NTNU – Trondheim
Norwegian University of
Science and Technology

Robust control solutions for stabilizing flow from the reservoir: S-Riser experiments

Mahnaz Esmaeilpour Abardeh

Chemical Engineering
Submission date: June 2013
Supervisor: Sigurd Skogestad, IKP
Co-supervisor: Ole Jørgen Nydal, EPT
Esmaeil Jahanshahi, IKP

Norwegian University of Science and Technology
Department of Chemical Engineering

Robust control solutions for stabilizing flow from the reservoir: S-Riser experiments and simulations

Mahnaz Esmaeilpour Abardeh

June 26, 2013

Preface

This thesis is written as the final part of my Master degree in Chemical Engineering at the Norwegian University of Science and Technology (NTNU), class of 2013.

I would like to express my greatest gratitude to my highly knowledgeable supervisor, professor Sigurd Skogestad, for all his helps, his good guidance and encouragements. I am also grateful to my co-supervisors, professor Ole Jorgen Nydal and PhD student Esmaeil Jahanshahi, who helped and supported me throughout my thesis. It has been a great opportunity and honor for me to be part of your team in generating new ideas and I am confident what I have learned through this thesis will be surely used in practice in my professional career.

Declaration of Compliance

I, Mahnaz Esmaeilpour Abardeh, hereby declare that this is an independent work according to the exam regulations of the Norwegian University of Science and Technology (NTNU).

Date and signature:

26/06/2013 Esmaeilpour

Abstract

One of the best suggested solutions for prevention of severe-slugging flow conditions at offshore oilfields is the active control of the production choke valve. This thesis is a study of robust control solutions for stabilizing multiphase flow inside the riser systems; through S-riser experiments and OLGA simulations. “Nonlinearity” as the important characteristic of slugging system poses some challenges for control. Focus of this thesis is on online tuning rules that take into account nonlinearity of the slugging system. The main objective has been to increase the stability of riser systems at higher levels of valve openings with more production rates.

Similar research has been done previously, but is repeated in this thesis using new systematic tuning methods. Three different tuning methods have been applied in this thesis. One is Shams’s set-point overshoot method developed by Shamsozzhoha (Shamsuzzoha and Skogestad 2010). The other is IMC- (Internal Model Control) based tuning method with respect to the identified model of the system from closed-loop step test. The last tuning method is simple PI tuning rules with gain scheduling for the whole operating range of the system considering the nonlinearity of the static gain. The two latter methods have been developed very recently by Jahanshahi and Skogestad (Jahanshahi and Skogestad 2013).

Two series of experiments have been carried out using a medium-scale two-phase flow S-riser loop. A single loop control scheme with riser-base pressure as the measurement was used. The robustness of different tuning methods was compared by slowly decreasing the set-point of the closed-loop system, which was the inlet pressure, until instability was reached. The choke valve opening was increasing gradually by decreasing the set-point. A control with a robust tuning method can maintain system stability in a large range of conditions. The choke valve was then replaced with a quicker valve after the first set of experiments. The same experiments were repeated and the effect of control valve dynamics was investigated thereafter.

The experiments were simulated in OLGA and the same control tests were performed. The OLGA case was constructed based on the first series of tests with valve 1 and the designed controllers with different tuning strategies were applied. Results of the experiments verified those of the simulations.

The tuning method with the highest robustness was thus the one which could stabilize the system at the largest choke valve opening (the lowest inlet pressure). The best tuning method, with respect to robustness is the simple PI tuning rules with gain scheduling for the whole operating range of the system. With this method, it was possible to stabilize the experimental riser system up to a choke valve opening of 37 % from an open-loop stability of 16 %. It was also able to stabilize the simulated riser system until a choke valve opening of 75 % from an open-loop stability of 26 %.

Top side measurements were in general difficult to use in anti-slug control. Measurement of the topside density using a conductance probe installation was not successful. Therefore, no cascade anti-slug control schemes could be tested.

Contents

1 Introduction

Multiphase pipelines are a common feature of offshore production in the North Sea. They connect subsea wells to the topside processing facilities or the platforms. In many points of transportation, these pipelines get the shape of L-shaped or S-shaped risers. The stability of multiphase flow inside these pipeline-riser systems is of great importance and many efforts have been spent on this issue so far. In low reservoir pressures or low flow rate conditions the liquid phases tend to accumulate in low points and form liquid slugs. This leads to the pipeline or riser blockage and can be more dangerous when the length of slugs is comparable to the length of the riser. This phenomenon is called Severe slugging (also Terrain slugging or Riser slugging) and is characterized by large oscillatory variations in pressure and flow rates (Storkaas 2005). These large variations lead to a poor separation, unwanted flaring and even a plant shutdown in the worst case.

Reducing opening of the topside choke valve has been a traditional way to suppress severe slugging. However, this increases the valve back pressure and therefore decreases the production rate from the well.

Active feedback control of the topside choke valve can make it possible to stabilize the flow at the conditions where normally severe slugging is predicted. This reduces the need for additional topside equipment and allows a higher rate of oil recovery. The control system is called anti-slug control and its main objective is to keep the multiphase flow as stable as possible by manipulating the topside choke valve using the parameters such as pressure or density as the control variables.

In the way of developing new technologies for stabilizing control of severe slugging in riser systems many researches have been done at the Norwegian University of Science and Technology. The work has been guided by Skogestad (Skogestad 2003; Storkaas 2005; Shamsuzzoha and Skogestad 2010; Jahanshahi and Skogestad 2011; Skogestad and Grimholt 2011; Jahanshahi and Skogestad 2013) and performed at the department of Chemical Engineering. Storkaas (Storkaas 2005), Sivertsen (Sivertsen 2008), Jahanshahi (Jahanshahi and Skogestad 2011) and numerous master students have worked on modeling and controlling of riser systems.

Companies like ABB (Havre, Stornes et al. 2000), Statoil and Total have all researched prevention of slugging and built installations at offshore locations. Statoil completed in 2001 their first slug control installation at the Heidrun oil platform. Siemens is also involved in slugging research and funds a PhD program, which this thesis is connect to.

In the anti-slug control system, it is very important that the controllers are fine tuned. Otherwise, the control system is not robust in practice and the closed-loop system becomes unstable after a plant change. The slugging system is highly nonlinear since the gain changes at different operating points. For such a system the controllers need to be retuned at each operating point.

1.1 Scope of the thesis

In this thesis three different tuning methods will be tested with experiments and simulations to find the most robust solution for anti-slug control system. High robustness will be obtained if the system can maintain stability at large deviations from open loop conditions. This means large choke valve openings. The tuning methods are systematic and have been developed very recently (Shamsuzzoha and Skogestad 2010; Jahanshahi and Skogestad 2013).

The experiments of this thesis will be carried out at the department of Energy and Process Engineering. Two series of experiments will be run using a medium-scale two-phase flow S-riser loop. The difference between the two series is the type of choke valve. The aim is to investigate the effect of control valve dynamics on performance of the control system in addition to robustness of the tuning methods. Possibility of different control structures will be also investigated.

The experiments will be simulated in multiphase flow simulator, OLGA, and the same control tests will be performed. Finally the simulated and experimental results will be compared.

2 Background

2.1 Multiphase transport

When it comes to offshore production of oil and gas, long transport of multiphase flow has recently become of great attention. Many pipelines and risers are carrying the combination of natural gas, condensate, oil and water from the North Sea to shore. Previously, large production platforms equipped with process facilities were built on the sea floor with the aim of separating gas, oil and water. Today this can be too expensive and multiphase transportation can save billions of dollars for the oil companies instead.

Design and operation of multiphase transportation systems raise many new challenges. These challenges could be either related to the flow, fluid or the pipe integrity. Pressure drop/ boosting, Slugging, liquid emulsion, temperature change, scaling, hydrate and wax formation can be examples of them. Overcoming these challenges and having a safe and uninterrupted multiphase flow refers to the term "flow assurance". This term was first used by Petrobras in the early 1990s and it originally referred to only thermal hydraulics and production chemistry issues encountered during oil and gas production (Fabre, Peresson et al. 1990).

One important issue in flow assurance is stabilizing the multiphase flow inside the pipeline-riser systems. From a control engineering point of view, this can be referred as control of the disturbances in the multiphase flow as the feed to the separation process. Avoiding variations in the flow entering the processing unit, at the outlet of the multiphase pipelines is the issue of interest for control (Bratland 2010). The ability of predicting the flow patterns and reserving a stable flow is of great importance, which is the objective of the thesis. Figures 2.1 and 2.2, adapted from Bratland (Bratland 2010), describe possible flow patterns inside the horizontal and vertical pipelines.

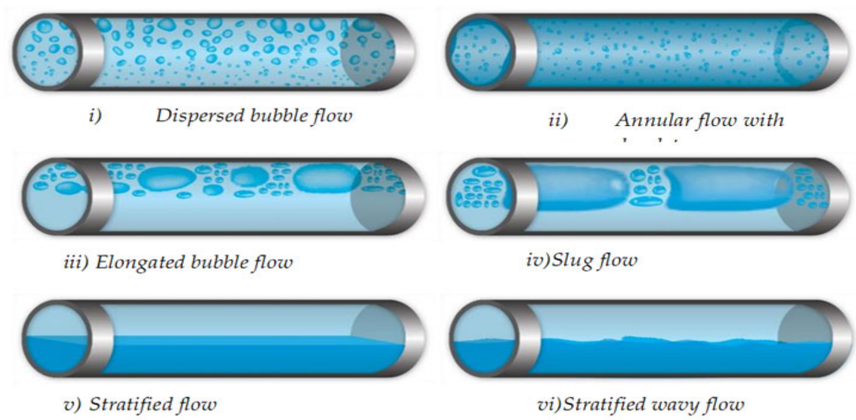

Figure 2.1: Gas-liquid flow regimes in horizontal pipes.

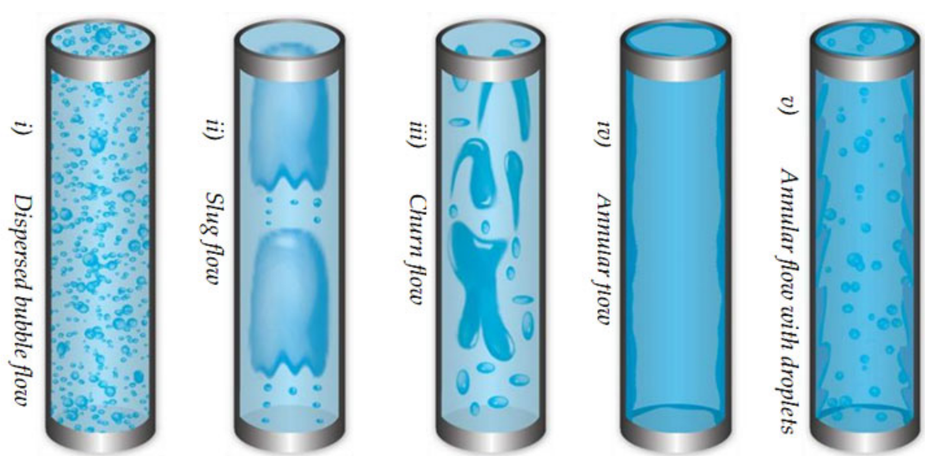

Figure 2.2: Gas-liquid flow regimes in vertical pipes. Slug flow is the point of interest in the thesis.

Changing multiphase flow between different flow regimes can be described by a typical flow regime map shown in figure 2.3, adapted from Taitel (Taitel 1986). The boundaries between stable and unstable regions are clearly shown in the flow regime map. With applying feedback control these boundaries can be moved and thereby the stable region can be increased.

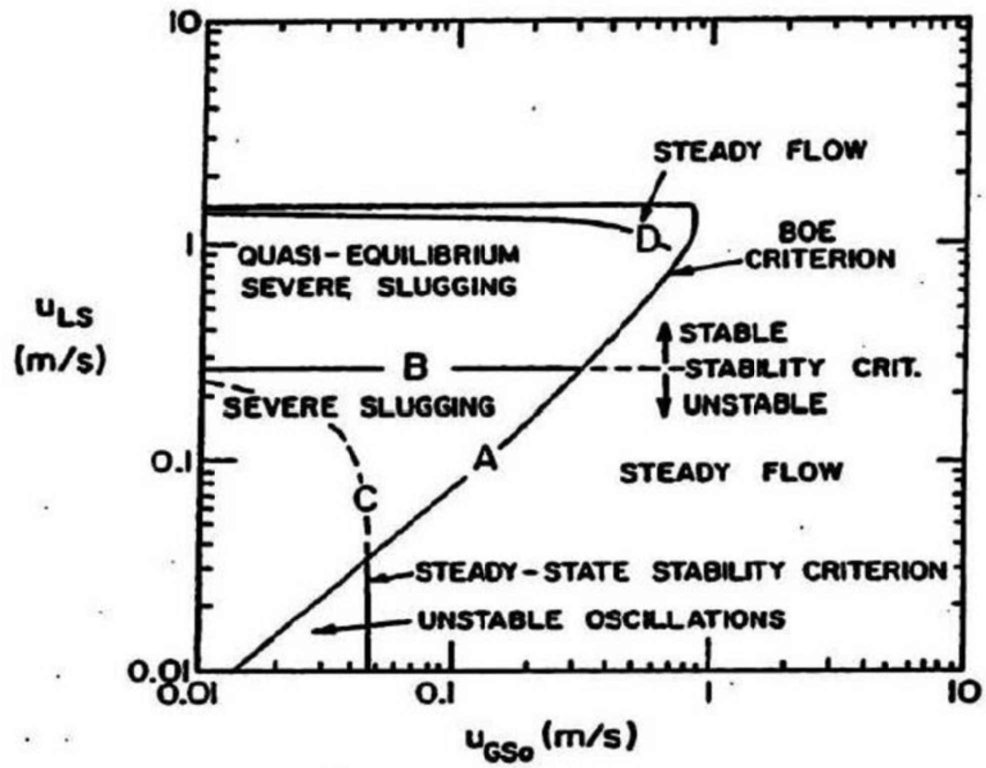

Figure 2.3: Stability map for multiphase flow (Taitel 1986). Stability boundaries are clearly shown in the map.

2.2 Slug flow

Among the flow assurance concerns, management of slugging in system deliverability has received much interest in recent years (Godhavn, Fard et al. 2005). Slug flow is one of the flow patterns characterized by alternating slugs of gas and liquid flowing in the pipes. In this type of flow regime, elongated bubbles of gas separated by "slugs" of liquid, travel from one end of the pipe to the other end. It can be either due to different velocities of gas and liquid phase which is referred as hydrodynamic slugging or pipeline geometry which is referred as terrain induced slugging. The latter one is common in risers and its main reason is the gravity. A schematic map of slug flow is shown in figure *2.4*, adapted from(Yan and Che 2011). The master unfavorable effect of slug flow is its instability that has a negative impact on the operation of offshore production facilities. The periodic oscillations of liquid and gas phases due to their inhomogeneous distribution cause oscillatory pressures and decreases the level of production as large as 50%. The average of these oscillations is lower than the equilibrium production and this gives the production losses. More over these oscillations can damage the pipe and the separation process. For these reasons, suppressing the slug flow is of dominant importance. A homogeneous steady flow with very small bubbles of gas well distributed in the continuous liquid phase is most desired. In such desired situation, the pressure remains constant over time.

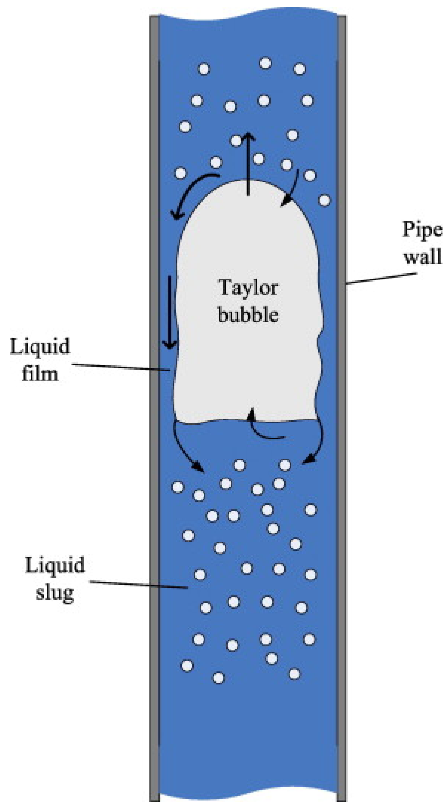

Figure 2.4: Schematic map of slug flow in a vertical pipe in a slug unit (Yan and Che 2011)

2.3 Risers containing multiphase flow

Risers are a special type of pipeline developed for vertical transportation of materials from seafloor to production and drilling facilities on the water's surface. They can be in types of rigid risers, flexible risers and hybrid risers that is a combination of the rigid and flexible. Risers can have many different configurations. However in this thesis all the S-shaped types are the point of interest regardless of their differences. The functional suitability and long term integrity of the riser system affects the selection of riser configuration (Bai 2001).Figure 2.5 shows prevalent riser configurations.

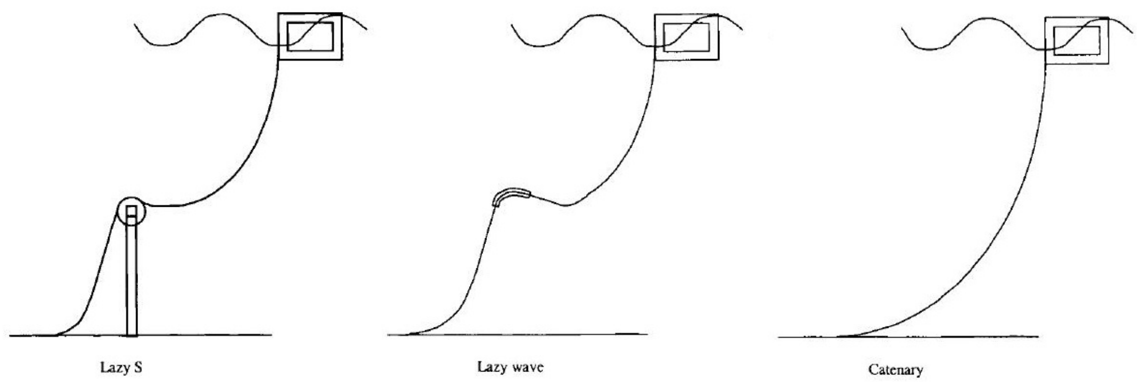

Figure 2.5: Common riser configurations applied in the oil and gas industry(Bai 2001)

2.4 Riser slugging

Riser slugging (also called severe slugging/ terrain induced slugging) is the toughest type of slugging happening in a pipeline-riser system where a downward inclined pipeline is connecting into an upward riser. Storkaas (Storkaas 2005) explains the cyclic behavior of riser slugging illustrated schematically in figure 2.6. It can be broken down into four steps. Step 1: Slug formation: gravity causes the liquid to accumulate in the low point and a prerequisite for severe slugging to occur is that the gas and liquid velocity is low enough to allow for this accumulation. Step 2: Slug production: The liquid blocks the gas flow, and a continuous liquid slug is formed in the riser. As long as the hydrostatic head of the liquid in the riser increases faster than the pressure drop over the riser, the slug will continue to grow. Step 3: Blowout: When the pressure drop over the riser overcomes the hydrostatic head of the liquid in the slug, the slug will be pushed out of the system and the gas will start penetrating the liquid in the riser. Since this is accompanied with a pressure drop, the gas will expand and further increase the velocities in the riser. Step 4: Liquid fall back: After the majority of the liquid and the gas has left the riser, the velocity of the gas is no longer high enough to pull the liquid upwards. The liquid will start flowing back down the riser and the accumulation of liquid starts again.

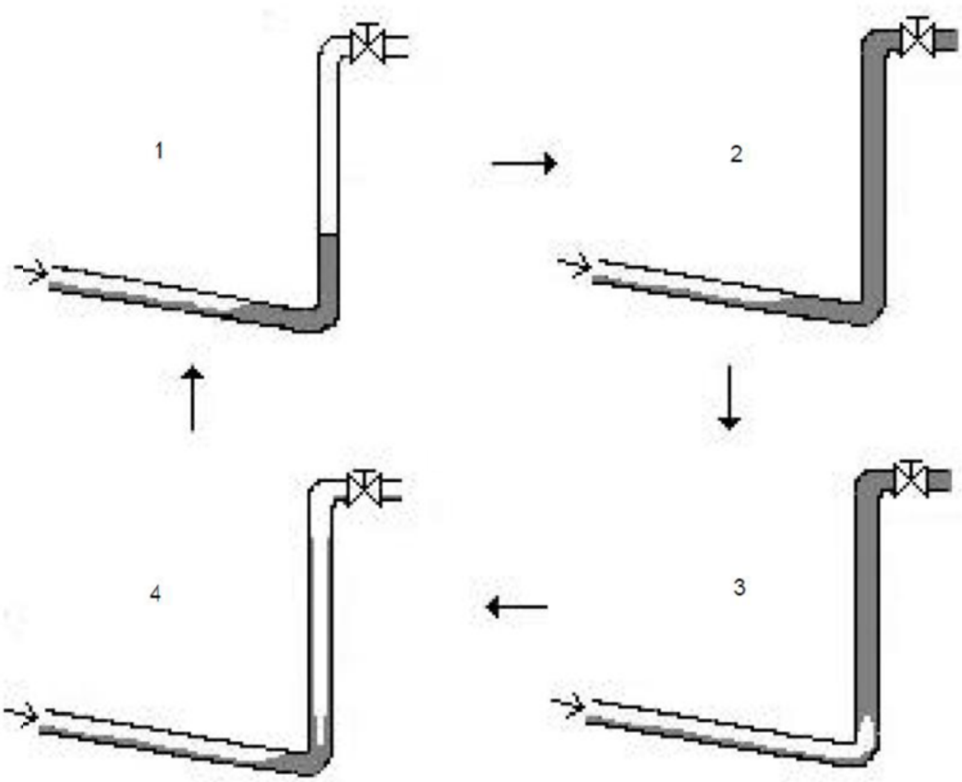

Figure 2.6: Graphical illustration of a slug cycle (Yan and Che 2011).Slug formation is shown in part 1. Slug production is shown in part 2, Blowout in part 3 and liquid fall back in part 4.

Severe slugging causes periods of no liquid or gas production into the separator followed by very high liquid and gas rates, when the liquid slug is being produced. Length of liquid slugs can be several times the length of the riser. This phenomenon is highly undesirable. The large liquid production might cause overflow and shut down of the separator. Fluctuations in gas production might cause operational problems during flaring, and the high pressure fluctuations might reduce the production capacity of the field (Jansen, Shoham et al. 1996). It can reduce operating capacity for separation and compression units. The reduced capacity is caused by the need of larger operating margins to handle the larger disturbances. Larger disturbances require a larger back-off from the optimal operation point, and thus reducing the throughput (Storkaas 2005).

Severe slugging can occur in two different modes of I and II. In type I of severe slugging the liquid fully block the bend while in type II there is a partial blockage at bend and gas passes through. The type I is characterized by large oscillations in pressure and accelerated blow out. In fact the pressure oscillations reflect static head of the riser. There are small pressure oscillations in the severe slugging of type II and the slug length is shorter than the height of the riser. But flow oscillations can be large. Type II slugging is not usually critical for a stable operation. Figures 2.7 and 2.8, adapted from Malekzadeh (Malekzadeh, Henkes et al. 2012), illustrate SS1 and SS2 respectively. Figure *2.7* is based on a measured cycle of the riser ΔP for SS1 corresponding to $U_{SL} = 0.20ms^{-1}$ and $U_{SGO} = 1.00ms^{-1}$. Figure 2.8 is based on the experimental cycle for the riser ΔP of SS2 corresponding to $U_{SL} = 0.10ms^{-1}$ and $U_{SGO} = 2.00ms^{-1}$.

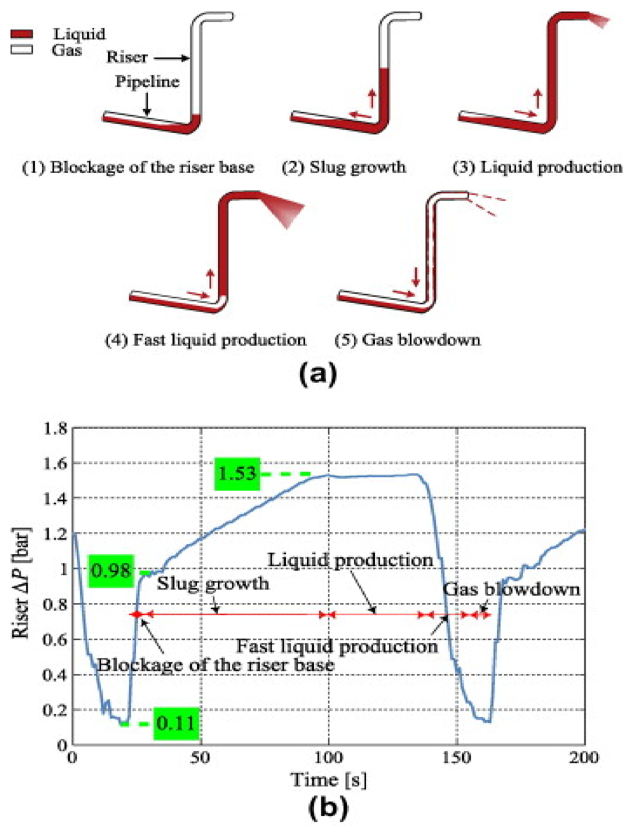

Figure 2.7: Stages for SS1 (a) a graphical illustartion (b) marked on a cycle of an experimental riser ΔP trace ($U_{SL} = 0.20ms^{-1}$ and $U_{SGO} = 1.00ms^{-1}$) (Malekzadeh, Henkes et al. 2012)

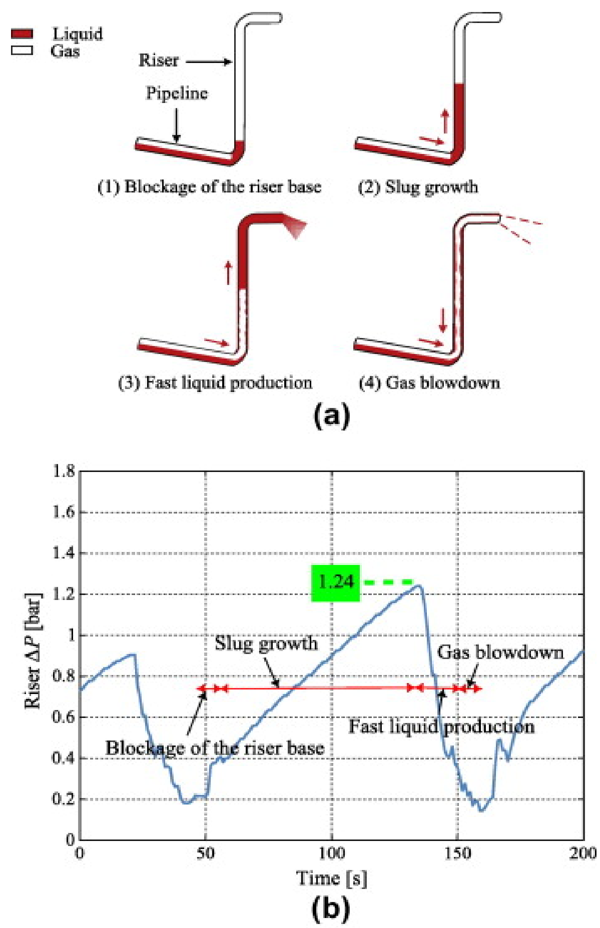

Figure 2.8: Stages for SS2 (a) a graphical illustartion (b) marked on a cycle of an experimental riser ΔP trace ($U_{SL} = 0.10ms^{-1}$ and $U_{SGO} = 2.00ms^{-1}$) (Malekzadeh, Henkes et al. 2012)

2.5 Anti-slug operations

As the fields become more mature the more advanced technology is demanded. The reason is that the energy of reservoir decreases due to its aging. This leads to lower pressure and temperatures in reservoir. The lower pressure of reservoir causes limited driving force to the flow and thereby lower phase velocities in result and finally more probable riser slugging formation. Low temperatures also increase the probability of solid formation. Changing the design of pipe- riser system to avoid slugging cannot be economically feasible. The most common methods for avoiding slugging are presented below.

2.5.1 Choking

Schmidt *et al.* (1979) first suggested choking (decreasing the opening Z) of the valve at the riser top as an elimination way of severe slugging. The theory behind this suggestion is that the steady flow is gained if the acceleration of the gas above the riser is stabilized before reaching the choke valve (Jansen, Shoham et al. 1996). This increases the back pressure and the velocity at the choke thereupon. The mechanism is explained as a positive perturbation in the liquid holdup in a pipeline-riser system with a stable flow will increase weight and will cause the liquid to "fall down". The result of this is an increased pressure drop over the riser. The increased pressure drop will increase the gas flow and push the liquid back up the riser, resulting in more liquid at the top of the riser than prior to the perturbation. With a valve opening larger than a certain critical value (Z_{crit}) too much liquid will leave the system, resulting in a negative deviation in the liquid holdup that is larger than the original positive perturbation. Thus, we have an unstable situation where the oscillations grow, resulting in slug flow. For a valve opening less than the critical value Z_{crit}, the resulting decrease in the liquid holdup is smaller than the original perturbation, and we have a stable system that will return to its original, non-slugging state (Storkaas 2005).

2.5.2 Gas lift

Gas lift has been suggested as another method of eliminating severe slugging. In this method the hydrostatic head of the riser is reduced with gas injection and thus the pipeline pressure will be reduced. The injected gas lifts the liquid towards up the riser. If sufficient gas is injected the liquid will be continuously lifted and a steady flow will occur. The drawback of gas lift is the large gas volumes needed to obtain a satisfactory stability of the flow in the riser and this is too expensive (Storkaas 2005).

2.5.3 Slug catchers

One other way to accommodate slugging is common to install a large separator as a slug catcher at the exit of the pipeline. The slug catcher is the first element in the processing facility and determining its proper size is vital to the optimal operation of the entire facility. The fundamental purpose of slug catcher is to remove free gas from the liquid phase and to deliver a relatively even supply of liquid to the rest of the production facility. An advantage of this set-up is that inspection and maintenance on the slug catcher can be done without interrupting the normal operation. There are mainly two types of slug catchers, the vessel and the multiple-pipe types and the use of each type depends on the type of flow stream. Multiple-pipe separators have been widely applied in gas-condensate processing facilities (Miyoshi, Doty et al. 1988).

Installing slug catchers has several drawbacks; it puts a lower bound on the operating pressure of the pipe, which again limits the flow from the reservoir. It also increases the mechanical wear of the pipeline due to large oscillations in pressure. The capital and maintenance costs of a slug catcher are relatively large (Olsen 2006).

2.5.4 Active control

Riser slugging can be prevented using stabilizing feedback control. An approach based on feedback control was first proposed by Shmidt (Shmidt et.al. 1978). The idea of paper was to suppress terrain slugging by using the top-side choke valve and a simple feedback loop, measuring pressure at the inlet and upstream the riser or the top pressure before the choke valve as inputs. With feedback control, the stability of the flow regimes can be changed to enhance operation. In fact the boundaries can be moved via feedback control, thereby stabilizing a desirable flow regime where riser slugging “naturally” occurs (Storkaas 2005). Anti-slug control can move the boundaries in flow regime map resulting in increased stable region. It sounds to be one of the best solutions for prevention of severe-slugging. Several models have been suggested by researchers to describe the system dynamics and several controllers have been designed. The models are meant to aid tuning of controllers which use the production choke valve as the actuator and try to stabilize the system with a more production rate in a higher valve opening. The objective could be defined as obtaining the most robustness for the system against large inflow disturbances. “Nonlinearity” as the important characteristic of slugging system provides some challenges for control. However, a good control system using a model that is most consistent with the plant could have good results in achieving desired stable flow regimes.

2.6 Modeling of riser systems

The main objectives of modeling of production flow in pipelines and risers are to predict the pressure drop, the phase distributions, the potential for unsteady phase delivery (slugging) and the thermal characteristics of the system (Pickering, Hewitt et al. 2001). The reliability of these simplified models is however questionable. The analysis and modeling of multiphase flows relies heavily on empirical correlations and the predictions for the models are only as reliable as the empirical data on which they are based. Therefore it can be questioned whether the models would be valid if applied to real systems. They are tested by the use of small diameters experimental risers and may be more than good enough for such systems, but they still may be invalid for use in larger systems (Pickering, Hewitt et al. 2001).

The tuning methods used in this work are provided via linear and nonlinear multiphase flow models based on the mass balances over the different sections of the pipeline-riser system. The simplified four-state mechanistic model made by Jahanshahi and skogestad (Jahanshahi and Skogestad 2011) uses simple relationships to calculate the phase distributions over the different sections of system. The model has been then linearized around an unstable operating point and a fourth-order linear model with two unstable poles, two stable poles and two zeros is produced. Since a model with two unstable poles is enough for control design, the model order is reduced by using balanced model truncation via square root method. This identified model of the system is then used for an IMC (Internal Model Control) design and finding new IMC-based tuning rules. (Jahanshahi and Skogestad 2013). Moreover, a simple model for the static nonlinearity of the system is proposed by Jahanshahi and based on this static model, simple PI tuning rules considering nonlinearity of the system are given (Jahanshahi and Skogestad 2013). These tuning rules have been used in the simulations and experiments of this thesis and a clear comparison of the results have been presented.

