Tuning and Control Strategy for an Offshore Process Subject to Minimum Environmental Impact

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November 26th 2007

Abstract

The aim of this project is to present a tuning strategy for minimizing the environmental impact of an offshore process. A simulink-model of the platform Oseberg East is used as a basis, and the amount of natural gas flared is used as a measurement of the environmental impact. Skogestad's simple and smooth tuning strategies have been reviewed as well as the works of general gas- and oil- production. A plantwide analysis has been conducted on the system. Different tuning strategies have been tested on the model and a few cases given a closer look. As a result a set of considerations have been proposed which should be taken into account subject to minimizing flaring. Some aspects of the problem have been viewed, and a suggestion to further work presented.

KEYWORDS: TUNING, ENVIRONMENT, SIMC, PLANTWIDE, PETROLEUM

Contents

1. Introduction

Today petroleum still stands as the world's most important source of energy due to its high energy density, easy transportability and relative abundance. In the latter years however the world has been faced with questions related to the overall good of employing fossil fuels, taking millions of years in the making, in just a couple of centuries.

In the 4th assessment report recently released by the Intergovernmental Panel on Climate Change it's stated that human activity is very likely, that is over 90% certain, to have something to do with the increased amount of carbon dioxide stored in the atmosphere and the related climate changes the world have experienced in the later years. [1]

Renewable energy still has a long way to go in becoming an effective alternative as far as energy resources go. A lot of research and development remains and will be in need of energy to get carried trough. This implies that in a realistic perspective the oil will keep coming up the wells.

Only about 70% of hydrocarbons extracted from the ground reach the private or industrial consumer. The rest are losses, flaring of associated gas and energy consumption for production systems, transportation, refining and distribution of oil and gas. The less natural gas released into the air, the more is left to enhance the oil production. [2]

The optimum will be found as a trade-off. As far as flaring is concerned it's a necessary safety measure, but should be avoided as loss of natural gas to the air, is loss of gas available to enhance oil production. At the same time operating close to the flaring limit will speed up production.

The importance of utilizing what's left of fossil fuels in a best possible manner as well as doing so minimizing the impact on the environment is getting more and more attention. This is an effect of the environment becoming a more integrated part of the petrol-industry's economy with the introduction of taxes on greenhouse gases and extending legislative measures.

Most industrialized countries have ratified the Kyoto protocol which states that the industrialized countries will reduce their collective emissions of greenhouse gases by 5.2 % compared to the year 1990.

Tuning is a simple grip which doesn’t include costly additional equipment. Obtaining a tuning strategy which minimizes environmental impact, but at the same time maintains the production level would be a measure easily implemented in industrial practise.

2. General review of oil and gas production

2.1 The products

Oil and natural gas originate from organic material deposited in earlier geological periods, typically 100 to 200 million years ago, under, over or with sand or silt, it's transformed by high temperature and pressure into hydrocarbons. The petroleum collects in crests under non permeable rock with gas at the top, then oil and fossil water at the bottom. [2]

2.1.1 Crude oil

Crude oil is a complex mixture of hydrocarbons of various lengths, the approximate range being C_5H_{12} to $C_{18}H_{38}$. [3]The oil is often characterized by the API (American Petroleum Institute) gravity grade which is a measure of the crude's specific gravity, or density. The higher the API number the less dense (heavier) the crude. Crude oil from different fields and from different formations within a field can be similar in composition or be significantly different. Other important characteristics describing a crude oil besides the API grade and hydrocarbons is the unwanted elements present like sulfur which is regulated and needs to be removed. [2]

2.1.2 Natural gas

Natural gas is composed of shorter length hydrocarbons, that is components shorter than C_5H_{12}. The main component is methane, but commonly existing in a mixture with other hydrocarbons, principally ethane, propane, butane and pentane, and also additional components such as water vapor, hydrogen sulfide, carbon dioxide and so on. Natural gas being lighter than air will naturally rise to the surface of a well. [2]

2.2 The different stages of the production

2.2.1 The oil-train – *the separation processes*

Most often the well gives out a combination of gas, crude oil, water, condensates and various contaminants which must be separated and processed. The process executing this stage of the production is often known as the Gas Oil Separation Plant (GOSP), and has the purpose of processing the well flow into clean marketable products: oil, natural gas or condensates. [2]

2.2.2 The gas-train - *preparation for further transport of the natural gas*

Gas coming from separators on it's way to further preparation has generally lost so much pressure that that it must be recompressed to be transported to succeeding use in a gas lift or gas injection, or to an onshore facility. The pressure drop in the separator is necessary to achieve the wanted composition of products, that is a low enough vapor pressure on the oil, and a light enough gas The task of recompressing is executed by the compressor/gas train. In addition to the actual compressors a large section of associated equipment such as scrubbers and heat exchangers are needed.

The compressors have a limited capacity range, represented by maximum and minimum values for the flow, max differential pressure achievable and a lower limit of the pressure differential where the compressor goes into surge. Surge is a state in which you'll find a imbalance between head and flow for the compressor. A head too high compared to the flow will lead to a gas stream temporarily going backwards out of the compressor. This is a problem occurring when there isn't enough gas going through the compressor to operate it, something that can be handled by recirculation.

The heat exchangers in this part of the process are there to cool down the gas stream between each compressor. The lower the temperature is the less energy will be used to compress the gas and achieve the wanted final pressure and temperature.

The scrubbers also have an important function in the gas train where they're working as demisters. Liquid droplets can be found in the gas coming from the oil-train or as a result of the cooling done by the heat exchanger where water or liquid hydrocarbons can form. Either way it has to be removed before it reaches the compressor, because of the possible erosion damage it can do on the fast rotating blades. [2]

2.2.3 Gas- injection and lift

When a well is drilled the hydrostatic formation pressure drives the hydrocarbons out of the rock and up into the well. When the well flows, gas, oil and water are extracted, and at the same time the reservoir composition changes. The recovery of an oil reservoir is typically around 40%, but using certain measures one can take it up to about 70%.

Gas or water injection is often used with the purpose of maintaining overall and hydrostatic reservoir pressure and in this way force the oil toward the production wells. A free flowing oil well has enough downhole pressure to reach a suitable wellhead production pressure and maintain an acceptable well-flow. On the other hand when the formation pressure is too low, and water or gas injection cannot maintain pressure or is unsuitable, then an artificial lift of the well is used.

Gas lift injects gas into the well flow. The downhole reservoir pressure falls off to the wellhead due to the counter pressure from weight of the oil column in the tubing. By injecting gas into the oil, the specific gravity is lowered and the well will start to flow. Gas lift can be controlled for a single well to optimize production, and to reduce slugging, but can also be used over several wells to use available gas in the most efficient way. [2]

2.2.4 Process control

The process control system is large and integrated with the purpose of reading values from a large number of sensors, run programs to monitor the process and control valve switches to control the process. At the same time values, alarms, reports and other information are presented to the operator and command inputs accepted. The main function of the control system is still to make sure the production, processing and utility systems operate efficiently within design constraints and alarm limits. [2]

2.4 Flaring

Faring is defined as controlled combustion of gas justified as a cause of safety.
The flare subsystem includes flare, atmospheric ventilation and blow down. The purpose of the system is to provide safe discharge and disposal of gases and liquids.

Flaring is not a part of the normal operation of a plant, but may still occur as a result of:

- Spill-off flaring from the product stabilization system.
- Production testing.
- Relief of excess pressure caused by process upset conditions and thermal expansion.
- Depressurisation either in response to an emergency situation or as part of a normal procedure.
- Planned depressurisation of subsea production flowlines and export pipelines.
- Venting from equipment operating close to atmospheric pressure.

The emergency valves will separate the process components and blow-down valves that will allow excess hydrocarbons to be burned off in the flare. These valves are operated if critical operating conditions are detected or on manual command, by a dedicated Emergency Shutdown System. This might involve partial shutdown and shutdown sequences since the flare might not be able to handle a full blow-down of all process sections simultaneously.

In an oil and gas facility the primary response to an emergency event is to isolate and depressurize. The typical action would be to close the inlet and outlet sectioning valves and open the blowdown valve. This will isolate the malfunctioning unit and reduce pressure by flaring of the gas.[2]

3. Tuning

This chapter will cover tuning strategies for tight as well as smooth tuning focusing on the implementation on pressure and level controllers, that is handling of first order plus time-delay (FOPTD) as well as integrating responses.

3.1 SIMC- Skogestad's/Simple Internal Model Control

Skogestad [4] present a two step procedure in which is easy to use and remember, but also result in good closed loop behaviour.

Step 1: Obtaining a first- or second-order plus delay model on the form

$$g(s) = \frac{k}{(\tau_1 s + 1)} e^{-\theta s} \tag{3.1}$$

or

$$g(s) = \frac{k}{(\tau_1 s + 1)(\tau_2 s + 1)} e^{-\theta s} \tag{3.2}$$

where

- g(s) denotes the process transfer function
- k is the plant gain.
- τ_1 is the dominant lag time constant.
- θ represents the effective time delay.[1]
- τ_2 denoted the second' order lag time constant for processes in which ~ $\tau_2 > \theta$.

Or for an integrating process where $\tau_1 \rightarrow \infty$

$$\frac{\mathrm{k}}{\tau_1 s + 1} e^{-\theta s} \approx \frac{k}{\tau_1 s} e^{-\theta s} = \frac{k'}{s} e^{-\theta s} \tag{3.3}$$

Where k' = k/ τ_1 is the slope of the ramp-response. This is also a good approximation for lag-dominant processes.

[1] The effective delay is an efficient way of integrating different lags into the simple model by approximation.

