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Control Structure and Tuning Method Design for suppressing Disturbances in a multi-phase Separator

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Abstract

In this project report the control structure for a three phase separator for a oil production unit is examined. The purpose is to analyse the control structure, and make the suggestion for how a control structure based on single loop control, can be designed to reject large disturbance occurring in the system. The second goal is to make a tuning method based on the plant properties to prevent process shutdown, that keeps the process running smooth. The assignment text states that the following tools can be used to develop the control structure: feed forward control, gain scheduling, first and second order filters, PI- and PID-controllers.

A process model containing all process variables is derived in the report. For verification purposes and development of control structure and tuning, the model is implemented in MATLAB and Simulink where simulations are done. The model is normalised such that the control structure is tested for the general case. Plant verification model is based on dimensions from a real plant, but process parameters are changed for comparing with other plants. Tools like gain scheduler, anti-windup, derivative filter and low pass filter are derived and designed for the purpose.

The report shows which parameters that limit the process performance and which techniques that should be used to reduce their impact to the system. There is also made analysis for how control loops and the controller can regard the limitations. An important part of this is the development of a signal filter to reduce measurement noise in the system.

As a basis for controller tuning SIMC is used. The system is analysed to find a method adapted for separator tuning. The customised method is tested for different plants and compared to SIMC to verify the abilities to control the plant. Analyses of the system, is done for a linear system, even though the system is nonlinear. To be able to use linear system techniques, analysis is done for the system linearised around setpoint and chosen points within the working area.

The result of the report, is a suggestion for how single loop control structure, and tuning can be performed for a separator based on the plant properties.

Sammendrag

Denne rapporten tar for seg reguleringsstrukturen til en trefaseseparator i et oljeproduksjonsanlegg. Formålet er å analysere strukturen for så å utvikle et forslag for hvordan reguleringsstrukturen, basert på enkeltsløyfer, kan designes for best mulig undertrykking av prosessforstyrrelse. Mål nummer to er å utvikle en optimaliseringsmetode, basert på anleggets egenskaper, som hindrer anleggsnedstengning og gir en rolig regulering. Oppgaveteksten oppgir as følgende reguleringsverktøy kan brukes til utvikling av reguleringsstrukturen: foroverkopling, gain scheduling, første og andre ordens filter, PI- og/eller PID-regulatorer.

En modell av prosessen som inneholder alle prosessvariable er utledet i rapporten. Modellen er implementert i MATLAB og Simulink for utvikling og verifisering av reguleringsstruktur og tuning. Modellen er normalisert slik at analysene kan gjøres for et generelt system. Dimensjonene som danner grunnlaget for verifikasjonen av resultatene, er hentet fra et eksisterende anlegg, men parameterne er endret for sammenlikning med andre anlegg. Verktøy som gain scheduler, anti-windup, filtering as regulatorens derivatledd og lavpassfiltrere er utviklet og designet til formålet.

Rapporten gir innblikk i hvilke forhold som gir begrensninger for systemet, hvilken påvirking de har på prosessen og hvilke metoder som kan brukes for å redusere begrensingene. Analyser viser også hvordan reguleringsstrukturen kan ta hensyn til systembegrensingene. En viktig del av dette er utvikling av signalfilter for å redusere effekten av støy i systemet.

Systemet er analysert for å finne en tilpasset metode for tuning der SIMC er brukt som basis. Tuningmetoden er testet på forskjellige separatorer og sammenliknet med SIMC for verifikasjon. Analysene er i hovedsak gjort for et lineært system, selv om systemet er ulineært. For å kunne bruke lineær systemteori er systemet linearisert rundt setpunkt og for flere punkt innen arbeidsområdet.

Resultatet av rapporten er et forsalg til hvordan reguleringsstrukturen for en separator bør utformes, samt en metode basert på prosessens egenskaper for tuning av regulatorene.

Preface

This master thesis is worked out by Matias Wilhelmsen at Norwegian University of Science and Technology, department of Engineering Cybernetics the spring of 2013. The report is the final result for the 10th semester project TTK4900 – Master Thesis.

In autumn 2012 I worked out a specialization project in cooperation with ABB AS section of Process Performance Solutions Integrated Operations / Oil, Gas & Petrochemicals. In the specialisation project the purpose was to investigate process behaviour and controller tuning, and suggest a controller for the oil level control that is able to reject any disturbances occurring for the oil phase in the separator. This project is an extension of this particular project. The purpose in this project is to examine all three phases to find a control structure and a tuning method for the plant. ABB AS has specified the task in cooperation with NTNU. In addition to be the principal, ABB AS has been an external part for help and solution verification.

Professor Lars Struen Imsland at NTNU has been the supervisor for the project and Ph.D Bjørnar Bøhagen has been the the external supervisor from ABB AS. I would like to thank both of them for good help, advices and high quality background material.

To be able to achieve full advantage of the report, the reader should have basic knowledge in cybernetics.

The following tools are used in the project work:

- LaTeX 2_ε with MikTex and TexStudio
- Microsoft Office Word, Excel, Visio and Project
- Microsoft Paint and Snipping Tool
- MathWorks MATLAB and Simulink
- Maplesoft Maple 14
- Adobe Acrobat X Pro

Trondheim, 5th June 2013

Matias Wilhelmsen

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Nomenclature

Table 0.1: Parameters and values

Symbol	Parameter	Value	Unit
A	Oil surface in separator		m^2
g	Acceleration of gravity	9.81	m/s^2
h	Separator liquid level		m
M	Molar gas constant		Mol
p	Separator pressure		Pa
R	Gas constant	8.3145	$\frac{J}{K \cdot Mol}$
T	Separator temperature		K
q_{i_w}, q_{o_w}	Inlet, outlet water volume flow		m^3/s
q_{i_o}, q_{o_o}	Inlet, outlet oil volume flow		m^3/s
w_{i_g}, w_{i_g}	Inlet, outlet mass flow rate of gas		kg/s
v	Process disturbance		m^3/s
ρ	Fluid density		kg/m^3
ω	Frequency		rad/s
ω_b	Filter bandwidth		rad/s
ω_c	Crossover frequency		rad/s
ω_t	Closed loop frequency at $-3[dB]$	$-3[dB]$	rad/s
ω_{180}	Phase crossing frequency	$-180°$	rad/s
τ	Time delay		s

Table 0.2: Variables

Symbol	Variable
d	Measurement noise / Diameter
C_v	Valve constant
h	Separator level
$h_a(s)$	Actuator transfer function
$h_f(s)$	Filter transfer function
$h_m(s)$	Transducer transfer function
$h_p(s)$	Process transfer function
$h_r(s)$	Controller transfer function
K_f	Filter gain
K_p	Controller gain
k	Desirable constant
k_g	Magnitude gas pressure loop
k_o	Magnitude oil level loop
k_w	Magnitude water level loop
$L(s)$	Loop transfer function
l	Total separator length
l_w	Separator water length
p	Separator pressure
r	Separator radius
$r(s)$	Process reference
T	Time constant
T_d	Derivative time
T_f	Filter time constant
T_i	Integral time
$T(s)$	Tracking transfer function / closed loop transfer function
u	Process input
$V_g(h)$	Function of gas volume
v	Disturbance
y	Process output / controlled variable
y_{h_o}	Normalised process output, oil level
y_{h_w}	Normalised process output, water level
y_p	Normalised process output, gas pressure
y_r	Normalised process reference
Z	Flashing constant
z	Controller manipulated value
α	Calculation variable for separator level
β	Calculation variable for separator width
$\beta(h)$	Function for cross-section calculation
ψ	Phase margin
ρ	Density
τ	Time delay

1 Introduction

This report is based on the task given in TTK4900 – Master-thesis given at NTNU in the spring of 2013. The project is a continuation of the TTK4551 – Specialization project – Control and tuning of a three phase separator with respect to disturbances. In this project we will examine all the three phases, derive a model of the process, suggest a control structure for the plant and develop a tuning method for the process.

To make the separation process work properly, water level, oil level and gas pressure should not be changed too fast. However changes have to be fast enough to reject any disturbance. Therefore well designed and over damped control loops is essential to the system. In this task we will pay attention to the control structure for each process variable and develop a structure that is able to reject large disturbances occurring in the system. For small disturbances the control structure should as far as possible keep the process rest.

Large disturbances can occur as a result of water plugs and gas bubbles, which is called slug, or from fast changes in the choke. Most frequently large disturbances in the separator, is caused by slug flow in the production pipe. The focus of the report is therefore on the separator control structures ability to handle slug flow.

Slug flow is in most cases a problem for oilfields that have been in production for several years. After some years in production the structure of the reservoir can be changed and lead to other production conditions. Water or gas penetration in the well may lead to changed flow and slug can be more frequently. Slug can also occur when production flow is led from one oil platform to another which is described in (Havre, Stornes & Stray 2000) or from hills and low-points in pipelines lying on the seabed which is outlined in (Jahanshahi, Skogestad & Helgesen 2012).

The desired outcome of the task contains a suggestion for control structure design. There is a guideline for the information needed for process tuning and suggestion for how it is done. The method is verified by analysis and simulations and compared to the SIMC tuning method.

1.1 The task

The purpose of the task is to suggest and analyse an appropriate control structure that prevent shut down for large disturbances. If there is impossible to find such structure that prevent shut down for large disturbances, the plant should be analysed to find the reason. There should also be made a tuning method for the control structure selected. The tuning method should be based on the separator properties. The tuning method has to make the controller able to reject any disturbance as well as make the process handling as calm as possible.

Suggested control structure should be based on single loops with PI or PID controllers. Measurements available is water level, oil level and gas pressure. Tools that can be used to make the control structure is: first or second order filters, feed forward between the loops and gain scheduled controller. Disturbance feed forward or cascade control where outflow is controlled in an inner loop is not possible to perform since there is no flow measurements in the plant.

The report contains analyses for how the structure can be designed and which control techniques that make the performance robust. The goal is to present a general procedure for control structure design and tuning to be able to prevent process shutdown if the plant is exposed to large disturbances. It should be able to perform the control structure and tuning both for existing plants and new constructions.

A model of the system is developed based on the separator geometry and the law of physics. ABB AS deliver dimensions from a real plant, but the values is normalised to make the conclusions general for separator processes. The purpose of the model is to test structures and tuning, simulate and verify the results of the analysis.

The tuning rule is based on known methods such as SIMC or Ziegler & Nichols method, but modifications is done to meet the control objectives of a separator. The process performance should prevent that waves and currencies occur in the separator. However the controller has to react sufficiently fast to prevent overflow for large disturbances.

Analysis of controllability and robustness is required for each control loop. It is also necessary to analyse how the process variables and control loops affect each other and which conditions the limits the process performance.

1.2 Earlier work

Three phase separators is an important plant in petroleum plants and it has been used as long as oil has been produced. Gas, oil and water separators are most important in all offshore oil and gas production facilities. All the large petroleum production companies like Shell, Exxon Mobil, British Petroleum, Statoil, etc. as well as the plant construction companies, works continuously with separator improvements.

In 2000 Kjetil Havre, Karl Ole Stornes and Henrik Stray wrote an article to ABB Review about slug flow in pipelines and how it affect the oil production. In this article several techniques to reduce the problem, based on research made by ABB, are presented. Benefits and drawback with different techniques are discussed, but no clear solution for every plant is presented.

At NTNU the three phase separator has been analysed in many master thesis by many disciplines. In 1998 Ragnhild Wilhelmsen analysed the fluid behaviour and the thermodynamic properties of the separator. Per Morten Hellervik did in 2002 an analyse of control structure and tuning of three phase separators in series in in collaboration with ABB AS, supervised by Professor Jan Tommy Gravdahl. In the autumn 2012 the writer of this report, analysed the control structure for oil level in a three phase separator.

Tor Steinar Schei, Peter Singstad and Aage Jostein Thunem examined, in 1991, transients in gas-oil-water separation plants. Their work covers the whole separation train, and examines the impact of disturbance form one separator to the other. In the article the disturbance from flow changes is discussed, which also is examined in this report. The article also examines disturbances from temperature change in such plants, which is not examined in this report.

1.3 Outline

This report is divided in 8 main parts. 7 Chapters and Appendix. Each Chapter has a main is main topic followed by sections where the topic is explained and deduced.

The report contains the following Chapters:

- Chapter 1: Introduction
- Chapter 2: Process description
- Chapter 3: The process model and aspects that limits the performance
- Chapter 4: Control structure design, controllers, filters and process tuning
- Chapter 5: Evaluation and process analysis
- Chapter 6: Results and further work
- Chapter 7: Conclusion

In the Appendix at the end of the report, relevant process plots, bode plots for the control loops, MATLAB code and Simulink diagrams is attached.

2 Process description

As the name says, separators are made to separate two or more products. For an oil production facility the products are gas, oil and water. Sand or other solid may also be included for some plants. Separators can be designed in many ways, dependent of the inlet flow component and the space available for the plant. Gravity based separators can be both lying and standing tanks. There may be one or more outlet for each product. In this report a separator for oil gas and water, constructed as a lying cylinder with one inlet flow and one flow outlet for each product, as shown in Figure 2.1, is examined.

The production separator her considered are based on the gravitational principle. It means that the product with highest density, in this case water, will sink to the bottom and the low density gas will rise to the top. The oil which will stay in the middle, flows over a wier and is taken out at the end of the separator.

At the inlet entrance, there is a slug catcher that reduces the effect of slug. Slug is large gas bubbles or liquid plugs that makes large disturbances to the process. The demister is designed to catch liquid droplets in the gas. Droplets in the gas can destroy the compressor taking out gas, and for later processing units it is important that the gas is as dry as possible. The purpose of the wier is to separate the oil inlet and the oil outlet chamber. At each liquid outlets a vortex breaker is mounted. The purpose of the vortex breaker is to prevent gas to escape through the liquid outlets if the level becomes low.

Product quality is directly dependent of the ability to keep the medium in rest. Therefore the control structure and the separator design should prevent waves and liquid currencies. Waves and liquid currencies generated form slug, or from too fast control, will blend gas, oil and water. This means that the outlet oil flow will contain more water, the water outflow will contain more oil and the liquids will secrete less gas.

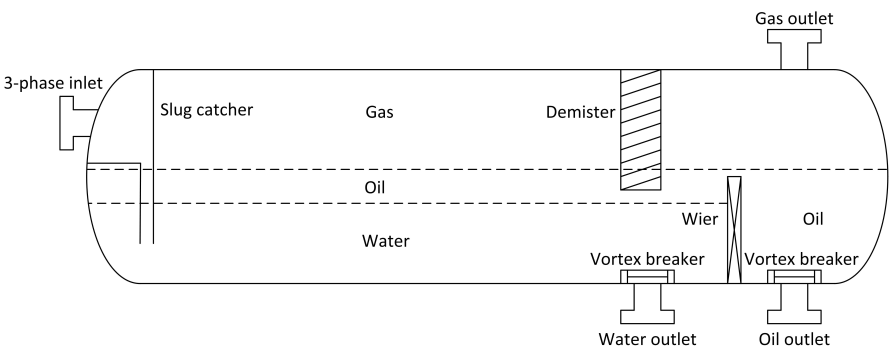

Figure 2.1: Drawing of one type of the structure inside a three phase separator constructed as a lying cylinder. Space for water, oil and gas is marked.

There are three control loops for system control, one for each process variable. In general there are level measurements for water and oil and pressure measurement for gas. For special plants there may be flow measurements for inlet and outlets. A piping and instrumentation diagram is shown in Figure 2.2 (Utdanningsdirektoratet 2011).

The main control loops are the oil level control loop and the gas pressure control loop (Devold 2010). Though in this report the water level is derived as well, and it is considered as important as the other control loops. The reason why water level is seen that important is that it influence oil level directly and gas pressure as well.

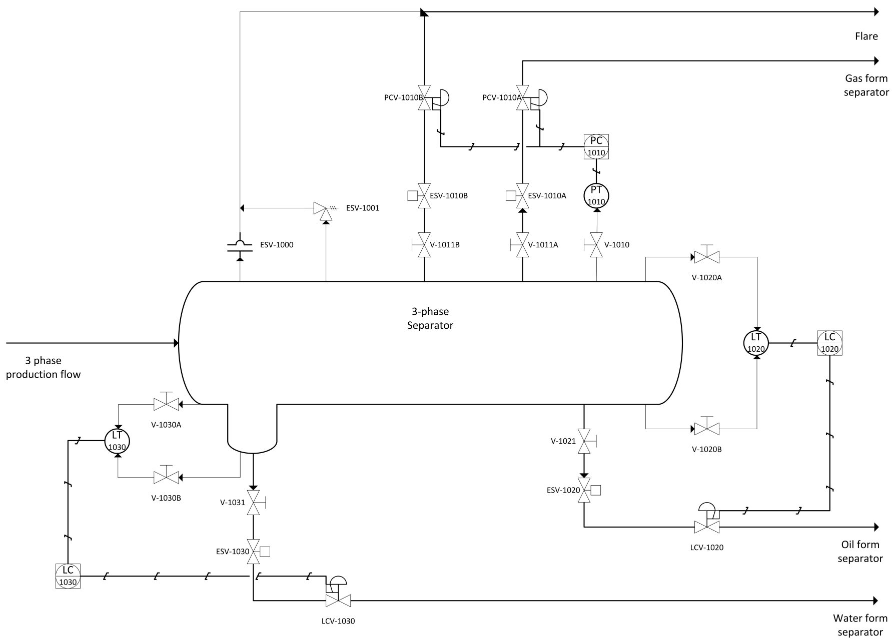

Figure 2.2: Piping and instrumental diagram for the three phase water, oil and gas separation process. The diagram contains the control loops for each process variable, illustrated as single loop feedback control structure. Differential pressure transducers for oil and water level and a pressure transducer for gas pressure are used for process measurement.

3 The process model and process limitations

Before the control structure can be developed, we have to know the process and which elements that limits the performance. In the first part of this Chapter we will derive the process model. It is shown simplifications which is done for model analysis. The second part give a description of which conditions that limits the process performance, and what it does to the separator process.

3.1 The process model

The analysed separator process contains three measurable process variables: gas pressure, oil level and water level. The model for each state is based law of physics and the separator geometrical shape. Sayda & Taylor (2007) and Schei, Singstad & Thunem (1991) are articles where similar separator models are derived, and those articles are used as basis for model in this report.

In the model flashing is not regarded, but it is show in the mass balance equation. For the process derived in this task net inflow is the same as net outflow for each component. In the mass balance model flashing is denoted as a factor Z. In the analysis done in Schei et al. (1991), transients due to flashing is discussed.

3.1.1 Separator geometry

Often a separator is constructed as a lying cylinder, which is the case for the analysed plant. In (Wilhelmsen 2012) the section dedicated for oil is assumed to have plain walls. For the considered separator in this report, the separator geometry is regarded, but the end sections are regarded to be plain walls. For a short separator with large radius, the end sections that in most cases is curved, should be taken into account.

Figure 3.1 shows a sketch for how the cross-section of the liquid level is calculated. Equations for calculating the liquid cross-section and calculation of a area of a circle segment. The cylinder radius is given by $r = \frac{d}{2}$.

We know that the liquid surface changes as a function of the liquid level. The width of the liquid surface can be calculated by the Pythagoras equation given by Equation (3.1). Half the width is denoted by β, radius by r, liquid level by h and space between liquid level and the tank middle by α (Kristensen 2008). Figure 3.1 shows a sketch of the separator cross-section showing the parameter objectives.

$$r^2 = \alpha^2 + \beta^2 \tag{3.1}$$

$$\beta^2 = r^2 - \alpha^2$$

$$\beta = \sqrt{r^2 - \alpha^2} \tag{3.2}$$

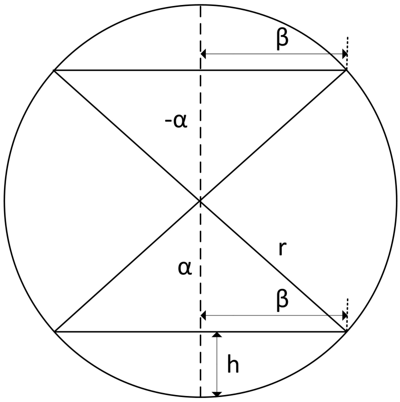

Figure 3.1: Sketch showing the cross-section of the liquid area, seen from the separator end.

In Equation (3.2) we change α to $r - h$ to make β as a function of h shown in Equation (3.3) and Equation (3.4).

$$\beta(h) = \sqrt{r^2 - (r-h)^2} \tag{3.3}$$

$$\beta(h) = \sqrt{2rh - h^2} \tag{3.4}$$

It seems like there will be some problems when the liquid level rises above the separator midpoint. In those cases α will be negative which is not legal in the calculations. However this is solved by symmetrical considerations. A prof showing that the calculations work despite negative α will not be shown here.

By the formula for the liquid surface width, we can find the area of the cross-section. It is found by integrating the line segments from the separator bottom $h = 0$ to the liquid level. The area of the liquid cross-section is called $A(h)$.

$$A(h) = 2 \cdot \int_0^h \sqrt{2rh - h^2} dh'$$

$$A(h) = 2 \cdot \int_0^h \beta(h') dh' \tag{3.5}$$

To find the volume of the liquid, the cross section area is multiplied with the length l of the area dedicated for each liquids. Calculations from Equation (3.6) show the derivative of volume. Tank construction and dimensions for volume calculation are shown in Figure 3.2.

$$
\begin{aligned}
V(h) &= 2l \cdot \int_0^h \beta(h')dh' \\
\frac{dV}{dt} &= \frac{d}{dt}(2l \cdot \int_0^h \beta(h')dh') \\
\frac{dV}{dt} &= 2l \cdot \frac{\partial}{\partial h}(\int_0^h \beta(h')dh') \cdot \frac{\partial h}{\partial t} \\
\frac{dV}{dt} &= 2l \cdot (\beta(h) - \beta(0)) \cdot \frac{\partial h}{\partial t}
\end{aligned} \tag{3.6}
$$

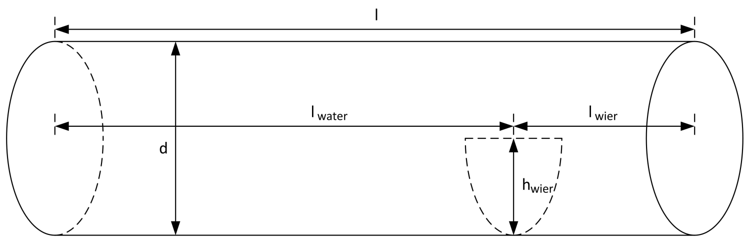

Figure 3.2: Simplified sketch of the separator tank, showing the objectives inside the tank.

For small differences in height in the middle of the tank, the width of the liquid surface will only make minor changes. It means that for small level changes the separator can be assumed to have plain walls. However for large changes we have to take into account the changes in width provided by the separator shape.

3.1.2 Water level model

The liquid level can be derived from mass balance (Egeland & Gravdahl 2003) (Sayda & Taylor 2007). From mass the balance in Equation (3.7) the change in water level is derived. We look at the water mass flow admitted to the process w_{i_w} and the water mass flow leaving the process w_{o_w}. Flashing is denoted as Z_w.

$$
\begin{aligned}
\frac{dm_w}{dt} &= w_{i_w}(t) - w_{o_w}(t) - Z_w \\
\frac{dV_w}{dt}\rho_w &= \rho_{w_i} q_{i_w}(t) - \rho_{w_o} q_{o_w}(t) - Z_w
\end{aligned} \tag{3.7}
$$

The water density ρ_w can be assumed constant, which leads to constant volume. Net water volume flow $(q_{i_w}(t) - q_{o_w}(t))$ is called $q_w(t)$. Since flashing is not regarded in the analysis, Z_w is not used in further calculations.

$$\begin{aligned}
\frac{dV_w}{dt} &= q_{i_w}(t) - q_{o_w}(t) \\
\frac{\partial}{\partial h_w}(2l_w \cdot \int_0^{h_w} \beta(h')dh') \cdot \frac{dh_w}{dt} &= q_{i_w}(t) - q_{o_w}(t) \\
\frac{dh_w}{dt} &= \frac{1}{\frac{\partial}{\partial h_w}(2l_w \cdot \int_0^{h_w} \beta(h')dh')} q_w(t) \\
\frac{dh_w}{dt} &= \frac{1}{2l_w(\beta(h_w) - \beta(h_0))} \cdot q_w(t)
\end{aligned} \tag{3.8}$$

The differential equation given in Equation (3.8) denotes the change in water level. For analyse purposes it is desirable to transform the model into the frequency domain, given in calculations from Equation (3.9). Since our model is nonlinear we have to linearise the model before we make a transfer function. It is desirable linearise around the water level setpoint $h_{w_{sp}}$.

$$\mathcal{L}\{\frac{dh_w}{dt}\} = \mathcal{L}\{\frac{1}{2l \cdot \beta_w(h_{w_{sp}})}(q_w(t))\} \tag{3.9}$$

$$sh_w(s) = \frac{q_w(s)}{2l_w \cdot \beta_w(h_{w_{sp}})}$$

$$h_w(s) = \frac{q_w(s)}{2l_w \cdot \beta_w(h_{w_{sp}})s} \tag{3.10}$$

When analysis is carried out, linearisation around various values of $B(h)$ is required for the nonlinear system. There may be some unwanted effects of nonlinearity in the system.

3.1.3 Oil level model

Now the model for the oil level will be derived. We consider the oil level h_o relative to the water level h_w. First the available oil volume in the separator is derived.

From Figure 2.1 and Figure 3.2 we se that the oil volume is divided into three parts which is given mathematical in Equation (3.11). The space between water level and top of wier has a length of l_w. From wier top to the oil surface has a length of l and behind the wier from separator bottom to the top of wier has a length of $l - l_w$. In our model the oil level is not allowed to be below the top of wier. That means the oil which is in the chamber below the weir and in the outflow side can be regarded as a constant volume. It is known that oil level is dependent of water level.

$$
\begin{aligned}
V_o(h_w, h_o) &= V(h_w) + V(h_o) + V_{wier} \\
\frac{dV_o(h_w, h_o)}{dt} &= \frac{\partial V_o}{\partial h_w} \cdot \frac{dh_w}{dt} + \frac{\partial V_o}{\partial h_o} \cdot \frac{dh_o}{dt} \\
\frac{dV_o(h_w, h_o)}{dt} &= \frac{\partial V_{o_1}}{\partial h_w} \cdot \frac{dh_w}{dt} + \frac{\partial V_{o_2}}{\partial h_o} \cdot \frac{dh_o}{dt}
\end{aligned} \tag{3.11}
$$

We divide the derivative of oil volume into volume below wier V_{o_1} and volume above wier V_{o_2}.

$$
\begin{aligned}
V_{o_1} &= 2l_w \cdot \int_{h_w}^{h_{wier}} \beta(h')dh' \\
\frac{\partial V_{o_1}}{\partial h_w} &= 2l_w \frac{\partial}{\partial h_w}(B(h_{wier}) - B(h_w)) \\
\frac{\partial V_{o_1}}{\partial h_w} &= 2l_w \cdot (-\beta(h_w))
\end{aligned} \tag{3.12}
$$

$$
\begin{aligned}
V_{o_2} &= 2l \cdot \int_{h_{wier}}^{h_o} \beta(h')dh' \\
\frac{\partial V_{o_2}}{\partial h_o} &= 2l \cdot \frac{\partial}{\partial h_o}(B(h_o) - B(h_{wier})) \\
\frac{\partial V_{o_2}}{\partial h_o} &= 2l \cdot \beta(h_o) \cdot
\end{aligned} \tag{3.13}
$$

The change in oil level is derived from mass balance given in Equation (3.14). Flashing is denoted as Z_o.

$$
\begin{aligned}
\frac{dm_o}{dt} &= w_{i_o}(t) - w_{o_o}(t) - Z_o \\
\frac{dV_o}{dt}\rho_o &= \rho_o(q_{i_o}(t) - q_{o_o}(t)) - Z_o
\end{aligned} \tag{3.14}
$$

Oil density ρ_o is assumed constant and we can therefore do the rest of calculations for volume flow. Flashing Z_o is not regarded in the analysis and is not used in later calculations. Net oil flow is called $q_o(t)$ and is $q_{i_o}(t) - q_{o_o}(t)$.

