



NTNU – Trondheim
Norwegian University of
Science and Technology

Model Predictive Control of Compressor Systems

Arthur Batalov

Master of Science in Engineering Cybernetics

Submission date: July 2012

Supervisor: Morten Hovd, ITK

Co-supervisor: Sigurd Skogestad, IKP
Marius Støre Govatsmark, Kårstø, Statoil

Norwegian University of Science and Technology
Department of Engineering Cybernetics

Abstract

The purpose of this thesis was to investigate the suitability of applying MPC for the performance control of compressor systems. Dynamic model of the Kårstø gas compression system was used as a process model in D-spice simulator. It consists of two stages with three parallel compressors on each and intercooling to reduce the temperature and total required compression work. The study has deepened into direction of optimization of operation for this system.

A plantwide control design procedure was applied for the systematic approach of the control problem. Economic potential for the optimization was examined for the different operational conditions and controlled variables were stated.

The main result of the study, a developed and tested MPC application for the optimization of operation of export gas compression system in Septic MPC. It was discovered significant nonlinear interactions in the plant during the modeling and simulation. Acquired step response models works properly only in a specific operational region where they taken.

Abstract

Formålet med oppgaven var å undersøke egnethet å anvende MPC til ytelsesregulering av kompressorsystemer. Dynamisk modell av Kårstø gasskompresjon system ble brukt som en prosess modell i D-Spice simulator. Den inneholder to trinn med tre parallelle kompressorer og mellomkjøling for å redusere temperaturen og total nødvendig arbeid. Studien har spisset seg i retning av drift optimalisering for dette systemet.

En plantwide kontroll design prosedyre ble brukt for å gjennomgå kontroll problemet. Økonomisk potensial for optimalisering ble undersøkt for de ulike drifts forhold og kontrollerte variabler ble foreslått.

Hovedresultatet av studien, er en utviklet og testet MPC applikasjon for drift optimalisering av eksport gass kompresjon system i Septic MPC. Det ble oppdaget betydelige ulineære interaksjoner i anlegget under modellering og simulering. Sprangrespons modeller fungerer bra, men bare i ett operativ området som de er ment for.

Acknowledgments

The following thesis is done in collaboration with Statoil, Kårstø. I would like to thank my contact person and co-supervisor Marius Støre Govatsmark from Statoil, for the guidance and motivation throughout the project. I would also like to thank my supervisor, professor Sigurd Skogestad for the assistance and inspiration. Also I would like to thank professor Morten Hovd for initial help with D-spice.

Contents

List of Figures	iii
Nomenclature	v
1 Introduction	1
1.1 General information	1
1.2 System description	1
1.3 Original task	3
1.4 Completed work	3
1.5 Outline	3
2 Theory	5
2.1 Natural gas	5
2.2 Compressors	6
2.2.1 General	6
2.2.2 Centrifugal compressor	7
2.2.3 Surge and Rotating stall. Performance Map	8
2.2.4 Cooling	10
2.2.5 Compressor Control	10
2.2.6 Multi-stage compressor system	11
2.3 Model Predictive Control	12
2.3.1 Principles of MPC	12
2.3.2 Basic formulation and prediction	13
2.4 SEPTIC	14
2.4.1 Application	14
2.4.2 Tuning	15
2.5 Dynamic simulation: D-Spice	16
2.5.1 OPC server	16
2.6 Summary	16
3 Matlab	17

4	Application to Kårstø	19
4.1	Plant	19
4.2	Plantwide control design	20
4.3	Implementation	25
4.3.1	D-SPICE	26
4.3.2	SEPTIC MPC	27
5	Simulation Results	31
5.1	Models	31
5.2	Scenarios	34
5.3	Discussion	43
6	Conclusion	45
6.1	Problems faced during the thesis	46
6.2	Further Work	46
A	Drawings	47
B	System responses	51
C	D-spice	62
C.1	Changes in model	62
C.2	Add a PID controller to D-spice	63
C.3	Comments	64
D	MATLAB	65
E	OPC Server configuration	71
E.1	XML file for OPC Server	73
E.2	Definition file for D-Spice	74
F	D-Spice	75
F.1	XML file for start-up configuration	75
F.2	Registry file	76
F.3	Operational points	78
G	SEPTIC configuration	79
	Bibliography	81

List of Figures

1.1	Gas Transport System	1
1.2	Statpipe Compression System	2
2.1	Multi Stage Centrifugal Compressor	7
2.2	Centrifugal compressor rotor	8
2.3	Performance map	9
2.4	Capacity and Antisurge control	11
2.5	Load Balancing Control	12
2.6	Basic scheme of MPC	13
3.1	Matlab GUI interface for MPC	17
4.1	Performance map	20
4.2	Potential for compression cost[NOK/hour] (2010 prices)	23
4.3	Septic Main Application	28
4.4	Modeling	30
5.1	System behavior on step in booster suction pressure	32
5.2	Application test	35
5.3	Minimizing consumed energy for Booster compressors	36
5.4	Medium rate optimization	39
5.5	Only control of DEV is active, DEV=0,15	40
5.6	Response on ramping of pipeline pressure	42
A.1	Statpipe Compression System	48
A.2	Plant Simulator	49
A.3	Compressor Instrumentation in D-Spice	50
B.1	Model for full rate, step in booster suction pressure	52
B.2	Model for full rate, step in feed rate	53
B.3	Model for full rate, step in output pressure	54
B.4	Modeled with high output pressure to provoke ASV on full rate. Higher ASV values then in other cases	55
B.5	Low rate, step in output pressure. One of the compressors take over the whole recycling. Model not suited for use	56

B.6	Response to fast disturbances. Flow rate $\pm 5\%$, pressure +1 bar	57
B.7	Optimization of TotalWorkBooster after significant flow drop(-25%)	58
B.8	Optimization of energy consumption for first stage compressors	59
B.9	Optimization of TotalWorkBooster on the high pipeline pressure(185 bar) .	60
B.10	Behavior on significant pipeline pressure increase. Full panel	61
F.1	Operational points	78

Nomenclature

ASV	Antisurge valve
Dev	Deviation of operation point from control line
F_{FG}	Fuel Gas flow to first stage [sm ³ /h]
F_{SG}	Sales Gas flow [sm ³ /h]
P_{El}	Electricity price
P_{SG}	Sales Gas price
$STPBC$	Statpipe Booster Compressor
$STPBCi$	Statpipe Booster Compressor where $i=\{A,B,C\}$
$STPSGC$	Statpipe Sales Gas Compressor
$STPSGCi$	Statpipe Sales Gas Compressor where $i=\{A,B,C\}$
W_{El}	Amount of energy consumed [kW/h]

Chapter 1

Introduction

1.1 General information

Kårstø gas processing plant has a major role in Norwegian and European gas processing industry. It is located at the west coast of Norway north for Stavanger and receives natural gas from several fields on the Norwegian continental shelf. Figure 1.1 shows the Norwegian transport system for natural gas.

At the plant the unprocessed gas is separated into different components, such as dry gas and natural gas liquids (butane, ethane, propane and naphtha). The dry gas is transported further to Europe through the underwater pipe system while the NGL and condensate products are exported with tankers by sea. There are 600-700 ship loads from Kårstø each year.

Gas products is required as raw material in various industries, such as chemical, plastic, paper, etc. It is also important as a source of energy since it is cleaner than other fossil fuels. These are the reasons why Norwegian gas exports is at a record levels.

1.2 System description

Rich gas from the North sea after some preprocessing and slug elimination goes into the Statpipe treatment facility. The lightest gas (methane) gets separated in a methane column

a major role in Norwegian and

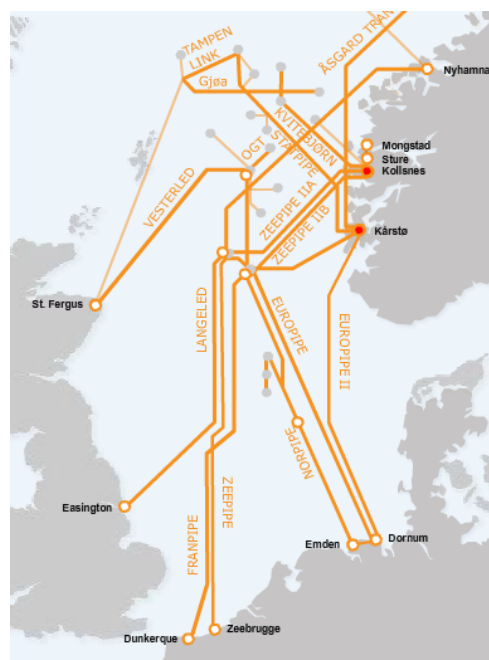


Figure 1.1: Gas Transport System

while the heavier components go to further fractionation. Before the sales gas can be sent to export it must go through the compression. Such compression system is illustrated in figure 1.2 and in appendix A.1. It consists of two stages with three parallel compressors on each and intercooling to reduce the temperature and total required compression work. First stage compressors were named as 'Statpipe Sales Gas Compressors' and will be referred to in the following text in abbreviated form as STPSGC. Similar principle apply to the 'Statpipe Booster Compressors' at the second stage, which will be called SPTBC. In addition, one more letter (A,B,C) may be added to address the specific compressor on the stage. First stage compressors runs on fuel gas while the remaining three are electrically powered.

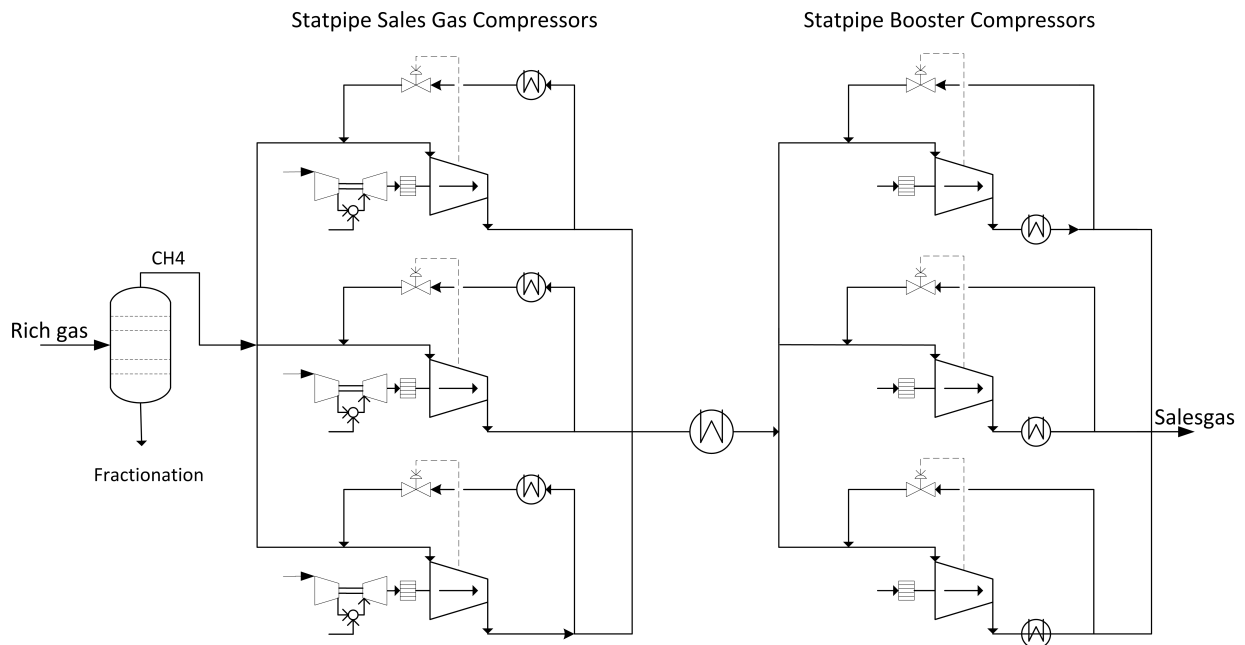


Figure 1.2: Statpipe Compression System

Each compressor has two control systems. Capacity control and antisurge control. First one take care of the load sharing between stages by manipulating the engine speed of the compressors and keep the suction pressure constant. Also it provides equal load sharing between machines in parallel on the same stage by holding the deviation from the control line equal. The antisurge control provides a stable operation and keeps the compressor from damage. It first become active when the flowrates are decreased. In that situation compressors tries to operate on the control line with equally open recycle valves on the stage. Compressor control will be covered deeply in the next chapter.

Dry gas(sales gas) gets compressed and sends from processing plant to the customers in Europe.

1.3 Original task

This is an original proposed task. See below for what is done.

The purpose of the thesis is to examine the suitability of MPC for capacity control of compressor systems.

1. Improve load distribution between different compressor stages by using MPC to manipulate the setpoint of suction pressures of each compressor stage for example to minimize the consumed electrical energy on second stage.
2. Consider to replace existing capacity control with MPC instead of PI(D) controllers and logic modules. Antisurge control remains unchanged.
3. Consider to improve load distribution between compressor stages and replace existing capacity control
4. Consider use of alternative variables to distribute load sharing between parallel compressors, e.g. marginal cost instead of equal distance to the surge control line.

1.4 Completed work

1. Developed plantwide control design with attempt to systematic approach of the control problem
2. Developed and tested MPC application for the optimization of operation of export gas compression system

1.5 Outline

Chapter 2 gives an introduction into compressor theory, its control methods and some of the associated challenges. Chapter 3 presents a Matlab application which was developed on the early stage of the project for communication with D-spice. Chapter 4 presents plantwide control procedure and practical implementation of the Septic MPC application. Chapter 5 unveils and discuss received results. Chapter 6 conclude on the project results and suggest further work.

Chapter 2

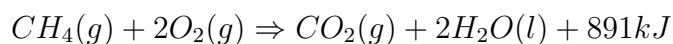
Theory

This chapter gives insight into necessary theory. First section introduce the term - natural gas. Second provides an introduction to the physical structure and working principles of compressors, introduces some challenges associated with their operation and shows basic mechanisms which can be applied to control such machinery. Next section describes Model Predictive Control and its working principles. Then, application of SEPTIC MPC is presented. Last section introduce dynamic process simulator D-Spice which was used through the whole working period.

2.1 Natural gas

Natural gas is a generic term for the gas that is formed by the breakdown and transformation of organic material beneath the earth's surface. It is an important source of the world's supply of energy and also used in a wide amount of industries such as chemical, pulp and paper, stone, glass, plastic, etc. It is colorless, shapeless and scentless in its original form.

The term natural gas implies by itself the mixture of the hydrocarbon gases. The composition of it, varies dependent on the production field but the main component is a methane. It can also include ethane, propane, butane and other gases. Methane molecule is denoted as CH_4 and formed by a one atom of carbon and four hydrogen atoms. It is combustible and while burning it produces high amount of energy. One molecule of methane combined with two oxygen molecules react to form one molecule of carbon dioxide, two molecules of water and 891kJ of energy.



Burning the methane is the cleaner way to produce the energy compared to other fossil fuels such as oil and coal since it emits fewer byproducts.

The formation of the natural gas under the ground took place over very long period. Basically it is a rests of the animals and plants who lived for many millions of years ago.

Such organic mass has formed today's oil and gas deposits by exposure to a very high pressure and temperature for a very long time.

There are two terms applied to the natural gas. 'Wet' and 'Dry'. 'Wet' gas implies the gas from reservoir to the processing plant which contains different components. At the plant the various products are separated and 'dry' gas is one of them. It is almost pure methane. Dry gas is distributed mostly through the pipes, but can also be transported with tankers, either in compressed form CNG (Compressed Natural Gas) or cooled to -162 degrees Celsius LNG (Liquid Natural Gas). The measurement standard employed at the Norwegian continental shelf is one standard cubic meter (1 Sm^3). It is defined at 15°C and 1 atm.

2.2 Compressors

2.2.1 General

Compressor is a device that increases the pressure of the gas by supplying the mechanical power. There are four general types of the compressors distinguished by their working principles. Centrifugal, axial, reciprocating and rotary. Centrifugal and axial compressors use the rotating elements inside to increase the velocity of the gas and then transform the kinetic energy to the pressure energy by changing the direction of the flow. Reciprocating and rotary compressors use a positive-displacement principle to increase the pressure by decreasing the volume of the trapped gas.

Centrifugal compressors use a rotating impeller to accelerate the velocity of the flow by forcing the gas to the edge of the impeller. Then it uses a diffuser to decelerate the flow and increase the pressure. Flow from the centrifugal compressor is perpendicular to the rotating shaft. Several impellers usually combined to form a multistage compressor. Axial compressors design contains a several rows of rotating rotor blades and casing with a static stator blades. The flow goes parallel to the rotating shaft. Kinetic energy is increased by the rotating blades and transformed to the higher pressure by the stator blades.

Reciprocating compressors operate in a cyclic manner. Each cycle the trapped amount of gas in the cylinder gets compressed by the piston which reduce the volume and increase the pressure. The gas is then discharge and a new volume of the gas is compressed by the piston again. Such cycle is executed several hundreds times per minute. Rotating compressors use two rotating elements to create pockets filled with gas. During the operation, volume of the gas pockets is decreased and pressure is increased. There are several types of the rotary compressors which utilize same general principle.

Axial compressors dominate in the jet engine industry because of the capability to deal with a higher flow rates related to the frontal area. Centrifugal compressors is best suited for the high flow rates and lower pressures. That's why they are preferred in the process industry and will be covered more deeply in the next section.

2.2.2 Centrifugal compressor

The main parts of a single stage centrifugal compressor are the impeller where the acceleration of the velocity take place and a diffuser where the kinetic energy transforms to the potential energy. Compressor design consist usually of the several combined stages. Where all impellers attached to the rotor shaft while the diffusers are the bended passage ways between nearby impellers. Figure 2.1 shows physical structure of a multi stage centrifugal compressor. The gas is sucked into the inlet section and sent further into the eye of the first impeller. There it is pressed by the rotational motion to the edge of the impeller and into the bended passage so called diffuser. The pressure gets increased and gas goes further to the next impeller and through the next stage in the similar manner. In such way each stage contributes to build up the overall discharge pressure.

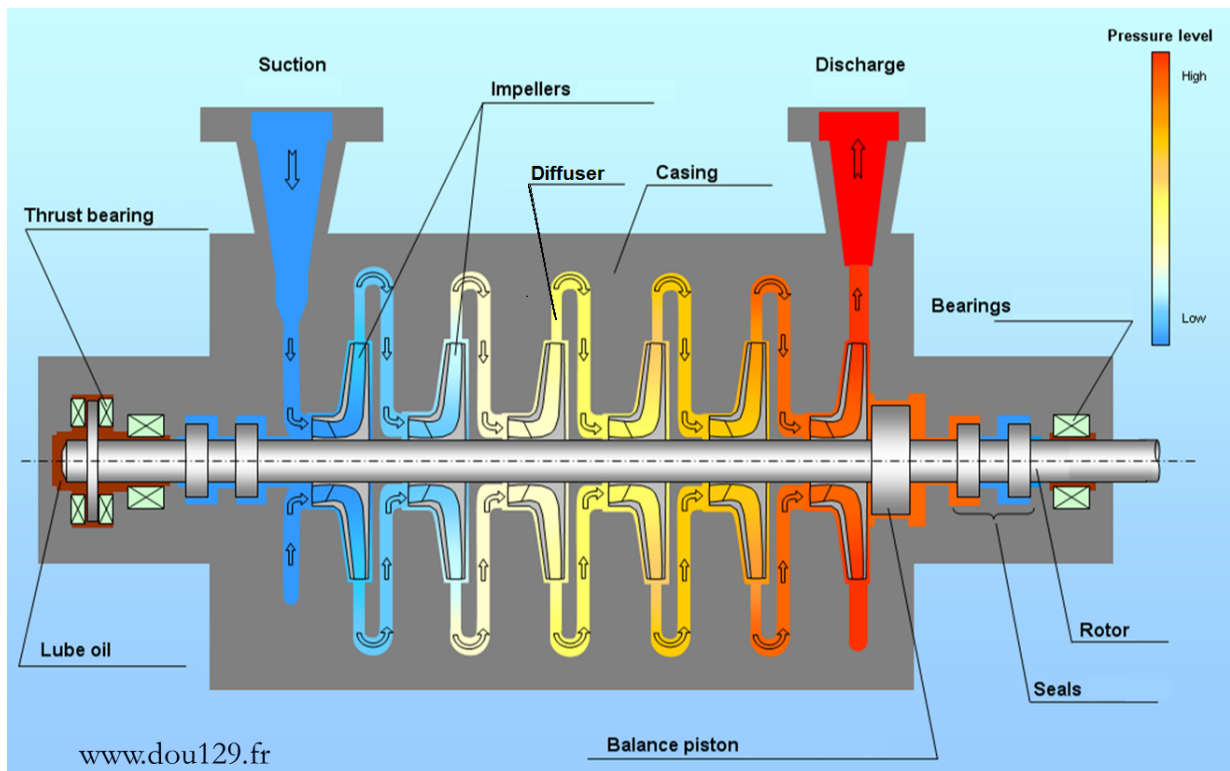


Figure 2.1: Multi Stage Centrifugal Compressor

The compressor performance is decided by the design of the impellers and diffusers. Figure 2.2 shows a rotor shaft with several impellers. Inlet guide vanes into the eye of the impeller, blade angles, vaned or vaneless diffuser are the most vital factors that determine the capacity, efficiency and stability of the compressor. As an any mechanical device, the compressor is not a perfect machine and there is always some inefficiency caused by the friction. Common parameter used for calculations is efficiency factor which determines

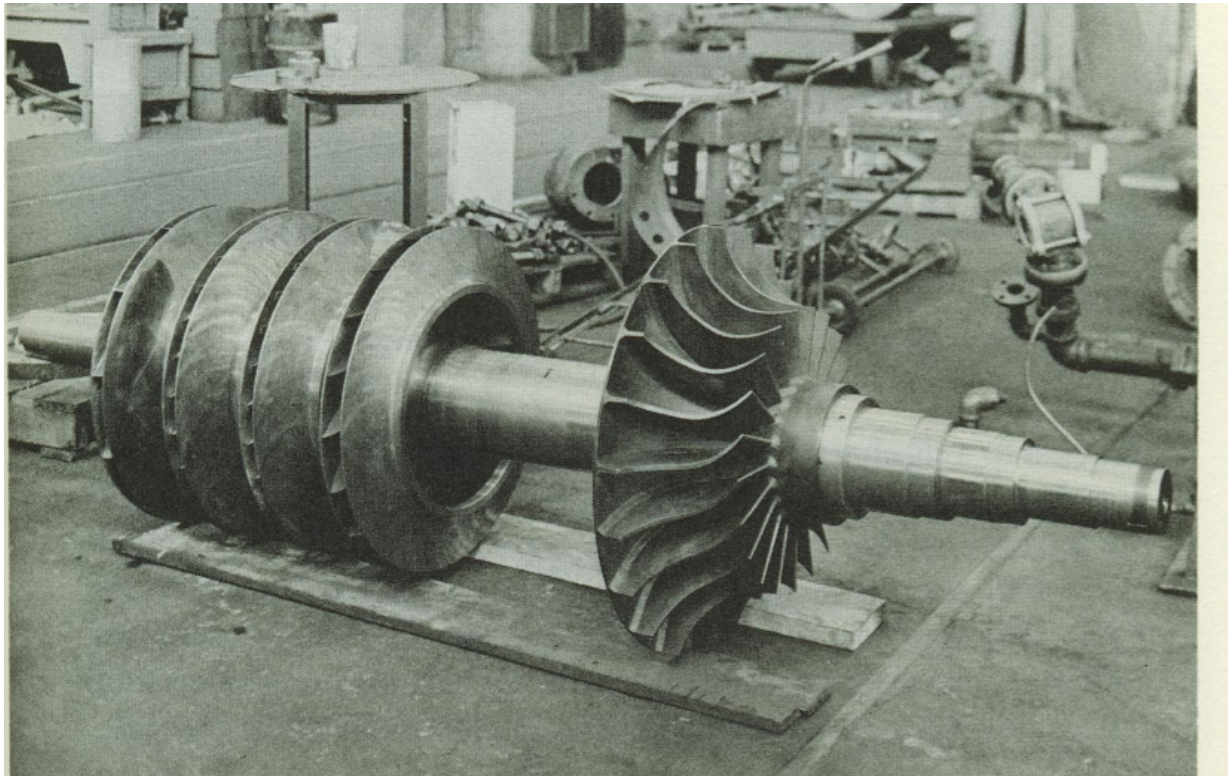


Figure 2.2: Centrifugal compressor rotor with five closed impellers and one open impeller (Nisenfeld 1982)

ratio of the supplied power to the output pressure. Usually is about 0,75. efficiency

Compressors in the process industry can be driven by gas turbines or electrical motors. Gas turbines use bypassing gas as a fuel and do not need an external supply. It consists of a compressor, combustion chamber and turbine. The compressor is responsible for the air supply into combustion chamber where the fuel gas is burned. The product of the combustion are forced further into the turbine section where the useful energy is extracted in form of the shaft power. The second alternative is to use electrical power. The electrical motors used in compressor stations can be driven with AC and DC current. AC motors is divided into three types: synchronous, asynchronous and linear. Synchronous motors are generally used with high power compressors. They have very high power factor and are not very sensitive to changes in load. Induction(or asynchronous) motors are mostly used in centrifugal pumps. They are used in low power networks and when the considerable starting torque is required. Linear motors are typically used with a reciprocating compressors.

