emphasizing that care should be taken in designing gas lift wells with large-diameter tubing, particularly when significant changes in the injected-gas flow rate are expected, and this can also be applied to large-diameter risers.

Much research investigation has not been carried out on expansion driven flow instability but density wave oscillation has gained substantial attention in the research world. This type of flow instability occurs in cooling systems for a nuclear reactor, oil wells and boilers. Models have been built especially for cooling systems and boilers to determine their susceptibility to density wave oscillations. This is done by determining the speed of void fraction perturbations in vertical bubbly flow/slug flow. Laboratory investigation of expansion driven flow instability in long wells or risers are carried out in the course of this thesis to ascertain parameters and conditions at which it can occur during production. The result is validated against numerical simulation using geometry and flow parameters considered during experiment.

Chapter 3

Experimental setup and Instrumentation

3.1 Experimental setup

All laboratory experiments were carried out using the EDI flow rig within the Multiphase laboratory under the Department of Energy and Process Engineering at the Norwegian University of Science and Technology, Trondheim, Norway. The EDI flow rig is a flow loop set up to demonstrate conditions for expansion driven flow instability in long risers.

Figure 9 below shows a schematic diagram of the flow loop set up to investigate the possibility of expansion driven flow instability in long risers under certain flow conditions using air-water system, with water as the production fluid and air as the lift-gas.

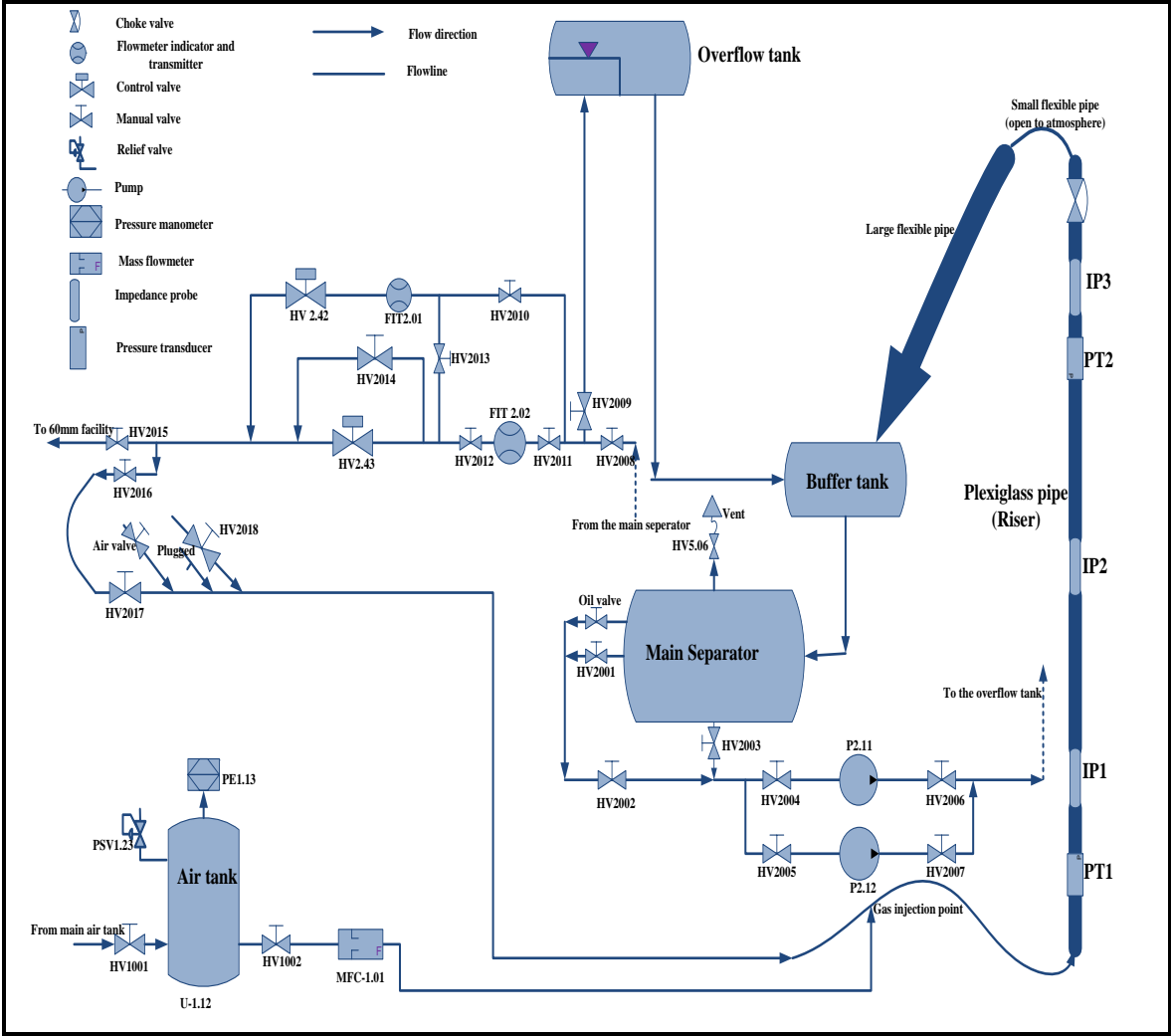


Figure 9: Layout of experimental flow loop

The flow loop consists of a flexible pipe representing a horizontal pipe connected to a vertical pipe considered as the riser. The laboratory installation represents a gas-lift system (or a natural flowing system with gas trapped upstream) with the gas injected upstream along the horizontal pipe segment, at a point approximately 0.8m from vertical pipe base. Compressed air is used as lift gas and water as produced fluid. The vertical pipe is a transparent acrylic flexiglass pipe, facilitating visual inspection of the flow development occurrence as gas is injected into the system. The flexible horizontal pipe having inner diameter of 32mm is inclined at 1.29m distance to the vertical pipe base which is adjustable at different inclination to achieve desired flow behaviour. The vertical pipe (B-3) consists of two separate plexiglass/acrylic plastic pipes connected via flanges at each end of the pipes, with the vertical pipe clipped at several points along its length to maintain a vertical position and isolate the vertical pipe from external vibration. Upstream of the vertical pipe inlet, is the horizontal pipe consisting of upward (along which is the point of gas injection) and downward inclined (inclined towards the vertical pipe inlet) pipe which are varied to different angle of inclinations from the horizontal.

The vertical pipe measures 9.12m in height and has an inner diameter of 32mm. The pressure of the flowing fluid in the horizontal pipe is given by the static height of the overflow tank connected to a flexible pipe. Fluid from the vertical pipe outlet is transported back to the main separator via the buffer tank. A thin flexible pipe connected to larger pipe (to maintain atmospheric pressure at the vertical pipe outlet) was used to channel fluid from the vertical pipe outlet to the buffer tank. Water is circulated via a pump in a closed loop. Oscillatory behaviour in the vertical pipe is monitored using two absolute pressure transducers connected to the vertical column. The bottom pressure transducer is at 0.87m above the vertical pipe inlet while the top pressure transducer is approximately 0.38m below the vertical pipe outlet. Only the signals for the bottom pressure transducer was processed and reported in this study.

In the experiment run, gas was injected upstream the vertical pipe inlet at varying desired rate (at different angle of inclination and liquid inlet pressure by varying the height of the overflow tank) to achieve expansion driven flow in the system. Input and output signals from the pipeline-riser system are handled by a computer system with the aid of a DAG system

Water from 3m³ main separator shown in B-3 is pumped to the riser inlet followed by continuous injection of air at low mass flow rate upstream the vertical pipe base from the air

tank. Air injection is done at an angle of inclination along the S-shaped pipe upstream to the riser inlet to achieve a maximum local accumulation. Water which is the continuous phase is pumped from the main separator into the vertical pipe as well as the overflow tank with the overflow routed back through the buffer tank shown in B-1 back to the main water separator. The overflow tank which has inner baffle is hung on an adjustable steel chain from a crane hooked to the wall. This enables the adjustment of the overflow tank height to achieve the desired hydrostatic inlet pressure of water. The overflow tank which can be likened to a weir minimizes fluctuation of hydrostatic pressure to the vertical pipe inlet. Water was pumped at low frequency of 26.8Hz through pipe into the vertical pipe and into the overflow tank (to maintain the water hydrostatic pressure). There the horizontal-vertical pipe pressure is given by the static height of water in the overflow tank. To get rid of air in the vertical pipe, water was pumped at high frequency of 35Hz to flash out air in the vertical pipe after which it was gradually reduced to 26.8Hz. Stable flow at very low flow rate was observed at the riser outlet. Pressurized air, the lift-gas, was routed from an air tank (shown in B-2) through a small diameter pipe and injected continuously through an injection orifice 0.8m upstream the riser base as shown in B-3. Then, water-air mixture from the riser outlet is routed back to the main separator through the buffer tank where they are separated by gravity.

The experiment was conducted at varying air volumetric flow rate, ranging from 0.25litres/minute – 4.0litres/minute. The experiment was recorded at varying mass flow rates with a digital video camera using a shutter speed of 1/1000s, conducted under atmospheric condition with maximum operating pressure of 2 bara and at room temperature, assuming isothermal condition. Physical properties of experimental fluids at average room temperature of 15°C are given in Table 1 below.

Table 1: Fluid properties

Fluid	Density [kg/m ³]	Viscosity [mPa.s]	Surface tension of water against air [mN/m]
Water	999	1.002	72.704
Air	1.2250	18.000	

3.2 Measuring instruments

The flow loop consists of the several manual valves, relief valves, pressure manometer and the following instrument:

1. Pumps (automatically controlled)
2. Flowmeter indicator and transmitter
3. Control valves (automatically controlled)
4. Pressure manometer
5. Mass controller(automatically controlled)
6. Pressure transducer
7. Impedance probe

3.2.1 Pumps

The flow loop consists of two sets (large and small) of centrifugal pumps (B-2) which supply water from the main water tank to the overflow tank and to through the waterline to the riser inlet. The large water pump was used for this experiment at pump frequency of 26.8Hz.

3.2.2 Flow meter indicator and transmitter/ control valve

This is shown B-2 assist in metering and transmitting while the control valve controls the flow rate of water. There are two sets of flow meter indicator/transmitter and control valve, each for the large diameter and small diameter pipe waterline respectively. The waterline used depends on the flow rate range needed for the experiment. In this experiment the large waterline was used to reduce friction effect due to large friction effect observed with the small waterline. The flowmeter indicator and transmitter was not used in this experiment due to the flow rate range been higher than the flow range used in this experiment giving rise to erroneous results.

3.2.3 Absolute pressure transducers

With range of 0-2.5bar shown in the B-3 were mounted at the base of the riser and upstream the choke valve (B-3) close to the riser outlet with the transducer inlets set flush the inside surface of the riser. The pressure transducer measures pressure fluctuations at the different positions.

3.2.4 Impedance probes

The impedance probe is mounted at 3 different locations along the riser as shown in Figure 10. It has an internal diameter of 32mm (same as the riser diameter) and set flushed the inside

surface of the riser. Variation of liquid holdup measurement was conducted using impedance probe which consist of two chromel (ring electrode) wires traversing the diameter of the vertical pipe (riser). The impedance probe was not calibrated for this experiment and measurements are referred to as impedance.

Input and output signals (voltage signals from pressure transducers and impedance probes) from the installation are logged by a National Instruments data acquisition (DAQ) system to which a computer is interfaced for running the experiment and presenting results.

Water is supplied from main water to a centrifugal pump which pumps the water to the overhead tank and through the water flowline to the flowline-riser. Air is provided from the central compressed air system to the buffer tank and then injected 0.8m upstream the riser base along the flowline.

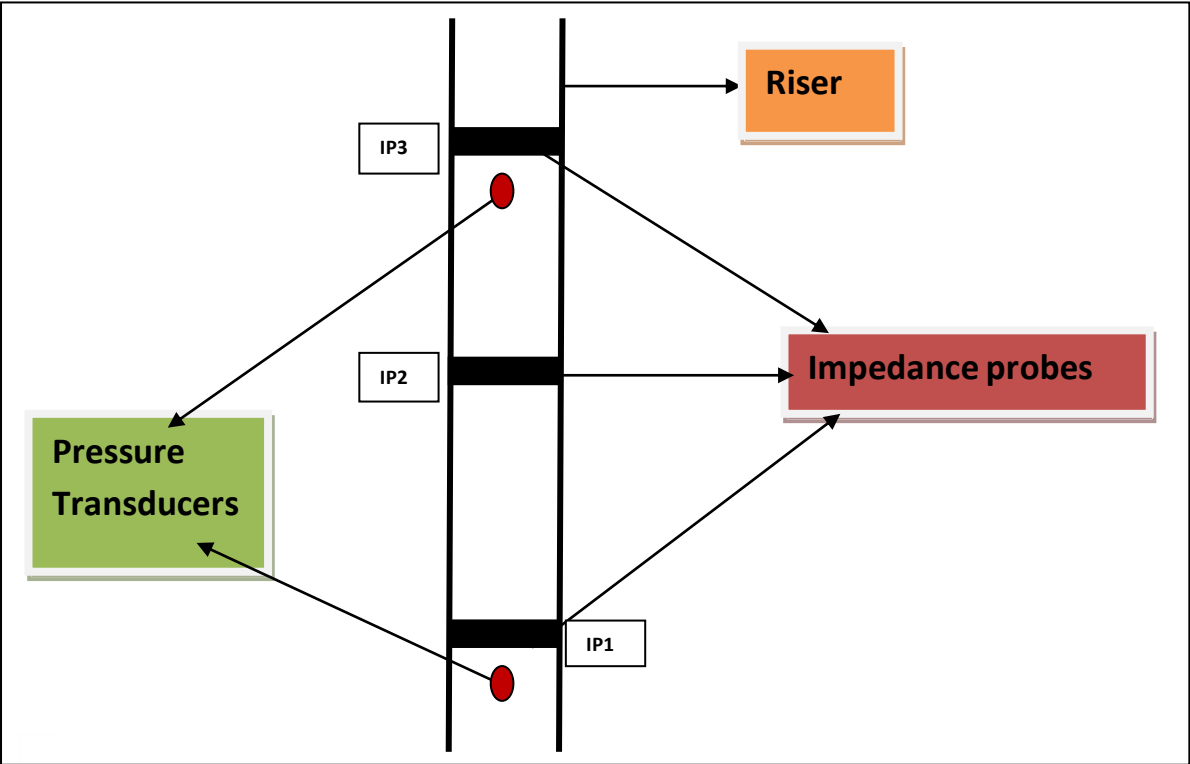


Figure 10: Impedance probe along the riser.

Chapter 4

4.1 Experimental Results

The riser was initially filled with air which was flashed out with water at high pump speed and then gradually reduced to a low speed of 26.8Hz. The continuous phase (water) pumped at of speed of 26.8 Hz was visually observed to flow at low rate out of the riser before lift gas injection. The riser column was then filled with water before gas injection and inlet pressure of the liquid is assumed to be the hydrostatic height of the overflow tank which was varied at different flow conditions. Pressure at the riser base is also assumed equivalent to the height of water column above the riser base. Compressed air was injected 0.8m upstream the riser inlet which resulted into a two-phase flow to the riser base. But due to the inlet flow pressure of the liquid and the mass flow rate of the injected gas, liquid was observed to flow down the downward inclined pipe blocking gas flow into the riser. Unless severe slugging where stratification is observed at the downward inclined pipe, slight stratification was observed before upstream liquid is completely blocked by the flowing gas. This leads to accumulation and expansion of the gas towards the riser inlet until the gas penetrates the riser thereby aerating the riser fluid. The penetration of the gas into the riser reduces the bottom pressure at the riser base enabling inflow of water as well as gas into the riser leading to unstable flow behaviour in the riser. This unstable flow behaviour is identified as expansion driven flow instability.

Unstable flow due to expansion driven flow instability was observed to be a factor of the inlet pressure of flowing fluid which depends on the height of the overflow tank, inclination angle and gas injection mass flow rate. These factors lead to pressure oscillation and holdup variation in the riser.

The height of the overflow tank was varied at different height to establish varying inlet pressure of the flowing fluid. The heights considered are measured with reference to the riser outlet. The height of the riser is 9.12m and the heights of the overflow tank reported in this work are:

- 213mm below the riser outlet (that is, height of 8.907m)
- 30mm above the riser outlet (that is, height of 9.15m)

- 210mm above the riser outlet (that is, height of 9.33m)

The experiment was carried out at atmospheric condition and the inlet pressure of the flowing fluid is dependent on the height of the overflow tank. Therefore the inlet pressures at the various height of the overflow tank considered in this experiment are:

At height of 8.907m, $P_{inlet} = \rho_{water}gh = 8.907m \cdot 9.81m/s^2 \cdot 998.2kg/m^3 = 87,220.39Pa = 0.872bar + P_{atm}(1.01bar) = 1.882bar$

At height of 9.15, $P_{inlet} = 9.15m \cdot 9.81m/s^2 \cdot 998.2kg/m^3 = 89599.9293Pa = 0.896bar + P_{atm}(1.01bar) = 1.906bar$

At height of 9.33, $P_{inlet} = 9.33m \cdot 9.81m/s^2 \cdot 998.2kg/m^3 = 91362.55Pa = 0.914bar + P_{atm}(1.01bar) = 1.924bar$

Where P_{inlet} = Inlet pressure of flowing fluid (water)

ρ_{water} = density of water [kg/m^3]

g = acceleration due to gravity [m/s^2]

h = height of the overflow tank [m]

At each height of the flow tank, the angle of inclination of the pipe upstream the riser inlet is varied at different angles and pressure fluctuations as well as holdup variation at these angles of inclination are recorded. The angles considered for the different heights of the overflow tank are:

At the height of 8.907m, the angles of inclination of the downward inclined pipe considered are:

- -25degrees
- -33degrees
- -45 degrees

At the height of 9.15m, the angles of the inclination of the downward pipe considered are:

- -21 degrees
- -33 degrees

- -45 degrees

At the height of 9.33m, the angles of inclination of the downward pipe considered are:

- -27 degrees
- -36 degrees
- -45 degrees

Also, at these varying heights of the overflow tank and inclination angles, different gas injection rates were implemented and their effect on pressure oscillation and holdup variation was noted. At average room temperature of 15°C, the mass flow rate of air at different volumetric flow rates as shown in the Table 1 below was obtained using:

$$M = \rho Q$$

Where, M = Mass flow rate of air

$$\rho = \text{density of air at } 20^{\circ}\text{C} = 1.204\text{kg/m}^3$$

$$Q = \text{Volumetric flow rate of air}$$

Table 2: Different volumetric and mass flow rate of gas at 15 degrees considered in this report

Q [Litre/mins]	M [kg/s]
0.25	5.1042E-06
1.0	2.0417E-05
1.2	2.45E-05
1.5	3.0625E-05
2.0	4.0833E-05
2.3	4.6958E-05
2.5	5.1042E-05
2.8	5.7167E-05
3.0	6.125E-05
3.3	6.7375E-05
3.5	7.1458E-05
4.0	8.1667E-04

Pressure oscillations graphs under the above flow configuration are shown below. In all the cases gas was injected upstream the riser base giving rise a two-phase (air-water) from the flowline to the riser base. Therefore, the pressure oscillation at the different heights and angles of inclination of the downward inclined are:

4.1.1 Case 1: At a height of 8.907m, that is 213 below the riser outlet

A. Inclination angle of -25 degrees

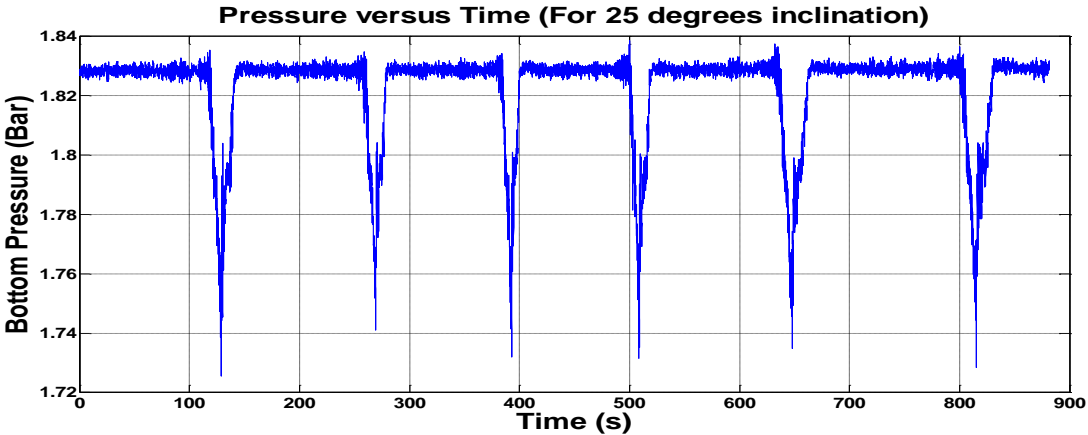


Figure 11: Gas injection rate = 5.1042E-06kg/s

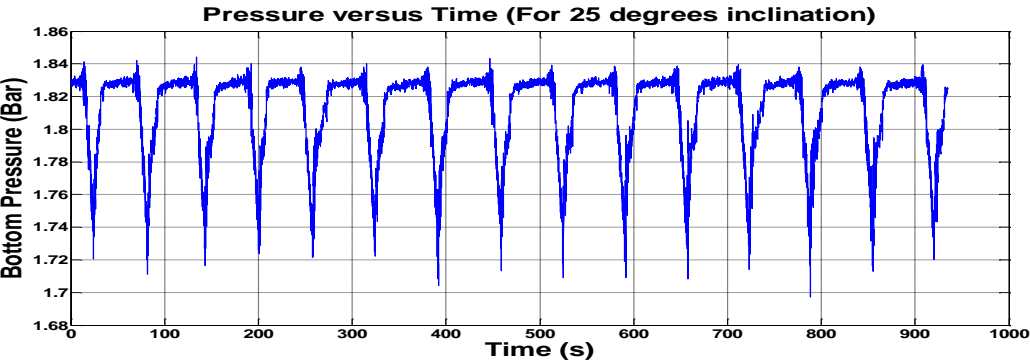


Figure 12: Gas injection rate = 3.0625E-05 kg/s

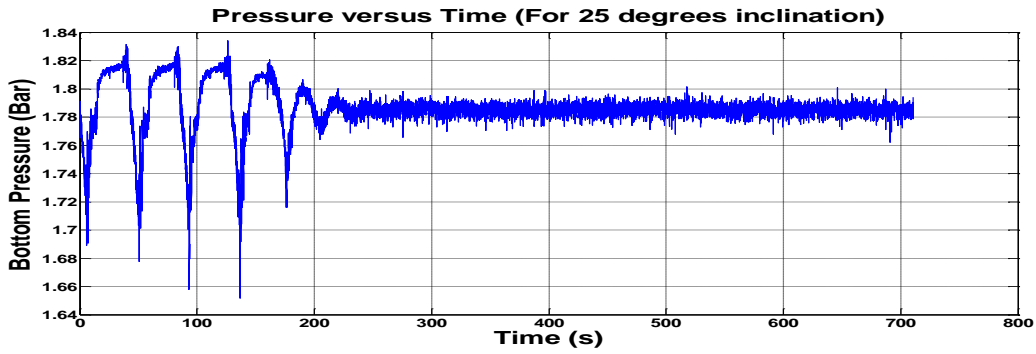


Figure 13: Gas injection rate = $5.1042E-05$ kg/s

Case 1A - The height of the overflow tank was lowered to 213mm below the riser outlet which also determine the inlet pressure of the flowing fluid given the hydrostatic head of the overflow tank. The inlet pressure at this condition is 1.882bar. With inclination angle of 25 degrees, pressure oscillation was observed at low gas injection rate and stability was attained at increased gas flow rate. At gas injection rate of $5.1042E-06$ kg/s (Figure 11) the period for a cycle was observed to be approximately 120s and gas penetrates in the riser the time taken for acceleration and expansion in the riser was approximately 20s. Also maximum pressure at the bottom for the liquid column before gas entrance was 1.83 bar which dropped to 1.73bar as gas penetrates the riser due to decrease in mixture density.

