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Advanced Control Structures for Balancing Supply and Demand in Steam Distribution Networks

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Abstract

This master thesis proposes a supervisory control structure based on PID controllers for balancing supply and demand in a steam distribution network. It is then compared to a model predictive control for performance comparison. The steam distribution network comprises six high-pressure supplier pipelines, directed through a main pressure pipeline diverted to six medium pressure consumer pipelines. For modeling, the programming software MATLAB with Simulink and the nonlinear optimization software CasADi were used.

The supervisory controller structure is used for controlling maintaining network pressure. It uses the idea of a time scale separated control system involving primary (fast), secondary (slower), tertiary (slow), and quaternary (slowest) control. The primary acting control consists of proportional pressure control on the network producer side. In the secondary acting control, proportional integral pressure control is used on the producer side. The tertiary control uses either parallel control, controllers with different setpoints, or valve position control on the producer side. Last, the quaternary control uses proportional integral control on the consumer side when primary, secondary and tertiary controllers are saturated.

Two remaining consumer types are proposed. That is the involuntary droop consumers who offer no control but are self-regulating due to network coupling and normal consumers, being single loop integral valve position controllers. Both involuntary droop and normal consumers act as a system disturbance.

The decentralized controller structure using controllers with a different setpoint for the tertiary control structure showed the most promising results regarding stability and performance. The valve position configuration showed promising results, however, responding slower than the different setpoints structure. The model predictive controller showed good initial results but had problems returning to nominal positions again. The parallel control configuration did not help control the modeled steam network as it did not fully utilize the tertiary controllers.

Sammendrag

Denne masteroppgaven presenterer en overvåkende kontrollstruktur basert på PID-kontrollere, for å balansere tilbud og etterspørsel i et dampnettverk. Deretter blir prestasjonen sammenliknet med en modellprediktiv kontroller. Dampnettverket består av seks høytrykksleverandører, som ledes gjennom et middels trykksatt hovedrør og deretter ut gjennom seks middels trykksatte forbrukerrør. For modellering av nettverket, ble programvaren MATLAB med Simulink og den ikkelinære programvaren CasADi brukt.

Den overvåkende kontrollstrukturen brukes til å kontrollere nettverkstrykket. Den bruker ideen om et tidsseparert kontrollsystem som involverer primær (rask), sekundær (tregere), tertiær (treg) og kvaternær (tregst) kontroll. Primær kontroll består av proporsjonal trykkontroll på nettverkets leverandørside. I sekundær kontroll blir proporsjonal-integral trykkontroll på nettverkets leverandørside brukt. Tertiær kontroll bruker enten parallell kontroll, kontrollere med forskjellige settpunkter eller ventilposisjonkontroll på leverandørside. Til sist bruker kvaternær kontroll proporsjonal-integral trykkontroll på forbrukersiden når primær, sekundær og tertiær kontroll er mettet.

To karakteristiske forbrukere uten kontroll presenteres også. Det er ufrivillige droop-forbrukere som ikke har noen kontroll involvert, men blir selv-regulert på grunn av nettverkskobling. Normale forbrukere er den andre typen, bestående av enkeltsløyfede integralkontrollere for ventilposisjon. Begge forbrukerne opptrer som systemforstyrrelser.

Den desentraliserte kontrollstrukturen med kontrollere med forskjellig settpunkt for den tertiære kontrollstrukturen viste de mest lovende resultatene for stabilitet og opptreden. Ventilposisjonkontroll viste også lovende resultater, men responderte saktere enn kontrollere med forskjellig settpunkt. Den modellprediktive kontrolleren viste gode intiell respons, men hadde problem med å returnere til normale kontrollerposisjoner. Parallell kontroll hjalp ikke for å kontrollere dampnettverket da det ikke klarte å fullt utnytte de tertiære kontrollerne.

Preface

I would like to thank my supervisor Sigurd Skogestad and co-supervisor Cristina Zotica for guiding me through this master thesis. I would like to thank Cristina in particular for helping me get through the entire process, providing much needed help and input to both my master thesis and my specialization project last semester.

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Abbreviations

The list presented in this section describes all abbreviations and acronyms that could be used within the body of this master thesis:

AE Algebraic equation

CV Controlled variable

DAE Differential algebraic equation

DCS Decentralized control structure

DOF Degrees of freedom

DV Disturbance variable

HP High pressure

LP Low pressure

MIMO Multiple input multiple output

MISO Multiple input single output

MPC Model predictive controller

MV Manipulated variable

ODE Ordinary differential equation

PID Proportional-Integrating-Derivative

SIMC Simple internal model control

SIMO Single input multiple output

SISO Single input single output

VPC Valve position control

Nomenclature

Table 1 presented in this section describes all nomenclature used within the body of this master thesis:

Table 1: Most important system parameters and descriptions.

Variable	Description
T	System Temperature
R	Universal Gas Constant
V	System Volume
P_{Prod}	Nominal Inlet Pressure
P^{nom}	Nominal Main Pressure
P_{Cons}	Nominal Consumer Pressure
$q_{SP,j}^{nom}$	Nominal Flow Swing Producer
$q_{DP,k}^{nom}$	Nominal Flow Droop Producers
$q_{EP,l}^{nom}$	Nominal Flow Extra Producers
$q_{SC,m}^{nom}$	Nominal Flow Swing Consumer
$q_{DC,n}^{nom}$	Nominal Flow Involuntary Droop Consumers
$q_{NC,p}^{nom}$	Nominal Flow Normal Consumers
$Cv_{SP,j}$	Valve Coefficent. Swing Producer
$Cv_{DP,k}$	Valve Coefficient Droop Producers
$Cv_{EP,l}$	Valve Coefficient Extra Producers
$Cv_{SC,m}$	Valve Coefficient Swing Consumer
$Cv_{DC,n}$	Valve Coefficient Involuntary Droop Consumers
$Cv_{NC,p}$	Valve Coefficient Normal Consumers
$z_{SP,j}^{nom}$	Nominal Valve Position Swing Producer
$z_{DP,k}^{nom}$	Nominal Valve Position Droop Producers
$z_{EP,l}^{nom}$	Nominal Valve Position Extra Producers
$z_{SC,m}^{nom}$	Nominal Valve Position Swing Consumer
$z_{DC,n}^{nom}$	Nominal Valve Position Involuntary Droop Consumers
$z_{NC,p}^{nom}$	Nominal Valve Position Normal Consumers
$\tau_{c,flow}$	Time Constant Flow Controllers
$\tau_{c,DP}$	Time Constant Pressure Controllers
$\tau_{c,SP}$	Time Constant Swing Producer Controllers
$\tau_{c,EP}$	Time Constant Extra Producer Controller
$\tau_{c,DC}$	Time Constant Swing Consumer Controller
SP	Number of Swing Producers
DP	Number of Droop Producers
EP	Number of Extra Producers
SC	Number of Swing Consumers
DC	Number of Involuntary Droop Consumers
NC	Number of Normal Consumers

Chapter 1

Introduction

This chapter introduces steam distribution networks and provides a brief presentation of a simplified steam distribution network. Then, control issues and challenges within steam networks are presented. Based on that, a motivation for implementing advanced control in steam distribution networks is stated, followed by the scope and outline of this thesis.

1.1 Steam Distribution Networks

Steam distribution networks were traditionally used for heating purposes and as an industry tool until the 1800s. Now, several other uses for steam networks are utilized, such as power production[3]. Today's steam distribution networks in industrial plants typically include heat, electricity, and mechanical power supply. Heat requirements usually come from heat exchanger demand. In contrast, electrical and mechanical demands usually arrive from power requirements in process unit operations such as distillation columns, compressors, pumps, and more[4].

Looking at how a steam network operates, it typically starts on a producer side with boilers producing steam at a high-pressure level. The boilers are referred to as producer $1, 2, ..., m$ in Figure 1.1. Multiple suppliers containing high-pressure steam meets in a common pipeline then divides to medium pressure levels on what is referred to as the consumer side. From the consumer side, steam is used by process equipment. Process equipment is referred to as consumer $1, 2, ..., n$ in Figure 1.1. Altogether the different presented pressure levels define as a steam distribution network. A steam distribution network usually only includes a few producers but could involve hundreds or thousands of consumers[5] [6].

On the producer side, the steam-producing boilers usually include the possibility to decrease or increase steam supply to ensure stable production, as steam distribution networks are subject to disturbances [5] [6]. The consumer side is possibly the most susceptible for disturbances as it is used to supply consumer demand from different consumers such as process equipment. These demands usually operate at a varying load, creating the disturbance [4]. In a worst-case situation, consumer demand could cause a costly situation with a lack of steam supply to the network, reducing plant production capacity. Control implementation could reduce the chances of this happening by possibly extending the plant operating range.

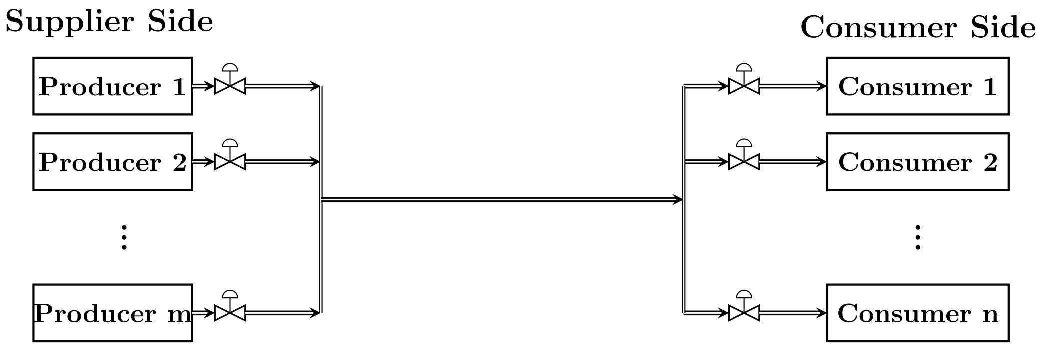

Figure 1.1: Simplified process flowsheet of a steam network with a high-pressure supplier side with suppliers 1,2,...,m moving through a medium pressure main pipeline to medium pressure consumer side with consumers 1,2,..,n.

1.2 Steam Distribution Network Costs

If general steam distribution network operating costs are high, they should also motivate implementing process control. The problem with generalizing steam network costs is that they are highly individualized and depend on the type of operation, equipment involved, size, and many more. It is, therefore, more suitable to look at relative steam distribution network costs in process plants.

Usually, a plant's most significant fixed investment cost using steam networks is the steam generation and distribution. Steam generation usually stands for $2.6 - 6.0\%$ of the total fixed-capital investment in a process plant, while steam distribution typically ranges somewhere between $0.2 - 2.0\%$ of the total fixed-capital cost. This includes planned maintenance which could be reduced if steam use is lessened [7].

Looking at a product cost in a typical plant, utility costs usually make up $5 - 10\%$ of the total product cost [8]. Utilities include costs for steam generation and distribution, cooling water, refrigeration, fuels, waste treatment, and more. Steam costs are usually the most expensive utility [7] [8]. With increasing energy prices and environmental concerns, there is naturally an industrial desire to reduce steam generation and distribution costs. Implementing process control in a steam generation network could optimize the use of process inputs, that is, steam generation and distribution, and thus reduce the use of steam. This could reduce maintenance requirements and product costs, possibly making plants more competitive. Process control could also improve the quality of the steam, that is supply it at more predictable conditions in terms of pressure and temperature, which would be more attractive for steam consumers.

1.3 Motivation, Scope and Earlier Work

Steam networks should be operated stably and efficiently to assure that they do not become too large a disturbance for the consumers in the downstream processes. Therefore, the scope of this work is to propose an advanced control structure based on decentralized controller structures that can balance and supply-demand in a steam network, focusing solely on the transition from the producer side valves to the consumer side valves in Figure 1.1, not the actual producers and consumers. Also, to possibly extend the operating range, a list of priorities is proposed in Chapter 4 to allocate the steam load to a set of suppliers and to a subset of consumers. In the end, the performance of the decentralized control structure will be compared to that of a model predictive controller.

The idea of the decentralized control structure would resemble frequency control in power grids where different power generators on the supplier side control the frequency at different time scales. That is primary (fast), secondary (slower), and tertiary (slowest) control [9] [10]. The difference from frequency control in the decentralized control structure proposed in this thesis would also be to make use of the consumers when the suppliers are at maximum capacity. This is currently not practiced in real electrical grids, though literature offers examples of demand (consumer) response. Examples include: Short and Infield (2010), which propose demand-side management of domestic electricity loads with emphasis on the UK [11]. Houwing et al. (2011) proposes an intelligent, price-based control concept based on demand response [12]. Garcia et al. (2011) shows that demand-side control could be successfully added to frequency control using decentralized control [13].

As it is proven for performance, model predictive control is implemented in comparison to the decentralized control structure. Gopalakrishnan and Biegler (2013) successfully implements nonlinear model predictive control to optimize operational costs in a gas pipeline network by minimizing compressor operating cost at the producer side as the controller objective[14]. Zhu et. al. (2001) implements a linear model predictive control strategy to a large-scale gas pipeline network, using a linearized plant model. They use both input and output variables subject to operational constraints, and shows that the chosen control implementation significantly improves operability[15].

In this work, a steam distribution network is analyzed. However, the control structures implementing optimal operation would possibly not limit to steam distribution networks. The United Kingdom grid system for home gas supply involves a high-pressure gas network supplier distributing to thousands of medium pressure splits, leading to household consumers[16]. Other forms of distribution networks other than gas and steam also exist. Examples include sewage treatment networks, water supply networks, and others[17].

1.4 Outline

The outline of this thesis will be such that Chapter 2 starts with a review of the most significant literature for the modeling and process control implementation possibilities of a steam distribution network. In Chapter 3, the steam distribution network model including the open loop model is proposed under study . Then, in Chapter 4, the implementation of advanced control to the model is presented, both regarding PID-based control and model predictive control. In Chapter 5, results from the implementation of both controller structures, including tuning results are presented. Then, Chapter 6 presents the closed-loop results. Chapter 7 discusses the performance of the individual control structures and other implementation issues. In Chapter 8, results and discussions are concluded and recommended future works are proposed. In the end, the appendices present the simulations excluded from the tuning and result chapters.

This thesis continues the project work presented by the undersigned in 2020[18]. Some of the sections in the theory part (2.1 through 2.4) in this thesis are therefore based on the work performed in 2020. This thesis, therefore, only projects minor changes from the project work in these sections.

Chapter 2

Theory

This chapter presents the most essential and relevant control theory for implementing process control in a steam distribution network. It first offers process control objectives, then the objectives and issues of regulatory and supervisory control, including decentralized and multivariable control ideas being stated.

2.1 Process Control

Chemical processes often require precise control of measured process data, such as flow rates, pressures, or valves, to keep the system at steady-state conditions. Process control is defined as the methods applied to control such processes in a process plant [19].

The control of a chemical process is usually divided into a hierarchy like the one seen in Figure 2.1. The control hierarchy is separated on a time scale basis into different control objectives. Those are longer-term economic optimization and shorter-term stability objectives. The upper, longer-term control layers receive measurements from the lower, shorter-term control layers to achieve these control objectives. The measurements are used in the upper layers to solve an optimization problem to change the setpoints in the lower control layers for optimal chemical process operation [1].

At the bottom of the hierarchy is the process with physical process units, only measuring and receiving setpoints from the higher control layers. In this thesis, only the control layer, including supervisory and regulatory control, which optimize and directly change the setpoints in the physical process units, will be considered.

The control layer outputs change the manipulated variables (MVs) or inputs u in the plant to keep the systems controlled variables (CVs) or outputs y at given setpoints.

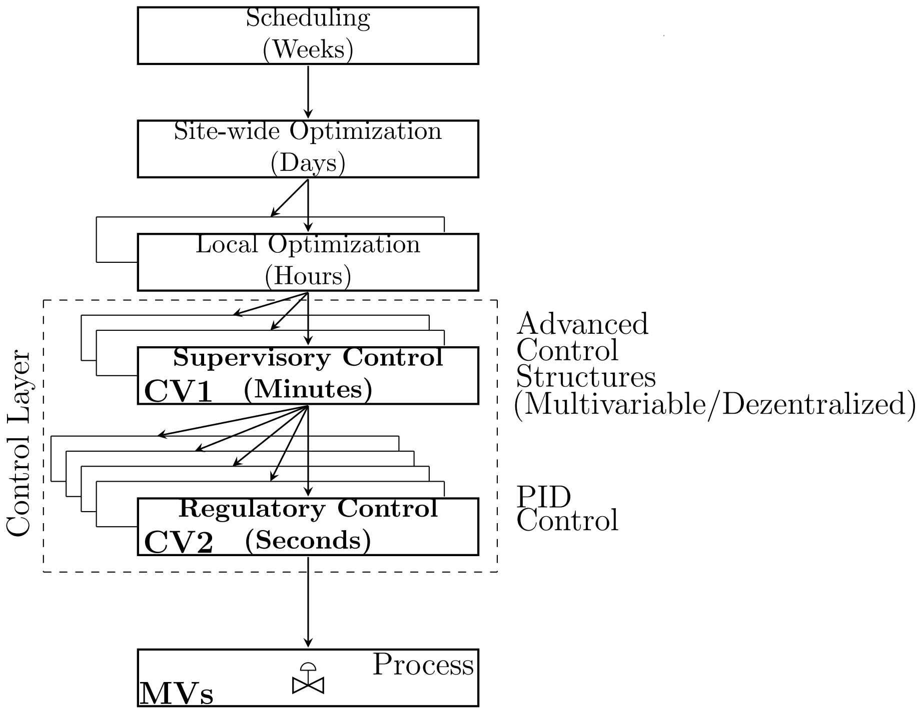

Figure 2.1: Typical control hierarchy in a process plant[1].

The manipulated variables are thus the physical process units used for system control. In this thesis, all manipulated variables are valve actuators. The controlled variables are the non-physical variables such as pressures, flows, and temperatures. It is usually desired to keep the process outputs at a constant setpoint to stabilize a process.

In the control layer, the CVs are usually divided into primary (CV1) and secondary controlled variables (CV2). This is because the supervisory control layer performs economic optimization using the most economically important, primary controlled variables. The regulatory layer performs shorter-term stability optimization, which the secondary controlled variables perform, selected for their stabilizing effect. That is, the upper supervisory control layer uses primary controlled variables to solve an optimization problem and change the setpoints for the secondary controlled variables in the lower regulatory control layer.

The following sections explain the regulatory and supervisory control layers in further detail, including theory and concepts.

2.2 Regulatory Control

As mentioned, regulatory control is the direct-acting control layer to the plant inputs u, as seen in Figure 2.1. The main objective of the regulatory control layer is to stabilize the process, avoiding drifting from a desired steady-state setpoint, and fast time-scale disturbance rejection. This is obtained using secondary controlled variables, selected for their stabilizing effect [20].

Implementing regulatory control to a system makes it a closed-loop system. Possibly the most common closed-loop system is referred to as a feedback system, shown in Figure 2.2. Feedback systems are based on measuring the process output y. Then the process error e is calculated from the output compared to a reference state y^s, yielding $e = y^s - y$. The system then tries to compensate to make the output equal to the reference input state. That is done using a controller, which sends the input u to the system. Using feedback control yields the advantage that the controller structure is error-driven, and no disturbance model is required.

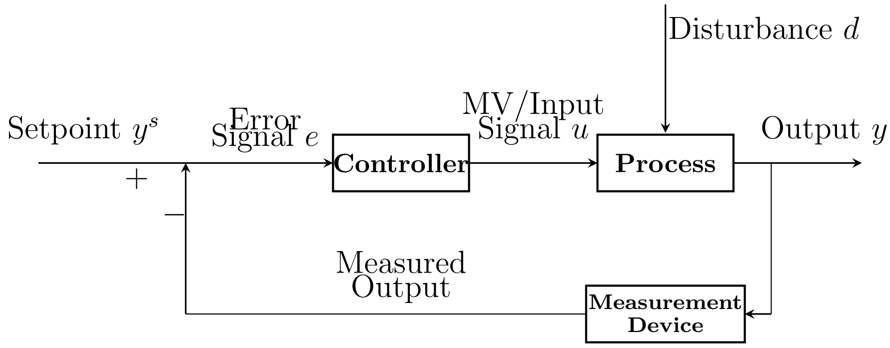

Figure 2.2: Closed-loop feedback control structure.

For processes where there are unknown disturbances, feedback would therefore be the obvious choice. The disadvantage for feedback control would be that the process only can take corrective measures once the disturbance is detected in the process. Therefore, this controller structure could be ineffective for processes with significant dead-time between the measured output variable and the input or manipulated variable [19].

2.3 Supervisory Control

While regulatory control itself should ensure plant stability, supervisory control should be added to ensure economic operation, prioritizing maintaining the setpoints for the primary controlled variables. For most processes, the supervisory control layer is therefore required for optimal process control.

The main objective of the supervisory control layer is to ensure that the primary controlled variables (CV1) are kept at optimal setpoints using degrees of freedom (DOF) in the system. This ensures keeping the primary controlled variables at optimal setpoint and if used correctly, could extend plant operating range. By doing this, the supervisory control layer also makes effective use of extra available inputs and measurements in the plant, making sure the plant is using all available resources for optimal operation.

As mentioned, the supervisory layer also changes the setpoints for the secondary controlled variables (CV2) in the regulatory layer. The reason the setpoints in the regulatory layer are changed is to keep the manipulated variables in the regulatory layer from saturating, ensuring more stable plant operation. This is because regulatory control is mainly used to stabilize operation, and if the manipulated variables in regulatory control layer are saturated, they are unable to maintain their related controlled variables at setpoint and the process could go unstable, potentially affecting optimal operation. Therefore, another important role of the supervisory control layer is avoiding saturation in the stabilizing, regulatory control layer.

Another important role of the supervisory control layer regards setpoint tracking. An important part of setpoint tracking involves switching between active constraints in the primary controlled variables. This is often required as the optimal controlled variables may change during operation because of changes in active constraints caused by disturbances. The active constraint changes can happen on the manipulated variables or on the controlled variables. Thus, there are three possible ways of constraint switching, that is: CV-CV switching, changing from one controlled variable to another, MV-MV switching, changing from one manipulated variable to another and the combination of the two, CV-MV switching.

Constraint switching is performed in practice by changing control objectives when constraints become active or inactive. That is, switching either between manipulated or controlled variable constraints when they become active or inactive. This is done using either single-loop classical advanced control, also known as decentralized control, or a multivariable control structures such as model predictive control. Decentralized and centralized control should in addition to constraint switching also implement the other objectives presented as objectives for the supervisory control layer. The next sections will emphasize the theory behind important concepts regarding decentralized and multivariable control structures.

2.4 Proportional-Integral-Derivative Control

In order to achieve satisfactory decentralized control, sufficient controller algorithms and tunings must be implemented. The PID-controller algorithm is one of the most commonly used controller algorithms in the industry today. It consists of three individual controller algorithm concepts, that is a proportional, an integral, and a derivative link which are summed together[19]. The algorithm is often written using the joint proportional gain K_c as shown in Equation 2.1.

$$u(t) = K_c\left(e(t) + \frac{1}{\tau_i}\int_0^t e(\tau)\mathrm{d}\tau + \tau_d\frac{\mathrm{d}e(t)}{\mathrm{d}t}\right) \tag{2.1}$$

In Equation 2.1, e is the controller error from the measured output y subtracted from the reference input y^s in Figure 2.2. K_c is the joint controller proportional gain, determining how much amplification each controller action part should receive. The last algorithm parameters, the integral time τ_i and derivative time τ_d represent first and second-order time delays in process open-loop step responses.

To obtain the tuning parameters, this thesis will use the an analytically derived, performance proven controller tuning method named simple internal model control (SIMC)[21].

The method involves only one tuning parameter, the desired first-order closed-loop time constant, τ_c. For processes requiring tight control, τ_c is often set equal to the process time delay θ. When needing more robust control, τ_c should be given a larger value than θ. In this thesis, τ_c is selected based on how fast the controllers react to changes in the system.

The other tuning parameters are decided by analytical derivation based on input step changes in the system. This thesis only involves two types of responses. Those are the responses of a first-order process and a static process. From these responses, information about the simple internal model control plant gain, k, the dominant time constant τ_1, and the effective time delay θ, can be obtained. These are tuning parameters for the proportional controller gain K_c. The simple internal model control tuning rule for K_c for a first-order process is found in Equation 2.2[21].

$$K_c = \frac{1}{k} \cdot \frac{\tau_1}{\tau_c + \theta} \tag{2.2}$$

The variables used in Equation 2.2 can be found using Figure 2.3. The effective time delay θ is the time from a step change on the input until an actual process response. The time constant τ_1 is defined as the time it takes for the response to reach 63 % of its new steady-state value. Plant gain k is defined as the change in the open-loop response $\Delta y(\infty)$ divided by the input step change Δu.

To add integral action, an integral time constant must be decided. The simple internal

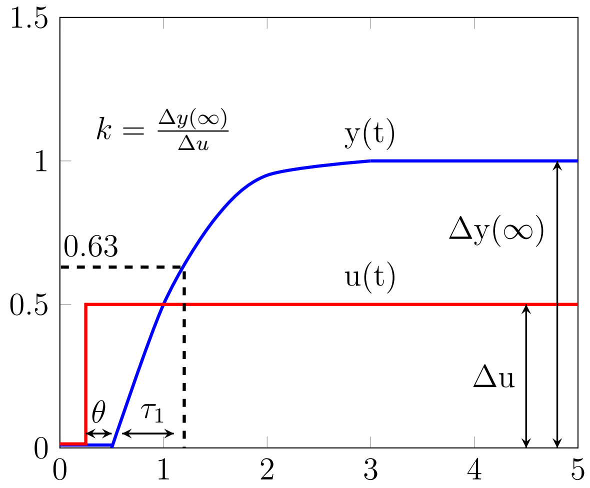

Figure 2.3: SIMC tuning parameters from open-loop response for a first order process.

model control rules state that integral time for tight control should be chosen such that τ_i is chosen as the minimum between the first-order time constant, τ_1, and $4 \cdot (\tau_c + \theta)$. This is expressed as a minimum function, shown in Equation 2.3[22].

$$\tau_i = min[\tau_1, 4 \cdot (\tau_c + \theta)] \tag{2.3}$$

The tuning rules for a first-order process are derived from its transfer function on a discrete-time domain. For a first-order process, the time delay process, the process transfer function $g(s)$ shown in Equation 2.4 is used to obtain the simple internal model control tuning parameters in a continuous-time domain[23].

$$g(s) = k\frac{e^{-\theta s}}{\tau_1 s + 1} \tag{2.4}$$

In Equation 2.4, k represents the plant gain, θ represents the effective time delay, τ_1 is the dominant lag time constant and s is Laplace-transformed variable from a continuous time domain where it would be denoted as t.

For a static process, as shown in Figure 2.4, the response is equal in shape to the one of $u(t)$, only varying in time delay θ and magnitude. Thus, τ_1 from the proportional gain K_c will approach zero, making K_c zero, shown in Equation 2.5.

$$K_c = \frac{1}{k} \cdot \frac{\tau_1}{\tau_c + \theta} \xrightarrow{\tau_1 \to 0} 0 \tag{2.5}$$

Having the proportional gain K_c approach zero means there is a need to derive a new

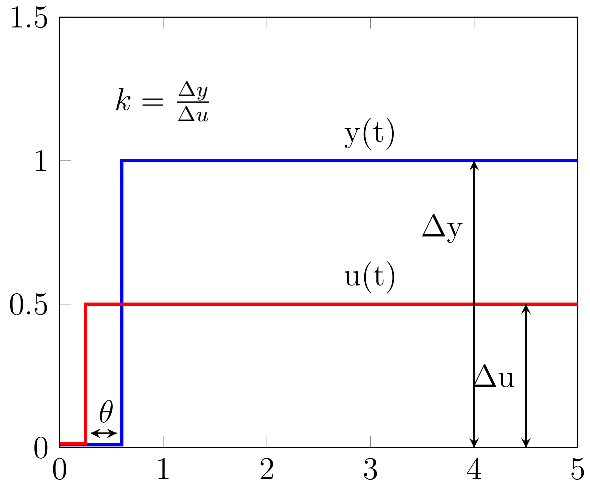

Figure 2.4: SIMC tuning parameters from open-loop response for a static process.

relation for the integral gain K_i independent of the proportional gain K_c. The integral gain for a static process is derived in Equation 2.6:

$$K_i = \frac{K_c}{\tau_i} = \frac{1}{k} \cdot \frac{\tau_1}{\tau_c + \theta} \cdot \frac{1}{\tau_1} = \frac{1}{k} \cdot \frac{1}{\tau_c + \theta} \tag{2.6}$$

The parameters for the integral gain K_i is obtained as shown in Figure 2.4. For k, the relationship between the magnitudes of $\Delta y(t)$ and $\Delta u(t)$ should be used. The remaining parameters are described previously.

Looking at the process transfer function $g(s)$ in the discrete-time domain for the static process yields the expression of that in Equation 2.7.

$$g(s) = ke^{-\theta s} \tag{2.7}$$

In Equation 2.7, θ, k and s refers to the same process parameters as in Equation 2.4 for a first order process. The difference for a pure time delay process the time constant is zero, simplifying the process transfer function $g(s)$ for the process.

Derivative action could also be required for the mentioned processes responses, when measured data is bad. Since this is not the case for a modeled scenario in this thesis, it is not relevant to introduce here.

When PID-controller output is not equal to the plant input, integral action will cause error accumulation. This would happen when the controller output signal is altered before it is inputted to the plant. Examples where this would happen, would include using selectors, meaning a logic block is outputting one of the multiple input signals, or in the case of valve saturation, which occurs when a valve reaches its physical limits, fully closed or fully opened.

Methods to avoid windup are commonly denoted as anti-windup. One anti-windup method is back-calculation. The back-calculation concept is shown in Figure 2.5.

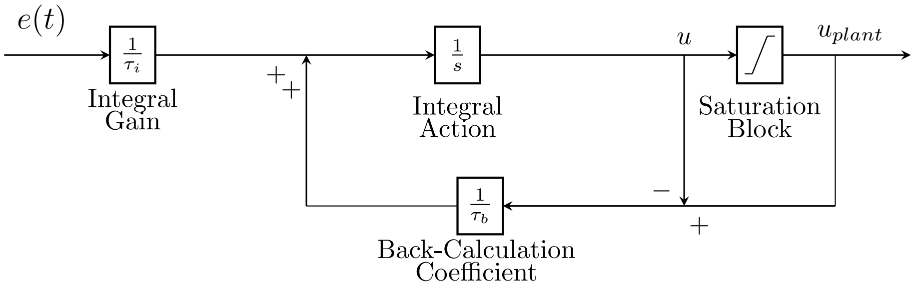

Figure 2.5: Illustration of the back-calculating anti-windup concept in Simulink.

In Figure 2.5, a new variable, the back-calculation coefficient τ_b is introduced. Usually, τ_b is selected such that is is equal to $\tau_b = 1/\tau_i$. Also note that in Figure 2.5, windup is caused by a saturation block. The saturation block could be exchanged with any other plant input altering block, such as a selector, and the back-calculating principle will be the same, using controller output u and the modified plant input u_{plant} to correct the controller signal.

2.5 Decentralized Control

Decentralized control is a form of supervisory control using single-loop classical advanced control structures. Advanced controller structures recognized by being based on simple, single loop controller elements, not involving any centralized controller, but using already existing controller structures. Thus, it is usually simple to implement decentralized control on top of an existing plant with only regulatory control implemented [24].

There are many different simple controller elements of supervisory control. Classical examples include cascade control, a common implementation, where the idea is to measure and control an internal variable y_2, as well as feedforward control, where the idea is to measure disturbance. However, none of these methods care for active constraint changes, which was mentioned as one of the objectives of the supervisory control layer.

For situations where active constraint switches would happen, many decentralized control structures have been proposed, according to which constraint switch is happening.

Reyes-Luá et al. (2018) propose that for CV-CV constraint switching, selectors should be used. Selectors involve the use of a minimum, mid, or maximum selector for a process with many controlled variables and a single manipulated variable in the process. For MV-MV switching, split range control, controllers with different setpoints, or input (valve) position control should be used. Split range control involves a controller sending an internal signal to a split range block. If the internal signal moves below or above a given split value, the input is switched, while the other inputs are fixed at limiting values. Controllers with different setpoints involves two controllers with the same output, using different setpoints to only actively use one input at a time, like split range control. Valve position control involves using the manipulated variable, the controller output to regulate the controlled variable, which is used as the input for the next controller. Usually for valve position control both valves are working at the same time[1].

In the last case, CV-MV constraint switching, no control structure is required, that is if the input saturation pairing rule is followed. The input saturation pairing rule is defined as when more than one input have an effect on a single controlled variable, a more important controlled variable should be paired with the input that is not likely to saturate[25]. If this is not the case, a MV-MV scheme should be used with a selector, taking over control when the main manipulated variable saturates[1].

In this thesis, the decentralized control structure is either based on or directly using cascade control, selectors, different setpoints and valve position control, each to be presented in detail in the next section, along with the control structures using them as a basis.

2.5.1 Cascade Control

Cascade control uses the output of an outer control loop as the setpoint for an inner control loop. It is often advantageous to add cascade control, as it has some linearizing effect, which could be ideal for a nonlinear process. Also, cascade is advantageous if there is possibility to use the inner loop controlled variable y_2 to faster reject the disturbance d_2 than using only the outer loop controlled variable y_1, shown in Figure 2.6. If a disturbance on the inner loop is not present, cascade control is usually not more efficient than not using a cascade loop[24].

A cascade loop containing only two controllers is often referred to as a single cascade. However, even more, controller loops can be added on top of the cascade loop. The cascade loop can be in parallel or a series configuration[24].

The single series cascade principle is illustrated in Figure 2.6 [24]. Here, the inner control loop is recognized by the first controller, c_1, and the plant g_1. The inner loop output signal is inputted to the plant for the outer loop, giving the desired output for y_1. The setpoint for the controller in the inner loop is y_2^s, coming from the controller output in the outer loop, that is, controller c_2. The setpoint for c_2 is y_1^s. The cascade principle will extend the system's operating range by using multiple inputs to control y_1.

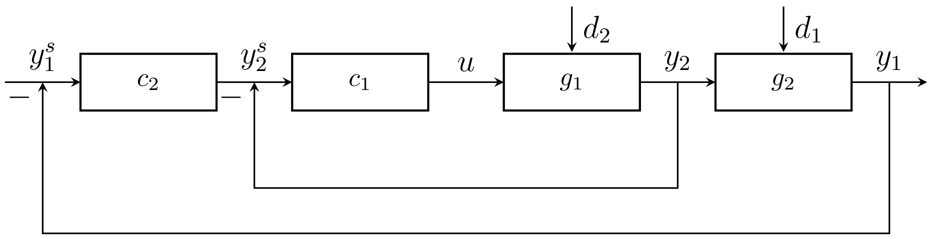

Figure 2.6: Block diagram for series cascade controller structure.

Figure 2.7 illustrates the parallel cascade principle. The difference from a series cascade is that there is only one process g outputting both y_1 and y_2. The principle still evolves around a fast, inner loop with controller c_1, and a slower, outer loop with controller c_2. This configuration does not have individual disturbances, meaning a process disturbance will affect both y_1 and y_2. This also means that the type of disturbance does not matter as it would for the series cascade. In the parallel cascade, it is more important that y_1 is closely related to y_2 [24]. This would be the case for a flow-pressure network coupled around a main pipeline pressure.

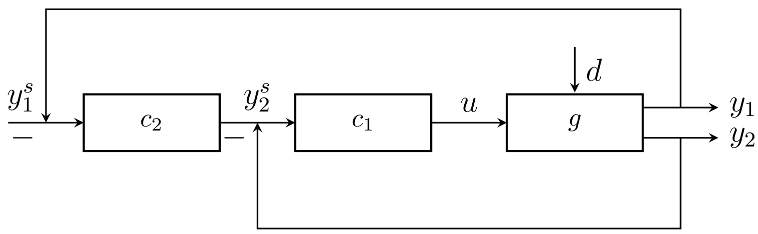

Figure 2.7: Block diagram for parallel cascade controller structure.

A common cascade problem is an interaction between the inner and outer loop. To avoid interactions between the inner and the outer loop, cascade control is based on a slow "master" controller in the outer loop and a fast "slave" controller in the inner loop. In this thesis, the analogy for that would be flow control as a quick inner loop and pressure control as a slow outer loop, controlling the flow setpoints. To ensure that interactions are minimal in the cascade loop, the outer loop time constant τ_c should increase at least five times per loop increment [26].

2.5.2 Parallel Control

Parallel control implies two controllers acting simultaneously. There are two options to implement parallel control. That is through the use of valve position control, and using one PI-controller and one P-controller. Used correctly, they can give the same performance.

Valve position control (VPC), also known as input resetting or mid-ranging control, uses switching between multiple input variables to control a single controlled variable, known as MV-MV switching. The idea of valve position control is to use a primary manipulated variable that should not saturate and is desired to keep at setpoint. The rest of the manipulated variables become secondary variables [1].

The configuration for valve position control would involve two inputs and one output. Here, both inputs u_1 and u_2 are used at the same time using their respective controllers c_1 and c_2. In this case, valve position control could improve control of y if the response from the input u_1 is slow compared to u_2.

Shown in Figure 2.8, the input u_2 is used directly to control y, while u_1 is used to return u_2 to steady-state nominal conditions. This works well to avoid saturation in u_2, especially if u_1 is a bigger, slower input compared to u_2, for example if the inputs are valve openings. From a process perspective this would mean u_1 is the main flow supply, while u_2 is a more fine-tuned and measured flow, to ensure greater accuracy in the output y. An industrial mixing process where output accuracy is important could be an example where this use would be relevant [1].

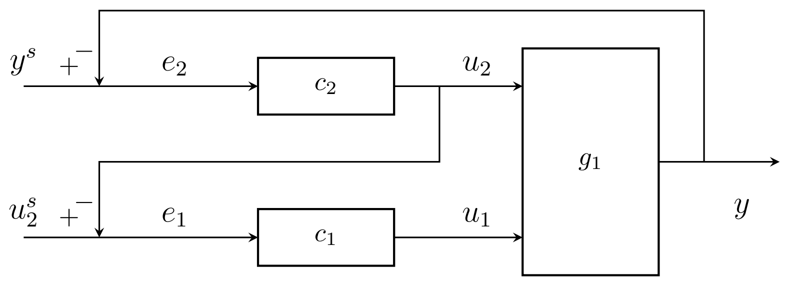

Figure 2.8: Block diagram of a system with two manipulated variables using valve position control to maintain one controlled variable.

The configuration in Figure 2.8 could also be set up such that the use of u_2 only occurs when u_1 saturates. To do this, the setpoint for u_2 is set as either $u_2^s = u_2^{min} + \Delta u_2$, or $u_2^s = u_2^{max} - \Delta u_2$. Δu_2 represents a small value close to the saturation limit, also referred to as back-off. This is required for the secondary controller to act appropriately and in time to maintain the output y, but will also create an operating region where both inputs are active at the same time.