2.2.3 Surge and Rotating stall. Performance Map

Surge and rotating stall are two instabilities associated with centrifugal compressors running at the reduced flowrates. Surge is a phenomenon that occurs when the flow throughput

is decreased below its minimum for a given speed. In such moment the head pressure generated by the compressor is less than downstream pressure and the reversal of flow takes place. Rotating stall is an aerodynamic disturbance which usually occurs in one or several stages when operating near the surge point. It describes a condition when the pressure ratio does not change in a stable manner with flow rate. During the surge, severe and rapid oscillations in flow and pressure disperse through the system. It causes loud noise and intense vibrations which can damage machinery. The task for the engineer is to develop an efficient control system that would prevent surge.

After the design and manufacture of the compressor the operational characteristics are fixed. The surge and choke (stonewall) points remain the same. Figure 2.3 illustrates the performance map of a centrifugal compressor with its limits. When operating to the left of the surge limit, the machine will suffer from flow reversal and become unstable. On the other end of the map is the choke limit. From this point, the compressor is unable to increase flow throughput and it can be seen as a capacity limit for the device. Curved green lines represent the variable speeds of the compressor from the lowest at the bottom to the highest on top. The abscissa is flow throughput in cubic meters per hour. The ordinate usually gives a pressure ratio of the compressor. Figure 2.3 is taken from D-spice where the Y-axis shows a polytropic head which denotes the required work in joules per kilogram gas. For a given constant speed, the polytropic head will decrease with an increase in flow rate and vice versa. To run the device safely, the operation point must be always inside the middle section.

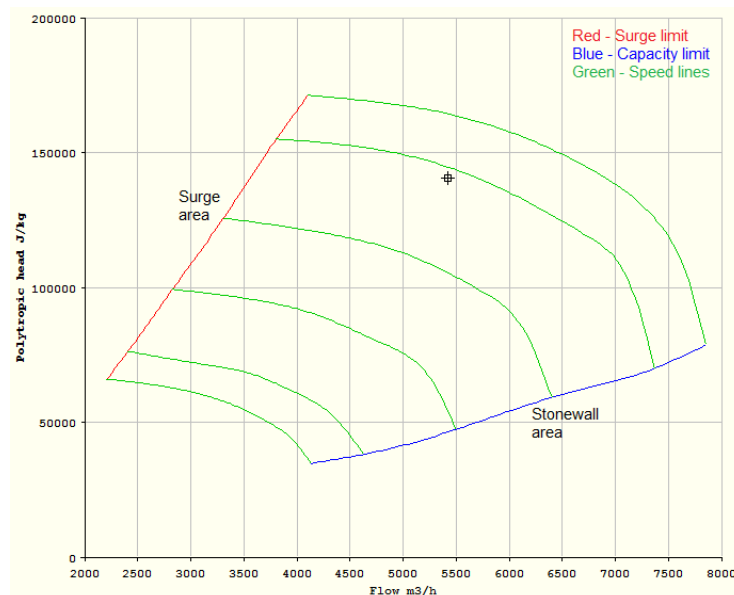


Figure 2.3: Performance Map(D-Spice)

2.2.4 Cooling

As stated and shown in the (Skogestad 2009, p. 144) to reduce total work during the compression it is optimal to cool as much gas as possible on the intermediate stage.

2.2.5 Compressor Control

Compressor in operation consumes large quantities of energy. That is why a well functioning controller will give energy and cost savings. There is always trade off between precise control and investment cost to achieve it but the main control objectives for the any compressor can be defined as:

- personal and environmental safety
- operation within limits and surge avoidance
- accurate CV control
- minimize energy consumption

Out of this follows two major elements within compressor control subject: a capacity control and antisurge control. First one takes care of operation such that the process requirements are met while the second one is related to stability and equipment protection.

There are several available variables which can be used to design this two controllers. The common concept today is to use either suction valve, discharge valve, rotational speed or inlet guide vanes as a manipulated variable for the capacity control. Each choice has its benefits and drawbacks and depends on the process to control. In such layout it is common to use the recycle valve as a manipulated variable for the antisurge control. It starts to act first when the flowrate through compressor falls to its minimum for a given speed by opening the recycle valve and returns back the necessary amount of gas to the compressor inlet. In such way it moves operation point to the right of the surge line. Described principle lies at the basis of today anti-surge controllers implemented in D-Spice process simulator. Figure 2.4 illustrate essence of it. Speed controller and recycle valve are used for capacity and antisurge control, respectively. Suction and discharge valves left fully open and not in use in automatic control but can be operated manually by for example an operator at the plant, in case of emergency situation or necessary shut down.

Both presented control functions affect the mass balance around any compressor in surge. That's why, an intense interaction between these two systems must be expected. Since an antisurge control is vital for the stable operation it must be given a priority over the capacity control to avoid fatal consequences and damage of the equipment. However the interaction between functions can be reduce by efficient decoupling. Such that both systems are aware of each other.

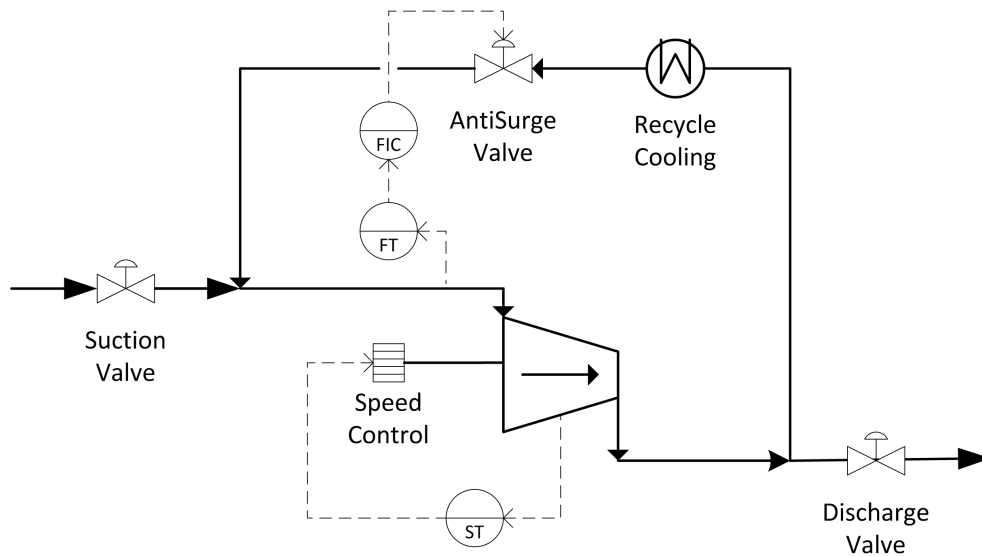


Figure 2.4: Capacity and Antisurge control

2.2.6 Multi-stage compressor system

The compressor system shown in figure 1.2 presents many challenges for the control engineer. Three compressors in parallel at the first stage provides suction pressure to the triplet of compressors at second stage. In addition to capacity and surge control previously discussed emerges a new challenge of load balancing among parallel compressors on first and second stages. The antisurge control becomes even more difficult since the gas recycling at one compressor affect both the suction conditions and amount of input flow to other two compressors. The start ups and shut downs of a single compressor must be performed very carefully. Such that discharge pressure is balanced before the compressor ties into the joint manifold and goes 'online'. If it is done too early it can lead to heavy surge. If a compressor failing out, the missing load must be taken over by others rapidly or it can lead to a gas starvation for the process. All things above is further aggravated by the fact that it does not exist two absolute identical compressors. Even two similar devices from the same manufacturer will have some minor differences which will only increase with time due to aging, wearing of various components, number of shut downs and etc.

Figure 2.5 illustrate the basic idea of load sharing system for the parallel compressors. The main controller distributes load between two machines by calculating velocity setpoint for each of them through some characterizer function $f(N)$. Such function is used in computation to represent difference between two devices. The output from capacity controllers goes straight to the gear of compressors. Whole control structure must be supplied with necessary measurements like suction and discharge pressures, flow rates in different loops, compressor speed, etc. Antisurge controls are not shown in figure but implies its existence.

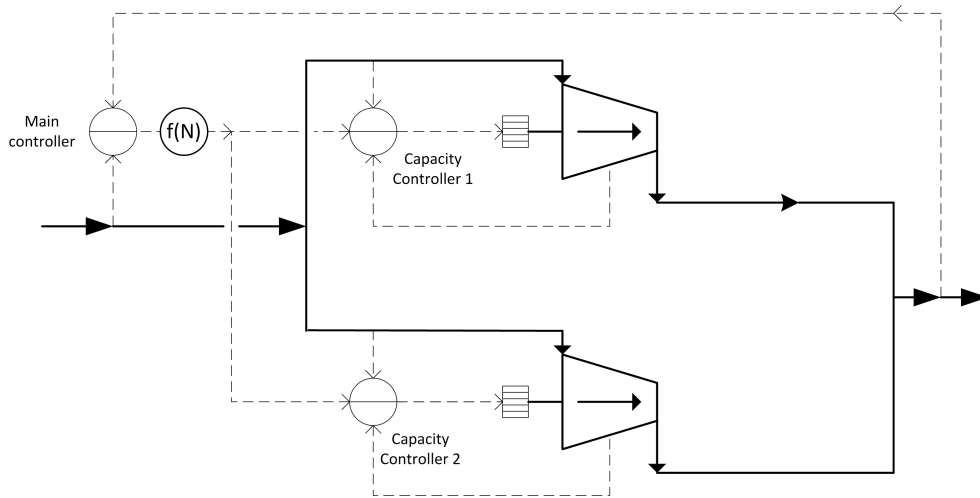


Figure 2.5: Load Balancing Control

2.3 Model Predictive Control

Model Predictive Control is an advanced method for process control. The scope of the practical applications of MPC is expanding and cover the variety of the industries such as petrochemical, light industry, aerospace and energy systems. Idea of this control strategy is to use a model of the process to predict future response of the system and to re-optimize the objective function regard to the future control inputs. The main advantages of MPC is its ability to manage multivariable control problems and to optimize processes in a real time with respect to constraints in both manipulated and controlled variables. It is quite significant, since working close to the active constraints leads often to the most profitable operation. In addition, the ability to deal with possible time delays, changing control objectives, equipment failures are reasons for success of the MPC in industry.

2.3.1 Principles of MPC

MPC algorithm is based on utilizing the model of the process to predict future trajectories of the system based on the past knowledge. Dependent on the available information of the process, the model can be given as a state space equations and transfer function or it can be obtain from the impulse and step responses. The resulting model is used by the optimization technique to decide the optimal input sequence needed in order to achieve desired setpoint trajectory.

Figure 2.6 illustrates the idea of the Model Predictive Control and the essence of the algorithm can be presented in 4 steps:

1. Calculate future output of the system from the process model with output measurements served as the initial conditions at time k .
2. Minimize the objective function to find input sequence which will drive the output to the reference trajectory on a given prediction horizon p . The optimization must consider entire range of the constraints imposed on both manipulated and controlled variables.
3. Apply the first sample of the input sequence to the process. Move prediction horizon one sample to the right.
4. Repeat steps 1-3 at time $k+1$.

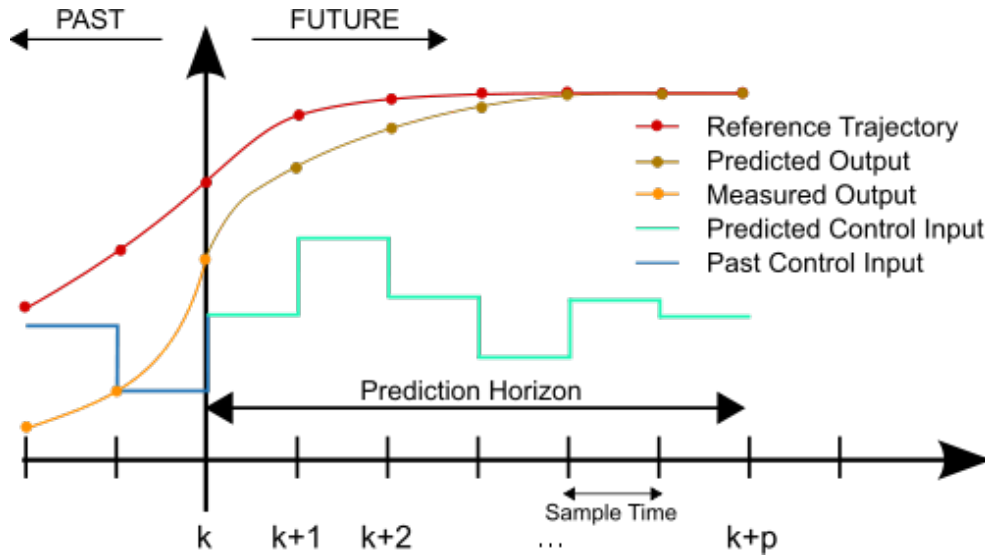


Figure 2.6: Basic scheme of MPC(wikipedia.org)

Such sliding concept where the prediction horizon moves towards in time if often called receding horizon strategy.

2.3.2 Basic formulation and prediction

QP problem presented below in 2.1-2.4 are solved each time step k to find optimal inputs. The cost function J penalizes deviations of the measured controlled variable values from setpoint trajectory, control input values from ideal values and changes in the input. It is also defines constraints for max/min control input, its rate of change and max/min limits for the controlled variables. The prediction model 2.5 calculates future output of the process.

$$\min J(x, u) = x_{dev}^T Q x_{dev} + u_{dev}^T R u_{dev} + \Delta u^T P \Delta u \quad (2.1)$$

s.t.

$$u_{min} \leq u \leq u_{max} \quad (2.2)$$

$$\Delta u_{min} \leq \Delta u \leq \Delta u_{max} \quad (2.3)$$

$$x_{min} \leq x \leq x_{max} \quad (2.4)$$

$$x_{k+1} = Ax_k + Bu_k + d_k \quad (2.5)$$

2.4 SEPTIC

Statoil R&D began developing SEPTiC (Statoil Estimation and Prediction Tool for Identification and Control) in 1996, with the purpose to create their own software for model based predictive control. The first application came in 1997 and since then has been implemented on several process plants. Including Kårstø gas processing plant to control of distillation columns. It is programmed in C++ and use text file for configuration. Most of SEPTIC implementations use experimental SISO step response models as it gives sufficient approximation to the process dynamics. Non-linear models can also be used.

2.4.1 Application

Implementation process starts with identifying process variables and related characteristics, such as:

- CVs with their setpoints, high and low limits;
- MVs with ideal values, rate of change, high and low limits;
- DVs ¹ .

Then, all this specifications must be transfer to configuration text file. Next step is to find experimental models by step-testing different variables. SISO identification algorithm is programmed into SEPTIC and obtained models are stored on hard drive. Often there is not enough MVs to control all CVs. Such issue is solved by introducing the priority order shown below. Low priority variables are respected as long as they are not conflicting with higher order variables.

Priority order with highest on top:

1. MV change (MaxUp/MaxDown),
2. MV high and low limits (High/Low)
3. Hard CV constraints (HighLimit/LowLimit = 0)
4. CV's high and low limits, CV setpoints, ideal MV values with a assigned number from 1 to 99 with 1 as a highest priority

¹CV - controlled variable, MV - manipulated variable, DV - disturbance variable

Before solving the dynamic optimization problem above, a steady-state optimization problem is solved to remove conflicting specified setpoints and limits achieving a feasible solution by adjusting the setpoints and/or limits with the lowest priority. The steady-state optimization tries to minimize the deviation from specified setpoint and/or violation of specified limits with low priority while fulfilling the process model and setpoints and/or limits with higher priority. The adjusted setpoints and/or limits are used in the dynamic optimization. E.g: If specifying setpoint (priority 2) equal 1.5 and high limit (priority 1) equal 1 for a given controlled variable, both the specified setpoint and high limit cannot be achieved. The steady-state optimization finds that the setpoint (lowest priority) must be reduced to 1 to fulfill the high limit (highest priority). Adjusted setpoint equal 1 and high limit equal 1 is used in the dynamic optimization.

Another useful concept implemented in SEPTIC is variable sub-grouping. It gives possibility to put related variables (CV, MV, DV) into the same group with one or several critical members. When one of CVs or MVs is disabled or failing, the whole group can be easily disabled also. It is practical with regard to parallel operating trains and total eight groups can be defined.

2.4.2 Tuning

Available set of the most relevant tuning parameters for the SEPTIC MPC:

- *Span*: internal scaling reflecting the 'acceptable' deviation of each variable. Rule of thumb: $Span = \text{'acceptable variation for the variable'}$
- *Fulf*: deviation penalty for the CV or for the ideal value of MV

$$w = \left(\frac{Fulf \cdot (pred - settpoint)}{Span} \right)^2$$

- *MovePnlty*: penalty on Mvr change from sample to sample

$$w = \left(\frac{MovePnlty \cdot \Delta u}{Span} \right)^2$$

- *HighOn/LowOn*: Upper and lower constraints for Cv's and Mv's
- *HighLimit/LowLimit*: defines how far the solution can be from the High/Low constraints, e.g. $HighLimit = \text{small/large}$ gives respectively a hard/soft upper constraint
- *HighPnlty/LowPnlty*: penalty on violation of the High/Low constraints
- *MaxUp/MaxDown*: maximum increase/decrease in Mv per sample
- *IvOn*: ideal value for Mv
- *IvRoc*: rate of change towards ideal value

2.5 Dynamic simulation: D-Spice

D-SPICE is a Dynamic Simulator for Process Instrumentation and Control Engineering programmed in C++. It uses numerical methods and algorithms to solve mass, energy and balance equations. It consists of several separate programs that communicate with each other through data files and a TCP/IP-based protocol. Large amount of common oil and gas operation units are included such as compressors, pumps, heat exchangers, distillation columns, pipes, valves and etc.

2.5.1 OPC server

OPC is a set of widely used and accepted specifications that provides universal methods of data exchange in the industrial instrumentation and control systems. OPC originally stands for OLE for Process Control. It is developed in collaboration of several major worldwide automation producers and Microsoft. OPC server is a program which translates data from a specific program format to common OPC format. Such that large amount of systems can be connected and exchange data between each other(D-spice and Matlab). Before existence of OPC, producers of instrumentation equipment had to create different interfaces towards various automation systems. By using this common specification only the single interface must be created to 'communicate with the rest of the world'. The OPC Foundation is organization that coordinate the development and support of the this technology.

2.6 Summary

The presented theory was meant to give a reader necessary insight into compressor control theme. Theory about compressors was collected from (Nisenfeld 1982) and (Gravdahl & Egeland 1999), about MPC from (Maciejowski 2002) and (Bao-Cang 2010), about Septic from (Strand & Sagli 2004) and about D-spice from (Fantoft 2004).

Chapter 3

Matlab

In the beginning of the project a creation of an MPC application in Matlab was considered. The idea was to use quadratic programming package(quadprog) in the implementation to optimize operation and control system through the OPC server. Communication between Matlab and D-spice was successfully established. It used a loop sequence for detecting server outputs. Issue with that was to be able to write values from user interface into Matlab while it was running. Thats how an idea of using a GUI(graphical user interface) showed up. Matlab contains a GUI package and figure 3.1 illustrates an application

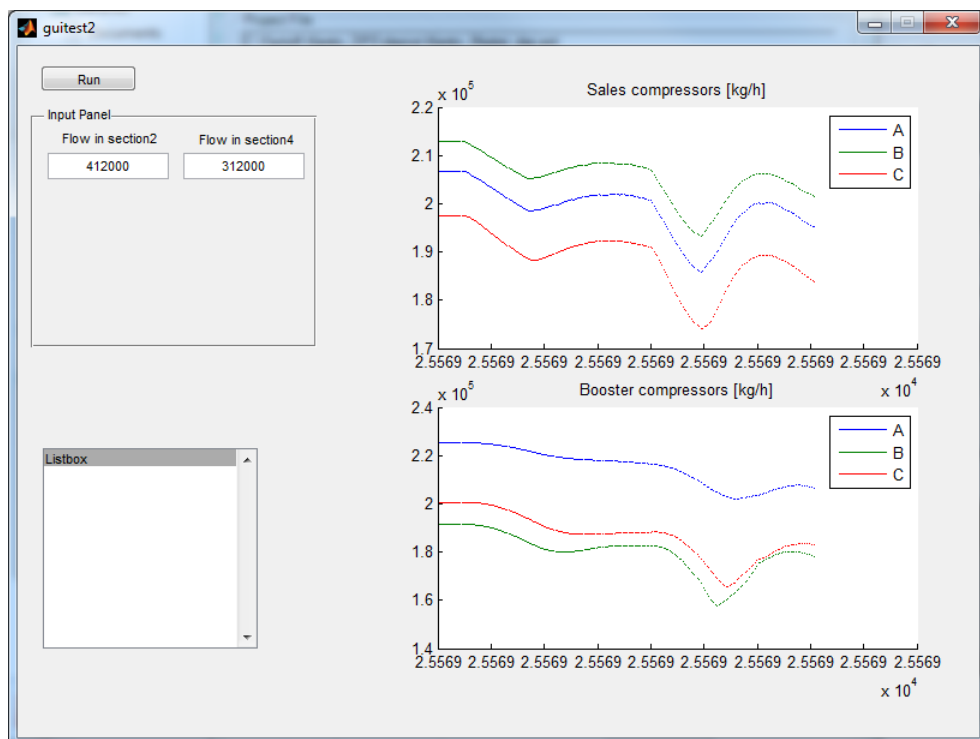


Figure 3.1: Matlab GUI interface for MPC

which was created to implement communication with a graphical interface. This program establishes connection to OPC server and is able to read and write defined data into it. In the input panel, section flowrates to the system can be set(.setp) while graphs shows measured throughput(.flow) for all six compressors. Strange values on the x-axis is a time format used in the OPC server. Decimal format in Matlab must be changed to 'Long' to be able to register changes. To start and stop application, 'Run' button must be pressed. With time this application could increase to handle additional number of variables and to have an optimization techniques in the bottom. But development of it was left in favor of Septic MPC. Automatic generated code and essential code for connection can be found in appendix D

Chapter 4

Application to Kårstø

This chapter will give insight into today's control structure of Statpipe compression system at Kårstø gas processing plant, suggests alternative control structure and present part of its implementation. It will reveal configuration procedure and practical aspects of working with D-spice and Septic MPC, present made choices and summarized changes applied to the plant model.

4.1 Plant

Dynamic model of the Kårstø processing plant implemented in D-Spice process simulator was used in this project. The overall model is build up of different sections and shown in appendix A.2. Only sections 2, 4 and 7 were activated for work with this thesis. Section 7 consists of the actual compression system where all 6 Statpipe compressors are implemented. Sections 2 and 4 are the inlet parts of the two separate Statpipe extraction trains. This two parts presented as a distillation column in figure 1.2 where methane is separated from natural gas and goes further to the sales gas compressors(STPSGC). After the first stage compression, gas goes to the intercooling. The heat exchangers is placed in sections 2 and 4 and exploits the low temperature of the methane column and seawater to reduce the gas temperature on the intermediate stage. That is a main reason why all three sections must be enabled and run simultaneously, otherwise the interactions between first and second compression stages would be missing.

As mentioned previously, the compression occurs in two stages where each of stage consists of a three parallel compressors. Figure in appendix A.3 shows sales gas compressor A with associated gas turbine, recycle gas heat exchanger, controller structure and required instrumentation. Equivalent equipment can be found around 5 other machines with only difference that Booster compressors(second stage) are driven by the electrical motors instead of turbines. Control structure is build up with a CCC controllers and logic modules. It consists of two controllers per compressor for capacity and antisurge control. In addition, there is a main performance controller on each stage for coordination of actions and suction pressure control.