As gas injection rate was increased to $3.062E-05$ kg/s (Figure 12) pressure oscillation was still observed but the acceleration time of the gas in the riser was increased to 25s. The period of oscillation was decreased to 65s as more gas was injected as well as the mixture density in the riser which further decreased indicated by the further decrease in pressure to approximately 1.71bar as gas penetrates the riser. Stability was attained as gas flow rate was increased to $5.1042E-05$ kg/s as shown in Figure 13.

B. Inclination angle of -33 degrees

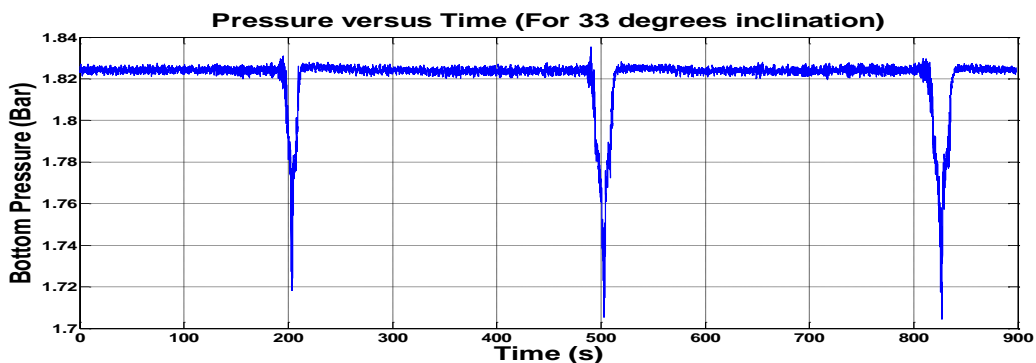


Figure 14: Gas injection rate = $5.1042E-06$ kg/s

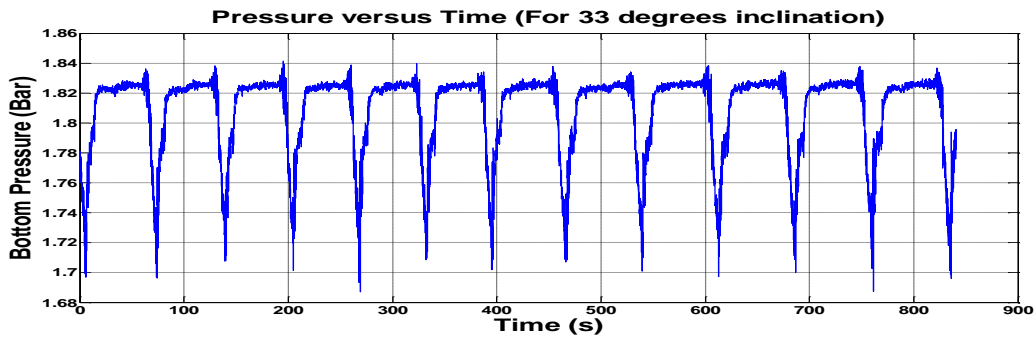


Figure 15: Gas injection rate = 3.0625×10^{-5} kg/s

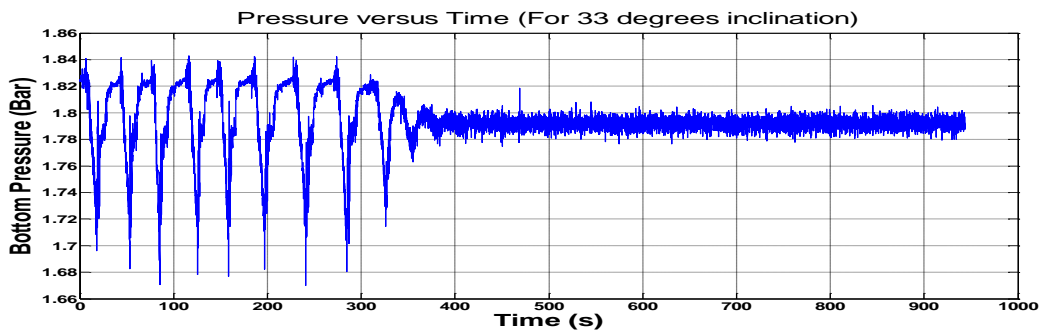


Figure 16: Gas injection rate = 5.7167×10^{-5} kg/s

Case 1B – Maintaining inlet pressure of 1.882bar and increasing the inclination angle to 33 degrees, the period for one oscillatory circle for expansion driven flow instability system increased at gas injection flow rate of 5.1042×10^{-6} kg/s (Figure 14) compared to when the pipe was inclined at 25 degrees. The period for one cycle was approximately with 200-280s. The amplitude of the oscillation is also observed to be the same at the given flow rate of gas injected.

At gas injection rate of 3.0625×10^{-5} kg/s (Figure 15), the period of a given cycle was decreased to 63s. Increase in gas injection as caused further decrease in pressure as gas penetrates the riser due to decrease in mixture density. Stability was attained at increased gas injection and the flow rate of gas required to attain stability for an inclination angle of 33 degrees is more than that requires for the 25degrees inclination as stability is attained at 5.7167×10^{-5} kg/s of gas as shown in Figure 16.

C. Inclination angle of -45 degrees

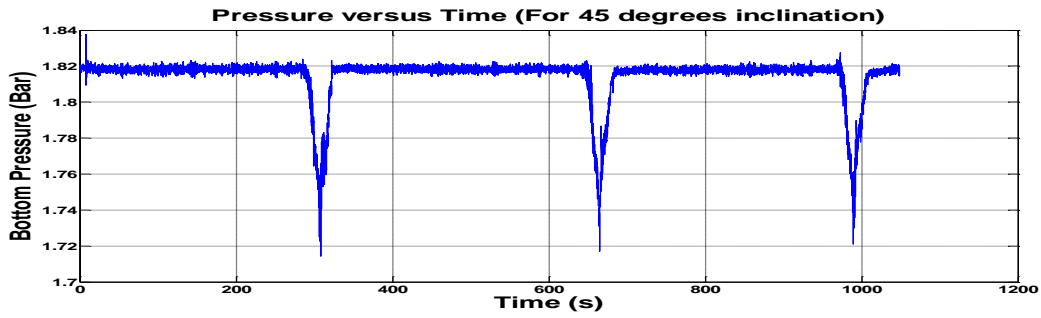


Figure 17: Gas injection rate = $5.1042E-06$ kg/s

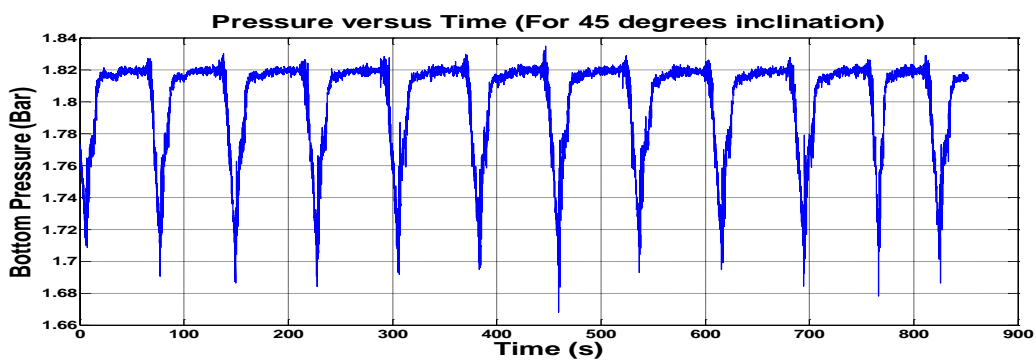


Figure 18: Gas injection rate = $3.062E-05$ kg/s

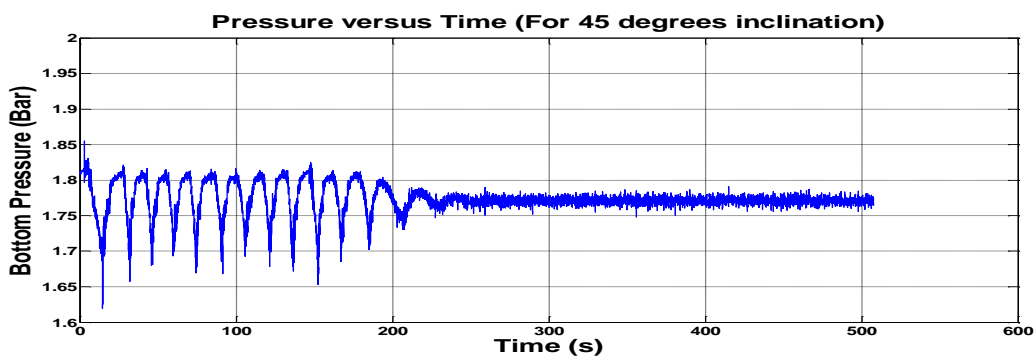


Figure 19: Gas injection rate = $8.1667E-05$ kg/s

Case 1C - Also maintaining the inlet pressure at 1.882bar, increase in inclination angle to 45degrees was observed to result in increase of the period of the cycle at low flow rate of $5.1042E-06$ kg/s (Figure 17) to approximately 320-360s compared to periods at inclination

angle of 25 and 23 degrees at the same flow rate of gas. The acceleration time at this flow rate of gas and angle of inclination was observed to be approximately 30s while the amplitude remains the same as with other inclination angle considered. The period for one oscillatory cycle was decreased approximately 77s as gas flow rate was increased to 3.062E-05kg/s (Figure 18). But this high compared to the period at inclination angle of -33degrees at the same flow rate of gas. The amplitude of oscillation was same as the cases above but stability was attained at higher gas injection rate of 8.1667E-05kg/s (Figure 19). This implies that with increased inclination angle large volume of gas is required to maintain stable flow in the flowing system.

4.1.2 Case 2: At a height of 9.15m, that is 30mm above riser outlet

A. Inclination angle of 21 degrees

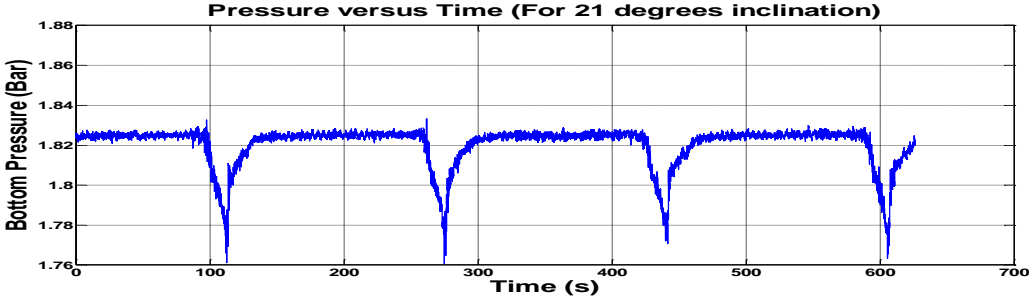


Figure 20: Gas injection rate = 5.1042E-06 kg/s

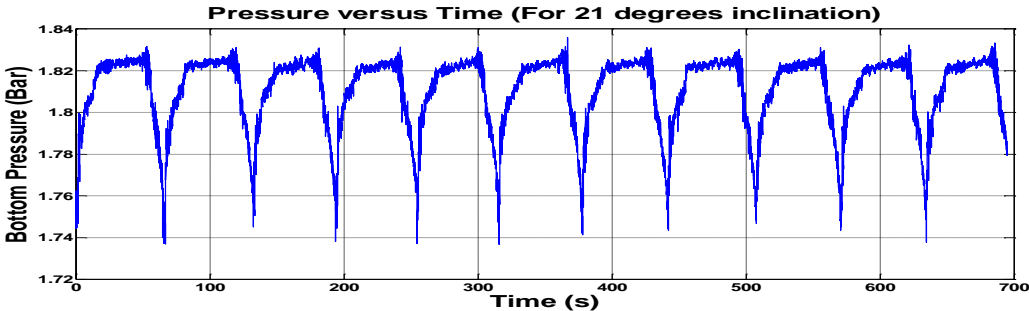


Figure 21: Gas injection rate = 2.0167E-05 kg/s

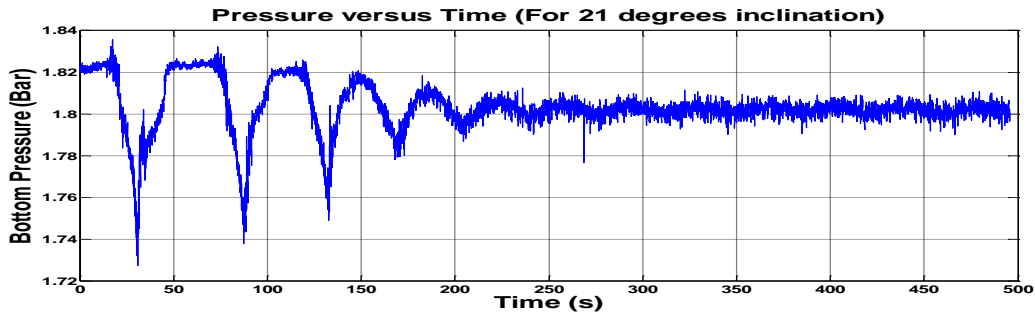


Figure 22: Gas injection rate = 3.0625×10^{-5} kg/s

Case 2A - The inlet pressure of the flowing fluid was raised to 1.906bar by increasing the height of the overflow tank to 30mm above the riser outlet. At this height the pipe was inclined to an angle of 21 degrees and pressure oscillation was observed at low gas injection rate of 5.1042×10^{-6} kg/s (Figure 20). The period for each cycle at this inlet pressure and angle of inclination was observed to be approximately 165s. Acceleration time of the gas in the riser was approximately 41s and decrease in pressure due to gas penetration into the riser was observed to be less compared to other cases considered above. Also at gas flow rate of 2.0417×10^{-5} kg/s (Figure 21) the period was decreased to approximately 63s. Increase in gas injection rate also resulted to decrease in pressure in the riser as gas penetrates the riser and resulted in mixture density decrease. Stability was attained at gas injection rate of 3.0625×10^{-5} kg/s (Figure 22) and this indicates that at slight increase of inlet pressure, not large volume of gas is required to attain stability.

B. Inclination angle of -33 degrees

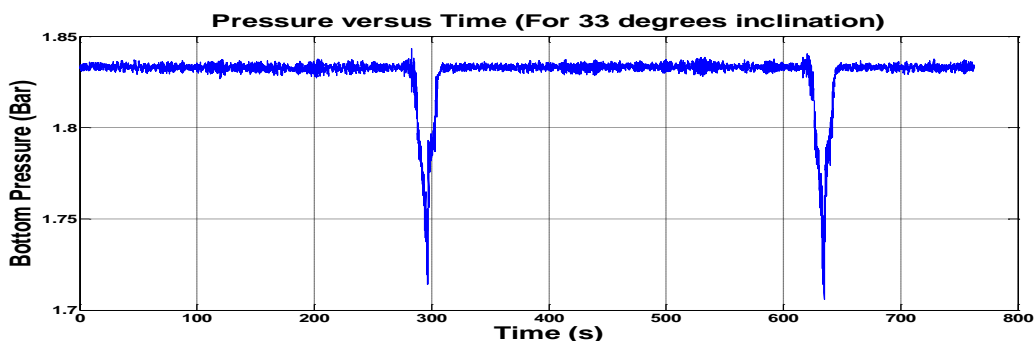


Figure 23: Gas injection rate = 5.1042×10^{-6} kg/s

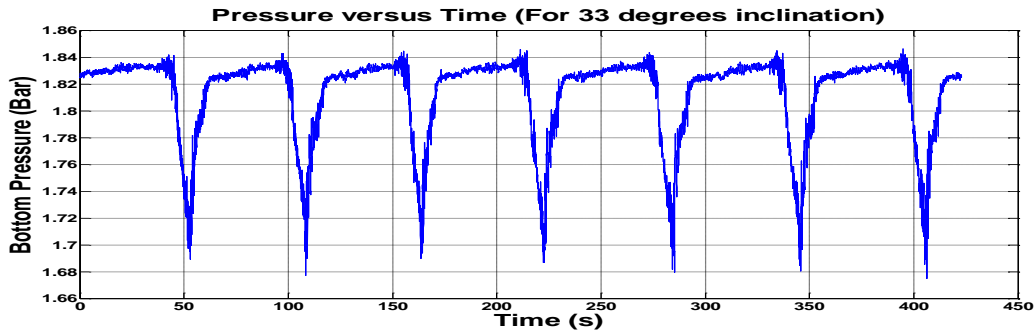


Figure 24: Gas injection rate = $4.0833\text{E-}05$ kg/s

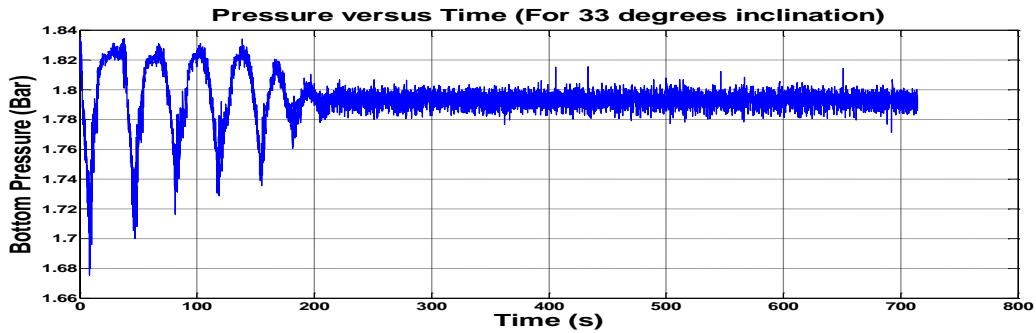


Figure 25: Gas injection rate = $6.7375\text{E-}05$ kg/s

Case 2B – With increase in inclination angle to 33 degrees the period for each cycle was increased to 338s at low gas injection rate of $5.1042\text{E-}06\text{kg/s}$ (Figure 23). Amplitude of each cycle was still the same as the cases above with the acceleration time during aeration of riser fluid at approximately 25s. The period was reduced to approximately 61s at increased gas injection rate of $4.0833\text{E-}05\text{kg/s}$ (Figure 24) which also leads to decrease in riser mixture density indicated by the low pressure rate of 1.69bar as gas penetrates the riser. Stability was attained as large volume of gas was injected into the riser at a mass flow rate of $6.7375\text{E-}05\text{kg/s}$ (Figure 25).

C. Inclination angle of -41 degrees

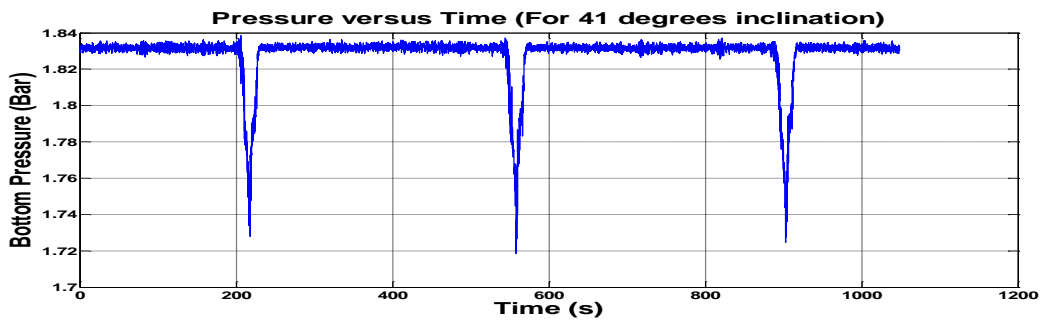


Figure 26: Gas injection rate = $5.1042\text{E-}06$ kg/s

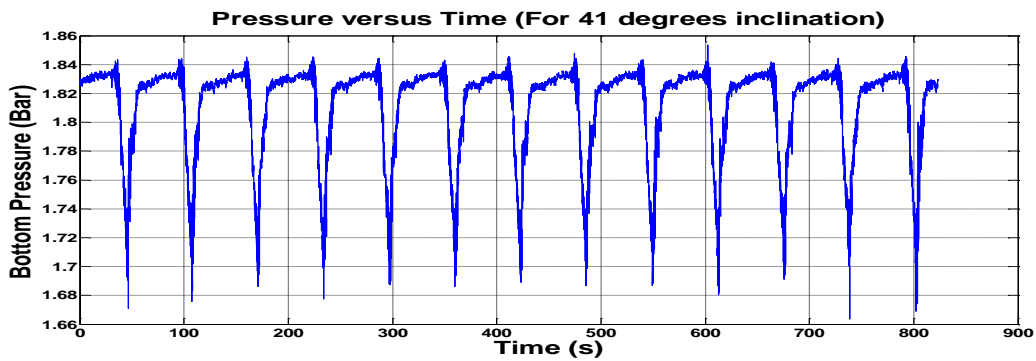


Figure 27: Gas injection rate = $4.0833\text{E-}05$ kg/s

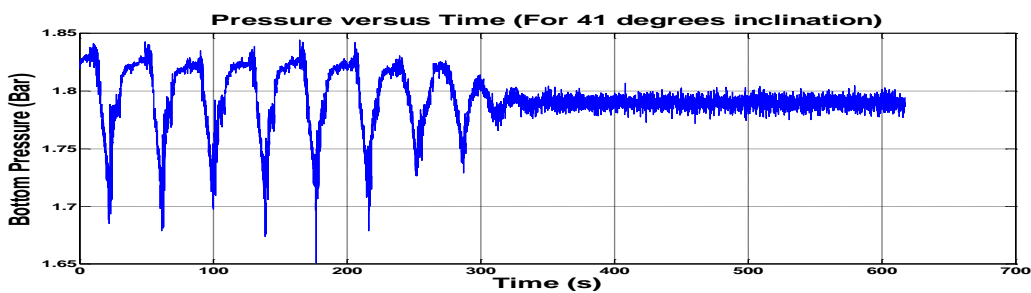


Figure 28: Gas injection rate = $7.1458\text{E-}05$ kg/s

Case 2C – The period for cycle at inclination angle of 41 degrees for gas injection rate of $5.1042\text{E-}06$ kg/s (Figure 26) was observed to be slightly higher at approximately 344s

compared to when the pipe was inclined at 33degrees. The amplitude of oscillation is the same as the cases considered above. At gas injection rate of $4.0833E-05\text{kg/s}$ (Figure 28), the period of oscillation was reduced to approximately 60s due to increased gas volume injected into the system. Stability was attained at higher gas injection rate of $7.1458E-05\text{kg/s}$ (Figure 28) which was higher compared to that required to achieve stability at inclination angle of 33 degrees.

