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EXPERIMENTAL INVESTIGATION OF GAS/LIQUID CYLINDRICAL CYCLONE SEPARATOR

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ABSTRACT

EXPERIMENTAL INVESTIGATION OF GAS/LIQUID CYLINDRICAL CYCLONE SEPARATOR

BY

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The Gas-Liquid Cylindrical Cyclone (GLCC) separators have proven themselves in laboratory and field, as a well alternative to the conventional gravity-based gas/liquid separator. This study presents investigation on effect of changes in physical aspects on GLCC performance. These changes are including increases in outlet length and reduction in gas body column length, inlet, body column, liquid and gas outlet diameter. Results show that reduction inlet diameter enhanced the GLCC performance but any reduction in diameter of body column and liquid outlet has negative effect on that. Also changes in gas outlet diameter doesn't influence on the GLCC flowrates domain. Any increases in length of outlets rises the friction force and diminishes the performance of separator.

Gas lock is one of the main problems of multiphase flow in separator risers. This thesis presents the results of comprehensive experimental studies on characteristics of gas lock in riser that has been conducted on a transparent pipe with 24.5 mm internal diameter for simulating riser with variation in angel for 3 different riser angle. Flow regimes map were presented for range of obtained experimental data to show in which settings gas lock has occurred in riser. An examination of the experimental data conducted that the gas lock repetition frequency and Taylor bubble velocity increased with the gas superficial velocity. The Taylor bubble passage duration was observed to increase as the liquid and gas superficial velocity increase in 83°, but was observed be weakly dependent on them in 90°. The slug passage duration was also observed to decrease with an increase in gas and liquid superficial velocity. All of the empirical correlations have overpredicted the actual pressure drop in the riser.

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NOMENCLATURE

c	Total
eq	Equilibrium
f	Friction Factor
g	Gas Phase
in	Inlet
l	Liquid Phase
m	Mixture
q	Flowrate
r	Radius
s	Superficial
v	Velocity
x	Quality Fraction
A	Area
Co	Flow Coefficient
D	Diameter
G	Gas
H	Holdup
L	Liquid
P	Pressure
Q	Flowrate
R	Total Radius
T	Total
U	Velocity
V	Volume
α	Void Fraction
ρ	Density
σ	Interfacial Tension
Δ	Differential
ϕ	Frictional Pressure Loss
$\dot{M}$	Mass Flowrate
2D	Two Dimensional
3D	Three Dimensional
CFC	Upstream Procorditionary Equipment

CFD	Computational Fluid Dynamic
ECT	Electrical Capacitance Technology
GLCC	Gas Liquid Cylindrical Cyclone
LDA	Laser Doppler Anemometer
WMS	Wire Mesh Sensor

CHAPTER ONE: INTRODUCTION

1.1 Background

Over the last few decades, the efficient separation of gas-liquid mixtures applicable to offshore and onshore petroleum industries have become increasingly important. The production of oil after exploration and effective transportation is inevitably accompanied by the presence of natural gas and water. To this end, it is necessary to employ the use of efficient separators to separate small bubbles/drops from the continuous phase which are formed due to shear [15].

Conventional separators have been used as the main separation methods or pre-conditioning equipment placed before gravity settlers. Some of these archaic vessel-type separators utilized in oil industry arc large, heavy and expensive to purchase and operate where the limitations arc most severely felt in offshore operations in cases of escalating platform costs.

Due to the high cost associated with these separators the oil industry have shown a great interest in the development of novel alternatives that are compact, low in weight and low in capital/operating costs. One of such firmly established alternatives is the Gas Liquid Cylindrical Cyclone (GLCC) separator [16].

The simple and compact arrangement as shown in Figure 1-1 consists of a vertical pipe (also known as an upstream pre-conditioner-CFC) with a tangential inlet and horizontal outlets for gas and liquid. The tangential inlet to the body of the GLCC induces a swirl to the flow thereby producing centrifugal force which is an order of magnitude higher than the force of gravity.

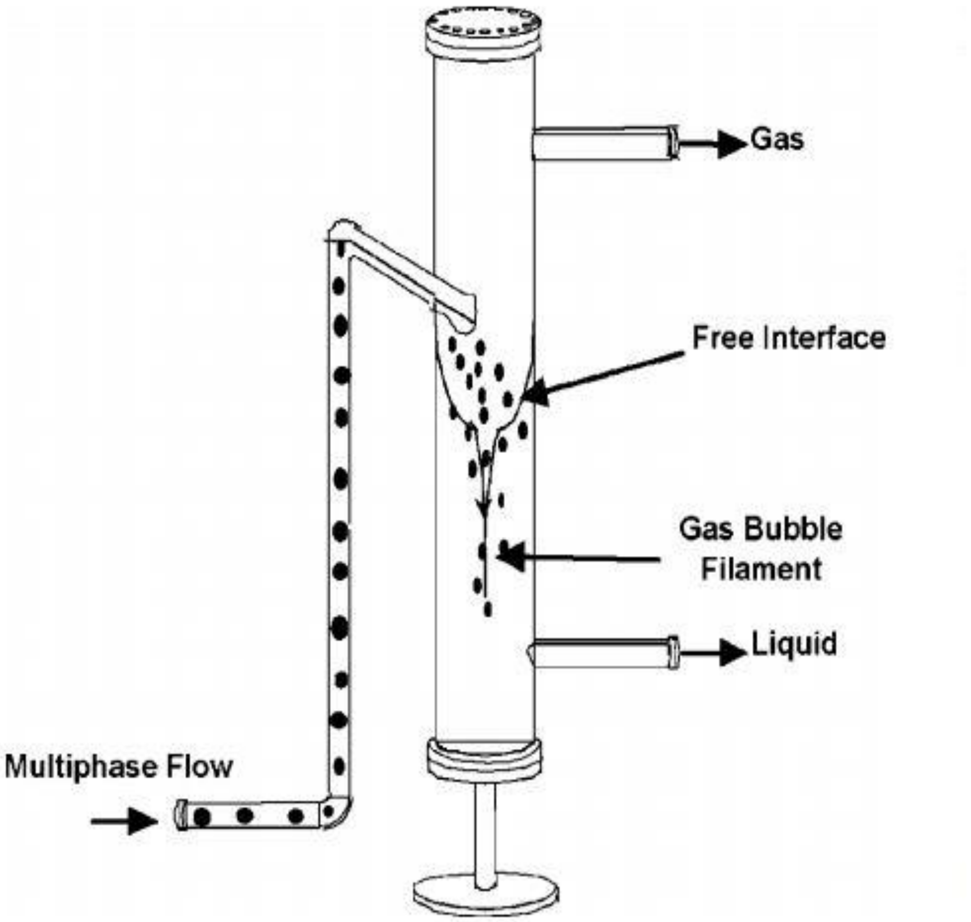

Figure 1-1 Schematic of GLCC separator configuration. [16]

The Gas-Liquid Cylindrical Cyclone (GLCC) separator is commonly used for the separation of oil and gas mixtures flowing from the well head. Similar to the design used by other separators, it has an inlet and two outlets for gas and liquid respectively. However, the inlet to the separator can either be single or dual type. The pipeline connection from the upstream preconditioning equipment (CFC) is inclined downwards and has a tangential inlet slot. The essence of having a downward inclination is to promote pre-separation of the fluid phases. On the other hand, a vertical riser is need to deliver the multi-phase flow to the inlet. Existence of a riser in upstream of a separator effects on the type of phases flow regimes in the inlet of separator [12].

The accurate prediction of slug flow characteristics has been the topic of research for many years. Under certain situation, an unsteady state operation may occur in pipelines and makes it hard to predict the flow regimes characteristics. One of this kind of unsteady state operation happens when a riser section exists in transportation lines. Poor design of the riser pipe connecting a two-phase pipeline can cause 50% reduction in flow capacity in production systems [5].

For Example when a pipeline pass through a steep incline before reaching to platform or when a pipe is connected to a vertical pipe to deliver the multiphase flow to separators, heat exchangers or other surface facilities. In this situation liquid is accumulated in lower part of pipe and blocks the gas way to pass through the riser. The blocked gas is compressed and its pressure increase until it overcomes the liquid gravitational head. Then the compressed gas pushes a long liquid slug through vertical pipe with increasing velocity.

Terrain Slugging is assumed as an unstable flow regime and known for its fluctuations in pressure and the gas and liquid flow rates in the outlet of pipe. A Terrain slug cycle usually divided to four steps. 1.Slug Formation 2.Slug Production 3.Gas Penetration 4.Gas Blow-down. These steps are illustrated in fig 1-2.

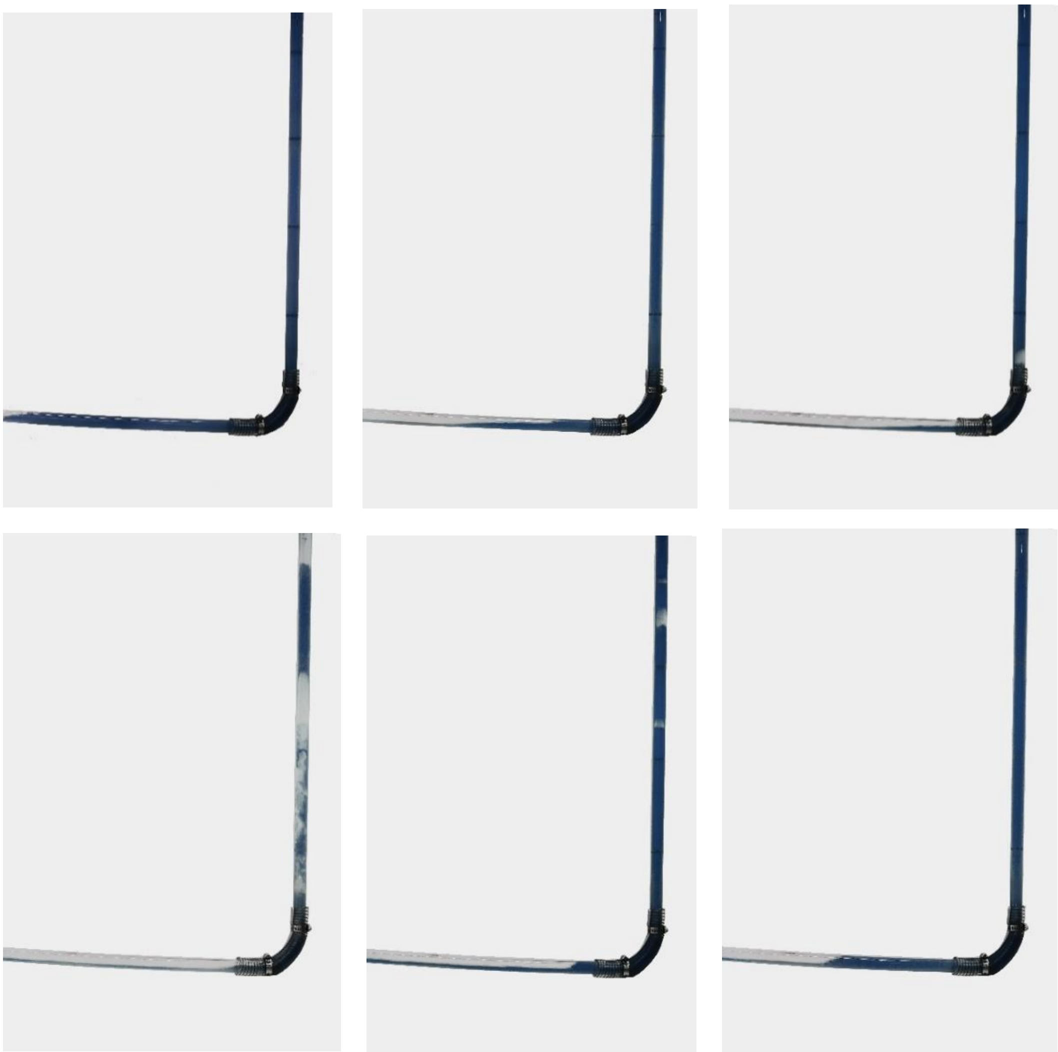

Figure 1-2 Process of severe slug formation

Production systems face with sever slugging when low flowrates of gas and liquid are passing through the system. Fig 1-3 shows the flow regime of experimental data for flow in upstream horizontal section of pipe. As could be seen all of experimental data predict stratified and elongated bubble flow pattern. Enhancing of terrain slugging occurs when the upstream of the horizontal line has a downward inclined to the base of the riser. Fig 1-4 presents the flow pattern in the vertical section of riser which predict bubble and slug flow in vertical line.

The following work attempts to conduct a series of an experiments over a simulated riser which its bend can change in some degrees to study of slug properties. All experimental data is collected by analyzing high speed camera that is the third method category. In following seven different physical properties of air-water mixture flow is investigated by these results. These properties include of slug period (frequency), flow map, Taylor bubble rise velocity, liquid passage duration, Taylor bubble passage duration, fluid passage ratio and pressure drop. All parameters are compared for two different riser angles.

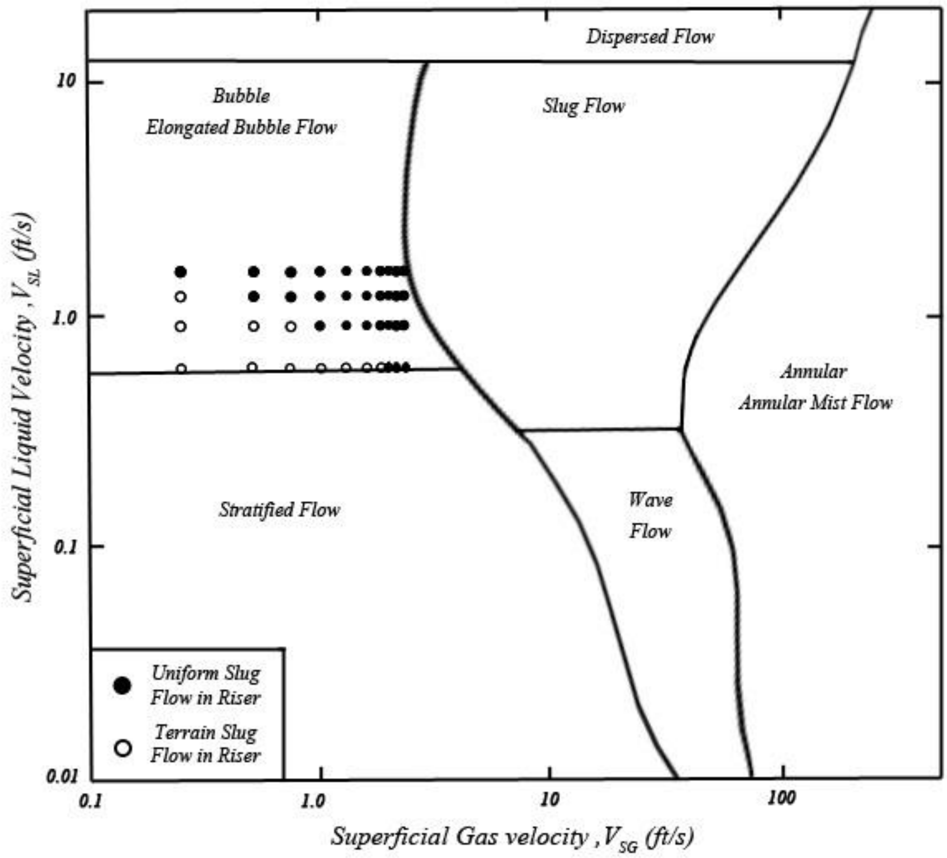

Figure 1-3 Horizontal pipe flow mapm [10]

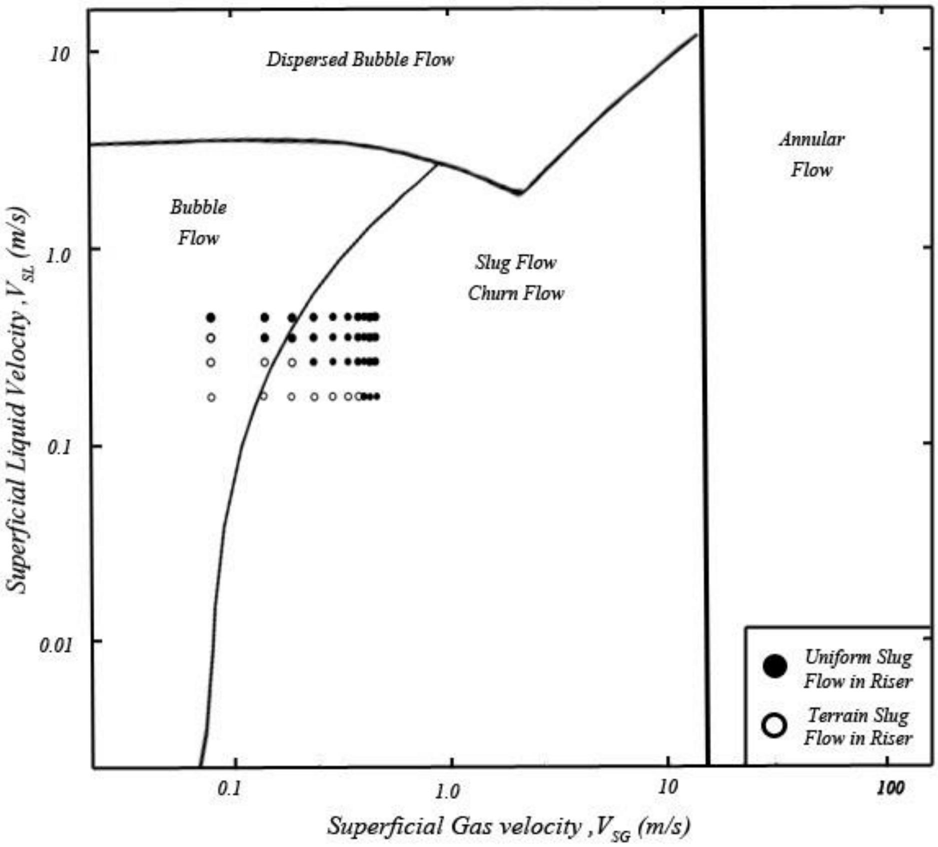

Figure 1-4 Vertical pipe flow map [10]

1.2 Objectives of Thesis

This study tries to investigate:

- The performance of a GLCC separator
- Effect of Changes in inlet diameter on operational flowrates domain of GLCC separator
- Effect of Changes in liquid outlet leg diameter on operational flowrates domain of GLCC separator
- Effect of Changes in gas outlet leg diameter on operational flowrates domain of GLCC separator
- Effect of Changes in length of outlet leg on operational flowrates domain of GLCC separator

- Effect of Changes in length of gas body column on operational flowrates domain of GLCC separator
- Effect of Changes in column diameter on operational flowrates domain of GLCC separator
- Terrain slug creation in separator riser
- Effect of riser angle and flowrates changes on terrain slug flow map
- Effect of riser angle and flowrates changes on terrain slug period
- Effect of riser angle and flowrates changes on terrain slug Taylor bubble rise velocity
- Effect of riser angle and flowrates changes on terrain slug fluid passage ratio
- Effect of flowrates changes on terrain slug pressure drop through vertical pipe of riser

1.3 Summary of Thesis Chapters

This thesis is prepared as follows:

1. A literature review on:

   Overview of Gas-Liquid Cylindrical Cyclone Separators and previous studies

   Field Application Design of GLCC

   Two Phase Flow (Separated Flow Concept)

   Introduction to Gas-Liquid Flow Regimes and previous studies

   Flow Pattern Maps

2. Experimental procedures and materials.
3. Results and discussion on measured data.
4. Conclusion and recommendations for future works.

CHAPTER TWO: LITERATURE

2.1 Overview of Gas-Liquid Cylindrical Cyclone Separators

Detailed literature reviews on compact separation technology research were even by Arpandi et al. by Mohan et al. and by Gomez et al. A summary of state of the art of cylindrical cyclone technology was recently presented by Shoham and Kouba. The Mowing is an update on cylindrical cyclone studies [4,10,20,25].

2.2 Hardware Development

Few systematic studies of design configurations of different GLCC mechanical features have been conducted. Recent laboratory observations and computer simulations indicate that hardware modifications to the GLCC can have a profound effect on GLCC performance. (Kouba & Shoham, 1996) discussed these in some detail. The following is a summary and update of the most important hardware improvements [15].

2.2.1 Inlet Design.

The inlet section determines the incoming gas/liquid distribution and the initial tangential-inlet velocity in the GLCC. Because GLCC performance is strongly dependent on the tangential-inlet velocity, the inlet has been the single most redesigned component of the GLCC [25].

2.2.1.1 Inclined Inlet

Conventional vertical separators typically use a perpendicular inlet. Recent studies on the GLCC have demonstrated that an inclined inlet improves GLCC performance by reducing liquid any-over in the gas stream through two mechanisms. First, the downward inclination of the inlet promotes stratification and provides preliminary separation at the inlet nozzle. Second, the downward inclination causes the liquid stream to spiral below the inlet after one revolution, preventing the liquid from blocking the flow of gas into the upper part of the GLCC [25].

2.2.1.2 Inlet Nozzle.

The nozzle is the last element of the inlet that influences flow distribution and the tangential velocity entering the GLCC body. The tangential inlet male is the most expensive pan of the GLCC to fabricate. Several nozzle configurations have been tested, aimed at optimizing hydrodynamic performance cost-effectively. The optimum configuration for hydrodynamic performance is a thin, rectangular tangential wall slot, which is difficult to fabricate. On the other hand, the concentric-circular tangential inlet is easy to fabricate but exhibits lower performance. A preliminary experimental comparison of three different inlet-slot configurations (rectangular, con-centric-circular, and crescent) with the same cross-sectional area found that the concentric-circular nozzle (reduced pipe) configuration had the poorest performance, while the crescent nozzle (tangential flat plate) performed closest to the rectangular slot [13,25].

2.2.1.3 Dual Inlet

Dual inclined inlets provide preparation of the inlet stream into a liquid-rich stream (lower inlet) and gas-rich stream (upper inlet). Testing of the dual inlet indicated a significant improvement in liquid-carry-over performance at low to moderate gas rates (slug flow dissipating to stratified flow at the inlet) with less discernible improvement at high gas rates (annular flow at the inlet) [13,25].

2.2.2 GLCG Body Configuration

Despite the simple design of the GLCC, several possible modifications to the body configuration can influence performance.

2.2.2.1 Inlet Location

For GLCC's without active liquid-level control, it is important to locate the inlet section just above the liquid level. Most testing to date indicates that the optimum liquid level in the single-inlet GLCC is approximately 1 to 3 L/d below the inlet. Liquid levels farther below the inlet than 3 L/d result in significant decay in the tangential-inlet velocity, which compromises the GLCC performance. If the liquid level is above the inlet, gas must blow through the liquid and is more likely to cause carry-over [25].

2.2.2.2 Optimum Aspect Ratio

The aspect ratio is the ratio of GLCC length to diameter. The dimensions of the GLCC influence performance and cost. For a given diameter, the length of the GLCC above the inlet provides liquid-surge capacity, while the length below the inlet determines residence time for separating bubbles from the liquid. In addition, centrifugal and buoyancy forces are inversely proportional to diameter and tangential-velocity decay is directly proportional to length. Because of the complexity of this phenomenon, a fundamental set of criteria to determine optimum aspect ratio has been proposed only recently) [25].

2.2.2.3 Cyclone-Body Taper

An investigation on diverging, converging, and cylindrical cyclones concluded that cylindrical walls are slightly superior to either converging or diverging walls for gas/Liquid separation?

2.2.3 Liquid-Level Controls

Active liquid-level control in a GLCC for a wide range of flow conditions is not straightforward owing to its compact size. Several different liquid-level-control strategies are being investigated, including flow control on the gas leg, flow control on the liquid leg, and flow control on both legs. Also being considered are combinations of backpressure control on the gas leg and liquid-level control on the liquid leg. Other issues of concern include power requirements, robustness, and cost. Several alternatives for GLCC liquid-level control have been implemented. For example, a commercial multiphase measurement system has used conventional control equipment successful-ly to maintain a tight control on liquid level by controlling the gas-outflow rate of the GLCC. Another project explored low-power alternatives to conventional level controls that exploit hydrostatic-head difference in the GLCC to operate the controls.8 A recent study examined GLCC performance with a passive control system that uses only the flow energy and no external energy.5 Crucial future work is to develop robust, active liquid-level-control strategies. Because of the smaller residence time of the compact separator and the stringent response time requirement of the control valve, this is not a simple extension of the control technology avail-able for large vessel-type separators. The strategies should enable the GLCC to handle slugging, surging, and a wide range of flow rates, from essentially full-gas-flow to full-liquid-flow conditions [25].

2.2.4 Integrated Separation System

Great economic incentives exist for the industry to move away from conventional gravity-based separators to compact separation systems. Depending on the application, the GLCC can be used for full or partial separation. Partial gas separation allows downstream equipment to be smaller (and therefore less expensive) and perform more efficiently. The GLCC has been particularly effective when combined with multiphase meters, desanders, and liquid/liquid hydrocyclones. Configured either alone or in combination with other equipment, the GLCC can reduce cost and weight significantly. This is particularly important in designing or retro-fitting offshore platforms, where savings in platform-construction costs may be many times greater than the cost of the separation equipment [25].