Antisurge controller is activated when the flow throughput decreased to its minimum. It then starts to recycle required gas amount to keep operational point on the surge control line. Load sharing between stages is done by holding suction pressures at constant setpoints by manipulating machines speed. Load sharing between parallel compressors on the same stage is done by manipulating the engine speed to equalize the deviation of the operating point from the surge control line for the high flow rates. For the low flowrates, operation point is kept on the control line with equal antisurge valve opening. Figure

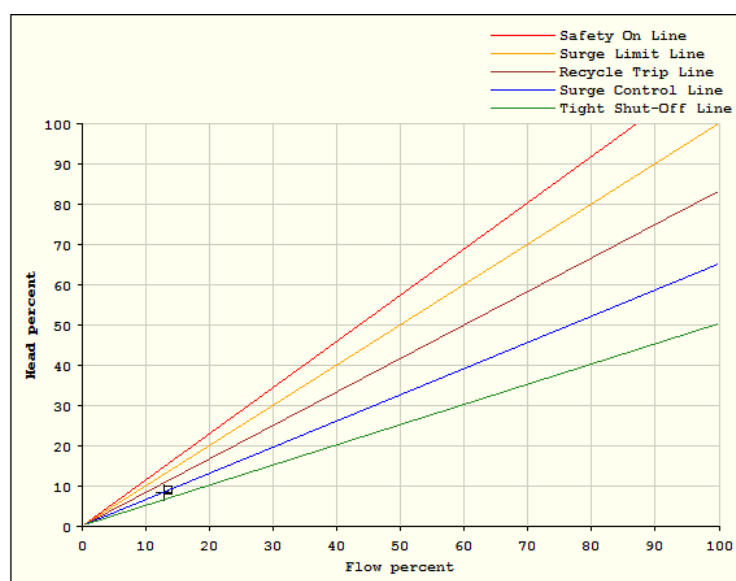


Figure 4.1: Compressor Control Lines(D-Spice)

4.1 illustrates important concept for the antisurge control implemented in CCC structure. When operation point is to the left for the Surge Control Line the recycle valve opening increases, when to the right - decreases. When operation point is to the right for Tight Shut-Off Line the recycle valve is fully closed. In order to maximize compressor efficiency, the control margin between the surge and the control line must be minimized. Compressor is deactivated if operation point cross the recycle trip line.

4.2 Plantwide control design

In the initial stage of the project many questions regarding control structure was raised. Such as: which variables should be controlled, how many manipulated variables, what are the system constraints and which of them are active, what can disturb the plant and etc. To answer these questions and develop control structure with a systematic approach a '*Plantwide Control Design Procedure*' (Skogestad 2004) was applied. Another related publication (Downs & Skogestad 2011) provided a good examples about application of the

procedure. Here the plantwide control design procedure is applied to compression system at Kårstø gas processing plant.

I. Top-down part

Step 1: Define operational objectives(economics) and constraints.

Since the export compressors are not limiting the plant throughput rate, the operational objective of the compression system is to minimize the compression costs.

Cost function:

$$Min Cost = J = P_{FG} * W_{1.stage} + P_{El} * W_{2.stage} \quad (4.1)$$

where: P_{FG} - fuel gas price per kWh produced, P_{El} - price for kWh of electrical power, $W_{1.stage}$ - energy consume on first stage, $W_{2.stage}$ - energy consume on second stage.

Operational constraints:

- maximum / minimum compressor speed.
 - Sales gas compressors(High/low): 5200/3200 RPM
 - Booster gas compressors(High/low): 12200/8166 RPM
- Distance to control line for all compressors must be positive, operation to the right of or at the control: $DEV > 0$
- Maximum/minimum suction pressure for the booster compressors: 126/100 barg
- Closed anti-surge valves(ASV) for all compressors ($ASV > 0$)
- Given sales gas compressor suction pressure(typically 41 barg).

Step 2: Identify degrees of freedom and optimize operation for important disturbances.

As described in section 2.2.5 there are two manipulated variables associated with each compressor. Compressor speed and recycle valve opening. There are a total of 6 compressors which gives 12 degrees of freedom(DOF) for the plant. At high rates all compressors are optimally operating with closed anti-surge valves ($ASV=0$), and at low rates all compressors are optimally operating at surge control line ($DEV=0$). 6 DOF is used. We expect that it is optimal that equal compressors in parallel should be operated equally. Equal operation of the sales gas compressors uses 2 DOF and equal operation of the booster compressor uses 2 DOF. Another DOF is used for sales gas compressors suction pressure as it affects the liquid recovery in the demethanizer (upstream for the export compressors) and must be kept at the constant value. There is only one degree of freedom left. The setpoint for pressure between stages(booster suction pressure setpoint) will be used as a manipulated variable for closer optimization.

The main disturbances for the compression system are:

- Input gas flowrate
- Export pipeline pressure
- Fuel gas prices and energy prices

Input gas flowrate is given by the production rate from the fields offshore and can vary with time. The discharge pressure is decided by the amount of sales gas delivered into the transport pipe and amount of taken gas in Europe. The prices vary with marked conditions.

Table 4.1 is meant to illustrate the challenges associated with control of compression system. It presents the energy consumption for first stage, second stage and total required power for all compressors in different operational conditions. First column contains flowrate percentage value, second is export pipeline pressure. Both values are disturbances as stated above. Third column is a booster suction pressure. Required work and load distribution between stages will vary with its value. Next three columns presents the required energy for first stage, second stage and total energy consumption. Last column contains active constraints.

Table 4.1: Steady state potential in different conditions

Rate [%]	Output Pressure [bara]	Booster Pressure [bara]	Sales gas Work[kW]	Booster Work[kW]	Total Work[kW]	Active Constraints
100	165	105	29006	13966	42972	
		110	30686	12344	43030	
		115	32404	10822	43226	
		120	34015	9432	43447	
		125	35475	8146	43621	
100	175	105	30795	14727	45522	a,c
		110	30798	14726	45524	a,c
		115	32377	13145	45522	
		120	33996	11631	45627	
		125	35480	10276	45756	
100	185	105	33086	15088	48174	a,c
		110	33092	15088	48180	a,c
		115	33117	15083	48200	a,c
		120	34006	14258	48264	a,c
		125	35470	12422	47892	
75	165	105	22978	12818	35796	c
		110	24356	10742	35098	c
		115	26168	9038	35206	b,c
		120	28605	7617	36222	b,c
		125	30780	6367	37147	b,c

- *Bold number indicate minimum compressor work (sales gas, booster or total) for given gas rates and export pipeline pressures.
- *a - indicate operation with maximum booster compressor speeds.
- *b - indicate operation of sales gas compressors at surge control line (DevSTPSGC=0)
- *c - indicate operation of booster compressors at surge control line (DevSTPBC=0).

As expected, high values for suction booster pressure decreases energy consumption on the second stage while low values - decreases energy consumption for the first stage. However, for high export pipe pressures, normal rate and low setpoint for suction booster pressure, second stage compressors are running on a max speed. Suction pressure saturates and unable to reach low setpoint. In the table it can be identified with same or very close booster energy consumption values.

Another set of active constraints is introduced in low rate case. With high setpoint for booster suction pressure, compressors on both stages operate on the surge control line with active recycling. With low setpoint for booster suction pressure, recycling is active only for compressors on second stage. But recycled flow has a quite large number about 25-30% of the methane flow rate and is probably on the boundary where one of the compressors can be turn down. It again will lead to a total different picture of power consume.

In addition, fuel gas prices and energy prices will affect the choice of variable to optimize. With high energy prices compared to gas prices, it is more profitable to minimize energy consumption on the second stage. Such case is presented below based on values given in table 4.1.

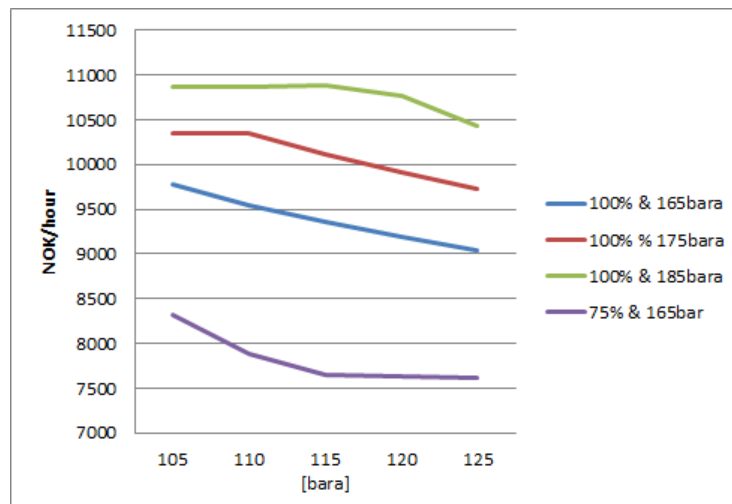


Figure 4.2: Potential for compression cost[NOK/hour] (2010 prices)

Cost estimate for potential optimization is created, figure 4.2. X-axis represent values for booster suction pressure, Y-axis - price in NOK per hour. Given curves represent various operational conditions with flow in % and export pipe pressure in bara. Consumed

power for the different operational points in table 4.1 multiplied with energy prices from 2010 ¹. Energy price to produce 1 kWh: from natural gas = 0,18 NOK/kWh, electrical power = 0,326 NOK/kWh. Figure shows that for high flow rate it is absolutely potential to decrease the cost by increasing the suction booster pressure. For the 75% flow rate, economical potential is quite small after reaching 115 bara. But at the same time, if recycling is large enough to turn down one of the compressors the curve will be totally different.

To summarize, active constraints will strongly depend on operational conditions and main disturbances such as feed rates, export pipeline pressure and energy/gas prices.

Step 3: Controlled variables.

Normally active constraints must be controlled. But which constraints are active will vary with different operational conditions. Here it is assumed high electricity prices such that operational cost for the gas driven compressors on first stage are lower than cost for electricity driven second stage compressors.

- **Condition 1:** High rate and low pressure increase

Sales gas compressors should run on a max speed. Suction pressure for first stage must be hold constant as it will affect the methane extraction.

Active constraints:

- max speed for the first stage compressors
- all recycle valves closed
- suction pressure for STPSGC

- **Condition 2:** High rate and high pressure increase

Booster compressors operate on the control line, recycling needed to avoid surge and if possible should be minimized.

Active constraints:

- max speed for the first stage compressors
- booster compressors operate on the control line
- suction pressure for STPSGC
- close antisurge valves for STPBC if possible

- **Condition 3:** Low rate and low pressure increase

Sales gas compressors operate on the control line. Recycling for booster compressors should be minimized.

Active constraints:

- sales gas compressors operate on the control line

¹<http://www.ssb.no/emner/10/07/indenergi/tab-2011-06-27-01.html>

- booster compressors operate on the control line
- suction pressure for STPSGC
- close antisurge valves for STPBC if possible

In addition, load sharing on the same stage is done by holding equal distance to control line for parallel compressors.

Step 4: Throughput manipulator

The throughput for the compression system is given upstreams and the main operation mode is to maximize energy efficiency.

II. Bottom-up part

Step 5: Regulatory control layer:

The basic compressor control system stabilize the sales gas and booster compressor suction pressure and avoid drift of load between parallel compressors by manipulating the compressor speeds. The load sharing between the parallel compressors are done by trying to keep equal distance to surge control at high throughput rates (load balancing) and equal recycle valve opening at the low throughput rates (recycle balancing) with operating point at or to the right of the surge control line.

Step 6: Supervisory control layer:

As presented previously, the system is characterized by changing active constraints and strongly interacting variables. MPC application suits well for control of such systems. Set-point for the booster suction pressure will be used as MV to optimize operation. Mentioned active constraints will be implemented in MPC application.

4.3 Implementation

D-Spice dynamic simulator and Septic MPC were connected together through the OPC server. This section will summarize configuration procedure and changes applied to the original model.

Another very useful program MatrikonOPC Explorer was used in configuration procedure and sometimes for simulations. When variables are connected to OPC server they become unavailable for direct change in D-spice, as it use the server values. This software gives easy and fast access to them through the server, such that one can change single variables manually and also it suits very well for troubleshooting of communication. Author will recommend to install this program for everyone working with OPC communication. It can be downloaded from <http://www.matrikon.com/> for free.

4.3.1 D-SPICE

Some changes to the model were applied. As originally, two of Statpipe booster compressors sends gas into Europe II while last one towards Draupner. It was decided to bring all compressors to the same output manifold and close valve in Draupner direction and open valve to Europe II for booster compressor C. It was assumed that it will simplify the system and bring it to the common output pressure. There were done some minor changes to compressors and control around them in attempt to equal their operation as much as possible. Also all side streams were deactivated. All specific changes can be found in appendix C

Operational points

Dspice allows to save various operational points so that repetitive simulations can start from the same conditions. Two files, one with system parameters(.prm) and another one with current values(.val) must be saved. Procedure for doing this can be found in appendix F.3. Three such points were created for the current project:

- Basecase/original rate - will be used later for comparing. It will be related to as a 100% rate even though that the max flow throughput can be increased with 10-15% at the feed inlet. Base case saved as master.prm/val
- Mediumrate - 75% of the basecase, saved as master_mediumrate.prm/val
- Lowrate - 50% of the basecase, saved as master_lowrate.prm/val

All three have 118 bara on the intermediate stage and 167,4 bara as an export pipeline pressure. In addition several operation points was saved for verifying system behavior. They can be found in same directory and have the following structure:

master_x_75rate_165utlop_105booster.prm/val, where '75rate' stands for - 75% flowrate, '165utlop' - 165 bara system output pressure, '105booster' - 105 bara suction pressure for booster compressors.

Change of flowrate was done in controllers 20_FC1001 and 20_FC2001 in section 2 and section 4, respectively. For a fast access to them a workspace with panels for this two controllers can be obtain from: File-> Restore Workspace-> flowrate_change_workspace.wsp. Output pressure changes can be done in section 7_1 in block: To_EPII.

OPC Server configuration

Configuration of OPC Server takes place in Excel file, see appendix E as an example with a part of tags used in this project. All tags for required measurements/values from/to the plant must be specified there. Then configuration file is transformed into OPC server settings(.xml) and DCS format file for Dspice with OPCconfig application.

4.3.2 SEPTIC MPC

Previously described plantwide control structure in section 4.2 covers both the optimization task and capacity control task. Septic MPC was implemented as a supervisory controller on top of today's structure with purpose to optimize operation. Implementation will be covered below.

Almost entire configuration of Septic took place in a text based file. CVs, MV and DVs were defined there. Each CV variable has a unique name, high/low boundaries with belonging priority and also setpoint value with its priority. It is similar concept for the MV but setpoint is called Ideal value(IV).

Table 4.2: Control configuration

Type	Variable Name	High Prio	SetPnt Prio	Low Prio	Low Limit	Set Point	High Limit	Group	Description
Cv	22SC0102	1	-	1	3200	-	5150	1*	Speed for Sales gas compressors
Cv	22SC0202	1	-	1	3200	-	5150	2*	
Cv	22SC0302	1	-	1	3200	-	5150	3*	
Cv	22SI0406	1	-	1	8166	-	12200	4*	Speed for Booster gas compressors
Cv	22SI0407	1	-	1	8166	-	12200	5*	
Cv	22SI0408	1	-	1	8166	-	12200	6*	
Cv	DevSTPSGCA	-	2	2	0	0	-	1	Deviation from control line Sales A,B,C
Cv	DevSTPSGCB	-	2	2	0	0	-	2	
Cv	DevSTPSGCC	-	2	2	0	0	-	3	
Cv	DevSTPBCA	-	2	2	0	0	-	4	Deviation from control line Booster A,B,C
Cv	DevSTPBCB	-	2	2	0	0	-	5	
Cv	DevSTPBCC	-	2	2	0	0	-	6	
Cv	ASVSTPSGCA	2	3	2	0	0	100	1	Antisurgevalve opening for Sales A,B,C
Cv	ASVSTPSGCB	2	3	2	0	0	100	2	
Cv	ASVSTPSGCC	2	3	2	0	0	100	3	
Cv	ASVSTPBCA	2	3	2	0	0	100	4	Antisurgevalve opening for Booster A,B,C
Cv	ASVSTPBCB	2	3	2	0	0	100	5	
Cv	ASVSTPBCC	2	3	2	0	0	100	6	
Cv	TotalWorkSales	-	4	-	-	-	-	All	Required Energy
Cv	TotalWorkBoosters	-	4	-	-	-	-	All	Total cost
Cv	TotalWorkCost	-	4	-	-	-	-	All	
			IvPrio			IV		Group	
Mv	22XIC0518SP		3		101	118	126	All	Setpoint for Booster pressure
Dv	EPIIP							All	Output pressure
Dv	Gasfode							All	Sum of input feeds

Table 4.2 contains all CVs, MV and DVs which is used in implementation together with essential parameters. In addition several Tvr variables were defined in configuration which were used for the intermediate calculations, e.g. sum of compressor power on the same stage. Not covered here. In the table the first column is a type of variable, then its name in application. Third column contains priority of the high(upper) constraint, fourth - priority of setpoint and fifth - bottom constraint priority. In the Low limit column, the values for the bottom constraint stated, next column contains setpoint values for variables and next - the upper constraint value. In the group column, the group membership for each variable is stated, will be covered more deeply later. In the last column a short description for variables is given.

Figure 4.3 shows the main window of Septic MPC with final application for the optimization. Setpoint for booster suction pressure is used as a manipulated variable with its low/high constraints and 'ideal' value to optimize operation. In addition to the operational constraints, implementation constraints on rate of control input change, Δu was also included here. High and low boundaries for speed controllers is implemented without any desired setpoint such that MPC is just aware of constraints. Deviation from control line and recycle valve opening are implemented with both low/high boundaries and desired setpoints in attempt to minimize their values. Relevant disturbances, such as input flowrates and pressures are shown below CVs. Optimization CV variables was implemented with their setpoints: total fuel gas consume by the first stage in Sm³/h, energy consume by each stage and its total cost. In order to start optimization of operation with respect to them, the specific variable must be activated and a low setpoint value must be set for them. In addition, their setpoint priority must be increased. As an example of Septic configuration, parameters for one of the variables is shown in appendix G.

The screenshot displays the 'STATPIPE GAS COMPRESSORS' section of the software. It contains two main tables: one for manipulated and regulated variables, and another for energy and cost variables.

Manipulerte	Desc	Unit	DS	FS	Value	LL	IV	HL	CTRL	Groups
22XIC0518SP	Main CCC, 2.stage	[bara]	Act/	ACT	118.00	101.00	118.00	126.00	COMP	1* 2* 3* 4* 5* 6*
Regulerte	Desc	Unit	DS	FS	Value	LL	SP	HL		Groups
22SC0102	Turtall Sales A	[RPM]	Act/	ACT	4887.3	3200.0		5150.0		1*
22SC0202	Turtall Sales B	[RPM]	Act/	ACT	4914.1	3200.0		5150.0		2*
22SC0302	Turtall Sales C	[RPM]	Act/	ACT	4855.0	3200.0		5150.0		3*
22SI0406	Turtall Booster A	[RPM]	Act/	ACT	10586	8166		12200		4*
22SI0407	Turtall Booster B	[RPM]	Act/	ACT	10530	8166		12200		5*
22SI0408	Turtall Booster C	[RPM]	Act/	ACT	10541	8166		12200		6*
DevSTPSGCA	Avstand til kontrollinje, Salgsgass Kompressor A		Act/	ACT	0.22238	0.00000	0.00000			1
DevSTPSGCB	Avstand til kontrollinje, Salgsgass Kompressor B		Act/	ACT	0.21492	0.00000	0.00000			2
DevSTPSGCC	Avstand til kontrollinje, Salgsgass Kompressor C		Act/	ACT	0.23047	0.00000	0.00000			3
DevSTPBCA	Avstand til kontrollinje, Booster Kompressor A		Act/	ACT	0.17968	0.00000	0.00000			4
DevSTPBCB	Avstand til kontrollinje, Booster Kompressor B		Act/	ACT	0.17966	0.00000	0.00000			5
DevSTPBCC	Avstand til kontrollinje, Booster Kompressor C		Act/	ACT	0.17966	0.00000	0.00000			6
ASVSTPSGCA	Antisurge ventilåpning, Salgsgass Kompressor A	[%]	Act/	ACT	0.00	0.00	0.00	100.00		1
ASVSTPSGCB	Antisurge ventilåpning, Salgsgass Kompressor B	[%]	Act/	ACT	0.00	0.00	0.00	100.00		2
ASVSTPSGCC	Antisurge ventilåpning, Salgsgass Kompressor C	[%]	Act/	ACT	0.00	0.00	0.00	100.00		3
ASVSTPBCA	Antisurge ventilåpning, Booster Kompressor A	[%]	Act/	ACT	0.00	0.00	0.00	100.00		4
ASVSTPBCB	Antisurge ventilåpning, Booster Kompressor B	[%]	Act/	ACT	0.00	0.00	0.00	100.00		5
ASVSTPBCC	Antisurge ventilåpning, Booster Kompressor C	[%]	Act/	ACT	0.00	0.00	0.00	100.00		6
Forstyrrelser	Desc	Unit	DS	FS	Value					Groups
20FC1001	Section 2, FØDE	[kg/h]	Trk/	TRK	412000.0					1 2 3 4 5 6
20FC2001	Section 4, FØDE	[kg/h]	Trk/	TRK	412000.0					1 2 3 4 5 6
22XIC0004SP	Sugetrykket, 1. trinnet	[bara]	Trk/	TRK	40.50					1 2 3 4 5 6
EPIIP	Utløpstrykket mot Europipe2	[bara]	Trk/	TRK	167.4					1 2 3 4 5 6
ENERGIFORBRUK og KOSTNAD										
Regulerte	Desc	Unit	DS	FS	Value	LL	SP	HL		Groups
TotalFuelSalesSm3	Samlet tilført drivstoff 1. trinnet	[Sm ³ /h]	Trk/	TRK	6152					1 2 3
TotalWorkSales	Samlet forbruk for 1. trinnet	[kW]	Trk/	TRK	33176		35000			1 2 3 4 5 6
TotalWorkBooster	Samlet forbruk for 2. trinnet	[kW]	Trk/	TRK	10456		6000			1 2 3 4 5 6
TotalWorkCost	Total kostnad for tilført energi	[kr/h]	Trk/	TRK	9380		10500			1 2 3 4 5 6

Figure 4.3: Septic Main Application

TotalWorkCost variable is calculated in the following way:

$$0,18 * TotalWorkSales + 0,326 * TotalWorkBooster \quad (4.2)$$

Where: TotalWorkSales and TotalWorkBooster is sum of the energy [kW] required to drive compressors on first and second stage, respectively. While 0,18 and 0,326 are the prices [NOK/kWh] for produced kilowatt of energy from gas and electrical power, as presented earlier.

Consumed fuel gas by the first stage compressors was also implemented but not used in a final optimization. Its value was transformed to standard cubic meters per hour as it is a common standard in the oil and gas industry for measurements of products and gas prices usually provided in NOK/Sm³. Value for the TotalFuelSalesSm³ [Sm³/h] variable is calculated in the following way:

$$\frac{\sum Turbine\ flows\ A,\ B,\ C[mol/s]}{42,292[mol/Sm^3]} * 3600[s/h] \quad (4.3)$$

Where: 1 Sm³ equal to 42,292 mol (Skogestad 2009, p. 11).

Another important aspect in practical implementation of MPC is prioritizing of variables. General priority rule can be summarized as follows with highest priority on top:

1. MV high and low limits(speed max/min)
2. CV hard constraints(antisurge valve opening, setpoint and low limit for distance from control valve, deviation from suction pressure setpoint on first stage)
3. CV setpoints(setpoint for antisurge valve)
4. Ideal setpoints(energy consumption)

More specifically, used variables for the optimization task with their type, name, low/high boundary priority, setpoint priority, group membership and short description summarized in the table 4.2. As it shown all variables is divided into six group, by the number of compressors. With a speed controllers as a critical variable(star). Now control structure for a single compressor can be turn down by just setting associated speed controller in tracking mode.

A panel with all essential variables for control was created and can be found in menu under Display group -> Basis. Models from MVs and DVs can be found in DisplayGroup -> Model_Gasfode and-> Model_EPIIP.

Modeling

Many of currently running applications at Statoil are built with experimental SISO step response models. They are easy to build, understand and maintain. Several SISO identification techniques is available in Septic. A single variable stepping was done to identify the dynamic influence from manipulated variables and disturbances on controlled variables and to acquire models for the simulated process. Figure 4.4 shows modeling panel with CV

response of speed controller for sales gas compressor A from step in the booster suction pressure. Similar modeling procedure was applied to all CVs. Each model then has a model file which is saved in the 'Models' folder.