4.1.3 Case 3: At a height of 9.33m, that is 210mm above riser outlet

A. Inclination angle of -27 degrees

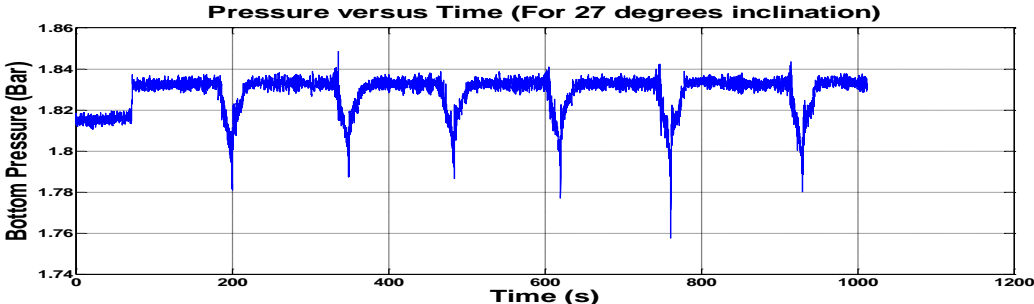


Figure 29: Gas injection rate = $5.1042E-06$ kg/s

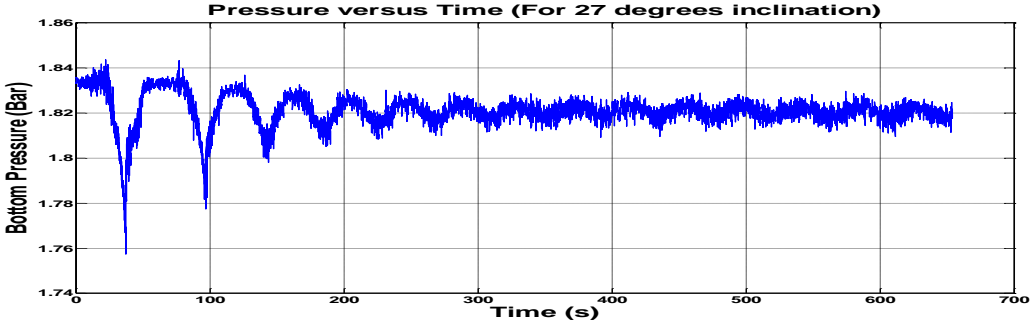


Figure 30: Gas injection rate = $2.04167E-05$ kg/s

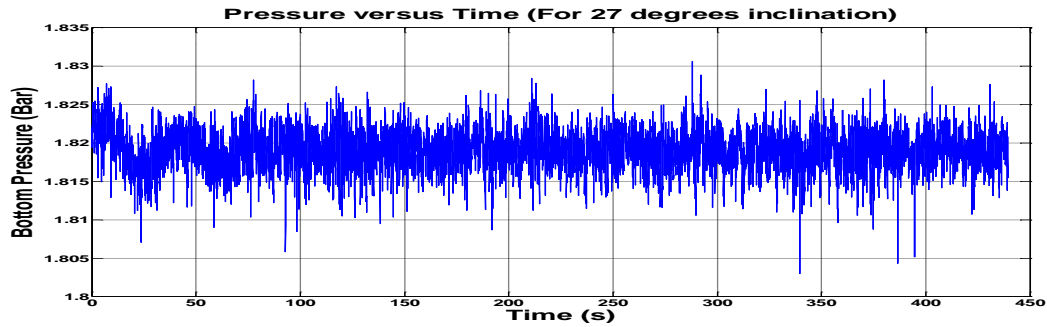


Figure 31: Gas injection rate = $2.45E-05$ kg/s

Case 3A – The inlet pressure of the flowing liquid was increased to 1.924bar by raising the height of the overflow tank 210mm above the riser outlet. And at an angle of 27 degrees and gas injection rate of $5.1042E-06$ kg/s (Figure 29), the period of one pressure oscillation cycle was observed to be approximately 136s and the ‘dip’ in pressure as a result of gas penetration was observed to be less at 1.7bar compared to case 1 and 2. The amplitude of oscillation at this condition is approximately the same as case 1 and 2. Stability was almost attained at gas injection rate of $2.04167E-05$ kg/s (Figure 30). Figure 31 shows stable flow at $2.45E-05$ kg/s.

B. Inclination angle of -36 degrees

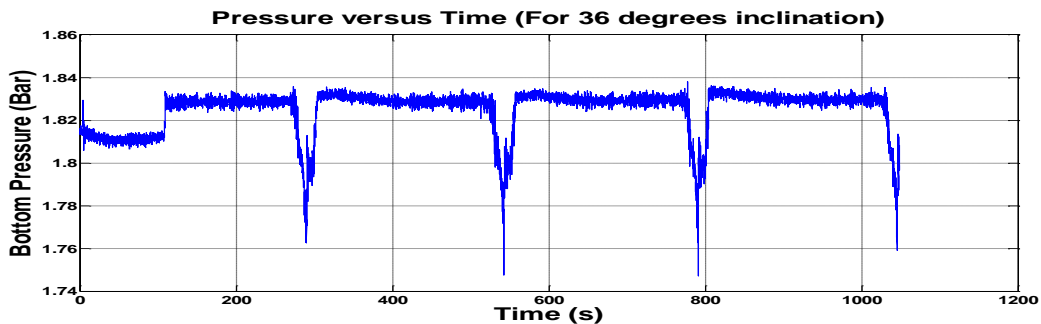


Figure 32: Gas injection rate = $5.1042E-06$ kg/s

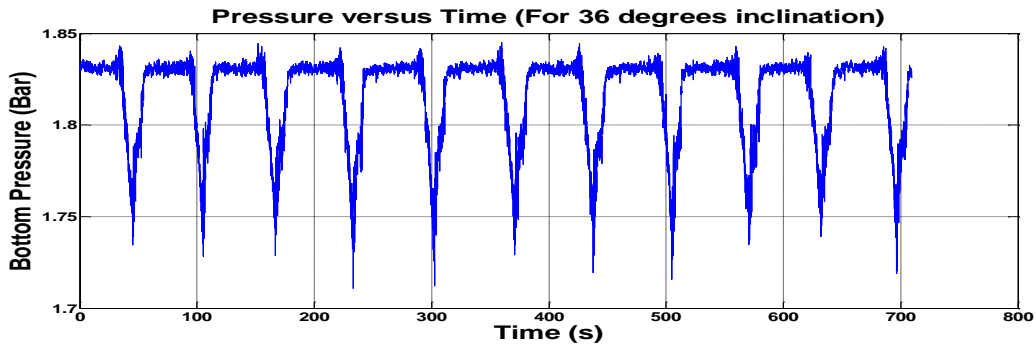


Figure 33: Gas injection rate = $3.0625E-05$ kg/s

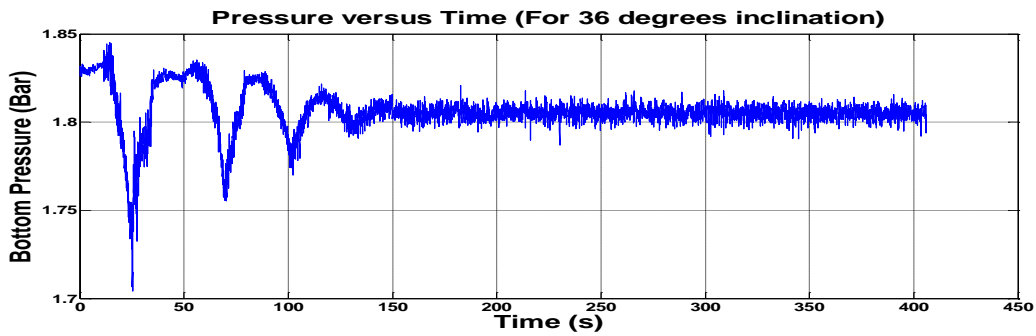


Figure 34: Gas injection rate is = $4.6958E-05$ kg/s

Case 3B - With increased angle of inclination of 36 degrees and gas injection rate of $5.1042E-06$ kg/s (Figure 32). The period for one cycle of unstable flow was increased to approximately 250s. The amplitude as well as the ‘dip’ in pressure was observed to be the same as the case above. The acceleration time of gas flow in the riser is approximately 30s. Increase in gas injection rate to $3.0625E-05$ kg/s (Figure 33) reduced the period to 66s and stability was attained at higher gas injection rate of $4.6958E-05$ kg/s (Figure 34) compared to that required at an inclination angle of 27 degrees.

B. Inclination angle of -45 degrees

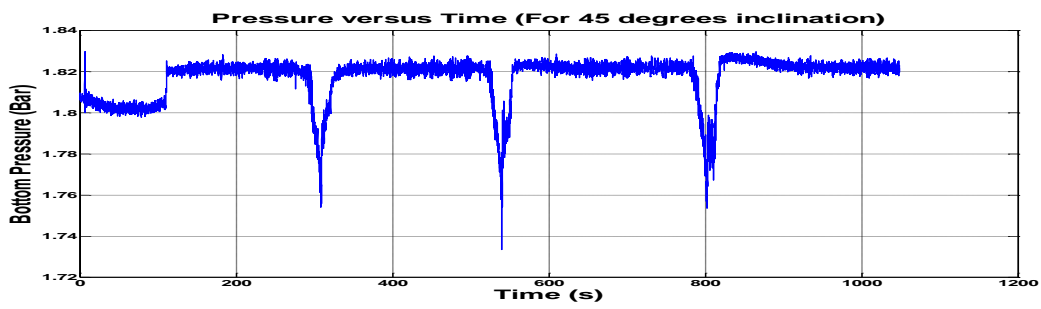


Figure 35: Gas injection rate = $5.1042E-06$ kg/s

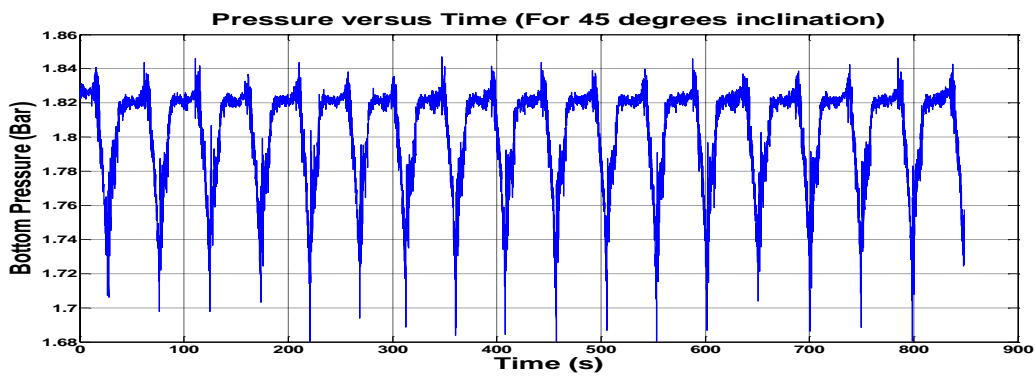


Figure 36: Gas injection rate = $5.1052E-05$ kg/s

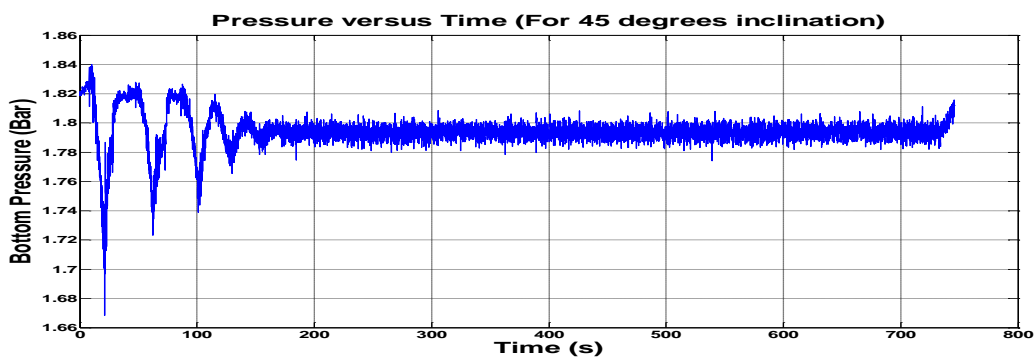


Figure 37: Gas injection rate = $6.125E-05$ kg/s

Case 3C – Increased inclination angle to 45 degrees at gas injection rate of 5.1042E-06kg/s (Figure 35) was observed to show an increase in the period of an oscillation cycle to approximately 264s compared to case 3A and 3B at the same gas injection rate. Oscillation amplitude was observed to be similar to cases above and acceleration time was observed to be approximately 33s. Increase in gas injection rate to 5.1052E-05kg/s (Figure 26) resulted in decrease of the oscillation to approximately 50s. Stability was attained at higher gas injection rate of 6.125E-05kg/s (Figure 37).

4.2 Effect of overflow tank height on pressure oscillation observed in the vertical pipe

Increase or decrease of the overflow tank height above or below the riser outlet increases or decreases the inlet pressure of the flowing fluid respectively. The pressure of the flowing fluid from the reservoir plays a significant role in providing the energy required lifting reservoir fluid to the surface and it also determine the stability of the flowing fluid even in an undulating plain. For natural-lift system or gas-lift systems, inlet pressure of the flowing fluid was observed to play a significant role in the stability of the system.

To determine the effect of inlet pressure on expansion driven flow instability in long vertical pipes, the height of the overflow tank was varied between above and below the riser outlet by using the riser outlet as the boundary condition. Increase in the overflow tank height above the riser outlet was observed to increase the inlet pressure of the flowing fluid which also enables stabilization of the riser fluid at slight inclination and slight increase in gas injection rate compared to decrease in the overflow tank below the riser outlet where sufficient gas is required for stability of flow at the slight inclination of the flowline close to the riser inlet.

Comparison of the effect of the height of the overflow tank (or the flow inlet pressure) on the stability of the gas injection system at different inclination is made in Table 3 below:

Table 3: Effect of overflow tank height on stability of system

Height of the overflow tank [mm]	Stability of the flowing fluid
213mm below riser outlet @	@ 5.1042E.05kg/s

-25deg.	
210mm above riser outlet @ -27deg.	@ 2.04167E-05kg/s

Despite slight increase in inclination angle at the height of 213mm above the riser outlet, stability of the riser fluid was attained at lower gas injection rate compared to the amount required to attain stability at 213mm below the riser outlet. This establishes the fact that the height of the overflow tank, and invariably the inlet flow pressure, determines the magnitude of expansion driven flow instability in a natural lift of gas-lift system.

4.3 Effect of inclination angle on pressure oscillation observed in the vertical pipe

To determine the influence of the angle of inclination on expansion driven flow instability, air and water are transported to the riser through a downward inclined pipe at different inclinations: 21,25,27,33,36,41 and 45 degrees. These inclination angles were considered for varying height of the overflow tank and gas injection rate. It was observed that inclination angle at varying overflow tank height and gas injection rate plays an important role in the extent of expansion driven flow instability in vertical pipe. Stability was observed to be attained faster at low inclination compared to high inclination of the downward pipe to the riser inlet. Below Table 4 and Table 5 are stability gas rates at different angle of inclination for different heights of the overflow tank:

Table 4: Stability gas rate at different inclination angles for 1.822 bar Inlet pressure

Inclination angle [Θ]	Stability gas rate [kg/s]
@ -27	2.04167E-05
@-36	4.6958E-05
@-45	6.125E-05

Table 5: stability gas rate at different inclination angles for 1.924 bar Inlet pressure

Inclination angle [Θ]	Stability gas rate [kg/s]
@-25	5.1042E-05
@-33	5.7167E-05
@-45	8.1667E-05

From the Table 4 and Table 5 above, large volume of gas is required to attain stability at increased angle of inclination. At overflow tank height of 213mm below riser outlet, approximately 60percent increase in gas mass flow rate is required at to attain stability at 45degrees in comparison to the amount required at -25 degrees. Also at an inclination angle of 45 degrees, approximately 33 percent increase in gas flow rate is required at overflow tank height of 213mm below riser outlet in comparison to the amount required at 210mm above the riser outlet. This implies that angle of inclination also plays a significant role in promoting expansion driven flow instability in long vertical pipes. This angle of inclination can be determined by the undulation of the plain on which a flowline is laid. .

4.4 Flow Regime

4.4.1 Flow regime classification

Flow regime are developed when gas and liquid flow in a pipe, at varying superficial velocity. Factors that determine flow regime pattern in a pipe include: geometry, flow rate, inlet conditions like the inlet pressure of the flowing fluid. Flow regime classification depends on whether the gas-liquid are flowing in horizontal or vertical pipes. Flow regimes observed in horizontal pipes can either be: stratified, bubble, annular or slug flow (Figure 38). While flow regimes observed in vertical pipes can be: churn, bubble, annular and slug flow as stratified flow does not occur in vertical pipe (Figure 39).

Stratified flow is a type of flow that occurs at low liquid and gas flow rates in horizontal and downward inclined pipe and has widely gained attention in the research world. Mathematical

models for stratified flow has been developed and depends on pressure drop, cross-sectional area of the pipe, interfacial friction between the fluid and the wall, interfacial friction between the liquid and gas, angle of inclination, acceleration due to gravity, shear stress and perimeter of the wall on which the stress acts on. For horizontal pipe inclination angle is not taken into consideration.

For the pipeline-riser system used for this experiment, slight initial stratification in the downward inclined pipe section to the riser inlet was observed then followed by subsequent blockage of the upstream liquid by gas as the gas accumulates and expands towards the riser inlet until gas penetrate the riser.

Slug flow is a type of flow pattern which occurs in both vertical and horizontal pipe in which large 'Taylor bubbles' with underlying liquid with tiny bubbles is followed liquid slug with tiny bubbles. This type of flow occurs in pipe at low liquid and gas rates. In horizontal pipes, slug flow occurs when Kelvin Helmholtz criteria is satisfied. It occurs when waves in a stratified flow grows and reaches the top of the pipe leading to the throttling of the gas and the liquid upstream. Slug flow occurs in many industrial processes such as the transportation of hydrocarbons in pipes, in air-lift systems, bioreactors, chemical reactors, etc. In horizontal pipes, slug flow is divided into separated and mixed flow region. The region of large 'Taylor bubbles' with underlying liquid with tiny bubbles is considered as separated region while the region of slug flow with tiny bubbles is considered as the mixed region. In vertical pipe, the long bubble completely or nearly fills the cross section followed by liquid slug. Bubble and liquid slug length depends on coalescing or fragmentation of long bubbles. Slug flow is the most common type of flow regime in vertical pipes.

In the experiment, expansion driven flow instability gave rise to a slug flow behaviour at low gas injection rate with long Taylor bubbles in the vertical pipe (riser) with bubbles coalescing and expanding as it rise in the liquid column. The bubble is followed by liquid slug with tiny bubbles. Taylor bubbles gets longer due to coalescing and expansion of bubbles. Stability was attained at increased gas injection rate which gave rise to slug flow with short bubbles as coalescing of bubbles is eliminated.

Annular flow occurs at low liquid flow rate and high gas flow rate with the liquid dispersed as liquid droplets in the core gas. In vertical pipes, at high gas and low liquid rate, the liquid is visualized a thin film on the wall of the pipe with gas core in the middle with liquid droplet entrainment. In horizontal pipes, flow can be asymmetric with gas and liquid droplet

entrainment in the core. Annular flow was not observed during the experiment as the experiment was not carried out at high gas and low liquid flow rate.

Churn flow is an unstable flow which occurs in vertical pipes in which Taylor bubbles are deformed which can lead to transition from slug flow to bubble flow. This type of flow behaviour was also not observed during the experiment.

Bubble flow is a type of flow pattern that occurs in horizontal and vertical pipes at either low or superficial velocities. Horizontal pipes, gas bubbles are dispersed in the continuous liquid phase with large part of the dispersed bubbles at the upper part of the tube due to buoyancy and are common when bubbles coalesce to form large bubbles there by concentrating at the upper part of the pipe. Dispersed bubble flow can also occur in horizontal pipe where tiny bubbles are dispersed in the continuous liquid phase at high flow rate of the liquid. In vertical pipes, discrete bubbles are dispersed in the continuous liquid phase having varying shapes and sizes. The flow in vertical pipes is normally symmetric. During the experiment, bubble flow was not observed.

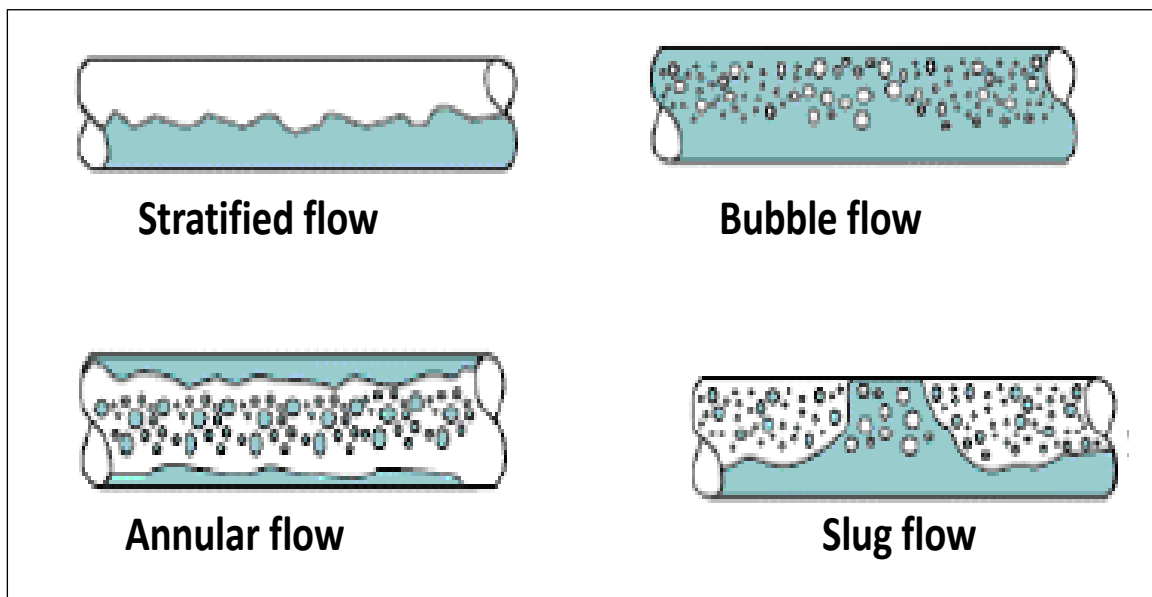


Figure 38: Flow regimes in horizontal pipes (Ali, 2009)

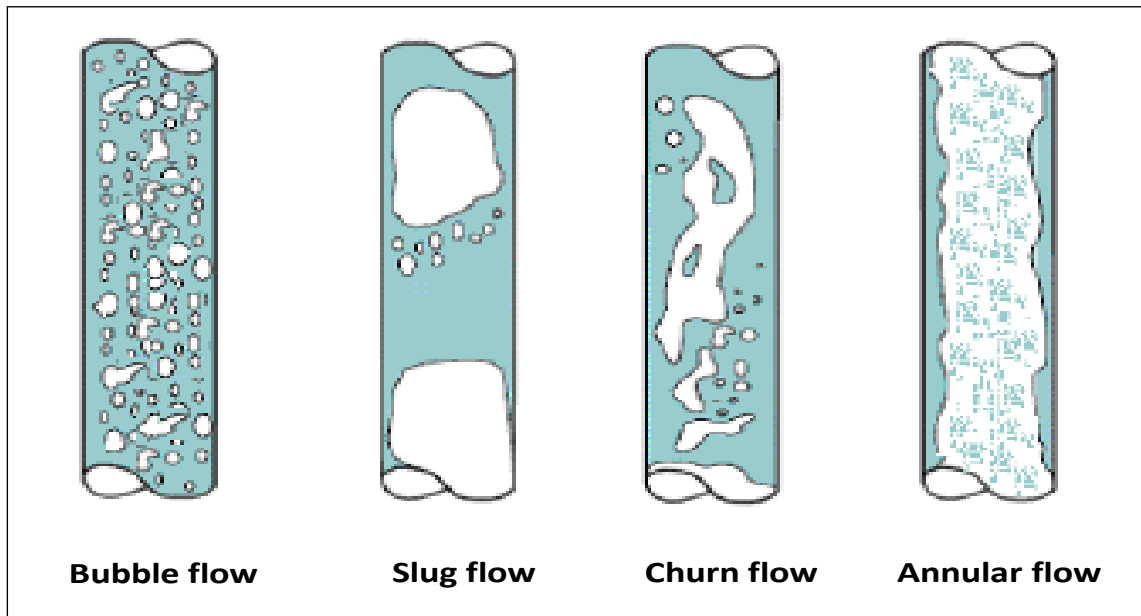


Figure 39: Flow regimes in vertical pipes (Ali, 2009)

4.4.2 Visual flow behaviour along horizontal-pipe

Flow behaviour of the gas-liquid mixture in vertical pipes was observed to depend on flow rates, inclination angle and inlet condition. The flow pattern observed during this experiment was slight stratification at the downward pipe which later disappears as gas accumulates and expands in the downward leading to blockage of upstream fluid. Flow instability was observed in the form of slugs whose length depended on the coalescing and expansion of bubbles during acceleration in the vertical pipe. Stable flow was observed as slug flow with short bubbles and longer residence time during acceleration leading to stable flow out of the vertical pipe.