2.2.5 Experimental Studies

Replan and Gauvins studied the behavior of confined vortex flow in conical cyclones. Their studies show that an increase in the magnitude of the inlet velocity does not change the shapes of the tangential velocity axial velocity and the static pressure profiles but increases their respective magnitudes [6].

Local Laser Doppler anemometer (LDA) velocity measurements in cylindrical cyclone separators were reported by Millington and Thew. These authors suggested the use of twin, diametrically opposite inlets for greater axisymmetric and gas core filament stability, leading to a much improved gas carry-under performance. They made the important observation that the vortex occurring in the cylindrical cyclone separator is a forced vortex with a tangential velocity structure.

Farchi conducted tangential velocity measurements in a cylindrical cyclone with static pitot tubes. His measurements confirmed that a forced vortex occurs in the cyclone. However as the diameter of the cyclone increases, the velocity distribution tends to match the free vortex profile [6].

Through a study on gas-liquid flow characteristics in a spiral horizontal cyclone with vortex generator. Kurokams and Ohtaik confirmed the existence of a complex velocity profile by accurate single-phase liquid flow measurements. The study distinguishes a forced vortex generating a jet region with extremely high swirl velocity around the pipe center from a second swirl region formed by a free vortex near the wall and also an intermediate region of backflow with high swirl velocity [6].

2.2.6 Mechanistic Modeling

Few mechanistic models have been developed recently to describe and predict the flow behavior in the cylindrical cyclone. A mechanistic model for predicting separation efficiency based on the analysis of droplet trajectories in liquid-liquid, oil/water hydrocyclones was presented by Wolbert et al. These trajectories were calculated through a differential equation, combining mechanistic models for the three bulk velocity distributions, namely radial and tangential [6].

Arpandi et al., based on experimental and theoretical studies performed at Tulsa U. Separation Technology Projects (TUSTP), have developed a mechanistic model capable of predicting the general hydrodynamic flow behavior in a cylindrical cyclone separator. This includes simple velocity distributions, gas-liquid interface shape, equilibrium liquid level, total pressure drop, and operational envelope [4,6].

An analysis of bubble trajectory for cylindrical cyclone separation was presented by Marti et al. The model predicts the gas-liquid interface (vortex) near the inlet as a function of the radial distribution of the tangential velocity. A bubble trajectory analysis enables the determination of separation efficiency, based on the gas bubble size [19,6].

New experimental data on the acts of geometry, fluid physical properties and pressure on the hydrodynamic flow behavior in cylindrical cyclone separators were presented by Movafaghian and by Movafaghian et al. The data were used to verify and refine the cylindrical cyclone mechanistic model developed previously by Arpandi et al [4,22,6].

Wang et al. developed a steady-state and a dynamic model as the framework for cylindrical cyclone passim and active control, respectively. The steady-state model was used to analyze the sys-tem sensitivity and the dynamic model was used to analyze the system stability by applying linear control theory. In this investigation, a preliminary control strategy was proposed for cylindrical cyclone control system design [6,28].

A set of correlations for the prediction of the velocity field in the cylindrical cyclone (tangential and axis' was presented by Mantilla et al. An improved bubble trajectory model was presented that uses the developed correlation and showed good agreement with the experimental data which was obtained previously [6,18].

Recently. Gomez et al. is developed a state-of-the-art computer simulator for cylindrical cyclone design in an Excel-Visual Basic platform that is capable of integrating the different modules of the mechanistic model. Model enhancements include a flow pattern dependent nozzle analysis for the cylindrical cyclone inlet, an analytical model for the gas-liquid vortex interface shape, an unified panicle trajectory model for bubbles and droplets, including a tangential velocity decay formulation, and a simplified model for the prediction of the gas and liquid cylindrical cyclone aspect ratio [6,10].

2.2.6.1 Hydrodynamic mechanistic model

The mechanistic model of Kouba et al. forms the theoretical foundation for the model developed in this section. Significant improvements have been made to the prediction of the GLCC hydrodynamic flow behavior. Also, the new model includes, for the first time, the prediction of the operational envelope of the GLCC [12,16]. This was made by addressing the following 3 questions concerning the onset to liquid carry-over: (1) How much liquid there is in the GLCC (Equilibrium liquid Level)? (2) How is the liquid distributed (Gas-Liquid Interface)? and (3) How much liquid can be tolerated in the upper part of the GLCC before liquid carry-over is observed (Zero-net Liquid Holdup)? The sub-models presented in the following sections provide the answers for these questions. Finally, these sub-models are combined for the prediction of the liquid carry-over operational envelope. The GLCC geometrical parameters and nomenclature for the model are given in Figure 2.1 [4,12,16].

2.2.6.1.1 Equilibrium Liquid Level.

Determination of the equilibrium liquid level is important for the prediction of both liquid carry-over and gas carry-under. For proper operation of the GLCC, the liquid level must be maintained below the inlet to avoid gas flowing through the liquid stream and carrying liquid into the gas leg. Also, the liquid level should be sufficiently high above the liquid exit at the bottom of the GLCC. This is done in order to avoid gas carry-under in the liquid stream and prevent gas liberation in the liquid meter. There-fore, it is essential to be able to predict the liquid level for proper operation of the GLCC [4].

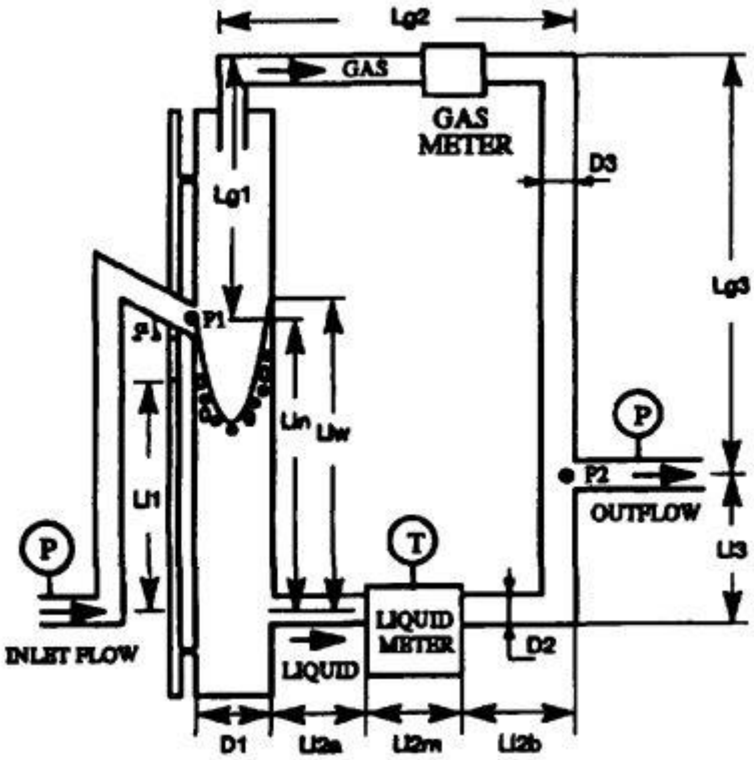

Figure 2-1 GLCC Nomenclature for the mechanistic model [4]

The liquid level can be determined for the metering loop configuration by balancing the pressure in the gas and the liquid legs, between the inlet and outlet of the GLCC (P1 and P2 in Figure 2.1). This model neglects any hydrodynamic inter-actions between the gas and the liquid phases. Following Kouba et al., the pressure drops in the liquid and gas legs are given, respectively, by [4,16]

$$\Delta P_l = \rho_l g\left(L_{eq} - L_{l_3}\right) + \rho_g g\left(L_{in} - L_{eq}\right) - \left(\phi_l + \frac{f_{l_1} L_{eq}}{D_1}\frac{\rho_l v_{l_1}^2}{2}\right) \qquad (2.1)$$

$$\Delta P_g = \rho_g g\left(L_{g_3} - L_{g_1}\right) - \phi_g \qquad (2.2)$$

Where ϕ_l and ϕ_g are the frictional pressure losses in the liquid and gas sections, as given:

$$\phi_l = \frac{\rho_l}{2}\left(\sum_{i=2}^{n}\frac{f_i L_i v_i^2}{D_i} + \sum_{i=1}^{n} K_i v_i^2\right)_l \qquad (2.3)$$

$$\phi_g = \frac{\rho_g}{2}\left(\sum_{i=1}^{n}\frac{f_i L_i v_i^2}{D_i} + \sum_{i=1}^{n} K_i v_i^2\right)_g \qquad (2.4)$$

The first terms in the parentheses of Equations (2.3) and (2.4) represent the frictional losses in the different pipe segments of the loop and the second terms represent the losses in the different piping fittings. Equating the pressure drops in the liquid and gas sections, as given by Equations (2.1) and (2.2), the liquid level can be solved as follows:

$$L_{eq} = \frac{\phi_l - \phi_g + \rho_l g L_{l_3} - \rho_g g\left(L_{in} + L_{g_1} - L_{g_3}\right)}{g\left(\rho_l - \rho_g\right) - \left(\frac{\rho_l v_{l_1}^2}{2}\frac{f_{l_1}}{D_1}\right)} \qquad (2.5)$$

2.2.6.1.2 Gas-Liquid Interface

The physical model for the determination of the gas-liquid interface shape is given in Figure 2.2. The main assumption is that the tangential flow from the inlet into the GLCC generates a forced vortex. This was substantiated by Millington and Thew [4].

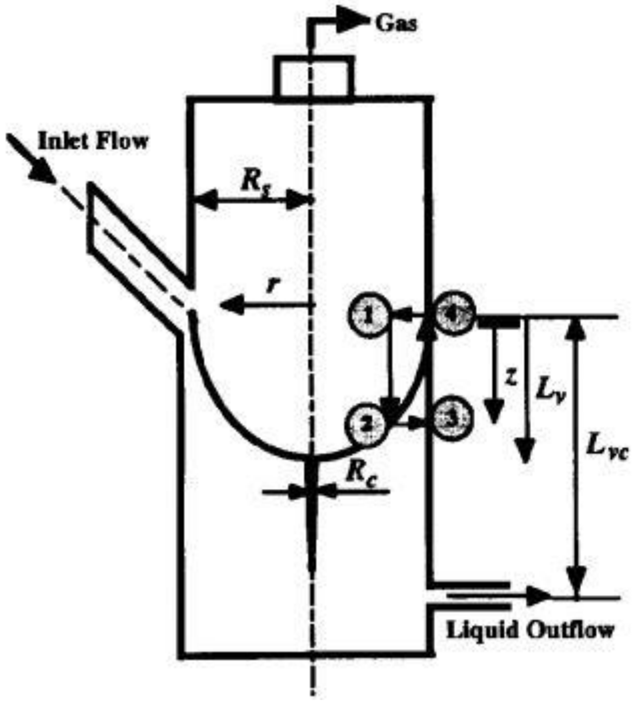

Figure 2-2 Gas-liquid interface geometry [4]

The model is essentially a pressure balance between points 1 to 4. The pressure drops between points 1 to 2 and 3 to 4 are simply due to the hydrostatic head in the gas and the liquid phases, respectively. No pressure change is assumed in the gas between points 4 to 1. To close the pressure "loop" the pressure change between points 2 to 3 is needed. This pressure difference is due to the centrifugal force acting on the two-phase mixture, and can be calculated as follows. The tangential velocity distribution for a forced vortex is of the form

$$v_t(r) = v_{t_{is}}(\frac{r}{R_s})^n \tag{2.6}$$

Where n=1 and v_{tis} is detrmined from the liquid celocity ar the inclind inlet, v_l, as follows

$$v_{t_{is}} = v_l \frac{A_{in}}{A_{is}} \tag{2.7}$$

The liquid velocity, v_l, is calculated using the Taitel and Dukler model. Similarly, the radial holdup or mixture density distribution is assumed to be of a similar form as the tangential velocity.

$$\rho_m(r) = \rho_g + \left(\rho_l - \rho_g\right)(\frac{r}{R_s})^m \tag{2.8}$$

In this model, it is assumed that only liquid present around the vortex, namely, m=0. At any given position, the radial pressure difference due to the centrifugal force, between points 2 to 3 as shown in figure 2.2, is given by

$$\Delta P(r) = \int_r^{R_s} \frac{\rho_m(r)[v_t(r)]^2}{r} dr \tag{2.9}$$

Finally, carrying out the pressure balance between points 1-4 results in an equation that gives the location of the interface at any axial position z as a function of the radial coordinate r, namely

$$\mathrm{z(r)} = \frac{\Delta \mathrm{P(r)}}{\mathrm{g}(\rho_\mathrm{l} - \rho_\mathrm{g})} \tag{2.10}$$

The total liquid volume displaced by the gas vortex and core is

$$\mathrm{V_g} = \int_{\mathrm{R_c}}^{\mathrm{R_s}} 2\pi \mathrm{rz(r)dr} + \frac{\pi}{4}\mathrm{D_c^2}(\mathrm{L_{vc}} - \mathrm{L_v}) \tag{2.11}$$

where Lv = z(Rc). The second term in Equation (2.11) is the volume of the core filament that extends front the bottom of the gas core vortex to the liquid exit, as shown in the figure. The height of the liquid, where the gas-liquid interface touches the wall, namely, the vortex crown, is calculated as if the total gas volume is submerged in the liquid, as follows

$$L_{vc} = L_{eq} + \frac{V_g}{A_s} \quad (2.12)$$

2.2.6.1.3 Zero-Net Liquid Holdup

For zero-net liquid flow, assuming churn/slug flow in the upper part of the GLCC, the gas velocity can be developed from a modified Taylor bubble rise velocity expression, namely

$$v_{g_o} = C_o v_{sg} + 0.35\sqrt{gD_s(\frac{\rho_g - \rho_g}{\rho_l})} \quad (2.13)$$

A constant value for the flow coefficient Co is assumed for slug/churn flow, as given by

$$C_o = 1.15 \quad (2.14)$$

The liquid holdup is given by

$$H_{l_o} = \left[1 - \left(\frac{v_{sg}}{v_{go}}\right)\right](1 - \frac{L_d}{L_{g_1}}) \quad (2.15)$$

where L_{g1}, is the total height of the GLCC above the inlet (see Figure 2.1). Churn/slug flow occurs only in the lower region, right above the inlet, while at the top region, liquid is present primarily in the form of droplets. The length of the droplet region, Ld, can he determined from a simplified droplet ballistic analysis. It is equal to the trajectory length of a fine droplet, assuming that the gas void fraction in this region is approximately one. This results in the upward gas velocity being approximately equal to the superficial gas velocity. Thus, the length of the droplet region, Ld, is given by

$$L_d = \frac{1}{\frac{2g_c}{v_{sg}^2} - \frac{C_d}{2}(\rho_g v_{sg})^2 \frac{3}{32\rho_l \sigma g_c}} \tag{2.16}$$

Note that Equation (2.16) can be rearranged to determine the blow-out velocity, v_{bo} . This is the droplet velocity (v_{sg} in Equation (2.16)) for which the length of the droplet region, Ld, is equal to the total height of the GLCC above the inlet, L_{g1} . Clearly, for these conditions the zero-net liquid holdup, as given by Equation (2.15), tends to zero [4].

2.2.7 Computational Fluid Dynamics Simulations

Computational fluid dynamics (CFD) simulations have been used to support the mechanistic modeling effort by investigating the detailed hydro-dynamics of the flow in the cylindrical cyclone. Bandyopadhyay at al. conducted a numerical study to investigate the mechanism of separating gas bubbles from a bulk liquid in a cyclone separator. Erdal et al. presented CFD simulations using a commercial code called CFX (CFX 4.1). The simulations presented details of the flow behavior in the cylindrical cyclone for single-phase and two-phase flow. The results verified that two-dimensional (2D) axisymmetric simulations (with three velocity components) gave similar results to the three-dimensional (3D) simulations. An expression was developed for the equivalent inlet tangential velocity for the axisymmetric model [6,8,9].

Motto et al. presented a simplified model, based on a CFD approach, for rotational two-phase flow in a cylindrical cyclone separator. The model assumes an axisymmetric flow with three velocity components, and is applicable to steady-state and isothermal conditions. As an example of a potential application of the proposed model for cylindrical cyclone design, the study combiner the gas cony-under and liquid carry-over envelopes to present the region of proper operation of the cylindrical cyclone [6].

The above overview of the state-of-the-art of Gas/Liquid Cylindrical Cyclone (GLCC) technology reveals that more studies need to be conducted in order to be able to design and operate the cylindrical cyclone properly. The present study includes a mechanistic model and for the first time new experimental data for liquid carry-over in the cylindrical cyclone beyond the operational envelope and extension of the model for high-pressure conditions [6].

2.2.8 Compact Separator and GLCC Separators Studies

The Gas-Liquid cyclone separator is a common type of separator used in the oil and gas industry. Much of the earlier work consists of reducing the size and increasing the efficiency of the separator.

Davies and Watson (1979) pioneered the use of gas-liquid cyclone separators for offshore production operations. They were successful in reducing the size and weight of offshore separators. Additionally, experiments were performed using a rectangular- tangential inlet by Zikarev and Kutepov (1985) and Nebrensky et al. (1980) for optimizing the separator's performance [6].

Due to the increasing number of offshore exploitations, there was a need to further reduce the cost and footprint of gas-liquid separators. The GLCC was a compact separator developed by Chevron and University of Tulsa investigators Kouba et al. (1995). Thus, earlier studies on behavior and optimal design of the GLCC was performed by Kouba et al. (1995). Utilizing the field and laboratory data, it was shown that inclined angle for inlet of the GLCC increases the operational envelope of separator liquid carry-over [16].

Marti et al. (1996) studied gas carry-under in the GLCC Separator. Arpandi et al. (1996) performed experiments in a 3-inch GLCC to measure the pressure loss and liquid carry over in the GLCC. Movafaghian (1997) performed experiments on different inlet geometry and liquid viscosity. Erdal (1997) performed CFD stimulation in the GLCC for single-phase flow and two-phase flow to understand the hydrodynamic behavior of flow in the GLCC. CFD simulation were performed for bubble carry-under in the GLCC [4,7,6,19].

Chirinos (1998) conducted experiments at different gas flow rata and liquid flow rates. It was observed that in chum flow, at low gas flow rates and high liquid flow rates, large quantities of liquid were carried over. While for annular flow, high gas flow rates and low liquid flow rates, large quantities of liquid droplet were carried through the gas outlet [6,22].

Chirinos and Gomez (1999) performed experiments in the GLCC for liquid carry-over in the separator. Kolla (2007) performed experiments on the GLCC to calculate the operational envelope of liquid carry-over for the gas-water-oil phase. The experiments were performed for heavy oil and light oil, and the effect of water cut on the liquid carry over was noted [6,10].

2.3 Field Application Design

The developed simulator has been utilized to design over 100 GLCC systems, the majority of which operating within the USA, and several in Canada, South America, Africa and the Far East. This section provides details of four typical GLCC field units that are now in operation [17].

2.3.1 GLCC Multiphase Metering Loop System

The GLCC, configured in a multiphase metering loop, offers an attractive alternative for production well testing and metering. As shown schematically in Figure 2-3, in this application, the separated gas exits from the top of the GLCC into the gas leg where it is measured with a single-phase gas flow meter. Similarly, the liquid phase exits from the bottom of GLCC and is metered with a single-phase liquid flow meter. The two phases are then re-combined in the recombination point downstream. A fill separation is required for this application avoiding gas carry-under in the liquid stream and liquid carry-over in the gas stream. Full separation can be enhanced with proper control systems (Wang et al., 1998, Mohan et al., 1998) [20,28].

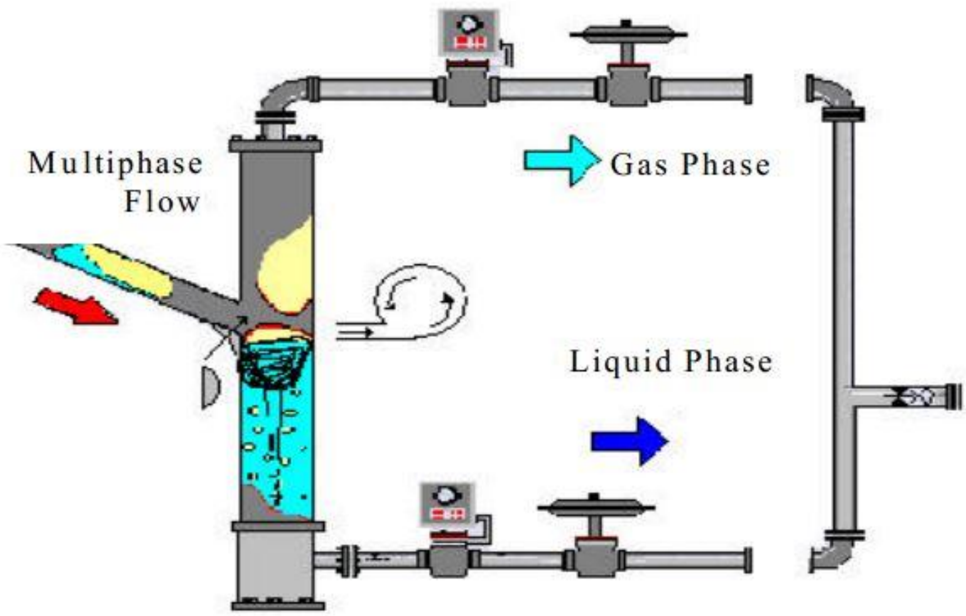

Figure 2-3 Real field design of GLCC [17]

The GLCC Metering loop system design was carried out for an offshore platform operated by PDVSA in Lake Maracaibo. As given in Table 2-1, production from 24 wells is gathered on the platform. The production data show a wide range of gas and liquid flow rates, namely: 683-4597 Mscf/d gas and 43-3033 bbl/d oil. The different wells have been routinely tested in the past utilizing a vertical test separator operating in a dump cycle and metered by an orifice meter. This previous system was large, and inconvenient, and has recently been replaced by a compact GLCC multiphase metering loop.

Table 2-1Design of flow situation [17]

Well #	Q_g (Mscf/d)	Q_t (Stb/d)	Q_o (Stb/d)	Q_w (Stb/d)	Equilibrium Liquid Level	
					d_g = 2.0”	d_g = 4”
1	1671.0	3033.0	182.0	2851.0	2.43	
2	757.0	1436.0	345.0	1091.0	4.49	
3	704.0	1047.0	272.0	775.0	3.06	
4	1591.0	2697.0	1187.0	1510.0	2.41	
5	682.0	1250.0	175.0	1075.0		5.16
6	3172.0	3207.0	641.0	2566.0		11.22
7	1282.0	1705.0	375.0	1530.0		7.07
8	1298.0	1413.0	480.0	933.0		5.41
9	919.0	608.0	365.0	243.0		3.89
10	1280.0	1697.0	1646.0	51.0		7.38
11	1618.0	1531.0	306.0	1225.0		5.46
12	1062.0	133.0	80.0	53.0		3.21
13	1970.0	396.0	238.0	158.0		2.69
14	1424.0	368.0	236.0	132.0		3.19
15	1440.0	595.0	500.0	95.0		3.57
16	1120.0	349.0	328.0	21.0		3.42
17	1096.0	175.0	168.0	7.0		3.24
18	2180.0	164.0	158.0	6.0		2.16
19	1668.0	491.0	481.0	10.0		3.19
20	1929.0	245.0	240.0	5.0		2.54
21	4597.0	50.0	50.0	0.0		-2.92
22	1594.0	43.0	43.0	0.0		2.72
23	2530.0	330.0	328.0	2.0		1.83
24	1006.0	231.0	230.0	1.0		3.35

The wide range of operational conditions made it difficult to design a single GLCC unit. The primary consideration in the design process was to maintain the liquid level in the GLCC below the inlet for all flow rate combinations. The proposed GLCC configuration, as shown in Figure 2-4, is 12 inch I.D. and 11 ft. high. The 3-inch diameter inclined inlet is located 5 ft. from the bottom of the GLCC.