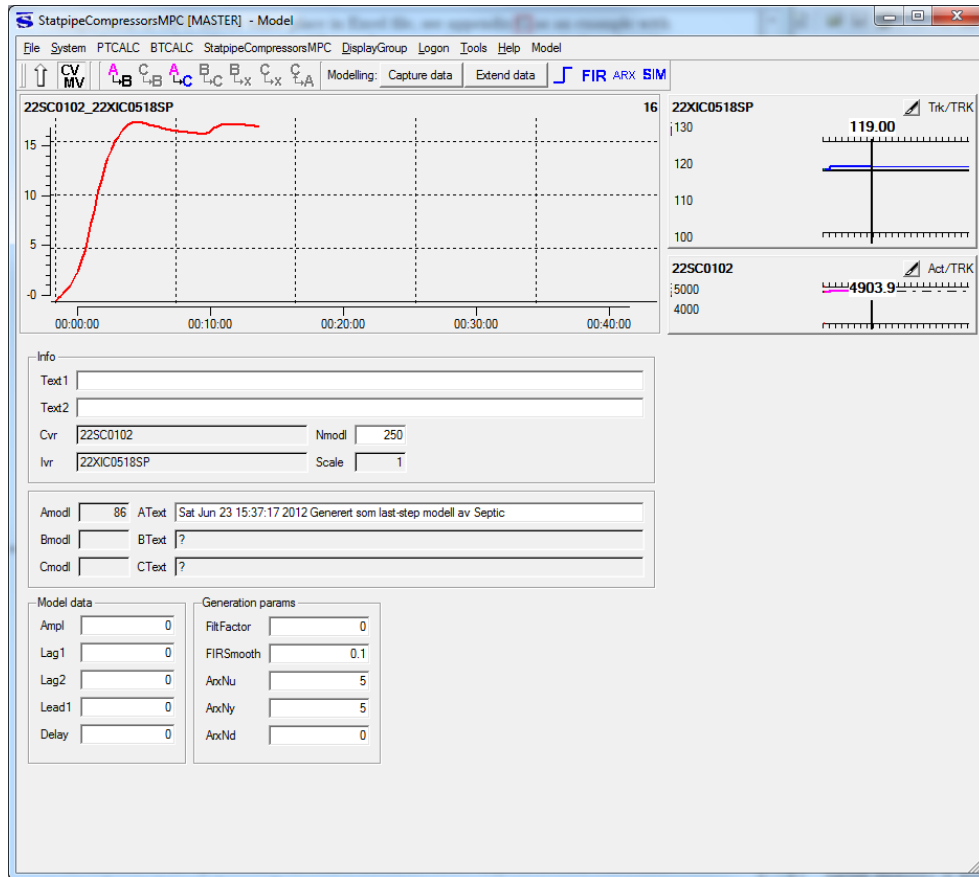


Figure 4.4: Modeling

Chapter 5

Simulation Results

Chapter will present simulation results after implementation of Septic MPC with purpose to optimize operation of Statpipe compression system. Models for the process behavior on different operational points were acquired and will be present first. Next, simulation results of different operational scenarios will be shown. Previously described process model in D-spice was used.

5.1 Models

Models of the system behavior were acquired for full rate, medium rate(75%) and low rate(50%). Steps were made in booster suction pressure, input flowrate and output pressure as it is a MV and two main disturbances. So total 3 model sets for each flowrate.

In a full rate conditions all compressors operate sufficiently far to the right from the surge control line such that all antisurge controllers are deactivated. This situation made modeling procedure quite straight forward without any extensive challenges. Model response on step in setpoint for booster suction pressure showed that speed for the first stage compressors will be adjusted upwards with decreased distance to control line while speed for the second stage machines will go downwards with increased distance to control line. Respectively the response for energy consume will increase for sales gas compressors and will decrease for booster. Response on stepping in the feed flow rate showed that increase in load will be distributed between all compressors quite similar and handled by speeding them up. Response on stepping in output pressure will be handled by second stage machines by increasing their speed. Response models for full rate case can be found in appendix B. But probably case with 75% flowrate gives best insight in system behavior and presented next.

In this case all compressors operate on the control line with open recycle valves and active antisurge control. Figure 5.1 shows a panel with response models from stepping in setpoint for the booster suction pressure on medium rate. Second, third and fourth columns contains six fields where each cell corresponds to specific compressor, from A compressor on first stage to C compressor on second stage. Second column is compressor speed, third

is distance from surge control line(DEV), fourth is antisurge valve opening(ASV). In the first column, on top is difference between setpoint and measurement for the sales gas compressors suction pressure. Next two are the energy consume for the first and second stage, respectively. Fourth cell is a response for the sum of operational cost for both stages.

To increase intermediate pressure, first stage compressors should do more work and second - less. That is handled by adjusting speed, similar as in the full rate case. But in addition, recycling is a new active variable. First stage ASV opens more to increase flow throughput. It then allows to move compressor operational point higher in the performance

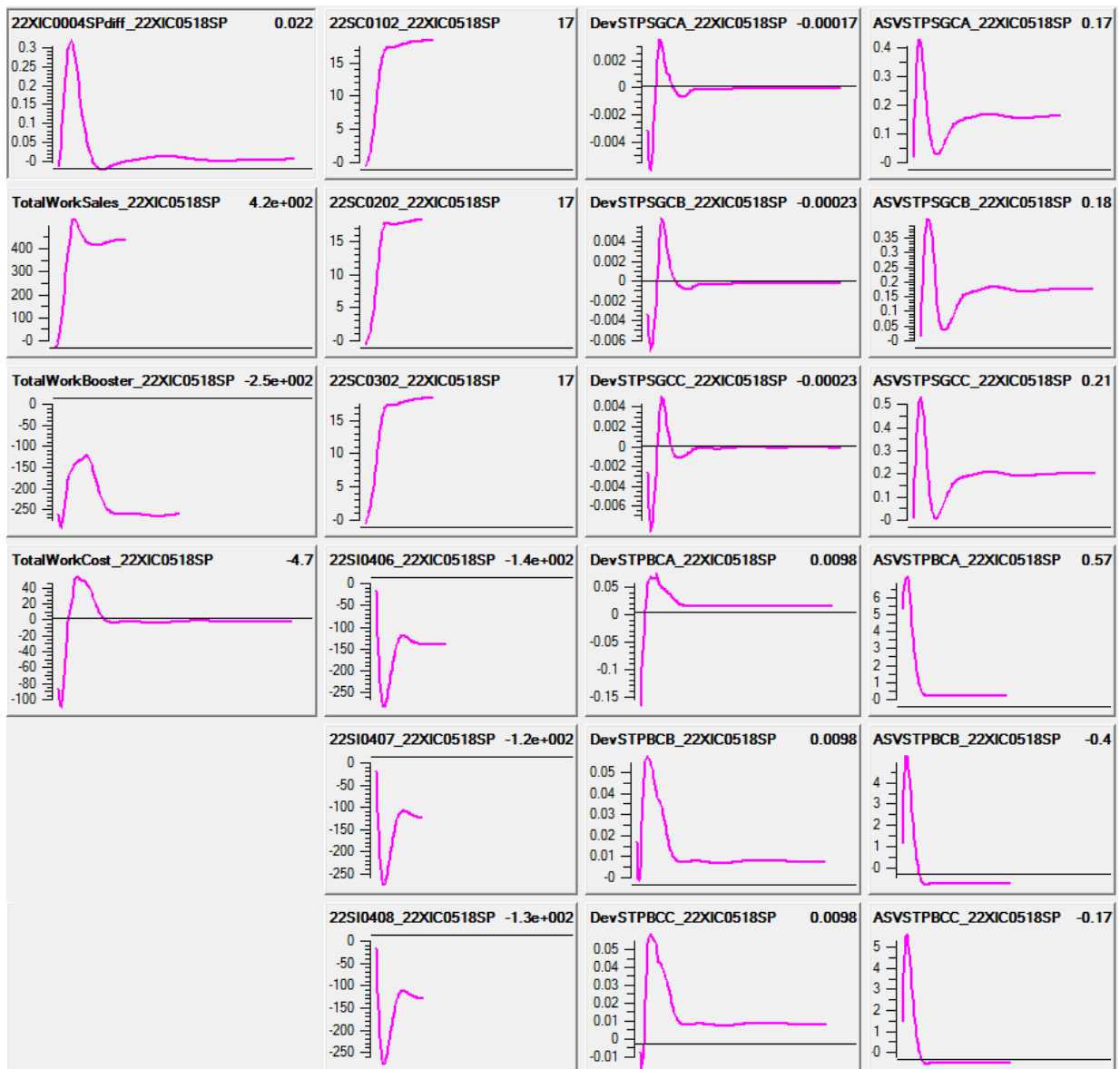


Figure 5.1: System behavior on step in booster suction pressure

map and be able to deliver higher intermediate pressure. The opposite takes place on second stage as its required differential pressure becomes less.

Response models for the low rate case were also acquired and can be found in appendix B. But they are not properly suited for simulation purposes. On rates close to 50%, something different happens then previously described. In D-spice one of the compressors take over almost whole recycle duty while two others operate on a surge control line with some small ASV opening. Practically this compressor does not contribute to the system and whole its throughput goes to recycling. It seems to be just a modeling solution in D-spice because at the real plant in time of low rates one of the compressors can be turn down manually by operator. In such case compressor control structure works as on medium rate but just with only two machines. That is a reason why later results will exclude simulations on low rates.

5.2 Scenarios

Large amount of simulations was done and five different scenarios are introduced below to present system behavior with implemented Septic MPC. First two cases was tested on a full rate, then third is on medium rate with operational point for compressors on the surge control line. Next one, illustrates optimization of distance to control line(DEV). Last one, unveils nonlinearity. Rate of change for the suction pressure MV was adjusted during experiments from 0,2 to 0,02 per sample since it had huge influence on the flows in the system. It is also more realistic value.

Following scenarios are presented below:

- Case 1: Application test. Variation of MV setpoint. Handling of step in export pipeline pressure +1 bar.
- Case 2: Startup of compressor MPC with low setpoint on the booster compressor work and full rate.
- Case 3: Startup of compressor MPC with low setpoint on the booster compressor work and low rate.
- Case 4: Start up of compressor MPC to control distance to the surge control line.
- Case 5: System behavior on significant export pipeline pressure increase.

Case 1: Application test. Variation of MV setpoint. Handling of step in export pipeline pressure +1 bar

First, the explanation to the window which shows response of several system variables will be given then case will be present. In the first row: setpoint for the suction pressure for second stage(MV), then gas feed rate(DV) and output pressure(DV). Second and third rows consists of CVs associated with first stage and second stage compressors, respectively. In the first cell - speed response, then distance to surge control line(DEV) and antisurge valve opening(ASV). Only compressors C is chosen from each stage, as a response from other machines on the same stage were similar. In the fourth row, first and second cells are energy consume for sales and booster gas compressors. Last cell contains total cost for both stages. Each window is scaled to 3 hours range. The vertical line with a value on top indicates current measured value, area to the left of it shows past variable response(2 hours), area to the right shows predicted response(1 hour). Indicator in the top right corner of each cell shows if the variable is active. Top and bottom constraints shown for those variables which have them.

Ideal value(IV) priority for booster suction pressure increased to 2. Manual step of 1 bar in MV was done to test application. On change in booster suction pressure up(1), speed for compressors on first stage was raised and reduced for second stage compressors. The opposite happened on step down(2). Step in the output pressure(3) was handled by increasing speed of second stage compressors as expected.

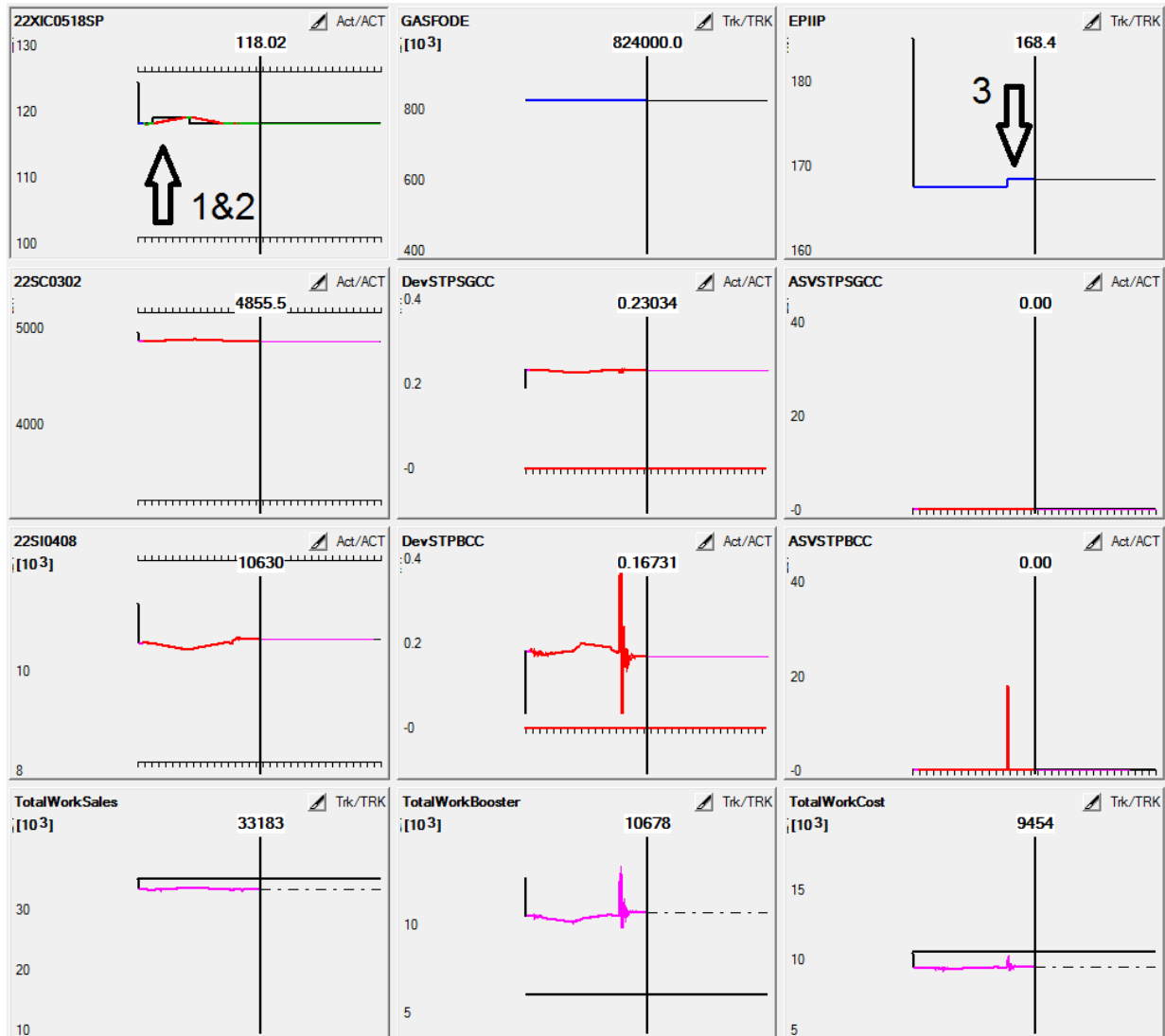


Figure 5.2: Application test

Case 2: Startup of compressor MPC with low setpoint on the booster compressor work and full rate

In this case booster suction pressure is used as a MV. Figure 5.3 shows system behavior when the goal for optimization is to minimize energy consume for the second stage. Setpoint for the 'TotalWorkBooster' is set to a low value and its priority increased to 2. Such that MPC can minimize energy consume and give up setpoints for DEV and ASV which will be handled by the lower layer control. Ideal value for MV is turned off, while other priorities kept the same as in first case.

After turning MPC on, system behave as expected. The booster pressure is driven up

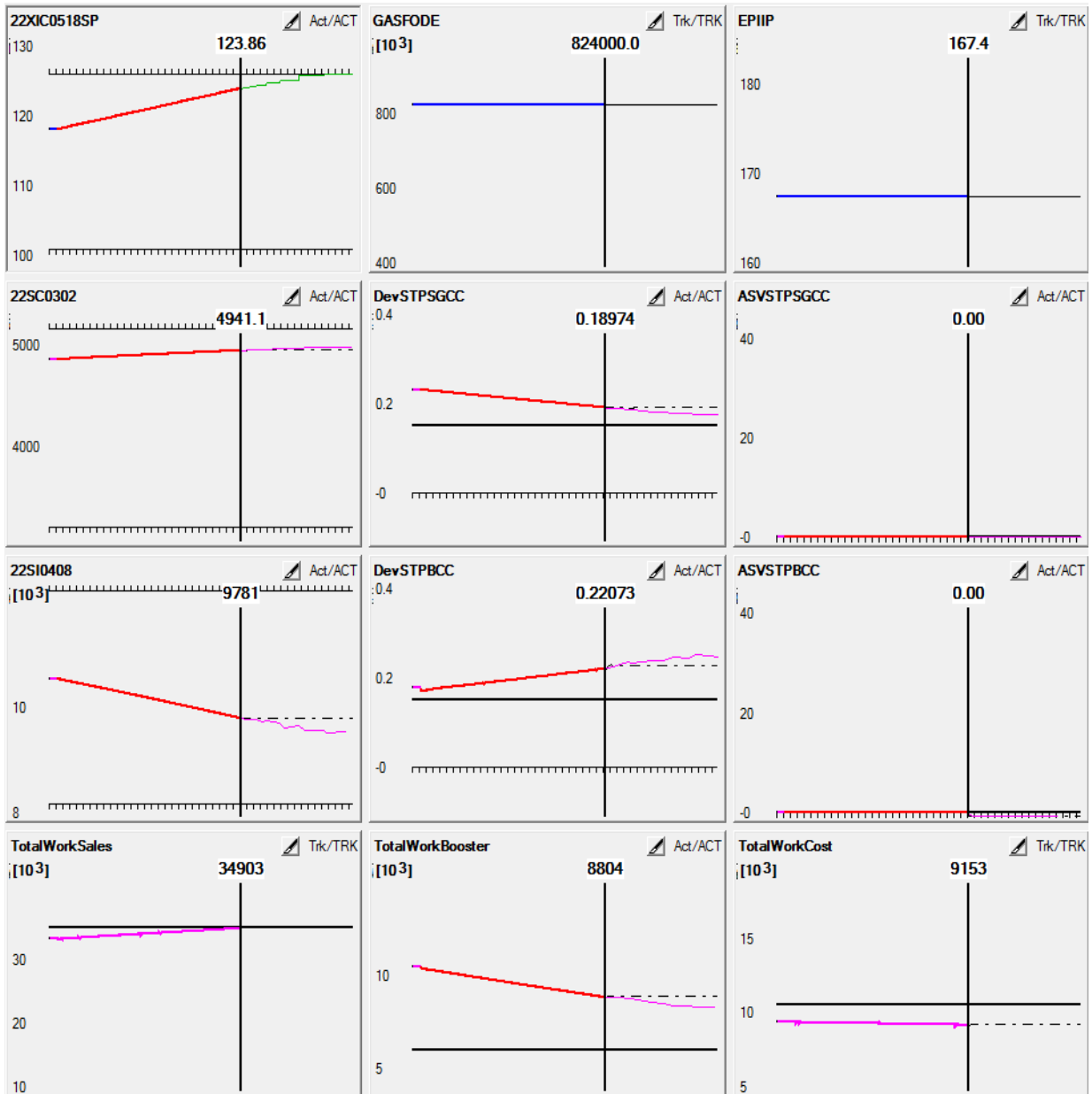


Figure 5.3: Minimizing consumed energy for Booster compressors

by raising compressors speed at the first stage and decreasing speed for the second stage compressors. It reaches the upper constraint of 126 bar and stay there. Not shown in figure but final values for required energy and its cost are summarized in table 5.1. Energy consumption increases for sales gas compressors and falls for booster compressors. Total cost with prices presented before goes down with 3%.

Several other simulations with other optimization variables were completed and summarized in table 5.1. A full rate case with 118 bara at booster suction pressure and 167,3 bara output pressure was used as a base case to compare with. Optimization with respect to TotalWorkCost gave same results but with considerably longer simulation time to reach same values. Minimizing the energy consume(TotalWorkSales) for the first stage gave higher cost values as was expected but also showed that control system is working. Another presented case in table 5.1 has the same conditions as mentioned but with high output pressure, 185 bara(low gas consume in Europe). It gave 4% decrease in cost with driving booster suction pressure to high constraint. Supplied fuel gas for first stage compressors is also shown in table but was not used in simulations.

Table 5.1: case2

Optimization Variable:	Fuelgas [Sm ³ /h]	Sales gas [kW]	Booster [kW]	TotalCost [kr/h]
Basecase	6152	33176	10456	9380
TotalWorkBooster	6481	35586	8344	9126
TotalWorkSales	5601	29459	14050	9883
TotalWorkCost	6480	35594	8342	9126
Basecase+Pout185	6150	33158	14827	10802
TotalWorkBooster	6480	35574	12093	10346

Case 3: Startup of compressor MPC with low setpoint on the booster compressor work and low rate

This case is similar to case 2 but compression takes place on medium rate. 'TotalWorkBooster' is chosen as an optimization variable again and medium rate response models are used. Priorities are kept the same as in case 2. Low limit for DEV for second stage compressors was changed to -0,02 as it showed that lower layer control does not react as expected. Usually, when the operation point cross the surge control line (DEV=0) the antisurge control must open the recycle valve. Antisurge control for booster compressors have a range of $\pm 0,02$ where nothing happens. So if MPC starts with DEV in this range, it can not start optimization. Otherwise, MPC works fine and able to minimize energy consumption by second stage. Final values are summarized in table 5.2 and compared to medium rate flow with 118 bara suction pressure and 167 bara pipeline output pressure.

Table 5.2: Medium rate optimization

	Fuelgas [Sm ³ /h]	Sales gas [kW]	Booster [kW]	TotalCost [kr/h]
Basecase75	5291	27581	8809	7836
TotalWorkBooster	5853	31237	6768	7829

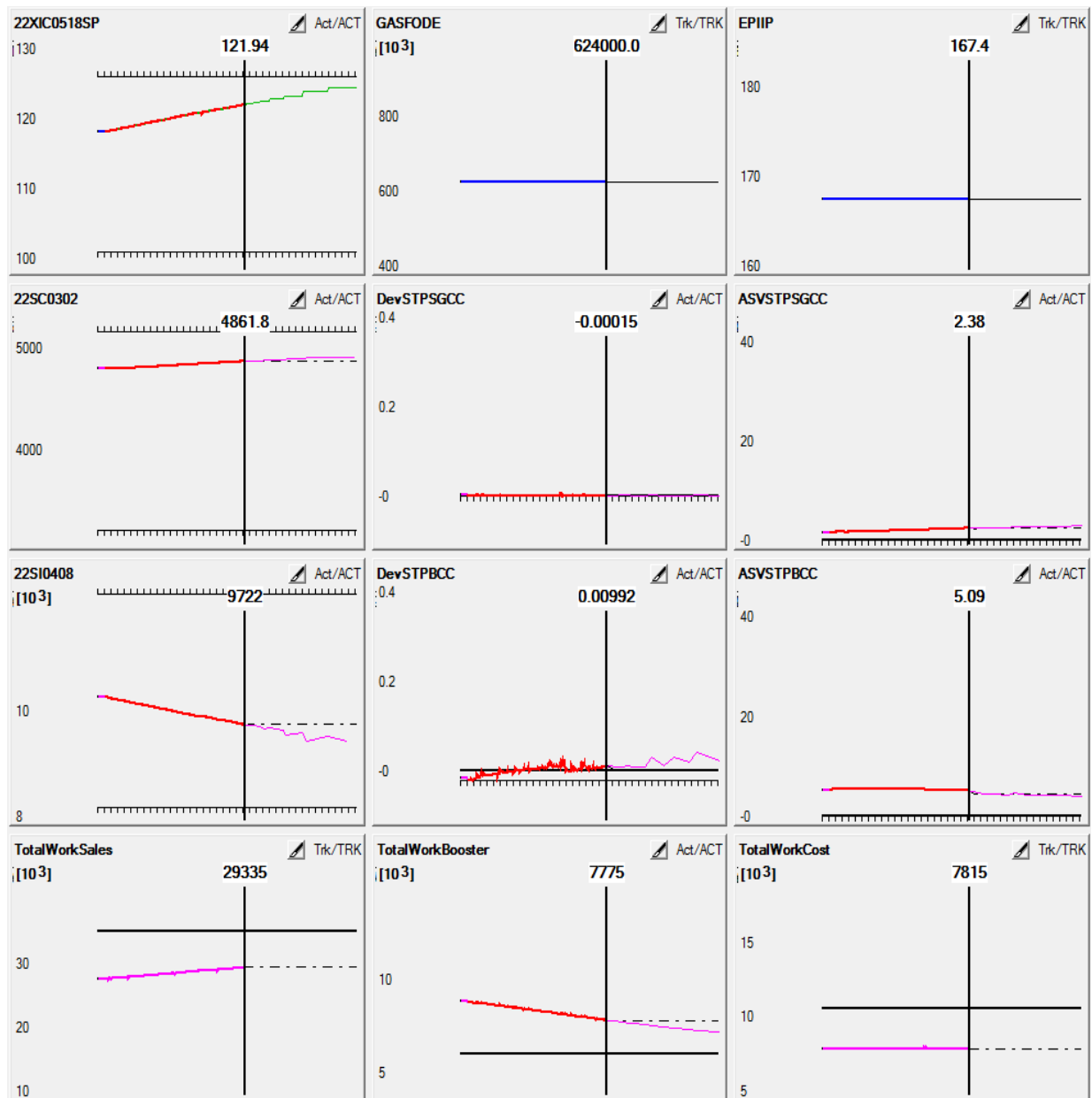


Figure 5.4: Medium rate optimization

Case 4: Start up of compressor MPC to control distance to the surge control line

This case was added in the late phase of the project when a new discovery was made. It is meant to show MPCs ability to control without any economic variables involved and was not seen first. All priorities are the same as stated in table 4.2 previously. Ideal value(IV) for the MV is turned off. The setpoints for all DEVs are changed to 0,15.

