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Modeling and Simulation of Anti-slug Control in Hydro Experimental Multiphase Flow Loop

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Problem Description

Handling unstable flow is important for most offshore oil and gas production units. Both production wells and transportation pipelines exhibit unstable flow leading to disturbances for the downstream process. Unstable flow leads to poor process efficiency and utilization, and economic losses.

In Porsgrunn, a pilot scale experimental rig has been built to simulate a real offshore production system. Three phases (model oil, water and air) are fed into a well section, followed by an approximately 100 m long 2 degrees declining transport line. The riser is 10.5m of height and ends up in a three phase separator. A complete description of the pilot plant will be provided separately.

For certain production rates the multiphase flow is unstable (slug flow). In real world systems, unstable multiphase flow leads to large disturbances to the downstream process yielding production losses and unwanted process shutdowns.

Stabilizing slug flow using automatic control is the preferred method for slug mitigation. Linear stabilizing controllers have been shown to have poor robustness. To improve robustness, simple nonlinear controllers should be tested by simulations on a model of the multiphase flow loop in Porsgrunn.

The following points should be addressed:

1. Literature review: A brief literature review on anti-slug control should be included.

2. Modelling: Develop a model of the pilot plant using the state of the art multiphase simulator OLGA 5. Data for the rig is provided by Hydro. Using the available tuning parameters in OLGA 5, the model should be fit to the observed experimental data provided by Hydro.

3. Stabilizing control design: Traditional anti-slug control using SISO linear controllers has been shown to have poor robustness and is difficult to tune to achieve good performance. This is believed to be mainly caused by the fact that the process gain changes as the operational conditions changes.

To improve robustness of the control system, an approach using gain scheduling control should be investigated. Different control strategies should be analyzed and compared to traditional SISO linear controllers. It is especially interesting to address how robust the gain scheduling approach is to rate changes in inflow, in addition to noisy measurements.

Assignment given: 18. January 2007
Supervisor: Ole Morten Aamo, ITK

Abstract

Handling slug flow is a matter of necessity in the oil and gas industry. Large slugs cause problems at the production platform by overfilling the first stage separator leading to poor separation and pressure increase, which again could lead to more flaring or even complete process shut-down.

Using the riser base pressure as the controlled variable has been shown in several academic papers to have a stabilizing effect on the slug flow. In this thesis it has been concentrated on designing robust anti-slug controllers using this variable as the controlled variable.

Experience from Hydro's multiphase rig at their Research Centre in Porsgrunn shows that re-tuning of anti-slug controllers during experiments is necessary to be able to keep the flow stabilized. This is caused by the change in process gain, which is large for operation around small valve openings and low for operation around large valve openings.

An OLGA model was developed for this multiphase rig. This model was shown to predict the amplitude of the slugs and their frequencies fairly well, however, the transition point from stable flow to slug flow differed to a larger extent.

Several gain scheduling controllers were designed and tested on this model. The best gain scheduling approach which was found was based on a gain scheduling regime controlling the riser base pressure in a secondary loop without integral action, with a slow primary loop controlling the valve opening (with integral action). The scheduling variable was a low-pass filtered measurement of the valve opening. This approach was also robust with respect to noise and lowered inlet rates.

Preface

This report is the final work for the M.Sc degree in Engineering Cybernetics at the Norwegian University of Science and Technology. This Master's thesis has been written in co-operation with Hydro's Research Centre in Porsgrunn, and is motivated by their interest in handling slug flow.

A special thanks to my supervisor at Hydro, Vidar Alstad, who gave me the opportunity to work with this subject and for his valuable help, especially for taking his time commenting the final report. I am also very grateful for all the help I have gotten from Professor Ole Morten Aamo, my supervisor at the Department of Engineering Cybernetics. A final thanks goes to Professor Sigurd Skogestad at the Department of Chemical Engineering, who first introduced me to the subject of anti-slug control.

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Einar Hauge

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Chapter 1

Introduction

Transporting multiphase flow is a complicated task for the oil and gas industry. In an attempt to utilize smaller and less valuable fields, these fields are connected as tie-ins to production platforms. One major problem with transporting oil, gas and water in the same pipeline over long distances is the possibility for introducing a flow regime called slug flow. This is a flow regime where the liquid (oil/water) flows separately from the gas through the pipeline. The slugs (existing of oil/water) can be initiated by several physical phenomena. The focus taken here is on riser-induced slug flow, caused by liquid accumulating at the bottom of the riser.

These slugs cause severe problems at the downstream facilities. The slugs can be so large that they can even fill up the separator completely. This will evidently lead to poor separation, in addition to increasing the separator pressure causing more flaring or even complete process shut-down.

Minimizing the consequences these slugs imply is therefore an important task in the oil industry. Historically these slugs have been coped with in two ways. Change of design or change of operation. A more recently adopted way of avoiding these slugs is to use control methods.

Using automated control methods is a superior solution to design changes and changes of operational conditions. As can be shown (and which will be further discussed and investigated), using control methods can suppress the slug flow almost without installing new equipment and at the same time operate at boundary conditions which would normally lead to slug flow. The manipulated variable is the choke-valve topside, while several physical measurements can be used as controlled variables.

1.1 Limitation of scope

Motivated by the importance of ensuring stable flow in the pipelines in the oil-and gas industry, Norsk Hydro ASA has build a multiphase rig at the Research Centre in Porsgrunn so that they are able to investigate the behavior of multiphase flow. Promising control strategies found by simulations can then be tested experimentally in the multiphase rig before actual implementation offshore.

One of the main tasks in this thesis was therefore to contribute to this research by developing a model for this rig, using the multiphase simulator OLGA 5. Verification and tuning of this model was done by comparing OLGA results to experimental data obtained from the rig.

A problem which has been experienced when testing various control strategies at the rig, is that these strategies are vulnerable to changes in operational conditions. In practice this means that the operator of the rig must re-tune the controllers during experiments in order to be able to keep the flow stabilized. This need for re-tuning is mainly caused by the fact that the process gain changes as the operational conditions changes. The consequence is that good control parameters at one operational point, does not work satisfactory at another operational point. Different nonlinear control strategies based on simple gain scheduling theory will therefore be investigated and tested on the OLGA model developed in this thesis. These controllers will be compared to linear SISO (Single Input, Single Output) controllers to show possible improvements with respect to robustness.

1.2   The structure of the report

Chapter 2 will start by describing the riser slug phenomena, before moving on to the brief literature review. Some comments which put this thesis in a perspective to the anti-slug control research will be given as well.

Chapter 3 will describe what is done in the Master's thesis in detail, and is divided in three main parts. OLGA modeling, quantification of the gain in the process and gain scheduling design.

Chapter 4 will present the results, and will be divided in the same three parts as **Chapter 3**.

Concluding remarks will be given in **Chapter 5**, in addition to some comments on further work.

Chapter 2

Background

Historically slug flow has been coped with by either change of design or operational changes. Inspired by Havre and Dalsmo (2001) it is distinguished between the following three methods for handling slugs:

1. Design changes

2. Operational changes and procedures

3. Control methods [1]

   - Feed forward control
   - Active feedback control

An example of design change is to increase the buffer capacity of the separators at the riser outlet. Often the slugging problematic gets tougher at the tail-end of production at an oil and gas field, so changing the design in the operation stage of a platform is costly and unwanted.

By changing the operational conditions it is meant increasing the down hole/pipeline pressure by decreasing the opening of the choke-valve topside. Decreasing this opening means lower production rate and evidently lesser income.

The focus here will be on using control methods to handle slug flow, by using active feedback. The advantage with using feedback control is that the accumulation of slugs can be prevented with little or no new equipment, and at the same time operate at boundary conditions which would normally cause slug flow. Most of the research relating to the control method approach has been developed during the 90's and up to now (2007), although it was shown as early as in 1979 by Schmidt et al. that this could be a viable approach.

Much research has been done since then, and it has been done several attempts on making a comprehensive literature review on the subject. In Havre and Dalsmo (2001) a very well written section on previous work on using control theory to cope with slug flow can be found. However, this paper was written in 2001 and naturally does not contain any references to work done for the last six

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[1] This is slightly different from the list found in Havre and Dalsmo (2001) where slug choking is also considered as a control method. However, it is regarded as more convenient to consider this as a method for changing the operational conditions rather than an actual control method.

years. In the doctoral thesis Storkaas (2005) a more recently written literature study can be found, although very brief. In Venås (2006) very much of the same literature is referred.

Havre and Dalsmo (2001) and Storkaas (2005) in addition to Venås (2006) will here be used as basis for this litterature review. However, before it is moved on to the literature review, the properties of slug flow will be explained in some more details.

2.1   More on the riser slug phenomena

Riser induced slug flow may occur in pipelines with multiphase flow. These multiphase pipelines can be found offshore between a production platform and subsea wells (or to a wellhead platform). A schematic view of a typical pipeline riser system is shown in figure 2.1.
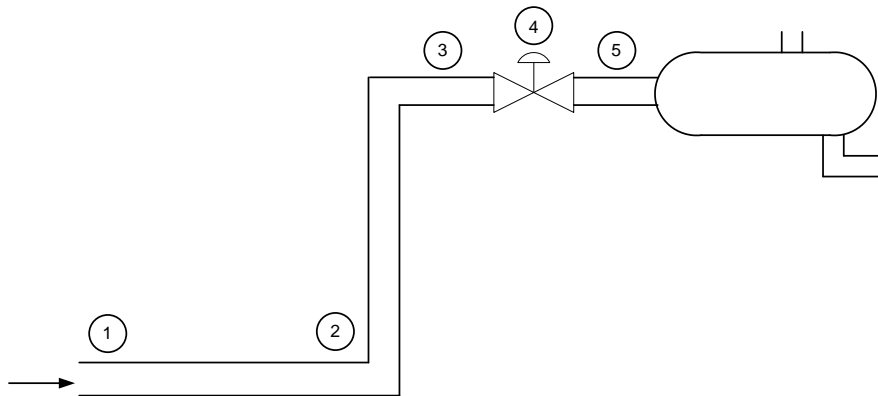
Figure 2.1: 1:Inlet   2:Riser base   3:Topside   4:Choke-valve   5:Separator

Slug flow is an unstable oscillating flow regime which often has a period in a matter of hours[2]. The slug is first initiated by liquid accumulating at the bottom of the riser blocking for the gas to pass through. The liquid starts to fill up the riser, while the gas pressure in the bottom pipeline starts to build up. Often the liquid column fills up the hole riser before the gas pressure overcomes the hydrostatic pressure in the riser. When this happens, the slug is pushed out the riser with an accelerating speed. While the liquid is being pushed out the riser, the pressure starts to fall in the riser causing the gas to expand. This causes an even bigger force pushing out the slug. After the slug is blown out, the pressure in the pipeline has fallen and the liquid starts accumulating at the bottom of the riser again. (Storkaas, 2005) and (Havre and Dalsmo, 2001).

Riser induced slug flow is first of all introduced by the geometry of the pipeline, which allows for accumulation of liquid (oil/water) at the riser bottom. In addition to this, slug flow arises often at low liquid and gas rates. This is not so surprising, as high liquid and gas rates would mean more kinetic energy in the system, which would be enough for the liquid to be forced up the riser without falling back causing the liquid accumulation. This explains also why slugging often is a larger problem at tail-end production of oil and gas fields, when the reservoir pressure has decreased

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[2]Note that the slug frequency of the multiphase pilot rig at Hydro's Research Centre, which is of much smaller scale than a pipeline-riser system offshore, has a slug period in matters of minutes.

causing a lower velocity for the multiphase flow.

In addition to this, the choke-valve opening also comes in to play whether slug flow is introduced or not. A lower valve opening increases the pipeline pressure, and also increases the differential pressure over the riser. This differential pressure must be larger than the hydrostatic pressure at all time to avoid slug flow to arise. This is a viable option to eliminate slug flow, however, not the best choice from an economical point of view, since a smaller valve opening necessarily means lower production[3].

The gas to oil ratio also has a significant impact on the existence of slug flow. The gas has a much lower density than the oil phase. If the gas to oil ratio is large, this imply that the density of the total flow is low compared to if the gas to oil rate is less. A larger gas to oil ratio imply that the system would be more resistant against slug flow.

2.2 Previous work

It seems that the research community often refer to Schmidt et al. (1979) as the first contributory research on using control theory to suppress the unwanted slug flow regime. This was an experimental approach using a pressure measurement upstream the riser in addition to a flow measurement of the fluid through the riser. Based on an algorithm the choke valve was adjusted automatically to cope with the slug flow.

In Taitel (1986) a more theoretical approach is adopted. From control theory it is a well known fact that to operate at an unstable equilibrium point feedback control is required because of disturbances acting on the system. (Disturbances will move the system from the equilibrium point, and since the equilibrium is unstable the system will not naturally fall back by itself to the equilibrium point). With this as a basis Taitel (1986) defined a theoretical control law.

In Hedne and Linga (1990) an experimental approach similar to Schmidt et al. (1979) was made, however, using a conventional PI-controller with only a pressure measurement upstream the riser.

One of the first industrial implementations of a slug controller known to the public was an implementation in connection with the Dunbar pipeline (Courbot, 1996). This is a 16" multiphase pipeline connecting the Dunbar field with the Alwyn platform at the British side of the North Sea. The slug flow was suppressed using the choke-valve as the manipulated variable to control the riser base pressure. In Storkaas (2005), however, it is criticized that the setpoint pressure for the riser base has been set so high. This makes it easier to stabilize the flow, and limits the production through-put. Storkaas argues that this actually only applies the idea of slug-choking to change the operational condition. That means decreasing the choke-valve opening, thus increasing the pipeline pressure to remove the development of slugs.

In Henriot et al. (1999) the same Dunbar pipeline is investigated using a multiphase simulator developed by IFP, TOTAL and ELF. Here the setpoint is set lower, and as Storkaas (2005) points out, this shows that the flow has been stabilized at an operating condition which would normally

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[3]Note that for the multiphase rig the inlet rates of air and water are constant. This means that the production will not increase when the choke-valve opening increases. However, the stabilizing effect of anti-slug control can be verified and in practice this will result in higher production in cases where the inlets are pressure dependent.

lead to slug flow.

In Havre et al. (2000) a successful implementation of a slug controller is documented for the multiphase pipeline between the platforms Hod and Vallhall where BP is operator. Hod is a wellhead platform and connected to the Vahall production platform with a 13-km long pipeline. Experimental results have shown that the formation of slugs has been eliminated by the slug controller. As emphasized by Storkaas, turning on and off the slug controller shows that the system has been stabilized at the unstable operating point. To determine mathematically that the flow regime is stable along the pipeline has not been done, however, profile plots with the flow simulator OLGA indicates that this is the case.

Godhavn et al. (2005) describes an anti-slug control system implemented in connection with the multiphase pipeline between the subsea field Tordis and the production platform Gullfaks C in the North Sea (operated by Statoil). The control system exists of a cascade slug controller with a secondary flow controller setting the reference for the topside choke valve. The primary loop consists of a measurement of the pipeline inlet pressure which then determines the setpoint for the secondary volume controller. This is combined with a MPC level controller for the first stage separator, to handle possible slugs entering the separator despite of the cascade controller.

All the industrial implementations commented here have available measurements of inlet pressure, or some sort of subsea pressure measurement (for example at the riser base). However, subsea measurements are often not available, in particular where older fields are operated. The use of topside measurements are therefore wanted from a practical (and economical) perspective.

In Storkaas (2005) a simplified model based on physical relationships has been developed, in order to try to investigate the problems relating to topside measurements mathematically. Transfer functions from the choke-valve opening to several topside measurements show fundamental control limitations when the system is linearized around a high choke-valve opening. It is shown that for the case defined in Storkaas a pressure measurement topside introduces right-half plane zeros (RHP-zeros). RHP-zeros cause an upper limit of the bandwidth (an upper limit of the control system's response). In contradiction to this, an unstable system needs a certain bandwidth to cope with the instability. Thus these types of systems are certainly hard to control, or even impossible in a practical sense using conventional PID-control. Theory of the limitations regarding RHP-zeros and RHP-poles can be found in Skogestad and Postlethwaite (2005). Using a flow measurement is therefore more promising, since it has been shown in Storkaas (2005) that this does not introduce unstable zero dynamics. Nevertheless, using a flow measurement as the controlled variable is limited by the fact that the steady-state gain is so low. This means that the system will most certainly drift away from the wanted operating point.

Using the volume flow in a secondary loop and the pressure measurement as the primary loop, has been shown to work theoretically. Although with poor setpoint tracking because of the RHP-zero in the pressure measurement. Design of an H_∞ controller using topside measurements only, was shown to improve the performance significantly.

Olsen (2006) investigated further the use of cascade control. He proposed two possible approaches. The first approach was to use cascade control with flow and pressure measurements topside. The second approach was using a flow measurement in a secondary loop, with a measurement of the valve opening in a primary loop. The first approach seemed to work well, and was also tested at

an experimental multiphase loop at the Norwegian University of Science and Technology (NTNU). However, it was hard to stabilize the flow using this method, so extra attention from the operator was required. Use of the valve opening z as the primary loop was not tested experimentally.

These approaches were all using the principle of feedback control. The problem with conventional feedback control is that action is taken after a deviation from the desired value of the controlled variable(s) has/have occurred. Dhulesia et al. (1997) describe an implementation of an acoustic slug-detection system using feed forward control. The idea is that this system will detect the slugs and some of the characteristics of the slugs (slug length, slug velocity and fluid density) approximately 2 minutes before they arrive at the first stage separator. It is then up to the operators to handle the information, and to take appropriate action. This method is realized using accelerometer sensors that measure the vibrations in the pipeline which are realized by the slugs. The vibration frequency generated during the slug period differs from the vibration frequency caused by the gas. Using two measurements along the pipeline, which is at least 20 meters apart, is enough for determine slug length and velocity. Determination of the density needs extra equipment.

2.3 Defining the thesis in its context

Much research has been done regarding anti-slug control. Several different control strategies have been investigated by simulations, but also tested in experiments or even implemented offshore. However, questions regarding what can be done using control theory still need answers and is one of the reasons why Norsk Hydro ASA has build an experimental multiphase loop at their Research Centre in Porsgrunn Norway. Simulations and testing of possible control structures before actual implementation is a matter of necessity.

A model of this experimental multiphase loop would therefore be very useful, and one part of this project is dedicated to model this loop in the multiphase simulator OLGA 5 and verify this according to experimental results. Then control structures can first be investigated using this OLGA model, before tests on the actual experimental rig can be done.

In the literature review it was emphasized that the use of topside measurements gets extra attention from the research community nowadays. The focus taken here, however, is not on using topside measurements but to find more robust control strategies using the riser base pressure as the controlled variable. Using the riser base pressure is well-known for stabilizing the slug flow.

Motivated by experience from the multiphase rig at Hydro, finding a more robust controller to operational changes is desired. Today, controller tuning of simple PID-controllers at the multiphase rig is based on trial and error around a desired operating point. The problem, however, is that due to disturbances the system will drift away from the operational point for which the controller was originally designed. When this happens, the controller might not longer be able to stabilize the flow, and the system will again exhibit the unwanted slug flow regime. Re-tuning of the controller must then be done manually by the operator of the multiphase rig.

Observations from the rig indicated that the need for re-tuning was mainly caused by nonlinear behavior of the process gain. This nonlinear behavior has also been observed in the OLGA model. Since the open-loop gain is the process gain times the controller gain, it is intuitive that the con-

troller gain must be changed as the process gain changes.

A well-known approach for extending the viability of linear controllers to nonlinear processes, is that of gain scheduling. Instead of only designing one linear controller at one operational point, a family of linear controllers are designed for a range of operational points. Often the only parameter which is changed between the controllers is the controller gain.

Neither the nonlinear behavior of the process gain, nor gain scheduling have gotten much attention by researchers. A motivation to the gain scheduling approach will be given in section 2.3.1 by an example which illustrates the importance of changing the controller parameters when the operational conditions changes.

2.3.1 OLGA example related to process gain

In the following a specific example using the OLGA model is given to show how the process gain affects the choice of controller parameters (i.e. the controller gain). Experience from the rig has shown that the process gain decreases as the choke-valve opening increases. The following idea is therefore applied:

- Find the best tuning parameters using simple PI-control for operation around two different valve openings (one small and one large)

- Investigate what happens when the controller parameters are swapped.