2.7 Bifurcation diagrams

Bifurcation diagrams have been used in this thesis in order to plot the values of pressure versus the values of valve opening for the slugging system either in open-loop position or in closed-loop position with different controllers. Bifurcation diagrams are the simplest way to illustrate the stability of the system. In the stable regions the plot consists of a unit curve showing the exact value of the pressure (in simulations) or the average of very small pressure oscillations (in experiments) while in the unstable regions the plot consists of three curves, one for steady state conditions and the two others showing the maximum and minimum of oscillations at each value of valve opening over the work range of choke valve.

2.8 PID and PI controllers

PI (proportional-integral) and PID (Proportional-integral-derivative) control are of the earlier control strategies. The PID controller includes the proportional action (P), integral action (I), and derivative action (D). The controller uses the error signal $e(t)$ to generate the proportional, integral, and derivative actions. A mathematical description of the PID controller is:

$$u(t) = K_p[e(t) + \frac{1}{T_i}\int_0^t e(\tau)d(\tau) + T_d\,\frac{de(t)}{dt}] \qquad \text{Equation 2.1}$$

Where $u(t)$ is the input signal to the plant model. The error signal $e(t)$ is defined as $e(t) = r(t) - y(t)$ and $r(t)$ is the reference input signal (Fabre, Peresson et al. 1990). After a Laplace Transform the controller can be shown as:

$$c = K_c(1 + \frac{1}{\tau_I s} + \tau_d s) \qquad \text{Equation 2.2}$$

Where K_c, τ_I and τ_d are the respective tuning parameters for the P, I and D actions. PI and PID controllers are the most widely used controllers in the industry. However, they need to be tuned appropriately for robustness against plant changes and large inflow disturbances. (Jahanshahi and Skogestad 2013) Thus finding the most appropriate amounts of K_c, τ_I and τ_d could be extremely required. A typical structure of a PID control system is shown in Figure 2.9.

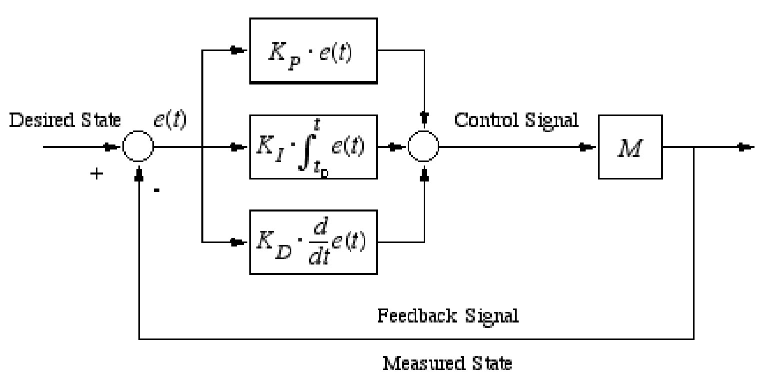

Figure 2.9: A typical PID control structure

2.9 Tuning of PID and PI controllers

Many tuning methods for different systems have been introduced so far by researchers and engineers. Depending on the characteristics of the system (plant), for instance nonlinearity and stability, different levels of robustness is achieved by different tuning methods. Three different tuning methods have been applied in this thesis. Two of them are quite new and have been recently developed (Jahanshahi and Skogestad 2013). They are specified for the slugging system. In fact, this thesis is a verification of these new methods.

2.9.1 Method 1: Shams's set-point overshoot method for closed-loop systems

Some systems like slugging system are originally unstable in open-loop. For these systems model from closed-loop response with P-controller can be used to find the appropriate tuning parameters. A method called "Shams's set-point overshoot method" was first constructed by Shamsuzzoha et al. (Shamsuzzoha and Skogestad 2010). Skogestad et al. (Skogestad and Grimholt 2011) developed this method further into a two-step closed-loop procedure. A step by step description of the two step closed-loop Shams's method is presented below.

The closed-loop system with P-controller should be at steady-state initially, that is, before the set-point change is applied. Then, a set-point change, Δy_s , is applied. The step change and the P controller gain (K_{c0}) should be adjusted in a way that the overshoot (D) is approximately 30 %. Figure 2.10 shows a graphical illustration and equation 2.4 finds the overshoot.

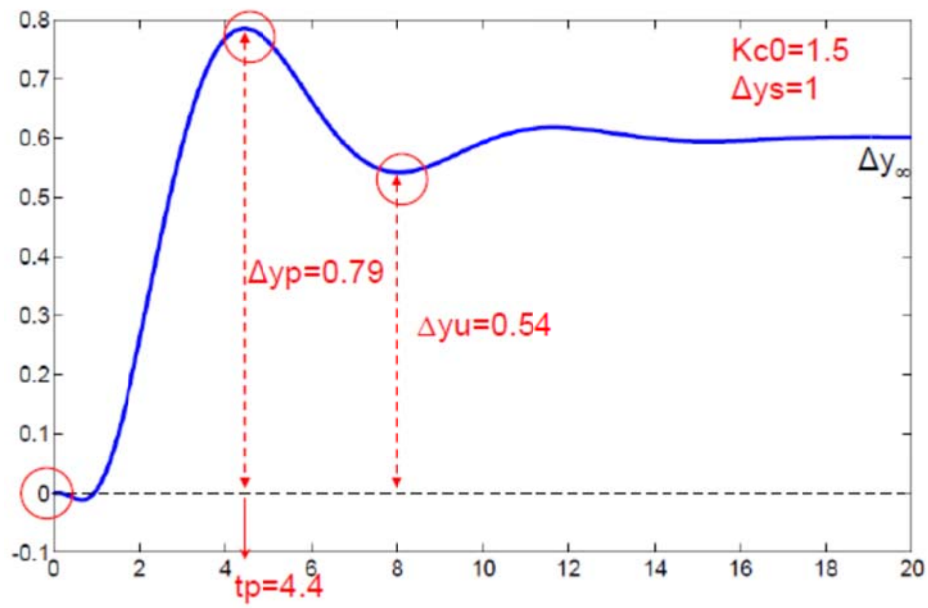

Figure 2.10: closed-loop set-point response with P-only controller (Skogestad and Grimholt 2011)

Extract information from the graphical step response:

- Time to first peak: t_p
- Maximum output change: Δy_p
- Relative steady state output change: Δy_∞
- Alternatively, Δy_∞ can be estimated from equation *2.6*, using the output change at first undershoot (Δy_u):

$$\Delta y_\infty = 0.45(\Delta y_p + \Delta y_u) \qquad \text{Equation 2.3}$$

- Overshoot:

$$D = \frac{\Delta y_p - \Delta y_\infty}{\Delta y_\infty} \qquad \text{Equation 2.4}$$

- Steady state offset:

$$B = \left| \frac{\Delta y_s - \Delta y_\infty}{\Delta y_\infty} \right| \qquad \text{Equation 2.5}$$

- The parameter (A):

$$A = 1.152D^2 - 1.607D + 1 \qquad \text{Equation 2.6}$$

- The parameter (r):

$$r = \frac{2A}{B} \qquad \text{Equation 2.7}$$

The first order plus delay model parameters:

- Steady state gain:

$$k = \frac{1}{K_{c0}B} \qquad \text{Equation 2.8}$$

- Delay:

$$\theta = t_p(0.309 + 0.209e^{-0.61r}) \qquad \text{Equation 2.9}$$

- Time constant:

$$\tau_1 = r\theta \qquad \text{Equation 2.10}$$

Now a first order plus delay model is found and with respect to this model, the tuning parameters are:

$$K_c = \frac{1}{k} \cdot \frac{1}{(\theta + \tau_c)} \qquad \text{Equation 2.11}$$

$$\tau_I = \min\left(\tau_1, 4(\tau_c + \theta)\right) \qquad \text{Equation 2.12}$$

In the paper by Skogestad (Skogestad 2003), it was recommended to use $\tau_c = \theta$ as a good compromise between performance and robustness.

2.9.2 Method 2: Tuning based on IMC design

The Internal Model Control (IMC) method was developed by Morari. *et.al*. (Morari and Zafiriou 1989) The method supposes a model, states desirable control objectives, and, from these, proceeds in a direct manner to obtain both the appropriate controller structure and parameters. For the objectives and simple models common to chemical process control, the IMC design procedure leads naturally to PID-type controllers, occasionally augmented by a first-order lag. (Rivera, Morari et al. 1986)

Consider the block diagram for the IMC structure *(See figure 2.11)*. Here, $\tilde{g}$ is model of the plant that in general has some mismatch with the plant. g_c is inverse of minimum phase part of $\tilde{g}$ and f (s) is a low-pass filter for robustness of the closed-loop system.

The goal of control system design is fast and accurate set-point tracking:

$$y \cong y_s \quad \forall t \;, \forall d \qquad \text{Equation 2.13}$$

Efficient disturbance rejection:

$$y' \cong y_s - d \quad \forall t \;, \forall d \qquad \text{Equation 2.14}$$

and insensitivity to modeling error.

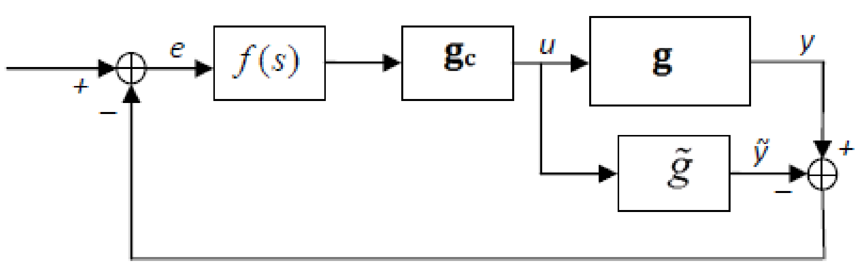

Figure 2.11: The internal model control (IMC) structure

Jahanshahi and Skogestad do not use this configuration for the unstable system; instead they use an equivalent as shown in figure 2.14, where:

$$C = \frac{g_c f}{1 - \tilde{g} g_c f} \qquad \text{Equation 2.15}$$

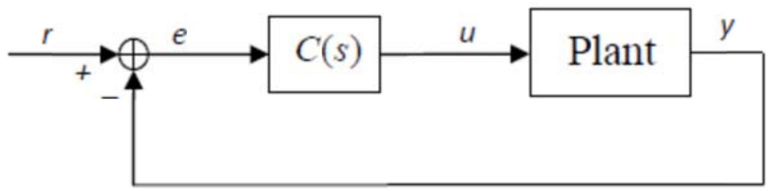

Figure 2.12: Closed-loop system with IMC controller (Jahanshahi and Skogestad 2013)

They propose online identification of linear model by a closed-loop step test. They design an IMC (Internal Model Control) based on the identified model. Then, they use the resulting IMC controller to obtain tuning parameters for PID and PI controllers. A summary of their work, which this thesis has been done based on that, will be presented below.

2.9.2.1 Model Identification

To identify process model ($\tilde{g}$), Jahanshahi and Skogestad use the step test information in a closed-loop stable system to do online model identification. The suggested model has two unstable poles and is in the form of:

$$\tilde{g}(s) = \frac{b_1 s + b_0}{s^2 + a_1 s + a_0} \qquad \text{Equation 2.16}$$

Four parameters, b_1, b_0, a_1 and a_0 need to be estimated by information extracted from closed-loop step response. Jahanshahi uses a systematic manner to find the related four parameters. In his method the loop is closed by a proportional controller with gain K_{C0}, to get the closed-loop stable system. For closed-loop transfer function from the set-point to the output one similar to the model used by Yuwana. *et.al*. (Yuwana and Seborg 1982) is considered:

$$G_{cl}(s) = \frac{K_2 (1 + \tau_z s)}{\tau^2 s^2 + 2\zeta \tau s + 1} \qquad \text{Equation 2.17}$$

The four parameters (K_2, τ_z, τ and ζ) are estimated by using six data ($\Delta y_p, \Delta y_u, \Delta y_\infty, \Delta y_s, t_p$, and Δt) observed from the closed-loop response (*see figure 2.10*). Then, they use a systematic procedure to back-calculate the parameters of the open-loop unstable model in equation 2.16 (Jahanshahi and Skogestad 2013).

2.9.2.2 IMC design for unstable systems

To design the IMC controller (C), the identified model ($\tilde{g}$) is used as the plant model.

$$\tilde{g}(s) = \frac{b_1 s + b_0}{s^2 + a_1 s + a_0} = \frac{k'(s+\varphi)}{(s-\pi_1)(s-\pi_2)} \qquad \text{Equation 2.18}$$

$$g_c(s) = \frac{(1/k')(s-\pi_1)(s-\pi_2)}{(s+\varphi)} \qquad \text{Equation 2.19}$$

They also design the filter $f(s)$ for robustness of the system as explained by Morari. *et.al*. (Morari and Zafiriou 1989). The filter is in the following form:

$$f(s) = \frac{\alpha_2 s^2 + \alpha_1 s + 1}{(\lambda s + 1)^n} \qquad \text{Equation 2.20}$$

λ is an adjustable filter time-constant. The coefficients α_1 and α_2 are calculated by solving the following system of linear equations:

$$\begin{bmatrix} \pi_1 & \pi_1^2 \\ \pi_2 & \pi_2^2 \end{bmatrix} \times \begin{bmatrix} \alpha_1 \\ \alpha_2 \end{bmatrix} = \begin{bmatrix} (\lambda\pi_1 + 1)^3 - 1 \\ (\lambda\pi_2 + 1)^3 - 1 \end{bmatrix} \qquad \text{Equation 2.21}$$

Filter only acts to the derivative action.

Finally the resulting IMC controller is found as the following:

$$C(s) = \frac{\left[\frac{1}{k'\lambda^3}\right](\alpha_2 s^2 + \alpha_1 s + 1)}{s(s+\varphi)} \qquad \text{Equation 2.22}$$

2.9.2.3 PID and PI tuning based on IMC controller

Jahanshahi writes the IMC controller of equation 2.22 in form of a PIDF controller and propose the tuning parameters based on that. PIDF is a PID controller which a low-pass filter has been applied on its derivative action.

$$K_{PID}(s) = K_p + \frac{K_i}{s} + \frac{K_d s}{\tau_f s + 1}$$ Equation 2.23

Where the tuning parameters are:

$$\tau_f = 1/\varphi$$ Equation 2.24

$$K_i = \frac{\tau_f}{k'\lambda^3}$$ Equation 2.25

$$K_p = K_i\alpha_1 - K_i\tau_f$$ Equation 2.26

$$K_d = K_i\alpha_2 - K_p\tau_f$$ Equation 2.27

An important point to be considered in tuning of PI/PID controllers based on IMC design is choosing an appropriate λ. It must be chosen in a way that the required following conditions are satisfied:

$$K_p \prec 0$$ Equation 2.28

Equation 2.29

$$K_d \prec 0$$

A PI controller has been also obtained by reducing the order of IMC controller to 1.

$$K_{PI}(s) = K_c(1+\frac{1}{\tau_I s}) \qquad \text{Equation 2.30}$$

And the suggested tuning rules are:

$$K_c = \frac{\alpha_2}{k'\lambda^3} \qquad \text{Equation 2.31}$$

$$\tau_I = \alpha_2 \varphi \qquad \text{Equation 2.32}$$

2.9.3 Method 3: Simple online PI tuning method with gain scheduling

One main part of the thesis is tuning the controller by a new method called "Simple online PI tuning rules" proposed by Jahanshahi and Skogestad (Jahanshahi and Skogestad 2013). One advantage of this method is that Nonlinearity of the slugging system has been considered when providing the tuning rules. Gain of the slugging system changes drastically for different operating conditions and as the source of nonlinearity, makes control of the system difficult. The method consists two parts:

First, a simple MATLAB static model for the static nonlinear gain is identified at each operating point (valve opening).

Then, the identified model at each operating point is used and simple PI tuning rules based on single step test but with gain correction to counteract nonlinearity of the system are proposed as functions of valve opening.

In this method of tuning, Jahanshahi and Skogestad have used gain-scheduling with multiple controllers based on multiple identified models. The MATLAB model and the obtained PI tuning rules for each controller will be explained below.

2.9.3.1 Simple MATLAB static model

The simple model for an L-shaped riser considering static nonlinearity was made by Jahanshahi (Jahanshahi and Skogestad 2013). The model is based on the mass balances and it calculates the phase distributions over the different sections. This model needed to be modified for an S-shaped riser to be used in the thesis.

A good assumption of valve equation is very important in using the simple model. The reason is that the slugging gain of the system as a function of valve opening, is derived based on this equation. Jahanshahi assumes the valve equation as the following:

$$w = K_{pc} f(z)\sqrt{\rho \Delta p} \quad \text{Equation 2.33}$$

Where w is the inlet mass flow rate to the riser, K_{pc} is the valve constant and $f(z)$ is the characteristics of the valve which is defined as the following for the linear valve used in experiments:

$$f(z) = z \quad \text{Equation 2.34}$$

and as follows for the OLGA valve model in simulations:

$$f(z) = \frac{z.cd}{\sqrt{1 - z^2.cd^2}} \quad \text{Equation 2.35}$$

Δp is the pressure drop over the valve and as it is clear in the valve equation, it's a function of valve opening that can be written in the following form:

$$\Delta p = \frac{1}{\bar{\rho}}\left(\frac{\bar{w}}{K_{pc}.f(z)}\right)^2 \quad \text{Equation 2.36}$$

Then the simple model for the inlet pressure is:

$$P_{in} = \Delta p + \bar{P}_{fo} \quad \text{Equation 2.37}$$

$\bar{P}_{fo}$ is the inlet pressure at fully open position of the valve and has been calculated from the below equation:

$$\bar{P}_{fo} = P^* - \Delta p^* \qquad \text{Equation 2.38}$$

P^* is a large enough inlet pressure to overcome the riser slugging:

$$P^* = \rho_L.g.L_r + P_s + P_{v,\min} \qquad \text{Equation 2.39}$$

Here ρ_L is the density of liquid which is water in our system. g is the gravity and L_r is the length of riser. P_s is the separator pressure in downstream and $P_{v,\min}$ is the minimum pressured drop over the valve and has been considered zero in the simulations.

Δp^* is the pressure drop over the valve at the critical valve opening of the system (bifurcation point).

Then based on the above equations, the static gain of the slugging system is derived as a function of valve opening by differentiating P_{in} with respect to *Z*. Finally the simple model for the static gain of the system is:

$$k(z) = \frac{\partial P_{in}}{\partial z} \qquad \text{Equation 2.40}$$

2.9.3.2 Simple PI tuning rules based on identified MATLAB model

Jahanshahi and Skogestad (Jahanshahi and Skogestad 2013) then perform a closed-loop step test with a P-only controller at the initial valve position of Z_0. The parameter (β) is then calculated by using data ($\Delta y_p, \Delta y_u, \Delta y_\infty$, t_p, and Δt) observed from the closed-loop response *(see figure 2.10)* and the static model given in equation 2.40.

$$\beta = \frac{-\ln\left(\dfrac{\Delta y_\infty - \Delta y_u}{\Delta y_p - \Delta y_\infty}\right)}{2\Delta t} + \frac{K_{c0}k(z_0)\left(\dfrac{\Delta y_p - \Delta y_\infty}{\Delta y_\infty}\right)}{4t_p} \qquad \text{Equation 2.41}$$

Where K_{c0} is the proportional gain used for the step test. The suggested PI tuning parameters as functions of valve opening are given as the following:

$$K_c(z_0) = \frac{\beta T_{osc}}{k(z_0)\sqrt{z_0 / z^*}} \qquad \text{Equation 2.42}$$

$$\tau_I\left(z_0\right) = 3T_{osc}(z_0 / z^*) \qquad \text{Equation 2.43}$$

T_{osc} is the period of slugging oscillations when the system is in open-loop position and z^* is the critical valve opening of the open-loop system (where slugging starts).

3 Experimental work

Control of Severe Slugging and creating a stable flow regime by applying control using new online tuning methods has been verified in this thesis. Air-water Sever slugging control experiments in S-shaped riser has been one of the main parts of this thesis in addition to modeling and simulations. A series of tests have been conducted at a medium scale setup located in NTNU multiphase flow laboratory at department of Energy and Process Engineering *(See figure 3.1).* It has been tried to evaluate the applicability of three tuning methods explained previously in different conditions. Experiments in this issue and comparing them with simulated results are also valuable in the way of approving prediction of simulations.

The experimental work include trying two different choke valves with different dynamics as the actuator and running series of control experiments for each valve separately. Series of control experiments have been in the following order: First the open-loop experiments have been run in order to make the open-loop bifurcation diagram of the system. Then a P-only controller has been used to close the loop and the set-point step change test has been run with the aim of finding appropriate tuning parameters. Finally, after calculating different tuning rules based on the data of step change test, closed-loop experiments were run and the closed-loop responses of different controllers tuned with different methods were evaluated. Buffer tank pressure (riser inlet pressure in the real systems) has been selected as the control variable (CV) in series of control experiments.

Moreover cascade control experiment using topside pressure combined with outflow density as the control variables has been tried.

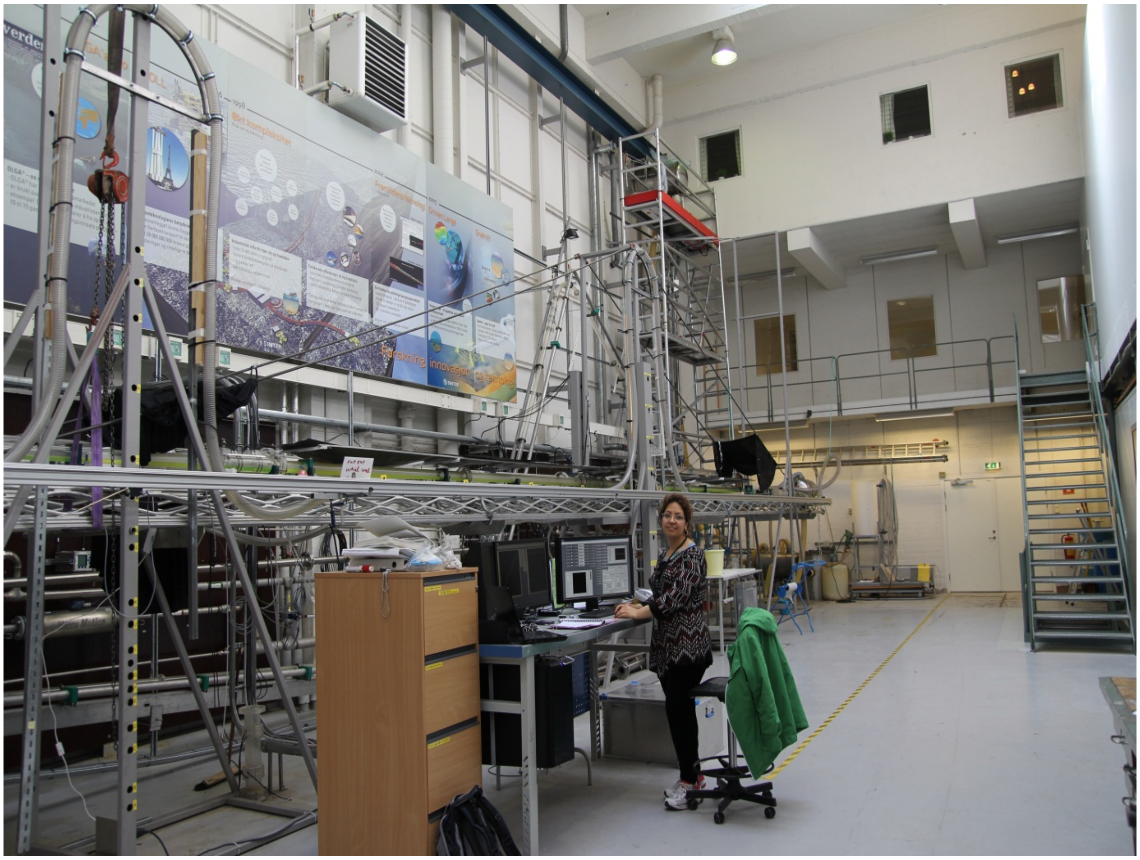

Figure 3.1: Medium scale experimental setup of multiphase flow laboratory located at department of Energy and Process Engineering of NTNU

3.1 Setup Description

The three-dimensional overview of the multiphase flow rig used to perform the series of experiments in this thesis is shown in figure 3.2. The flow loop was consisting of water and compressed air supply.

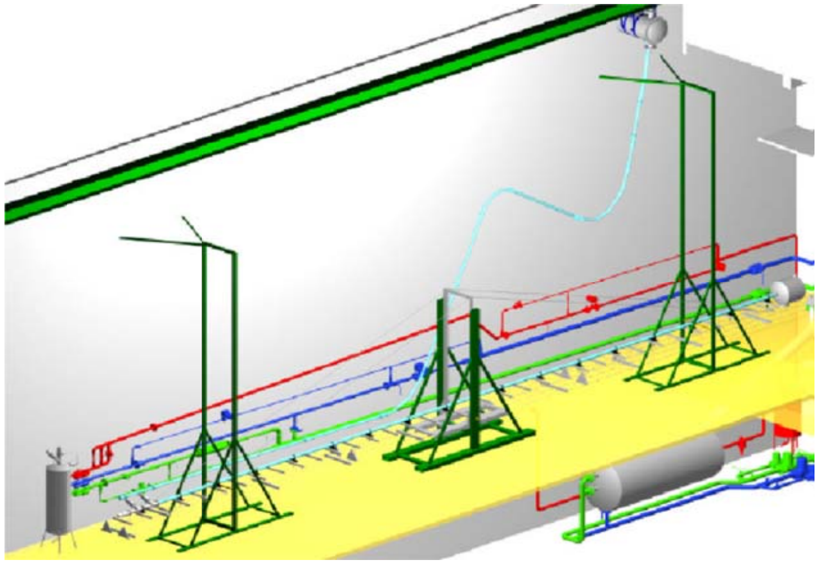

Figure 3.2: Multiphase Test Rig Layout, NTNU (Lilleby 2003)

Figure 3.3 shows a schematic overview of the experimental setup with more details. The whole system is placed at two levels. Large storage water and pressurized air tanks (T1 and T2) and water pump (P1) were placed at basement. Flow lines continued to lab-level and all flow meters, control valves, horizontal test section and S-riser were placed at this level. The flow line of test with inner diameter of 50mm was connected to a mixer/inlet section containing the air/water supply and the multiphase flow was forced up the S-riser. The air buffer tank (T3) was installed upstream the mixing point to increase the air volume and emulate a long pipeline. The air volume should be large enough to force the liquid up the riser and cause slugging to occur.

As one of the most important equipment, choke valve (V) was mounted at top of the S-riser. It was used as the control actuator for controlling the inlet pressure/ top pressure and outlet flow density as the control variables. It was also possible to adjust it manually while running the system in open-loop position. Pressure transmitters (PT1 and PT2) and the conductance probe as the density meter (C) were installed at various places in the setup, and were used to construct a number of different control structures.

After the S-riser, air and water were entered into an overflow tank (T4), then moved into a small separator (T5) through a large flexible pipe made of hoses, and were separated there. The water is then returned from the test section back to the water large storage tank in the basement. The air is vented out without further treatment.

The dimensions of the experimental setup are illustrated in figure 3.4. The length scale is given in meters.

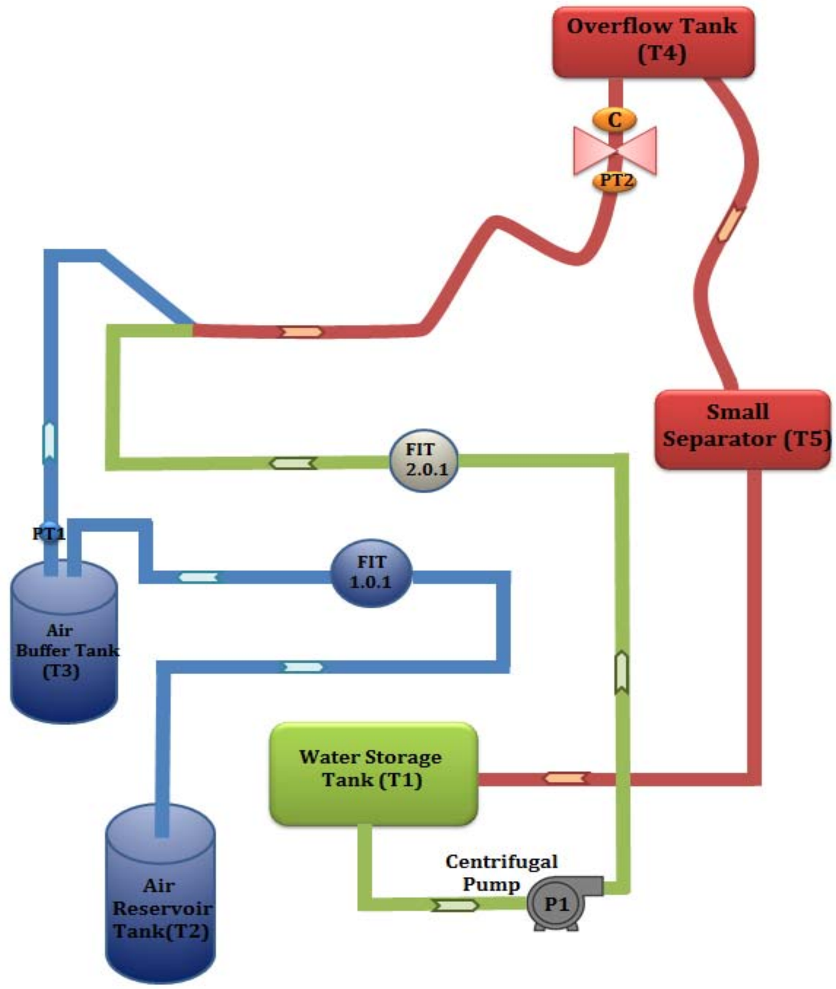

Figure 3.3: Medium scale Test Rig Layout with more details, NTNU

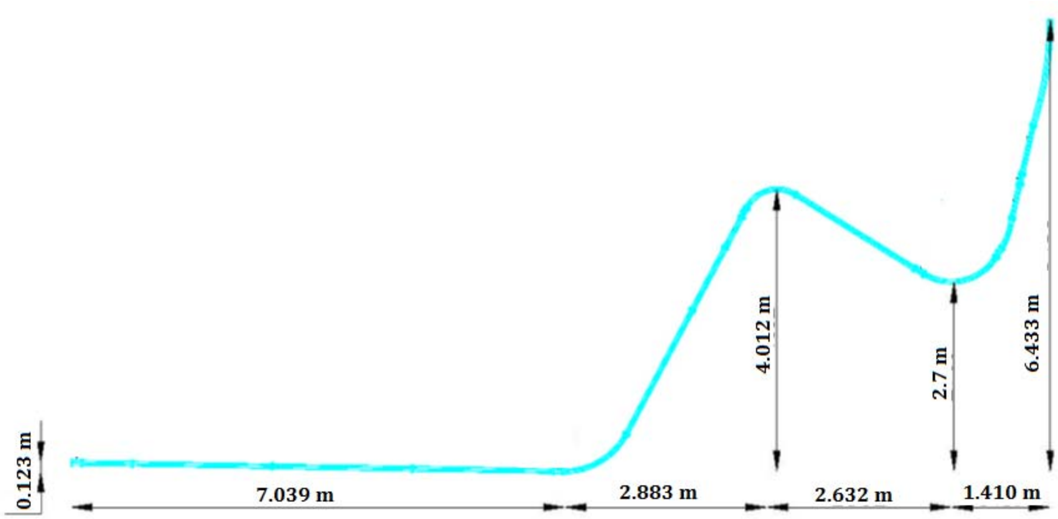

Figure 3.4: Configuration of the S-shaped riser test section (Lilleby 2003)

3.2 Equipment

In this section properties and purpose of the main equipment are given. All the pipes, bends and other connections are made of acid-proof steel, AISI 316L. This is the case for the entire piping up to the test sections. The valves are made of treated brass, and are quite resistant to corrosion.

3.2.1 Main water storage tank

Water is filled in a separator (T1). It is a $3\,m^3$ acid proof tank placed in the basement. From the separator, water is pumped through the infrastructure, into the test section and returned to the separator again.