One of the advantages of using valve position control is that the controlled variable can be controlled continuously through MV-MV switching to achieve tight CV control. This could as mentioned also be achieved through the use of split range control, which is not considered in this thesis.

Parallel could in addition to valve position control be implemented as one PI-controller and one P-controller. This is another control structure which can be used for two manipulated variables acting on one controlled variable, shown in Figure 2.9. The control requires different time constants for the controllers[27], and only one of the controllers can involve integral action. The reason only one controller can involve integral action, is because with two integral actions in the loop, there will be no unique steady-state solution for the manipulated variables[28].

The selection of controllers should be such that the manipulated variable with the largest steady state effect on the output should use PI-control, while the P-controller should be used on the remaining manipulated variable. The idea is thus to use the first manipulated variable with the largest steady state effect, u_1, to stabilize and return the system output back to nominal value. Then, once the error is zero, the other manipulated variable, u_2 is returning to its nominal value[29].

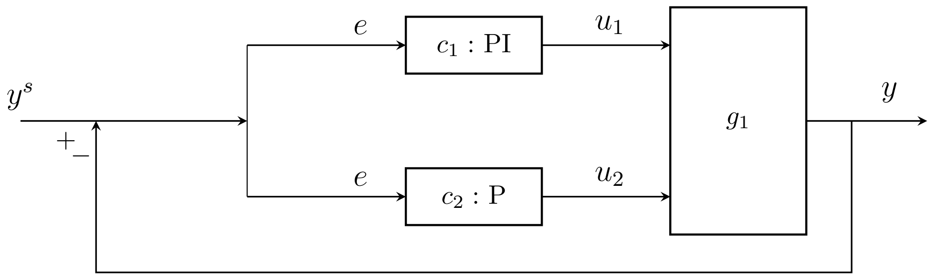

Figure 2.9: Block diagram of the parallel controller structure.

Selecting the controller structure is also a question about the time constants in the system, that is the time constants from u_1 and u_2 to y. The system should be brought slowly to steady state, so the PI-controller should be used for the process with the slowest time constant, meaning it will have the largest time constant. Thus, the PI controller should ideally be chosen for the input with the largest steady state effect and largest time constant.

2.5.3 Different Controllers with Different Setpoints

Controllers with different setpoints, a form of input sequencing like split range control, is used for extending operating range using MV-MV switching in a multiple input single output process[1].

The concept of different controllers with different setpoint is illustrated in Figure 2.10. For this structure, two controllers c_1 and c_2 and two inputs u_1 and u_2 are used to control the same output, y. However, the setpoint for each controller is different. The first controller uses the desired output setpoint y_1^s, while the second controller uses an offset Δy^s as setpoint. This setpoint offset, Δy^s, should be either added or subtracted from the setpoint for u_1, y_1^s, based on process knowledge of whether Y will drift above or below setpoint if u_1 saturates. For example, in a steam distribution network, it is likely with a pressure drop in the system because of lack in supply, thus the setpoint for u_2 should be $y_2^s = y_1^s - \Delta y^s$. Likewise, for a process where a positive drift would occur with saturation in u_1, a positive offset for the setpoint on u_2 should be selected[1].

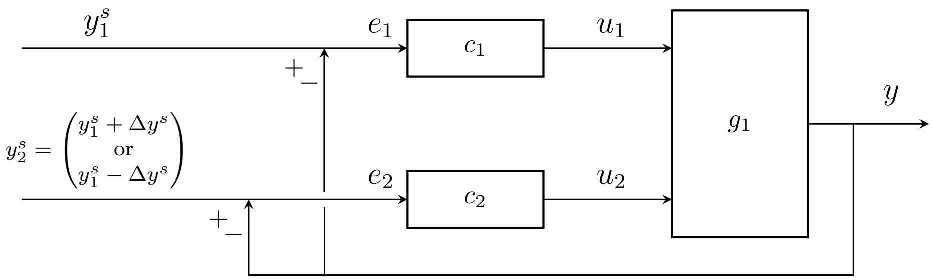

Figure 2.10: Block diagram of different controllers different setpoints controller structure.

The disadvantage using different controllers with different setpoint is that there has to be an offset in the output from the setpoint before a MV-MV switch happens. This effect could be minimized by selecting a small value for Δy^s. However this is a trade-off, as a to small offset value could cause both controllers to act at the same time or enable the second controller when it is not really required to[1].

2.5.4 Selectors

Selectors are a logic block often used when implementing advanced control structures, used to switch controlled variables for a single input system, also known as CV-CV switching. This involves that one manipulated variable controls multiple controlled variables. To obtain this, one controller is required for each controlled variable in the system. The selectors, which are usually maximum, minimum, or mid-range selectors, should be designed such that the single manipulated variable is controlling only one controlled variable at a time. The principle behind selectors is shown in Figure 2.11.

Figure 2.11 shows a process with two outputs y_1 and y_2, and two controllers c_1 and c_2 each processing one output signal each. The selector then chooses either the minimum, the maximum or the middle value based on a logic statement. Note that the middle selector can only be used if there is three a minimum of three controller outputs. The use of selectors are feasible when all outputs can be acceptably controlled at any given time with a single output. However, there are cases where this form of CV-CV switching has been shown to not be feasible[30].

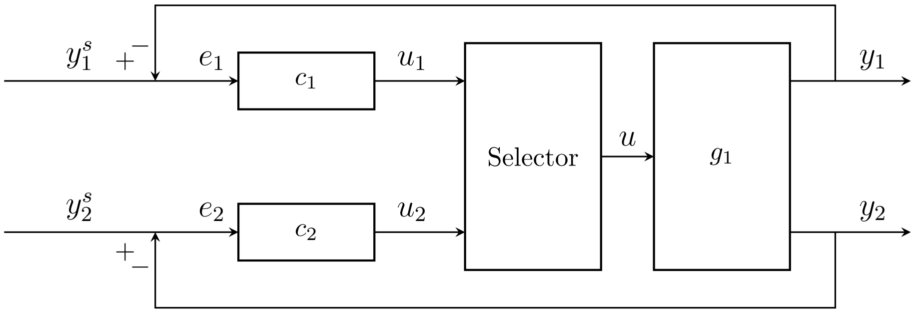

Figure 2.11: Block diagram of a system with one manipulated variable using a selector to maintain two controlled variables.

Usually, when designing selectors, the input pairing rule should be used. The rule states that manipulated variables that are likely to saturate should be paired with a CV that can be given up. This could be the less important producer or consumer flows or pressures in this thesis[24].

2.6 Droop Control

In a power network, frequency is a variable measuring the imbalance between supply and demand in the same way as the pressure would measure the imbalance in a steam distribution network. For this reason, as mentioned in the introduction, the control structure proposed in this work has some similarities with how frequency control is done in an electric grid.

Most electrical networks utilizes the control idea frequency control. Electrical frequency is a continuously changing variable controlled by second-to-second balance between demand and generation. If generation is greater than demand, system frequency increases, and opposite. To ensure keeping the frequency as close to nominal value as possible, sufficient reserves must be made available such that producer supply and consumer demand can be balanced. To achieve a sufficient reserve to maintain frequency, frequency control is built on the principle of using a control hierarchy, that primary, secondary, and possibly tertiary control. The control is separated on different time scales with primary being the fastest acting, and tertiary the slowest[31]. This principle is later utilized to prioritize consumers and producers in the modeled steam network in this thesis.

The primary reserve is an automatic change in active power output as response to a frequency change. This is done by synchronized generators making use of automatic speed governors denoted by a characteristic droop, expressed in Equation 2.8 [32].

$$s_G = -(\Delta f / f_n)/(\Delta P_G / P_n) \tag{2.8}$$

In Equation 2.8, Δf is the steady-state frequency deviation from the nominal frequency f_n, ΔP_G is the change in the power generation, P_n is the nominal power generator output power [32].

The droop itself is thus defined as the ratio between change in steady-state frequency from nominal to a new steady-state and the steady-state change in power output from nominal to a new steady-state. From a control perspective, the droop controller is a proportional controller using the droop gain and will thus create an offset when change in frequency happens. The primary control is therefore only meant to limit and stop frequency excursion from setpoint value, but will reach a new steady state because of proportional offset [32].

2.7 Centralized Control - Model Predictive Control

Multivariable control differs from decentralized control as it can work with more than one control objective at the same time. This means that the process can use multiple inputs and multiple outputs, simultaneously. Multivariable supervisory control is usually perceived to be synonymous with model predictive control, but many multiple multivariable control methods exist today, such as neural networks. This thesis will also only explore model predictive control.

Model predictive control is perhaps the most common multivariable control structure. It has been developed and used since the 1970s, but decentralized control has been preferred. This is because model predictive control requires powerful computing capabilities. However, with increasing computing capacities, use in the industry has seen an increase since the 2000s [33].

Model predictive control has multiple advantages and is best for using in interacting processes where inactive constraints change. It is good for the handling of feedforward control where disturbances are measured, as well as constraint changes. This is because the process model captures the dynamic and static interactions between input, output, and disturbance variables, and constraints on inputs and outputs are considered systematically[34].

Even though model predictive control is theoretically promising, it faces some issues. The implementation requires a multivariable dynamic model. This makes tuning or weighting the model predictive controller optimally difficult. In a gas network, this requires information about all consumer and producer flow-rates, which would usually not be available, especially if there are many consumers[16]. Also, the objective function should be selected in such a manner that it represents the control objective. If the model, objective function, and tuning for the model predictive controller is not sufficient, performance could, in the worst case, become worse than no control. Related installation costs are also high, as an extra controller with communication between the already existing controller structure is required[33].

Mayne et al. (2000) defines model predictive control as the following:

> "*Model predictive control is a form of control in which the current control action is obtained by solving, at each sampling instant, a finite horizon open loop optimal control problem, using the current state of the plant as the initial state; the optimization yields an optimal control sequence and the first control in this sequence is applied to the plant.*" [35]

2.7.1 Objective Function

To obtain the optimal control sequence, an optimal control problem must be defined. There are many ways to define optimal control problems, but the objective function should be a scalar function which describes a property that is desired to minimize or maximize[2]. In this thesis the objective will be stated as a minimum function, as the CasADi framework is based on minimizing the objective function[36].

Further, there exist many types of objective functions. In this thesis a combination of a tracking and an economic objective function is used. A tracking objective function tracks the deviation from the actual values for the states, input and input usage, compared to their respective nominal values. The economic objective function is more abstract, but should reflect the plant costs through a custom cost function as a function of states, inputs and possibly input usage. Equation 2.9 presents the combined objective function for a discrete time system, with the first link being the tracking part, and the second being the economic part.

$$\begin{aligned} min\sum_{t=1}^{t=N}(x_t - x_t^s)'\mathrm{W}(x_t - x_t^s) + \sum_{t=1}^{t=N} cost(x_t, u_t) & \\ \text{s.t.} \quad \dot{x} = f(x, u, d) & \\ g(x, u) \leq 0 & \\ W \geq 0 & \end{aligned} \tag{2.9}$$

This objective function spans on the time horizon from $t = 0$ to $t = N$ which is the prediction horizon. The value W is a weight or cost number used to prioritize the most important setpoint tracking, and should therefore match the dimensions of the state x_t. It thus is a matrix with a quadratic dimension if the dimension of x_t is greater than one. If $\mathbf{dim(}x_t\mathbf{)}= 1$ the problem could be simplified to that in Equation 2.10. The economic part for now only illustrates a general cost function based on inputs and states, as this part needs individualization for each process.

$$\begin{aligned} min\sum_{t=0}^{t=N} W(x_t - x_t^s)^2 + \sum_{t=0}^{t=N} cost(x_t, u_t) & \\ \text{s.t.} \quad \dot{x} = f(x, u, d) & \\ g(x, u) \leq 0 & \\ W \geq 0 & \end{aligned} \tag{2.10}$$

A nonlinear model predictive controller assumes that the system objective function is based on states, obtained from differential model equations, $\dot{x}$. In this thesis, the differential equation should be a function of states x, inputs u and disturbances d. As for the constraints, there could be equality constraints, meaning $g(x, u) = 0$, and inequality constraints, $g(x, u) \leq 0$. In practice all constraints are equality constraints when they are active, for example in a case with a zero pressure, the physical minimum constraint for any pressure. Further, for the problem to be feasible, the weight W cannot be negative.

2.7.2 Model Predictive Controller Algorithm

The model predictive controller algorithm works by using objective function to find and use the numerical approximation of the optimal feedback control input u_t for a given current state x_t, as shown in Figure 2.12. The predicted future optimal input is then used to predict the system's future behavior. This is done for each step on the prediction horizon. This means that for one moving horizon step, predictions for future states $x^*_{t+1}, ..., x^*_{t+N}$ are obtained. Thus, when moving one step further on the moving horizon, the optimization is done all over again, obtaining new optimal states $x^*_{t+2}, ..., x^*_{t+N+1}$. The optimal states for each prediction horizon is used to obtain the optimal system inputs u_t. The procedure is then performed over and over again on the moving time horizon, which in theory could be endless. However, for any change in the plant, the model needs to be updated and tested, indicating that by practical means, model predictive controllers are not on an endless time horizon[2].

The model in this thesis will be nonlinear. Therefore the model predictive controller will be required to handle a nonlinear system. The only difference between a nonlinear and linear model predictive controller is the nonlinear model and objective function, making the problem nonlinear and non-convex. This complicates solving the optimization problem, as a nonlinear solver will have to be used[2]. The next section presents the basics for minimizing or maximizing a constrained nonlinear objective function using collocation methods.

2.7.3 Nonlinear Optimization - Collocation methods

In order to solve the constrained nonlinear ordinary differential equation $\dot{x} = f_c(x, u)$ a numerical solver is required. For model predictive control, a popular choice is a subclass of implicit Runge-Kutta methods formed by the name of collocation methods. The collocation methods approximate the nonlinear differential state equation using linear polynomials, efficiently approximating the problem to an algebraic optimization problem. The optimization problem can then be solved analytically for each time step, yielding an approximated new state for use in the next time step calculation. This method is proved for high accuracy and relatively low calculation costs if matrix sparsity is exploited[34]. A popular collocation method is direct collocation, which among others is used by the nonlinear optimization tool package CasADi for MATLAB or Python uses[36].

For direct collocation, the time horizon $[0, T]$ should first be divided into a large number N of collocation intervals $[t_i, t_{i+1}]$, with $t < t_1 < ... < t_N = T$. Each of these intervals

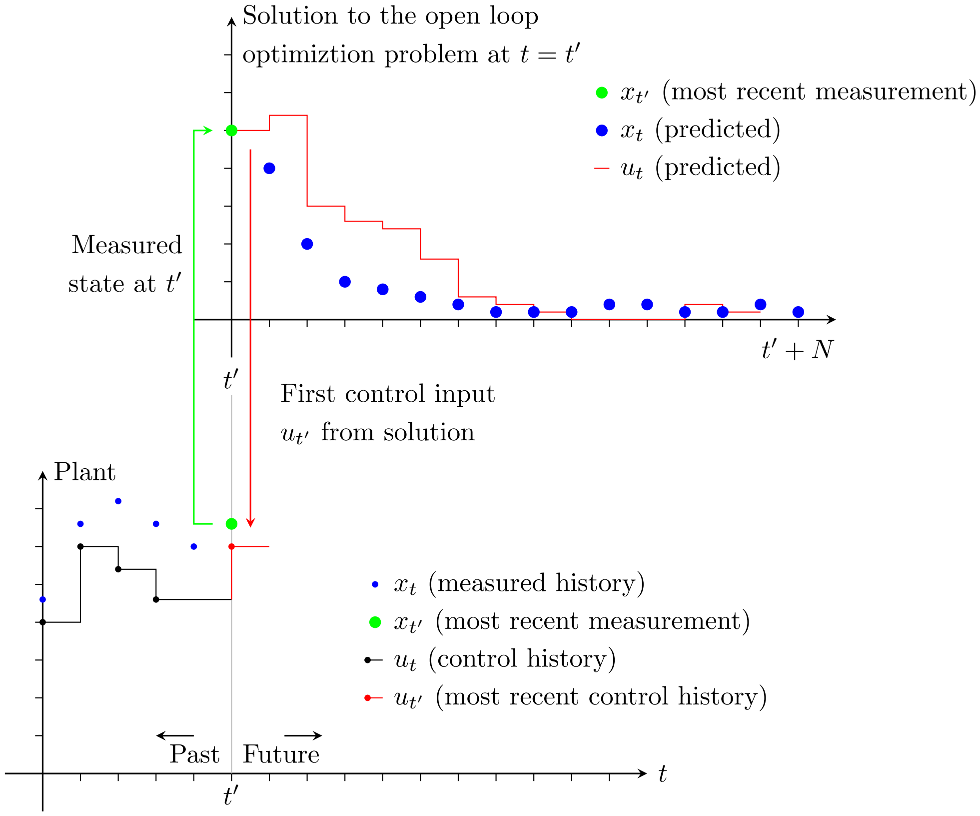

Figure 2.12: Illustration of the MPC principle[2].

include the use of an implicit Runge-Kutta integration rule of collocation type to transcribe the ordinary differential equation $\dot{x} = f_c(x, u)$ to a finite set of nonlinear equations[34].

To achieve the creation of a finite set of nonlinear equations, states $s_i = x(t_i)$ at the time points t_i should be introduced. These should regard the implicit Runge-Kutta equations with M stages on the interval with length $h_i = (t_{i+1} - t_i)$, which create an implicit relation between s_i and s_{i+1}. Furthermore, introducing variables $K_i = [k'_{i,1}...k'_{i,M}] \in \mathbb{R}^{nM}$ where $k_{i,j} \in \mathbb{R}^n$ is the state derivative at the collocation time point $t_i + c_j h_i$ for $j = 1, ..., M$. K_i are uniquely defined by the collocation equations if s_i and the control value $q_i \in \mathbb{R}^m$ are given. The collocation equations are presented in 2.11[34].

$$G_i^{RK}(s_i, K_i, q_i) = \begin{bmatrix} k_{i,1} - f_c(s_i + h_i(a_{11}k_{i,1} + ... + a_{1,M}k_{i,M}), q_i) \\ k_{i,2} - f_c(s_i + h_i(a_{21}k_{i,1} + ... + a_{2,M}k_{i,M}), q_i) \\ \vdots \\ k_{i,M} - f_c(s_i + h_i(a_{M1}k_{i,1} + ... + a_{M,M}k_{i,M}), q_i) \end{bmatrix} \tag{2.11}$$

The collocation points give a transition to the next system state by calculating $s_i =$

$F_i^{RK}(s_i, K_i, q_i)$ where Equation 2.12 unfolds the expression for $F_i^{RK}(s_i, K_i, q_i)$[34].

$$F_i^{RK}(s_i, K_i, q_i) = s_i + h_i(b_1 k_{i,1} + ... + b_M k_{i,M} \tag{2.12}$$

With direct collocation, a separate control parameter within every collocation time point is possible. Doing this would cause the maximum number of control degrees of freedom compatible with direct collocation, and they could be interpreted as a piecewise polynomial control parameterization of order $(M-1)$. This parameterization could be solved analytically[34].

Chapter 3

Process Modeling

3.1 Process Description

The process considered in this master thesis models a theoretic steam distribution network using MATLAB, MATLAB extension Simulink, and numerical optimization software CasADi. The system does not model the actual producers nor consumers, only the network between them. Shown in Figure 3.1, the network consists of a high-pressure supplier side, including six high-pressure suppliers with different flow specifications. The suppliers are categorized into three categories, namely droop producers, swing producers and extra producers.

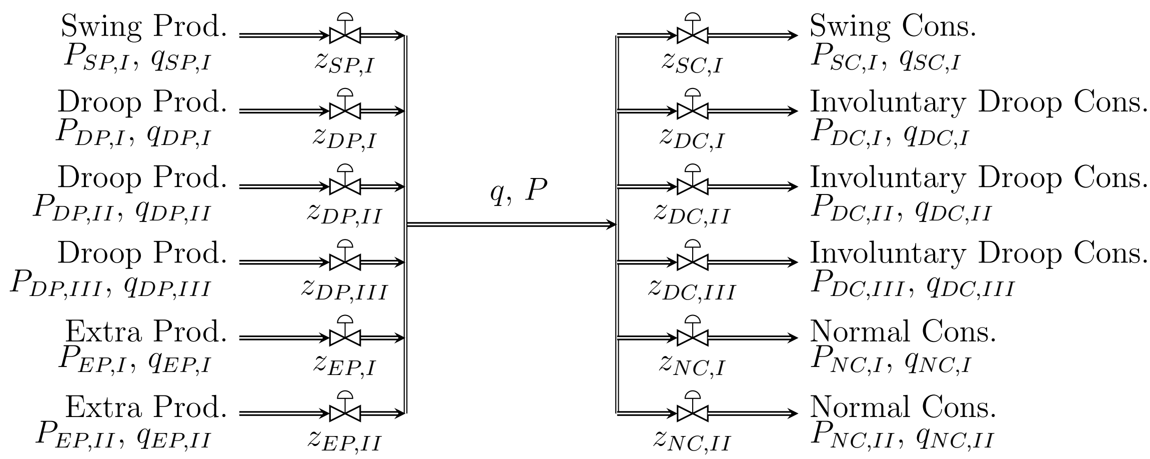

Figure 3.1: Illustration of a theoretical gas network with six consumers and six suppliers with different dynamics, including process variables.

Three droop producers resemble the idea of primary proportional acting primary control from frequency control in electrical distribution networks. Therefore, droop producers are the fastest acting, stabilizing system controllers.

The one swing producer in the system has its name from swing control operation around boilers and level control. That is, the controller has a delay or a threshold where control action is not taken reducing wear and tear an also minimizing energy loss from the supplier. Therefore the swing producers are used after the droop producers, if their supply is not sufficient. The swing producers are thus the secondary control structure in the steam distribution network, acting on a slower timescale than the droop producers.

The producer side also involves two extra producers, which act as backup control, not activated until the primary and secondary droop and swing producers are insufficient, thus making them a tertiary control structure. They work on an even slower timescale than the droop and swing producers. Their supply could originate from extra boilers in a steam distribution network.

The suppliers are separated using linear valves, losing pressure before meeting in a medium pressure frictionless main pipeline. Then, the main pipeline is separated into six medium pressure consumers separated into three different categories. That is, swing consumers, involuntary droop consumers and normal consumers. The one swing consumer resemble the same idea as the swing producer, but should act as a quaternary control, meaning it should not act before the primary, secondary and tertiary control is insufficient.

Three involuntary droop consumers originate their name from the droop producers, because they involve no control structure, but are directly affected by the droop producers because of coupling in the network around the main pipeline. They are therefore directly affected by disturbances and regulated from corrective action from the droop producers.

The consumer side also involves two normal consumers. They are named normal as they involve only regulatory control. They are therefore not involved in the supervisory control structure, but rather acting as a system disturbance. Last, it should be noted that all consumer flows are separated using consumer valves, losing pressure through the valves.

The supplier and consumer flows are named according to what controller principle they should be implemented with, emphasised in Chapter 4. In total, the system involves six types of consumer and supplier valves intended for different control purposes. Those are one swing producer, SP, I, one swing consumer, SC, I, three droop producers, DP, I, DP, II, and DP, III, three involuntary droop consumers, DC, I, DC, II, DC, III, two extra producers, EP, I, and EP, II and two normal consumers, NC, I, and NC, II. The consumer and producer pressures, flows, and valve openings, denoted $P_{XX,i}$, $q_{XX,i}$, and $z_{XX,i}$ respectively, $\forall XX \in DP, SP, EP, SC, DC, NC$. The flow and pressure in the main pipeline is denoted q and P with no subscripts.

Each of the control structures defining the consumer and producer flows are to be described in detail in Chapter 4. For now, the following sections will look at the nominal system without control, including assumptions and constraints, possible manipulated variables, and controlled variables, as well as model limitations and considerations.

3.2 Assumptions and Constraints

Starting with some clarification in the modeled network, all parameters and nominal values for the modeled system are kept equal for both the decentralized and the multivariable control simulations, to keep them consistent. It should also be mentioned that the steam network is modeled without process equipment, except for linear valves and frictionless pipes, as the focus and scope of this thesis lies within optimal process control, not process optimization.

As for system considerations: Most importantly, the gas network is assumed to be represented by water in pure gas form (steam) for modeling simplicity. This is ensured by keeping the system above its dew point. The dew point defines as where the first drop of condensate is formed. One way to ensure this is by using a temperature-pressure diagram. By assuming constant system temperature, the diagram can be read such that the system temperature in the system is set above dew point according to the pipeline with the highest pressure. Therefore, the temperature in the system is assumed constant [37].

The reason for selecting the pipeline segment with the highest pressure is that this segment will also have the dew point occurring at the highest temperature. To ensure a temperature above the dew point at a nominal 15 bar producer pipeline pressure, a temperature of at least 198 °C should be maintained. As a buffer because of possible increased producer pressure due to disturbances in the system, the actual modeling temperature is set slightly above this, at 200°C [37].

For a real system requiring pure gas without condensate, keeping close to the dew point is economic, as maintaining a higher system temperature requires more heating costs. However, staying closer to the dew point conditions would make the system more vulnerable to pressure increasing disturbances. Therefore, the modeled system will not be realistic for a large pressure increase, as condensate would form, and in that case a two-phase system would have to be modeled to represent system behavior, which is not within the scope of the system modelled in this thesis.

The assumption about a pure steam system follows that the equation of state for an ideal gas is valid. The ideal gas law is used to relate pressure P and the volume V with temperature T, the amount of gas in moles n, using the universal gas constant R in $m^3atm/Kmol$ to relate the equation variables. The relation is presented in the network model section [38].

The basis for using the ideal gas law is that the velocities and pressures in the flow network are assumed to vary only in the flow direction, with no condensate forming within the control volume. No condensate formation is ensured through maintaining the system conditions above dew point conditions. Variation only in the flow direction is a direct assumption, but this should yield an appropriate approximation for large flows [38].

For convenience and model simplicity, it is also assumed that the system is isolated, with no heat loss from the system to the environment is assumed. The valves in the system are also assumed all to be linear valves. This simplifies the relationship between the valve pressure drops and the system pressures in the modeling section.

The system itself involves physical constraints. The valves are physically constrained from a fully closed position to a fully opened position, ranging from zero to one, $z \in [0, 1]$. The pressures are physically limited from zero to infinity, $P \in [0, \infty)$. Flow, in theory, has no limitations, as reverse flow is possible if the pressure drops before and after a valve are negative. This would cause opposite direction flow. However, it was assumed earlier that there would only be flow in one direction. Therefore, the minimum flow should also be zero, meaning flow would go from zero to infinity, $q \in [0, \infty)$. In order for the minimum flow assumption to be true, it constrains the pressures to decrease along the flow direction in the network, $P_{Prod} > P > P_{cons}$.

The physical system constraints presented could also be viewed as input or output constraints from a control perspective. The system also involves control constraints and assumptions from a control perspective. All of this is elaborated in Chapter 4. First, the system network model with no control whatsoever is presented.

3.3 Network Model

A mathematical model based on the proposed model requirements can be obtained by knowing the system variables, model assumptions, and physical constraints. The model should include a relationship between process variables: the flows, pressures, and valve positions in the system. This relation should be obtained while maintaining the assumptions and physical constraints for the system. The book Chemical and Energy Process Engineering (2009) proposes that two things should be considered when modeling a system [39].

1. What is the control volume of the system?
2. Which type of balance should we use (Mass, component or energy)?

Looking at the overall system in Figure 3.1, this entire figure will represent the system control volume, that is, the system between the steam producing boilers and process equipment consumers. Since all the flows of the system are through the main pipeline, a mass balance would be a good starting point. Relating the flows through a molar mass balance would yield the expression found in Equation 3.1 [39].

$$\frac{dn}{dt} = \sum_{j=1}^{1} q_{SP,j} + \sum_{k=1}^{3} q_{DP,k} + \sum_{l=1}^{2} q_{EP,l} - \sum_{m=1}^{1} q_{SC,m=1} - \sum_{n=1}^{3} q_{DP,n} - \sum_{p}^{2} q_{NC,p} \tag{3.1}$$

Equation 3.1 states that the change of n, which is the holdup in $kmol$, is the subtraction of the total producer flow into the control volume minus the total consumer flow out of the control volume. That is, the sum of all producer flows, $q_{SP,j}$, $q_{DP,k}$ and $q_{EP,l}$, subtracted by the sum of all the consumer flows $q_{SC,m}$, $q_{DC,n}$ and $q_{NC,p}$. All flows are using the units $kmol/s$. Thus, $\frac{dn}{dt}$ represents the total molar mass flow change per time.

Since the flows should be related to the pressure P, a relation between P and n is required. To create a relation, the equation of state for an ideal gas could be used to express P as a function of n. The equation of state for an ideal gas is presented in Equation 3.2 [38].

$$P = \frac{nRT}{V} \tag{3.2}$$

The ideal gas law states the pressure P in terms of temperature T, system volume V and the number of moles particles n in the system using the universal gas constant R to relate it. Using the equation of state for an ideal gas, the molar mass balance can be reformulated in terms of pressure and the flows relevant from the problem formulation, expressed in Equation 3.3.

$$\frac{dp}{dt} = \frac{RT}{V}\left(\sum_{j=1}^{1} q_{SP,j} + \sum_{k=1}^{3} q_{DP,k} + \sum_{l=1}^{2} q_{EP,l} - \sum_{m=1}^{1} q_{SC,m} - \sum_{n=1}^{3} q_{DP,n} - \sum_{p=1}^{2} q_{NC,p}\right) \tag{3.3}$$

Now, expressions for the system flows $q_{SP,j}$, $q_{DP,k}$, $q_{EP,l}$, $q_{SC,m}$, $q_{DC,n}$ and $q_{NC,p}$ are required. They can be expressed using the general valve equation for a flow q, presented in Equation 3.4[26].

$$q = Cvf(z)\Delta P = Cvf(z)\sqrt{\frac{(P_{in} - P_{out})}{\rho}} \tag{3.4}$$

Equation 3.4 raises some new parameters to the problem. ρ is a simplified expression representing the average density between the flow before and after the valve. Cv is the valve coefficient. It is a constant parameter representing flow capability through the valve. That is, at a given pressure differential, a larger coefficient would mean a larger flow. Traditionally, a Cv value of one is defined as the Cv value required to flow one gallon per minute of water at 60°F with a pressure drop of one PSI. The value for Cv is often chosen based on system size, which would also be the case in this master thesis[26].

P_{in} is defined as the pressure before the valve, while P_{out} defines as the pressure after the valve. The remaining coefficient, $f(z)$, represent the valve characteristic. There are different types of valves with different characteristics, but linear valve characteristics are used in this thesis. Linear valves have the simplest form of valve dynamics, with $f(z) = z$, meaning the valve dynamics are taking a linear value between the physical limits decided in this thesis to range from zero to one, $z \in [0, 1]$. Valve characteristics could also be semi-linear or nonlinear[26].

Using the assumption that an average density can be used for the pressure drop in the valve and the assumption that the system follows the ideal gas law principles, the density can be simplified and expressed in other terms. This yields a relation with the flow only depending on the pressures entering and exiting the valve and the valve opening. The general, simplified, density-independent expression for the valve equation is presented in Equation 3.5[26].

$$q_i = C_{v,i} z_i \sqrt{P_{i,in}^2 - P_{i,out}^2} \tag{3.5}$$

The relation in Equation 3.5 yields the algebraic equations for the flows $q_{SP,j}$, $q_{DP,k}$, $q_{EP,l}$, $q_{SC,m}$, $q_{DC,n}$ and $q_{NC,p}$, presented in Equations 3.6-3.11.

$$q_{SP,j} = Cv_{SP,j} z_{SP,j} \sqrt{P_{Prod}^2 - P^2} \tag{3.6}$$

$$q_{DP,k} = Cv_{DP,k} z_{DP,k} \sqrt{P_{Prod}^2 - P^2} \tag{3.7}$$

$$q_{EP,l} = Cv_{EP,l} z_{EP,l} \sqrt{P_{Prod}^2 - P^2} \tag{3.8}$$

$$q_{SC,m} = Cv_{SC,m} z_{SC,m} \sqrt{P^2 - P_{Cons}^2} \tag{3.9}$$

$$q_{DC,n} = Cv_{DC,n} z_{DC,n} \sqrt{P^2 - P_{Cons}^2} \tag{3.10}$$

$$q_{NC,p} = Cv_{NC,p} z_{NC,p} \sqrt{P^2 - P_{Cons}^2} \tag{3.11}$$

Taking a glance at Equations 3.6 through 3.11 it becomes clear that the system is highly interactive, as P relate all the system flows. This means that even a tiny change in a consumer or producer would change the main network pressure P, disturbing the other flows in the network.

This section has now presented a relation for the flows $q_{SP,j}$, $q_{DP,k}$, $q_{EP,l}$, $q_{SC,m}$, $q_{DC,n}$ and $q_{NC,p}$ using pressure differences, as well as a mass balance represented in terms of pressure. The mass balance in terms of pressure is an ordinary differential equation, while the flow equations are algebraic as a function of the pressure solution of the differential equation. In order to implement the presented model of the system, control objectives, including choosing the systems manipulated variables and controlled variables, should be stated. They are presented in the next chapter.

3.4 Nominal Values

This section presents all system nominal values for the inputs, outputs and parameters described in the previous section. They are shown in Table 3.1.

It should be noted that for nominal values, producer pressure is equal for all producers, as well as nominal consumer pressure for all consumers, meaning $P_{SP}^{nom} = P_{DP}^{nom} = P_{EP}^{nom} = P_{Prod}$ and $P_{SC}^{nom} = P_{DC}^{nom} = P_{NC}^{nom} = P_{Cons}$. The same is not the case for the valve openings. On the consumer side, the swing consumer, the involuntary droop consumers and the normal consumer has a nominal valve opening of $z_{SC}^{nom} = z_{DC}^{nom} = z_{NC}^{nom} = 0.1$, while on the producer side the swing producer and the droop producers have a nominal value of $z^{nom}_{SP} = z^{nom}_{DP} = 0.5$.

For the extra producer, the nominal flow and valve position would be zero, $z^{nom}_{EP} = 0$. This is since the extra producer should enact as the tertiary backup from frequency control if a consumer disturbance causes supply to be insufficient. Therefore, at nominal conditions, only four of the six producer side valves will be open.

Table 3.1: A complete depiction of the simulation parameters and nominal values.

Variable	Value	Unit
T	200	[°C]
R	$8.314*10^{-5}$	[m^{3}bar/kmol]
V	150	[m^3]
P_{Prod}	15	[bar]
P^{nom}	10	[bar]
P_{Cons}	2	[bar]
$q^{nom}_{SP,j}$	1	[kmol/s]
$q^{nom}_{DP,k}$	1/3	[kmol/s]
$q^{nom}_{EP,l}$	0	[kmol/s]
$q^{nom}_{SC,m}$	0.6	[kmol/s]
$q^{nom}_{DC,n}$	2/15	[kmol/s]
$q^{nom}_{NC,p}$	0.5	[kmol/s]
$Cv_{SP,j}$	0.1789	[(kmol/s)/bar]
$Cv_{DP,k}$	0.0596	[(kmol/s)/bar]
$Cv_{EP,l}$	0.1789	[(kmol/s)/bar]
$Cv_{SC,m}$	1.3765	[(kmol/s)/bar]
$Cv_{DC,n}.$	0.3059	[(kmol/s)/bar]
$Cv_{NC,p}$	1.1471	[(kmol/s)/bar]
$z^{nom}_{SP,j}$	0.5	[-]
$z^{nom}_{DP,k}$	0.5	[-]
$z^{nom}_{EP,l}$	0	[-]
$z^{nom}_{SC,m}$	0.1	[-]
$z^{nom}_{DC,n}$	0.1	[-]
$z^{nom}_{NC,p}$	0.1	[-]
SP	1	[-]
DP	3	[-]
EP	2	[-]
SC	1	[-]
DC	3	[-]
NC	2	[-]

3.5 Open Loop Model

The model implemented in this thesis is built around the ordinary differential equation in the system, that is dP/dt presented in Equation 3.3, as well as the algebraic flow equations making up the main part of the differential equation. For the open loop model, these are embedded in the MATLAB file OLODEFILE.m. The equations are calculated using four inputs vectors. The vectors consists of a state vector x, an input vector u, a disturbance vector d, and a constant parameter vector $params$. These values are inputted to the OLODEFILE.m using a Simulink Model block.

In this the thesis, the state would be the solution to the ordinary differential equation, P, the inputs the consumer and producer valve openings z_{vec}, the disturbances the consumer and producer side pressures, P_{vec}. The system constant parameters, making up $params$, would be system size V, the universal gas constant R, system temperature T, the number of each consumer and producer $SP, SC, ..., EP$, and all CV values, $Cv_{SP,I}, Cv_{SC,I}, ..., Cv_{EP,II}$.

The output of the Simulink Model block is a vector containing P and the algebraic calculations of all the consumer and producer flows, based on the state input. However the current pressure is outputted as an ordinary differential equation and must therefore be solved using a MATLAB nonlinear solver. To solve the system, however, the algebraic equations must be split from the differential equations. This is done using MATLAB Gain blocks. Two gains are defined to separate the equations, K_{Diff} and K_{Alg}. Their structures are presented in Equation 3.12 and 3.13.

$$K_{Diff} = \begin{bmatrix} 1 & 0 & \dots & 0 \end{bmatrix} \tag{3.12}$$

$$K_{Alg} = \begin{bmatrix} 0 & 1 & 0 & & \dots & 0 \\ \vdots & 0 & 1 & 0 & \dots & 0 \\ \vdots & & & \ddots & & \vdots \\ \vdots & & & & \ddots & 0 \\ 0 & & \dots & \dots & 0 & 1 \end{bmatrix} \tag{3.13}$$

Having separated the ordinary differential equation and algebraic equations, the differential equation can be solved using a Simulink Integrate block. The Simulink Integrate block envelopes many solvers, depending on system dynamics. Since the system solved is highly coupled and has many inputs and outputs, a stiff solver is required. The stiff solver from MATLAB used in this thesis is $ODE15s$.

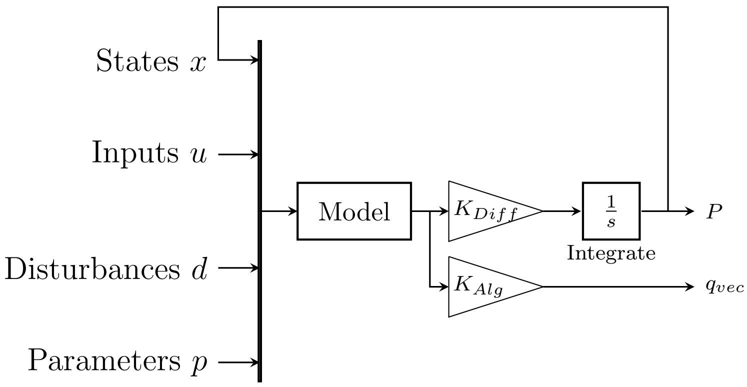

Figure 3.2: Illustration of the open loop system in Simulink, using the states, MVs,DVs and constants as inputs, outputting states, CVs.