Step 2: Deriving model-based controller settings. PI-settings result if we start from a first-order model given in equation (3.1), whereas PID-settings result from a second-order model given in the equation (3.2). The SIMC PID controller settings can be presented by equation (3.4), (3.5) and (3.6) given below

$$K_c = \frac{1}{k} \frac{\tau_1}{\tau_c + \theta} = \frac{1}{k'(\tau_c + \theta)} \tag{3.4}$$

$$\tau_I = \min(\tau_1, 4(\tau_c + \theta)) \tag{3.5}$$

$$\tau_D = \tau_2 \tag{3.6}$$

Skogestad goes on to propose a use of $\tau_c = \theta$. For the integrating process τ_I will as $\tau_1 \sim \infty$ always be given as $\tau_I = 4(\tau_c + \theta)$ as a result of equation (3.5).

3.2 Smooth tuning

Skogestad [5] provides a tuning strategy for obtaining robust control, subject to achieving acceptable performance in terms of disturbance rejection, something preferred in the industry. Whereas tight control is used subject to achieving acceptable robustness using as fast control as possible, smooth control consists of a as slow as possible control.

3.2.1 Introducing an higher limit for τ_c

SIMC presents a lower limit for $\tau_c \geq \theta$, and now a higher bound is given as well resulting in a preferred area for τ_c given by

$$\tau_{c,\min}(\text{"}tight\text{"}) \leq \tau_c \leq \tau_{c,\max}(\text{"}smooth\text{"}) \tag{3.7}$$

Where $\tau_{c,min}$ represents the limit for tight control and $\tau_{c,max}$ the limit for smooth control.

As presented earlier the limit $\tau_{c,min}$ depends mainly on the robustness requirements with respect to the delay θ. Skogestad now wants to include a larger value for τ_c in order to obtain smoother control with

(i) less input usage

(ii) less sensitivity to measurement noise

(iii) better robustness, and

(iv) less disturbing effect on the rest of the plant

A problem with using a too large τ_c is that the resulting disturbance response may be unacceptable, a higher bound, $\tau_{c,max}$, is therefore necessary. Finding this value is done by deriving a lower limit on the lower gain $K_{c,min}$, which is given by rearranging the SIMC' equation (3.4) and writing it for $\tau_{c,max}$ and $K_{c,min}$ in the way presented below

$$\tau_{c,\max} = \frac{\tau_1}{k} \cdot \frac{1}{K_{c,\min}} - \theta \qquad (3.8)$$

We find that the requirement for acceptable disturbance rejection is

$$K_c \geq K_{c,\min} = \frac{|u_0|}{|y_{\max}|} \qquad (3.9)$$

where $|u_0|$ is the required input magnitude for disturbance rejection and $|y_{max}|$ is the maximum allowed deviation in the output y.

Substituting $K_{c,min}$ into equation (3.8) one can obtain $\tau_{c,max}$, and end up with the wanted region for the desired closed-loop time constant given in (3.7)

4. Model development

The basis of this project is a model of the platform Oseberg East developed in Simulink in Matlab. The process design drawing is presented in figure 4.1 below, and the model attached on a separate disc to the original report.

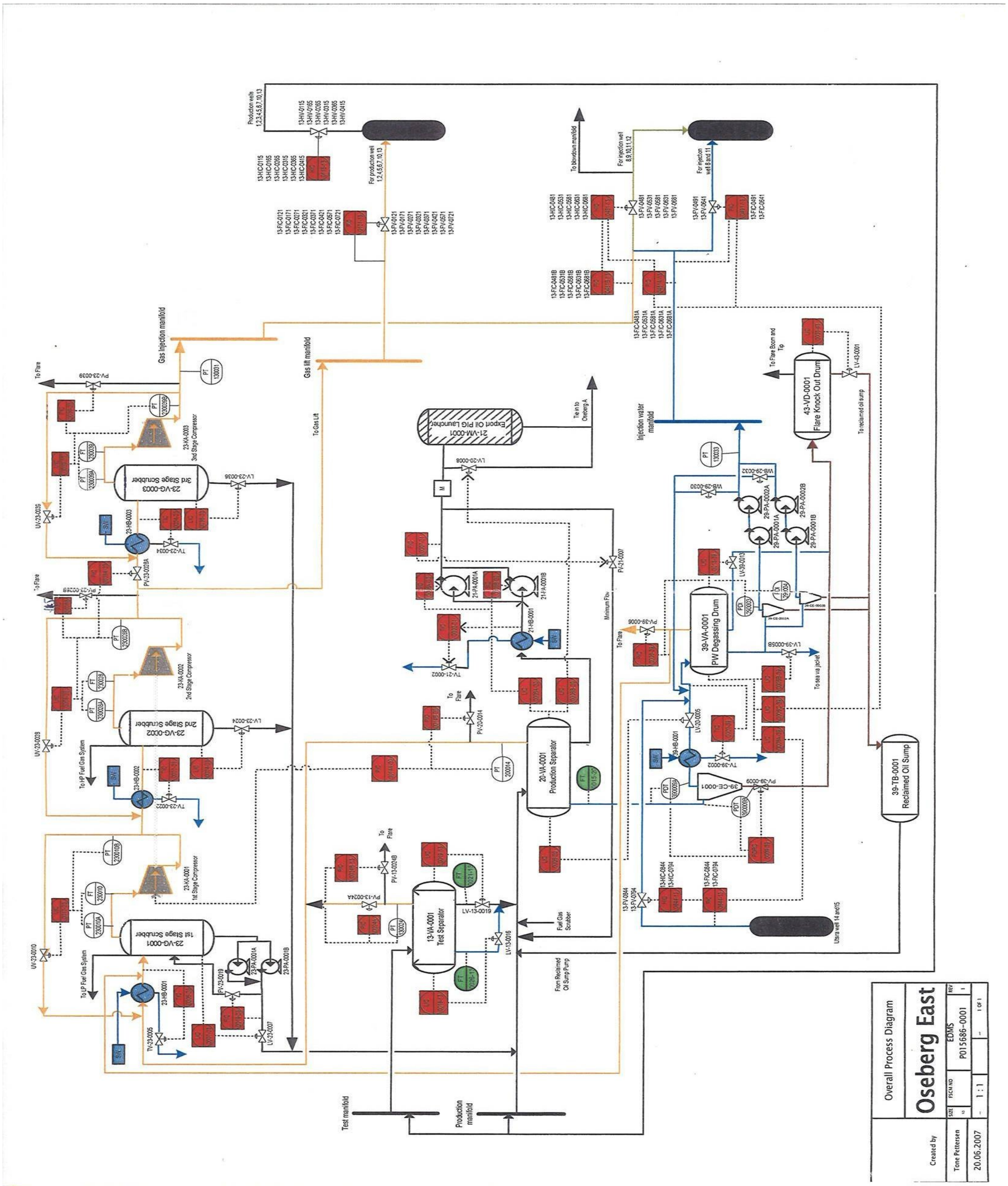

Figure 4.1 Design drawing of the process.

4.1 Stage 0 – The state of the model after ending the summer internship

A simulink-model of the platform Oseberg East was made as part of an internship with ABB's department Enhanced Operation and Production over the summer of 2007, and became the basis of this successive project.

The model's state prior to project start-up was one complete with a working gas- and oil-train. Some important characteristics and assumptions that were made in the making are listed below:

- Energy-balances are neglected.
- The waterside is left to be modelled at a later stage, maybe as part of a succeeding diploma-thesis discussing water quality.
- Flow, level and pressure controllers implemented, but consistent tuning is left to be worked out.
- Spin control on the oil-export pumps neglected.
- The interactions between the gas returning to the oil wells to optimize production and the inflows to the system are disregarded.

4.2 Stage 1 – A stable starting point

The model was first tuned to represent the process data and P&IDs for the real platform as well as possible still being able to counteract disturbances introduces to the system. A practical but unsystematic tuning was implemented.

First a look at the effect of step-changes on the closed-loop system was conducted as a way of gaining a better understanding of the process control of the process. A succeeding look at step-changes on the manipulated variables of the open-loop system was necessary in finding the parameters representing process characteristics presented in section 3.1.

A consistent choice of the desired closed loop time constant τ_c as the only available tuning parameter as presented by SIMC has been performed throughout the scope of this project. A choice made out of practical reasons, subject to tuning the whole process in a consistent manner, but also wanting to continue Skogestad's ideas of keeping the tuning-rules simple and in this way possibly easing introduction to the industry.

A base-tuning originating from the SIMC rules with $\tau_c > \theta$ was chosen after first testing the system for $\tau_c = \theta$, resulting in a too aggressive control base. The choice of τ_c was done for each of the controllers separately on a base of practical values which had already presented a flexible and robust system when running the simulation. A robust base tuning was wanted.

The tuning was performed loop by loop for the different controllers with an assumption of low interactions in the system. This assumption was later tested and shown to be valid in the case of two of the most important pressure controllers in the system, something presented in section II.1 of Appendix II.

4.3 Stage 2 – Upgrade and simplification

The biggest difference between the model presented and the original replication of the design drawing presented in figure 4.1 is that further simplifications have been done. The reason for this was that the simulations were running too slowly, which is unacceptable in such a time limited project. To have the time to simulate and observe different scenarios a decrease in the simulation time was a necessary measure, and something done most efficiently by further simplifying the model. If this project had been part of a larger study the simplification should maybe have been avoided to keep the model as close to the real process as possible. But as energy-balances have already been neglected the effect of additional changes, like removing the coolers, are minor. The test-separator was also removed, subject to it only being put to use every third month and thereby not being part of a day to day operation of the plant.

4.4 Stage 3 – Introducing disturbances, scaling and finding the base case

Disturbance series for the gas-, oil- and water inlet were collected from platform data and imported into the model along with an amplification factor, AF. The amplification factor provides a measure of enhancing or decreasing the amplitude of the disturbances put on the system. In the model it works as a proportional factor added to the bias of the given disturbance sequence from the mean inflow of the system. The enhanced/reduced disturbances are subsequently added to the mean inflow again which now provides the inlet flow of the separator.