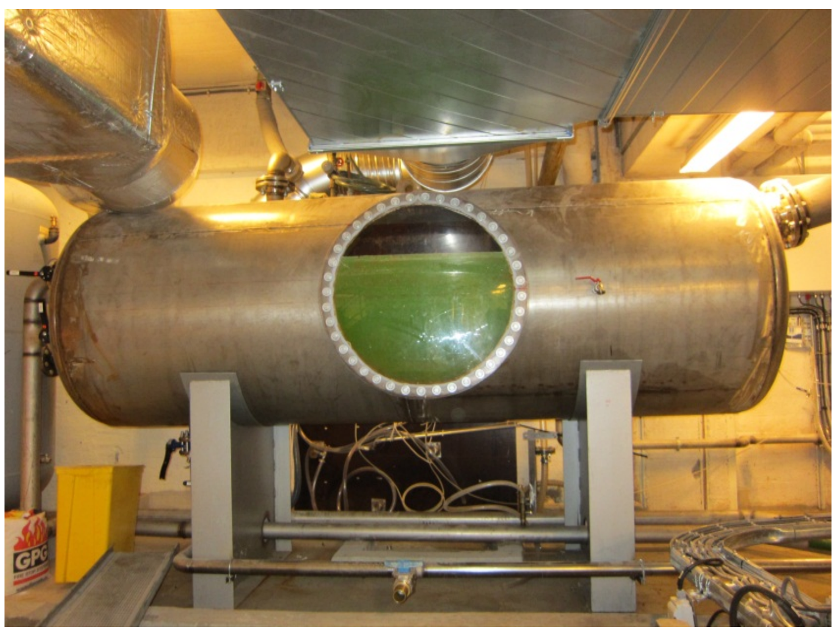

Figure 3.5: Main water storage tank located in basement

3.2.2 Air reservoir tank

The air supply (T2) is connected to the central high-pressure supply. This supply is a pressure vessel made by *Nessco* and gives a pressure of 6-7 bars, which is then reduced through a pressure reduction valve to the operational pressure of (usually) approximately 3 bars.

Figure 3.6: Air reservoir tank located in basement

3.2.3 Air buffer tank

The air buffer tank (T3) with a volume of 200 liters and the type “DN50 flange” has been made by the company “Laguna”. It is installed before the mixing point. To make slugging possible, a large pipe volume for pressure buildup is necessary. The buffer tank is used to emulate this large pipe volume. The maximum pressure the buffer tank can withstand is limited. For safety, the tank has been equipped with a safety valve, to ensure that the pressure not will exceed 3 Bars.

Figure 3.7: Air buffer tank

3.2.4 Overflow tank

An overflow system is made to achieve pressure dependent liquid flow. It is a vented steel tank (T4) filled with water. Flexible pipes connect the tank to the separator. A bypass flow will flow into the tank and back to the separator and maintain a constant liquid level inside the tank. The pressure at the overflow tank will be constant equal to the hydrostatic pressure of the liquid column from the tank. This will simulate a constant reservoir pressure and make the inflow to the test section dependent on the inlet pressure. The supply pipes for the plastic overflow tank are small, so it will only work properly if the flow through it is very low.

Figure 3.8: Over flow tank at top of riser

3.2.5 Pressure transmitters

Pressure transducers (PT1 and PT2) made by *Siemens* were installed on the buffer tank and riser to measure the buffer pressure and top pressure respectively. They have a working range of *0-4 bars*.

3.2.6 Small separator

The flow from the overflow tank (T5) is moved into a small separator located down the hoses pipe. A picture of the separator is shown underneath in figure 3.9*.* The air from the riser is released from the top outlet. The bottom outlet is used for the water recycle and returns the water to the water storage tank.

Figure 3.9: Small separator

3.2.7 Centrifugal water pump

A large centrifugal water pump (P1) of the type *DN100 flange* made by *Wilo Norge AS* was used to push the water into the system. In order to prevent water flow oscillations the centrifugal water pump was run in a very high level of power (80% of the maximum). However to get the desired flow rate of water which was not high (0.39 kg/sec) the water control valve was open in small values, instead.

Figure 3.10: Centrifugal water pump

3.2.8 Air flow meter

The vortex flow meter of type *DN40 wafer* manufactured by *JF Industrisensorer* was used to measure the air flow rate (FIT1.01). The number that it gave was in the unit of *Kg/hour* and needed to be converted into the desired unit (kg/sec). It was located upstream the air buffer tank. The working range of the air flow meter was *5-2180 kg/h.*

3.2.9 Water flow meter

The Electro-magnetic water flow meter of type *1/2'' union*, manufactured by *JF Industrisensorer* was located upstream of the mixing point (FIT2.01). It has a working range of *0.19-6.4* m^3/h.

Figure 3.11: Air flow meter

Figure 3.12: Water flow meter

3.2.10 Choke valves

Two different choke valves (V) have been used in this thesis and the series of experiments have been run with both. First a *slow valve* was used as the actuator to run the control experiments and then it was replaced with a *fast valve*. The effect of their dynamics was then investigated. They are angle seat valves located on the top of the riser upstream of the separator. The choke valve is operated by pressurized air (*4 bars*) supplied from the pressurized air system in the laboratory, through the valve positioner. The specifications of the old slow valve were not available, while the specifications of the fast valve are as follows:

Manufacturer: ASCO
Material: Stainless Steel
Pilot Pressure: 4-10 bar
Operator Diameter: 90 mm
Opening Time: 2 sec

Diameter: 2 inch
Operation: NC (Normally Closed)
Maximum Working Pressure: 6 bar
Signal: 4-20 mAmp
Closing Time: 2.5 sec

Figure 3.13: Choke valves; left: Fast valve, Right: Slow valve and its positioner

3.2.11 Conductance probe (C)

In the second series of experiments with the fast valve a cascade control structure was used with outflow density and the top pressure as the control variables. Conductance probe was applied to measure the density of the outflow from the riser. The probe has been calibrated by Kazemihatami (Kazemihatami 2012) very recently. The output of the probe was in the form of voltage. The calibration curve presented by Kazemihatami was used to find the relation between voltage and holdup. Equation *3.1* shows this relation. *H* means holdup and *V* means voltage.

$$H = 0.9857V \qquad \text{Equation 3.1}$$

The density of mixed flow is found from the equation *3.2*:

$$\rho_m = \rho_{water}.H + \rho_{air}.(1 - H) \qquad \text{Equation 3.2}$$

After inserting the related values in the above equation, the density of mixed flow is found as a function of voltage:

$$\rho_m = 984.513V + 1.204 \qquad \text{Equation 3.3}$$

Figure 3.14: The conductance probe

3.2.12 LabVIEW

The Laboratory Virtual Instrumentation Engineering Workbench (LabVIEW) software developed by National Instruments was used for instrumentation control and data logging. The user interface is illustrated in figure *3.15*. The pressures, flow rates and valve position could be monitored directly from the interface. In addition it was possible to run the loop manually by manipulating choke valve opening, or automatically by setting tuning parameters for PID/PI/P controllers. Some modifications were applied in case of control. Two modes of control were implemented in the program; a single mode and a cascade mode. The single mode used buffer pressure as control variable and the cascade mode was using top pressure and outflow density as control variables. A schematic view of control modes are presented in figure *3.16.*

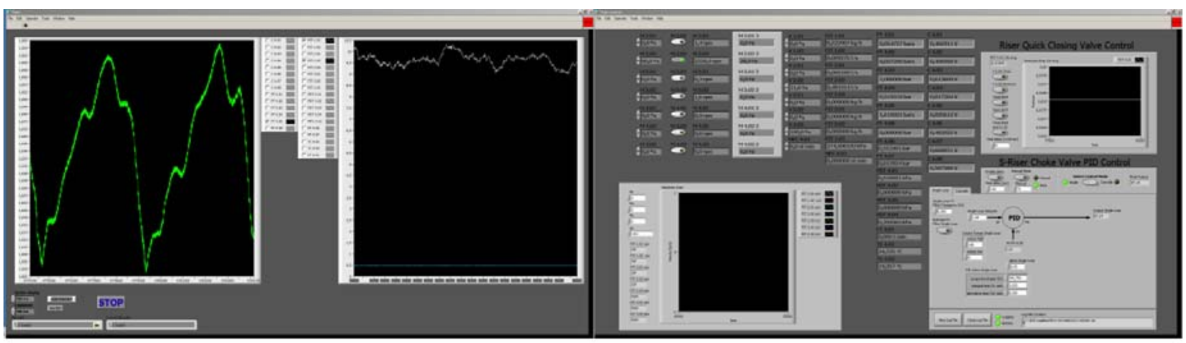

Figure 3.15: LabVIEW user interface

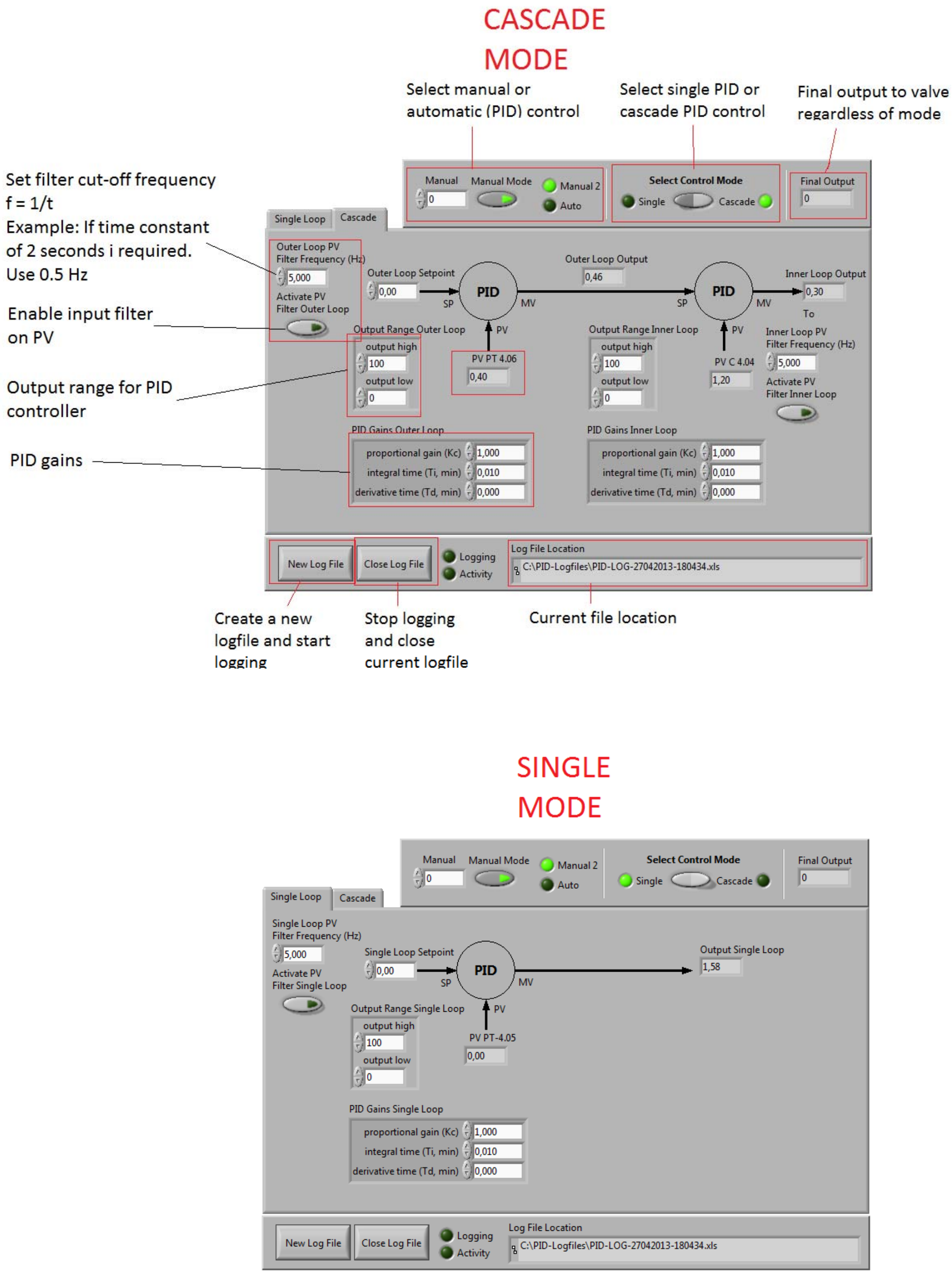

Figure 3.16: Implemented control modes in LabVIEW

4 Simulation of experimental cases

4.1 OLGA®, multiphase simulation tool

OLGA® (OiL and GAs simulator) is a commercial multiphase flow simulator widely used in the oil and gas industry. It solves many numerical equations to simulate the flow by considering the system dynamics and offers heat and mass transfer models.

The experimental case was constructed in OLGA. The designed controllers with different tuning strategies were used and the results were compared. In order to fit the OLGA model with the MATLAB models and experiments some of the parameters were manipulated within limited ranges. OLGA® version 7.1 was used for the simulations.

In this chapter the case construction with implementing the S-shaped riser geometry, fluid properties, numerical settings and boundary conditions is explained stepwise.

4.2 Construction of the case

Establishment of a good case with appropriate particular items such as fluid properties, numerical settings, initial and boundary conditions and flow path geometry, was the initial step for simulation process. The "S-riser simple" case made by Jahanshahi (Jahanshahi and Skogestad 2011) was basically used for the open-loop simulations. Some improvements and modifications were applied after the file was received. For open-loop simulations the modifications were in terms of numeric and for the closed-loop simulations they were related to implementing the PID controller into the case. In terms of numeric some Integration parameters were manipulated in Properties window of the program.

4.2.1 Flow path geometry

The “S-riser simple case” with a geometry based on the experimental set-up at the Department of Energy and Process Engineering was used. The reason to use such geometry is that the simulation results are to be compared with the experimental results in the thesis. The exact geometry is presented in table 4.1.

The *X*-*Y* coordinates have been calculated with respect to table 4.1and the resulting geometry has an overview of the figure 4.1.

According to the experimental setup in multiphase flow laboratory, the sources of air and water are placed in the beginning and the end of the buffer tank respectively.

Table 4.1: The geometry of the S-riser experimental set-up

Pipe	L [m]	D[m]	θ [$\circ$]	θ_{out} [$\circ$]	θ_{in} [$\circ$]
1	8.125	0.20	-45.0		
2	3.000	0.05	-10		
3	6.050	0.05	-4.0		
4	1.200	0.05	-1.8		
5	1.106	0.05		-1.8	-61.8
6	4.110	0.05	61.8		
7	0.709	0.05		61.8	-32.0
8	2.160	0.05	-32		
9	1.716	0.05		-32	79.0
10	1.820	0.05	79.0		
11	1.150	0.05	90		

Figure 4.1: Geometry of S-riser in OLGA

4.2.2 Fluid properties

All fluid properties had been written in PVT file by (Jahanshahi and Nilsen 2012). It is a table of phase compositions at different temperatures and pressures and is made by a program called PVT-Sim. By specifying temperature and pressure limits and the compositions of the fluids involved, the program calculates the values for the phase compositions. Heat transfer and temperature change were not important in simulations due to experimental condition. Water was assumed as an incompressible flow. Heat transfer and temperature related properties such as enthalpy or entropy were filled with dummy numbers.

4.2.3 Boundary and initial conditions

The types of the air and water sources as inlet nods were defined as inlet mass flow. The flow rates were fixed for all simulations. The volume fractions were established to 1 for both nodes since only water or air was injecting through the node. The outlet nod type was selected to pressure type and it has been set to atmospheric pressure.

4.2.4 Numerical setting

The numerical setting specifications such as simulation time and time step were adjusted in different numbers from case to case. This is due to the diversity of phase velocity in different cases.

4.3 Implementing PID controller in OLGA

In order to implement a PID controller in OLGA first a positive check valve was placed right after the water source in pipe 2, section 1 of the case. The reason was to make sure that the flow will move only in the defined direction. Then a pressure transmitter was located in pipe 2, section 2 that is the inlet of the riser, right after the buffer tank. It was aimed to measure the buffer pressure and send the pressure signal into the PID controller. The PID controller was used in a way that it received the measurement signal from the pressure transmitter and sent the output signal into the choke valve located at top of the riser (Pipe 8, section 3). Choke valves can be simulated by selecting the *Hydrovalve* for the valve model in OLGA.

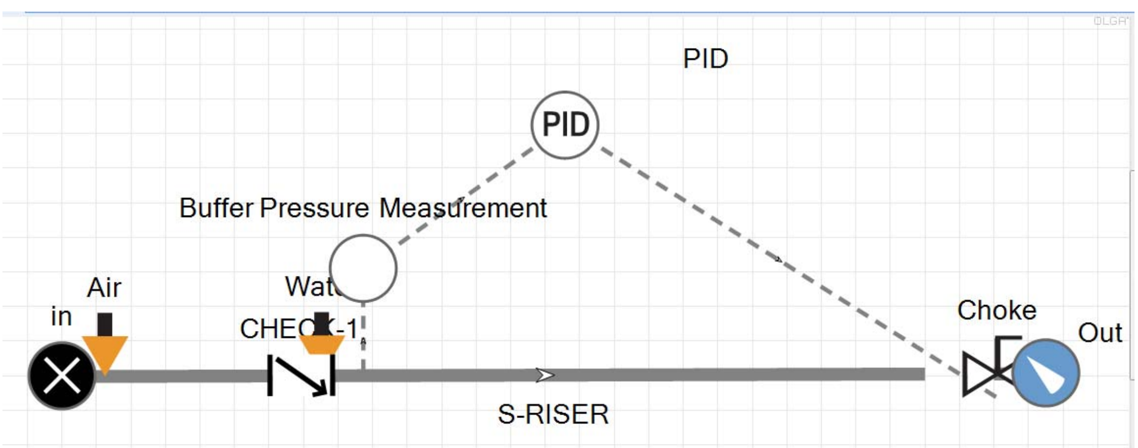

Figure 4.2: OLGA case with PID controller. The controller receives the measurement signal from the pressure transmitter and sends the output signal into the choke valve located at top of the riser.

When applying a PID controller in OLGA several specifications need to be established by user, depending on the desired conditions and results. The more important specifications that have been manipulated many times during simulations are the PID parameters and the time varying specifications. When it comes to PID parameters in property window of the simulator, *AMPLIFICATION* refers to the gain of the controller; *BIAS* is the desired initial output value (it was used as the desired valve opening in our simulations); DERIVATIVECONST is the time constant for the derivative action and INTEGRALCONST is the time constant for the integral action. As the time varying specifications the MODE was set to AUTOMATIC and the SET-POINT values were changed from one simulation to another.

5 Results and discussion

The purpose of this chapter is to present the results from experiments and simulations and a clear comparison of them. The experimental results from two series of experiments using a slow and a fast choke valve will be presented in section 5.1. The effort of cascade control experiment using top pressure combined with density as measurements and the faced issues has been also mentioned there. Section 5.2 evaluates the effect of control valve dynamics through comparing results of slow valve with those of fast valve. The simulated results will be explained in section 5.3. In section 5.4 the experimental results are compared with simulated results. In section 5.5 the three different used tuning methods have been compared and the best tuning method has been investigated.

5.1 Experimental results

The operating procedures and the results from experimental activities done at NTNU multiphase flow laboratory are discussed in this section.

For each series of experiments with valve 1 (slow valve) or valve 2 (fast valve) the open-loop system with basic conditions would be explained first. Then the procedure of implementing closed-loop step test and calculating the tuning parameters by using different tuning methods will be discussed. The results of tuning in the form of tuning rules are explained thereafter. Finally the closed-loop responses using calculated tuning parameters will be presented as the main results of the experimental work.

5.1.1 Series of experiments with valve1 (slow choke valve)

The experimental work in this thesis started with using slow choke valve as the actuator. The goal was to repeat the same series of tests with a slow and a fast choke valve and then evaluate the effect of control dynamics on the final results.

5.1.1.1 *Open-loop experiments*

The starting point in the experiments was running the loop in manual mode. The tests were run in different valve openings with fixed liquid and gas flow rates while no controller was implemented in the system. It was aimed to present the system behavior in natural conditions without control. The inflow conditions and the related bifurcation diagram are presented below.

5.1.1.1.1 *Inflow conditions*

The applied fixed flow rates have been $w_l = 0.3927\ [kg / sec]$ for water and $w_g = 0.0024\ [kg / sec]$ for air *(See figure 5.1.)* These flow rates correspond to $U_{sl} = 0.2$ $[m / sec]$ and $Usg = 1\ [m / sec]$ as the liquid and gas superficial velocities. The water flow rate could be set in lab view by adjusting the pump frequency and the control valve, while the air flow rate needed to be set with a manual valve in the path of the flow. The reason was that the control valve for the air was broken. The manual valve was far from the screen and this made it difficult to obtain the exact flow rate.

The water flow rate was not also easy to set. Large variations in the flow rate were eliminated by running the pump with a high frequency and opening the control valve in a small value. The more opening the choke valve, the more slugging the flow regime and the more unstable the flow rates were resulted. In the following series of experiments, a constant flow rate of air and water was used. As a result, the water and air flow rates needed to be readjusted when the valve opening in open-loop was changed. However, when using a controller in closed-loop mode, it was considered not to be reasonable to readjust the inflow conditions. Figure 5.1 compares variations of flow rates and pressure in two different valve openings.

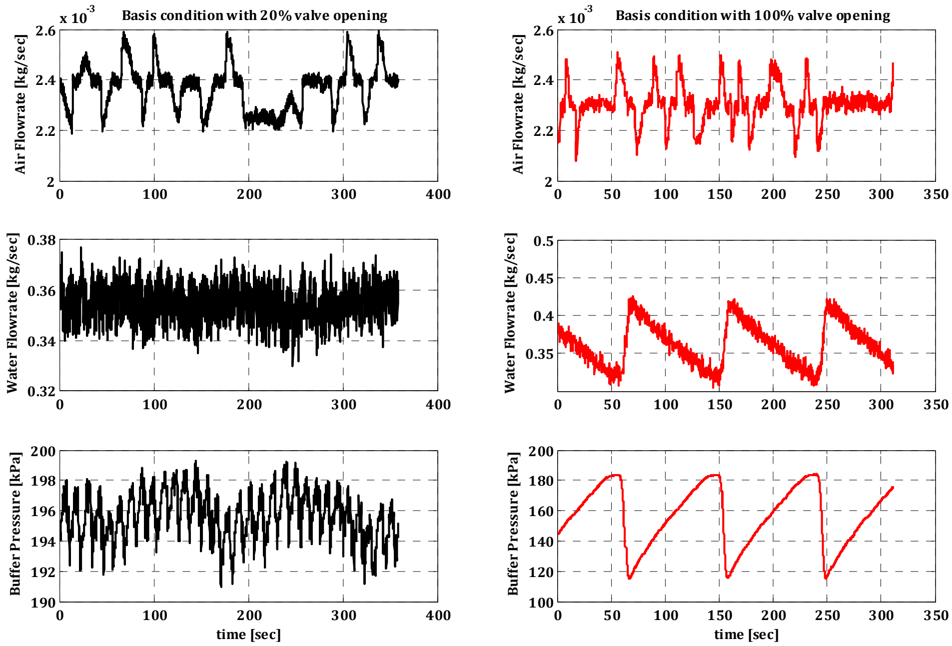

Figure 5.1: Illustration of basis open-loop conditions in case of flow rates and pressure. The left series of plots are illustrating the system with valve opening Z=0.2 that is related to the stable region while the right side plots present the system with valve opening Z=1 that is related to the unstable region. Large oscillations are clear signs of instability at Z=1.

5.1.1.1.2 Bifurcation diagram

The experiments were started with the valve opening of Z=0.2. Then the valve was open stepwise until it was fully open. The results of buffer pressure were logged and the related bifurcation diagram was plotted, presented in Figure 5.2. The critical stability point (the bifurcation point) is the maximum choke valve opening the system can have while being stable. In the presented bifurcation diagram, the top line tracks the maximum values of pressure at each operating point, the bottom line presents the minimum values of pressure and the middle line shows the average values of the buffer pressure at different valve openings. As clear in the figure the critical stability point was found to be at approximately 26% choke valve opening (Z= 0.26).

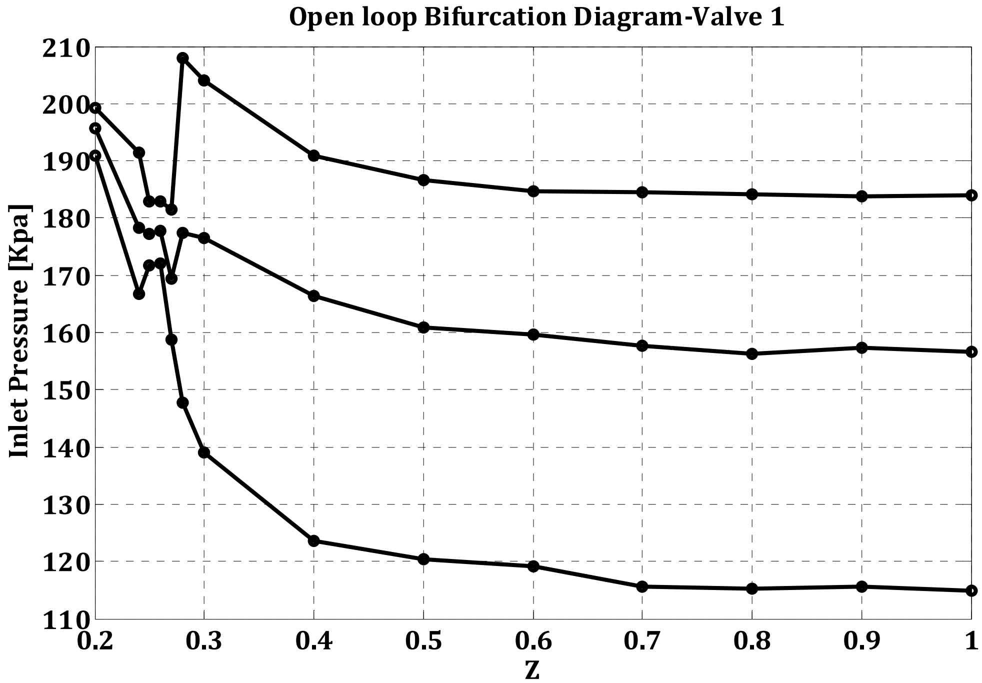

Figure 5.2: Open-loop bifurcation diagram from the slow choke valve experiments. The bifurcation point occurs at valve opening of Z=0.26. The top and bottom line illustrate the maximum and minimum values of oscillations for inlet pressure respectively at each operating point. The middle line shows the average values of pressure.

5.1.1.2 Closed-loop step test

In order to apply each of tuning methods to get an appropriate controller for the slugging system a closed-loop step test is required with a step change in set-point (the buffer pressure). To do this it was tried to control the system by trial and error. A P-only controller was selected and as the initial guess for the gain, a big value of 100 was tried. The reason was that the set-point value was a small number (pressure in bars) and the gain had to be selected in a way that it could change the output (Z) in a large range after a small change in set-point. Increasing the gain resulted in a more stable flow with smaller pressure variations or smaller amplitude of slugs. Finally a high value of $K_{c0} = 220$ was selected to perform the step test. Set-point was manipulated to get the average valve opening higher than 0.26 and the obtained value of 0.29 was satisfying. It was aimed to do the test in a region that is unstable in open-loop position. After the system was stabilized, four step tests were implemented and data were logged. The related specifications are presented in table 5.1 and the related diagrams are shown in figure 5.3.

Table 5.1: Closed-loop step test specifications run with slow choke valve

<table>
<tr><th></th><th>K_{c0}</th><th>τ_I</th><th>Initial set-point</th><th>Final set-point</th></tr>
<tr><td>Test_1</td><td rowspan="4">220</td><td rowspan="4">∞</td><td>1.52</td><td>1.72</td></tr>
<tr><td>Test_2</td><td>1.73</td><td>1.54</td></tr>
<tr><td>Test_3</td><td>1.54</td><td>1.73</td></tr>
<tr><td>Test_4</td><td>1.49</td><td>1.70</td></tr>
</table>

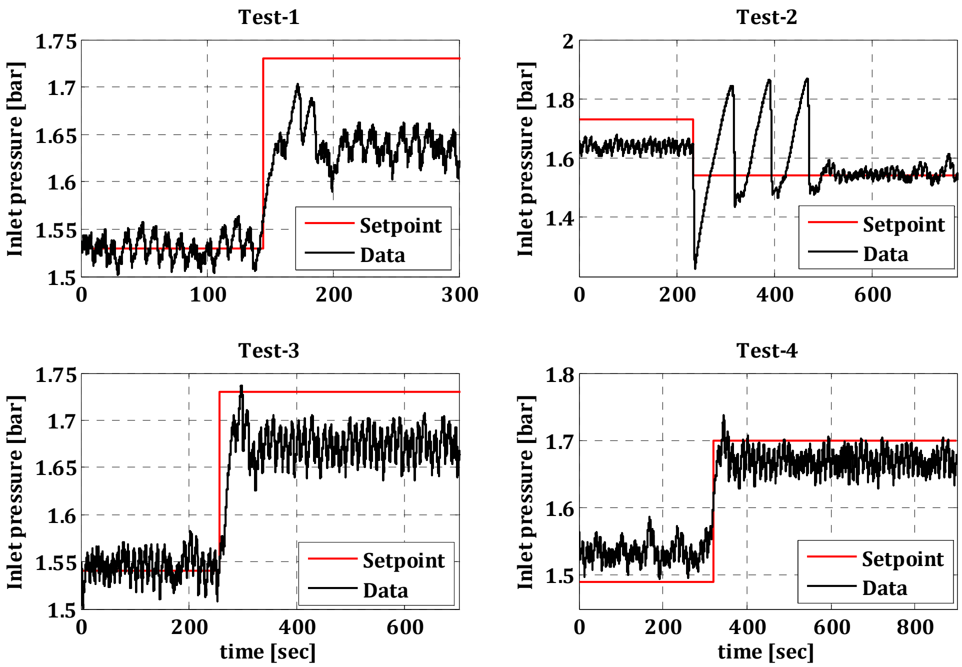

Figure 5.3: Presentation of different tests of set-point step change for a closed-loop feedback experiment with a P_only controller using inlet (buffer) pressure as control variable. Test-4 shows the best characteristics in case of desired overshoot and steady state gain required for tuning the controller.

After evaluating data from step tests it seemed that the last one (test_4) has better characteristics compared to the others with respect to the point that a unit step test was going to be used for all tuning methods. It was decided to use test_4 in the tuning of controller by different methods. Some important considerations in selecting the best step test were:

1. For the step test to be used in Shams's method the recommended 0.3 overshoot was desired.

2. The steady state gain of the system must be smaller than one ($\frac{\Delta y_\infty}{\Delta y_s} \prec 1$) to be used in *IMC-based tuning method*.

Since the response was noisy, a low-pass filter in MATLAB from the type of Simple infinite impulse response filter was used to reduce the noise effect. A smoothing factor of $\alpha = 0.001$ was used to smooth the signal as well as required ($\alpha = 1$ means no filtering). Figure 5.4 illustrates the step response used in the tuning methods.

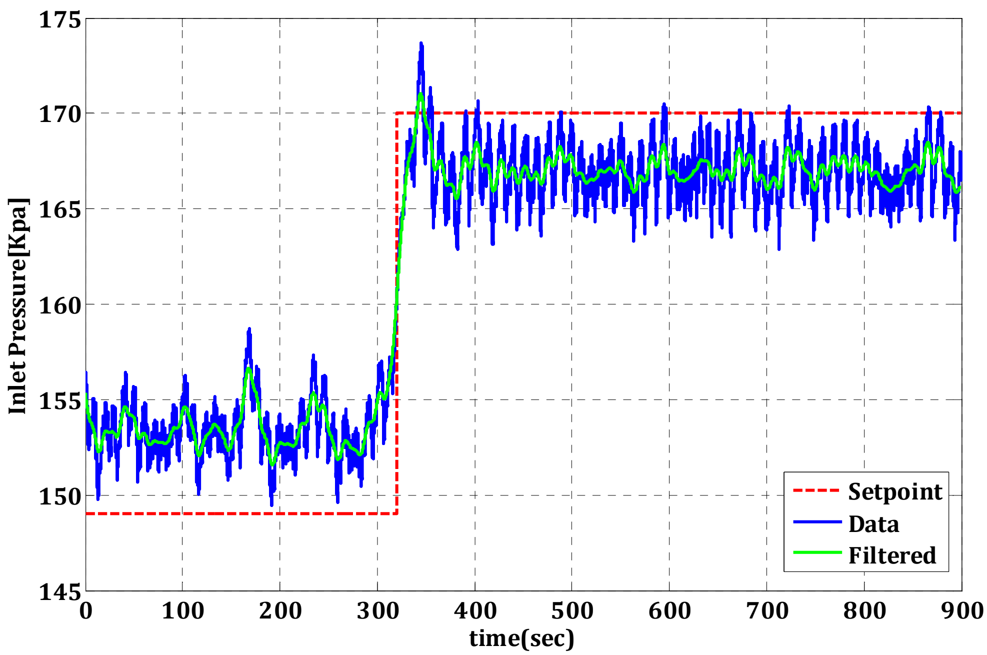

Figure 5.4: Set-point step change for a closed-loop feedback experiment with a P_only controller using inlet (buffer) pressure as control variable. A low pass filter with a smoothing factor of $\alpha = 0.001$ was used to remove the noise effect from the response.

5.1.1.3 Tuning the controller

The tuning methods explained in section 2.9 have been used to tune the controller using buffer (inlet) pressure as the control variable and slow choke valve as the actuator. The tuning procedure and the related results are explained in the following.

5.1.1.3.1 Tuning by Shams's closed-loop method

The first method to be used for tuning of the controller was Shams's method developed by Shamsuzzoha (Shamsuzzoha and Skogestad 2010). In order to tune by Shams's method, explained in section 2.9.1 the information from the step test explained in previous section *(See figure 5.4)* were used. Then, the overshoot was calculated and the appropriate tuning parameters were found. Table 5.2 shows the resulted tuning parameters by Shams's method. K_{c0} is the initial gain used in the step test, K_c is the calculated proportional gain, and τ_I is the integral tuning parameter. The system has been considered as a first order plus delay model.