The complete open loop model is presented in Figure 3.2. The closed loop system is based on this, which in addition uses Simulink step blocks for step responses and disturbance implementation, PI-controller blocks for control, and a minimum selector block for quaternary control implementation. The calculated pressure and flows for a given time step is given as feedback signals to their respective PI-controller blocks.

Chapter 4

Control Structures

4.1 Control Objectives

It is important to emphasize that even without control, the modeled system in this thesis will self-regulate and reach some steady-state because all system flows depend on the main network pressure P. However, control is required for two reasons. First, the self-regulating effect would be too slow for practical purposes of delivering desired flow and pressure, which would not be sufficient for an actual plant requiring fast responses to demand changes. Second, because the process is very interactive, a tiny disturbance on a consumer or producer would quickly affect the entire network, which could cause instability. Also, with only self-regulation, the supply pressure might become too small, leading to insufficient pressure supply to consumers.

In order to provide a sufficient supervisory control structure, control objectives must be defined. The control objective in this master thesis is to implement simple control policies that balance the generation and demand in a steam distribution network. Because all the network flows are related by P, this should be the main supervisory controlled variable. To maintain P, a prioritization for use of the producers and consumers is also defined as a control objective. Since steam distribution networks usually prioritize maintaining consumer flows at setpoint, this thesis will prioritize such that the producer flows should deviate from their nominal values before the consumer flows, using supervisory control. Implementing control with this focus should stabilize the consumer flows in the occurrence of disturbances and possibly extend the system operating range.

The producer side prioritization is using inspiration from frequency control in electrical grid networks. Recall, the idea involves the use of primary, secondary and tertiary control. For the system in this thesis, the primary acting control is the droop control the producer side. This would mean using droop proportional control action ultimately reducing or stopping setpoint drift using proportional control. This producer should therefore also be used first in the occurrence of a disturbance.

Secondary control would be recognized on a slower timescale and involves the swing producer. The swing producer control would involve a combination of proportional and integral control, ultimately stabilizing using proportional control and returning the consumer flows to setpoints using integral action.

Tertiary control is represented by the extra producers, which are not intended for use unless primary and secondary control is insufficient [32]. They are therefore not used at nominal operation.

Being the tertiary control, the extra producer should be used among the producers and therefore act on an even slower timescale than the primary and secondary control. Thus, the producer side prioritization becomes using the droop producers first (fast), then the swing producer (slower) and then the extra producers (slowest).

Still one supervisory control structure remains, this one being on the consumer side. That is, the added quaternary control or the swing consumer control. The swing consumer control should only be used if primary, secondary and tertiary control fails, meaning swing consumer usage should receive the lowest prioritization for input use when balancing supply and demand.

This leaves two consumer types. First, the involuntary droop consumers, which are regarded as non-controllable, only being controlled by the system's self-regulating effect, with no control added to the actual valves. Self-regulating consumers are realistic for real networks, where there could be thousands of consumers and most of them involve no implemented control of them for cost reasons [16]. Therefore they are outside the scope and possibility of prioritization. This means the involuntary droop consumers rely entirely on proper network control for adequate performance.

The last consumer type, the normal consumers, are assumed to have implemented only regulatory control, making them self-regulatory and acting as a disturbance. Therefore, the involuntary droop consumers and the normal consumers are not degrees of freedom and not part of the prioritization. This gives the prioritization list found in Table 4.1.

Table 4.1: Input use priority list.

Input Use Priority List for Pressure Control
1. Use primary control (droop producers).
2. Use secondary control (swing producer).
3. Use tertiary control (extra producers).
4. Use quaternary control (swing consumer).
Non-prioritizable: Involuntary droop consumers and normal consumer.

4.2 Variable Selection

To implement the system control proposed in the previous section, possible manipulated and controlled variables must be identified. Figure 4.1 shows the chosen manipulated and controlled variables in the modeled system of this thesis.

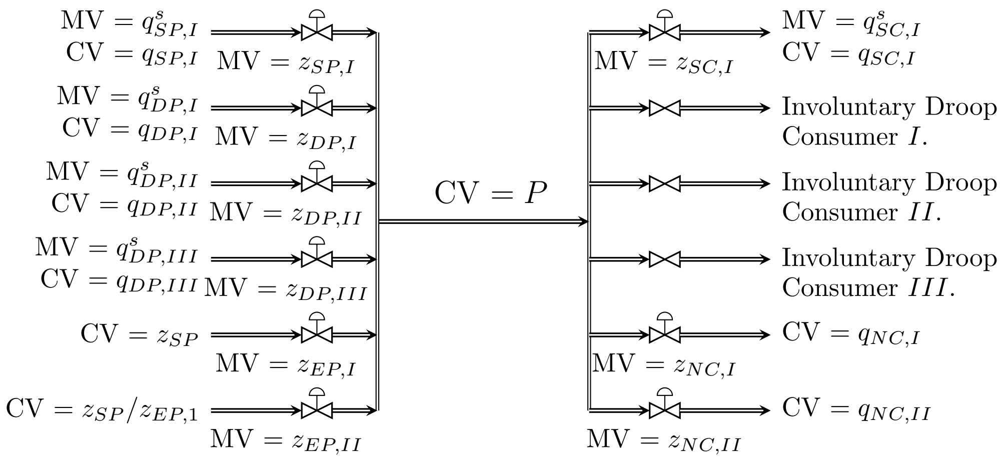

Figure 4.1: Illustration of a the modelled steam network, indicating possible manipulated and controlled variables.

First, the controlled variables and manipulated variables should be divided based on whether they belong to the regulatory or supervisory control layer, starting with the regulatory control. This should be fast acting, stabilizing control. It would therefore be a good idea to implement regulatory control on the controllable flows in the system.

The "pair close"-rule states that one should use an input-output pair with a small phase lag or equivalently a small effective time delay. By effective time delay it is meant the sum of the apparent time delay caused by dead time, inverse responses and high order lags[40]. Using the "pair close"-rule the flow outputs should be paired with their related valve positions as inputs. This is valid both location-wise and in term of timescale dynamics, as both the flow and valve dynamics are fast, resulting in a small effective delay in corrective action taken by the controller.

Thus, all flows that can are paired with their respective valve position controllers. However, not all flows can or should be paired with their respective valve position controller. These are the involuntary droop consumers, which are non-controllable, and the extra producers. Since the extra producers are off at nominal value, regulatory control would be possible, but not ideal. The result would be the valve positions always staying closed, at least without supervisory control to change the flow setpoints if needed. Since the flow setpoints are already utilized as a manipulated variable in both the droop producer and swing producer, using the flow setpoints as an input for the extra producer would make this structure equal to one of them. Therefore, valve position controllers are used instead.

The idea for the extra producers is therefore to use the valve position from the swing producer as a controlled variable, as this is the producer that should be used just before the extra producers. This way, the extra producers would not act unless the swing producers are close to saturation.

For the extra producer II, it would be possible to use both the swing producer valve position as an input, but also the extra producer I, in the same fashion as the swing producer valve position is used for the first extra producer. This yields three different potential controller structures for the extra producers, which are all discussed in the controller implementation section.

Because the main objective is to maintain P, this becomes the primary controlled variable for the supervisory layer. The remaining degrees of freedom in the system are now the flow setpoints for the droop producers, the swing producer and the swing consumers. They should all be used in order to control the main supervisory controlled variable, the pressure P.

Using the system variables in Figure 4.1, the now determined system inputs and outputs, and the control priority list, the overview presented in Table 4.2 is constructed. The table separates between what variables belong to the supervisory control structure and regulatory control. It also prioritizes the controlled variables in the system, independently of the regulatory and supervisory control, with $CV1$ being the most important and $CV6$ being the least important. Note that even though the extra producers are a part of the supervisory control structure, the controlled variables are not, as they do not change the setpoints for a lower control layer.

Using the pairings proposed in Table 4.2, practical implementation of each controller structure is presented in the next sections.

Table 4.2: A controlled variable priority list, stating the controlled variable priorities from highest ($CV1$) to lowest ($CV6$).

Supervisory Control		
CV	MV	Description
$CV1 = P$	$MV = q^s_{DP}$	Balance supply and demand.
$CV1 = P$	$MV = q^s_{SP}$	Balance supply and demand.
$CV1 = P$	$MV = q^s_{EP}$	Balance supply and demand.
$CV1 = P$	$MV = q^s_{SC}$	Balance supply and demand.
Regulatory Control		
CV	MV	Description
$CV2 = q_{SC}$	$MV = z_{SC}$	Maintain nominal swing consumer flow.
$CV4 = z_{SP}/z_{EP,I}$	$MV = z_{EP,II}$	Make sure $z_{SP}/z_{EP,I}$ does not saturate.
$CV5 = z_{SP}$	$MV = z_{EP,I}$	Make sure z_{SP} does not saturate.
$CV3 = q_{SP}$	$MV = z_{SP}$	Maintain nominal swing producer flow.
$CV6 = q_{DP}$	$MV = z_{DP}$	Maintain nominal droop producer flows.

4.3 Decentralized Control Implementation

The next sections look at the implementation specifics for each of the proposed controller structures. The controller structures follows the hierarchy shown in Table 4.3, including primary (fast), secondary (slow), tertiary (slower) and quaternary control (if all else fails).

Table 4.3: A depiction of the controller hierarchy in the system

Decentralized Supervisory Controller Hierarchy	
1. Primary Control (Fast)	Droop Producer Control.
2. Secondary Control (Slow)	Swing Producer Control
3. Tertiary Control (Slower)	Extra Producer Control
4. Quaternary Control (If all above fails)	Swing Consumer Control

4.3.1 Primary Control - Droop Producers

The droop producer control structure in this work is based on primary electrical network frequency control. If frequency control were to be implemented in a steam network exactly like in electrical network control, the primary control objective should be fast, stabilizing proportional acting control.

The primary control in the steam network in this thesis, builds around on an inner, regulatory flow control loop. That is, the inner flow control loop ensures flow stability and setpoint tracking. Then, the outer, primary control loop changes the flow setpoint in order to maintain the network pressure P as a supervisory control structure. So for primary control, the primary manipulated variable is $MV1 = q^s_{DP}$, and the primary controlled variable is $CV1 = P$.

The regulatory control layer is not a part of the primary control, but is affected by the flow setpoint changes. This is due to the fact that th regulatory control uses what becomes the secondary controlled variable, is $CV2 = q_{DP}$, and the secondary manipulated variable $CV2 = z_{DP}$, which are affected by changes in the flow setpoint.

Thus, for primary droop control presented in Figure 4.2, the control idea would be to use proportional control action $c_{Primary}$ for pressure control using the droop producer flow setpoints q^S_{DP} as supervisory control in a cascade loop. Since the cascade only involves a single process g, it is a parallel cascade.

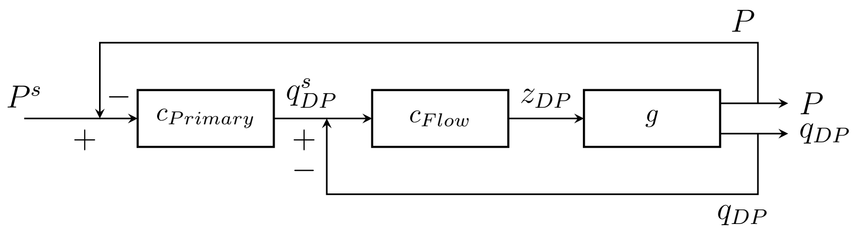

Figure 4.2: Block diagram for proposed droop producer control.

In order to avoid interaction in the parallel cascade loop, the time constants of the inner and outer loops need to be sufficiently separated on different time scales, with the valve position controllers being on the fastest time scale as they should represent a fast, inner loop. As the none of the flow controllers will interact with each other, except for the extra producers, all flow controllers except the extra producers should use the same time constant tuning parameter.

The primary controller, being the outer part of a cascade loop, should use a time constant at least five times larger than the flow controllers. Note that the primary controller will also in fact interact with the secondary, tertiary and quaternary supervisory control if they are on the same timescale, as they all use the same controlled variable, pressure. Therefore, the secondary control, presented next, is separated on an even slower timescale than $c_{Primary}$.

4.3.2 Secondary Control - Swing Producer

The secondary control structure, the swing control structure also involves a cascade like structure like the droop control structure, shown in Figure 4.6. The major difference from the droop producer control structure is that the outer, supervisory control loop controller, $c_{secondary}$, involves both proportional and integral action. The supervisory control on the swing producer should also act on a slower timescale than the droop producer, to avoid being utilized too soon and avoid interaction.

Thus, for the swing producers, the control idea would be to use combined proportional and integral control action $C_{cecondary}$ for pressure control using the droop producer flow setpoints q_{SP}^{S} as the outer control loop, and integral control $c_{Primary}$ controlling the flow q_{SP} using valve position z_{SP} as the main action in an inner cascade. In a real plant this would require measurement of both the flow q_{SP} and the pressure P, which both would be involved in feedback loops creating error signal to their respective controllers, as shown in Figure 4.3.

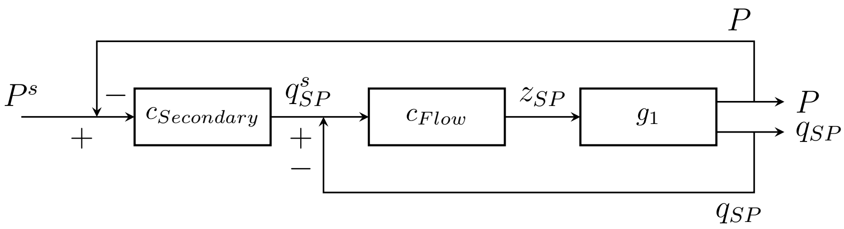

Figure 4.3: Block diagram for proposed swing producer control.

For this parallel cascade loop, it is even more important to have sufficient difference between the system time constants in the cascade loop. This is because there are two proportional controllers in a single control loop, which would interact if not given enough difference in response time. If interactions occur, the controllers could in a worst-case scenario work against each other. In addition, the outer loop constant should be larger than the one for the droop producers, to ensure the swing producers only act when the droop producers are about to saturate.

4.3.3 Tertiary Control - Extra Producers

Because of the prioritization made, and in order to correspond with the idea behind tertiary frequency control, the valves on the first extra producer, extra producer I, should act only when the swing producer valve is about to saturate. If supply from swing producer I is inadequate, swing producer II should act. Thus, the extra producers should be used for operating range extension and MV-MV switching.

The theory part proposes multiple ways to implement MV-MV switching control. Whatever control structure discussed, a good starting point is using valve position control with the swing producer valve position as an indication for when the first extra producer I should act. This because the extra producers are next in line to act according to the selected prioritization when the swing producers are saturated. Thus, using the swing producer valve position as an indication for when the extra producer I should be used as an input would be a good idea. For the second extra producer II, the options are more wast, and there are three alternatives explored in this thesis.

The first concept of proposed extra producer II configuration would involve using the swing producer valve position for both extra producer I and II. This configuration is illustrated in 4.4. The configuration gives the possibility of implementing two discussed controller implementations from the theory chapter. One would be the implementation of different setpoints on the extra producer controllers. This would enable the controllers to be equal PI-controllers using different setpoints for the controller input error from the swing producer valve position.

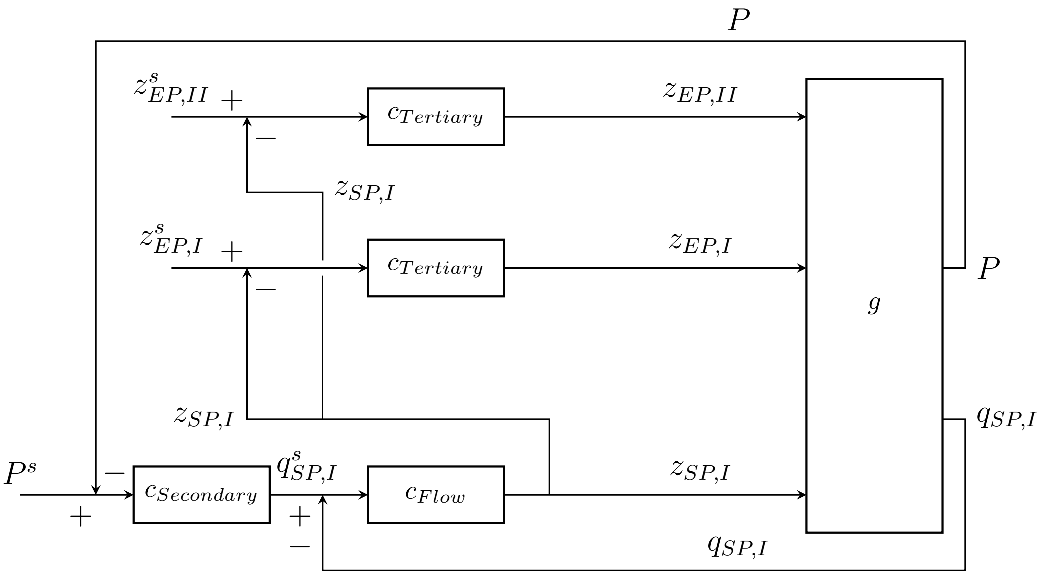

Figure 4.4: Block diagram for structure of extra producer control with the use of swing producer valve position as input

The second configuration would involve the use of parallel control, that is proportional-integrative control on one controller and only integral action on the other controller. Recall that the controller with the largest steady state effect should be chosen as the PI-controller, and the remaining should use a P-controller. However, since both extra producers are equal, which controller involves PI-control and which includes P-control does not matter.

The third control option, illustrated in Figure 4.5, uses valve position control, utilizing different controlled variables for each controller. For this controller structure, the swing producer valve position is used as a controlled variable only for the first swing producer I. The second swing producer II instead uses the valve position of extra producer I

as a controlled variable.

Valve position control would require some back-off to work efficiently and make sure the swing producer valve position does not saturate before the extra producer supply is secured. This means the extra producer I would actually turn on before the swing producer valve position is fully saturated.

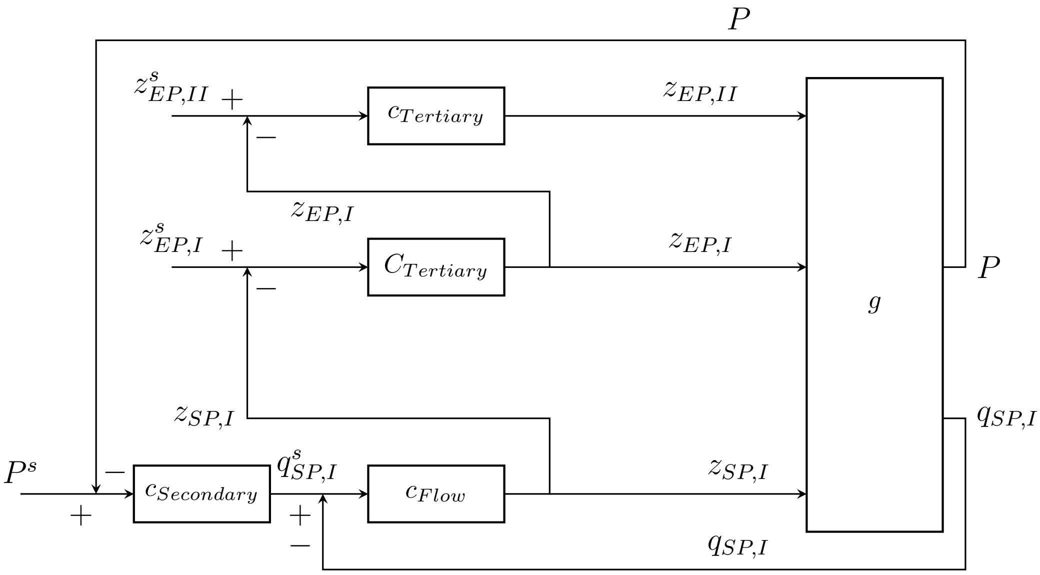

Figure 4.5: Block diagram for structure of extra producer control with the use of swing producer valve position as input for extra producer I, and extra producer I valve position as input for extra producer II.

Summarizing, there are in total three different extra producer configurations to try out in this thesis, presented in Table 4.4. The table presents the potential controller structures with their respective controlled variables and setpoint as well as controller type for each of configuration. All three will be implemented to find which structure performs the best in the decentralized control structure.

Table 4.4: Possible extra producer controller configurations.

$Controller_{EP,I}$	$Controller_{EP,II}$	$CV_{EP,I}$	$CV_{EP,II}$	$Setpoint_{EP,I}$	$Setpoint_{EP,II}$
PI-control	PI-Control	$z_{SP,I}$	$z_{SP,I}$	0.9	0.85
PI-control	P-control	$z_{SP,I}$	$z_{SP,I}$	0.9	0.9
PI-control	PI-control	$z_{SP,I}$	$z_{EP,I}$	0.9	0.9

All the control concepts should, based on the theory section, be able to work such that one extra producer is activated, at a time, when the swing producer is close to saturating. All three implementations will be implemented and discussed based on performance.

The time constants for the swing consumers should be equal, but slower than the already proposed droop and swing producer controllers, to further enact a slow, tertiary response from the extra producers. Also, since the valve input from the swing producers is used as an input for at least one of the controllers, the timescale needs to be much slower than the swing producer controller to avoid interactions.

4.3.4 Quaternary Control - Swing Consumer

The quaternary control, the swing consumer control, resemble the swing producer, however the swing consumer should only act if all producer flows are saturated and the pressure in the network drops. The easiest way to ensure this is by using a lower pressure setpoint, $P^s - \Delta P$ for the swing controller.

Using a lower setpoint could pose an issue as this would initially make the quaternary controller work towards lowering the pressure in the system, thus working against the entire producer side, which would cause unwanted instability in the system. This can fortunately be fixed, by implementing a minimum selector after the quaternary controller $c_{quaternary}$, as illustrated in Figure 4.6.

The minimum selector is set to choose the minimum of the nominal flow setpoint $q_{SC}^{s,2}$, and the setpoint from the supervisory controller $q_{SC}^{s,1}$. Since the setpoint is a lower pressure than the nominal network pressure, the quaternary controller will naturally increase the flow setpoint in order to reduce the network pressure. This is stated in the valve equation, increased flow means lower pressure. Thus, until the pressure in the network actually drops because of producer side saturation, the nominal flow setpoint will be selected. When saturation on producer side is reached, and the pressure drops below threshold, the flow setpoint will decrease in order to balance the supply and demand.

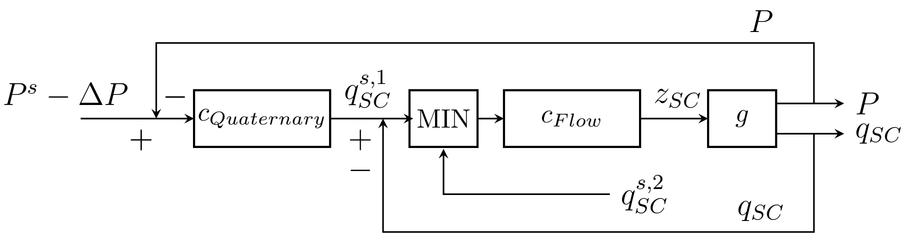

Figure 4.6: Block diagram for proposed swing consumer control.

Because of the minimum selector, the swing consumer control will not work for saturation on the producer side causing an increase in pressure, which could be an issue. However, this situation is regarded highly unlikely, and is therefore not considered.

4.3.5 Normal Consumers

The normal consumers are not implementing any supervisory controller structure elements. The normal consumers does however, implement regulatory control for the related consumer flows, which will act as disturbances to the system. As is the case for the other flow controllers in the system, the normal consumer valve positions are used as manipulated variables, and the related normal consumer flows as controlled variables. Implementing integral action alone, the controllers will ensure good setpoint tracking, as long as there is error in the controller. This will only be the case for a setpoint change if only integral action is involved for the consumer. Thus, the controller is not really regulating anything unless a disturbance is directed specifically at the normal consumer flow.

This same idea makes the normal consumer ideal for use as a system disturbance. A flow setpoint change would make the controller act accordingly and increase until the new setpoint is reached, but then stop acting again as the integral error would be zero. The concept of normal consumer control is illustrated in a block diagram in Figure 4.7.

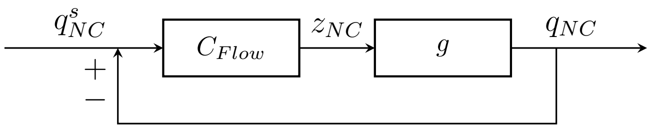

Figure 4.7: Block diagram for normal consumer control using simple flow control in steam networks.

4.3.6 Involuntary Droop Consumers

The involuntary droop consumers involve no control structure, but are merely a representation of the possible hundreds or even thousands of consumers that could exist in an actual network structure[16]. They will in the modeled steam network act as consumers that are "paid" to adjust demand in relation to the change in network pressure. Therefore the involuntary droop consumers act as disturbances as the normal consumers, and will respond according to how well the network is regulated. This is due to the fact that the network is highly coupled through the valve equation, which relate flow and pressure before and after valve opening, effectively relating all flows to the main pipeline pressure P.

Because of the relation between all consumers and producer through P, a change on any consumer or producer would therefore disturb the droop consumers without corrective action from the implemented control structures. The involuntary droop consumers therefore rely entirely on producer side control and swing consumer control. However, it is regarded less important to keep the involuntary droop consumer flows at setpoint, and therefore they are not controlled actively.

4.3.7 Combined Control Structure

The full consumer and supplier model is depicted in Figure 4.8. Note that for the droop producers and the normal consumers, only one control loop is shown to save figure space. Involuntary droop consumers are left out entirely as they are not a part of the control structure.

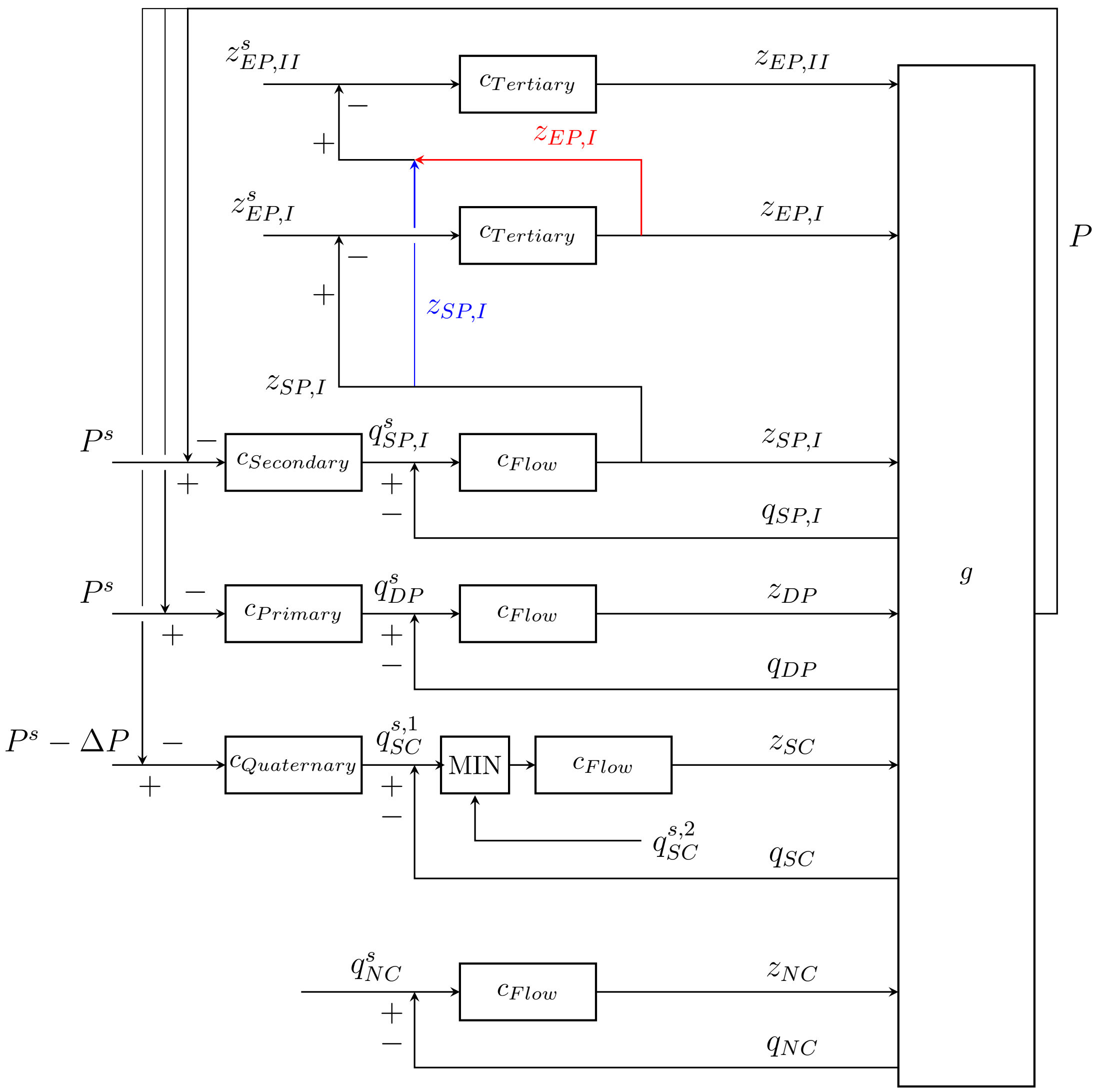

Figure 4.8: Block diagram showing a complete depiction of the decentralized control system used in this thesis. Possible configurations for the extra producers are shown in blue and red.

Figure 4.8 uses different colors representing the different extra producer configurations. The two configurations using the swing producer valve position as an input is for extra producer II is shown in blue. The extra producer configuration using the valve position for extra producer I as input for extra producer II, is shown in red.

With proper tunings for the controllers, the combined control system should be able to perform stable operation with considerable extended operating range compared to a system with no control. Open loop responses and tuning parameters are shown in the open loop and tuning implementations chapter. In the next sections, multivariable control implementation is discussed before MATLAB implementation for both controller structures are discussed.

4.4 Multivariable Control

For the multivariable control structure, the same objectives and priorities from the decentralized control structure will be utilized. Since multivariable control handles both multiple inputs and multiple outputs (MIMO) at the same time, the first priority when constructing the multivariable control structure will be defining inputs, outputs and disturbances. Then, an objective function representing the same ideas presented in the previous sections should be made. The model and objective function can then be implemented using the CasADi software.

4.4.1 Inputs and Outputs

The multivariable control structure should be as equal as possible to the decentralized control structure. This would mean using the same inputs and outputs as for the decentralized control structure. Since the model predictive control is minimizing an objective function, this is where the majority of setpoint tracking and controller objectives such as input usage prioritization are defined in this thesis.

The model predictive controller is defined such that only the pressure and the swing consumers are tracked as inputs in the objective function, meaning only they will have a setpoint. The other producers will not have an explicit setpoint but are used in an input usage cost function.

The normal consumers are also defined such that they are integral error controllers as in the decentralized structure, and they will therefore also involve setpoint tracking, however outside the objective function.

For the system outputs, the multivariable system will be somewhat different than for the decentralized control structures. Since CasADi has no way of implementing integral controllers, which should be used for the normal consumers in order to make them act as disturbances, their error signal will have to be explicitly stated as a system output. Thus, the process states in the multivariable control system will be x, stated in equation 4.1.

$$x = \begin{bmatrix} P \\ e_{NC,I} \\ e_{NC,II} \end{bmatrix} \tag{4.1}$$

In Equation 4.1, P represents the main pressure in the pipeline, $e_{NC,I}$ is the error signal from the integral flow controller for normal consumer flow I, and $e_{NC,II}$ is the error signal from the normal consumer flow II.

The flow equations are stated as algebraic equations in both the decentralized and multivariable control systems, meaning they are calculated based on the selected input variables. They are therefore also process outputs in multivariable control, but do not differ from the decentralized control structure. What differs is that since flows setpoints are not explicitly defined in this implementation except for the extra producers and the pressure, the flows themselves should be stated in an objective function to make use of the flows as outputs. The objective function is explained in detail in the next section.

4.4.2 Objective Function

The multivariable system objective function should represent the same process control objectives as the decentralized control structure. This means following the priority list where droop producers should deviate from setpoint first, followed by the swing producers, then the extra producers and at last the swing consumers. In addition the normal consumers should be implemented as simple integral action controllers, and the involuntary droop consumers should be modelled as a system calculation based on the current main system pressure P.

Because the normal consumers and the involuntary droop consumers should be treated as disturbances in the model predictive controller, they should not be a part of the objective function neither. An objective function which matches the control objectives from the decentralized controller structure and is excluding involuntary droop consumer and normal consumer control is depicted in Equation 4.2.

$$L = w_0 \cdot (P - P^s)^2 + w_1 \cdot (q_{SC,I} - q^s_{SC})^2 + w_2 \cdot (z_{DP,I})^2 + w_2 \cdot (z_{DP,II})^2 \\ + w_2 \cdot (z_{DP,III})^2 + w_3 \cdot (z_{SP,I})^2 + w_4 \cdot (z_{EP,I})^2 + w_5 \cdot (z_{EP,II})^2 \tag{4.2}$$

In Equation 4.2, $w_0, w_1, .., w_5$ represent the system weights. The objective function should be minimized, and the weights could therefore represent the cost of either setpoint deviation or the cost of input usage. They are tuning parameters, and should thus be tuned bearing in mind the input usage prioritization from the control objectives section. Both input usage cost weighting and setpoint deviation cost is proposed in the objective function in this thesis.

Setpoint deviation cost is used for pressure and swing consumer flow to ensure tight setpoint control. This is equivalent to the tracking part of the objective function. They are proposed such that a deviation from nominal pressure P^S should be minimal from the actual system pressure P using w_0, and that the swing consumer flow $q_{SC,I}$ should be minimal from that of the nominal swing consumer flow $q^S_{SC,I}$ using w_1. For this type of weighting, a higher weight value on the deviation would mean that keeping setpoint has a higher prioritization than with a lower weight value.

The remaining weights $w_2 - w_5$ are on input usage, and represent the economic cost function part of the objective function. These weights are used on all the producers, and is explicitly stating how high the cost of input usage is. That is, a higher weight on the input usage means a higher cost of use. Therefore, the extra producers valve positions $z_{EP,I}$ and $z_{EP,II}$ should have the highest weight here because the associated cost are highest, as they should be used last among the producers. The swing producers valve positions z_{SP} should have a lower weight because they should be used more easily, and last, the droop producer valve positions $z_{DP,I}$, $z_{DP,II}$, and $z_{DP,III}$ should have the lowest weights as they should have the lowest cost of input usage. In total, the input weights for the producer valve positions should be much lower than the pressure and flow setpoint deviation weights. This is because input usage of the producers should occur before the use of the swing producer.

Knowing the system inputs and outputs and the objective function, implementation in MATLAB using CasADi can be performed. In the next chapter, all open loop implementations and tuning parameters are considered for the decentralized and multivariable control structure.

Chapter 5

Control Implementations

This chapter presents the implementation specifics regarding the MATLAB and Simulink closed loop control implementation, including a system presentation and tuning results. The chapter starts by discussing the decentralized control model and selected parameters. Then, using the model and parameters, tuning results are presented first for the open and closed loop tunings from the decentralized controller structure. Last, the multivariable controller structure CasADi implementation is presented. The results from the closed loop tuned systems are presented in Chapter 6.

5.1 Decentralized Control - Time Constants

Before tuning and implementation of the closed decentralized model can commence, a selection for the τ_C values in each of the individual controller loops has to be determined to separate the primary, secondary, tertiary and quaternary control structures on different time scales. The time scale separation of the system is depicted in Table 5.1.

Table 5.1: A depiction of the decentralized control system time scale separation using different time constants.

Usage	Symbol	Value	Unit
Flow controllers (Except Extra Producers)	$\tau_{c,flow}$	6	[s]
Droop Producer controllers	$\tau_{c,DP}$	30	[s]
Swing Producer controller	$\tau_{c,SP}$	60	[s]
Extra Producer controllers	$\tau_{c,EP}$	300	[s]
Swing Consumer controller	$\tau_{c,DC}$	60	[s]

The time scaling is based on the fastest controllers, the regulatory flow controllers.

Knowing that time constants for a typical valve position controller usually lies within the range $5 - 10$s, according to the book "Process control: A practical approach" (2011)[19], the flow controller time constants with exception of the extra producers are set to 6s.

For the primary droop controller, the time constant should be five times larger than this to avoid interaction, meaning it should be at least 30s. This is because it is involved in a parallel cascade loop with the flow controllers, which requires the outer loop to be at least five times slower than the inner loop.

Since the swing producers represent secondary control, they should be slower than the droop controllers. Since they are not in a direct loop with the droop producer, they are not required to be at least five times slower as the cascade loop would require them to. Thus, the swing producer time constant is only set two times higher than the droop producers, at 60s. Note that even though they do not directly interact in a loop, they use the same controlled variable, and should therefore be separated on different time scales. They should also be separated according to the prioritization list presented in the control objectives section.

The same time constant used for the swing producer is used for the swing consumer, as they are modelled such that the time constant does not act before all other control is lost and pressure in the system drops. Therefore, it does not matter that for the extra producers, the time constant $\tau_{c,EP}$ is set five times higher than the swing consumer, at 300s.

The reason that the extra producers use a five times larger time constant than the swing producer, is that at least one of the extra producers will use the swing producer as an input. This creates a loop which could cause interaction if there is not sufficient time scale separation between the swing producer controller and the extra producer controllers.[26]

5.2 Decentralized Control - Tuning

The results from the open- and closed-loop step responses during the tuning process were analysed to determine the type of process the system dynamics represent, and thus which tunings should be used. Since the flow controllers represent the fastest controller dynamics in the system based on the time scale separation, their tuning parameters were obtained first. After obtaining all the open loop flow controller responses, tunings were implemented. The time scale separation arrangement were then used to choose the order of tuning for the next controllers.

Since the primary, secondary and quaternary controllers all share the same controlled variable, their respective closed loop responses and tunings were all obtained before tuning implementation commenced. The extra producers are on a slower timescale and uses the swing producer as an input, therefore they were tuned last, after all other controller implementations.

Each of the tuning steps are systematically presented through the next sections. For the regulatory flow controllers, and the primary, secondary, and quaternary controller tuning steps, only the tuning step response for the swing producer controllers are shown, whereas the rest of the tunings were similar in response appearance and are therefore only shown in the appendices. For the extra producer flow tunings, only the first extra producer is shown as they both had an equal response appearance.

5.2.1 Flow Controller Tunings

The first tunings were performed on the open loop model system, where a +10% step on each controllable valve position were performed at $t = 1000$s and the respective flow response were analyzed. Figure 5.1 shows the open loop tuning response for a +10% step on the swing producer valve position $z_{SP,I}$ at $t = 1000$s.

Note that the y-axis in Figure 5.1 for the swing producer valve position subplot is scaled to better display the step performed. The y-axis scale is therefore different for the remaining tunings and results. Also note that all the flow subplots and the pressure subplot are scaled to better show the flow response, differentiating those scales from the upcoming results. Scaled y-axes are used in the rest of the open loop flow responses and the rest of the tuning results. The rest of the open loop responses are shown in the appendices, which involves steps on the valve positions for the swing consumer, $z_{SC,I}$, the droop producer, $z_{DP,i}$, and the normal consumer, $z_{NC,i}$.