Oil volume given in Equation (3.12) and Equation (3.13) put together with mass balance, leads to the model for oil level change given in Equation (3.15).

$$
\begin{aligned}
\frac{dV_o}{dt} &= q_{i_o}(t) - q_{o_o}(t) \\
\frac{\partial V_{o_1}}{\partial h_w} \cdot \frac{dh_w}{dt} + \frac{\partial V_{o_2}}{\partial h_o} \cdot \frac{dh_o}{dt} &= q_{i_o}(t) - q_{o_o}(t) \\
\frac{\partial V_{o_2}}{\partial h_o} \cdot \frac{dh_o}{dt} &= q_{i_o}(t) - q_{o_o}(t) - \frac{\partial V_{o_1}}{\partial h_w} \cdot \frac{dh_w}{dt} \\
\frac{dh_o}{dt} &= \frac{q_{i_o}(t) - q_{o_o}(t) - \frac{\partial V_{o_1}}{\partial h_w} \cdot \frac{dh_w}{dt}}{\frac{\partial V_{o_2}}{\partial h_o}} \qquad (3.15)
\end{aligned}
$$

As for the water level, we transform the system into the frequency domain for analyse purposes. To be able to make a transfer function we have to linearise the system given in Equation (3.15), around a working point. For simplicity the water level is assumed constant with a magnitude of k_w at a working point when analysis is performed. Equation (3.16) denotes the linearised oil level model in time domain.

$$
h_o(s) = \frac{q_O(s) - (\beta(h_{w_{sp}}) \cdot k_w)}{2l(\beta h_{o_{sp}}))s} \qquad (3.16)
$$

3.1.4 Model of the separator gas pressure

The gas pressure in the separator is dependent of volume available for gas, mass balance and gas density. In fact it is also dependent on temperature and flashing but it is not regarded here even though flashing is shown in the mass balance equation. The gas pressure model will now be derived, and we start by defining the volume available for gas, and the derivative of gas volume.

$$
\begin{aligned}
V_g(h_o) &= 2 \cdot l \cdot \int_{h_o}^{2r} \beta(h')dh' \qquad (3.17) \\
\frac{\partial V_g}{\partial h_o} &= \frac{\partial}{\partial h_o}(B(2r) - B(h_o)) \cdot 2l \\
\frac{\partial V_g}{\partial h_o} &= -\beta(h_o) \cdot 2l
\end{aligned}
$$

$$
\begin{aligned}
\frac{dV_g}{dt} &= \frac{\partial V_g}{\partial h_o} \cdot \frac{dh_o}{dt} \\
\frac{dV_g}{dt} &= -2l \cdot \beta(h_o) \cdot \frac{dh_o}{dt}
\end{aligned}
$$

Since the volume is not constant, we can not calculate the process as a volume balance. Though the process can be modelled as a mass balance. As a basis for our model we use the ideal gas law given in Equation (3.18) and mass balance given in Equation (3.19). Flashing can be taken into account if the factor Z_g is added to the mass balance given in Equation (3.19). However in this case it is not used in further calculations.

$$pV_g = m_g RT \tag{3.18}$$
$$\frac{dm_g}{dt} = w_{i_g}(t) - w_{o_g}(t) - Z_g \tag{3.19}$$

The gas density is not constant, therefore a differential equation for gas density is needed.

$$\begin{aligned} m_g &= \rho_g V_g \\ pV_g &= m_g RT \\ p &= \rho_g RT \\ \frac{dp}{dt} &= \frac{d\rho_g}{dt} RT \end{aligned}$$

$\frac{d}{dt}(RT) = 0$ and therefore the derivative of pressure is as given in Equation (3.20)

$$\frac{d\rho_g}{dt} = \frac{1}{RT} \cdot \frac{dp}{dt} \tag{3.20}$$

Now the derivative of mass is derived. The temperature is assumed constant in the separator. The final model for change in gas pressure is given by Equation (3.21).

$$\begin{aligned} \frac{dm_g}{dt} &= \frac{d\rho_g}{dt} V_g + \frac{dV_g}{dt} \rho_g \\ \frac{dm_g}{dt} &= \frac{V_g}{RT}\frac{dp}{dt} + \rho \frac{dV_g}{dt} \\ \frac{dm_g}{dt} &= \frac{V_g}{RT}\frac{dp}{dt} + \frac{p}{RT}\frac{dV_g}{dt} \\ w_{i_g}(t) - w_{o_g}(t) &= \frac{V_g}{RT}\frac{dp}{dt} + \frac{p}{RT}\frac{dV_g}{dt} \\ \frac{dp}{dt} &= RT(\frac{w_{i_g}(t) - w_{o_g}(t)}{V_g}) - p\frac{\frac{dV_g}{dt}}{V_g} \end{aligned} \tag{3.21}$$

3.2 Actuator models

A control valve is an actuator that reduces a flow by mechanically reduce the valve cross section, dependent of the control signal. The model for at frictionless and incompressible flow though a restriction is given by Equation (3.22) (Egeland & Gravdahl 2003).

$$q(z,p) = C_v f(z) \sqrt{\frac{2\Delta p}{\rho(1 - (\frac{A_2}{A_1})^2)}} \tag{3.22}$$

The valve opening A_2 is dependent of the manipulated value from the controller z. When the area of the pipe A_1 becomes much bigger than valve areal A_2, we can simplify the flow though a valve to be given by Equation (3.23).

$$q_o = C_v f(z) \sqrt{\frac{2\Delta p}{\rho}} \tag{3.23}$$

When the gas pressure is close to the setpoint we can make $C_v \sqrt{\frac{2\Delta p}{\rho}}$ to a constant k_v.

Large valves often have both time constant and time delay. For this plant, the given values for dead band is $\tau = 5$ seconds and time constant is $T = 6$ seconds for liquid valves. Time constant is modelled as first order dynamic. It is assumed that both valves and pipes are frictionless. The Equation (3.24) gives the valve equation in the time domain for the nonlinear valve linearised around a working point z_{wp} and gas pressure at setpoint p_{sp}.

$$\begin{aligned} h_a(s) &= \frac{1}{T_a s + 1} e^{-\tau s} \\ q_o(s) &= \frac{k_v f(z_{wp})}{Ts + 1} e^{-\tau s} \\ h_a &= \frac{q}{Ts + 1} e^{-\tau s} \end{aligned} \tag{3.24}$$

The nonlinear term $f(z)$ is the equal percentage term given in Equation (3.25)

$$\begin{aligned} f(x) &= \alpha^{x-1} \cdot 100 \\ \alpha &= 35 \\ x &= \frac{z}{100} \end{aligned} \tag{3.25}$$

In the model the valve acts like a low pass filter with a nonlinear gain and time delay. However it can not be used as a low pass filter since small signal changes will tear the valve and make leakage and uncertain opening, which means that the valve requires a signal with little noise.

3.2.1 Water valve model

The basis for water valve is the equation for flow though a restriction given in Equation (3.26). In the equation $C_v A_{v_w} \sqrt{\frac{2}{\rho_w}}$ is a constant which is called k_{v_w}. The equal percentage nonlinearity for valve opening is given by Equation (3.27).

$$q_w = C_v f(z_w) \sqrt{\frac{2}{\rho_w}(p_1 - p_2)} \tag{3.26}$$

$$q_w = k_{v_w} f(z_w) \sqrt{(p_1 - p_2)}$$

$$f(z) = 35^{\frac{z}{100}-1} \tag{3.27}$$

We obtain the equation for output flow dynamics in the time domain by combining Equation (3.26) with the first order time constant plus delay given in Equation (3.28). T_{a_w} is the water valve time constant.

$$q_{w_o}(z_w) = C_v f(z_w) \sqrt{\frac{2}{\rho_w}(p_1 - p_2)}$$

$$h_{a_w}(s) = \frac{1}{T_a s + 1} e^{-\tau s} \tag{3.28}$$

$$h_{a_w}(s) = \frac{q_{w_o}}{T_{a_w} s + 1} e^{-\tau s} \tag{3.29}$$

The output flow will be dependent of separator pressure. Large pressure differences will make the flow rate change. A simplification done here is only to take separator gas pressure into account. Actually the liquid column generates pressure but it is small compared to the gas pressure. Therefore pressure generated by the liquid column is neglected.

3.2.2 Oil valve model

The basis for the oil valve is the same as for the water valve Equation (3.30). For the oil valve the constant value for $C_v f(z) \sqrt{\frac{2}{\rho_o}}$ calculated to one value called k_{v_o}.

$$
\begin{aligned}
q_o &= C_v f(z_o)\sqrt{\frac{2}{\rho_o}(p_1 - p_2)} \\
q_o &= k_{v_o} f(z_o)\sqrt{(p_1 - p_2)}
\end{aligned}
\tag{3.30}
$$

As well as for the water valve the equation for oil flow dynamics in the time domain, is obtained by combining Equation (3.30) and Equation (3.28) and linearising the model around a working point.

$$
\begin{aligned}
h_{a_o}(s) &= \frac{C_v f(z_{o_{sp}})\sqrt{\frac{2}{\rho_o}(p_1 - p_2)}}{T_{a_o}s + 1} e^{-\tau s} \\
h_{a_o}(s) &= \frac{q_{o_o}}{T_{a_o}s + 1} e^{-\tau s}
\end{aligned}
\tag{3.31}
$$

The difference between the oil valve and the water valve is the size of the restriction, the time constant and the valve time delay. Also for the oil valve the pressure generated by the liquid column is neglected.

3.2.3 Gas actuator model

The gas output can be controlled by different actuators in different plant. It may be a compressor or a valve. In this task we use a valve with a first order characteristic but without time delay. However the time constant is large for the used valve.

$$
q_{g_o} = k_{v_g} \cdot \sqrt{(p_1 - p_2)} \tag{3.32}
$$

The gasflow out of the valve is obtained when the flow equation in Equation (3.32) is put together with the first order dynamic in Equation (3.33).

$$
h_a(s) = \frac{1}{T_a s + 1} \tag{3.33}
$$

$$
h_{a_g}(s) = \frac{k_{v_g} \cdot \sqrt{(p_1 - p_2)}}{T_{a_g}s + 1}
$$

$$
h_{a_g}(s) = \frac{q_{g_o}}{T_{a_g}s + 1} \tag{3.34}
$$

There is not modelled any time delay for the gas actuator. There may be some delay but it is small compared to the time constant. If we had used a compressor, the time constant could have been as long as 2 minutes, for the used valve $T_{a_g} = 60$ seconds for the given case.

3.3 System scaling

To make the model analysis simpler and to make the analysis more general, the system models should be scaled. An other positive impact is the the engineer has to make judgement about the required performance of the system (Skogestad & Postlethwaite 2005).

Skogestad & Postlethwaite (2005) suggests the following method for SISO system normalization:

$$\hat{y} = \hat{h}\hat{u} + \hat{h_d}\hat{d}; \qquad \hat{e} = \hat{y} - \hat{r} \tag{3.35}$$

The system is given by the Equation (3.35) where hat ($\hat{}$) denotes the reel value. A common method for variable scaling is to make the values be between zero and one. This is done by in the suggested method in Skogestad & Postlethwaite (2005).

$$y = \frac{\hat{y}}{\hat{e}_{max}}, \qquad r = \frac{\hat{r}}{\hat{e}_{max}}, \qquad e = \frac{\hat{e}}{\hat{e}_{max}}$$
$$d = \frac{\hat{d}}{\hat{d}_{max}}, \qquad u = \frac{\hat{u}}{\hat{u}_{max}}$$

- $\hat{d}_{max}$ - maximum disturbance value
- $\hat{u}_{max}$ - largest allowed input value
- $\hat{e}_{max}$ - largest allowed control value
- $\hat{r}_{max}$ - largest expected change in reference

For the separator model this method is used, and the signals is normalized between $[0, 1]$. The signals could have been between $[0, 100]$ to describe the process values in percent, this is done by multiplying all normalized signals by 100.

3.3.1 Model normalisation

For one special case, we may use real dimensions and flows. In most cases however, measurements already are scaled by the control system, and therefore it makes sense to make a normalised model based on scaled measurements. For system normalisation some normalisation constants a and b is used to describe liquid levels and gas pressure and. A constant C to describe flow to the separator. The normalised process variable is denoted y, and the normalised setpoint is denoted y_r.

From the equations derived from Equation (3.36) the relation between liquid level or gas pressure and the constants a and b are given.

$$\begin{aligned}
y_h &= ah + b \\
\frac{dy_h}{dt} &= a\frac{dh}{dt} \\
\frac{dh}{dt} &= \frac{\frac{dy_h}{dt}}{a} \\
h &= (y - b)\frac{1}{a}
\end{aligned} \tag{3.36}$$

Flow equations

Also the inflow q_i and the outflow q_o have to be normalised. Outflow is normalized with a constant C for the valve flow, and it is given by Equation (3.37). Inflow is given by Equation (3.38).

$$q_o = Cf(z)\sqrt{\frac{\Delta p}{\rho}} = Cf(z)\sqrt{\frac{\Delta p_{nominal}}{\rho}}\sqrt{v_p} \tag{3.37}$$

$$q_i = C\sqrt{\frac{\Delta p_{nominal}}{\rho}}v_q \tag{3.38}$$

$C\sqrt{\frac{\Delta p_{nominal}}{\rho}}$ represents the maximum valve flow at nominal pressure. Percentage deviation from differential pressure is given by $v_q \in [0, \rightarrow]$, and $v_p = 1$ gives the nominal pressure. Divination in inflow is given by $v_q \in [0, \rightarrow]$. Equal percentage valve opening is given by $f(z) \in [0, 1]$, where $z \in [0, 1]$ is the controller manipulated value.

$$\begin{aligned}
q &= q_i - q_o \\
q &= C(\sqrt{\frac{\Delta p_{nominal}}{\rho}}v_q - f(z)\sqrt{\frac{\Delta p_{nominal}}{\rho}}\sqrt{v_p}) \\
q &= C(v_q - f(z)\sqrt{v_p})
\end{aligned} \tag{3.39}$$

The normalised inflow outflow representation is given by Equation (3.39)

Normalised water level model

The normalisation constants, a_w given in Equation (3.41) and b_w given by Equation (3.42), are used to scale the system . In the calculations y_{h_w}, given by Equation (3.40), is the normalised water level, which is derived from actual water level h_w.

$$y_{h_w} = a_w h_w + b_w \tag{3.40}$$
$$a_w = \frac{y_{w_2} - y_{w_1}}{h_{w_2} - h_{w_1}} \tag{3.41}$$
$$b_w = y_{h_w} - a_w h_{w_0} \tag{3.42}$$

Now we resume the water level model given in Equation (3.8). The normalised model can be described as a function $c_w(\cdot)$. When violating different separators, the magnitude of the function differs, or the factors are weighted different.

$$\frac{dh_w}{dt} = \frac{1}{2 \cdot l_w \cdot (\beta(h_w) - \beta(h_0))} \cdot (q_{i_w} - q_{o_w})$$
$$\frac{dy_{h_w}}{dt} = \frac{dh_w}{dt} \cdot a_w$$
$$\frac{dy_{h_w}}{dt} = (\frac{C_w(v_{q_w} - f(z_w)\sqrt{v_{p_w}})}{2l_w\beta(h_w)}) \cdot a_w$$
$$\frac{dy_{h_w}}{dt} = (\frac{q_w}{2l_w\beta(\frac{y_{h_w} - b_w}{a_w})}) \cdot a_w \tag{3.43}$$
$$\frac{dy_{h_w}}{dt} = c_w(q_w, y_{h_w}) \tag{3.44}$$

Equation (3.44) give the normalised water level model. Calculations based on Equation (3.8) shows the factors in the equation.

Normalised oil level model

For the oil level model we first have to derive normalisation constants a_o given in Equation (3.46) and b_o given in Equation (3.47). Normalised level y_{h_o} is given in Equation (3.45) is derived from actual oil level h_o.

$$y_{h_o} = a_o h_o + b_o \tag{3.45}$$
$$a_o = \frac{y_{o_2} - y_{o_1}}{h_{o_2} - h_{o_1}} \tag{3.46}$$
$$b_o = y_{h_o} - a_o h_{o_0} \tag{3.47}$$

The normalised oil level model is derived from the oil level model given by Equation (3.15). We derive a normalised model function $c_o(\cdot)$ which is based on the real value model. The model gain and parameter weight differs from one plant to another. To analyse different magnitude and parameter weight is changed.

$$
\begin{aligned}
\frac{dh_o}{dt} &= \frac{q_{i_o} - q_{o_o} - \frac{\partial V_{o_1}}{\partial h_w} \cdot \frac{dh_w}{dt}}{\frac{\partial V_{o_2}}{\partial h_o}} \\
\frac{dy_{h_o}}{dt} &= \frac{dh_o}{dt} \cdot a_o \\
\frac{dy_{h_o}}{dt} &= \left(\frac{C_o(v_{q_o} - f(z_o)\sqrt{v_{p_o}}) - \frac{\partial V_{o_1}}{\partial h_w} \cdot \frac{dh_w}{dt}}{\frac{\partial V_{o_2}}{\partial h_o}}\right) \cdot a_o \\
\frac{dy_{h_o}}{dt} &= \left(\frac{q_o + (2 \cdot l_w \cdot \beta(h_w) \cdot \frac{dh_w}{dt})}{2 \cdot l \cdot \beta(h_o)}\right) \cdot a_o \\
\frac{dy_{h_o}}{dt} &= \left(\frac{q_o + (2 \cdot l_w \cdot \beta(\frac{y_{h_w} - b_w}{a_w}) \cdot \frac{dy_{h_w}}{dt} \cdot \frac{1}{a_w})}{2 \cdot l \cdot \beta(\frac{y_{h_o} - b_o}{a_o})}\right) \cdot a_o \qquad (3.48) \\
\frac{dy_{h_o}}{dt} &= c_o(q_o, y_{h_o}, y_{h_w}, \frac{dy_{h_w}}{dt}) \qquad (3.49)
\end{aligned}
$$

From Equation (3.49) we have the normalised oil level model.

Normalised gas pressure model

Normalised gas pressure y_{p_g} is given by Equation (3.50). Normalisation constants a_g given by Equation (3.51) and b_g given by Equation (3.52) are derived below.

$$
\begin{aligned}
y_{p_g} &= a_g p_g + b_g \qquad (3.50) \\
a_g &= \frac{y_{g_2} - y_{g_1}}{p_{g_2} - p_{g_1}} \qquad (3.51) \\
b_g &= y_{p_g} - a_g p_{g_0} \qquad (3.52)
\end{aligned}
$$

To find the normalised gas pressure model we resume Equation (3.21). Calculations show how the model can be normalised and Equation (3.54) gives a function $c_g(\cdot)$ for the normalised gas pressure derivative. The normalised gas pressure model is given in Equation (3.54).

$$\frac{dp}{dt} = RT(\frac{w_{i_g}(t) - w_{o_g}(t)}{V_g(h_o)}) - p\frac{\frac{dV_g(h_o)}{dt}}{V_g((h_o))}$$
$$\frac{dy_p}{dt} = \frac{dp}{dt} \cdot a_p$$
$$\frac{dy_p}{dt} = (\rho_g RT(\frac{C_g(v_{q_g} - f(z_g)\sqrt{v_{p_g}})}{V_g(\frac{y_{h_o}-b_o}{a_o})}) - p\frac{\frac{dV_g(\frac{y_{h_o}-b_o}{a_o})}{dt}}{V_g((\frac{y_{h_o}-b_o}{a_o}))}) \cdot a_p$$
$$\frac{dy_p}{dt} = (\rho_g RT(\frac{q_g}{V_g(\frac{y_{h_o}-b_o}{a_o})}) + p \cdot \frac{2 \cdot l \cdot \beta(\frac{y_{h_o}-b_o}{a_o}) \cdot \frac{dy_{h_o}}{dt} \cdot \frac{1}{a_o}}{V_g((\frac{y_{h_o}-b_o}{a_o}))}) \cdot a_p \tag{3.53}$$
$$\frac{dy_p}{dt} = c_p(g_g, y_{h_o}, \frac{dy_{h_o}}{dt}) \tag{3.54}$$

3.4 Process limitations

For every process there are some limitations. In this section the main limitations for the separator process and process control will be derived. This section is largely inspired by Skogestad & Postlethwaite (2005) and Wilhelmsen (2012).

3.4.1 Impact due to noise

In all real processes there will be noise which influence the process, in this case noise influence the measurement signal. Even if the the wiring is screened and we introduce signal filters, we are not able to get rid of the whole problem. In most cases the noise will be a random signal with zero mean value, called white noise (Brown & Hwang 1997). Therefore band limited white noise is introduced to the measurement to represent noise in the separator model.

The problem introduced by noise is measurement uncertainty. If the noise effect is large, the real process value can not be known. An other problem is that noise reinforced by the controller can make the actuator constantly work and tear mechanic parts. For instants the stem in a globe valve can be worn and make leakage through the packing box.

For the separator control the noise will limit the controller gain. The noise effect in the given process is chosen to be 1 % of the measurement value, and there is only noise on the process measurements. In our case 1 % peak to peak value of noise is the maximum for all signals and the process outputs. This means that the controller gain will be limited to the highest value of one without filtering the noise.

To reduce the impact due to noise, filters like first or second order filters can be introduced to the control system. Though the control structure performance will be limited by the filter character. For instance a low pass filer will introduce a new time constant to the system, which may cause phase lag in the system. A well working filter has to be designed from plant properties. This is shown in Section 4.2.6 and Section 4.4.

3.4.2 Limitations imposed by time delay

Time delay causes serious limitations to control performance. The mathematical description of the time delay is given by Equation (3.55). The problems that time delay causes is that every change has to wait in a given time before something happens. It is easy to understand that this reduces the speed of the control performance (Skogestad & Postlethwaite 2005).

$$h(s) = e^{-\tau s} \tag{3.55}$$

Figure 3.3 shows a block diagram of transport delay in a feedback controlled loop.

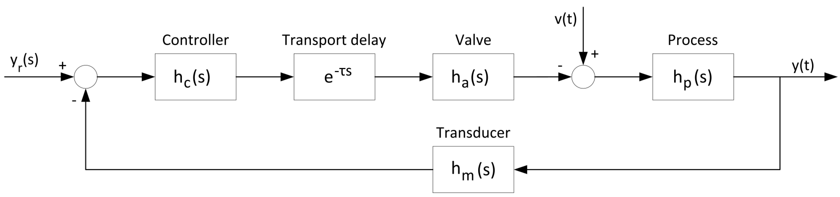

Figure 3.3: Block diagram of a feedback controlled loop with transport delay.

Figure 3.4 shows the phase diagram where the function given in Equation (3.55) is plotted. The time delay does not affect the loop magnitude which is 0 $[dB]$ regardless frequency. However it affects the loop phase. When there is a time delay in the process, it is not possible to maintain phase lead for frequencies over a certain frequency.

Skogestad & Postlethwaite (2005) suggests an upper band limit of $\omega_c < \frac{1}{\tau}$ for a closed loop control system.

3.4.3 Limitations caused by phase lag

Skogestad & Postlethwaite (2005) mention that in theory there are no fundamental limitations caused by phase lag resulting from minimum-phase elements. Though in practice there are often some limitations.

For the minimum-phase system given by Equation (3.56) the gain will be strongly reduced for high frequencies, $|h(j\omega)| \approx (\frac{k}{\prod T_i})\omega^{-n}$. In practise a large phase lag at high frequencies for the plant given in Equation (3.56), poses a problem independent of k, even when the saturation is not an issue. It is the case since positive phase margin is needed for $\omega_c > -180°$ for the system. The stability criteria for the system is $\omega_c < \omega_{180}$.

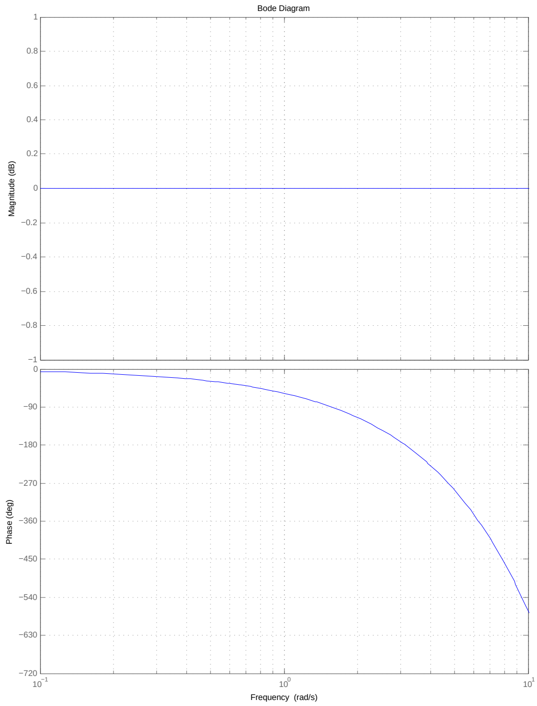

Figure 3.4: Bode diagram of the time delay function $e^{-\tau s}$ where $\tau = 1$.

$$h(s) = \frac{k}{(1+T_1 s)(1+T_2 s)(1+T_3 s)\cdots} = \frac{k}{\prod_{i=1}^{n}(1+T_i s)} \tag{3.56}$$

Stability bounds for P or PI control: $\omega_c < \omega_u$ which also is the practical stability bound for PID control (Skogestad & Postlethwaite 2005).

A P or a PI controller will have stability bounds imposed phase lag. For a proportional P controller we have $\omega_{180} = \omega_u$ and for PI controller we have $\omega_{180} < \omega_u$ The fundamental limitation stability bound for a P or a PI controller is given by Equation (3.57) (Skogestad & Postlethwaite 2005).

$$\omega_c < \omega_u \tag{3.57}$$

To extend the gain and crossover frequency ω_c beyond ω_u, zeroes must be placed in the controller to provide a phase lead. This can be done by a derivative action which counteracts the negative phase in the plant. A normal PID controller has a maximum phase lead of 90 degrees at high frequencies, but in practice it is less. The practical performance bound for a PID controller is given by Equation (3.57).

3.4.4 Limitations imposed by input constraints

All physical systems, as well as the separator process, are limited by constraints. In this case the input is a limit for how large the changes can be for the manipulated value. In Section 3.3.1 the system is normalised between zero and one, which means that the manipulated value always will be $|u(t)| \leq 1$. The analysis of the plant has to be done, to find out if the control is able to reject any disturbances and achieve perfect control $|e| < 0$ or acceptable control $|e| < 1$.

For the separator process input stabilisation is required. The process is unstable and feedback control is required to stabilise the plant. Input constraints, large disturbances and noise may make stabilisation difficult. Under all conditions the magnitude of manipulated value u should be $|u| < 1$ for the disturbance $|v| = 1$ to reject the disturbance. Otherwise u will exceed 1 for a periodic disturbance and the controller may not be able to stabilize the plant (Skogestad & Postlethwaite 2005).