Table 5.2: Tuning parameters from Sham's method for the slugging system

K_{c0}	Z_{ave}	Overshoot	Offset	K_c	τ_I
220	0.29	0.3846	0.6501	121.5189	224.3679

It was tried to control the system by the related tuning parameters seen in table 5.2. Yet, the mentioned tuning parameters couldn't work; meaning that the PI controller with these parameters was not able to stabilize the system and severe slugging was not eliminated. We may say that the Sham's tuning method is not a suitable approach for the slugging system.

5.1.1.3.2 Tuning based on IMC design

Next method applied in tuning of controller in experiments was the IMC-based tuning described in section 2.9.2. To do this, it was tried to identify the closed-loop stable system with respect to the data from step test and according to the method proposed by Jahanshahi (Jahanshahi and Skogestad 2013) explained in section 2.9.2.1. The identified model of closed-loop system was in the form of:

$$G_{cl}(s) = \frac{11.74\ \mathrm{S} + 0.606}{96.38S^2 + 10.88S + 1} \qquad \text{Equation 5.1}$$

The identified closed-loop transfer function is shown by the black line in figure 5.5.

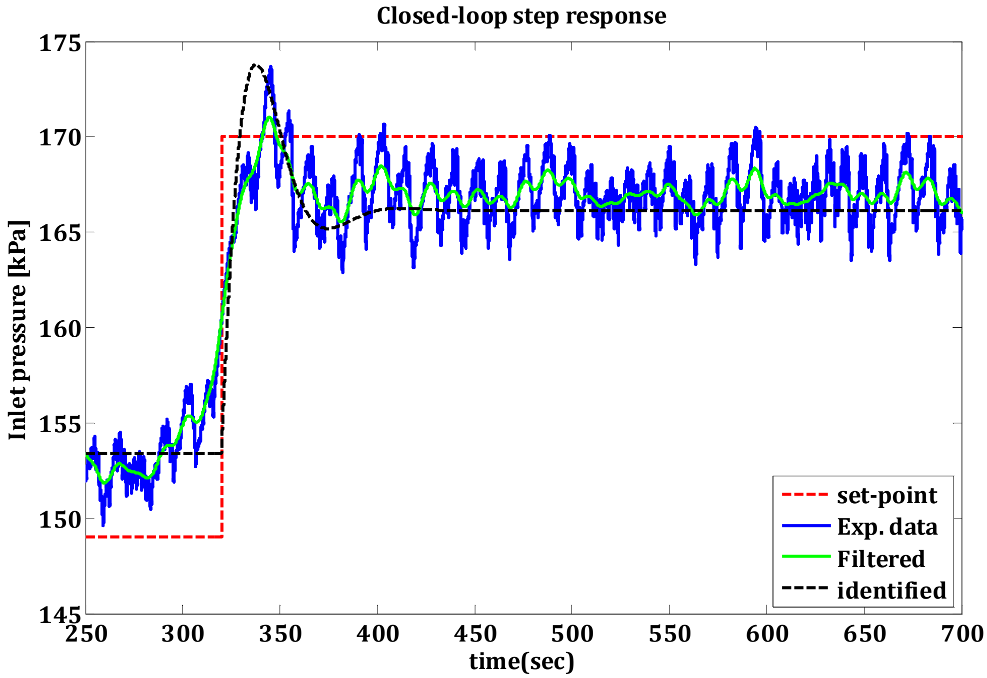

Figure 5.5: Presentation of identified closed-loop step response. The dashed black line shows the identified closed-loop transfer function obtained from IMC design.

Then, the open-loop unstable system has been back calculated by using the procedure proposed by Jahanshahi (Jahanshahi and Skogestad 2013). The open-loop unstable system has the form of:

$$\tilde{P}(s) = \frac{-0.0005538\ S - 2.858e\text{-}05}{S^2 - 0.008984S + 0.004088} \qquad \text{Equation 5.2}$$

Then the IMC controller (C) is designed by using the method explained in section 2.9.2.2. The time constant of the closed-loop system is an important manipulated parameter and has been selected as $\lambda = 20$. This number was obtained by trial and error and experiencing different results. The designed IMC controller is:

$$C(s) = \frac{-287.0673(S^2 + 0.02146S + 0.0007862)}{S(S+0.05161)} \qquad \text{Equation 5.3}$$

The IMC controller is a second order transfer function which can be written in form of a PIDF controller. PIDF is a PID controller which a low-pass filter has been applied on its derivative action. It will be mentioned as PID controller.

A PI controller was also obtained by reducing the order of IMC controller to 1.

The related tuning parameters have been obtained and are shown in table 5.3.

Table 5.3: IMC-based PID and PI tuning parameter

	K_{c0}	K_c	τ_I	τ_D	τ_F
PID	220	34.6387	7.92	141.2113	19.3773
PI	220	287.0673	65.6371	_	_

The approach of implementing the low pass filter in the experiments is described in appendix A.

To find the control results all related tuning parameters were implemented in LabVIEW and the loop was run in the stable region with an average valve opening of Z=25%.Then it was tried to decrease the set-point value in a stepwise manner. At each step it was waited until the steady state was reached and then a new step of reduction was done. Figures 5.6and 5.7 describe the results of control using the IMC-based PID and PI controllers respectively. The experimental slugging system could be stabilized up to Z= 40% with IMC-based PID controller and up to Z= 38.4% with IMC-based PI controller even though the controllers have been designed at valve opening of Z=28%.

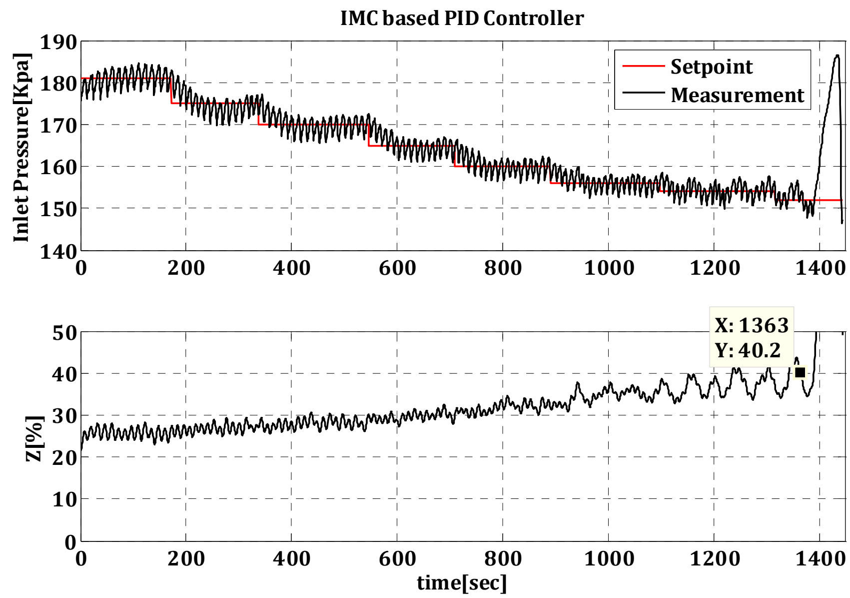

Figure 5.6: Result of control using the IMC-based PID controller. The controller has been able to move the bifurcation point from Z=26% up to Z=40.2%.

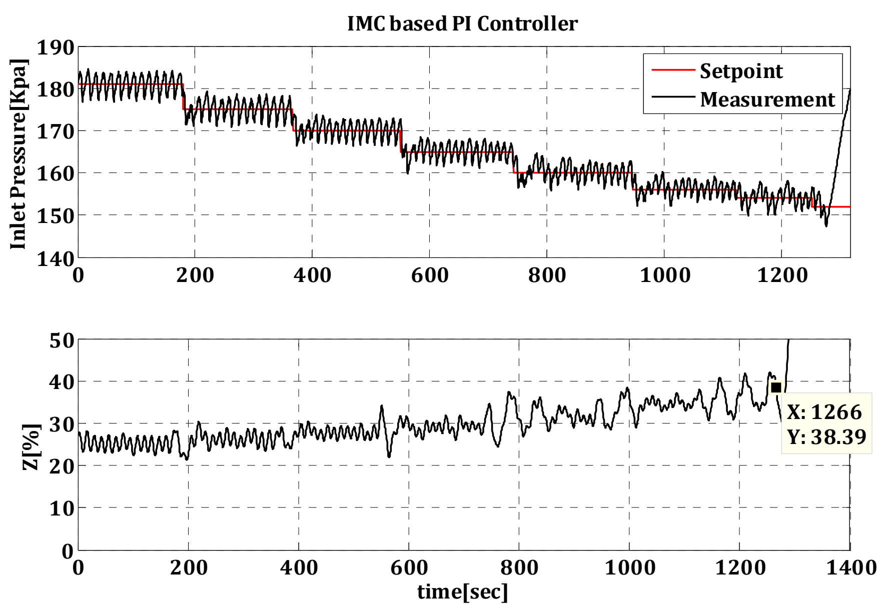

Figure 5.7: Result of control using the IMC-based PI controller. The controller has been able to move the bifurcation point from Z=26% up to Z=38.39%.

5.1.1.4 Inconclusive efforts and the related practical issues

When working with the first valve, some efforts were inconclusive and no results were produced. Below some explanations are given.

5.1.1.4.1 Tuning the controller by Simple online method based on identified MATLAB model of the system

As the last method of tuning it was tried to use simple PI tuning rules described in section 2.9.3. The method has been proposed by Jahanshahi (Jahanshahi and Skogestad 2013) and is based on the identified MATLAB static model of the system. To implement this method, first the simple static MATLAB model of the system which tuning rules are based on needed to be modified and fit to the experimental steady state model. For a reasonable result, it was required to have an accurate model of the experiments. Though, right in that time the lab technician replaced the current valve with the fast valve since he was going to vacation and this couldn't be done for a long time. Therefore this tuning method was tried only by the second valve.

5.1.1.4.2 Applying time delay in the controller

One important issue regarding the slow valve tests that needs to be mentioned is about applying time delay. It was aimed to check the robustness of control system by implementing delay on measurement. In order to do this, an algorithm was implemented in LabVIEW by one of the lab technicians. It was a digital delay line which delayed the samples of the measured data by a desired given time. The desired delay time could be set from the front panel. Yet the delay setting couldn't work in a desired way, meaning that even for very small values of delay the system switched to severe slugging and the control was impossible. It was clear that for such a long pipeline riser system small values of delay in the range of milliseconds couldn't crash the control and the reason of inconveniency may be from LabVIEW. It might be because of mistakes in the algorithm or in the connections inside LabVIEW. Since the system was in medium scale and no one else except for the lab technicians was able to do modifications in the system or LabVIEW and also due to time issues it was decided to ignore implementing time delay after counseling with my supervisor. It was an extra work to be done in the thesis while the next required experiments were not started yet at that time.

5.1.2 Series of experiments with valve 2 (fast choke valve)

The next series of experimental work in this thesis was repeating the first series of tests with a new fast valve as the actuator. A new method of tuning has been used here in addition to the tuning methods of previous section.

5.1.2.1 Open-loop experiments

The loop was run in manual mode with fixed flow rates of $w_l = 0.3927\ [kg/sec]$ for water and $w_g = 0.0024\ [kg/sec]$ for air. These flow rates are the same values used for the slow valve. The related inflow conditions have been fully described in section 5.1.1.1. The tests were run in different valve openings with fixed liquid and gas flow rates without applying control. The system behavior in natural conditions was then presented with the related bifurcation diagram as seen in figure 5.8.

5.1.2.1.1 Bifurcation diagram

The starting point was the valve opening of Z=0.1. Then the valve was open stepwise until it was fully open. The results of buffer (inlet) pressure were logged and the related bifurcation diagram was plotted. The critical stability point (the bifurcation point) is the maximum choke valve opening the system can have while being stable and is located at Z=0.16 for the system with valve 2. In the presented bifurcation diagram, the top line tracks the maximum values of pressure at each operating point, the bottom line presents the minimum values of pressure and the middle line shows the average values of the buffer pressure at different valve openings. Small pressure oscillations before the bifurcation point are due to hydrodynamic slugs and are not the signs of instabilities.

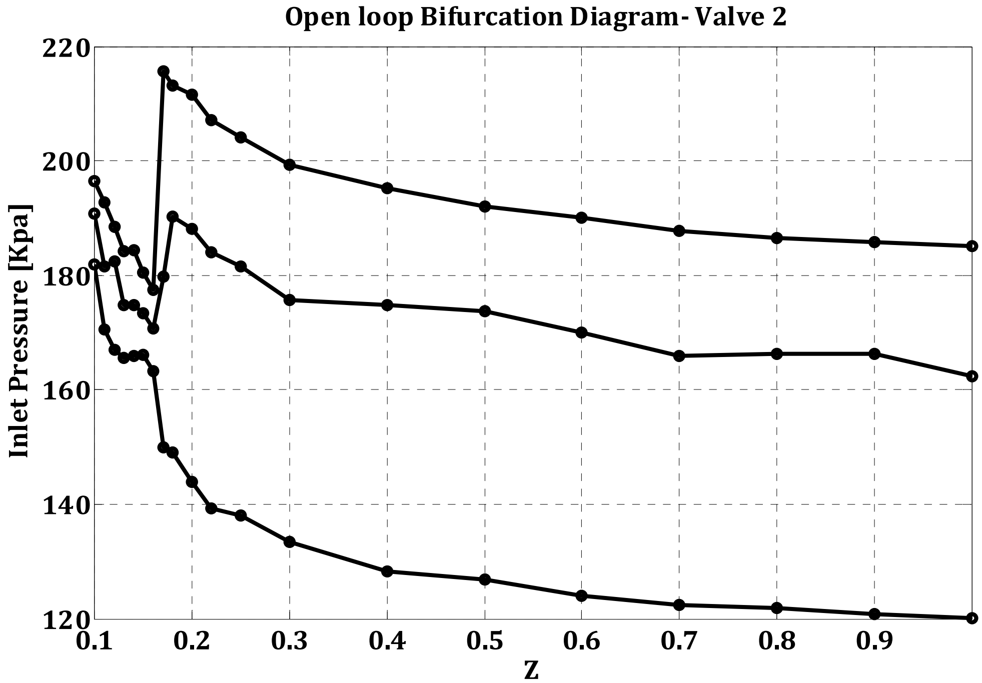

Figure 5.8: Open-loop bifurcation diagram from the fast choke valve experiments. The bifurcation point occurs at valve opening of Z=0.16. The top and bottom line illustrate the maximum and minimum values of inlet pressure respectively at each operating point. The middle line shows the average values of pressure.

5.1.2.2 Closed-loop step test

Just like the experiment series with valve 1, the first step to tune the controller with any tuning method was a closed-loop step test with a step change in set-point (the buffer pressure). The loop was closed with a P-only controller with a gain value of $K_{c0} = 250$. The step change was done in a region that is unstable in open-loop position. The average of valve opening was $Z = 0.18$. The related plot is shown in figure 5.10. Since the response was noisy, a low-pass filter in MATLAB from the type of Simple infinite impulse response filter was used to reduce the noise effect. A smoothing factor of $\alpha = 0.25$ was used to smooth the signal as well as required ($\alpha = 1$ means no filtering).

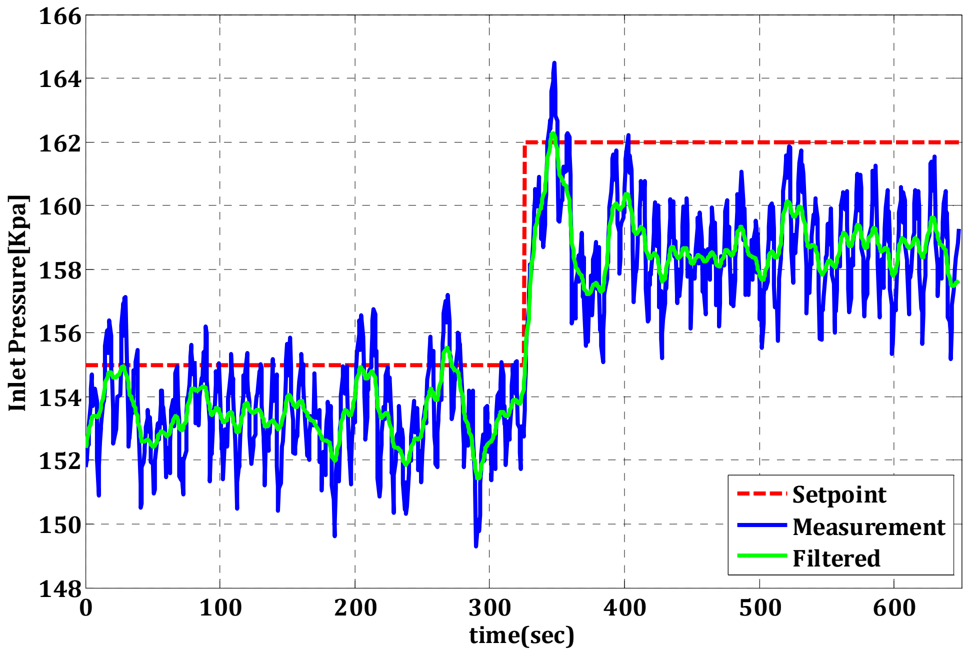

Figure 5.9: Set-point step change for a closed-loop feedback experiment with a P_only controller using inlet (buffer) pressure as control variable. A low pass filter with a smoothing factor of $\alpha = 0.25$ was used to remove the noise effect from the response.

5.1.2.3 *Tuning the controller*

Three different methods explained in section 2.9 have been used to tune the controller using buffer (inlet) pressure as the control variable and fast choke valve as the actuator. The tuning procedure and the related results will be presented below.

5.1.2.3.1 *Tuning by Shams's closed-loop method*

Shams's tuning method developed by Shamsuzzoha (Shamsuzzoha and Skogestad 2010) was used as the first tuning method. Table 5.4 shows the resulted tuning parameters by Shams's method. The system has been considered as a first order plus delay model. The information from the closed-loop step test *(See figure 5.9)* was used to find the tuning parameters.

Table 5.4: Tuning parameters from Sham's method for the slugging system

K_{c0}	Z_{ave}	Overshoot	Offset	K_c	τ_I
250	0.18	1.5738	1.3823	331.0775	246.7640

As expected, according to results of valve 1, the PI controller with these tuning parameters couldn't stabilize the system, meaning that Shams's method is not a suitable method to tune the slugging system controller.

5.1.2.3.2 Tuning based on IMC design

IMC-based tuning method described in section 2.9.2 was applied as the next method to tune the system with fast valve. Data from step test *(See figure 5.9)* were used and The model of closed-loop system was identified as explained in section 2.9.2.1.

$$G_{cl}(s) = \frac{9.076\ S + 0.7406}{64.76S^2 + 4.635S + 1} \qquad \text{Equation 5.4}$$

The identified closed-loop transfer function is shown by the black line in figure 5.10.

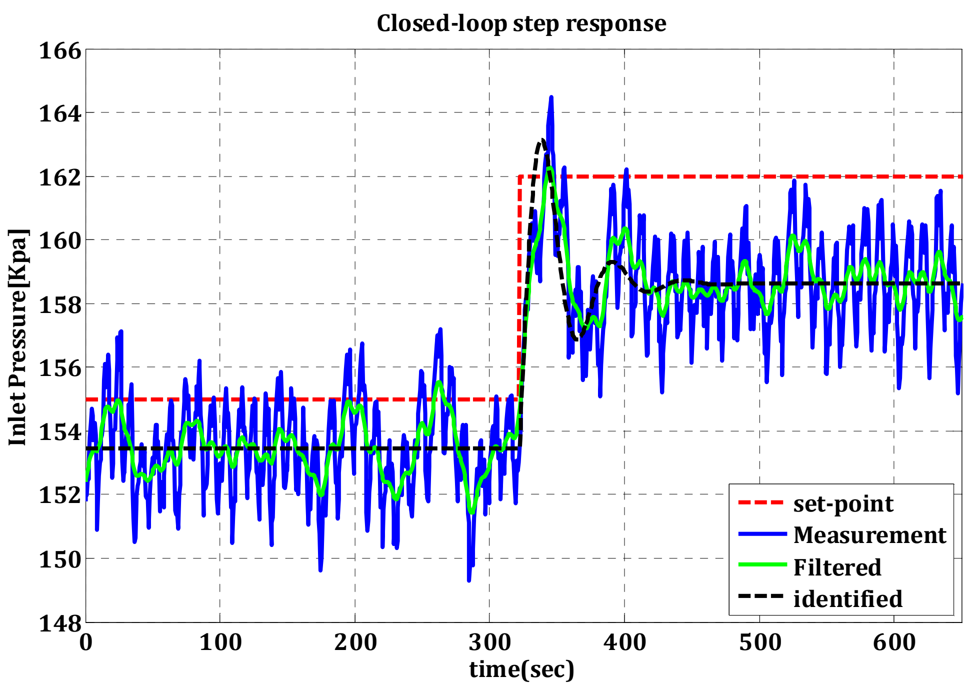

Figure 5.10: Presentation of identified closed-loop step response. The dashed black line shows the identified closed-loop transfer function obtained from IMC design.

The open-loop unstable system was then calculated as the form of equation 5.5 by using the procedure proposed by Jahanshahi (Jahanshahi and Skogestad 2013).

$$\tilde{P}(s) = \frac{\text{-0.0005606 S - 4.574e-05}}{S^2 - 0.06858S + 0.004006} \qquad \text{Equation 5.5}$$

The IMC controller (C) was obtained then as the equation 5.6 *(See section 2.9.2.2)*. A value of $\lambda = 24.5$ was used for the time constant of the closed-loop system. This value was manipulated by trial and error until a satisfying gain, phase and delay margin was obtained for the controller.

$$C = \frac{-340.7491(S^2 + 0.005194S + 0.000356)}{S(S + 0.0816)} \qquad \text{Equation 5.6}$$

The IMC controller as a second order transfer function was then written in form of a PID controller with a low-pass filter applied on its derivative action *(We may say a PIDF controller).*

A PI controller was also obtained by reducing the order of IMC controller to 1.

The related tuning rules are shown in table 5.5.

Table 5.5: IMC-based PID and PI tuning parameters

	K_{c0}	K_c	τ_I	τ_D	τ_F
PIDF	250	3.4736	2.3368	1189.9378	12.2552
PI	250	340.7491	229.2276	_	_

The function "*PID Advanced VI"* from LabVIEW was used to implement the low-pass filter in the experiments *(see appendix A).*

The PID tuning parameters were implemented in LabVIEW. First the system was run in open-loop manner with a manual valve opening of Z= 0.2 and data were logged. Then the loop was closed with a set-point P=170 kPa that results in an average valve opening of Z=0.16. After couple of minutes it was tried to decrease the set-point value in a stepwise manner. At each step it was waited until the steady state was reached and then a new step of reduction was applied. The same was done with PI tuning parameters. Figures 5.11 and 5.12 describe the results of control using the IMC-based PID and PI controllers respectively. The experimental slugging system could be stabilized up to Z=0.30 with IMC-based PID controller and up to Z=0.29 with IMC-based PI controller.

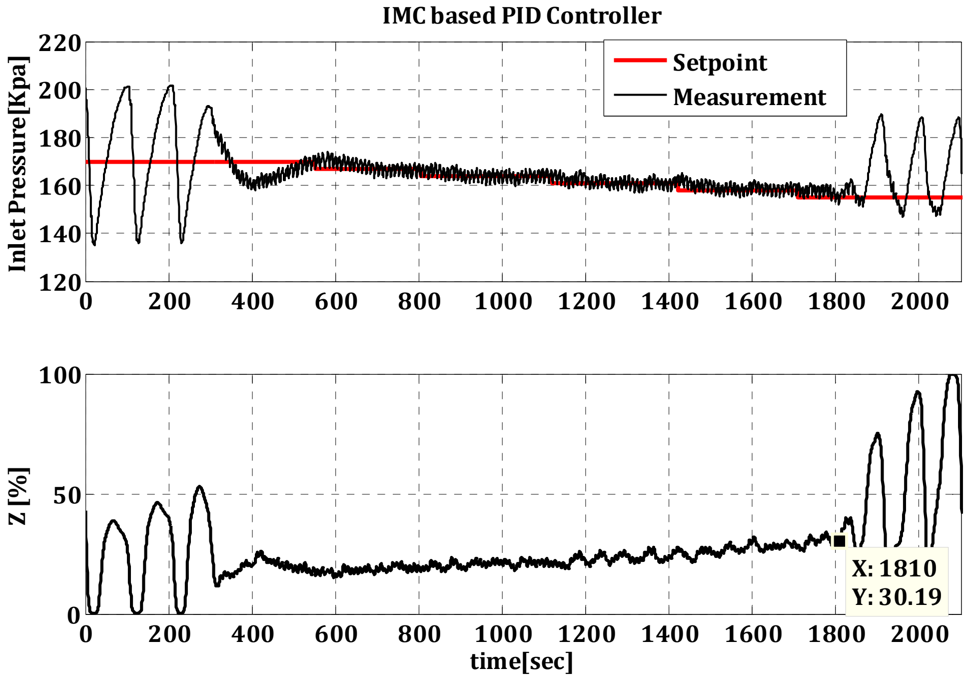

Figure 5.11: Result of control using the IMC-based PID controller. The controller has been able to move the bifurcation point from Z=16% up to Z=30.19% (about the double value).

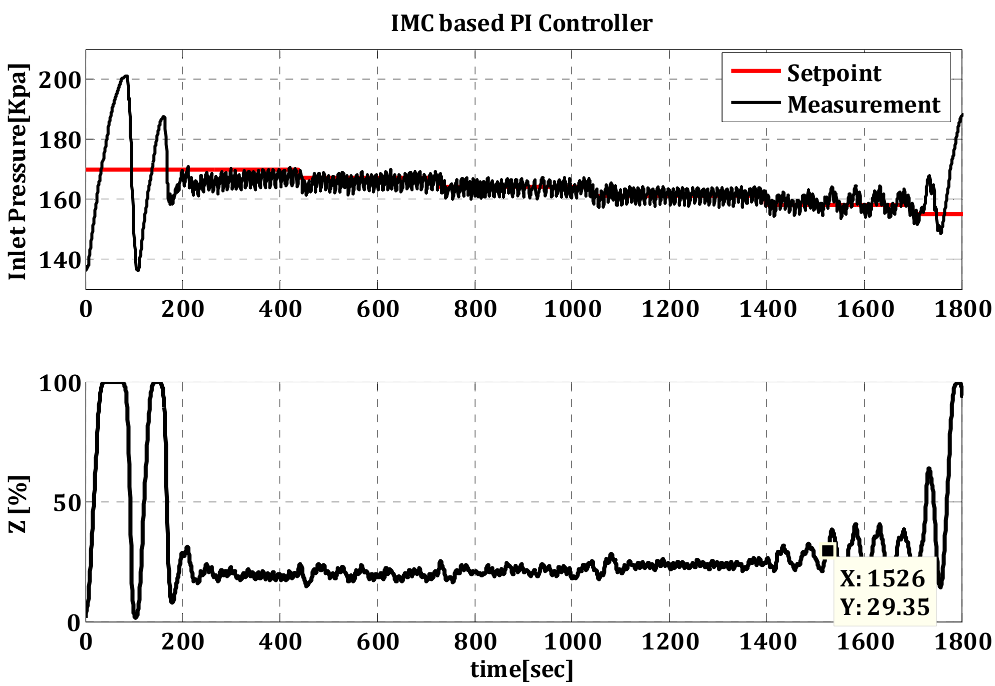

Figure 5.12: Result of control using the IMC-based PI controller. The controller has been able to move the bifurcation point from Z=16% up to Z=29.35%.

5.1.2.3.3 Simple online PI tuning based on MATLAB model with gain scheduling

Simple PI tuning rules described in section 2.9.3 was used as the last method of tuning the controller. Since the method is based on the simple static MATLAB model of the system, the MATLAB model needed to be identified and fit to the experimental steady state model.

For a reasonable result, it was first required to have an accurate model of the experiments. To find the model the loop was closed with a PI controller and was run in the region after stability point in open-loop bifurcation diagram. This was done such that set-point was set to a value lower than the corresponding value of the bifurcation point, then it was waited until steady state was reached and data were logged. The average of valve opening was found from logged data and the obtained point was located on the steady state experimental model. By repeating this for some other set-point values the steady state line of experimental model was found. It is shown in figure 5.13 with the black midline.

Next step was to modify the MATLAB static model and fit that with the experimental model. As described in section 2.9.3.1 the MATLAB model is derived based on the valve equation and is a function of valve opening. Therefore a good assumption of valve equation is very important in using the simple model. The valve equation is as the form:

$$w = K_{pc} f(z) \sqrt{\rho \Delta p} \qquad \text{Equation 5.7}$$

The valve is linear and its characteristic is defined as:

$$f(z) = Z \qquad \text{Equation 5.8}$$

The simple model for the inlet pressure is as defined in section 2.9.3.1 and the static gain of the system becomes in form of:

$$k(z) = \frac{-2\overline{w}^2}{\rho . z^3 . K_{pc}{}^2} \qquad \text{Equation 5.9}$$

Since the tuning parameters are found based on this MATLAB model, a good match between this model and the experimental model is very important meaning that the values of inlet pressure and the static gain obtained by the model needed to be true values. The parameters L_r (length of riser), $P\_V_{\min}$ (minimum Pressure drop over the valve) and K_{pc} (the valve constant) were manipulated many times until the desired match with the experimental model was reached. Below is a discussion of these parameters.

Length of riser

In MATLAB model length of riser is directly used to calculate the static pressure of the riser when it is filled with liquid and thereafter this static pressure is used to find the inlet pressure at any level of valve opening. Therefore manipulating of that could be very helpful in producing desired results. The exact length of riser in the experimental setup has been 6.433 m*.* Though, it was changed to 6.7 m in model to provide the best results.

Minimum pressure drop over the valve

This parameter is used in several calculations in the model. The most important one is the value of inlet pressure in the fully open position of the valve that uses $P\_V_{\min}$ directly *(See section 2.9.3.1*). Level of the curve in the inlet pressure plot of the model was quite affected by inlet pressure at fully open position of the valve. A value of $P\_V_{\min} = 3kPa$ was used to get the best fitness of the models.

Valve constant

The valve constant K_{pc} has a major effect on the slope of the curve in the inlet pressure plot of the model. A value of $K_{pc} = 1.6 \times 10^{-3}$ was used in the MATLAB model.

Figure 5.13 compares simple static MATLAB model to the experimental model. As clear in the figure there is a good match between the two models. The MATLAB model is attached in Appendix C.5. The black midline in the figure presents the steady state values of the inlet pressure and the red midline is the values of inlet pressure from the MATLAB model. The top and bottom black lines show the maximum and minimum values of pressure oscillations at each operating point in the open-loop system.

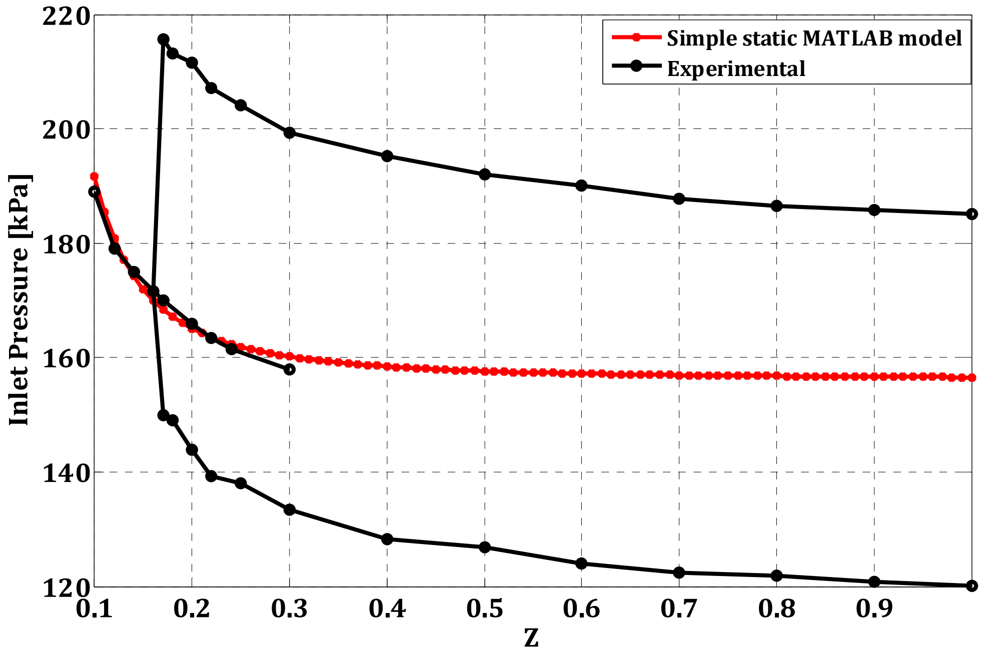

Figure 5.13: Simple static MATLAB model compared to the experimental model. The black midline in the figure presents the steady state values of the inlet pressure and the red midline is the values of inlet pressure from the MATLAB model. The top and bottom black lines show the maximum and minimum values of pressure oscillations at each operating point in the open-loop system.