Analysing the result from Figure 5.1, the initial open loop response on the swing consumer flow $q_{SP,I}$ is static. The tunings were therefore performed accordingly to the simple internal model control rules for a static process, described in the theory chapter. Because the flow response is initially static, only integral action is added.

For the remaining open loop responses shown in the appendices, a static response is also observed. Therefore, all flow controllers involve only integral controller action. The tuning parameters obtained and implemented using the simple internal model controller algorithm are presented in Table 5.2.

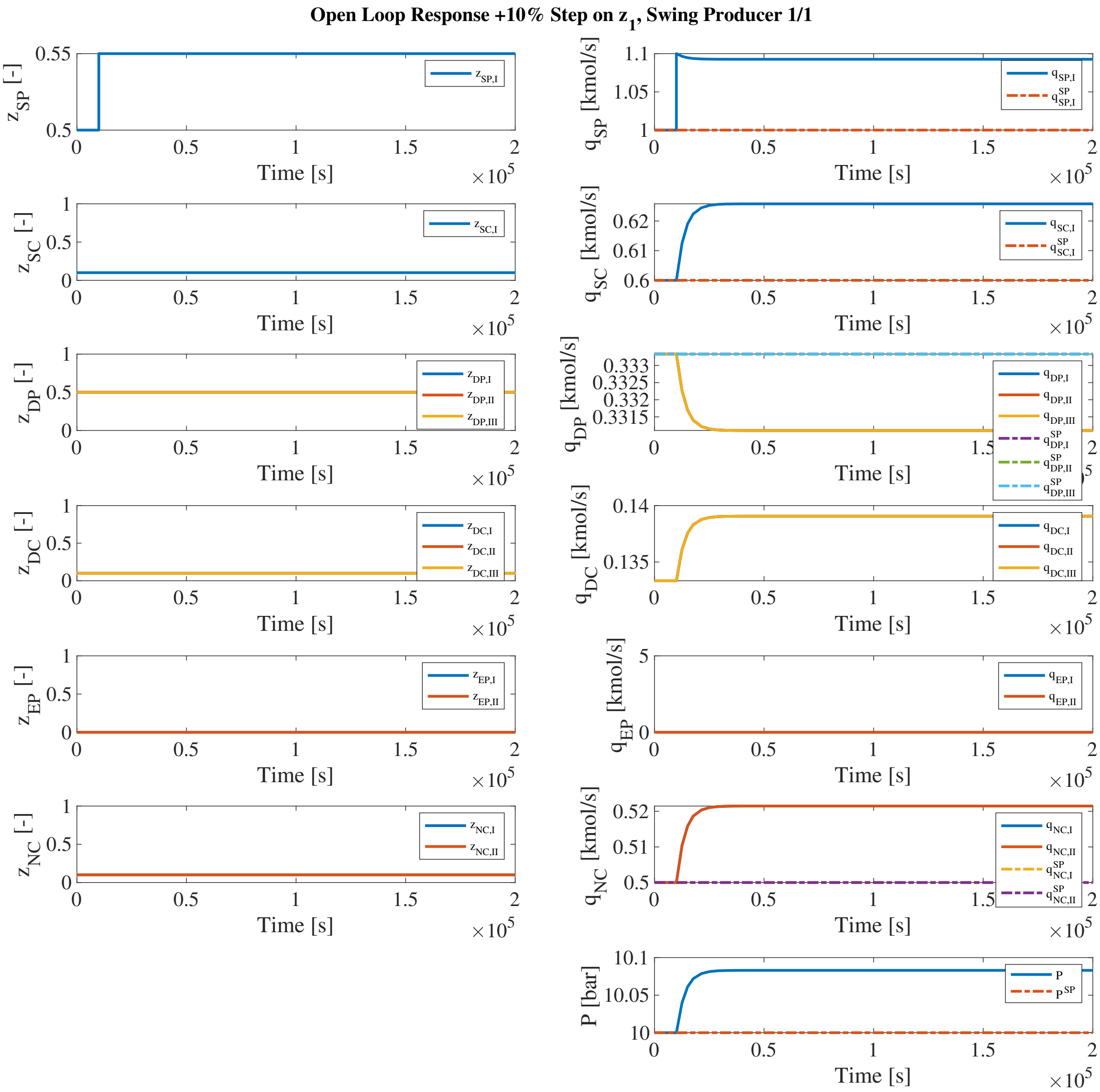

Figure 5.1: Open loop system response for a +10% step on $z_{SP,I}$, that is valve position for swing producer I, at $t = 1000$s.

Table 5.2: SIMC tuning parameters for flow controllers.

Input	Controller Type	K_p	K_i	τ_1	τ_c [s]	τ_i [s]	k
$u_1 = z_{SP,I}$	Swing Producer (1/1)	0	0.0833	0	6	24	2
$u_2 = z_{SC,I}$	Swing Consumer (1/1)	0	0.0278	0	6	24	6
$u_3 = z_{SP,I}$	Droop Producer (1/3)	0	0.250	0	6	24	0.667
$u_4 = z_{DP,II}$	Droop Producer (2/3)	0	0.250	0	6	24	0.667
$u_5 = z_{DP,III}$	Droop Producer (3/3)	0	0.250	0	6	24	0.667
$u_6 = z_{EC,I}$	Extra Consumer (1/2)	0	0.0333	0	6	24	5
$u_7 = z_{EC,II}$	Extra Consumer (2/2)	0	0.0333	0	6	24	5

Table 5.2 uses the MATLAB Simulink controller definition, which uses individual pro-

portional, derivative, and integral gains. This differs from the controller definition proposed by the simple internal model control definition, which uses a combined gain in a series form. The simple internal model control parameters are given in Equation 5.1 [23].

$$u(s) = K_c \left(1 + \frac{1}{\tau_i s}\right)\left(y_s(s) - \frac{\tau_d s + 1}{(\tau_d/N)s + 1} y(s)\right) \tag{5.1}$$

MATLAB Simulink however, represents the series controller with individual gains, rearranging Equation 5.1 on the form shown in Equation 5.2.

$$u(s) = \left(K_p + K_i \frac{1}{s} + K_d \frac{N}{1 + N\frac{1}{s}}\right) y_s(s) \tag{5.2}$$

Comparing the MATLAB Simulink and the simple internal model controller definitions, the result becomes that the proportional gain K_p is equal to K_c from the simple internal model control derivation, and the integral controller gain K_i becomes equal to K_c/τ_i. The MATLAB Simulink gain is used in all decentralized controller tunings in this thesis.

5.2.2 Supervisory Controller Tunings

After applying the tunings for the flow controllers, the primary, secondary and quaternary controller tunings should commence. They should all be tuned before the tertiary control because it is the slowest acting control in the time scale separation, however as mentioned it is not the slowest in practice, because of the minimum selector on the swing consumer. Also, the extra producer uses the valve input from the swing producer, making it an outer loop on the swing producer control. Therefore, swing producer control should be fully tuned and active before implementing the extra producer control.

The now closed loop step simulations on the primary, secondary and quaternary control involve doing a step on the pressure controller inputs, that is the flow setpoints. Then, the system state, the pressure P can be used to obtain tuning parameters.

The closed loop step responses does a +10% step on each flow setpoint for the swing producer, the swing consumer and the droop producers t = 1000s. The closed loop response for a +10% step on the swing producer flow setpoint $q^s_{SP,1}$ at t = 1000s is shown in Figure 5.2. The remaining step responses are presented in the appendices.

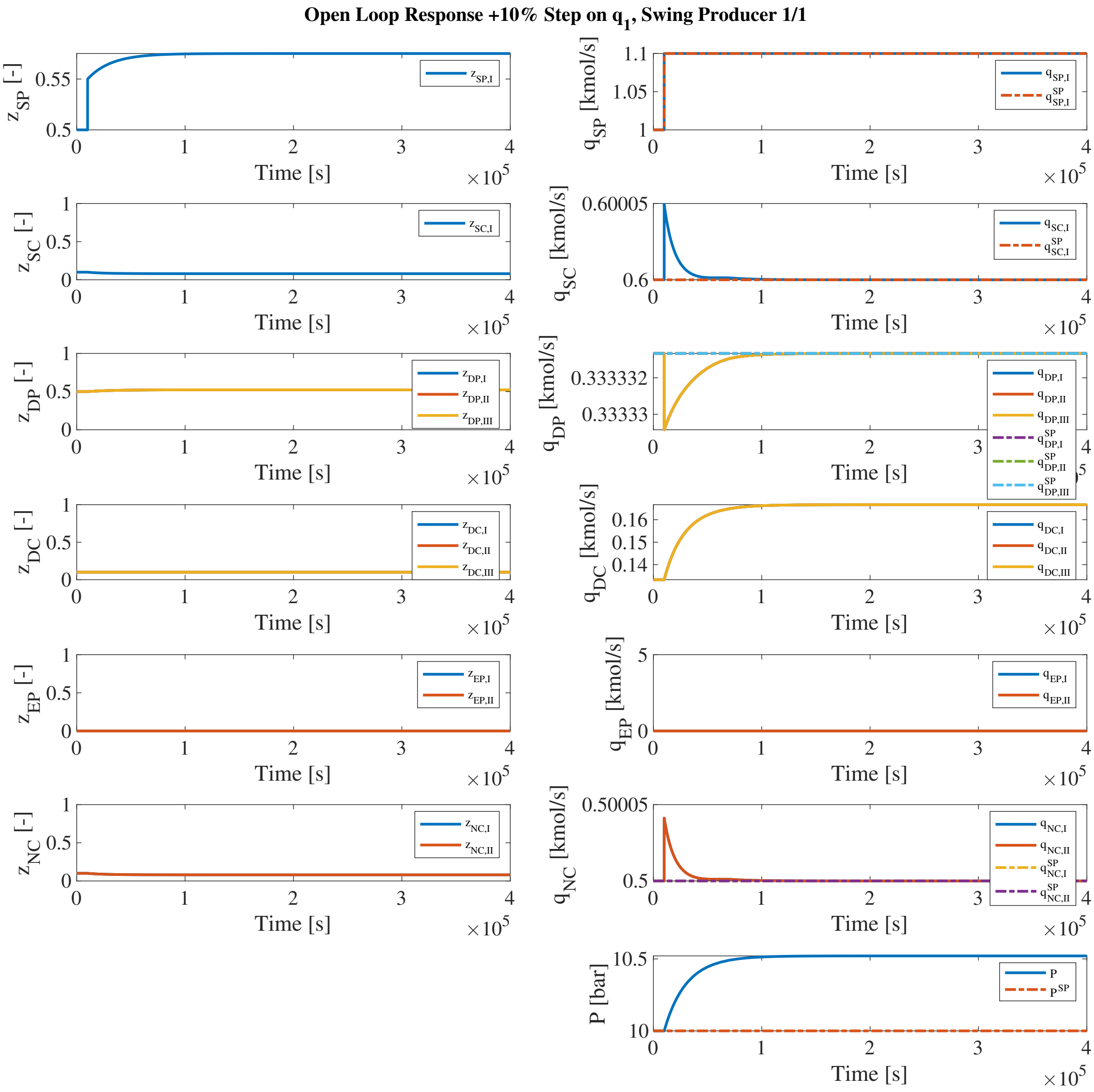

Figure 5.2: Open loop system response for a +10% step on $q_{SP,I}^{SP}$, that is the flow setpoint for swing producer I, at $t = 1000$s.

In Figure 5.2, the y-axis are again scaled for all the flows and the pressure to better display the step responses. The remaining open loop responses in the appendices are also scaled.

As for the response, a first order process response is obtained. This remains for the swing consumers and droop producers as well, however for the swing consumer the closed loop first order response for the pressure has a negative magnitude. Therefore, the sign of the tuning for the swing consumer is negative.

For the swing producer and consumer, proportional and integral action is implemented on the controllers, whereas for the droop producers, only proportional control is used, as discussed in the control implementation chapter. The tuning parameters obtained from the closed loop responses are presented in Table 5.3.

Table 5.3: SIMC tuning parameters for primary, secondary and quaternary controllers.

Input	Controller Type	K_p	K_i	τ_1	τ_c [s]	τ_i [s]	k
$u_1 = q_{SP,I}^{SP}$	Swing Producer (1/1)	65	0.271	2.03e+04	60	240	53
$u_2 = q_{SC,I}^{SP}$	Swing Consumer (1/1)	-63	-0.262	1.70e+04	60	240	47
$u_3 = q_{SP,I}^{SP}$	Droop Producer (1/3)	121	0	1.77e+04	30	120	41
$u_4 = q_{DP,II}^{SP}$	Droop Producer (2/3)	121	0	1.77e+04	30	120	41
$u_5 = q_{DP,III}^{SP}$	Droop Producer (3/3)	121	0	1.77e+04	30	120	41

After applying the tunings for the pressure and flow controllers, the extra producers tuning could commence. There are three different configurations for the extra producers, however they all use the same tunings. That is, by doing a step on the valve position on either the extra producer I or II. Then, the initial response from the swing producer is used to obtain tuning parameters. Since both the extra producers maintains equal dimensions, their tunings parameters are expected to be equal.

Thus, the tuning results were obtained by doing a +10% step on each valve position for extra producer I and II at t = 10s. Both closed loop responses were obtained before tuning implementations. The closed loop step response for a +10% step on extra consumer I valve position, $z_{EP,I}$, at t = 10s is shown in Figure 5.3. The response for a step on extra consumer II were equal and is only shown in the appendices.

To better show the initial step response, Figure 5.3 scales the swing producer valve position. Also, the y-axis for all the flow subplots are scaled, as well as the y-axis for the main pipeline pressure P subplot.

The initial response is approximately equal to a first order step response with a time delay. This means that for this process, the time delay has to be added according to the simple internal model control algorithm. The tuning parameters obtained are presented in Table 5.4. As expected they are equal for both the extra producers as the same controlled variable is used and the producers themselves include the same sizing and physical parameters. Note that for the parallel control case, the integral parameter K_p is set to zero, making it a P-controller only.

Table 5.4: SIMC tuning parameters for tertiary controllers.

Input	Controller Type	K_p	K_i	τ_1	τ_c [s]	τ_i [s]	k	θ
$u_1 = z_{EP,I}^{SP}$	Extra Producer (1/2)	-0.45	-0.019	24	300	24	-0.18	1
$u_2 = z_{EP,II}^{SP}$	Extra Producer (2/2)	-0.45	-0.019	24	300	24	-0.18	1

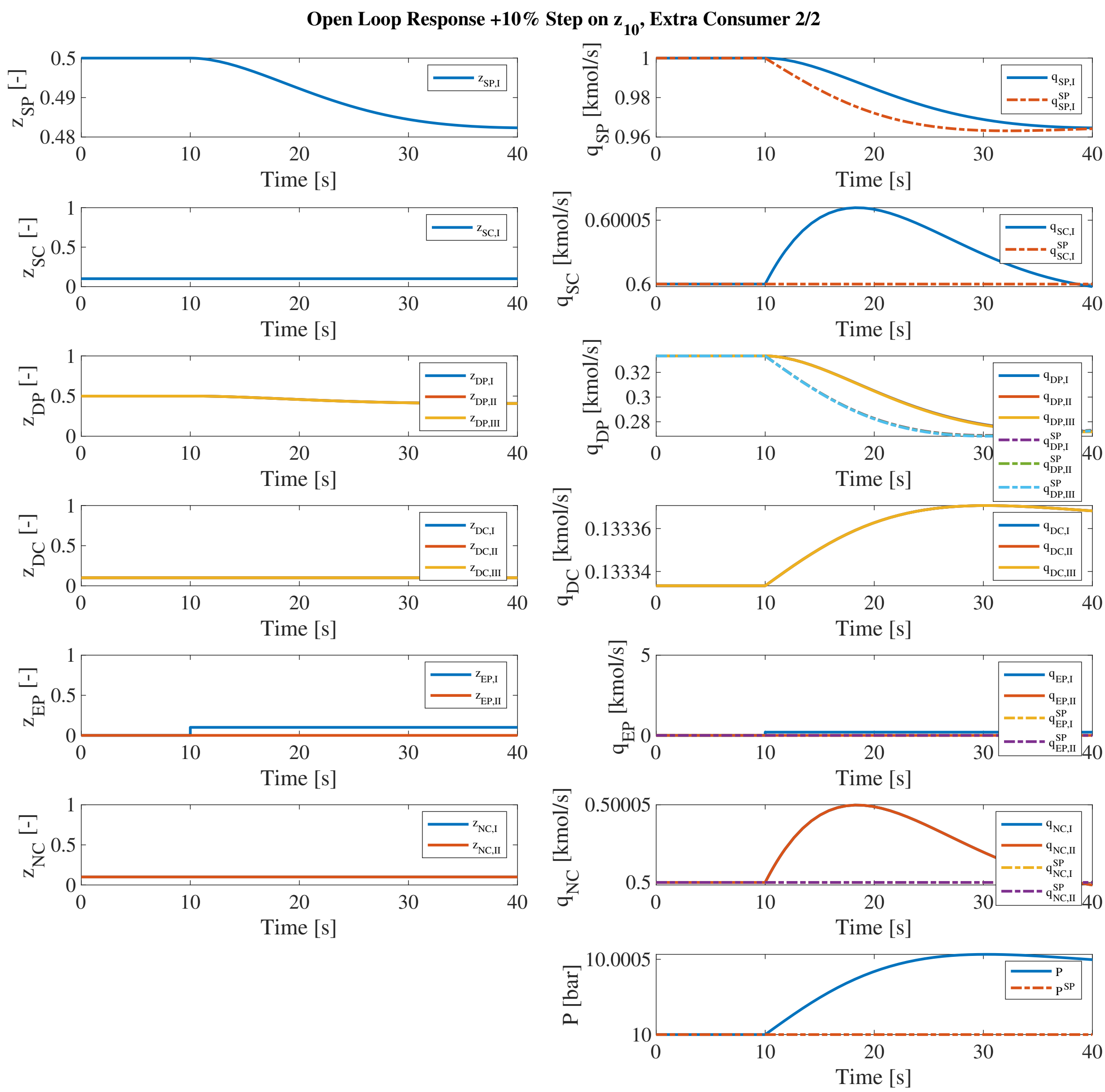

Figure 5.3: Open loop system response for a +10% step on $z_{EP,I}$, that is valve position for extra producer I, at $t = 10$s.

It should be noted that for the valve position controller, which configuration uses the swing producer valve position as an output for extra producer I and valve position for extra producer I as an output for extra producer II, another tuning is possible in theory, but not in practice.

The alternative tuning procedure involves tuning extra producer I as described above. Then the tuning should be implemented on extra producer I. After the implementation, a step on extra producer II should commence before analysing the response in the now tuned extra producer I instead of the swing producer response. This was tried out, however because extra producer I is the quaternary control, the response in the extra producer I was in fact still zero for a +10% step on extra producer II, as the droop and swing producer acted before extra producer I was required in the time scale separated control hierarchy. Therefore, all configurations use the initial swing producer valve position response as an output for controller tuning, making all the extra producer tunings equal for all configurations.

5.3 Model Predictive Control - Implementation

In order to implement the model predictive controller, two things need to be implemented. That is, first, the process plant model, which integrates the differential equation in the system and solves it to find the main pressure and consumer and producer flows. Then, a nonlinear solver must be implemented, which first should build the collocation equations for the system, then solve the objective function and output new predicted system states.

For the plant model, every fundamental dynamic variable that will change with the system or could be changed manually during simulation, should be defined as a MX expression[36]. This includes the states, the inputs, the flows, the disturbances, as well as the algebraic equations in the system.

Recall that since the extra consumers should only act as a disturbance, they are used as integral controllers. This should therefore also be the case for the model predictive controller. Recall that integral controllers integrate the received error signal. Since using CasADi makes it necessary to manually define the integral controllers, the in-built integrative functionality is used to do the integration. This means the normal consumers flow error signal should therefore be a part of the system states in addition to the main pressure. The differential equations describing the system is therefore a vector containing three elements, presented in Equation 5.3. Note that the indexing on the error signals $x(2)$ and $x(3)$ is different because the disturbances are not defined for the involuntary droop consumers.

$$\dot{x} = \frac{d}{dt}\begin{bmatrix} x(1); \\ x(2); \\ x(3); \end{bmatrix} = \begin{bmatrix} P; \\ e_{q_{NC,1}}; \\ e_{q_{NC,2}}; \end{bmatrix} = \begin{bmatrix} (R * T/V) * (q(1) - q(2) + q(3) + q(4) + q(5) - q(6) \\ -q(7) - q(8) - q(9) - q(10) + q(11) + q(12)); \\ d(6) - q(9); \\ d(7) - q(10); \end{bmatrix} \tag{5.3}$$

The system involves the previously defined ordinary differential equation for the system as well as the errors for the integral controllers. That is $d(7)$ and $d(8)$ represents nominal flows, while $q(9)$ and $q(10)$ is the actual flows for the normal consumers.

The integral controllers should use the integrated states $x(2)$ and $x(3)$ to control valve position, using a tuning gain $K_{i,1}$ and $K_{i,2}$ found in the decentralized control structure. They should also use the nominal valve position values $u_0(9)$ and $u_0(10)$. This will affect the flow because of the flow-valve relation in the model valve equations. The expressions for the valve positions for the normal consumer are presented in Equation 5.4.

$$\begin{bmatrix} z_{NC,1} = u_0(9) + K_{i,1} * x(2) \\ z_{NC,2} = u_0(10) + K_{i,2} * x(3) \end{bmatrix} \tag{5.4}$$

All consumer and producer flows are also defined as MX symbols, as they are later used to define the algebraic flow equations. The inputs, on the other hand, for the valve openings only include all nominal consumer and producer valve openings except on the normal consumers. This is because the normal consumer valve positions already are CasADi variables as the are determined by their respective system states, the integrated error signal.

5.3.1 Model Parameters

To solve the system on a moving time horizon in CasADi, some other system parameters than in the decentralized controller structure need implementing. The first thing that should be implemented is tunings for the extra controllers, then the sampling time, the prediction horizon and the number of control intervals should be defined. The parameters used in the simulations are presented in Table 5.5.

Table 5.5: Model predictive controller weight tunings.

Parameter	Description	Value	Unit
dt	Sampling Time	30	[s]
Ph	Prediction horizon	120	[s]
N	Number of Control Intervals	4	[-]
K_i	Tuning Extra Producer Flow Controllers	0.033	[-]
P_{Low}	Soft Lower Constraint Pressure	9.1	[bar]
ϵ	Back-Off from Physical Pressure Constraint	0.1	[bar]

The system involves integral action flow controllers for the normal producers, which receive the same tuning as in the decentralized controller structure, to resemble the decentralized control system as much as possible. These use a time constant of $\tau_{c,flow} =$ 6s. In order for the system to not interact with the flow controllers, the system sampling time should be slower than this, at least five times. Thus, the sampling time dt is set at 30s, meaning the system does one calculation per 30s.

The prediction horizon, which will move as the controller solves one iteration, is set at 120s, meaning that for each optimization iteration, the problem is optimized 120s forward in time. Thus, the plant can predict behaviour 120s forward in time at each time step. However, as $dt = 30s$, this only give 4 calculations per iteration. This number is also known as the number of control intervals N for the system.

Having defined the time scales of the system, nominal values and bounds need to be explicitly defined. That is physical constraints presented previously, as well as newly defined constraints. The constraints are defined in a function g in CasADi. Since flow is determined to only flow in one direction, one important constraint is a minimum main pipeline pressure of 9 [bar]. While a strict constraint of 9 [bar] could be feasible, it will most likely not be if the solver gets close to this limit. Therefore, some back-off $\epsilon = 0.1$ is added, and the minimum limit is set at $P_{Cons} + \epsilon$, at 9.1 [bar]. Using a soft constraint with back-off means the solver could actually at some points iterate at a lower value than this minimum, but are punished severely for doing so, and will correct itself if iterating below this threshold.

Having implemented all nominal values and defined the time interval, a collocation basis is implemented by constructing an empty polynomial basis. Then, an empty nonlinear problem is constructed, matching the dimensions and constraints of the problem. The empty nonlinear problem is used with the collocation basis to construct the nonlinear problem which can then be tuned and solved for closed loop disturbances.

5.4 Model Predictive Control - Tuning

Model predictive controller tuning does not have a systematic tuning procedure like decentralized control. Therefore, most of the tuning is left to trial and error. Still, some of the tuning can be determined using intuition.

$$\begin{aligned} L = w_0 \cdot (P - P^s)^2 + w_1 \cdot (q_{SC,I} - q^s_{SC})^2 + w_2 \cdot (z_{DP,I})^2 + w_2 \cdot (z_{DP,II})^2 \\ + w_2 \cdot (z_{DP,III})^2 + w_3 \cdot (z_{SP,I})^2 + w_4 \cdot (z_{EP,I})^2 + w_5 \cdot (z_{EP,II})^2 \end{aligned} \tag{5.5}$$

Recall the objective function from the controller structures chapter, restated in Equation 5.5. Using the controller objectives and the objective function, an idea of the tuning parameter dimensions can be constructed using intuition. For example, the tuning parameters for the tracking part of the model predictive controller, determining how important pressure setpoint and swing consumer setpoint tracking is, should be given a larger weight than the economic tracking, as the inputs used in economic tracking should be given up more easily. The weights obtained are presented in Table 5.6.

Table 5.6: Model predictive controller weight tunings.

Weight	Related Coefficient	Value	Description
w_0	$(P(k) - P^s(k))^2$	100	Pressure Setpoint Tracking
w_1	$(q_{SC,I}(k) - q^s_{SC}(k))^2$	10	Swing Producer Setpoint Tracking
w_2	$(u_{DP,II}(k))^2$	0.0001	Droop Producer Input Cost
w_3	$(u_{SP,I}(k))^2$	0.0003	Swing Producer Input Cost
w_4	$(u_{EP,I}(k))^2$	0.1	Extra Producer I Input Use Cost
w_5	$(u_{EP,II}(k))^2$	1	Extra Producer II Input Use Cost

The tuning parameters shown in Table 5.6 were obtained both using an open loop simulation involving no steps, as well as a closed loop simulation involving steps. Intuition would say that the pressure tracking weight should be the largest, the swing consumer weight the second largest, and the input usage weight lowest. That is, the droop producer should have the lowest weight of them all, the swing producer the second to lowest, and the extra producer should have weights in between the setpoint tracking and the swing and droop producers. This is in accordance to the prioritization made in the control objectives section.

The intuitive results did fit in the trial and error simulations. In order for the closed loop system not to give up pressure control completely, the pressure weight w_0 had to be the largest weight by far, or else the system would give up pressure control instantly, leading it towards the minimum constraint of 9.1 $[bar]$. The rest of the coefficient also fitted with intuition, with $w_1 >> w_5 > w_4 >> w_3 >> w_2$. That is, the extra producers requires have individual weighting, as they should not be used simultaneously if they are not both required. Since the decentralized control structure used equal tunings for the droop producers, all the droop producers use the same weightings.

Using the tuning weights, a model predictive controller using the input prioritization list was created. Figure 5.4 shows the open loop response with no steps performed on the model predictive controller.

For a decentralized control structure, with no steps performed, it is expected that the system is at steady state conditions from the simulation start. However, because of

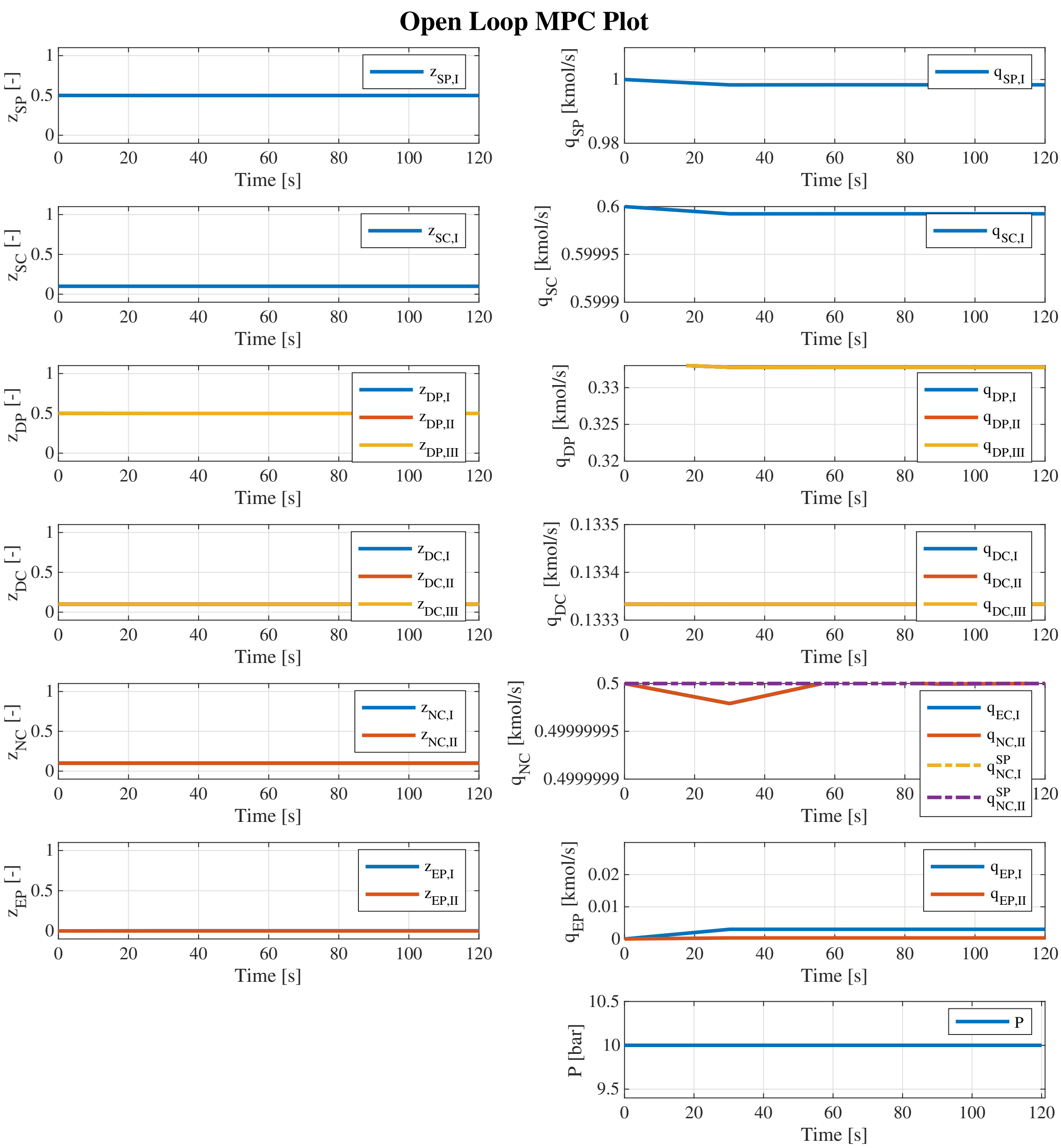

Figure 5.4: No step simulation for the tuned model predictive controller.

numerical errors in the solver in the nonlinear model predictive controller, some time is required before steady state conditions are reached, as observed in Figure 5.4.

Note that the errors are minimal, and the y-axis for the flows are all scaled to show the minimal numerical error. For the decentralized control structure, the numerical steady state error is however zero, and cannot be shown by scaling the plot like with the model predictive controller.

Note that the extra producers have the largest numerical errors. This is because the model predictive controllers tries to make use of all available inputs at all times. Using controllers that should be off at nominal point can therefore be hard when using model predictive control.

Having implemented and tuned all controller structures, the next chapter presents the closed loop results for disturbances on the decentralized and multivariable control structures.

Chapter 6

Results

The result chapter is divided into two parts. The first part presents the results obtained from the decentralized control structure, that is the closed loop responses with disturbances. Then, the centralized control structure closed loop disturbance responses are presented with the same disturbances as for the decentralized control structure. In the last part, a comparison between the decentralized and centralized controller structure are discussed.

6.1 Decentralized Control Performance

The decentralized control performance section is split in different parts, each testing the system for disturbance rejecting abilities. The first part changes the normal consumer flow setpoints, causing a system disturbance, to see the system response for a different disturbances to help determine how the extra producer control structure performs according to the prioritization list from the control objectives section. Then, the control systems ability to reject pressure disturbances from the droop producer, the swing producer, the swing consumer, and last a combined producer side disturbance.

For disturbances from the droop producer and the swing consumer, no extra producer action is involved. Therefore, the closed loop responses are equal independently of what extra producer controller structure is involved. Because of that, only one simulation plot is shown for those disturbances.

For the remaining disturbance simulations, the extra producers are utilized and therefore all three extra producer controller structure responses are shown, denoted by parallel control, controllers with different setpoint and valve position control.

6.1.1 Small Normal Consumer Flow Disturbance

This section presents results originating from closed loop step simulations where a +4.5 $[kmol/s]$ disturbance on the flow setpoint for normal consumer I and a +0.5 $[kmol/s]$ disturbance on the normal consumer II are introduced. Both disturbances are introduced at $t = 1000$s, and reverted at $t = 2000$s. The results are presented for all three extra producer configurations, including parallel control, controllers with different setpoints and valve position control. They are presented in Figures 6.1 through 6.3. The results are discussed in the next paragraphs for each controller structure.

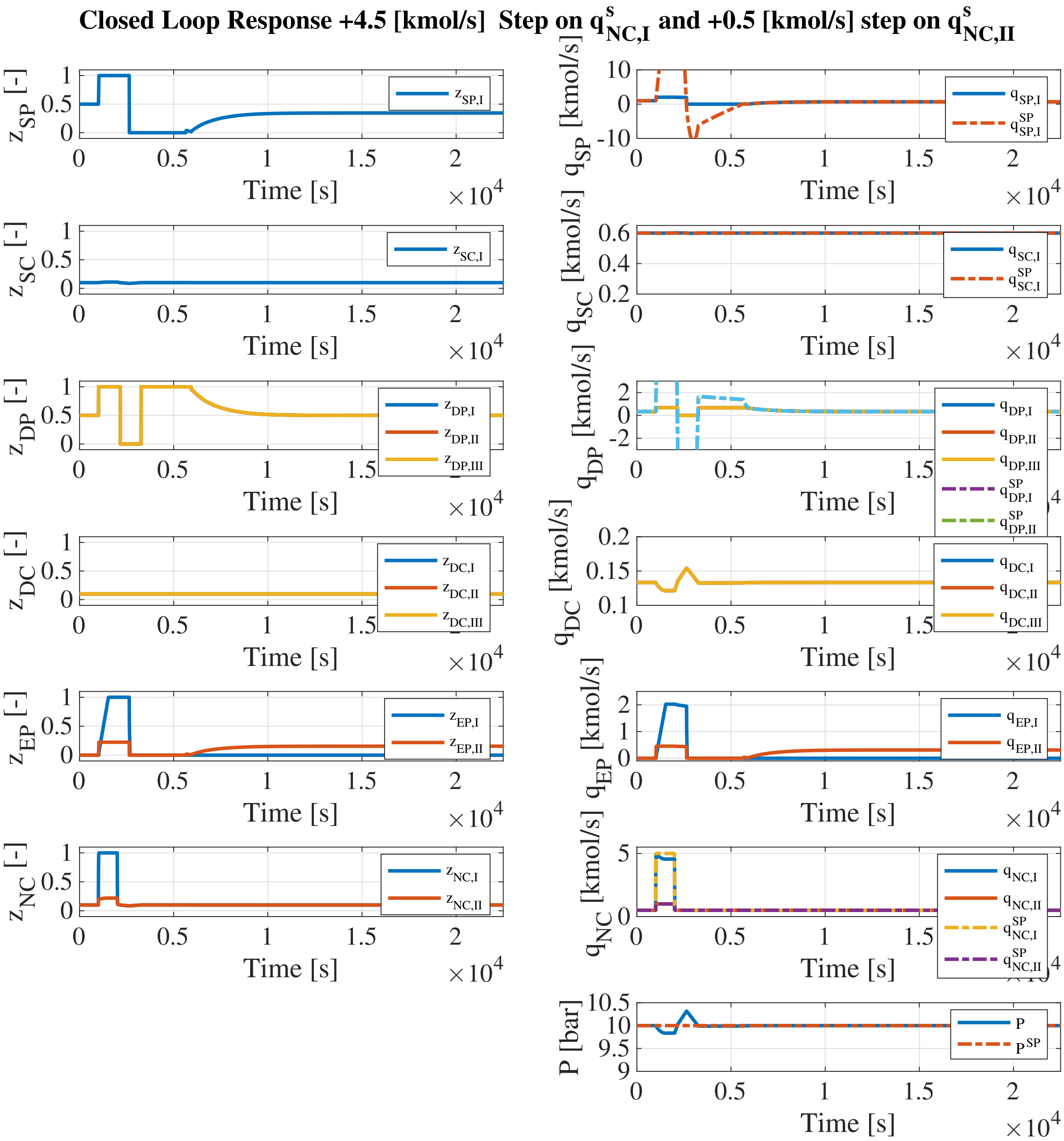

Figure 6.1: A +4.5 $[kmol/s]$ disturbance on $q^s_{NC,I}$ and +0.5 $[kmol/s]$ on $q^s_{NC,II}$ at $t = 1000$s, both reverting at $t = 2000$s using parallel control.