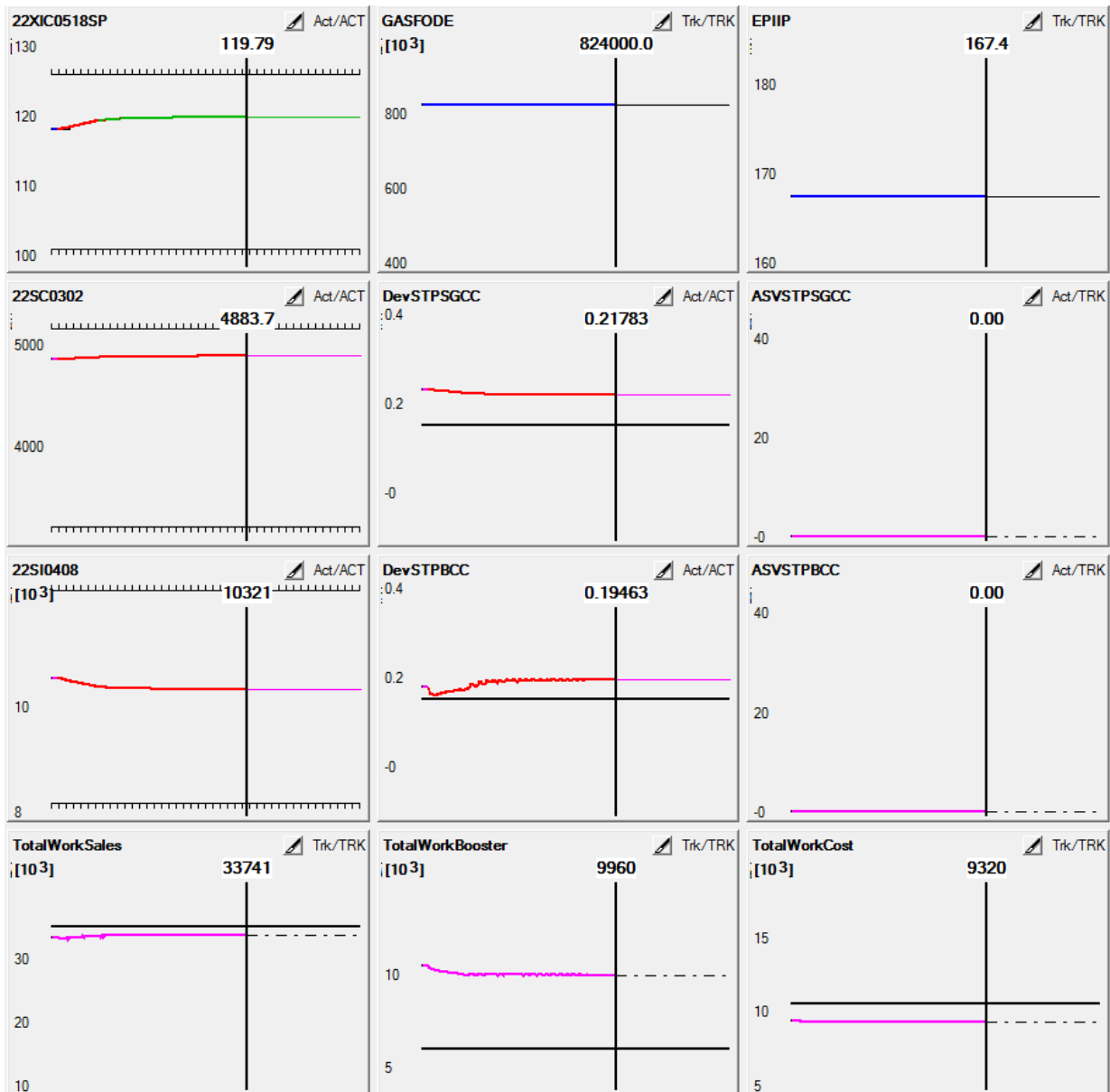


Figure 5.5: Only control of DEV is active, DEV=0,15

It can be seen from the figure that on start up, MPC increase suction pressure to about

119,7 and stabilizes there. This is done to optimize most important CVs which in this case are DEV setpoints(priority=2). Through simulation work it was observed several times that MPC tries to drive suction pressure from ideal value when its activated. It was solved by increasing priority for IV to 2, the same as a DEV priority. After that the IV value was perfectly hold. But it seems to be totally wrong to do it that way. The reason for drifting must be equal penalty weighting on all DEVs. Such that when compressors on one stage operate close to surge control line (DEV=0), other stage compressors may operate far from surge line. The MPC observe potential to minimize cost in QP problem by driving far variables in the direction of setpoint.

$$w = \left(\frac{Fulf \cdot (pred - settpoint)}{Span} \right)^2$$

This was difficult to see in cases which involved variables with priority 2, such as minimizing energy on a single stage. Increasing the setpoint value from 0 to 0,15 makes the penalty cost more equal. Compressors on first and second stage are controlled to keep almost equal distance to the surge control line. It confirms the ability of MPC to control load distribution.

Case 5: System behavior on significant export pipeline pressure increase

This case is meant to simulate the falling withdrawal of gas from the export pipeline. It can be seen as an extreme case as normally it takes more than 2 hours to increase pressure with 20 bar. All priorities and setpoint values kept the same as in previous case. The MPC

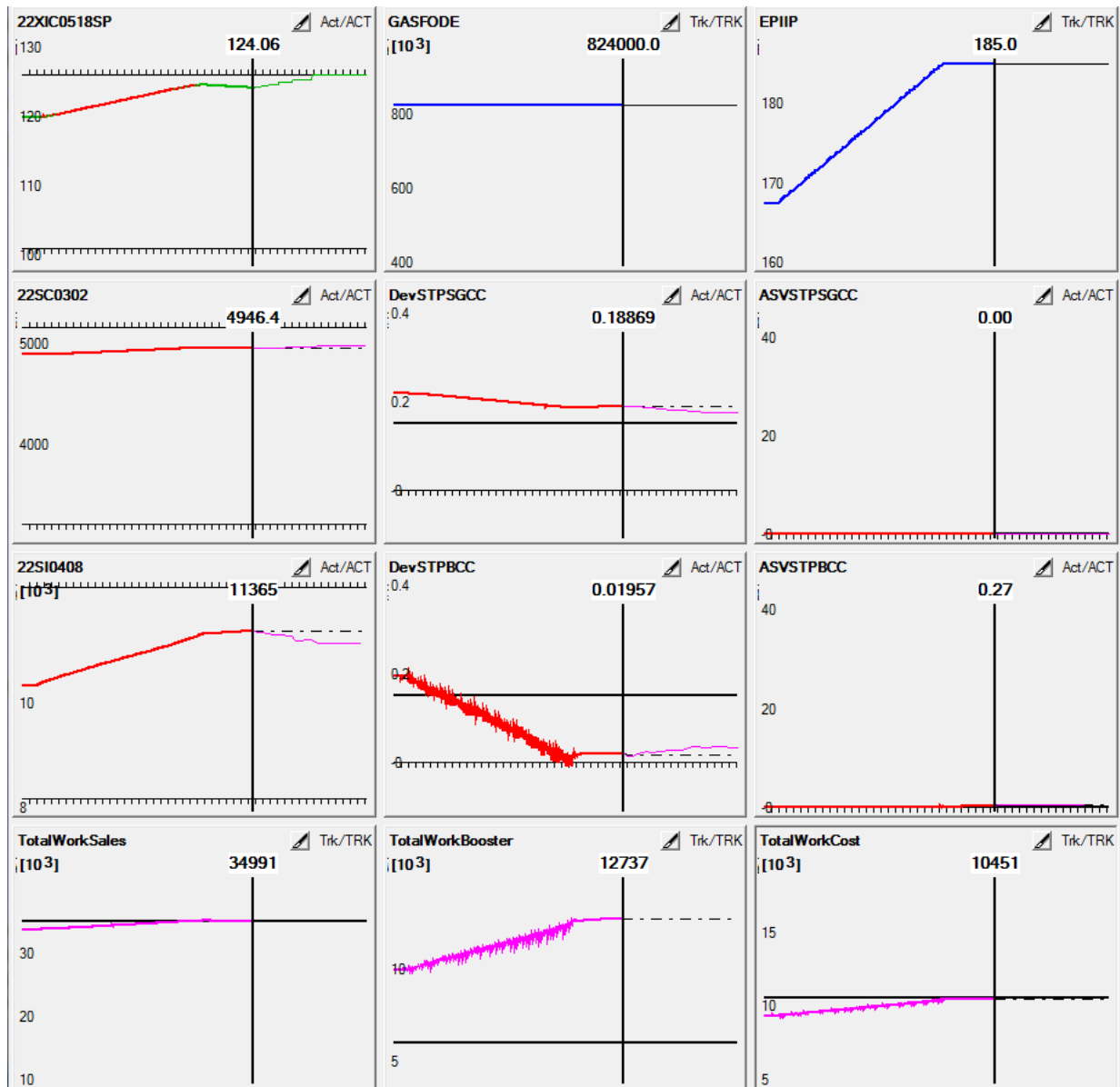


Figure 5.6: Response on ramping of pipeline pressure

struggling to reach equal DEV setpoints this time. The suction pressure is raised to almost its maximum to supply required pipeline pressure. Booster compressors operate on a surge control line. Although, MPC tries to equalize the DEV values. It will be not possible,

as to decrease the DEV value for the booster compressors the suction pressure must be increased even more which will not be allowed by the suction pressure upper constraint.

Another major issue is that system goes from one operational region to another and such situation creates problems for the supervisory MPC. As it can be seen from figure, the operation starts on the high rate and low pipeline pressure with closed antisurge valves. Corresponding models are used with zero ASV prediction. As operation reaches the max pressure, the antisurge control becomes active. But the models used here are not suited for ASV prediction and MPC gets stuck with wrong picture of underlying control structure. This non linearity creates problem for MPC operating with a same set of linear models in different regions.

A few more examples of system behavior can be found in appendix B, figures B.6- B.10. Read introduction to appendix on how to read response plots.

5.3 Discussion

Above, it was attempted to illustrate the ability of the developed compressor MPC to control and optimize the system with respect to different objectives. Also the major challenge of the linear supervisory compressor MPC was shown.

The application is able to minimize the booster compressors energy consumption which become a main objective of this paper. Shown pattern from case 2 was confirmed by simulations in all possible operational regions. The same way procedure was applied to try optimization of other variables such as energy consumption for the first stage(appendix B.8) and total cost function with assumed high energy prices. In addition, it was attempted to make an optimization of booster energy with considerable falling in the supplied flow rate, can also be found in appendix.

Cases 4 and 5 presented the ability of MPC to optimize with respect to economic independent variables. Load sharing between stages was optimized with respect to the recycling as by keeping setpoint values 0,15 on all DEVs the recycling is zero. This was found in the very late project stage and any different weighting was not tried. But it seems promising and should be considered for further testing. Weighting and optimal setpoint values probably will be challenging parameters as they should vary with operational points.

In addition, case 5 unveiled probably the main weaknesses of applied linear response models. In transition from one operational region to another MPC stops to work properly. Active constraints varies with different disturbances as it was stated in the plantwide control procedure.

Modeling outcome varies significantly depended on operational conditions. The effort was applied to make good models for each region. Response models for high rate and low pressure were acquired by a straight forward procedure of stepping variables and recognizing the response. Other types, for example response models for low rate uncovered new and unexpected behavior of underlying system. Response modeling with high pipeline pressure gave the total different antisurge response on high rates, than with low pipeline pressure. High rate and medium rate models gave various responses, as in first case, compressors

operate far from surge control line, while in the second case, operation is at the control line. The underlying overall compressor control structure is strongly non linear for the supervisory compressor MPC.

Chapter 6

Conclusion

The main outcome of this work is a developed and tested MPC application for the optimization of operation of export gas compression system. Another important achievement is a plantwide control design suggestion with attempt to systematic approach of the control problem. At last but not least, operation of the D-spice OPC server is fully tested with Septic MPC and simple communication between Matlab and D-spice is established.

Strong nonlinear interactions in the plant uncovered during the modeling procedure and simulation of operation. Acquired linear models, works fine only in a specific operational regions where they were taken. A gain scheduling or a nonlinear MPC needed to be implemented to handle transitions between various regions and to collect prediction models to one complete application.

This thesis will probably does not make a huge breakthrough into compressor control subject but author hopes that it will be a useful supplement for the future work. Specially for students who might be working further with Kårstø compression system or another Kårstø application. All data files in current state will be hand over together with this thesis. It includes: D-spice model with updated parameters, Septic MPC application, OPC server configuration, several Septic models and D-spice operational points for the different conditions. It will hopefully give the opportunity to proceed straight to the further control structure development and testing.

Through the whole project period, several major breakthroughs can be highlighted such as: successfully running of the Kårstø master model, established connection between D-spice and Matlab, development of the first simple but workable Septic application, plantwide control structure, building of final MPC application.

This paper showed economical potential for minimizing electrical energy consumption based on a recently energy prices when the gas produced energy has a significant lower cost compared to electrical power price. But it must be kept in mind during development and implementation of the final plant application that energy prices are the uncertainty factor here. It is not impossible that cost situation will totally change. With for example, raise in the CO₂ quota prices, which is politically indicated to take place over the next decade. Another thrilling aspect with maintaining and optimization of throughput for compression system is potential connection of new offshore fields to Kårstø processing plant.

6.1 Problems faced during the thesis

In the early beginning, some local PC problems with D-spice installation and model configuration parameters were faced. The troubleshooting resulted in changed model configuration and a register file with keys which must be applied for the proper master model operation in D-spice, can be find in appendix F.2. D-spice simulator showed stable operation through the whole semester, until 2-3 last weeks before deadline. In this period, almost daily restarts of the xkeys server was performed. The reason for that, was not identified but some of the faced problems are summarized in appendix C.3.

6.2 Further Work

The scope for the further work is wide. Probably the next step is to extend MPC application to work as a performance controller, instead of todays PI(D) controllers and associated logic. Decoupling between capacity and antisurge controllers assumed to be one of the possible challenges for this work. Another one, is possible reconfiguration of implemented antisurge control as some of the links from replaced capacity control might be broken. As from work experience with this thesis, it can be surely stated that todays underline control structure is very complex and some insight into it is absolutely needed.

On a large scale, the implementation of a nonlinear MPC or a gain scheduling is needed to handle non linearities in the system presented before.

Another possible direction for further work is implementation of antisurge model predictive control which seems to be a very challenging and popular field among researchers.

Interconnection of D-spice OPC server and Matlab as well as its configuration can be interesting for those who will work with possible project in development of MPC for the butane splitter at the department of Cybernetics.

Appendix A

Drawings

Figure A.1 illustrate compression system.

Figure A.2 shows the overall model of the Kårstø processing plant implemented in D-Spice process simulator. In this particular project only section 2, 4 and 7 were activated.

Figure A.3

Figure A.1: Statpipe Compression System

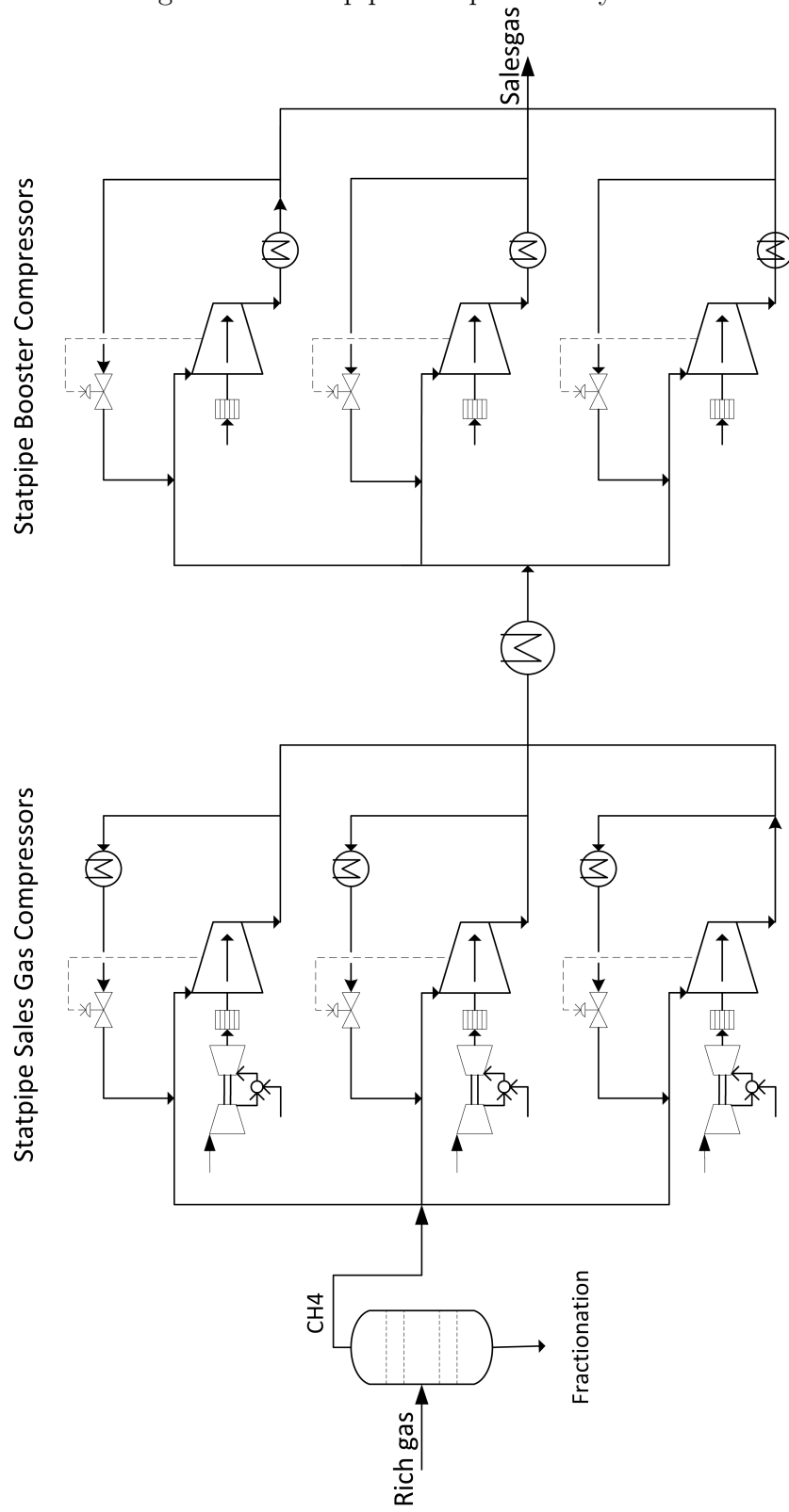


Figure A.2: Kårstø Dynamic Plant Simulator

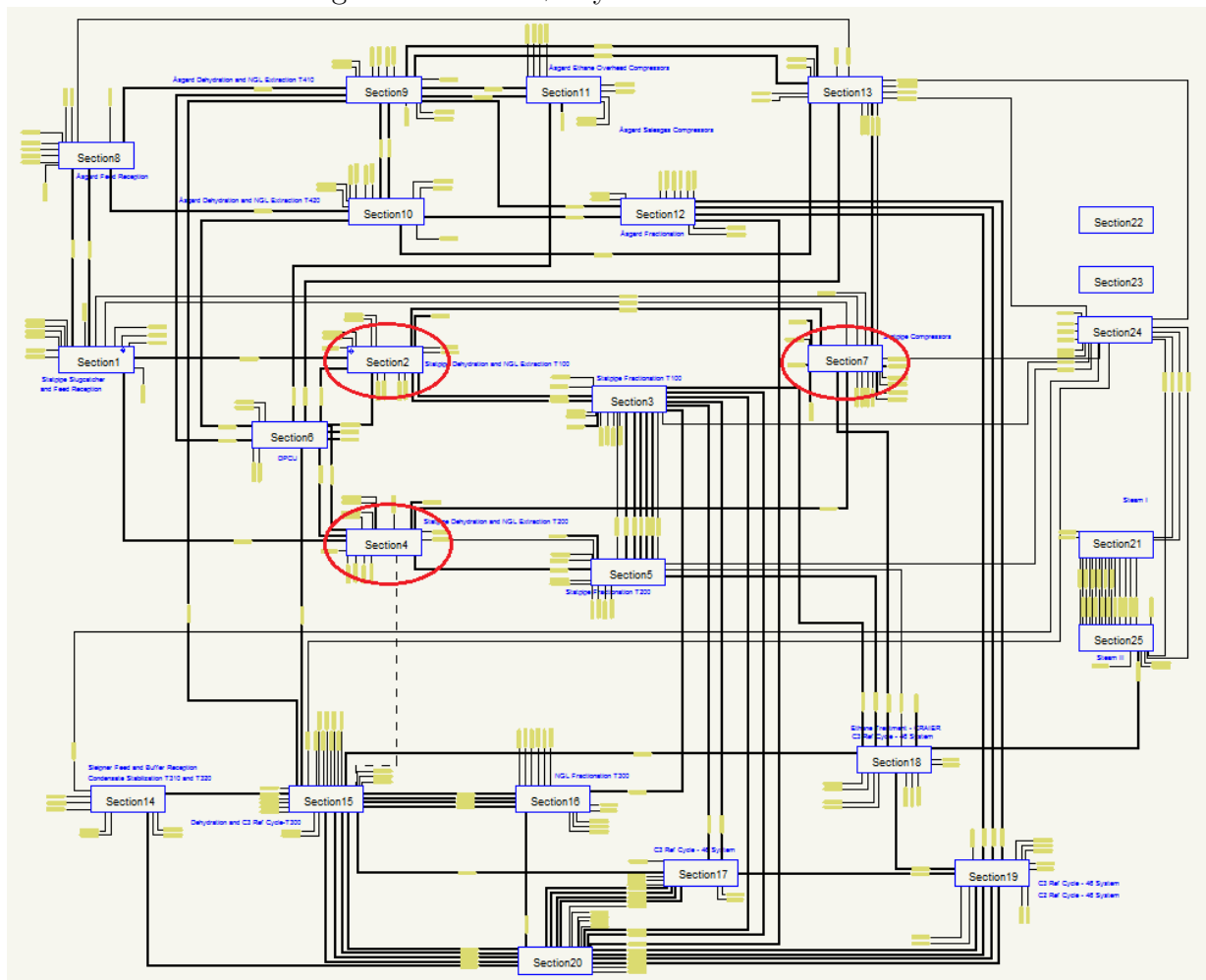
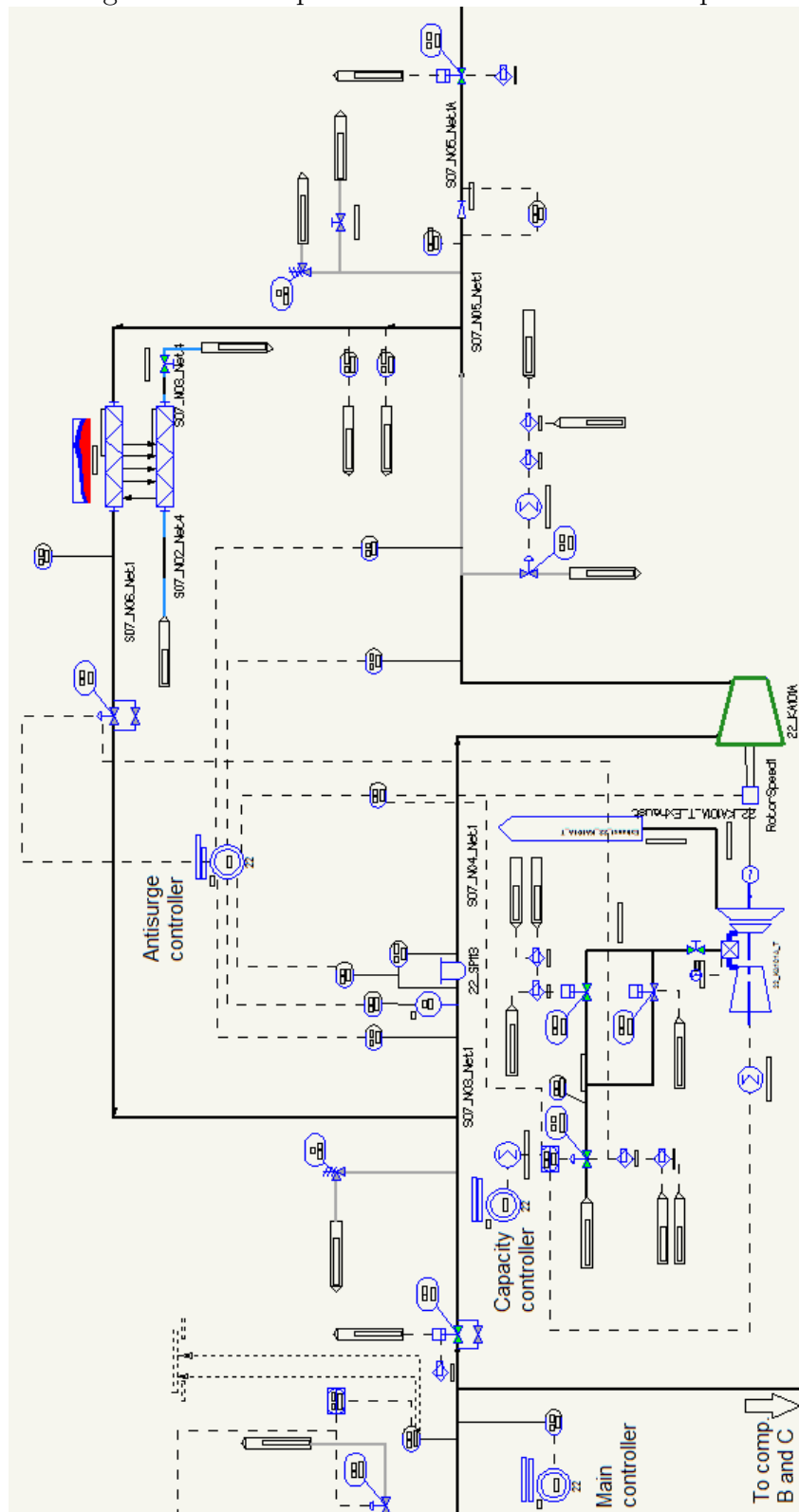


Figure A.3: Compressor Instrumentation in D-Spice



Appendix B

System responses

Figures below shows the response models and responses for the main variables at the plant.