In order to find the appropriate tuning parameters based on the identified MATLAB model a closed-loop test with step change of set-point is required *(See section 2.9.3.2)*. Data from the step test explained in section 5.1.2.2 was used *(See figure 5.9)* and the parameter β was found from the equation 2.41 as β = 0.038. The period of slugging oscillations in open-loop experiments has been T_{osc} = 90 Sec. When running the model at a special operating point the parameters $K_c(z)$ and $\tau_I(z)$ were found for the specified operating point as functions of valve opening (Z) by the equations 2.42 and 2.43.

By running the model with a loop for a wide range of Z values, it was possible to find multiple tuning parameters each for a controller at a specified operating point. Then gain-scheduling with multiple controllers was used to stabilize the system. To do this in the experiments, five PI controllers were used with the related found tuning parameters. Table 5.6 shows the resulted controller for each operating point of valve opening.

In order to perform the gain scheduling between the controllers the corresponding value of inlet pressure to that specific operating point of valve opening was determined from the model and then this pressure value as the set-point together with the related tuning values were entered in LabVIEW. The closed-loop was run and it was waited until the system was in steady state. Then the next pressure value (set-point) corresponding to the next valve opening was tried and the new tuning values were entered in LabVIEW very fast (I was working as the control woman!). This action was repeated until the instability was appeared.

Figure 5.14 shows the results of control with gain scheduling tuned by simple online tuning method. The controllers could stabilize the flow up to a valve opening of Z=0.35. Changing bifurcation point from Z=16 % into Z=35 % could be a very good result.

Table 5.6: PI tuning values and the corresponding operating points from simple online tuning method based on MATLAB model. These five controllers were connected and performed gain scheduling with multiple controllers for the nonlinear slugging system.

K_c	τ_I	Set-point	Valve opening
360.6481	320.625	166	0.19
511.9335	354.375	165	0.21
816.9311	405	161	0.24
2000.8383	523.125	157	0.31
4080.2676	641.25	156	0.38

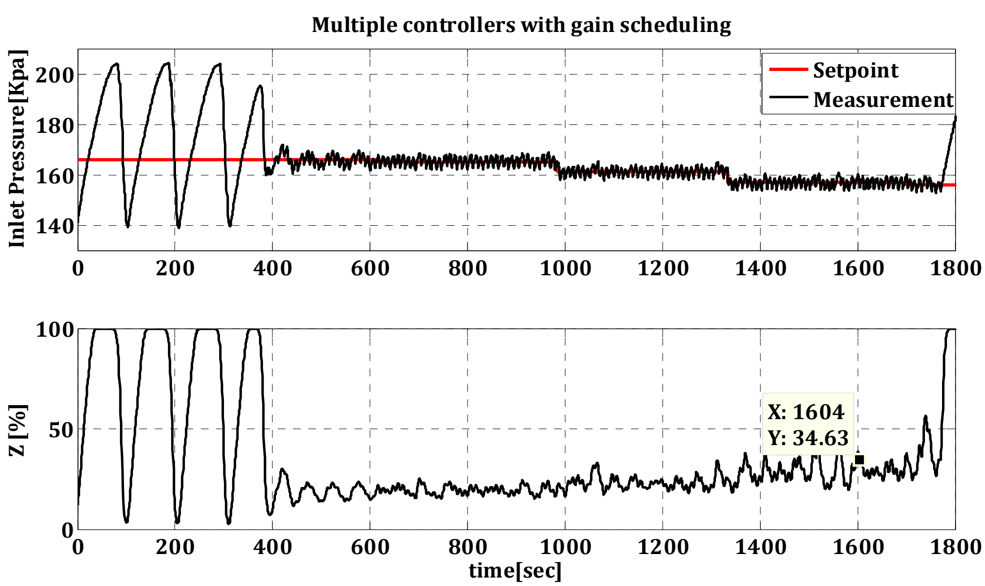

Figure 5.14: Results of gain scheduling using four PI controllers. When the system was switched into the fifth controller the instability was appeared; meaning that the maximum level of stability was reached with four controllers tuned by simple online tuning rules. The controllers have been able to move the bifurcation point from Z=16% up to Z=35%.

5.1.3 Cascade Control using top pressure combined with density

One of the tasks in the thesis has been running the closed-loop using a cascade control structure with selecting top pressure and density of outflow as the control variables. The intentions were tuning this control structure by trial and error and then analyze the controllability characteristics in comparison with the single loop structure.

5.1.3.1 The test practical issue – test incomplete

The outflow density was used as the control variable of inner loop and the top pressure was selected as the control variable for the outer loop. Having a right measure of density could be very important in controlling slugging system with the cascade structure. Figure 5.15 illustrates a schematic overview of cascade structure in LabVIEW. The device used for measuring the outflow density was the conductance probe explained in section 3.2.11.

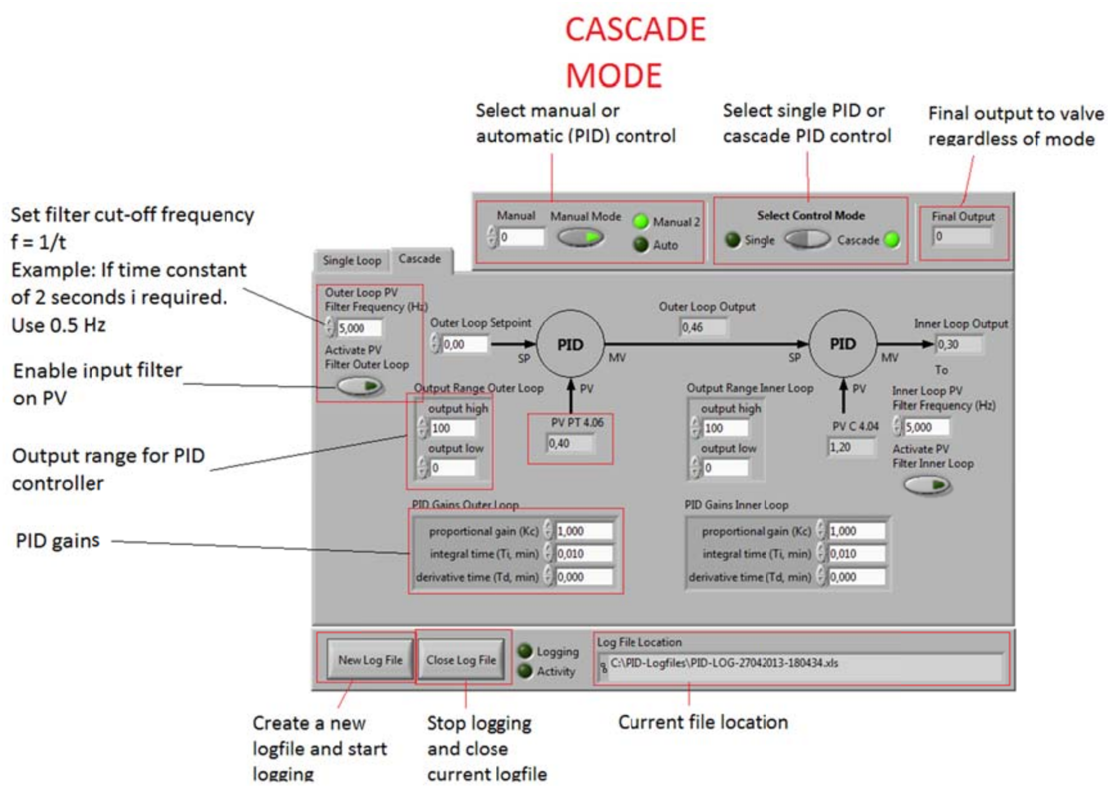

Figure 5.15: A schematic view of cascade control structure in LabVIEW. The outer loop receives signal from top pressure sensor as the measurement and produces the set-point signal for the inner loop. The inner loop uses the density signal from the conductance probe installed in the outlet of the riser.

However it seemed that the conductance probe is not a good measuring device for the density. After some unsuccessful tries to control the system by tuning the loops with trial and error, it was decided to perform a step test in open-loop situation of the system and evaluate the open-loop step response of the conductance probe. It was aimed to check the applicability of probe as an appropriate sensor to measure the density. To do this, the loop was run in manual mode at the stable region with a valve opening of Z=7%. Data from density meter (Conductance probe) and top pressure sensor was logged. After some minutes the valve was changed to Z=12% while it was tried to keep the system in the stable region. Figure 5.16 presents the open-loop step response of the probe as the density meter.

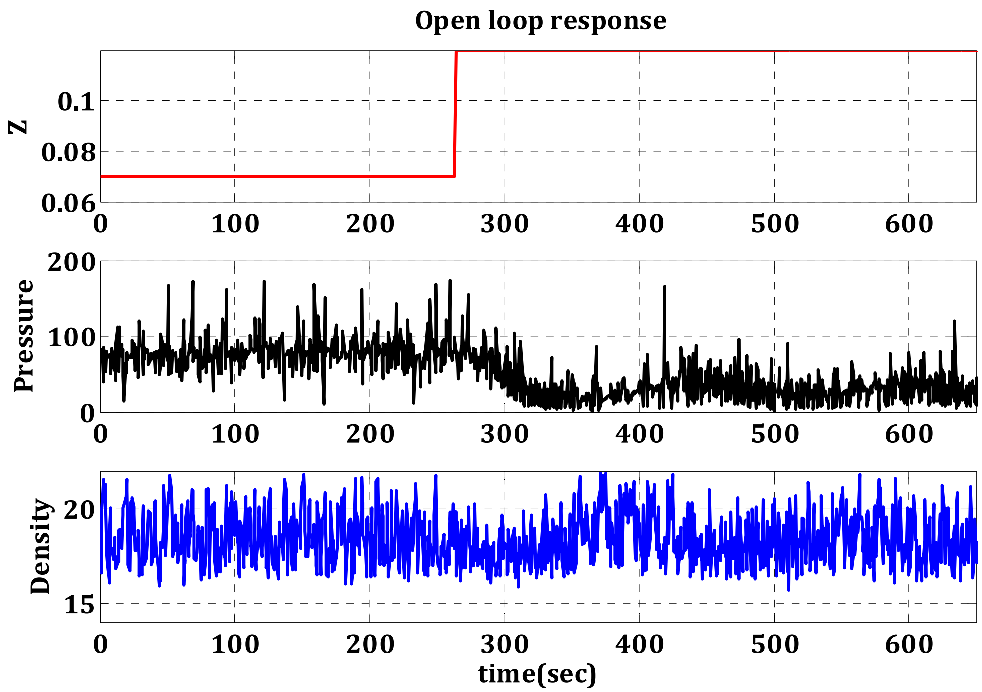

Figure 5.16: Open-loop step response of conductance probe (density meter) in the stable region. The first plot (Z) shows the step change in valve opening, the second plot presents the step response of top pressure in Kilo Pascal and the third plot illustrates the step response of the outflow density in kg/sec.

As seen in the figure, the density signal does not show a clear response to the step change and is very noisy. This signal couldn't be a suitable measurement for the control targets. In order to have an efficient cascade control with density as the inner loop control variable, more accurate signals of density are required.

Trying each loop separately to evaluate their response independently, could be considered as an alternative work. But this was not practical since the backup of the LabVIEW file was lost and the compiled file couldn't be manipulated or modified. Making a new file was not possible due to time issues.

5.2 Comparison of Slow valve and Fast valve

In this section it has been tried to compare the dynamics of the applied control valves (slow valve and fast valve) by investigating their related results. For this target the open-loop and closed-loop results of the two valves were compared. Our criterion to evaluate control loop is the stability. For a fast stability the dynamic response of the valve is important. It means a small dead time for the valve.

The criterion for evaluating the stability of the slugging control loop has been usually the level of valve opening (Z). However, this criterion couldn't be useful for comparing control valves with each other. The reason gets back to the valve inherent characteristics that will be explained in the following.

The relation between the flow rate and the level of valve opening is an inherent characteristic of the valve that has been defined as the valve equation:

$$q_{mix} = K_{pc} f(z) \sqrt{\frac{\Delta P}{\rho_{mix}}} \qquad \text{Equation 5.10}$$

Here q_{mix} is the volumetric flow rate, K_{pc} is the valve constant, ΔP is the pressure drop over the valve and ρ_{mix} is the mixed density of outflow.

Both valves have been considered linear with $f(z) = z$. But in reality valve 2 (fast valve) could be nonlinear to some extent, meaning that it produces the same flow rate as the slow valve even with lower levels of valve openings. Figure 5.17 describes this concept more clearly by illustrating the characteristic curves for the two valves. If we specify a level of flow rate and try to find the corresponding levels of valve opening for each of the valves, it will be seen that the slow valve may give higher level of valve opening for the same flow rate.

The main desired result that can be affected by the valve dynamics is the minimum inlet pressure the system could obtain. For the open-loop system, this is defined as the minimum inlet pressure at fully open position of the valve and for the closed-loop system it will be defined as the minimum set-point the controller can stabilize the system. Figure 5.18 compares the open-loop behavior of the system for the slow valve with that of fast valve. As clear in the figure, the slow valve gives lower inlet pressures at most operating points of valve opening including the fully open position of the valve (Z=1).

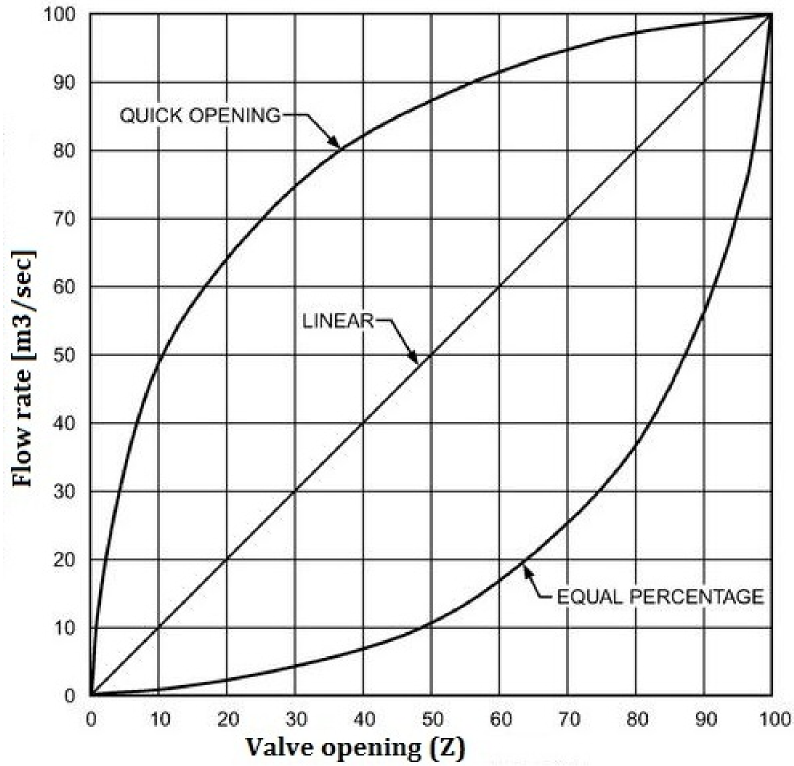

Figure 5.17: Characteristic curves for slow (linear) and fast (quick opening) valves. There is a lower level of valve opening for the fast valve at a specific flow rate (for instance 50), meaning that the fast valve can produce the same flow rate as the slow valve even at lower levels of valve opening.

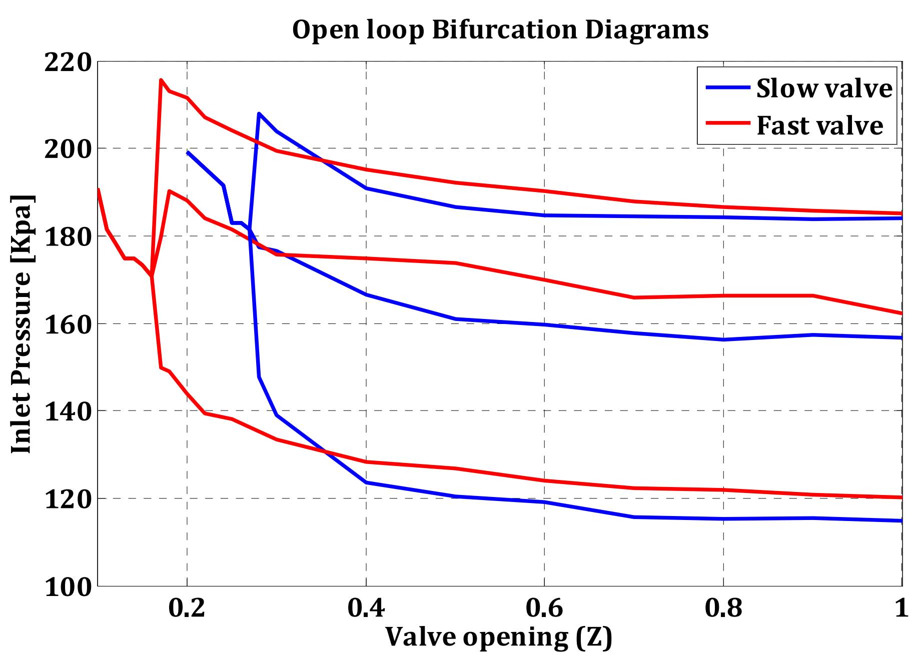

Figure 5.18: Comparison of inlet pressure between the slow valve and the fast valve at their different operating points for the open-loop system. At a certain level of valve opening, the slow valve gives a lower inlet pressure.

Based on the previous descriptions, it was decided to compare the minimum achievable set-points in the closed-loop responses. Figure 5.19 present the control results with IMC-based PI and IMC-based PID controllers. The valve opening is also presented, just in case, and is not a point of interest to compare the results.

From the figures it can be said that the slow valve has had a better performance compared to the fast valve. This means that the slow valve has been already fast enough for our control targets and there has been no need to valve 2 (faster control valve). In other words the stability of the slugging system is more affected by the tuning parameters for the controller instead of control valve dynamics.

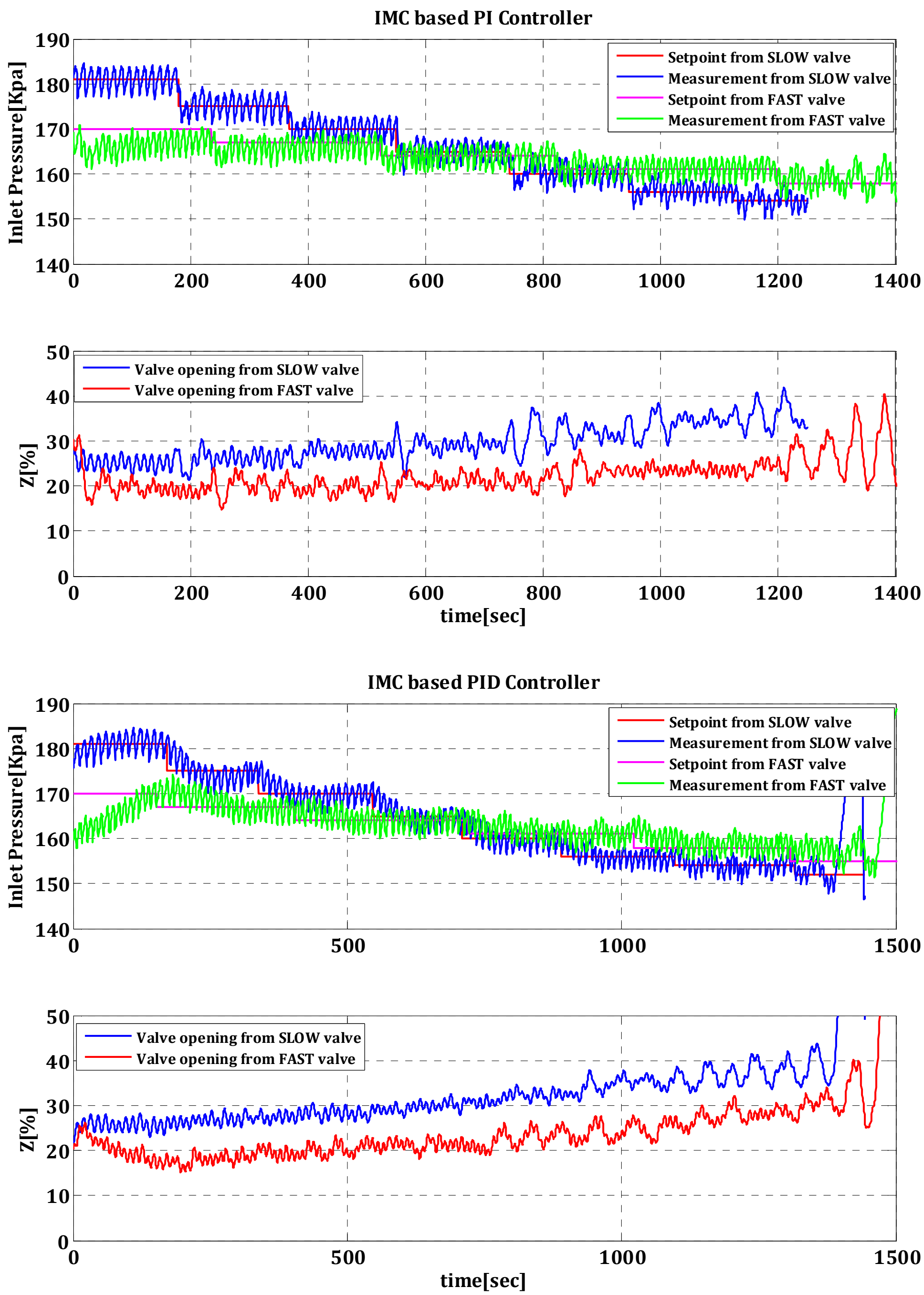

Figure 5.19: Comparison of IMC-based control results from slow valve with those of fast valve. With the slow valve it has been able to decrease set-point in a wider range to a lower level. With the fast valve, the open-loop system switched to slugging at P = 170 kpa while with the slow valve the instability started at P=180 kpa in the open-loop system. These are the initial points, respectively, to start control. The minimum achievable set-point has been P=154 kpa for the slow valve and P=158 kpa for the fast valve. This means that the slow valve has shown a better performance for the slugging system and has been already fast enough as well.

5.3 Simulated results from OLGA model

In this chapter simulation of the experimental cases in OLGA® are presented. The simulations have been matched with the experimental models from valve1 (Slow valve). The open-loop simulations are discussed in section 5.3.1. In section 5.3.2 results of control simulations by trial and error would be explained. Section 5.3.3 deals with finding the appropriate tuning rules based on the methods explained in section 2.9. Results of control by applying the calculated tuning parameters are also discussed in this section.

5.3.1 Open-loop simulations

The first step before implementing the controller is running simulations for different valve openings with fixed liquid and gas flow rates. The values of Z and the related flow regime types are presented by table 5.7.

Table 5.7: Different values of valve opening (Z) used in open-loop simulations

Z	Flow regime stability
0.20	Stable
0.25	Stable
0.26	Stable
0.27	Unstable
0.28	Unstable
0.30	Unstable
0.40	Unstable
0.50	Unstable
0.60	Unstable
0.70	Unstable
0.80	Unstable
0.90	Unstable
1.00	Unstable

Figure 5.20 describes the open-loop bifurcation diagram from simulations. The diagram shows the maximum, minimum, average and steady state values of buffer pressure versus the valve openings. The applied fixed flow rates have been $w_l = 0.3927$ $[kg / sec]$ for water and $w_g = 0.0024\ [kg / sec]$ for air (*The same as experiments*). These flow rates correspond to $U_{sl} = 0.2\ [m / sec]$ and $Usg = 1\ [m / sec]$ as the liquid and gas superficial velocities. The critical stability point (the bifurcation point) is the maximum choke valve opening the system can have while being stable. In a bifurcation diagram, the critical stability point is where the maximum and minimum pressures approach a finite value. In the presented bifurcation diagram, the red line shows the steady state

values of the buffer pressure at different valve openings and the average values of the pressure are on the mid black line that is higher than the steady state line. The coefficient of discharge was changed to c_d=0.34 in order to manipulate the placement of the critical valve opening (the bifurcation point) based on the experimental result of valve 1 and also fit the steady state OLGA values with the steady state values from experiments and models. The diagram comparing steady state values from OLGA with that of the model will be presented in section 5.3.3. As clear in the figure the critical stability point was found to be at approximately choke valve opening of Z=26%.

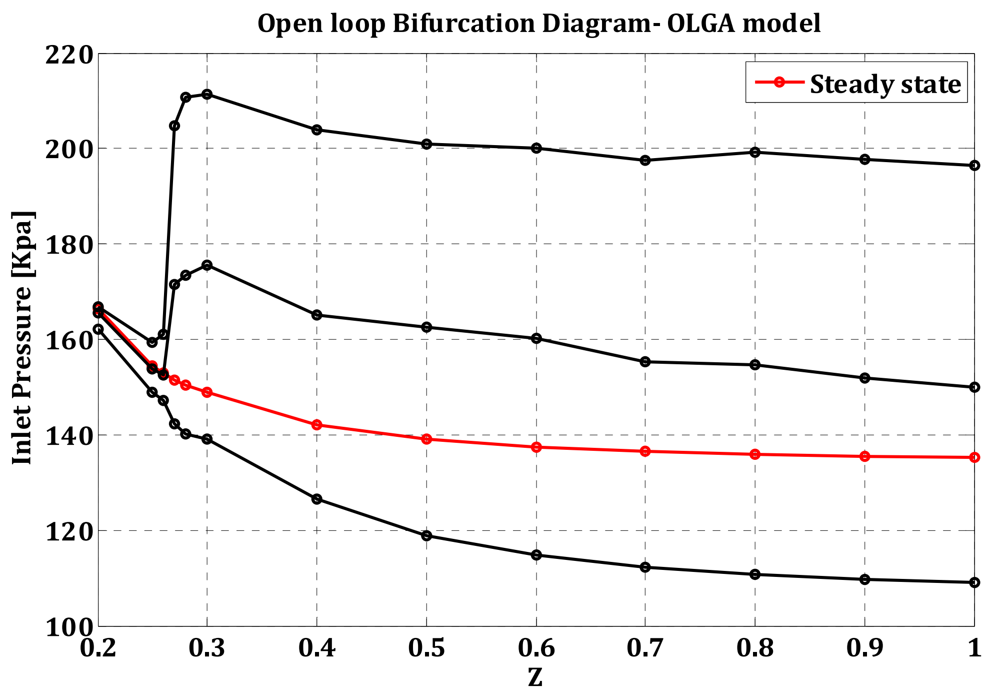

Figure 5.20: Open-loop bifurcation diagram from OLGA simulations. The bifurcation point occurs at valve opening of Z=0.26. The top and bottom line illustrate the maximum and minimum values of oscillations for inlet pressure respectively at each operating point. The mid-black line is the shows the average values of pressure.

5.3.2 Control by trial and error

As the first work after implementing PID controller in OLGA, it was tried to stabilize the flow by trial and error. Two types of controller including P-only controller and PI controller were tried to be tuned by trying many different values as the related tuning parameters.

5.3.2.1 P-only controller

As the first try a P-only controller was used to stabilize the system. P-only controller has been designed by inserting $\tau_D = 0$ and $\tau_I = 10^{10} \simeq \infty$.A point in unstable region with $Z = 0.3$ was selected and different values of gain parameter were tried to check which gain can create stability with the highest level of valve opening (Z). For each gain value it was tried to find the minimum amount of buffer (inlet) pressure as Set-point or in other words the maximum level of valve opening as manipulated variable by stepwise reduction of the Set-point. Table 5.8 shows different values of gain that have been tried and the corresponding minimum value of Set-point and maximum value of Z.

Table 5.8: different tried P-only controllers

initial valve opening	K_c	Minimum Set-point value (P)	Maximum Manipulated variable (Z)
0.3	0.01^*	_	_
	0.05	141	0.3686
	0.1	138	0.4497
	0.5	135.6	0.6454
	1	136	0.6305
	2	_	_
	5	_	_
	10	_	_

The best controller that gives stability with the highest level of Z and the lowest level of achievable set-point is the one with $K_c = 0.5$. With this controller, the bifurcation point was moved from Z=26 % into Z=65 %. Figure 5.21 shows the result of control by P_Only controller for $K_c = 0.5$. For the controller with $K_c = 0.01$, specified by the star in the table, the simulator could converge in some values of Set-point. However, the result was not good and there were many oscillations in pressure and valve opening. It was almost impossible to make a reduction in the Set-point.

For the controller with $K_c = 1$, control was difficult and the Set-point reduction was challenging. Figure 5.22 shows the result of control by for $K_c = 1$. The steps of reduction had to be selected very small and the simulator could not converge with a larger step than it is observed in the figure. For the values filled with dash the simulator could not converge for any values of Set-point, meaning that it was impossible to control the system with the gain values higher than 1. A P-only controller with $0.05 \le K_c \le 1$ can stabilize the system.

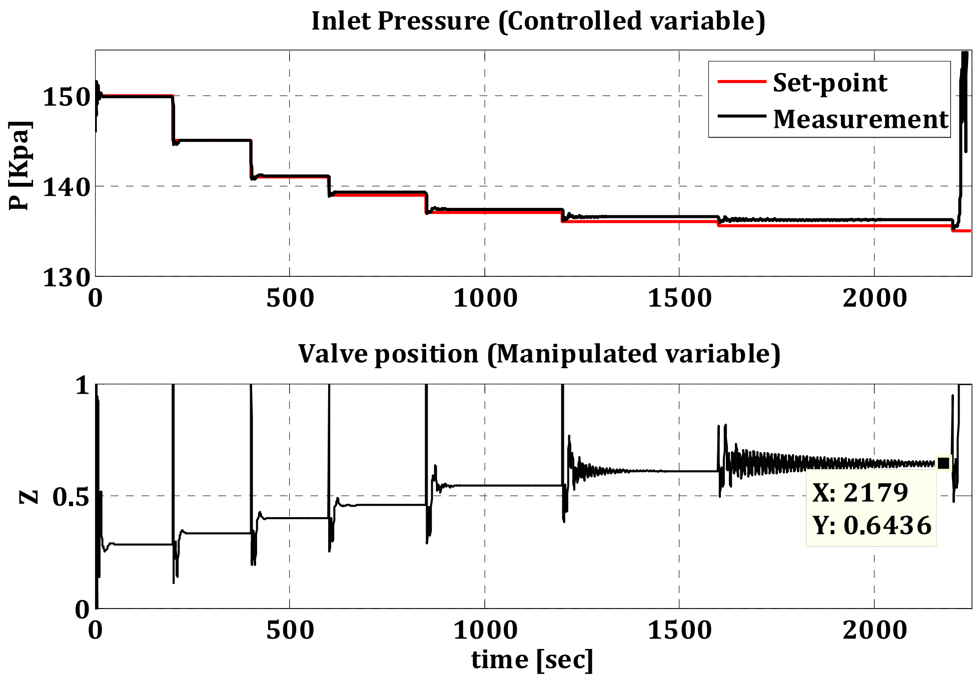

Figure 5.21: Simulation result of control by P-Only controller for $K_c = 0.5$ with OLGA. This has been the best result from trial and error due to the lowest achievable set-point or in other words the highest level of valve opening.

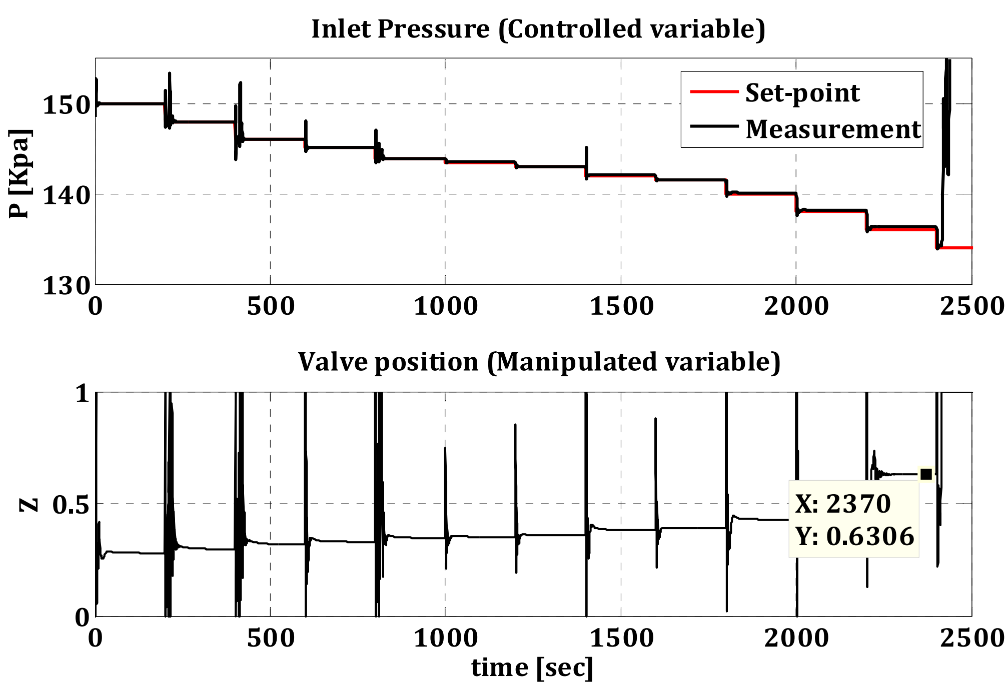

Figure 5.22: Simulation result of control by P-Only controller for $K_c = 1$

5.3.2.2 *PI controller*

PI controller was used to stabilize the system in the second series of simulations by trial and error. The controller was designed by inserting $\tau_D = 0$ and trying different values for K_c and τ_I. The aim as the previous part was to tune the controller to create stable flow with the highest production rate (the highest level of valve opening (Z)). Stepwise reduction of the Set-point was implemented as the one for P-only controller. Table 5.9 shows different values of tuning parameters that have been tried and the corresponding minimum value of Set-point and maximum value of Z.