3.4.5 Impact due to disturbances and commands

In this part the requirements for how fast the control system has to be in order to reject disturbances and track commands, will be derived. The model for the disturbance model is given by the transfer function $h_d(s)$ and the magnitude for the reference is Y_r.

Disturbance rejection

We look at the disturbance v in a system with constant reference $y_r = 0$. If the system given in Equation (3.58) is without control and it is scaled properly, the sinusoidal response is given by Equation (3.59). For the properly scaled plant, the worst case disturbance will be $v(t) = \sin(\omega t)$.

$$y(\omega) = h(j\omega) \cdot u(\omega) \tag{3.58}$$

$$e(\omega) = h_d(j\omega) \cdot v(\omega) \tag{3.59}$$

Skogestad & Postlethwaite (2005) states that: "*No control is needed if $h_d(j\omega) < 1$ at all frequencies (in which case the plant is said to be self-regulated).*" Control is needed when $h_d(j\omega) > 1$ for some frequencies and we consider feedback control and the control error equation given by Equation (3.60). The requirement is that the process is in minimum-phase.

For our plant, which is an integrator, we know the system is not a minimum-phase system and that $|h_d(j\omega)| \geq 1$ for some frequencies $|e(\omega)| \geq 1$. That means we need feedback and/or feed forward control.

$$e(s) = S(s) \cdot h_d(s) \cdot v(s) \tag{3.60}$$

The performance requirement $|e(\omega)| < 1$ and a disturbance $|d(\omega)| \leq 1$ at any frequency is satisfied if and only if:

$$|Sh_d(j\omega)| < 1 \quad \forall\omega \qquad \Leftrightarrow \qquad \|Sh_d\|_\infty < 1 \tag{3.61}$$

$$\Leftrightarrow \qquad |Sh_d(j\omega)| < 1/|h_d(j\omega)| \quad \forall\omega \tag{3.62}$$

$$\omega_B > \omega_d \qquad \text{where } \omega_d \text{ is defined by } |Sh_d(j\omega_d)| = 1 \tag{3.63}$$

Skogestad & Postlethwaite (2005) claims that a plant with a small $|h_d|$ or small ω_d is preferable since the need for feedback control is less.

3.4.6 Limitations caused by variations in dependent process variables

Often process variables are dependent of each other in the processes. Time constants, delay or fast response in process variables, that each process are dependent of, may limit the speed of the control performance. Most of all variations in linked process variables requires smaller variation tolerance for each process variable. When many process variables are linked together, it is required to keep less variations in each one of them to keep the whole system within the process limits.

Linked process variables often introduce larger nonlinear effect to the system. The nonlinearity is determined by the plant structure and the physical dependence between the process variables. Analysis of dependence between process variables, may lead to use of nonlinear techniques or linearisation around a given points, which is done in this report.

Dependence between process variables, does not always introduce a nonliear effect. For instant for to unmixed liquids in a tank with strait vertical walls, the level of upper liquid medium will be linearly dependent of the level of the lower liquid medium. However the gas pressure for a closed tank is not linearly dependent of liquid level, since the pressure is not linearly dependent of volume.

3.5 Slug in pipelines

Large disturbances in the separator process, may often be caused by slug flow. Therefore we will pay extra attention to this phenomenon. The disturbance used in simulations is in fact a slug flow. It is the first separator in the chain that is directly exposed to slug flow, but often the effect of slug flow propagate in the system.

Variations in flow will often appear in pipelines with multiphase flow connecting oil wells to offshore or onshore processing units. *Slug flow* or *slug* is a form for flow variations, in which the liquid flows intermittently along pipes in a concentrated mass, shown in Figure 3.5. Gas pockets or liquid plugs in the pipe form the slug and make the flow arrive intermittent to the plant.

Frequently and large rapid flow variations may causes unwanted flaring and it can even cause plant shut down (Havre et al. 2000). Impact on the process due to slug, increases the need for large operating margins, which may reduce the plant throughput.

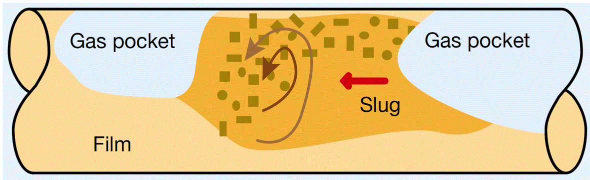

Figure 3.5: Variable pipeline flow

The seabed where oil an gas pipes are lying, is not even. Hills and valleys makes low-points in the pipeline which tends to accumulate liquid plugs that block the gas flow. For low flow rates this leads to formation of a slugging flow regime called terrain-slugging (Jahanshahi et al. 2012).

Havre et al. (2000) describes the slug build up process like this:

"Slug flow starts with an accumulation of oil and water in low-lying parts of the pipeline. Gas collects downstream of this growing slug, causing an increase in pressure. When the pressure reaches a certain level, the slug begins to move towards the pipeline outlet, followed by the gas. This process repeats itself."

(Havre et al. 2000) suggests some methods to reduce the impact of slug on the process.

- Increase the first stage separator to provide a necessary buffer.
- Implement disturbance feed forward control to prepare the process for slug.
- Detecting the slug build-up and use the choke to reduce the impact on the separator unit.

A slugging well can be characterized as a stable limit cycle (Kaasa, Alstad, Zhou & Aamo 2008). Figure 3.6 shows a typical slug build-up in a riser for a offshore oil production unit. For control structure analysis slug flow can be modelled as a square pulse with constant frequency. First oil will arrive, then water and at last gas. After a while the input to the separator will be low, since the slug is building up in the riser. Often slug flow is much more complicated, but this is a approximation for analyse purposes.

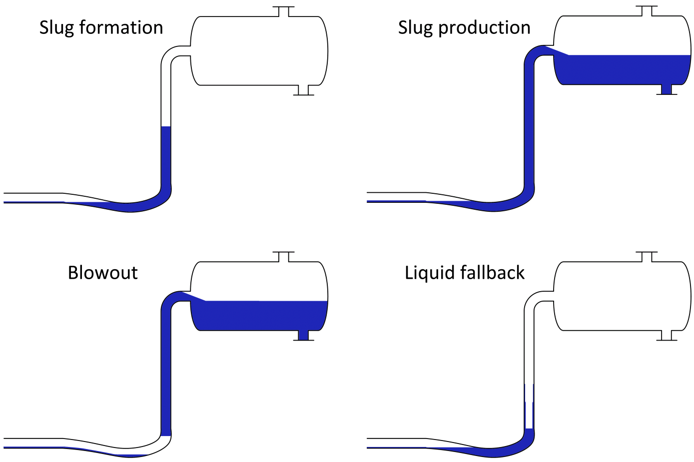

Figure 3.6: Schematics of the severe slug cycle in flow line riser systems. In the beginning a plug of oil and water is formed in the pipe. When the pressure in the pipe becomes too large the liquid will be pushed into the tank, oil first and later water. When most of the liquid has been push into the separator, gas will arrive and blow into the separator which is called "blow-out". When the pipe pressure drops, the liquid will fall back down and form a new liquid plug.

4 Control structure tools, design and tuning methods

In this Chapter we will look at different control structures, tools for making the structure able to control the process and tuning. In the first part control tools is discussed and developed for the purpose. The second part examines the construction of a suitable PID controller for the simulation model. In the last part controller tuning is derived, and calculations for the specific plant is done.

Extract of the Simulink model of the most important plant parts are given in Appendix D, and the script for initialising the model is given in Appendix C.

4.1 Controller tools

This section concerns the tools that make the controller work properly and mathematics of the controller will be derived here. The basis for loop control is a feedback controlled process with a continuous time PID controller. To make the controller work properly in the system, some functions are implemented like anti integral windup and derivative filter.

4.1.1 PID controller

The controller used to control the process is the PID controller given by Equation (4.1) and Equation (4.2). Haugen (2010) states that SIMC controller tuning method assumes a serial PID controller given by the transfer function in Equation (4.1). However, often the parallel PID controller given in Equation (4.2) is implemented in the simulation model, and that is also the case here. SIMC is derived in Section 4.3.1 later in this Chapter.

$$u(s) = K_{p_s} \frac{(T_{i_s} s + 1)(T_{d_s} s + 1)}{T_{i_s} s} \tag{4.1}$$

$$u(s) = [K_{p_p} + \frac{K_p}{T_{i_p} s} + K_{p_p} T_{d_p} s] e(s) \tag{4.2}$$

The calculated PID parameters by SIMC has to be recalculated for parallel PID controlled by the equations from Equation (4.3) to Equation (4.5).

$$K_{p_p} = K_{p_s}(1 + \frac{T_{d_s}}{T_{i_s}}) \tag{4.3}$$

$$T_{i_p} = T_{i_s}(1 + \frac{T_{d_s}}{T_{i_s}}) \tag{4.4}$$

$$T_{d_p} = T_{d_s} \frac{1}{1 + \frac{T_{d_s}}{T_{i_s}}} \tag{4.5}$$

Figure D.2 shows the parallel PID controller with anti-windup used in the verification model in Simulink.

4.1.2 Anti-Windup

In the ideal case there should not be necessary to stop the integral term form increasing the controller output. In practice every controller output is limited to an upper and a lower value. It means that the integral term is able to increase the output beyond the limits without affecting the process. This is called integrator windup (Bohn & Atherton 1995).

When the integrator performers a large windup it can be impossible for the controller to get back within working range. This makes the controller perform bad or even unstable. Anti-windup is a collective term for different techniques to prevent integrator windup. The method implemented in the controller is the modified tracing anti-windup described in (Bohn & Atherton 1995) and (Wilhelmsen 2012). The technique is based on feedback from the controller output to the integral term to reduce the integral input when the output reaches the limits. Figure 4.1 shows a Simulink diagram for how tracking anti-windup is implemented in a PID controller.

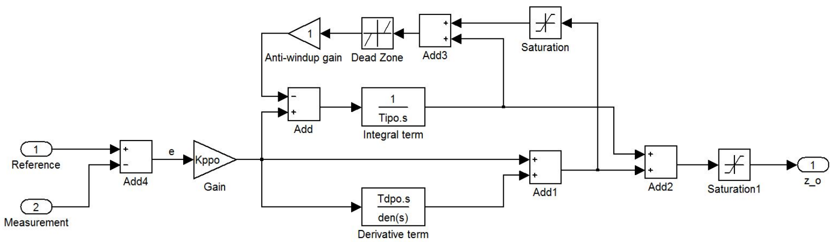

Figure 4.1: Simulink diagram showing the modified tracking anti-windup

4.1.3 Derivative filter

In most practical cases the derivative term is switched off in the controller. There are many reasons why, some of them is that it is the most difficult part to tune, it makes the stability region of the controller more complex and it is very sensitive to noise (Visioli 2006). To reduce the effect and noise amplification caused by derivative term, a derivative filter can be introduced. For the separator process control derivative action, to reduce the phase lag, may be an opportunity. Since there is measurement noise in the control loop we have to introduce a filter in connection with the derivative action.

The derivative filter introduces a new tuning constant N_p to the controller, which is the filter constant. Equation (4.7) to Equation (4.10) gives the recalculated control parameters where K_p', T_i', T_d', N_p' are initial tuning parameters (Visioli 2006) (Wilhelmsen 2012). Equation (4.6) gives the equation for filtered derivative term. The structure of the PID controller is given in Figure 4.2.

$$T_{df} = \frac{T_d}{\frac{T_d}{N_p}s + 1} \tag{4.6}$$

$$K_{pf} = K_p' \frac{T_{if}}{T_i'} \tag{4.7}$$

$$T_{if} = T_i'(1 + \frac{1}{N_p'}) \tag{4.8}$$

$$T_{df} = T_d'(\frac{T_i'}{T_{if}} - \frac{1}{N_p'}) \tag{4.9}$$

$$N_p = \frac{T_{df}N_p'}{T_d'} \tag{4.10}$$

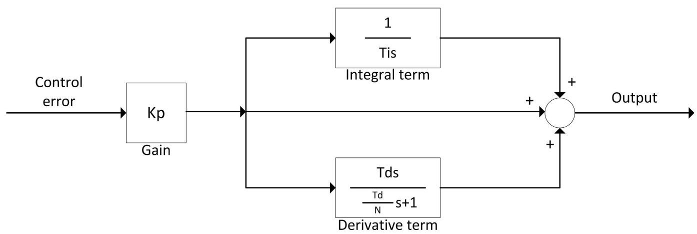

Figure 4.2: Block diagram for the derivative filter design in a PID controller

4.2 Control structure design

Plant control is done by three SISO loops, one for each process variable that should be controlled. This section covers choices for control structure, and how they work in the system.

The assignment text states that following specifications for measurements. There is only measurements for water level, oil level and gas pressure. No flow measurements are available, which means that disturbance feed forward and cascade control where the actuator is in the inner loop, can not be implemented.

4.2.1 Single loop PI/PID feedback control

Single loop feedback control is the basis for separator control. A block diagram showing the structure is given in Figure 4.3. The structure suggested in Chapter 5 is based on single loop control and is compared to this structure.

The basic idea of feedback control is to compare the process variable y to a reference signal y_r. The control error is the difference between those signals given in Equation (4.11) and it is used by the controller to control the process through an actuator.

$$e = y - y_r \tag{4.11}$$

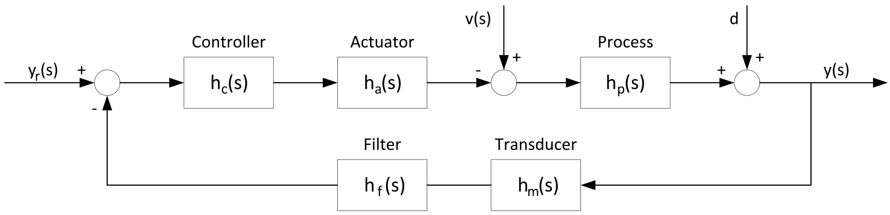

Figure 4.3: Block diagram of the feedback controlled loop. Disturbances given by the function $v(s)$ and measurement noise is given by d.

Water loop

In Figure 2.2 a piping and instrumental diagram for all the control loops are presented. The structure for each of the tree control loops are given by Figure 4.3. The piping and instrumental diagram of the water level loop is given in Figure 4.4. In this case a differential pressure transducer is used to measure water level and globe valve is used to control water outflow.

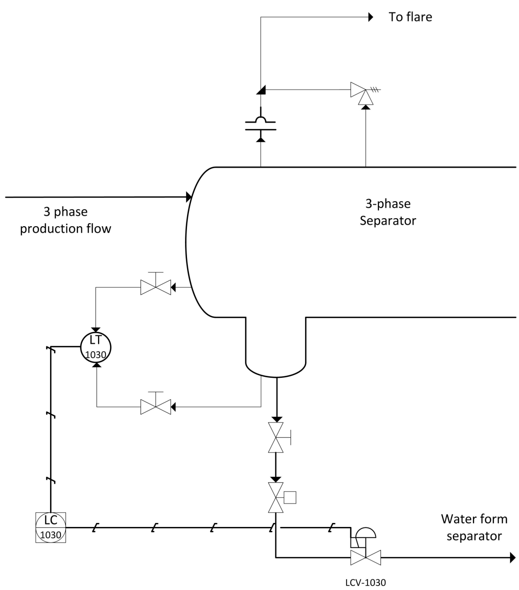

Figure 4.4: Piping and instrumental diagram for the feedback controlled water level loop.

Oil loop

The piping and instrumental diagram for the feedback controlled oil level loop is presented in Figure 4.5. Oil level is measured by a differential pressure transducer and it is fed back to the controller. The controller compares reference and measurement, and controls the outflow from the separator, which means that the loop is reversed. The actuator controlled by the controller is a nonlinear globe valve.

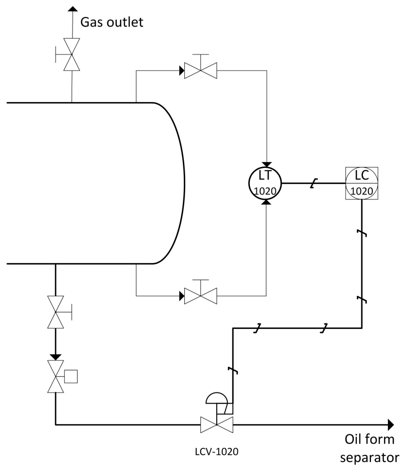

Figure 4.5: Piping and instrumental diagram for the feedback controlled oil level loop.

Gas loop

Figure 4.6 shows a piping and instrumental diagram of the gas pressure loop. Also the gas loop is controlled by the outlet as well as oil level and water level. That means it is a reversed loop. The transducer is a pressure transducer and the actuator is described as a simple globe valve, in other cases the actuator could have been a compressor.

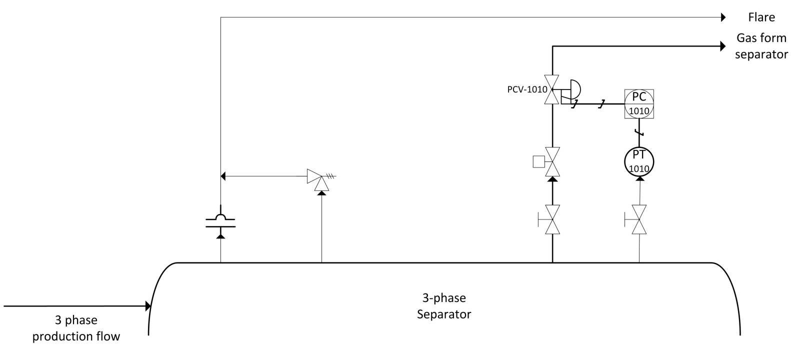

Figure 4.6: Piping and instrumental diagram for the feedback controlled gas pressure loop.

4.2.2 Feed forward control

Feed forward control has the advantage that it makes the control system counteract a disturbance before it affects the system. Ideally the feed forward signal cancels the disturbance perfectly. However no feed forward control can be expected to be perfect. To successfully implement disturbance feed forward, both the disturbance transfer function and the transfer function for the process model have to be known with reasonable accuracy.

Feed forward can not by itself cause instability to a system, but it cannot by itself be used to stabilise an unstable system system. Feedback has to be implemented as well (Hovd 2011) . Most frequently feed forward control is used to improve control performance at high frequencies. Often the method is used to achieve bandwidth beyond achievable bandwidth for closed loop feedback control (Hovd & Bitmead 2012).

A most important criteria of disturbance feed forward is that the disturbance can be measured. For the process, given in this task, there is no flow measurement neither for inflow or outflow from the separator. That means there is impossible to implement disturbance feed forward control.

Feed forward may be implemented even though disturbance feed forward is impossible to implement. Feed forward between the control loops may be a possibility. The most interesting would be a feed forward solution from the water level loop to the oil level loop. Since oil level is directly influenced by water level, and a feed forward function will be relatively simple to develop.

Feedback becomes less efficient when the loop gain becomes less than one. This often happens for frequencies higher than crossover frequency ω_c, which also is the case for the separator control loops. In systems where the feedback control is effective, feed forward becomes less effective. Often feed forward is most interesting for systems where there is impossible to achieve large bandwidth, for instance systems where time delay is prominent (Balchen, Andresen & Foss 2003) (Hovd 2011).

Balchen et al. (2003) claims that feed forward is a model based method since a mathematical model of the process has to be develop before implementation. The model is used to calculate the admission required to counteract the measured disturbance. Haugen (2009) explains that the feed forward function F_f, usually consists of a sum of calculated reference feed forward and disturbance feed forward. Three methods for developing the function is suggested:

- F_f developed from a differential process equation.
- F_f developed from a transfer function process model.
- F_f developed from experimental data. This method is model free and should be regarded as an approximative method.

The control structure solution containing floating reference, derived later in the report, is a method inspired by both feed forward and cascade control. The idea behind floating reference is to make the control loop regard changes in other control variables, like disturbance feed forward regard flow disturbances.

4.2.3 Cascade control

Cascade control for oil level control, is derived in (Wilhelmsen 2012). The conclusion in this report is that it makes the level control even slower than simple feedback control. Improvements like actuator linearising and disturbance rejection within the inner loop gave an unsatisfactory result.

The dynamic of the control loops derived in this report, is not largely different form the oil loop derived in Wilhelmsen (2012). All the control loops contain a large time constant in the actuator which means that the inner loop will be slow compared to an outer loop. Disturbances also appear in the outer loop, which means no advantage with disturbance rejection in the inner loop. Based on results derived in (Wilhelmsen 2012) cascade control will not be analysed in this report. An other reason for not deriving cascade control is that there are no flow measurements available in the analysed plant.

4.2.4 Gain scheduling

In cases where the dynamics changes with the process, we may want to change controller parameters as a function of the process. When the changes are known, there is possible to change the parameters in a preprogrammed way. This is called gain scheduling. Gain scheduling based on process measurements and reference can perform a well for variations in process parameters and known nonlinearities.

Åström & Wittenmark (2008) describes gain scheduling as follows:

"Gain scheduling is a nonlinear feedback of a special type; it has a linear controller whose parameters are changed as a function of operating conditions in a preprogrammed way."

A simple sketch showing the gain scheduling principle is shown in Figure 4.7. The block digram shows that the gain scheduling function uses the process conditions to calculate new controller parameters. This method can be used if it is possible to find auxiliary parameters that correlate well with process parameters.

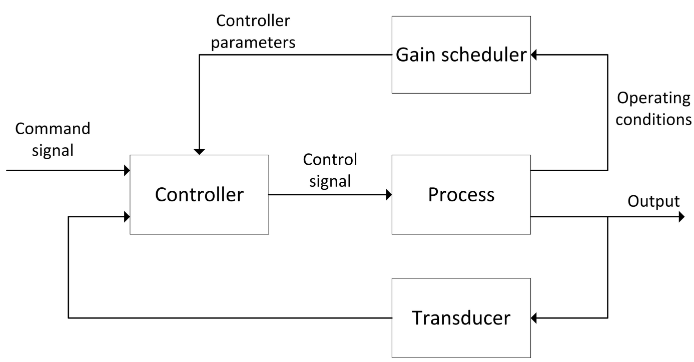

Figure 4.7: Block diagram showing the gain scheduling principle, with feedback from the process to the gain scheduling function.

Åström & Wittenmark (2008) states the following general ideas as useful when designing gain scheduling controllers:

– Linearisation of nonlinear actuators.
– Gain scheduling based on measurements of auxiliary parameters.
– Time scaling based on production rate.
– Nonlinear transformations.

From the beginning only the controller gain was changed due to the process, which also gave the name gain scheduling (Åström & Wittenmark 2008). Now it is common to change the integral and derivative term as well.

In Wilhelmsen (2012) gain scheduling and integral scheduling for the oil level control in a separator process is derived. In this case gain scheduling is used to prevent overflow in a separator for large disturbances. To prevent large gain when it is needed, a nonlinear function which regard process error changes the controller parameters. This method provides low gain for small disturbances and high gain when it is needed, Figure 4.8 shows the principle.

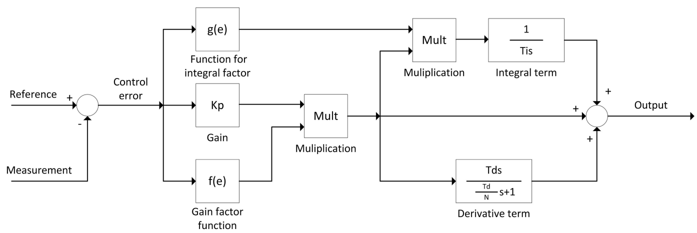

Figure 4.8: Block diagram showing how gain and integral scheduler can be implemented in the PI controller.

4.2.5 Model reference adaptive control

An alternative is to introduce the model predictive adaptive control (MRAC). To be able to implement this structure a reference model has to be made. The reference model is chosen to generate the desired trajectory y_m, which the plant output y_p has to follow. By the function $e_1 = y_p - y_m$ the deviation between the process output and the reference model is calculated (Ioannou & Sun 2003). This is used to calculate the controller parameter either by the *direct* or the *indirect* method.

The direct MRAC method may be performed for the nonlinear plant. Calculations is done by the same principle as for a linear plant (Ioannou & Sun 2003). Since it is not an important part of the task, calculations is not shown in this report.

Figure 4.9 shows the basic structure of a MRAC scheme. The closed loop plant is made up of an ordinary feedback law that contains the plant, a controller and an adjustment mechanism. The adjustment mechanism generates the controller parameters on-line.

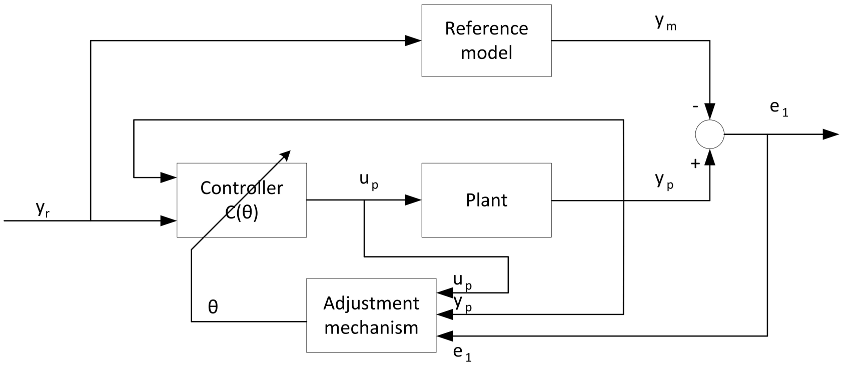

Figure 4.9: Basic block diagram of the direct Model Reference Adaptive Control (MRAC) principle.

4.2.6 Signal filter design

Signal noise and peaks in measurements may limit the controller performance. In Section 3.4.1 impact to the process due to noise on the measurement is derived. To avoid large measurement peaks and measurement uncertainties that limits the controller gain, a low pass filter can be introduced to the system.

By introducing a signal filter, it is possible to attenuate (ideally: remove) certain frequencies (Haugen 2010). A low pass filer is designed to reduce noise with high frequencies, but keep dynamic acting with low frequencies. A disadvantage with signal filters, is that they introduce a new time constant to the system and some phase lag. Therefore filters should not be designed to remove more than necessary.

Equation (4.12) is the the standard formula for first order low pass filter in time domain. The bandwidth of the filer is given by ω_b. The bandwidth defines the upper limit of the passband. Haugen (2010) claims that it is common to say that the frequency where the filter gain is $\frac{1}{\sqrt{2}} \approx 0.71 \approx -3\text{dB}$ is the bandwidth.

$$h_f(s) = \frac{1}{T_f s + 1} \tag{4.12}$$

$$h_f(s) = \frac{1}{\frac{s}{\omega_b} + 1} \tag{4.13}$$

Filter gain is $K_f = 1$ and time constant $T_f = \frac{1}{\omega_b}$. Filter frequency response i given by Equation (4.14).

$$h_f(j\omega) = \frac{1}{\sqrt{(\frac{\omega}{\omega_b})^2 + 1}} \tag{4.14}$$

Phase lag is given in Equation (4.15). Equation (4.16) give the filter bandwidth.

$$\arg h_f(j\omega) = -\arctan(\frac{\omega}{\omega_b}) \tag{4.15}$$

$$f_b = \frac{\omega}{\omega_b} \tag{4.16}$$

4.3 Controller tuning

There are many methods for controller tuning, and some methods work well in for many kinds of systems. However it is not expected that they work perfectly for any the given process. In this section we will look at some well known methods that may be used as a startpoint for our tuning.

Separator controller tuning may be based on known techniques like Ziegler & Nichols tuning method (ZN), model based methods like SIMC also called Skogestad's tuning rule, or the method based on phase amplitude diagram and ZN derived in Balchen et al. (2003). In this section we will derive the controller parameters based on SIMC and the method described in Balchen et al. (2003).