The riser base pressure is the controlled variable, and the integral time is 50 seconds for both controllers. See appendix A for how the controllers are implemented in MATLAB. Figure 2.2 and figure 2.3 show simulations for two different setpoints for the riser base pressure (1.54 barg and 0.5657 barg), which correspond respectively to two different valve openings (8% and 40%). The simulations start at an unstable valve opening (30%), and the controllers are turned on after 2.6 minutes, and start at the lowest pressure in the slug cycle which is indicated by a black vertical line in the figures. The simulation lasts for a total of 20 minutes, and are run without any disturbances or noise.
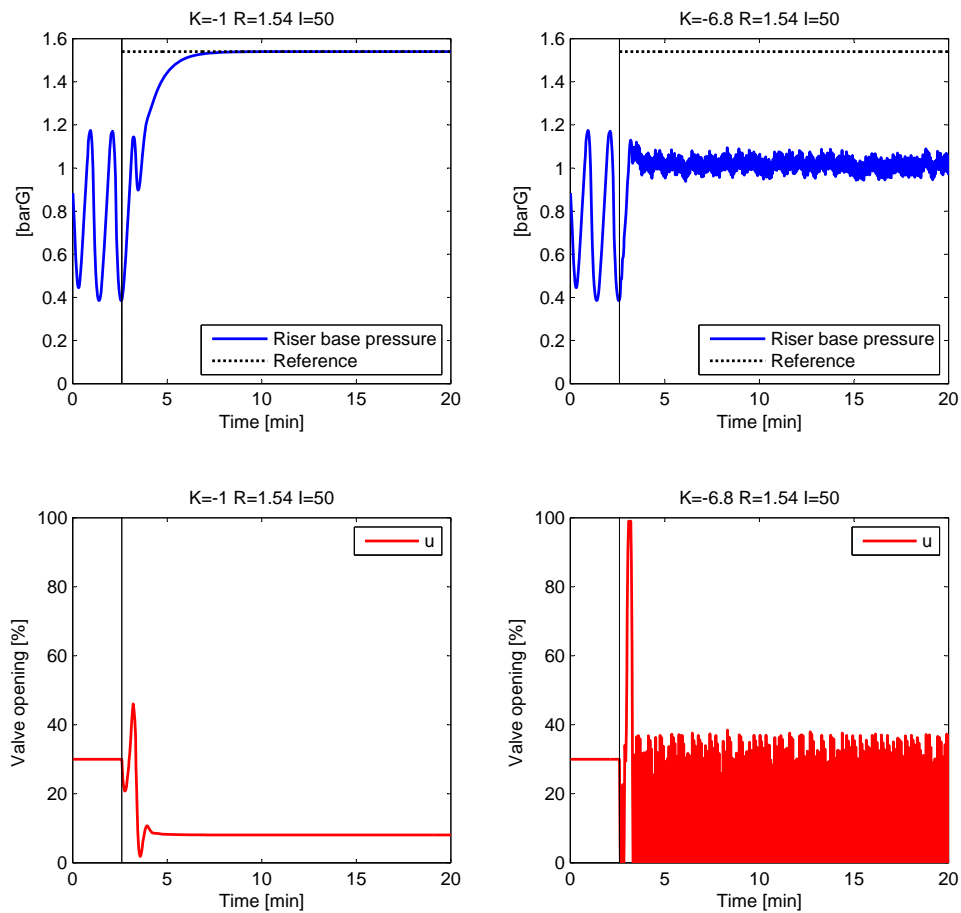
Figure 2.2: Reference = 1.54 barg. The controller is turned on at the minimum pressure in the slug cycle

Figure 2.2 shows the use of two different controllers for a reference for the riser base pressure of 1.54 barg. The reference corresponds to a small valve opening of 8%. At this valve opening the process gain is high. Based on trial and error a low control gain equal to -1 was found out to give good control performance. On the right in figure 2.2 shows a simulation with a higher control gain, equal to -6.8. As can be seen from the plot, the control system is not able to stabilize the system at this reference due to disturbances introduced by the actuator. This is because a control gain equal to -6.8 is too large at this operatinal point because of the high process gain.
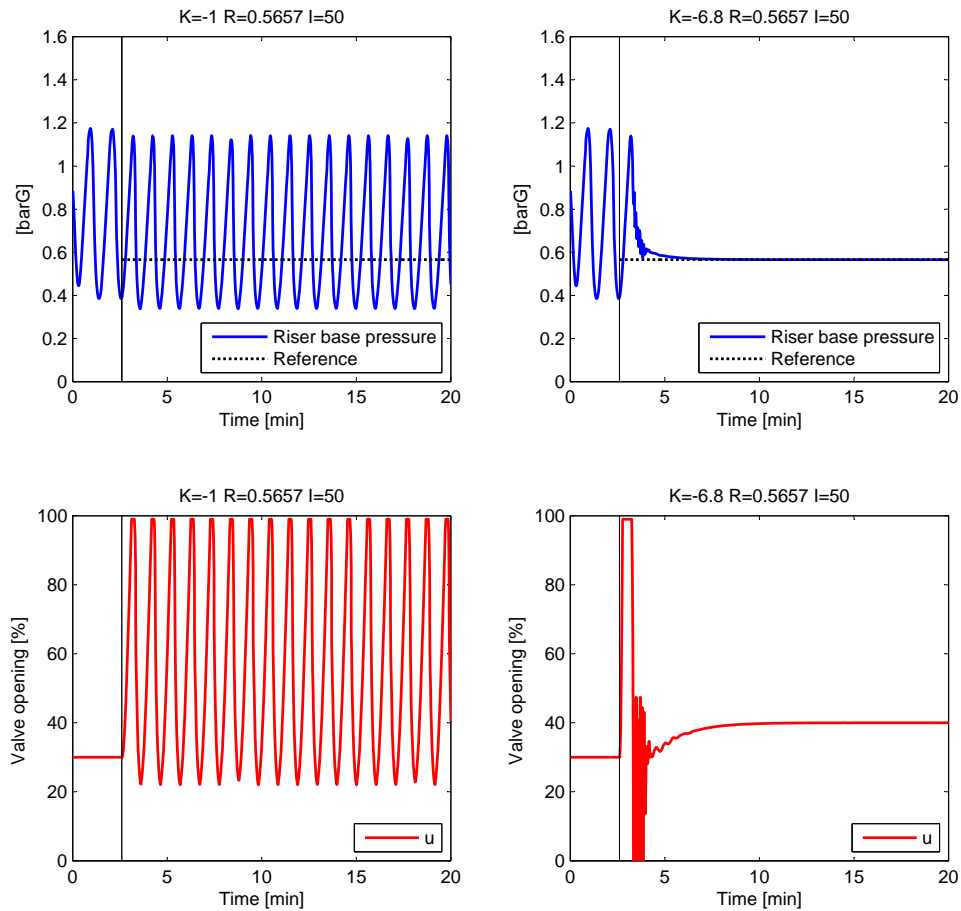
Figure 2.3: Reference = 0.5657 barg. The controller is turned on at the minimum pressure in the slug cycle

Figure 2.3 shows the use of two different controllers for a reference for the riser base pressure of 0.5657 barg. The reference corresponds to a high valve opening of 40%. At this valve opening the process gain is low and the unstability is severe. Based on trial and error a high control gain equal to -6.8 was found out to give good control performance. On the left in figure 2.3 shows a simulation with a lower control gain equal to -1. With this gain the control system is not able to stabilize the system at the reference since the process gain is so low. (need high gain because of low process gain and since the operational point is at an unstable valve opening)

2.4 Background summary

Use of control theory as a way to stabilize slug flow has been shown to work well, and has been documented in several academic papers. It is well known that using the riser base pressure as the controlled variable has a stabilizing effect on the slug flow. On the other hand, using nonlinear control as a way to make the slug controllers more robust to disturbances and measurement noise has not gotten so much attention. By a simple example it has been shown that when the operational conditions changes, the controller gain should also be changed. This validates the idea of designing

a more robust anti-slug controller based on a gain scheduling regime.

Chapter 3

Process description and theory

This chapter describes in detail what is done in the thesis, in addition to explaining necessary theory. The first part gives a brief description of the physical rig at the Research Centre and describes the case which is build in OLGA. The second part describes some methods for finding the process gain. In the third part it is outlined theory and rule of thumbs for the gain scheduling approach. It also discussed different approaches for this specific case considering anti-slug control.

3.1 OLGA modeling

The experimental multiphase rig at the Research Centre in Porsgrunn works as a test facility for unstable flow and separator control. It aims to simulate the part of oil and gas production which relates to the production from the well, through the pipe and up to the first stage separator at the platform. In order to have a background for testing interesting control approaches a model of this experimental rig has been designed using the multiphase simulator OLGA. Building and tuning of this model will be discussed in the following.

3.1.1 Describing the experimental multiphase loop

The multiphase rig is designed for three-phase flow, where the water oil and gas are fed to the system using different inlets and mixed in a well section. The three-phase flow then follow an almost horizontal pipeline section, which is about 100 meters long. At the end of the pipeline there is an approximately 10 meter high riser which ends in a three-phase separator tank with atmosphere pressure. The air flows out into the environment through a hole in the separator tank and is not recycled. The water and oil, however, are sent to a buffer tank before they are pumped back to the well section. Although the multiphase rig is designed to deal with three-phase flow, it is here only considered experiments with two-phase flow with water and air.

The profile plot of the experimental rig used in OLGA is given in figure 3.2. It should be noted that the rig is modeled in the x-y plane only. The third dimension is not taken into consideration in the OLGA model. The actual rig can be seen in figure 3.1. For further description of the rig see Hestetun et al. (2006).

Figure 3.1: The multiphase rig at the Reasearch Centre

3.1.2 Building the OLGA model

Necessary data for developing an OLGA model for the multiphase rig was provided by Hydro. Most of the geometrical data can be found in Hestetun et al. (2006). Hydro also provided a pvt-file for fluid characterizations which is used by OLGA, in addition to open-loop experimental data.

Although the multiphase rig is designed for three-phase flow, it is here considered experiments with two phases, air and water. The multiphase rig is designed for constant inflow of the fluids, and the separator at the riser top has a pressure equal to the atmosphere. Important data are given in table 3.1.

Air	27 kg/h
Water	7 Sm^3/h
Separator pressure	1 atm
Roughness	$4.5 \cdot 10^{-5}$ m
Stroke time choke-valve	5 s
Diameter pipe	7.62 cm
Riser height	10.5 m

Table 3.1: Important data from the multiphase rig

Stroke time is the time used by the valve to go from fully closed to fully opened. The geometrical data are plotted in figure 3.2.
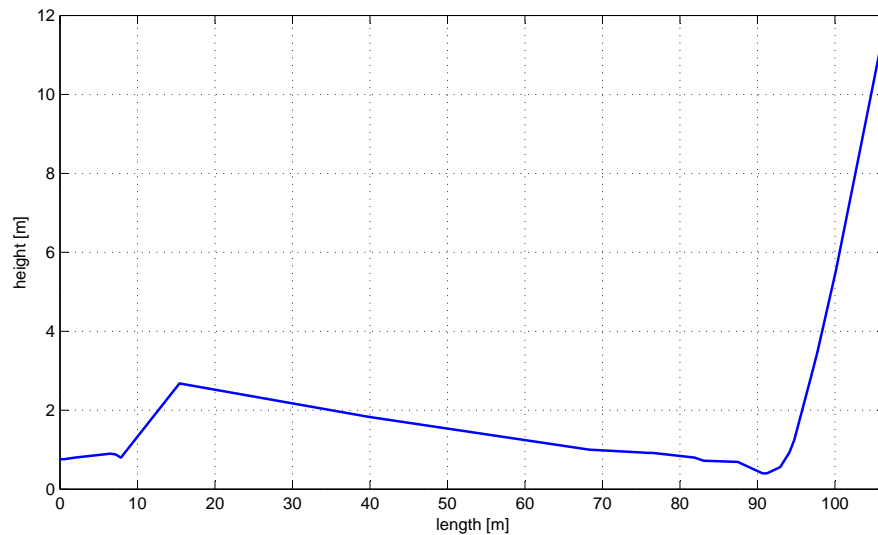
Figure 3.2: Profile plot of the experimental multiphase loop

3.1.3   Fitting the model to experimental data - tuning

OLGA 5 is an advanced simulator for modeling multiphase flow. Nonetheless can a perfect description of the flow in the multiphase rig be expected. Therefore, it will be outlined some tuning approaches which are used on the model. Obviously it is a desire to have a model which describes the multiphase rig in the best possible way. The following properties are of special interest when characterizing slug flow.

- The amplitude of the slugs

- The slug frequency

- Occurrence of slug flow as a function of the choke-valve opening

- Steady-state flow description

All these properties are important when characterizing slug flow. However, a well described steady-state flow description is perhaps the most important, since this model will be used for testing controllers which operate around steady-state. In addition to this the transition point from stable flow to unstable slug flow is of great concern. The model will be compared to experimental data from the rig, and all of these four properties will be addressed explicitly.

To obtain the best possible model, it was used two main tuning procedures:

- Tuning on the choke-valve discharge coefficient

- Tuning on the interfacial friction parameter in the OLGA tuning option

A choke-valve is used for modeling the valve, and has a discharge coefficient (CD-value) which must be decided. The effect of the discharge coefficient is as follows. An increase in the choke-valve

discharge coefficient gives a decrease in the pressure loss over the valve. The pressure downstream the choke-valve is fairly constant since the outlet separator has a pressure equal to 1 atm. The consequence is that when increasing the discharge coefficient the pressure topside (see figure 2.1) will decrease for a corresponding valve opening. This parameter would therefore influence at which valve opening the instability occurs. A lower pressure in the pipeline makes that pipeline more exposed to slug flow than a higher pipeline pressure. This means (in theory) that increasing the discharge coefficient will give a transition between the stable flow regime and slug flow regime at a lower valve opening.

The second tuning approach was to explicitly use the tuning option in OLGA which makes it possible to tune some parameters in the model. The interfacial friction parameter is in this case the most interesting. The friction between the phases directly influence the behavior of the system. A larger friction between air and water will have the effect that more water is pushed up the riser without accumulating at the bottom. A lower friction between air and water will on the other hand increase this accumulation. A lower value of the interfacial friction parameter will move (in theory) the transition point between the stable flow and unstable slug flow to a lower valve opening.

3.2 Process gain

Experiments from the multiphase rig has shown that the process gain decreases as the valve opening increases. Quantifying this gain for the model can be very useful, either as a further motivation for the gain scheduling approach or used more explicitly when designing controllers. First a definition of the steady-state process gain will be given, before some methods for quantifying the process gain are outlined.

3.2.1 Process gain - a nonlinearity of the process

The steady-state process gain is defined as:

$$K_p = \frac{\Delta y_{ss}}{\Delta u_{ss}} \tag{3.1}$$

where Δy_{ss} is the steady-state change in the process variable given a steady-state change in the manipulated variable Δu_{ss}. The process gain says how much a process variable (for example pressure) changes, as the manipulated variable changes. For this riser-pipeline system the manipulated variable is the choke-valve. The process variable can be any variable in the system. However, in this case the process variable will be the riser base pressure, since this variable is used as the controlled variable in the gain scheduling controllers.

Observations from the multiphase rig indicated that the process gain decreases as the valve opening increases. This means that for the same percentual opening of the choke-valve, the process variable will change less when operating at a high valve opening than for a small valve opening.

3.2.2 Methods for quantifying the process gain

Three methods for quantifying the process gain will be outlined. The first method is a practical method which requires simulation from a model or the actual rig. The second method is more theoretical and is based on a steady-state model of the flow in the pipeline-riser system. The last method is just a combination of the two methods already mentioned.

Finding an approximation of the process gain for the OLGA model

Finding an approximation of the process gain as a function of the valve opening can be found by performing simulations on the OLGA model or the actual multiphase rig. For stable systems, an approximation of the process gain can be easily found by using the formula given in (3.1). The method can be done on a step by step basis:

Procedure 3.1 *Finding the process gain for open-loop stable systems*

- *Step up to the valve opening where the process gain is to be found*

- *Find the steady-state process variable which corresponds to this valve opening*

- *Make a new **small** perturbation from this valve opening*

- *Find the new steady-state process variable which corresponds to this valve opening*

- *Use equation* (3.1) *to compute the process gain.*

This method, however, is only applicable to stable systems where the steady-state process variable can be easily found. For the riser base system which is investigated here, the system is unstable for large valve openings. To bypass this problem a simple cascade controller[1] was designed in order to first perform a stabilization before the process gain could be found. Figure 3.3 shows the control configuration for finding the process gain in the unstable area.
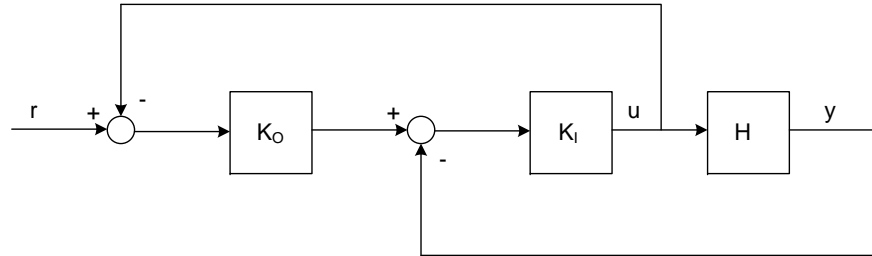
Figure 3.3: Control configuration for finding the process gain for unstable valve openings

Figure 3.3 shows a simple cascade controller where the riser base pressure is controlled in the secondary loop, while the manipulated variable (i.e. choke-valve opening) is controlled in the primary loop. The idea is to find the process gain by performing steps on the reference to the cascade controller, instead of directly changing the manipulated variable. It is here assumed that the secondary loop is controlled by a proportional controller only, while the primary loop also has integral action. Since the choke-valve is controlled in the primary loop with integral action, it can be observed from the block diagram in figure 3.3 that at steady-state the following formula applies:

$$\frac{\Delta y_{ss}}{\Delta r_{ss}} = \frac{\Delta y_{ss}}{\Delta u_{ss}} \tag{3.2}$$

where Δy_{ss} is the steady-state change in the process variable, Δu_{ss} is the steady-state change in the manipulated variable and Δr_{ss} is the steady-state change in the reference to the cascade controller. The consequence of this formula is rather interesting, since it means that the process gain can be found by dividing the change in the process variable by the change of the reference given to the cascade controller, irrespective of the controller parameters (obviously as long as the given controller parameters are actually able to stabilize the system). This can also be shown mathematically:

Starting by finding the closed loop transfer function $\frac{y}{r} = H_{cl}$ using the notation in the block diagram given in figure 3.3.

$$H_{cl}(s) = \frac{y(s)}{r(s)} = \frac{K_I K_O(s) H(s)}{1 + K_I K_O(s) + K_I H(s)} \tag{3.3}$$

Finding the steady-state value after an input change can be found by applying the final value theorem which is defined in equation (3.4) (Balchen et al., 2004).

$$\lim_{t \to \infty} f(t) = \lim_{s \to 0} s f(s) \tag{3.4}$$

Applying procedure 3.1 corresponds to the mathematical relationship:

---

[1] A cascade controller consists of two loops, where the outer loop's manipulated variable is the inner loop's reference. The inner loop is often called the secondary or slave loop, while the outer loop is called the primary or master loop

$$\Delta y_{ss} = y_2 - y_1 = \lim_{s \to 0} sH_{cl}(s)\frac{r_2}{s} - \lim_{s \to 0} sH_{cl}(s)\frac{r_1}{s}$$

$$\Delta y_{ss} = \lim_{s \to 0} H_{cl}\Delta r_{ss}$$

$$\Delta y_{ss} = \lim_{s \to 0} \frac{K_I K_O(s)H(s)}{1 + K_I K_O(s) + K_I H(s)}\Delta r_{ss}$$

Inserting a PI-controller for K_O:

$$\Delta y_{ss} = \lim_{s \to 0} \frac{K_I K_o(1 + \frac{1}{T_i s})H(s)}{1 + K_I K_o(1 + \frac{1}{T_i s}) + K_I H(s)}\Delta r_{ss}$$

$$\Delta y_{ss} = \lim_{s \to 0} \frac{K_I K_o(T_i s + 1)H(s)}{T_i s + K_I K_o(T_i s + 1) + K_I T_i s H(s)}\Delta r_{ss} \qquad (3.5)$$

$$\Delta y_{ss} = \frac{K_I K_o K_p}{K_I K_o}\Delta r_{ss}$$

where K_p is the steady-state process gain of $H(s)$. Solving the last relationship in equation (3.5) with respect to K_p yields:

$$\frac{\Delta y_{ss}}{\Delta r_{ss}} = K_p \qquad (3.6)$$

Equation (3.6) shows that the steady-state process gain can be found by dividing the change in the process variable on the change in the reference of the cascade controller, as was argued for by only investigating the block diagram in figure 3.3.