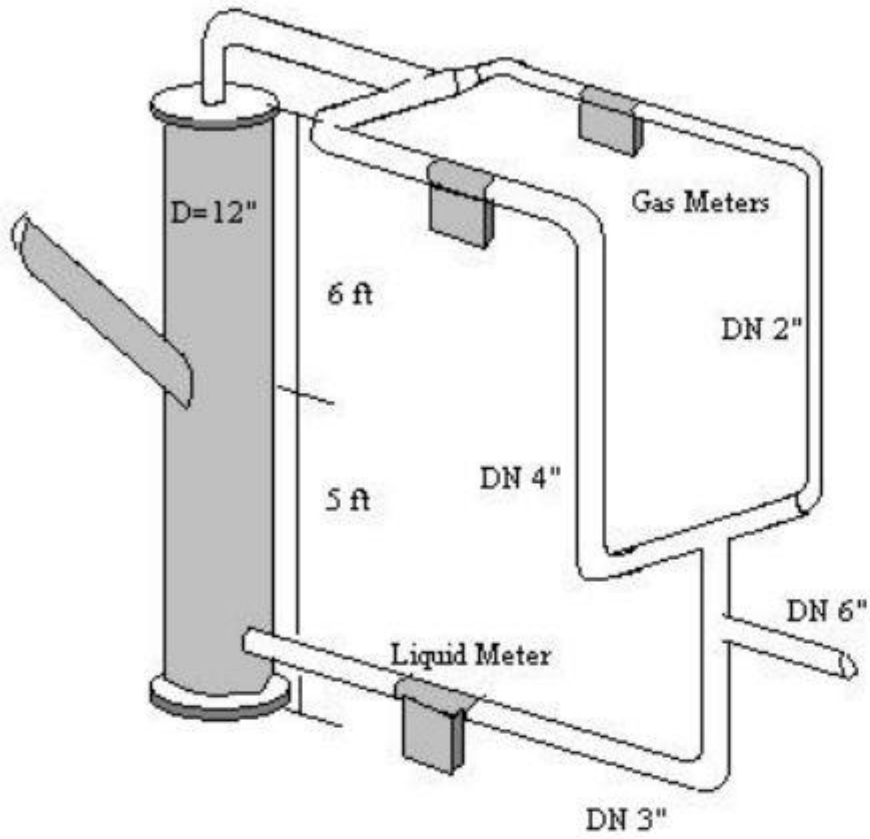

Figure 2-4 GLCC application in a multiphase metering loop [17]

Initial simulation showed that the results were not sensitive to the liquid leg dimensions (since the liquid meter dominates the pressure drop), while the gas leg diameter had a more significant effect. As can be seen no single gas leg diameter is capable of tolerating the production from all the wells without application of a control system. The proposed alternative is to use two different gas legs in parallel, as shown in Figure 2-4.

The proposed liquid leg is 3 inch I.D and 8 ft. long, equipped with a Micro Motion mass flow meter that gives the total flow rate and the water cut. Since control was not considered, a dual gas leg was designed, 2 inch and 4 inch in diameter, each with a gas flow meter. The gas can flow through either legs or both legs. The 2-inch gas leg is used for the wells with low gas and high liquid production, while the 4-inch gas leg is used for the high gas and low liquid production wells. For the highest gas and lowest liquid production rates, the gas can flow simultaneously through both legs.

Table 2-1 gives the simulated results for the equilibrium liquid level in the GLCC for the 24 wells. As can be seen, the first 4 wells have a gas and liquid rate combinations that require flowing through the 2-inch gas leg, while the other 20 flow through the larger gas leg. The simulator can be used to identify the wells that will not fall within the proper operational envelope of the GLCC metering loop. For this dual gas leg configuration, only wells 6 and 21 fall out of the proper operational range of the system, as denoted by very high liquid level of 11.22 ft and negative liquid level of 2.99 ft., respectively. With the 4 inch gas leg, well 6 will experience liquid carry-over while well 21 will experience gas carry-under. If desired, this can be resolved by utilizing a control system, whereby, the gas or liquid legs are choked appropriately. The gas line should be choked for well 6 (in order to lower the liquid level in the GLCC and the liquid line should be choked for well 21 (in order to increase the liquid level in the GLCC. Alternatively, the restriction in the gas leg for well 21 could be overcome by opening both the parallel gas legs during the testing of this well.

This field application successfully demonstrates that the GLCC can be configured in a multiphase metering loop utilizing off-the-shelf technology single-phase flow meters on the gas and the liquid legs, resulting in significant savings as compared to utilization of a full bore three phase meter [17].

2.3.2 GLCC External Pre-Separation System

The GLCC can be utilized as a pre-separator upstream of conventional vessel-type separators, as shown schematically in Figure 2-5. This configuration provides considerable advantages extending the operating range, capacity and efficiency of the conventional separator and eliminating the need for utilization of slug catchers or vessel type separators upstream of the primary conventional separators. This application does not require a full separation in the GLCC.

The gas stream from the exit of the GLCC, which might contain some entrained liquid, flows into the upper part of the vessel separator, while the liquid stream, with some entrained gas, flows to the lower part of the vessel separator. The pre-separation in the GLCC ensures "quieter" operation of the vessel separator, reducing the inlet momentum and turbulent mixing, resulting in lower emulsion generation. Also, the liquid level in the GLCC is dictated by the liquid level in the vessel separator. Thus, there is no need for control system since the vessel separator serves as a liquid level control for the GLCC.

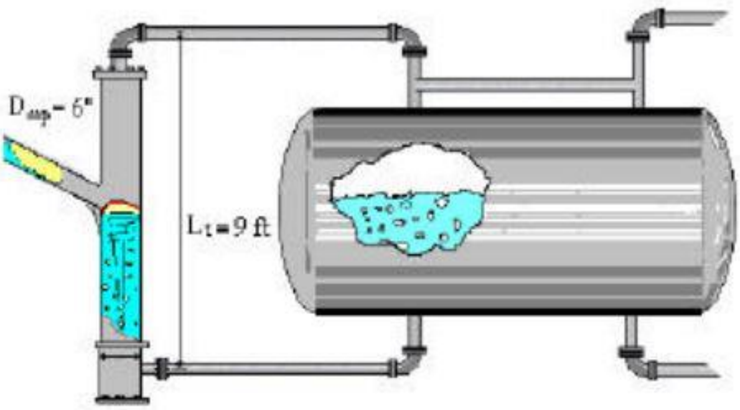

Figure 2-5GLCC application in pre-separation configuration [17]

A GLCC external pre-separation system was designed for Petrobras, Brazil, to be operated in a wet gas field in the Amazon jungle, with operating condition of 1786 bpd, and 461 Mscfd at 611 psia and 59°F. For this particular application, due to increasing water production, terrain slugging has occurred and the existing vessel type separator did not function properly. The GLCC, as a pre-separator, was designed and installed as a slug breaker, to attenuate the effect of slugging and ensure more even flow rates into the existing vessel separator.

The GLCC (schematic shown in Figure 2-5) is 6-inch in diameter and 9-ft high. In order to avoid the loading effect of the vessel separator on the pre-separator, the GLCC is designed such that the inlet is located 0.5 to 1-ft above the liquid level in the conventional vessel separator. This application resulted in significant savings, eliminating the need to replace the existing separator with a larger one [17].

2.3.3 GLCC Gas Knockout System

The GLCC can be used as a gas knockout drum in partial processing applications. An example is gas knockout upstream of de-sanders. Most of the gas can be diverted upstream of the de-sander and recombined downstream, after the removal of the sand. This results in a smaller unit and more efficient operation of the de-sander. Another example is multiphase metering of a high GOR stream, where the gas is knocked out upstream of a multiphase meter and metered separately by a single-phase meter. Again, utilization of the GLCC results is a smaller and less expensive multiphase meter with an improved performance.

A GLCC was designed (not yet installed in the field) for ARCO Alaska to operate as a gas knockout drum to remove partially the gas from a high GOR crude oil system upstream of a multiphase flow meter. This system, too, did not need complete separation of gas and liquid. The requirement for the knockout system was to keep the in-situ gas volume fraction below 10%.

Table 2-2 provides 5 different cases chosen for the simulation and design of the GLCC system. These 5 cases indicate the expected flow rate conditions over a 20-year period. The design was aimed at eliminating the carry-over of liquid in the gas stream completely in the upper part of the GLCC, so that a single-phase meter could be used in the gas leg. A schematic of the GLCC design proposed for this application is shown in Figure 2-6. The designed GLCC is 30 inch in diameter and 8-ft high. The inlet is located 3 ft from the bottom of the GLCC in order to avoid liquid carry-over in the upper part of the GLCC.

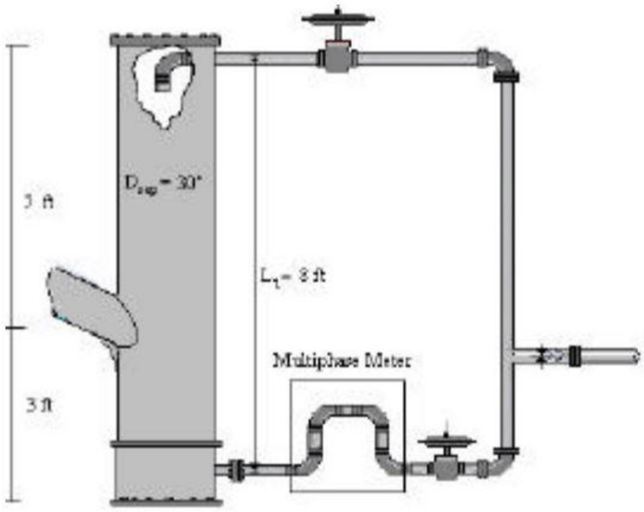

Figure 2-6 GLCC application as a gas knockout system [17]

The gas and the liquid outlet flow rates are controlled using the respective control valves so as to maintain the liquid level in the GLCC below the inlet for various inflow conditions. The percentage valve opening predicted by the simulator is represented as the ratio d_{min}/d_{max} , as shown in Table 2-2, which could be used to set the control valve position corresponding to the respective production forecast for the different years [17].

Table 2-2 Design of GLCC in Alaska [17]

Case	Year	Q_o (bbl/d)	GOR (scf/stb)	Q_g (MMscf/d)	V_l (ft/s)	V_G (ft/s)	P (psi)	d_{min}/ d_{max}	Comment
1	5	1500	25,000	37.5	5.3	34.3	1000	0.55	High qo
2	1	2000	20,000	40	7.0	36.6	1000	1.0	Max qo
3	1	2000	6,000	12	7.0	10.7	1000	1.0	High qg
4	5	1500	10,000	15	5.3	13.6	1000	1.0	
5	20	250	50,000	12.5	0.9	11.5	1000	0.55	High GOR

2.3.4 GLCC Bulk Separation/Metering System

The largest GLCC designed utilizing the developed simulator has been recently installed by CPI in Indonesia (Kouba and Marrelli, 1999). The GLCC, shown in Figure 2-7 is 5 ft. ID and 20 ft tall. It is installed for bulk separation/metering of the production from the Minas Light Oil Steam Flood (LOSF) field. The GLCC operates at 160 psia and 360°F, handling liquid and gas production rates of 200,000 bpd 17 MMscfd, respectively. The GLCC is equipped with control valves on the gas and liquid legs and a sophisticated control system for liquid level control. This GLCC (with additional 3 36 inch GLCC improved the metering accuracy and saved about $3 million over conventional separators for Minas field [17].

Figure 2-7 GLCC application in a bulk separation/metering loop [17]

2.4 Two Phase Flow (Separated Flow Concept)

When more than one phase lbws in a conduit, namely: gas, liquid and solid, this occurrence is considered as multiphase flow. In consideration of the flows in industrial equipment such as in the production of hydrocarbons, power generation and those in chemical industries, multiphase flow occurs in all facets of these industrial applications. The flow can be of various forms, that is, by the combination of the phases above: Gas-Liquid-Solid, Gas-Liquid-Liquid and Solid-Liquid-Liquid. However, for the remit of this work, this thesis concentrates on an aspect of multiphase flow, two phase gas-liquid flows.

When these two phases, gas and liquid, flow in the same pipe for instance, a deformable interface is formed between them. Furthermore, the gas or liquid occupies a certain fraction of the pipe cross-sectional area. From general consensus, the fraction of the pipe elms-section occupied by the gas phase is known as the void fraction, (Azzopardi. 2006). The term liquid holdup or liquid fraction is given to the fraction of the pipe cross-section occupied by the liquid phase. The section below sheds more light on related terms as regards two-phase gas-liquid flow.

2.4.1 Void Fraction

From the discussion above, the void fraction is mathematically given below as. (Azzopardi. 2006):

$$\alpha_g = \frac{A_g}{A_c} \tag{2-17}$$

where A_g and A_c are the area occupied by the gas phase and the cross-sectional area respectively. The liquid holdup on the other hand is given as:

$$\alpha_L = 1 - \alpha_g \tag{2-18}$$

When these two phases flow in the pipe as shown in Figure 2.8, they travel at a particular mass flowrate referred to as the total mass flowrate which is the sum of the flowrate of the phases.

$$\dot{M}_T = \dot{M}_g + \dot{M}_l \tag{2-19}$$

where $\dot{M}_T$ $\dot{M}_g$ and $\dot{M}_l$ and M are the total mass flowrate and mass flowrate of the gas and liquid phases respectively.

The fraction of the flow travelling as gas/vapor is called the quality and is given as:

$$x_g = \frac{\dot{M}_g}{\dot{M}_g + \dot{M}_l} \tag{2-20}$$

The volume fluxes of the phases give the gas and liquid superficial velocities. This represents the velocity at which each phase will travel as if occupying the entire pipe cross section.

Gas superficial velocity,

$$U_{gs} = \frac{\dot{m} x_g}{\rho_g} \tag{2-21}$$

Liquid superficial velocity,

$$U_{ls} = \frac{\dot{m}(1 - x_g)}{\rho_l} \tag{2-22}$$

From equations (2.17) and (2.21), the mean gas velocity is given in equation (2.23). On the other hand, the mean liquid velocity is given in equation (2.24) from equations (2.18) and (2 22).

Mean gas velocity,

$$U_g = \frac{U_{gs}}{\alpha_g} \tag{2-23}$$

Mean liquid velocity,

$$U_l = \frac{U_{ls}}{\alpha_L} \tag{2-24}$$

In consideration of the mass balances for each phase. the void fraction can also he defined as:

$$\alpha_g = \frac{1}{(1 + \frac{U_g}{U_l} \frac{(1 - x_g)\rho_g}{x_g \rho_l})} \tag{2-25}$$

where (U_g/U_l) is the ratio of the mean velocities for the gas and liquid phases and is known as the slip ratio. U_g. When $U_g = U_l$, so that $U_R = 1$, this is known as homogenous flow. In this case the equation (2.9) becomes:

$$\alpha_{gH} = \frac{1}{(1+\frac{(1-x_g)\rho_g}{x_g\rho_l})} \qquad (2\text{-}26)$$

Predicted correlations also exist for void fraction [26].

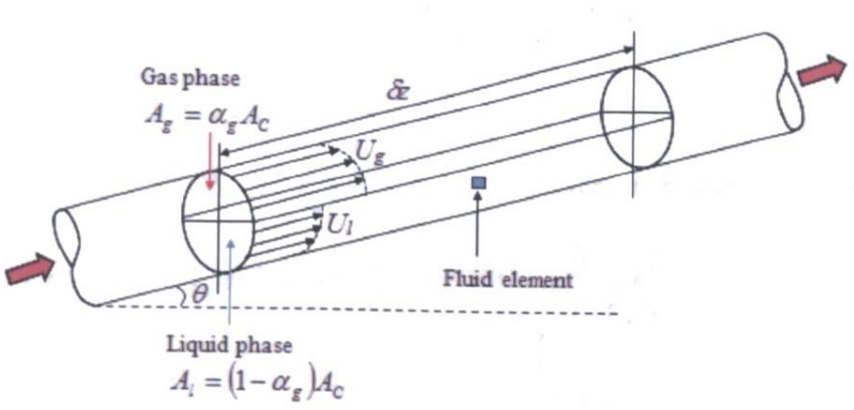

Figure 2-8 Concept of gas- liquid flow in a pipe [20]

2.5 Introduction to Gas-Liquid Flow Regimes

In a well shaken flay drink, there is the presence of gas represented as bubbles and the surrounding liquid conceiving the existence of two fluid phases. When closely observed, the lighter of the two phases, which are the gas bubbles, rise quickly and are arranged almost uniformly within the liquid. A close interaction between the gas bubbles and the liquid can also be observed. Similarly, when gas-liquid mixtures or liquid-liquid mixtures flow through a pipe, the two phases arrange themselves in a variety of patterns known as flow regimes, (Azzopardi, 2006). These flow regimes are found in most industrial processes and are of four major types: bubbly, slug, chum and annular flow patterns where the fizzy drink example may probably be classified as bubbly flow regime. The flow regimes present in vertical two-phase upward flow are discussed below.

2.5.1 Bubbly Flow

In bubbly flow• the gas phase flows as discrete bubbles in a continuous/continuum phase. This occurs at very low gas superficial velocities. The gas bubbles rise with a velocity greater than that of the liquid. Figure 2.9 (a) shows a schematic of the bubbly flow regime from the work of (McQuillan and Whalley, 1985).

2.5.2 Slug Flow

As the gas superficial velocity increases, the bubble number density increases accordingly. The largest bubbles are of the same order of size with the diameter of pipe otherwise known as "Taylor bubbles". By definition, a Taylor bubble is a constant pressure surface, whose shape is that of a cylinder bounded on top by a bullet shaped nose and at the bottom by a distorted flat tail, (Chen& 1997). Leading and trailing Taylor bubbles are separated by structures similar to bubbly flow beneath them commonly known as liquid slugs. Also, contributing to its formation is the presence of a downward moving thin liquid film between the Taylor bubbles and the pipe wall. Figure 2.9 (b) from the work of McQuillan and Whalley (1985) shows the slug flow regime as one of the two phase gas-liquid flow regimes.

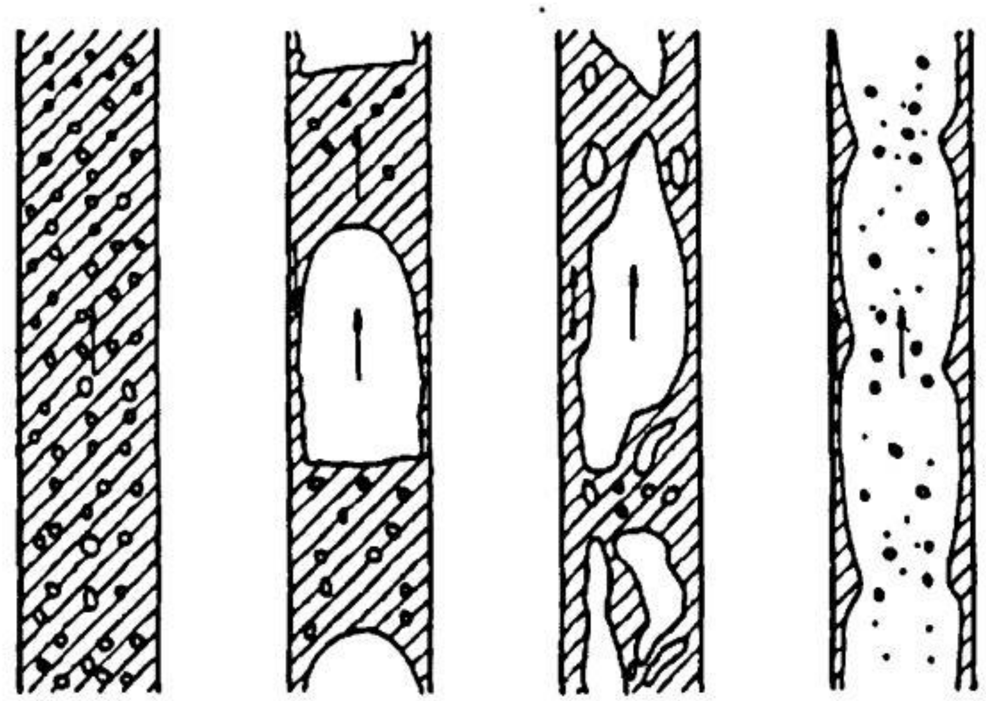

Figure 2-9(a) Bubbly (b) Slug (c) Chum (d) Annular flow regimes in upward two-phase flow in vertical pipes (McQuillan and Whalley, 1985). [1]

2.5.3 Chum Flow

Figure 2.9 (c) shows the schematic of the chum flow regime. Therefore, from the slug flow regime, as the gas flowrate increases bubbles become narrower and more or less irregular in shape. The bullet shape nose of the Taylor bubbles is suppressed to form large irregular shaped bubbles and the continuity of liquid slugs between successive Taylor bubbles is repeatedly destroyed by the high gas inertia pertaining to the flow. This occurrence causes the liquid slug to fall, thereby accumulating a volume of liquid with entrained bubbles that bridge the pipe. This is occasionally lifted by the fast moving gas phase giving an oscillatory behavior. In addition, the falling liquid film previously surrounding the Taylor bubbles is no longer observed.

Hewitt and Hall-Taylor (1970) initially identified that the behavior above is a well-marked region between slug and annular flow as shown in Figure 2.9 (c). Therefore, they were the first to ascribe the name "chum flow" to this flow behavior. In large diameter pipes, they added that this instability eventually results in the complete destruction of the slug flow thereby translating into a direct transition from bubbly to churn flow accompanied with 'churning' or oscillatory motion. However, this may depend on the viscosity of the continuous phase. Some other workers have referred to this flow regime as 'semi-annular flow', (Nicklin and Davidson. 1962). However, Hewitt and Hall-Taylor (1970) stated that the appellation, chum flow, should be given to it as it encompasses the whole flow region. McQuillan and Whalley (1983) stated that the chum flow regime is a highly disordered regime where the vertical motion of the liquid is oscillatory. Azzopardi and Wren (2014) attributed the flow regime as the least understood of alf of the flow patterns as regards vertical two phase upward flow.

2.5.4 Annular Flow

As the direct opposite to the bubbly flow regime, here, the gas flows continuously along the core of the pipe. The more or less dispersed liquid phase floss partially as liquid films along the pipe walls moving upwards in a wavy manner and as droplets in the gas core. The schematic of the flow regime is shown in Figure 2.9 (d). The liquid film may or may not contain gas bubbles and the continuous gas core which occupies most of the pipe cross-section may not contain entrained droplets. (Hewitt and Hall-Taylor, 1970). In general, the liquid film is typically uniform about the pipe cross-section.

2.5.5 Wispy-annular Flow

Mother interesting behavior that has been observed in vertical upward two phase flow is the wispy-annular flow regime. The appellation 'wispy-annular flow' was initially given its name by (Bennet et at, 1965). They stated that:

“.....wispy annular regime was characterized by the nature of the entrained phase. The phase appeared to flow in large agglomerates somewhat resembling ectoplasm."

In agreement with Bennet et al. (1965), Hewitt and Hall-Taylar (1970) stated that the entrained phases is agglomerated into large lumps or 'wisps' and the size of these 'wisps' are dependent on the gas velocity. This is because when the latter increases, the size decreases, Hewitt and Hall-Taylor (1970) added that this regime may occur as a result of the breakdown of slug flow at high mass velocities. In this velocity range the behavior below the large gas bubbles tend to become unstable and a frothy linger is formed around the bubble axis. As the velocity increases, annular flow is entered but the "fingers- still exist and require a finite distance to breakup. From another perspective, as a result of the instabilities of the shear and gravity forces which develop at the gas-liquid interface, this forms liquid structures in the core. Hewitt and Hall-Taylor (1970) added that the wispy-annular flow regime can be entered as a result of droplet coalescence when the gas velocity is reduced for any reason.

Hemandez-Perez et al. (2010) agreed with Hewitt and Hall-Taylor on the fact this regime occurs at high flow rates in what should be the typical annular flow. They added that the annular flow regime is made more complex by the presence of wisps in the gas core. They identified them from visual observation of which they appear as dark patches when viewed through a transparent pipe wall. Hawkes et al. (2001) pointed out that the views of these wisps through the transparent pipe wall are blurred by the wavy liquid film interface.

McQuillan and Whalley (1985) suggested that the annular flow pattern can be subdivided into two regimes: wispy-annular flow and non-wispy annular flow. In corroboration to the above workers. they stated that the wispy annular flow occurs as a result of the agglomeration of droplets in the gas cow to loan streaks of liquid or wisps. Figure 2.10 shows the wisps observed by (Hewitt and Roberts, 1969 and Hernandez et al., 2010)[1].

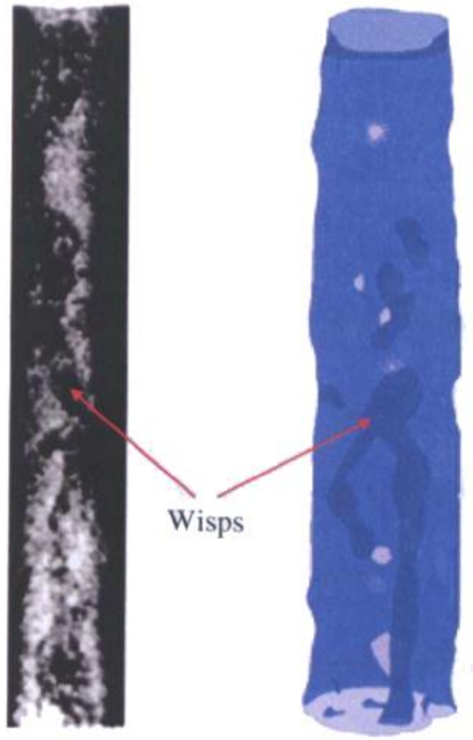

Figure 2-10 (a) Wisp recorded by X-ray photography by Hewitt and Roberts (1969) 32mm pipe diameter (left) (b) Type of wisps front Hernandez el al. (2010) 67mm pipe diameter revealed by wire mesh sensor studies (right).[1]

Froth and mist flow are other regimes that exist in venical two phase gas-liquid flows. The froth flow is covered partly by chum flow and annular flow. The mist flow regime is defined as one of complete dispersion of the liquid in the gas phase.

2.5.6 Train Slug Flow

Steady state operation of two-phase flow in pipes usually means that the flow rate of liquid and gas are constant. As a result, the conditions at any point in the pipe remain constant; namely, the flow pattern, average void fraction, average pressure drop and average local flow rates do not vary with time. The term average is used here because two-phase flow is seldom a truly steady state flow and averaging values are used over a time period characteristic of the flow pattern. A typical example is the slug flow pattern, for which average values are taken during one or a few slug passages.