4.3.1 Skogestad/Simple IMC

When the process model is known, it is possible to use SIMC (Skogestad/Simple IMC) to control the process (Haugen 2010). SIMC is developed by the NTNU professor Sigurd Skogestad, and it is claimed to be the best PID tuning rule in the world (Skogestad & Postlethwaite 2005).

When optimising a controller it is common to approximate the process as a first order plus time delay model given in Equation (4.17). This is the basis for many tuning rules, like for instance Ziegler & Nichols open loop method. However when the Ziegler & Nichols open loop method is based on a step response, the SIMC method is based on a calculated process model, and it may not only be a first order plus delay model. The SIMC tuning rule require an accurate model of the process before tuning starts (Skogestad 2004).

$$h_s(s) = \frac{k}{Ts+1}e^{-\tau s} \tag{4.17}$$

In Table 4.1 the formulas for PID parameter calculation based on type of process can be found. Skogestad (2004) suggests using $c = 4$, but Haugen (2010) argues that it will make the disturbance rejection a bit poor and suggest using $c = 2$. However in calculations shown later in the report $c = 4$.

The article Ruscio (2010) derives SIMC, Ziegler & Nichols method and a few other known methods to find a better tuning method for integrator plus delay systems

Table 4.1: Calculation of PID parameters based on Skogestad's tuning rule (SIMC).

Process type	$h_s(s)$	K_p	T_i	T_d
Integrator + delay	$\frac{K}{s}e^{-\tau s}$	$\frac{1}{K(T_C+\tau)}$	$c(T_c+\tau)$	0
Time constant + delay	$\frac{K}{Ts+1}e^{-\tau s}$	$\frac{T}{K(T_C+\tau)}$	$min[T, c(T_c+\tau)]$	0
Integr + time const + del.	$\frac{K}{(Ts+1)s}e^{-\tau s}$	$\frac{1}{K(T_C+\tau)}$	$c(T_c+\tau)$	T
Two time const + delay	$\frac{K}{(T_1s+1)(T_2s+1)}e^{-\tau s}$	$\frac{T_1}{K(T_C+\tau)}$	$min[T_1, c(T_c+\tau)]$	T_2
Double integrator + delay	$\frac{K}{s^2}e^{-\tau s}$	$\frac{1}{4K(T_C+\tau)^2}$	$4(T_c+\tau)$	$4(T_c+\tau)$

as given in Equation (4.18). Both Haugen (2010) and Ruscio (2010) claim that controllers tuned with SIMC do not perform well for disturbance rejection. However setpoint tracking capabilities should be better.

In our case disturbance rejection is important in the way that the level limits can not under any circumstances be broken. However it is not important that the liquid level is at setpoint all the time, and therefore good setpoint tracking capabilities is not an important property for the controller. A more important property is to make the controller performance well damped.

$$h_p(s) = \frac{k}{s}e^{-\tau s} \tag{4.18}$$

4.3.2 Choice of parameters and the Balchen method

To choose parameters that should be used in the controller may not be clear. SIMC give us guidelines for which parameters that should be used, but Ziegler & Nichols method is not that precise. In many practical implementations the derivative term is not in use as mentions in Section 4.1.3.

Balchen et al. (2003) gives some guidelines for how control parameters could be calculated and which parameters that could be used by the controller, shown in table 4.2. The following principles are used to derive the suggestion for parameters:

– Relative noise damping $\zeta \approx 0.5$.
– Gain margin $\Delta K \approx 2 \approx 6[dB]$.
– Phase margin $\psi \approx 45°$.
– Ziegler Nichols method.

The statements in (Balchen et al. 2003) are used to develop the tuning method given in table 4.2. For some plant models the bandwidth has to be desired by the engineer. In these cases "*dbw*" is written in table 4.2 which means "*desired bandwidth*".

Table 4.2: PID parameter formulas from the Balchen tuning method, based on Ziegler & Nichols method and amplitude phase diagram.

$h_s(s)$	P	PI	PID
$\frac{K}{s}$	$K_p = \frac{1}{K}\omega_c,\ \omega_c = dbw$	$K_p = \frac{1}{K}\omega_c,\ T_i = \frac{5}{\omega_c},\ \omega_c = dbw$	Not applicable
$\frac{K}{Ts+1}$	$K_p = \frac{1}{K}T\omega_c,\ \omega_c = dbw$	$K_p = \frac{1}{K}T\omega_c,\ T_i = T,\ \omega_c = dbw$	Not applicable
$\frac{K}{(Ts+1)s}$	$K_p = \frac{1}{KT}$	$K_p = \frac{1}{KT\sqrt{\beta}}\ T_i = \beta T,\ \beta \approx 10$	$K_p = \frac{1}{K}\frac{T}{T_d}\omega_c,\ T_i = T,\ T_d = T,\ \omega_c = dbw$
$\frac{K}{s}e^{-\tau s}$	$K_p = \frac{1}{K}\frac{\pi}{4\tau}$	$K_p = \frac{1}{K}\frac{1}{2\tau},\ T_i = 10\tau$	Not applicable
$\frac{K}{Ts+1}e^{-\tau s}$	$K_p = \frac{1}{K}\frac{t+\frac{\tau}{2}}{\tau}$	$K_p = \frac{1}{K}\frac{\pi T_i}{4\tau},\ T_i = T$	Not applicable
$\frac{K}{(Ts+1)s}e^{-\tau s}$	$K_p = \frac{1}{K\tau}\frac{1+2\frac{T}{\tau}}{1+4\frac{T}{\tau}} \leq \frac{1}{KT}$	$K_p = \frac{1}{KT\sqrt{\beta}},\ T_i = \beta\tau,\ \beta = 20$	$K_p = \frac{1}{K\tau},\ T_i = \beta\tau,\ \beta = 20\ T_d = T$
$\frac{K}{(T_1s+1)(T_2s+1)}e^{-\tau s}$	$K_p \approx \frac{1}{K}\frac{T_1+T_2}{2\tau}$	$K_p = \frac{1}{K}\frac{\pi T_i}{4\tau},\ T_i = T$	$K_p = \frac{1}{K}\frac{\pi T_i}{4\tau},\ T_i = T_1,\ T_d = T_2$

In Section 3.4.3 limitations imposed by phase lag is derived. Skogestad & Postlethwaite (2005) claims that frequencies $\angle h(j\omega) \to -n \cdot 90^\circ$ poses a problem for plant stability, because positive phase margin is needed.

In cases where we want to extend the crossover frequency beyond the ω_{180} frequency, we must place a zero in the controller to provide a phase lead. Often this is done by introducing the derivative term in the PID controller. The maximum phase lead the derivative term can provide is 90°, but often less. The practical performance bound for a PID controller is $\omega_c < \omega_{180}$ (Skogestad & Postlethwaite 2005).

To introduce the derivative term to provide a phase lead without taking other factors into account is not a good idea. Often it is a time delay that bounds the bandwidth. In this case we have to analyse the plant to find out how large the derivative phase effect will be to the plant, and how large the derivative effect can be. At a certain point there is impossible to provide a large enough phase lead.

We will now calculate the PI and PID parameters for each controller. Both SIMC and the Balchen method will be used. To be able to calculate parameters we have to have a linear model, therefore each control loop is linearised around the setpoint, which is 0.5 for all loops.

4.3.3 Tuning of water level control loop

To tune the controller we have to know the loop dynamics, gain, time constant and time delay. We resume the separator model for water level from Section 3.3.1 given by Equation (3.43) and Equation (3.44) and the actuator model for water valve given by Equation (3.29).

$$\begin{aligned}\frac{dy_{h_w}}{dt} &= \left(\frac{q_w}{2l_w\beta(\frac{y_{h_w}-b_w}{a_w})}\right)\cdot a_w \\ \frac{dy_{h_w}}{dt} &= c_w(q_w, y_{h_w}) \\ q_{w_o} &= k_{v_w} f(z_w)\sqrt{(p_1-p_2)}\end{aligned}$$

By inserting the values for the process around setpoint in the normalised model, the gain of the water level derivative is found, which is the velocity of the level change. The gain around setpoint is denoted $k_{w_{sp}}$.

$$h_s(s) = h_p(s)\cdot h_a(s)$$

$$h_p(s) = \frac{k_{w_{sp}}}{s} \tag{4.19}$$

$$h_a(s) = \frac{q_{w_o}}{T_w s+1}e^{-\tau_w s} \tag{4.20}$$

$$h_s(s) = \frac{q_{w_o}\cdot k_{w_{sp}}}{(T_w s+1)s}e^{-\tau_w s} \tag{4.21}$$

The overall model for the water level control is obtained by combining the separator transfer function model in Equation (4.19) and the valve transfer function model given given by Equation (4.20). The open loop water level model is given in Equation (4.21).

Simplifications to make tuning easier have to be made for the water loop controller. As well as for oil level, water level depends on liquid surface area, variable valve gain and separator pressure. It is not wrong to assume that both liquid level and gas pressure is at setpoint. This assumptions means that the water surface area can be seen constant and the outflow is only dependent of the valve opening. Based on these assumptions we can calculate PID parameters for different valve gain.

Formulas for loop tuning by SIMC is given in table 4.1. In this specific case we use the formulas for integrator plus time constant and delay. The SIMC tuning results are presented in table 4.3. The same calculations as done with SIMC are done by the method described in (Balchen et al. 2003). The results are presented in table 4.4.

Table 4.3: Calculated PID parameters for the water level controller by the SIMC method for different valve gain.

$f(z)$ value	$k_{w_{sp}}$	K_p	T_i	T_d
0.33	$4.48 \cdot 10^{-4}$	223.1	40	6
0.67	$9.10 \cdot 10^{-4}$	110.0	40	6
1.0	$1.36 \cdot 10^{-3}$	73.6	40	6

Table 4.4: Calculated PID parameters for the water level controller by the Balchen method for different valve gain.

$f(z)$ value	$k_{w_{sp}}$	K_p	T_i	T_d
0.33	$4.48 \cdot 10^{-4}$	371.8	120	6
0.67	$9.10 \cdot 10^{-4}$	183.1	120	6
1.0	$1.36 \cdot 10^{-3}$	121.8	120	6

4.3.4 Tuning of oil level control loop

Also for the oil level we have to derive the dynamics around setpoint before tuning the controller. Therefore we resume the normalised oil level model given in Equation (3.48) and Equation (3.49).

$$\begin{aligned}
\frac{dy_{h_o}}{dt} &= \left(\frac{q_{i_o}(t) - q_{o_o}(t) + (2 \cdot l_w \cdot \beta(\frac{y_{h_w} - b_w}{a_w}) \cdot \frac{dy_{h_w}}{dt} \cdot \frac{1}{a_w})}{2 \cdot l \cdot \beta(\frac{y_{h_o} - b_o}{a_o})}\right) \cdot a_o \\
\frac{dy_{h_o}}{dt} &= c_o(q_o, y_{h_o}, y_{h_w}, \frac{dy_{h_w}}{dt}) \\
q_o &= k_{v_o} f(z_o) \sqrt{(p_1 - p_2)}
\end{aligned}$$

Since both oil level and valve gain are varying, some simplifications have to be made. The first thing we assume is that the oil level is at setpoint and therefore the liquid surface is constant. The other assumptions is that the valve gain is one, which is a worst case scenario for fully open control valve. Calculations are also done for some other working points for the valve.

When inserting the values for setpoint, we find the magnitude of the level change around that point and $k_{p_{sp}}$ is calculated. From the process transfer function in Equation (4.22) and the linearised actuator in Equation (4.23), we can calculate the transfer function Equation (4.24) required for tuning with SIMC. Results of the SIMC tuning is presented in table 4.5.

$$h_p(s) = \frac{k_{p_{sp}}}{s} \tag{4.22}$$

$$h_a(s) = \frac{k_{a_o}}{T_o s + 1} e^{-\tau s} \tag{4.23}$$

$$h_s(s) = h_p(s) \cdot h_a(s)$$

$$h_s(s) = \frac{k_{p_{sp}} k_{a_o}}{(T_o s + 1)s} e^{-\tau s} \tag{4.24}$$

Formulas for loop tuning by SIMC is given in table 4.1. In this specific case we use the formulas for integrator plus time constant plus delay.

Table 4.5: Calculated PID parameters for the oil level controller by SIMC for different valve gain.

$f(z)$ value	$k_{w_{sp}}$	K_p	T_i	T_d
0.33	$3.28 \cdot 10^{-3}$	30.5	40	6
0.67	$4.62 \cdot 10^{-3}$	21.6	40	6
1.0	$5.92 \cdot 10^{-3}$	16.9	40	6

In table 4.6 tuning results by the tuning rule described in (Balchen et al. 2003) is presented.

Table 4.6: Calculated PID parameters for the oil level controller by the Balchen method for different valve gain.

$f(z)$ value	$k_{w_{sp}}$	K_p	T_i	T_d
0.33	$3.28 \cdot 10^{-3}$	50.8	120	6
0.67	$4.62 \cdot 10^{-3}$	36.1	120	6
1.0	$5.92 \cdot 10^{-3}$	28.1	120	6

4.3.5 Tuning of gas pressure controller

We will now calculate the PI and PID parameters for the gas pressure controller. Both SIMC and the Balchen method is used. First we resume the normalised model from Equation (3.53) and Equation (3.54) and linearise the model around the setpoint.

$$\frac{dy_p}{dt} = (RT(\frac{w_{i_g}(t) - w_{o_g}(t)}{V_g(\frac{y_{h_o}-b_o}{a_o})}) + p \cdot \frac{2 \cdot l \cdot \beta(\frac{y_{h_o}-b_o}{a_o}) \cdot \frac{dy_{h_o}}{dt} \cdot \frac{1}{a_o}}{V_g((\frac{y_{h_o}-b_o}{a_o}))}) \cdot a_p$$

$$\frac{dy_p}{dt} = c_p(w_g, y_{h_o}, \frac{dy_{h_o}}{dt})$$

$$q_{o_g} = k_g \cdot \sqrt{(p_1 - p_2)}$$

Linearised around the setpoint, the transfer function for the separator pressure is given by Equation (4.25) and the actuator by Equation (4.26).

$$h_p(s) = \frac{k_{p_{sp}}}{s} \tag{4.25}$$

$$h_a(s) = \frac{k_{a_p}}{Ts+1} \tag{4.26}$$

$$h_s(s) = h_p(s) \cdot h_a(s)$$

$$h_s(s) = \frac{k_{p_{sp}} k_{a_g}}{(Ts+1)s} \tag{4.27}$$

The dynamic of the gas pressure model is a bit complex, but it is possible to simplify without changing the main dynamic radically. When all process variables operate around the setpoint, we can assume the gas volume as constant and therefore the pressure is only dependent of gas flow.

From the simplified and linearised model given in Equation (4.27) we find a process that works like an integrator plus time constant function. The factor k_{a_g} is calculated the actuator gain plus the process gain. Formulas for loop tuning by SIMC is given in table 4.1. In this specific case we use the formula for integrator plus time constant plus delay for SIMC. Results for controller tuning from SIMC is given in table 4.7.

Table 4.7: Calculation of PID parameters for the gas pressure controller for different gas volumes, based on SIMC (Skogestad's tuning rule).

Volume	$k_{w_{sp}}$	K_p	T_i	T_d
$0.8 \cdot V(h_{o_{sp}})$	$9.02 \cdot 10^{-3}$	1.84	240s	60s
$1.0 \cdot V(h_{o_{sp}})$	$7.21 \cdot 10^{-3}$	2.31	240s	60s
$1.2 \cdot V(h_{o_{sp}})$	$6.01 \cdot 10^{-3}$	2.77	240s	60s

For the Balchen method we have to choose ω_c and we use the formula for integrator plus time constant. From the bode plot of the open loop it is here chosen to be $\omega_c = 0.10 rad/s$. The tuning result is presented in table 4.8.

Table 4.8: Calculated PID parameters for the gas pressure controller by the Balchen method for different volumes of gas.

Volume	$k_{w_{sp}}$	K_p	T_i	T_d
$0.8 \cdot V(h_{o_{sp}})$	$9.02 \cdot 10^{-3}$	11.1	60	60
$1.0 \cdot V(h_{o_{sp}})$	$7.21 \cdot 10^{-3}$	13.9	60	60
$1.2 \cdot V(h_{o_{sp}})$	$6.01 \cdot 10^{-3}$	16.6	60	60

4.4 Filter tuning

In Section 4.2.6 filter design was derived. We will now calculate the filter constants suitable for this system.

When the controller tuning was derived, the gain was calculated to be higher than one for all controllers. It means that measurement noise also will be reinforced by the controller. A simple first order filter low pass filter, filtering the measurement signal, can reduce the noise effect in the system.

Often the frequency (ω_t) at $|T(j\omega)| = \frac{1}{\sqrt{2}} \approx 0.71 = -3dB$ is used as the bandwidth definition for the low pass filter in the system (Haugen 2010). From closed loop bode plot given in Appendix B, the value is found to be approximately $0.015 rad/s$ for all loops. If we use $\omega_t = \omega_b$ the filter constant can be calculated to be:

$$T_f = \frac{1}{0.015} \approx 66.7 \tag{4.28}$$

From the open loop bode plot when the low pass filter is added in the system, it can be seen that the filter has too large impact on the phase for each loop. The gain margin is significantly reduced and the filter effect is too prominent. It is therefore recommended to make the filter frequency at least a decade slower than ω_t, which means that maximum value of $T_f = 6.7$ for all control loops. Verifications done by simulations and bode plots show that the following filter give good results:

$$h_f(s) = \frac{1}{5s+1} \tag{4.29}$$

The impact on the system phase is reduced significantly compared to the filter given in Equation (4.28). When the filer in Equation (4.29) is implemented, the phase margin is reduced with approximately 3° for each loop at ω_c. Simulations of the system shows minor changes in the process dynamics. The only change in process performance when filter is added, is that noise is largely reduced.

5 Results and analysis

In Chapter 3 we derived the process model and looked at conditions that limit the performance. In this Chapter we will look at the preferred control structure and a tuning method based on separator characteristics. The result is a suggestion for a the control structure and a tuning method. As a method for comparison the SIMC method is used for tuning, and single loop feedback control for the control structure.

5.1 Control structure

There are many criteria that have to be fulfilled for the control structure. The most important is that the control structure has to provide stable control within the process limits for any disturbance under all conditions. That means the controllers has to be tuned with respect to disturbance rejection. However it is not important to maintain setpoint all the time. The goal is not to achieve as fast control as possible. It is much more important to keep the process as calm as possible.

5.1.1 Conventional feedback control

One choice of control structure is to design simple feedback controlled loops for every process variables. In this case there will be three control loops as shown in Figure 5.1, since there are three process variables we want to control. Figure 2.2 shows the piping and instrumental diagram for this structure.

The advantages with this control structure is that it is easy to implement and understand. Implementation of single loop feedback control for each loop is shown in piping and instrumental diagrams in Section 4.2.1.

However there are some drawbacks. The most important is that disturbances in other control loops are not taken into account by the other control loops. Since both oil level and gas pressure are dependent on more variables than the respectively net inflow of oil or gas, the action trigged by the respectively control loop, may be wrong for future operation. This may happen because the control structure only rejects the disturbance without violating the reason. There is no solution for this problem without implementing a more complex control structure to the process.

5.1.2 Floating reference feedback control

Floating reference is designed to improve the control loops by taking into account changes in other process variables. For instance when well slug appear, oil, water and gas will often enter the separator in a pattern described in Section 3.5. First the oil flow is largely increased and the oil level rises. When the oil flow is largely reduced and the oil level still rises cause water is filling the separator, the oil level controller will keep the oil control valve open and reduce the amount of oil. What

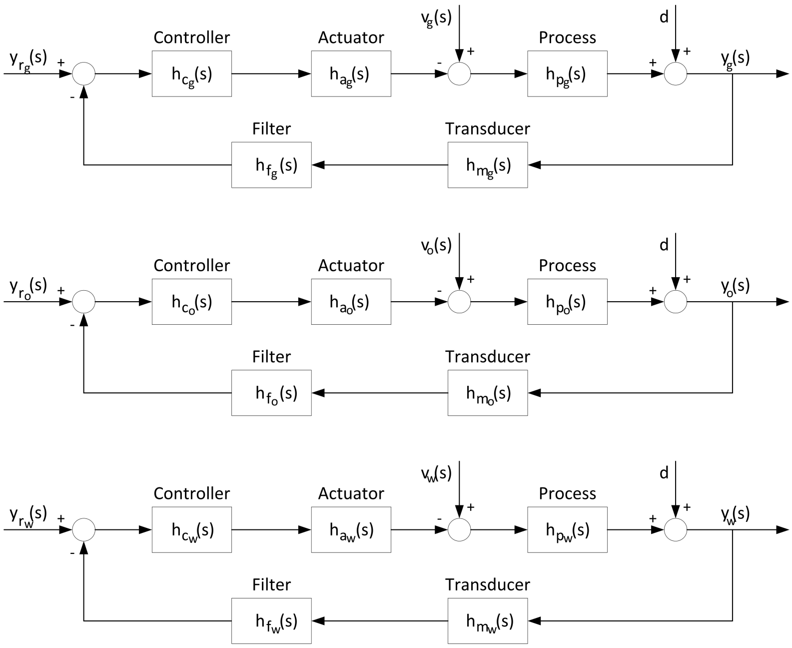

Figure 5.1: Block diagram showing conventional feedback loop control for the three controlled process values in the separator process.

happens when the water flow is reduced, is that the oil layer has become thin. The problem is when the water level is back at the setpoint, and the oil level is below the lower limit, which may happen if the oil layer has been too thin.

An improvement can be to make the reference value be specified by the process variable that influence the other. From the models for each process variables, given in Section 3.1, we know that oil level is directly affected by water level and that gas pressure is affected by liquid level. Since oil level is directly affected by water level, it should be a connection between oil level and water level. In this structure water level is used as oil level reference ($y_w = y_{r_o}$), without weighting compared to a fixed oil level reference. For a more accurate setpoint calculation, it could have been made a function for weighting the measured water level to the fixed oil level setpoint and the separator geometry could be regarded.. Figure 5.2 shows a block diagram for a floating reference implementation.

The purpose for making the oil reference dependent on water level, is to make a practical oil volume control that takes the level limits into account. An implementation where for oil volume control without considering the level limits may be made. However there is some problems according to converting level limits to volume limits for oil since the water level determine the volume limits.

When oil volume control is implemented without considering the actual level, there is no guarantee for oil level to stay between the limits for all conditions. An option is to implement two controllers, one for oil volume control and one for oil level

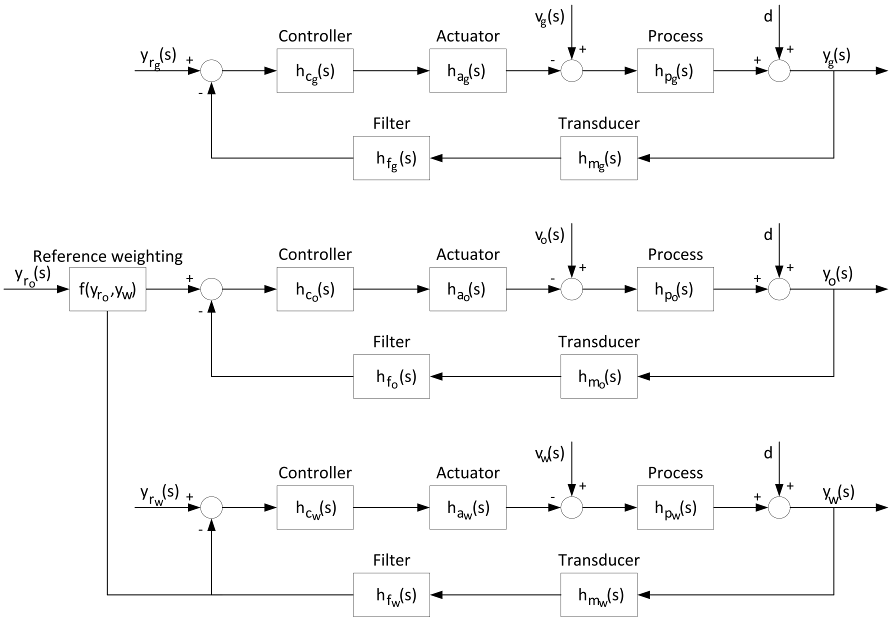

Figure 5.2: Block diagram showing implementation where oil level reference dependent of water level.

control. For this implementation a function for control signal weighting has to be made. A simpler but still well working method is the floating reference method.

An implementation where gas reference is given by oil level directly is not a good option compared to a fixed setpoint. Gas pressure is dependent on oil level, but it is gas flow in and out that makes the largest pressure changes in the process. From the process equations derived in Section 3.1.4, it is clear the gas pressure is not linearly dependent on oil level. If gas pressure reference should be dependent on oil level, the equation should be a nonlinear weighting the oil level. Without a complex nonlinear function, weighting of oil level to a fixed gas pressure setpoint, makes no benefit to the system. A fixed setpoint makes the gas pressure control perform acceptable within the limits, and therefore it is used in the control structure.

5.1.3 Discussion about control structure

More than one control structure may satisfy the control requirement in some manner. The purpose of this task is though to suggest a control structure that work well for most plants exposed to large disturbances. The following requirements have to be fulfilled for the control structure:

- Rejection of any kind of disturbance without leading to process shutdown.
- Significantly less oil level variation compared to fixed setpoint single loop feedback control.

– Less or equal use of the actuator compared to single loop control.

There is possible to operate the base case system within the process limits for the given plant using PI controllers. In this case there is a dampening effect of the disturbances for single loop PI control with fixed setpoint for all loops. However the floating setpoint leads to significantly less variations in oil level and gas pressure and less use of both actuators. The disturbance dampening and speed of process changes are more or less the same for both structures. Most significant is the improvement for oil level variations. Floating setpoint leads to significantly less oil level undershoot, for slug flow, when water inflow is largely reduced.

The improvement of floating setpoint increases when the size and lime period of the disturbance increases. For separators were liquid level rises faster than in the specific case examined, the floating reference structure performs better for the lower oil level limit. Plants with c value 20 % higher than the given case, for each loop, are still able to control the process within limits for floating oil level structure while it is a problem for fixed setpoint structure. Due to larger margins introduced by floating oil setpoint control, there is possible to control the process even though the controllers are not perfectly optimised. This is a result of less variation in oil level and gas pressure.

Floating oil reference will not give accurate volume control in this case, due to the separator shape. Still within the operating area for oil and water level, this is a well suited method. However the method would have given accurate volume control if the water cross-section had been dependent on water level only.

The effort of floating oil reference compared to fixed reference is largest when the control tuning give a bit sluggish response, disturbances are large or the period between each disturbance is long. In fact if the length of each disturbance is very short, there would be possible to have constant valve opening all the time even if the magnitude of the disturbance is high. When disturbances are large and the period between each disturbance is long, the floating reference for oil level, reduces the variation of oil level distinctly without significantly more use of the actuator. The largest benefit is obtained for the lower limit. To maintain the upper limit the controller has to be tuned properly.

5.1.4 Suggestion for control structure

The suggestion is to implement the control structure derived in Section 5.1.2, where oil level setpoint is given directly or by a function from water level. When this method is used, the oil level control loop will regard changes in water level, which is critical to maintain the lower oil limit. The requirement for this structure is that water level can not run close to its limits.