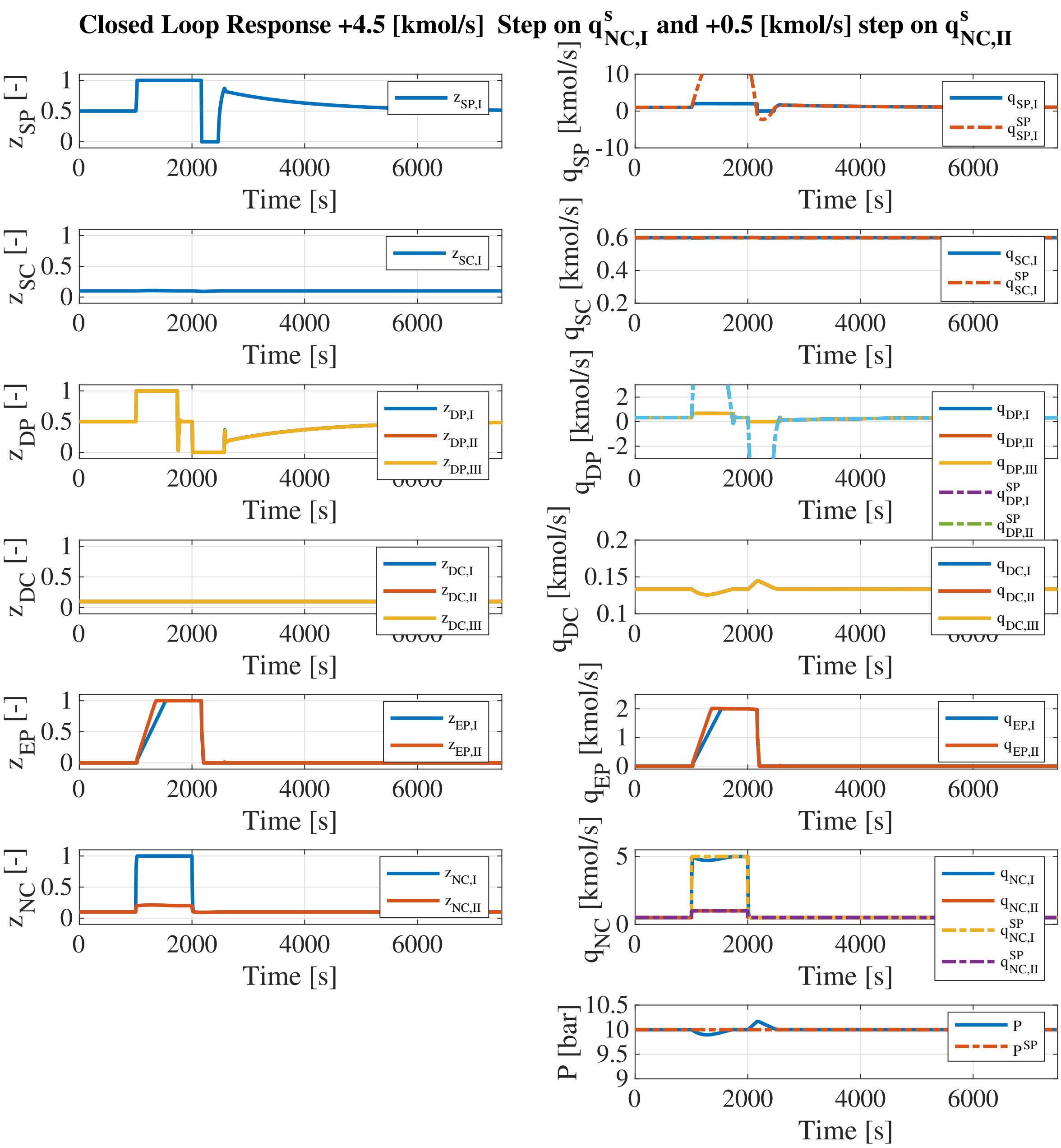

Figure 6.2: A +4.5 $[kmol/s]$ disturbance on $q^s_{NC,I}$ and +0.5 $[kmol/s]$ on $q^s_{NC,II}$ at $t = 1000$s, both reverting at $t = 2000$s using controllers with different setpoints.

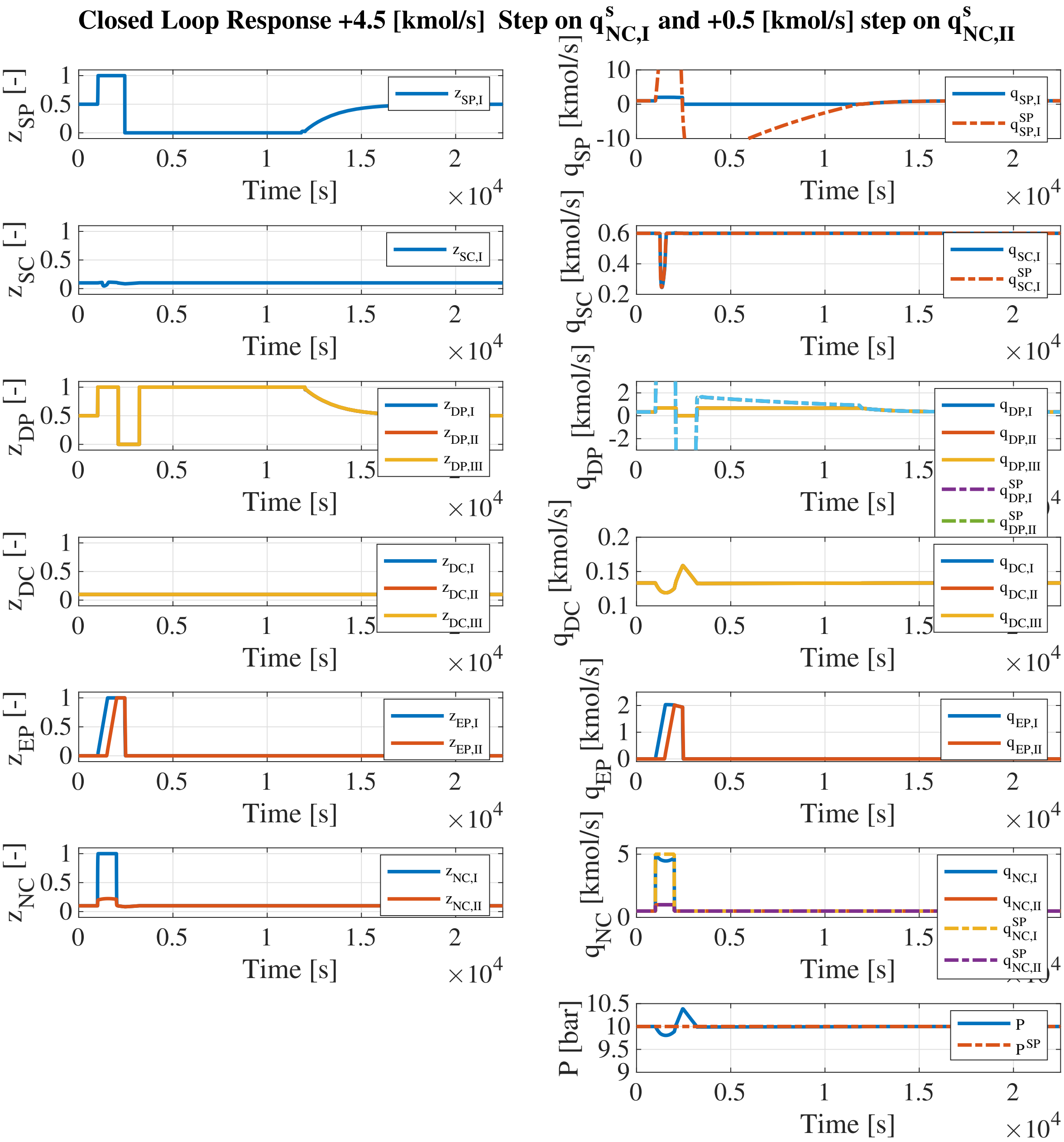

Figure 6.3: A +4.5 $[kmol/s]$ disturbance on $q^s_{NC,I}$ and +0.5 $[kmol/s]$ on $q^s_{NC,II}$ at $t = 1000$s, both reverting at $t = 2000$s using valve position control.

Parallel Control Responses

Figure 6.1 uses parallel control with one P- and one PI-controller with the same set-point, both using the swing producer valve position as a controlled variable. The initial valve position responses from Figure 6.1 are shown in Figure 6.4. Since the step is performed at $t = 1000$s, no controller action is performed before this. From $t = 1000$s it can be observed that as expected, the first producer to act in order to restore nominal system values, will in fact be the droop producers. That is, because all three droop producers are identical both in nominal system sizes and tunings, they will react identical to the disturbance and there is no difference between them whatsoever.

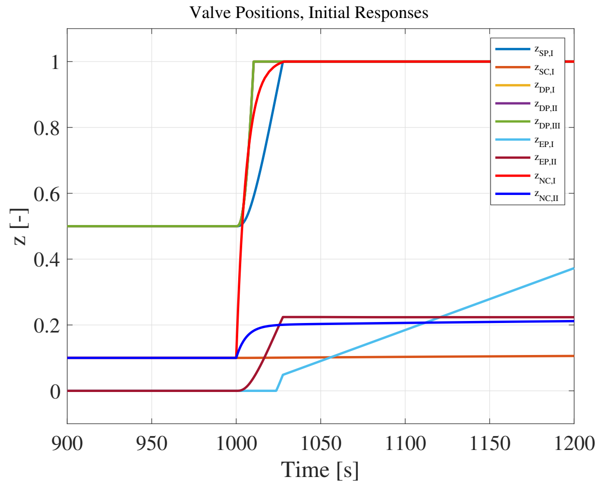

Figure 6.4: Enlarged initial response for a +4.5 $[kmol/s]$ disturbance on $q^s_{NC,I}$ and +0.5 $[kmol/s]$ on $q^s_{NC,II}$ at $t = 1000$s, both reverting at $t = 2000$s using parallel control.

Because all the droop producers all saturate almost immediately, that is within 20 seconds, due to the magnitude of the disturbance, the swing producer starts acting rather quickly too, and saturates within 30 seconds. Since the disturbance size is large, the swing producer will quickly saturate as well, encouraging the extra producers to start acting.

The extra producers are in reality also acting before the extra producers reach the valve position setpoint for the extra producers, since error is also accumulated when the swing producer is below the extra producer setpoint, 0.9. However this causes a negative deviation, and therefore the controller will try to do a negative valve position which is not possible because of the implemented controller saturation limit.

The configuration however, is not working well for the intended use. The intended use is for both the extra producers to operate at maximum capacity before utilizing the swing consumer. Using one PI-controller and one P-controller does not initiate this behaviour. The PI-controller works as expected. It will gradually act and increase the use of its input, valve position for extra producer I. This is because the accumulated error integral will grow as long as the swing producer is saturated and a positive deviation from the setpoint is caused, which will accumulate the error signal, increasing the flow setpoint for the swing producer indefinitely as long as the swing producer is saturated and the main pressure is not at nominal conditions.

So, the PI-controller is not really a problem, the problem lies within the proportional controller controlling valve position on the extra producer II. Since it is only a proportional controller with a proportional gain, it will act according to the error from setpoint, which is in fact $e = 1 - 0.9 = 0.1$. For proportional action this error will only be multiplied with the proportional gain, and proportional action is taken for this change. However, when the swing producer is saturated, the error does not change, which for proportional control causes no further action to be taken, causing an offset from utilizing extra producer II.

Due to the proportional action, this controller structure is not very useful for two reasons. Number one is that both the extra producers will start acting almost at the same time, even when not required. Since operating extra producers could be very costly and using them should be limited, if possible, only one extra producer should be utilized. However this is not the behaviour of this control structure. Here, both extra producers will start acting at the same time to different extents.

The other reason is the fact that using only proportional gain on a variable which should be close to or actually saturating, will cause poor performance, as in this case, where it causes the second extra producer flow not to be fully utilized. The utilization could in this case have been better using a larger proportional gain, but that would only improve this specific case and could cause issues for cases with other disturbances. This controller structure is therefore not very useful for further research.

It should also be mentioned that the swing consumer actually does a minimal valve opening correction in this configuration, using the regulatory control layer because of imbalance in the supply. The deviation is however very small, as can be seen on the y-axis scale for the swing producer valve position.

Opening the swing consumer valve is not normally expected to happen, because of the minimum selector, which prevents the swing producer from acting based on pressure correction unless it drops below the setpoint for the swing consumer. The concept is the same as for the extra producers, but instead of saturation limits the minimum selector is used.

That is, for a main pipeline pressure above threshold for the swing consumer pressure controller setpoint, which is 9.8 $[bar]$, the pressure controller will try to increase the swing consumer flow setpoint. However, because of the minimum selector, the controller cannot obtain a higher setpoint than the nominal flow, which is 0.6 $[kmol/s]$. But, once the setpoint is below 9.8 $[bar]$, the sign of the error will change and the pressure controller will decrease the flow setpoint, making the selector choose the minimum setpoint. Thus, the only swing consumer controller action in this simulation is corrective action from the flow controller to maintain nominal flow.

It is therefore, for the other simulations instead more likely that the valve will decrease valve opening, that would be if the supply is not enough to maintain the threshold setpoint set for the swing consumer. However in this case, the swing consumer flow drops minimally, and a minimal corrective action is taken by the flow controller.

For the droop and swing producers, the pressure controllers are what mainly forces the valve saturation, as seen by the increased flow setpoints the droop producer and swing producer. The setpoint increases also indicate that in order for the droop and swing producers alone to bring the system back to nominal conditions, a large increase in producer flow is required. This is the reason why the extra producers are modelled with large available capacity.

It is observed that the setpoint for the swing producer keeps increasing, seemingly in a linear fashion. This is caused by the integral action on the pressure PI-controller. Since the pressure is below nominal value, the error keeps accumulating in this controller and thus the setpoint keeps adding corrective measures. The setpoint for the droop producer flows on the other hand, stabilizes after a while, as the controllers only involve proportional action, which does not accumulate the error.

When the disturbances from the normal consumers returns to nominal values in Figure 6.1 at t = 2000s, the controller undershoots instead of overshooting. This happens because initially, the swing and droop producers closes their valves fast because of the sudden lack of demand from the normal consumers. However, the extra producers senses the lack in demand, causing them to fully shut, too. Thus, the entire producer side is shut for a moment. Since the undershoot causes a rather large pressure increase, the negative error in the swing producer becomes large, keeping it shut longer than the droop producer, which acts faster as it only involves P-control compared to PI-control on the swing producer. The droop producers are also tuned on a faster timescale.

When the error signal returns to nominal value for the swing producer, the valve opens again. However, for the P-controller on extra producer I, which only detects a sudden change from setpoint in the swing producer valve position, this will also react. Thus, the end state of the system has the swing producer valve position deviating from nominal value and the extra producer also deviating from nominal value. This is really undesired, again proving that this extra producer configuration is in fact insufficient when controlling the steam distribution network.

Controllers with Different Setpoints Responses

The response for the controllers with different setpoints configuration, using the swing producer valve position as a controlled variable for both input and two PI-controllers with different setpoints is presented in in Figure 6.2.

Enlarging the initial valve position responses in Figure 6.2, displayed in Figure 6.5, shows that as expected, the droop producers act to maintain system nominal conditions, followed by the swing producers, then the extra producers, and last a minimal action on the swing consumer, which is then returning to nominal when both extra producers are saturated. The discussion for the droop producer and swing producer initial action follows the same arguments as the previous section and are not discussed in this and the following sections.

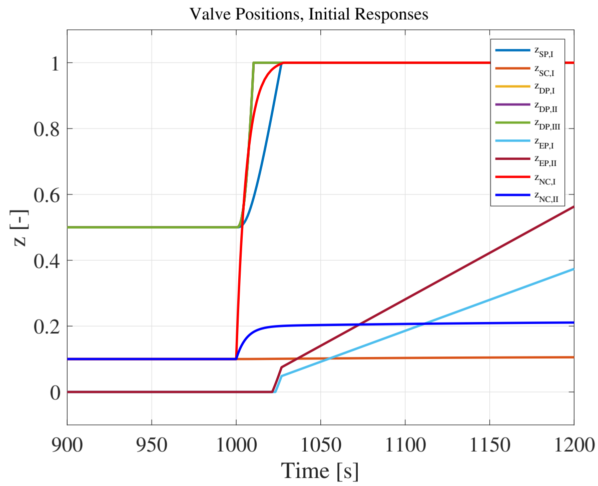

Figure 6.5: Enlarged initial response for a +4.5 $[kmol/s]$ disturbance on $q^s_{NC,I}$ and +0.5 $[kmol/s]$ on $q^s_{NC,II}$ at $t = 1000$s, both reverting at $t = 2000$s using controllers with different setpoints.

The fast response in the system poses some problems. Because the producer side act very quickly, an overshoot is caused, forcing a too large supply in the system. This causes the droop producers to return to nominal value just before the reverted disturbance at $t = 2000$s, but in fact the extra producers should return to nominal value before the droop producers. The cause for this is in fact the proportional controllers on the droop producers, and the fact that they are the fastest acting controllers in the system.

The overshoot also causes problems with the extra producers, as they should not remain open if there is available flow from the swing producers and the droop producers, as stated in the prioritization list. The problem with this is in fact that the swing producer acts to slow to revert back after the overshoot, so the droop producer creates a new nominal system. And as long as the swing producer is saturated, so will the extra producers because the swing producer valve position is used as a setpoint for them.

A solution to this could be to make the swing producers even faster compared to the primary control. However this would pose an issue for normal operation as it would make the swing producer act faster for smaller disturbances as well. Thus, the prioritization list would not be feasible.

Another, better solution solution to this could be the implementation of a maximum limit and minimum limits on the swing producer flow setpoint, as the error gets fairly large which takes longer time to revert, say for a supply overshoot. This could make the swing producers return faster to nominal value, making also the extra producers return to nominal values easier.

Maximum and minimum limits on the flow setpoints for the droop producers and a minimum limit to the swing consumer could also be a good idea, since the flow setpoints are reaching multiple times higher than the flow maximum value. Implementing maximum and minimum limits would require adding back-calculation to the primary and secondary and controllers as well, as well as modifying the minimum selector on the swing consumer.

Since the controllers have different setpoints, the error will be different for saturation in the swing producer, that is, for the extra producer I, which has a setpoint of 0.9, the error will be smaller than for the extra producer II, which has a setpoint of 0.85. Therefore, as observed in Figure 6.2, the extra producer II will actually saturate faster than extra producer I, as the controller error is in fact larger.

The disturbances revert at $t = 2000$s. The system now undershoots, as it previously did an overshoot, which causes for the droop and swing producers to saturate in closed positions. This is because as for the initial disturbance, the reverted disturbance is large. The undershoot causes error to accumulate in the controllers because the network pressure becomes to high. When the pressure network is stabilized again, the swing producer controllers will slowly revert their setpoint to nominal values, making the system return to nominal value.

The reason for the imbalance and undershoot is caused by the extra producers. Because the swing producer closes immediately along with the droop producer as the supply demand is rapidly decreased, the error signal input to extra producer I and II is very large, making it act very quickly. Thus, the system suddenly undershoots. This causes for large error in the swing producer and droop producer controller, however larger in the swing producer as it also involves integral error which accumulates to a larger sum than the droop producers.

Now, after the extra producers close, the system is able to revert to nominal conditions. As mentioned, the imposed disturbance is large. Therefore the system takes a long time to revert. It is however a system weakness that the system takes very long to return to nominal values after a large negative disturbance. However, this is not expected, especially not a sudden drop in the extra producer supply. The disturbances are rather expected to increase slowly over time, and the cases are only meant to show the handling of the disturbances.

The reason this structure reverts both the extra producers are that they both involve PI-controllers. This structure works better than the parallel control configuration because of this, which does not utilize the P-controller. Recall that for parallel control, the system is not able to revert itself after the disturbance is induced or reverted.

Valve Position Control Responses

The results in this subsection uses the configuration with two PI-controllers with the same setpoint, where extra producer I are using the swing producer valve position as a controlled variable and extra producer II uses the valve position for extra producer I as a CV. The result for a +4.5 $[kmol/s]$ disturbance on the flow setpoint for normal consumer I and a +0.5 $[kmol/s]$ disturbance on the normal consumer II at $t = 1000$s, both reverting at $t = 2000$s is presented in 6.3.

Zooming in on the initial valve position responses in Figure 6.3, shown in Figure 6.6, it is observed that the system does as expected for using the droop producer and then the swing producer, as for the other configurations. The structure then utlizes both the extra producers. However, this controller structure has the largest deviation in the swing consumer nominal flow, which is not optimal as this should be the last input used, and completely avoided from use if possible.

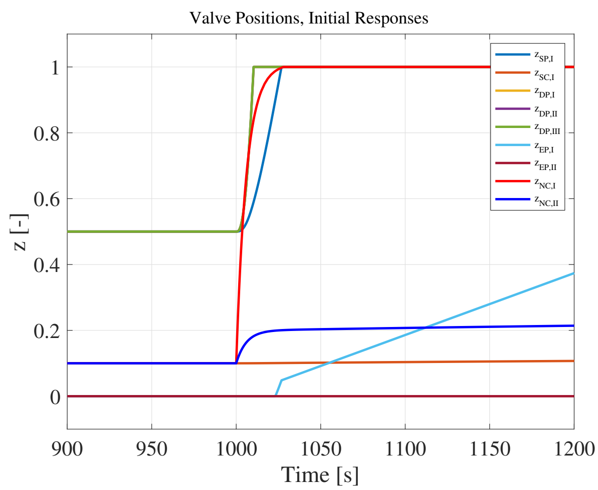

Figure 6.6: Enlarged initial response for a +4.5 $[kmol/s]$ disturbance on $q^s_{NC,I}$ and +0.5 $[kmol/s]$ on $q^s_{NC,II}$ at $t = 1000$s, both reverting at $t = 2000$s using valve position control.

The reason the swing consumer is used more than for any of the controller structures is the extra producer configuration. The valve position control on the extra producers are configured such that one manipulated variable should be utilized at a time, with the exception of the required back-off in the controllers. This makes them very slow acting when both are required in an instant, and for a very large disturbance, the control structure is to slow to instantly handle the disturbance. The valve position controller does however, fully utilize both extra producers, returning the swing producer to nominal value, so it works better for this system than parallel control.

Solutions to the slow extra producer controller action could be larger controller gains on the extra producers, but that could cause interaction with the droop and swing producer control, causing the extra producers to act too early. This is therefore a trade-off, because in this structure the extra producers are performing best of all the controller structures. The first extra producer I starts acting when the swing producer valve position is about to saturate. Then, when extra producer I is about to saturate, extra producer II starts production, effectively doing MV-MV switching with a small buffer zone where both manipulated variables are active. This is done when the droop producer and swing producer saturate, which is what the control structure intends to do. However, in the model in this thesis, the controllers with different setpoints would work better.

When the disturbance reverts to nominal values, the response is equal to the one in the controllers with different setpoints. However, the controllers act on a slower timescale as the negative swing producer error signal is actually much larger than for the different setpoint configuration, thus using more time to return to nominal values.

The reason why the magnitudes for the error in the swing producer is larger for valve position than when using different setpoints, is because the offset in the pressure P is larger. The offset is larger because the extra producers act slower in utilizing their full supply, causing a system imbalance for a longer time, giving more error accumulation time for the swing producer PI-controller. Thus, the different setpoints configuration is preffered over the valve position controller.

6.1.2 Large Normal Consumer Flow Disturbance

This section presents an even larger disturbance to the normal consumer flow setpoints. A +4.5 [kmol/s] disturbance on both the normal consumers is introduced at $t = 100$s, which is not reverted due to the fact that the disturbance is so large that is not expedient to show the reverting process. The reasons why it is not reverted is that 1) the reverting time process takes thousands of seconds, which would make the plot very hard to read and 2) the reverting process is equal to the one in the first disturbance plot, on a longer timescale. The results are discussed in the next section.

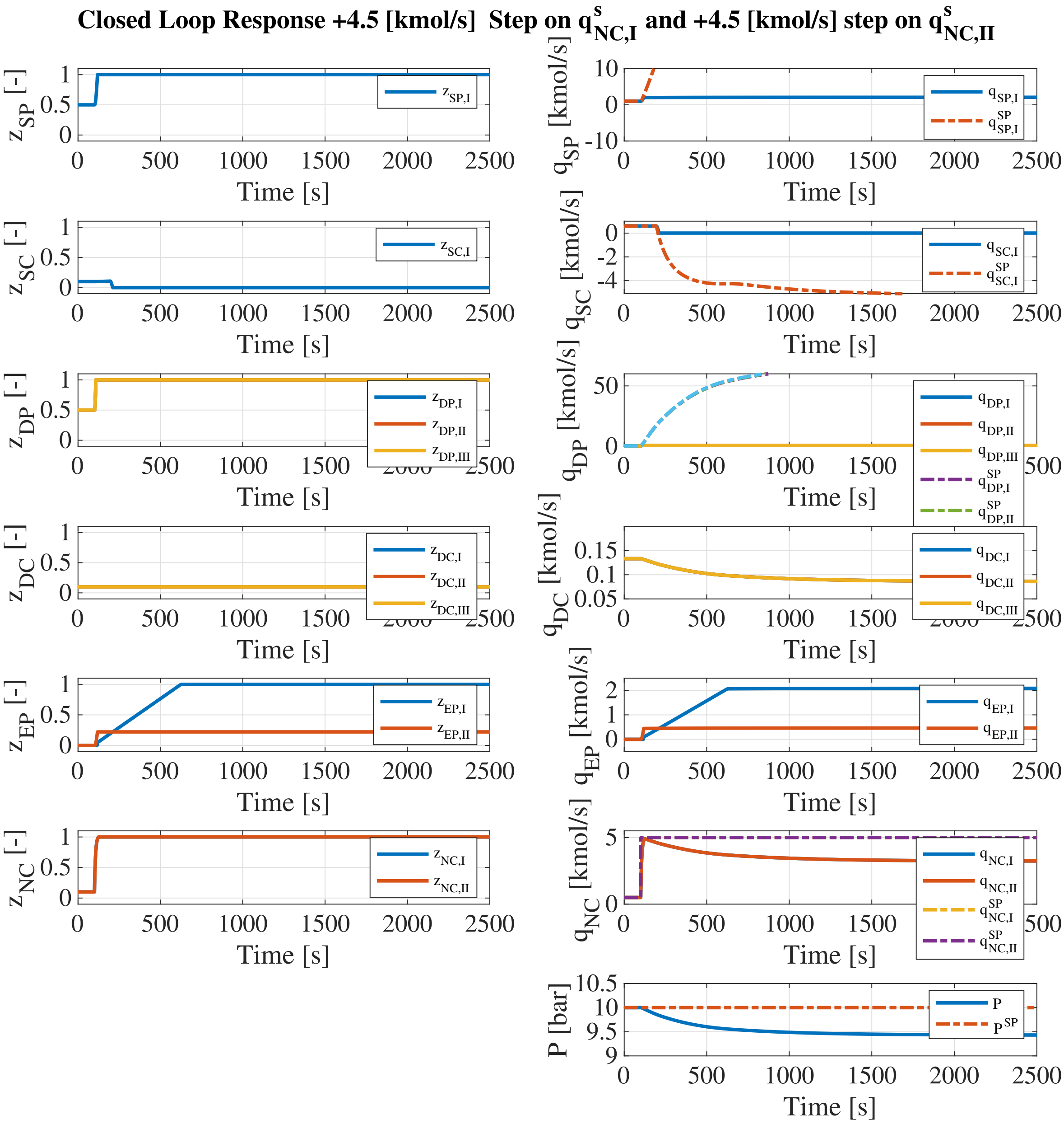

Figure 6.7: A +4.5 $[kmol/s]$ disturbance on $q^s_{EC,I}$ and $q^s_{EC,II}$ at $t = 100$s using parallel control.

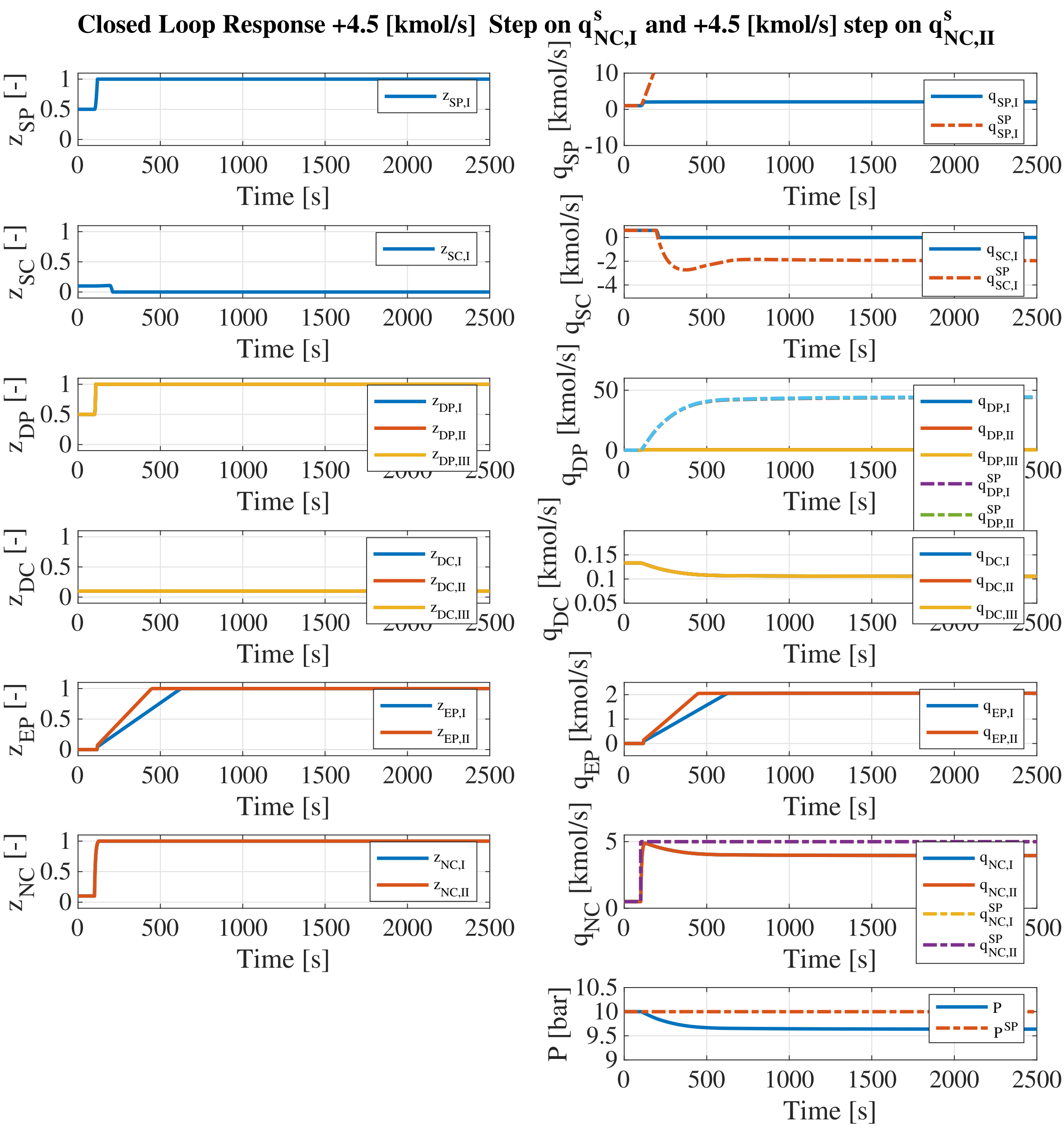

Figure 6.8: A +4.5 $[kmol/s]$ disturbance on $q^s_{EC,I}$ and $q^s_{EC,II}$ at $t = 100$s using different controller setpoints.

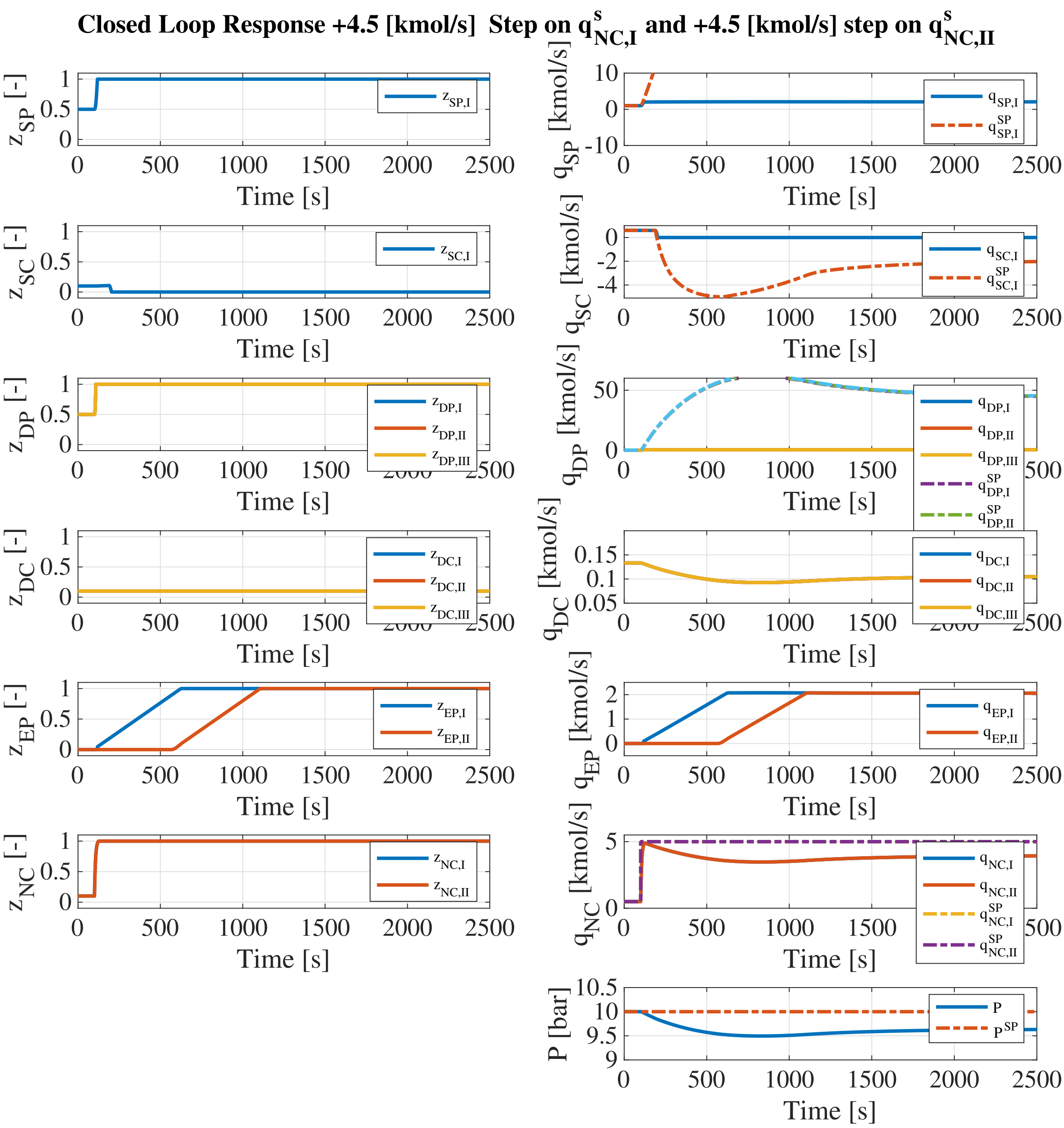

Figure 6.9: A +4.5 [$kmol/s$] disturbance on $q^s_{EC,I}$ and $q^s_{EC,II}$ at $t = 100$s using valve position control.

Parallel Control Response

Figure 6.7 shows a large normal consumer disturbance for the parallel controller configuration. The response is equal to the initial response in Figure 6.1, except for the fact that the disturbance now is so large that the pressure in the network now drops below the setpoint for the swing consumer, which is 9.8 [bar], causing the swing consumer to fully close.

The behaviour from the swing consumer is as expected. It is the slowest acting controller in the system, because of the shift in pressure setpoint, and is not acting before all other flows are used at maximum capacity. That is, with exception of the extra producer II, which is not working properly because of the proportional action.

Since the swing consumer flow is rather small compared to the entire flow system, the closing of the swing consumer is not enough to restore system balance. However, the process has a self stabilizing effect, which happens some time after all producers and the swing consumer are saturated.

Controllers with Different Setpoints Response

Figure 6.8 shows the response for a +4.5 $[kmol/s]$ change in flow setpoints on both normal consumers at $t = 100$s using controllers with different setpoints. The response is equal to the one in Figure 6.2, except for the fact that the pressure in the network now drops below the setpoint for the swing consumer, which is 9.8 [bar], causing the swing consumer to fully close.

The response fully utilizes the extra producers, but the disturbance is so large that all control is lost because of controller saturation. The structure is working according to the control objectives, as the swing consumer saturates when all control is lost to increase network pressure. The disturbance is however so large that the system is not restored to nominal conditions. Comparing to the parallel control, the new steady state pressure is closer to nominal for this configuration, making it a better control alternative.

Valve Position Control Response

Figure 6.9 shows a a +4.5 $[kmol/s]$ change in flow setpoints on both normal consumers at $t = 100$s. The response is equal to the initial response in Figure 6.3, except for the fact that the pressure in the network now drops below the setpoint for the swing consumer, which is 9.8 [bar], causing the swing consumer to fully close.

Figure 6.9 confirms that also for this controller structure, the extra producers are fully utilized, however on a much slower timescale than the controllers with different setpoints structure, and they are still only being used one at a time.

6.1.3 Droop Producer Pressure Disturbance

To verify the decentralized controller structures system stability, disturbances should also be performed on the pressures on the producer and consumer sides. Therefore, disturbances on the droop producer, the swing producer, the swing consumer and last a combined producer side disturbance is performed. For the first disturbance simulation, poses a -4.5 $[bar]$ disturbance on droop producer I at $t = 5000$s. Then, at $t = 20000$s, the pressure is increased by $+4.5$ $[bar]$ on droop producer I, allowing for the system to retain to nominal values. Figure 6.10 shows the concept for a drop in the producer pressure.

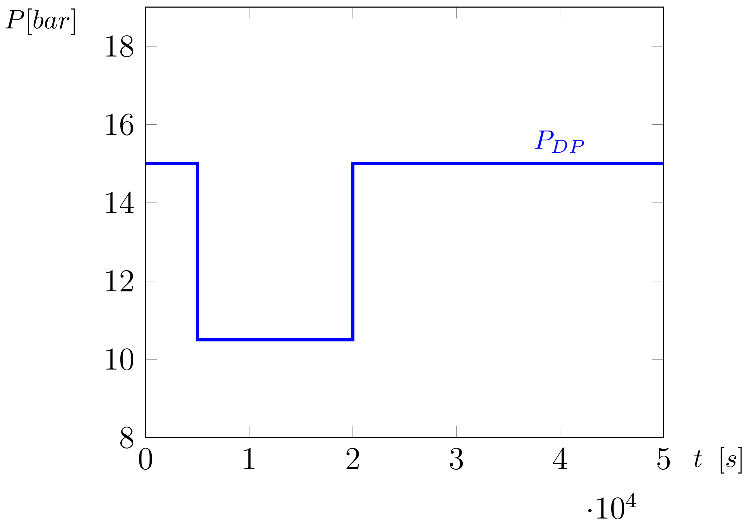

Figure 6.10: Illustration of a negative producer side pressure step on a droop producer, followed by an equal step with opposite sign to return to nominal value.

The result from doing this step change on droop producer I pressure is shown in Figure 6.11. As mentioned, only this disturbance did not cause any controller action for any of the extra producer controllers, meaning that all the controller structures responded equally. The reason for this is the fact that the swing producer valve position, lower than 0.85, which is the lowest setpoint for the controller structures using the swing producer valve position as an input. Thus, the controller error is negative, and is limited by the minimum saturation limit at zero opening for the valve position.

Figure 6.11 shows that the droop producer experiencing the disturbance, droop producer I, is in fact responding by saturating for the pressure disturbance. However, droop producer II and droop producer III are doing minimal controller action, as is the same for the swing producer. The reason for this is that that even though the droop producers are separated from the swing producer on a different time scale, they both interact because they control the same controlled variable and uses proportional control action.