2.5.6.1 Terrain Slug Formation

However, under certain situations a steady state operation is not possible. For example, when a subsea line with downwards inclination ends with a vertical riser to a platform, or when a pipe is laid in a hilly terrain, under certain conditions the lower section of the pipe accumulates liquid and blocks the gas passage. The gas upstream is compressed until it overcomes the gravitational head of the liquid, thereby creating a long liquid slug that is pushed in front of the expanding gas upstream. Under such conditions a cyclic operation is obtained, termed severe or terrain slugging [26].

For Example when a pipeline pass through a steep incline before reaching to platform or when a pipe is connected to a vertical pipe to deliver the multiphase flow to separators, heat exchangers or other surface facilities. In this situation liquid is accumulated in lower part of pipe and blocks the gas way to pass through the riser. The blocked gas is compressed and its pressure increase until it overcomes the liquid gravitational head. Then the compressed gas pushes a long liquid slug through vertical pipe with increasing velocity.

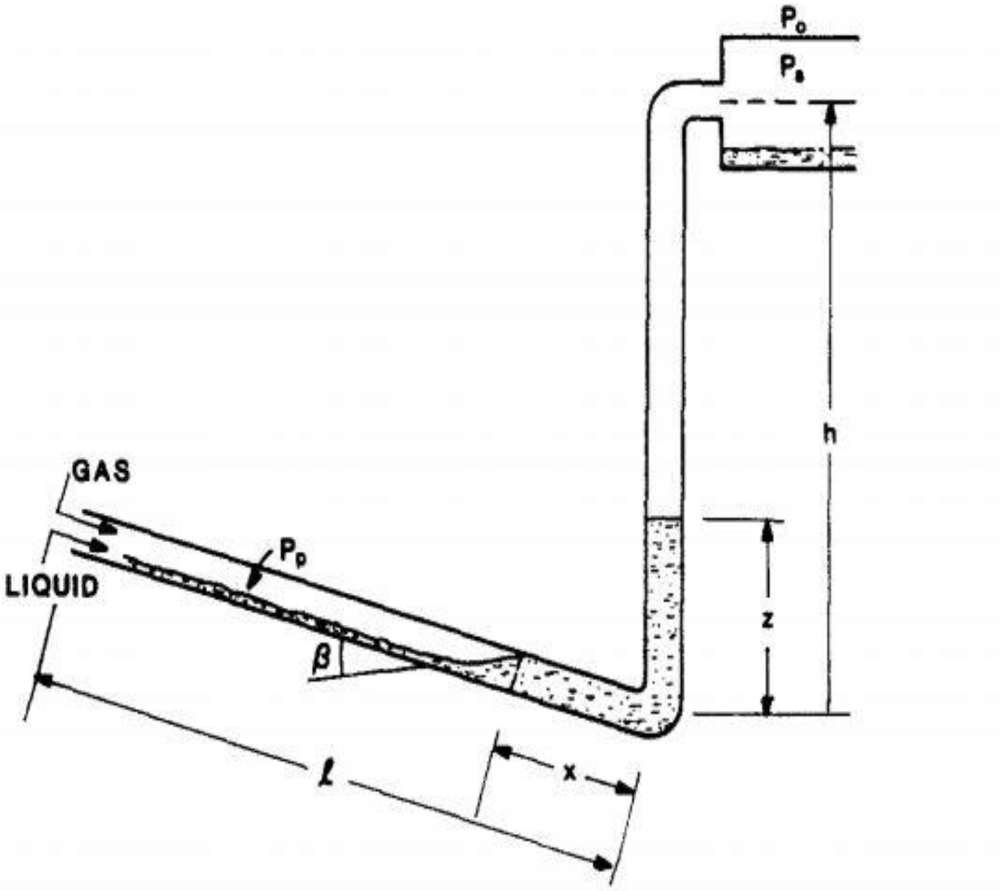

Figure 2-11 Slug formation [26]

Severe slugging is considered to be an unstable flow regime in the sense that it is associated with large and abrupt fluctuations in the pipe pressure and in the gas and liquid flow rates at the outlet. The process of severe slugging formation can be described as taking place according to the following steps. The first step is the slug formation (figure 2-11). In this step liquid entering the pipeline accumulates at the bottom of the riser, blocking the gas passage and causing the gas to compress. When the liquid height in the riser, z reaches the top of the riser. z= h. the second step of slug movement into the separator starts (figure 2-12) [26].

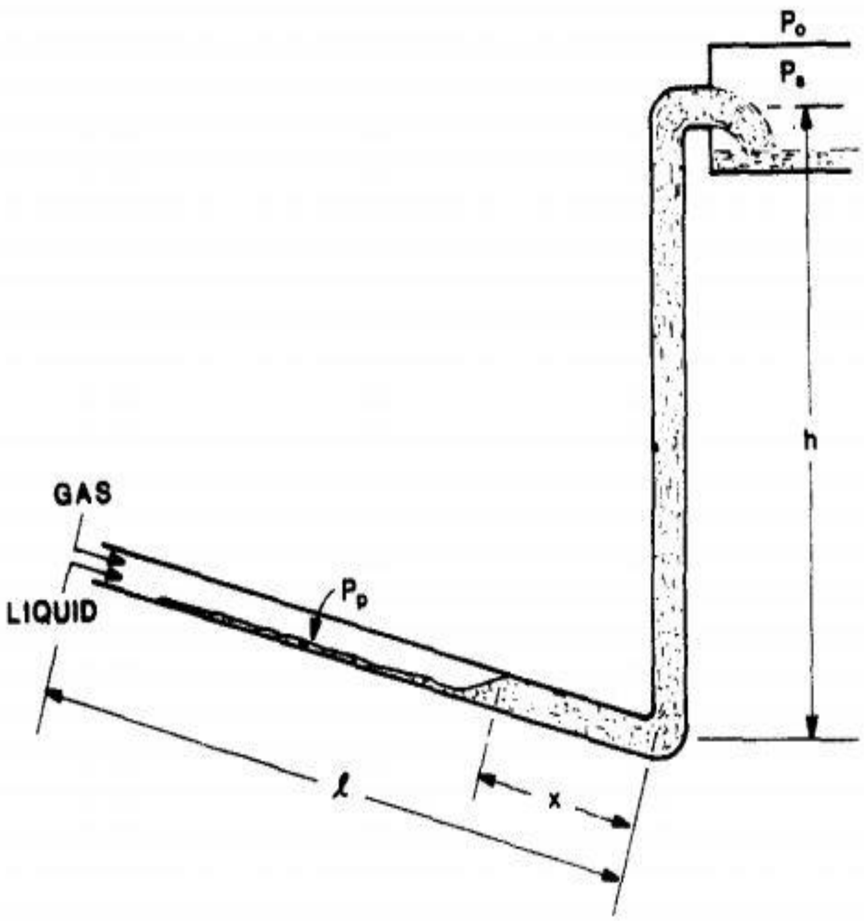

Figure 2-12 Slug movement into the separator [26]

After the gas that is blocked in the pipeline reaches the bottom of the riser, the liquid slug continues to flow into the separator with a rather fast velocity, termed blowout (figure 2-13). In the last step, figure 2-14, the remaining liquid in the riser falls back to the bottom of the riser and the process of slug formation starts again. The severe slugging pattern is typical of relatively low liquid and gas flow rates. It requires that the flow pattern in the pipeline be stratified. In addition, it requires that the liquid reaches the top of the riser pipe before the gas reaches the bottom of the riser during slug formation [26].

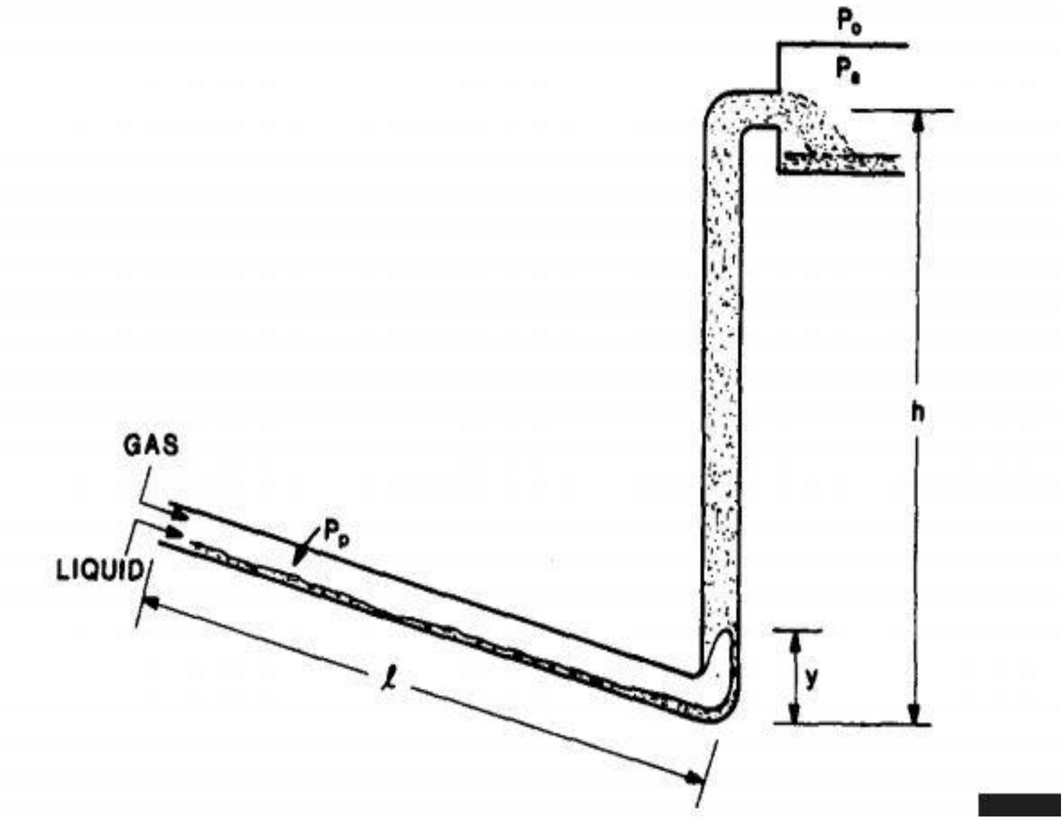

Figure 2-13 Blowout [26]

The latter condition can be calculated using the Schmidt et al. model (1980). A simplified version of the Schmidt model is used here to determine the flow rate of liquid and gas at which severe slugging will not occur. Severe slugging is an undesired phenomenon. One of the methods of alleviating severe slugging is by increasing the separator back pressure (Yocum 1973). Choking the flow (Schmidt et al. 197913. 1980) was also found to alleviate severe slugging with minimal increase in the pipeline pressure (for the same flow rates of liquid and gas) [26,30].

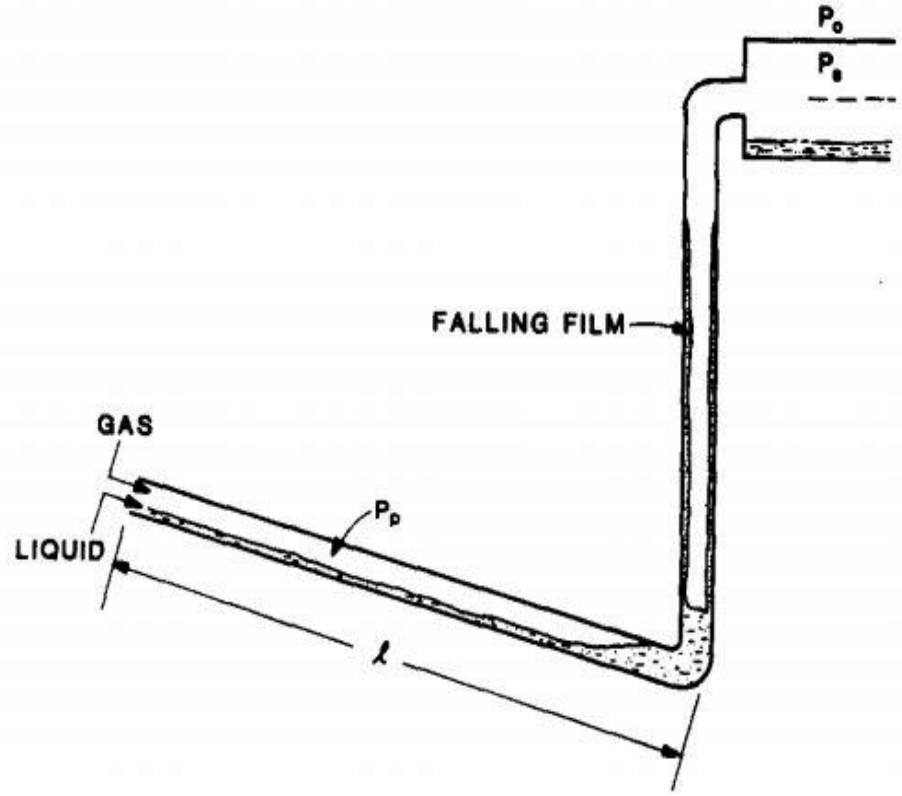

Figure 2-14 Liquid fallback [26]

Once severe slugging was eliminated, a steady state operation was achieved as shown in figure 2.15. In this steady state operation the pipeline is in stratified flow while the riser is in bubble or slug flow. The pressure of the pipeline remains constant and the liquid does not penetrate upstream into the pipeline to form the long liquid slug. In spite of the progress achieved in eliminating severe slugging, it scans that this process is not well understood and the conditions under which severe slugging can be transformed into steady state flow are still not clear.

The statment that *"the process in which severe slugging has been eliminated successfully has been repeated often enough to prove the value of choking as probably the most practical method of eliminating slugging"*

(Schmidt et al. 1980), reveals the need for a better understanding of this process. In this work we examine the conditions under which severe slugging will take place and find under what conditions and how severe slugging could be eliminated and transformed into steady state operation. Furthermore, the stability of steady state operation is analyzed and the conditions under which steady state operation will take place are established [26].

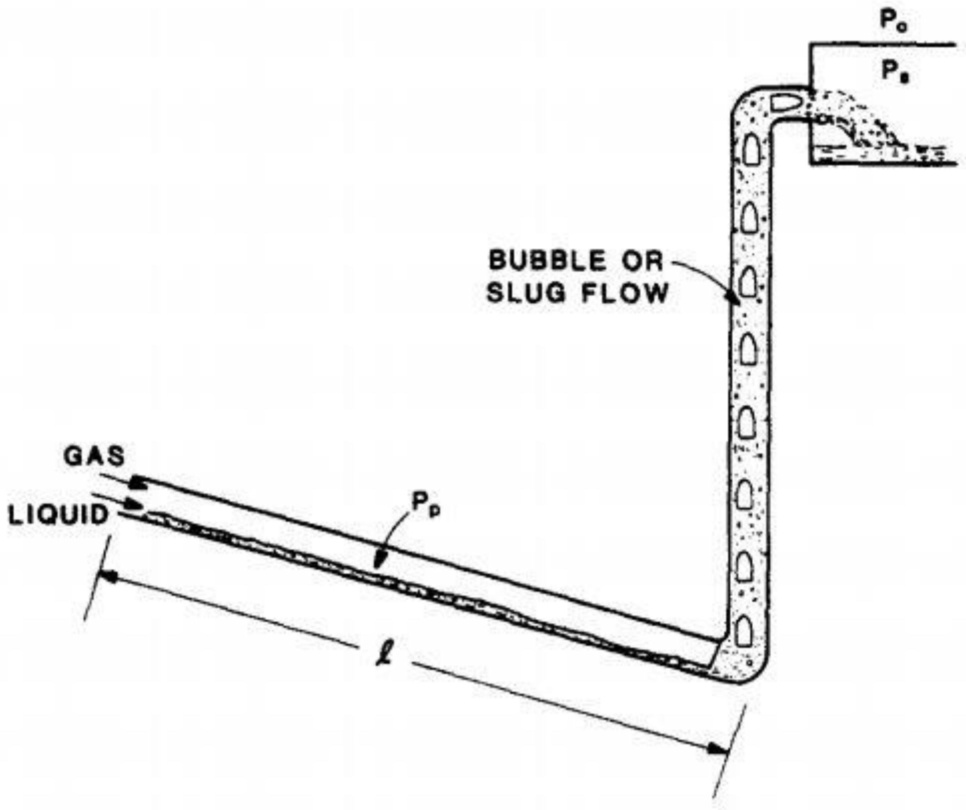

Figure 2-15 Steady state operation [26]

2.5.6.2 Terrain Slug Studies

Terrain Slugging is an undesirable phenomena. extensive research have been tried to investigate and create a reliable model to predict its behavior and characteristics like slug frequency, average pressure drop and average void fraction. For the first time Yocum (1957) mentions that the riser systems have suffered losses in flow capacity because of poor design and the main reason of design uncertainty is sever slug happening in production system.

After that Griffith and Wallis (1961) offered their flow pattern map for vertical two-phase which is connected to a horizontal pipe at the bottom. That flow map shows most of gas and liquid flowrates create slug flow pattern in the riser system and indicates that intense of sever slug problem in risers [11,30].

Yocum (1973) used a friction loss calculation and Flanigan's (1958) elevation correction to present a method for calculation pressure losses in riser. While both of these calculations were established for pipelines, their combination had not accurate result for riser system. Schmidt et al. developed in 1980 a new model to determine the flow rate of liquid and gas at which Terrain slugging will not happen. Taitel (1985) investigated the stability of sever slugging. He presented new solutions to control the sever slugging and limit its creation flow rates boundaries. Zabaras (2000) and Schulkers (2011) proposed two empirical correlations for slug frequency calculation. Their correlations predict the slug frequency more accurately than other methods studied [24,26,31,32].

Many studies are done over different pipeline shapes which are dealing with sever slug problem. A famous example of tis shapes is hilly terrain pipeline that consists of connected horizontal downhilly inclined and uphilly inclined pipe sections. Zheng (1994) proposed a slug-track model that follow the behavior of individual slugs in a hilly terrain pipline. This model considered two cases: when a steady slug flow maintains its identity and a more complex case when new slugs are created and vanished. Zheng (1995) conducted many experiments to understand slug flow behavior in hilly terrain pipelines He observed complex physical phenomena like variation of slug length along the pipe and persistence existence of slug flow in downhill sections. After that De Henau (1996) suggested a transient two-flow model that is validated foe conditions of terrain –induced slugging. New correlations for drag coefficient and virtual mass force were presented in this model. Taitel (2000) extended Zheng model and developed a new model to predict the slug behavior in low and top elbows [7,27,33,34].

Also the effect of compressibility was conducted in that model. In 2002 Al-Safran developed a transient slug tracking model based on quasi-equilibrium formation to calculate maximum slug length and accurate slug length distribution. Also Al-Safaran (2005) reported updated experimental data about slug flow characteristics in hilly terrain pipelines to improve the current phenomenological understanding. Yang (2017) used volume of fluid model and RNG k-ε turbulence model and presented a new numerical model to simulate liquid slug formation in a hilly terrain pipeline. He discussed about influence of pipe geometric structure and flow condition on liquid slug formation in his work [2,29].

Several experimental studies have appeared recently which evaluate multi-phase flow through bend with different angle. Omerbere-lyari and Azparandi (2007) started a wide experimental investigations on vertical riser and collect appropriate data series. In their study a wire mesh Sensor was used to obtain size distribution data over slug behavior. Saidj et al. (2014) tested air-water phase flow through vertical 90 degree bend and reported their conclusions over different flow regimes that may occur in a riser. They have employed ring-shaped plate conductance probes to measure their desired parameter. Mokhatab (2007), Loannou (2012) showed accuracy of OLGA codes to predict slug flow behavior in real field facilities especially offshore platform risers [14,21,23].

2.6 Flow Pattern Maps

The only way to represent results of observation of flow patterns described those is to plot them on a graph where the x-y coordinates are represented by the gas and liquid superficial velocities respectively of the two phases. When the observations arc recorded, transition lines am drawn on this graph to represent the extent of the boundaries between the flow regimes. From general consensus, this is known as "flow pattern map".

These flow pattern maps can be of two forms. (McQuillan and Whalley, 1985). They are given as: (a) Experimental flow pattern map (b) Theoretical flow pattern map. This will be discussed in the sections below.

2.6.1 Experimental Flow Pattern Map

These flow pattern maps arise from experimental work done by researchers for a particular fluid pair and pipe geometry. Baker (1954), Hewitt and Roberts (1969) and Taitel et al. (1980), proposed their respective coordinate systems as shown in Table 2.3 below. Taitel et al. (1980) concluded that all the flow patterns in Figure 2-16 cannot be represented by a single coordinate pair. However. Weisman (1979) have identified different scaling parameters that may be used to overcome this problem. An example of an experimental flow pattern is shown in Figure 2-16 from the work of Hewitt and Roberts. (1969) [26].

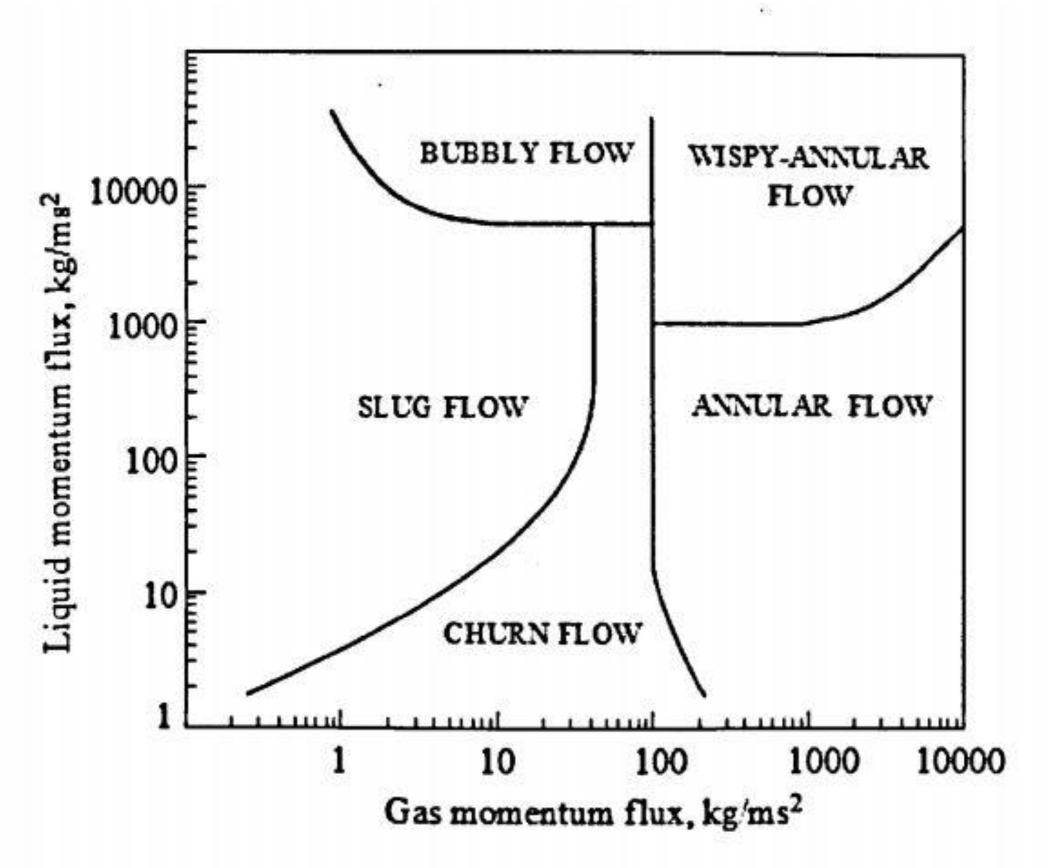

Figure 2-16 Experimental flow pattern map of Hewitt and Roberts, (1969) [27]

Table 2-3 Review on flow pattern map

Author	Year of Publication	Map Coordinates	
		X-axis	Y-axis
Baker	1954	U_{gs}	U_{ls}
Hewitt	1969	$\rho_g U_{gs}^2$	$\rho_g U_{ls}^2$
Taitel et al.	1980	U_{gs}	U_{ls}

2.6.2 Theoretical Pattern Map

As an alternative to the experimental flow pattern maps, previous workers such as Taitel et al. (1980) and Mishima and Ishii (19134) obtained theoretical flow pattern maps by initially considering the conditions necessary for the existence of each of the flow pattern. This basis allowed them to postulate mechanisms by which the transitions between the various flow patients might occur. Afterwards, these transitions were modelled to produce a series of equations. Therefore, when the phase physical properties and pipe diameter are known, this enabled the flow pattern boundaries to be calculated. Figure 2-17 shows an example of a theoretical floe pattern map by Taitel et al., (1980) [27].