Procedure 3.1 can now be slightly changed in order to be valid for open-loop unstable systems.

Procedure 3.2 *Finding the process gain for open-loop unstable systems*

- *Find a stabilizing cascade controller on the form in figure 3.3*

- *Step up the reference to the corresponding valve opening where the process gain is to be found*

- *Find the steady-state process variable which corresponds to this valve opening*

- *Make a new **small** perturbation in the reference from this valve opening*

- *Find the new steady-state process variable which corresponds to this valve opening*

- *Use equation (3.6) to compute the process gain.*

Method based on a steady-state model

A method for finding the process gain based on a steady-state model is outlined in this section. This model can be found in Alstad (2007) and will be restated here. While the first method is applicable to various process variables, this method only considers the riser base pressure.
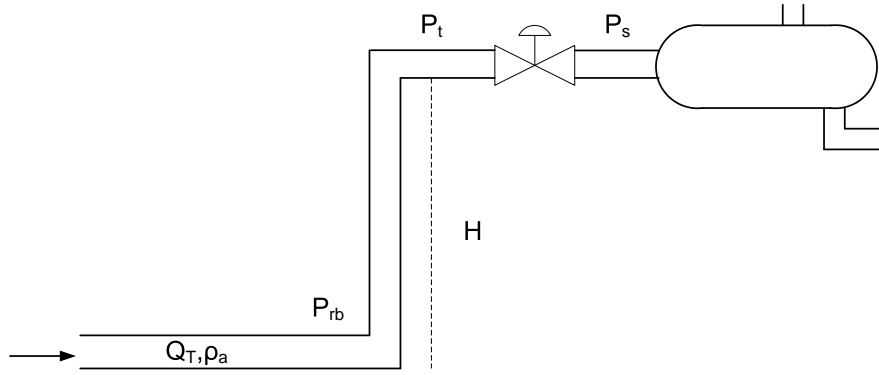
Figure 3.4: Schematic view of the pipeline-riser system

Q_T is the total volume flow, ρ_a is the average density, P_{rb} the riser base pressure, P_t the topside pressure, P_s the separator pressure and H the height of the riser.

Two equations are used for finding an analytical expression for the steady-state process gain. The first is a simple valve model.

$$Q_T = C_v u \sqrt{\frac{P_t - P_s}{\rho_a}} \tag{3.7}$$

And the second is an equation relating the topside pressure and the riser base pressure.

$$P_{rb} - P_t = \rho_a g H \tag{3.8}$$

Note that the equation given in (3.8) only model the pressure drop due to gravity (not pressure loss due to frictional forces). It is assumed that the density upstream the choke-valve is equal to the density at the riser base. Inserting equation (3.8) into (3.7) and solving it with respect to P_{rb} gives

$$P_{rb} = \left(\frac{Q_T}{C_v u}\right)^2 \rho_a + \rho_a g H + P_s \tag{3.9}$$

Differentiate equation (3.9) with respect to u gives the steady-state process gain for the riser base pressure.

$$K_p(u) = -2 \left(\frac{Q_T}{C_v}\right)^2 \rho_a u^{-3} = -2K_1 u^{-3} \tag{3.10}$$

Which shows that the absolute value of the process gain decreases as the valve opening increases.

Least square method - combination of two methods

In (3.10) a simple analytical model of the process gain was found based on a steady-state relationship. However, this require the computation of a constant which exists of the parameters Q_T, C_v and ρ_a. Finding the correct values for this could in some cases be difficult. Assume that process gain data

is available [2], the constant K_1 could be found by fitting the model to this data using a least-square approach. This least-square problem could be formulated as

$$\min_{K_1} \sum_{i=1}^{N} (y_i - K_{p,i})^2$$
$$s.t \qquad\qquad (3.11)$$
$$K_{p,i} = -2K_1 u_i^{-3}, i = 1 \ldots N$$
$$K_1 \geq 0$$

where y_i could be the approximated gain from the OLGA model or gain data from the rig for the i'th valve opening, and $K_{p,i}$ is the corresponding gain from the model. K_1 is the scalar factor which is to be found.

This problem can be rewritten in the form:

$$\min_{K_1} 4Z^T Z K_1^2 + 4Y^T Z K_1 + Y^T Y$$
$$s.t \qquad\qquad (3.12)$$
$$K_1 \geq 0$$

where

$$Z = \begin{bmatrix} u_1^{-3} \\ u_2^{-3} \\ \vdots \\ u_n^{-3} \end{bmatrix}, Y = \begin{bmatrix} y_1 \\ y_2 \\ \vdots \\ y_n \end{bmatrix}$$

Since the minimization problem is optimized with respect to K_1, the last term in the object function in equation 3.12 can be ignored since it is independent on K_1. It can then be identified as a quadratic programming problem, which can be solved easily in MATLAB.

Least square method with modified weights

If some of the data is believed to be more correct than others, a least square method with modified weights could be used. The problem formulation in equation (3.11) can be modified to

$$\min_{K_1} \sum_{i=1}^{N} w_i (y_i - K_{p,i})^2$$
$$s.t \qquad\qquad (3.13)$$
$$K_{p,i} = -2K_1 u_i^{-3}, i = 1 \ldots N$$
$$K_1 \geq 0$$

This problem can be rewritten in the form:

---

[2]Type of data which can be found by procedure 3.2

$$\min_{K_1} 4Z^T W Z K_1^2 + 4Y^T W Z K_1 + Y^T W Y$$

$$s.t \tag{3.14}$$

$$K_1 \geq 0$$

where

$$Z = \begin{bmatrix} u_1^{-3} \\ u_2^{-3} \\ \vdots \\ u_n^{-3} \end{bmatrix}, Y = \begin{bmatrix} y_1 \\ y_2 \\ \vdots \\ y_n \end{bmatrix}$$

and

$$W = \begin{bmatrix} w_1 & 0 & 0 & 0 \\ 0 & w_2 & 0 & 0 \\ 0 & 0 & \ddots & 0 \\ 0 & 0 & 0 & w_n \end{bmatrix}$$

which is also a quadratic programming problem which can be solved in MATLAB.

3.3 Gain scheduling - robustifying the controller

The approach taken will be on designing a simple nonlinear controller, based on a gain scheduling regime. A gain scheduling controller is shortly explained as a family of linear controllers for different operational conditions. As the operational conditions changes, the gain scheduling regime will cope with this by changing the parameters of the linear controller, typical the controller gain.

This section will start with an introduction of existing gain scheduling theory and describing state of the art design of these types of controllers. Next some different gain scheduling strategies relevant to the slug flow case will be discussed.

3.3.1 Gain scheduling theory

Figure 3.5 is a standard block diagram which illustrates a typical gain scheduling controller.
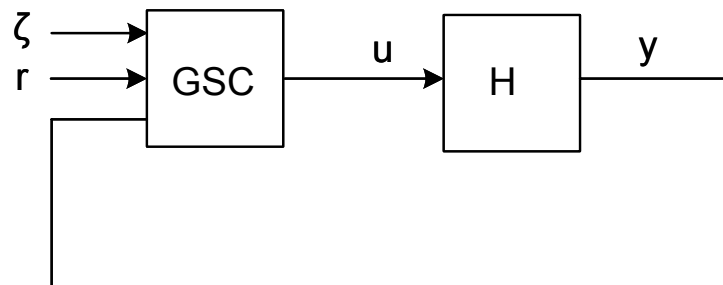
Figure 3.5: Block diagram gain scheduling controller

where u is the manipulated variable, y the controlled variable, r the reference and ζ the scheduling variable. A gain scheduled PI-controller has in continuous time the form

$$u(t) = K(\zeta)(e(t) + \frac{1}{T_i} \int_{t=0}^{t} e(\tau)\mathrm{d}\tau) \tag{3.15}$$

where ζ is the scheduling variable. The controller gain will change as a function of this variable, which for instance could be a reference to the control system, the position of the actuator or a measurement of some physical values in the system. One of the most important problems which must be solved when designing a gain scheduling controller, is to find the scheduling variable ζ which ensures stability and acceptable performance. Shamma and Athans (1992) points out the two following guidelines for the choice of the parameter ζ.

- The scheduling variable, ζ, should capture the plant's nonlinearities

- The scheduling variable, ζ, should vary slowly.

The wish for finding a scheduling variable which captures the nonlinearities is self-explanatory. However, finding such a variable could in some cases be difficult, and extensive physical insight of the system may be required. In this case the gain scheduling approach will be used on anti-slug control, and the nonlinear process gain property can be seen by several variables. This could be the valve opening, reference to the controller and different types of pressure measurements. Different

types of scheduling variables will be investigated and discussed later on.

When using a gain-scheduling regime, the resulting closed loop system is a so called linear parameter-varying (LPV) system. This is a linear system which depends on exogenous parameters with values which are not known a priori. Such a parameter-varying system could be stable when the parameters are frozen, but if the parameters vary with time the system might become unstable. In a system with a gain scheduling controller, frozen controller gains are found for different operational points in such a way as to ensure stability and performance locally around these operational points. However, the resulting gain scheduling controller might go unstable since the controller gain changes in time. Fortunately this problem can be avoided if the scheduling variable varies slowly (Shamma and Athans, 1992).

In Bequette (1998) the following usual gain scheduling design procedure is given.

Procedure 3.3 *Gain scheduling design procedure*

1. *Develop a linear process model for a set (usually a discrete number) of operating conditions.*

2. *Design linear controllers for each operating condition (model)*

3. *Develop a schedule for the controller parameters*

4. *Implement the parameter-scheduled controller on the nonlinear plant*

In this report a nonlinear model of the flow has been developed using the multiphase simulator OLGA. A linear process model on the other hand has not been looked into. However, the non-linearities of the process model has been quantified and investigated by performing simulations on the OLGA-model, in addition to using a simplified steady-state model of the flow. So the control parameters will not be decided explicitly on a linear model, but on trial and error in addition to knowledge of the variations of the process gain.

3.3.2 Selecting the scheduling variable ζ

As mentioned, the scheduling variable should capture the plant's nonlinearities, and it should vary slowly. In section 2.3.1 it was shown that a high control gain was suitable for operational conditions corresponding to a high valve opening, while a low control gain was desirable for operation around small valve openings.

Intuitively, the valve opening points itself out as a scheduling variable which capture nonlinearity of the plant. But since the valve is also the manipulated variable, rapid changes of this variable would be expected, which contradicts the demand for a slow varying scheduling parameter to ensure stability of the closed-loop system. A solution to this problem is to use a filtered version of this valve opening as a scheduling parameter, instead of the actual valve opening.

Another idea is to use the reference to the controller as a scheduling parameter. The problem by using the reference is that it does not cope with dynamic changes in the process gain, only the steady-state process gain. Using the reference as the scheduling variable will therefore not be looked further into.

Another option is to use the topside pressure as a scheduling parameter. A high topside pressure, means high process gain which indicate a demand for low controller gain. A low topside pressure means low process gain which indicate a demand for high controller gain. This also counts for the riser base pressure, which also capture the nonlinearity of the plant in the same manner.

3.3.3 Selecting the scheduling function

Deciding the scheduling variable is important. But this raises a new question on how the actual relationship between the controller gain and the scheduling variable should be. Bequette (1998) defines three following different options for the scheduling of the controller.

- Switch parameters at discrete values of the scheduling variable

- Interpolate parameters as a function of the scheduling variable

- Vary parameters continuously with the scheduling variable

The approach taken her will be varying parameters continuously with the scheduling variable. This function, however, could be implemented with various forms. Two different strategies will be explained. The simplest is a linear scheduling function, which changes the controller gain linearly as a function of the scheduling parameter. The second strategy is to try to keep the open-loop gain at steady-state (controller gain times process gain) constant by using the inverse of the process gain as a scheduling function. The first strategy will be referred to as the Linear scheduling function, the second as the OLGC-scheduling function (Open-Loop Gain Constant).

Linear scheduling function

A simple scheduling function is to let the controller gain $K(\zeta)$ be linearly dependent on the scheduling variable ζ:

$$K(\zeta) = \alpha + \beta\zeta \tag{3.16}$$

The two unknown parameters α and β can be decided based on at least two control gains at two different operational conditions. This could be the valve opening or a pressure measurement. Here it will be used well tuned control gains for two different operational conditions corresponding to a small and a high valve opening.

To give an example. Assume that two control gains have been found for two different valve openings. K_{min} is the gain for the small valve opening u_{min}, while K_{max} is the gain for the high valve opening u_{max}. The parameters α and β can be found by the relation:

$$\beta = \frac{K_{max} - K_{min}}{u_{max} - u_{min}}$$
$$\alpha = K_{min} - u_{min}\beta \tag{3.17}$$

This simple regime is not dependent on the scheduling variable which is used, and a similar approach can be used for scheduling on a pressure measurement. Note that when using this method, one need to make sure that the gain does not change sign. To avoid this an upper or lower bound on the gain might be necessary when implemented.

OLGC-scheduling function

The Linear scheduling function adopts the knowledge of the change in process gain qualitatively. In section 3.2.2 a steady-state model of the process gain change was outlined. The idea is to try to use this model explicitly in a gain scheduling regime. By scheduling the control gain as the inverse of the process gain, one would be able to keep the open-loop gain at steady-state constant(controller gain times process gain). Keeping the open-loop gain constant is also mentioned in Bequette (1998), as a way to keep the stability margins constant. This is obviously not the case when considering the steady-state gain only, but this could still be a viable approach as a way to keep the amplification through the system fairly constant.

The equation for the process was originally stated in equation (3.10), and will be restated here.

$$K_p(u) = -2K_1 u^{-3} \tag{3.18}$$

This process gain was found for the riser base pressure, so this means that this scheduling regime can only be used if the riser base pressure is the controlled variable. It can also only be used if the valve-opening is used as the scheduling variable.

The scheduling function $f(u)$ can be found by multiplying it with the process gain function, and require that it should be equal to a constant open-loop gain:

$$2K_1 u^{-3} f(u) = C \tag{3.19}$$

solving with respect to $f(u)$ gives

$$f(u) = \frac{1}{2K_1} C u^3 \tag{3.20}$$

The problem with using directly the function for the process gain, in such a way as to keep the open-loop gain constant, is that there is only one degree of freedom. The only degree of freedom is the choice of the open-loop gain C.

3.3.4   Gain scheduling in a cascade control loop

In traditional gain scheduling (see figure 3.5) the gain scheduling algorithm takes care of changes in control gain, as well as integral action. The integral action is necessary in order to remove steady-state offset, however, it is destabilizing for the plant (Balchen et al., 2004). An interesting approach is then to remove the integral action in the gain scheduling regime, and instead have integral action in the primary loop in a cascade fashion. The idea is illustrated in figure 3.6, where the primary controller controls the valve opening.
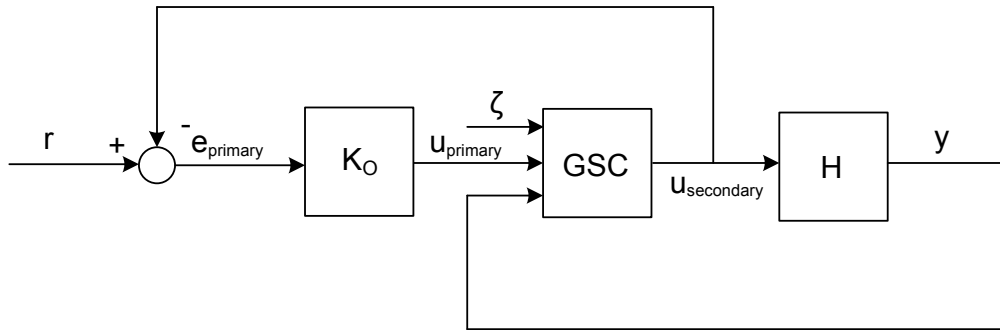
Figure 3.6: Block diagram with gain scheduling in a cascade control regime

The gain scheduling controller in the secondary loop simplifies to

$$u_{\text{secondary}}(t) = K(\zeta)e_{\text{secondary}}(t) \tag{3.21}$$

where $e_{\text{secondary}} = (u_{\text{primary}} - y)$. The primary controller is an ordinary PI-controller on the form

$$u_{\text{primary}}(t) = K_o(e_{\text{primary}}(t) + \frac{1}{T_i}\int_{t=0}^{t} e_{\text{primary}}(\tau)\mathrm{d}\tau) \tag{3.22}$$

where $e_{\text{primary}} = (r - u_{\text{secondary}})$.

This approach is also interesting because it is often more convenient for an operator to use a setpoint for the valve opening, rather than a pressure setpoint. Removing the destabilizing integral action from the secondary stabilizing loop is especially interesting since it is dealt with an open-loop unstable process.

3.3.5 Filtering the scheduling variable

To avoid rapid changes in the gain, which could cause instability, the scheduling variable will be filtered in the gain scheduling algorithm. It is used a simple first order filter, which in continuous time have the form

$$\zeta_f = \frac{1}{1 + Ts}\zeta \tag{3.23}$$

Since it is operating in discrete time, equation (3.23) is discretized using the backward euler method

$$\dot{\zeta_f} = \frac{\zeta_{f,n} - \zeta_{f,n-1}}{\Delta t} \tag{3.24}$$

Combining equation (3.23) and (3.24) gives a discrete filter on the form

$$\zeta_{f,n} = \frac{\Delta t}{\Delta t + T}\zeta_n + \frac{1}{\Delta t + T}\zeta_{f,n-1} \tag{3.25}$$

which can be simplified to

$$\zeta_{f,n} = \psi\zeta_n + (1 - \psi)\zeta_{f,n-1} \tag{3.26}$$

where

$$\psi = \frac{\Delta t}{\Delta t + T}$$

The sampling time, Δt, will always be equal to one second, but the time constant T could be different for different scheduling variables.

3.3.6 Rate changes

The system is originally investigated with constant rates which are given in table 3.1. However, it is of special interest to see how robust the gain scheduling controllers are to changes in the inflow. The approach taken here is to reduce the rates by 10% and 20% and see how robust the controllers are to these changes. As already explained in section 2.1, lower inlet rates means lower kinetic energy in the system, and the pipeline-riser system would be more exposed to slug flow. On the other side, if the inlet rates are increasend, the system will at a certain rate not exhibit slug flow since the kinetic energy in the system will be enough for avoiding liquid accumulation at the riser base. This will only be investigated for the most promising controllers.

3.4 Summary of process description and theory

Geometrical data and fluid characterizations were provided by Hydro. The model could then be built, and two main tuning procedures were outlined. Tuning on the choke-valve discharge coefficient and the interfacial friction parameter.

Since the steady-state process gain has such an impact on controller tuning, several methods for examining the behavior of process gain quantitatively have been outlined. The first method was based on performing simulations on the model or the rig only. The second approach used a simple mathematical steady-state relationship to find a model for the process gain for the riser base pressure. Finding the variables which define the constant in this model may not be easy, and it might be easier to find this constant by using process gain data found by simulations and fit this data to the model with a least-square method.

The gain scheduling chapter discussed various scheduling variables and scheduling functions which can be used in an anti-slug gain scheduling controller. In table 3.2 the various approaches that will be investigated in this thesis are shown.

	Topside pressure	Riser base pressure	Valve opening
Linear scheduling	X	X	X
Cascade scheduling			X
OLGC scheduling			X

Table 3.2: Matrix showing combinations of scheduling functions and scheduling variables considered in this thesis.

Chapter 4

Results

Results from the methods and approaches which were explained and outlined in chapter 3 will be shown. Verification and tuning of the OLGA model will be performed by several plots which show how well the flow in the pipeline-riser system is described. The use of all three process gain methods are demonstrated. Finally simulation results for several gain scheduling algorithms are shown, and how their performance are when exposed to noise and rate changes. (see table 3.2)

4.1 OLGA model

In this section the results of using the tuning approaches which were explained in section 3.1.3 will be shown. The most important properties of the model was to predict the steady-state behavior, in addition to the transition from stable to unstable slug flow in the best possible way.

The tuning was done on a step-by-step basis. Figure 4.1, 4.2 and 4.3 show bifurcation plots [1] for different tuning approaches. The bifurcation plots for the OLGA model was found in two different ways. Either by starting from stable or unstable valve openings. The first method is referred to as 'step-up' in the label of the bifurcation plots, while the latter is referred to as 'step-down'.