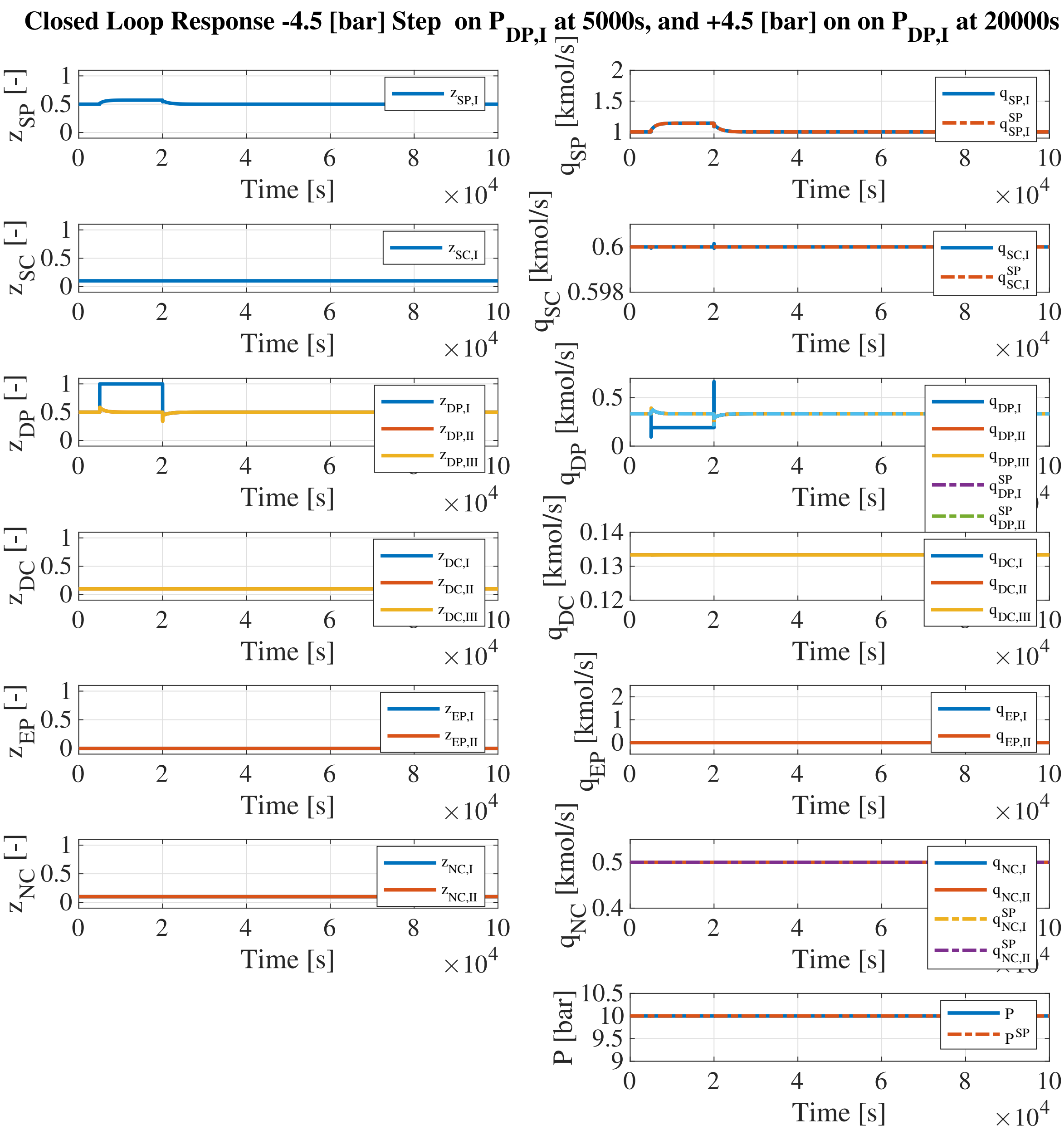

Figure 6.11: Closed loop response for a -4.5 $[bar]$ disturbance on droop producer I at $t = 1000$s, before returning the droop producer to nominal pressure at $t = 2000$s.

The issue in why only the droop producer experiencing a disturbance is saturated, lies within the fact that the pressure disturbance is local. Therefore, the total flow disturbance is rather small, making the main pressure deviation is minimal. This means the proportional action on the droop producers will only act at a minimum, changing the setpoint for the droop producer II and III minimally. However, there is still a minimal error accumulating, and the PI-controller on the swing producer catches this by increasing the swing producer flow setpoint. This means the pressure error signal will decrease again and the droop producer controller II and III will return to nominal. However, to balance the system inflows and outflows, the swing producer maintain the new valve position in a new steady state position. This is unfavorable, but is an effect of the fact that the swing producers use PI-controller action, while the droop controllers only use P-controller action.

It should also be noted that since the swing producer are barely acting, no controller action is done by the extra producers or the swing consumer. The swing consumer, as well as the normal and droop consumer, do have some very small flow deviations, however this is only caused by the coupling in the system. Since the deviations are so small, no visible corrective action is taken by the controllers. If the deviations were to become larger, it is expected that flow controller action would be taken for the swing consumer and the normal consumer.

When returning to nominal values, the droop producer is the fastest acting controller between the swing producer and the droop producers. This goes both ways, since the droop producers have the controller that should act fastest on a disturbance, they will give up their nominal flow setpoints first, but they will also return to nominal flow first when the disturbance is over.

6.1.4 Swing Producer Pressure Disturbance

In the second pressure disturbance simulation, the same concept as for the droop producer disturbance is utilized, only now the disturbance is on the swing producer. A -4.5 [bar] disturbance is posed on the swing producer at $t = 5000$s. Then, at $t = 20000$s, a $+4.5$ [bar] disturbance is added again on the swing producer, allowing for the system to reach nominal values. The results are shown in Figure 6.12 through 6.14.

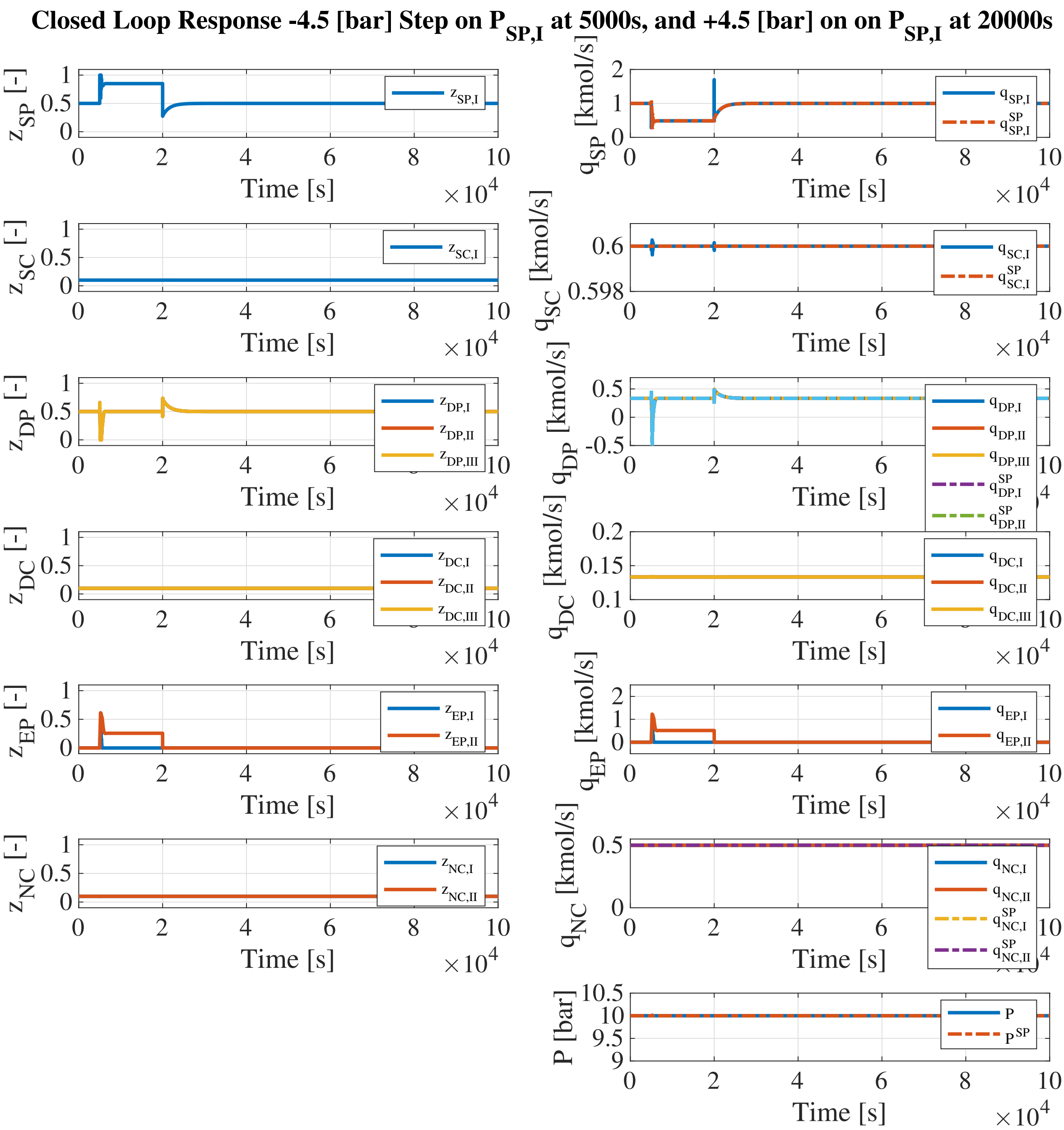

Figure 6.12: A -4.5 [bar] disturbance on swing producer I at $t = 1000$s, returning to nominal pressure at $t = 2000$s using parallel control.

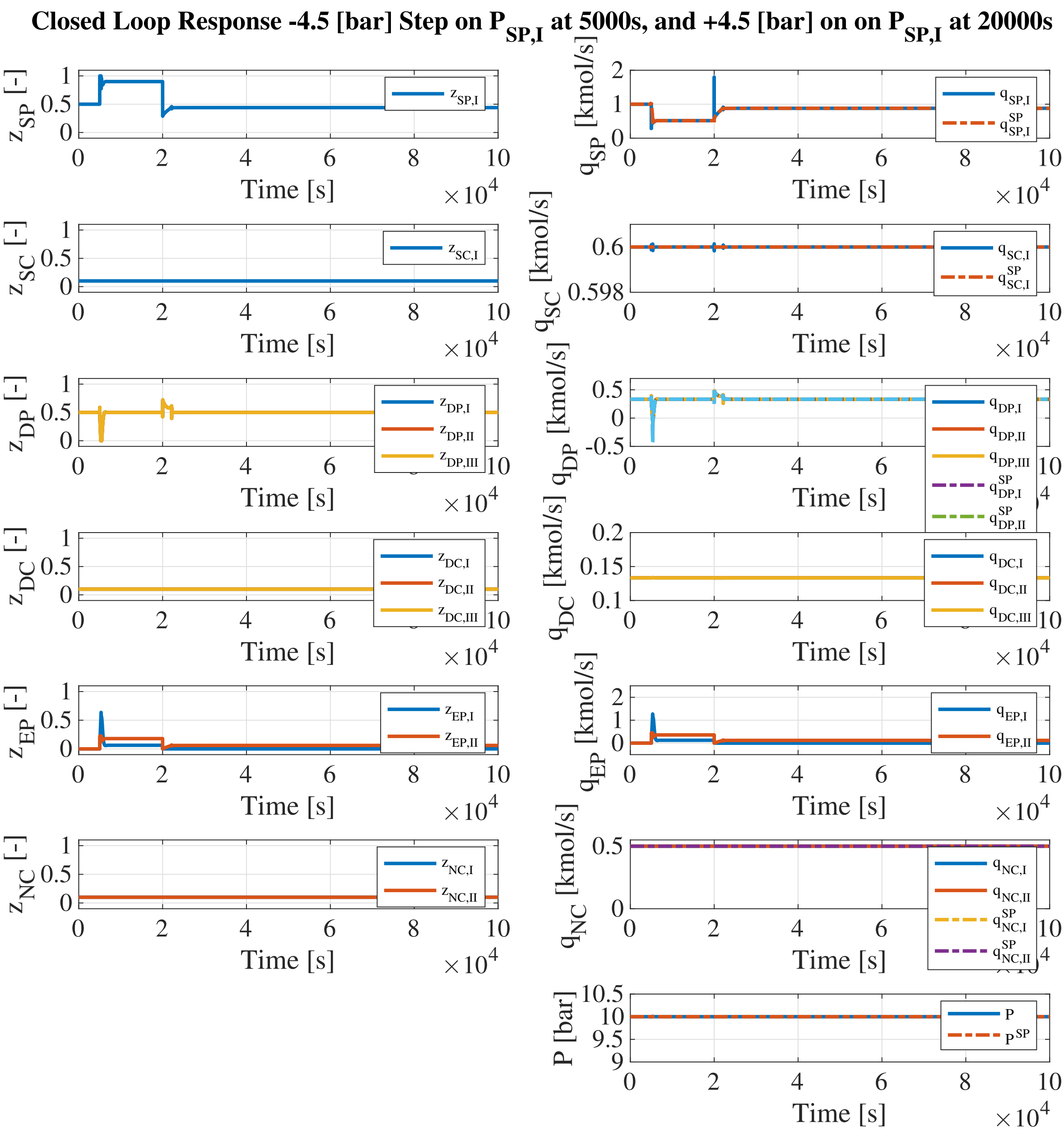

Figure 6.13: A -4.5 [bar] disturbance on swing producer I at $t = 1000$s, returning to nominal pressure at $t = 2000$s using controllers with different setpoints.

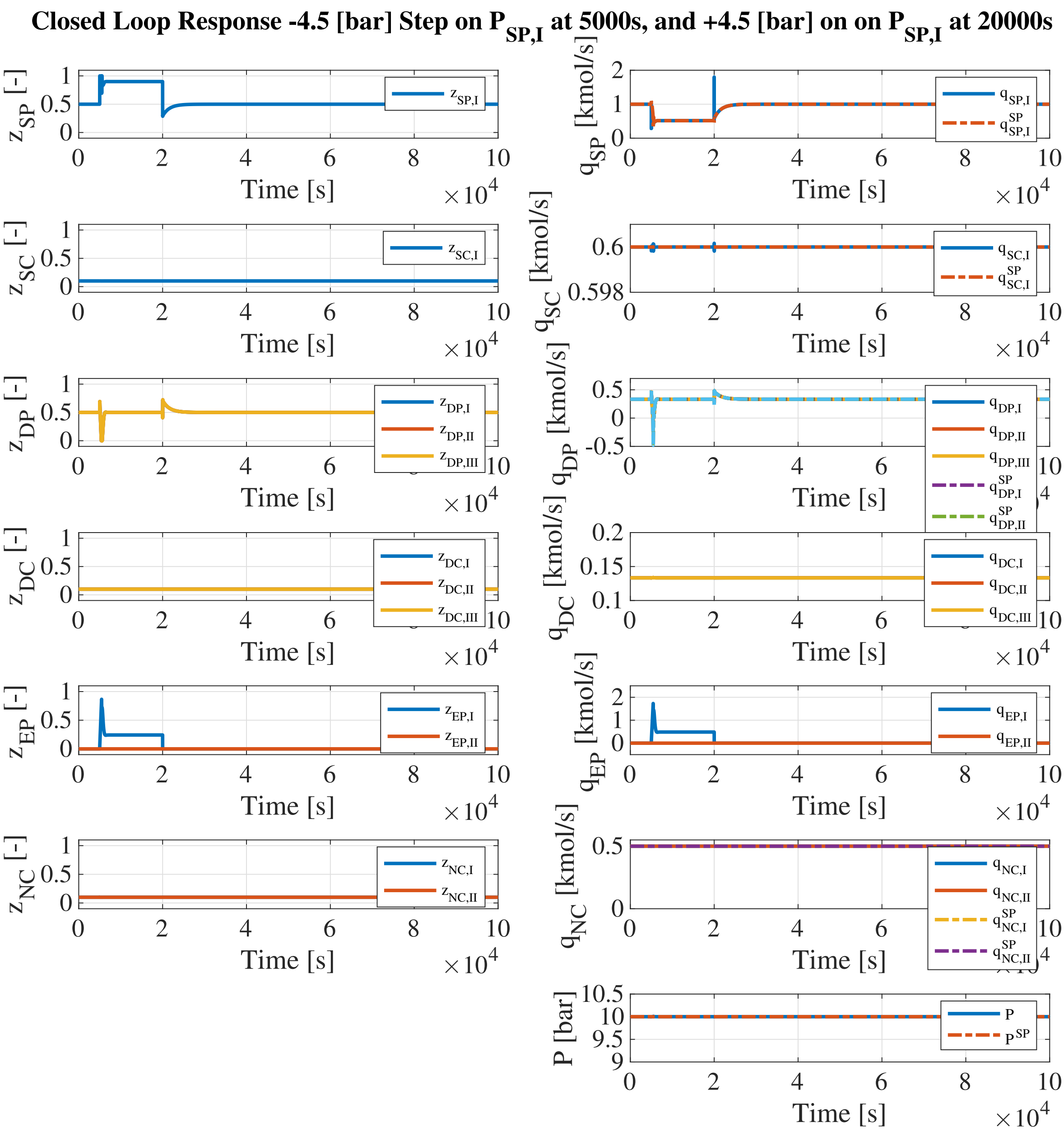

Figure 6.14: A -4.5 $[bar]$ disturbance on swing producer I at $t = 1000$s, returning to nominal pressure at $t = 2000$s using valve position control.

Parallel Control Response

Figure 6.12 uses the same controlled variable but different setpoints configuration for the extra producers. The response for this configuration shows that for a direct disturbance on the swing producer, it will cause a small time period of saturation in the controller. However, because the droop producers are fast acting, they will try to reverse act this. The reason for that is that when the swing producer saturates, the main system pressure will increase in the system, and the droop producer pressure controllers will try to decrease the pressure by decreasing the flow setpoints, efficiently forcing the flow controller to reduce flows.

Because the droop producers are equally tuned, when the disturbance is not directly on any of them, they will act equally, as previously seen for a disturbance on the normal consumers. This could pose a disadvantage, if say all the droop producers were to fully close, leading no flow through the droop producers. However, this is not the case here, as the system regulates such that the swing producer goes from fully open to almost fully open.

Another thing causing the droop producers to almost fully close is the extra producers. Initially, both extra producers respond as the swing producer saturates, causing the supply to be too large. Because the valve position for the swing producer dips below 0.9, but not below 0.85, extra producer I will turn off, while extra producer II will remain on. This is yet another example of why this extra producer controller structure is not ideal, as only one extra producer should act at a time, ideally.

The swing producer and normal consumer flows are minimally affected here as they were for the droop producer disturbance. Again, this is caused by the coupling in the system, and since the disturbance is so small, minimal corrective controller action is taken.

Controllers with Different Setpoints Response

Figure 6.13 uses the same controlled variable and equal setpoints configuration for the extra producers. The response is equal as in the configuration used above, with controllers with different setpoints, except for the extra producer response. The arguments are therefore equal except for the arguments regarding the extra producers.

In this controller structure, as well as the one with different setpoints, both extra producers are used initially, which is undesirable. Because, for this structure, one controller uses P-controller action while the other uses PI-controller action, this will happen when using the same setpoint and controlled variable. As seen for a disturbance on the normal consumers, the issue is this, in addition to the utilization of extra producer II, which should be larger.

In fact, when the swing producer saturates for a brief time in this simulation, the extra producer II, only involving P-control, initially has the same magnitude as for the disturbance in the normal consumers, that is $z_{EP,II} = 0.22$.

Thus, neither of the configurations using the same controlled variable seem to be very efficient when using the extra producers, as they use both extra producers without one of them being close to saturation.

Valve Position Control Response

Figure 6.14 shows the results from using the different controlled variable and equal setpoints configuration for the extra producers. In the figure, the response from the droop producer and the swing producer is equal to those in Figure 6.12 and 6.13, and is therefore not discussed further.

The extra producer responses however, is a different story. Unlike the other extra producer controller structures, only an initial response from one of the extra producers is observed. This proves again, that this structure is better for using only one manipulated variable at a time in a MV-MV switching controller structure.

The disadvantage with this structure is that extra producer I does a relatively large overshoot, causing the droop producer valves to almost fully shut closed for a moment of time. This is however not worse than for the other controller structures.

6.1.5 Swing Consumer Pressure Disturbance

The third simulation poses a disturbance on the swing consumer. Here, the concept is equal to that in Figure 6.10, but instead of a pressure drop, a pressure increase is performed. Therefore, a +0.5 [*bar*] disturbance is posed on the swing producer at $t = 5000$s. Then, at $t = 20000$s, a -0.5 [*bar*] disturbance is added again on the swing producer, allowing for the system to reach nominal values.

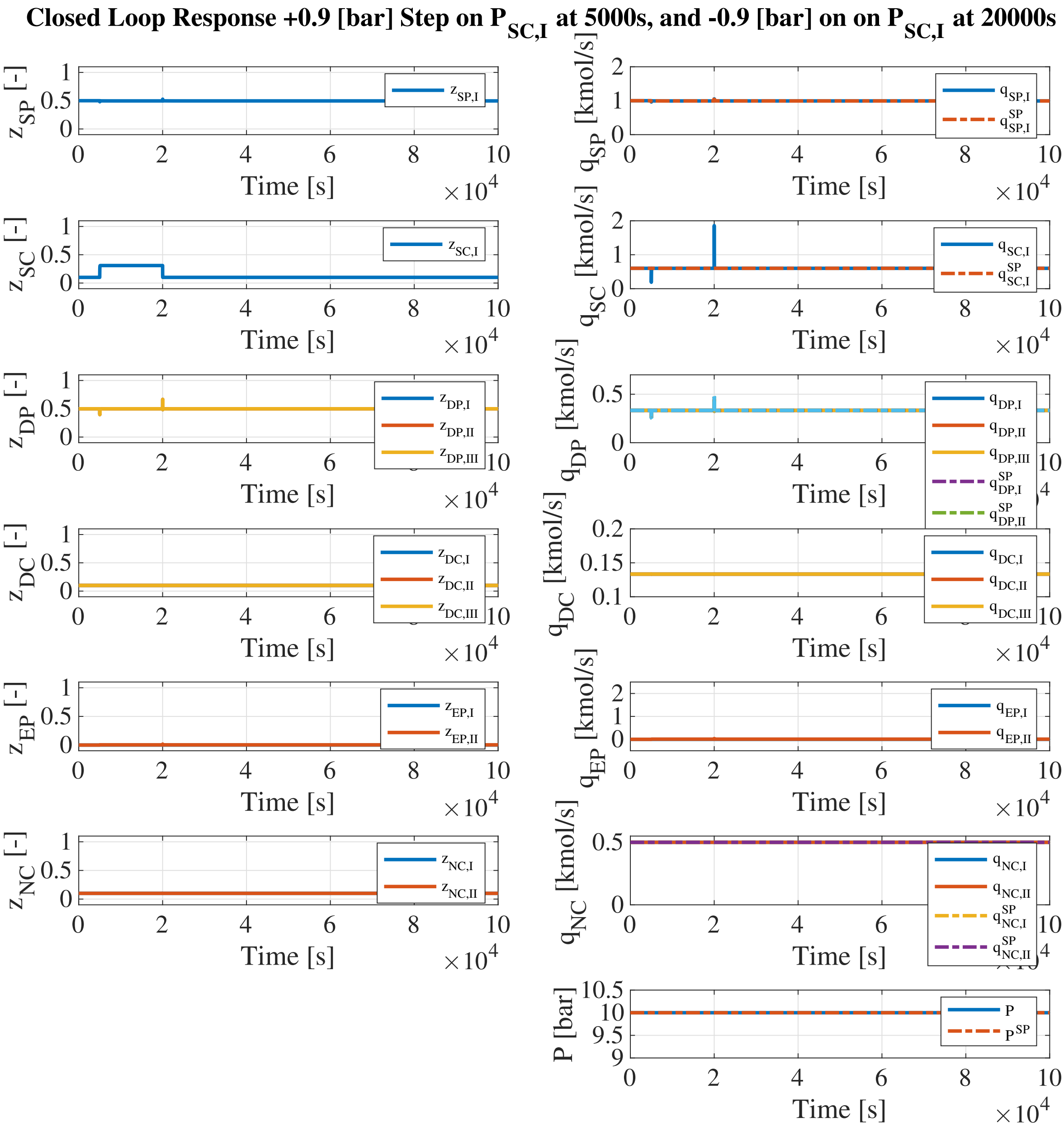

Figure 6.15: Closed loop response for a -4.5 [*bar*] disturbance on swing consumer I at $t = 1000$s, before returning the swing consumer to nominal pressure at $t = 2000$s.

Since the extra producers are not utilized for this disturbance, only one result figure is shown. It is clear in Figure 6.15 that the disturbance on the swing consumer is not large enough to affect the other flows to any noticable extent. Both the droop producers and the swing producer have minimal valve actions, the droop producers showing a larger magnitute in valve action than the swing producer valve, as expected. However, the largest controller action is performed by the swing consumer flow controller.

It should be noted that the flow variations in the swing consumer are quite large, the initial response drops the flow from 0.6 [$kmol/s$] to 0.2 [$kmol/s$]. This is however because the pressure step is really large, making the swing consumer pressure 9.9 [bar] while the nominal main pipeline pressure is at 10 [bar]. The swing controller is actually correcting this very fast, and nominal flow is restored within 60s.

Another solution would be to make the swing consumer controller faster, causing the response to the disturbance even faster. However, as a disturbance this large this fast is unlikely, the controller performance is considered satisfactory.

6.1.6 Combined Producer Pressure Disturbance

The fourth, last simulation poses a combined disturbance from the droop producer and the swing producer. First, a -4.5 $[bar]$ disturbance on the droop producer at $t = 5000$s. Then, a -4.5 $[bar]$ disturbance is posed on the swing producer at $t = 15000$s. When $t = 20000$, a $+4.5$ $[bar]$ change in the droop producer pressure is introduced, returning the droop producers to nominal supply pressure. Lastly, a $+4.5$ $[bar]$ change in the swing producer is introduced at $t = 30000$s, returning the system to nominal values. The results are discussed in the next section.

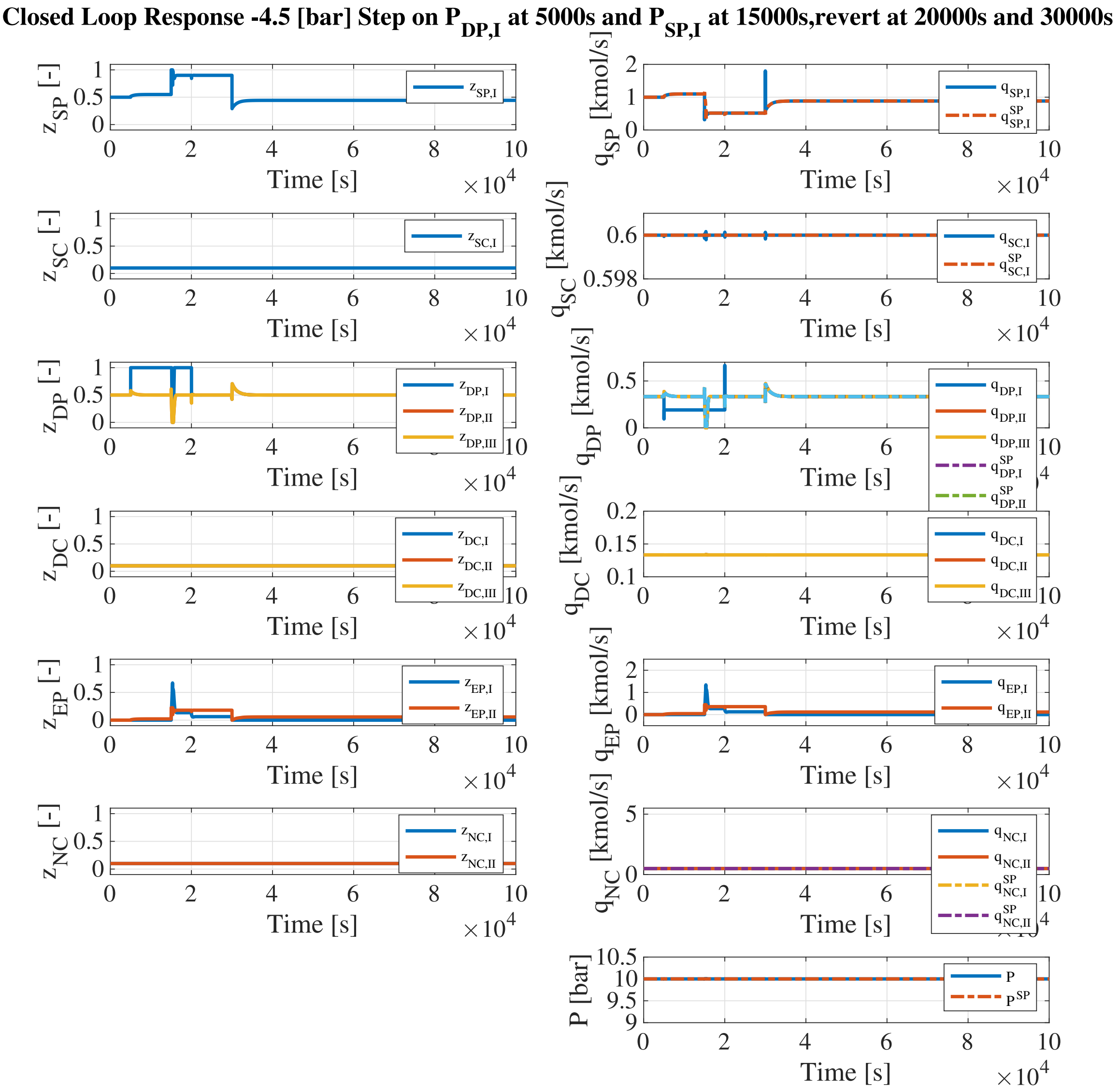

Figure 6.16: Closed loop response for a -4.5 $[bar]$ disturbance on droop producer I at $t = 5000$s, and then swing producer I at $t = 15000$s, before returning the droop producer I at 20000s and then swing producer I at 30000s to nominal pressure using parallel control.

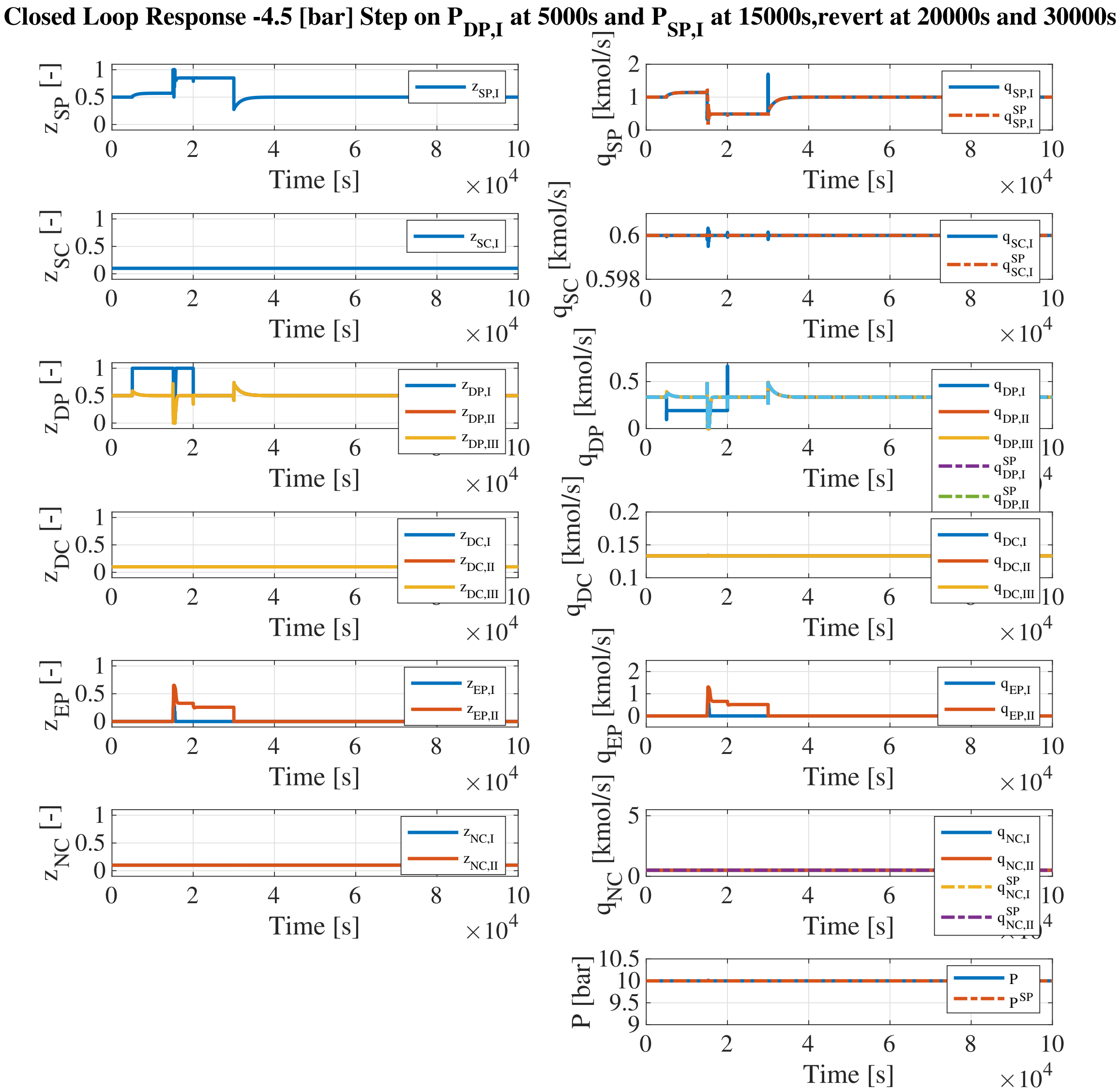

Figure 6.17: Closed loop response for a -4.5 $[bar]$ disturbance on droop producer I at $t = 5000$s, and then swing producer I at $t = 15000$s, before returning the droop producer I at 20000s and then swing producer I at 30000s to nominal pressure for controllers with different setpoints.

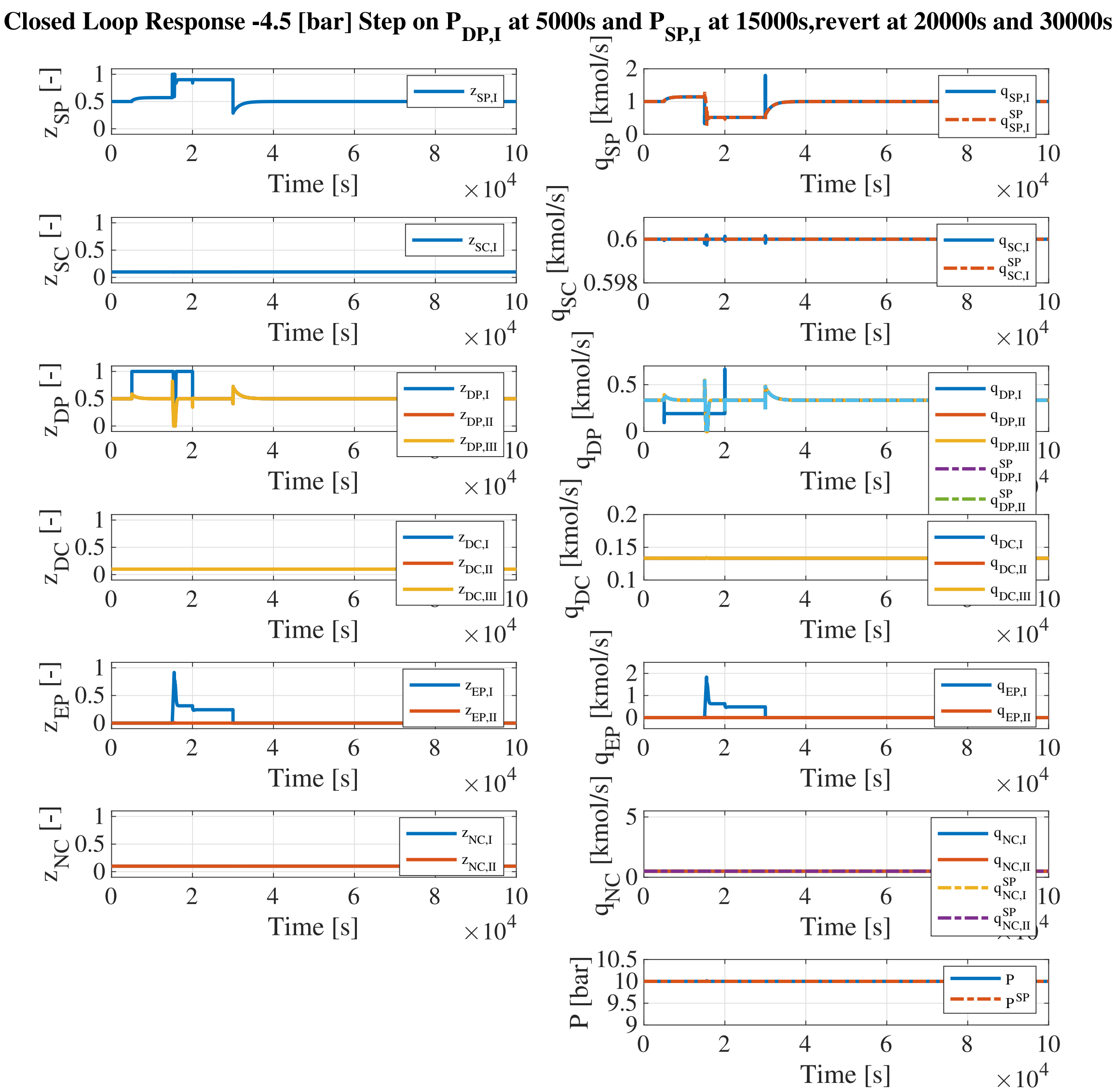

Figure 6.18: Closed loop response for a -4.5 [bar] disturbance on droop producer I at t = 5000s, and then swing producer I at t = 15000s, before returning the droop producer I at 20000s and then swing producer I at 30000s to nominal pressure for valve position control.

Parallel Control

Figure 6.16 uses the same controlled variable but different setpoints configuration for the extra producers. Initially, only the droop producer disturbance is introduced, and the response is equal to the one presented in Figure 6.11. The difference occurs when the swing producer disturbance is introduced simultaneously.

As the swing producer disturbance is introduced, the system uses a brief moment to stabilize as the extra producers activate. During this time, the extra producers overshoot, as seen previously, while the droop producers undershoot. Then, the droop producers return to the new nominal values they reached after the droop producer disturbance was introduced. The valve position for the swing producer reaches the value it had for a disturbance on the droop producer alone, and the extra producers stabilize at a larger total valve opening than previously, because the supply required now is larger than for the single disturbances.

However, it goes to show that this extra producer controller structure is not working well. In the period where both disturbances are simultaneous, both extra producers are almost equally open at around 20%. Since the extra producer dimensions are equal, that would make a 40% total valve opening if only one extra producer were utilized. Thus, only one extra producer should be open at this point.

Other than that, the disturbances are handled well, and since not both of the extra producers are saturated, no controller action is taken on the swing consumer, which is expected according to the proposed prioritization list.