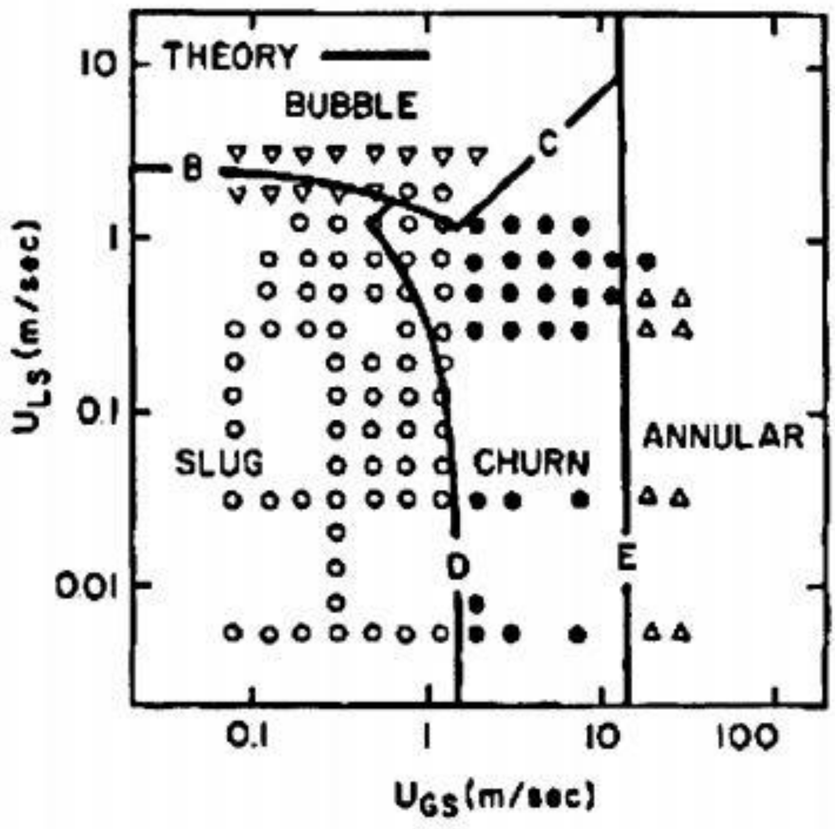

Figure 2-17 Theoretical flow pattern map of Taitel et al., (1980) [27]

CHAPTER THREE: EXPERIMENAL ARRANGMENTS

3.1 Overview of Experimental Facilities

All experiments were carried out on a two phase flow set-up within the Research Center of the Ahwaz Faculty of Petroleum. The experiment set-up is shown in Figure 3-1. Details of the experimental facilities configuration and their function are presented in following. This experimental three phase loop has consisted of two connected section, 1.Metering Section and 2.Riser Test Section 3.GLCC separator Test Section.

Figure 3-1 Experimental set-up

3.1.1 Mixer Section

This part of the loop is consists of two parallel single phase liquid and gas stream lines. For Gas line, Air which is selected for gas fluid, is provided by an air compressor with capacity of 150 ft^3/min at 80 psi. The gas flow in the line is controlled by a control valve and is measured by a rotameter which measures from 0.1 up to 1 m^3/hr. the liquid phase is supplied from a 80-liter storage tank at atmospheric pressure, and pumped to the liquid line by a centrifugal pump at maximum rate of 120 lit/min. Similar to the gas phase the liquid phase is metered by a rotameter with domain of 2 to 22 lit/min and controlled by a control valve. Also in every knee and downstream of any facilities a ball valve is located for safety and better control over the loop system. In addition two check valves are used after any feeder to prevent any backflow.

Figure 3-2 Mixer section

Figure 3-3 Liquid pump

Two single-phase gas and liquid stream are combined in the mixing part as shown in Figure 3-5. Then they delivered to the Riser Test Section. The mixing part is consisted of a section of a hose wall with 45 holes with 1 mm diameter spaced equally in 3 columns over a length of 100 mm. The water was introduced into an annular chamber surrounding this section of hose to create a better circumference mixing effect. As it shown in figure 3-6, a static mixer which is consisted of three spiral plates, is placed counter wise serial before test section for uniform mixing. Then the two-phase flow regimes are created due to the flow rates of fluids in the Riser Test Section. After that the two-phase mixture is separated by a conventional separator. The gas is escaped to the atmosphere and the water is returned to the storage tank to complete the cycle.

Figure 3-4 Air comprressor

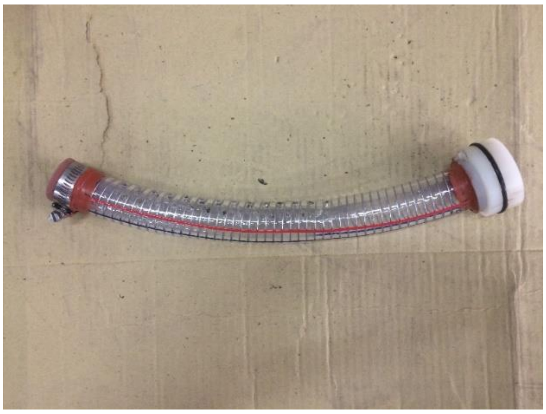

Figure 3-5 Mixer hose

3.1.2 The Riser Test Section

The riser test section is divided to three parts: 1. horizontal pipe 2. bend 3. vertical pipe. Two transparent pipes with internal diameter 24.5 mm and 2 m long are used as horizontal and vertical pipes. For angle sensitivity test the horizontal pipe is allowed to incline between -7 to 7 degrees. Also a transparent hose is employed to connect these two pipe as bend. In addition the average pressure drop over the vertical pipe is measured by a pressure transmitter that its pressure sensors are located in the beginning and end of the pipe. Density and temperature of the liquid phase are also measured and recorded by the system.

Figure 3-6 Static mixer

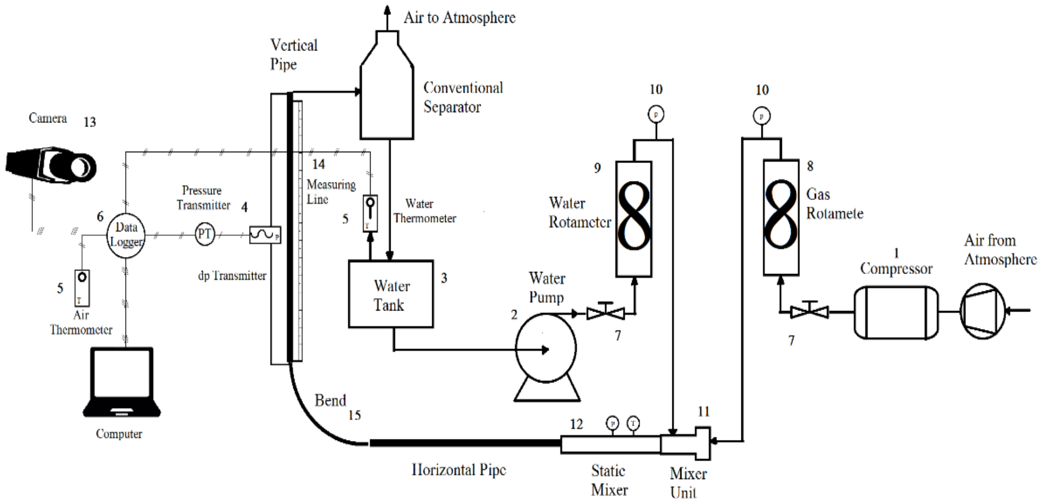

Figure 3-7 Riser test section schematic

3.1.3 GLCC Test Section.

The test section consists of a cylindrical cyclone separator, as shown in Fig. 3-6. The test section is divided into four parts:

1. the inlet section;
2. the cylindrical cyclone body;
3. the gas leg, which includes the liquid carry-over trap;
4. the liquid leg;

3.1.3.1 Inlet

The inlet of the cylindrical cyclone is an inlet pipe section, 3-in. diameter, connected to the cylindrical cyclone through a sector-slot/plate configuration with a nozzle area 25% of the inlet pipe cross-sectional area. The cylindrical cyclone can be configured with a single inlet or a dual inlet by using the appropriate inlet valves.

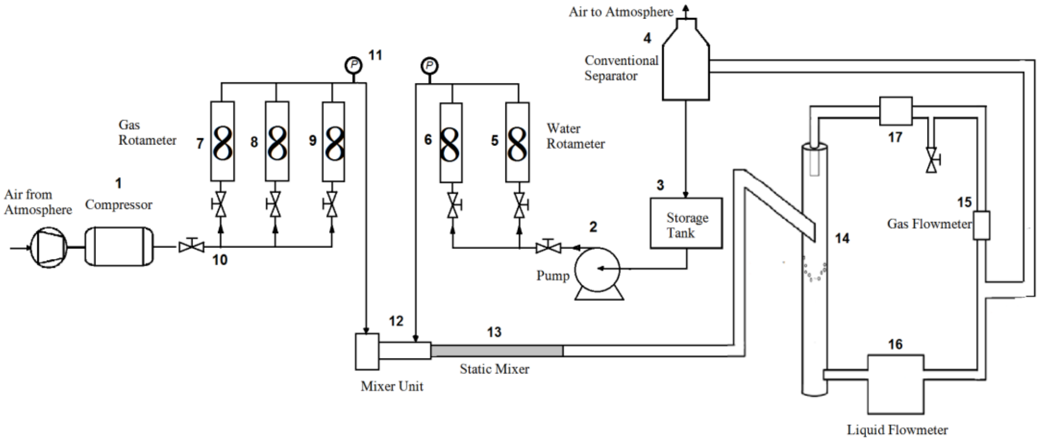

Figure 3-8 GLCC test section schematic

3.1.3.2 GLCC Body Gas and Liquid Legs and Liquid Trap

The cylindrical cyclone body is 3 in. in diameter and 7.4 ft tall. The gas leg is 2-in. in diameter, and includes a the liquid trap. A schematic of the liquid trap is shown in Figure 3-9. The liquid trap is a 6-in. pipe section expansion with a 2-in. pipe connection to the gas leg. It allows accumulation and measurement of liquid carry-over for conditions beyond the operational envelope. A mesh is installed at the exit of the liquid trap to trap fine liquid droplets in the gas stream. On the other hand, the liquid leg consists of a combination of 2-in.

Figure 3-9 Liquid trap

Figure 3-10 GLCC Test Section

3.1.4 Data Acquisition

By the advantage of transparency of riser test section all the process of experiment through the riser are recorded with a high speed camera. The camera was able to record 160 fps videos. Later the recorded videos were investigated to define the behavior of gas lock in riser. In each frame of recorded video position and length of any liquid slugs and Taylor bubbles could be determined as shown in Figure 3-11. Also velocity of any slug units and Taylor bubbles could be calculated by knowing of each frame time. AVS Video Editor software was used to analyze the slow motion videos.

Figure 3-11 Recorded Frame from Riser Test Section

3.1.4.1 Method of Data Acquisition

The usual approaches to collect physical behavior for adjusted gas and liquid flow rates can be placed in one of principally three categories. In X-Ray Radiography method high frequency X-Ray is emitted through the gas-liquid mixture flow in particular pipe section. Then passed rays are detected by an X-Ray detector which is placed on the other side of pipe.

Output signals from detector define amount of amount of cumulated gas inside of pipe as a function of time. Thus gas specific mass can be determine as a function of gas holdup in the pipe. In the other hand in Electric or Magnetic Probe method different conductance probes have been used to measure the conductivity of gas-liquid mixture flow. A conductive rod in center of pipe, a mesh in the pipe section or a sleeve over inner pipe surface are typical kinds of probes to quantify conductivity as a function of time. An

Oscilloscope assistances to analyze the quantities for further flow behavior studies. Electrical Capacitance Tomography (ECT) and Wire Mesh Sensor (WMS) are noble examples of this prime method category [23].

But in High Speed Film-taking, the proper method has been to visually observe and multi-phase phenomena record through a test section window when transparent apparatuses have been used in experiment set-up. Recent technology offers super high speed cameras which can take a frame in micro second with high resolution. Also with simple video editor software there is possibility to evaluate the recording videos. By this method multi-phase structural properties of any bubble or liquid droplets and liquid slug can be investigated simply.

Bubble formation or deformation in its shape, phase’s movement along each other through various shape sections and changes in flow regime are visible and investigable in this method. Figs show that how this method works in real experiment situation. For better evaluation a colorant material like Bromothymol Blue is added to the water. This helps to preference between phases and enhanced observation.

3.2 Procedure of Experiment

3.2.1 GLCC separator Experiments

As mentioned before a GLCC consists of different parts such as a two-phase inlet, a vertical column body, two horizontal outlet for gas and liquid single phase flow. Quantity of diameter and length of each part can effect on performance of the GLCC and its flow domain. Field situation describes the operational domain so an applicable and suitable separator must be designed for that condition.

Thus a test GLCC separator is designed and built in laboratory to determine its domain and understanding its function. The best operational domain is where the equilibrium liquid level placed below the inlet and between 1 L/D and 3 L/D of separator column. If it pass the inlet it causes liquid carry over and if it settles below the 3 L/D it will create gas carry under in the separation.

Thus the equilibrium liquid level was measure for different range of liquid and gas flowrates. In this work the gas superficial velocity was set for 1, 100 and 200 ft. per second and for each gas superficial velocity, liquid superficial velocity was changes between 1 to 11 ft. per second. After that any part of this test separator was changed and its effect on the separator operational domain was observed.

These changes are 0.5 inch reduction in inlet diameter size, 0.2 inch reduction in liquid outlet diameter size, 0.2 inch reduction in gas outlet diameter size, 0.4 ft. reduction in gas column length, 1 inch reduction in column diameter size and 4.6 ft. increment in outlet length. The changes in equilibrium liquid level proves which changes have enhancement on separator performance and which ones can couldn't help that. After writing a software based on the hydrodynamic model which is mentioned in previous chapter by Visual Basic the changes in physical dimension of GLCC are investigated.

3.2.2 Riser Experiments

The following experiment procedure was carried out to find domain and evaluate physical characteristics of gas lock phenomena in riser. Different characteristics parameters have been studied including period of repetition, Duration of Slug and Taylor bubble passage, Taylor bubble rise velocity and pressure drop in vertical column.

1. Specific superficial velocity was set for the liquid equal to 0.1867 m/s. (0.2801, 0.3735 and 0.4669 m/s for other experiment series.)

2. For a series of experiments the gas superficial velocity was set at the least equal to 0.0778 m/s and data were recorded.
3. After that, gas superficial velocity was increased to 0.1556 m/s and data were recorded again.
4. gas superficial velocity was increased to reach the following quantities respectively 0.2334, 0.3112, 0.3890, 0.4668, 0.5447, 0.6225, 0.7003, 0.7781 m/s
5. For next experiment series step 4 was repeated for other liquid superficial velocities which were mentioned in step 1.
6. After all these steps, the riser angle was changed and steps 1 to 5 were repeated for the new riser angle.

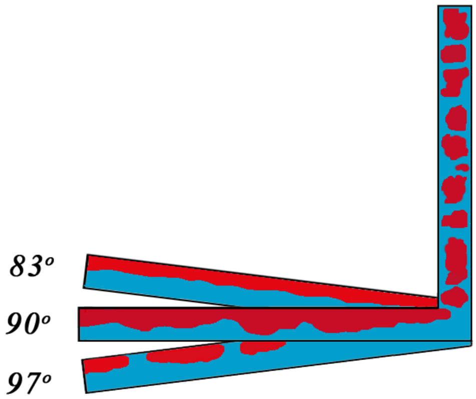

Figure 3-12Different angles of riser bend

In a specific time duration many slug units pass the riser. So, for finding the exact and correct value of physical behaviors of flow regime, Regression Analysis was conducted. The methodology of Regression Analysis, PA, was used to remove outlier points among experiment data. The mean value of the remaining data was reported in following.

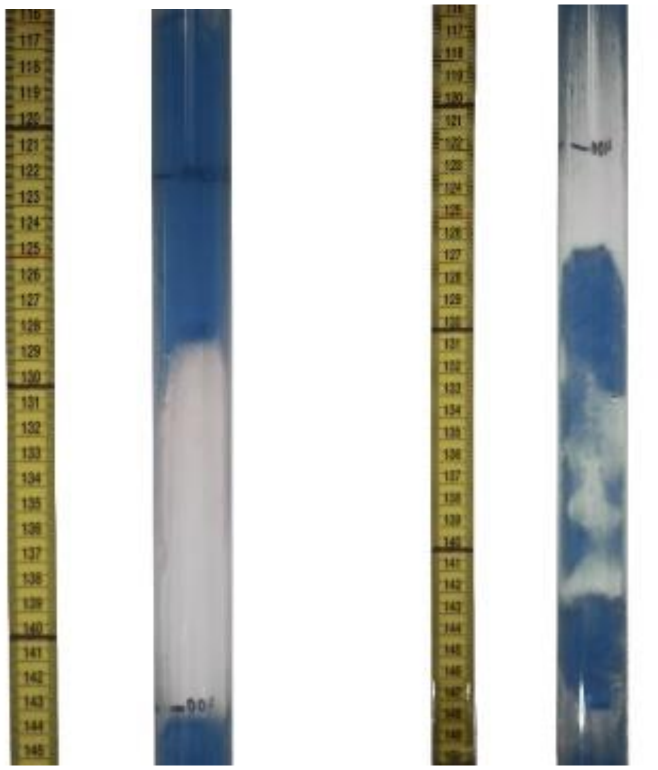

Figure 3-13 Bubble Deformation

CHAPTER FOUR: RESULTS AND DISSCUSIONS

4.1 GLCC Investigation Results

In following the effects of changes in different parts of GLCC separator are investigated and discussed about their reasons.

4.1.1 Original test GLCC

18 different gas and liquid flowrates are selected to examine the equilibrium liquid level of our test GLCC. For the test GLCC with 4.2 ft. length of body column below the inlet, the equilibrium level must stays between 1.4 to 2.8 ft. of the body column. Liquid level for some of these flowrates are placed between these domains but others are not suitable flowrates for this GLCC operational domain.

It is clear increasing in gas flowrate decreases the equilibrium level in body of GLCC. Accumulated gas in body push the liquid level down and it may cause gas carry through the liquid outlet. On the other hand increasing in liquid flowrate increases the accumulated liquid volume in the GLCC body and rises the liquid level. The tolerance of liquid and gas flowrates must stay balanced.For better GLCC operation the changes must reduce slop of these curves and make them be closer to each other. After that the changes separator could accept more gas and liquid flowrates and its operational domain is extended. Understanding the effect of these changes in physic of GLCC helps us to design a proper separator according our operational situation.

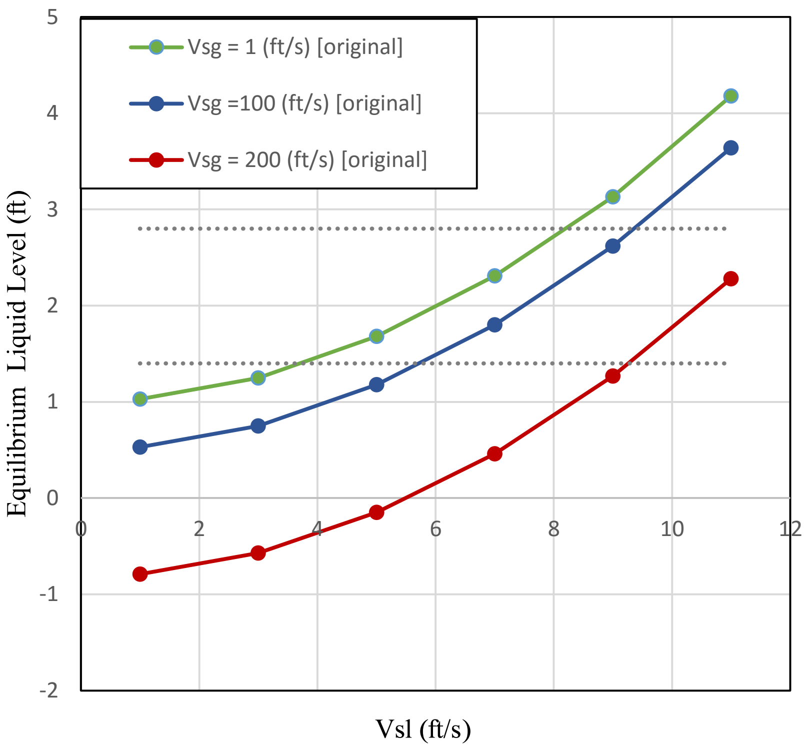

Figure 4-1 Equilibrium level liquid for original test GLCC

4.1.2 Reduction in Inlet Diameter

In this section 0.5 inch reduction in inlet diameter is evaluated. Reduction in inlet diameter make the multiphase flow stream be more effective in where it enters to the GLCC body column. The reduced inlet diameter increases the phase velocity and grows the centrifugal force. That helps the phase separation in GLCC.

In the figure 4.2 the effect of reduction in outlet diameter is illustrated. The curves become closer to each other and their slop is reduced. The hydrodynamic model can predict this change correctly. The enhanced effect is more obvious in high gas and liquid flowrates. Because high centrifugal force has an essential effect in those flowrates. This proves reduction in inlet diameter has a positive in GLCC performance and inflect its domain to accept more gas and liquid flowrates.

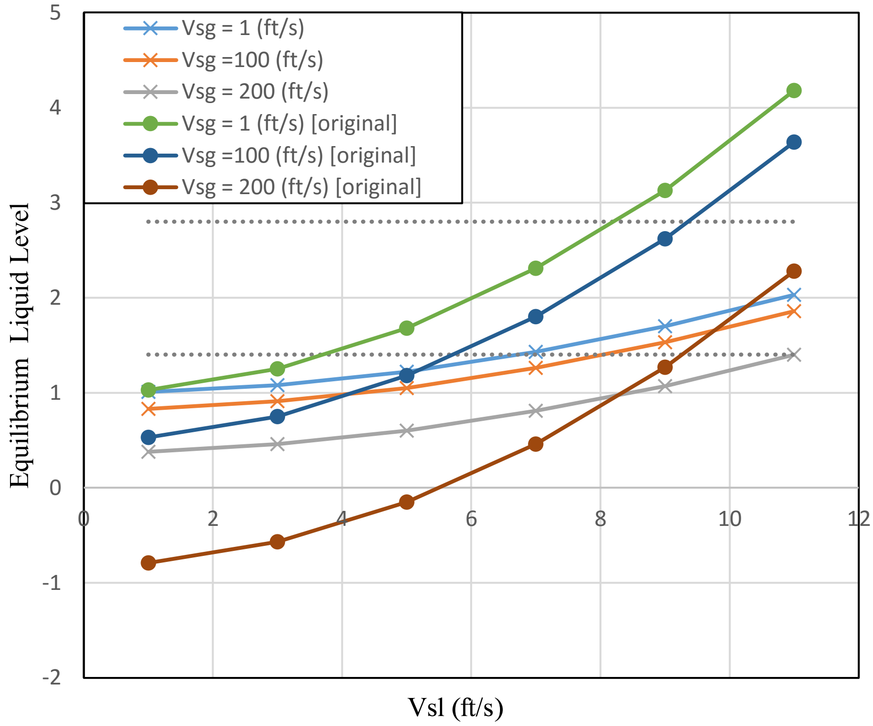

Figure 4-2 Equilibrium liquid level in test GLCC with 0.5 in. reduction in inlet diameter

4.1.3 Reduction in Liquid Outlet Diameter

As it is discussed accumulated liquid in GLCC body raises the equilibrium liquid level. Decreasing the liquid outlet diameter make the liquid volume evacuated slower than normal condition. Normal gate valve can does this responsibility to become a simple passive control for equilibrium liquid level. But it can be effective only when gas flowrate are high. Otherwise in high liquid flowrates it can cause liquid carry over in the separator system.

In the figure 4.3 0.2 inch decreasing in liquid outlet diameter is illustrated. This change helps the high gas velocity curve move and placed in the acceptable liquid level domain but this fact has a negative effect on low gas flowrates with high liquid flowrates points. Totally this change weakens the separator performance. It can be used in special situation to control liquid level.

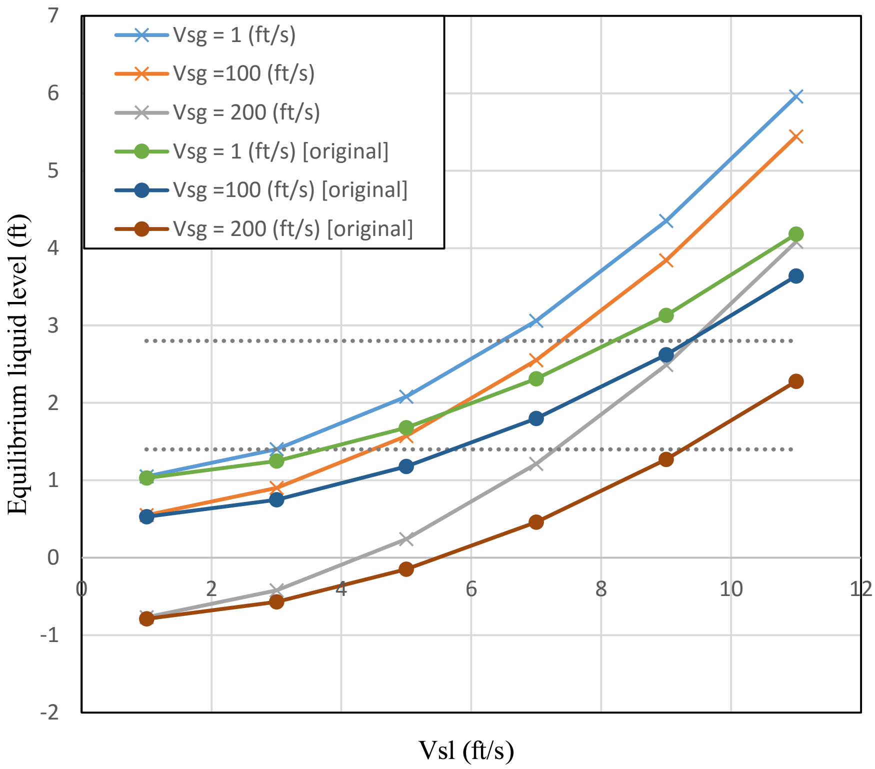

Figure 4-3 Equilibrium liquid level in test GLCC with 0.2 in. reduction in liquid outlet diameter

4.1.4 Reduction in Gas Outlet Diameter

This change increases the accumulated gas volume in the GLCC body column. As mentioned before this accumulation push the liquid level down. Again a gate valve placed on the gas outlet leg can play the role of a passive control system simply. This fact is shown in the figure 4.4. This change pushes the curves down except the low gas flowrate curve. Because in low gas flowrate there is no amount force to accumulate the gas in the separator. Of course if the diameter reduces more it can push this curve down like other curves.