The disturbance series for the three inlets are presented in the figures 4.2, 4.3 and 4.4 below.

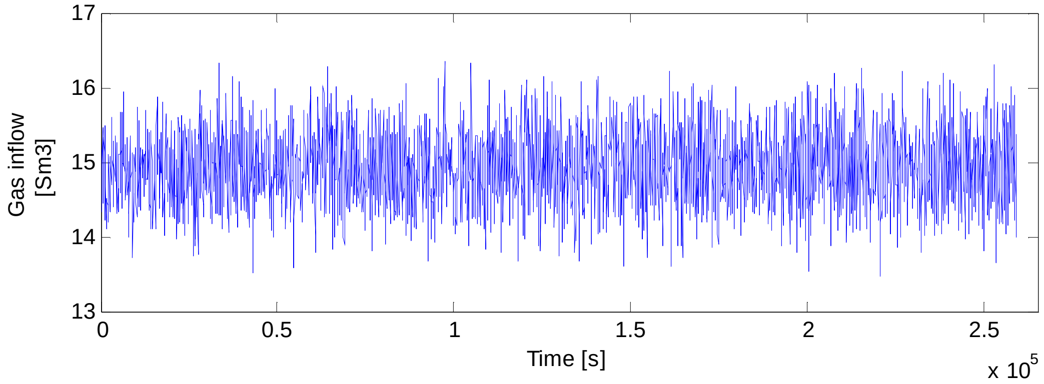

Figure 4.2 Gas inlet disturbances

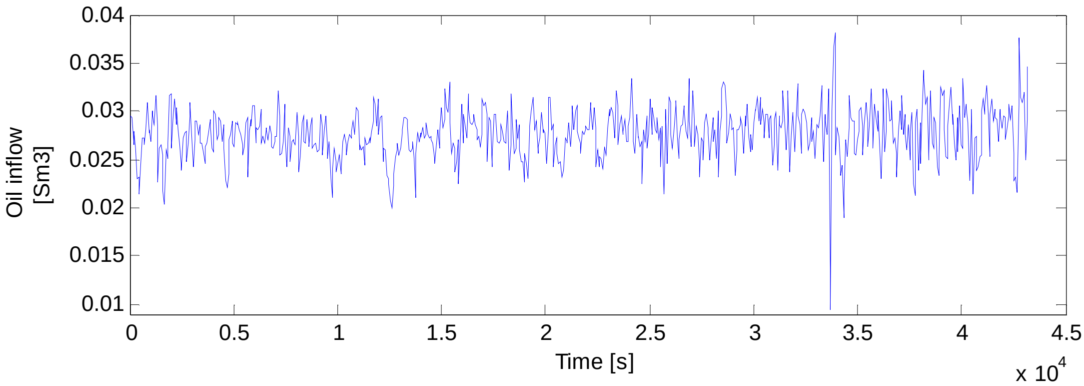

Figure 4.3 Oil inlet disturbances

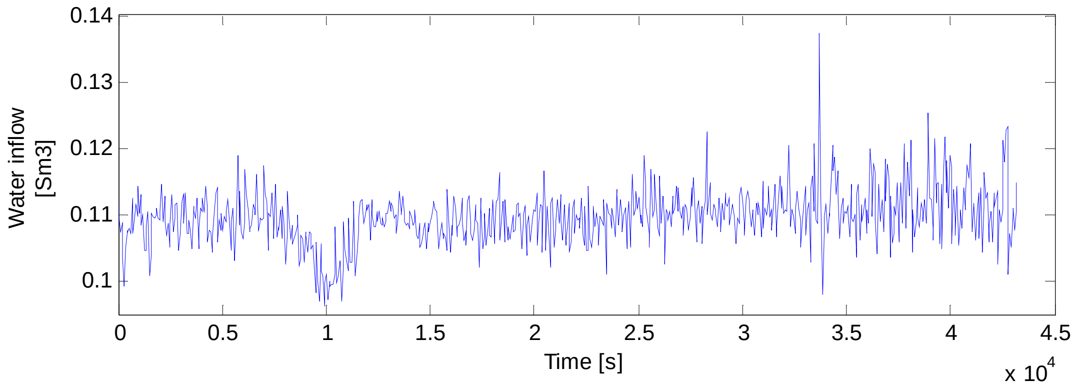

Figure 4.4 Water inlet disturbances

As an effect of introducing the disturbances a scaling of the model valves was necessary as the flows for the real case were a bit smaller than for the approximated one. A retuning of the base system succeeded.

The base case for running the simulations was first chosen to be a system borderline to flaring which was found by trial and error introducing different AFs to the inlets. This was disregarded eventually and replaced by an AF providing flaring to a larger extent. The gas inlet disturbances were amplified by a factor AF of 7, whereas an AF of 3.5 was chosen on the liquid inlets. Amplifications of this size provide a clear look at the effect of the disturbances for most cases, and also ensure that the minimum found is an evident minimum for a normal scenario.

5. Plantwide control

In this section control structure of the complete chemical plant as well as of the main components are discussed. Skogestad [6] suggest several ways of finding the degrees of freedom of a process, which in this section will be used as a measure of obtaining a larger understanding of the manipulated and controlled variables of the process. As this is a working plant the implementation of plantwide control is of course already done, but knowing the economical degrees of freedom, DOFs, of the process is still an important factor subject to finding the best control strategy for the process. Which variables are accessible in obtaining the cost function, that is which variables are left after stabilizing the process?

5.1 Cost function

How to minimize impact on the environment, number of flarings, subject to satisfying a certain level of production and also complying with safety regulations?

Constraints:

$$\mathrm{P}_{\min} \leq P \leq P_{\max} \quad \text{for the flares as well as for the gas lift/injection pressures.} \tag{5.1}$$

$$L_{\min} \leq L \leq L_{\max} \quad \text{for the liquids in the tanks of the system.} \tag{5.2}$$

$$\text{Feed flow}_{\min} \leq \text{Feed flow} \leq \text{Feed flow}_{\max} \tag{5.3}$$

5.2 The valve counting method

The valve counting method presented by Skogestad [6] consist of a two step procedure in finding the steady-state degrees of freedom N_{ss}.

1. Counting the number of dynamic degrees of freedom, N_{valves}, that is anything we can manipulate.
2. Finding all the manipulated variables which don't represent steady-state degrees of freedom, and subtracting them from the number found in 1.

The variables in 2 consist of valves that have no steady state effect, N_{0ss}, and also the number of equality specifications if specified, N_{specs}, for example given pressure.

N_{ss} can be written as:

$$N_{ss} = N_{valves} - N_{0ss} - N_{specs} \tag{5.4}$$

Where N_{0ss} consist of either controlled variables, N_{0y} or purely dynamic ones, $N_{0,valves}$.

$$N_{ss} = N_{valves} - (N_{0y} + N_{0,valves}) - N_{ecs} \tag{5.5}$$

5.3 Steady-state degrees of freedom of the system

In bringing about the valve counting method on the system, the main focus will be on the existing controllers in the simplified model used in the simulations, presented in figure 5.1. The method is employed not to implement controllers on the system as this is already done, but with the purpose of finding out which manipulated variables can be used as an advantage in minimizing the number of flarings.

Using the valve-counting method we find that the system has 15 valves and one adjustable compressor speed, that is we have N_{valves} = 16 we can manipulate. These are shown in the figure 5.1.

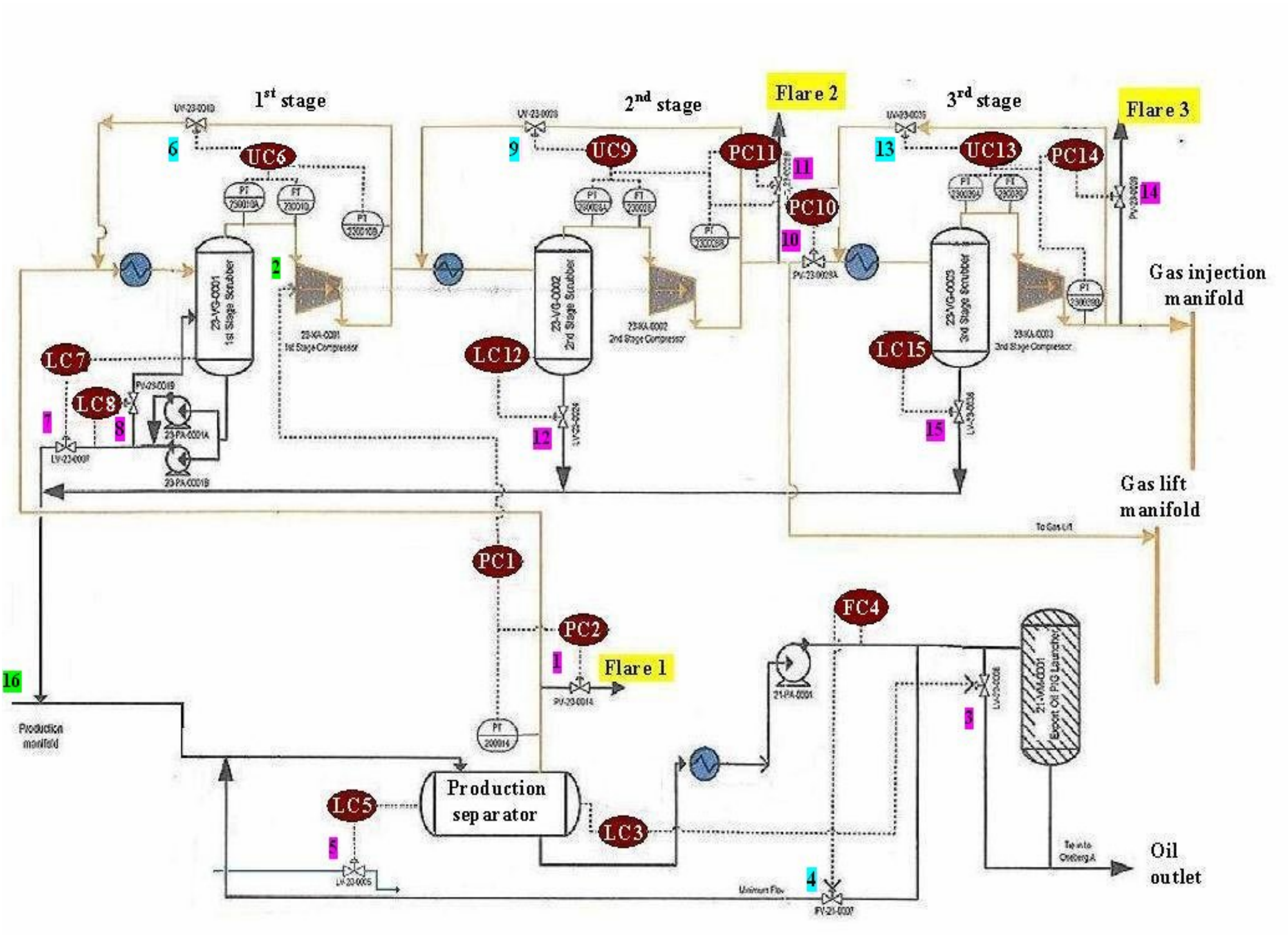

Figure 5.1 Process flow sheet for the simplified system used for running the cases.