Model panels contain: in the first column, on top is difference between setpoint and measurement for the sales gas compressors suction pressure. Next two are the energy consume for the first and second stage, respectively. Fourth cell is a response for the sum of operational cost for both stages. Second, third and fourth columns contains six fields where each cell corresponds to specific compressor, from A compressor on first stage to C compressor on second stage. Second column is compressor speed, third is distance from surge control line(DEV), fourth is antisurge valve opening(ASV).

Plant response panels contain: in the first column on top setpoint for booster suction pressure(MV), next is feed rate of rich gas from offshore(DV), then methane gas rate into compression system after ethane is removed from rich gas. Fourth cell shows suction pressure to first stage, then output pressure(DV) and last one is difference between setpoint and measured value for input pressure. Second, third and fourth columns contains six fields where each cell correspond to one of the six compressors with sales gas machines on top. Second column is speed, third - distance to control line, fourth ASV opening. Fifth column - on top is gas flow output[kg/h], then required power for the first and second stage[kW]; next one is cost per hour[kr/h] for the total power consume of both stages, TotalFuelSalesSm3 shows required gas input for turbines at first stage in Standard cubic meters. Last one is not used for the optimization but shows the value of the produced gas[kr/h].

Parameters:

- Blue - calculated response
- Yellow - steady state value
- Red - measured response
- Green - difference between yellow and blue
- Violet - close loop response
- Stippled - open loop response

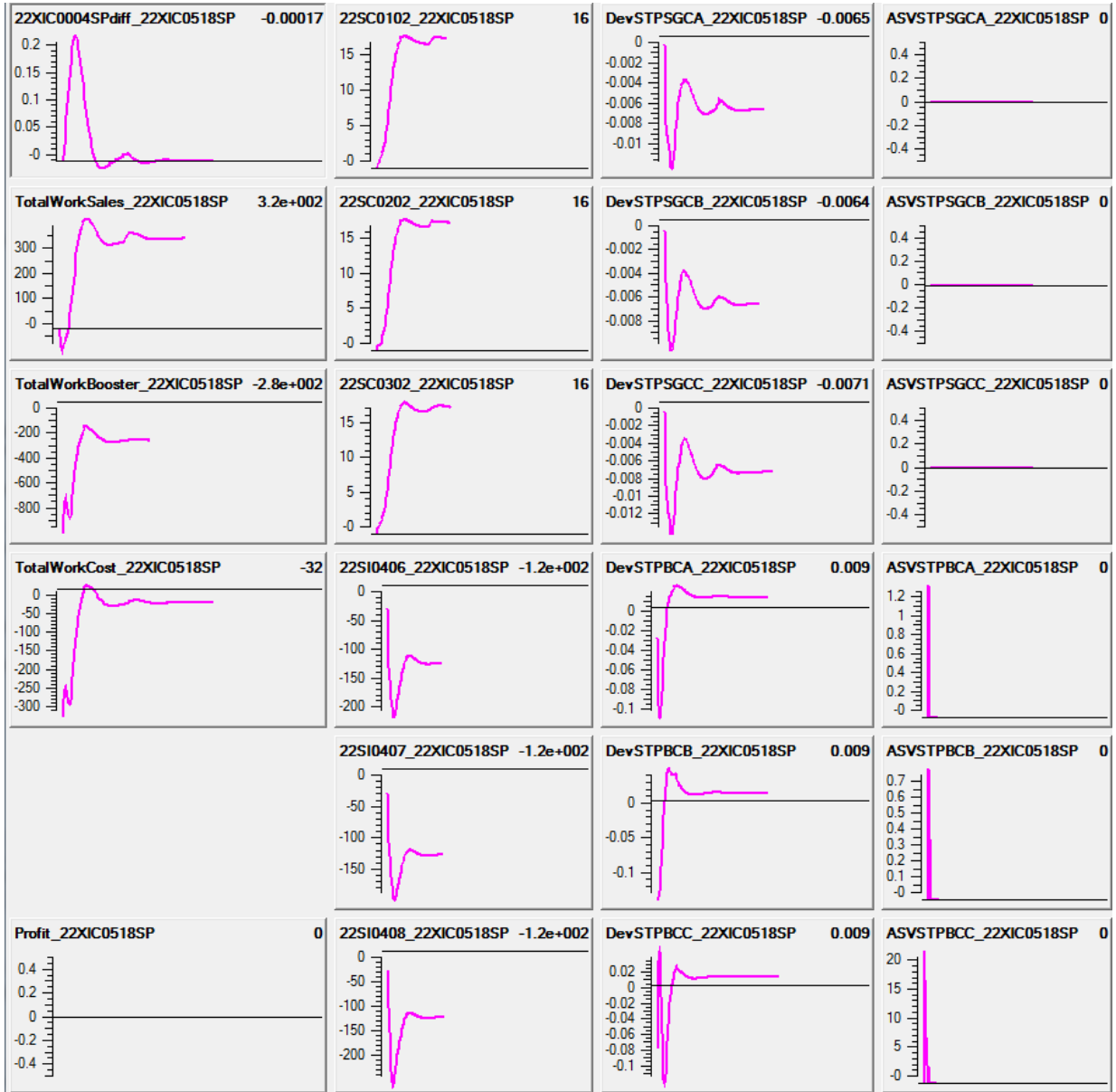


Figure B.1: Model for full rate, step in booster suction pressure

APPENDIX B. SYSTEM RESPONSES

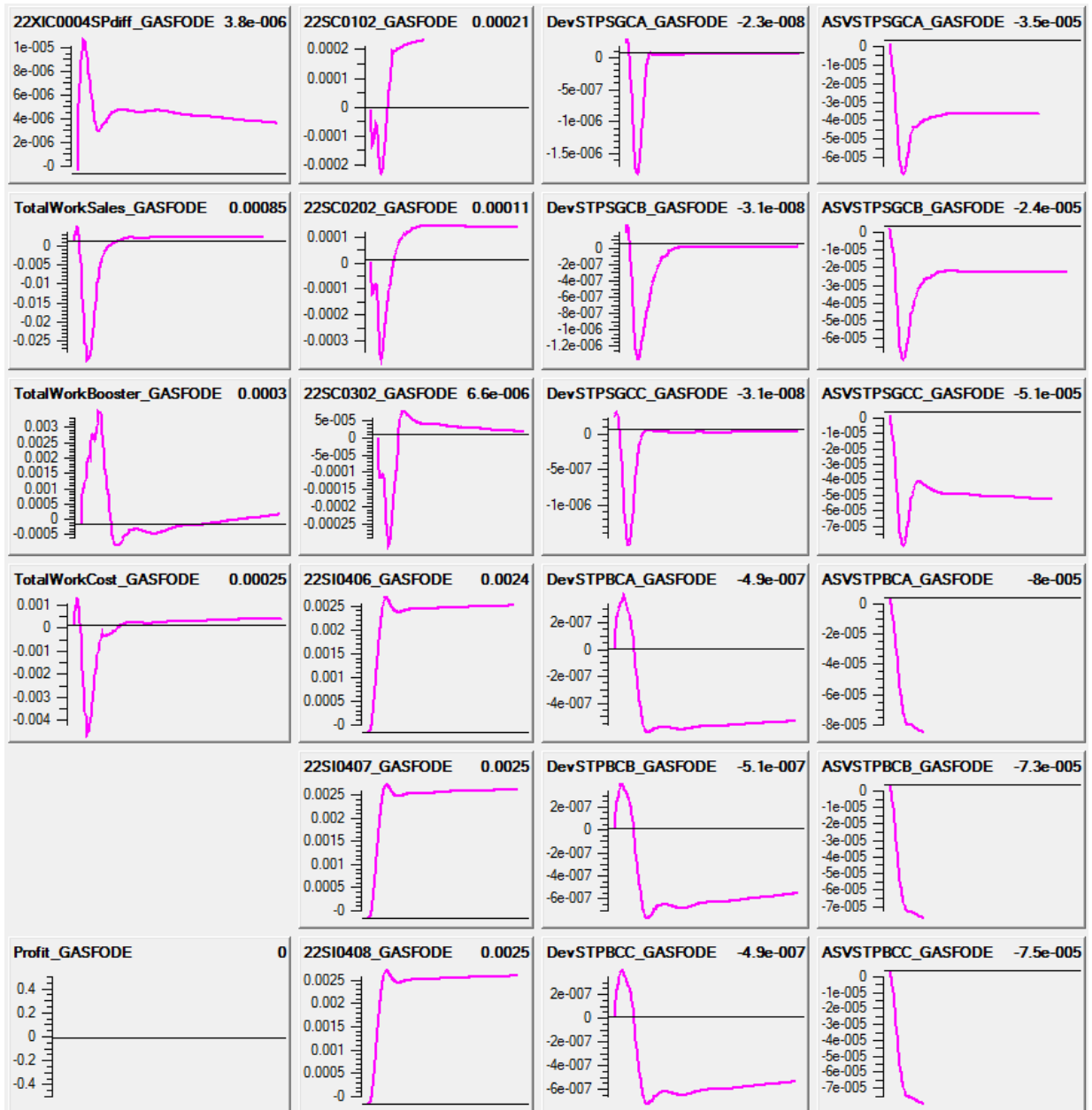


Figure B.2: Model for full rate, step in feed rate

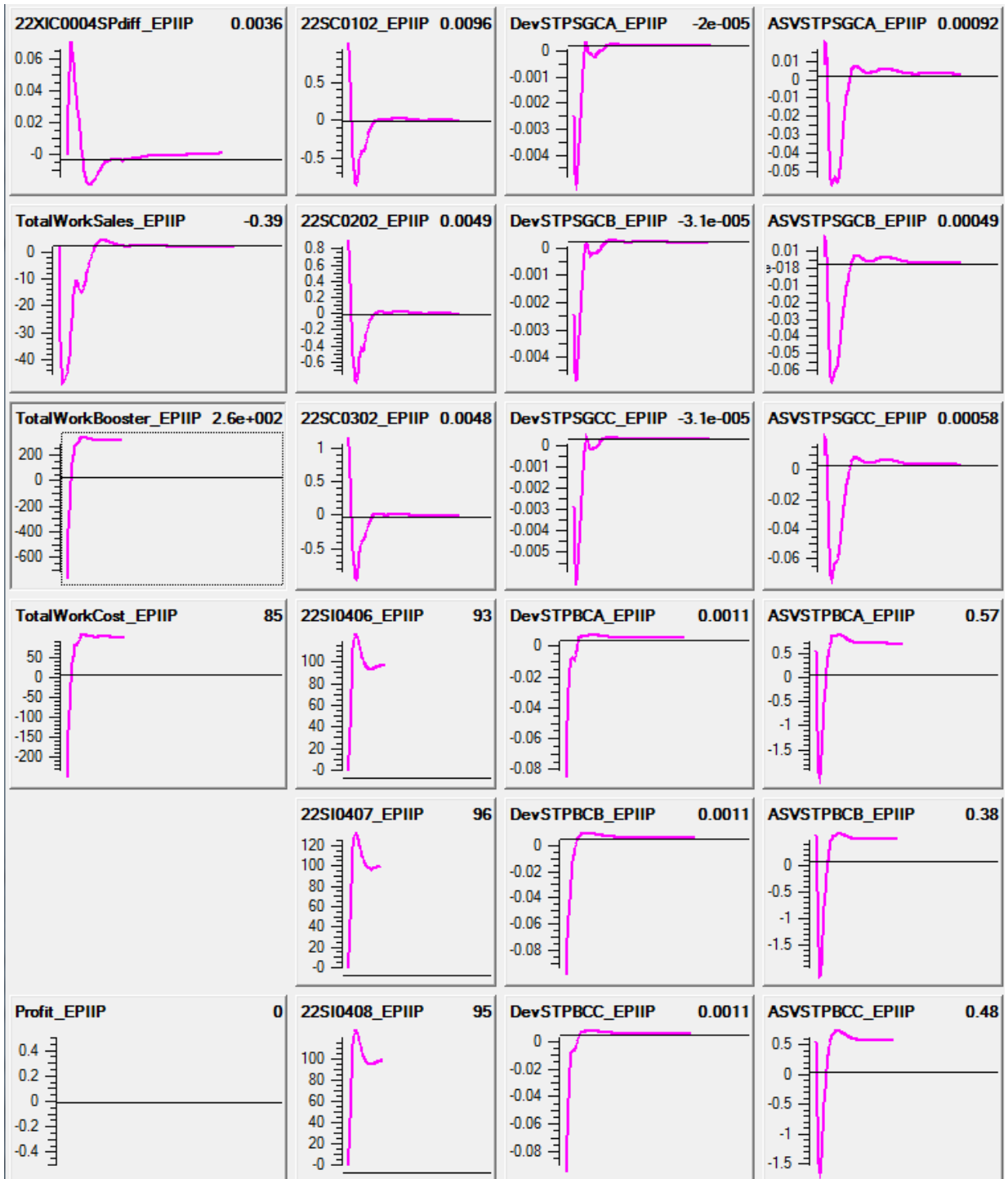


Figure B.3: Model for full rate, step in output pressure

APPENDIX B. SYSTEM RESPONSES

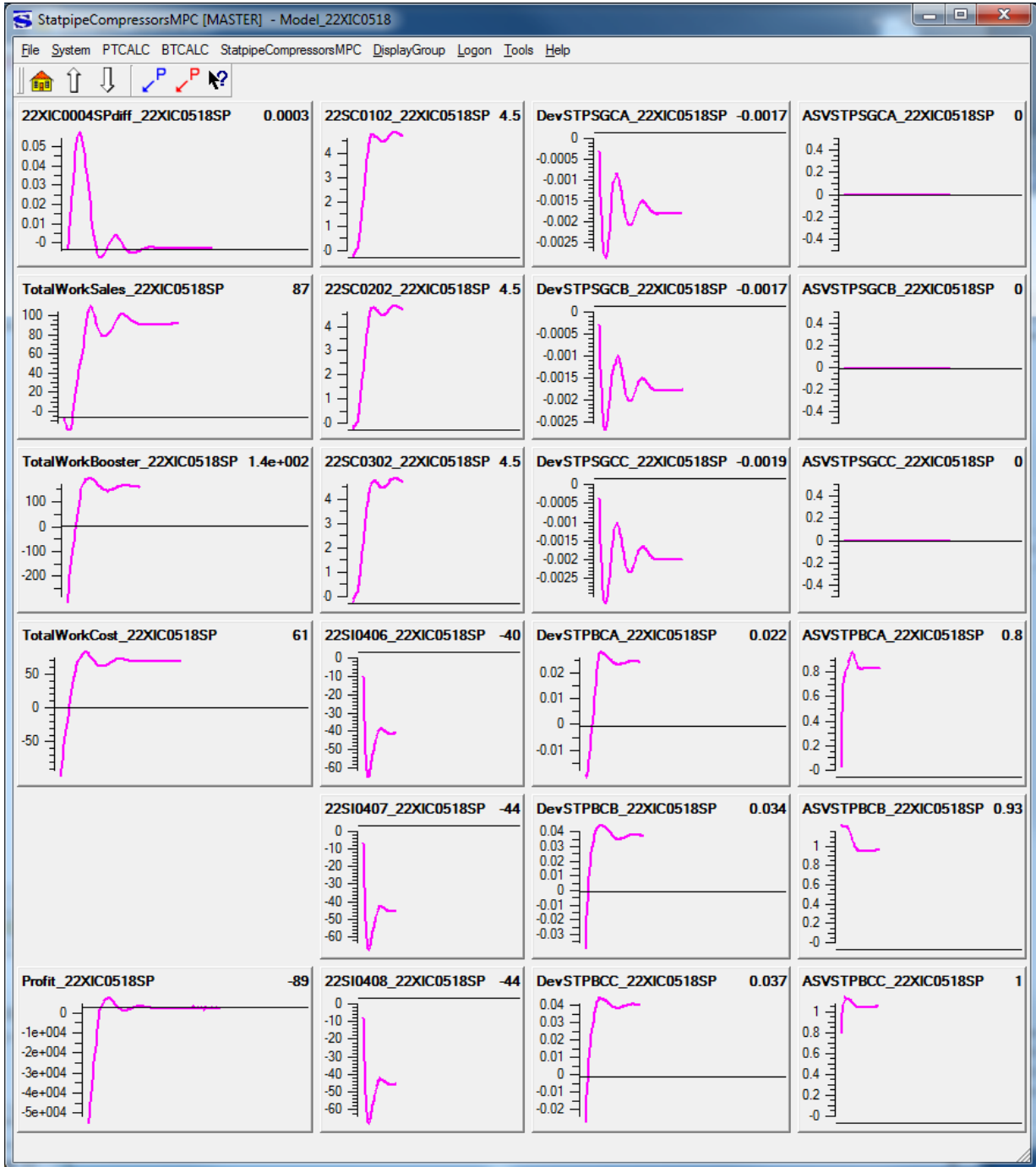


Figure B.4: Modeled with high output pressure to provoke ASV on full rate. Higher ASV values then in other cases

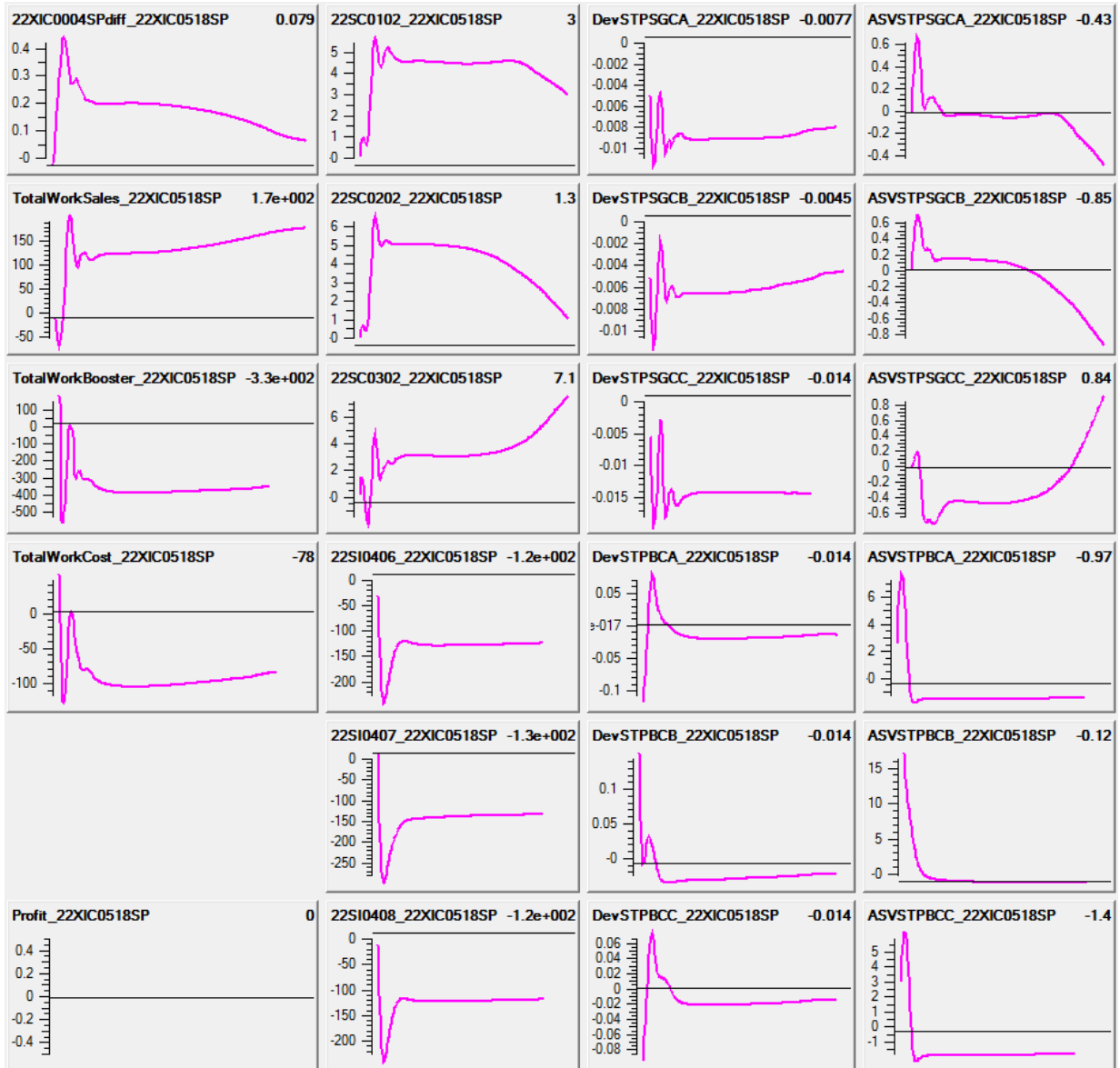


Figure B.5: Low rate, step in output pressure. One of the compressors take over the whole recycling. Model not suited for use