Required controller for each loop is specified by the process dynamics. In most cases a PI controller is sufficient, but a PID or a gain scheduled PID controller can be used for processes with fast dynamics. A first order low pass filter which is developed in Section 4.2.6 and Section 4.4 should be used to reduce the noise effect on the process measurement signal.

5.1.5 Choice of controllers

Choice of controllers are in this case dependent of how fast the process variables changes. In Wilhelmsen (2012) the change in oil level happens so fast that there had to be implemented a gain scheduled controller to reject the largest disturbances. In that specific case the measurement noise restricted the controllers gain, and higher gain could only be used when oil level was close to the level limits.

For a plant where the process gain is not high or there is possible to reduce the noise effect by a first og second order filter, PI or PID control can be adequate to reject disturbances. For this specific case PI or PID controllers yields good result. In our loop the measurement signal is filtered, which means that the controller gain can be more than one without reaching the maximum noise of 1 % effect given for the plant. For plants where changes in level happens faster, gain scheduled PID controller can be required to guarantee that the limits are not broken. A nonlinear gain scheduling function, that may be used for the controller, is described in Wilhelmsen (2012).

5.2 Controller tuning method

Controller tuning rules like Ziegler & Nichols methods, IMC and SIMC are all tuning rules that makes guidelines for parameter tuning. A common thing for most of the methods, is that they try to make the response as fast as possible within the stability limits. Some methods like Ziegler & Nichols method, often makes the response a bit too aggressive (Haugen 2010) (Ruscio 2010). Another commonality is that they try to make the loop both follow the setpoint and reject disturbances effectively at the same time, which is difficult to achieve.

In this section controller tuning for each controller is calculated and verification of the tuning is shown. Table 5.1 contains setpoints and $c(y)$ values at setpoints for each control loop are given.

Table 5.1: $c(y)$ values calculated for setpoints for each control loop.

Control loop	Setpoint	$c(y_{sp})$
Water level loop	0.5	$1.36 \cdot 10^{-3}$
Oil level loop	0.5	$5.93 \cdot 10^{-3}$
Pressure loop	0.5	$7.23 \cdot 10^{-3}$

5.2.1 Controllers tuned by SIMC

The expectations for parameters calculated with SIMC were that they would give a good basepoint for how the tuning rule should be preformed but no perfect result. Skogestad (2004) claims that the rule is developed to make fast response, which may not be the goal for the separator tuning. Simulation also showed that the response achieved when SIMC is used, is faster than desirable. In table 5.3 and table 5.4 values calculated with SIMC is showed.

From the process plots in Figure A.12 to Figure A.14 the response process variables are given for large disturbances when SIMC is used for tuning. Without detuning, it is clear that the gain is much too high for all three control loops. The integral time should also be longer to make the system better damped.

5.2.2 Tuning based on separator dynamics

The tuning method which in this report is called separator dynamics method, is derived from SIMC. SIMC is made to regard the actual process dynamics, but it is not developed specifically for separator control. Calculations of controller parameters has to be changed to achieve the control objective for separators. Before presenting the calculation formulas for separator dynamics method we resume SIMC from table 4.1.

Requirements for the tuning rule is that it has to keep the process within limits, prevent currencies in the separator and dampen the disturbances. A controller that keeps the process at setpoint for any disturbance is not required, calm process handling is more important. To maintain process limits for all kind of disturbances, means that the controller has to be sufficiently fast, but not so fast that disturbances is amplified. Therefore the controller gain is tuned to be low as possible but still able to reject any disturbance, which means largely reduced gain compared to SIMC.

The ability to maintain calm process handling, is given by the integral term. Integral time T_i is increased to make a better damped system compared to SIMC. A over damped system will prevent liquid currencies in the separator. The ability to keep setpoint at all time will be reduced for longer integral time.

Derivative action is not changed compared to SIMC. Derivative action make the process react faster and provides phase lead. The problem with the derivative term is that it is sensitive to noise, which should not be amplified more than necessary by the controller. Derivative action also make unrest for the process, disturbed by small disturbances, which is not desirable. If a phase lead is required in the controller, derivative action should not be tuned more aggressive than given by SIMC.

It is required for the separator dynamics tuning method that the process model is properly scaled such as in Section 3.3. Basis for process tuning is the linearised models for each loop in time domain. Linearised models for water and oil loop is given by Equation (5.1) and the linearised gas pressure loop is given by Equation (5.2). In table 5.2 the calculation formulas for separator dynamics method is presented.

$$h_s(s) = \frac{k}{s(Ts+1)}e^{-\tau s} \tag{5.1}$$

$$h_s(s) = \frac{k}{s(Ts+1)} \tag{5.2}$$

Table 5.2: Calculation formulas of PID parameters by separator dynamics method.

Process type	$h_s(s)$	K_p	T_i	T_d
Integr + time const.	$\frac{k}{(Ts+1)s}$	$min[\frac{1}{kT}, \frac{1}{kc_1(T_c+\tau)}]$	c_2T	T
Integr + time const + del.	$\frac{k}{(Ts+1)s}e^{-\tau s}$	$\frac{1}{kc_1(T_c+\tau)}$	$2c_1(T_c+\tau)$	T

In table 5.2 $c_1 = 10$ and $c_2 = 4$, as for SIMC $T_c = \tau$. For SIMC Skogestad (2004) suggests c = 4 and Haugen (2010) claims that $c = 2$ make the disturbance rejection better. For the integrator plus time constant function $T_c + \tau$ is given from the longest time delay in the related control loops.

5.2.3 Tuning verification

For tuning verification, the process response and phase margins for separator dynamic and SIMC tuned controllers are compared to each other for each loop. In Appendix A process plots for each loop for both PI and PID controller are presented for disturbances of 90 % of maximum magnitude and three different period times for a simulated slug flow. Plots of oil level control for PI and PID controllers tuned with SIMC are also presented. Bode diagrams for each loop both with PI and PID controller are presented in Appendix B.

Both PI and PID controllers keep the process within the limits for any disturbances when they are tuned with separator dynamics method as seen in the process plots. When the period time is long, the controller give maximum output to reject the disturbance, but for shorter periods when maximum valve opening is not required, and the tuning leads to disturbance dampening.

In table 5.3 to table 5.6 the controller parameters for both PI and PID controllers tuned with both SIMC and separator dynamics method are presented.

Table 5.3: Tuning results using SIMC for water, oil and gas PI controllers.

Control loop	$h_s(s)$	K_p	T_i	T_d
Water level loop	$\frac{1.36\cdot10^{-3}}{(6s+1)s}e^{-5s}$	73.6	40	0
Oil level loop	$\frac{5.93\cdot10^{-3}}{(6s+1)s}e^{-5s}$	16.9	40	0
Gas pressure loop	$\frac{7.22\cdot10^{-3}}{(60s+1)s}$	23.1	240	0

Gain and phase margins for PI controllers

An important part of control analysis for single loop control, is to calculate gain margins and phase margin to analyse the stability. Haugen (2010) states that a phase margin $\psi \geq 45°$ and a gain margin of minimum 6 [dB] is required to achieve good process stability. In this report bode analysis is used to calculate loop margins.

When process stability is examined, the the process is linearised. In this case, each loop is linearised around the setpoint, and each process variable is said to only

Table 5.4: Tuning results using SIMC for water, oil and gas PID controllers.

Control loop	$h_s(s)$	K_p	T_i	T_d
Water level loop	$\frac{1.36\cdot 10^{-3}}{(6s+1)s}e^{-5s}$	73.6	40	6
Oil level loop	$\frac{5.93\cdot 10^{-3}}{(6s+1)s}e^{-5s}$	16.9	40	6
Gas pressure loop	$\frac{7.22\cdot 10^{-3}}{(60s+1)s}$	23.1	240	60

Table 5.5: Results of controller tuning using the method based on separator dynamics method for PI control.

Control loop	$h_s(s)$	K_p	T_i	T_d
Water level loop	$\frac{1.36\cdot 10^{-3}}{(6s+1)s}e^{-5s}$	7.36	400	0
Oil level loop	$\frac{5.93\cdot 10^{-3}}{(6s+1)s}e^{-5s}$	1.69	400	0
Gas pressure loop	$\frac{7.22\cdot 10^{-3}}{(60s+1)s}$	2.31	240	0

change with their respective net flow. The actuators have to be linear, and for the water and oil level loops, time delay is taken into account. The PI controller given by Equation (5.3), is used as the controller. The open loop transfer function for water and oil is given by Equation (5.4) and the open loop transfer function for gas is given by Equation (5.5).

$$h_c(s) = \frac{K_p(T_i s + 1)}{T_i s} \tag{5.3}$$

$$L(s) = h_c(s) \cdot h_s(s) = \frac{kK_p(T_i s + 1)}{T_i s^2(Ts + 1)} e^{-\tau s} \tag{5.4}$$

$$L(s) = h_c(s) \cdot h_s(s) = \frac{kK_p(T_i s + 1)}{T_i s^2(Ts + 1)} \tag{5.5}$$

The open loop amplitude is calculated in decibel from Equation (5.6) or in real numbers by Equation (5.7). To find the cross frequency ω_c the magnitude has to be set to 0 in decibels og 1 in real numbers. For calculations of margins for the gas pressure loop, the $e^{-\tau s}$ term is not regarded in the calculations since there is no time delay in this loop.

$$\begin{aligned} |h(j\omega)|[dB] = 20(&\lg|kK_p| + \lg|1 + T_i j\omega| \\ &- \lg|j\omega| - \lg|T_i j\omega| - \lg|1 + Tj\omega| - \lg|e^{\tau j\omega}|) \end{aligned} \tag{5.6}$$

Table 5.6: Results of controller tuning using the method based on separator dynamics method for PID control.

Control loop	$h_s(s)$	K_p	T_i	T_d
Water level loop	$\frac{1.36\cdot10^{-3}}{(6s+1)s}e^{-5s}$	7.36	400	6
Oil level loop	$\frac{5.93\cdot10^{-3}}{(6s+1)s}e^{-5s}$	1.69	400	6
Gas pressure loop	$\frac{7.22\cdot10^{-3}}{(60s+1)s}$	2.31	240	60

$$A(\omega) = |h(j\omega)| = \frac{kK_p\sqrt{1+(T_i\omega)^2}}{\sqrt{1+(T\omega)^2}\cdot T_i\omega} \tag{5.7}$$

Phase for the open loop water level and oil level system is found by Equation (5.8). As well as for the magnitude, phase calculations has to be done in radians. To find ω_{180} the phase is the equal $-\pi$. The only difference is for gas pressure loop is that $\tau = 0$ in the equation.

$$\begin{aligned}\arg(h(j\omega)) &= \arg(kK_p) - \arg(\omega^2) - \arg(1 - Tj\omega) + \arg(1 + T_ij\omega) - \tau\omega \\ \arg(h(j\omega)) &= 0 - \phi - \arctan(Tj\omega) + \arctan(T_ij\omega) - \tau\omega\end{aligned} \tag{5.8}$$

Table 5.7 contains crossover frequencies ω_c and ω_{180} as well as gain and phase margins for the open loop with controller parameters calculated with SIMC. In table 5.8 margins and crossover frequencies for open loop transfer function for the loop with parameters calculated with separator dynamics method are presented.

Table 5.7: Gain and phase margins for each PI controlled loops based on SIMC tuning.

Process type	Gain margin [dB]	Phase margin (ψ)	ω_{180}	ω_c
Water level loop	5.05	20.9°	0.1394	0.0088
Oil level loop	5.05	20.9°	0.1394	0.0088
Gas pressure loop	35.9	17.0°	0.3580	0.0390

For loops controlled by a SIMC tuned controllers, both gain and phase margin are low which may lead to a stability problem for the process when large disturbances occur and the process gain is high. Process plot in Appendix A show that stability for high process gain may be a problem for loops tuned with SIMC. Process dynamics method give large gain and phase margins, and the plots also shows that there are no stability problems for loops tuned with this method.

Gain and phase margins for PID controllers

Table 5.8: Calculated gain and phase margins for each PI controlled loops, based on tuning by separator dynamics method.

Process type	Gain margin [dB]	Phase margin (ψ)	ω_{180}	ω_c
Water level loop	26.8	70.9°	0.1588	0.0103
Oil level loop	26.9	70.9°	0.1588	0.0103
Gas pressure loop	inf	34.0°	inf	0.0135

We will now derive the stability for the process when the loops are controlled by PID controllers. Equation (5.9) describe the transfer function for the PID controller. The open loop system for water and oil is given by Equation (5.10) and the gas pressure loop i given by Equation (5.11).

$$h_c(s) = \frac{K_p(T_i s + 1)(T_d s + 1)}{T_i s} \tag{5.9}$$

$$L_{liq}(s)| = h_c(s) \cdot h_s(s) = \frac{kK_p(T_i s + 1)(T_d s + 1)}{T_i s^2(Ts + 1)} e^{-\tau s} \tag{5.10}$$

$$L_g(s) = h_c(s) \cdot h_s(s) = \frac{kK_p(T_i s + 1)(T_d s + 1)}{T_i s^2(Ts + 1)} \tag{5.11}$$

Just as for PI controlled loops the magnitude has to be calculated to find the process stability. From Equation (5.12) the magnitude of the system, for given frequencies, can be calculated in decibel or in real number Equation (5.13). By setting the magnitude to 1 in real number, or 0 in decibel, crossover frequency ω_c in radians per second is found. For calculations for gas pressure loop, the time delay term $e^{-\tau s}$ in the equations is not regarded.

$$\begin{aligned} |h(j\omega)|[dB] = 20(&\lg|kK_p| + \lg|1 + T_i j\omega| + \lg|1 + T_d j\omega| \\ &- \lg|j\omega| - \lg|T_i j\omega| - \lg|1 + Tj\omega| - \lg|e^{\tau j\omega}|) \end{aligned} \tag{5.12}$$

$$A(\omega) = |h(j\omega)| = \frac{kK_p\sqrt{1 + (T_i\omega)^2}\sqrt{1 + (T_d\omega)^2}}{\sqrt{1 + (T\omega)^2} \cdot T_i\omega} \tag{5.13}$$

Phase calculations for open loop water and oil control is calculated by Equation (5.14). The gas pressure loop is calculated by the same formula but $\tau = 0$ in this case. To find ω_{180} we have to set $\arg(h(j\omega)) = -\pi$. Calculations are done in radians.

$$\arg(h(j\omega)) = \arg(kK_p) - \arg(\omega^2) \\ - \arg(1 - Tj\omega) + \arg(1 + T_i j\omega) + \arg(1 + T_d j\omega) - \tau\omega \quad (5.14)$$

$$\arg(h(j\omega)) = 0 - \phi - \arctan(Tj\omega) + \arctan(T_i j\omega) + \arctan(T_d j\omega) - \tau\omega$$

In table 5.9 cross frequencies ω_c and ω_{180} and margins for the PID controlled system tuned with SIMC are presented. Table 5.10 presents the same values for PID controllers tuned using separator dynamics method.

Table 5.9: Gain and phase margins for each control loops based on SIMC tuning and PID controllers.

Process type	Gain margin [dB]	Phase margin (ψ)	ω_{180}	ω_c
Water level loop	9.43	47.3°	0.2974	0.100
Oil level loop	9.44	47.5°	0.2974	0.099
Gas pressure loop	inf	87.6°	inf	0.100

Table 5.10: Calculated gain and phase margins for each PID controlled loops, based on tuning by separator dimensions.

Process type	Gain margin [dB]	Phase margin (ψ)	ω_{180}	ω_c
Water level loop	29.9	75.2°	0.3126	0.0103
Oil level loop	29.9	75.2°	0.3126	0.0103
Gas pressure loop	inf	76.0°	inf	0.0173

For the PID controller, SIMC performers gain and phase margin within the requirements of Haugen (2010). However for the largest disturbances, simulations show that there are some stability problems because the process gain will be higher than for the linearised system. The PID controllers tuned with separator dynamics method, performs large gain and phase margins and there are no stability problems when this method is used.

5.2.4 Discussion about separator dynamics method and comparing to SIMC

From the stability analysis it is shown that it yields good results for loop margins. However the following has to be fulfilled to have satisfying result for the tuning rule:

- The tuned controller has to be able to reject any disturbance.
- Better damped control loop compared to controller tuned with SIMC or Ziegler & Nichols method.

- Less and more smooth use of the actuator compared to SIMC or Ziegler & Nichols method.

As mentioned earlier, the SIMC method is developed to make fast control and provide the loop to have sufficient gain and phase margin at the same time. For separator control, the controller tuned by the SIMC rule, will make the process change too fast, but maintain the setpoint much better than separator dynamics method. Even though the margins for the open loop is within the acceptable area, there will be some unrest in the process. However in general the SIMC tuning rule performs a good starting point for the tuning.

The lowered gain and the increased integral time compared to both SIMC and Ziegler & Nichols method leads to more smooth use of the actuator. As for the controller tuned with SIMC or Ziegler & Nichols methods, the controller tuned with separator dynamics method will saturate the controller output. However for smaller disturbances the controller will not make the actuator work as hard a a SIMC or Ziegler & Nichols tuned controllers.

The main property of the separator dynamics method is to make smoother disturbance rejection, but still reject any disturbance regardless size and period time. A controller tuned with SIMC will in most cases be able to reject large disturbances, but the rejection is not smooth. For the process analysed here, the disturbance rejection is close to unstable, cause the gain tuned is tuned too high using SIMC.

For separator dynamics method the integral time is increased by a factor of 20 and gain is reduced by a factor of 10 for loop given by Equation (5.1), which is the water and oil loop in this particular case. For the process given by Equation (5.2) the process gain is multiplied with the time constant T for controller gain calculation. The integral time is the time canstant multiplied with $c = 4$, which is the same as for SIMC. This method may also be used for processes dominated by time constant.

Simulations presented in Appendix A show that the controller tuned with the separator dynamics method for separator control is able to reject any disturbance for the given case. For verification purposes, the process gain has been reinforced to test for a plant where the liquid level rises faster. The result is that the separator dynamics tuning rule gives good result also for faster changing plants. For plant where the gain is high og the noise effect is significant, the calculated gain may be too high in terms of amplifying the noise. Therefore the controllers may be gain scheduled to avoid too large noise effect when high controller gain is not needed.

5.3 How the limits affect the process

In this section we will look at how the process limitations, derived in Section 3.4, actually affects the process. We will also look at what is done to reduce the impact of control performance.

Impact due to change in other process variables

In Section 3.4.6 we looked at limitations due to changes in other process variables. For the separator process, all the controlled process variables affect each other in some manner. Outflow for gas, oil and water are all dependent on the separator gas pressure. Oil level is directly affected by water level, and oil level affects the gas pressure as a nonlienar function.

The process limitations is mostly related to how well the system has to follow the setpoint, and not how fast it can operate. The fastest loop is the gas pressure loop, which means that slower changes in water and oil level make less limitations for how fast the gas pressure loop can be controlled. However gas pressure affects the rate of oil and water flow leaving the separator. Steady gas pressure leads to steady outflow, and that decides how well the process can reject disturbances.

For the single loop control, the controllers only evaluate the process variable controlled in the specific loop. A disturbance in one loop, will lead to reaction in all the other loops. That means that the loop where disturbances occur, has to make the change in the process variable as small as possible to provide unwanted reaction in the affected control variables. A rule of thumbs is that more process variables depending on each other requires less variation in each one.

Control structure based on floating setpoint, for oil level loop, is a method for reduce impact due to change in other process variables. By letting oil level setpoint be decided by water level, makes the oil level loop regard the water level implicitly. This method provides the oil level to react less for disturbance occurring in the water level loop.

Impact due to noise

Our process has some noise, which is modelled as white noise with a peak to peak value of 1 % of the measurement signal. The way noise affect the process is that it limits the controller gain. The process loop is not allowed to reinforce the noise over 1 % , and that means the maximum controller gain has to ensure that Equation (5.15) is fulfilled.

$$|h_c(j\omega)| \cdot |h_a(j\omega)| \cdot |h_p(j\omega)| \cdot |h_m(j\omega)| \leq 1 \tag{5.15}$$

To increase the controller gain and also the total control loop gain, a filter is introduced to reduce the noise effect on the measurement signal. In Section 4.2.6 and Section 4.4 the design of the low pass filer is derived.

When a low pass filer is introduced in the system the gain and phase margin will be changed. The separator process is relatively slow. A filter that reduces the effect of noise, does not have to have cross frequency close to the process closed loop cross frequency without filter, and still reduce the high frequently noise effectively. The well designed filter makes it possible to rise the controller gain and still provide good gain margin and phase margin in the system. Raised total gain leads to better disturbance rejection in the control loop.

Impact due to time delay

In the water and oil level loop, time delay is prominent. For both water level loop and oil level loop, there is time delay affecting the actuator response. The gas pressure loop is not modelled with time delay, only a large time constant. In Section 3.4.2 impact due to time delay was derived.

For systems where the time delay is very prominent, it will be the main limit for achievable bandwidth in the system. To increase the bandwidth a bit, there is possible to introduce a zero in the time delayed loop. By introducing the derivative term in the controller it is to the system to achieve a phase lead. However at a certain point there is not possible to have phase margin in the system.

Time delay is a problem when separator starts filling. First it takes some time before the actuator reacted. When the actuator finally reacts, the process input compensate for the time the actuator did not react. If the controller is tuned badly the process can start swinging uncontrolled.

Impact due input constraints

In Section 3.4.4 input constraints was derived. The main topic in this section is input for stabilisation. It was stated that $|u| < 1$ for $|v| = 1$ is necessary to stabilise the plant and to reject disturbances.

For the largest disturbances in the separator process, the controller output and the actuator will saturate. However there is a rate of 1.2 between input and output flow, for the process running around setpoint. Even though there is a buffer, the rate between input and output can become less than 1. It may happen if the gas pressure becomes low, and it makes the controller unable to entirely reject disturbances. However the gas pressure dynamic will maintain the outflow flow capacity again when gas volume has become smaller and gas pressure has increased.

Impact due to phase lag

We resume Section 3.4.3 where limitation due to phase lag was derived. For the given system, tests and bode analysis show that the system is not limited by phase lag. However there may be a problem for plants with higher magnitude. For these plants there may by implemented a PID controller to provide phase lead in the system.

For the tuning performed both by SIMC and separator dynamics method, there is no problem to achieve the stability bound of $\omega_c < \omega_u$. Especially separator dynamics method performs large gain an phase margin. However if the plant magnitude is high, there may not be possible to both achieve the stability bound and control the process within limits. In this case phase lag limits the process performance. A controller where the derivative term is introduced, may help provide a phase lead, and if it is not sufficient the control structure may be changed. Feed forward control derived in Section 4.2.2 may be used for systems where there is difficult to achieve phase lead.

5.4 Resonances

In this section we will derive the resonance frequencies and analyse the impact to the system. The resonance frequency will give an upper limit for the bandwidth.

There may be no resonances in the system which is preferable.

5.4.1 Resonance calculation

Balchen et al. (2003) shows a method for resonance frequency calculation based on a second order system. An example based on the mass spring damper system, is shown to illustrate the calculations given by Equation (5.16). This method also works for systems of order higher than two.

$$
\begin{aligned}
\frac{1}{m}u(t) &= \ddot{y}(s) + \frac{f}{m}\dot{y}(t) + \frac{k}{m}y(t) \\
h(s) &= \frac{\frac{1}{m}}{s^2 + \frac{f}{m}s + \frac{k}{m}} = \frac{\frac{1}{m}}{(s-\lambda_1)(s-\lambda_2)} = \frac{\frac{1}{m}}{(s+\alpha+j\beta)(s+\alpha-j\beta)} \\
h(s) &= \frac{\frac{1}{m}}{s^2 + 2\alpha s + (\alpha^2+\beta^2)} = \frac{\frac{1}{m}}{s^2 + 2\zeta\omega_0 s + \omega_0^2} = \frac{\frac{1}{k}}{1 + 2\zeta\frac{s}{\omega_0} + (\frac{s}{\omega_0})^2} \qquad (5.16)
\end{aligned}
$$

Table 5.11 contains resonance parameters, how to calculate each value and which function each parameter have.

Table 5.11: Formulas for resonance calculation.

Symbol	Formula	Description
α	$\zeta\omega_0$	Absolute damping factor
β	$\sqrt{\omega_0^2 - \alpha^2}$	Frequency of oscillation
ω_0	$\sqrt{\alpha^2 + \beta^2}$	Undamped resonance frequency
ζ	$\sin\varphi$	Relative damping factor
φ	$\arcsin(\zeta)_0$	Phase lag

To find the resonance frequency and its property, we have to calculate the poles of the system. All processes do not have complex poles, but they have to be complex to make the system swing. In Figure 5.3 a graphical representation for poles and related values for resonance is shown.

We will now look at the simplified and linearised process description for gas pressure. In this case the gas volume is constant which means that the pressure only depends on net gas flow. Below the simplified system is given in the time domain.

$$
\begin{aligned}
h_{c_g}(s) &= \frac{(T_i s + 1)K_p}{T_i s} \\
h_{p_g}(s) &= \frac{k_{p_g}}{s} \\
h_{a_g}(s) &= \frac{k_{a_g}}{Ts + 1}
\end{aligned}
$$

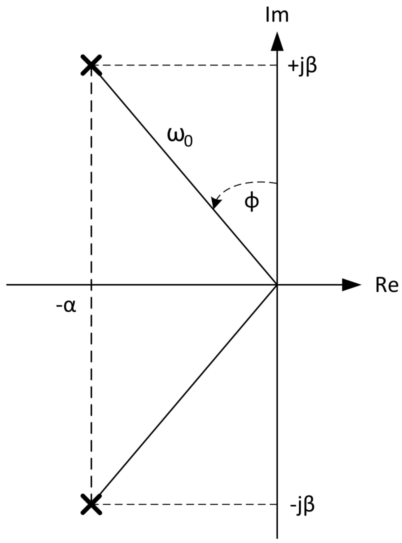

Figure 5.3: Graphical representation of complex conjugated poles and values related to resonance calculation.

Now we calculate the open loop transfer function based on the process model given above.

$$L_g(s) = h_{c_g}(s) \cdot h_{a_g}(s) \cdot h_{p_g}(s)$$
$$L_g(s) = \frac{K_p(T_i s + 1)}{T_i s} \cdot \frac{k_{a_g}}{Ts + 1} \cdot \frac{k_{p_g}}{s}$$
$$L_g(s) = \frac{k_{p_g} k_{a_g} K_p (T_i s + 1)}{(Ts + 1) T_i s^2}$$

We are most interested in resonance in the closed loop system when searching for resonances. Below calculations for the closed loop system is done based on the open loop transfer function given for gas pressure. Resonance frequency is found by calculating the poles of the system. The complex conjugated pair give the resonance factors shown in Figure 5.3.

$$T_g(s) = \frac{L(s)}{1 + L(s)}$$
$$T_g(s) = \frac{\frac{k_{p_g} k_{a_g} K_p (T_i s+1)}{(Ts+1) T_i s^2}}{1 + \frac{k_{p_g} k_{a_g} K_p (T_i s+1)}{(Ts+1) T_i s^2}}$$
$$T_g(s) = \frac{k_{p_g} k_{a_g} K_p (T_i s + 1)}{(Ts + 1) T_i s^2 + k_{p_g} k_{a_g} K_p (T_i s + 1)}$$

The method used for resonance calculations for the gas pressure, can not be used for the liquid levels since there is a time delay in the model. The reason is that the time delay give a infinite number of poles and zeros, which makes it impossible to make calculations for these systems. Though if the time delay can be neglected, it is possible to use the described method. For the given process, time delay is prominent. That means we have to make a bode digram of the closed loop model to find the resonance frequency and the magnitude of resonance.