---

[1] Bifurcation means in nonlinear systems theory a fundamental change in the properties of the system. In this specific pipeline-riser system it means the transition from a stable equilibrium point to a nonlinear limit cycle (Khalil, 2000)
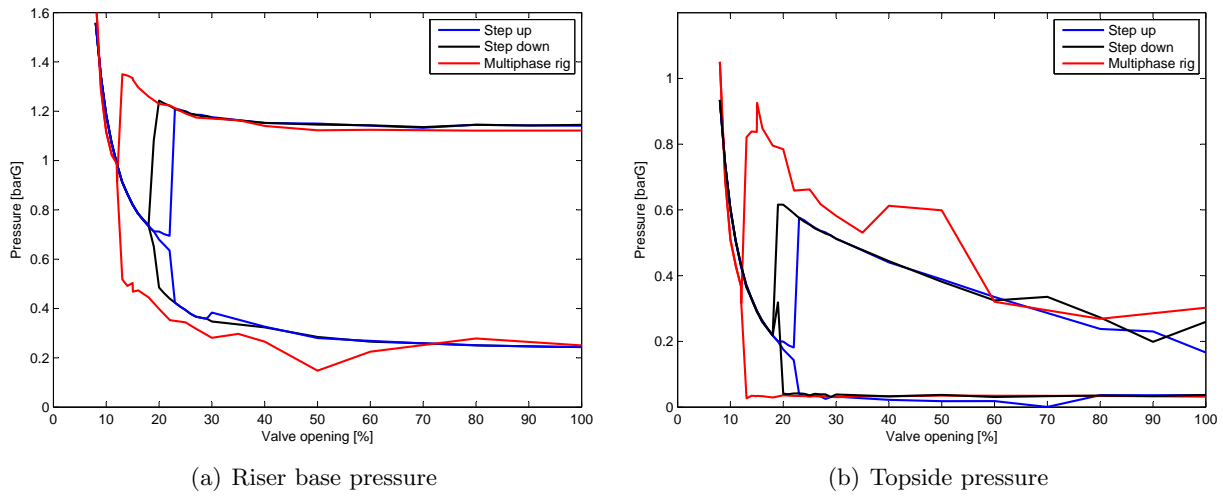
(a) Riser base pressure

(b) Topside pressure

Figure 4.1: Bifurcation plots with CD=0.84

Figure 4.1 shows a bifurcation plot when the discharge coefficient (CD) of the valve is equal to 0.84. This shows that the OLGA model predicts the steady-state behavoir fairly well, since the blue,black and red line lies on top of each other in the stable area. However, the multiphase rig will exhibit the unwanted slug flow regime at a lower valve opening than the model.

(a) Riser base pressure

(b) Topside pressure

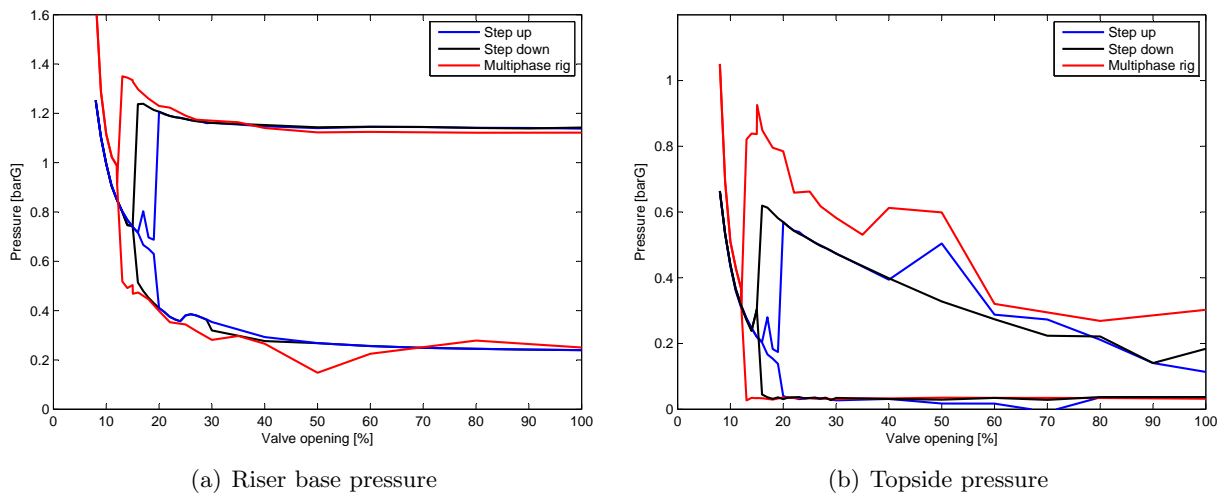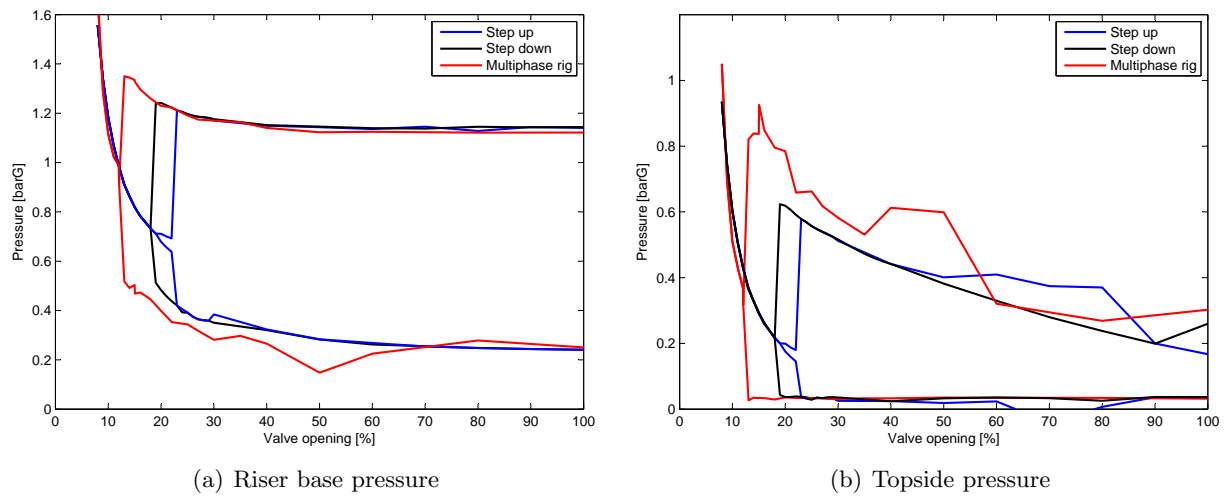**Figure 4.2:** Bifurcation plots with CD=1.0

An attempt to move the transition point for the model to a lower valve opening was performed by increasing the discharge coefficient. A bifurcation plot of this can be seen in figure 4.2. By comparing figure 4.1 with figure 4.2 the transition point has been moved to a lower valve opening. However, this has been achieved on the expense of the steady-state behavior.

(a) Riser base pressure

(b) Topside pressure

Figure 4.3: Bifurcation plots when the interfacial friction parameter is reduced by 20%. CD=0.84

Using the tuning option in OLGA, the interfacial friction parameter was reduced by 20% (The discharge coefficient was kept at 0.84). The bifurcation plot for this case is shown in figure 4.3. By comparing figure 4.1 to figure 4.3 the transition point has not changed significantly. Reducing this factor further resulted in simulation problems.

Figure 4.4 and figure 4.5 show frequency plots for two different unstable valve openings, 30% and 100%. The discharge coefficient is equal to 0.84 with no change in the interfacial friction parameter.

(a) Valve opening = 30%
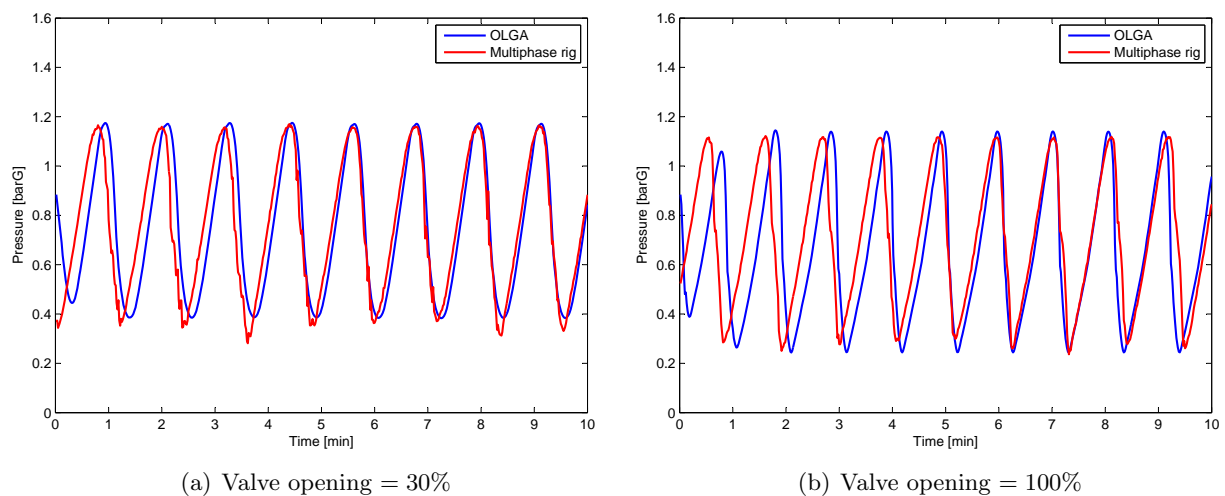
(b) Valve opening = 100%

Figure 4.4: Open-loop simulations for unstable valve openings. Riser base pressure.

(a) Valve opening = 30% (b) Valve opening = 100%
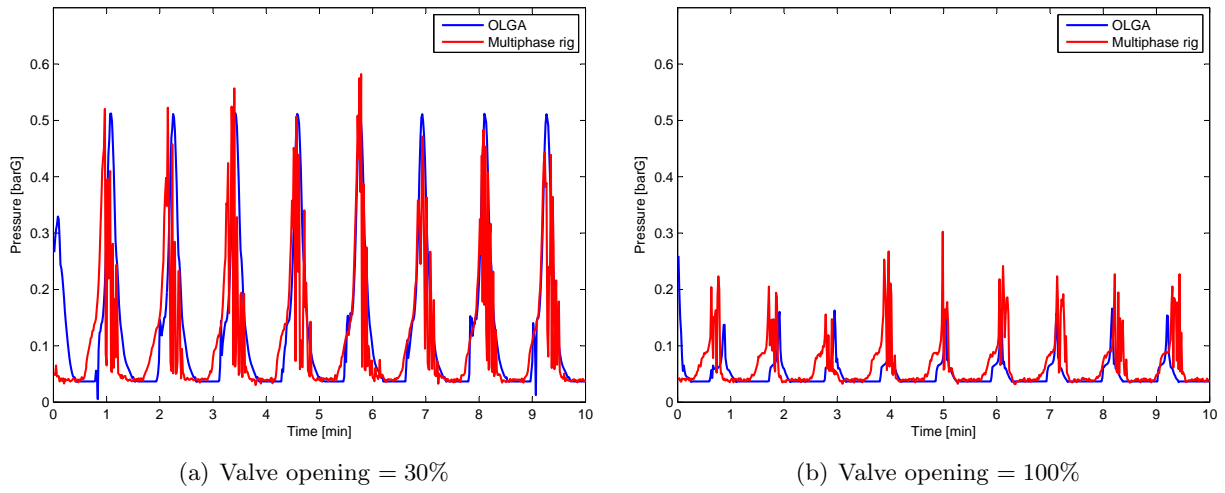
Figure 4.5: Open-loop simulations for unstable valve openings. Topside pressure.

The slug frequency and amplitude of the oscillations are well predicted for the two valve openings. This applies for both the topside and riser base pressure.

4.2 Applying the process gain methods

In this section the three methods outlined in section 3.2.2 will be applied to the OLGA model. First the method based on applying step-changes to the OLGA model will be investigated. This will be compared to the process gain predicted by the steady-state model. Then the least-square method for finding the constant K_1 in the steady-state model based on experimental data from the OLGA model will be looked into. Note that the process gain plots show the absolute value of the gain. The process gain for this system is actually negative, which means that the process variable (riser base pressure) decreases when the manipulated variable increases.

Finding an approximation for the process gain based on the OLGA model

In section 3.2.2 it was outlined a way for finding an approximation to the process gain both for open-loop stable and unstable systems. While finding the process gain for stable systems is straight-forward, finding it for unstable systems requires that the system is stabilized first, using a controller. The process gain for a few selected valve openings were found for each valve opening by perform-ing a step change for the choke-valve (or the reference to the cascade controller for unstable valve openings) of 0.5%. The process variable used here is the riser base pressure. Figure 4.6 shows the resulting process gain plot.
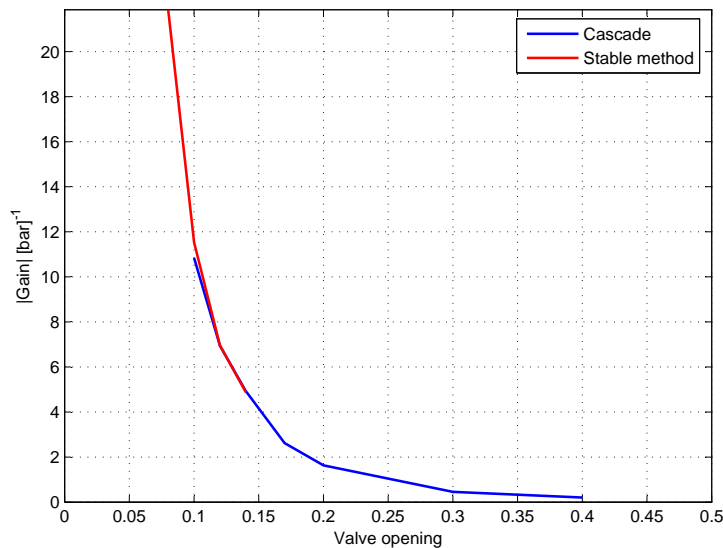
Figure 4.6: Process gain plot, found by performing procedure 3.1 and 3.2.

The stable method was used for stable valve openings, while the cascade method was used for unstable valve openings as well as for some stable valve openings for verification of the method. From figure 4.6 it can be seen that the cascade method predicts the same process gain as the stable method. The process gain was computed for the following valve openings.

$$U^T = \begin{bmatrix} 0.08 & 0.1 & 0.12 & 0.14 & 0.17 & 0.2 & 0.3 & 0.4 \end{bmatrix}$$

It was here used a step change of 0.5%. A smaller step could give a better description of the process gain, but at the same time it would be vulnerable to numerical issures, since the process gain is

found by dividing on the step change (see equation (3.1) and (3.6)). If the step change approach zero, the change in the process variable will altså approach zero, and equation (3.1) and (3.6) will approach a $\frac{0}{0}$ expression. This is obviously not very fortunate. A step change of 0.5% was for this purpose found out to be a good trade-off between avoiding an inaccurate description of the process gain and numerical difficulties.

Method based on a steady-state model

Below a restatement of equation (3.10) is given.

$$
\begin{aligned}
K_p(u) &= -2\left(\frac{Q_T}{C_v}\right)^2 \rho_a u^{-3} = -2K_1 u^{-3} \\
K_1 &= \left(\frac{Q_T}{C_v}\right)^2 \rho_a
\end{aligned}
\tag{4.1}
$$

To find K_1 it is required to know the total volume flow Q_T, the average density ρ_a and the C_v value for the valve model. Q_T and ρ_a could be found through OLGA simulations. Since the choke-valve model in OLGA was used, and the discharge coefficient (CD) was found through tuning, the actual C_v-value was not known. The C_v value was therefore computed via the following relationship.

$$
C_v = \frac{Q_T}{u}\sqrt{\frac{\rho_a}{\Delta p}}
\tag{4.2}
$$

To compute the C_v value Q_T, ρ_a and Δp was found via measurements from OLGA. A few open-loop simulations for stable valve openings were performed in order to compute the C_v-value. The following table shows the necessary values for computing K_1, for different valve openings.

%	C_v	ρ_a	Q_T
8%	0.0012	225	0.0020
13%	0.0012	225	0.0019
15%	0.0012	229	0.0020
18%	0.0012	229	0.0020

Table 4.1: Necessary variables for computing K_1

Taking the average of the variables in table 4.1 and inserting them into the formula (4.1) gives the following value for K_1.

$$
K_1 = \left(\frac{Q}{C_v}\right)^2 \rho = 0.0061 \ [\text{bar}^{-1}]
\tag{4.3}
$$

The function in equation (4.1) is plotted with the value for K_1 given in (4.3) together with the method based on simulations only.
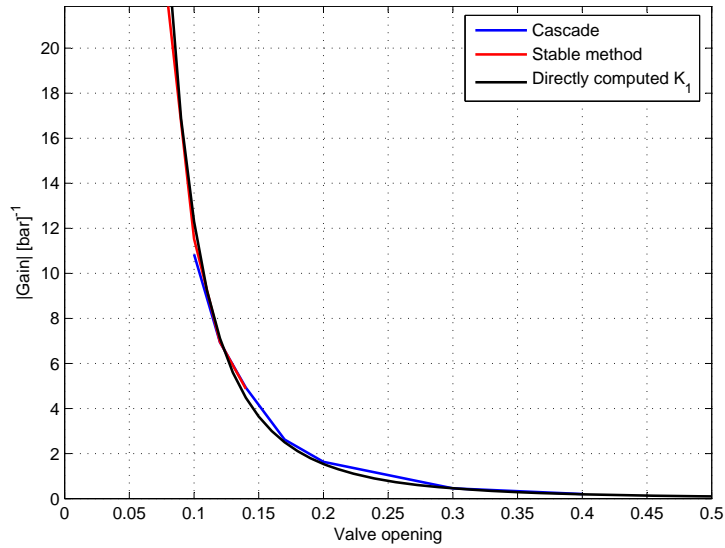
Figure 4.7: Process gain found by using the simplified steady-state model. Compared to the process gain found by procedure 3.1 and 3.2.

Least square-method

K_1 in equation (3.10) could be found by inserting steady-state values for the physical values. But finding the steady-state value for for example ρ is hard. Another approach is to use the data that was found for the process gain with procedure 3.1 and **??**. In practice the following valve openings where used.

$$U^T = \begin{bmatrix} 0.08 & 0.1 & 0.12 & 0.14 & 0.17 & 0.2 & 0.3 & 0.4 \end{bmatrix}$$

And the resulting K_1 was

$$K_1 = 0.0056 \tag{4.4}$$

The least square method is plotted together with the cascade and stable method. The different data points are equally weighted.
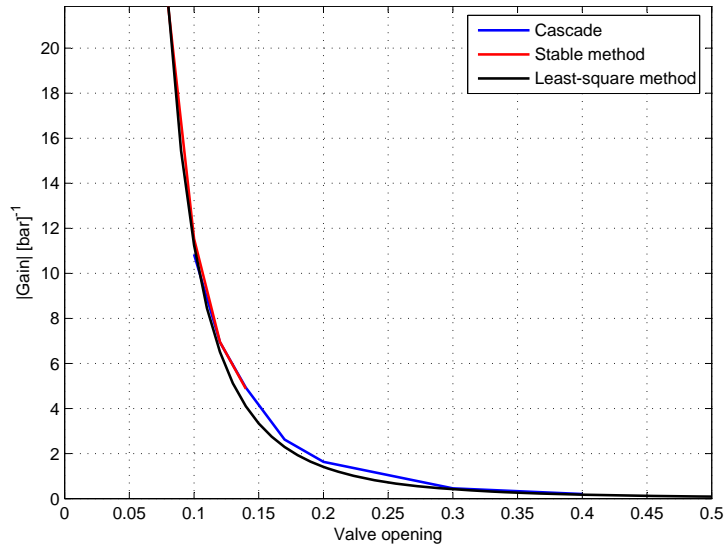
Figure 4.8: Process gain found by the least-square approach. Compared to the process gain found by procedure 3.1 and 3.2.

Least square method with modified weights

Figure 4.8 showed the least square approach where all the data points were equally weighted. In this section the following weights were chosen:

$$w_i = 1 \text{ for } i = \{1, 2, 3, 8\} \text{ and } w_i = 80 \text{ for } i = \{4, 5, 6, 7\}$$

where the valve openings for which gain data existed were

$$U^T = \begin{bmatrix} 0.08 & 0.1 & 0.12 & 0.14 & 0.17 & 0.2 & 0.3 & 0.4 \end{bmatrix}$$

The weights were chosen to find a K_1 that fitted better for valve openings of 14%, 17%, 20%, 40%. The resulting K_1 was

$$K_1 = 0.0064 \tag{4.5}$$

In figure 4.9 the resulting function from the least-square method with modified weights is plotted together with the original data.