Table 5.9: Simulation Results of different tried tuning parameters for PI controller

K_c / τ_I	0.01		0.05		0.1		0.5		1	
	Min. P	Max. Z	Min. P	Max. Z	Min. P	Max. Z	Min. P	Max. Z	Min. P	Max. Z
80	_	_	143.2	0.368	140	0.416	136.5	0.621	_	_
130	_	_	141.9	0.379	140	0.434	136.5	0.627	_	_
180	_	_	141.7	0.388	139.6	0.457	136.2	0.639	_	_
300	_	_	141.5	0.395	139.6	0.460	136.2	0.644	_	_
800	_	_	141.3	0.397	139.4	0.463	136.2	0.646	_	_

As it is observed in the table, the best tuning parameters are $K_c = 0.5$ and $\tau_I = 800$. Higher values of 800 were also tried for the parameter τ_I and no difference was made in result. Result of control by PI controller with the best tuning parameters is presented in figure 5.23. Increasing the parameter τ_I decreased the system oscillations very well and even eliminated it in some cases. However, it caused a longer time to be required for the output to track the Set-point in each step of Set-point reduction. This important effect of applying integral time constant could be verified by comparing figures 5.23 and 5.22. As it is observed, a less oscillatory system with longer simulation time is the result of PI controller compared with P-only controller.

The same as the one for P-only controller happened for the PI controllers with the gain values of $K_c = 0.01$ and $K_c = 1$. The simulator could not converge for any values of Set-point, meaning that it was impossible to control the system with the tuning parameters filled with dash.

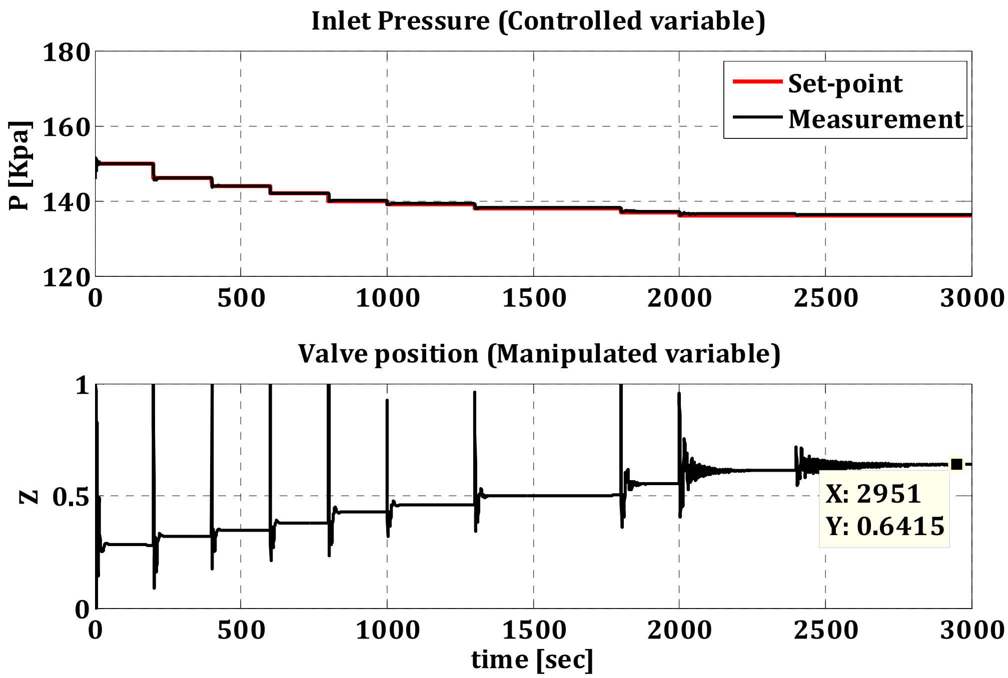

Figure 5.23: simulation results of control by PI controller for $K_c = 0.5$ and $\tau_I = 800$

5.3.3 Tuning the controller

Three tuning methods explained in section 2.9 were used in simulations. The aim has been to compare the tuning methods based on simulations as well as experiments.

5.3.3.1 Tuning using Shams's closed-loop method

In order to tune the controller with Shams's method a closed-loop step test was required. As explained in section 2.9.1 a P-only control is required to determine the optimal tuning parameters. The P-only controllers described in section 5.3.2.1 were used to achieve a step response close to the recommend 0.3 overshoot. First the system was set with the choke opening at Z=30 % where it is unstable in open-loop position and stable in closed-loop position. It was at steady-state initially and then a step change was applied in set-point. Different values of step change were tried to get a step response close to the recommend 0.3 overshoot. Then the system was set with the choke opening at Z=40 % and the same tries were implemented. Value of the resulting overshoot was highly depended on the initial gain and the amount of step change. In some cases with the same initial gain several tests with different amounts of step change were run in order to get the desired 0.3 overshoot. All simulations run to get the desired overshoot at different basis conditions of the controller are presented in Appendix B.

When the desired overshoot was achieved the Shams's method for closed-loop systems explained in section 2.9.1 was used to find the appropriate tuning parameters. Table 5.10 shows the resulting tuning parameters by Shams's method at different initial positions of choke valve. K_{c0} is the initial gain used in the tuning simulation, K_c is the calculated proportional gain, and τ_I is the integral tuning parameter.

Table 5.10: Tuning parameters from SIMC method for the slugging system

Initial valve position	K_{c0}	Overshoot	Offset	K_c	τ_I
0.3	0.1	0.3085	0.0787	0.0614	34.5702
0.4	0.15	0.3210	2.1132	0.0904	3.1150

Using PI controllers with the parameters found in table 5.10, the system became unstable at a choke valve opening of approximate Z= 38.84 % with the controller tuned at the initial position of 30 % and at a choke valve opening of approximate Z=39.45 % with the controller tuned at the initial position of 40 %.

Figures 5.24 and 5.25 show the step test using initial choke valve opening of 30% as the basis inflow condition and the result of control by Shams's method respectively.

Figures 5.26 and 5.27 are presented for the initial choke valve opening of 40%.

Two initial points were used for tuning to improve the results. However as it is clear from the figures, no notable change is observed in the results of control. Decreasing set-point even for a very small value more than the final value shown in the figures caused system to become unstable. Severe slugging occurred and the simulator could not converge.

As seen in the results, the second controller tuned at the initial point of Z=40% hasn't been able to stabilize the system for any further valve openings. It hasn't been able even to achieve the point that has been tuned for. This may not be strange since the Shams's method has been designed for the stable systems while the slugging system is unstable.

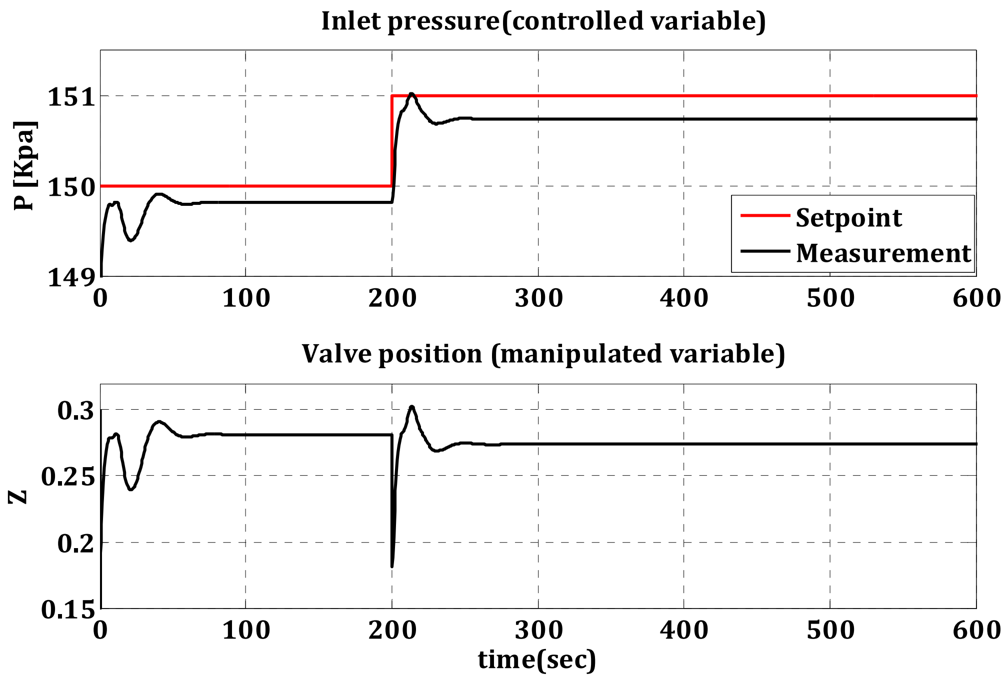

Figure 5.24: Set-point step change using initial choke valve opening of 30% and the initial gain of $K_{c0} = 0.1$. An overshoot of D=0.3, as the recommended value by Shams has been achieved.

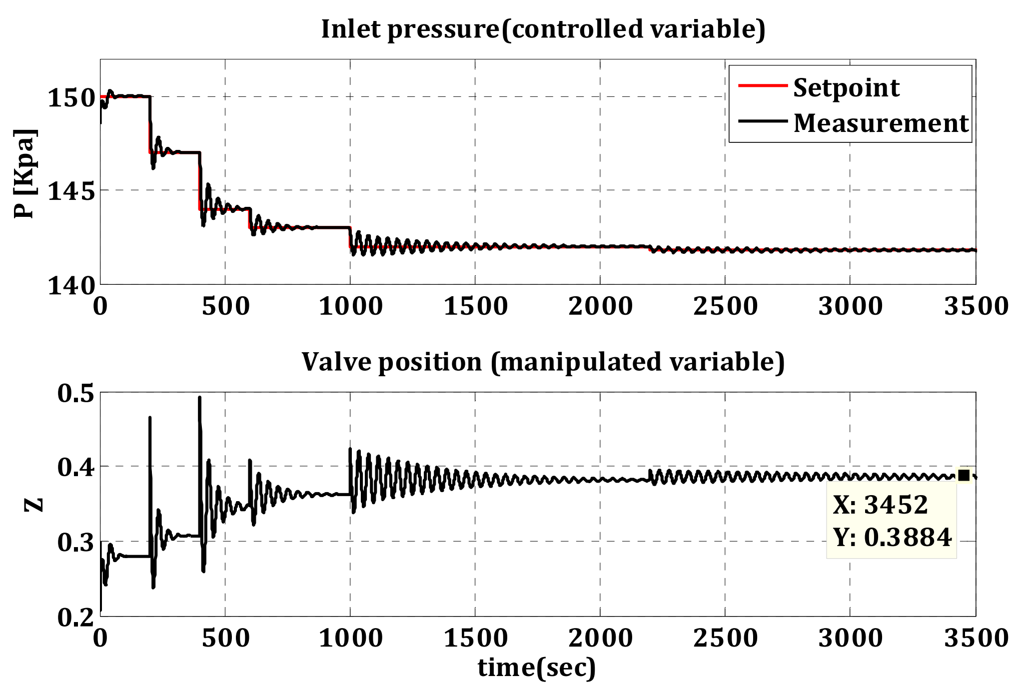

Figure 5.25: Simulation result of control by Shams's closed-loop method for the initial choke valve position of 30%. The values of Z=0.389 and P=142 kpa have been the maximum achieved valve opening and the minimum achieved set-point, respectively.

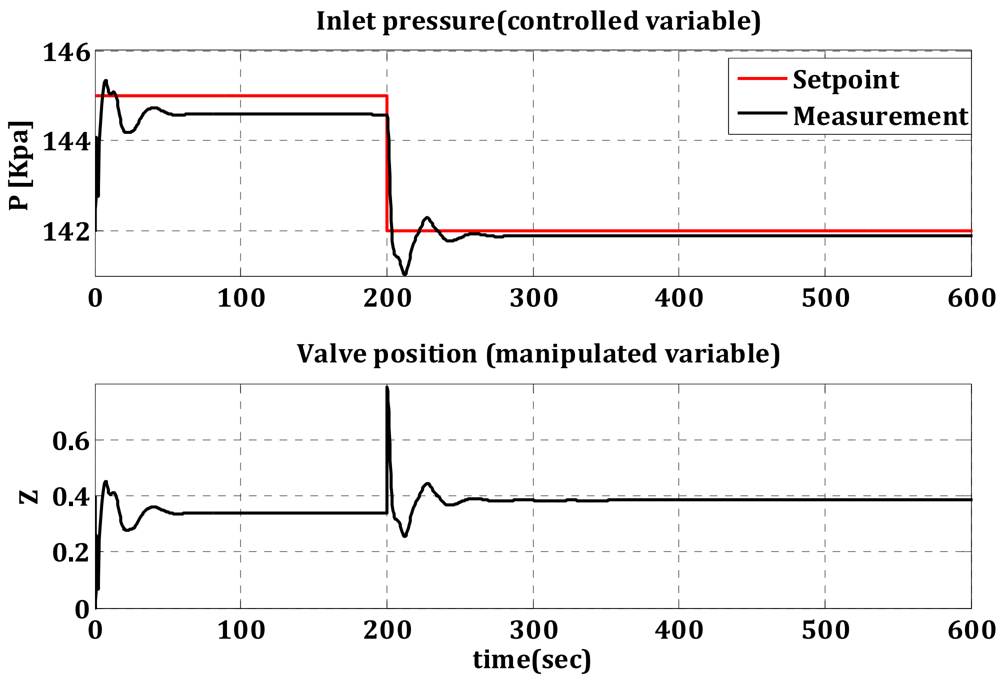

Figure 5.26: Set-point change using initial choke valve opening of 40% and the initial gain of $K_{c0} = 0.15$. An overshoot of D=0.32, near to the recommended value by Shams has been achieved.

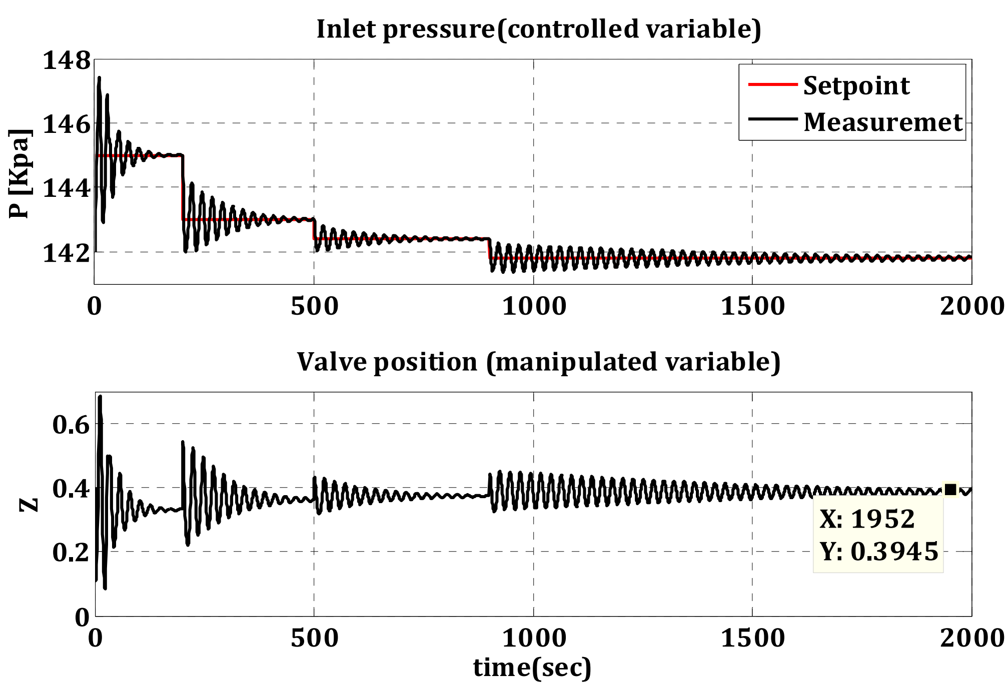

Figure 5.27: Simulation result of control by Shams's method for the initial choke valve position of 40%. The values of Z=0.395 and P=142 kpa have been the maximum achieved valve opening and the minimum achieved set-point, respectively.

5.3.3.2 Tuning based on IMC design

Next method used in tuning of controller in simulations was the IMC-based tuning described in section 2.9.2. As explained before, the open-loop system switches to slugging flow at valve opening of Z=26% and it is unstable at Z=30% or 40%. Tuning by this method was done for two different operating points of the system; Z=30% and Z=40%. Both simulations as well as their results are presented in this section.

5.3.3.2.1 IMC-based tuning at Z=30% as the initial valve position

The loop was closed by a P-only controller with an initial gain $K_{c0} = 0.1$ and set-point was changed by 2 kPa, at Z=30 %. Then with respect to the data from step test and according to the method proposed by Jahanshahi (Jahanshahi and Skogestad 2013) explained in section 2.9.2.1, closed-loop stable system was identified as the following:

$$G_{cl}(s) = \frac{8.105\ \mathrm{S} + 0.919}{17.73S^2 + 3.765S + 1} \qquad \text{Equation 5.11}$$

Figure 5.28 illustrates the implemented step change and the identified closed-loop transfer function shown by the black line.

Then, the open-loop unstable system has been back calculated by using the procedure proposed by Jahanshahi. The open-loop unstable system has the form of:

$$\tilde{P}(s) = \frac{-4.572\ \mathrm{s} - 0.5184}{s^2 - 0.2448s + 0.00457} \qquad \text{Equation 5.12}$$

Then the IMC controller (C) is then designed by using the method explained in section 2.9.2.2. The time constant of the closed-loop system has been selected as $\lambda = 10$. This number was obtained by trial and error and experiencing different results. The designed IMC controller is:

$$C(s) = \frac{-0.11916(S^2 + 0.04668S + 0.001835)}{\mathrm{S}\ (\mathrm{S}+0.1134)} \qquad \text{Equation 5.13}$$

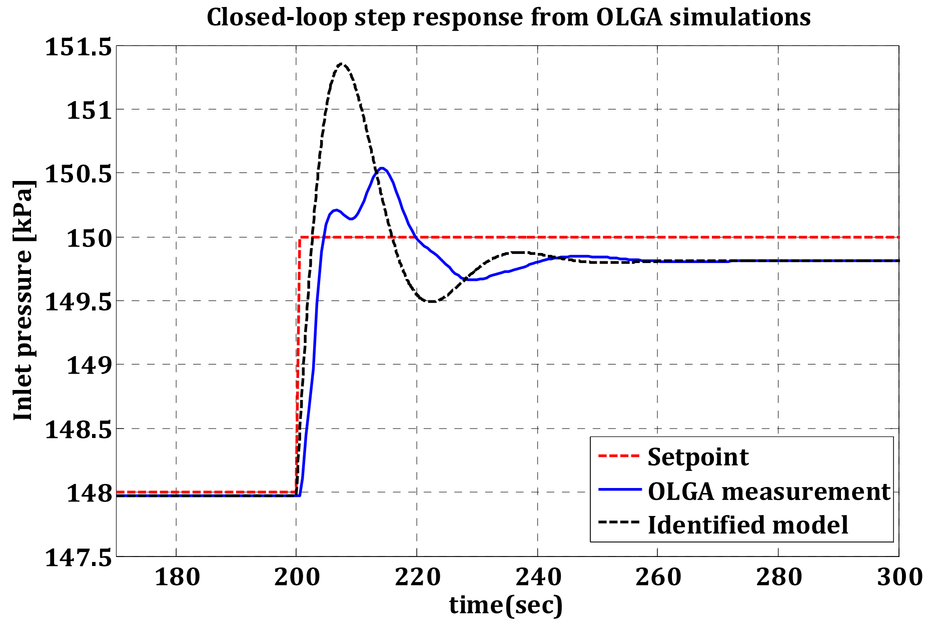

Figure 5.28: Closed-loop response of step change at initial valve opening Z=0.3. The dashed black line shows the transfer function of the IMC-based identified model.

The IMC controller is a second order transfer function and can be written in form of a PIDF controller. PIDF is a PID controller which a low-pass filter has been applied on its derivative action. It will be mentioned as PID controller.

A PI controller has been also obtained by reducing the order of IMC controller to one. The related PID and PI tuning parameters have been calculated as described in section 2.9.2.3 and are shown in table 5.11.

Table 5.11: IMC-based PID and PI tuning parameters tuned at the initial choke valve position of 30%

	K_{c0}	K_c	τ_I	τ_D	τ_F
PIDF	0.1	0.03204	16.6113	23.9802	8.8191
PI	0.1	0.11916	61.7797	-	-

Implementing low pass filter was not possible in OLGA. Therefore, despite the fact that the filter time constant was an important part of tuning parameters, it was neglected in simulations and a PID controller was using instead.

Figures 5.29 and 5.30 describe the results of control using the IMC-based PID and PI controllers respectively. The controllers were tuned for valve opening of Z=30%. But, they can stabilize the system up to very larger valve openings. The PID controller could stabilize the flow with a maximum of 50.27% valve opening and the PI controller could stabilize the system up to valve opening of Z=47%. The PID controller has shown a better performance compared to the PI. A lower set-point as well as a higher level of valve opening has been achieved with PID controller. In addition, the output from the PID controller is less oscillatory.

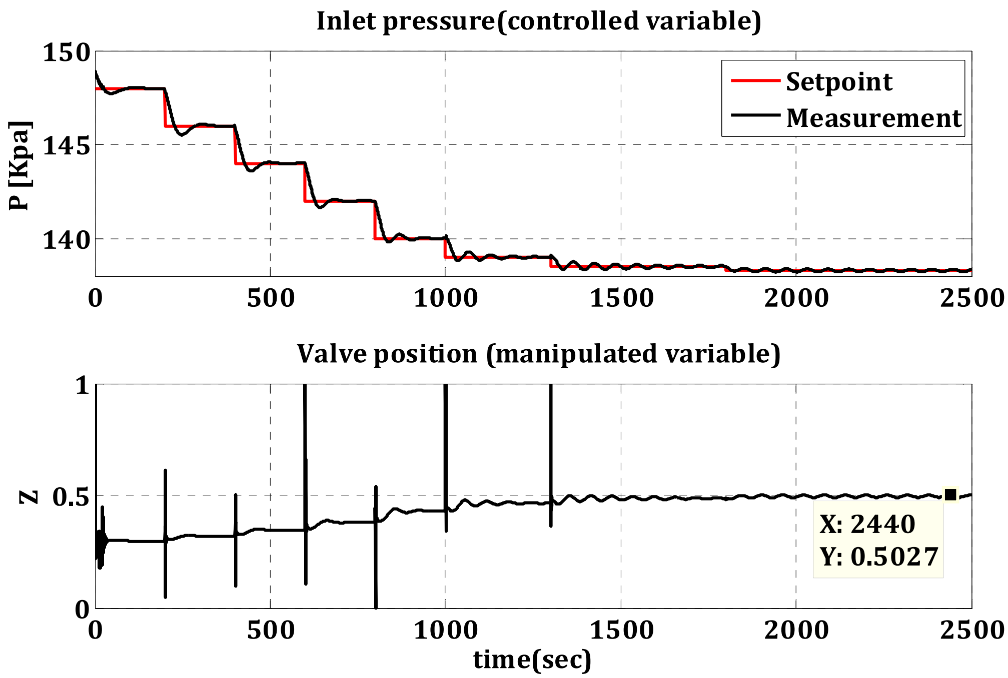

Figure 5.29: Simulation result of control using the IMC-based PID controller tuned at the initial choke valve position of 30%.

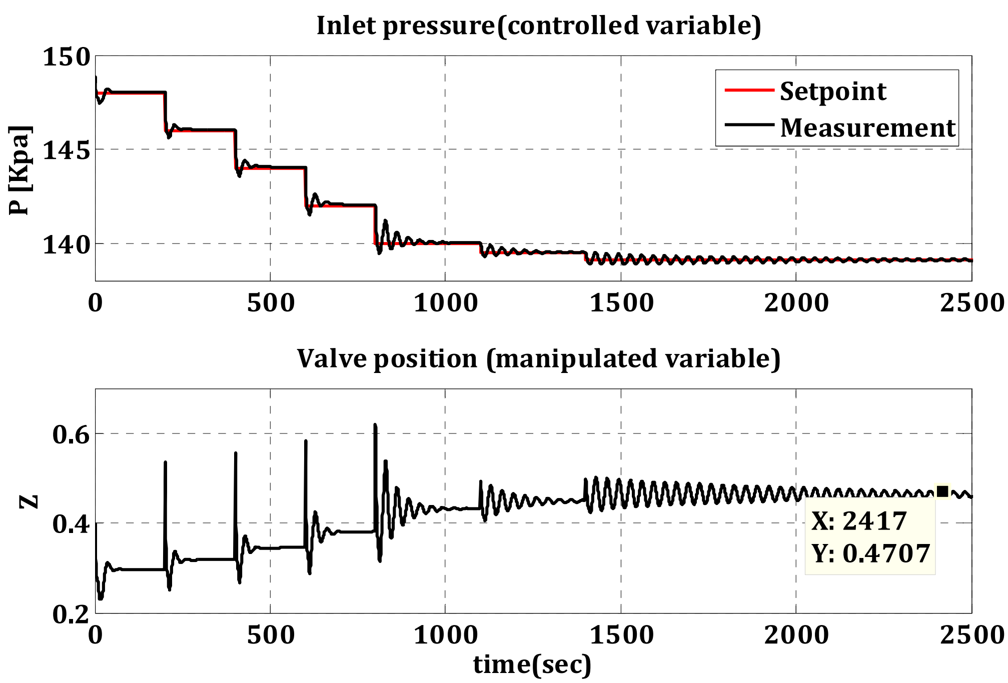

Figure 5.30: Simulation result of control using the IMC-based PI controller tuned at the initial choke valve position of 30%.

5.3.3.2.2 IMC-based tuning at Z=40% as the initial valve position

The same simulation as the one explained in previous section was run at Z=40%. The loop has been closed by a P-only controller with an initial gain $K_{c0} = 0.15$ and set-point has been changed by 1 kPa, at Z=40 %. The same procedure and calculations as described in previous section was used to find IMC-based PID and PI tuning parameters.

The closed-loop stable system was identified as the following:

$$G_{cl}(s) = \frac{7.011\ S + 0.805}{23.64S^2 + 2.27S + 1} \qquad \text{Equation 5.14}$$

The implemented step change and the identified closed-loop transfer function are illustrated in figure 5.31.

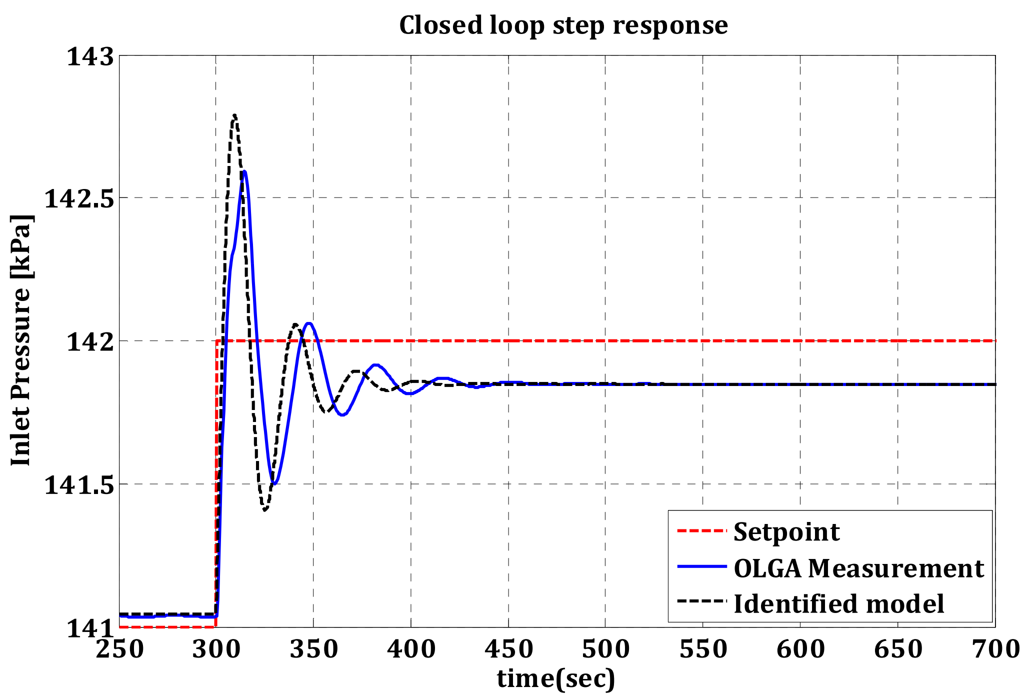

Figure 5.31: Closed-loop response of step change at initial valve opening Z=0.4. The dashed black line shows the transfer function of the IMC-based identified model.

The open-loop unstable system has the form of:

$$\tilde{P}(s) = \frac{-2.966\ S - 0.3405}{S^2 - 0.2006S + 0.00825} \qquad \text{Equation 5.15}$$

The designed IMC controller is:

$$C(s) = \frac{-0.16877(S^2 + 0.04345S + 0.001998)}{\text{S (S+0.1148)}} \qquad \text{Equation 5.16}$$

And finally the related PID and PI tuning parameters have been calculated as shown in table 5.12.

Table 5.12: IMC-based PID and PI tuning parameters tuned at the initial choke valve position of 40%

	K_{c0}	K_c	τ_I	τ_D	τ_F
PIDF	0.15	0.038293	13.0406	29.6774	8.7097
PI	0.15	0.16877	57.4753	-	-

Just like the previous part, the filter time constant was neglected due to impossibility of applying low-pass filter in OLGA and a PID controller was used instead.

Figures 5.32 and 5.33 describe the results of control using the IMC-based PID and PI controllers respectively, tuned for Z=40%. The controllers were tuned for valve opening of Z=40%. The PID controller could stabilize the system up to Z=54.61%. In fact with this controller, the bifurcation point has been moved from Z=26% into Z=54.61%. The PI controller could stabilize the system up to Z=51%.

As well as the result for the initial point of Z=30%, the PID controller shows a better performance with less oscillations in output and a higher level of valve opening has been achieved.

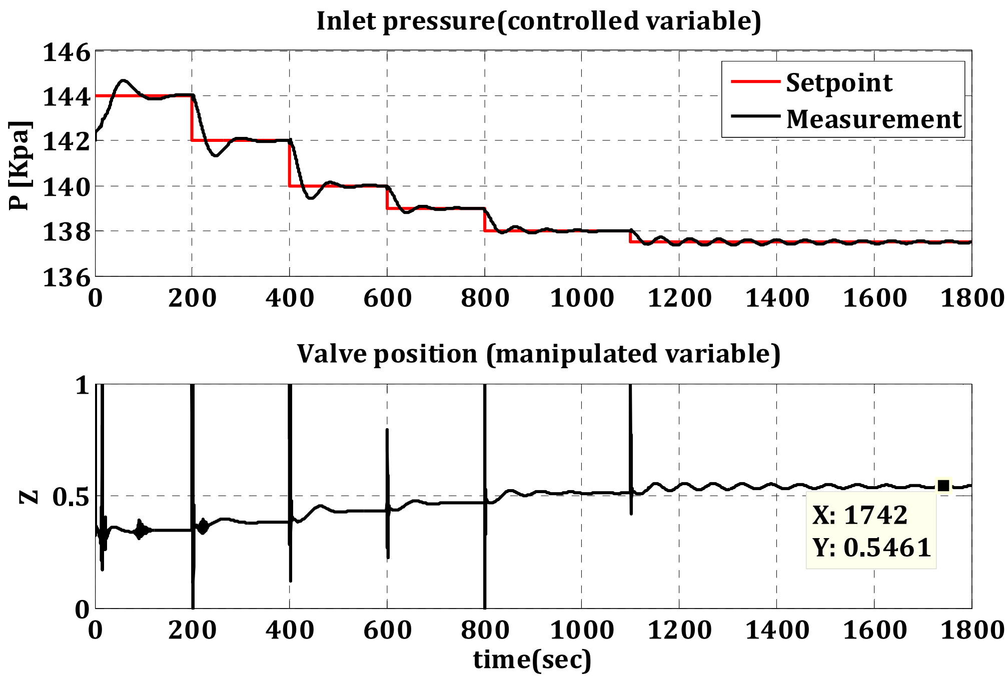

Figure 5.32: Simulation result of control using the IMC-based PID controller tuned at the initial choke valve position of Z=40 %.