5.3.1 Degrees of freedom analysis on the system locally

To get a better look at the variables available in attaining the cost function, the main components of the system were reviewed from a DOF point of view.

The Production Separator: Typically an adiabatic flash tank has zero steady-state degrees of freedom. In this separation-system we have three possible variations that can influence the control system, two liquid levels and one pressure. The pressure in the tank could be thought of as a steady-state degree of freedom, but it would require a local pressure valve as we're looking at the separator as a unit. There is actually a manipulated variable connected to the pressure variation, but this lies in the speed setting of compressors 1 and 2, that is, the degree of freedom transfers to this unit. Liquid levels are normally considered as controlled variables without any steady-state effect, but in this case we also have a liquid/gas split which has to be taken into consideration. The levels may have an impact on the pressure in the separator determining the split, previously presented as a possible s.s. degree of freedom, by causing the gas to compress/decompress as the levels rice or go down.

Scrubber 1, 2 and 3: Have no steady-state degrees of freedom.

Compressor 1 and 2: Have a joint degree of freedom in the speed-input coming from controlling the pressure out of the production separator. Handling of the compressors will be limited. If the speed is at its maximum we have an active constraint, but if not there will be an optimal speed limited by the compressor curves. We can compare a compressor with a car in that it has a curtain function in a specific area. You're only supposed to drive a car on one side of the road within the given speed limit. As for the compressor you want it to give you a specific pressure, in this case one high enough to either carry the gas on to gas lift or for the last compressor through to gas injection. A too high pressure out of one of the compressors can cause strain on the succeeding equipment, whereas a too low pressure may result in never reaching the pressure required at the outlets of the gas-train.

Compressor 3: Is run at a fixed speed, and does not have any steady-state DOFs.

The split between the 2nd and 3rd stage of the compressor-train. That is the split of the gas going to the lift-manifold from what's going on to the 3rd stage of the gas-train and continuing on to the injection-manifold. The split can be viewed as a degree of freedom to some extent when you look at it in connection with the valves placed on the many pipe-lines to the wells from the two manifolds. There is an optimal speed to gas lift, and the higher the gas lift the larger the production, but this degree goes beyond the limits of this simplified model.

The feed: Here we find a degree of freedom which is typically used in maximizing production, and which therefore will be the most dangerous with regards to the object of obtaining minimum impact on the environment. The desire to have this degree of freedom as an active constraint has to be weighted with the loss of robustness, stabilization and in the end the number of flarings it will be responsible for. In reality there are many wells and optimization with regards to production will often be dependant on which wells we choose to produce from and how much we choose to obtain from each. For example if the production-stream is rich on gas it would be preferable to include a stream from a well that has a larger oil-constituent. This aspect is beyond the capacities of the model.

5.3.2 Steady-state degrees of freedom of the system as a whole

In the model we have 16 manipulated variables. These are presented in figure 5.1, along with the controllers of the simplified system.

As stated earlier there are:

N_{valves} = 16 including everything that can be manipulated.

Out of these,

N_{0ss} =14are not steady-state DOF's.

- N_{0y} = 6 levels + 4 pressures keeping the process stable, marked with a pink color in figure 5.1.
- $N_{0,valves}$ ~ 4 . The anti-surges (UC) [2] and the minimum recycle flow are strongly integrated in the system but aren't necessary in a day to day operation, but present as important safety precautions. As they're already available in the system we can make use of them in improving the dynamic response. These are marked with a turquoise color in figure 5.1.

As a result of the plantwide analysis it was discovered that the production separator has one steady state degree of freedom located in the speed setting of the 1st and 2nd stage compressors, which is given by the oil-gas split provided by the pressure. There is also an economical degree of freedom in the feed. The suggested steady state DOFs are marked with a green color in the figure 5.1. The rest of the manipulated variables are used for stability purposes. The economical degrees of freedom are still very limited in obtaining the cost function stated in section 5.1. The best way of controlling the feed for flaring purposes would be to lower it, which would soon mean meeting the production lower limit given by equation 5.3. The degree available for controlling the speed of the compressors is also limited as stated in section 5.3.1.

The limitations in the number of economical freedoms in the system indicate that a good tuning strategy will be determining in obtaining the cost function.

[2] The anti-surge controllers use an input of combined pressure and flow measurements.

6. Model operation

6.1 Assumptions and explanations

The gain K and the process time-constant τ_1 describing the process transfer function were found on a basis of the response of the main controllers to set-point changes in the manipulated variable u. These along with an approximated value for the time delay and a chosen value for the desired closed loop time constant, τ_c, were gathered in a separate script. A dependence of the parameters in the script was introduced for the values of K_c and τ_I in the model-controllers. The values for the base parameters are presented in table 6.1.

The limitations presented in 5.2 and 5.3 were also included in the script for the tanks of the system, stopping the simulation at a given alpha if violated.

6.1.1 Active or constant tuning parameters

The controllers were chosen as either active (A), or represented by constant tuning parameters (C). Active control is represented by a dependence of τ_c on a proportional amplification factor alpha, α, as stated below,

$$\tau_c = \alpha\tau_{c0} \tag{6.1}$$

τ_{c0} was as τ_c in stage 1 chosen as a practical value $\tau_{c0} > \theta$.

Most of the controllers were chosen to be active, with some exceptions including controllers that aren't put to use in a day to day operation, and/or are unavailable for tuning. Hence for the system presented the anti-surge and minimum recycle controllers were chosen to be passive. The flow control on the water return to the first stage scrubber was also given constant control parameters. There is generally a very small liquid contribution from the demisters, and so the effect of controlling this recycle is negligible. The water recycle system in the other stages was given a fairly smooth base control as it doesn't have a very large impact on the simplified system presented excluding the waterside.

6.1.2 Introducing limitations

Time delays didn't exist in the model at this stage, but wanting to keep the system as realistic as possible, an implementation was provided. Limitations were introduced as part of the controller base parameters as well as in the most important valves of the system, making the case more realistic and the results a bit more conservative. Ranging the dead-time up provides more uncertainty to the system, and thereby safer controllers.

The time-delay was chosen to be either fast or slow for each of the controllers and valves. A choice made depending on the positioning in the process for the valves, but for the controllers mainly on what was the controlled object. Levels react more slowly to change than the pressures given the smaller theta (thetaFast) and were therefore described by the larger theta (thetaSlow). All the flow and anti-surge controllers had at this stage already been provided by constant tuning parameters.

6.1.3 The controllers of the system

The different controllers and their base parameters are presented below:

Table 6.1. Controller information

Controller	Full name[1]	k/k'[2]	T_1[3]	θ[4]	τ_{c0}	Status[5]	K_c	τ_I
Oil train								
PC 1	20-PIC-0014a	-0.15	30.044	Fast	15	A		
PC 2	20-PIC-0014b	-0.1	60	Fast	30	A/C[6]	-20	15
LC 3	20-LIC-0008b	-0.0031	∞	Slow	150	A		
FC 4	21-FIC-0007					C	-0.4	30
LC 5	20-LIC-0005	-0.1165	∞	Slow	100	A		
Gas train *1st stage*								
UC 6	20-UIC-0010					C	-3	10
LC 7	23-LIC-0007	-0.1165	∞	Slow	20	A		
FC 8	23-FIC-0019					C	-1	60
2nd stage								
UC 9	23-UIC-0028					C	-3	15
PC 10	23-PIC-0028a	-0.27	16	Slow	50	A		
PC 11	23-PIC-0028b	-0.1	30	Fast	10	A/C[6]	-20	15
LC 12	23-LIC-0024	-0.1165	∞	Slow	20	A		
3rd stage								
UC 13	23-UIC-0039					C	-5	15
PC 14	23-PIC-0039	-0.0044	3.75	Fast	10	A/C[6]	-6	15
LC 15	23-LIC-0036	-0.1165	∞	Slow	20	A		

1 This is the name presented in the design drawing in figure 4.1 as well as in the model

[2] Use of k vs. k' for FOPTD vs. integral responses according to the SIMC rules, see ….

[3] Level controllers show an integral response which implies a integral time constant $\tau_c = \infty$.

[4] Fast vs. Slow referring to a small vs. a larger time delay

[5] A vs. C referring to active controllers vs. inactive controllers given constant values for K_c and τ_I.

[6] The flare controllers were active for most of the cases presented in the succeeding section, but kept constant for some.