It is clear that this change make the curves further in compression with normal condition and move them below their primary place. It can be useful to control the equilibrium liquid level when the separator is dealing with high liquid flowrates.

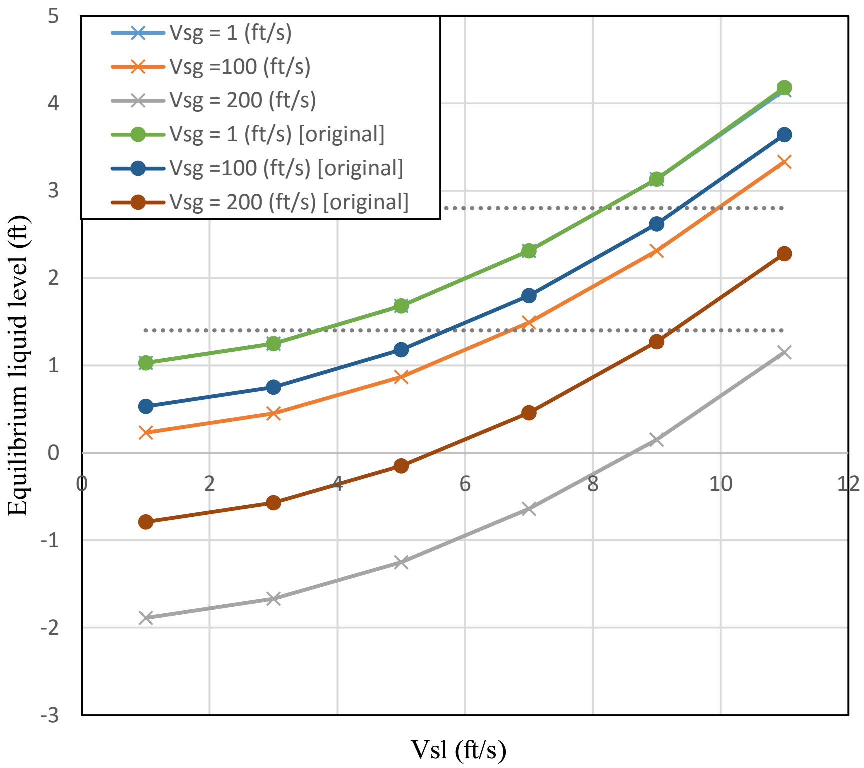

Figure 4-4 Equilibrium liquid level in test GLCC with 0.2 in. reduction in gas outlet diameter

4.1.5 Reduction in Gas Column Length

In figure 4.5 the effect of 0.4 ft. reduction in gas column length is illustrated. However there is no obvious change in curves and it indicates that changes in gas column have no effect on separator performance but it is not a quite true state. This hydrostatic model assumed any phase as a rigid phase and couldn't predict any distribution in operation. If any distribution happens it will throw liquid droplet to gas column and gas will carry that into the gas outlet. This change increase the chance of liquid carry over. So this phenomena is harsher in high gas flowrates (high gas superficial velocity) and when the equilibrium liquid level is close to the inlet. Because of puffing of gas into the liquid level this problem shows itself. However compactness of a separator always is important topic but reduction in gas column length is not recommended in any situation.

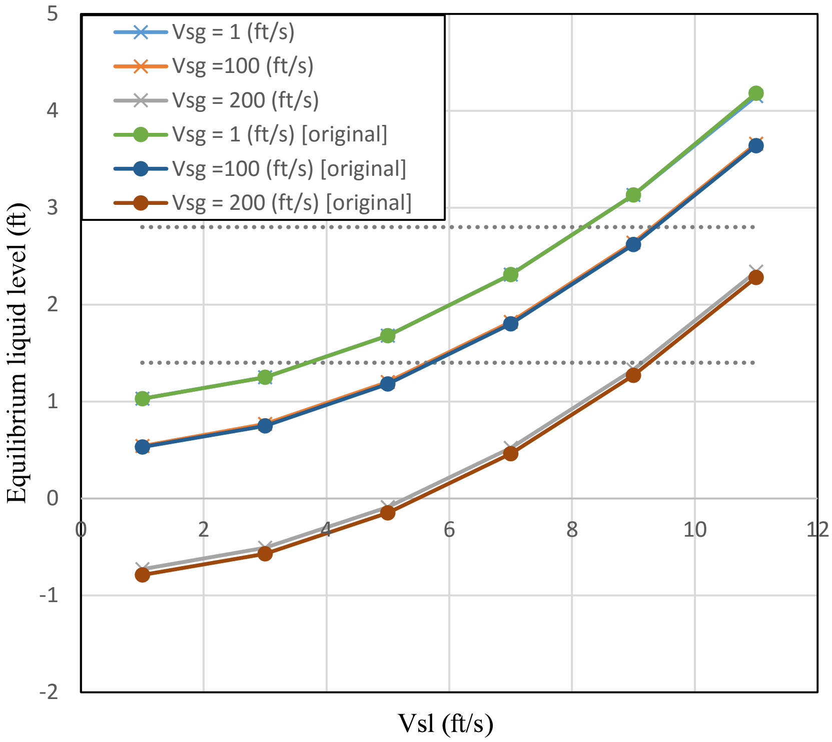

Figure 4-5 Equilibrium liquid level in test GLCC with 0.4 ft. reduction in gas column length

4.1.6 Increment in Outlet Leg Length

In this section the effect of increasing in the length of outlet leg is studied. The only thing that matters when length of horizontal outlet pipe increases is that the friction force rises and try to overcome and resistant against the flow movement in the pipelines. So this change absolutely is a negative variation in GLCC physical size.

This fact is visible in the figure 5.6. Any increasing in length of outlet leg increases the curves slop. This slop change is not seen in low gas and liquid flowrates because of low friction in pipelines. Other interesting fact is that in high gas flowrates and low liquid flowrates the gas friction has greater effect that liquid friction and increases accumulated gas volume in GLCC body and pushes the equilibrium liquid level down. Else increasing the length of outlet leg rises the equilibrium liquid level generally.

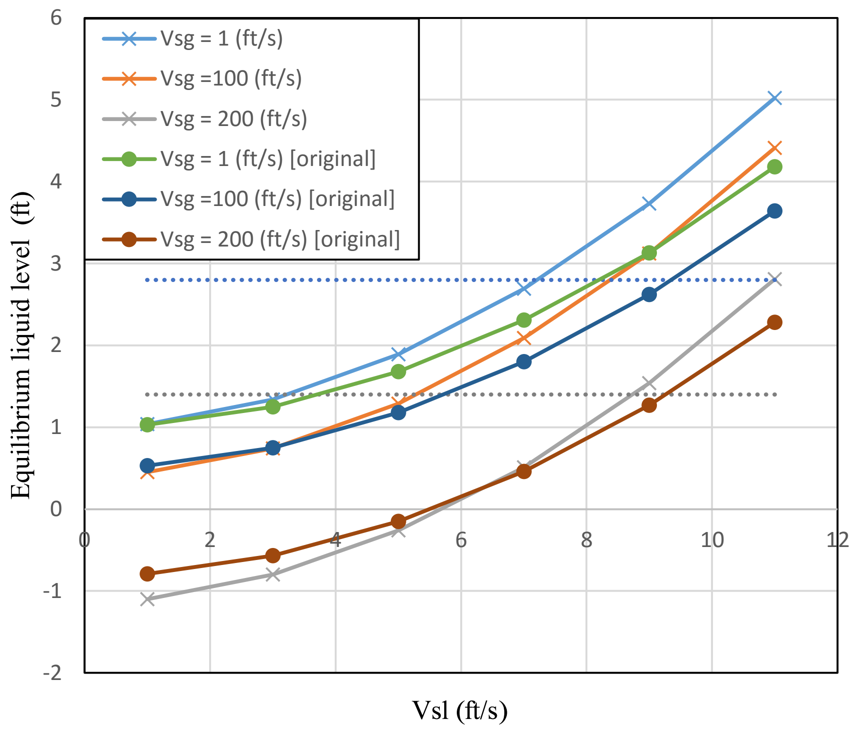

Figure 4-6 Equilibrium liquid level in test GLCC with 4.6 ft. increment in outlet leg length

4.1.7 Reduction in Column Diameter

The effect of reduction in column diameter is shown in figure 4.7. It seems any reduction has negative effect because of vortex flow nature. A vortex is defined as a circular liquid streams with narrow gas core in center of liquid streams. In high liquid flowrates, any decreases in column diameter causes the gas core penetrates more into the liquid phase. Thus in the worst condition the gas phase can reach the liquid outlet and gas carry through occurs.

The figure 4.7 shows that in high gas and liquid flowrates equilibrium liquid level is raised. Totally this change in column diameter has negative effect on GLCC performance but it is not quite sure any increasing in GLCC body diameter helps its performance. The nature of vortex flow is much complex and further studies are doing through this issue.

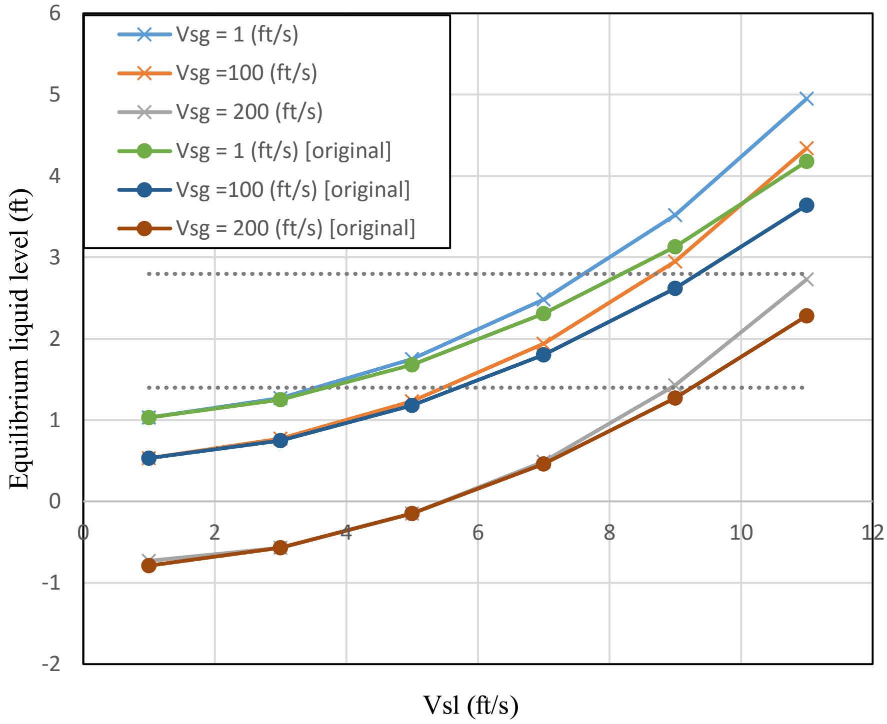

Figure 4-7 Equilibrium liquid level in test GLCC with 1 in. reduction in column diameter

4.2 Riser Investigation Results

4.2.1 Flow Map

Flow map, a map with transient lines to separate the pattern regions is the simplest way to predict flow regime for specific flow situation. The experiment results for gas lock domain is shown in Figures 4.8 to 4.10 called as regime flow map for different riser angles. Identification of flow pattern was done upon visual observation from transparent test section. The most important specification of gas lock phenomena is generation in low gas and liquid flow rate. Liquid superficial velocity has the greatest effect on limiting the gas lock region.

Experiment results state that for liquid superficial velocity as 0.4669 m/s or more there is no possibility for gas lock creation in riser .It is obvious the gas lock region which is presented as blue points is extended by decreasing riser angle. Also experiments show that after increasing only 2 degree in riser angle, the gas lock phenomena is not observed anymore in the processing flow domains. As a result angle of riser has a great role in creation of gas lock in risers.

After the threshold liquid flowrate the liquid phase gain the ability to carry accumulated gas and carry that through the bend. After that gas can pass the riser in stable situation and slug flow occurs. So this the reason why in literature mentioned sever slug flow (terrain flow) as unstable flow. Instability of this flow regimes shows its effects on fluctuation in phase, properties like pressure, flowrates, line kicks and noises. Although in slug flow these parameters are fluctuating but these are more extreme in sever slug flow.

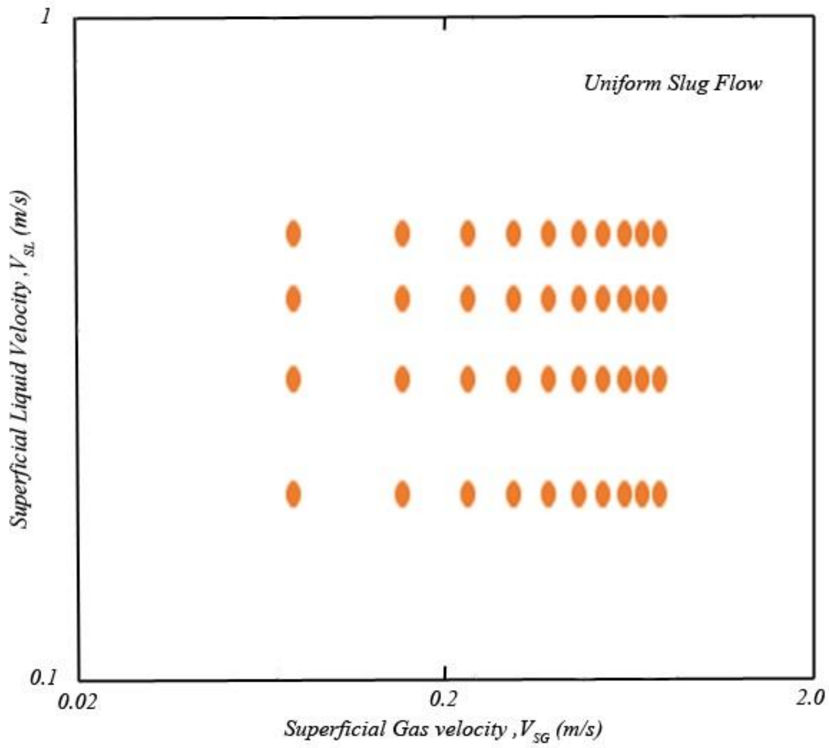

Figure 4-8 Flow map for 97 degree bend

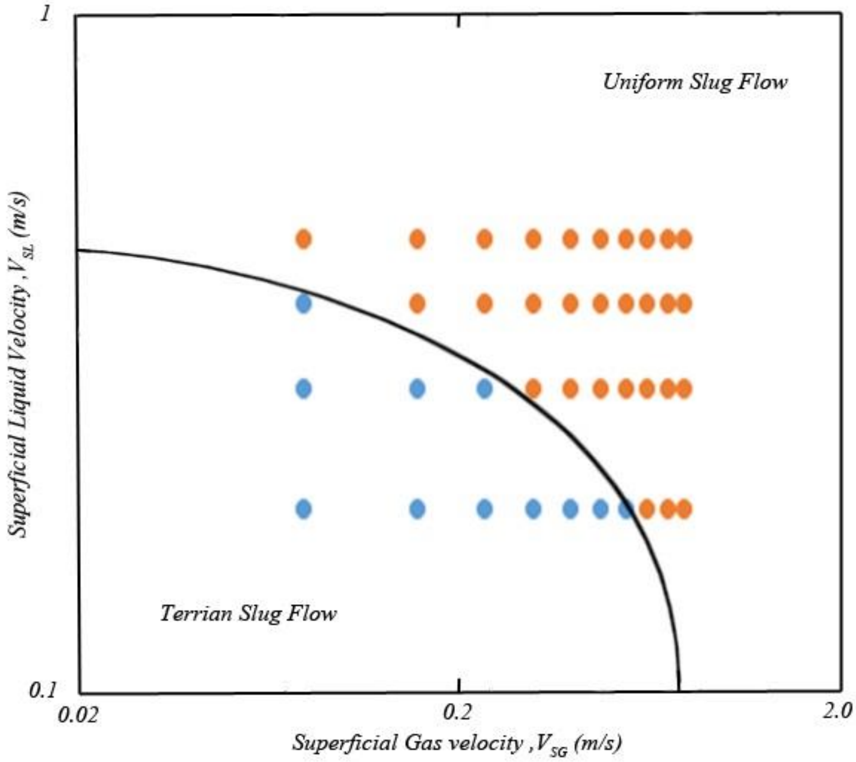

Figure 4-9 Flow map for 90 degree bend

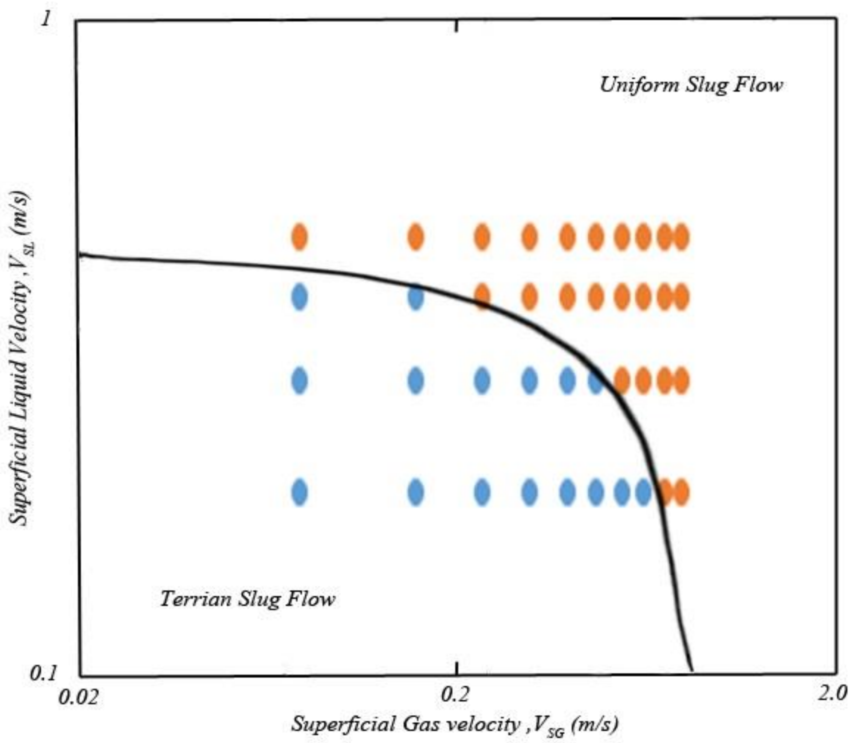

Figure 4-10 Flow map for 83 degree bend

4.2.2 Slug Period

The results of runs with creation gas lock phenomena are presented here. They were conducted according to the method described in last section figures 4.11 and 4.12 show that the period of every released gas lock decreases with increasing superficial gas velocity. Also increasing superficial liquid velocity decreases period of gas lock release. These observations are evident at both riser angles (87 and 90 degree). By comparison of results between these two angles in figures 4.13 to 4.15 it can be observed that angle has no significant effect on the gas lock release periods. The period according to Fig is varying between 15.9 to 2.0 s.

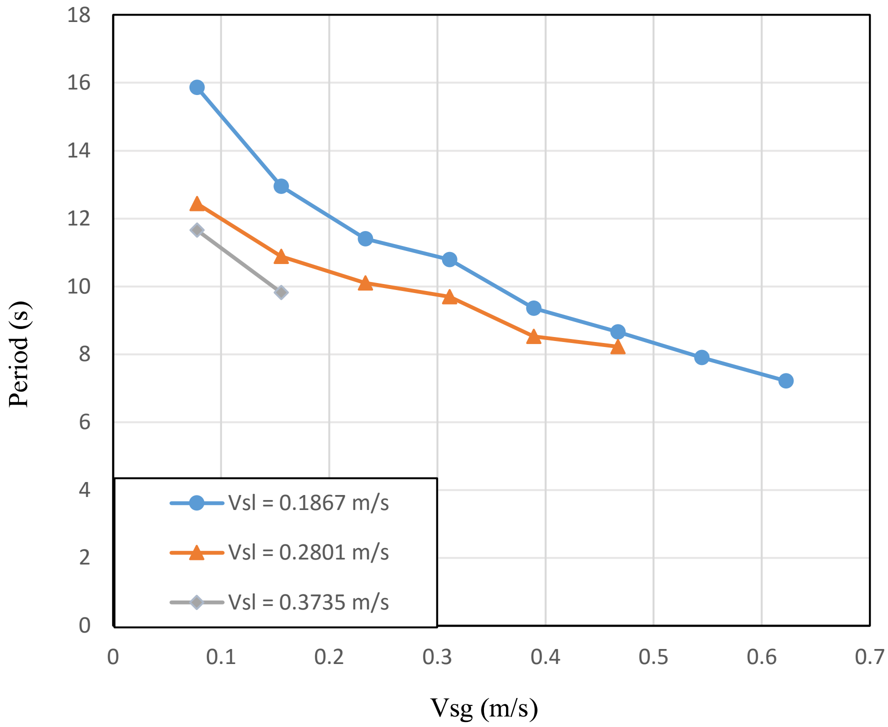

Figure 4-11 Effect of liquid and gas superficial velocity on severe slug period (83degree bend)

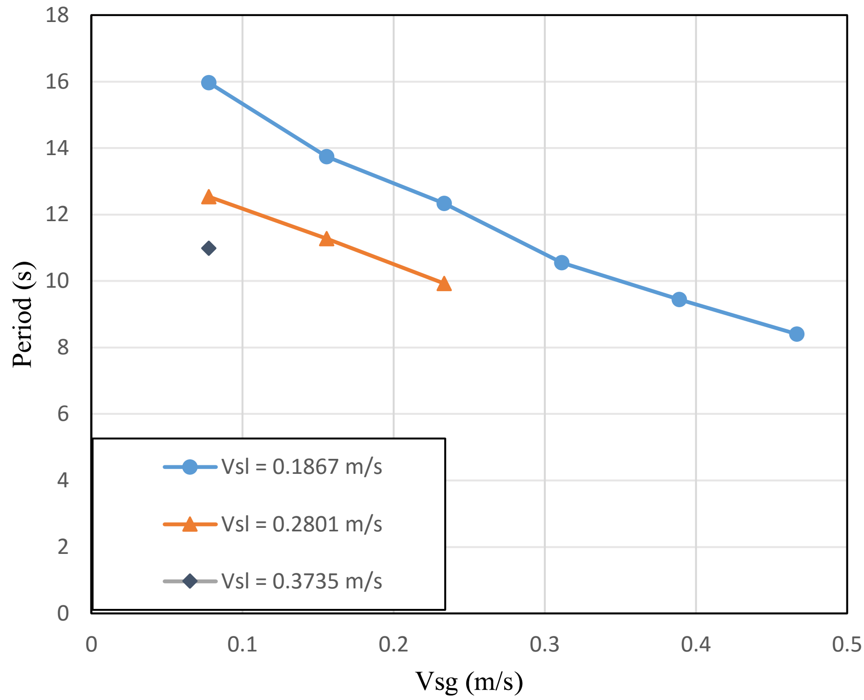

Figure 4-12 Effect of liquid and gas superficial velocity on severe slug period (90degree bend)

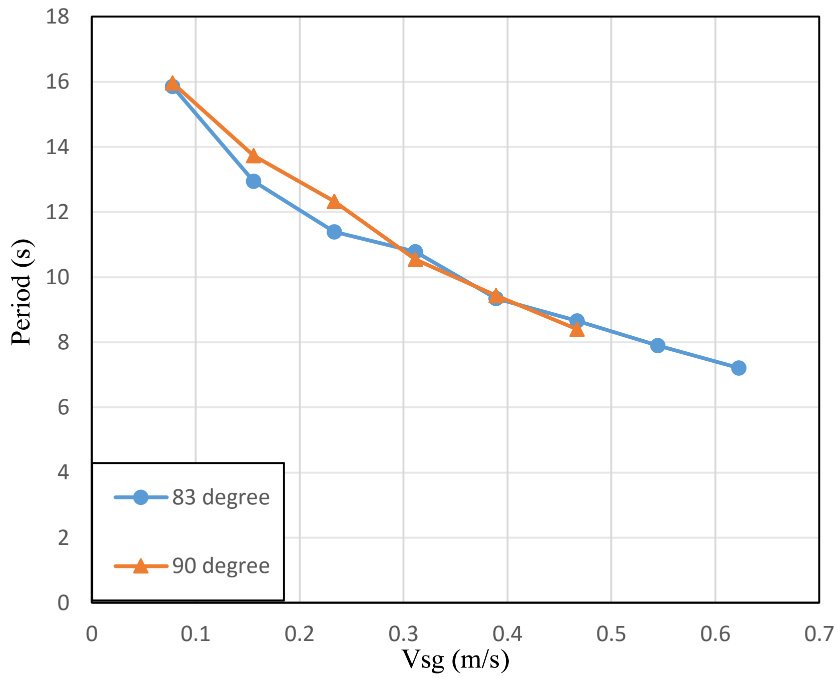

Figure 4-13 Effect of bend angle on severe slug period (Vsl = 0.1867 m/s)