The effect of different gas injection rate on the stability of the horizontal-vertical pipe system was recorded and has been attached as appendix D. The effect of different gas flow rate at an angle of 45 degrees and inlet pressure of 1.924bar was observed. The flow rate considered for the video recording are:

- Gas injection rate = 5.1042E-06 kg/s
- Gas injection rate = 3.0625E-05 kg/s
- Gas injection rate = 4.0833E-05 kg/s

Snapshots from the video showing the stages during one expansion driven flow instability cycle was taken and shown in Figure 40 - Figure 47 below. Also, flow behaviour in the riser is shown in Figure 48 and Figure 49 below. Expansion driven flow instability was observed to be of cyclic behaviour in the following manner:

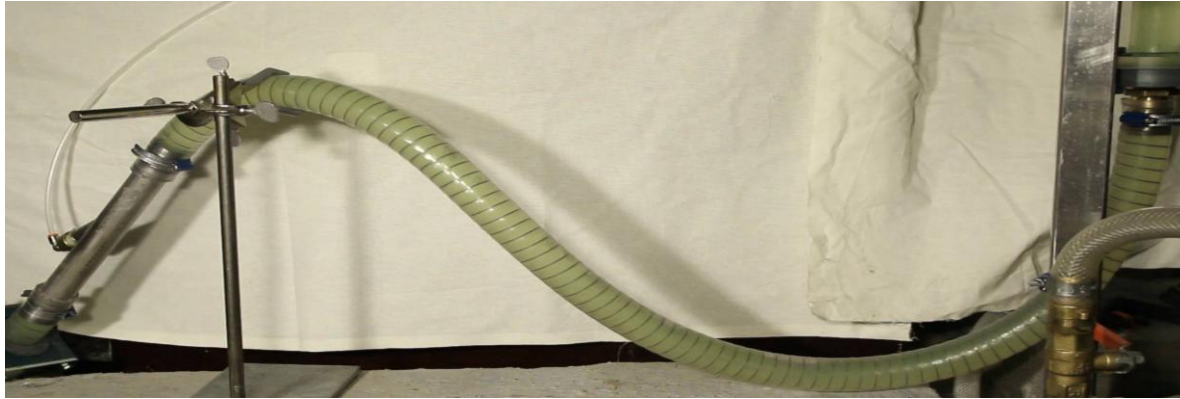


Figure 40: Initial filling of the riser with liquid at low inlet pressure.

Figure 40 shows the beginning of the cycle where liquid was initially pumped and filled the horizontal-vertical pipe. Liquid flow rate out of the riser was observed to be low at low pressure of the flowing fluid using pump frequency of 26.8Hz to maintain low water flow rate. This low inlet pressure was observed to be insufficient to lift the fluid out of the riser outlet.

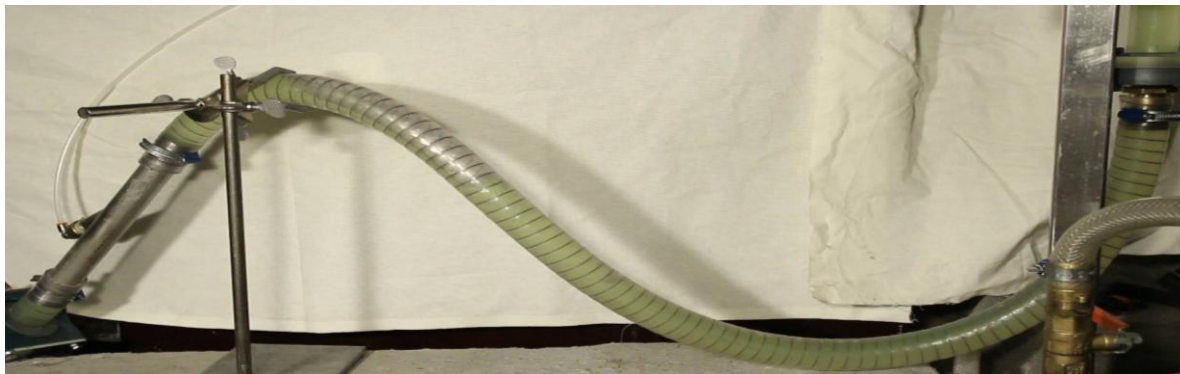


Figure 41: Flow behavior at low gas injection

Figure 41 shows initial stratification as gas was injected upstream riser inlet. Gas was trapped due to blockage by liquid column in the riser.

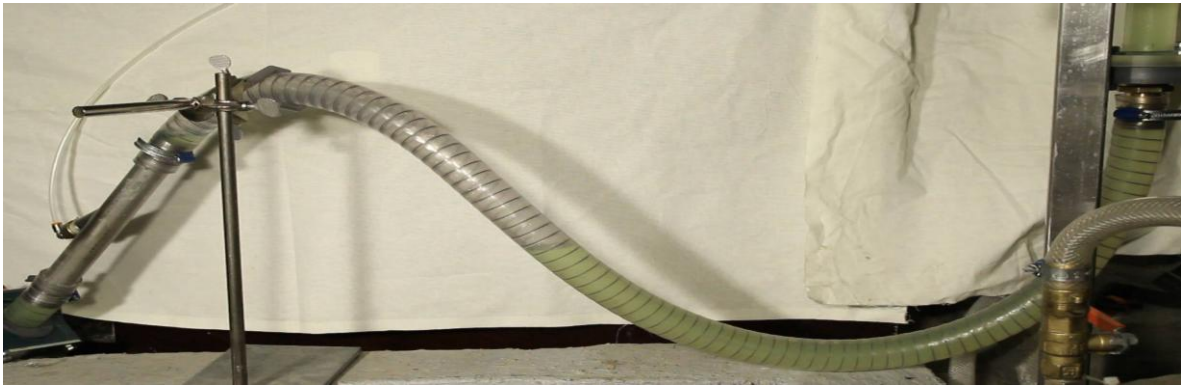


Figure 42: Gas accumulation and expansion.

Figure 42 shows further accumulation and expansion of gas trapped. The initial stratification has disappeared as gas pushes the upstream fluid backward as it expands.

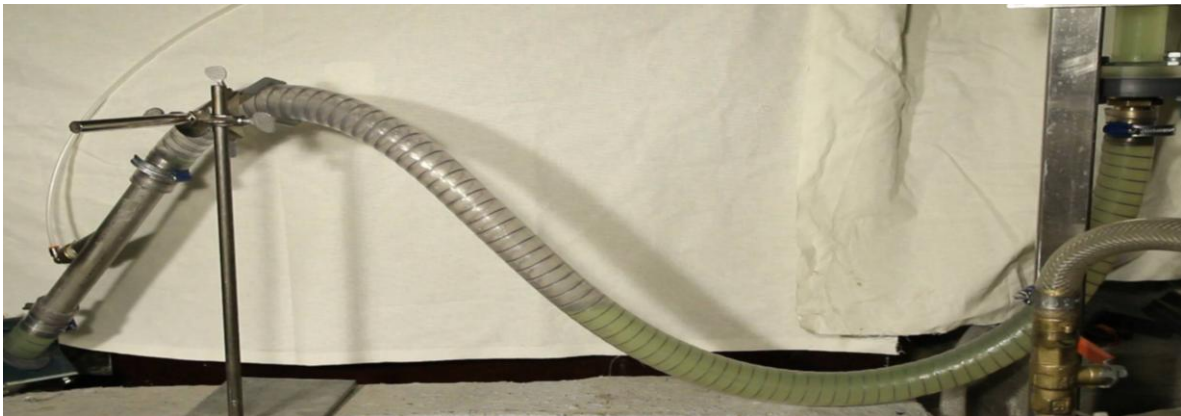


Figure 43: Further gas accumulation and expansion as well as pushing back of upstream fluid

Figure 43 shows further accumulation and expansion of gas towards the riser inlet as well as further push of the upstream fluid.



Figure 44: Gas penetration into the riser

Figure 44 shows point of gas penetration into the riser leading to the aeration of the riser fluid. This resulted in decrease of riser pressure as upstream fluid was observed to rising to flow towards the downward inclined pipe.

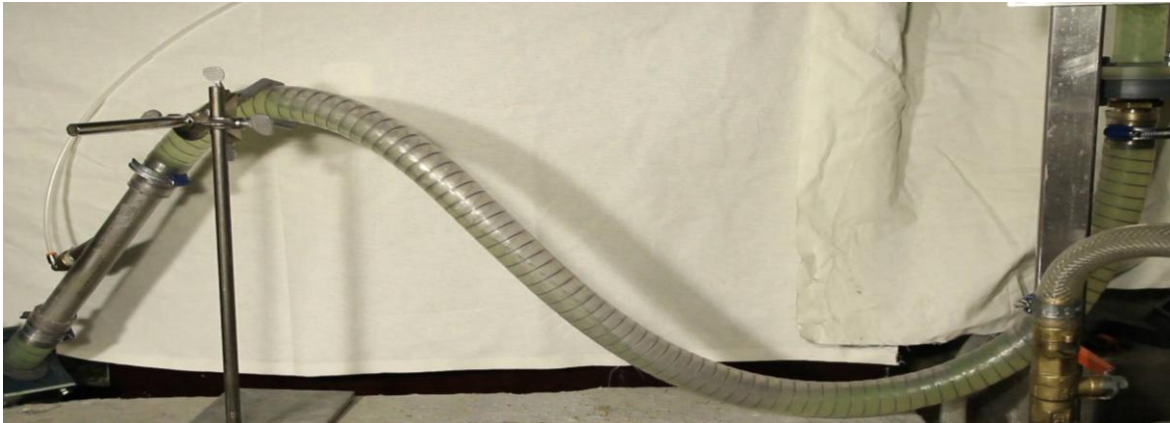


Figure 45: Further penetration due to aeration of riser fluid

As shown in Figure 45, further penetration of gas into the riser resulted in aeration of riser fluid as well as decrease in riser pressure as riser mixture density is decreased. This lead to increased inflow of upstream fluid to the riser as the bottom pressure is decreased.

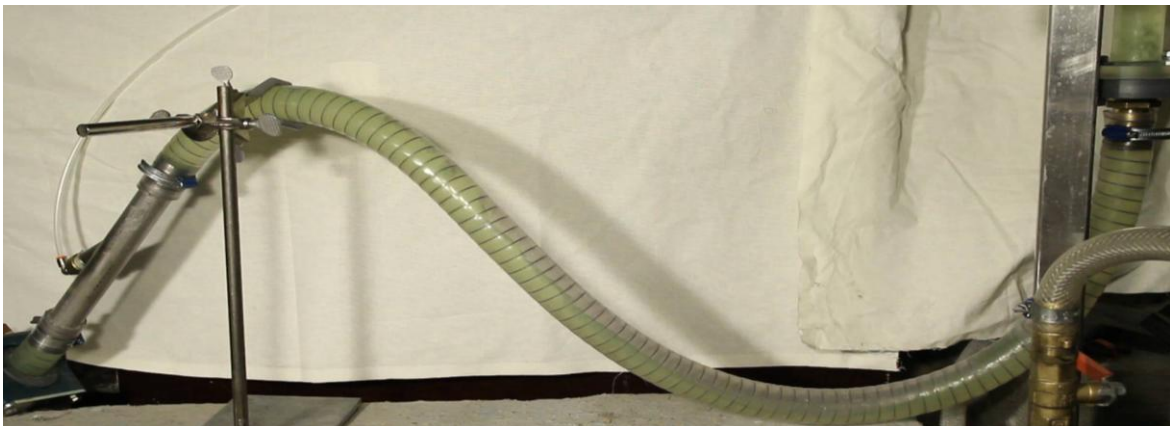


Figure 46: Inflow of upstream fluid due to aeration

Figure 46 shows further inflow of upstream fluid as a result of increased aeration in the riser.

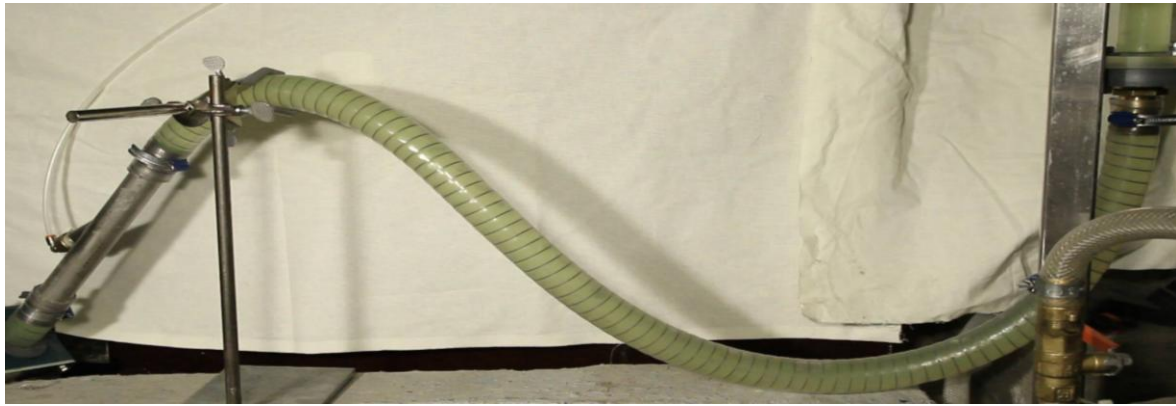


Figure 47: The beginning of another cycle

Upstream fluid as displaced all the gas in the downward pipe filling it up with liquid as well as inclined displacement of gas in the riser as the mixture density rises due to increased liquid flow. Increased mixture density also leads to increased bottom pressure at the riser base. And the cycle begins again. This is shown in Figure 47.



Figure 48: riser with water and bubble behavior due to instability

Figure 48 shows the riser filled with liquid initially then due to entrapment of gas in the downward inclined horizontal pipe instability is established resulting in slug flow behaviour in which bubbles in the riser coalesce as they rise.

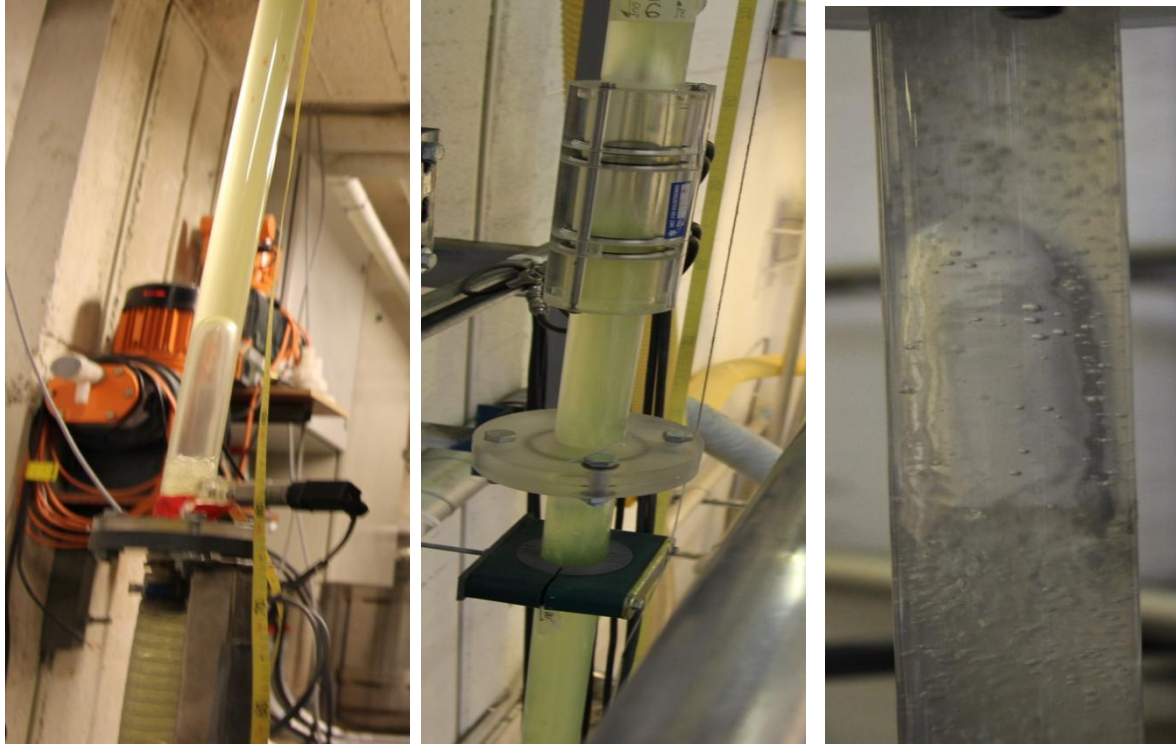


Figure 49: bubble length as a result of coalescing and bubble passing through impedance probe

Coalescing of bubbles resulted in longer bubbles as shown in Figure 49. This variation in bubble behaviour is captured as the bubble passes through the impedance probe.

4.4.3 Stability

In the experiment stability is attained as smooth flow pattern transition occurs from slug flow to bubbly flow. The void fraction waves are dampened and the damping increases as the mean void fraction increases. At high inlet pressure and low or no inclination of the flowline pipe, stability was observed in the long vertical pipe. Also at either high liquid flow rate or high gas flow rate the oscillatory behaviour experienced for expansion driven flow instability is strongly dampened. In this work, gas injection rate, angle of inclination of the downward pipe to the vertical pipe and inlet pressure of the flowing fluid were observed to influence the stability of the flowing fluid.

A test of the pipe at overflow height of 213mm below the riser inlet was carried out to test stability of the system at no inclination angle. Different gas injection rate were tested and all gave stable flow at the given condition. Below are graphs for different gas injection at no inclination angle and height of 213mm below riser outlet.

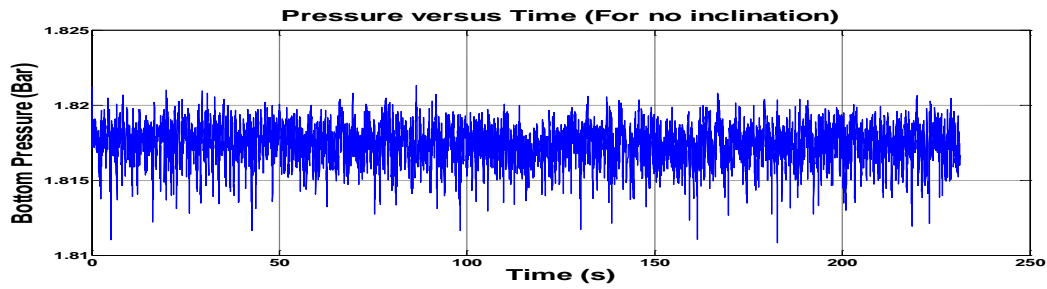


Figure 50: Gas injection rate = 5.1042E-06 kg/s

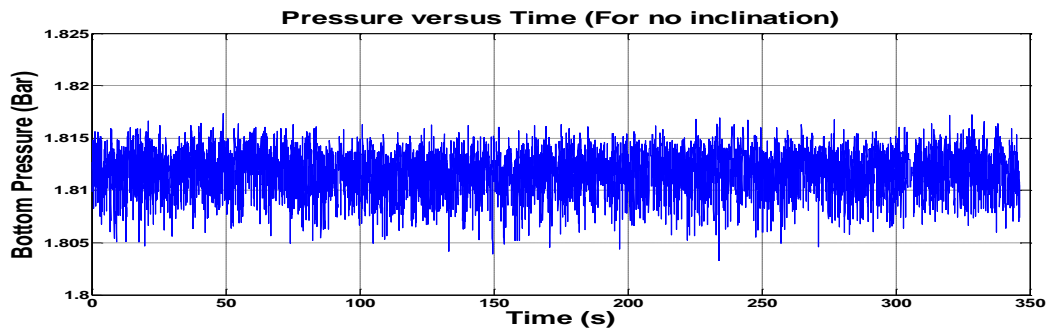


Figure 51: Gas injection rate = 2.04167E-05 kg/s

From Figure 50 and Figure 51 above, regardless of gas injection rate, flow stability was attained even at low gas flow rate. This implies that amongst inlet pressure, gas injection rate and inclination angle, inclination angle plays a vital role in determining the stability of the flowing fluid in the vertical pipe.

Expansion driven flow instability in long vertical pipes experiment involves oscillatory behaviour at no or low friction dampening, but most flow meters designed to measure low liquid flow rate has low internal flow diameter, and using a flexible pipe of 32mm across the flow meter results in pressure drop induced friction which results in oscillation dampening. Less friction means less dampening of flow variations which can lead to stability of flow system. Friction was observed to strongly dampen the oscillatory behaviour in the vertical pipe giving rise to stability of the system. Different small flow meter (coriolis flow meter, venturi tube, rotameter) for low liquid flow rate were used for measurement of the liquid flow rate, but due to the internal diameter of the flow meter, pressure drop was observed across the flow meters. This friction effect is transported down the flowline system resulting in oscillation dampening even at low gas injection rate. Therefore, stability was observed to be attained at even very low gas flow rate.

4.5 Gas Injection characteristics

In this work, gas was injected upstream the riser base at varying gas injection rates, depending on the inclination angle and inlet pressure of the flowing fluid. Gas rate at each flow condition was kept constant and pressure behaviour in the vertical pipe at each gas flow rate - recorded. The logging in amperes at one of the gas flow rate under flow condition in consideration (gas injection rate of $5.1042\text{E-}06\text{kg/s}$) is shown in Figure 52 below to ascertain that it is constant. 1% variation was observed from the graph and this might be attributed to systematic errors such as in the form of electrical noise, accuracy of the air flow meter etc.