Plotting the function together with the original data
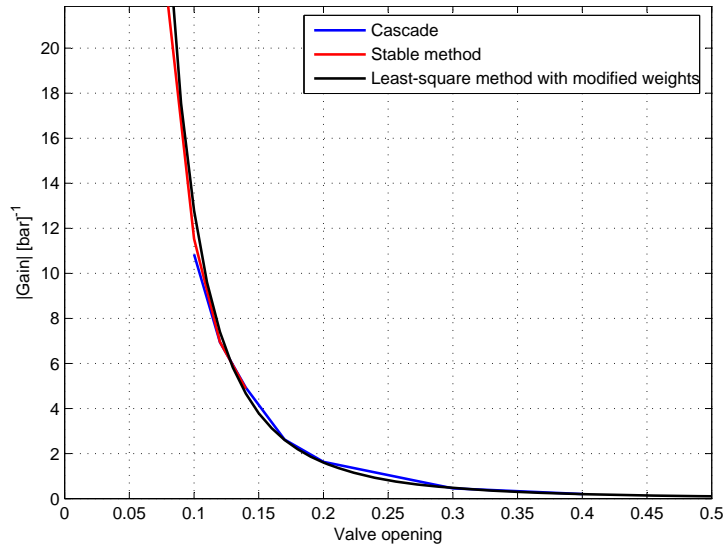
Figure 4.9: Process gain found by the least-square approach with modified weights. Compared to the process gain found by procedure 3.1 and 3.2.

4.3 Simulations of different gain scheduling strategies

In section 3.3 gain scheduling theory was presented. Simulation of different types of scheduling variables and scheduling functions will be shown. The controllers which will be investigated were presented in table 3.2, and will be restated here for convenience.

	Topside pressure	Riser base pressure	Valve opening
Linear scheduling	X	X	X
Cascade scheduling			X
OLGC scheduling			X

Table 4.2: Matrix showing combinations of scheduling functions and scheduling variables considered in this thesis.

Equal for all the controllers based on a linear scheduling function is that they are tuned without noise for two valve openings 8% and 40%. The OLGC approach has only one tuning parameter, and is tuned for the valve opening of 40%.

To avoid having too many plots for each controller, only plots with noise will be presented if the controller is able to stabilize the flow without noise. Each controller will be presented with two setpoints, corresponding to operational conditions at 8% and 30% with noise. If they are plotted without noise it will be for operational conditions corresponding to 8% and 40% valve openings. All the gain scheduling controllers are plotted against constant gain controllers with gain corresponding to those valve openings which are presented.

Simulations will last for 20 minutes without noise and 60 minutes with noise. The noise is implemented as white noise in the interval ±0.05 barg. Please note that for some of the simulations there

are some vertical pressure spikes. The reason for these spikes is that OLGA have fluid characterizations in a table file (pvt-file), and if the pressure in the model is outside the interval in this file, these spikes occur. Note that these spikes occur only for controllers which are shown to not work satisfactory.

All the controllers are turned on at the lowest pressure in the slug cycle (riser base pressure). This is because that this position is the worst to stabilize the system from. In appendix B simulations show that it matters when the controller is turned on for stabilization. When the controller is turned on is shown in the plots by a vertical black line (do not mistake this for being a pressure spike).

The results will be presented here with some comments. The discussion and conclusion, however, will be presented in chapter 5.

4.3.1 $\zeta = u$ - Linear approach

Linear approach using a low-pass filtered measurement of the valve opening as the scheduling variable ζ. Riser base pressure is used as the controlled variable. Equation (3.15) shows the continuous equivalent of the gain scheduled PI-controller used. T_i is 50 seconds, while the cut-off frequency T for the first-order low-pass filter in equation (3.23) is 20 seconds.

(a) Gain and valve (ζ) plot

(b) Riser base pressure

(c) Gain and valve (ζ) plot

(d) Riser base pressure
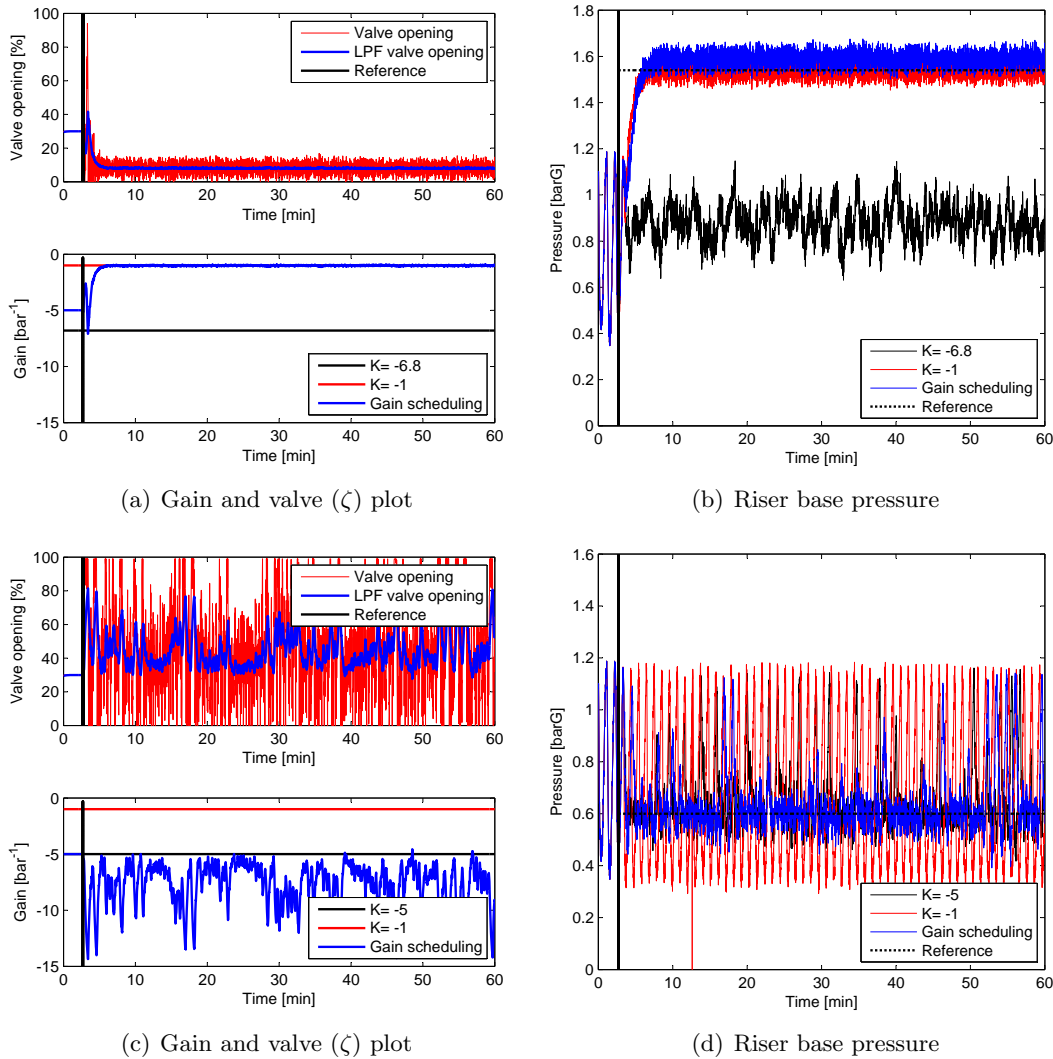
Figure 4.10: a) and b) - reference 1.54 barg. c) and d) reference 0.6 barg

Valve opening [%]	Gain [bar^{-1}]
8	-1
40	-6.8

Table 4.3: Tuning variables used for the linear scheduling function, where ζ is a low-pass filtered measurement of the valve opening

4.3.2 $\zeta = u$ - Cascade approach

Cascade approach using a low-pass filtered measurement of the valve opening as the scheduling variable ζ in the secondary loop without integral action, where the riser base pressure is used as the controlled variable. The primary loop controls the valve opening using a conventional PI-controller. See (3.21) and (3.22). T_i is 50 seconds, while T for the first-order low-pass filter is 20 seconds. See equation (3.23).
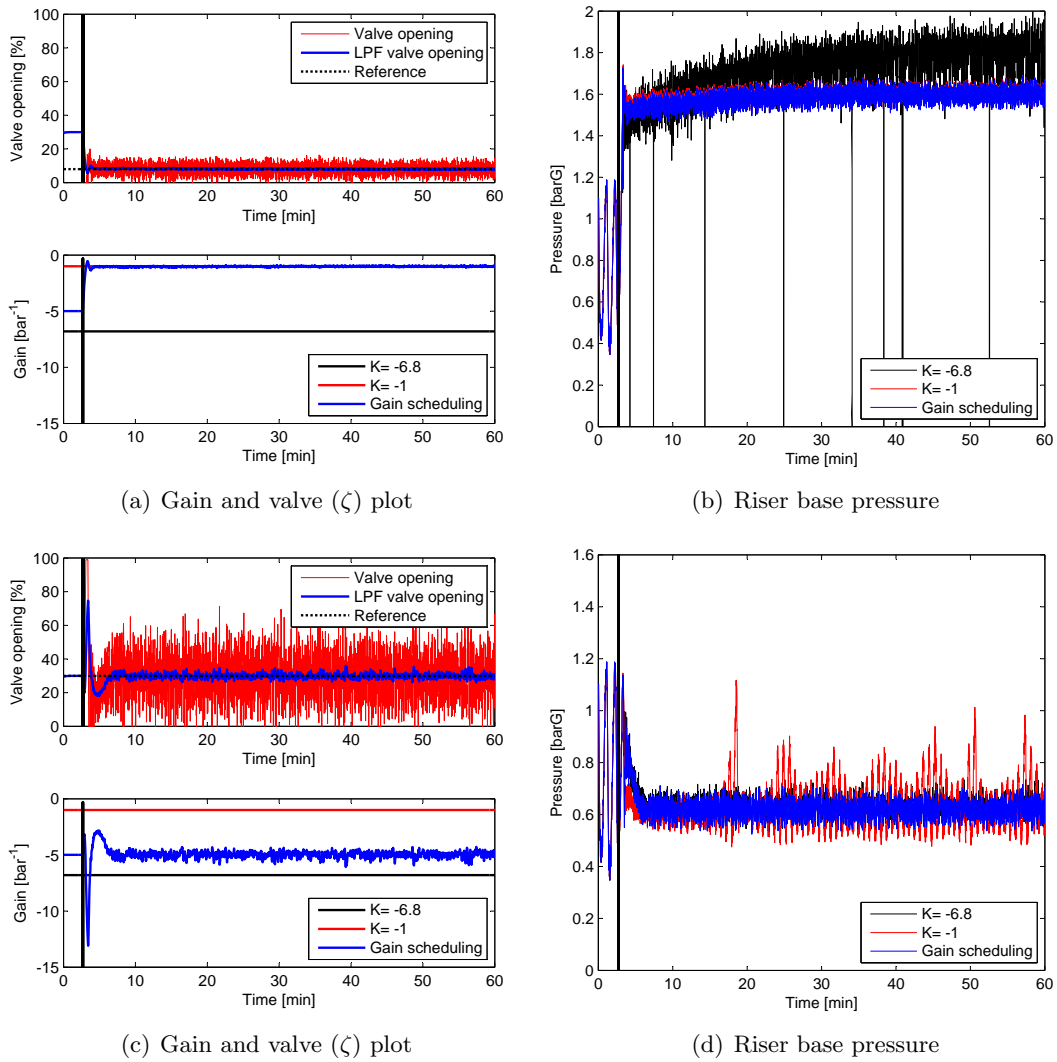
Figure 4.11: a) and b) - reference 8%. c) and d) reference 30%

Valve opening [%]	Gain [bar^{-1}]
8	-1
40	-6.8

Table 4.4: Tuning variables used for the linear scheduling function in the secondary loop

4.3.3 $\zeta = u$ - OLGC approach

OLGC approach using a low-pass filtered measurement of the valve opening as the scheduling variable ζ. Riser base pressure is used as the controlled variable. Equation (3.15) shows the continuous equivalent of the gain scheduled PI-controller used. T_i is 50 seconds, while the cut-off frequency T for the first-order low-pass filter in equation (3.23) is 30 seconds. Simulations were performed without noise.


(a) Gain and valve (ζ) plot

(b) Riser base pressure

(c) Gain and valve (ζ) plot

(d) Riser base pressure
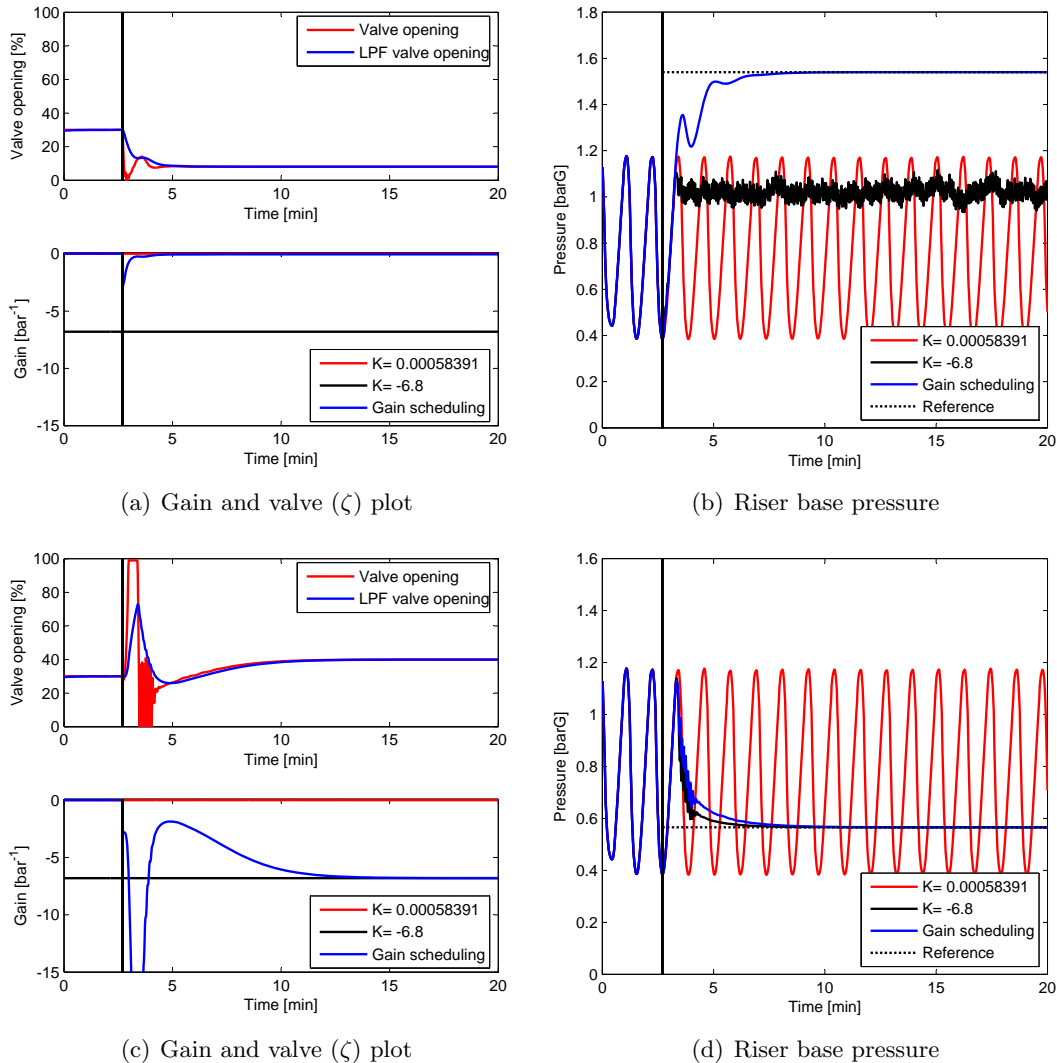
Figure 4.12: a) and b) - reference 1.54 barg. c) and d) reference 0.5657 barg

Valve opening [%]	Gain [bar^{-1}]
40	-6.8

Table 4.5: Tuning variable used for the OLGC gain scheduling approach, where ζ is a low-pass filtered measurement of the valve opening

4.3.4 ζ=Topside pressure - Linear approach

Linear approach using a low-pass filtered measurement of the topside pressure as the scheduling variable ζ. Riser base pressure is used as the controlled variable. Equation (3.15) shows the continuous equivalent of the gain scheduled PI-controller used. T_i is 50 seconds, while the cut-off frequency T for the first-order low-pass filter in equation (3.23) is 20 seconds.

(a) Topside pressure (ζ)

(b) Riser base pressure

(c) Gain plot

(d) Valve plot

Figure 4.13: Reference 1.54 barg

(a) Topside pressure (ζ)

(b) Riser base pressure

(c) Gain plot

(d) Valve plot
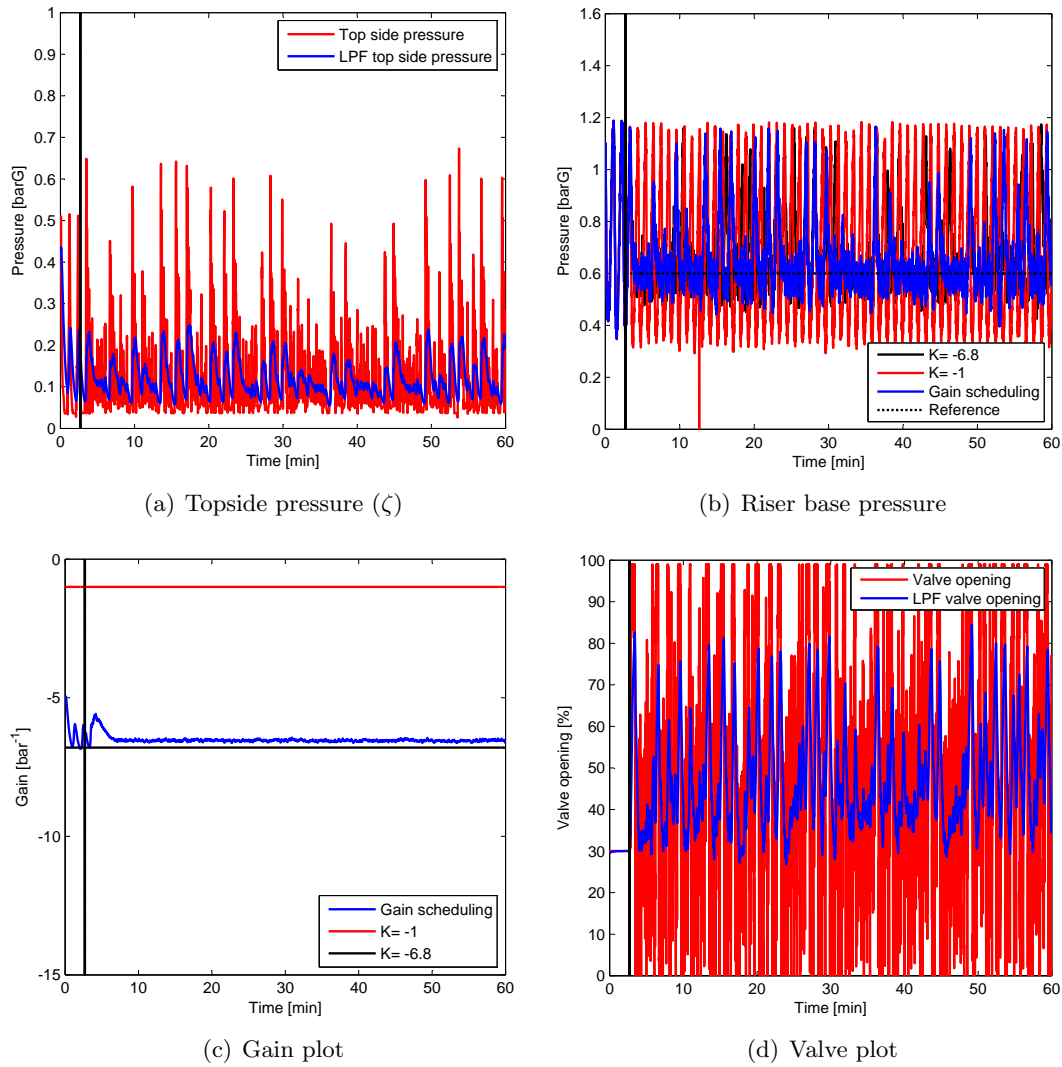
Figure 4.14: Reference 0.5657 barg

Topside pressure barg	Gain [bar^{-1}]
0.9186	-1
0.0777	-6.8

Table 4.6: Tuning variable used for the linear gain scheduling approach, where ζ is low-pass filtered measurement of the topside pressure.