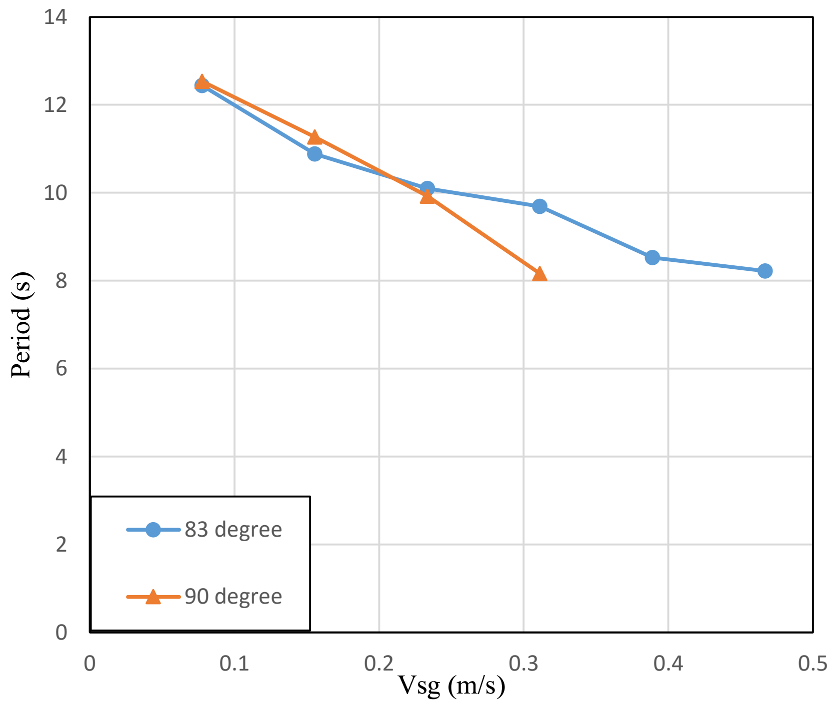

Figure 4-14 Effect of bend angle on severe slug period (Vsl = 0.2801 m/s)

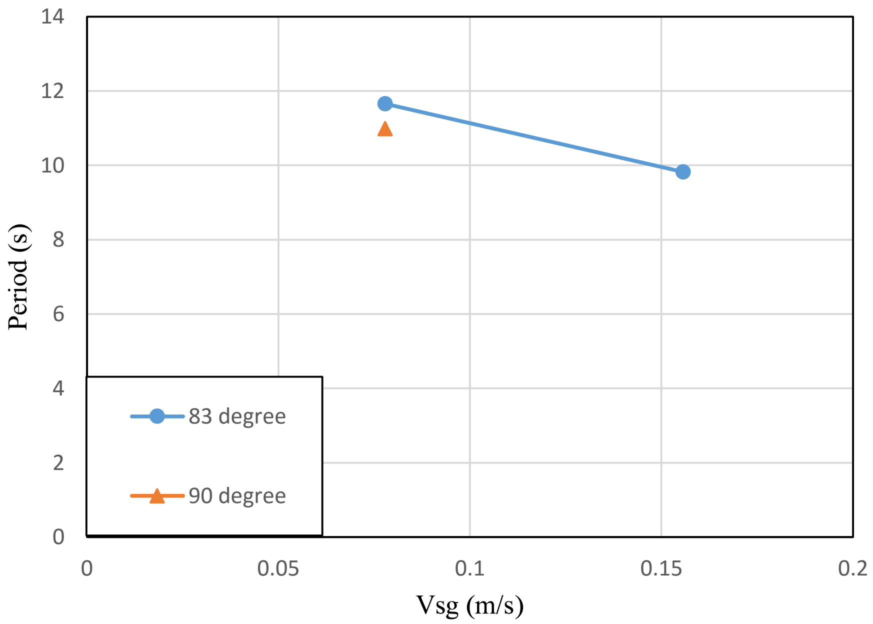

Figure 4-15 Effect of bend angle on severe slug period (Vsl = 0.3735 m/s)

4.2.3 Taylor Bubble Rise Velocity

Figures 4.16 to 4.2 show that the Taylor bubble velocity increases with gas superficial velocity at constant liquid superficial velocity. But in high gas flowrates there, this trend has changed and faced a hump in graphs. In low gas flowrates any increasing in liquid superficial velocity decrease Taylor bubble velocity. In other hand in high gas flowrates this parameter turns reversely. Also this kind of behavior is evident in angle sensitivity test. In low gas flowrates any decreasing in riser angle shows its effect on Taylor bubble velocity increasing. But in high gas flowrate Taylor bubble velocity increases with riser angle.

Increasing in Taylor bubble velocity is because of increasing in gas flowrates. The velocity of Taylor bubble in sever slug regimes is not constant in a cycle. Bubbles leak and pass through the bend and slowly after that they can overcome the hydrostatic pressure of the liquid column. From the begging of gas leak until total gas passage in gas phase has an increasing and then decreasing behavior. The Taylor bubble velocity that is presented here is mean gas velocity of gas passage cycles.

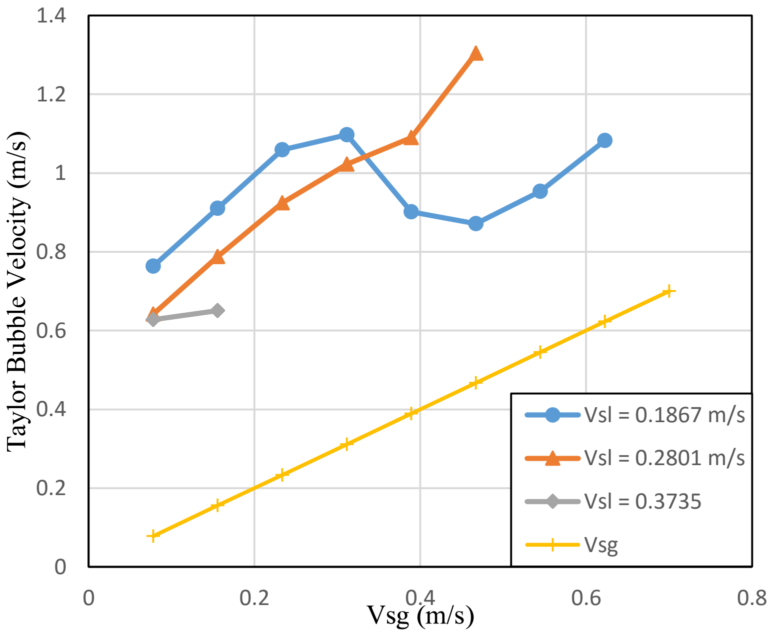

Figure 4-16 Effect of liquid and gas superficial velocity on Taylor bubble velocity (83 degree bend)

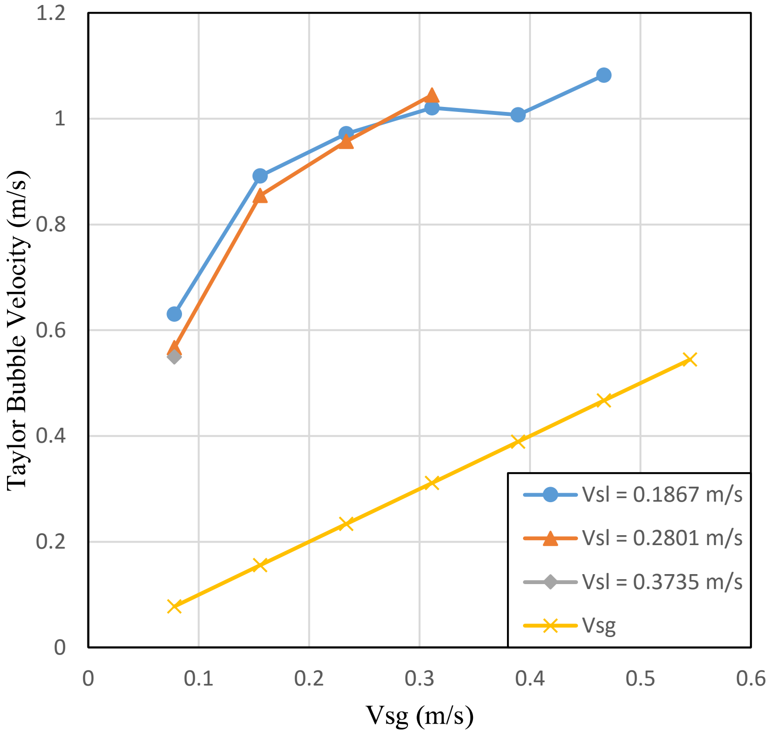

Figure 4-17Effect of liquid and gas superficial velocity on Taylor bubble velocity (90 degree bend)

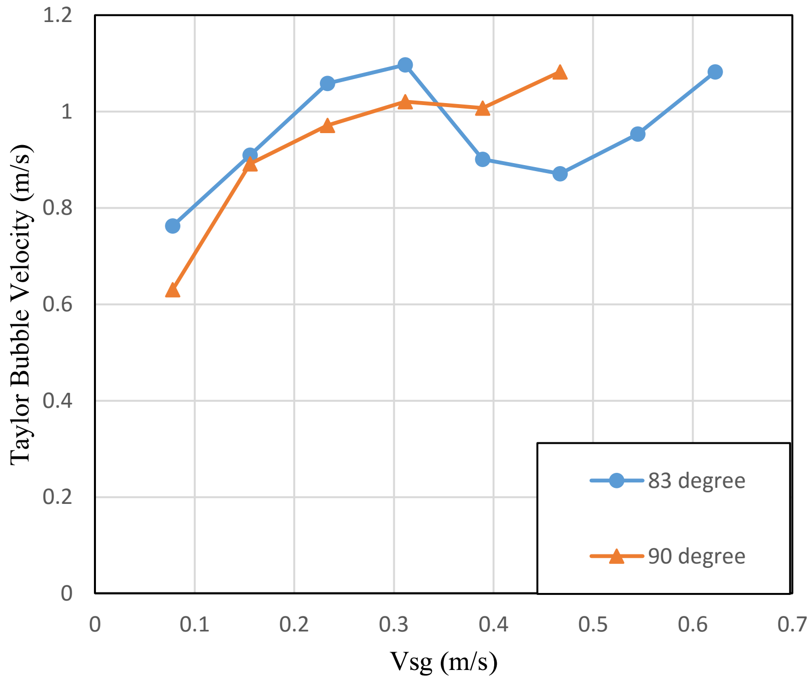

Figure 4-18Effect of degree on Taylor bubble velocity (Vsl = 0.1867 m/s)

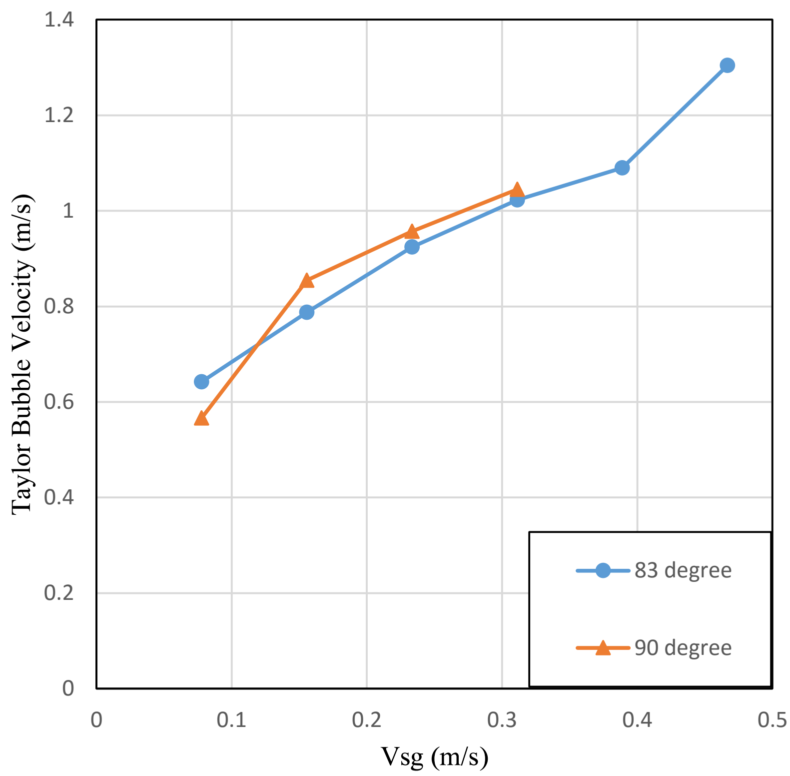

Figure 4-19Effect of degree on Taylor bubble velocity (Vsl = 0.2801 m/s)

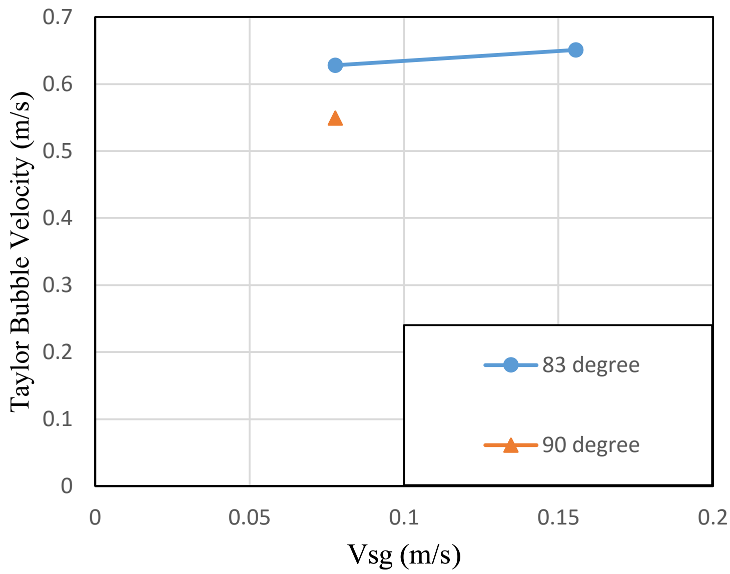

Figure 4-20Effect of degree on Taylor bubble velocity (Vsl = 0.3735 m/s)

4.2.4 Fluids Passage Ratio (Slug and Taylor Bubble Passage Duration)

Analysis of gas lock releasing periods shows interesting results. This period time can be divided to two separated times. 1. Liquid passage duration 2. Taylor bubble passage duration. A ratio is defined as a fluids passage ratio and equals to Taylor bubble passage duration divided by the liquid passage duration. The fluids passage ratio is found to increase with gas superficial velocity and liquid superficial velocity. This can be explained by that fact that increases in liquid and gas superficial velocity decreases liquid passage duration. Meanwhile there is no clearly defined trend for Taylor bubble passage duration with liquid and gas superficial velocity.

In the transition area of flow map between gas lock and slug regions this fluid passage ration for both regimes are nearly close. So the main reason of changing type of flow regimes is that increasing of any fluids flowrates can change the gas lock flow regimes to slug one. In addition decreasing of gas lock releasing period helps to slug regimes occurs in the riser. Figure 4.21 to 4.25 to confirm these facts and also figure 4.26 to 4.29 show that fluid passage ratio decreases with riser angle. This fact verify that reduction of riser angle prolong the gas lock effect in higher flowrates and extends the gas lock flow region in flow map.

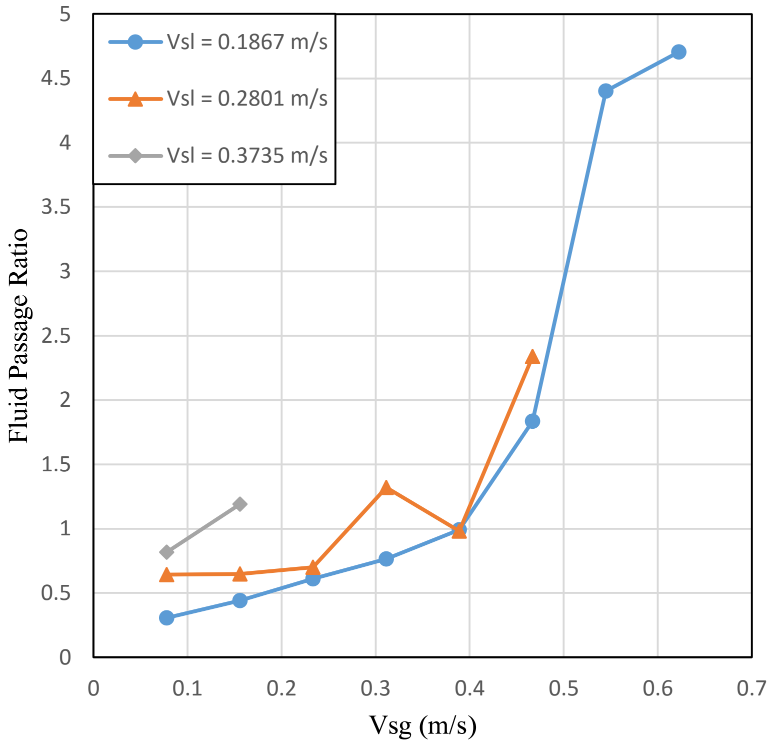

Figure 4-21Effect of liquid and gas superficial velocity on fluid passage ratio (83 degree bend)

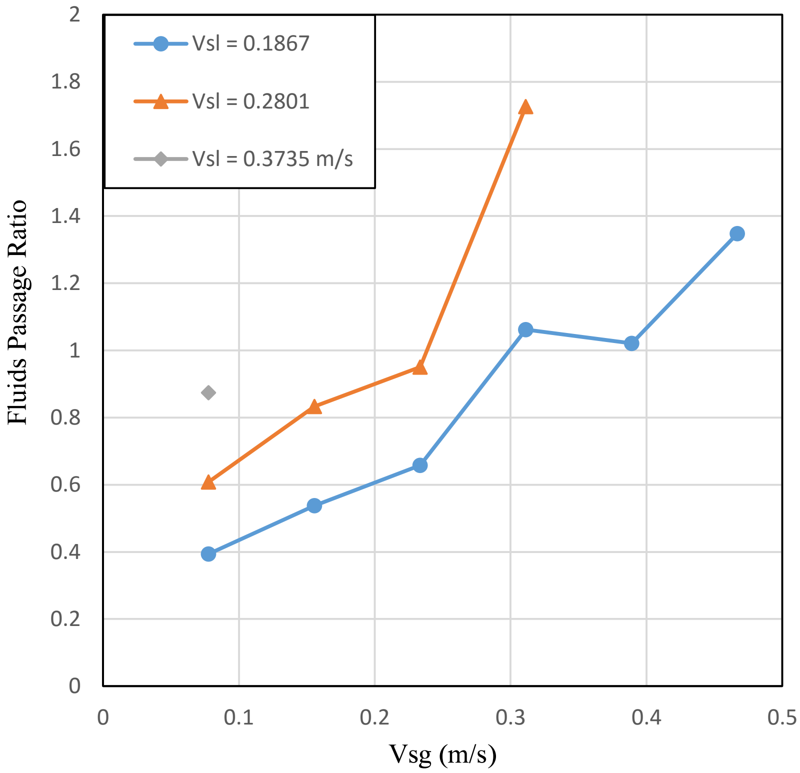

Figure 4-22Effect of liquid and gas superficial velocity on fluid passage ratio (90 degree bend)

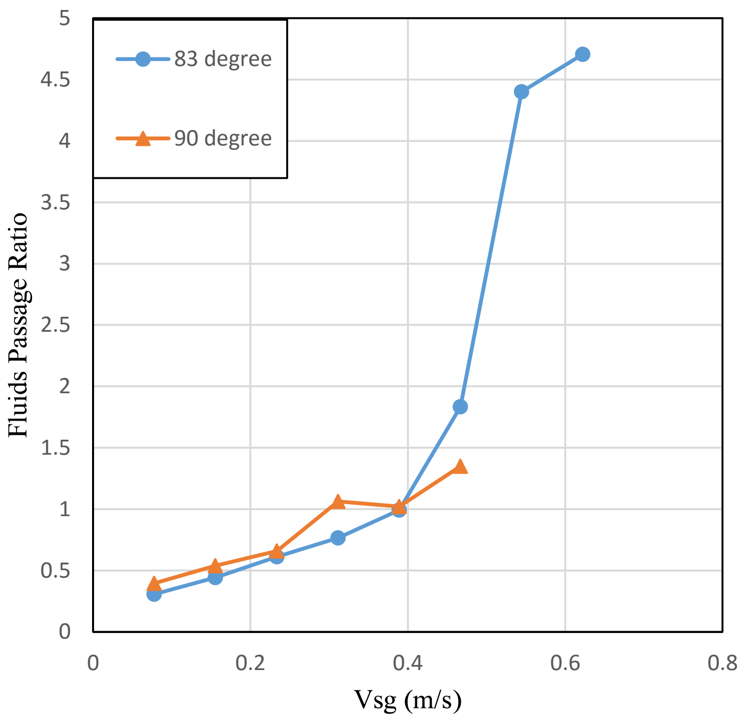

Figure 4-23Effect of degree on fluid passage ratio (Vsl = 0.1867 m/s)

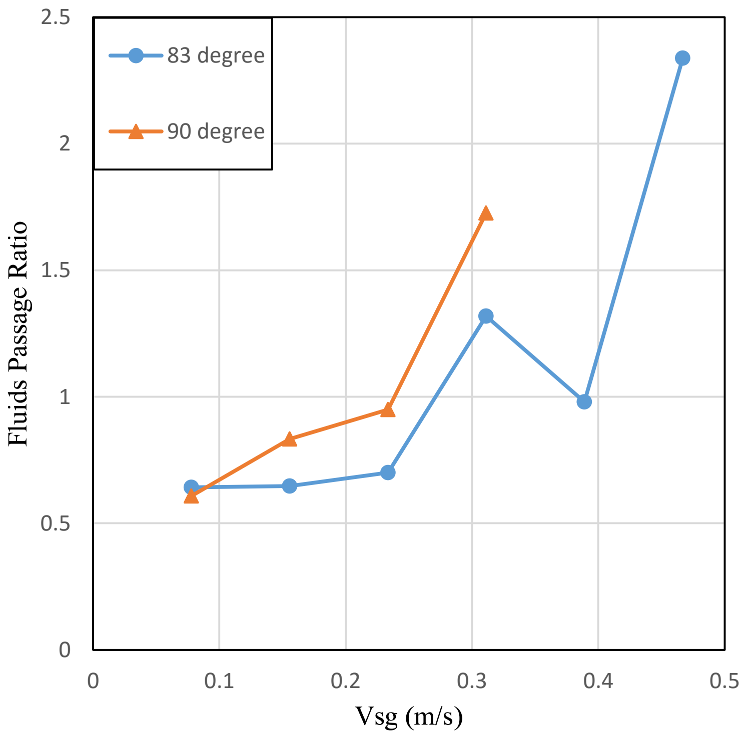

Figure 4-24Effect of degree on fluid passage ratio (Vsl = 0.2801 m/s)

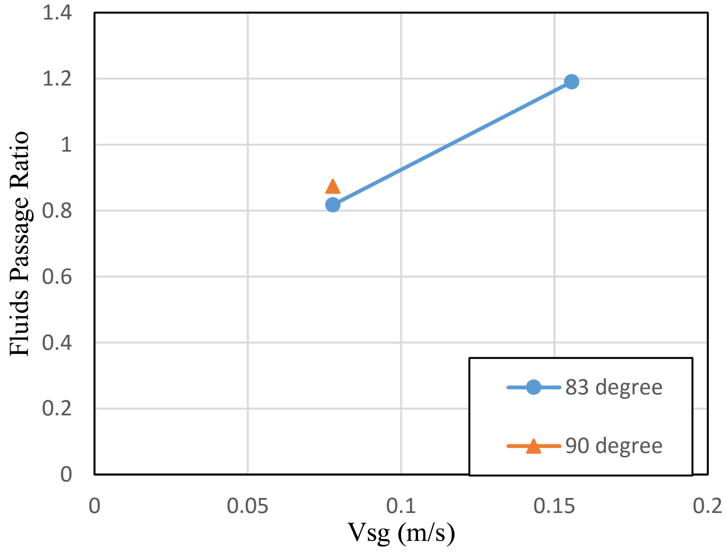

Figure 4-25Effect of degree on fluid passage ratio (Vsl = 0.3735 m/s)