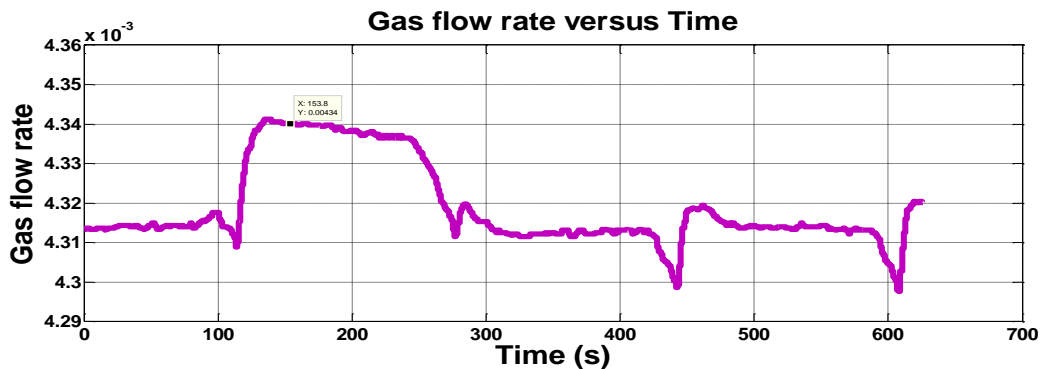


Figure 52: Gas injection rate = $5.1042\text{E-}06\text{ kg/s}$

Gas injected in the flowline at some distance upstream the riser inlet can be likened to the effect of varying inlet gas rate to the overall behaviour of a flowing system. At low gas injection rate, gas was observed to accumulate and expand in the downward pipe to the riser inlet. This accumulation and expansion of the injected gas lead to the blockage of the upstream fluid which also leads to back-pressure as the flow meter (in use) indicator along the liquid line indicates ‘no liquid’ flowing along the waterline. Gas penetration into the vertical pipe aerates the fluid by decreasing mixture density in the pipe leading to sudden inflow of liquid into the riser. As gas accumulates and expands, pressure sensor at the riser detects an increased pressure due to increasing water head. But gas penetration into the vertical pipe lead to a sharp pressure decrease detected by the pressure transducer. Also, at low gas injection rate slightly higher values of oscillation periods were observed for a flow with smaller gas slugs. Increasing gas injection rate was observed to increase stability at the varying inclination angle and inlet pressure of the flowing system as oscillation amplitude was observed to decrease with increasing gas injection rate. This implies that gas lift or natural lift increases stability along the axis of superficial gas velocity.

4.5.1 Void fraction characteristics

Void wave propagation was observed to control the EDI phenomenon. In this experiment gas flow rate was observed to affect void wave propagation as the rise in velocity of the bubble decreases with increasing volume fraction of the bubbles. These were observed at high flow rate of gas injection leading to sufficient residence time required to optimize production. In contrast, at low gas flow rate, coalescing of bubbles in the presence of other bubble leads to slug flow (Taylor bubbles with underlying liquid having tiny bubbles followed by liquid slug having tiny bubbles of varying sizes), low residence time and rise in velocity of the bubble occurs in a cyclic oscillatory manner giving rise to expansion driven flow instability. The gas bubbles were observed to be approximately the diameter of the pipe and the nose of the bubble has a characteristic spherical cap. Coalescing of bubbles to form long Taylor bubbles can be related to strong pressure oscillation in the system. Increase in the length of the bubbles can be directly related to the expansion of the gas phase which is a function of the hydrostatic pressure acting on the bubbles.

Also, at low gas injection rate, the slug length was observed to be long compared to when the gas flow rate was increased to the optimum desired rate, where the slug length was reduced and the flow was observed to be bubbly causing a more continuous production at the riser outlet and also a lower system pressure. The gas bubbles were observed at low gas injection rate is approximately the diameter of the pipe and the nose of the bubble has a characteristic spherical cap. Coalescing of bubbles to form long Taylor bubbles can be related to strong pressure oscillation in the system. And increase in the length of the bubbles can be directly related to the expansion of the gas phase which is a function of the hydrostatic pressure acting on the bubbles.

Void fraction fluctuations caused by unstable flow are measured by processing the signals of the impedance probes with its electrode rings (two each with distance apart) mounted at three locations along the riser (vertical pipe). Six impedance probes in all are flush mounted on the wall but only the signals of four probes are processed in this study. Probe 8 seems to have systematic errors; therefore probe 7 and 8 was not processed. The impedance probe signals are processed to obtain the void fraction fluctuations. Logging speed was made to be to be 10milliseconds as the duration for one cycle during gas penetration into the riser is short. Figure 53 - Figure 55 shows the void wave propagation with respect to impedance variation at its normalised signal, inlet pressure of 1.882bar, inclination angle of 33 degrees. Figure 53 shows impedance variation at gas injection rate of $5.1042E-06$ kg/s. At this gas injection rate,

gas penetration into the riser showed 70% void fraction as gas, coalesces, expands and accelerates through the riser. Void fraction disturbance was observed to be initiated at some critical value of void fraction. Increase in gas injection as shown in Figure 55 shows a stable flow with increased residence time and ‘no’ coalescing which can be attributed to increase in void fraction.

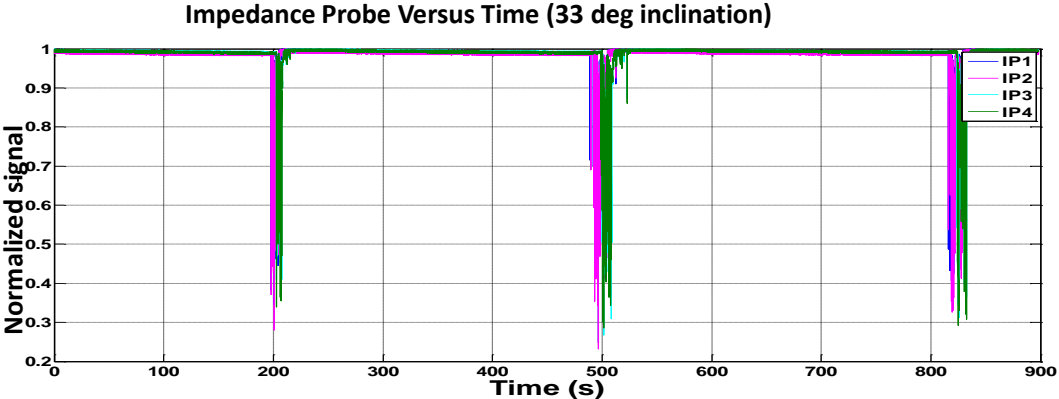


Figure 53: At gas injection rate = $5.1042E-06$ kg/s, inclination angle of -33 degrees and inlet pressure of 1.882bar

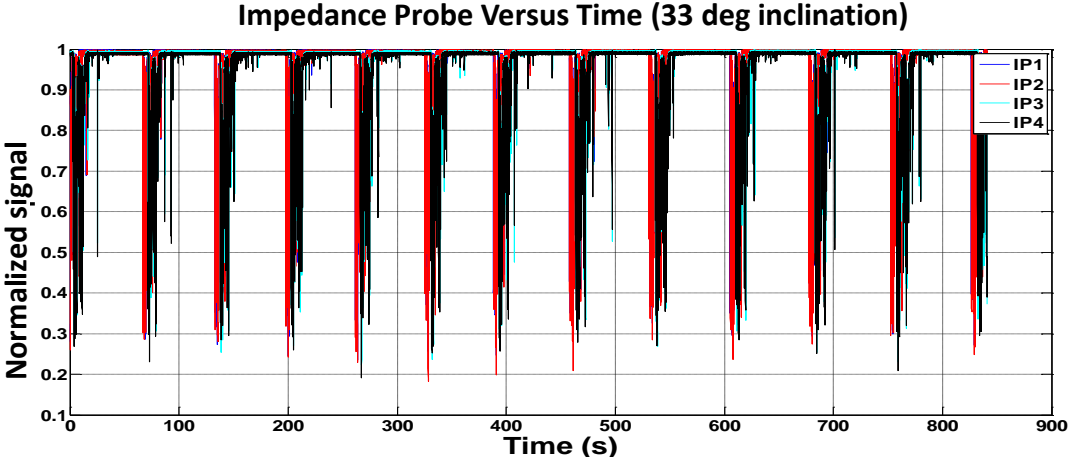


Figure 54: At gas injection rate = $3.0625E-05$ kg/s

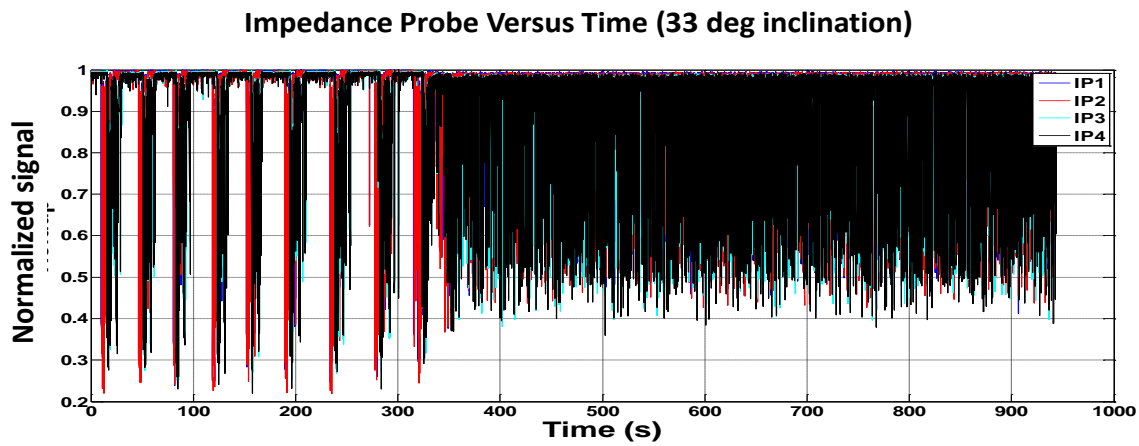


Figure 55: At gas injection rate = $5.7167\text{E-}05$ kg/s

4.5.2 Vertical pipe-base pressure characteristics

Pressure drop fluctuation in the vertical pipe as recorded using data logging furnishes a convenient means of analysing expansion driven flow instability in the vertical pipe. Pressure fluctuation in the vertical pipe was captured with the aid of pressure transducer flush mounted on the wall of the vertical pipe at the bottom and top of the vertical pipe. The pressure fluctuation at the base of the vertical pipe was reported in Figure 11 - Figure 37. Pressure gradient was observed to depend largely on flow regime, gas injection rate, inlet pressure of flowing fluid and inclination angle amongst others.

As liquid fills the vertical pipe column, pressure build-up was observed in the vertical pipe that is equivalent to the hydrostatic pressure above it. This maximum pressure in the riser results in gas blockage of the two-phase fluid flowing down the downward inclined pipe at low gas injection rate to the vertical pipe inlet as the pressure in the vertical is above the inlet pressure of the flowing fluid. Gas blockage, accumulation, expansion and subsequent penetration into the vertical pipe leads to decrease in the bottom (base) pressure which enhances inflow of the upstream fluid into the vertical pipe. Mixture density in the vertical pipe decreases as gas penetrates the vertical but this regained with inflow of upstream fluid. Also, it was observed that as gas bubbles move up the vertical pipe column at low gas injection rate, the pressure acting on each of the bubble decreases as it coalesce and expand along the column.

4.6 Summary of experimental results

The three cases considered during experimental analysis can be summarised as follows:

Case 1: Inlet pressure at 1.882bar

This was achieved by lowering the overflow tank height 213mm below the riser (9.12m) outlet. Varying inclination angles as well as flow rates were considered under this case. This was achieved by subdividing the case into three sub-cases (Case 1A, Case 1B and Case 1C). Inclination angles considered for the cases are: 25, 33, and 45 for cases Case1A, Case 1B and Case 1C respectively. Instability was observed at very low gas injection rates for the three cases. But, comparison of the three cases (Figure 13, Figure 16 and Figure 19) shows that not much gas injection is required to achieve stability at low inclination angle and the period of oscillation increases at increasing inclination angle considering same gas flow rates (Figure 11, Figure 14 and Figure 17).

Case 2: Inlet pressure at 1.906bar

Height of the overflow tank was raised 30mm above riser outlet. Similarly varying inclination and gas flow rates were considered. This was done by subdividing the case into three sub-cases (Case 2A, Case 2B and Case 2C). Inclination angles considered are: 21, 33, and 41 for Case 2A, Case 2B and Case 2C respectively. Gas was injected at varying flow rates for the three sub-cases. Instability was observed at low gas injection rates in the three cases. High amount of gas rate is required to attain stability at increased inclination angle (Figure 22 and Figure 28) and period of oscillation at high inclination angle is higher compared to low inclination angle at same flow rate. Increase in the height of the overflow tank amounted to an increase in the inlet pressure of the flowing fluid and compared to Case 1 this amount to attaining stability faster (Figure 20 and Figure 26).

Case 3: Inlet pressure at 1.924bar

Height of the overflow tank was raised to 210mm above the riser outlet. Varying inclination angles and gas flow rate were considered under sub-cases (Case 3A, Case3B and case 3C). Inclination angles considered respectively for the sub-cases are 27, 36 and 45. Varying gas rates was used for each sub-case and instability was observed at low gas injection rate for each of the case. Instability was observed to be longer for a wide gas rate range at high

inclination angle (comparing Figure 29 - Figure 31) thereby requiring large volume of gas. Also period of oscillation was observed to be longer at high inclination angle compared to low inclination at the same gas injection rate Figure 29 - Figure 35. Stability was attained faster in Case 3 compared to Case 2 at the same inclination angle (Figure 37 and Figure 28) this is due to increased inlet pressure.

Chapter 5

5.1 Simulation of horizontal-vertical pipe system

Simulation using OLGA software was carried out to determine expansion driven flow instability in long risers/wells using air-water system and the experimental flow conditions. Results from the simulation were compared with results obtained from the experimental studies. The OLGA model simulated a complete expansion driven flow cycle in a horizontal-vertical pipe system where the pressure of flowing liquid is modelled by an overflow tank system, with a short horizontal pipe of varying inclination to the riser and also point of gas injection. Point of gas accumulation and expansion (onset of expansion driven flow instability), back pressure build-up upstream the flowline, liquid slug build-up in the riser and subsequent penetration of gas and production are described. The pressure at the top of the riser is prescribed as atmospheric.

OLGA model was developed by the Norwegian Institutes SINTEF and IFE. It is a dynamic multiphase flow simulator. OLGA software uses one-dimensional implicit two-phase flow model concept and find its origin in computer programs that were originally developed for the nuclear industry. It accounts for flow regimes, droplet entrainment and deposition. It consist of three separate continuity equations for gas, liquid bulk and liquid droplets and two momentum equations and one mixture energy equation. The momentum equations consist of one for the liquid film at the wall and another combined one for gas and liquid droplets. Having six equations and several unknowns, closure laws are used to solve the conservation equations. The closure laws consist of relation for wall friction, liquid holdup, or interface friction. The closure laws are flow regime dependent and are selected based on flow regime considerations. Depending on the information fed, OLGA choose flow pattern from the two basic flow regimes: The separated flow regime which consist of stratified and annular flow and the mixed flow regime which consist of slug and bubble flow. Minimum slip concept is used to account for transition from one flow regime to another.

5.2 Comparison of experimental and simulation results

Using the same experimental flow data, the flow simulator was used to determine the pressure variations, liquid and gas holdup and flow rate variation in the vertical pipe and results are

compared against experimental results. Different gas mass flow rates were used and the point of instability in flow was established. The simulation setup consists of five straight pipes with different lengths connected to each other and the geometrical profile of the five pipes of one of the cases considered (for 213mm below vertical pipe outlet) is shown in the Figure 56 below:

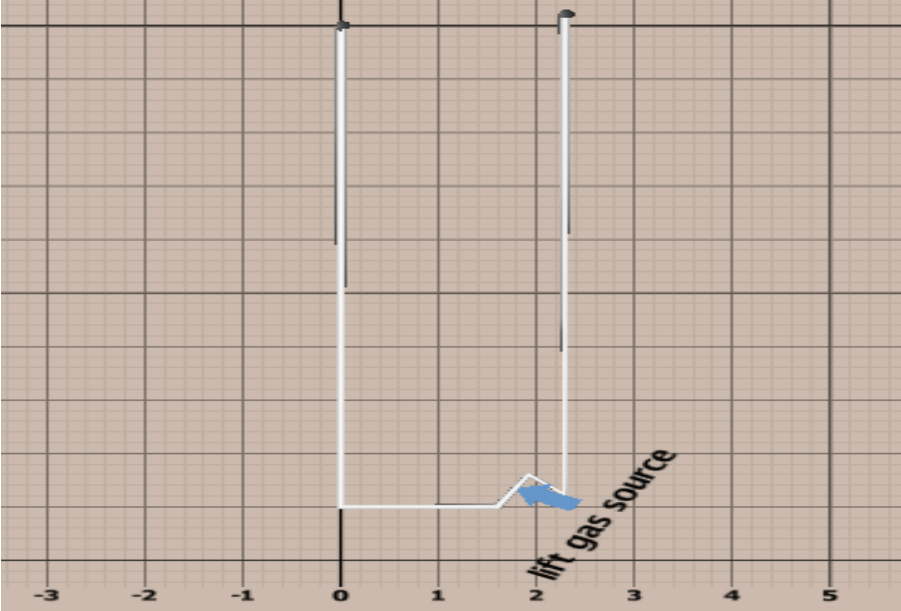


Figure 56: Simulation geometry at an angle of 45 degrees, inlet pressure of 1.882bar

These geometrical profiles are used as input data for space discretization (grid generation). The length of the horizontal pipe is cut into short pipe of lengths; (1.6m, 0.68m, 0.52m) so as to fully represent the pipe. And a property file (PVT) (shown in appendix C-2) was generated for the air- water system and is used by OLGA for the numerical simulation.

Information about the simulation geometry for which the variables have been chosen for presentation is given in Table 6 below

Table 6: Pipe Geometry.

Pipe	Cells	Diameter [m]	Roughness [m]	Inclination Θ [o]
1	250	0.032	2.00E-06	-90
2	50	0.032	2.00E-06	90
3	50	0.032	2.00E-06	62
4	25	0.032	2.00E-06	-45
5	250	0.032	2.00E-06	90

Table 6 above shows the geometry of the experimental Case 1: Overflow tank height at 213mm below vertical pipe outlet (inlet pressure of 1.882bar), inclination angle of -45 degrees and varying gas flow rates from 5.1042E-06kg/s to the flow rate at which stability is attained.. Pressure variation at the bottom, holdup variation, and liquid flow rate graphs at the vertical pipe base under varying inclination angles and gas injection rate at inlet pressure rate of 1.882bar are shown in Figure 29 - Figure 34. Attached in Appendix C-1 is an example of input file used during simulation.

5.2.1 Case 1: Inclination angle of 25 degrees

A. Gas injection rate = 5.1042E-06kg/s

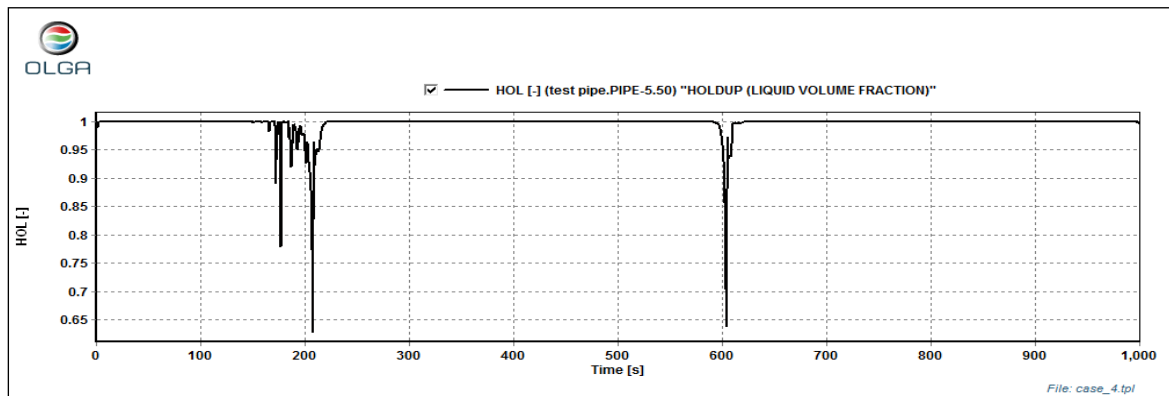


Figure 57: Holdup variation at 1.824m above riser inlet

Holdup variation was not captured during the experiment as impedance probe was not calibrated. Figure 57 shows variation of holdup at gas injection rate of 5.1042E-06 kg/s and inclination angle of 25 degrees.

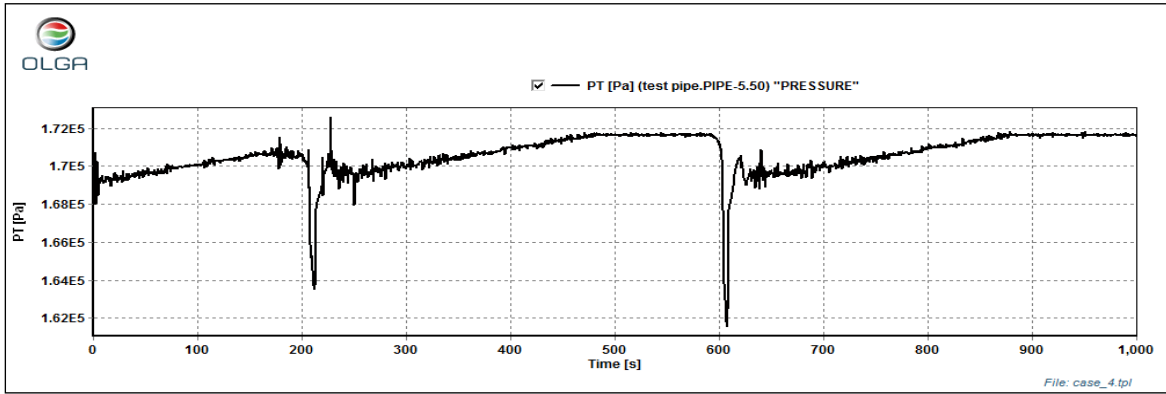


Figure 58: Pressure oscillation at 1.824m above riser

Compared to experimental result shown in Figure 11, Figure 58 pressure oscillation reading was taken at 1.824m above riser inlet compared to 0.87m as is the case of experimental result. Discrepancies in the pressure results can be attributed to simplicity of flow geometry during simulation. The period of one oscillation cycle in Figure 11 is 120s compared to Figure 58 that is approximately 200 – 400s.