4.3.5 ζ=Riser base pressure - Linear approach

Linear approach using a low-pass filtered measurement of the riser base pressure as the scheduling variable ζ. Riser base pressure is used as the controlled variable. Equation (3.15) shows the continuous equivalent of the gain scheduled PI-controller used. T_i is 50 seconds, while the cut-off frequency T for the first-order low-pass filter in equation (3.23) was 20 seconds.
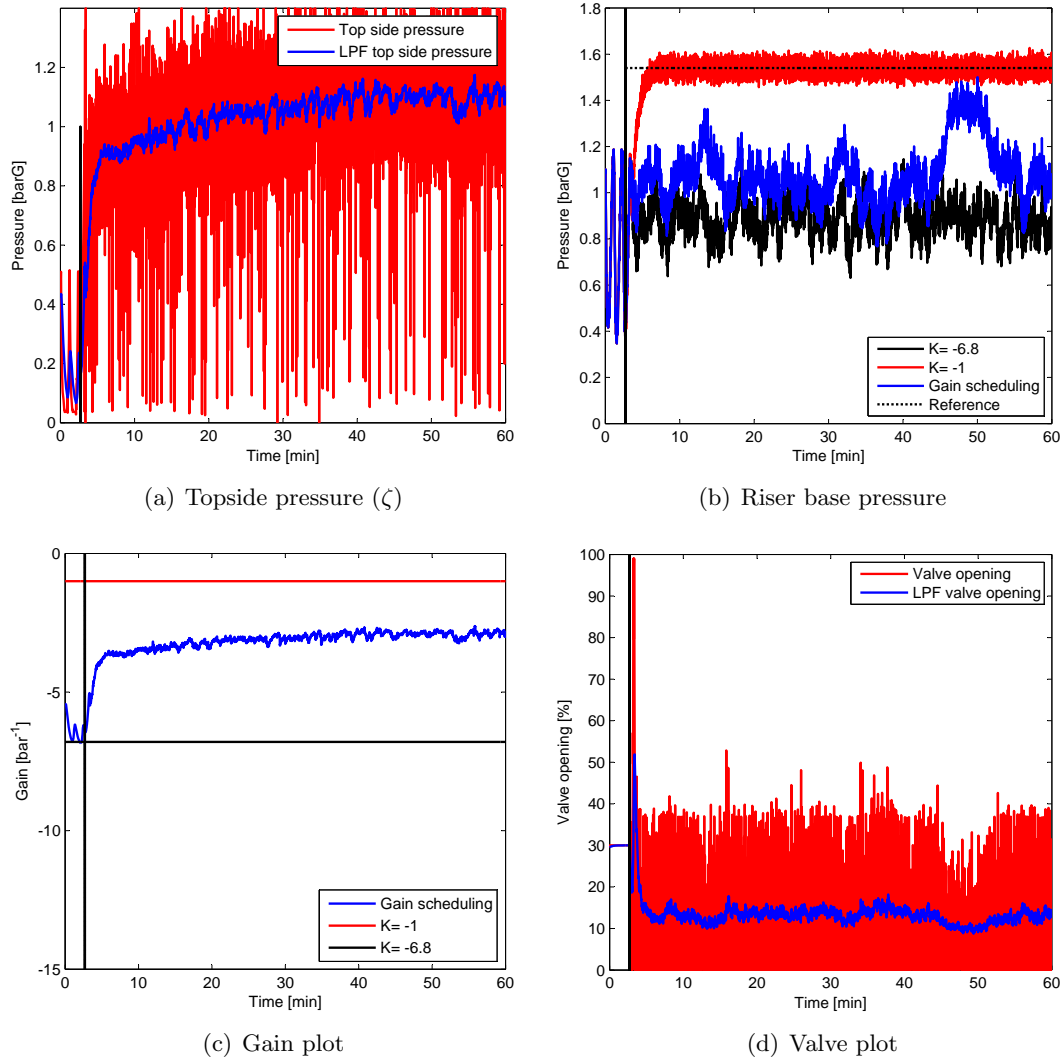

(a) Topside pressure

(b) Riser base pressure (ζ)

(c) Gain plot

(d) Valve plot
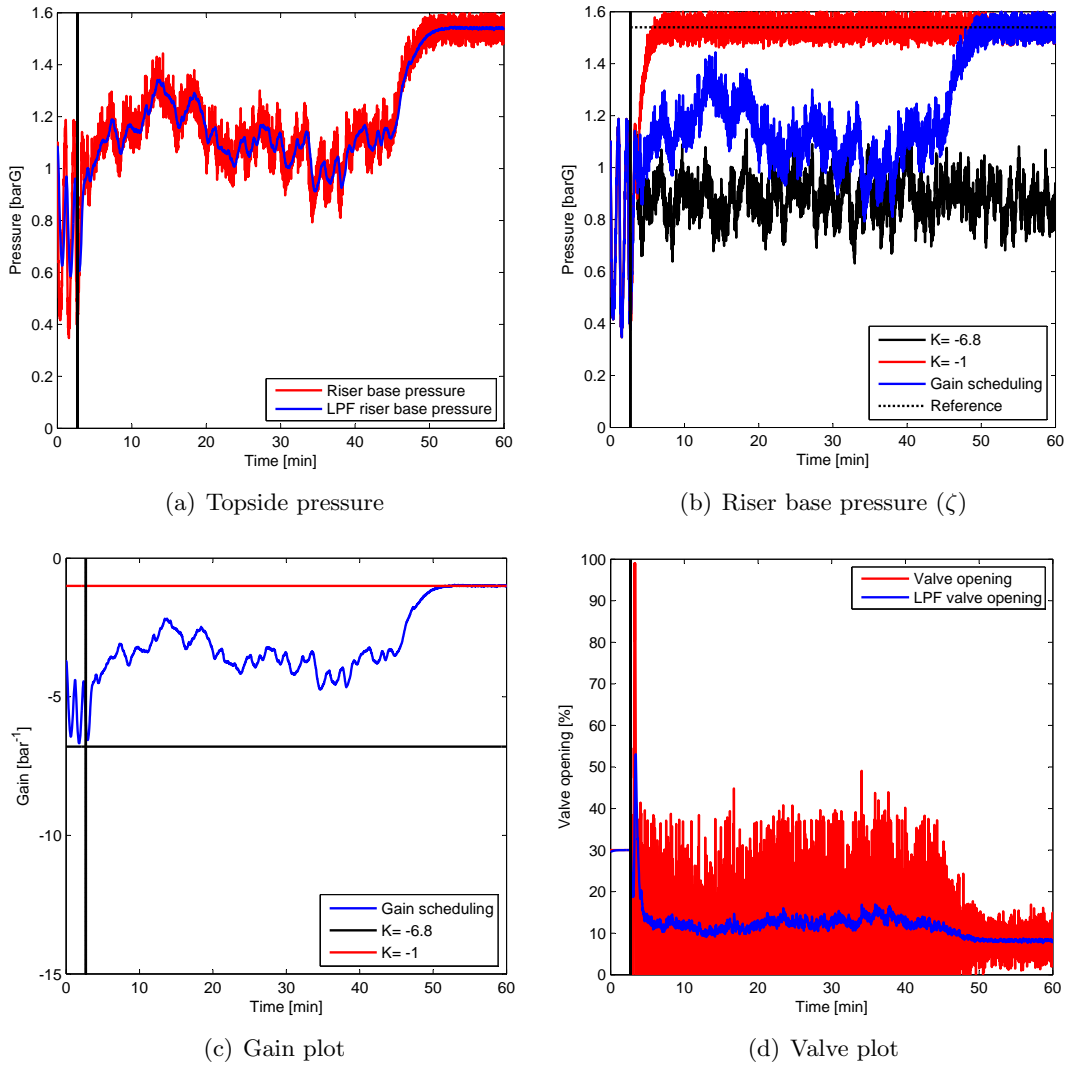
Figure 4.15: Reference 1.54 barg

(a) Topside pressure

(b) Riser base pressure (ζ)

(c) Gain plot

(d) Valve plot
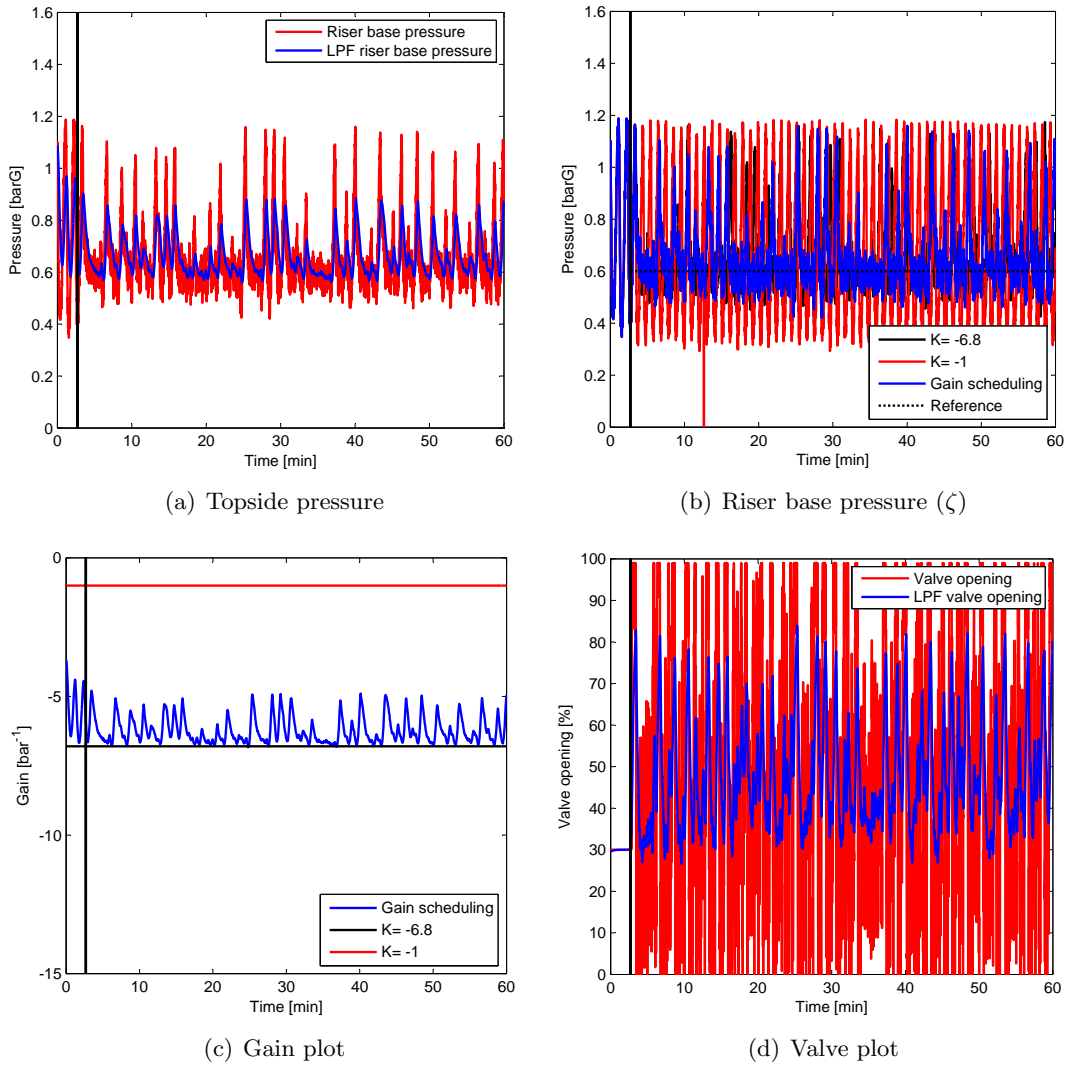
Figure 4.16

Topside pressure barg	Gain [bar^{-1}]
1.54	-1
0.5657	-6.8

Table 4.7: Tuning variables used for the linear gain scheduling approach, where ζ is a low-pass filtered version of the riser base pressure.

4.3.6 Rate changes

Lowering the rates by 10%

Simulations with the inlet rates lowered by 10% were performed for the combined gain scheduling and cascade controller, using a low-pass filtered measurement of the valve opening as the scheduling variable. Figure 4.17 shows a new bifurcation plot after a 10% change in the inlet rates. Figure 4.18 shows simulations of the cascade controller.

(a) Riser base pressure

(b) Topside pressure
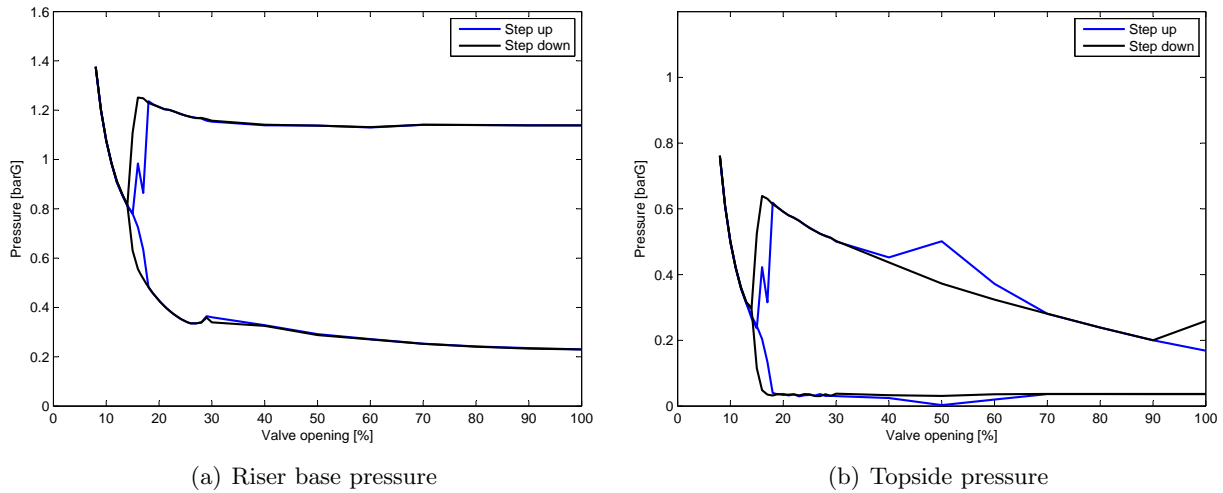
Figure 4.17: Bifurcation plots with inlet rates lowered by 10%

(a) Gain and valve (ζ) plot

(b) Riser base pressure

(c) Gain and valve (ζ) plot

(d) Riser base pressure
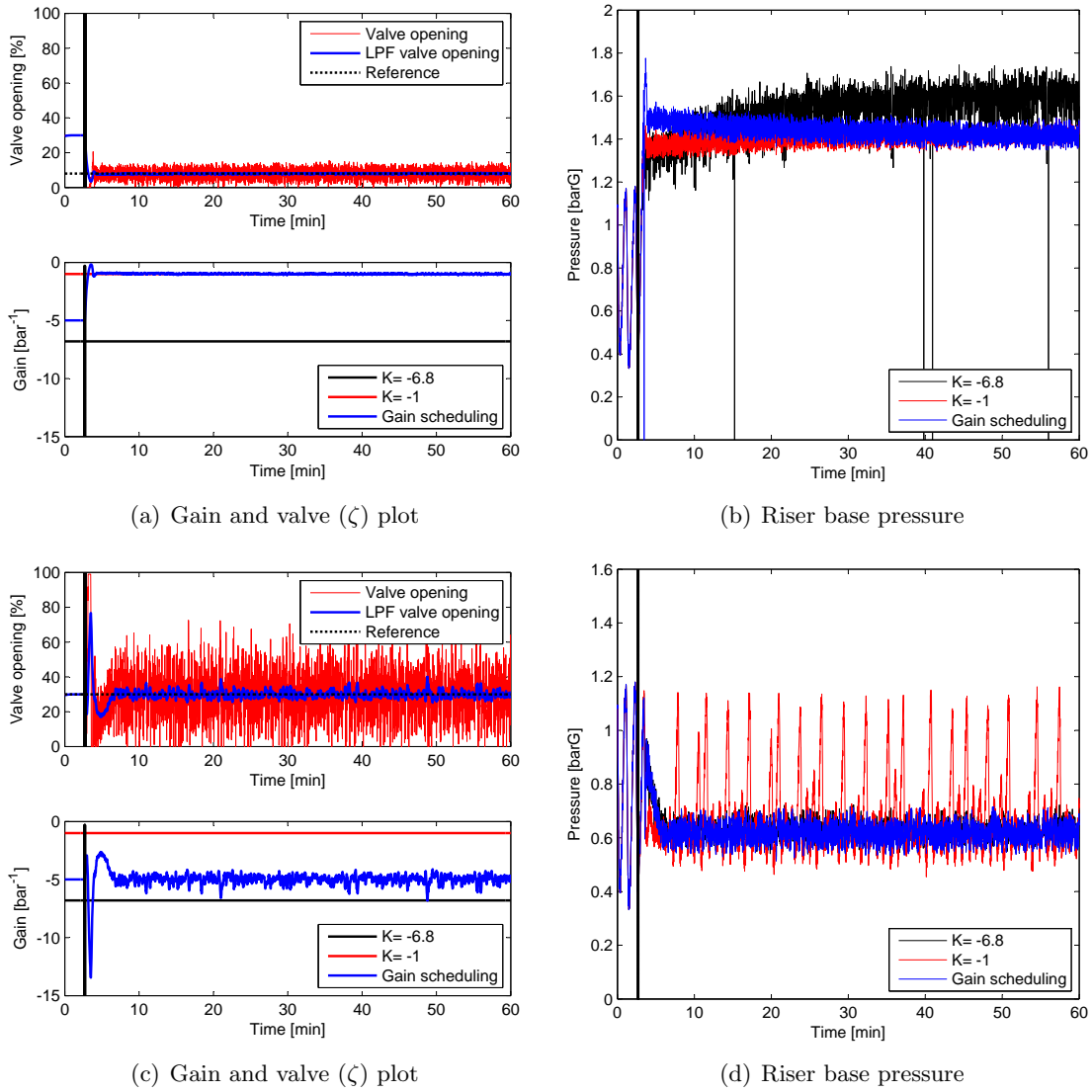
Figure 4.18: a) and b) - reference 1.54 barg. c) and d) reference 0.6 barg

Lowering the rates by 20%

Simulations with the inlet rates lowered by 20% were performed for the combined gain scheduling and cascade controller, using a low-pass filtered measurement of the valve opening as the scheduling variable. Figure 4.19 shows a new bifurcation plot after a 20% change in the inlet rates. Figure 4.20 shows simulations of the cascade controller.