6.2 Effect of disturbances on the system

6.2.1 Focusing on the disturbances on the measurement

The disturbances introduced on the inlets to the oil-train will affect the whole system, but in the cases presented in the next section a focus will be on the main effect it has on the controller inputs. The effect of disturbances on the controller output is still important in providing the best tuning strategy for the system, and a review of the effects expected for this case is therefore presented in Appendix III. Running cases providing a measure of the effects these have on the system would be interesting if work on this project is to be resumed.

6.2.2 The chosen disturbance range

The disturbance range I've chosen to focus on is one starting 15000 seconds into the sequences and going on for three hours. The choice of time interval was done in connection with the choice of the AF described in section 4.4. The disturbances in the time period chosen, though fluctuating, doesn't include any extreme spikes or tendencies of the whole inflows going in one direction or the other. Looking back at the figures 4.2, 4.3 and 4.4 we see that such cases can be found for the water input around 10000 seconds and for the oil-inlet at about 33000 seconds. As the amplification factor was chosen quite large an additional extreme to the disturbances induced is unnecessary.

The disturbance inlets for the range chosen are presented below in figures 6.1, 6.2 and 6.3.

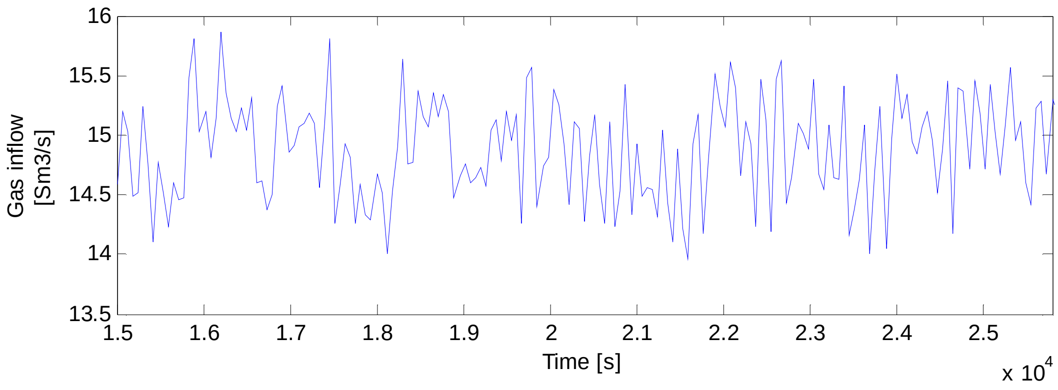

Figure 6.1 Gas disturbance inlet

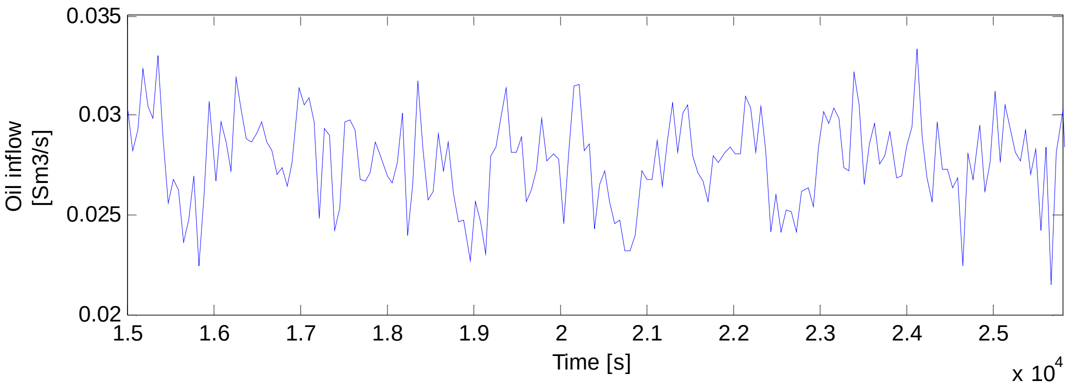

Figure 6.2 Oil disturbance inlet

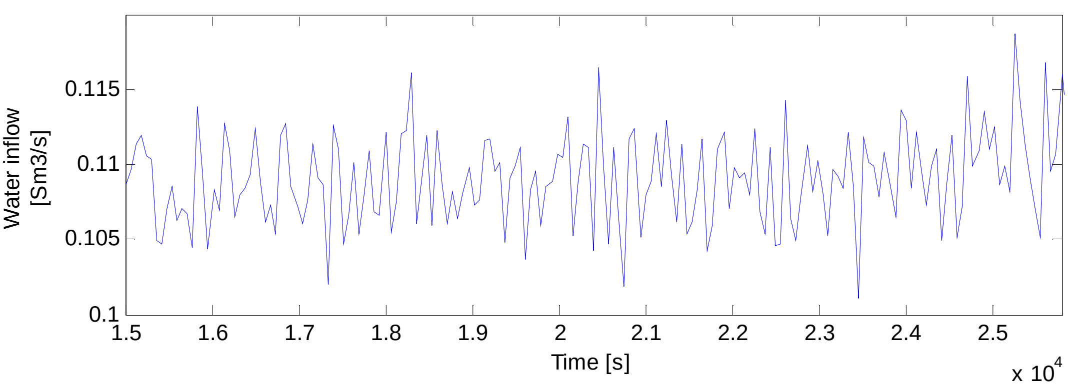

Figure 6.3 Water disturbance inlet

7. The cases

There are several aspects to this problem, but as this is a time limited project only the effect of a few have been given a closer look in this report. Several cases were tested though presenting the effect of introducing different disturbance intervals, amplification factors and so on. A summary of these are presented in appendix IV, and the results handed in on a separate disc.

7.1 The effect of limitations

As presented by the SIMC tuning rules an upper bound due to effective time delay is expected.

7.1.1 Case 1- Small limitations

Only a small time delay contribution, represented by a time delay of 1 second (thetaFast) for the pressure controllers and fast working valves, and one of 2 seconds (thetaSlow) for the slower level controllers and valves.

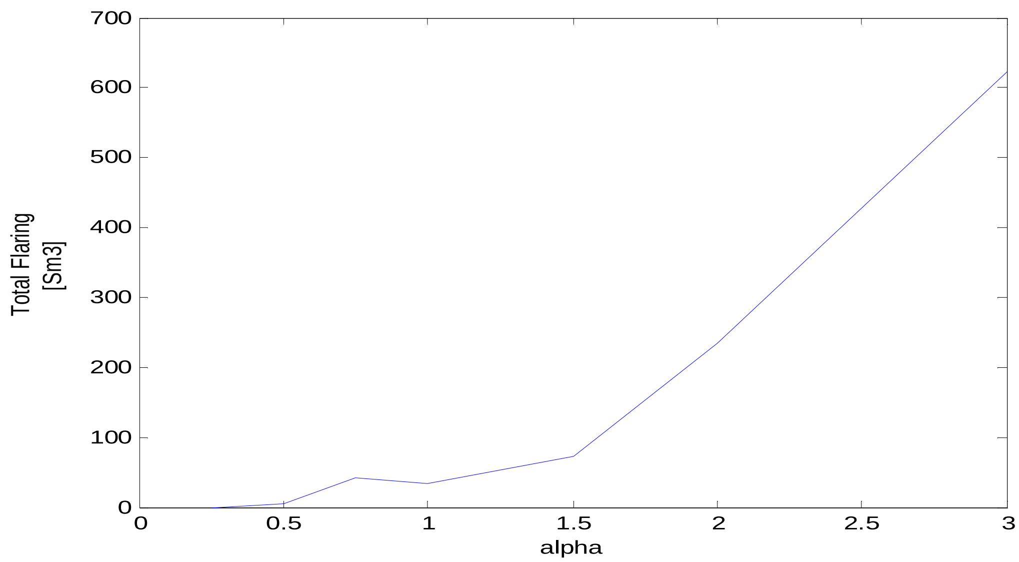

Figure 7.1 Small limitations on the system.

All 1st and 2nd order systems without time delays are inherently stable. The phase margin will never reach -180°. This coincides with the results presented in figure 7.1. With small limitations it looks as if we can speed up the control as much as we'd like still avoiding flaring. It is actually preferable to use an alpha lower than 0.25.

7.1.2 Case 2 – Limitations included

Resistances in the control structure included, represented by delays in the process as well as in the controllers giving a more realistic case. A time delay of 5 seconds (thetaFast) was induced for the faster controllers and valves and one of 10 seconds (thetaSlow) for the slower controllers and valves.

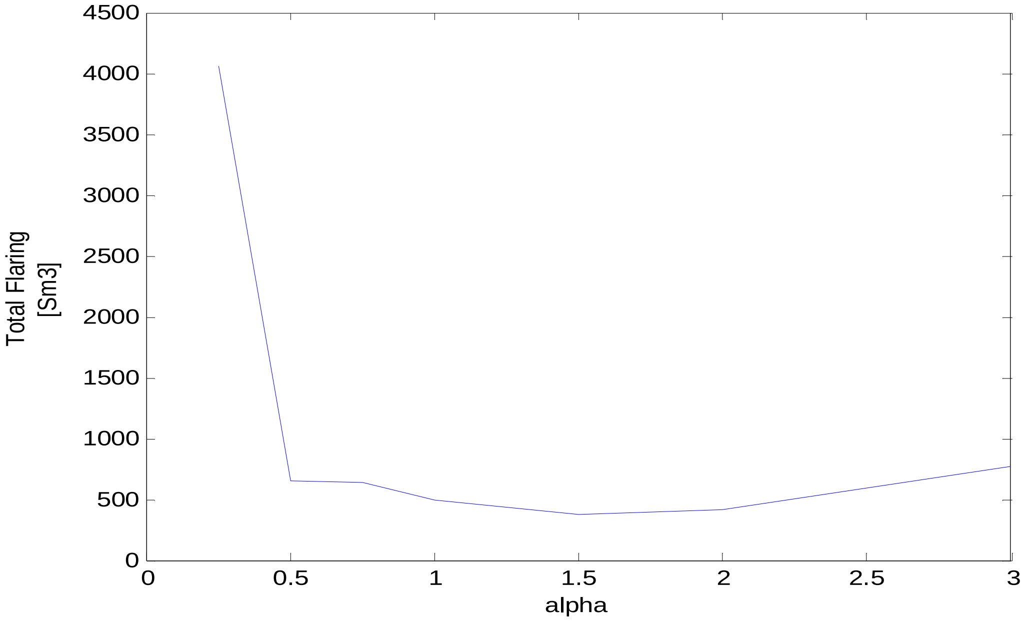

Figure 7.2 Limitations on the system.