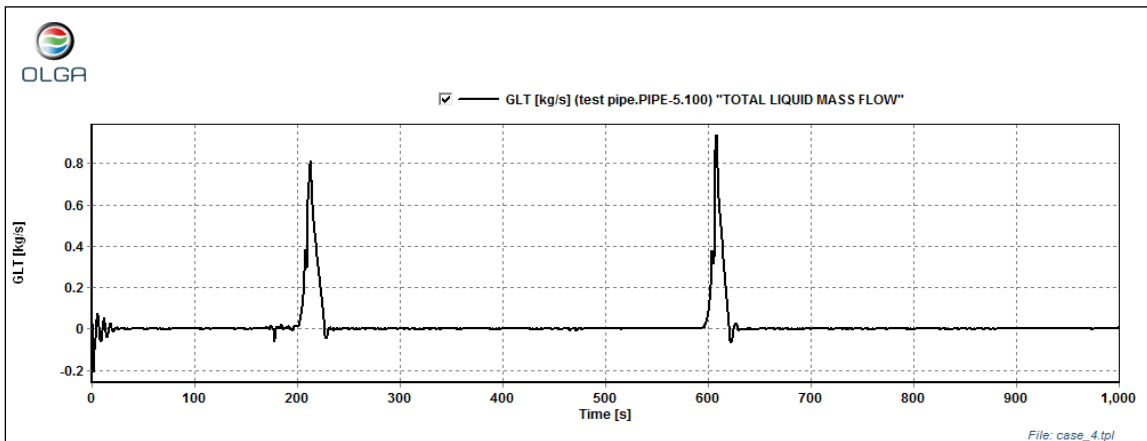


Figure 59: Liquid flow rate variation in riser

Figure 59 was plotted to determine the mass flow rate of liquid at inlet pressure of 1.882bar as this was not established during laboratory experiment due to incompatibility of the flow meter to flow range used. The mass flow rate from the graph is approximately 0.8-0.9 kg/s

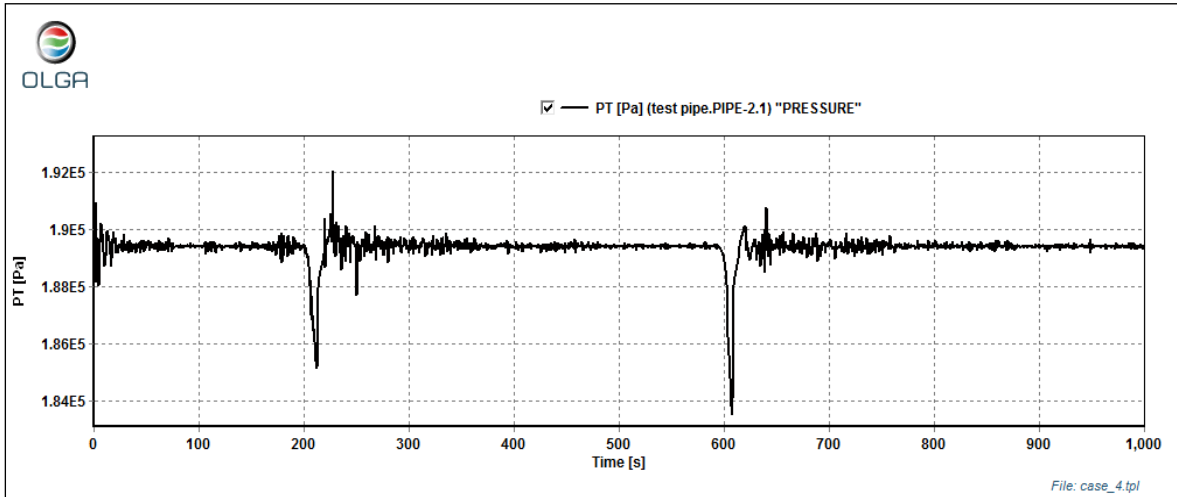


Figure 60: Inlet pressure variation due to expansion driven flow instability.

The inlet pressure rate under experimental condition was 1.882bar and this is approximately the same as the simulation pressure maximum as shown in Figure 60.

B. Gas injection rate = 5.1042E-005 kg/s

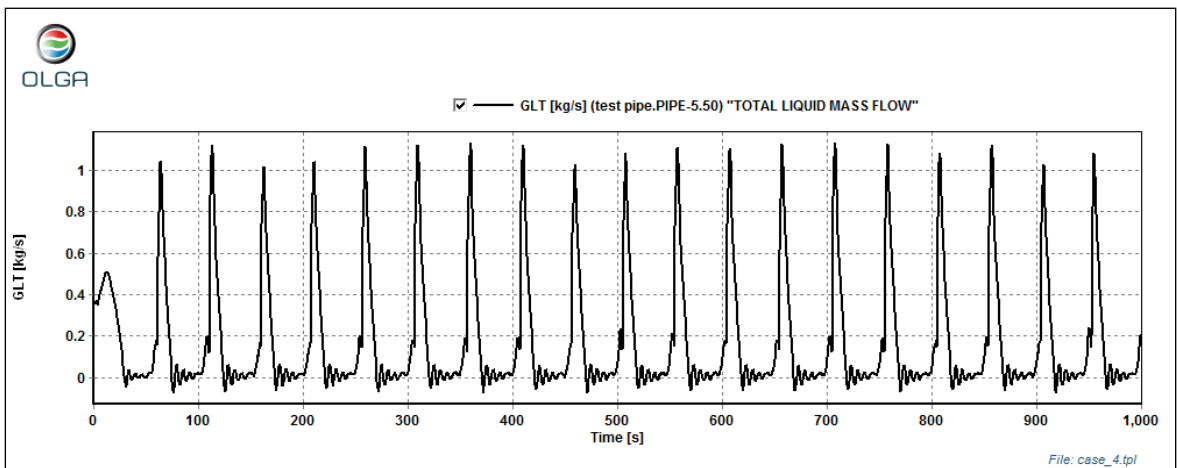


Figure 61: Liquid flow rate at 1.824m above riser inlet

Stability was attained at gas injection rate 5.1042 E-005 kg/s in Figure 11. Flow instability was observed to be the case at this flow rate under simulation condition. This discrepancy can be attributed to simplification and assumption made during simulation.

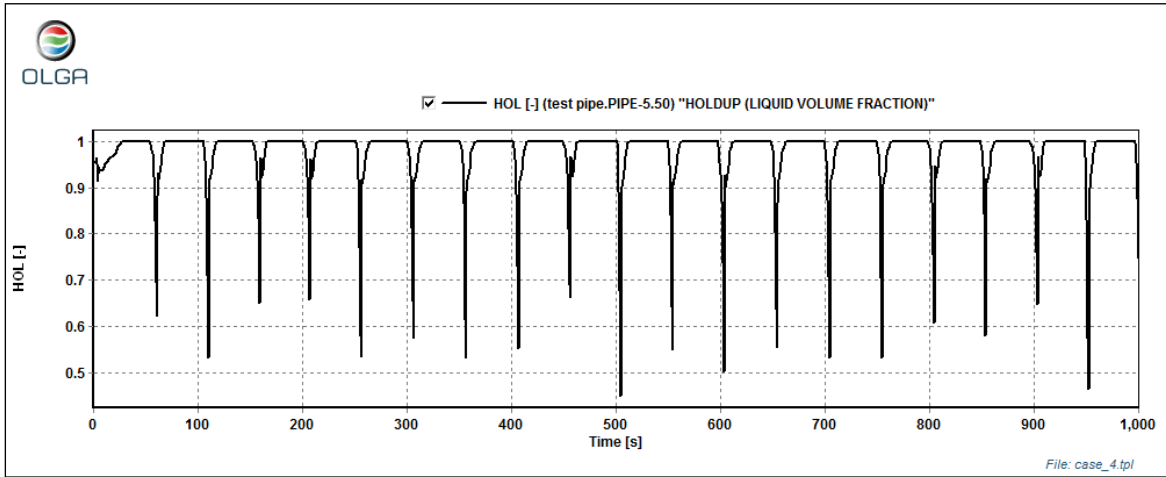


Figure 62: Holdup at 1.824m above riser inlet

Holdup variation was observed in Figure 62 at gas injection rate of $5.1042\text{E-}05$ kg/s.

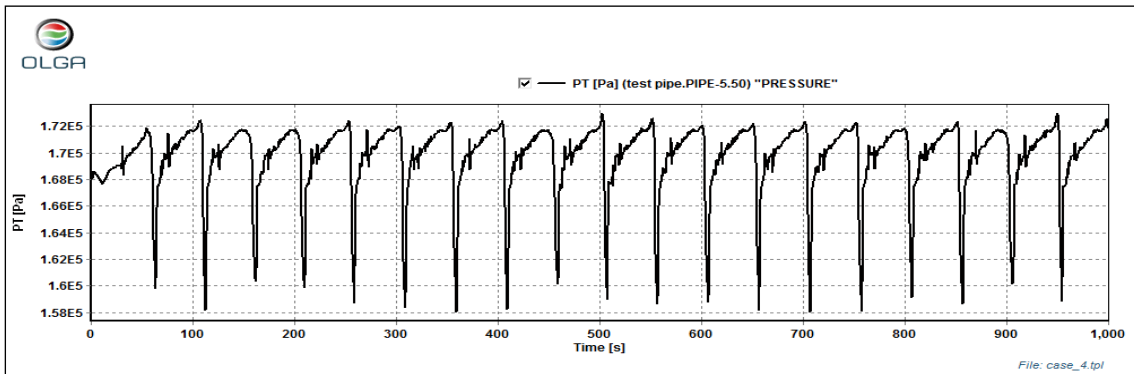


Figure 63: Pressure variation at 1.824m above riser inlet

Pressure was observed to be stable in the riser in Figure 13 at gas injection rate of $5.1042\text{E-}05$ kg/s under experimental variation. This is in contrast with pressure result obtained in Figure 63 at same gas injection rate.

5.2.2 Case 2: Inclination angle of 33 degrees

A. Gas injection rate = $5.1042\text{E-}06$ kg/s

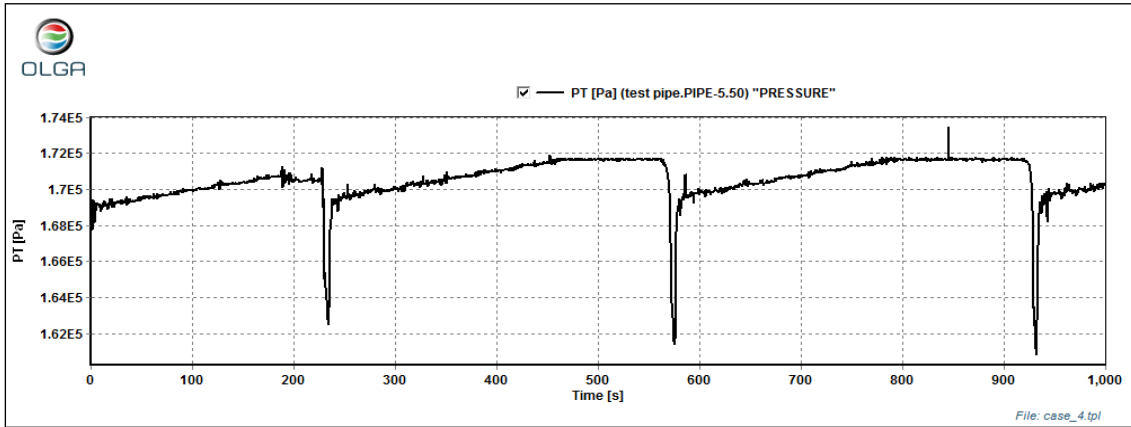


Figure 64: Pressure variation at 1.824m above riser inlet

Slight variation of about 15% was observed between experimental bottom pressure plot in Figure 14 and simulation plot in Figure 64. The period of one oscillatory cycle in Figure 64 is approximately 230-240s at low gas injection of $5.1042\text{E-}06$ kg/s.

B. Gas injection rate = $5.7167\text{e-}005$ kg/s

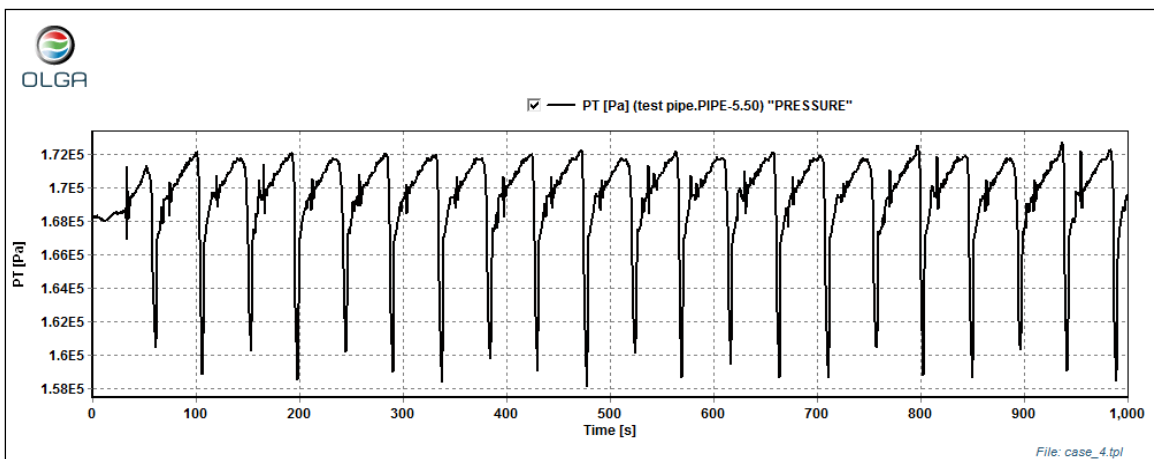


Figure 65: Pressure variation at 1.824m above riser inlet

Stability was attained approximately 400s after gas injection at $5.7167\text{E-}05$ kg/s in Figure 16. The flow behaviour in Figure 65 depicts the initial behaviour before 400s but did not stabilize afterwards.

5.2.3 Case 3: Inclination angle of 45 degrees

A. Gas injection rate = $5.1042\text{E-}06$ kg/s

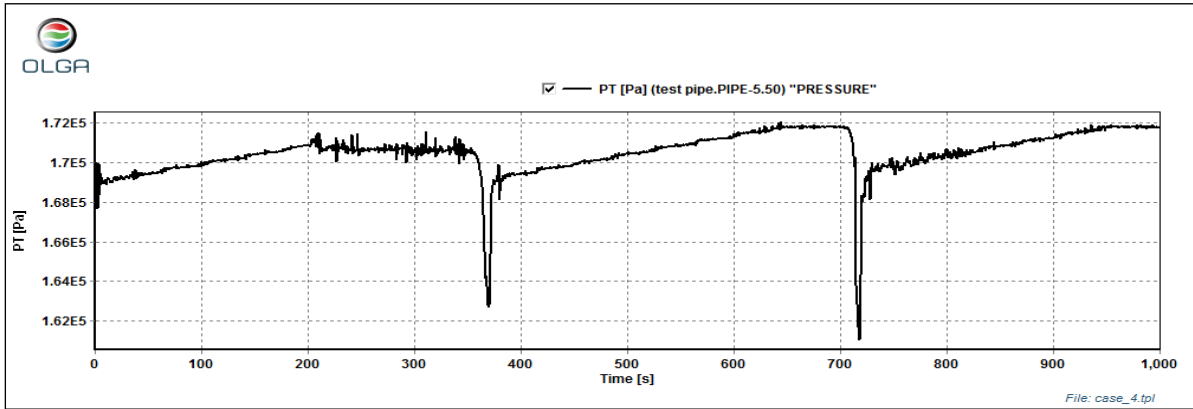


Figure 66: Pressure variation at 1.824m above riser inlet

Pressure variation in Figure 17 and Figure 66 are almost similar with slight variation of approximately 12%. The period of oscillation at gas rate of $5.1042E-06$ kg/s is approximately 370s.

B. Gas injection rate = $8.1667E-05$ kg/s

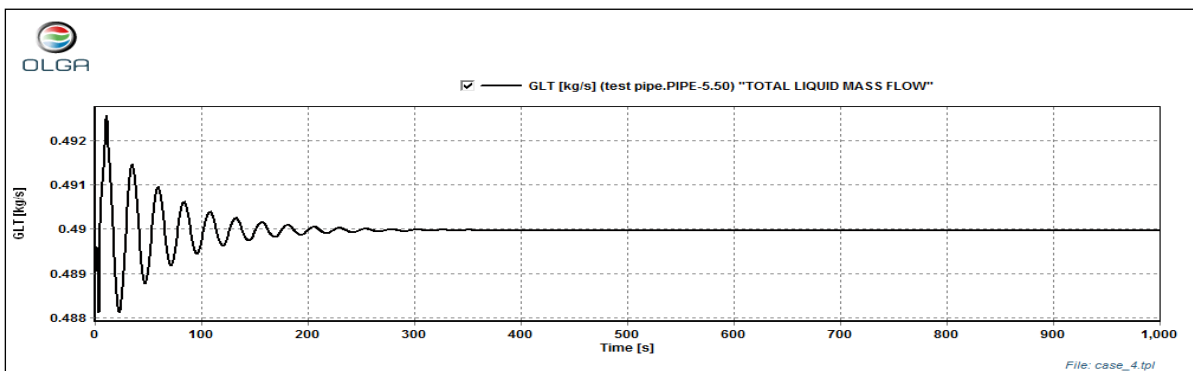


Figure 67: Liquid flow rate behavior at riser base

Stability was attained at $8.1667E-05$ kg/s as shown in Figure 67. And the liquid flow rate at stability was approximately 0.49 kg/s.

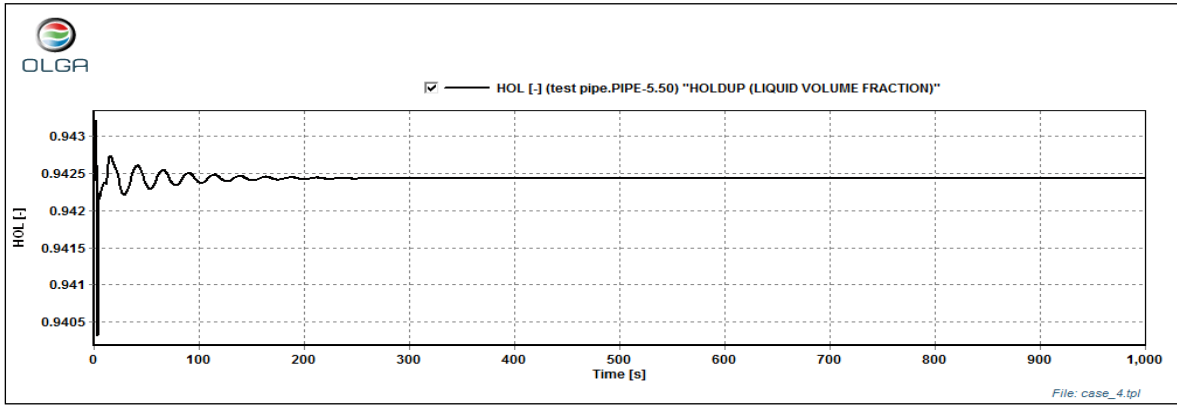


Figure 68: Holdup at the riser base

Figure 68 showed stability at gas injection rate of $8.1667\text{E-}05$ kg/s and holdup at the riser base was approximately 0.94.

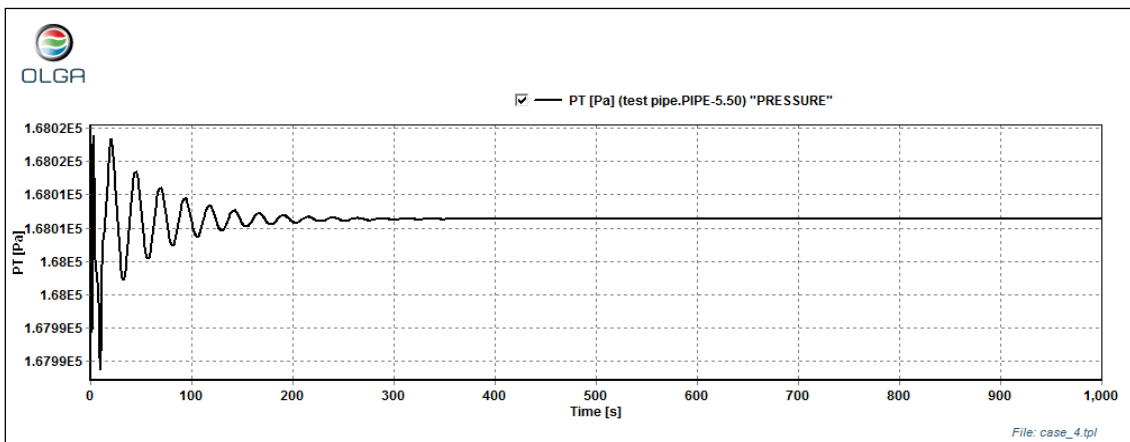


Figure 69: Pressure variation at 1.824m above riser inlet

Pressure stability at the riser base in Figure 69 is similar to that obtained in Figure 19 at gas injection rate of $8.1667\text{E-}05$ kg/s. Pressure stability point in Figure 69 is lower at approximately 1.68bar compared to 1.77bar (5% variation) obtained in Figure 19. This may be attributed to different measurement point used in both cases.

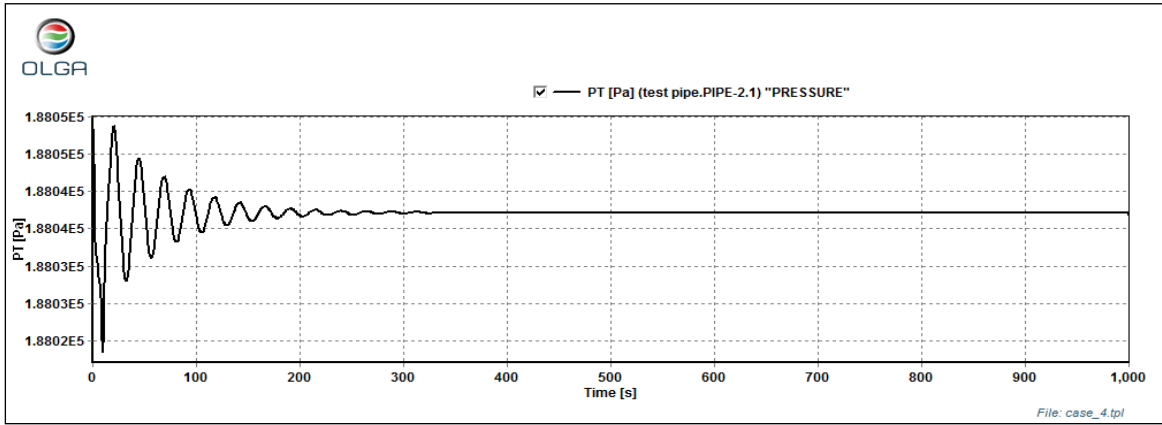


Figure 70: Inlet pressure of the flowing fluid

Simulation result shown in Figure 70 at stability attest to the fact that the inlet pressure rate of 1.882bar used under experimental condition was also observed during simulation at the given geometry and flow conditions.

Chapter 6

Conclusions

A flow loop was setup at the Multiphase Laboratory at NTNU to demonstrate expansion driven flow instabilities in long wells and risers. The loop was used to investigate expansion driven flow instability in natural and artificial lift systems in which gas is trapped upstream of a well or riser inlet. Experiments were conducted by varying system geometry, inlet pressures and gas flow rates. Expansion driven flow instability was observed to exhibit a cyclic behaviour as follows:

1. Gas accumulates in even moderate pipe bends - upstream the riser base.
2. The gas builds up until it starts to penetrate the riser.
3. Gas entry into the riser decreases the fluid density which in turn decreases the hydrostatic head on the riser base. A decreasing pressure in the riser gives higher flow rate. The liquid column in the riser is aerated and flows out of the riser at an increased rate.
4. The increasing flow rate due to gas expansion along the riser carries the accumulated gas across the bend.
5. A new cycle is initiated after the gas has been released.