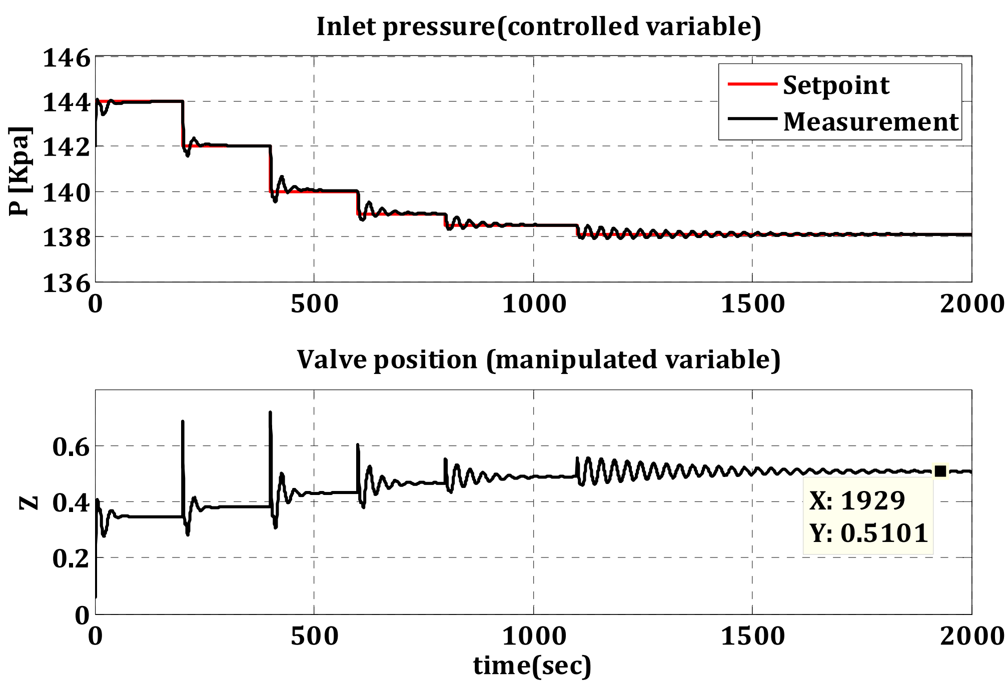

Figure 5.33: Simulation result of control using the IMC-based PI controller tuned at the initial choke valve position of Z=40%.

5.3.3.3 *Tuning using simple online method with gain scheduling*

Simple PI tuning rules based on identified MATLAB static model of nonlinear part of the system was used as the last tuning method in the simulations. The method has been proposed by Jahanshahi (Jahanshahi and Skogestad 2013) and described in section 2.9.3.

5.3.3.3.1 *Modifying MATLAB model*

As the first step in implementing this method the simple static MATLAB model of the system which tuning rules are based on, needed to be modified to be similar to the OLGA case used in the simulations of the thesis. As explained before, the OLGA case was the pipeline-S-shaped riser setup located at multiphase laboratory of NTNU.

As described in section 2.9.3, the simple model is based on the valve equation:

$$w = K_{pc} f(z) \sqrt{\rho \Delta p} \qquad \text{Equation 5.17}$$

For the valve used in OLGA simulations the valve characteristic is defined as:

$$f(z) = \frac{z.cd}{\sqrt{1 - z^2.cd^2}} \qquad \text{Equation 5.18}$$

Here *cd* is the discharge coefficient of the valve and had an important role in fitting the MATLAB model to the simulations. K_{pc} was considered as:

$$K_{pc} = \sqrt{2} A \qquad \text{Equation 5.19}$$

A is the cross sectional area of the pipe and finally the model was found as the following for the simulations:

$$k(z) = \frac{-2\overline{w}^2}{\rho.z^3.cd^2.K_{pc}{}^2} \qquad \text{Equation 5.20}$$

The model is a function of valve opening and therefor the value of inlet pressure and the static gain achieved at a specified operating point (valve opening) was different from the one in another operating point. Since the tuning parameters are found based on this model, it is very important that the model to be realistic, meaning that the values of inlet pressure and the static gain obtained by the model needed to be true values. In order to make a good match between the model and the OLGA case the geometry was changed to suit the experimental setup. However it soon became clear that the model needed to be manipulated to achieve the desired results. As the manipulated parameters, length of riser and the discharge coefficient of the valve were quite effective. A description regarding this issue will be presented below.

Length of riser as the first manipulated parameter

In MATLAB model length of riser is directly used to calculate the static pressure of the riser when it is filled with liquid and thereafter the static pressure of the riser is used to find the inlet pressure at any level of valve opening. Therefore manipulating that could be very helpful in producing desired results. The exact length of riser that was used in simulations is 7.7054 m*.* Though, it was changed to 5.15 m in model to provide the best results.

Discharge coefficient of choke valve (cd) as the second manipulated parameter

The coefficient of discharge in the valve equation is a constant which depends on the pressure drop over the valve. In order to fit the simple static MATLAB model to the OLGA case this parameter was manipulated. Decreasing the value of cd caused the model to have a better match with the simulations. The parameter *cd* used in OLGA case was 0.34 while a value of 0.31 was implemented in MATLAB mode.

The simplest way to check if the model is correct is comparing the values of inlet pressure and static gain from the MATLAB model by the same values from OLGA simulations. To do this, the steady state values of inlet pressure from OLGA simulations were used. The simulator gives the steady state values as the initial value of any variable including inlet pressure in the simulations. Therefor the initial value of the inlet pressure at each open-loop simulation for a specified valve opening was used to be compared with those of obtained from the model. Figure 5.34 compares simple static model to the OLGA case. As clear in the figure there is quite a good match between the model and OLGA. The MATLAB model is attached in Appendix C.7.

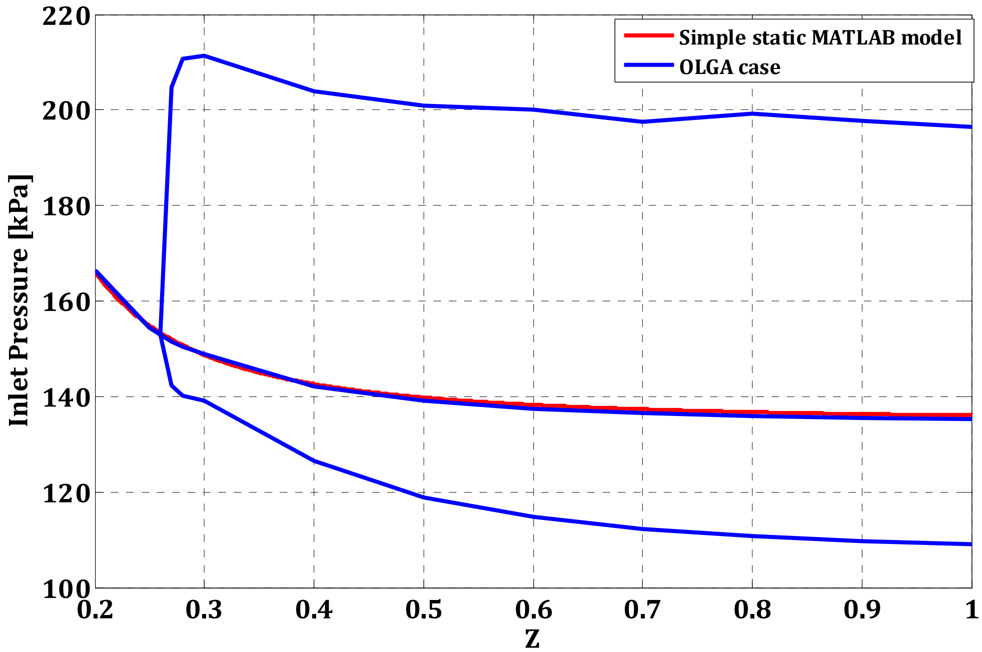

Figure 5.34: Simple static MATLAB model compared to the OLGA model. The blue midline in the figure presents the steady state values of the inlet pressure from OLGA simulations and the red midline is the values of inlet pressure from the MATLAB model. The top and bottom blue lines show the maximum and minimum values of pressure oscillations at each operating point in the open-loop system.

5.3.3.3.2 Calculating Tuning Parameters based on MATLAB model

In order to find tuning parameters based on the identified MATLAB model a closed-loop test with step change of set-point was required. The step test was done by a P-only controller as it was proposed by Jahanshahi (Jahanshahi and Skogestad 2013). The same step tests applied in section 5.3.3.2 were used here too. Two different step tests, one with the gain value of $K_{c0} = 0.1$ at the initial valve position of $Z_0 = 0.3$ and the other with the gain value of $K_{c0} = 0.15$ at the initial valve position of $Z_0 = 0.4$ were used to find two sets of tuning parameters. The method of how to find the tuning parameters has been described in section 2.9.3.2.

5.3.3.3.3 Results of tuning using initial valve position of Z_0=0.3

With respect to the information extracted from the step test, the parameter β has been found from the equation 2.41 as β = 0.2848. The period of slugging oscillations in open-loop simulations have been T_{osc} = 140 Sec. The model has been run for each operating point separately, meaning that the parameter Z has been changed after each running of the MATLAB model. The parameters $K_c(z)$ and $\tau_I(z)$ have been found as functions of valve opening *(Z)* by the equations 2.42 and 2.43 and are presented in table 5.13.

Table 5.13: PI tuning values in OLGA simulations with initial choke valve position of 30%

K_c	τ_I	Set-point (Inlet pressure) [kpa]	Valve opening
0.0499	484.6154	148.5	0.3000
0.0774	549.2308	145.5	0.3244
0.1142	613.8462	143	0.3616
0.1622	678.4615	141	0.4033
0.2229	743.0769	140	0.4307
0.2985	807.6923	138.5	0.4846
0.3908	872.3077	138	0.5084
0.5650	969.2308	137	0.5673
0.7477	1050	136.5	0.6061
0.9691	1130.8	136	0.6539
1.2338	1211.5	135.5	0.7145
1.5465	1292.3	135.3	0.7562

Then gain-scheduling with multiple controllers based on multiple identified models was used to stabilize the system. To do this in the simulations, 12 PI controllers were implemented in OLGA with the related found tuning parameters. The controllers could stabilize the flow up to 75.5 % of valve opening. Changing bifurcation point from Z=26 % into Z=75.5 % could be a very good result. Figure 5.35 illustrates the result of control using gain scheduling between PI controllers tuned for the initial choke valve position of Z=30%.

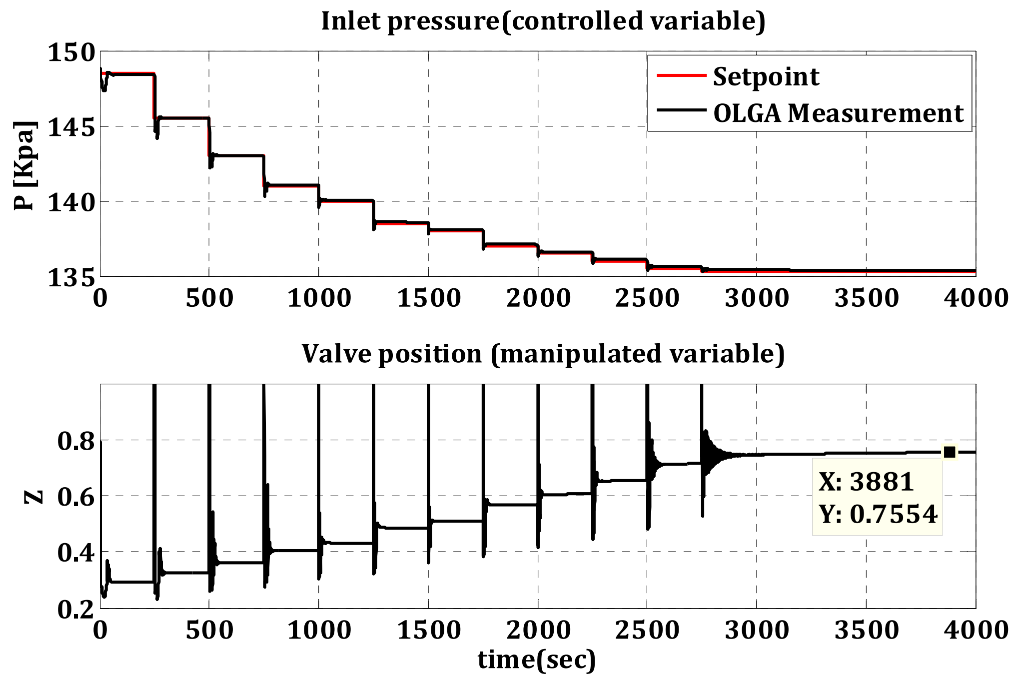

Figure 5.35: Simulation result of control using gain scheduling between PI controllers tuned for the initial choke valve position of Z=30%.

5.3.3.3.4 Results of tuning using initial valve position of Z_0=0.4

Everything has been done in the same way as explained in previous section for Z_0=0.3 except for the step test that has been run for Z_0=0.4. With respect to the information extracted from the step test, the parameter β has been found as: β = 0.8183. In order to do gain-scheduling with multiple controllers in the simulations, eight PI controllers were implemented in OLGA with the related found tuning parameters. The controllers could stabilize the flow up to Z=66.34 % of valve opening. Table 5.14 and Figure 5.36 describe the result of tuning and control using control using gain scheduling between eight PI controllers tuned for the initial choke valve position of Z = 40%.

Table 5.14 PI tuning values in OLGA simulation with initial choke valve position of Z=40%

K_c	τ_I	Set-point	Valve opening
0.3927	646.1538	141	0.4024
0.5482	710.7692	140	0.4323
0.7434	775.3846	138.5	0.4871
0.9838	840	138	0.5111
1.2751	904.6154	137	0.5713
1.8207	1001.5	136.5	0.6111
2.2661	1066.2	136.2	0.6398
2.7843	1130.8	136	0.6633

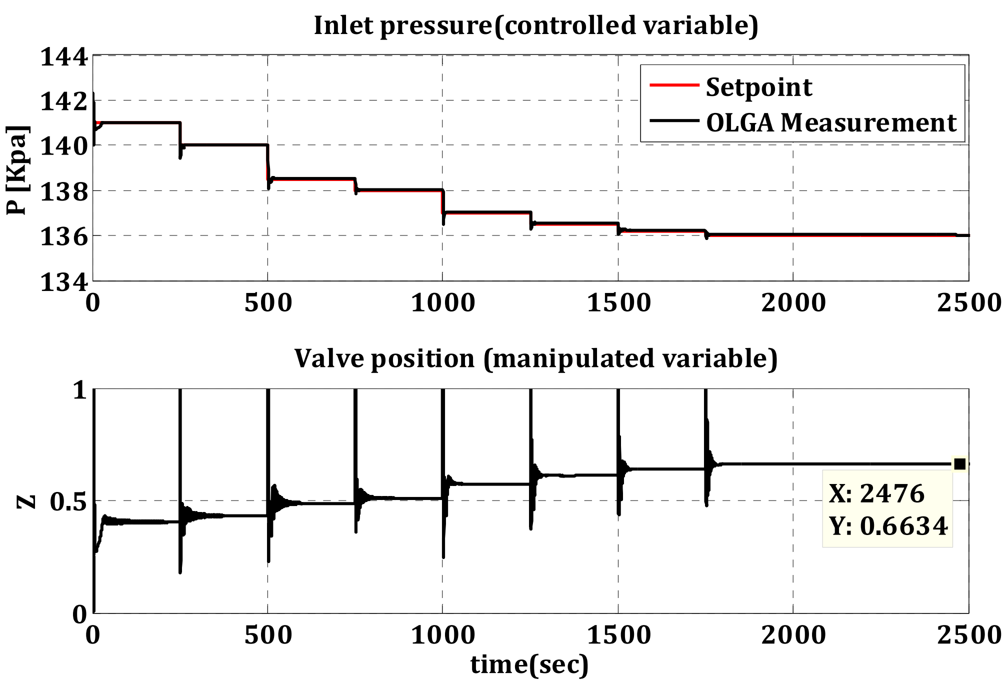

Figure 5.36: Simulation result of control using gain scheduling between PI controllers tuned for the initial choke valve position of Z = 40%.

5.4 Comparison of experimental and simulated results

The simulations in this thesis have been matched with the experimental models from valve1 (Slow valve). Therefore in case of numerical comparison, it is reasonable to compare simulated results with experimental results from valve1. But generally in case of comparison of different tuning methods and finding the best tuning approach for the slugging system the simulated results do agree with the experimental results from the both valves. In this section each tuning method would be discussed separately and the result of control from simulations and experiments will be compared. Finally a comparison of all tuning methods, used in the thesis, based on the simulations and both valves experiments will be presented.

5.4.1 Open-loop bifurcation diagrams

A comparison of the simulated open-loop results from the OLGA case and the experimental results from valve1 is shown in figure 5.37. It can be seen in the figure that the bifurcation point is fairly the same for the both models. It occurs at the same valve opening of Z=0.26 for both models but at a higher pressure for the experiments. Models are slightly deviated from each other. For the OLGA simulations the maximum of inlet pressure oscillations are located at higher values.

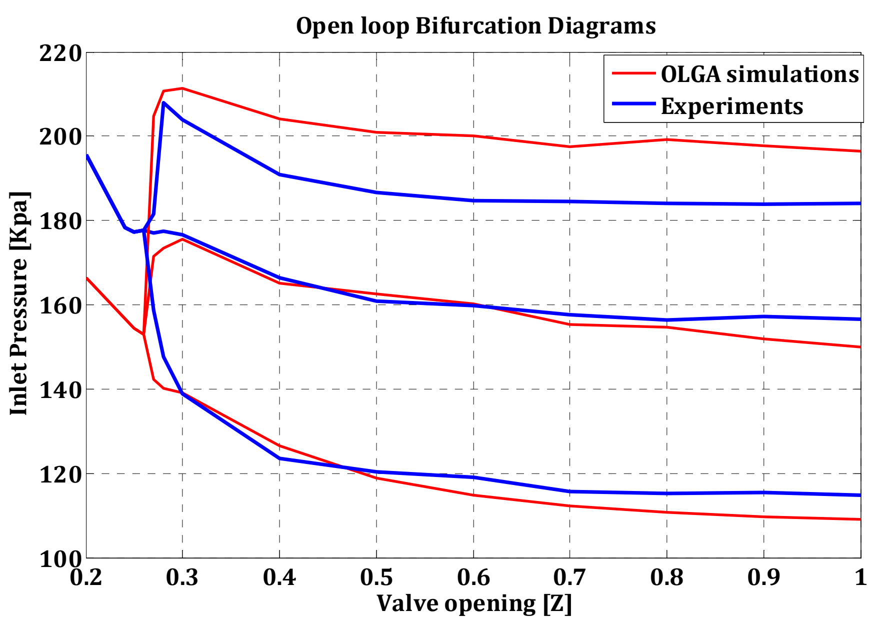

Figure 5.37: simulated results from the OLGA case compared with the experimental results from valve1. The bifurcation point is fairly the same for both models.

5.4.2 Comparison of control results from IMC-based tuning method

A comparison of simulated and experimental closed-loop responses from controllers tuned with IMC-based method is presented in table 5.15. Max Z shows the maximum valve opening achieved with that controller and Min P presents the minimum value of set-point that is inlet pressure in kilo Pascal. The numbers are the rounded values. The controllers have been tuned at the initial valve position of Z = 0.30.

Table 5.15: Comparison of simulated and experimental results from controllers tuned with IMC-based method. Z is the level of valve opening and P is the inlet pressure in KPa.

	Stability before control		Stability after control with IMC-based PI controller		Stability after control with IMC-based PID controller	
	Max Z	Min P	Max Z	Min P	Max Z	Min P
Experiments	0.26	177.8	0.38	154.5	0.40	154
Simulations	0.26	153	0.46	139.5	0.50	138.5

Although the simulated closed-loop results show a higher level of valve openings, still the amount of set-point reduction is larger for the experiments. Both models confirm that the IMC-based tuning method is a fine approach for the slugging system. Moreover they agree that the IMC-based PID controller has a better performance compared to the PI.

5.4.3 Comparison of control results from Simple online tuning method

Simple online tuning method based on MATLAB model was not tried in the series of experiments with valve1 *(See section 5.1.1.4 for more explanations)*. Instead it was tried with valve2. Therefore the results can't be numerically compared since the OLGA simulations are based on the experiments with valve1. However, the experimental and simulated results do agree on confirmation of this method as the best method of tuning with the highest level of stability for the slugging system. This will be seen more clearly in the next section.

5.4.4 Comparison of tuning methods

An overview of all experimental and simulated results from the applied tuning methods is presented in table 5.16 in numeric form. The maximum valve opening achieved as well as the minimum obtained set-point for each closed-loop test or simulation is illustrated.

It can be said that the best tuning method for the slugging system is the simple online PI tuning rules with gain scheduling for the whole operating range of the system based on MATLAB model. Tuning based on IMC design also works very well for the slugging system. These tuning methods are able to move the critical stability point significantly and considerably increase the production rate as a result.

It was also tried to make a clear comparison between the applied tuning methods by using figures. To do this the open-loop and the closed-loop bifurcation diagrams were plotted for the simulations and each series of experiments. Figure 5.38 compares the results of stabilizing control simulations by different tuning methods. Figures 5.39 and 5.40 do the same for the results of control experiments.

The bifurcation point as a sign of stability level is shown before and after control with each tuning method. The rightmost bifurcation point is related to the best tuning method that provided the most stability in each series.

It should be noted that Shams's method didn't work in the experiments. This is not surprising since it has been developed for the systems that are stable in open-loop while the slugging system is highly unstable. Simple online tuning method based on MATLAB model was not tried in the series of experiments with valve1 *(See section 5.1.1.4 for more explanations)*.

Table 5.16: Comparison of simulated and experimental results from all tuning methods. Z_{Max} is the maximum level of valve opening and P_{Min} is the minimum set-point achieved that is the riser inlet pressure in Kilo Pascal.

		Open-loop stability limit	Shams's set-point overshoot method	IMC-based PI tuning method	IMC-based PID tuning method	simple PI tuning with gain scheduling
Set 1 of Experiments with slow valve	P_{Min}	177.8	-	154.5	154	NOT Performed
	Z_{Max}	0.26	-	0.38	0.40	
Set 2 of Experiments with fast valve	P_{Min}	170.8	-	158.5	157.5	156
	Z_{Max}	0.16	-	0.29	0.30	0.35
OLGA simulations	P_{Min}	153	142	139.5	138.5	135.5
	Z_{Max}	0.26	0.38	0.46	0.50	0.75

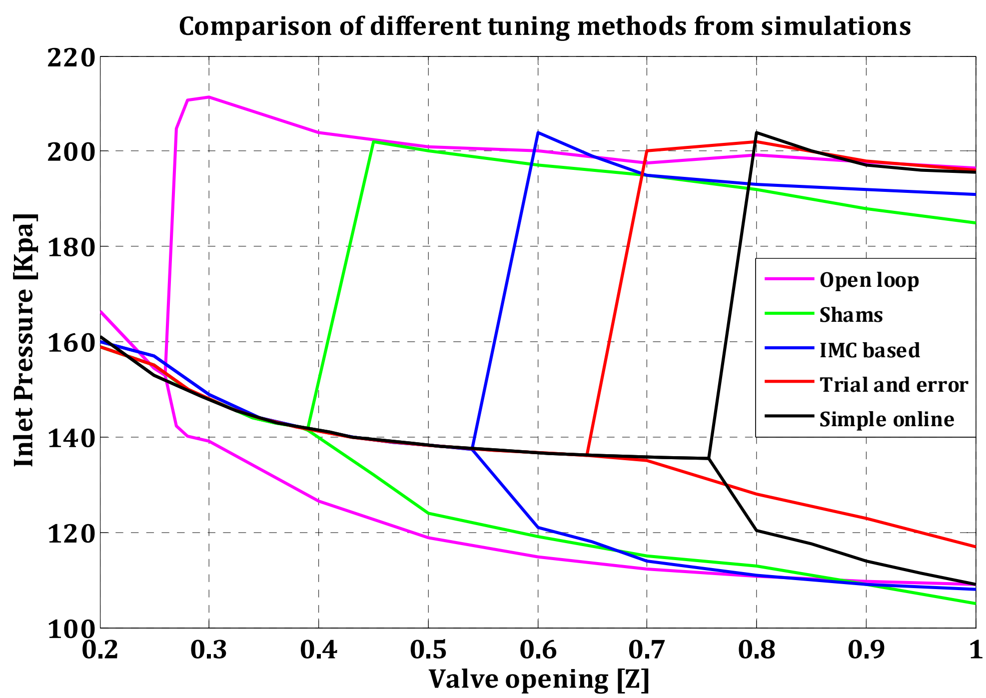

Figure 5.38: Comparison of stabilizing control results from different tuning methods applied in the simulations. It can be said that simple online method with gain scheduling is the most stabilizing and the IMC-based designed method is the second best as systematic manners to tune the controllers.

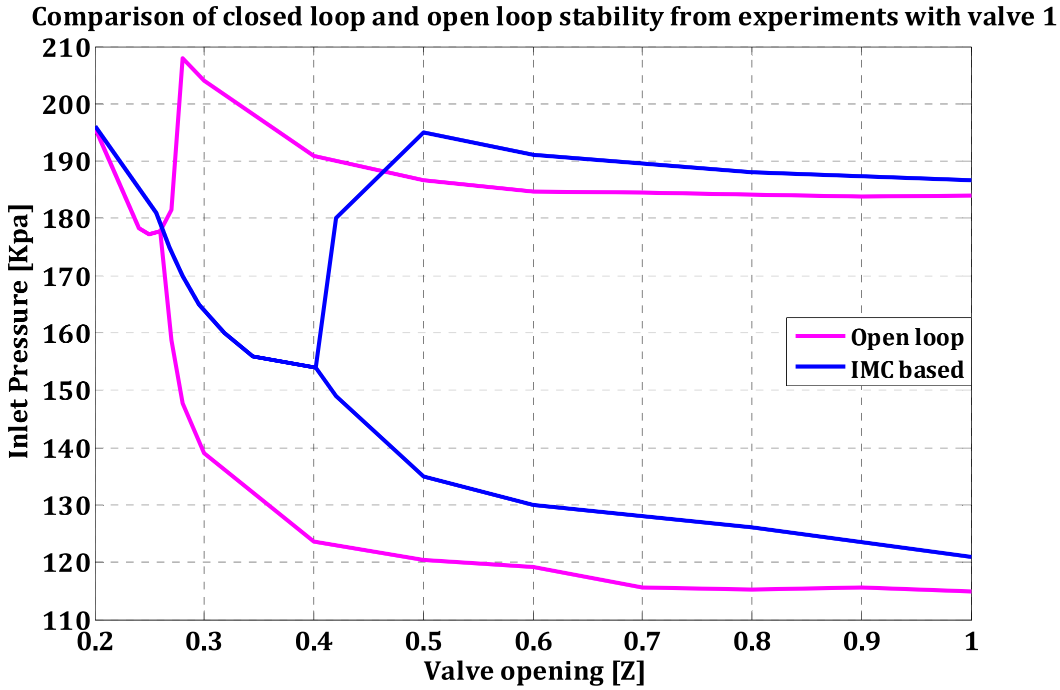

Figure 5.39: Comparison between the stabilizing control results from IMC-based tuning method and the open-loop system for the experiments with valve 1.

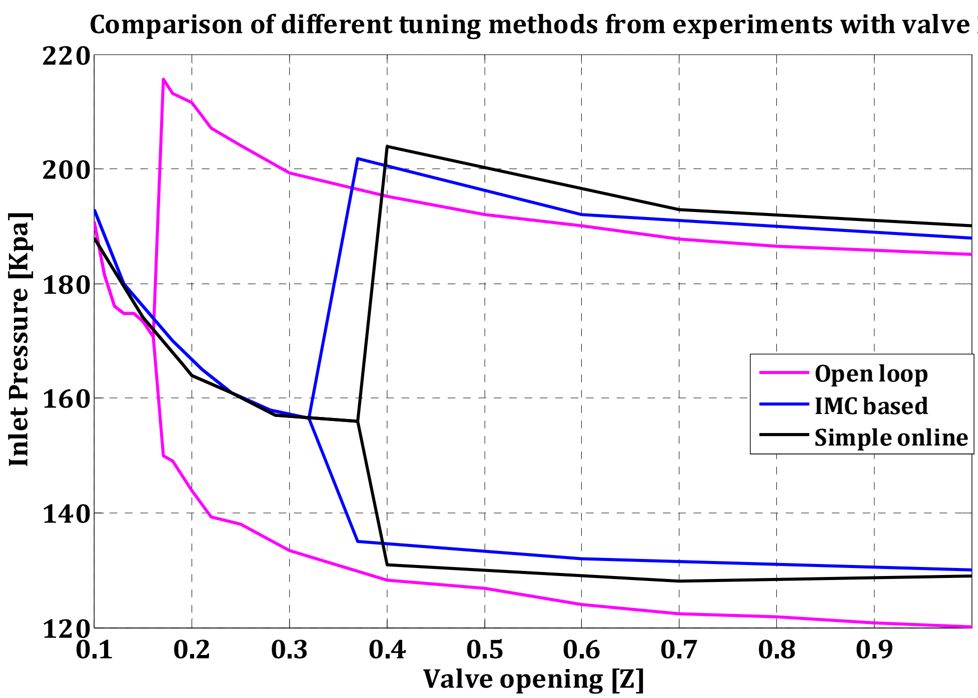

Figure 5.40: Comparison of stabilizing control results from different tuning methods applied in the experiments with valve 2. Simple online method with gain scheduling shows the best performance.

6 Discussion and further works

6.1 Tuning methods

The main objective in this thesis was to verify the very recently developed tuning methods (Jahanshahi and Skogestad 2013) by medium scale experiments and OLGA simulations and identify the most robust tuning method for the slugging system. From the results it can be seen that the highest level of stability is related to the controllers tuned by simple online method based on MATLAB model. It should be noted that in this thesis wherever simple online method has been applied gain scheduling between multiple controllers has been also performed. Since the slugging system is nonlinear and the gain of system changes drastically with changing level of valve opening, tuning the controllers at each operating point and then connecting them via gain scheduling has a huge effect on the control performance. It can be said that applying gain scheduling between the IMC-based controllers may also lead to a higher level of stability compared to applying a single IMC-based PID/PI controller. This can be tried in the future.

Generally, both IMC-based method and simple online method are very useful systematic approaches to tune the controllers for the slugging system. Previously, trial and error was mostly used for tuning the controllers in the slugging system. It can be said that the tuning rules used in this thesis are from the first systematic rules for anti-slug control and give very clear fine results.

In future works time delay can be added to the measurements in the experiments in order to have a better investigation of the system robustness. Actually this was tried in this thesis as an inconclusive effort *(See section 5.1.1.4.2).*

One important point needs to be mentioned in relation with tuning based on IMC design. The IMC-based PID tuning rules include a filter time constant that means an IMC filter must be implemented on the derivative action of the PID controller. This was impossible in OLGA and therefore had to be neglected. Although the simulation results do agree with the experimental results it can't be denied that neglecting filter action deviates the simulated results from the reality. This may be possible in future versions of the simulator.

About simple online method, the MATLAB model is discussable. From the results it can be observed that the MATLAB model did fit to the simulated results from OLGA and also the experimental results. However, the manipulations done to fit the model to the simulations and experiments may lead to the inaccuracy of the results. Also, in the MATLAB model the valve is assumed to have a linear characteristic; however this may not be the case for the valve in the experiment *(See section 5.2).*

Shams's tuning method has been designed for the stable systems while the slugging system is unstable. Therefore it may not be far from the expectation that it couldn't work for the slugging system. This method didn't work in any of the experimental series. Though it worked in simulations but didn't give very good results. This even small stability found with this method in OLGA simulations may deviate from the reality. This deviation may be due to inappropriate assumptions or inaccurate initial and boundary conditions in OLGA model. The overall result can be that this method can't be a suitable one to tune the controllers in the slugging system. Instead, the recently developed IMC-based and simple online methods perform much better.

6.2 Control structures

The control structure used in the series of experiments and simulations was a SISO control with buffer pressure as the control variable. This measurement is the riser inlet pressure in real subsea systems and may not be very easy to measure. However it has been proved previously that it's the best control variable for the active control of severe slugging (Jahanshahi, Skogestad et al. 2012) (Meland 2011) . In the article by Jahanshahi (Jahanshahi, Skogestad et al. 2012) one pressure measurement from the pipeline combined with choke flow rate has been suggested as the best measurements for a multivariable structure. At the beginning of the thesis it was decided to try a similar structure with top pressure combined with riser outflow density as the measurements. But this didn't become practical during the thesis due to the inconvenient density sensor *(See section 5.1.3).* The new tuning methods applied in this thesis can be tried by other measurements and control structures in the future. As the first step an accurate density sensor shall be used to give correct measurements of densities. Then it can be used in the new control structures.

6.3 Discussable issues related to experimental activities

6.3.1 Oscillations in flow rates

In order to have a fixed U_{sl} and U_{sg} in each test it was important to have constant and consistent flow rates. The air and water flow rates had many oscillations and it was very difficult to set the exact required flow rates. Specially, for the case of air this problem was more challenging. The reason was that the control valve for the air was broken and the air flow rate had to be set with a manual valve far from the screen. The manual valve made a big change in air flow rate even when it was tried to open or close it very little. It was necessary to go and come many times to make a flow rate close to the desired value. For water flow rate the centrifugal water pump was the reason of oscillations. However, it was tried to deal with this issue through running the pump in a very high level of power (80% of the maximum) and opening the water control valve in small values, instead.