A minimum is found for the total flaring an $\alpha \sim 1.5$, and a broader minimum in the range of alphas going from 1 to 2. The effect of the limitations put on the system can be seen as the process is unstable for $\alpha < 0.5$.

7.1.3 A comparison of cases 1 and 2

For small alpha values the amount of flaring will depend on the system limitations. For alphas in the higher range one would expect the total flaring of the system to show almost identical responses independent of the size of the limitations on the system. As alpha becomes bigger, $\tau_c >> \theta$ and we can write equations 3.4 and 3.5 as

$$K_c = \frac{1}{k} \cdot \frac{\tau_1}{\tau_{c0}\alpha + \theta} \approx \frac{1}{k} \cdot \frac{\tau_1}{\tau_{c0}\alpha}$$

and $\tau_I = \min(\tau_1, 4(\tau_{c0}\alpha + \theta)) \approx \min(\tau_1, 4\tau_{c0}\alpha)$

A similarity of the responses at the higher alpha values can be found to some extent in the results presented in figures 7.1 and 7.2. Running the cases at an even higher alpha range would provide a better look at the potential equality, but was in this case limited by the restrictions included for the levels given by equation 5.2.

We've found that for a more realistic system there will be an upper bound on τ_c due to the effective time delay. Keeping within this bound is the object of a robust plant. This is given by the SIMC-rule $\tau_c \geq \theta$.

7.2 Separate flaring control

As a safety measure it's important to keep the flaring controllers alert to sudden disturbances induced on the system at all times. As a consequence a set of tight tuning parameters was implemented on the three flare-controllers PC 2, PC 11 and PC 14.

7.2.1 Case 3 – Tight flaring control

The flare controllers were switched from active to running with constant tuning parameters for the system including limitations (Case 2).

The values chosen for the constant tuning of the flares correspond to alphas in the lower range of active control. The integral time constant of 15 actually represents an $\alpha < 0$ for the previous cases run. The choice of tuning parameters was based on the need of a rapidly responding controller. As flaring should only occur when needed out of safety measures, a response in the flare-controller should be provided as soon as it's required. A too large controller gain is unwanted though as it can cause the controller to overreact resulting in a flaring initiated before the triggering set-point is reached. The most important measure of tightening flare control is in lowering τ_I, that is providing a larger integral action. The choice of K_c has at a later stage maybe proved to be a bit too aggressive as some flaring has been detected ahead of set-point triggering.

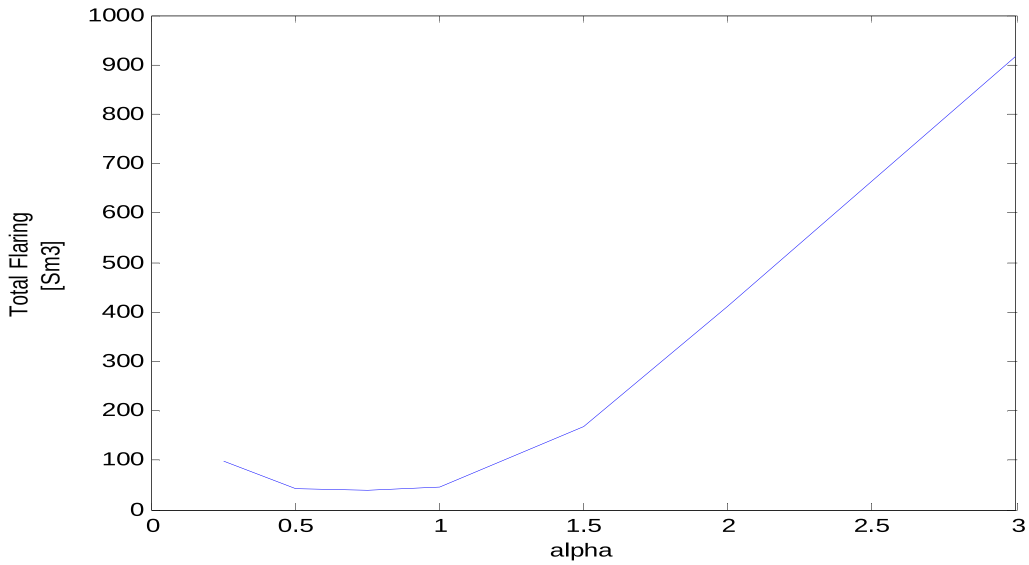

Figure 7.3 Constant tuning of the flare controllers.

After pacifying the flare controllers figure 7.3 presents the alphas providing the best control of the remaining active system mostly consisting of level controllers, but also included two very important pressure controllers. The alpha giving the least flaring is lower for case 3 than for case 2, which may be an effect of having introduced a tighter level control.

One would think that as the flares controllers are tuned tighter, they have a more rapid turn on resulting in a higher amount flared in total. The results presenting the opposite indicate that this effect is made up for by the presence of a tighter level control. One may argue that the pressures remaining in the system are also tuned tighter, but as the integral control of the levels has a stronger dependence on the tuning variable at higher alpha values it's assumed that the effect of a tuning being too smooth will be greater on these. The constant dependence on τ_c for the levels is presented in section 3.1 and the different effect of tightening control on pressures compared to levels tested in an example provided in section II.2 of the appendix II.

As far as tightening the control of the remaining pressures go, this can be seen as a possible advantage for the controller PC 10 on the pressure entering stage 3 of the compression-train. A tight control here will cause the disturbances to move further down the system and possibly going all the way back into the wells where it will be met by a lot of dynamics providing damping of the effect. For the model disregarding the outlet – inlet connection this possibly represent the ideal case, but as far as for the real process this might cause for a disturbance coming from one well to be introduced to several as an effect of the manifolds.

The other pressure controller left in the system, that is the control of the pressure in the separator will be discussed in section 7.4 concerning buffer capacities of the tanks.

Introducing a separate control of the flaring has resulted in a lower minimum of the total amount of gas released to the atmosphere for case 3 than for case 2. A tighter control on the levels may be advantageous. A range of alphas from 0.5 to 1 gives the best results.

7.3 Case 4 – Introducing the SIMC choice of $\tau_{c0} = \theta$

7.3.1 Case 4.1 - $\tau_{c0} = \theta$ introduced for case 1 with low restrictions

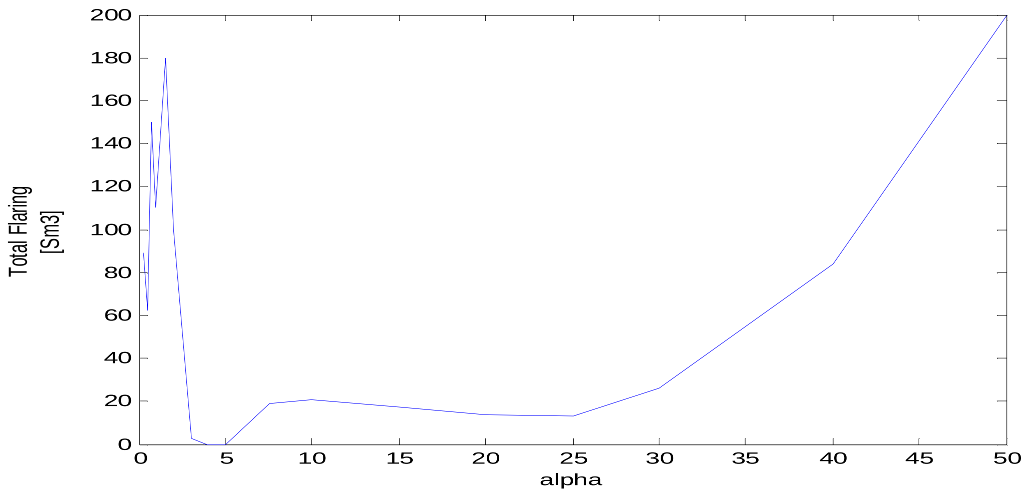

Figure 7.4 Case 4 on case 1.

The smaller τ_{c0} for the SIMC system[3] provides a look at an even tighter control scheme than the one provided by case 1, and subject to this the effect of a too tight control is seen also for the case of low limitations. This lower limit has its base in the disturbances induced as well as in the levels being integrating processes.

For the SIMC case with small limitations a $\tau_c \sim 5$ would be good choice.

[3] Compare the values presented in table 6.1

We also have a higher limit for smooth control for the low-limited case given by the SIMC rule $\tau_c = \theta \approx 0$ and combining the equations 3.8 and 3.9 for smooth control obtaining a τ_c approximately given by

$$\tau_c \leq \frac{\tau_1}{K}\frac{|y_{max}|}{|u_0|} = \frac{|y_{max}|}{|y_0|}\tau_1$$

where $|y_0|$ is the expected variation in $|y|$ with no control. This is the speed-up required for disturbance rejection. This effect can also be seen for case 1.