(a) Riser base pressure

(b) Topside pressure
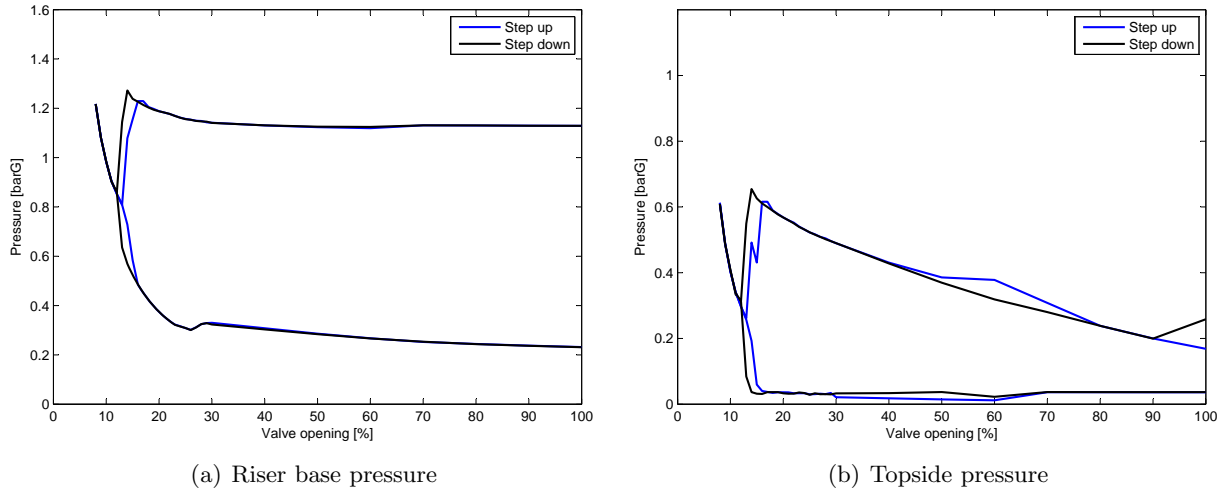
Figure 4.19: Bifurcation plots with inlet rates lowered by 20%

(a) Gain and valve (ζ) plot

(b) Riser base pressure

(c) Gain and valve (ζ) plot

(d) Riser base pressure
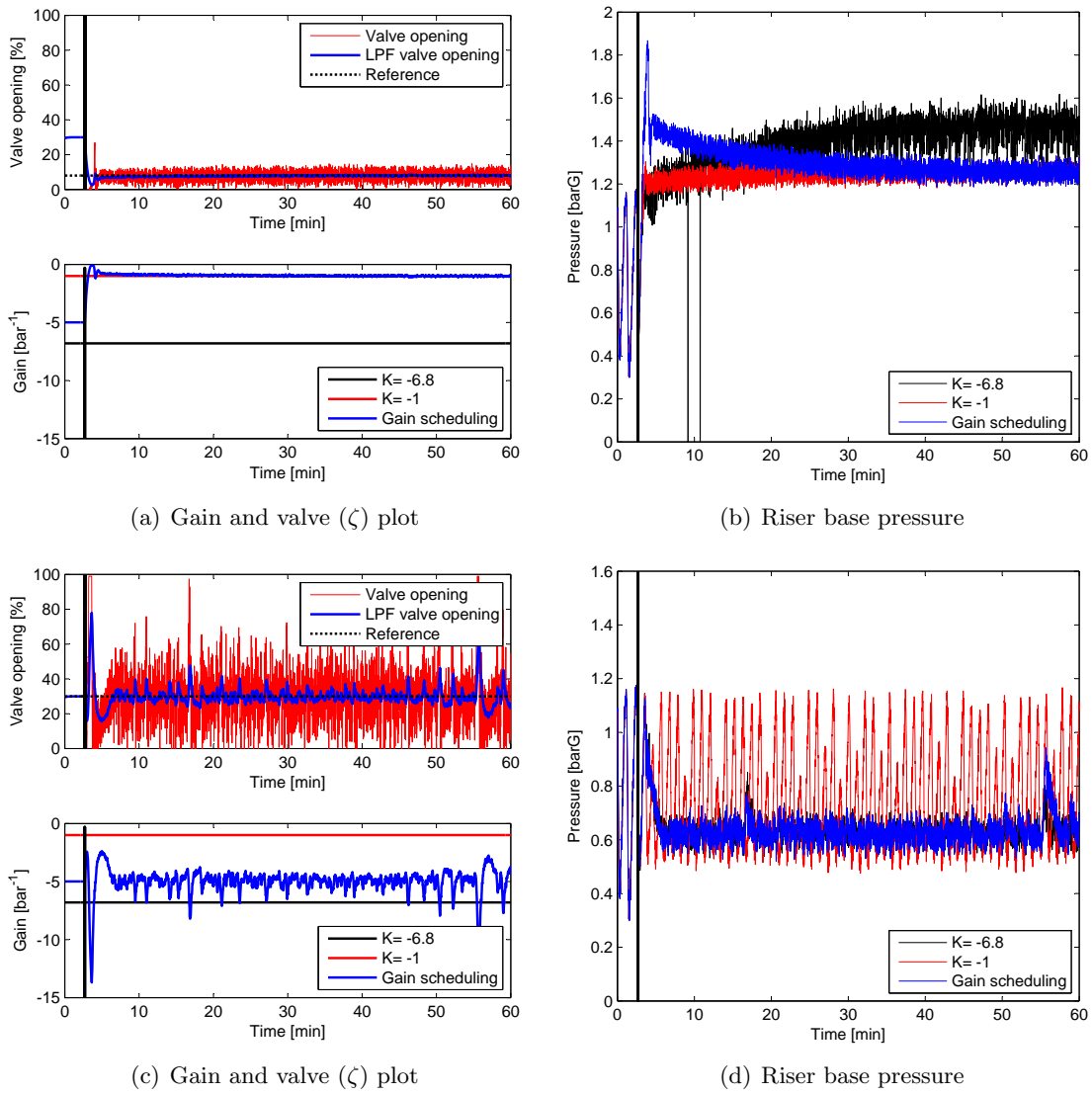
Figure 4.20: a) and b) - reference 1.54 barg. c) and d) reference 0.6 barg

4.4 Summary of results

Simulations from the OLGA model were plotted against process data from the rig in bifurcation diagrams comparing the amplitude of the slugs and the transition point from stable to unstable slug flow. Bifurcation plots for different discharge coefficients were shown, in addition to bifurcation plots when changing the interfacial friction between the phases. Neither tuning on the discharge coefficient nor the interfacial friction parameter were able to predict the transition point between the stable and unstable flow better, while at the same time predict the steady-state flow satisfactory. On the final model, a transition point for the OLGA model of 19% had to be accepted altough the transition point for the multiphase rig was 12%. Despite this, the open-loop plots showed that the slug frequency were predicted very well for the valve openings of 30% and 100%.

The process gain was found by three different methods and compared. Please note that the gain is actually negative, however, it is the absolute value of the gain which is presented in the process gain figures. The first method based on procedure 3.1 and 3.2 compared to the steady-state model gave almost the same process gain. The least-square method was just a combination of these two methods.

Simulations of several gain scheduling approaches have been shown, and robustness with respect to rate changes were shown for the combined cascade and gain scheduling controller. The linear approach using the valve opening as the scheduling variable was able to stabilize the system at the small valve opening. At the 30% valve opening some pressure oscillations can be observed. This was also the case when using the riser base pressure as the scheduling variable. Using the topside pressure as the scheduling variable, the gain scheduling approach was not able to stabilize the system at any of the valve openings.

The OLGC scheduling were only shown without noise. In appendix C a simulation study shows that this approach is not robust with respect to the filtering coefficient T. In addition to this the OLGC approach has only one tuning factor which resulted in a very low gain for the low valve opening (8%). This gain was not enough for stabilizing the system around this valve opening when turning on the controller from an unstable condition. This approach lacks robustness properties and therefore simulations with noise are not shown. The OLGC algorithm is neither used in a cascade manner.

Using a linear scheduling function with a low-pass filtered measurement of the valve opening as the scheduling variable, in a cascade control fashion was able to stabilize the system for both valve openings. This approach was also robust with respect to 10% rate changes.

Chapter 5

Discussion and concluding remarks

Use of control theory as a way of stabilizing slug flow has been shown to work well, and has been documented in several academic papers. The advantage by using control methods is that little or no new equipment is needed to suppress the slug flow, while at the same time operate at boundary conditions which would normally cause the unwanted slug flow regime.

Using the riser base pressure as the controlled variable has proven to have a stabilizing effect on the slug flow. However, since this requires a measurement at the bottom of the sea, using topside measurements is desirable. However, these measurements are either limited by unstable zero dynamics or have low steady-state gain.

The view which has been taken here, however, is to further investigate the riser base pressure as a controlled variable and try to make this controller more robust. This approach has been motivated by experience from the multiphase rig at Hydro's Research Centre, where it has been discovered that the controllers need to be re-tuned during experiments because of changes in operational conditions caused by noise and the fact that the process gain changes as the operational conditions changes. As a part of the contribution to solving this problem, an OLGA model of the multiphase rig has been developed, and simple nonlinear controllers have been simulated on this very model.

5.1 OLGA modeling

The OLGA model was tuned using two approaches. Tuning on the discharge coefficient of the valve, in addition to changing the interfacial friction parameter.

Increasing the discharge coefficient was believed to move the transition point from stable flow to unstable slug flow to a lower valve opening. However, the problem with increasing this value, was that at the same time the prediction of the steady-state behavior worsened. A trade-off between the prediction of the transition point and the steady-state behavior had to be made. Since the controllers will be operating around steady-state, this behavior was regarded as the most important, and the value of the discharge coefficient was set to 0.84.

Lowering the interfacial friction between the air and water was also believed to move the transition point to a smaller valve opening. However, after reducing it by an amount of 20% no significant improvement could be seen. Reducing it even further only led to simulation problems. In the OLGA manual it is stated that tuning should be applied with great care, as the validation and verification

of the OLGA model may not be valid for such cases.

Consequently, it was decided to keep the discharge coefficient at the 0.84 value and no changes to the interfacial friction parameter was made on the final model. The multiphase rig has a transition point at 12% opening, while the OLGA model has a transition point at 19%. The rates are implented as constant inputs to OLGA through the OLGA MATLAB toolbox. Pump dynamics has not been included in the model, and could perhaps affect this transition point.

Lowering the inlet rates showed that the transition point in the bifurcation plots moved to a lower valve opening. Consequently the rates could have been used as a tuning parameter. But since the rates are well known from measurements at the rig, using the rates as a tuning factor could not have been explained from physical insight.

It has been used two methods for finding the bifurcation plots for the OLGA model. Either by performing steps on the model starting from stable valve openings or unstable valve openings. Note that for the multiphase rig there was only data starting from unstable valve openings. From all the bifurcation plots it can be seen that the transition point seems to be at a larger valve opening if one start at unstable valve openings. Logically it should have been the same valve opening. Imagine that a linear model is found around every valve opening for the pipeline-riser system. At small valve openings the system would have all the poles in the left-half plane which means that the system is stable. For higher valve openings at least some of the poles would be in the right-half plane forcing the system to converge to the nonlinear limit cycle. Somewhere between the stable and unstable valve openings there should exist exactly *one* valve opening which had all the poles on the imaginary axis. However, the system can still exhibit a limit cycle although the equilibrium point is stable. So whether the system converges to the limit cycle or the stable equilibrium point depends on the initial condition.

Motivated by the construction of the multiphase rig at the Research Centre, constant inlet rates of water and gas are used. The drawback by using constant rates is that the flow is the same at steady-state irrespective of the valve opening. However, stabilizing the system at a higher valve opening would in practice lead to higher production throughput for pressure dependent inlet rates.

5.2   Process gain

The steady-state process gain has a huge impact on the choice of the control parameters. Finding this gain has therefore been an important basis when gain scheduling controllers were to be designed.

Perhaps the easiest way to find the process gain, is to perform small steps of the manipulated variable for the system and measure what the process variable is at steady-state for that particular value of the manipulated variable. The process gain would then simply be $\frac{\Delta y_{ss}}{\Delta u_{ss}}$. However, since the pipeline-riser system is unstable for valve openings larger than a critical valve opening, the system must be stabilized first to be able to find the steady-state process variable. Such a method has been outlined in this thesis. Using this method is easy when an OLGA model has been developed, but if the process was to be found by performing steps on the actual rig the result might be inaccurate. The problem lies in the fact that the system has to be stabilized first if the process gain is to be found.

Therefore, using the steady-state model outlined in section 3.2.2 as a basis for finding the process

gain, might be a better approach. However, the resulting steady-state gain function had a parameter K_1 which had to be found. Finding this constant analytically required the knowledge of ρ_a, Q_T and C_v. The C_v value for a valve is often known, but knowing the average density ρ_a and the total volume flow Q_T which both depend on the pressure in the pipeline could be difficult. Whether finding process gain data from the rig by performing steps is more difficult than estimating ρ_a and Q_T for finding K_1, is hard to say without actually testing this method on the rig.

It is very useful to know the form of the process gain. And K_1 could instead be found by fitting K_1 to process gain data if this was considered easier to find, than the values for Q_T and ρ_a. It could even be found if only gain data for stable valve openings existed. A least square approach with modified weights has also been shown, if some of the data is believed to be more correct than others.

From all of the process gain plots it can be seen that the gain increases incredibly for small valve openings. It should be noted that this is also caused by the case definition where the boundary condition at the inlet is constant rates. If the inlet flow had been pressure dependent, the steady-state process gain would most certainly be more even.

5.3   Gain scheduling

A gain scheduling algorithm can be explained as a family of linear controllers for different operational points. Changing between the linear controllers are done based on the scheduling variable, which here is defined as ζ. Shamma and Athans (1992) outlined two important properties for the scheduling variable. It should capture the plant's nonlinearities and it should vary slowly.

It is obvious that this scheduling variable must capture the nonlinearities of the plant, since coping with the nonlinearities is the reason for taking this approach in the first place. To avoid instability of the closed-loop system, it is important that the process gain varies slowly and therefore should also this scheduling variable vary slowly. But this variable can on the other hand not vary too slowly either, since the system in the unstable area requires high bandwidth in order to stabilize the system. To avoid too rapid changes in the gain, the scheduling variable has been low-pass filtered. The time constant of the low-pass filter has been found by extensive simulations. The scheduling variables considered are measurements of the topside pressure, riser base pressure and the valve opening. Using the reference as a scheduling variable could also be a viable approach, but since this scheduling regime would not take into consideration changes in the process gain dynamically, it has not been looked further into.

In addition to the scheduling variable, the scheduling function must also be chosen. Two approaches were considered in this thesis, a linear scheduling function and the OLGC-function. In addition to this, the linear scheduling function has been combined in a cascade control loop with the valve opening controlled in a primary loop. Table 3.2 summarizes the combinations of scheduling variables and scheduling functions which were investigated.

Comparing all the different gain scheduling controllers, the use of a low-pass filtered version of the valve opening as a scheduling variable in a cascade manner with the valve opening in a primary loop stand forward as the best approach according to the simulations. There are several reasons for this.

First of all, integral action is destabilizing for the plant and removing this from the secondary loop has a positive effect on the controller performance. In addition to this the valve opening is intuitively a very good indicator of the gain in the process, both dynamically and at steady-state.

Using a pressure measurement as the scheduling variable did not work satisfactory. This applies to using both the riser base pressure and the topside pressure. For a given unstable valve opening in the slug cycle, the pressure vary in a large interval. Using a pressure measurement is therefore not such a good idea, since the pressure does not relate as good as the valve opening to the gain in the process.

The OLGC approach was tuned for a valve opening of 40% without noise. Since this approach only consists of one tuning parameter, this resulted in very low gain for the small vale opening (8%) which is considered here. In addition to this it was shown in appendix C that the OLGC approach was little robust to changes in the filtering coefficient. The reason for this is the high derivative of the process gain. There is a trade-off between a rapid change in gain to be able to cope with the instability, and at the same time the gain must not change too rapid according to Shamma and Athans (1992). This method obviously lacked robustness properties and was not investigated more extensively.

It was also investigated how robust the system was with respect to changes in the inlet rate. The rate was reduced by 10% and 20%. The new bifurcation plots after the rate changes show that the instability occurs at a lower valve opening, as predicted. Lower rates mean lower kinetic energy, which would cause more liquid accumulation at the riser base. It is interesting to see that the best gain scheduling algorithm (linear scheduling in a cascade loop with a low-pass filtered version of the valve opening) successfully stabilized the system for a valve opening equal to 30% even with a 10% reduction of the rates. Reducing it further to 20% is about the maximum change in rates before the controller is completely unable to stabilize the system. (some pressure spikes can be observed, in addition to some gain spikes). Another advantage of using a valve opening setpoint instead of directly setting the pressure setpoint when changing the rates, is that this controller setup autmatically changes the pressure setpoint to the secondary loop.

In appendix B it is shown that where in the slug cycle the controller is turned on, impact whether the system is stabilized or not. It is shown that stabilizing the system from the bottom of the slug cycle is the most difficult, so this starting point has therefore been used for all the other simulations. In addition to this the initial starting value for the incremental controller algorithm also impacts whether the system can be stabilized with a small gain for a large valve opening. However, the choice of this starting value is not important as long as the gain is high enough. (This has not been shown by simulations). The important lesson from this simulation study is that to make sure the system is stabilized around a large valve opening, high gain is absolutely necessary.

5.4   Further work

This thesis has only used open-loop data from the multiphase rig in the development of the model. None of the controllers designed and investigated in this thesis has been tested in practice. Even though the simulations of the cascade controller with a gain scheduling algorithm as the secondary loop seems promising, conclusions on its robustness properties can not be concluded fully until it is tested at the multiphase rig.

Appendix A

Comment on controller implementation

In this section comments on the controller implementation will be given, and some information about the OLGA Matlab communication.

A.1 Comment on controller design and OLGA communication

The controllers are implemented in Matlab. Measurements and inputs to the OLGA server are passed through the OLGA Matlab toolbox. A sampling time equal to $1s$ is used. The PI-controllers which are used here is in continuous form:

$$u(t) = K(e(t) + \frac{1}{T_i} \int_{t=0}^{t} e(\tau)\mathrm{d}\tau) \tag{A.1}$$

and in incremental form which is outlined in Balchen et al. (2004):

$$u[k] = u[k-1] + \frac{K(1+T_s)}{2T_i}e[k] - \frac{K(1-T_s)}{2T_i}e[k-1] \tag{A.2}$$

The gain scheduling version of the incremental form is

$$u[k] = u[k-1] + \frac{K(\zeta)(1+T_s)}{2T_i}e[k] - \frac{K(\zeta)(1-T_s)}{2T_i}e[k-1] \tag{A.3}$$

A.2 Anti-windup

Use of PI-controllers when the input saturates introduce the problem of integrator windup. The problem is that when the actuator saturates, the integrator will keep on integrating causing the controller output to increase. An increase of the controller output, however, will certainly not help since the system has already saturated. But the actual problem with integral-windup is when the error signal changes sign, and it is time for the controller output to decrease. The controller output will be much larger than the saturation value of the actuator, and it will take some time before the actuator will be able to get out of saturation modus. The effect is poor performance (i.e overshoot and long settling time).