6.3.2 Water flow back into the buffer tank

When the buffer pressure became lower than the pressure inside pipeline, water did flow back into the buffer tank. This reduced the volume of buffer tank and caused the buffer pressure deviates from the real values. This discrepancy could distract the controller performance and therefore it was very important to remember to drain the buffer tank between the experiments. Installing an automatic sensor to quickly sense the water in the buffer tank could be very helpful to overcome this issue.

6.3.3 Leakage in steel connection

The laboratory facility was in a way that a single pipeline needed to be connected to any of the risers (Steel S-riser, Hose L-riser or Horizontal pipeline). On the other hand several people were working in the lab and on the different setups during the semester. This caused the pipeline-risers connections needed to be changed several times a week. This was not a very easy job and sometimes the connection couldn't be fitted quite well even with trying many different sealing rubber O-rings, screws and nuts. Therefore there was some flow leakage from the connection during the work. This could affect the accuracy of the experiments since the flow meters were located before this connection.

However, the flow meters themselves were not of the best quality and their numbers may be also inaccurate. One way to overcome this occasional leakage is to make a multiple connection between the pipeline and all risers with the manual valves for each connection. Then the valves can be manipulated to change flow directions instead of the time consuming change of the connections by mechanical work.

7 Conclusion

This chapter is organized based on the tasks defined in the thesis description. These tasks have been followed and the desired results have been obtained mostly.

7.1 Stabilizing control experiments using bottom pressure

Stabilization control experiments using the medium scale S-riser setup proved that the severe slugging phenomena can be delayed to a large extent by active control of production choke valve and using the bottom pressure (buffer tank pressure) as the control variable. Two sets of experiments with two different choke valves showed that the anti-slug control structure using bottom pressure as measurement and a good tuning method as well, the stability region could be extended widely.

7.2 Testing online tuning rules on S-riser experiments

Three different tuning methods for anti-slug control were tested online and their robustness was compared with respect to the stability limits they provided *(See table 5.16 and also figures 5.39 and 5.40)*.

The Shams's set-point overshoot method (Shamsuzzoha and Skogestad 2010) failed to stabilize the system in both sets of experiments. This was not far from the expectation, since Shams's method has been developed for the stable systems while the slugging system is highly unstable.

For implementing the IMC (Internal Model Control) based tuning method (Jahanshahi and Skogestad 2013) the model of the system was identified from a closed-loop step test. The identified model was used for an IMC design, and then PID and PI tunings were obtained from the resulted IMC controller. The IMC-based PID tuning rules could increase the stability limit from 26 % to 40 % of choke valve opening in the first set of experiments using the slow valve and from 16 % to 30 % of choke valve opening in the second set of experiments using the fast valve.

The simple PI tuning rules with gain scheduling for the whole operating range of the system, was used as the last tuning method and proved to be the best tuning approach for the slugging system. To implement this method, a MATLAB model was modified and fitted to the steady state model of experiments. Then based on this MATLAB model and also a single closed-loop step test, the simple PI tuning rules were found. This tuning method could increase the stability limit from 16 % to 35 % of choke valve opening in the second set of experiments using the fast valve.

7.3 Control using top pressure combined with density

Measurement of the topside density using a conductance probe installation was not successful. The open-loop step test proved that the probe is not applicable as an appropriate sensor to measure the flow density. The probe signal couldn't show a clear response to the step change and therefore was not a suitable measurement for the control targets *(See figure 5.16).* In order to have an efficient cascade control with density as the inner loop control variable, more accurate signals of density are required. Therefore no cascade anti-slug control schemes could be tested.

7.4 Investigating effect of control valve dynamics

The criterion to evaluate the slugging control loop is the stability and since the valves' inherent characteristics are different, the level of valve opening can't be used to compare the valves' performance in the control loop. Instead, the minimum achievable set-points in the closed-loop responses and also the achieved range of set-point reduction were used to compare the valve behaviors. From the closed loop responses, it was proved that the slow valve has a better performance compared to the fast valve. This means that the slow valve has been already fast enough for our control targets and there has been no need to valve 2 (faster control valve). In other words the stability of the slugging system is more affected by the tuning parameters for the controller instead of control valve dynamics. Figure 5.19 compares the closed loop response of IMC-based controller for the two valves. With the slow valve, the IMC-based controller has been able to decrease set-point in a wider range, down to a lower level.

7.5 Control simulations using OLGA

The OLGA model was developed based on the first series of experiments with valve 1 and the implemented PID controller was fine-tuned using the different tuning strategies. Results of the experiments verified those of the simulations.

In open-loop condition there was a good match between the OLGA model and the experimental model of valve 1 *(see figure 5.37)*.

The same as the experimental results, the simulated ones proved that simple PI tuning rules with gain scheduling for the whole operating range of the system (Jahanshahi and Skogestad 2013) is the best tuning method providing the largest stability region for the slugging system. The PI controller in the simulations, tuned by this method, could increase the stability limit up to the valve opening of Z= 75 % from the open loop stability of Z= 26 % *(see table 5.16 or figure 5.38).*

From the simulation results it can be said that the IMC-based tuning method (Jahanshahi and Skogestad 2013) is the second best systematic manner to tune the controllers for the slugging system. The PID controller tuned by this method, increased the stability limit from 26 % to 50 % of choke valve opening.

The Shams's set-point overshoot method (Shamsuzzoha and Skogestad 2010) was used to tune the PI controller in two initial points of Z=30% and Z=40% in the simulations. The one tuned at the initial point of Z=30% could surprisingly stabilize the system up to the valve opening of Z=38 %. However, the other one tuned at the initial point of Z=40% wasn't able even to achieve the stability for the point that has been tuned for.

8 References

Bai, Y. (2001). Pipelines and risers, Elsevier Science.

Bratland, D. O. (2010). "The Flow Assurance Site." from http://www.drbratland.com/PipeFlow2/chapter1.html.

Fabre, J., L. Peresson, et al. (1990). "Severe slugging in pipeline/riser systems." SPE Production Engineering **5**(3): 299-305.

Godhavn, J.-M., M. P. Fard, et al. (2005). "New slug control strategies, tuning rules and experimental results." Journal of process control **15**(5): 547-557.

Havre, K., K. O. Stornes, et al. (2000). "Taming slug flow in pipelines." ABB review **4**: 55-63.

Jahanshahi, E. and S. Skogestad (2011). Simplified dynamical models for control of severe slugging in multiphase risers. World Congress.

Jahanshahi, E. and S. Skogestad (2013). Closed-loop model identification and PID/PI tuning for robust anti-slug control. Mumbai, India, 10th IFAC International Symposium on Dynamics and Control of Process Systems.

Jahanshahi, E., S. Skogestad, et al. (2012). Controllability analysis of severe slugging in well-pipeline-riser systems. IFAC Workshop-Automatic Control in O shore Oil and Gas Production, Trondheim, Norway.

Jansen, F., O. Shoham, et al. (1996). "The elimination of severe slugging—experiments and modeling." International journal of multiphase flow **22**(6): 1055-1072.

Kazemihatami, M. (2012). Experiments on Liquid Flushing in Pipes. Master Project. Trondheim, Norwegian University of Science and Technology.

Lilleby, K. (2003). User's Manual for Multiphase Flow Loop. Trondheim, Norwegian university of science and technology.

Malekzadeh, R., R. Henkes, et al. (2012). "Severe Slugging in a Long Pipeline-Riser System: Experiments and Predictions." International journal of multiphase flow.

Meland, K. O. (2011). Stabilization of two-phase flow in risers from reservoirs. Chemical Engineering. Trondheim, Norwegian University of Science and Technology. **Master**.

Miyoshi, M., D. Doty, et al. (1988). "Slug-catcher design for dynamic slugging in an offshore production facility." SPE Production Engineering **3**(4): 563-573.

Morari, M. and E. Zafiriou (1989). Robust process control, Morari.

Olsen, H. (2006). Anti-slug control and topside measurements for pipeline-riser system, Master's thesis, Norwegian University of Science and Technology.

Pickering, P., G. Hewitt, et al. (2001). The prediction of flows in production risers-truth & myth. IIR Conference.

Rivera, D. E., M. Morari, et al. (1986). "Internal model control: PID controller design." Industrial & engineering chemistry process design and development **25**(1): 252-265.

Shamsuzzoha, M. and S. Skogestad (2010). "The setpoint overshoot method: A simple and fast closed-loop approach for PID tuning." Journal of process control **20**(10): 1220-1234.

Sivertsen, H. (2008). Stabilization of desired flow regimes. Department of Chemical Engineering Norwegian University of Science and Technology. **Master**.

Skogestad, S. (2003). "Simple analytic rules for model reduction and PID controller tuning." Journal of process control **13**(4): 291-309.

Skogestad, S. and C. Grimholt (2011). The SIMC method for smooth PID controller tuning. PID Control in the Third Millennium, Springer**:** 147-175.

Storkaas, E. (2005). Control solutions to avoid slug flow in pipeline-riser systems. Chemical Engineering. Trondheim, Norwegian University of Science and Technology. **PhD**.

Storkaas, E. (2005). Stabilizing control and controllability. Control solutions to avoid slug flow in pipeline-riser systems, Norwegian University of Science and Technology.

Taitel, Y. (1986). "Stability of severe slugging." International journal of multiphase flow **12**(2): 203-217.

Yan, K. and D. Che (2011). "Hydrodynamic and mass transfer characteristics of slug flow in a vertical pipe with and without dispersed small bubbles." International journal of multiphase flow **37**(4): 299-325.

A. Low pass filter in LabVIEW

In order to implement the low-pass filter in the experiments the function "*PID Advanced VI"* from LabVIEW was used. The function implements a PID controller using a PID algorithm with advanced optional features. Figure 5.6, adapted from the National Instruments' website, shows the block diagram of related the function.

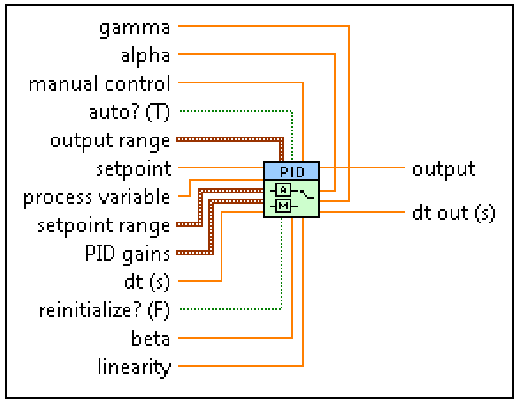

Figure A.1: PID Advanced (DBL)

In the presented figure *alpha* specifies the derivative filter time constant and can be a value between 0 and 1. The default is NaN, which specifies that no derivative filter is applied. The relation between τ_F from the method and α from LabVIEW is as follows:

$$\alpha = \frac{\tau_f}{\tau_D} \qquad \text{Equation A.1}$$

B. Simulated results to get the best step tests for Shams's method

In this appendix all simulations run to get the desired overshoot at different basis conditions of the controller are presented. The simulations are related to tuning of the controller by Shams's method (*See section 5.3.3.1*). Table B.1 presents the initial and final values of buffer (inlet) pressure used as control variable in simulations before and after step change and the resulting overshoot. The units are in kilo Pascal. K_{c0} is the initial gain used in the tuning simulations. The values specified by the red color are the best results those were used to find tuning parameters.

Table B.1: Resulting overshoots to the different step tests at different initial positions of choke valve

Bias	K_{c0}	Initial set-point value	Final set-point value	Overshoot
0.30	0.5	142	143	3.4058
		142	144	1.7426
		145	147	2.1049
	0.1	142	143	0.5741
		142	144	0.6318
		142	145	1.2219
		143	144	0.5402
		143	145	0.5609
		144	145	0.4894
		149	144	0.3108
		149	150	0.3308
		150	151	0.3085
0.40	0.1	141	142	0.8839
		145	142	0.4471
		145	146	0.5028
		148	149	0.3535
	0.15	145	142	0.3210

C. Some examples of MATLAB scripts

C.1 Tuning by Shams's method

```
clc
clear all

load Data

Kc0 = -0.1;
dy_s = 1;
t_init = 200;
dt = 0.1;

t = Data(:,1);
r = Data(:,2);
y = Data(:,3);
u = Data(:,4);
%%
figure(1)
clf
subplot(2,1,1)
plot(t,r,'r',t,y,'k','Linewidth',1.5)
xlim([0,1000])
title('Inlet pressure(controlled variable)')
ylabel('P [Kpa]')
legend('Setpoint','Data',2)
grid on
hold on

subplot(2,1,2)
plot(t,u,'k','Linewidth',1.5)
xlim([0,1000])
xlabel('time(sec)')
ylabel('Z')
title('Valve position (manipulated variable)')
grid on

%%
i_init = find(t==t_init);
y_plant = y(i_init:end);
t_plant = t(i_init:end);
u_plant = u(i_init:end);
y_init = y(i_init-200);
u_init = u(i_init-200);
yp = max(y_plant);
dy_p = abs(yp - y_init);
i_yp = find(y_plant==yp);
t_p1 = mean(t_plant(i_yp));
yu = min(y_plant(i_yp:10*i_yp));
dy_u = abs(yu - y_init);
i_yu = find(y_plant==yu);
t_u = mean(t_plant(i_yu));
y_inf = y_plant(end);
dy_inf = abs(y_inf - y_init);
tp = t_p1 - t_init;
```

```
Overshoot = abs((dy_p-dy_inf)/dy_inf);
D = Overshoot
Offset = abs((dy_s-dy_inf)/dy_inf)
B = Offset;
A = 1.152*D^2 - 1.607*D +1;
r = 2*A/B;
K = 1/(Kc0 * B);
Tetha = tp*(0.309 + 0.209*exp(-0.61*r));
tau1 = r*Tetha;
Kc = tau1/(K*2*Tetha)
tauI = min (tau1, 8*Tetha)
Tau_c=Tetha
```

C.2 Model identification based on IMC-design

```
clc
clear all
close all

load z30_148_150
Kc0 = -0.1;
dy_s = 2;
t_init = 200;
dt = 0.1;

t = z30_148_150(:,1);
y = z30_148_150(:,2);
r = z30_148_150(:,3);
u = z30_148_150(:,4);

%%

figure(1)
plot(t,r,'--r','LineWidth',2.25);
hold on
plot(t,y,'b','LineWidth',2.25);
xlabel('time(sec)');
ylabel('Inlet pressure [kPa]');
xlim([170 300]);
title('Closed-loop step response from OLGA simulations');
grid on
hold on

%%
i_init = find(t==t_init);
y_plant = y(i_init:end);
t_plant = t(i_init:end);
u_plant = u(i_init:end);
y_init = y(i_init-100);
u_init = u(i_init-100);
yp1 = max(y_plant);
dy_p1 = abs(yp1 - y_init);
i_yp1 = find(y_plant==yp1);
t_p1 = mean(t_plant(i_yp1));
yu = min(y_plant(i_yp1:10*i_yp1));
dy_u = abs(yu - y_init);
i_yu = find(y_plant==yu);
t_u = mean(t_plant(i_yu));
```

```
yp2 = max(y_plant(i_yu:2*i_yu));
dy_p2 = abs(yp2 - y_init);
y_inf = y_plant(end);
dy_inf = abs(y_inf - y_init);


D0 = (dy_p1 - dy_inf)/dy_inf;
deltaT = t_u - t_p1;
v1 = (dy_inf - dy_u)/(dy_p1 - dy_inf);
z = -log(v1)/sqrt(pi^2+(log(v1))^2);
Tau = (deltaT/pi)*sqrt(1-z^2);
K = dy_inf/(dy_inf-dy_s);
K2 = K/(K-1);
alpha = (K+1)/(K-1);
Tau1 = 2*z*Tau*(K-1)+sqrt(4*z^2*Tau^2*(K-1)^2+(K+1)*(K-1)*Tau^2);

tp = t_p1 - t_init;
Phi = atan((1-z^2)/z)-tp*sqrt(1-z^2)/Tau;
E = sqrt(1-z^2)/Tau;
D1 = D0/(exp(-z*(tp)/Tau)*sin(E*(tp)+Phi));
Tauz = z*Tau + sqrt(z^2*Tau^2-Tau^2*(1-D1^2*(1-z^2)));

s=tf('s');
disp('The identified closed loop model:')
G2 = K2*(1+Tauz*s)*exp(-0*s)/(Tau^2*s^2 + 2*z*Tau*s + 1)

u = [zeros(1,round(t_init/dt)) dy_s*ones(1,round((3600-t_init)/dt)+1)];
t = 0:dt:3600;
y1 = lsim(G2,u,t);
plot(t,y1+y_init,'--k','LineWidth',2.25);
legend('Setpoint','OLGA measurement','Identified model',3);

%%
%BACK CALCULATION OF THE OPEN LOOP UNSTABLE SYSTEM%%

A0 = 1/Tau^2;
A1 = 2*z/Tau;
B0 = K2/Tau^2;
B1 = K2*Tauz/Tau^2;

Kp = dy_inf/(Kc0*abs(dy_s-dy_inf));
a0 = A0/(1+Kc0*Kp);
b0 = -Kp*a0;
b1 = -B1/Kc0;
a1 = -A1-Kc0*b1;

s = tf('s');
disp('Identified model:')
Ge = (-b1*s-b0)/(s^2-a1*s+a0)
gcl = feedback(Kc0*Ge,1);
```

C.3 Design of Internal Model Controller

```
%% Internal Model Controller (IMC)
% Plant Information
[Zero,Pole,Gain,Ts] = zpkdata(Ge,'v');

indRHPzero = (real(Zero)>0);                    % indices of open RHP
zeros
indRHPpole = (real(Pole)>0);                    % indices of open RHP
poles
RHPpoles = Pole(indRHPpole);                    % RHP poles
NumRHPzeros = sum(indRHPzero);                  % number of open RHP
zeros
NumRHPpoles = sum(indRHPpole);                  % number of open RHP
poles

Tauc = 10; % Tuning parameter: time constant of the closed-loop system

% for MP systems
q_tilde = zpk(Pole,Zero,1/Gain);

k = NumRHPpoles+1; % since Vm always contains an pole at origin for step
input
m = max(length(zero(q_tilde))-length(pole(q_tilde)),1); % make sure
q=q_tilde*f is proper
filterOrder = m+k-1;

% 3. calculate filter as sum(a(k)s^k)/(tau*s+1)^filterOrder
coefficients = ones(1,k);

if NumRHPpoles>0
    A = zeros(NumRHPpoles,NumRHPpoles);
    for ctRHPpole = 1:length(RHPpoles)
        A(ctRHPpole,:) = RHPpoles(ctRHPpole).^(1:NumRHPpoles);
    end
    b = (Tauc*RHPpoles+1).^filterOrder-coefficients(1);
    coefficients(2:end) = (real(A\b))';
end
% computing f
num = fliplr(coefficients);
den = fliplr(poly(repmat(-Tauc,1,filterOrder)));
f = tf(num,den);

q = minreal(q_tilde*f);
C = feedback(q,Ge,+1);
disp('The IMC controller:')
C = minreal(C)
L1 = C*Ge;
allmargin(L1)
```

C.4 Finding PID/PI tuning rules based on IMC-design

```
disp('IMC based PID tuning:')

[Kc_PID,Ki_PID,Kd_PID,Tf_PID] = piddata(C)

Ti_PID = Kc_PID/Ki_PID;
Td_PID = Kd_PID/Kc_PID;


disp(['Kp = ' num2str(Kc_PID)])
disp(['Ti = ' num2str(Ti_PID)])
disp(['Td = ' num2str(Td_PID)])
disp(['Tf = ' num2str(Tf_PID)])
disp('FEED TO OLGA AND CLOSE THE LOOP!')

C2 = Kc_PID*( 1 + 1/(Ti_PID*s) + Td_PID*s/(Tf_PID*s+1));
L2 = C2*Ge;
allmargin(L2)

%%

%Reduce to PI Controller
C3 = balancmr(C,1);
[Kc_PI,Ki_PI] = piddata(C3);
Ti_PI = Kc_PI/Ki_PI;
disp('IMC based PI tuning:')
disp(['Kp = ' num2str(Kc_PI)])
disp(['Ti = ' num2str(Ti_PI)])

disp('FEED TO OLGA AND CLOSE THE LOOP!')

C3 = Kc_PI*(1+1/(s*Ti_PI));
L3 = C3*Ge;
allmargin(L3)
```

C.5 Simple static model fitted to experiments

```
%%%%Simple Static Model%%%
clc
clear all


g = 9.81;                           %Gravity (m/s2)
Wg_in=0.0024;                       %Inlet mass flow rate of gas (Kg/sec)
Wl_in=0.39298;                      %Inlet mass flow rate of liquid(Kg/sec)
W = Wg_in+Wl_in;                    %Inlet mass flow rate (Kg/sec)
R = 8314;                           %Gas constant (J/(K.Kmol))
M_g = 29;                           %Molecular weight of Gas (kg/kmol)

p_s=101325;                      %Separator pressure (pa)
p_vmin = 0;                      %minimum Pressure drop over the valve (Pa)
T=15+273.15;                     %Riser temperature (K)
par.r2 = 0.025;                  %Radius of riser (m)
par.A2 = pi*par.r2^2;            %Cross section area of riser (m2)
rho_g= (p_s+p_vmin)*M_g/(R*T);   %Average gas density at the outlet(Kg/m3)
```

```
rho_l=1000;                          %Liquid density (Kg/m3)

alpha_g = Wg_in/(Wg_in+Wl_in);   %Average gas mass fraction
alpha_l=(1-alpha_g)*rho_g/((1-alpha_g)*rho_g+alpha_g*rho_l); %liquid volume
fraction

rho = alpha_l*rho_l+(1-alpha_l)*rho_g;     %Average mixture density
L_r = 5.15;                                %Length of riser
z_star= 0.26;                              %Bifurcation point
cd = 0.31;                                 %Discharge coefficient of valve
K_pc = sqrt(2)* par.A2;                       %Valve constant (m2)
a = (1/rho)*((W/K_pc)^2);                     %Constant parameter
p_star= (rho_l*g*L_r)+p_s+ p_vmin;
fz_star = z_star*cd/sqrt(1-z_star^2*cd^2);
delta_p_star = a/fz_star^2;
p_fo = p_star-delta_p_star;                   %inlet pressure at fully open
position of the valve(at z=1) (pa)


z_t = 0.2:0.001:1;                %Different valve openings
n = length(z_t);
Pin = zeros(1,n);                 %Inlet Pressure
K_z_t = zeros(1,n);               %Static gain of the system (pa)



for i = 1:n
    fz = z_t(i)*cd/sqrt(1-z_t(i)^2*cd^2);
    Pin(i) =(a/fz^2 + p_fo)/1000;
    K_z_t(i) = (-2*a/z_t(i)^3*cd^2)/1000;
end

figure(1)
clf
subplot(2,1,1);
plot(z_t,Pin,'k','LineWidth',2);
xlabel('Z');
ylabel('Inlet Pressure [kPa]');
hold on
grid on

subplot(2,1,2);
plot(z_t,K_z_t,'k','LineWidth',2);
xlabel('Z');
ylabel('K(z)');
hold on
grid on
```

C.6 Online tuning based on simple static model and a closed loop step test

```
clc
clear all

load z40_141_142
z0 = 0.4;             %Initial valve position in step-test
Kc0= -0.15;           %gain used for the step test
dy_s = 1;
t_init = 300;
dt = 0.1;

t = z40_141_142(:,1);
y = z40_141_142(:,2);
r = z40_141_142(:,3);
u = z40_141_142(:,4);


i_init = find(t==t_init);
y_plant = y(i_init:end);
t_plant = t(i_init:end);
u_plant = u(i_init:end);
y_init = y(i_init-10);
u_init = u(i_init-10);
yp = max(y_plant);     %Step change is positive
dy_p = abs(yp - y_init);
i_yp = find(y_plant==yp);
t_p1 = mean(t_plant(i_yp));
yu = min(y_plant(i_yp:10*i_yp));
dy_u = abs(yu - y_init);
i_yu = find(y_plant==yu);
t_u = mean(t_plant(i_yu));
y_inf = y_plant(end);
dy_inf = abs(y_inf - y_init);
tp = t_p1 - t_init;

deltat = t_u - t_p1;

%%
%%%%%%MODEL%%%%%%

z = 0.3                            %The operating point
g = 9.81;                          %Gravity (m/s2)
Wg_in=0.0024;                      %Inlet mass flow rate of gas (Kg/sec)
Wl_in=0.39298;                     %Inlet mass flow rate of liquid
(Kg/sec)
W = Wg_in+Wl_in;                   %Inlet mass flow rate (Kg/sec)
R = 8314;                          %Gas constant (J/(K.Kmol))
M_g = 29;                          %Molecular weight of Gas (kg/kmol)

p_s=101325;                        %Seperator pressure (pa)
p_vmin = 0;                        %minimum Pressure drop over the valve
(Pa)
T=15+273.15;                       %Riser temprature (K)
par.r2 = 0.025;                    %Radius of riser (m)
par.A2 = pi*par.r2^2;              %Cross section area of riser (m2)
```

```
rho_g= (p_s+p_vmin)*M_g/(R*T)       %Average gas density at the
outlet(Kg/m3)
rho_l=1000;                         %Liquid density (Kg/m3)

alpha_g = Wg_in/(Wg_in+Wl_in)       %Average gas mass fraction
alpha_l=(1-alpha_g)*rho_g/((1-alpha_g)*rho_g+alpha_g*rho_l)        %liquid
volume fraction

rho = alpha_l*rho_l+(1-alpha_l)*rho_g      %Average mixture density
L_r = 5.30;                                %Length of riser
z_star= 0.26;                              %Bifurcation point
cd = 0.31;                                 %Discharge coefficient of valve
K_pc = sqrt(2)* par.A2                     %Valve constant (m2)
a = (1/rho)*((W/K_pc)^2)                   %Constant parameter
p_star= (rho_l*g*L_r)+p_s+ p_vmin
fz_star = z_star*cd/sqrt(1-z_star^2*cd^2)
delta_p_star = a/fz_star^2
p_fo = p_star-delta_p_star                 %inlet pressure at fully open
position of the valve(at z=1) (pa)

fz = z*cd/sqrt(1-z^2*cd^2)
K_z = -2*a/z^3*cd^2                        %static gain of the system (pa)
K_z0 = -2*a/z0^3*cd^2

%%
%%%PI tuning parameters%%

T_osc= 140  %period of slugging oscillations in sec in the open loop system
Betha=-log((dy_inf-dy_u)/(dy_p-dy_inf))/(2*deltat)+ Kc0*K_z0 *((dy_p-
dy_inf)/dy_inf)^2/(4*tp)
Kc=Betha*T_osc/K_z*sqrt(z/z_star)
tauI_z=3*T_osc*(z/z_star)
disp('FEED TO OLGA AND FIND THE MAXIMUM STABILITY!')

%%
%%%%%%%%%%Plot%%%%%%%

%MODEL
z_t = 0.2:0.001:1;
n = length(z_t);
Pin = zeros(1,n);
K_z_t = zeros(1,n);


for i = 1:n
    fz = z_t(i)*cd/sqrt(1-z_t(i)^2*cd^2);
    Pin(i) =(a/fz^2 + p_fo)/1000;
    K_z_t(i) = (-2*a/z_t(i)^3*cd^2)/1000;
end


figure(1)
clf
plot(z_t,Pin,'r','LineWidth',2.5);
xlabel('Z');
ylabel('Inlet Pressure [kPa]');
hold on
grid on
%%
```

```
%%%%%%OLGA MODEL

load Openloop
z_olga = Openloop (:,1);
P_max = Openloop (:,2);
P_min = Openloop (:,3);
P_ss = Openloop (:,4);


figure (1)
plot (z_olga,P_ss,'b','LineWidth',2.5);
hold on
legend('Simple static MATLAB model','OLGA case',2)

plot (z_olga,P_max,'b','LineWidth',2.5);
hold on
plot (z_olga,P_min,'b','LineWidth',2.5);
grid on
```

C.7 Simple static model fitted to the OLGA simulated model

```
clc
clear all

%%
%%%%%%%STEP TEST INFORMATION%%%%%%%

% load z30_148_150
% z0 = 0.3;               %Initial valve position in step-test
% Kc0= -0.1;              %gain used for the step test
% dy_s = 2;
% t_init = 200;
% dt = 0.1;
%
% t = z30_148_150(:,1);
% y = z30_148_150(:,2);
% r = z30_148_150(:,3);
% u = z30_148_150(:,4);

load z40_141_142
z0 = 0.4;               %Initial valve position in step-test
Kc0= -0.15;             %gain used for the step test
dy_s = 1;
t_init = 300;
dt = 0.1;

t = z40_141_142(:,1);
y = z40_141_142(:,2);
r = z40_141_142(:,3);
u = z40_141_142(:,4);


i_init = find(t==t_init);
y_plant = y(i_init:end);
t_plant = t(i_init:end);
u_plant = u(i_init:end);
```

```
y_init = y(i_init-10);
u_init = u(i_init-10);
yp = max(y_plant);     %Step change is positive
dy_p = abs(yp - y_init);
i_yp = find(y_plant==yp);
t_p1 = mean(t_plant(i_yp));
yu = min(y_plant(i_yp:10*i_yp));
dy_u = abs(yu - y_init);
i_yu = find(y_plant==yu);
t_u = mean(t_plant(i_yu));
y_inf = y_plant(end);
dy_inf = abs(y_inf - y_init);
tp = t_p1 - t_init;

deltat = t_u - t_p1;

%%
%%%%%%MODEL%%%%%%

z = 0.3                             %The operating point
g = 9.81;                           %Gravity (m/s2)
Wg_in=0.0024;                       %Inlet mass flow rate of gas (Kg/sec)
Wl_in=0.39298;                      %Inlet mass flow rate of liquid
(Kg/sec)
W = Wg_in+Wl_in;                    %Inlet mass flow rate (Kg/sec)
R = 8314;                           %Gas constant (J/(K.Kmol))
M_g = 29;                           %Molecular weight of Gas (kg/kmol)

p_s=101325;                         %Seperator pressure (pa)
p_vmin = 0;                         %minimum Pressure drop over the valve
(Pa)
T=15+273.15;                        %Riser temprature (K)
par.r2 = 0.025;                     %Radius of riser (m)
par.A2 = pi*par.r2^2;               %Cross section area of riser (m2)
rho_g= (p_s+p_vmin)*M_g/(R*T)       %Average gas density at the
outlet(Kg/m3)
rho_l=1000;                         %Liquid density (Kg/m3)

alpha_g = Wg_in/(Wg_in+Wl_in)       %Average gas mass fraction
alpha_l=(1-alpha_g)*rho_g/((1-alpha_g)*rho_g+alpha_g*rho_l)        %liquid
volume fraction

rho = alpha_l*rho_l+(1-alpha_l)*rho_g        %Average mixture density
L_r = 5.30                                   %Length of riser
z_star= 0.26;                                %Bifurcation point
cd = 0.31;                                   %Discharge coefficient of
valve
K_pc = sqrt(2)* par.A2                       %Valve constant (m2)
a = (1/rho)*((W/K_pc)^2)                     %Constant parameter
p_star= (rho_l*g*L_r)+p_s+ p_vmin
fz_star = z_star*cd/sqrt(1-z_star^2*cd^2)
delta_p_star = a/fz_star^2
p_fo = p_star-delta_p_star                   %inlet pressure at fully open
position of the valve(at z=1) (pa)

fz = z*cd/sqrt(1-z^2*cd^2)
K_z = -2*a/z^3*cd^2                        %static gain of the system (pa)
K_z0 = -2*a/z0^3*cd^2

%%
```

```
%%%PI tuning parameters%%

T_osc= 140  %period of slugging oscillations in sec in the open loop system
Betha=-log((dy_inf-dy_u)/(dy_p-dy_inf))/(2*deltat)+ Kc0*K_z0 *((dy_p-
dy_inf)/dy_inf)^2/(4*tp)
Kc=Betha*T_osc/K_z*sqrt(z/z_star)
tauI_z=3*T_osc*(z/z_star)
disp('FEED TO OLGA AND FIND THE MAXIMUM STABILITY!')

%%
%%%%%%%%%%Plot%%%%%%%

%MODEL
z_t = 0.2:0.001:1;
n = length(z_t);
Pin = zeros(1,n);
K_z_t = zeros(1,n);


for i = 1:n
    fz = z_t(i)*cd/sqrt(1-z_t(i)^2*cd^2);
    Pin(i) =(a/fz^2 + p_fo)/1000;
    K_z_t(i) = (-2*a/z_t(i)^3*cd^2)/1000;
end

figure(1)
clf
plot(z_t,Pin,'r','LineWidth',2.5);
xlabel('Z');
ylabel('Inlet Pressure [kPa]');
hold on
grid on


%%
%%%%%%Openloop and Steady-state
load Openloop
z_olga = Openloop (:,1);
P_max = Openloop (:,2);
P_min = Openloop (:,3);
P_ss = Openloop (:,4);


figure (1)
plot (z_olga,P_ss,'b','LineWidth',2.5);
hold on
legend('Simple static MATLAB model','OLGA case',2)

plot (z_olga,P_max,'b','LineWidth',2.5);
hold on
plot (z_olga,P_min,'b','LineWidth',2.5);
grid on
```