APPENDIX B. SYSTEM RESPONSES

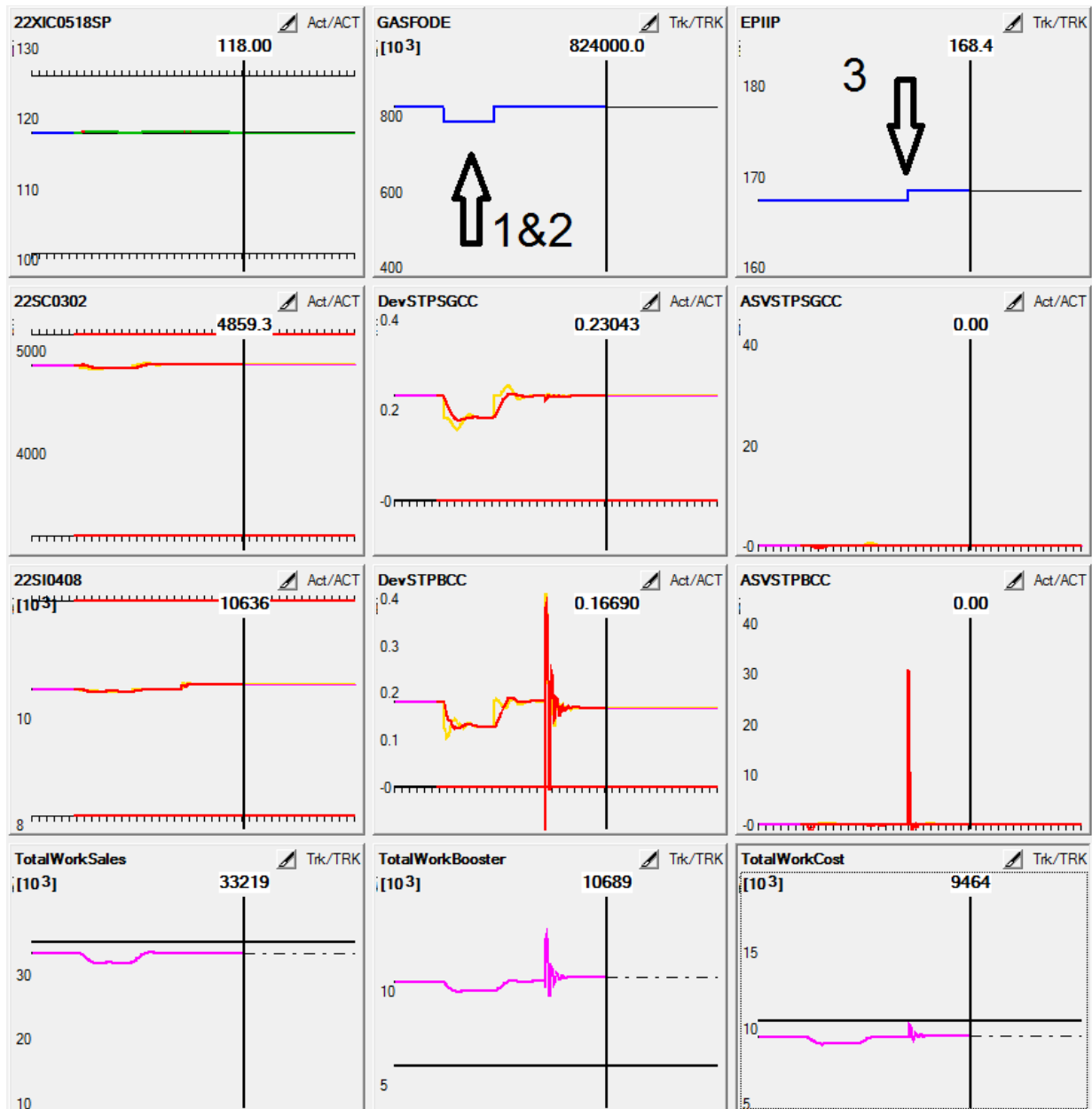


Figure B.6: Response to fast disturbances. Flow rate $\pm 5\%$, pressure +1 bar

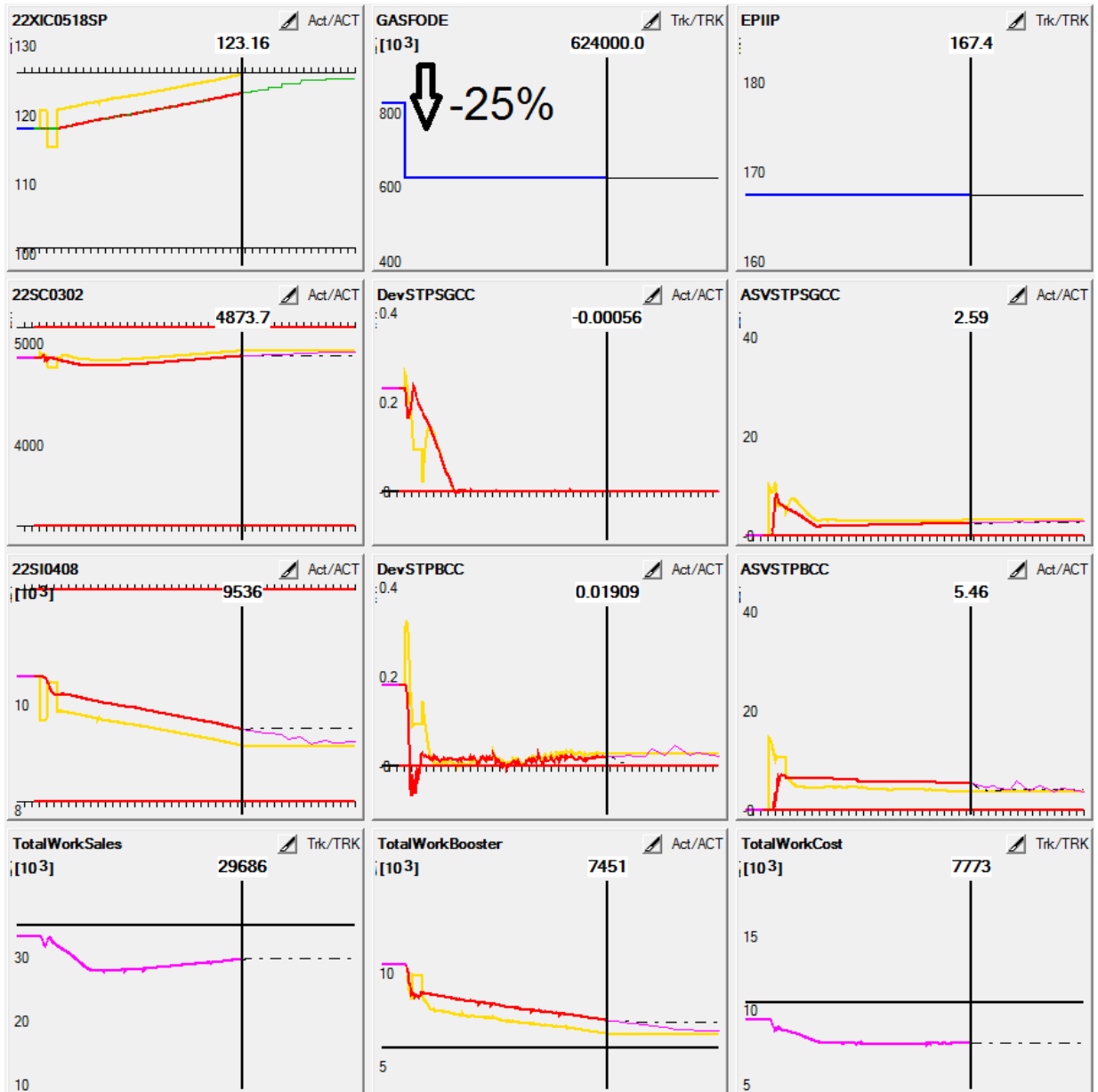


Figure B.7: Optimization of TotalWorkBooster after significant flow drop(-25%)

APPENDIX B. SYSTEM RESPONSES

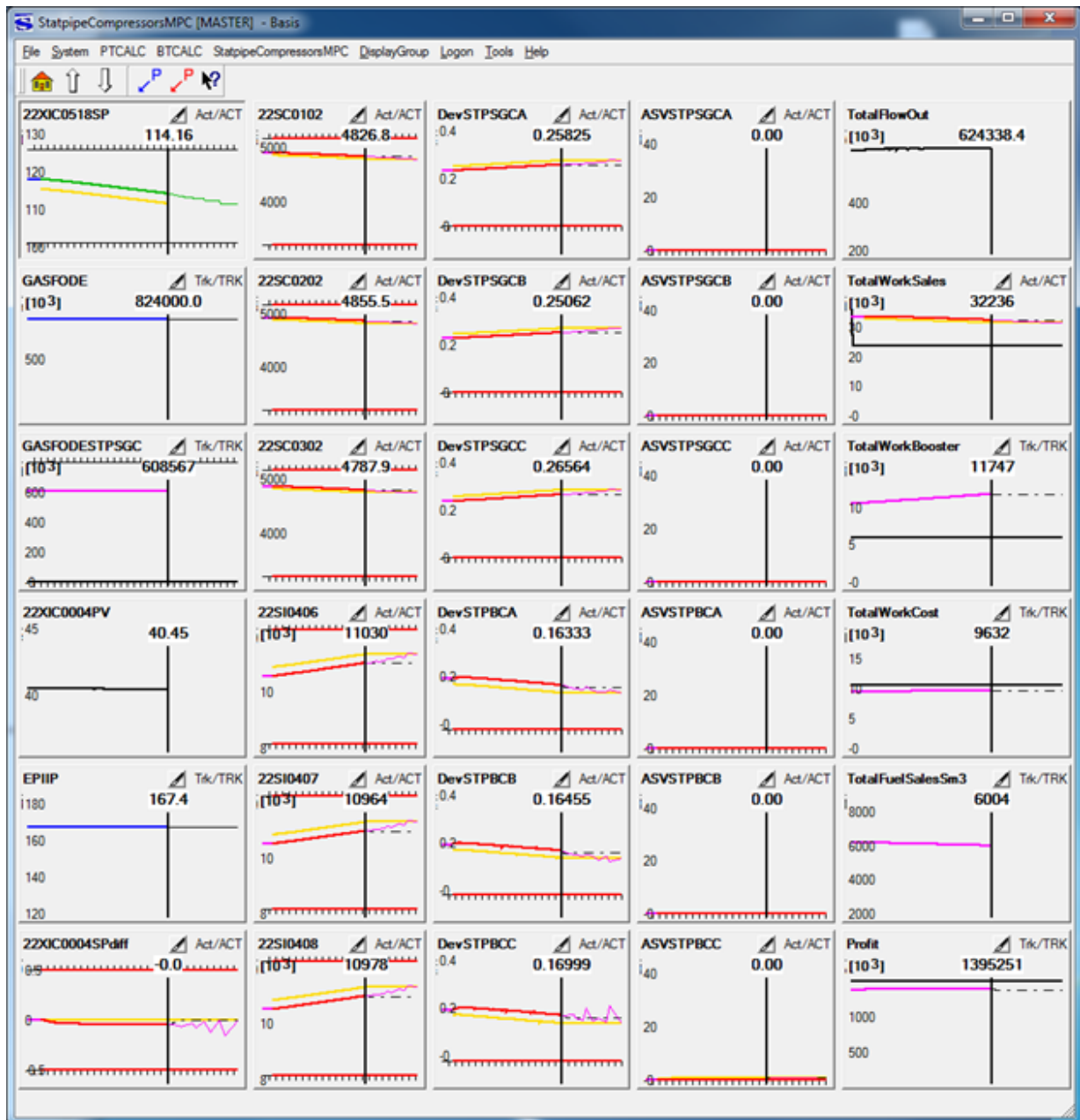


Figure B.8: Optimization of energy consumption for first stage compressors

APPENDIX B. SYSTEM RESPONSES

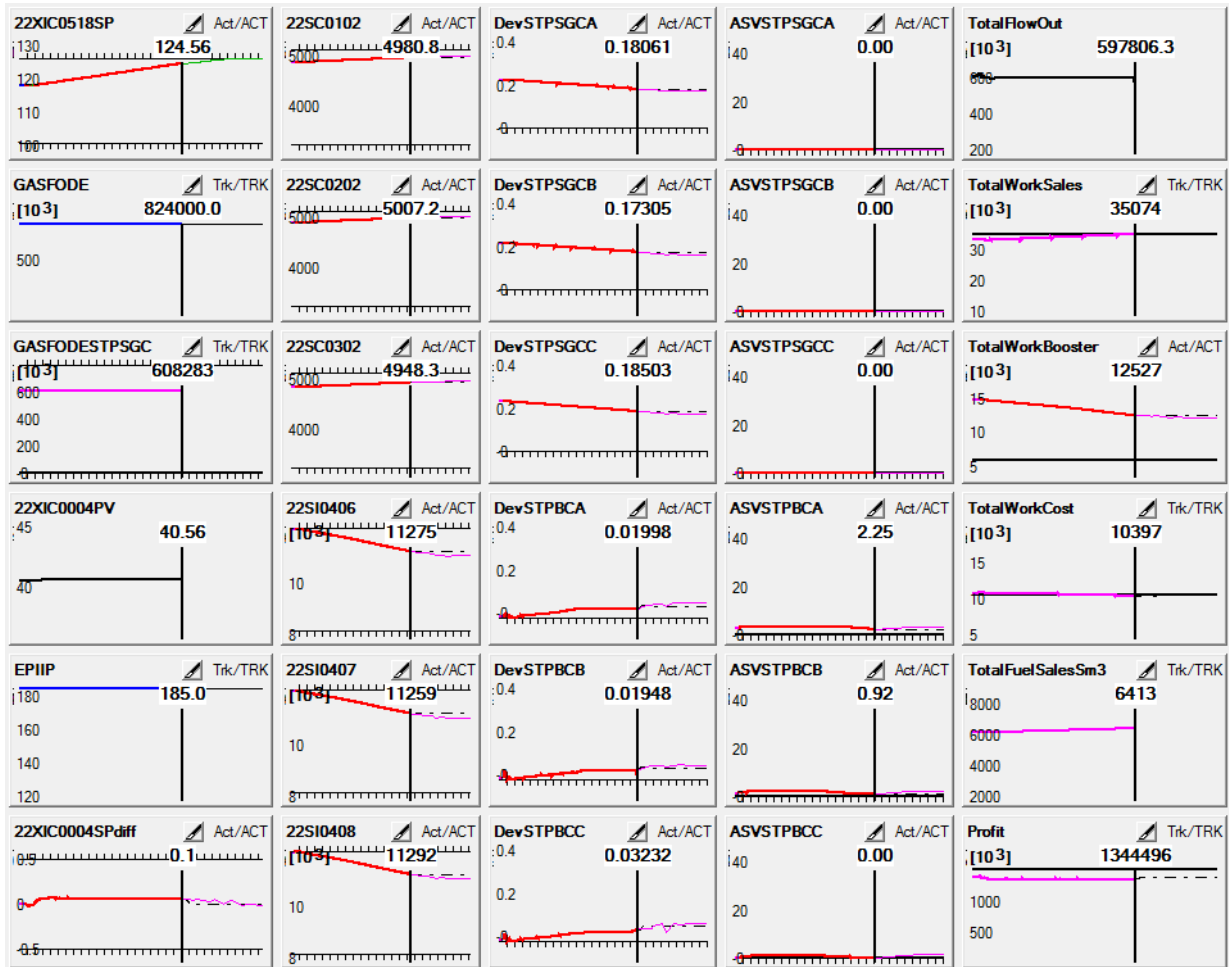


Figure B.9: Optimization of TotalWorkBooster on the high pipeline pressure(185 bar)

APPENDIX B. SYSTEM RESPONSES

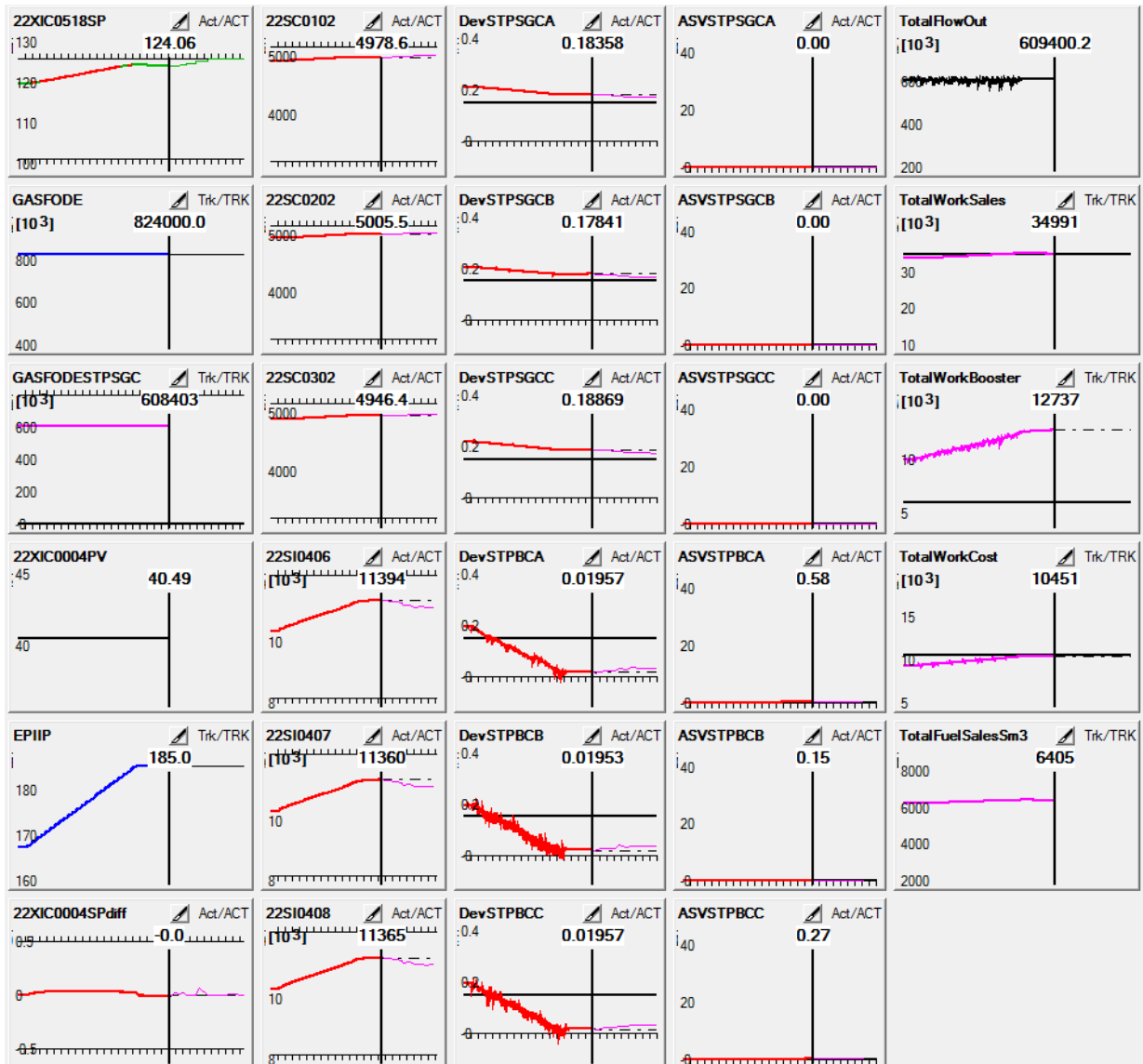


Figure B.10: Behavior on significant pipeline pressure increase. Full panel

Appendix C

D-spice

C.1 Changes in model

This section summarizes changes in the received D-spice model of compression system. All antisurge controllers were set into Auto-mode as they were deactivated by default. All side streams in section 2 and 4 were deactivated due to the testing of the system. As it showed that rapid increase in pressure could cause the flow reversals.

At the plant and originally in the model, $STPBC_A$ and $STPBC_B$ deliver sales gas into EuropipeII while $STPBC_C$ is connected to the pipe towards Draupner. It was changed by closing valve to Draupner and open valve in direction of EuropipeII for Booster compressor C. The decision was made to simplify the system and to bring all compressors to the common output manifold with the same pressure. In addition alarm value for high output pressure from C compressor was increased.

It showed to be that distribution of load among compressors was not perfectly balanced and effort was put into finding of differences together with co-supervisor from Kårstø. 'Moment of inertia' for $STPSGC_C$ was changed from 100-10 to 100-100 to match other two compressors in parallel. Also 'Synchronous speed at 100% control signal' for driven motor engine (22_KA_102CM) for Booster C compressor was changed to 4083rpm. After that distribution became more equal but still with some differences at the recycling responses.

Pressure values in CCC stage controllers

Shown setpoint in CCC main controller for suction pressure for the sales gas compressors has following formula: $(P - 1) * 10$ where P is a measured pressure in bar, e.g. $(41,5 - 1) * 10 = 405$, such that shown setpoint is 405. Setpoint for the CCC main controller for booster compressors has a scaling factor of 1,1967 between measured and shown, e.g. SP=98,5 in controller correspond to 118 bara measured pressure.

C.2 Add a PID controller to D-spice

Due to the fact that the D-spice manual does not contain a detailed description for procedure of installing a new PID controller, such description is shown below.

1. Add a PID controller from palette to the left.
2. Connect PID controller to desired transmitter and control element with dash line.
3. Copy "Max/Min limit for setpoint" values from transmitter to controller.
4. Insert parameters in to controller:
Range offset = Min limit for setpoint
Range span = Max - Min (limit for setpoint)
Output range offset = 0
Output range span = 1 (100%)

C.3 Comments

It could be interesting to look at communication between local D-Spice and server to try find possible issues for server access feiler.

1. Known D-spice problems/ assumed reasons:

- D-spice can run forward from OPC server. Probably, synchronization between clocks in D-spice and OPC server has a problem. Simple test by taking note of two timestamps unveiled the issue. First sample was taken at the 11:40.50 D-spice model time, corresponding value was read from OPC server with Matri-conOPC explorer and was 0,0081018519. That's gives a ratio or value of 1second is equal to 0,0000115658. Second time sample was taken at the 1:08:14.50 in D-spice and has corresponding value of 0,0473842593 which gives a 1 second value of 0,0000115727 which is not the same as the first one. This is probably not the most critical problem as it will make difference in simulations with a quite long time period but should be considered under development of own applications.
- No response from D-spice server operated by Dep. of cybernetics. Possible caused by the open but not active D-spice client on the user machine. Has experienced that problem after holding D-spice open over night one time and at another time, during the period of a couple hours at the daytime.
- Popup Warning: S04 SubModel-E-0010 Connection failed. Disable model S04 (Timestamp). May occur after Dspice freezing and incorrect closing procedure. Typically requires restart of the license server 'xkeys' to be fixed.
- Possible problems with OPC server for windows 64-bit computers. Author has used windows 7, 32-bit edition for work with this thesis and has not experienced any trouble with OPC server. But know about another user who did not get the OPC server to work properly on a 64-bit machine. Some days of joint troubleshooting, identical installation procedure and identical registry did not solved the problem.
- If running speed exceed 15 sec., connection between sections in master model is lost. It was easily tested by observing output/input in two connected sections. But works fine for the single stage.

Appendix D

MATLAB

General tips for those who will try to create applications which will communicate with D-spice OPC server:

To read time values from a D-spice OPC server, decimal tall in MATLAB must be extended with 'Long' command.