Controllers with Different Setpoints

Figure 6.17 uses the same controlled variable and equal setpoints configuration for the extra producers. The response is more or less equal to the response presented in Figure 6.16, except for the extra producer responses.

This controller structure utilizes the extra producers better than the case with equal setpoints. Initially, both extra producers are acting. However, one of the extra producers is only slightly used, at around 10%, and this is only for a brief moment, then it is turned off completely. The reason for this is a slight controller overshoot. It is the same behavior as seen for this controller structure previously, strengthening the expected controller behaviour. It also indicates that the tunings might be too aggressive.

Noted that the pressure deviation from nominal value is about twice as large for this structure, as compared to the structure in the same setpoint configuration. The magnitude is however very small for both structures, in the scale of 10e-5. Therefore, this deviation is likely caused by numerical error in the solver.

Valve Position Control

Figure 6.18 shows the results from using the different controlled variable and equal setpoints configuration for the extra producers. The response is more or less equal to the responses presented in Figures 6.16 through 6.17, except for the extra producer responses.

For this disturbance, valve position control actually utilize the extra producer controllers best. This is because, as expected, only one input is used because only half of extra producer I capacity is required. Apart from that, the controller acts much like the different setpoints controller. Extra producer I overshoots then stabilizes in the same fashion as the different setpoints controller structure.

The next section presents the results from the model predictive controller, including the tuning results, and disturbance rejection for the same disturbances as the decentralized control structures. Then, the two controller structures are compared to each other, before results are discussed.

6.2 Model Predictive Control

The results for the multivariable model predictive controller is presented in the same order as for the decentralized controller structure. First, the a no step weighted response is presented, then the closed loop system is presented. For the closed loop system, a disturbance on the normal consumer flow is presented first. Then, pressure disturbances are introduced. That is, first a disturbance on the droop producer is presented, followed by a disturbance on the swing producer, then the swing consumer, and last a combined producer side pressure disturbance is presented, as for the decentralized control structure.

6.2.1 Normal Consumer Flow Disturbance

This section presents a +4.5 $[kmol/s]$ disturbance on the flow setpoint for normal consumer I and a +0.5 $[kmol/s]$ disturbance on the normal consumer II, introduced at $t = 1000$s, and reverted at $t = 2000$s for the model predictive controller.

The response differs from the decentralized control structures on multiple levels. First of all, the controller enforces action on the droop producer, swing producer and the extra producer all at once. The droop producer is not slightly utilized before action on the swing producer commences, everything happens at once. The same happens when the disturbance reverts to nominal values. Everything happens at once when the system reverts, whereas for the decentralized controller structure this takes a very long time compared to the model predictive controller.

The model predictive controller also manages to balance supply and demand without using the swing consumers. This is because the system does in fact prioritize that swing consumers should not be used unless absolutely necessary. Even though the main priority is in fact maintaining the pressure P in the network, the model predictive controller does do a better job at saving the swing consumer usage.

The system should in fact, have utilized extra producer II even more, because as seen in Figure 6.19, the extra producer II is not fully utilized, which it could have been to increase the pressure network. This is one of the flaws of using model predictive control: The controller will calculate the cheapest use of the inputs which in this case meant giving up the pressure from nominal value. The results however, are considered better than for the decentralized controller structures, as the response is faster, and the controller return to system nominal values is way faster than any of the decentralized control structures.

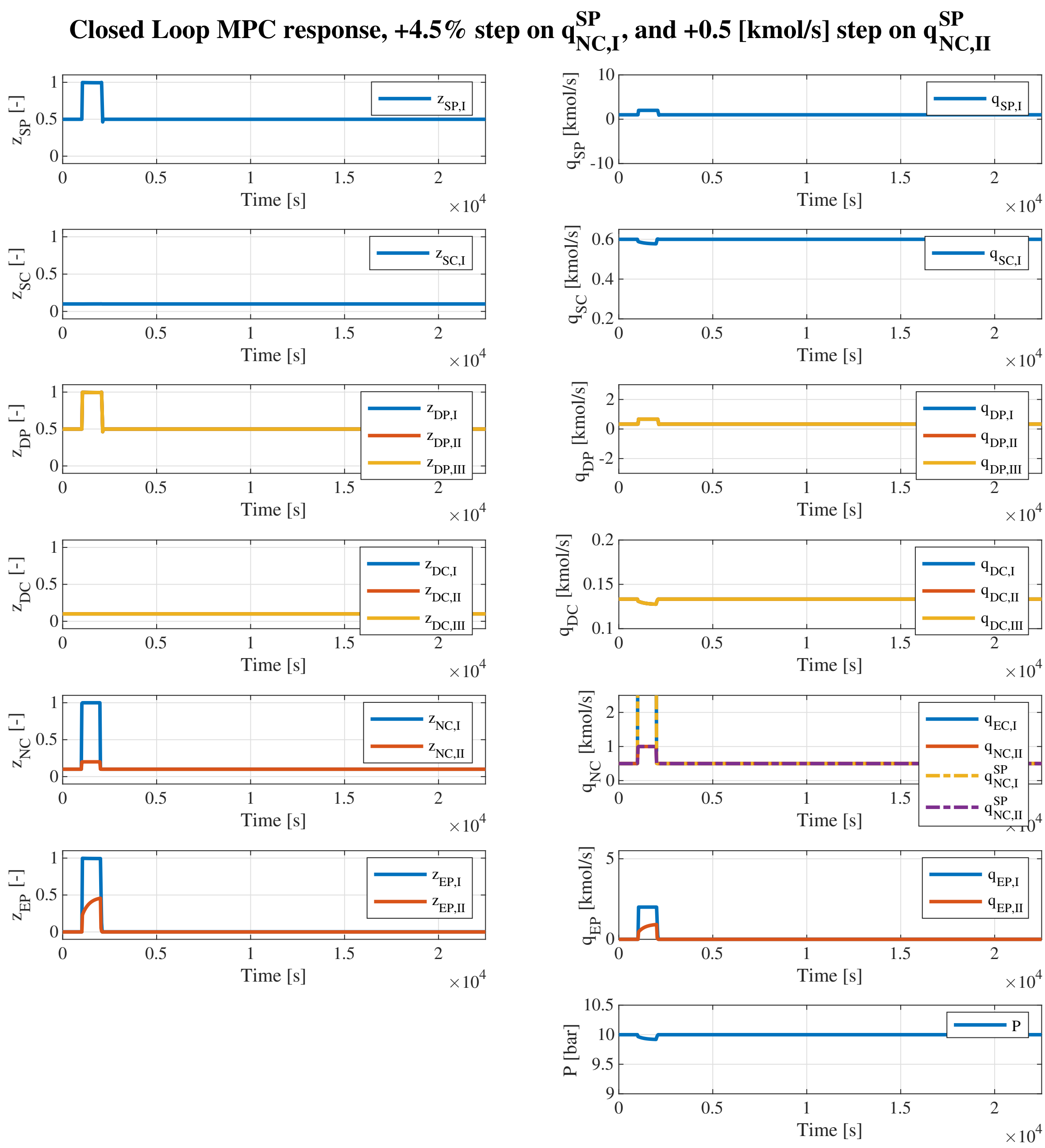

Figure 6.19: A +4.5 $[kmol/s]$ flow disturbance on normal consumer I and a +0.5 $[kmol/s]$ disturbance on normal consumer II at $t = 1000$s, reverting at $t = 2000$, using model predictive control.

Figure 6.20 shows a a +4.5 $[kmol/s]$ flow disturbance on normal consumer I and II at $t = 100$s, without reverting to nominal values. For this simulation, the results are in fact much worse than the decentralized control, for the same reason that pressure is given up in Figure 6.19. What happens is that initially, controller response is good. Every producer is utilized, making it unnecessary for the swing producer to do any controller action.

However, to do this, the model predictive controller finds it cheapest to let pressure

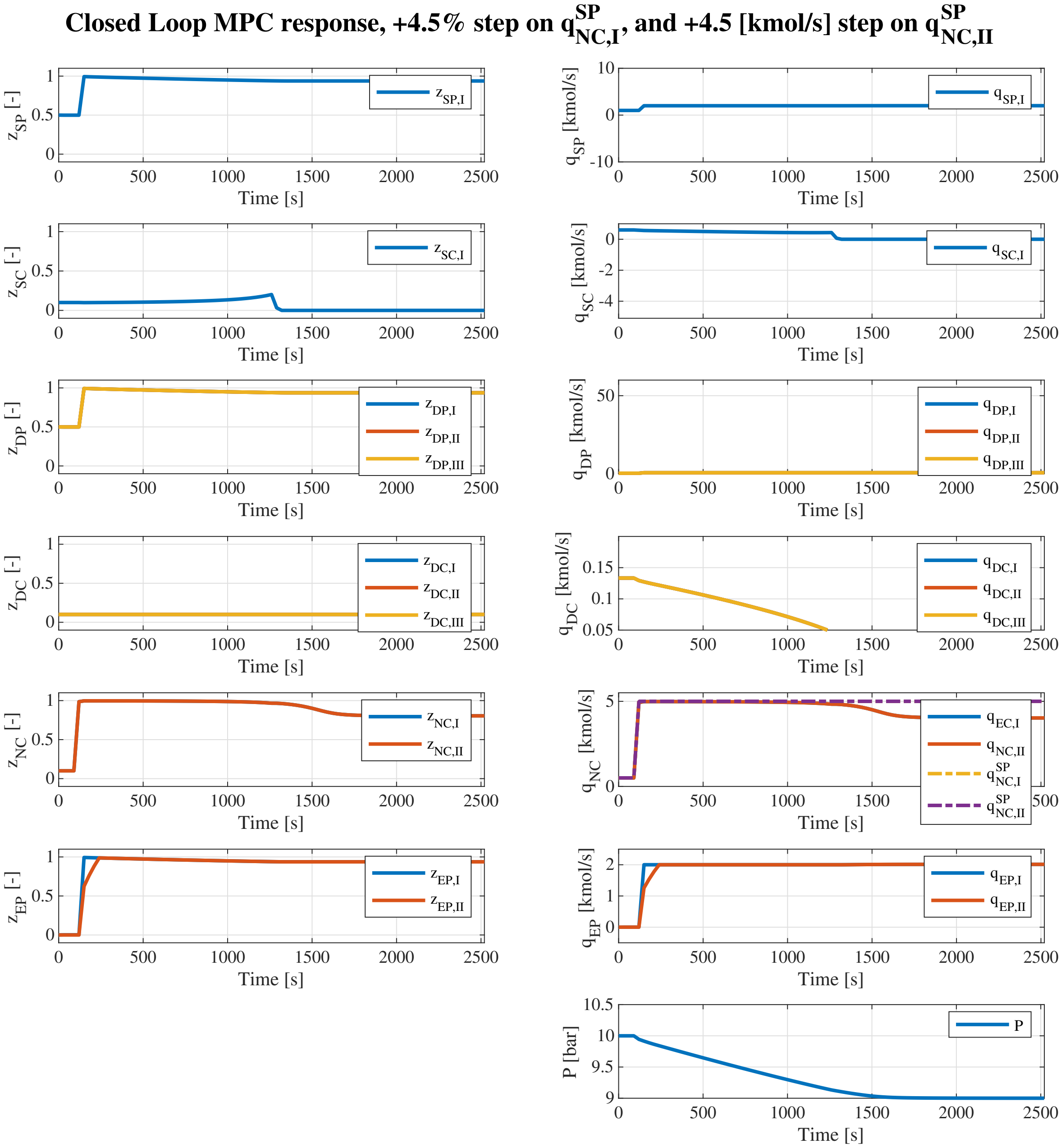

Figure 6.20: Closed loop response for a +4.5 $[kmol/s]$ flow disturbance on normal consumer I and II at $t = 100$s, using model predictive control.

drift from setpoint. This causes the pressure in the system to drop forcing the swing producer to open its valve to maintain flow setpoint. It also makes the producer side close their valves slightly, trying to maintain the main pressure in the network, as smaller producer flow means higher main network pressure.

At one point this causes the model predictive controller to calculate a new steady state, which happens when the pressure reaches its minimum constraint. This new steady state is actually not a desired state, as the desired state would be maximizing producer flows and minimizing controllable consumer flows. However the only thing actually controlled as desired is the swing consumer closing, even though this should not happen when the producer side is not saturated.

All in all, the normal consumer responses indicate that the model predictive controller performs worse than the different setpoints controller structure. Before conclusions can be drawn, pressure disturbances should be introduced to the process as well.

6.2.2 Droop Producer Pressure Disturbance

The droop producer pressure disturbance poses a -4.5 $[bar]$ disturbance on droop producer I at $t = 5000$s. Then, at $t = 20000$s, the pressure is increased by $+4.5$ $[bar]$ on droop producer I, allowing for the system to retain to nominal values. Figure 6.21 shows the response results obtained for the producer pressure disturbance on the model predictive controller structure.

Contrary to the decentralized control structure, the model predictive controller acts at an instant. The controller action is however not as good as for the decentralized control structure. First of all, the disturbance causes droop producer I, which is experiencing the disturbance, to almost fully close. This is reasonable to maintain system pressure. However, droop producer II and III are doing the opposite, actually opening up more than nominal. Ideally, the controller structure would close droop producer I more, then minimize the action on the remaining droop producer. This response is caused mainly by the fact that all three droop producers are tuned equally.

For this response, the swing producers is also utilized somewhat, which is expected, since the input costs in the objective function is almost equal. The biggest issue by far with this response, is what happens after the disturbances revert.

When the disturbances revert, the model predictive controller finds an entirely new steady state, doing large changes especially for the swing producer and the droop producers, which deviate over 20% from their respective nominal values.

This steady state deviation is likely because of the objective function. Several modifications were performed to try and make the steady state deviation when reverting to nominal value were performed, but none that would not change the objective were useful. Among the methods tried were using a linear objective function and different controller tunings. Other things that could help the performance is changing the objective function. However, doing this would change the process control objectives, thus it is not a viable option for this thesis.

It is observed that also minimal controller action is posed on the extra producers.

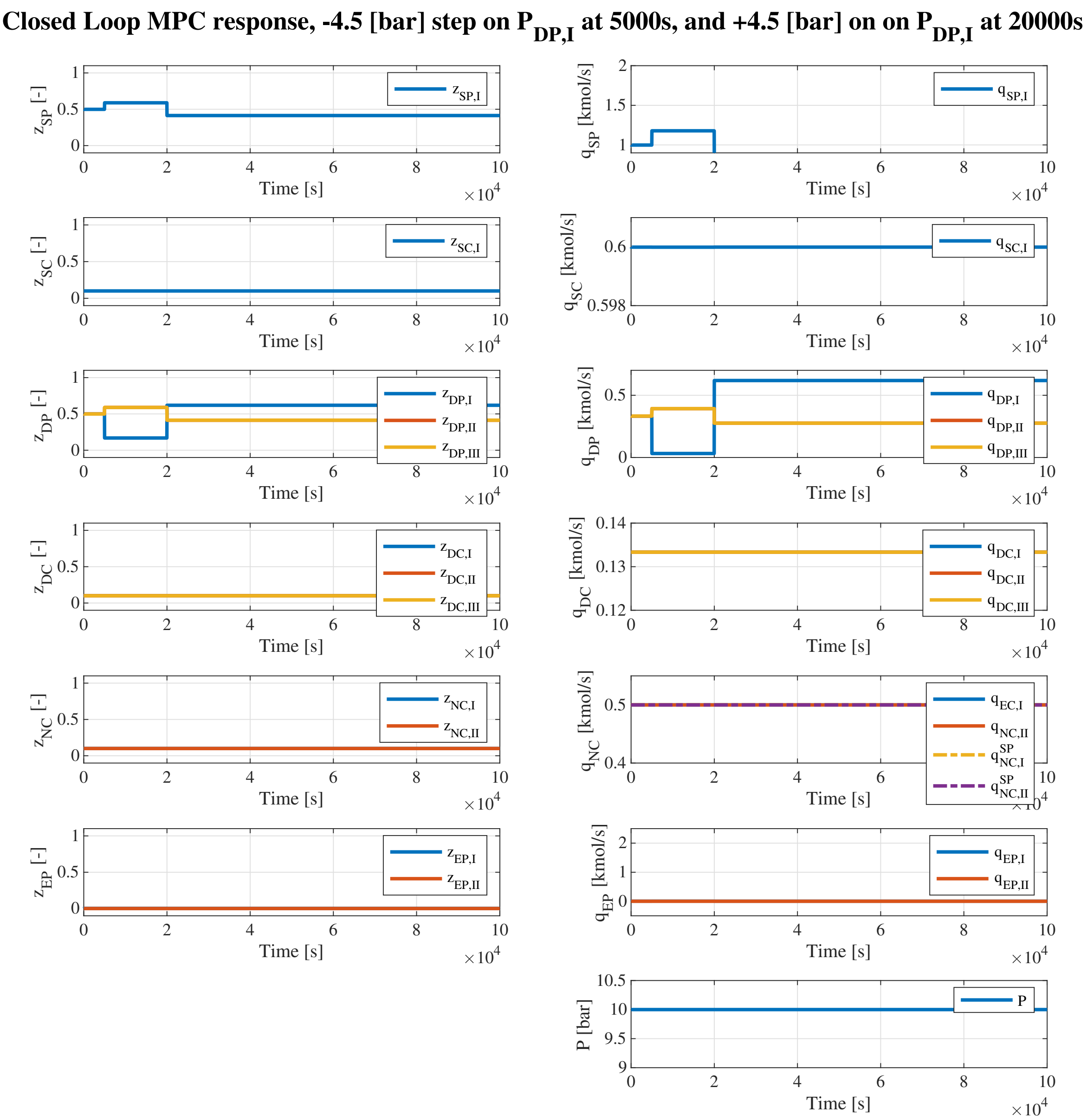

Figure 6.21: Closed loop response for a -4.5 $[bar]$ pressure disturbance on droop producer I at $t = 5000$s, then reverting at $t = 20000$s, using model predictive control.

However, since this is only about 1% valve capacity, it would be practical means be considered that the valve position is closed.

6.2.3 Swing Producer Pressure Disturbance

The swing producer pressure disturbance involves a -4.5 $[bar]$ disturbance posed on the swing producer at $t = 5000$s. Then, at $t = 20000$s, a $+4.5$ $[bar]$ disturbance is added again on the swing producer, allowing for the system to reach nominal values. The result is shown in Figure 6.22.

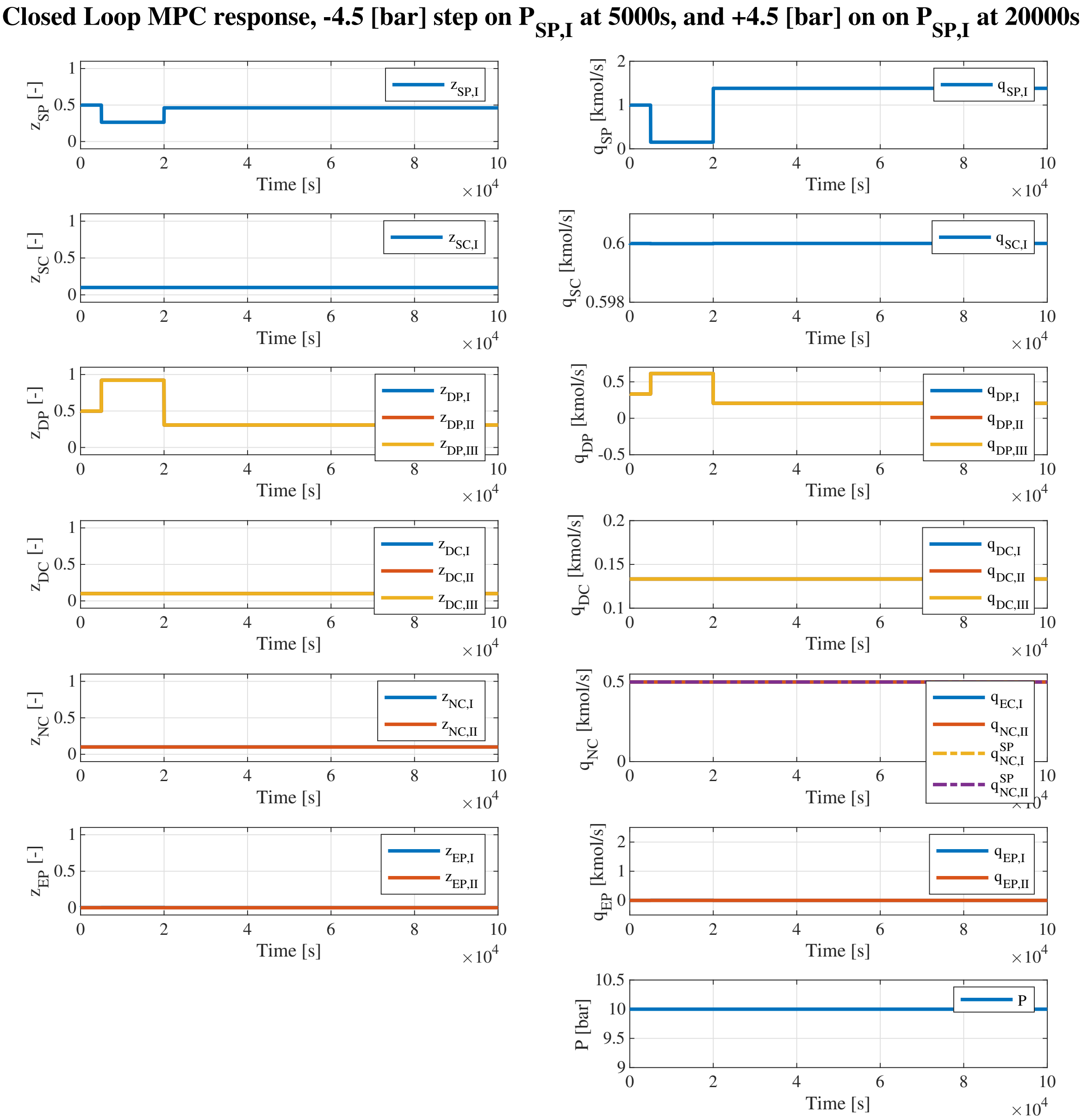

Figure 6.22: Closed loop response for a -4.5 $[bar]$ pressure disturbance on swing producer I at $t = 5000$s, then reverting at $t = 20000$s, using model predictive control.

For this controller structure, all droop consumers act close to saturation. However, this causes to much supply in the system, forcing the swing producer to actually close its valve slightly to restore balance in the control system. Again, the model predictive controller is much faster acting than the decentralized control structure, however the prioritization made by the controllers are not what is to be expected from the weighting and definition of the objective function.

This remains true also for the return to nominal value, in which the swing producer does not return to the given nominal values either. Also different actions were tried on this controller structure without succeeding at creating a better control structure.

Also for this disturbance, the extra producers are slightly utilized. However by practical means they would be zero, since the valve position still is only around 1% open.

6.2.4 Swing Consumer Pressure Disturbance

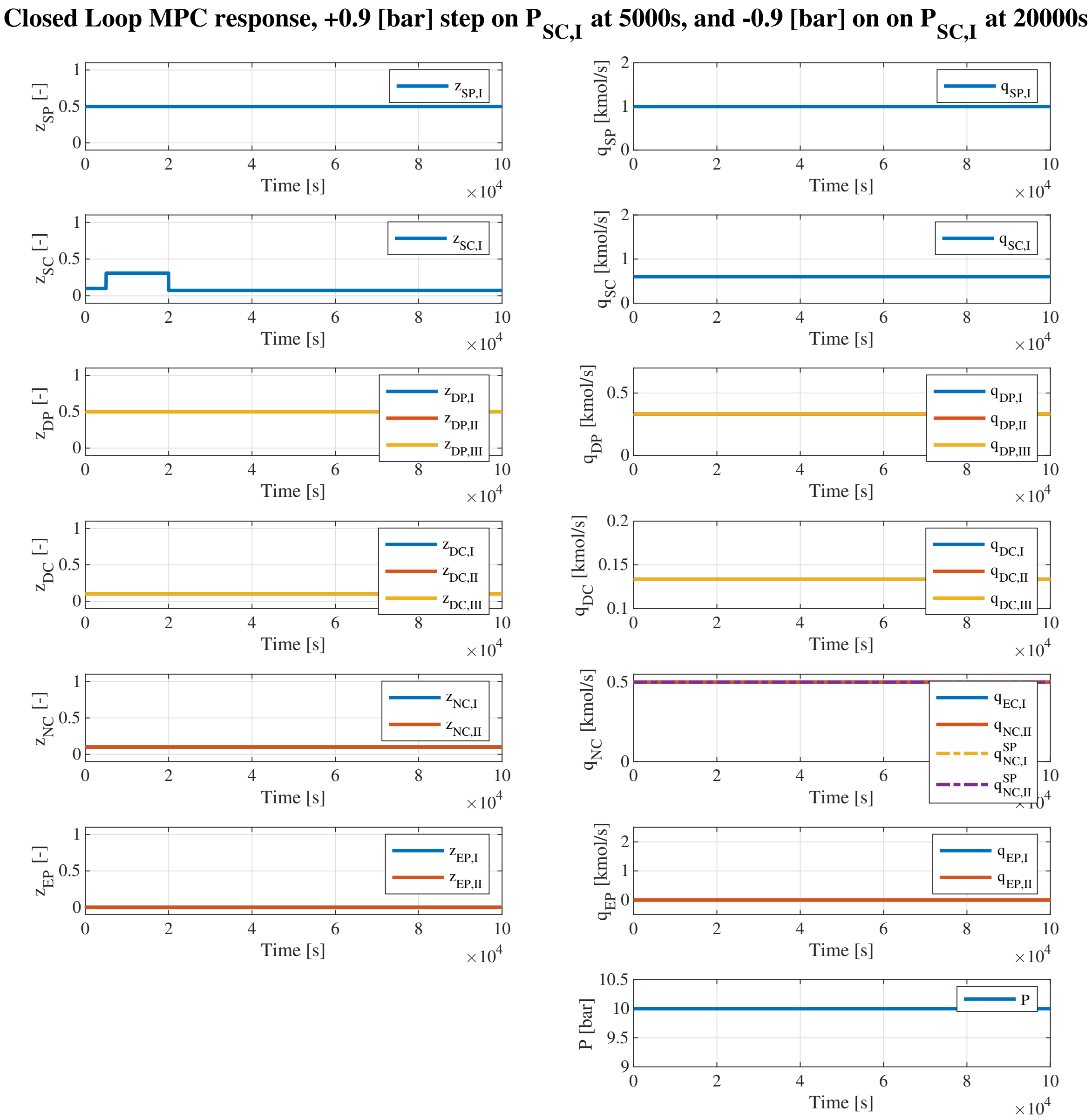

Figure 6.23: Closed loop response for a +0.9 $[bar]$ pressure disturbance on swing consumer I at $t = 5000$s, then reverting at $t = 20000$s, using model predictive control.

The third pressure disturbance simulation poses a disturbance on the swing consumer. A +0.9 $[bar]$ disturbance is posed on the swing producer at $t = 5000$s. Then, at $t = 20000$s, a -0.9 $[bar]$ disturbance is added again on the swing producer, allowing for the system to reach nominal values. Figure 6.23 shows the response obtained from the disturbance response.

For this disturbance, thew only response change is that on the swing consumer. This is likely since the swing consumer flow is so high prioritized, that the model predictive controller finds the most economic way to counteract the disturbance is by using swing consumer valve position to maintain the swing consumer at nominal value.

The disturbance only being local makes the controller structure actually return to nominal value after the disturbance is reverted. This is why it is important to test all aspects and possible disturbances when implementing model predictive control: The response can be good for some disturbances, while requiring more model tunings or model modifications for some disturbances.

6.2.5 Combined Pressure Disturbance

The last simulation poses a combined disturbance from the droop producer and the swing producer. First, a -4.5 $[bar]$ disturbance on the droop producer at $t = 5000$s. Then, a -4.5 $[bar]$ disturbance is posed on the swing producer at $t = 15000$s. When $t = 20000$, a $+4.5$ $[bar]$ change in the droop producer pressure is introduced, returning the droop producers to nominal supply pressure. Lastly, a $+4.5$ $[bar]$ change in the swing producer is introduced at $t = 30000$s, returning the system to nominal values. Figure 6.22 shows the results for the combined disturbance.

Initially the model predictive controller acts the same as for the disturbance on the droop producer. However, when the combined disturbance enacts, the response is actually just the sum of the two disturbances from the response on the single disturbance result plots. This is much the same as the results in the decentralized control structure, where the combined disturbance was also about the sum of the two disturbances for the producer and consumer responses.

Because the combined disturbance only acts as a sum of the individual disturbances and controller actions, the response when the droop producer disturbance reverts is the same as the response for only doing a disturbance on the swing producer.

In order to match the decentralized controller scaling for the response in Figure 6.24, it is not visible that there actually is minimal controller action on the extra producers. However, as for the other pressure disturbances, the magnitude is minimal, and in practice the controllers would be considered to be off.

Thus, for the model predictive controller, the initial responses are fast and accurate, however over a longer time period the controller loses track of the original nominal values and chooses its own nominal values. This is not optimal if the original nominal

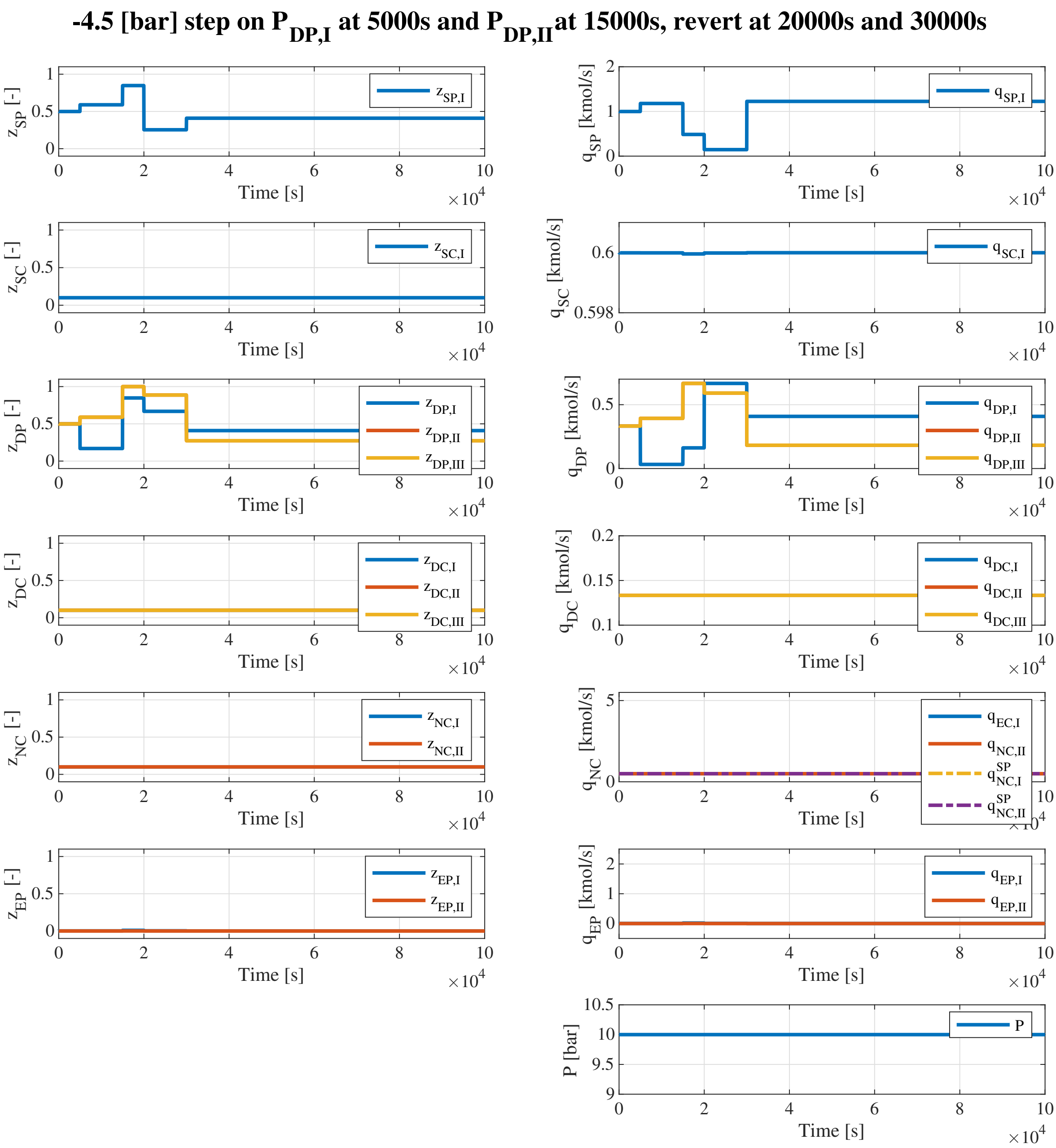

Figure 6.24: Closed loop response for a -4.5 $[bar]$ pressure disturbance on droop producer I at $t = 5000$s, and a -4.5 $[bar]$ disturbance on swing producer I at 15000s, then reverting at $t = 20000$s and $30000s$ respectively, using model predictive control.

values are calculated for economic optimality. Next, the key points from the model predictive controller is compared against the decentralized controller structures.

6.3 Performance Comparison

Starting with the initial responses of the decentralized controller structures and the model predictive controller, it is clear that the model predictive controller is initially faster than the decentralized control structures, for any of the configurations. This remains true for pressure disturbances as well as normal consumer disturbances.

In addition to the model predictive controller being initially faster, it is also more accurate. While some of the decentralized controller structures actually does small overshoots, the model predictive controller is very accurate in using exactly the required input amounts initially.

The reason why the model predictive controller is very fast and accurate, is because it receives perfect measurements and uses a perfect model. Therefore, it effectively acts as an ideal feedforward controller structure, predicting future variables perfectly. Since the decentralized controller structures can only act on measured variables, they will always perform worse than the model predictive controller for a perfect model with perfect disturbance measurement. However, for a real case, it is likely that neither the model or the measurements are perfect, which could make the model predictive control performance much worse. The decentralized controller performance is not expected to decrease in performance for a real plant operation, as it only depends on tuning and measurements.

Looking at the responses after a time period, the model predictive controller does not perform that well however. Because the model predictive controller calculates its input use based on an objective function, this is what determines the steady state for the model predictive controller. Over time, the model predictive controller finds new steady state values in which minimize the objective function even more than the given nominal values.

This remains especially true for when disturbances are removed. Whereas the controllers with different setpoints and the valve position controller configurations returns the controllers to steady state values, the model predictive controller fails completely at doing so.

Changing the objective function and redoing the tuning process could probably improve the model predictive controller behaviour, however this issue is common in model predictive control engineering, and only shows the complexity required when implementing the controller. Every detail about what the controller should do needs to be carefully implemented, if else the performance could be worse than no control.

Among all controller structures, the control structure using controllers with different setpoints perform the best. This structure utilized primary, secondary, tertiary and quaternary control fairly quickly, and returns the system back to nominal conditions. This is also the case for the valve position control, however valve position control acts on a slower timescale. The model predictive controller has the potential of being better than the decentralized controller structure, however that would require a larger time scope than that of this master thesis. Last, the parallel control structure is not recommended to use for steam network control.

Chapter 7

Discussion

Some subjects require further discussion than what is presented in the results chapter. Therefore, they are further discussed in this chapter. First, the tunings and implementations are discussed. Then, the performance of each controller structure for the decentralized control are discussed before the model predictive controller implementation is discussed. Last, system disturbances and simplifications are discussed.

7.1 Tunings

This section discusses the tuning implementations, simplifications made, and issues that are not emphasized in Chapter 5.

7.1.1 Decentralized Controller Tunings

The tuning method selected in this thesis is simple internal model control. There are, however, many other tuning methods presented, which could also have worked to implement. However, the simple internal model control method is algebraically derived and therefore expected to perform well, which is shown in the overall results as there is minor controller instability and fast disturbance rejection overall.

The decentralized controller using the simple internal model tuning rules makes some simplifications. For example, in the flow controller tunings, only the initial response is used. This is also the case for the extra producer tunings. However, this is not considered to affect the results as the controller behaviors are stable.

Another simplification is made in the initial response in the extra producer controller tunings. The response here is not entirely first order but simplified as one, as the time considered as a delay is when the swing producer valve position response is slightly changing. This adjustment is, however, minimal and should not have an impact on the results.

7.1.2 Model Predictive Controller Tunings

Whereas the decentralized controllers use the simple internal model controller tuning method, few or no working tuning methods have been proposed for model predictive control. Therefore, model predictive control tuning is mainly left to trial and error and engineering insights. This was also the case for the tunings presented in this thesis: They are based on engineering insight, first of all, using the system goals to achieve a coarse tuning for the controllers. Then, trial and error were used for fine-tuning, making the nominal system act as close to the decentralized controller system as possible.

There is, however, a slight difference in the nominal system compared to the decentralized controller structure. That is because the model predictive controller finds the new steady-state more economical following the objective cost function. The difference is because of fine-tuning minimal. The most significant difference is on extra producer I, which is actually at 0.1% valve opening. This is regarded as so minor that in an actual plant, the effect would be that the valve position is still fully closed, but it shows that it is harder to make a model predictive controller follow desired controller objectives.

7.2 Closed Loop Performance

This section discusses performance and issues with the different controller structure performances not addressed in Chapter 6.

7.2.1 Primary Control

The droop producers, acting as the primary control reserve, were on top of the prioritization list. They did perform as expected following the primary, secondary, tertiary, and quaternary prioritization order. However, being the fastest acting control, they faced issues. The main problem was that when the system responded to disturbances, sometimes an overshoot in supply would happen. This happened when the droop producer, the swing producers, the extra producers, and the swing consumer acted, for example, in Figure 6.2. Initially, the primary controllers worked the fastest, which is the desired response. However, because of the overshoot, production had to be limited, and with the droop producers being the most rapid-acting controllers, they would also return to nominal value the fastest. Thus, the system would still use the extra producers after the swing, and droop producers return to nominal values when the extra producers should rather be the controllers turning off first.

This issue could be fixed by implementing slower-acting controllers to avoid the supply overshoot. However, this overshoot only happens for an immense, unrealistic disturbance caused by the normal consumers and is therefore not considered realistic to occur in a real plant. Thus, in conclusion, the primary acting controllers act as expected.

7.2.2 Secondary Producer Performance

The swing producer, acting as the secondary control reserve, should work after the primary controllers in the control system. They performed as expected. However, they suffered from the fast-acting primary control returning to nominal value faster when a system overshoot happened.

Another issue with the swing producer was the error accumulation when saturation in the controller happened. Because the swing producer used PI-control, unlike the primary control only using P-control, the error kept accumulating when pressure was below nominal value and stabilized. This led the controller error to move towards infinity, which is not desirable. This would also cause the controllers to take longer to return to nominal values if an overshoot in supply happens.