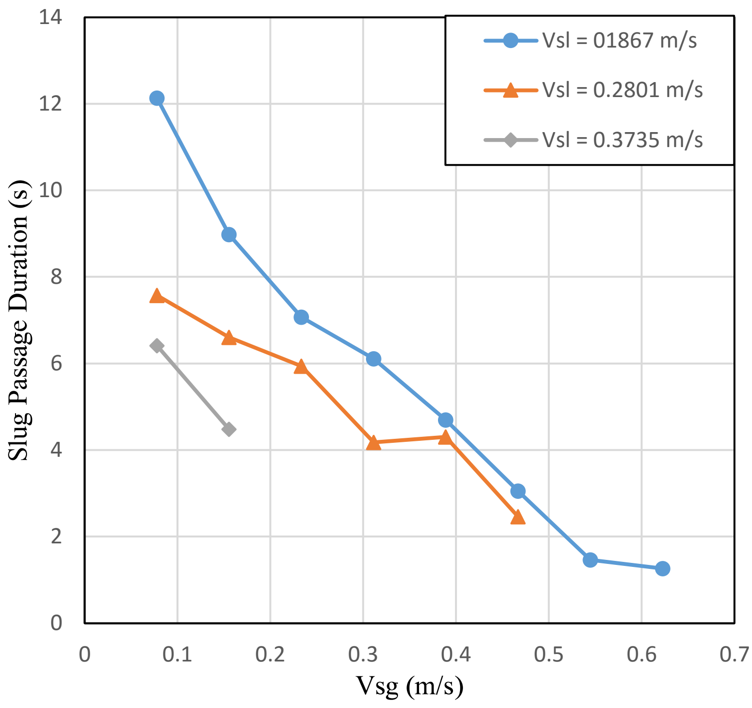

Figure 4-26Effect of liquid and gas superficial velocity on slug passage duration (83 degree bend)

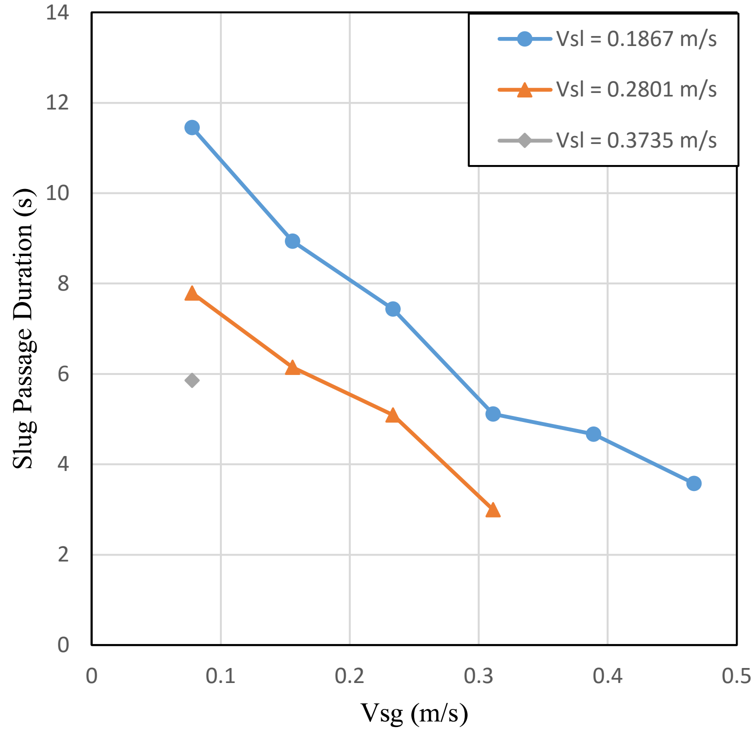

Figure 4-27 Effect of liquid and gas superficial velocity on slug passage duration (90 degree bend)

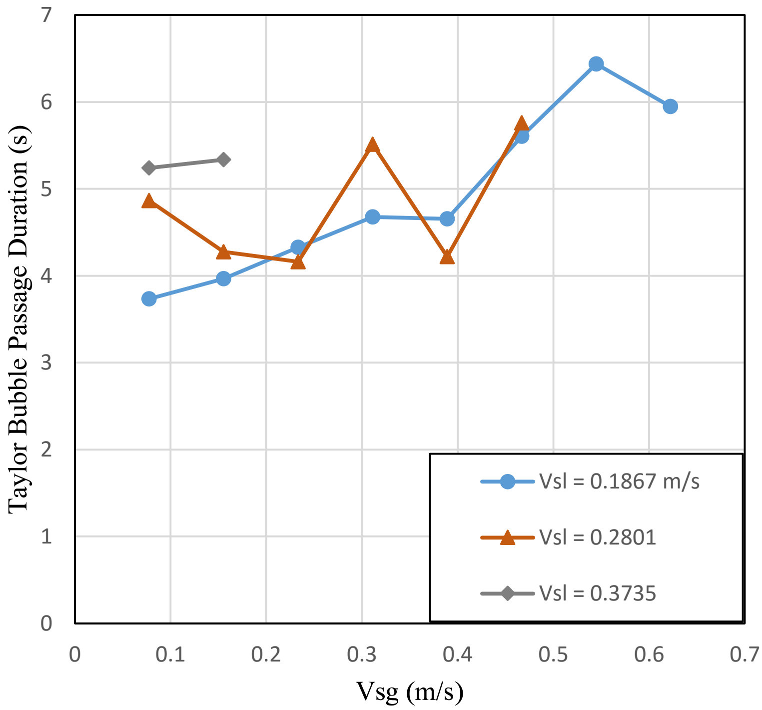

Figure 4-28Effect of liquid and gas superficial velocity on Taylor bubble passage duration (83 degree bend)

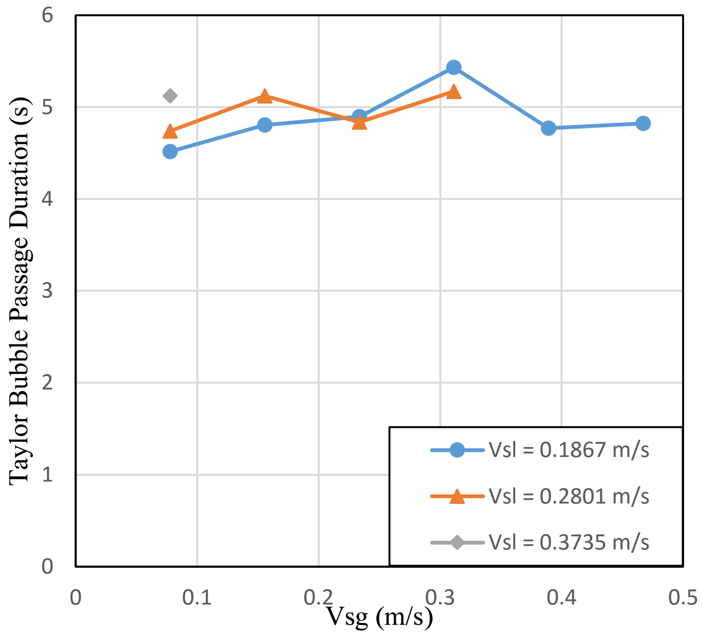

Figure 4-29Effect of liquid and gas superficial velocity on Taylor bubble passage duration (90 degree bend)

4.2.5 Pressure Drop

Pressure drop through vertical line of riser is measured for different gas and water and superficial velocity. Air superficial velocity was varied between 0.15 m/s and 0.46 m/s and water was just set for 0.23 and 0.46 m/s which velocity of 0.23 m/s was represented gas lock regime on the other hand velocity of 0.46 m/s was represented slug flow regime. Most well-known and applicable pressure drop empirical correlation are used and the error of their results are shown in Figures 4.30 and 4.31. As it is shown the error of results for 0.46 m/s are acceptable. The corrected begs & brill correlation has reached the best answers but all of them have overpredicted. This is similar to the result that reported by other researchers such as Lawson and Brill (1974) and Vohra et al. (1973) for first time. But for liquid superficial velocity of 0.23 it is clear these empirical correlation are not useful for vertical pipe which is used in riser. They have not considered gas lock phenomena if a bend was placed before the vertical pipe [5].

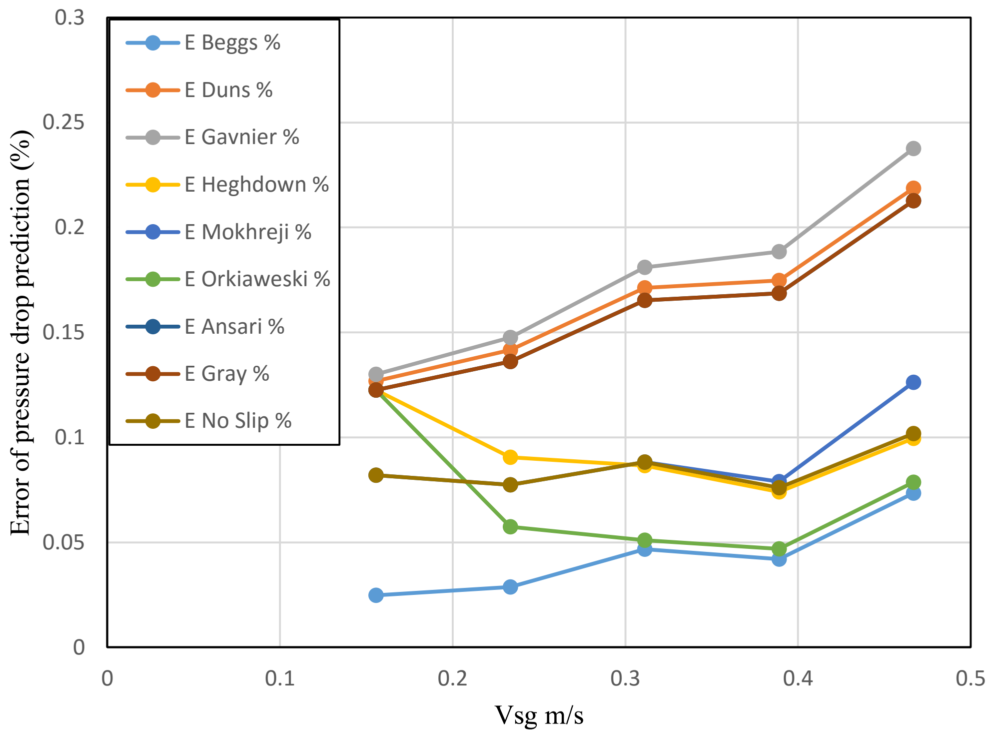

Figure 4-30Error of pressure drop prediction (Vsl=0.23 m/s)

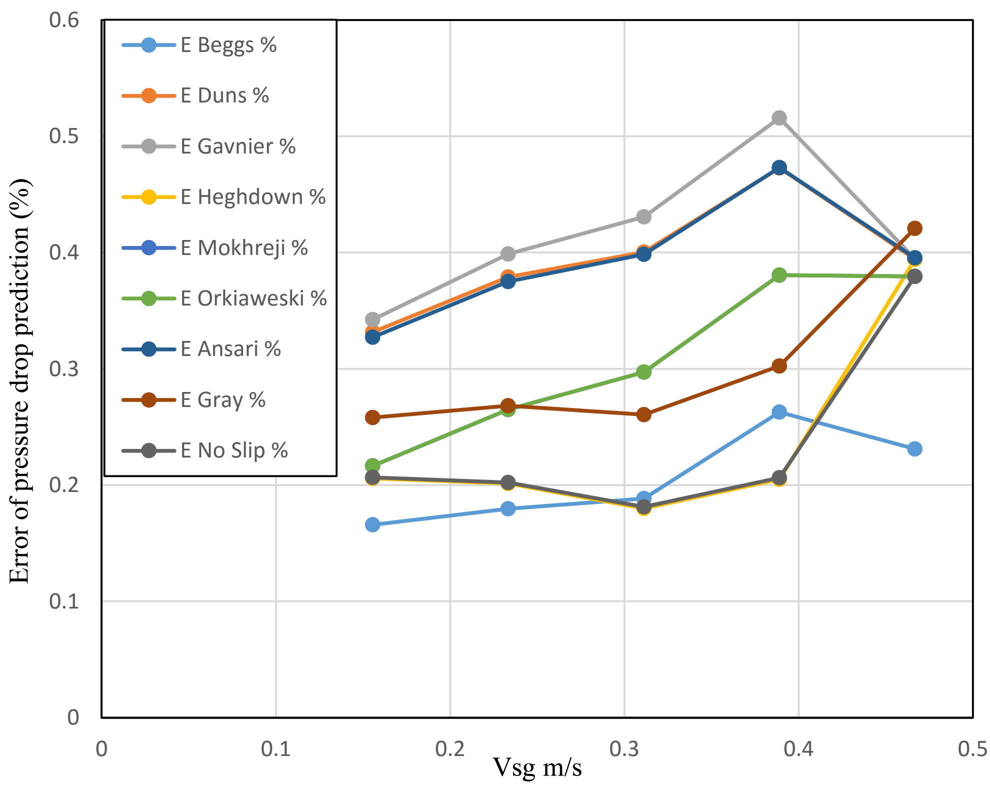

Figure 4-31Error of pressure drop prediction (Vsl=0.46 m/s)

CHAPTER FIVE: CONCLUSION & RECOMMENDATION

In this chapter, we first present summary and conclusions of this thesis. Then we conclude this work by providing recommendation for future study.

5.1 Conclusion

The performance of a test GLCC was investigated by changing in different part of separator. Following conclusion are conducted through these changes.

- Reduction in inlet diameter helps the GLCC separator performance. It allows more gas and liquid flowrates enter the separator for total separation by improving the centrifugal effect on liquid and gas phase.
- Reduction in liquid outlet diameter has negative effect in GLCC flowrates domain but this reduction can be used to control the equilibrium liquid level by a gate valve in liquid outlet leg.
- Also any reduction in gas outlet diameter has negative effect on GLCC performance. But in specific situation controlling the amount of accumulated gas in GLCC can avoid liquid carry over in the system.
- Reduction in gas column length shows no effect on the separator flowrates domain.
- Increasing in length of outlet legs increases the friction force and limited the separator performance.
- Reduction in separator body diameter raises the chance of liquid carry over and gas carry under and has negative effect on flowrates domain.

The characteristics of gas lock phenomena has discussed and explained in this work in details. From the present work one can conduct that from the flow behavior:

- The flow maps of a riser with three different angles are plotted and defined slug and gas lock flow regions on them. No gas bubble can be trapped in riser for liquid superficial Velocity of 0.4669 m/s or higher. Also decreases in riser angle extend the gas lock flow region on the flow map.
- Increases of liquid and gas superficial velocity decrease gas lock releasing period. Riser angle has no effect on this parameter.
- Tylor bubble rise velocity trend has hump in high gas flowrate but totally it increases with gas superficial velocity.
- Fluids passage ratio increases with gas and liquid superficial velocity and riser angle. The Tylor bubble passage duration strongly affects the fluid passage ratio. Meanwhile liquid passage duration has no any strong effect on that.

- Decrease of gas changing tock releasing period and increase of fluid passage ration change the flow regime from gas lock to uniform slug in riser at transition flowrates.
- The usual pressure drop empirical correlations have not considered gas lock phenomena in vertical pipe. The Beggs & Brill correlation predict the closest answers to the experiment ones but its answers has insignificant error.

5.2 Recommendation

- CFD studies on vortex flow in GLCC can solve and explane complex flow phenomena
- Using transparent instruments helps to understand the multiphase separation in GLCC better.
- Inclination in vertical pipe for riser can be investigated in further stuies.
- Inclination in both pipe at the same condition is a appropriate recommended topic for study.
- Further optimization of the internal designs should be achieved with CFD modeling.
- To investigate the effect of pressure and temperature for using at industrial applications.
- The reservoir oil can be used as the liquid phase and methane, ethane and propane can be applied as the gas phase.
- To investigate the other flow regime, it is better that the length of bubble column become taller in design of bubble column system.

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چکیده

تفکیک گر های ســـایکلون های اســـتوانه ای گاز و مایع توانســـته اند خود را به عنوان جایگزین مناســـبی به جای تفکیک گر های معمول مایع و گاز که بر اســـاس نیروی جاذبه کار می کنند، چه در آزمـایشـــگـاه و چـه در مناطق عملیاتی ثابت کنند. در این پژوهش به بررســـی تغییرات در فیزیک و ســاختمان این گونه از تفکیک گر ها و تاثیر آن بر عملکردشان بررسی شده است. این تغییرات فیزیکی شامل افزایش در طول پایه خروجی، کاهش در ستون گازی بدنه اصلی، قطر ورودی، قطر خروجی های گاز و مایع و بدنه ی اصـلی تفکیک گر می باشـد. نتایج این بررسی نشان داده است که کاهش در قطر ورودی تفکیک گر باعث بهبود عملکرد آن می شود اما هر گونه کاهش در قطر بدنه اصلی و یا خروجی مایع تفکیکگر تاثیر منفی بر روی عملکرد دســتگاه دارد. همچنین تغییر در قطر خروجی گازی دستگاه تاثیر نحســـوســـی ندارد. افزایش طول پایه های خروجی گاز و مایع به علت با بردن اثر اصـــطکاک در تفکیک گر نا مطلبوب است.

گیر افتادن گاز یکی از مشـکلات اصـلی در بالابرنده ها در صـنعت و به ویژه در تاسـیسـات تحت الارضـی مانند تفکیک گر ها است. در این تحقیق نتایج کاملی از مطالعات و آزمایش های انجام گرفته بر روی یک بالابرندهی شفاف به قطر 24.5 میلی متر و که در سه زاویه مختلف جهت بررسی پدیده ی تجمع گاز قابل تنظیم اسـت، گزارش شـده است. نقشه های جریانی در دامنه ی جریانی آزمایش شده در هر سـه زاویه ارائه شـده اند تا نشـان دهنده محدوده ی تشـکیل این پدیده در بالابرنده ها باشـند. همچنین اطلاعات مخصـوص این پدیده مانند تکررپذیری، سـرعت جباب های تیلور و نسـبت گذردهی سیالات در بالابرنده نسبت به سرعت ظاهری گاز و مایع و همچنین زاویه گزارش شده است. با بالا رفتن ســـرعت ظاهری گاز تکرر پذیری و ســـرعت حباب های تیلور افزایش می یابند. زمان عبور حباب های تیلور در زاویه 83 درجه با افزایش سرعت مایع و گاز بیشتر می شود ولی این رابطه در زاویه های بالاتر مانند 90 درجه مشـاهده نشـده اسـت. همچنین افزایش در سرعت گاز و مایع باعث کاهش زمان عبور

فاز مایع در بالابرنده می شـود. تمامی روابط تجربی معمول برای پیش بینی افت فشـار در لوله یعمودی بالابرنده دقیق نبوده و بیشتر از مقدار حقیقی گزارش کرده اند.

مقدمه

در دهه گذشته جدایش کامل مخلوط گاز و مایع در صنعت نفت و گار بسیار با اهمیت شده است. به همین علت پیدا کردن روش های جدایش کم هزینه تر و مفیدتر بیشتر نیاز می شود. یکی از این روش ها استفاده از تفکیک گر های سایکلون استوانه ای گاز و مایع است. برای جدا سازی فاز های مختلف سیال تولیدی از چاه از وسیله ای به نام تفکیک گر استفاده می شود. تفکیک گر های مرسوم مورد استفاده در صنعت نفت بسیار بزرگ، گران و سنگین می باشند به همین علت توجه صنعت به طراحی و استفاده از راه حل های جایگزین معطوف شده است. یکی از راه حل های مناسب جهت رفع این مشکلات استفاده از تفکیک گر گاز و مایع سایکلون استوانه ای است. این نوع تفکیک گر، ساده، کوچک و سبک و به علت هزینه های عملیات، نگهداری و تعمیر پایین از لحاظ اقتصادی بسیار مقرون به صرفه است. به همین دلایل استفاده از این نوع تفکیک گر به سرعت در صنعت گاز و نفت در حال رشد و توسعه است.

هدف اصلی تفکیک گر سایکلون استوانه ای، جداسازی گاز و مایع از یکدیگر است. اخیراً تفکیک گر های کوچک مانند تفکیک گر گاز و مایع سایکلون استوانه ای به علت آسان کردن مراحل عملیاتی بسیار محبوب شده اند. استفاده از روش های جایگزین جداسازی نشان از این دارد که این گونه راه حل های جایگزین به مراحل کاربردی قابل قبولی رسیده اند. از جمله ی این روش ها می توان به استفاده ی این دستگاه در فرآیند جداسازی و کنترل سیستم چندفازی اشاره کرد. اندازه گری جریان های چندفازی، کمک به پمپ سیالات چندفازی ،چاه آزمایی چاه های گازی و نفتی ،کنترل نسبت گاز به مایع ،کنترل جریان لخته ای و ... از کاربردهای تفکیک گر سایکلون استوانه ای می باشد.

عملکرد این نوع تفکیک گر بر اثر دو پدیده محدود می شـود: خروج سیال مایه از مسیر گازی به صـورت قطرات ریز و خروج سـیال گاز از مسـیر مایع به صـورت حباب. پیش بینی وقوع این دو پدیده اجـازه ی طراحی منـاسـب تفکیک گر ایده آل را به ما خواهد داد. هدف اصـلی این پروژه بررسـی آژمایشگاهی عملکرد دستگاه و این دو پدیده و تاثیر آنها بهروری علمکرد تفکیک گر خواهد بود.

از طرف دیگر وجود یک بالابرندهی افقی برای رســاندن جریان چندفازی به درون تفکیک گر الزامی اسـت. توصیف خصوصیات جریان لخته ای در بالابرنده ها یکی از موضوعات اصلی در مباحث دوفازی اسـت. در شرایط خاصی عملیات انتقال جریان چندفازی ناپایدار شده و جریان لخته ای به لخته ای حاد تغییر می کند. این تغییر شـرایط عملیاتی را برای پیش بینی صحیح جریان سیالات در لوله دشوار می کند. طراحی ضعیف و ایجاد این ناپایداری می تواند باعث افت تولید به میزان 50 درصد شود. تشکیل و ایجاد این رژیم ناپایدار در بالابرنده ها شامل 4 مرحله کلی می شود. 1. تشکیل لخته 2. تولید لخته 3. نفوذ گاز در بالابرنده 4. تولید گاز. این مراحل با تغییرات ممتد فشار و سرعت گاز و مایع در خروجی همراه است.

دبی کم گاز و مایع عبوری از درون بالابرنده یکی از دلایل عمده تشــکیل جریان لخته ای حاد است. در تمامی نقسه های جریانی لوله های عمودی و افقی ایجاد این نوع از رفتار جریانی پیش بینی نشـده اسـت. در این مطالعه تلاش شـده است تا با انجام سری آزمایشات مشخص و کاربردی در یک بالابرنده ی شـفاف که می تواند در ســه درجه ی متفاوت تنظیم شـود، نحوه تشـکیل و خصـوصیات این نوع از رژیم جریانی بررسی شود. این خصوصیات جریانی بحث شده شـامل نقشــه جریانی، سـرعت حرکت حباب های تیلور، زمان عبور و تکررپذیری جریان، زمان عبور سـیال مایع، زمان عبور سیال گازی، نسبت گذردهی سیالات و افت فشار می باشد. تمامی این خصوصیات در سه زاویه متفاوت بررسی شده اند.

این مطالعه به منظور انجام بررسی‌های زیر انجام شد:

- عملکرد تفکیک گر های سایکلون استوانه ای گاز و مایع
- تاثیر تغییرات بر روی قطر ورودی بر روی عملکرد و دامنه ی کاری تفکیک گر های سایکلون استوانه ای گاز و مایع

- تاثیر تغییرات بر روی قطر خروجی گازی بر روی عملکرد و دامنه ی کاری تفکیک گر های سایکلون استوانه ای گاز و مایع
- تاثیر تغییرات بر روی قطر خروجی مایع بر روی عملکرد و دامنه ی کاری تفکیک گر های سایکلون استوانه ای گاز و مایع
- تاثیر تغییرات بر روی قطر بدنه اصلی بر روی عملکرد و دامنه ی کاری تفکیک گر های سایکلون استوانه ای گاز و مایع
- تاثیر تغییرات بر روی طول لوله خروجی بر روی عملکرد و دامنه ی کاری تفکیک گر های سایکلون استوانه ای گاز و مایع
- تاثیر تغییرات بر روی طول ستون گازی و قطر بدنه اصلی بر روی عملکرد و دامنه ی کاری تفکیک گر های سایکلون استوانه ای گاز و مایع
- خصویات رژیم جریانی لخته ای حاددر باالبرنده های تفکیک گر
- تاثیر تغییرات در زاویه باالبرنده بر روی خصوصیات فیزیکی رژیم جریانی لخته ای حاد
- تاثیر تغییرات در دبی ورودی گاز و مایع به باالبرنده بر روی خصوصیات فیزیکی رژیم جریانی لخته ای حاد

1) پیشینه تحقیق در موضوعات:

- دسته‌بندی رژیم‌های جریانی
- تحول در ساخت و استفاده از تفکیک گر های سایکلون استوانه ای گاز و مایع در صنعت
- خصوصیات رژیم لخته ای حاد در شرایط محیطی متفاوت

2) مواد به کار برده شده و روش انجام آزمایش

3) نتایج و بحث بر روی داده‌های اندازه گیری شده

4) نتیجه‌گیری کار و پیشنهادات برای کارهای آینده