Three cases at varying inlet pressure, geometry and gas flow rate were studied. Inclination angle of the bend was observed to have the strongest effect on flow behaviour leading to expansion driven flow at low inlet pressures and gas rates. Instabilities were observed at all inclination angles considered, regardless of increase in inlet pressure or gas flow rate. Highest effect of inclination angle was observed at -45 degrees angle of inclination and inlet flow pressure of 1.882bar and at maximum flowrate of gas required to attain stability i.e. $8.1667E-0.5$ kg/s.

Snapshots from video were taken and have been explained. Video recordings show the accumulation and expansion of gas in the downward inclined pipe towards the riser inlet. Also, blockage of upstream fluid was captured as gas accumulates and expands..

Stability test for a horizontally-laid flow line at low gas injection rates was carried out and stable flow was observed even at low gas injection rate.

Inlet pressure test case of 1.822bar (Case 1) was modelled and simulated in OLGA. Results of the simulation model were compared against results of the experimental investigation. Discrepancies between the two sets of results were observed in some cases and captured. These discrepancies could be attributed to simplifications and assumptions made during the simulation case modelling.

The results from experimental and numerical simulation results demonstrated that expansion driven flow instabilities can occur in laboratory/pilot setups and so can possibly occur in deep water natural-lift as well as gas-lifted wells and risers, under certain flow conditions and flowline inclinations.

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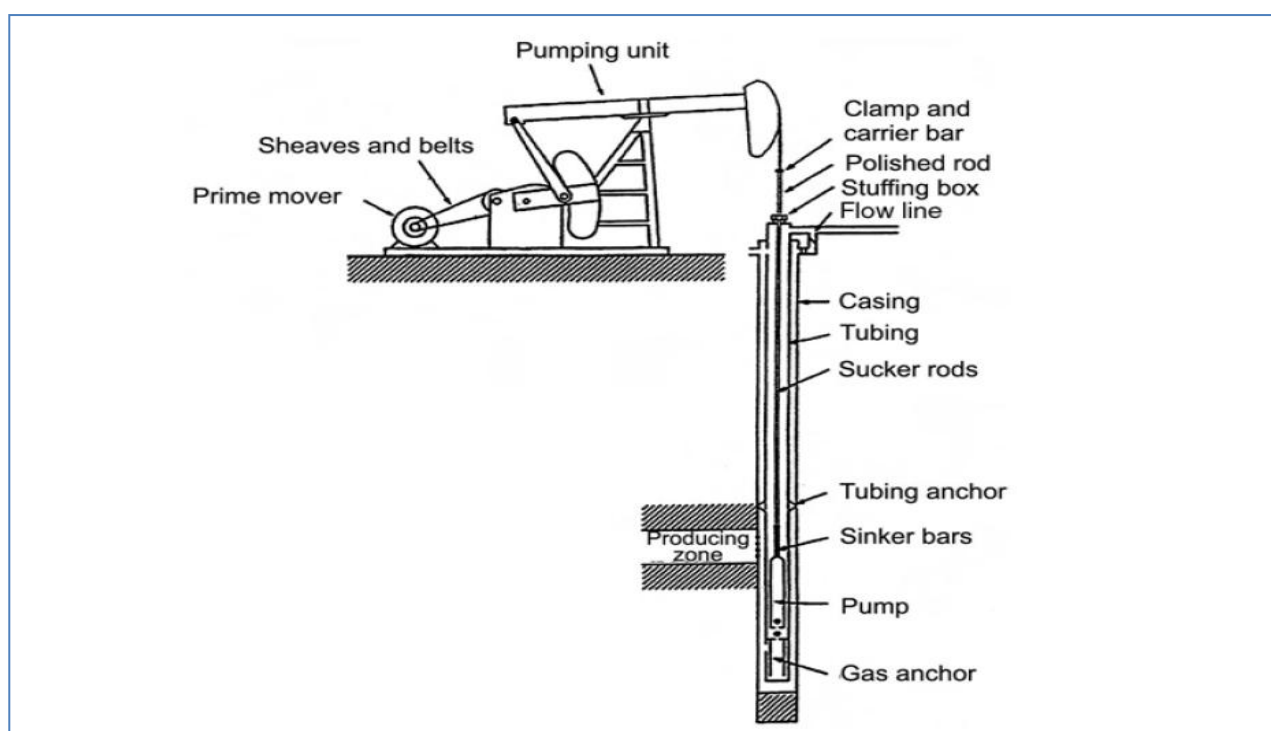
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Appendix A

A-1. Sucker-Rod Pumping

Sucker-rod pumping systems are the oldest type of artificial lift used in oil wells. It is mostly used in onshore operations for lifting small-moderate volumes from shallow–intermediate well depths. It is possible to lift up to 1,000 bbl/day from approximately 7,000 ft and 200 bbl/day from approximately 14000 feet (Lewis, 2007). In general, for offshore and higher-rate wells, use of other methods of artificial lift is recommended.

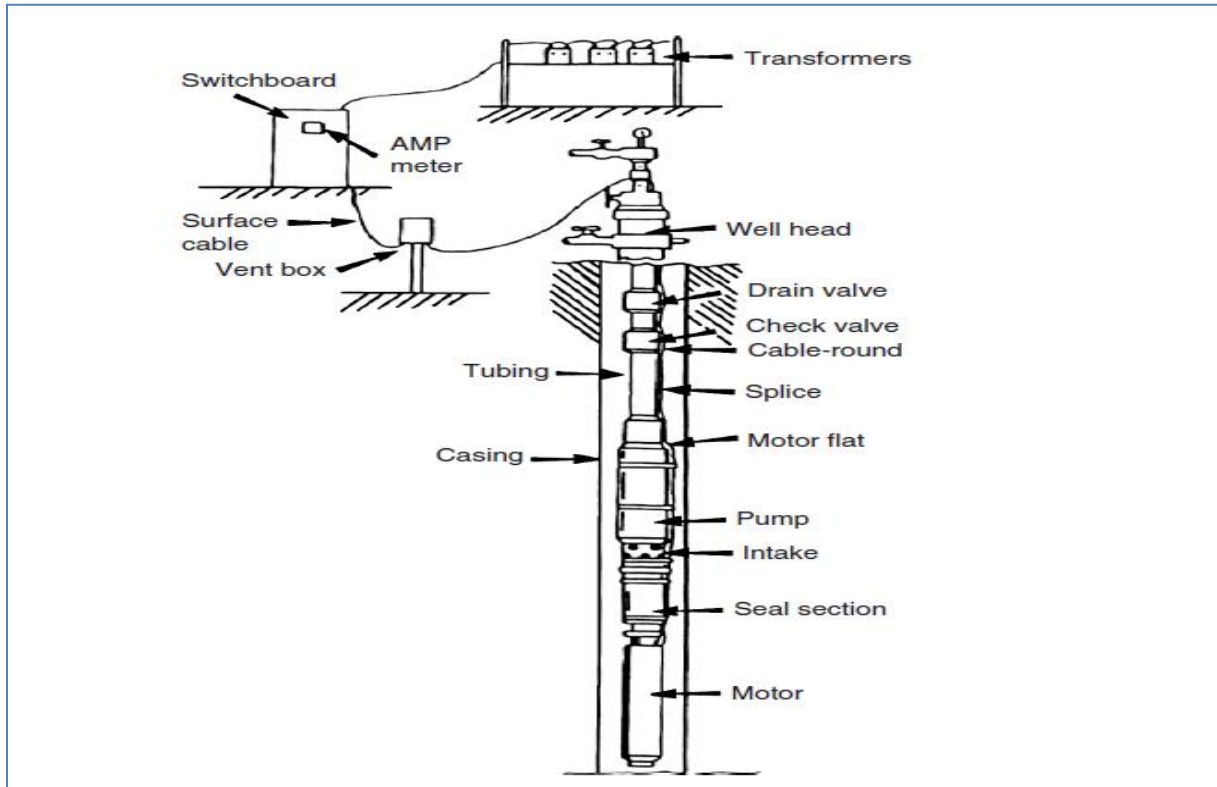


Schematic of a beam (or rod)-pumping system. (Courtesy of Harbison-Fisher)

A-2. Electrical Submersible Pumping

Electrical Submersible Pumping (ESP) has been used to lift fluids from highly deviated wells. The system consists of subsurface (downhole) and surface components. ESP pumps are made of dynamic or centrifugal pumps connected directly to an electric motor. ESP is considered a high-volume pump that provides for increased production volumes and water cuts. It is also suitable for offshore as well as simultaneous operations whereby lifted wells can be placed on production while drilling and working over wells in immediate vicinity. However ESPs are limited by inefficiency in low volume production. Below 400 bbl/day power efficiency drops sharply and at production rates lower than 150 bbl/day ESPs are inapplicable. Also, as in

rod/beam pumping systems, ESPs have low tolerance for gas and solids production, thus allowing only minimal percentages of sand production. Consequently the system's failure rate can be high, leading to production losses and need for expensive pulling operations to correct downhole failures, especially in offshore environment. Other limitations of ESP include high voltage demand.



Schematic of typical ESP system [Courtesy of Schlumberger (REDA)]

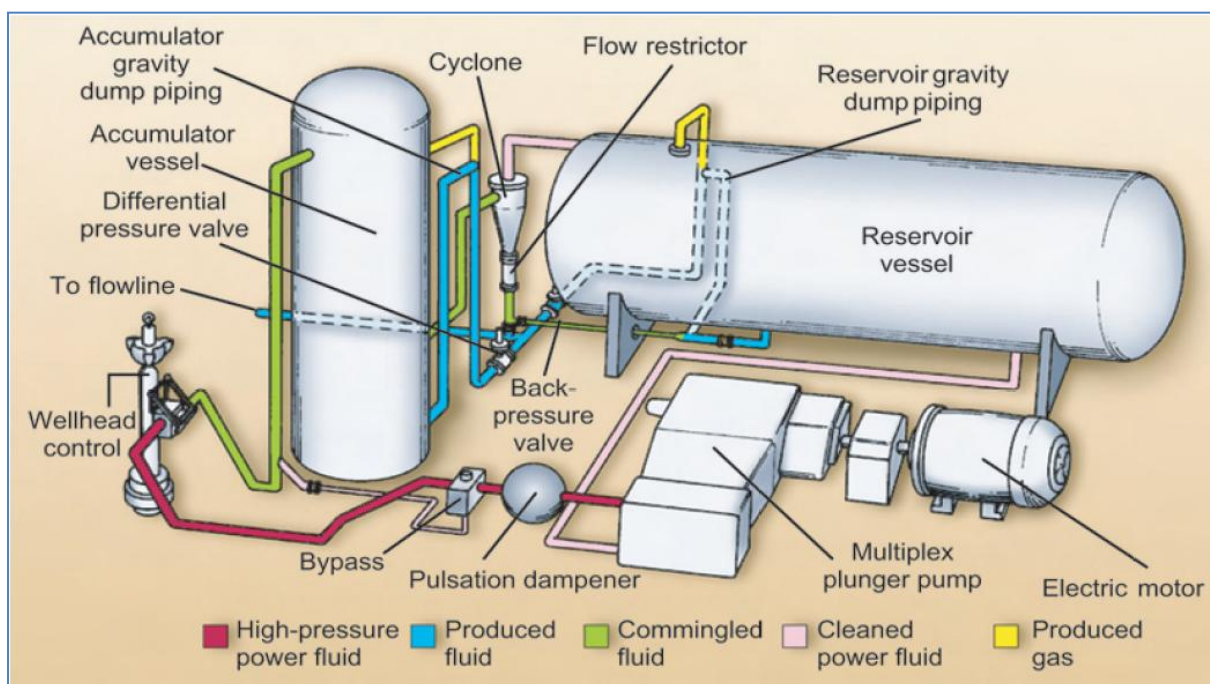
The limitations of ESP such as sand and gas problems has led to new products/methods such as progressive cavity pumping (PCP) and electrical submersible progressive cavity pump (ESPCP). The former can be installed in both deviated and horizontal wells and can be used for lifting highly viscous oils at varying flow rates without solids and free gas production posing serious problems. Similarly, use of ESPCPs eliminates the need for rotating rods and the associated problems with rods rotating in a deviated well. The disadvantages of PCP & ESPCP include:

- The stator material has an upper temperature limit and may be subject to H₂S and other chemical deterioration.
- Frequent stops and starts of the PCP pumps often can cause several operating problems.

- The gearbox in an ESPCP is another source of failure if wellbore fluids or solids leak inside it or if excessive wear occurs.

A-3. Hydraulic Pumping

Hydraulic pumping systems transmit power downhole by means of a pressurised power fluid (oil or water). Hydraulic transmission of the power-fluid can be accomplished by means of either a reciprocating positive displacement pump or a jet pump. A typical positive displacement hydraulic system consists of engine and pump pistons connected by a short shaft. The high pressure power fluid is injected downhole from the surface into a tubing string and is returned to the surface, either commingled with the produced fluid in the production string (open system) or separately through another tubing (closed system). The operating pressures in hydraulic pumping systems usually range from 2,000 to 4,000 psi (Lewis, 2007). Advantages of hydraulic pumping systems include ability to lift large volumes of production from very deep wells, up to 17,000 ft and deeper (Lewis, 2007). Hydraulic system is normally employed where other types of artificial lift fail as a result of well conditions or geometry.



Schematic Hydraulic pump system

Other benefits of hydraulic lift systems include:

- Use of jet pumps in sandy corrosive wells
- Use of reciprocating pumps in deep wells with low bottomhole pressure

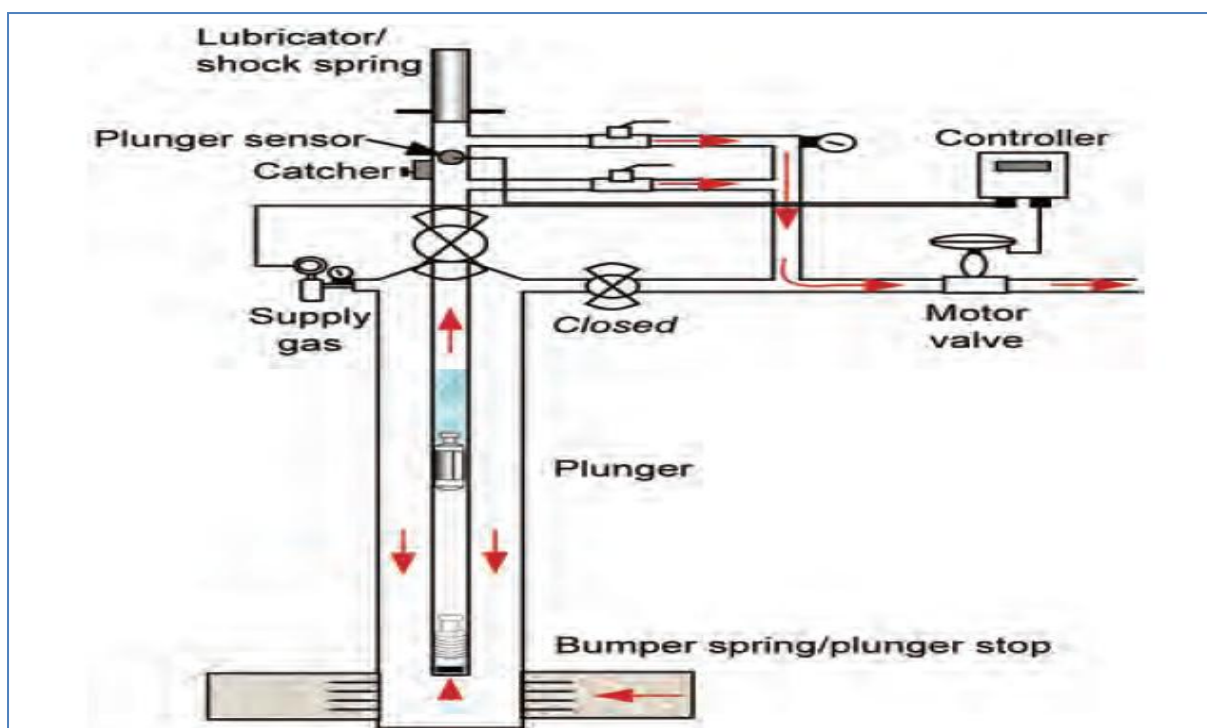
- Wells with rapidly changing producing volumes and highly viscous wells

Disadvantages of hydraulic pump lifts include the following:

- Shorter pump life between repairs, compared to sucker rod and ESPs
- Jet pumps typically have lower efficiency and higher energy costs
- Positive-displacement pumps generally require more maintenance than jet pumps and other types of artificial lift because pump speed must not become excessive.
- Power-fluid-cleaning systems require frequent checking to keep them operating at their optimum effectiveness
- Well testing is more difficult.

A-4. Plunger Lift

Plunger lift uses a free piston that moves up and down in the well's tubing string using the well's gas pressure as the energy source. Therefore it is especially applicable to wells with high gas-liquid (GLR) ratio. Like all artificial lift methods the plunger works to move liquids from the wellbore so that the well can be produced at the lowest bottomhole pressures. It provides a piston-like interface between liquids and gas in the wellbore and prevents liquid fallback. A plunger cycle starts when it drops from the surface to the bottom of the well. At the same time, the well builds gas pressure that will provide energy to lift both the plunger and the liquid slug to the surface. Therefore the plunger lift relies much more on the well's ability to store enough gas pressure to lift the plunger to the surface, and less on critical flow rates.

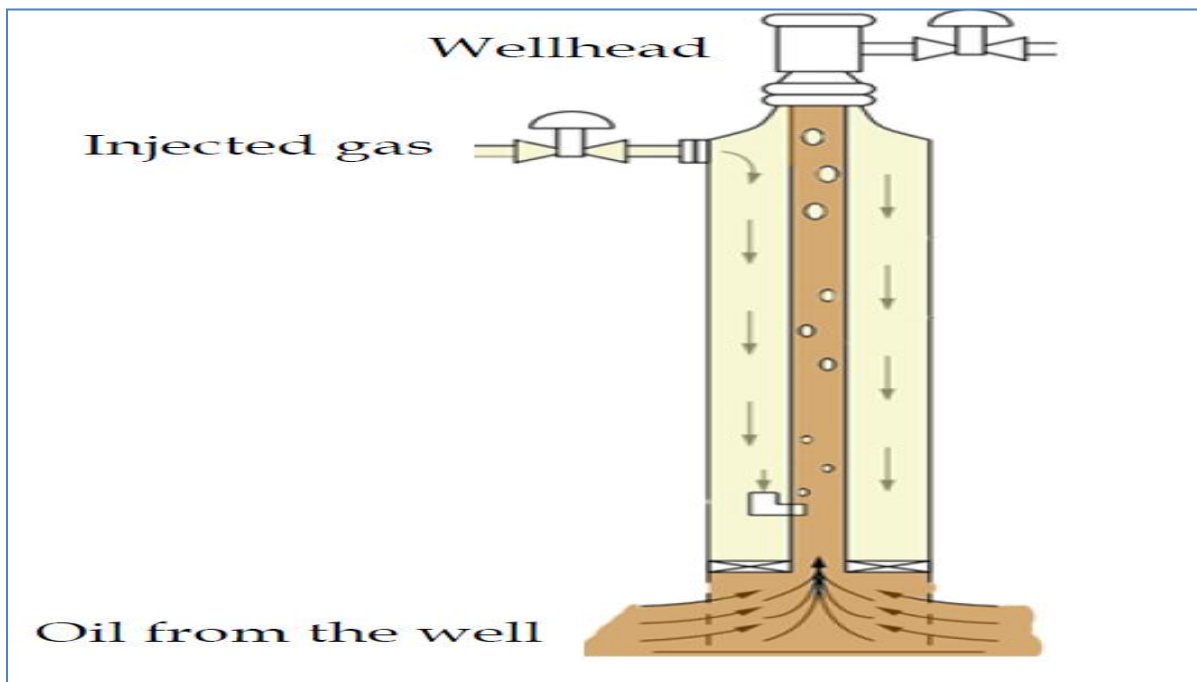


Schematic of Plunger-lift systems

Plungers have been installed on wells for the sole purpose of preventing paraffin or hydrate buildup, thereby reducing paraffin scraping or methanol injection (Beauregard & Ferguson, 1982) (Ferguson & Beauregard, 1983). The most common plunger-lift applications are used in liquid removal in gas wells, but plungers also are used successfully for oil recovery in high-GLR oil wells, mainly in combination with intermittent gas lift operations (Brown, 1980) (Hall & Bell, 2001).

A-5. Gas Lift

The basic principle of gas lift is that gas injected into a column of liquid reduces the effective density of the fluid, thereby reducing the pressure differential between point of gas injection and the surface of the fluid. Gas lift has long been applied as a method of artificial lift in oil wells, where it is used to increase production by injecting high pressure gas into the lower section of the well tubing through the annulus and a downhole valve. The potential of a gas lift well is controlled by the gas injection rate or the fluids gas-liquid-ratio (GLR) (Lyons & Ghalambor, 2007) (Lewis, 2007); whereby a maximum-efficiency gas lift system requires use of a minimum gas injection rate to lift a given amount of liquid. Consequently a well potential (expected liquid production rate) from gas lift, at a given injection rate, is determined by the GLR of the reservoir fluids.



Oil producing well with gas-lift (from PipeFlow 2, Multiphase flow assurance)

Gas lift operation has been classified into continuous and intermittent. Continuous gas lift operation involves steady-state flow of the aerated fluid in a column, whereas intermittent gas lift is characterized by periodic displacement of liquid slugs in the column, as a result of injecting high pressure gas at the base of the column. Continuous gas lift has been successfully used in wells with a high production index (PI) and a reasonably high bottomhole pressure (BHP), while intermittent gas lift method is typically used on wells with high PI and low BHP or low PI and low BHP.