7.3.2 Case 4.2 - $\tau_{c0} = \theta$ introduced for the limited case 2

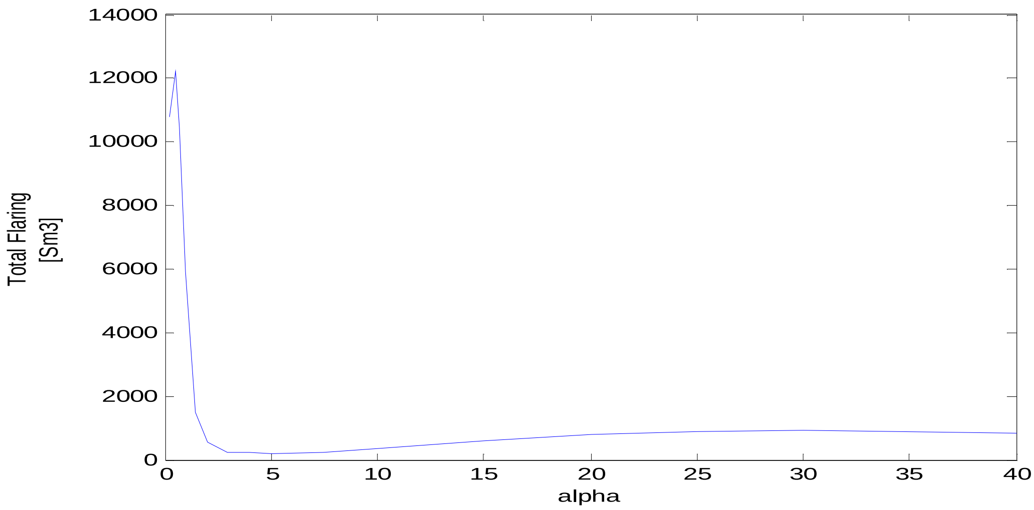

Figure 7.5 Case 4 on case 2.

The results are similar to the ones presented by case 2, but with a lower minimum. Again a τ_c = 5 seems to be a good choice.

7.3.3 Case 4.3 - $\tau_{c0} = \theta$ introduced for case 3 – including tight flare control

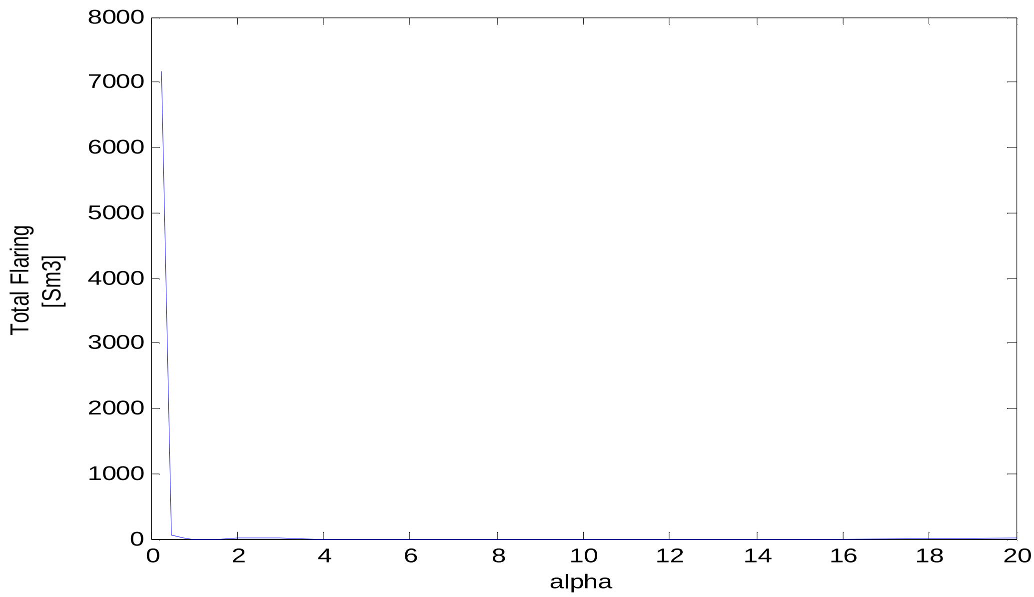

Figure 7.6 Case 4 on case 3.

$\tau_{c0} = \theta$ provide a tighter control on the system for a larger range of alphas as thetaFast and thetaSlow will be smaller than the practical values chosen for τ_{c0}[4] in the cases 1 and 2. That is for example for an alpha of 0.25, we will have an initial level control τ_c for the SIMC system of 2.5 on LC 5 which is quite small compared to one of 25 provided by the practical τ_{c0} values used.

The advantage of tuning the flaring system separately is showed also for this case.

7.3.4 A review of the results for SIMC

Looking at a sufficiently tight control range the effect of a too tight control is experienced also for the case of small limitations.

The improved result for $\tau_{c0} = \theta$ for SIMC shows that a tighter control and a degree of tight/smooth control action put on the levels compared to the pressures closer in range should be considered. This based on the improved results originating from introducing a less conservative level control compared to the pressure control and $\tau_c = \theta$ being quite a bit smaller than the practical values chosen.

[4] Compare the values presented in table 6.1

It would be interesting having a look at a case in which included the time-delay as part of the tuning parameter. Looking at the dependence of $(\tau_{c0} + \theta)$ on alpha, that is $\alpha(\tau_{c0} + \theta)$.

$$(\tau_c + \theta) = \alpha(\tau_{c0} + \theta)$$

This would be more consistent with the use of SIMC. Time delays loosing their effect at high alphas would be avoided in this manner, which would provide a better base for comparing cases 1 and 2.

7.4 Buffering capacities of the tanks

Another interesting measure in finding the optimal tuning for the process is the degree of buffering capacity of the tanks. This is a discussion on if the pressure in the separator, previously stated as possibly providing a steady state of degree freedom on the system should be given a tight or a smoother control.

A tight control provides a value for the controlled variable close to or at its set-point, but this isn't always the optimal choice. Especially when differently tuned variables affect and depend on each other, or when the available buffering-volumes are large enough, it might be favourable trying to smooth out variations/disturbances inside the tank producing a smooth varying manipulated flow. Averaging control will prevent the outlet flow from varying suddenly. [7]

To be able to buffer disturbances, volumes are needed of a size large enough to provide a significant residence time for the disturbance inflow. The volume capacity provided by the system is given by the tanks, that is the separator and the scrubbers.

As a tight level control has shown good results, and as the flaring has a more direct connection to the gas flow an averaging pressure control may be a good choice if sufficient volumes exist.

The residence time provided by the separator and the scrubbers for the gas is given as the volume of the gas in the tank divided by the gas-inflow rate. Which assuming that initial conditions are valid and using the mean inflow rate of the gas result in a residence time in the two tanks of 2.7 seconds for the separator and 1.9 seconds for the scrubber. These are small values, indicating limited buffer capacities in the system.

7.5 Overall discussion

To get a better picture of how varying the degree of tight control on the levels correspond to doing the same for the pressures would be interesting, that is the effect of introducing two alpha-values.

For the SIMC case we see that a $\tau_c = 5\theta$ is the best choice. As an alpha of 5 for this case represents a tighter tuning than an alpha of the same value for the system in the first cases the minimum found for τ_c for this case may actually be the same as the one given by an alpha of 1.5 for the previous cases. There wasn't enough time left to test this assumption.

A higher bound for the tuning is presented by all of the cases to a smaller or larger extent. This is a bound given by the smooth tuning rules, and which is presented by combining equations 3.8 and 3.9 resulting in $\tau_c \geq \frac{|y_{max}|}{|y_0|}\tau$ where $|y_0|$ is the output magnitude without control due to disturbances.

Concluding remarks and suggestions to further work

The best choice for τ_c is given as a trade-off between tight and smooth control.
An upper bound of $\tau_c \geq \theta$ has to be maintained for robustness purposes as well as a lower bound required for disturbance rejection given by $\tau_c \geq \frac{|y_{max}|}{|y_0|}\tau$ where $|y_0|$ is the output magnitude without control due to disturbances. Introducing a tighter control on the levels as well as on the pressure connecting stages 2 and 3 in the gas-train may be preferable but the effects should be studied further. The same goes for the effect of employing the buffer volume of the separator to a larger extent. This could be implemented to the model by running cases operating with several alphas, at least including one for the pressure and one for the level control, but maybe also separating the two pressure effects based on the results presented here.

A choice of τ_c equal to 5θ provides minimum flaring for the SIMC tuning.

Appendix I - Laws and international agreements on CO_2

The laws on CO_2 taxes and climate quotas as well as the petroleum act are the most important means we have controlling CO_2 emissions from the petroleum sector in Norway. The authorities also have a couple of other measures available, e.g. conditions stated in PUD (Planned Unit Development) and PAD (Plan for Installation and Operation of pipelines), and in emission and production permissions which also covers flaring. [8]

It's stated in the Petroleum Act that flaring is not permitted without consent from the Ministry of Petroleum and Energy except for what is necessary in terms of safety measures.

The Norwegian petroleum industry is subject to the CO_2-tax law of January 1^{st} 1991, which imposes fees on CO_2 from petroleum burned and natural gas released into air, as well as CO_2 secreted from processed petroleum and released into the atmosphere. Since January 1^{st} 2007 the tax is 80 øre per Standard Cubic Meter (Sm^3) gas, which corresponds to approximately 330 kr per tonn CO_2. [9]

The established CO_2 quota duty system today only covers some installations onshore, but including offshore instalments is a proposal of high importance when the act is up for revision some time later this year (2007). [10]

The Kyoto protocol, an agreement made under the UNFCCC (United Nations Framework Convention on Climate Change), has the objective to achieve stabilization of greenhouse gas concentrations in the atmosphere and keeping them at a level that would prevent dangerous anthropogenic interference with the climate system on a global scale. This protocol has bee signed and ratified by Norway where we have agreed to only increase our CO_2 emissions by 1% above 1990 level by 2012.[11]

Appendix II– Testing assumptions and providing supporting examples

II.1 The assumption of low interactions between the loops

The assumption of low interaction provides the basis for tuning the system loop-by-loop, each controller separate of the others. The assumption was tested for two of the most important pressure controllers in the system. That is interactions between the pressure control in the separator and control of the pressure going out of the 2nd stage compressor. This was done by having both the controllers in open loop, first implementing a step-change on the output of the controller in the oil–train observing the effect on both processes and then doing the same for the controller in the gas train. As for the first case the step provided in the lower part of the system showed an effect on the process further up, but not the other way around. Closing the valve connecting the two last stages of compression a little by introducing a step of -1 had no effect on the pressure in the separator and we therefore have an a RGA = 1 representing no interactions between the loops. There may still be interactions present between other loops, but this example at least gives an indication of the loop-by-loop tuning being somewhat sufficient.