Most likely, it is not possible to start/stop D-spice through OPC, must be done separately

Matlab code for Guitest2 application presented in chapter 3 shown below. It creates a graphical user interface. It is also possible to change flow rates in section4 and read flow throughput for all six compressors.

```
%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%%  
function varargout = guitest2(varargin)  
% GUITEST2 MATLAB code for guitest2.fig  
%     GUITEST2, by itself, creates a new GUITEST2 or raises the existing  
%     singleton*.  
%  
%     H = GUITEST2 returns the handle to a new GUITEST2 or the handle to  
%     the existing singleton*.  
%  
%     GUITEST2('CALLBACK',hObject,eventData,handles,...) calls the local  
%     function named CALLBACK in GUITEST2.M with the given input  
arguments.  
%  
%     GUITEST2('Property','Value',...) creates a new GUITEST2 or raises  
the  
%     existing singleton*. Starting from the left, property value pairs  
are  
%     applied to the GUI before guitest2_OpeningFcn gets called. An  
%     unrecognized property name or invalid value makes property  
application  
%     stop. All inputs are passed to guitest2_OpeningFcn via varargin.  
%  
%     *See GUI Options on GUIDE's Tools menu. Choose "GUI allows only
```

```

    one
%       instance to run (singleton)".
%
% See also: GUIDE, GUIDATA, GUIHANDLES

% Edit the above text to modify the response to help guitest2

% Last Modified by GUIDE v2.5 11-Apr-2012 11:16:19

% Begin initialization code - DO NOT EDIT
gui_Singleton = 1;
gui_State = struct('gui_Name',       mfilename, ...
                  'gui_Singleton',   gui_Singleton, ...
                  'gui_OpeningFcn',  @guitest2_OpeningFcn, ...
                  'gui_OutputFcn',   @guitest2_OutputFcn, ...
                  'gui_LayoutFcn',   [], ...
                  'gui_Callback',    []);
if nargin && ischar(varargin{1})
    gui_State.gui_Callback = str2func(varargin{1});
end

if nargout
    [varargout{1:nargout}] = gui_mainfcn(gui_State, varargin{:});
else
    gui_mainfcn(gui_State, varargin{:});
end
% End initialization code - DO NOT EDIT

% --- Executes just before guitest2 is made visible.
function guitest2_OpeningFcn(hObject, eventdata, handles, varargin)
% This function has no output args, see OutputFcn.
% hObject    handle to figure
% eventdata  reserved - to be defined in a future version of MATLAB
% handles    structure with handles and user data (see GUIDATA)
% varargin   command line arguments to guitest2 (see VARARGIN)

% Choose default command line output for guitest2
handles.output = hObject;

% Update handles structure
guidata(hObject, handles);

% UIWAIT makes guitest2 wait for user response (see UIRESUME)
% uiwait(handles.figure1);

% --- Outputs from this function are returned to the command line.
function varargout = guitest2_OutputFcn(hObject, eventdata, handles)
% varargout  cell array for returning output args (see VARARGOUT);
% hObject    handle to figure

```


APPENDIX D. MATLAB

```
% eventdata reserved - to be defined in a future version of MATLAB
% handles structure with handles and user data (see GUIDATA)

% Get default command line output from handles structure
varargout{1} = handles.output;

% --- Executes on selection change in listBox1.
function listBox1_Callback(hObject, eventdata, handles)
% hObject handle to listBox1 (see GCBO)
% eventdata reserved - to be defined in a future version of MATLAB
% handles structure with handles and user data (see GUIDATA)

% Hints: contents = cellstr(get(hObject,'String')) returns listBox1
% contents as cell array
% contents{get(hObject,'Value')} returns selected item from
% listBox1

% --- Executes during object creation, after setting all properties.
function listBox1_CreateFcn(hObject, eventdata, handles)
% hObject handle to listBox1 (see GCBO)
% eventdata reserved - to be defined in a future version of MATLAB
% handles empty - handles not created until after all CreateFcns
% called

% Hint: listBox controls usually have a white background on Windows.
% See ISPC and COMPUTER.
if ispc && isequal(get(hObject,'BackgroundColor'), get(0,'
defaultUicontrolBackgroundColor'))
set(hObject,'BackgroundColor','white');
end

% --- Executes on button press in run_button.
function run_button_Callback(hObject, eventdata, handles)
% hObject handle to run_button (see GCBO)
% eventdata reserved - to be defined in a future version of MATLAB
% handles structure with handles and user data (see GUIDATA)

% Hint: get(hObject,'Value') returns toggle state of run_button

%% connection to the OPC server
da = opcda('localhost', 'Fantoft.DSPICEOPCServer.DA.1')
connect(da)
da;
ns=getnamespace(da)
Items = serveritems(ns)
grp = addgroup(da)
set(grp, 'UpdateRate', 1)
```

```

%% objects to read
itm1=additem(grp, 'dspice.model.time');
itm2=additem(grp, 'section2.20_FC1001.measure');
itm3=additem(grp, 'section4.20_FC2001.measure');
itm4=additem(grp, 'section7.22_KA101A.flow');
itm5=additem(grp, 'section7.22_KA101B.flow');
itm6=additem(grp, 'section7.22_KA101C.flow');
itm7=additem(grp, 'section7.22_KA102A.flow');
itm8=additem(grp, 'section7.22_KA102B.flow');
itm9=additem(grp, 'section7.22_KA102C.flow');

%% objects to write to
itm10 = additem(grp, 'section4.20_FC2001.setp');

while get(hObject,'Value')

    %% reading values
    read(itm1)
    read(itm2)
    read(itm3)
    read(itm4)
    read(itm5)
    read(itm6)
    read(itm7)
    read(itm8)
    read(itm9)

    %% plot values
    axes(handles.axes1);
    hold on;
    plot(itm1.Value, itm4.Value, itm1.Value, itm5.Value, itm1.Value,
         itm6.Value);
    legend('A', 'B', 'C');
    %       axis([25569 25570 100000 220000]);
    title('Sales compressors [kg/h]');
    hold off;

    axes(handles.axes2);
    hold on;
    plot(itm1.Value, itm7.Value, itm1.Value, itm8.Value, itm1.Value,
         itm9.Value);
    legend('A', 'B', 'C');
    %       axis([25569 25570 150000 400000]);
    title('Booster compressors [kg/h]');
    hold off;

    %% writing values
    flow_input4=str2double(get(handles.flow_input4, 'String'));
    write(itm10, flow_input4);

    pause(0.1);

```

APPENDIX D. MATLAB

```
end
%% close connection to OPC
disconnect(da);
delete(da);
clear da grp;
%opcreset;

function flow_input2_Callback(hObject, eventdata, handles)
% hObject    handle to flow_input2 (see GCBO)
% eventdata  reserved - to be defined in a future version of MATLAB
% handles    structure with handles and user data (see GUIDATA)

% Hints: get(hObject,'String') returns contents of flow_input2 as text
%        str2double(get(hObject,'String')) returns contents of
%        flow_input2 as a double

% --- Executes during object creation, after setting all properties.
function flow_input2_CreateFcn(hObject, eventdata, handles)
% hObject    handle to flow_input2 (see GCBO)
% eventdata  reserved - to be defined in a future version of MATLAB
% handles    empty - handles not created until after all CreateFcns
%            called

% Hint: edit controls usually have a white background on Windows.
%       See ISPC and COMPUTER.
if ispc && isequal(get(hObject,'BackgroundColor'), get(0,'
defaultUicontrolBackgroundColor'))
    set(hObject,'BackgroundColor','white');
end

function flow_input4_Callback(hObject, eventdata, handles)
% hObject    handle to flow_input4 (see GCBO)
% eventdata  reserved - to be defined in a future version of MATLAB
% handles    structure with handles and user data (see GUIDATA)

% Hints: get(hObject,'String') returns contents of flow_input4 as text
%        str2double(get(hObject,'String')) returns contents of
%        flow_input4 as a double
flow_input4=str2double(get(hObject,'String'));
if isnan(flow_input4)
    set(hObject, 'String', 0);
    errordlg('Input must be a number','Error');
end

% % Save the new flow input value
% set(hObject, 'String', flow_input4);
```

```
% --- Executes during object creation, after setting all properties.
function flow_input4_CreateFcn(hObject, eventdata, handles)
% hObject    handle to flow_input4 (see GCBO)
% eventdata  reserved - to be defined in a future version of MATLAB
% handles    empty - handles not created until after all CreateFcns
    called

% Hint: edit controls usually have a white background on Windows.
%       See ISPC and COMPUTER.
if ispc && isequal(get(hObject,'BackgroundColor'), get(0,'
    defaultUicontrolBackgroundColor'))
    set(hObject,'BackgroundColor','white');
end
```

Appendix E

OPC Server configuration

Part of the OPC server configuration .xml file is shown below to give an insight into configuration procedure. Such code is generated based on the .xls file table E.1 and the whole configuration procedure takes place in the Microsoft Excel. Only small part of the configuration code is shown here cause the entire list would take several pages but entire code can be find on the attached CD.

OPC server not allows not write and read tags with the same name by default. So one must find another tag with same values, e.g. setp and last_setp. Sometimes an extra tag is not available so one can workaround such challenge by introducing a new tag name in configuration. Such new tag will show up in the server and will have corresponding value from an originally tag in D-spice. See table E.1, specially controller 22XIC0518 with a new tag SP_eu[1]_ext.

Table E.1: OPC Server Configuration

SERVER: section	module	tag	unit	type	D-SPICE: section	module	tag	type
DSPIECE_GROUP	name=input	direction=INPUT	sample=1					
section7	22XIC0004	SP_eu[1]	bar	VT_R8	section7	22XIC0004	SP_eu[1]	VT_R8
section7	22_SC0102	setp	RPM	VT_R8	section7	22_SC0102	setp	VT_R8
section7	22_SC0202	setp	RPM	VT_R8	section7	22_SC0202	setp	VT_R8
section7	22_SC0302	setp	RPM	VT_R8	section7	22_SC0302	setp	VT_R8
section7	22XIC0518	SP_eu[1]_ext		VT_R8	section7	22XIC0518	SP_eu[1]	VT_R8
section7	22XIC0518	PV_eu[1]		VT_R8	section7	22XIC0518	PV_eu[1]	VT_R8
section7	22_Speed2A	setp	RPM	VT_R8	section7	22_Speed2A	setp	VT_R8
section7	22_Speed2B	setp	RPM	VT_R8	section7	22_Speed2B	setp	VT_R8
section7	22_Speed2C	setp	RPM	VT_R8	section7	22_Speed2C	setp	VT_R8
DSPIECE_GROUP	name=output	direction=OUTPUT	sample=1					
section7	From_Extr_T100	flow	kg/h	VT_R8	section7	From_Extr_T100	flow	VT_R8
section7	From_Extr_T200	flow	kg/h	VT_R8	section7	From_Extr_T200	flow	VT_R8
section7	22XIC0004	PV_eu[1]	bar	VT_R8	section7	22XIC0004	PV_eu[1]	VT_R8
section7	22XIC0004	SP_eu[1]_ext	bar	VT_R8	section7	22XIC0004	SP_eu[1]	VT_R8
section7	22XIC0004	out_disp		VT_R8	section7	22XIC0004	out_disp	VT_R8
section7	22_SC0102	measure	RPM	VT_R8	section7	22_SC0102	measure	VT_R8
section7	22_SC0202	measure	RPM	VT_R8	section7	22_SC0202	measure	VT_R8
section7	22_SC0302	measure	RPM	VT_R8	section7	22_SC0302	measure	VT_R8
section7	22_SC0102	last_setp	RPM	VT_R8	section7	22_SC0102	last_setp	VT_R8
section7	22_SC0202	last_setp	RPM	VT_R8	section7	22_SC0202	last_setp	VT_R8
section7	22_SC0302	last_setp	RPM	VT_R8	section7	22_SC0302	last_setp	VT_R8
section7	22XIC0400	S		VT_R8	section7	22XIC0400	S	VT_R8
section7	22XIC0401	S		VT_R8	section7	22XIC0401	S	VT_R8
section7	22XIC0402	S		VT_R8	section7	22XIC0402	S	VT_R8
section7	22XIC0401	DEV		VT_R8	section7	22XIC0401	DEV	VT_R8
section7	22XIC0401	DEV		VT_R8	section7	22XIC0401	DEV	VT_R8
section7	22XIC0402	DEV		VT_R8	section7	22XIC0402	DEV	VT_R8
section7	22XIC0400	out_disp		VT_R8	section7	22XIC0400	out_disp	VT_R8
section7	22XIC0401	out_disp		VT_R8	section7	22XIC0401	out_disp	VT_R8
section7	22XIC0402	out_disp		VT_R8	section7	22XIC0402	out_disp	VT_R8
section7	22_KA101A	Psuct	bara	VT_R8	section7	22_KA101A	Psuct	VT_R8
section7	22_KA101A	Pdsch	bara	VT_R8	section7	22_KA101A	Pdsch	VT_R8
section7	22_KA101A	flow	kg/h	VT_R8	section7	22_KA101A	flow	VT_R8
section7	22_KA101B	Psuct	bara	VT_R8	section7	22_KA101B	Psuct	VT_R8
section7	22_KA101B	Pdsch	bara	VT_R8	section7	22_KA101B	Pdsch	VT_R8
section7	22_KA101B	flow	kg/h	VT_R8	section7	22_KA101B	flow	VT_R8
section7	22_KA101C	Psuct	bara	VT_R8	section7	22_KA101C	Psuct	VT_R8
section7	22_KA101C	Pdsch	bara	VT_R8	section7	22_KA101C	Pdsch	VT_R8
section7	22_KA101C	flow	kg/h	VT_R8	section7	22_KA101C	flow	VT_R8
section7	22_KA101A_T	Fuel.f	kg/h	VT_R8	section7	22_KA101A_T	Fuel.f	VT_R8
section7	22_KA101B_T	Fuel.f	kg/h	VT_R8	section7	22_KA101B_T	Fuel.f	VT_R8
section7	22_KA101C_T	Fuel.f	kg/h	VT_R8	section7	22_KA101C_T	Fuel.f	VT_R8

E.1 XML file for OPC Server

```

<?xml version="1.0"?>
<OPCSERVER>
  <NODE name="section7">
    <NODE name="22XIC0004">
      <TAG name="PV_eu[1]" type="VT_R8" />
      <TAG name="SP_eu[1]" type="VT_R8" />
      <TAG name="SP_eu[1]_ext" type="VT_R8" />
      <TAG name="out_disp" type="VT_R8" />
    </NODE>
    <NODE name="22_KA101A_T">
      <NODE name="Exhaust">
        <TAG name="f" type="VT_R8" />
        <TAG name="x[1]" type="VT_R8" />
      </NODE>
      <NODE name="Fuel">
        <TAG name="f" type="VT_R8" />
      </NODE>
      <TAG name="Power" type="VT_R8" />
    </NODE>
    <NODE name="22_KA101B">
      <TAG name="Pdsch" type="VT_R8" />
      <TAG name="Psuct" type="VT_R8" />
      <TAG name="flow" type="VT_R8" />
    </NODE>
    <NODE name="22_KA_102AM">
      <TAG name="mtr_pow" type="VT_R8" />
    </NODE>
    <NODE name="22_KA_102BM">
      <TAG name="mtr_pow" type="VT_R8" />
    </NODE>
    <NODE name="22_KA_102CM">
      <TAG name="mtr_pow" type="VT_R8" />
    </NODE>
    <NODE name="22_QSV0864C">
      <NODE name="strm_ot">
        <TAG name="m" type="VT_R8" />
      </NODE>
    </NODE>
  </NODE>
</OPCSERVER>

```

E.2 Definition file for D-Spice

Definition file for section 7(Section7.def) to run OPC Server properly. Similar code must be applied to all sections including the master definition file.

```
$trend 4 200 16 200 64 200
$alarms 0 0 0 0 0 2.5e+006 1000 2000000 100
$logging mins 1 0 0 0 200
hours 4 0 0 0 200
days 16 0 0 0 200
weeks 64 0 0 0 200
months 256 0 0 0 200
$snap 0 3600 30 1 1 ASCII WITHOUT_TRENDS MCL ERASE_SNAPS
$sounds "action=StopSound;" "action=PlaySound,_%_SET_SOUND"
",repeat;"

$dcs_defs = [ LookAhead_dcs="0" alarm_delay="5" alarms="OFF"
auto_save="OFF" dcs_enable="1" deadband_def="0" max_speed="100000"
op_mode="1" read_delay="0" upd_max_cnt="0" ]
$dcs_cmd_prop = [ def_react="0" file_tmo="0" loc_tmo="0"
report_lev="0" step_tmo="0" ]

$dcs_driver "OPC" "dspice_opc" attr = [ ForwardOutputs="1"
OTS_plugin="" allow_tag_resize="0" browse_machine=""
connect_cyc="10" da_ProgId="Fantoft.DSPICEOPCServer.DA.1"
da_clsId="{7CF26890-D2CD-47E9-AD57-729898FC8328}"
da_prog_desc="D-SPIICE OPC DA Server" init_dcs_out="1"
io_conn_cyclic="0" io_mode="0" machine_name="localhost"
mdl_time_ident="dspice.model.time" mdl_time_send="1"
ots_conn_cyclic="0" owner="Section7" serial_step="0"
unit_update="0" ]

$DCS_conf_file "C:/Fantoft/Karsto_DPS/dspice/model/
dcs/section7_OPcserver.dcs"
```


Appendix F

D-Spice

F.1 XML file for start-up configuration

Code below in xml format was created to load master model on a D-spice start up.

```
<?xml version="1.0" encoding="utf-8" standalone="yes"?>
<DSPICE xmlns="http://tempuri.org/test_project_dsp.xsd">
<PROJECT FILE="C:\Fantoft\Karsto_DPS\dspice\Karsto_Master_dsp.xml"
  VERSION="3">
<FANTOFT_ROOT DIR="C:\Fantoft\Karsto_DPS" />
<DSPICE_ROOT DIR="C:\Fantoft\Karsto_DPS\dspice" />
<DMMI_ROOT DIR="C:\Fantoft\Karsto_DPS\DMMI" />
<PROJECT_ROOT DIR="C:\Fantoft\Karsto_DPS\dspice" />
<SERVER NAME="DSPICE" EXE="C:\Fantoft\Karsto_DPS\dspice\
bin\dengine.exe"
LOAD="False" TOGETHER="False" MODE="0" MODEL="" PARAMETERS=""
CONDITIONS="" START="False" LOADHISTORY="False"
INSTANCE="0" DEF_MID="" DEFINITIONS="" ARGS="" CONSOLE="False"
LOADDEFS="False" /><CLIENT NAME="Master"
EXE="C:\Fantoft\Karsto_DPS\dmmi\bin\dmmi.exe"
INSTANCE="0" KERNEL="0" HOST=""
GRAPHIC="C:\Fantoft\Karsto_DPS\dspice\model\master.xdr"
APPSHEET="idcad.dap"
MODEL="C:\Fantoft\Karsto_DPS\dspice\model\master.mdl"
PARAMETERS="" CONDITIONS="" ARGS="" CONFIG=""
SPLASH="False" LOADMODEL="True" LOADTOGETHER="True" />
<OLGA HOME="" LICENSE="" SERVER="" SCPSS="" />
<THERMODYNAMICS LIQDEWPOINT="False" ONEPHASE="False"
  TWOPHASE="False" BUBBLEPOINT="False" SYSTEM="False" />
<DEBUG FLAG="" /><REGISTER TYPE="CurrentUser" />
</PROJECT></DSPICE>
```

F.2 Registry file

Necessary register keys to run Kårstø master model properly.

```
[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft]

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\D-SPICE]
"Directory"="C:\\Program Files\\D-SPICE"
"Version"="3.8.0.3"

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DCAD]

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DCAD\environment]
"DSPICE_MODEL"="C:\\Fantoft\\Karsto_DPS\\dspice\\model"
"GFX_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data\\gfxdata"
"XC_MENU"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data\\mmi_data\\gmenu"
"XC_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data\\xcdata"
"XC_SYMDIR"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\symbols;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\symbols;"
"HOME"="C:\\Fantoft\\Karsto_DPS\\dspice\\model"

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DMMI]

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DMMI\environment]
"DSPICE_MODEL"="C:\\Fantoft\\Karsto_DPS\\dspice\\model"
"DSPICE_HOME"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std"
"DSPICE_PROJECT"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\mmi_alarm;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\mmi_blk;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\mmi_sys;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\mmi_sys;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\mmi_blk;"
"GFX_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data\\gfxdata"
"DSPICE_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data"
"DSPICE_BIN"="C:\\Fantoft\\Karsto_DPS\\DMMI\\bin"
"XC_MENU"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data\\mmi_data\\gmenu"
"XC_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data\\xcdata"
"XC_SYMDIR"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\symbols;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\symbols;"
"HOME"="C:\\Fantoft\\Karsto_DPS\\dspice\\model"

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DMMI\settings]
"markers"="off"
"animation"="on"
"state_animation"="on"
"target_block"="on"
"target_conn"="on"
```

```
[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DSPICE]

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\DSPICE\environment]
"DSPICE_MODEL"="C:\\Fantoft\\Karsto_DPS\\dspice\\model"
"DSPICE_HOME"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std"
"DSPICE_PROJECT"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng
\\mmi_alarm;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\mmi_blk;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng\\mmi_sys;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\mmi_sys;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\mmi_blk;"
"GFX_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data
\\gfxdata"
"DSPICE_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data"
"DSPICE_SNAPS"="C:\\Fantoft\\Karsto_DPS\\dspice\\model
\\snapshots"
"XC_SYMDIR"="C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_eng
\\symbols;
C:\\Fantoft\\Karsto_DPS\\dspice\\mmi\\mmi_std\\symbols"

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\MCL]

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\MCL\environment]
"DSPICE_DATA"="C:\\Fantoft\\Karsto_DPS\\DMMI\\dspice_data"

[HKEY_LOCAL_MACHINE\SOFTWARE\Fantoft\Server output]
```

F.3 Operational points

File-> Save/load ancillary files: chose 'Parameters' and 'Conditions' files. OBS! When working with master model, operational points should be saved in the main directory. Then corresponding .prm/.val files for various sections will be saved in respectively directories.

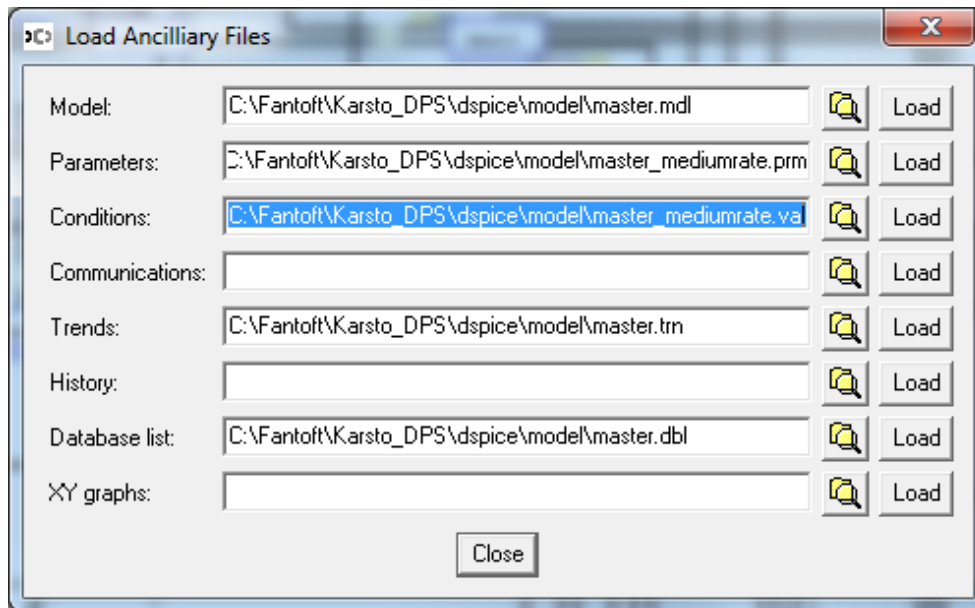


Figure F.1: Operational points

APPENDIX G. SEPTIC CONFIGURATION

```
Fulf= 1
HighPnlty= 5
LowPnlty= 5
HighLimit= 0.005
LowLimit= 0.005
RelxParam= 3 1 30 80
FulfReScale= 0.1
SetpTref= 0
BiasTfilt= 0
BiasTpred= 0
Constfilt= -1
Integ= 0
TransformType= NOTRANS
BadCntLim= 0
DesHorz= 0
Neval= 5
EvalDT= 0
KeepTargets= OFF
```

Bibliography

- Aske, E., Strand, S. & Skogestad, S. (2006). Coordinator mpc with focus on maximizing throughput, *Computer Aided Chemical Engineering* pp. 1203 – 1208.
- Bao-Cang, D. (2010). *Modern Predictive Control*, Boca Raton, Fla. : CRC Press.
- Downs, J. J. & Skogestad, S. (2011). An industrial and academic perspective on plantwide control, *Annual Reviews in Control* **35**(1): 99 – 110.
- Fantoft (2004). Full user guide.
- Golden, S., Fulton, S. A. & Hanson, D. W. (2002). Understanding centrifugal compressor performance in a connected process system, *Process Consulting Services Inc., Houston, Texas* .
- Gravdahl, J. T. & Egeland, O. (1999). *Compressor Surge and Rotating Stall*, Springer.
- <http://www.naturalgas.org> (2011). An educational website, Website. <http://www.naturalgas.org/overview/background.asp>.
- Maciejowski, J. M. (2002). *Predictive Control with Constraints*, Pearson Education Limited.
- Nisenfeld, A. E. (1982). *Centrifugal Compressors*, Monograph series/Instrument Society of America.
- Skogestad, S. (2004). Control structure design for complete chemical plants, *Computers and Chemical Engineering* **28**(1-2): 219–234.
- Skogestad, S. (2009). *Prosessteknikk*, Tapir Akademisk Forlag.
- Skogestad, S. & Postlethwaite, I. (2005). *Multivariable Feedback Control*, Wiley.
- Strand, S. & Sagli, J. (2004). Mpc in statoil - advantages with in-house technology, *AD-CHEM 2003*, p. 97–103. International Symposium on Advanced Control of Chemical Processes.

www.engineeringtoolbox.com (2011). Resources, tools and basic information for engineering and design of technical applications, Website. <http://www.engineeringtoolbox.com>.

www.instrumentation.co.za (2002). Advanced compressor control, Website. <http://www.instrumentation.co.za/article.aspx?pk1articleid=1801>.

www.opcfoundation.org (2012). Opc foundation, Website. <http://www.opcfoundation.org>.