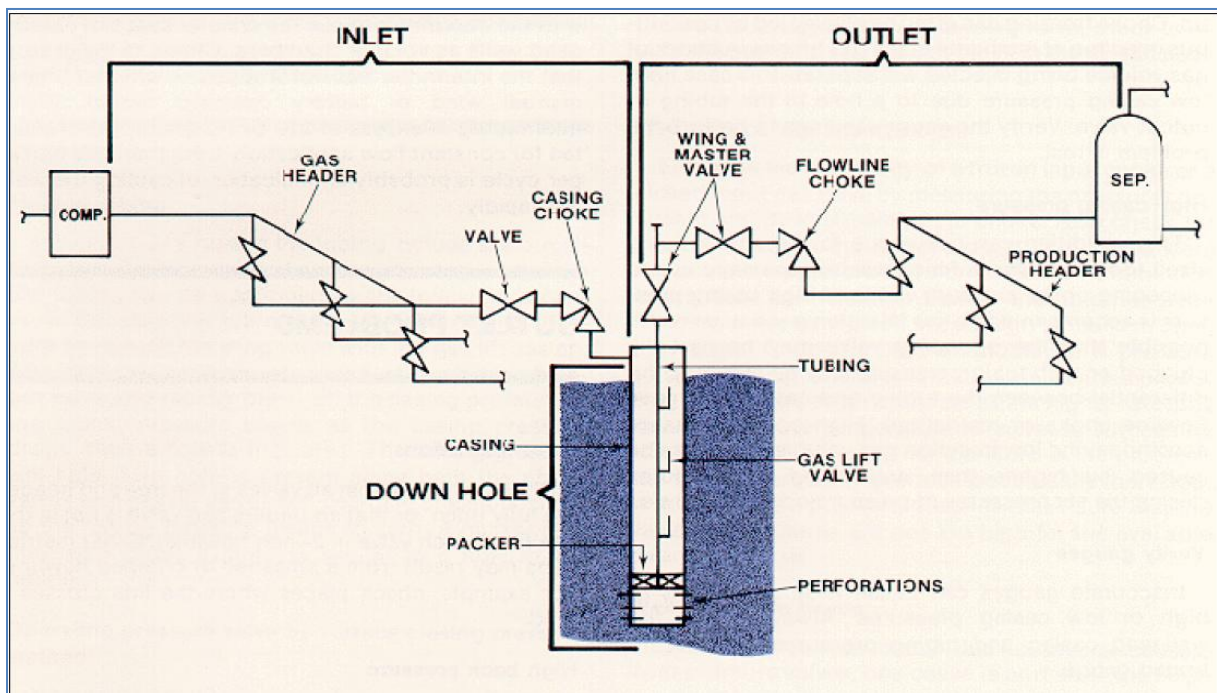


Diagram of a typical gas lift system. (from The Lease Pumper's Handbook)

Advantages of well gas-lift systems include the following:

- Best artificial lift method for handling sand/solids in well fluids.
- Deviated or directional wells can be lifted easily with gas lift.
- Gas lift permits the concurrent use of wireline equipment, allowing routine repairs through the tubing.
- Whereas high GLRs hinder other artificial lift systems, it is very helpful for gas-lift
- Produced gas means less injection gas is required; whereas, in all other pumping methods, pumped gas reduces volumetric pumping efficiency drastically.
- A wide range of volumes and lift depths can be achieved with essentially the same well equipment.
- A central gas-lift system easily can be used to service many wells or operate an entire field, lowering life-cycle costs.

Disadvantages of gas lift include:

- Gas lift can often result in large capital investments and high energy operating costs.
- The compressor takes up space and weight when used on offshore platforms.
- Adequate gas supply is needed throughout life of project. If the field runs out of gas, or if gas becomes too expensive, it may be necessary to switch to another artificial lift method.
- There is increased difficulty when lifting low gravity (less than 15°API) crude because of greater friction, gas fingering, and liquid fallback.
- Potential gas-lift operational problems that must be resolved include freezing and hydrate problems in injection gas lines, corrosive injection gas, and severe paraffin problems. Other problems that must be resolved are changing well conditions, especially declines in BHP and PI.

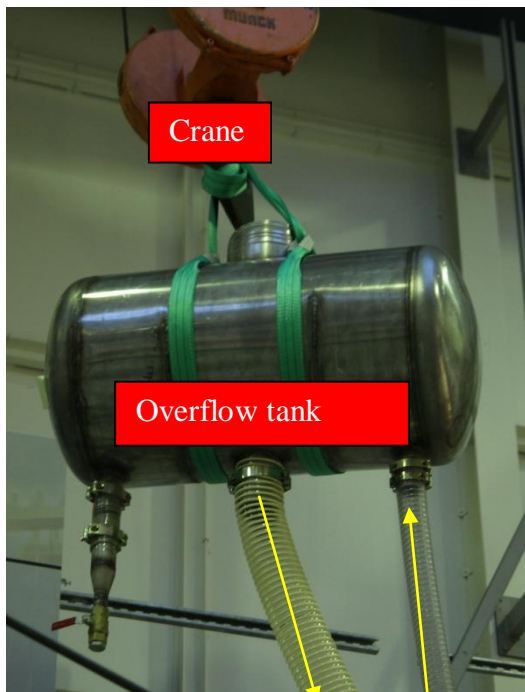
Gas lift has found wide application in deepwater operations as with increasing depth, static head will restrict flow and reservoir will require artificial lift to enhance production. More so, application of gas lift to increase field production has also found application in production systems other than the well. Recent research and studies have focused on use of gas lift in pipeline-riser systems. These includes injecting gas at the riser base or slightly upstream the riser base along the flowline. Riser base gas lift is generally perceived as less effective than compared to well gas lift and pumping systems (Bass, 2006). Also cooling effect due to gas expansion in the riser limits its applicability at greater water depths due to the risk of hydrate formation. However, increase in gas-liquid-ratio due to the riser base gas lift allows production systems to produce at additional amount at a given reservoir pressure. Therefore, the effectiveness of riser base gas lift increases as reservoir pressure declines.

The application of gas-lift has proved to be an efficient method of extending oil field life but not without some drawbacks particularly flow instability. Vertical flow of liquid-gas can be unstable in gravity dominated gas lifted wells. Continuous gas-lift is used in most high flow producers, but unstable flow conditions may occur in these systems as a result of small perturbations that can degenerate into huge oscillations, leading to intermittent flow (Alhanati, Schmidt, & Lagerlef, 1993) (Hu & Golan, 2003). Major flow instabilities associated with gas lift systems are casing heading and density wave instabilities. Expansion driven flow instability can be identified as a type of instability also associated with gas lifted vertical pipes which will be investigated in this project as to conditions at which it occurs.

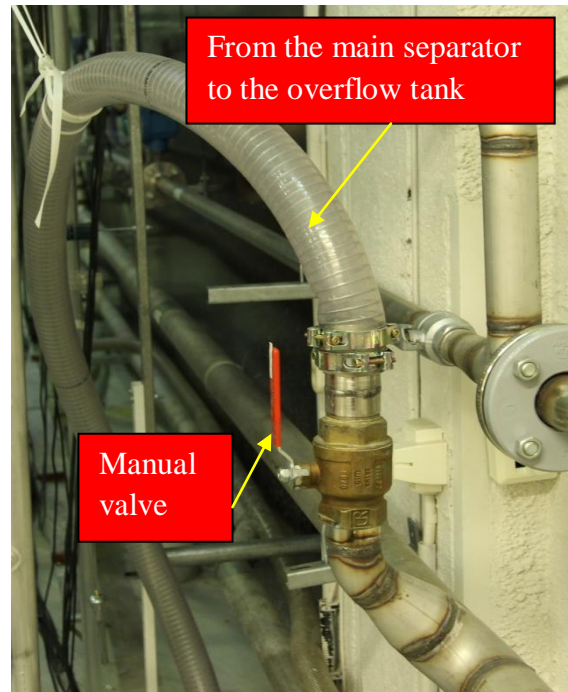
Appendix B

B-1. Experimental facilities

Figure [A] shows the overflow tank hanging on a crane. Water flows from the main separator [B] to the overflow tank and back to main separator through the buffer tank shown in [C] and [D]. This is done to achieve constant pressure flow to the riser. Also fluid from the riser outlet flows back to the main separator through the buffer tank from where it is circulated back.



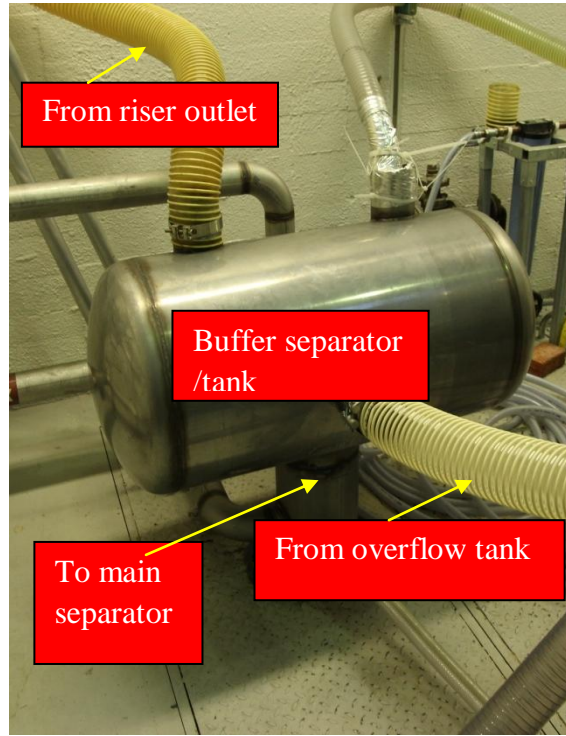
A



B



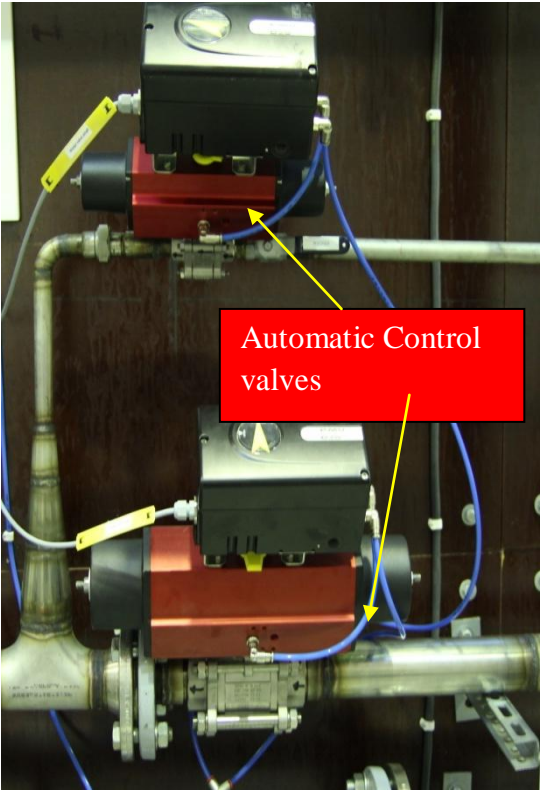
C



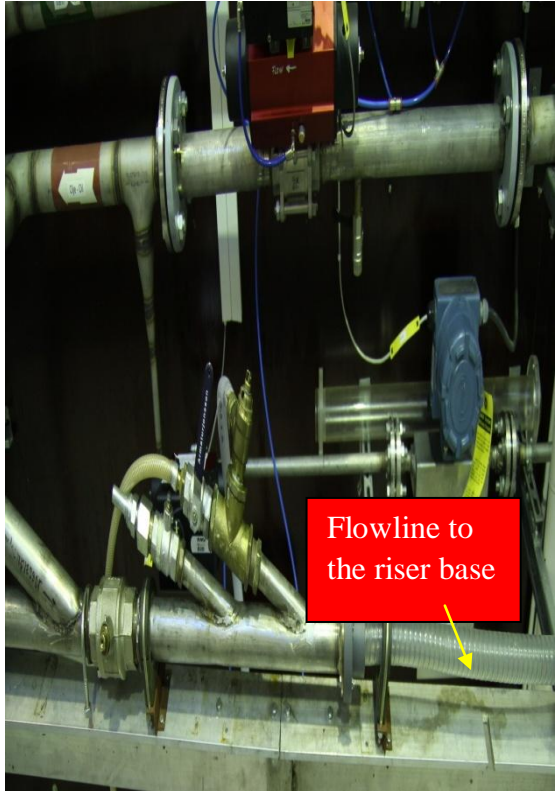
D

B-2. Main water tank, air tank and other facilities

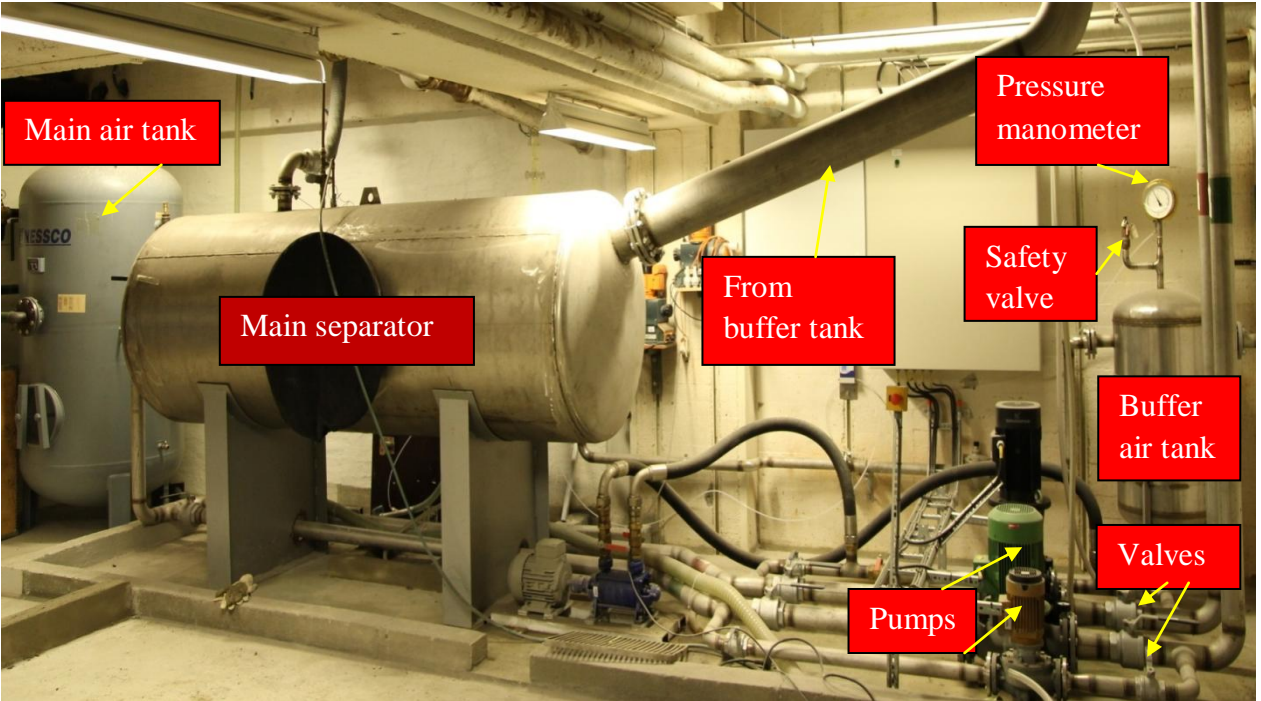
The automatic control [A] valve mounted along waterline used to control flow was fully opened to avoid the effect of friction. Figure [B] shows waterline to the riser inlet. Water is pumped from the main tank using a centrifugal pump to the riser and overflow tank in a closed loop [C]. Air from air buffer tank [C] is injected upstream the riser inlet.



A



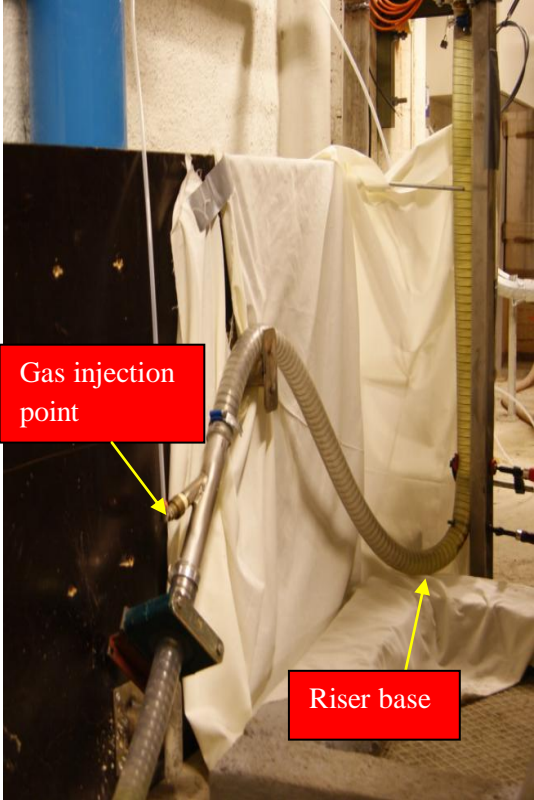
B



C

B-3. Riser base and riser experimental facility

Gas is injected upstream the riser base [A)] and choke valve at the top of the riser [43] was fully opened.



A



B

Appendix C

C- 1. Simulation input file

INPUT FILE

```
!*****
```

```
*****
```

```
! CASE
```

```
!*****
```

```
*****
```

```
!
```

```
!*****
```

```
*****
```

```
! OPTIONS
```

```
!*****
```

```
*****
```

```
OPTIONS DEBUG=ON, SLUGVOID=AIR, TEMPERATURE=OFF,  
STEADYSTATE=NOTEMP
```

```
!
```

```
!*****
```

```
*****
```

```
! FILES
```

```
!*****
```

```
*****
```

```
FILES PVTFILE="./air_water.tab"
```

```
!
```

!*****

! INTEGRATION

!*****

INTEGRATION DTSTART=4 S, ENDTIME=1000 s, MAXDT=10 S, MINDT=0.00324 s,
STARTTIME=0 S

!

!*****

! GEOMETRY

!*****

GEOMETRY LABEL="case_4 tt", XSTART=0 M, YSTART=0 M, ZSTART=0 M

PIPE LABEL=PIPE-1, ROUGHNESS=2E-06 M, XEND=5.70667E-16 M, YEND=-9 M,
DIAMETER=0.032 M,\

NSEGMENT=250, LSEGMENT=(0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,
0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036, 0.036,\

NSEGMENT=50, LSEGMENT=(0.0135906, 0.0135906, 0.0135906, 0.0135906,
0.0135906, 0.0135906,\

0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906,
0.0135906,\

0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906,
0.0135906,\

0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906,
0.0135906,\

0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906,
0.0135906,\

0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906, 0.0135906,
0.0135906,\

0.0135906, 0.0135906, 0.0135906, 0.0135906) M

PIPE LABEL=PIPE-4, ROUGHNESS=2E-06 M, XEND=2.286 M, YEND=-8.767 M,
DIAMETER=0.032 M,\

NSEGMENT=25, LSEGMENT=(0.0207607, 0.0207607, 0.0207607, 0.0207607,
0.0207607, 0.0207607,\

0.0207607, 0.0207607, 0.0207607, 0.0207607, 0.0207607, 0.0207607, 0.0207607,
0.0207607,\

0.0207607, 0.0207607, 0.0207607, 0.0207607, 0.0207607, 0.0207607, 0.0207607,
0.0207607,\

0.0207607, 0.0207607, 0.0207607) M

PIPE LABEL=PIPE-5, ROUGHNESS=2E-06 M, XEND=2.286 M, YEND=0.213 M,
DIAMETER=0.032 M,\

NSEGMENT=250, LSEGMENT=(0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592, 0.03592,
0.03592,\

0.03592, 0.03592, 0.03592) M

!

!*****

! NODE

!*****

NODE LABEL=inlet, TYPE=TERMINAL

NODE LABEL=outlet, TYPE=TERMINAL

!

!*****

! BRANCH

!*****

BRANCH LABEL="test pipe", FROM=inlet, TO=outlet, GEOMETRY="case_4 tt",

FLUID="luft_vann"

!

!*****

! BOUNDARY

!*****

BOUNDARY NODE=inlet, TYPE=PRESSURE, TEMPERATURE=20 C,

PRESSURE=101300 Pa, GASFRACTION=0

BOUNDARY NODE=outlet, TYPE=PRESSURE, TEMPERATURE=20 C,

PRESSURE=101300 Pa

!

!*****

! INITIALCONDITIONS

!*****

INITIALCONDITIONS BRANCH="test pipe", INTEMPERATURE=20 C,
OUTTEMPERATURE=20 C

!

!*****

! CONTROLLER

!*****

!

!*****

! SOURCE

!*****

SOURCE LABEL="lift gas source", TIME=0 s, TEMPERATURE=20 C, BRANCH="test
pipe", PIPE=PIPE-3,\

SECTION=30, MASSFLOW=5.1042e-006 kg/s, GASFRACTION=1

!

!*****

! OUTPUT

!*****

OUTPUT DTOUT=0.1 s

!

```
!*****
*****

!   TREND

!*****
*****

TREND DTPLOT=0.5 s

TREND BRANCH="test pipe", VARIABLE=(AL, GLT, HOL, ID, PT), PIPE=PIPE-5,
SECTION=(50,\

    100, 150, 200)

TREND BRANCH="test pipe", VARIABLE=PT, PIPE=PIPE-2, SECTION=1

!

!*****
*****

!   PROFILE

!*****
*****

PROFILE DTPLOT=1 s

PROFILE BRANCH="test pipe", VARIABLE=HOL

ENDCASE
```

C-2. PVT file

```
PVTTABLE LABEL = "luft_vann",PHASE = TWO,\n\nCOMPONENTS = ("H2O","AIR"),\n\nMOLWEIGHT = (18.02,28,97) g/mol,\n\nSTDPRESSURE = 1.0 ATM,\n\nSTDTEMPERATURE = 293.16 K,\n\nGOR = -999 Sm3/Sm3,\n\nGLR = -999 Sm3/Sm3,\n\nSTDGASDENSITY = 1.204 kg/m3,\n\nSTDOILDENSITY = 998 kg/m3,\n\nMESHTYPE = STANDARD,\n\nPRESSURE = ( 1.00E+04 , 3.00E+05) Pa,\n\nTEMPERATURE = ( 18 , 22 ) C,\n\nCOLUMNS = ( PT , TM , ROG , ROHL , DROGDP , DROHLDP ,\nDROGDT , DROHLDT , RS , VISG , VISHL , CPG , CPHL , HG , HHL ,\nTCG , TCHL , SIGGHL , SEG , SEHL )\n\nPVTTABLE POINT = ( 1.00E+04 , 18 , 1.196759E-01 , 998 , 1.196759E-05 , 0.0 , -\n4.110315E-04 , 0.0 , 100 , 1.83E-05 , 1.00E-03 , 1.01E+03 , 4.18E+03 , 0 , 0 , 2.51E-02\n, 5.98E-01 , 7.29E-02 , 0 , 0 )\n\nPVTTABLE POINT = ( 1.00E+04 , 22 , 1.180541E-01 , 998 , 1.180541E-05 , 0.0 , -\n3.999664E-04 , 0.0 , 100 , 1.83E-05 , 1.00E-03 , 1.01E+03 , 4.18E+03 , 0 , 0 , 2.51E-02\n, 5.98E-01 , 7.29E-02 , 0 , 0 )\n\nPVTTABLE POINT = ( 3.00E+05 , 18 , 3.590278E+00 , 998 , 1.196759E-05 , 0.0 , -\n1.233094E-02 , 0.0 , 100 , 1.83E-05 , 1.00E-03 , 1.01E+03 , 4.18E+03 , 0 , 0 , 2.51E-02\n, 5.98E-01 , 7.29E-02 , 0 , 0 )
```

PVTTABLE POINT = (3.00E+05 , 22 , 3.541622E+00 , 998 , 1.180541E-05 , 0.0 , -
1.199899E-02 , 0.0 , 100 , 1.83E-05 , 1.00E-03 , 1.01E+03 , 4.18E+03 , 0 , 0 , 2.51E-02
, 5.98E-01 , 7.29E-02 , 0 , 0)

Appendix D

D-1. Video recording

Video recording of expansion driven flow instability at the following gas injection rate is attached below:

- Gas injection rate = $5.1042\text{E-}06$ kg/s
- Gas injection rate = $3.0625\text{E-}05$ kg/s
- Gas injection rate = $4.0833\text{E-}05$ kg/s