II.2 The effect of tightening control on levels versus pressures

This is an example supporting the suggestion of introducing a tighter control on the levels. Comparing cases 3 and 2 as well as the results of case 4 compared to the three first cases it's shown that a tighter control provides a lower minimum for the total flaring produced. In this section an example will be provided supporting that a tightening of the levels in the system is the main reason for the improved case. An effect provided by pressure control tightening is not disregarded though. The difference in the effect of a tightening grip provided on an integrating process than a first-order one is given by the SIMC rules presented in 3.1 showing that the levels are much more dependant on the tuning parameter τ_c than the pressures for large values. The reason for this is the dependence of the integral action on τ_c. For pressures at larger τ_c values the integral action will be given by the time-constant of the process, which is independent of the tuning implemented, whereas it for the levels will be strongly dependant. This effect on the system is shown in an example comparing the effect of an alpha = 0.5 to an alpha = 1 for the pressure, PC 1 to the oil-, LC 3, and water-, LC5, level control of the separator.

Table II.1. The effect of tightening control on levels vs. pressures

Alpha		PC 1	LC 3	LC 5
1	K_c	-10	-2	-0.78
	τ_I	30.044	640	440
0.5	K_c	-16	-3.8	-1.4
	τ_I	30.044	340	240

As can be observed from the values presented in table II.1 a halving of the alpha/τ_c represent almost a doubling of the integral action put on the system, whereas it present no change for the pressure. As for the proportional action the dependence will be the same.

Appendix III – Disturbances on the manipulated variable

The effect of disturbances on the manipulated variable in the system hasn't been tested separately, but has been viewed as part of predicting possible effects of disturbances introduced to the system. Having a look at this case presents a greater perspective of the results, but cases should be run looking at these measures these effects have on the system.

III.1. Disturbances on the input u

An upper bound on u due to input saturation

When input saturates we've used all our range of control e.g. when a valve is fully open, it won't respond to signals from the controller telling it to open further and there exist a chance of process build up in the system.

We generally want to avoid a too large u because it causes less wear and saves input energy, and also because the manipulated variable often can be a disturbance to other parts of the system. In particular we usually want to avoid fast changes in u. [12]

On a system you have the possibility of a disturbance entering the system on the input or on the output of the process as presented below in figure III.1.

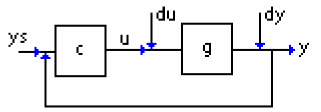

Figure III.1 Two types of disturbances on the system

III.1.1 For a signal ys (set-point change) going through the system

we have:

$$\frac{\mathrm{y}}{\mathrm{ys}} = \frac{gc}{(1+gc)} \tag{III.1}$$

which will be our reference case.

III.1.2 For disturbances introduced on the input of the controller

due to input saturation we have:

$$\frac{\Delta u}{-du} = \frac{cg}{(1+gc)} = \frac{1}{\tau_c s+1} \tag{III.2}$$

which is equal to $\frac{y}{ys}$ stated in equation (III.1), and will therefore not have any affect on the tuning strategy. It will not lead to instabilities.

The negative sign on du in the expression isn't of importance as the controller easily can be turned around, but as the expression is stated here, the result of a disturbance du going through the system will actually be somewhat like what's presented below

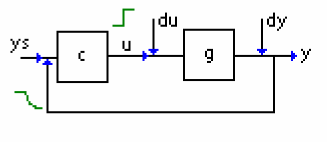

Figure III.2 Disturbance on u due to input saturation.

III.1.2.1 Filtering of input disturbances

We generally want to avoid a too large u because it causes less wear and saves input energy, and also because the manipulated variable often can be a disturbance to other parts of the system. In particular we usually want to avoid fast changes in u. [12]

Filtering of du may be wanted, that is there may be a requirement $\tau_c \geq \tau_{c,\text{input}}$ because fast changes of u are not desired.

III.1.3 Effect of disturbances on the output, dy, on the input u

Disturbances dy on the output y or set point changes ys's effect on u

Can be presented as shown below:

$$\frac{\Delta u}{-dy} = \frac{c}{(1+gc)} = \frac{1}{g}\frac{gc}{(1+gc)} = \frac{1}{g}\frac{1}{\tau_c s+1} \qquad \text{(III.3)}$$

The expression differs from the one given in equation (III.1) indicating that instabilities may occur.

Looking at the effect on the system both initially and at s.s we get,

a) At steady-state, given by the final value theorem stating that as $t \to \infty$, $s \to 0$, we have;

$$\frac{u}{dy} = \frac{1}{g(0)} = \frac{1}{k} \qquad \text{(III.5)}$$

which is something the system has to be able to handle. It's not a tuning related problem.

b) Initially as given by the initial value theorem, that is $s \to \infty$ and assuming the process transfer function can be given as a first order with out time delay,

$g(s) = \dfrac{k}{\tau_1 s + 1}$ we have

$$g(\infty) = \frac{k}{\tau_1 s} \Rightarrow \frac{u}{dy} = \left[\frac{1}{g(s)} \frac{1}{\tau_c s}\right]_{s\to\infty} = \frac{1}{k}\frac{\tau_1}{\tau_c} \qquad \text{(III.6)}$$

Which implies that we have an overshoot in u initially given by the speed-up $\dfrac{\tau}{\tau_c}$.

If we try to speed up the process when a disturbance, here represented by a step-change in figure I.3, is on the output, that is we try to reach the (upper) red line on the figure, from an original position of the (lower) black line

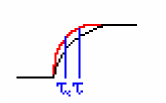

Figure III.3 Effect of a disturbance on y

we'll get a system looking somewhat like

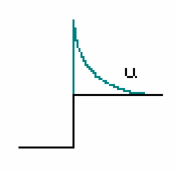

Figure III.4 Effect of dy on u

which is unwanted.

The additional effects of disturbances on the manipulated variable result in some additional comments on the choice of the tuning parameter τ_c for the system.

1. Maximum speedup to avoid input-saturation due to output disturbances

   $$\tau_c \leq \frac{|u_y|}{|u_{max}|}\square\tau \qquad \text{(III.7)}$$

   where $|u_y|$ is the input change required to reject output disturbance or set-point change.

2. The time required for filtering input disturbances $\tau_{c,input}$

   $$\tau_c \geq \tau_{c,input} \qquad \text{(III.8)}$$

3. The time required for acceptable setpoint tracking, $\tau_{c,setpoint}$

   $$\tau_c \leq \tau_{c,setpoint} \qquad \text{(III.9)}$$

   which is generally wanted as small as possible.

Appendix IV Additional results and cases

Table IV.1 presents an overview of all the cases run. Additional results for the cases presented in the main report as well as a couple of additional cases are presented. The supplementing cases may be of interest with regards to further work on the topic. The results are found on an attached disc.

Abbreviations used in the table:

- AF – Amplification factor.
- DS – Disturbance start, seconds into the sequence.
- ST – Simulation time in hours.

IV.1 Additional cases:

Some additional cases have been run, testing different aspects of the minimum flaring problem. The cases are presented below with a short comment on possible utilization in further work.

Case 5: represent the case of only having constant tuning parameters provided on the flare-controller in the 2nd stage of the gas-train. The constant tuning parameters being $K_c = -6$ and $\tau_c = 15$ and introduced to a system

a) With low limitations
b) Including limitations

Can be used in showing the effect of introducing constant tuning parameters only on the most active flare-controller, and relating this effect to the degree of limitations in the system.

Case 6: tests $K_c = -20$ for case 5b showing the effect of introducing a higher controller gain on the controller.

Case 7: Introducing constant tuning parameters for the flare controllers as for case 3, but using a proportional gain of $K_c = -6$ instead of -20. Showing the effect of introducing a smoother controller gain on the whole system as well as compared to case 6, the effect of tuning all the flaring controllers passively to just the one.

Case 8 – Introducing the SIMC choice of $\tau_{c0} = \theta$ disregarding limitations

Case 9 – The early cases before including restriction on the levels and pressures of the tanks

a) including start-up
b) starting at a stable state

Table IV.1 Additional results and cases

Case viewed	Case number	DS [s after start]	AF	ST [hours]
Case 1	30	0	7	3
	31	5000	7	3
Case 2	23	5000	7	3
	25	0	7	3
	26	18600	7	3
	32	15000	3	3
	39	15000	6	3
Case 3	40	0	7	3
	41	5000	7	3
Case 4.1	28	5000	7	3
Case 4.2	35	5000	7	3
	42	0	7	3
	43[5]	5000	7	3
Case 4.3	38	0	7	3
	44[6]	0	7	3
Case 5a	9	5000	7	3
	10	18600	7	2
	11	18600	6	2
	12	5000	7	12
	13	-	-	-
	14	15000	7	3
	15	15000	6	3
Case 5b	16	15000	7	3
	17	18600	7	3
	18	15000	8	3
	19	18600	6	3
	20	15000	6	3
	21	5000	7	3
	22	5000	8	3
Case 6	45	15000	7	3
	46	5000	7	3
Case 7	34	15000	7	3
Case 8	47	0	7	3
Case 9a	0	18600	6	1
Case 9b	1	18600	6	1
	2	15000	6	1
	3	15000	6	1
	4	15000	8	1
	5[7]	0	6	24
	6	0	6	5
	7	0	6	12
	8	43200	6	12

[5] Redoing case 42
[6] Redoing case 38
[7] Memory allocation error. The dime aspect of 24 hours was most likely too long

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