In Appendix B closed loop bode diagrams for all three loop are presented. There are no signs of any peaks in magnitude of any loop. The controller gain can be increased without signs of peaks in the magnitude. That means resonances will not be a problem for the process.

Equation (5.17) gives the closed loop transfer function for the liquids. In this case the nonlinear valve is linearised, and we operate at a certain level which means that the liquid surface is close to constant.

$$T_l(s) = \frac{k_{p_l} k_{a_l} K_p (T_i s + 1) e^{-\tau s}}{(Ts+1) T_i s^2 + k_{p_l} k_{a_l} K_p (T_i s + 1) e^{-\tau s}} \tag{5.17}$$

6 Discussion and further work

In this Chapter we will discuss the work done in the project, the validity of solutions and usefulness of the suggestion for how the process control for a separator should be designed and tuned. The last part contains a section that discuss further work to be done in later projects.

6.1 Discussion

What is the value and usefulness of the solutions found in this project? For which plants may the control structure be used, is the tuning rule useful in practice and is the process model valid for the purpose? That is questions discussed in this section.

The model used for control structure development is not accurate. More uncertainties there are in the model leads to larger uncertainty in the result. The problem is that it is impossible to make a perfect model, since some uncertainty can not be regarded. In special cases temperature changes, flashing, more complex flow models and a more complex disturbance model should be regarded. In Schei et al. (1991) flashing and large changes in temperature is described as a challenge for some processes where disturbances occur as a large transient. However to find a tuning method for a unspecified lying separator, the model used in this case, which based on mass balance, ideal gas law and plant geometry, is sufficient for analysis.

Disturbance caused by slug flow, which is the main focus in this report, is not easy to model perfectly. Slug can differ from plant to plant. Similarities between the different slug flows is in the flow pattern. Riser slugging, which often leads to the most prominent slug flow, often occur in a specific pattern and a given period time. Disturbance, which is modelled as square pules in this report, is a approximation of the riser slugging flow pattern.

Usefulness of the results can be seen both from a economical and a environmental point of view. The suggested control structure and tuning, which provides better disturbance handling compared to single loop feedback control with fixed setpoints and controllers tuned with SIMC, make process shutdown less frequently when large disturbances occur. Longer time in production and a plant less influenced by disturbance means better profit. From an environmental point of view it also means that there will be less unnecessary emissions from process start up and shut down.

The suggested control structure and tuning, may be introduced in both new and existing systems. Changes in control structure compared to fixed setpoint feedback control are not large, but in cases where disturbances occur with long amplitude and high magnitude the changed structure performs a major improvement. The filter introduced in the loop will also reduce the amplification of measurement noise in the system. Limitation imposed by measurement noise will in this case be reduced.

Separator dynamics tuning method may be used both for ordinary feedback control and floating setpoint for oil level loop. The advantage of this method compared to tuning rules like SIMC or Ziegler & Nichols is that is is developed for the separators specificity and not for a general plant.

Model adaptive control or MRAC was derived in Section 4.2.5. An adaptive controller based on a nonlinear MRAC scheme may be used to control the system. The method requires an accurate reference model for the system, which means that the model may have to regard more dynamic. This method may work as a more complex type of gain scheduled controller, but is may be a bit to complex to calculate for the pressure controller.

For single loop control, each process variable is controlled by one loop only. When introducing MPC there is a possibility to weight more than one process variable at the time to control the process. A large benefit would be to both weight both volume and liquid level in the controller. This is a problem for single loop PID control. In Imsland, Kittilsen & Schei (2010) a similar problem to the problem derived in this project is solved using a nonlinear MPC control. For this solution there may be possible to encounter both oil volume and oil level at the same time, which is impossible for single loop PI(D) control.

6.2 Further work

There are still a lot of work to be done by separator control with respect to large disturbances. More solutions may be found using feed forward control between SISO control loops. More or less the same task could also be given for a MIMO system. Controllers used for such control could be MPC or LQ-controller in cooperation with a Khalman filter which is called LQG.

Even though the plant and tuning dampens disturbances, large disturbance rejection leads to disturbances also for process parts later in the plant. A control structure designed for disturbance damping in a separator train could be analysed in a later project. However there have been done analysis of this in earlier projects, like for instance the master thesis of Per Morten Hellervik from 2002.

In this project the analysis regards the process variables, water level, oil level and gas pressure. The rate of separation is not considered, the only demand was to keep the process calm the maintain separation. For a later project, there may be interesting to analyse the effect on the separation rate when slug flow arrive. Maybe the tuning should be a bit different if separation rate is regarded.

Changes in temperature and flashing is not regarded in the analysis and is a remaining task. It is known that it will make some changes to the dynamic, and in some plant it is important dynamic. A tuning method that take these parameters into account may perform even better. Temperature changes may also lead to a more complex gain scheduling function if the control loop requires a gain scheduled controller.

An economic analysis for the benefit of introducing the tuning method for the industry, and introducing the floating setpoint for the oil controller is not presented here. The industry may require economical analysis before it can be introduced.

7 Conclusion

Control structure and tuning for a three phase separator has been examined in this project. A suggestion for how the control structure and a tuning method should be developed for rejection of large disturbances have been developed. Both the examined and developed structure is a SISO system developed for a a separator constructed as a lying cylinder.

The process model is derived from mass balance, ideal gas law and separator geometrical shape. Phenomenons that is not taken into account is flashing and temperature changes. This would lead to a more complex model, which is not required for control structure and process tuning development.

An ideal PI/PID controller described in the control literature, is not able to control a system in a real plant. Without the tools integrator wind-up and derivative filter, the PID controller would not be able to control the systems. Gain scheduler may be required if the process reacts fast and the control loop is a bit slow. Wilhelmsen (2012) derives a method where gain scheduler is used to prevent overflow in the separator, when large disturbances occur and a PI /PID controller can not work with high gain cause the noise effect is prominent. For most of the cases analysed in this report, controller gain is not reduced by the reinforcement of noise. The introduction of the low pass filtered measurement signal, make the controller in most cases able to run with sufficiently high gain. In most cases PI or PID controller is sufficient for process control.

The low pass filer introduces a new time constant to the system. The filter should therefore not remove more noise than necessary to reduce the phase lag introduced by the filter. Other limitations that may limit the process performance, if the process work fast, are time delay and input constraints.

Single loop feedback control with fixed setpoint is in many cases sufficient to control the process. Problems arise for plants where process changes is fast, or the limits for process variables are strict. The suggested control structure, where oil level setpoint is given by actual water level, makes a practical volume control for oil level that accounts process limits. For water level and gas pressure control loop, fixed setpoint feedback control works well.

The separator process requires calm handling, and both SIMC and the method described in Balchen et al. (2003), provide fast control within the stability limits but no sufficient control. Both methods make the control loop work to fast, and with gain and phase margin within the stability limits for the linearised system. But under extreme conditions margins are not sufficiently large. However both methods give good guidelines for how a tuning method can be derived.

Separator dynamics method is derived from SIMC. The tuning method meets the requirement of a well damped system and the required gain to reject any disturbances. It provides good phase and gain margin for both PI and PID controllers. The tuning rule is tested and verified for more than one separator, and it works well for the single loop control as well as controllers with floating setpoint.

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A Plots

Process plots

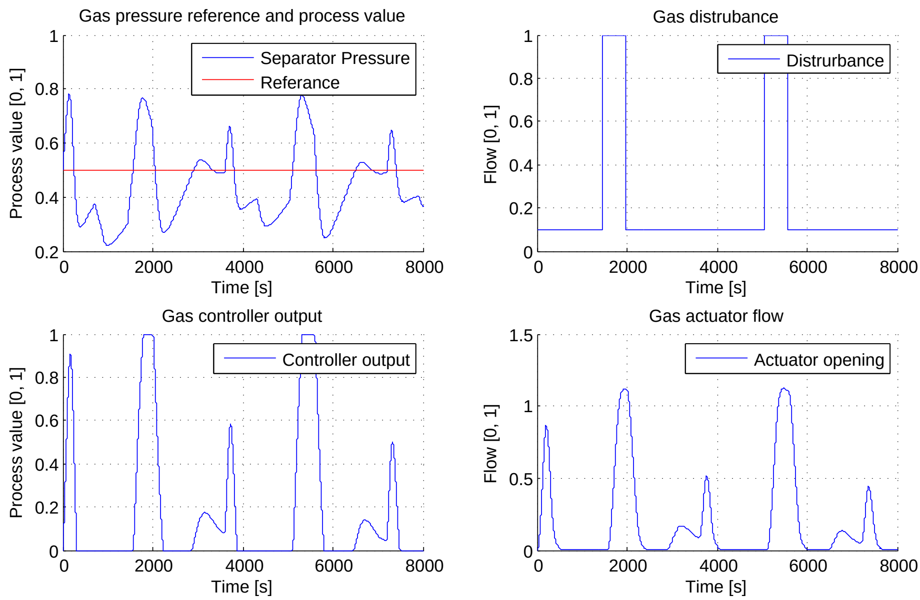

Figure A.1: Plot of the separator pressure when the process is exposed to large disturbances with period time of one hour and a variation of 90 % of maximum disturbance magnitude. The process is controlled by the floating oil reference structure.

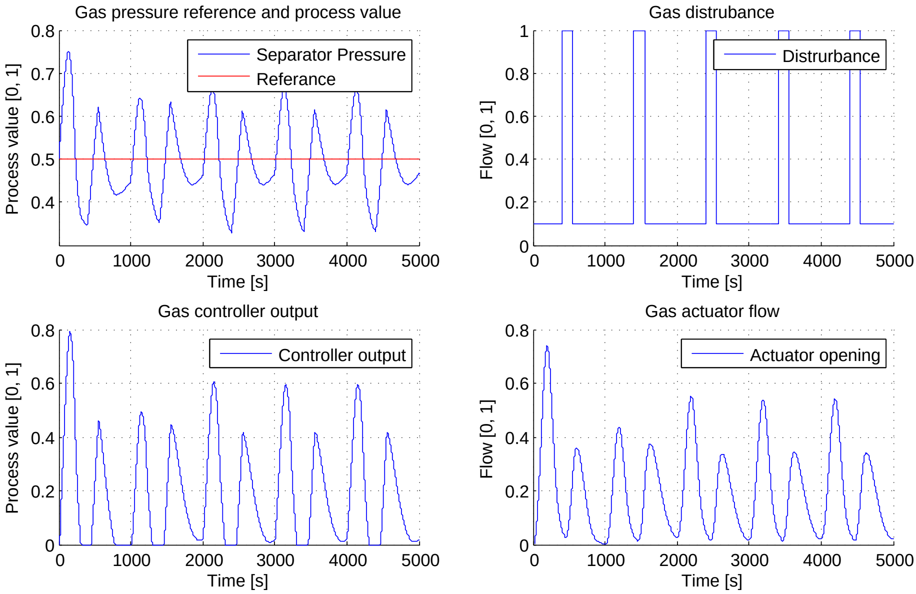

Figure A.2: Plot of the separator pressure for the process exposed to large disturbances with period time of 1000 seconds and a variation of 90 % of maximum disturbance magnitude. The process is controlled by the floating oil reference structure.

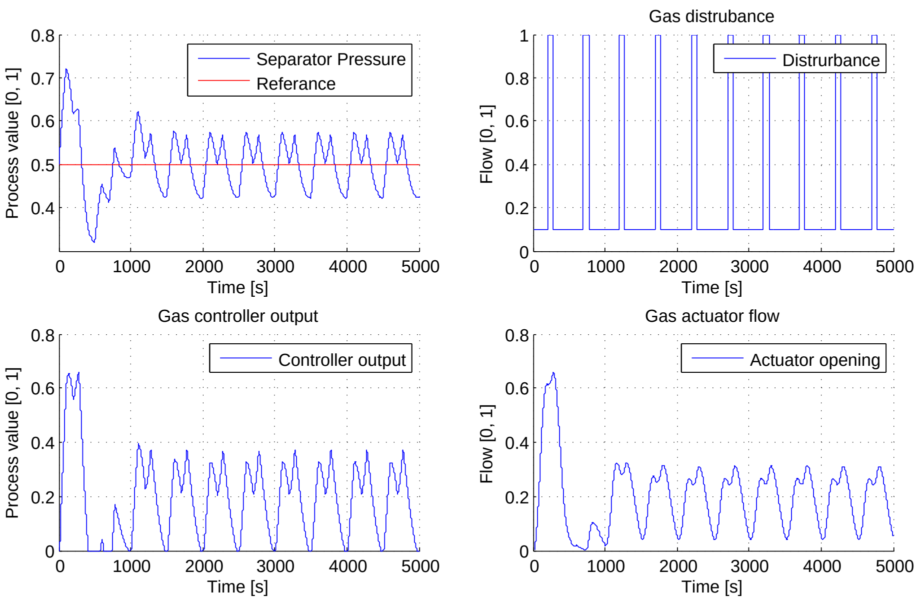

Figure A.3: Plot of the separator pressure for the process exposed to large disturbances with period time of 500 seconds and a variation of 90 % of maximum disturbance magnitude. The process is controlled by the floating oil reference structure.

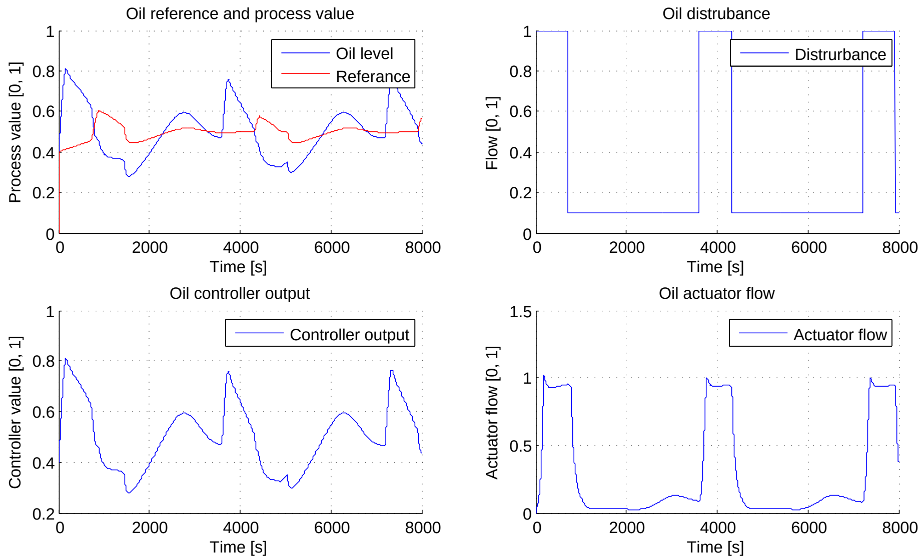

Figure A.4: Plot of the oil level in the separator for the process exposed to both oil and water disturbances of 90 % variation of maximum disturbance magnitude and a period time one hour. The process is controlled by the floating oil reference structure.

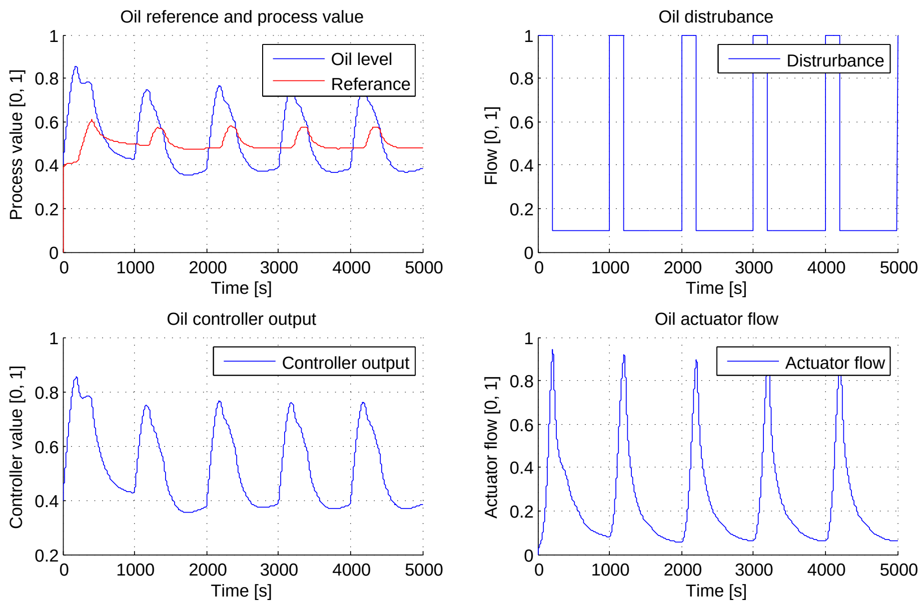

Figure A.5: Plot of the oil level in the separator for the process exposed to both oil and water disturbances of 90 % variation of maximum disturbance magnitude and a period time 1000 seconds. The process is controlled by the floating oil reference structure.

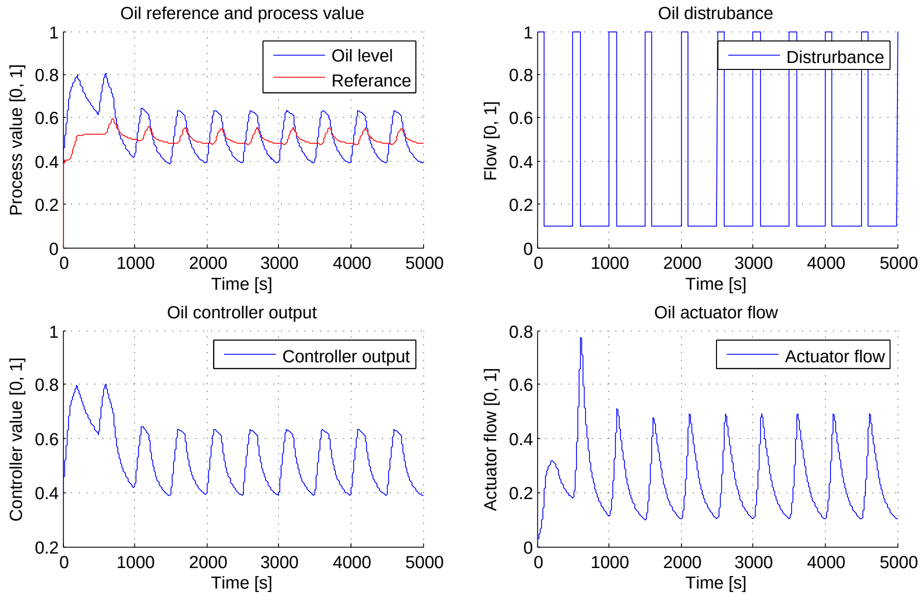

Figure A.6: Plot of the oil level in the separator for the process exposed to both oil and water disturbances of 90 % variation of maximum disturbance magnitude and a period time 500 seconds. The process is controlled by the floating oil reference structure.

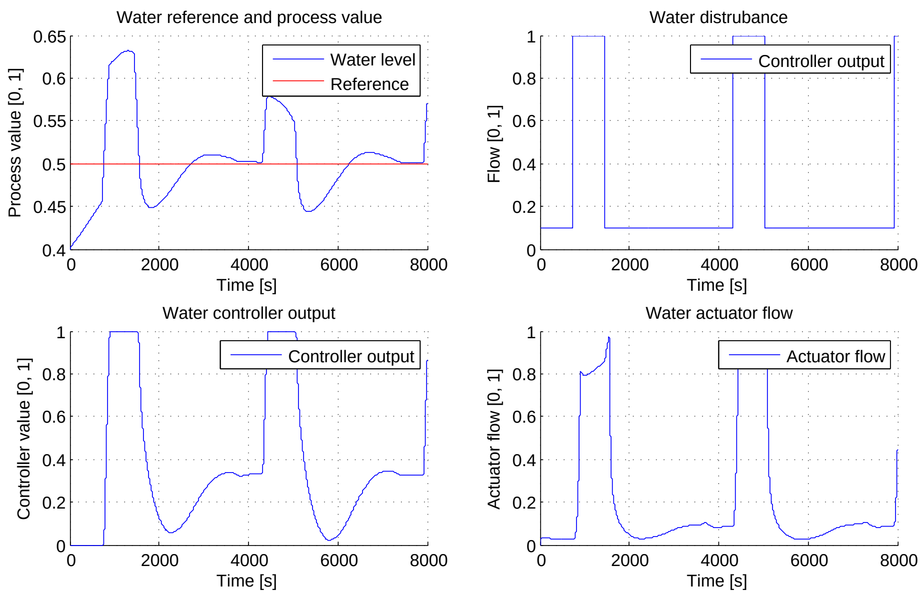

Figure A.7: Process plot of the water in the separator when the process is exposed to disturbances of 90 % variation of maximum disturbance magnitude and a period time of one hour. The process is controlled by the floating oil reference structure.

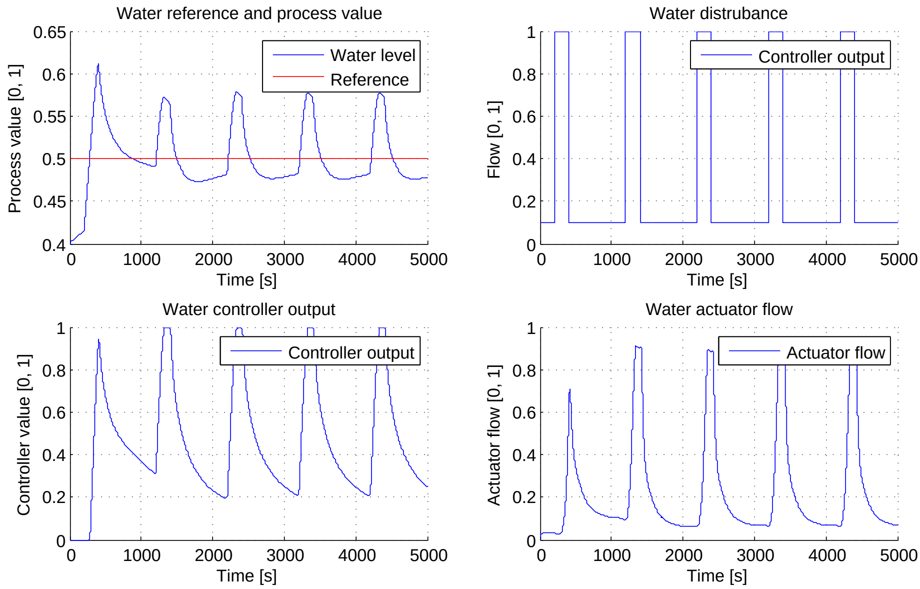

Figure A.8: Process plot of the water in the separator when the process is exposed to disturbances of 90 % variation of maximum disturbance magnitude and a period time of 1000 seconds. The process is controlled by the floating oil reference structure.

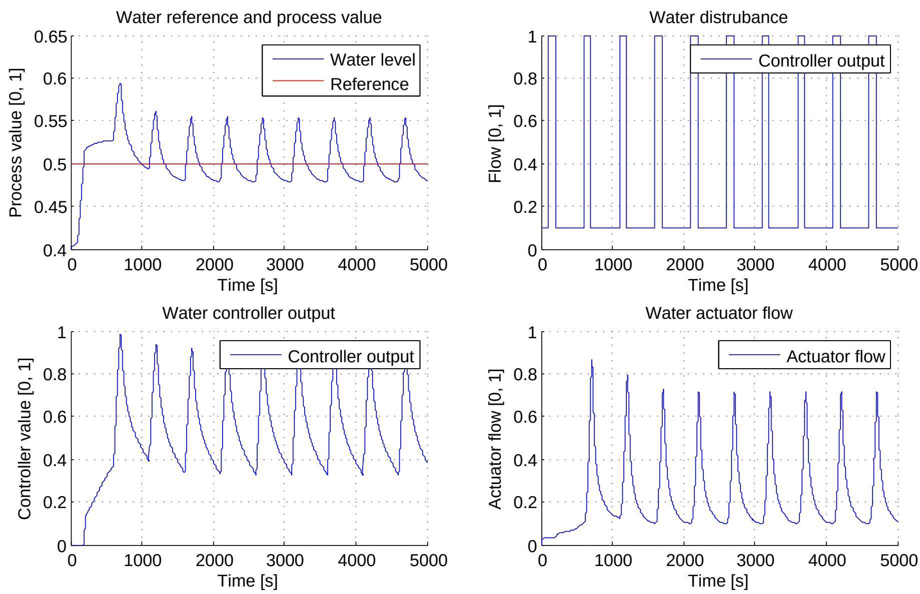

Figure A.9: Process plot of water level when the process is exposed to disturbances of 90 % variation of maximum disturbance magnitude and a period time 500 seconds. The process is controlled by the floating oil reference structure.

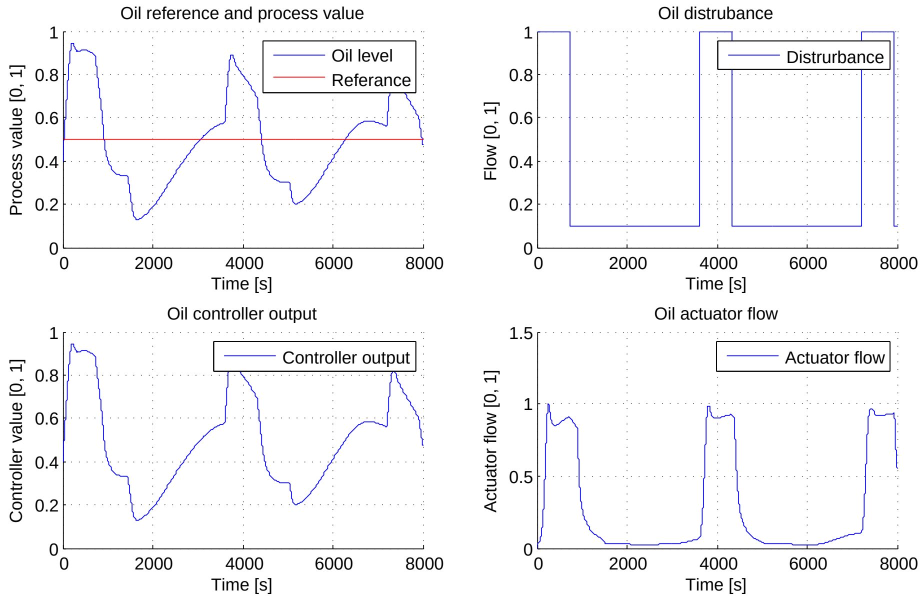

Figure A.10: Plot of the oil level in the separator for the process exposed to both oil and water disturbances of 90 % variation of maximum disturbance magnitude and a period time one hour. The control structure is the base case single loop fixed setpoint.