A very simple anti-windup regime is to detect when saturation occurs, and when this happens set the integrand $e(\tau)$ to zero (i.e. stop integrating). Se below for the discrete anti-windup regime used for the controllers in this thesis. Figure A.1 shows the control configuration of a controller with anti-windup.

$$u(t) = \begin{cases} 0 & \text{if } u_c < 0 \\ 1 & \text{if } u_c > 1. \\ u_c(t) & \text{if } 0 \leq u_c \leq 1. \end{cases}$$
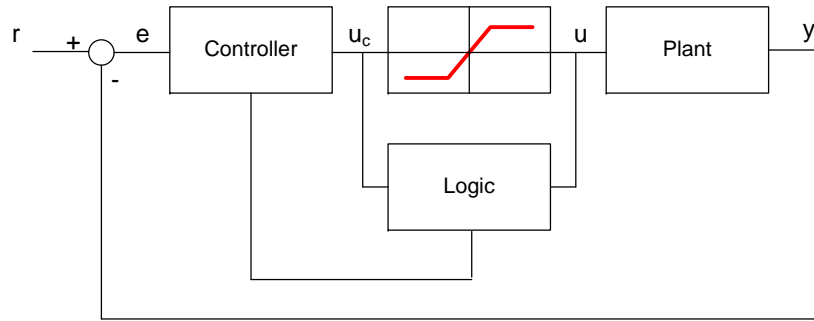

Figure A.1: Control configuration with anti-windup

A.3   Matlab code

The following MATLAB-code is generally what is done with all the controllers. This version is a stripped version. There may be some variations due to different versions of the controller.

```
%---------------------------------------------------------------------------\
% Try to start OLGA server
%---------------------------------------------------------------------------/
errmsg=1;
while(errmsg~=0)
    dos('start /low olga-5.1.exe -server olga2000');
    [errmsg]=OLGAConnect('localhost olga2000');
end


%---------------------------------------------------------------------------\
% Starts the olga-server using input-file - defining variables
%---------------------------------------------------------------------------/
[errmsg] = OLGAStart('slug_modell_EH.geninp');
[errmsg] = OLGASetTrendVar('$BHP PT PTMAX PTMIN $UCV PT PTMAX PTMIN $CHOKE VALVOP');
[errmsg] = OLGASetInputVar('$WATER SOURCE $AIR SOURCE $CHOKE-CONTROLLER CONTROLLER', [5 5 1]);


%---------------------------------------------------------------------------\
% Define inputs for simulation
%---------------------------------------------------------------------------/
```

```
WaterGasfraction=0;
WaterMassflow=liste_waterfeed_alstad(startoftable);
WaterPressure=0; %is ignored later, and is therefore set to zero (0)
WaterTemperature=20;
WaterWaterfraction=0;

AirGasfraction=1;
AirMassflow=liste_gasfeed_alstad(startoftable);
AirPressure=0; %is ignored later, and is therefore set to zero (0)
AirTemperature=20;
AirWaterfraction=0;

WaterSourceTotal=[WaterGasfraction WaterMassflow WaterPressure...
WaterTemperature WaterWaterfraction];
AirSourceTotal=[AirGasfraction AirMassflow AirPressure AirTemperature AirWaterfraction];

inpdata=[WaterSourceTotal AirSourceTotal ValveOpening];

%------------------------------------------------------------------\
% Start simulating
%------------------------------------------------------------------/
for jj=1:length_simulation
    [t, y, p, errmsg] = OLGASimStep(T(jj),inpdata);

%--------------------------------------------------------\
% Save data for later use.
%--------------------------------------------------------/
    pressure_RB_index=OLGAVariableIndex(y,'BHP','PT');
    pressure_UCV_index=OLGAVariableIndex(y,'UCV','PT');
    choke_actual_opening_index=OLGAVariableIndex(y,'CHOKE','VALVOP');

    data_BHP(jj)=y(pressure_RB_index).Data/1e5-atm; %[barG]
    data_UCV(jj)=y(pressure_UCV_index).Data/1e5-atm; %[barG]
    data_actual_choke(jj)=y(choke_actual_opening_index).Data;
    data_t(jj)=t;
    data_valveopening(jj)=ValveOpening;

%--------------------------------------------------------\
% Lowpass filtering the valve opening
%--------------------------------------------------------/
    if(jj==1)
        valve_lowpass(jj)=data_actual_choke(jj); %
    else
        valve_lowpass(jj)=alpha*data_actual_choke(jj)+(1-alpha)*valve_lowpass(jj-1);
    end

%--------------------------------------------------------\
% Turn gain-scheduling controller on if time greater than
```

```
% start_controller_at_point
%------------------------------------------------------/
    if(jj>=start_controller_at_point)

%----------------------------------------------------------------------\
%Finding the controller_gain as a function of the filtered valve opening
%----------------------------------------------------------------------/
        controller_K(jj-start_controller_at_point+1)=return_gain(valve_lowpass(jj),...
        gain_at_valve_min,gain_at_valve_maks);

        g0=controller_K(jj-start_controller_at_point+1)*(1+Ts/(2*Ti));
        g1=-controller_K(jj-start_controller_at_point+1)*(1-Ts/(2*Ti));

        e_error=controller_R-data_BHP(jj);
        e_error_m1=controller_R-data_BHP(jj-1);
        ValveOpening=ValveOpening+g0*e_error+g1*e_error_m1;

%----------------------------------------------------------------------\
%Saturation
%----------------------------------------------------------------------/
        if(ValveOpening>1)
        u_out=1;
        elseif(ValveOpening<0)
            u_out=0;
        else
            u_out=ValveOpening;
        end

%----------------------------------------------------------------------\
%Anti-windup
%----------------------------------------------------------------------/
        if(ValveOpening>1)
            ValveOpening=1;
        elseif(ValveOpening<0)
            ValveOpening=0;
        else
            ValveOpening;
        end

%----------------------------------------------------------------------\
%Computing new inpdata-file for simulation in next iteration
%----------------------------------------------------------------------/
        inpdata=[WaterSourceTotal AirSourceTotal u_out];

    else

%----------------------------------------------------------------------\
%If the controller is not turned on, use the same valve opening in inputfile
```

```
%--------------------------------------------------------------------/
        ValveOpening=ValveOpening_start;
        inpdata=[WaterSourceTotal AirSourceTotal ValveOpening];
    end

end

%--------------------------------------------------------------------\
%Disconnect the OLGA-server
%--------------------------------------------------------------------/
[errmsg]=OLGADisconnect;
```

Appendix B

Discussion of when to start the controller

B.1 Introduction

In this section it will be shown three slightly different simulation studies. All of the simulations consider a setpoint for the riser base pressure equal to 0.6 barg. This corresponds to a valve opening of 30%. It will be shown that it matters if the controller is turned on at a stable valve opening (13%), or at an unstable one (30%). If the controller is turned on when slug flow has already developed, it also matters where in the slug cycle it is turned on. The time when the controller is turned on is indicated by a black vertical line in the plots.

The controller is implemented in the incremental form shown in appendix A. The initial value of the controller is set to the same valve opening as the valve opening the controller is started from (13% and 30% respectively).
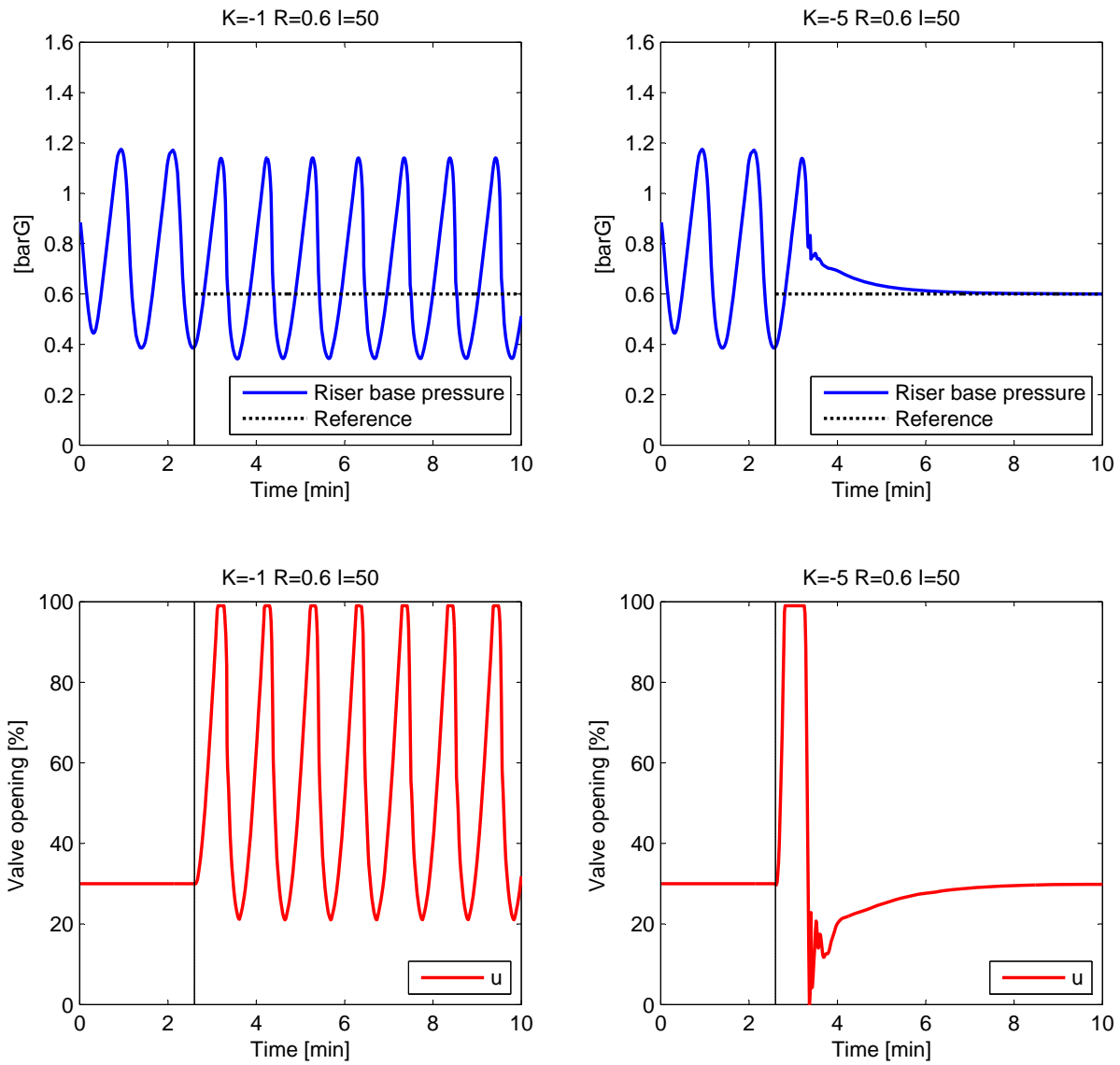
B.2   Simulations

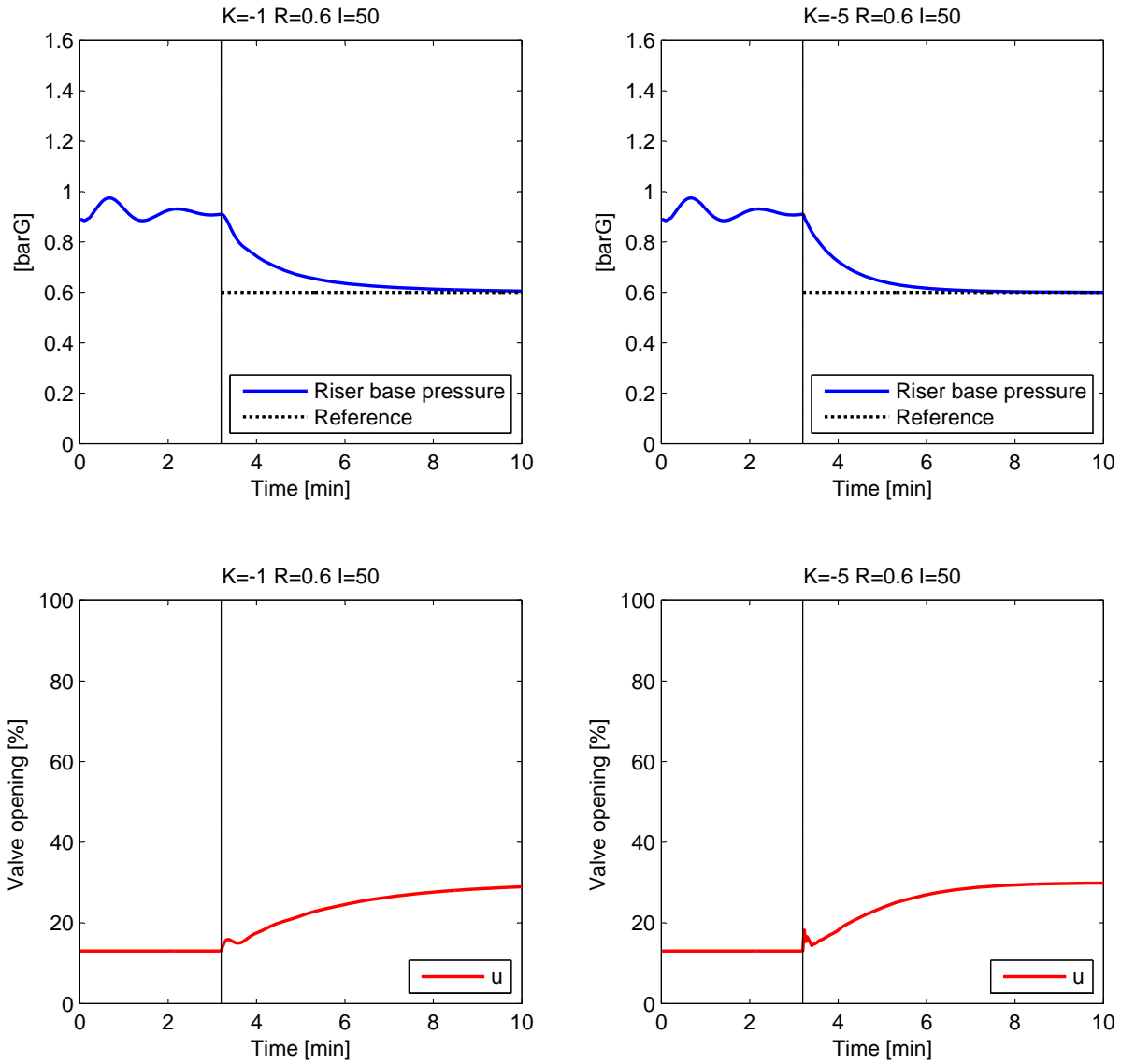### B.2.1   Reference $= 0.6$ barg - controller turned on at maximum pressure
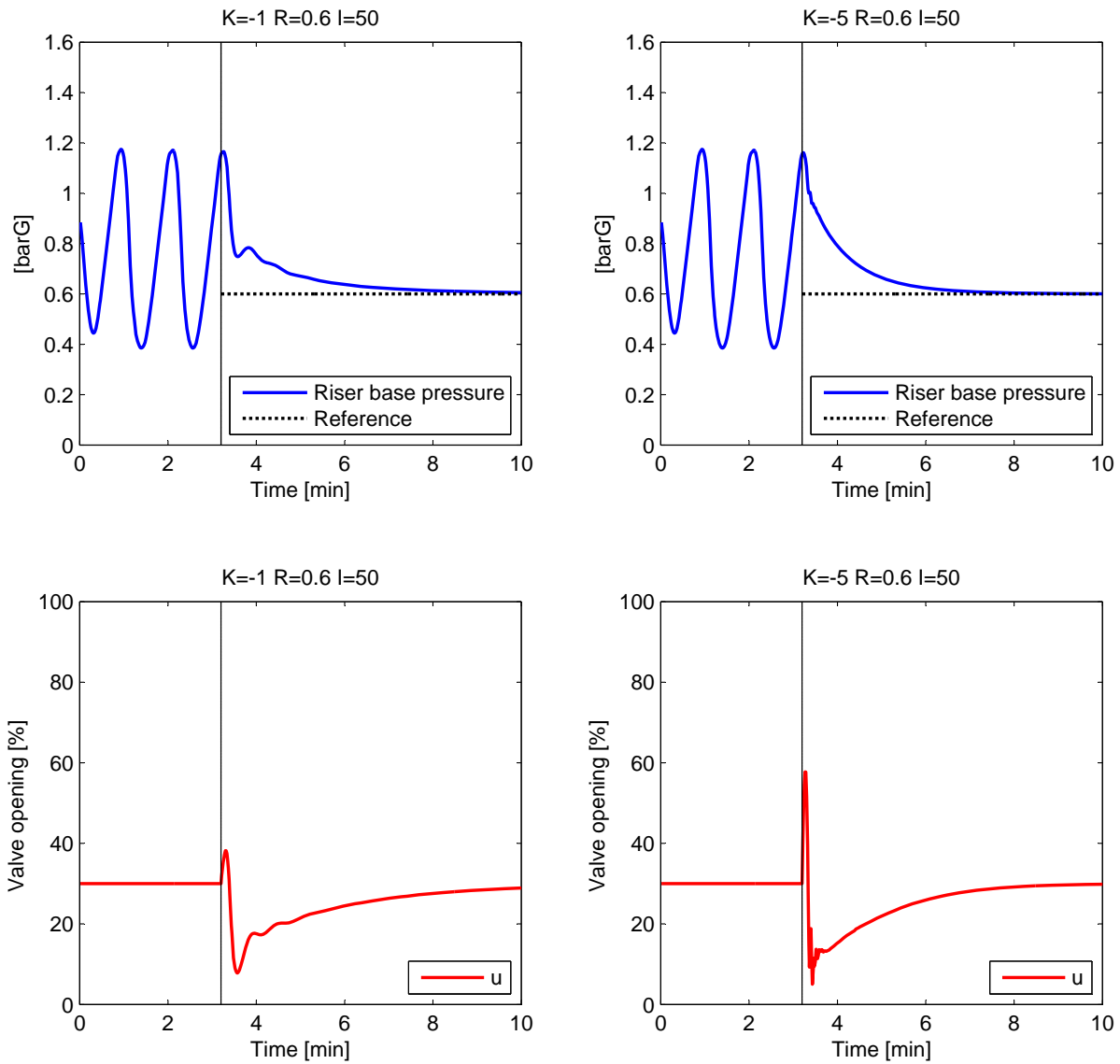
Figure B.1: Reference = 0.6 barg. The controller is turned on at the maximum pressure in the slug cycle

B.2.2 Reference $= 0.6$ barg - controller turned on at minimum pressure

Figure B.2: Reference = 0.6 barg. The controller is turned on at the minimum pressure in the slug cycle

B.2.3   Reference $= 0.6$ barg - controller turned on from stable valve opening

Figure B.3: Reference = 0.6 barg. The controller is turned on at stable valve opening

B.3 Summary of these simulation studies

It is interesting to see that if the controller starts at a stable valve opening, or starts when the pressure in the slug cycle is at its maximum, the controller is able to stabilize the flow with a lower gain than what was expected.

This simulation study considered the case were the initial value of the controller was set to the valve opening the controller was started from. It has been discovered that whether the system can be stabilized with a low gain for a large valve opening also depends on this initial value. This will not be shown in detail, however. The important lesson from this simulation study is that to make sure the system is stabilized around a large valve opening, high gain is absolutely necessary.

Appendix C

# Discussion of the low-pass filter's impact on the OLGC algorithm

In the result chapter it was shown that the OLGC algorithm was able to stabilize the system with a time constant for the low-pass filter equal to 30 seconds. However, this algorithm is little robust with respect to this constant, as is shown in figure C.1 (Only simulations with a reference of 0.5657 barg are shown). The system goes unstable both at a time constant equal to 20 and 30 seconds. This method is therefore not investigated further, as the robustness properties of the algorithm are not satisfactory.
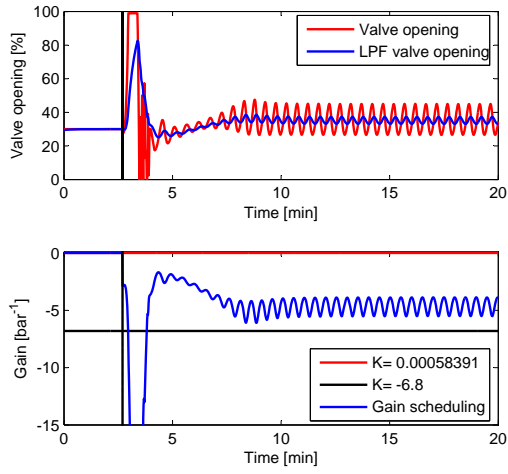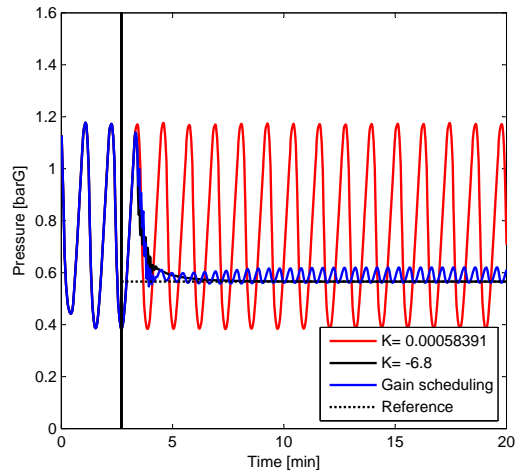
(a) Gain and valve (ζ) plot
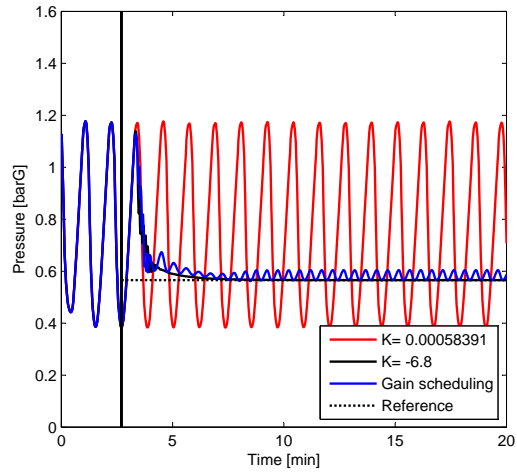
(b) Riser base pressure
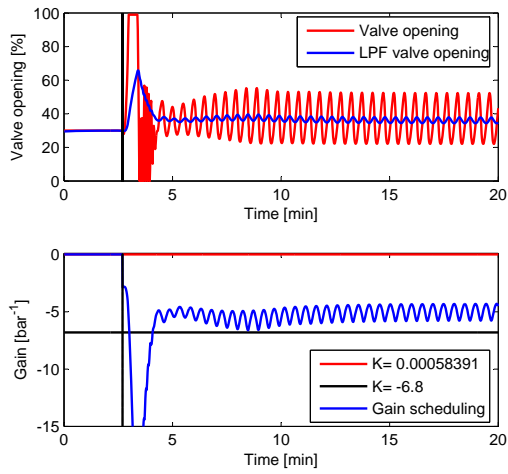
(c) Gain and valve (ζ) plot

(d) Riser base pressure

Figure C.1: a) and b) - T=20, reference 0.5657 barg. c) and d) - T=40 reference 0.5657 barg

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