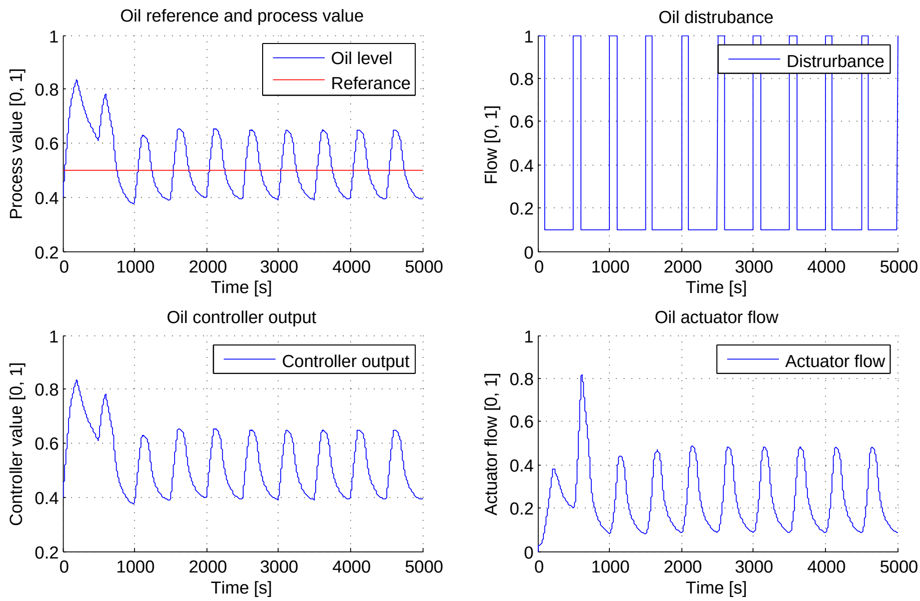

Figure A.11: Plot of the oil level in the separator for the process exposed to both oil and water disturbances of 90 % variation of maximum disturbance magnitude and a period time 500 seconds. The control structure is the base case single loop fixed setpoint.

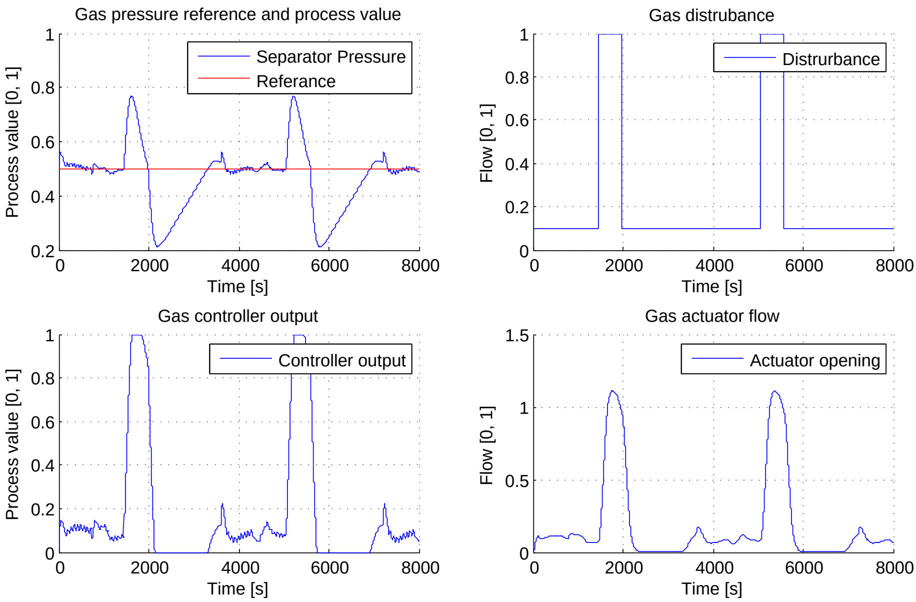

Figure A.12: Plot of the separator pressure for the process exposed to large disturbances with period time of one hour and a range of 90 % variation of maximum disturbance magnitude. The controller is tuned with SIMC. The process is controlled by the floating oil reference structure.

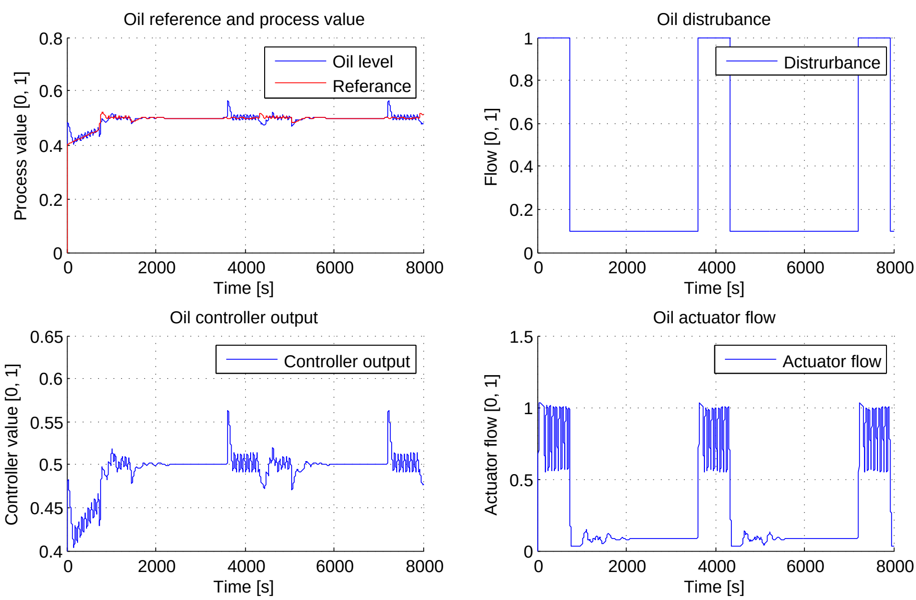

Figure A.13: Plot of the oil level in the separator for the process exposed to both oil and water disturbances of 90 % variation of maximum disturbance magnitude and a period time one hour. To tune the controller SIMC tuning rule is used. The process is controlled by the floating oil reference structure.

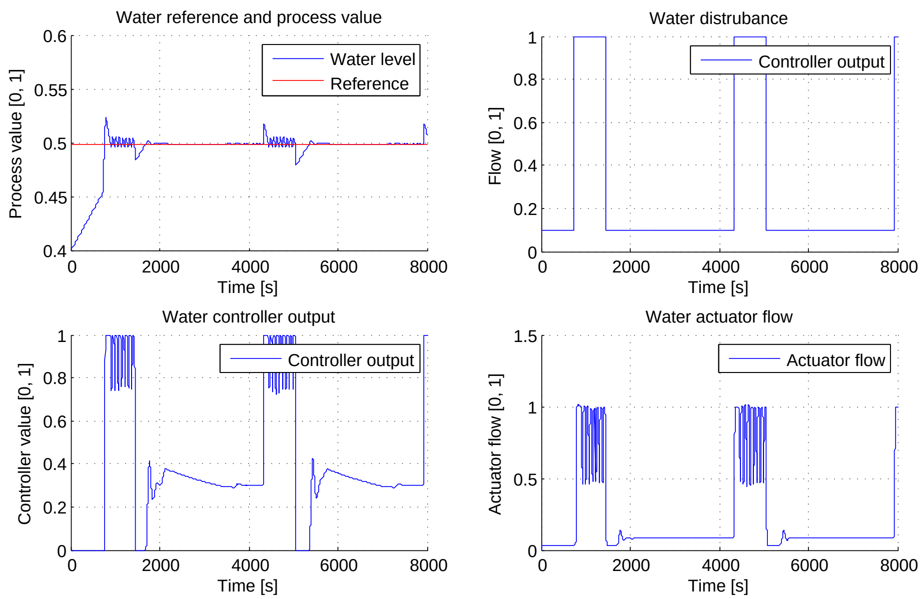

Figure A.14: Process plot of water level when the process is exposed to disturbances of 90 % variation of maximum disturbance magnitude and a period time of one hour. In this case the SIMC tuning rule is used to tune the controller. The process is controlled by the floating oil reference structure.

B Bode diagrams

Open and closed loop bode plot for each control loop

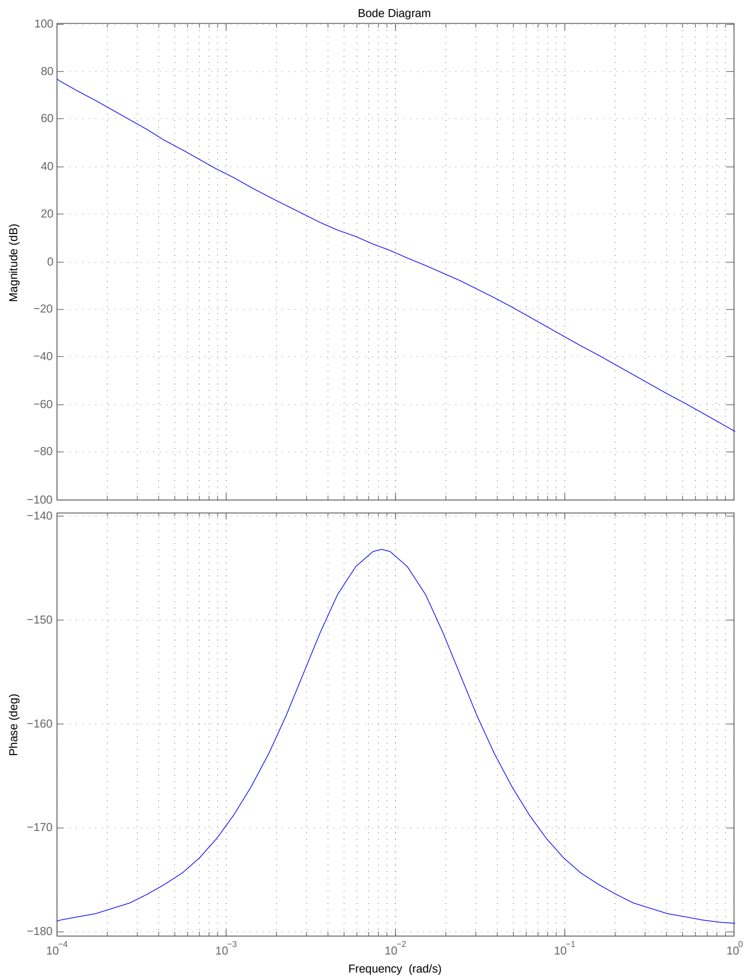

Figure B.1: Bode diagram of the open loop gas pressure transfer function. The controller is a linear PI controller tuned with separator dynamics method.

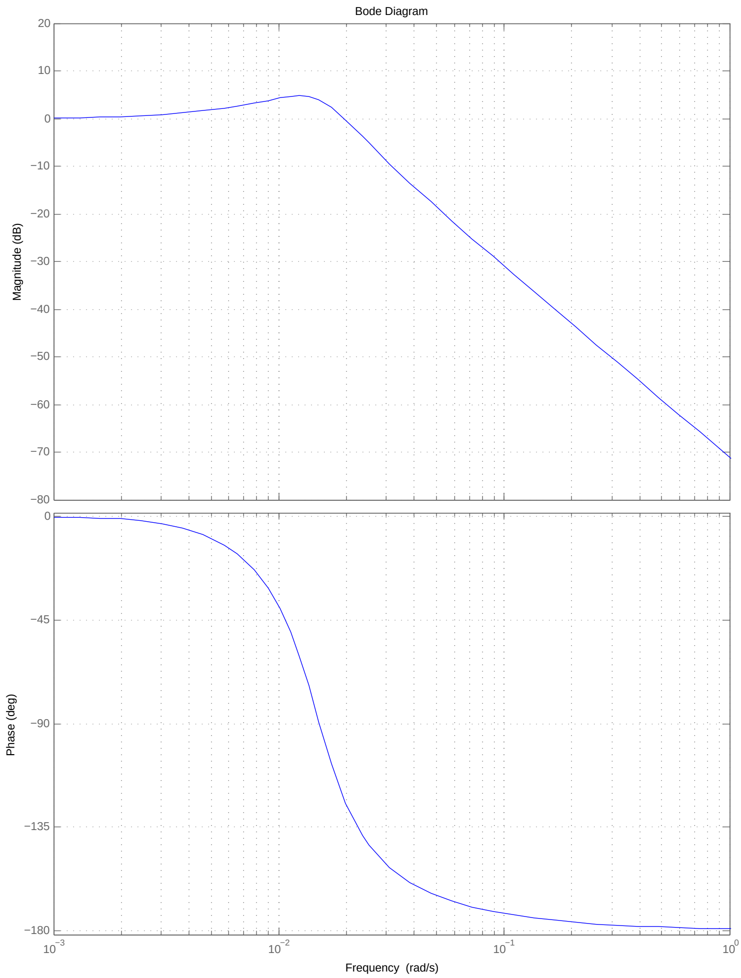

Figure B.2: Bode diagram of the closed loop transfer function for the gas pressure loop. The system is controlled by a linear feedback PI controller tuned with separator dimensions method.

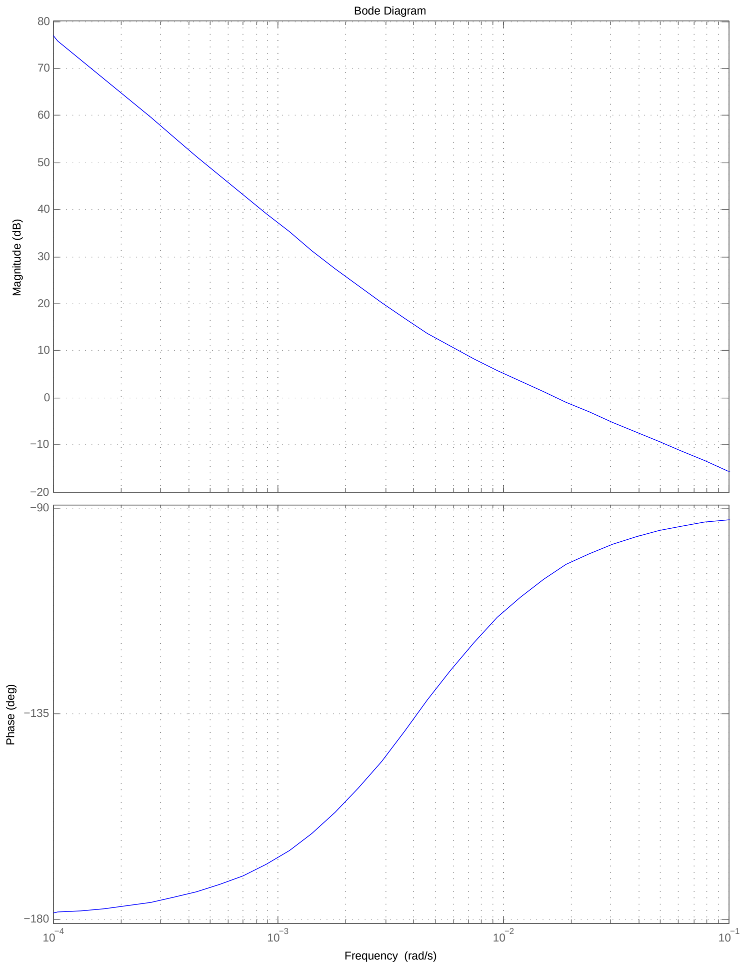

Figure B.3: Bode diagram of the open loop gas pressure transfer function. The controller is a linear PID controller which is tuned with separator dynamics method.

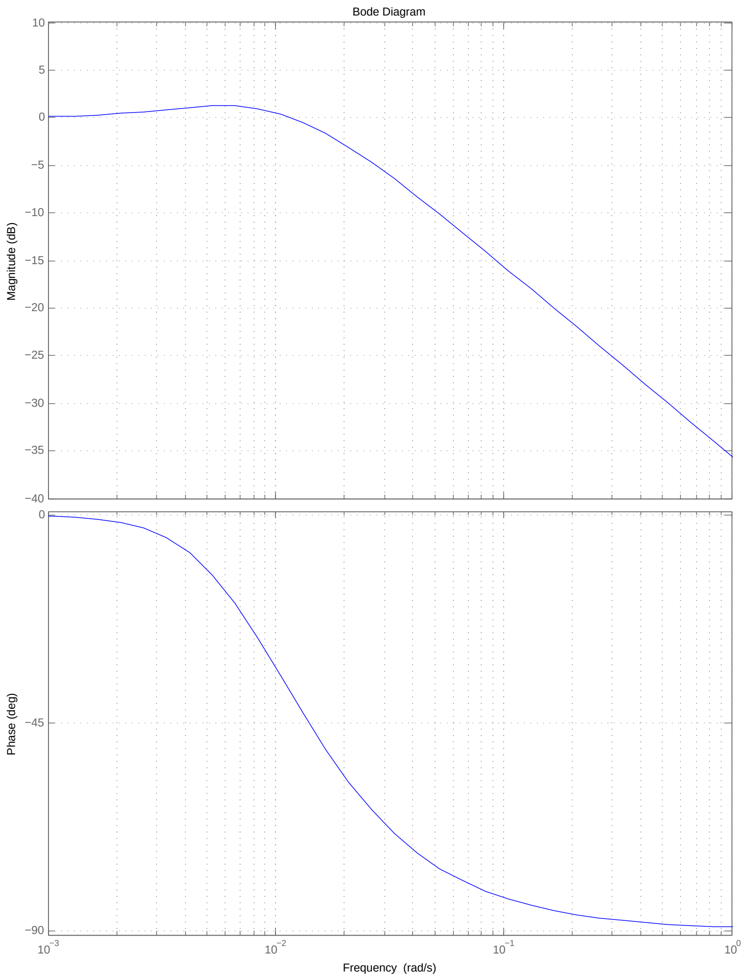

Figure B.4: Bode diagram of the closed loop transfer function for the gas pressure loop. The system is controlled by a linear PID controller tuned with separator dynamics method.

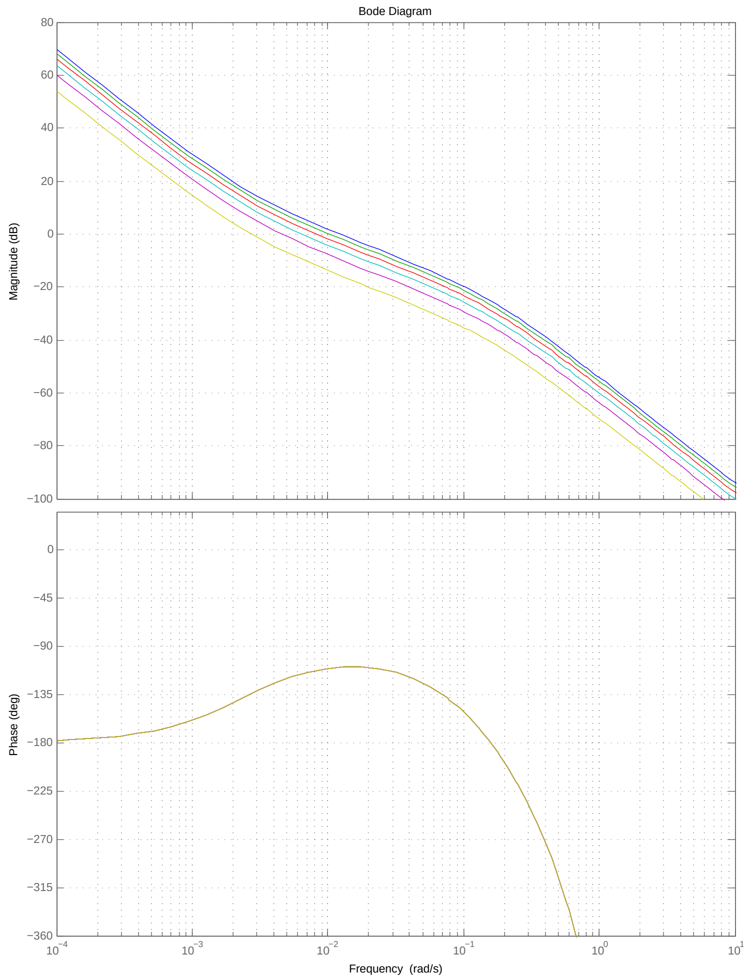

Figure B.5: Bode diagram of the open loop transfer function for oil level control. A linear PI controller, tuned with separator dynamics method, is implemented to the system. The plot shows the loop for different valve magnitude within the working area.

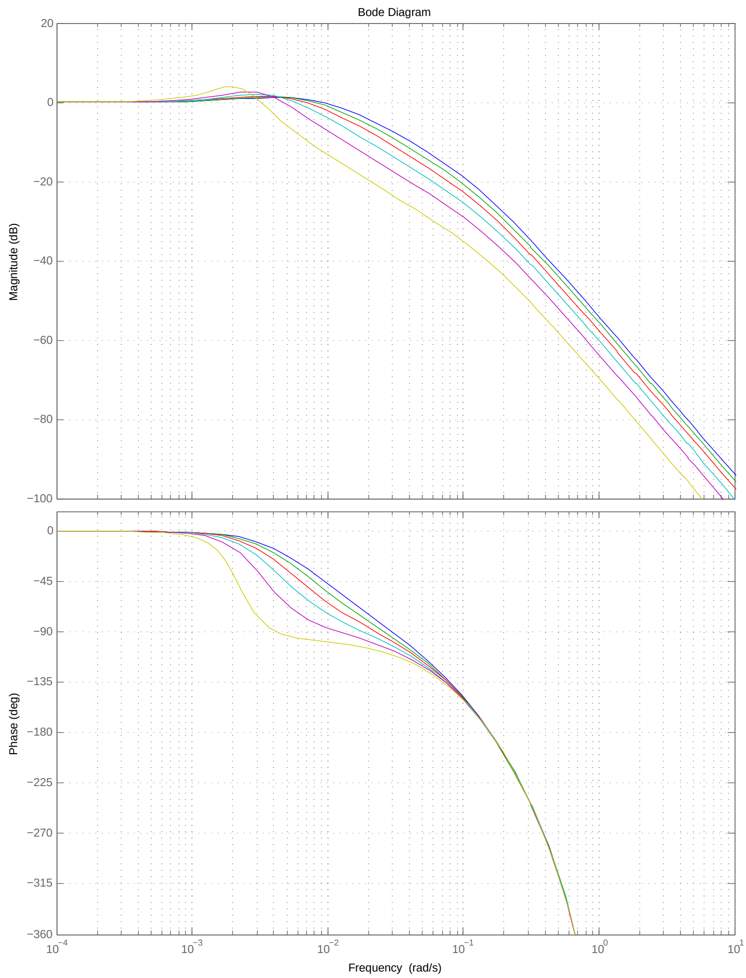

Figure B.6: Bode diagram of the closed loop transfer function for oil level control. The tuning is performed by the separator dynamics method, and the system is controlled by a linear PI controller. The plot shows the loop for different valve magnitude within the working area.

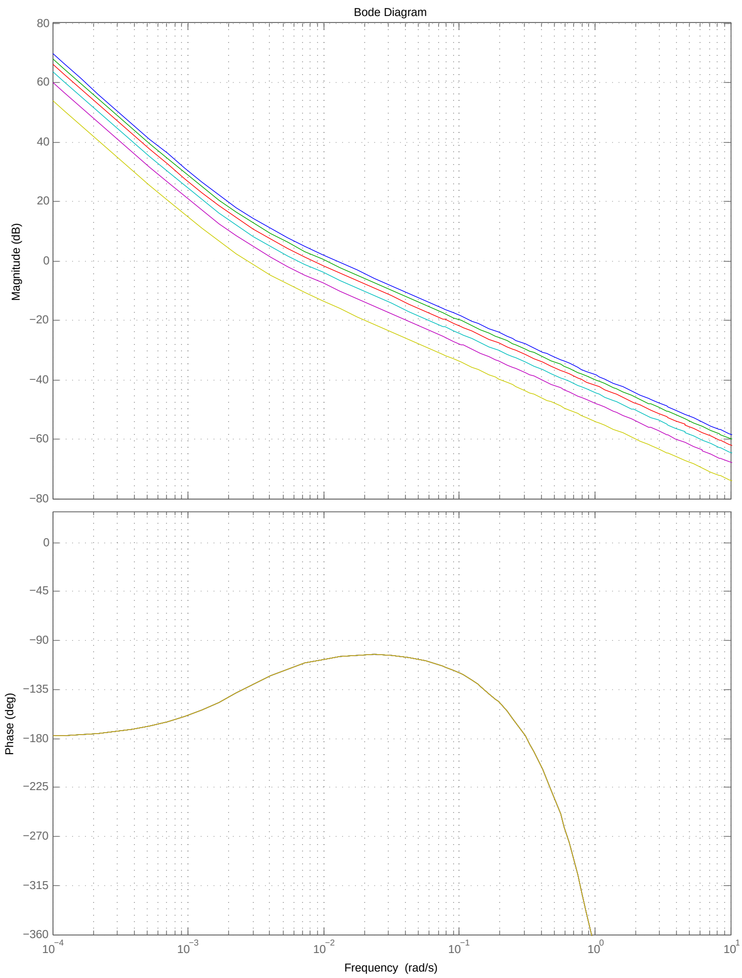

Figure B.7: Bode diagram of the open loop transfer function for oil level control. A linear PID controller, tuned with separator dynamics method, is implemented to the system. The plot shows the loop for different valve magnitude within the working area.

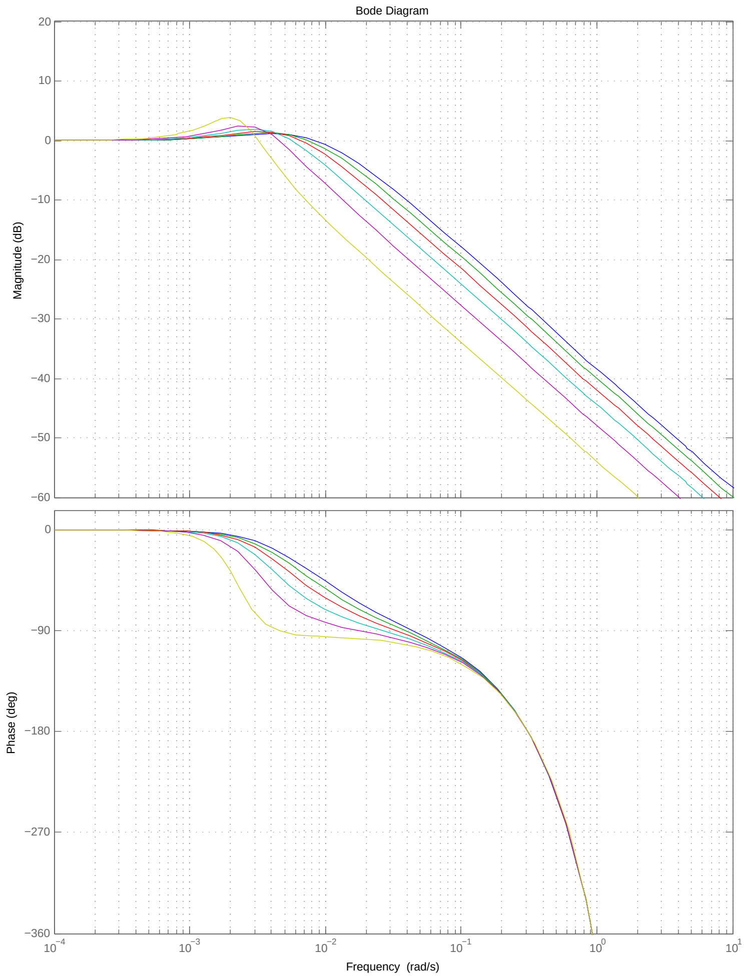

Figure B.8: Bode diagram of the closed loop transfer function for oil level control. The system is controlled by a linear PID controller tuned with separator dynamics method. The plot shows the loop for different valve magnitude within the working area.

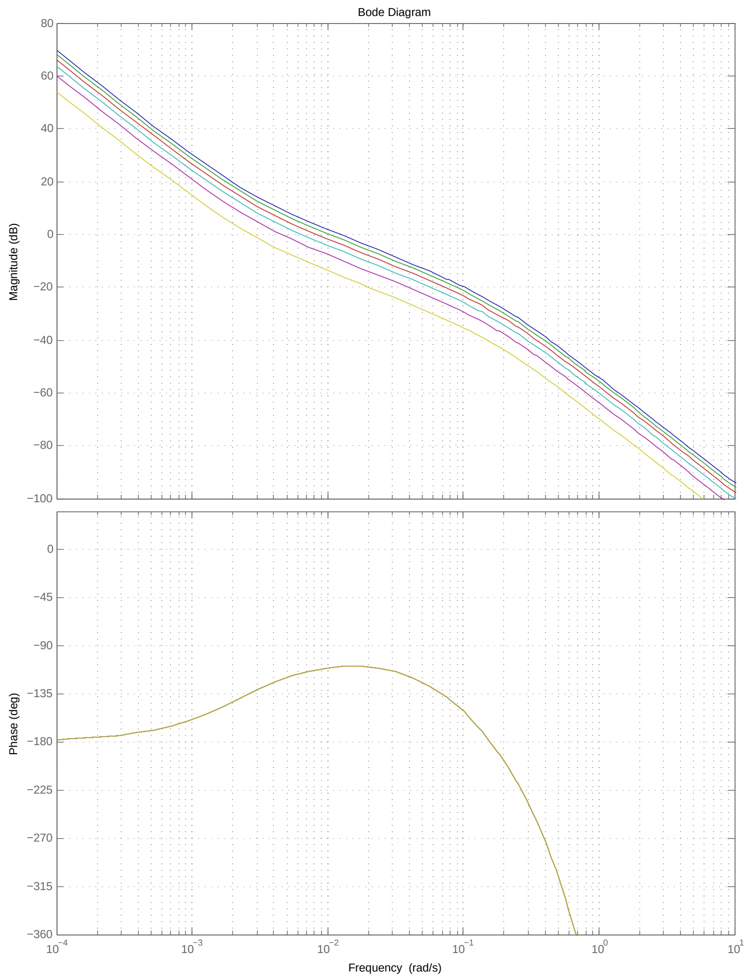

Figure B.9: Bode diagram of the open loop transfer function for water level control. A linear PI controller, which is tuned with separator dynamics method, is implemented to the system. The plot shows the loop for different valve magnitude within the working area.

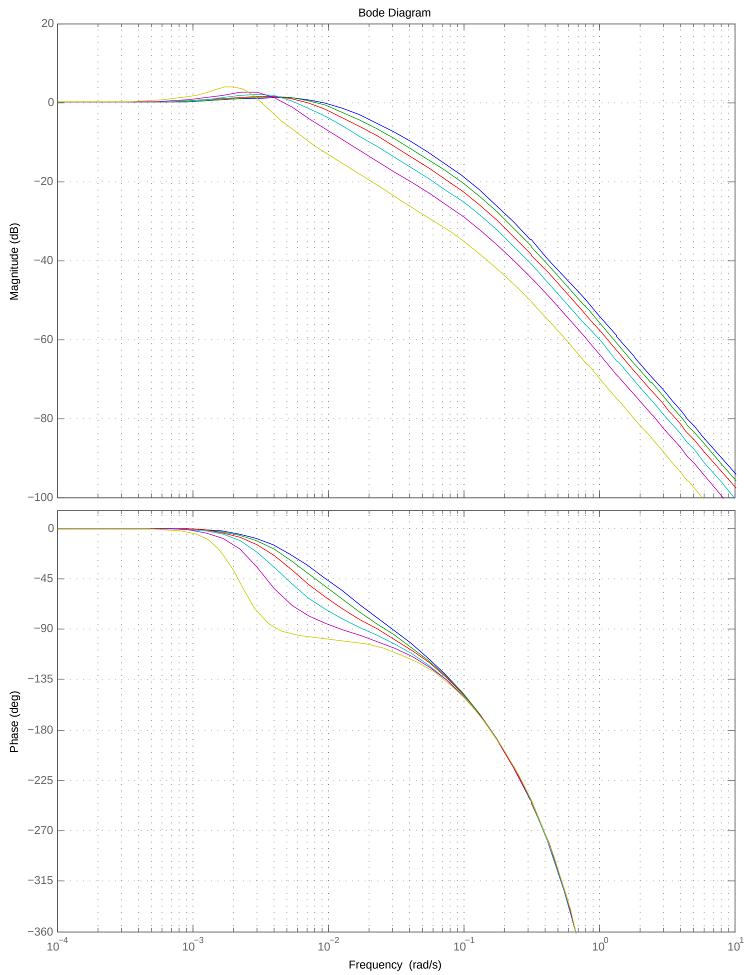

Figure B.10: Bode diagram of the closed loop water level transfer function. The system is controlled by a linear feedback PI controller which is tuned with separator dynamics method. The plot shows the loop for different valve magnitude within the working area.

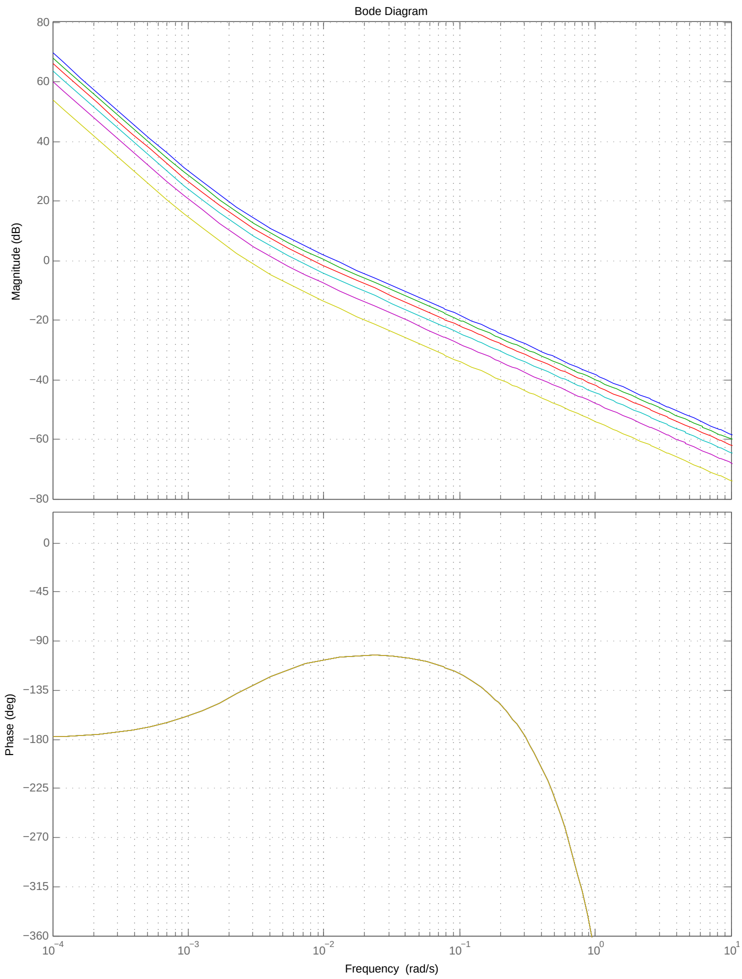

Figure B.11: Bode diagram of the open loop transfer function for the water level control. A linear PID controller, tuned with separator dynamics method, is implemented to the system. The plot shows the loop for different valve magnitude within the working area.

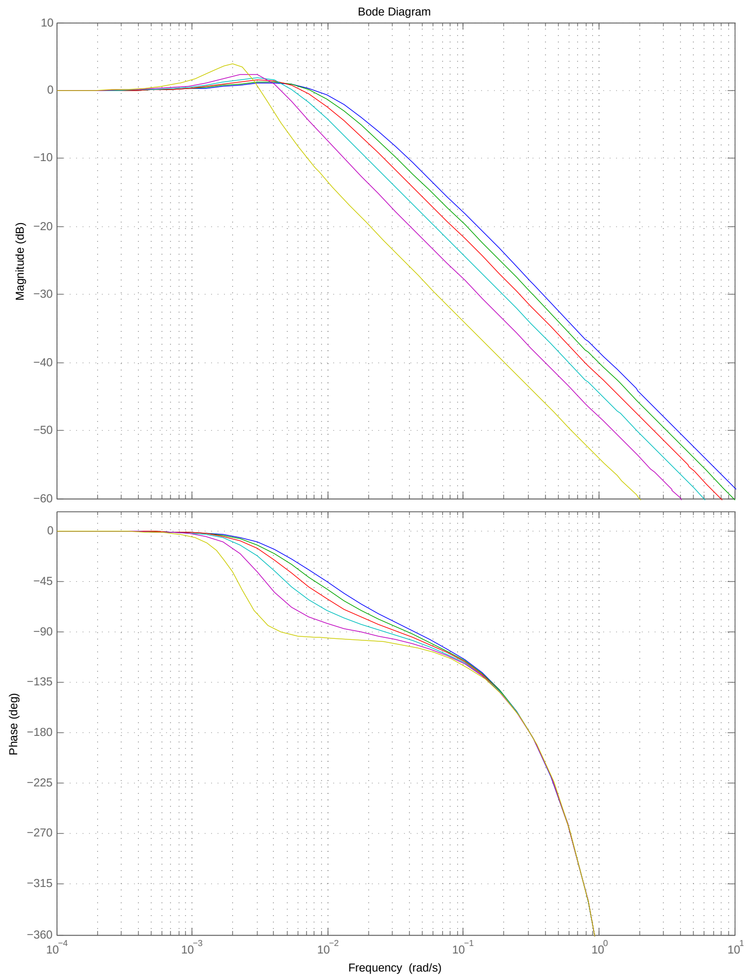

Figure B.12: Bode diagram of the closed loop water level transfer function. The system is controlled by a linear PID controller tuned with separator dynamics method. The plot shows the loop for different valve magnitude within the working area.

C MATLAB codes

Matlab script for initialising the Simulink model

```
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

% Configuration script for Separator model
% Master thesis 2013

%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%

clear all;
clc;

%*****************************************************************





%*****************************************************************
% System dimentions
%*****************************************************************

% Dimentions
L = N/A;                                % Separator length in mm
D = N/A;                                % Separator width in mm
Weir = N/A;                             % Wier height in mm
ScaleFactor = 1000;                     % m to mm
pi = 3.14159265;


% Scaled dimentions
l = L/ScaleFactor;
d = D/ScaleFactor;
wl = (3*l)/4;
weir = Weir/ScaleFactor;
innflow = N/A;                          % Maximum inflow (m^3/h)
Gasrho = N/A;                           % Gas density (kg/m^3)
Oilrho = N/A;                           % Oil density (kg/m^3)
Waterrho = N/A;                         % Water density (kg/m^3)
g = 9.81;                               % Gravitational force
v_z = 0.95;
Mg = 58.12 / 1000;                      % Molar mass in kg/mol
RT = (8.314472 * 293) / Mg;
TotVol = (d/2)^2*pi*l;
Po = N/A;                               % Pressure next level in Pascal


% Process constraints
MaxGas = N/A;                           % In MPa
MinGas = N/A;                           % In MPa
Gasset = N/A;                           % In Pascal
GasTravel = N/A;                        % In MPa
GasInit = N/A;                          % Initial gas pressure in Pa

MaxOil = N/A;                           % In mm
MinOil = N/A;                           % In mm
Oilset = N/A;                           % In mm
OilTravel = N/A;                        % In mm
OilInit = N/A;                          % Initial oil level in m

MaxWater = N/A;                         % In mm
MinWater = N/A;                         % In mm
Waterset = N/A;                         % In mm
WaterTravel = N/A;                      % In mm
```

```
WaterInit = N/A;                        % Initial water level in m

MaxGasflow = N/A;                       % Maximum gas inflow (m^3/h)
MaxOilFlow = N/A;                       % Maximum oil inflow (m^3/h)
MaxWaterFlow = N/A;                     % Maximum water inflow (m^3/h)


%Setpoints
GasSet = 0.50;                          % [0, 1]
OilSet = 0.50;                          % [0, 1]
WaterSet = 0.50;                        % [0, 1]


% Normatization values
a_g = (1/((MaxGas-MinGas)*(1e6)));      % Gas pressure constant
a_o = (1000/(MaxOil-MinOil));           % Oil level constant
a_w = (1000/(MaxWater-MinWater));       % Water level constant

b_g = -(MinGas/(MaxGas-MinGas));        % Gas pressure constant
b_o = -(MinOil/(MaxOil-MinOil));        % Oil level constant
b_w = -(MinWater/(MaxWater-MinWater));  % Water level constant

C_g = 1/(MaxGasFlow * 1.2);             % Gas flow constant
C_o = 1/(MaxOilFlow * 1.2);             % Oil flow constant
C_w = 1/(MaxWaterFlow * 1.2);           % Water flow constant




%*****************************************************************
% Actuators
%*****************************************************************

% Transferfunction
s=tf('s');

% Gas Actuator
GasNumval1 = [1];
GasDenval1 = [60, 1];
% GasDelay = 5;
% GasSys = tf(GasNumval1,GasDenval1,'InputDelay',GasDelay)
GasSys = tf(GasNumval1,GasDenval1)


% Oil Control valve
OilNumval1 = [1];
OilDenval1 = [6, 1];
OilDelay = 5;
OilSys = tf(OilNumval1,OilDenval1,'InputDelay',OilDelay)


% Water control valve
WaterNumval1=[1];
WaterDenval1=[6, 1];
WaterDelay=5;
WaterSys = tf(WaterNumval1,WaterDenval1,'InputDelay',WaterDelay)




%*****************************************************************
% Low pass filters
%*****************************************************************

% Filter for feed forward oil loop control
GasFilterNumval=[1];
GasFilterDenval=[0.5, 1];
GasFilterSys = tf(GasFilterNumval, GasFilterDenval)

% Filter for feed forward oil loop control
```

```
OilFilterNumval=[1];
OilFilterDenval=[1, 1];
OilFilterSys = tf(OilFilterNumval, OilFilterDenval)

% Filter for feed forward oil loop control
WaterFilterNumval=[1];
WaterFilterDenval=[0.1, 1];
WaterFilterSys = tf(WaterFilterNumval, WaterFilterDenval)




%*****************************************************************
% PID parameters
%*****************************************************************

% PID controller gas pressure loop
Kpg=-2.90;
Tig=240;
Tdg=60;

Kppg=Kpg*(1+(Tdg/Tig));
Tipg=Tig*(1+(Tdg/Tig));
Tdpg=Tdg*(1/(1+(Tdg/Tig)));
Npg=1;


% PID controller oil loop
Kpo=-1.68;
Tio=200;
Tdo=0;

Kppo=Kpo*(1+(Tdo/Tio));
Tipo=Tio*(1+(Tdo/Tio));
Tdpo=Tdo*(1/(1+(Tdo/Tio)));
Npo=1;


% PID controller water loop
Kpw=-7.36;
Tiw=200;
Tdw=0;

Kppw=Kpw*(1+(Tdw/Tiw));
Tipw=Tiw*(1+(Tdw/Tiw));
Tdpw=Tdw*(1/(1+(Tdw/Tiw)));
Npw=1;




%*****************************************************************
% Disturbance values
%*****************************************************************

% Gas disturbances
GasAmp=0.90;                                % GasAmp in [0, 1]
GasBias=0.10;                               % GasBias in [0, 1]
GasFreq=1/(60*60);                          % In Hz

% Oil disturbances
OilDist=0.50;                               % OilDist in [0, 1]
OilAmp=0.90;                                % OilAmp in [0, 1]
OilBias=0.10;                               % OilBias in [0, 1]
OilFreq=1/(60*60);                          % In Hz

% Water disturbances
WaterDist=0.50;                             % WaterDist in [0, 1]
WaterAmp=0.90;                              % WaterAmp in [0, 1]
WaterBias=0.10;                             % WaterBias in [0, 1]
```

```
WaterFreq=1/(60*60);                            % In Hz




%*****************************************************************
% Transducer models
%*****************************************************************

% Gas transducer
PTgas = 1;

% Oil transducer
LToil = 1;

% Water transducer
LTwater = 1;
```

D Simulink Diagrams

Simulink diagrams

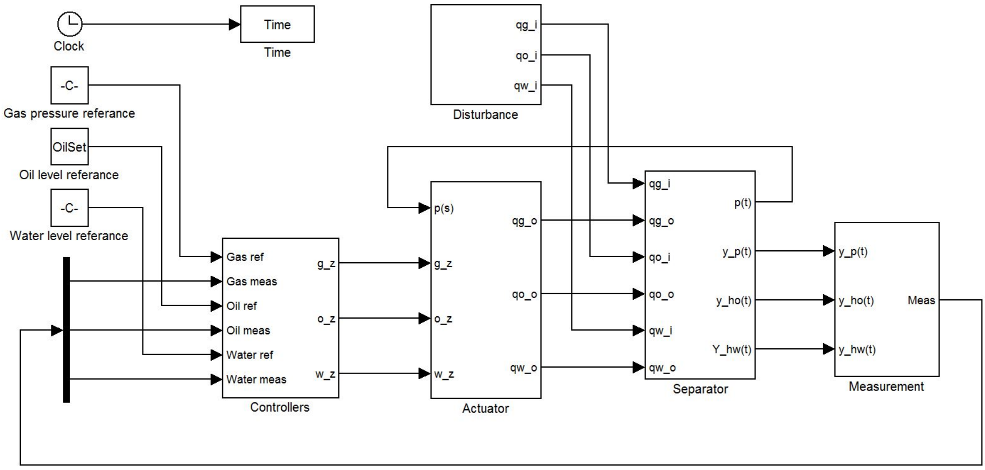

Figure D.1: Simulink diagram for the overall separator model. Each subsystem contains different parts of the overall system.

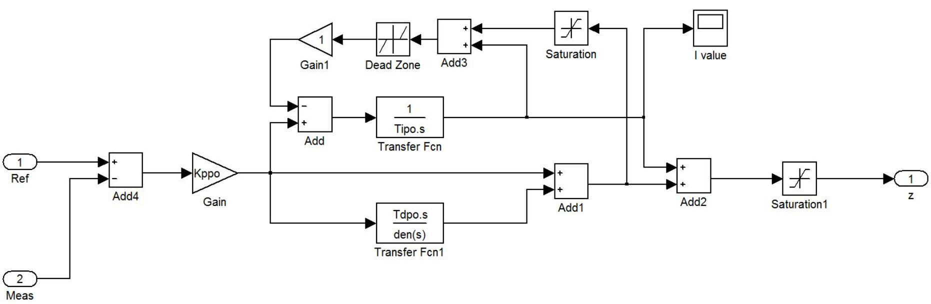

Figure D.2: Simulink diagram for the constructed PID controller where integrator anti-windup and derivative filter is implemented.

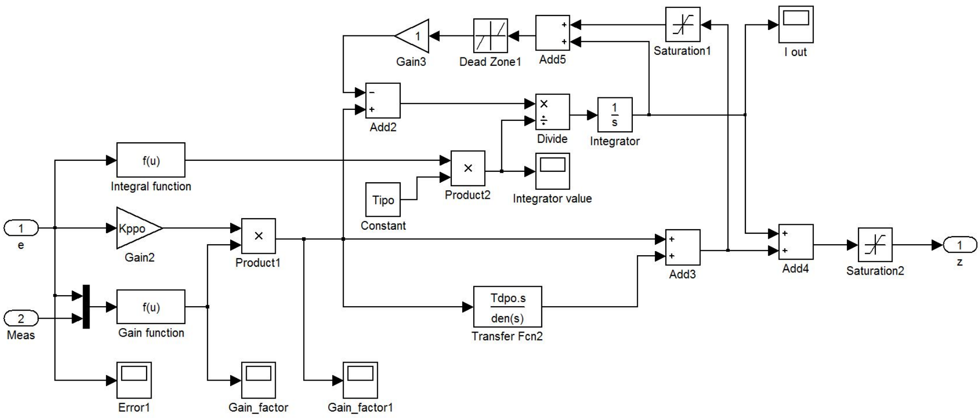

Figure D.3: Simulink diagram for the gain and integral scheduled PID controller. Integrator anti-windup and derivative filter is implemented in the controller.

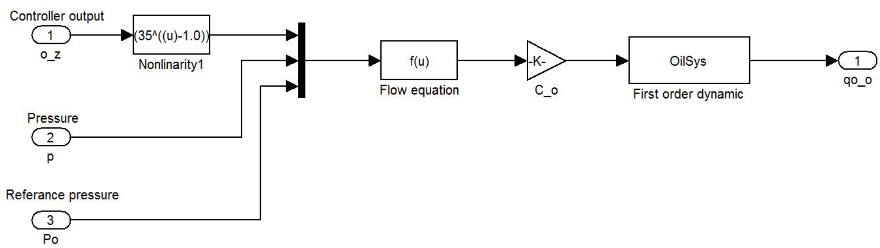

Figure D.4: Simulink diagram for the actuator structure in the separator system. The actuator function is dependent on three variables, signal from controller, separator pressure and reference pressure. The equal percentage nonlinearity is attached to the control signal as a mathematical function.

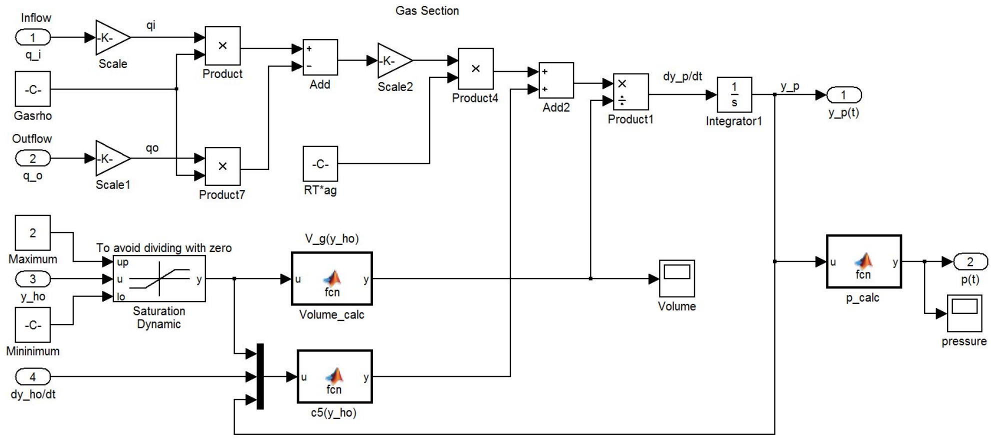

Figure D.5: Simulink diagram for the implemented gas pressure model. In the diagram, c_g is divided into two parts, volume calculation and mass balance. A function for calculating the real pressure is also added to the implementation. The model is based on ideal gas law and mass balance.

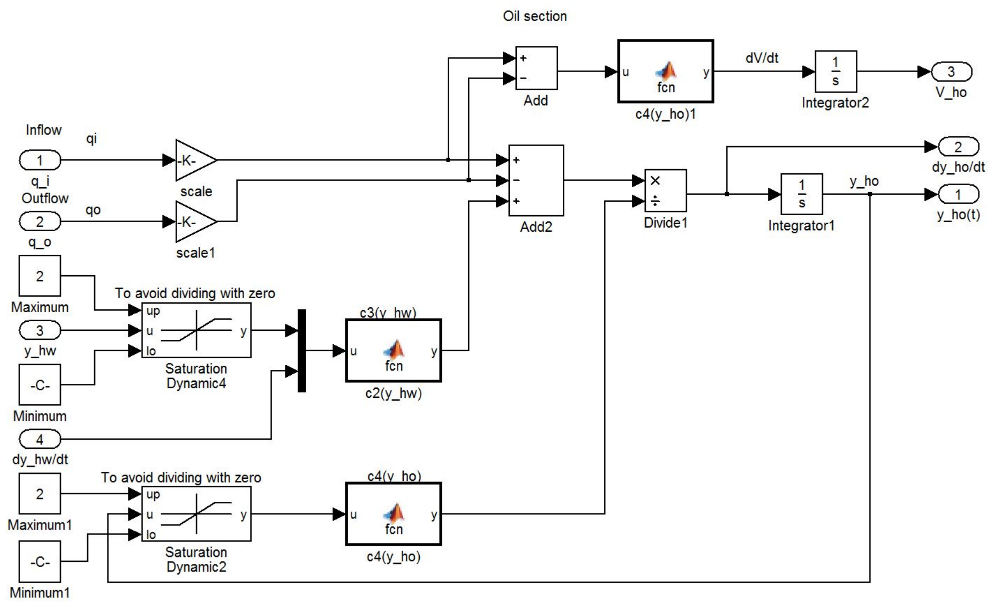

Figure D.6: Simulink diagram for the separator oil level model based on mass balance. The model is divided into two main parts, which take separator geometry into account. A function for oil volume calculation is added to the model.

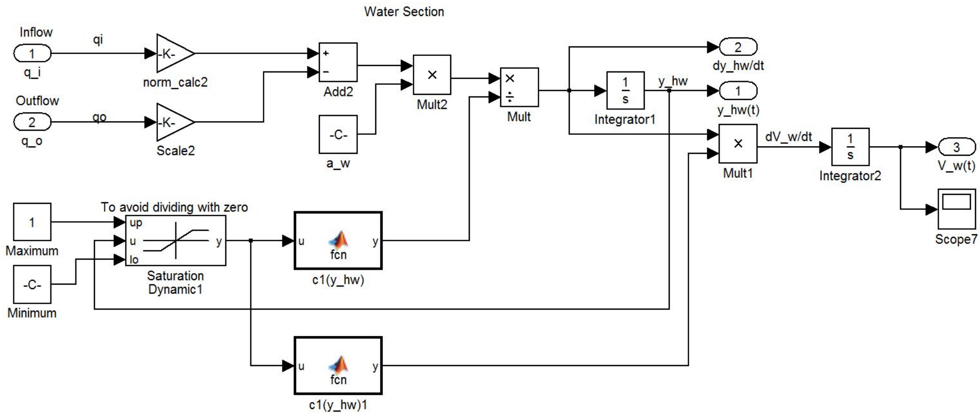

Figure D.7: Simulink diagram for the separator water level model based on mass balance. A function for volume calculations is added to the model. The model takes the separator geometry into account.

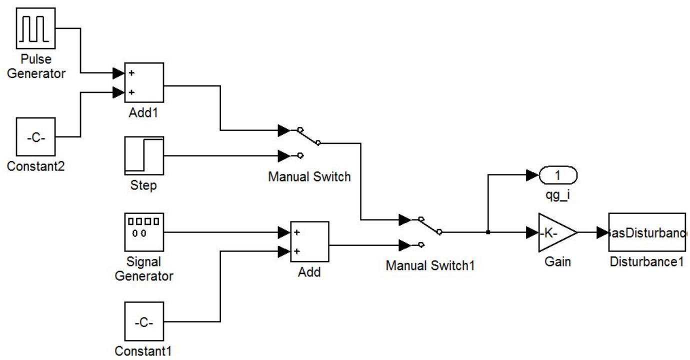

Figure D.8: Simulink diagram showing the possibility to chose different disturbances for the separator inflow. Signal generators are used to make different disturbance scenarios.

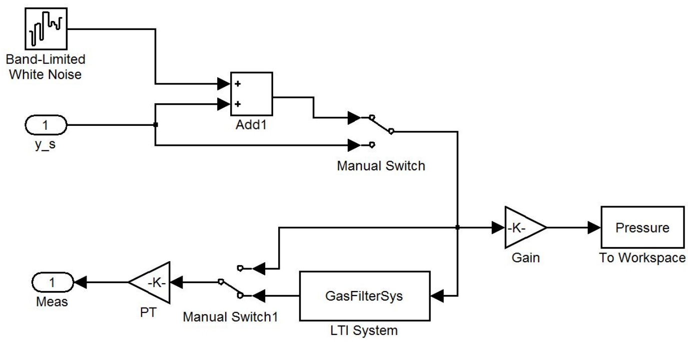

Figure D.9: Simulink diagram of the process variable measurement. The function contains a block for adding noise to the measurement signal and a filter to reduce the noise effect.