A solution to the error accumulation could be to define a max flow value, in addition to the nominal flow value and the setpoint sent from the secondary controller. Then, a middle selector could be implemented, which efficiently at nominal value would implement the nominal value. Then, if the dynamic setpoint from the secondary controller would increase, the selector would choose the median value, the dynamic setpoint, until the point where the dynamic setpoint would be higher than the maximum setpoint value. However, this configuration would not work if the dynamic setpoint were to decrease below the nominal value. In that case, only the nominal value would be selected. However, since lower demand is not likely, this structure could be an idea to implement to see whether the balance between supply and demand for a considerable disturbance could be improved. The system would do an overshoot in a best-case, and then the swing consumers would return to nominal value before the droop producers.

7.2.3 Tertiary Control Performance

The tertiary control was split into three different configurations: parallel control, different setpoints, and valve position control. A common issue for all the control structures was that when the system did a supply overshoot, none of the controller structures decreased the supply first, which they should. The reason for this is first because the tertiary controllers were on the slowest timescale in the system. Secondly, at least one of the extra producers relies on the swing producer valve position as an input, meaning that as long as the swing producer stays saturates, at least one of the extra producers will be active.

Therefore, there is no easy fix for making the tertiary controllers revert when there is an overshoot in supply. However, the controllers do revert when the system returns to nominal values, making this issue minimal.

7.2.4 Parallel Control Performance

The parallel control structure was the worst-performing tertiary control structure. The reason for this is likely that the system model is not intended for parallel control. As stated in the theory section, the primary manipulated using a PI-controller should be chosen as the variable with the most significant steady-state effect. However, in this thesis, both manipulated variables have an equal steady-state impact and are equivalent by definition. Therefore, the parallel control performs poorly.

Another thing to mention is the fact that the P-controller should only have a stabilizing effect. In this thesis, we want to extend the operating range using all available input. Using P-control alone causes the extra producer II not to be fully utilized and only implement the stabilizing effect. For a specific disturbance, the P-controller gain could be changed such that it would, in fact, fully use extra producer II. However, this could cause the controller action to be too immense for a lesser disturbance, causing a supply overshoot. Therefore, this is not a good solution, and parallel control should not be selected for its purpose in the modeled steam distribution network in this thesis.

7.2.5 Different Setpoints Performance

The different setpoint controllers both works and are utilized entirely at an instance for a considerable disturbance. This works well for the intended use in the steam distribution network. However, there are some issues. For a smaller disturbance where the extra producers are utilized, some controller action will be taken on both the extra producers initially. This is as long as the valve position is above the given setpoints for the extra producers.

One way to make a larger distinction between the controllers, thus making them act slower, is by increasing the difference in setpoints. This could separate the use further, and using both at an instance could be avoided. The setpoints used were 0.9 and 0.85 for extra producer I and II, respectively. By increasing and decreasing the setpoints to 0.95 and 0.8 on extra producer I and II respectively, faster action would be taken on extra producer II, possibly reducing the initial overshoot caused in the configuration now. However, if operating only one manipulated variable at a time is the ultimate process control objective for the extra producers, then valve position control or split range control should instead be used.

7.2.6 Valve Position Control Performance

Valve position control works as intended, increasing operating range by using one manipulated variable at a time, except for a small time frame where both manipulated variables are active. This controller structure works well for its intended use. However, there is an issue caused by the saturation in the secondary controller structure.

As mentioned, a small supply overshoot could happen for a large disturbance, which the droop producer quickly balances out since they are the fastest controllers. However, this causes the swing producer valve position, the output for the extra producer I in the valve position controller, to remain saturated. Thus, the extra producer I valve position, which is the output for the extra producer II valve position, will also remain saturated, forcing saturation in the extra producer II valve position. So the issue lies within the primary controller, which might take advantage of being slower or on a more similar time scale with the secondary controller to maintain the input usage prioritization list.

7.2.7 Quaternary Control Performance

The quaternary swing consumer control works as expected. However, when the system pressure drops below the threshold, the quaternary controller will be shut down the entire consumer supply, which could be unfortunate if a consumer relies on a steady supply from the quaternary consumer. However, this could be easily fixed using a larger flow through the swing consumer pipeline. Thus, more flow could be given up before the valve would have to shut entirely to balance supply and demand.

7.2.8 Model Predictive Controller Performance

The initial model predictive controller responses were perfect according to the controller objectives. This is because the MPC includes disturbances measurements. More importantly, the MPC model and the plant mode are identical, which causes perfect future predictions. In addition, the disturbances step time coincides with the MPC sampling time. Thus, the model predictive controller in this thesis is an ideal feedforward control structure.

The model predictive controller structure experienced some issues. First off, the tunings could not create the same nominal system as the decentralized controller structure. This is likely because of numerical errors in the system calculations. Also, the objective function did not include any constraints on the involuntary droop consumers.

The fact that the objective function did not include the involuntary droop consumers could cause the system to create a slightly different nominal system than the one in the decentralized control structure. This is shown especially for a step-change back to nominal values, say in Figure 6.21. When the system in the response figure reverts the disturbance, different nominal values are obtained, which could be caused by the fact that the objective function does not include all system inputs.

Not using all inputs in the objective functions could also be a cause for issues. The model predictive controller usually wants to control all inputs simultaneously, which is also the case for the extra producers. Even in the tuned system without any disturbance, the extra producers are slightly turned on, which is caused by the system trying to use all available inputs. Therefore, using controllers that should be off at nominal point can be hard when using a model predictive controller.

An idea to cope with this problem could be to use a tracking model predictive controller for the extra producers and increase the weight. This would, however, cause the entire process to be re-tuned, costing much time and effort, which is probably the biggest issue. For every small change made to the plant or model, loads of work must implement even small changes to the model predictive controller.

7.3 Disturbances

The system disturbances were changes in the setpoint for the normal consumers and pressure disturbances on the consumer and producer side. Thus, a real steam network would represent two different types of disturbances: unstable supply and unstable consumption for pressure disturbances. For normal consumers, the disturbances would represent changes in demand based on consumer requirements. Both types of disturbances are discussed in the following subsections.

7.3.1 Pressure Disturbances

Step pressure disturbances for droop producer I, the swing producer, and the swing consumer were defined and implemented. The consumer and producer side pressure disturbances did not affect the system in noticeable amounts, causing minimal controller actions. Therefore, they were not added in the main discussion, as they showed only the same as the main disturbance from the normal consumers to a smaller extent.

The pressure disturbances giving only a small effect on the system are expected since the pressure disturbances would: 1) Stay very local to the consumer or producer where they were enacted. 2) The pressure disturbances were forced to be very small compared to the flow setpoint changes in the normal consumers, as the pressure difference in the network is very small between the main pipeline pressure P and the consumer and producer side. Furthermore, since a reverse flow is deemed impossible, the supplier pressure cannot drop below the main pressure P, and the main pressure P cannot drop below the consumer pressure, limiting the magnitude of the pressure disturbances.

However small the effect from the pressure disturbances was, they were essential to analyze along with the normal consumer disturbances, as a real network is highly likely to experience pressure disturbances. Boilers supplying steam will operate with smaller pressure variations, and equipment can change the pressure on the demand side depending on use. Therefore it is important to know that the system functions well for smaller pressure disturbances. This is also why a combined disturbance plot is included, to see how the system responds to smaller disturbances at the same time in the system.

In conclusion, the system performs well for stabilizing when smaller pressure disturbances occur. For most pressure disturbances, the effect will be treated locally on the consumer or producer where the pressure disturbance happens. This is as mentioned because the regulatory controller layer will act to stabilize the system if it can. If the effect is not local, and the consumers or producers experiencing the disturbance cannot locally stabilize the disturbance on a regulatory level, the supervisory prioritization list is enacted to stabilize the system.

7.3.2 Flow Disturbances

The normal consumers work as a disturbance by only being regulatory controllers. Therefore, for a setpoint change, the controllers would only act locally, disturbing the whole system. This is realistic for a real plant as well, where they're often most likely would be some controllers only involving regulatory control.

Compared to pressure disturbances, normal consumers enforce large disturbances on the system. This is because the flows through the normal consumers are designed to be very large at maximum capacity compared to the total system flow. This means that the effect for a maximum normal consumer disturbance also will be much larger than for a maximum pressure disturbance, and not only local as most of the pressure disturbances.

For a real steam distribution network, the largest disturbances caused by the normal consumers performed in this thesis are not likely to happen. In this thesis, the idea of enacting them was merely to see the effect of primary, secondary, tertiary, and quaternary control structures and comparing the response to a model predictive controller. However, as mentioned, smaller disturbances are likely for regulatory level controllers in a real steam distribution network.

7.3.3 Model Predictive Control Disturbances

For the model predictive controller, there is two ways to implement disturbances. That is, both in the plant model solver, which integrates the system for a given set of current states and directly to the nonlinear solver, which solves the collocation problem.

In this thesis, only a disturbance directly to the nonlinear solver is given, which in future works should also be given to the plant, as in reality, plant disturbances are what would happen. This could cause the model predictive controller to enact the disturbances slower and make the performance slightly worse than the initial performance.

7.4 Simplifications and Other Considerations

Multiple simplifications are made to the modeled system. This section discusses the most critical simplifications and their impact on the results obtained in this thesis.

7.4.1 System Solvers

A stiff system involves both fast and slow model dynamics, which is true for the model in this thesis. There are fast flow dynamics and slower pressure dynamics. Also, the controllers are separated on faster and slower timescales. Therefore, a stiff solver is required for the models. For the decentralized system, MATLAB Simulink gives the choice of using ODE15s, which is MATLAB's recommended solver for differential-algebraic equations, which this system embraces. Therefore, this is probably the most efficient solver that could have been selected for the decentralized control structure.

For the model predictive controller, the solver used by CasADi is IPOPT (Interior Point Optimizer). It is an open-source software package for large-scale nonlinear optimization. For the problem encountered in this thesis, the solver performs well. However, it is not the same solver used in the decentralized control structure, and minimal numerical errors could make the result minimally different from if the same solver was used. However, this is not considered to have affected the results noticeably.

7.4.2 Ideal Gas Law

The arguments presented for the ideal gas law are valid only for high temperatures and relatively low pressures. Thus, they should be valid for this system. However, if it turns out that the ideal gas law is not valid, we should also consider density in the valve equation, and we would have to reformulate our entire problem structure, complicating the model to the extent.

7.4.3 Pure Gas System

The assumption that the system models a pure gas system is realistic as long as supplier pressure does not exceed nominal value at 15 $[bar]$, which it is not assumed to do. For a positive disturbance in P_{Supply} however, exceeding 15 $[bar]$, the system dew point temperature would increase, and the system temperature would have to be increased to avoid condensate in the system. With condensate in the system, the assumptions for the ideal gas law would most likely be invalid, and a more sophisticated gas behavior model would be required.

7.4.4 Valve Dynamics

Valve dynamics used in this thesis are linear. Linear valves for gas handling do exist, so this assumption is valid. Using another type of valve would also be possible without changing the conclusions drawn from the closed-loop responses because the valve dynamics are not the determining factor for the controller responses in closed-loop simulations. The time scale separation is.

7.4.5 Time Delays

Time delays were not included in the model, except for the extra producers, as their response procured a time delay. A real process would almost always include some form of time delay in the process. However, this would not change the end results other than the tunings, which would have to be redone. The closed-loop responses would show the same results.

7.4.6 Alternative Controller Structures

An alternative to the extra producer configuration using valve position control is split range control. Split range control could also have been tried out. However, the results obtained in this thesis using valve position control are somewhat similar to what has been found from split range performance[41]. Reyes-Lúa et al. (2019) show that the main difference between split range control and the valve position control used in this thesis is that for split range control, strictly one manipulated variable at a time. Then, MV-MV switching is performed once one manipulated variable saturates. On the other hand, valve position requires back-off, meaning there is a short period where both manipulated variables will be active before the MV-MV switching occurs. Therefore, split range control implementation on the tertiary controller is expected to perform relatively equal to the one using the valve position control.

Chapter 8

Conclusion and Future Work

This master thesis aimed to use supervisory control structures to balance supply and demand in a steam distribution network. The thesis modeled a simplified steam distribution network in which it implemented both a decentralized control structure with different configurations and a multivariable control structure.

The control objective for the decentralized and multivariable control structure was to balance supply and demand for disturbances caused in the network, maintaining main network pressure P. This was done using a specific priority list for input usage utilizing the idea of frequency control. That is, primary droop producer control (fast), secondary swing producer control (slow), tertiary extra producer control (slower), and quaternary swing consumer control (if all else fails) as stated in Table 4.1. In addition to that, the network was modeled with involuntary droop consumers, not involving any control structure. They were instead self-regulating because of interactions in the network. Last, normal consumers were involved, using regulatory control to control flow and thus enacted disturbances to the modeled network.

Separating primary, secondary, tertiary, and quaternary control on different time scales worked well for regulating a steam distribution network. The structure faced some issues, especially for extensive disturbances causing more supply than required because of controller overshooting, causing periods of imbalance to the system. Because the primary droop producers were set to act fastest, they were also quick to return to their nominal values when the system did an overshoot. Thus, initially, for immense disturbances, the system corresponded fast but not that accurately, causing prioritization problems when reaching a new steady-state. However, the decentralized control structure returned to nominal conditions without issues whatsoever for a return to nominal values.

The decentralized control structure was simulated for three different configurations on the extra producer control structure: parallel control, controllers with different setpoints, and valve position control. It turned out that parallel control performed worst in terms of utilizing the extra producers fully, as observed in Figure 6.1 and 6.7. This was because both extra producers were equal in terms of steady-state effect and should not be used in such a configuration. Parallel control should instead be used when one input has a more considerable steady-state impact than the other.

The controllers with different setpoints and valve positions both utilized the extra producer control fully but used the manipulated variables in different ways, as shown in Figure 6.2 and 6.3. The different setpoints configuration used both extra producers initially and simultaneously, while the valve position control utilized one manipulated variable except a small time frame when the MV-MV switch happened.

For the use in a steam distribution network, the controllers with different setpoints performed best as they utilized both extra producers to their full potential on the fastest time scale, however for an actual plant operation, the need might be different, and the valve position control could be more beneficial.

The model predictive controller was designed to resemble the same dynamics as the decentralized control structure. However, the control structure differs from the decentralized control structure as it used all required inputs at the same time, thus responding much faster than the decentralized control, as seen in Figure 6.19. The model predictive controller had a perfect initial response because it utilized a theoretical model, using a model with no imperfections. In a real plant, the model would have flaws, and a worse performance would be expected.

For a return to nominal values in the model predictive controller, the model inputs did not return to initial steady-state values, considered the cheapest operation window. The controller instead calculated the most reasonable new steady-state for given conditions according to its related objective function. Therefore, with more tuning and a more complex model predictive controller, results could have been better.

This shows that a model predictive controller requires more planning and implementation considerations and the required modeling, extra implementing costs, and operating costs. Therefore, it is recommended to explore decentralized control structures for steam distribution network control before considering model predictive control. Specifically, it is recommended to prioritize control using time scale separation control to ensure enough backup control for considerable disturbances. Also, for a network where fast response is critical, controllers with different setpoints are recommended above valve position control.

8.1 Future Work

It is recommended to expand the modeled steam distribution network in this thesis also to include process equipment. Process equipment would be boilers on the producer side and consist of distillation columns, heat exchangers, and more on the consumer side.

After adding complexity to the model, imperfect measurements, noise, and time delays should be added to make the model represent a realistic process plant. Then, both the decentralized and multivariable controller structures should be implemented on the more realistic network.

The more realistic network with the proposed control structures in this thesis should verify the results obtained in this thesis or possibly disprove the conclusions when process equipment, delays, and imperfect measurements are added.

There are also supervisory control structures for the decentralized control structure that are not utilized in this thesis, such as split range control. Future works should include a possibility analysis for whether other decentralized control structures could be helpful in steam distribution network control. If it turns out there could other usable control structures, their implementation and performance should also be put to trial.

Disturbances should also be implemented directly in the plant model in the model predictive controller instead of the nonlinear solver, as this is more realistic for a real plant. This could show that the performance of the model predictive controller is, in fact, worse than what is found in this thesis.

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Appendices

Plot Results

This section presents the plots not shown in the results part. This includes the open-loop system responses and closed-loop simulations not shown in the thesis.

Flow Controller Tunings

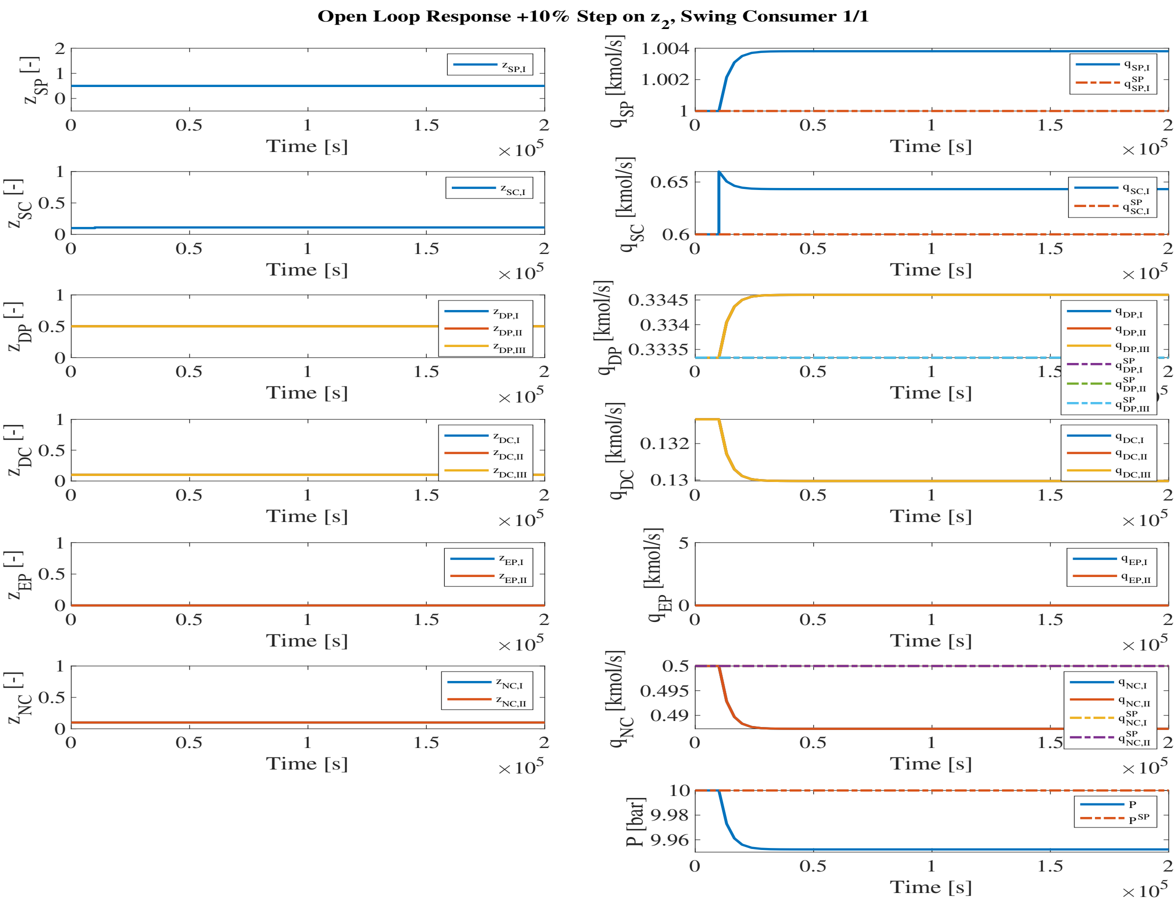

Figure 1: Open loop system response for a +10% step on $z_{SC,I}$, that is valve position for swing consumer I at $t = 1000$s.

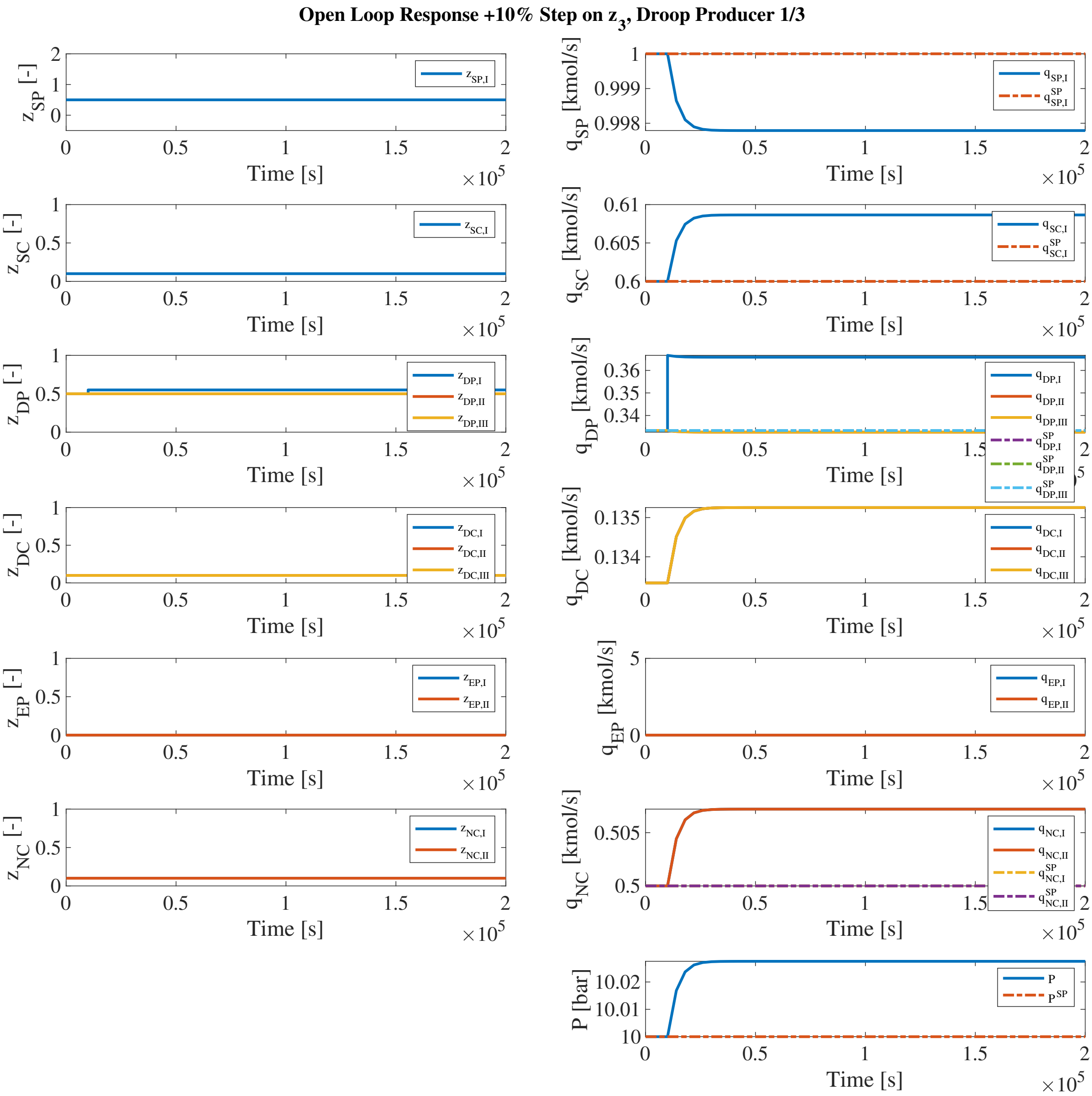

Figure 2: Open loop system response for a +10% step on $z_{DP,I}$, that is valve position for droop producer I at $t = 1000$s.

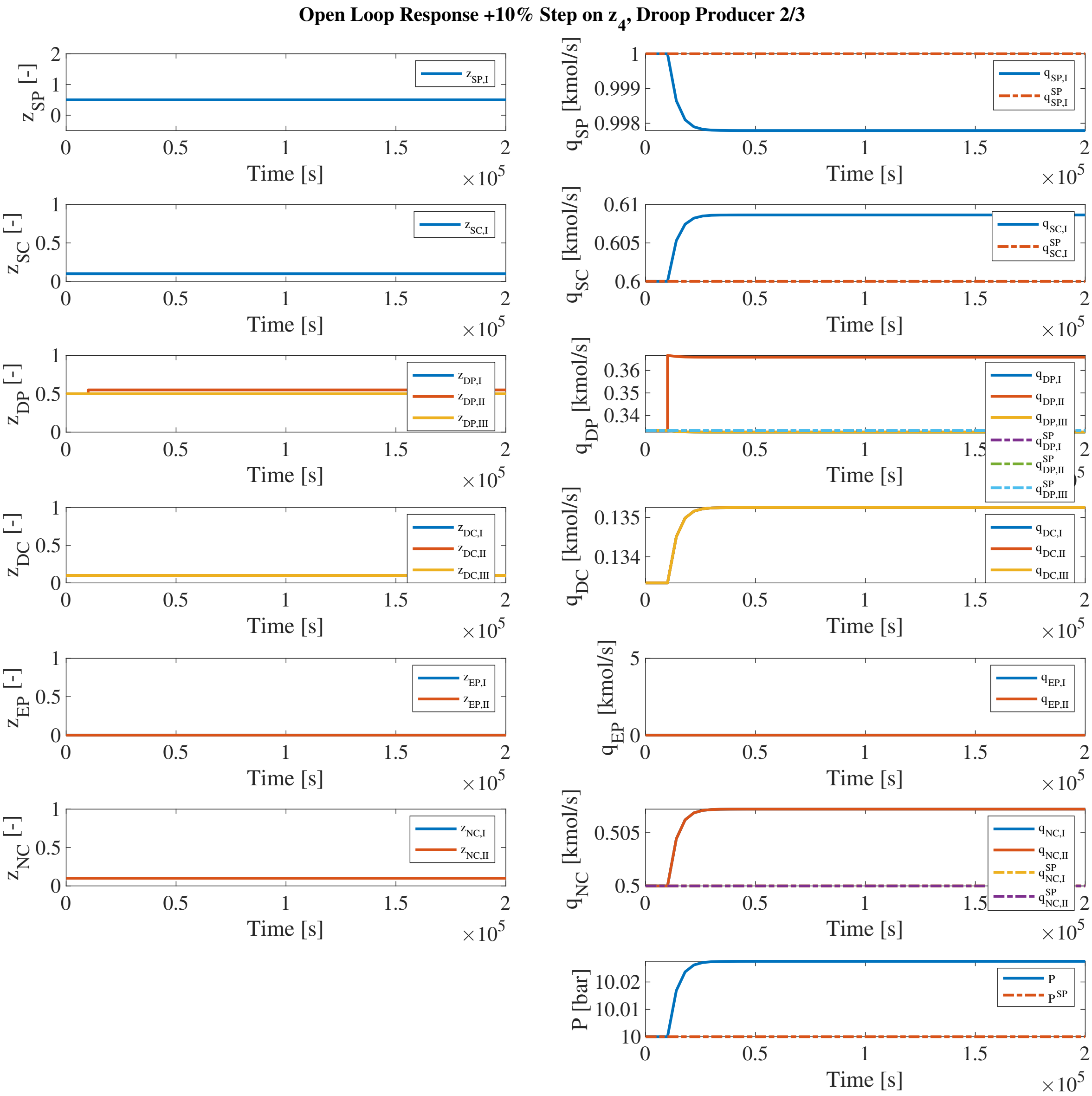

Figure 3: Open loop system response for a +10% step on $z_{DP,II}$, that is valve position for droop producer II at $t = 1000$s.

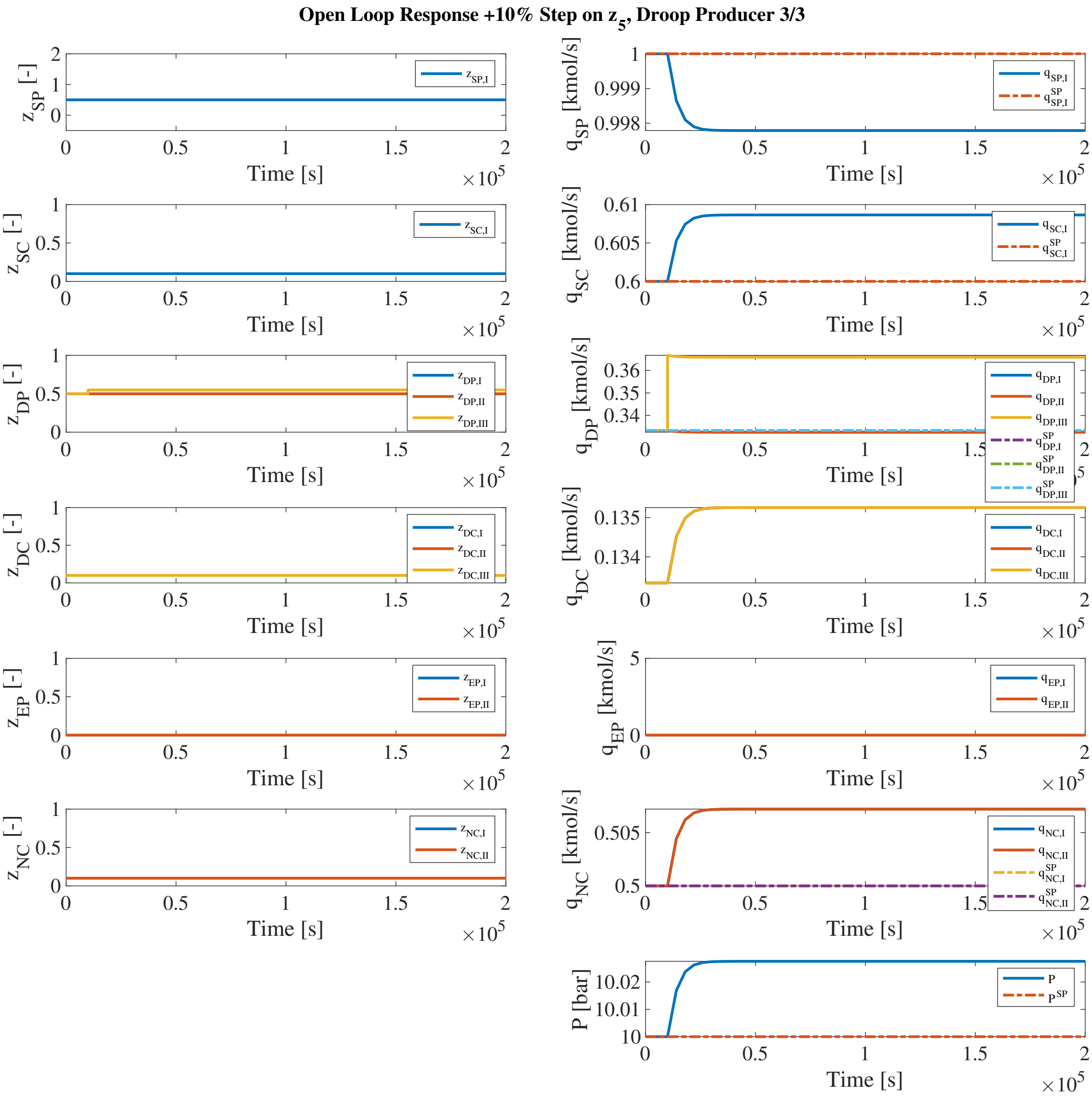

Figure 4: Open loop system response for a +10% step on $z_{DP,III}$, that is valve position for droop producer III at $t = 1000$s.

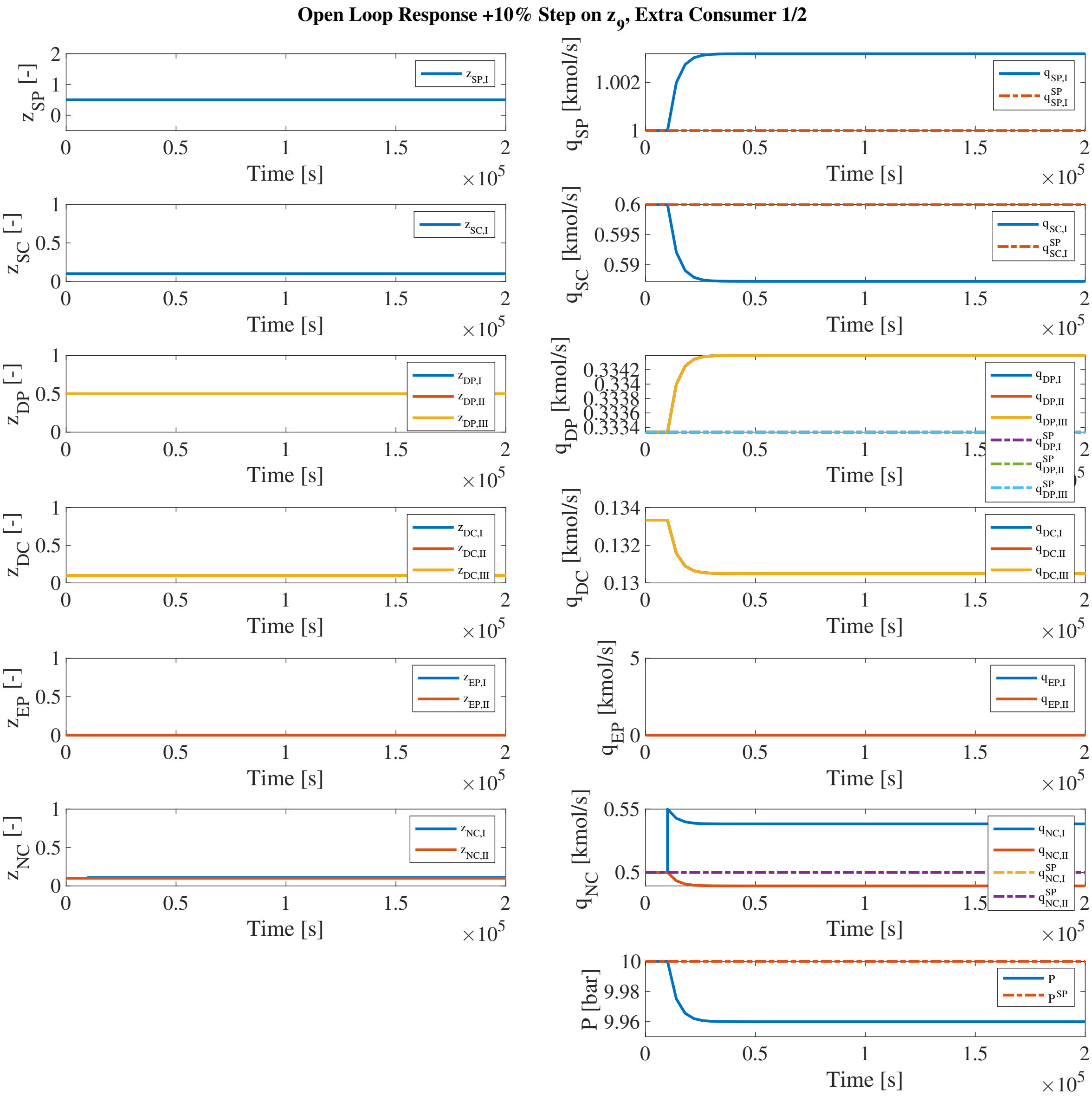

Figure 5: Open loop system response for a +10% step on $z_{NC,I}$, that is valve position for normal consumer I at $t = 1000$s.

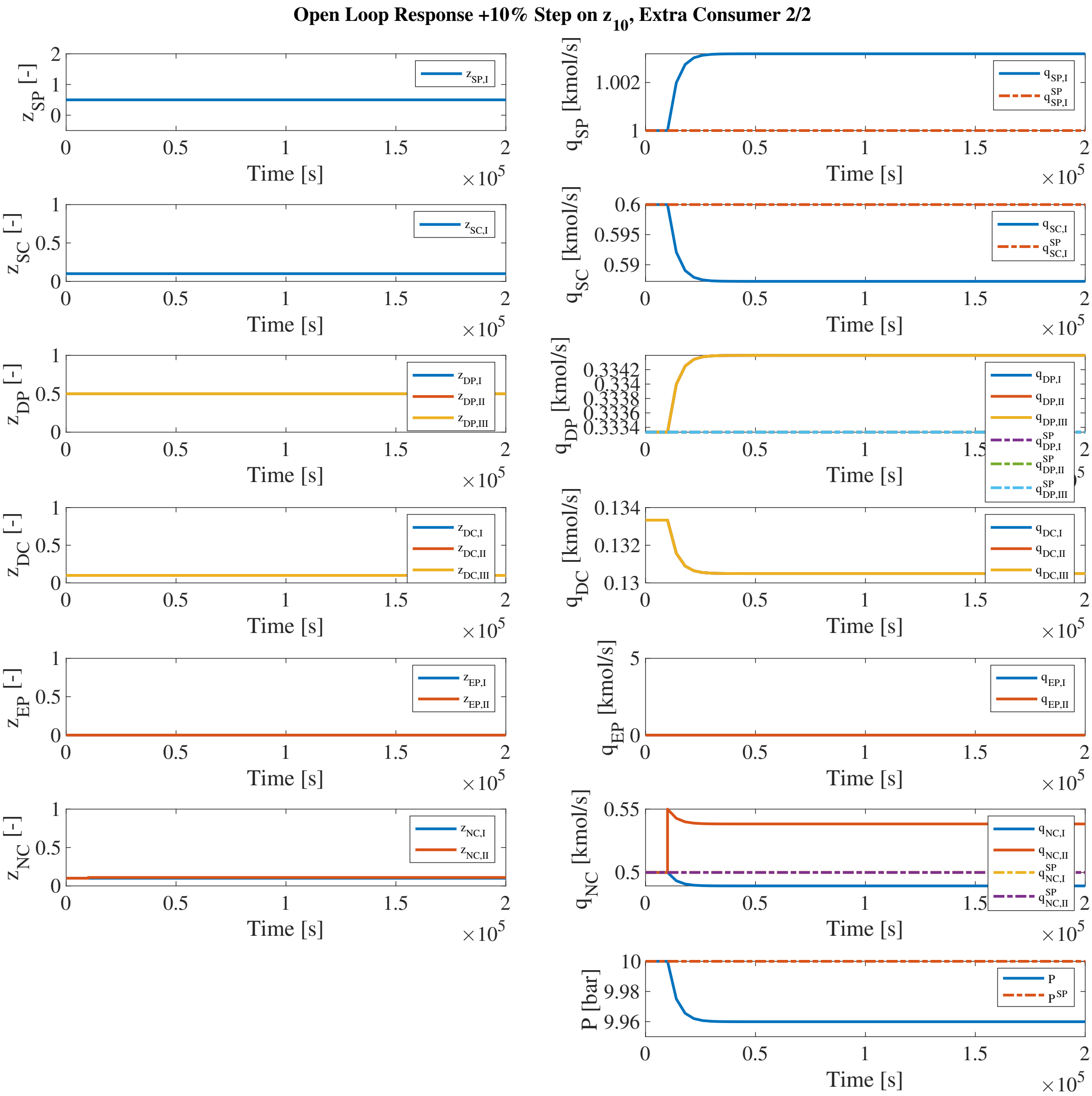

Figure 6: Open loop system response for a +10% step on $z_{NC,II}$, that is valve position for normal consumer II at $t = 1000$s.

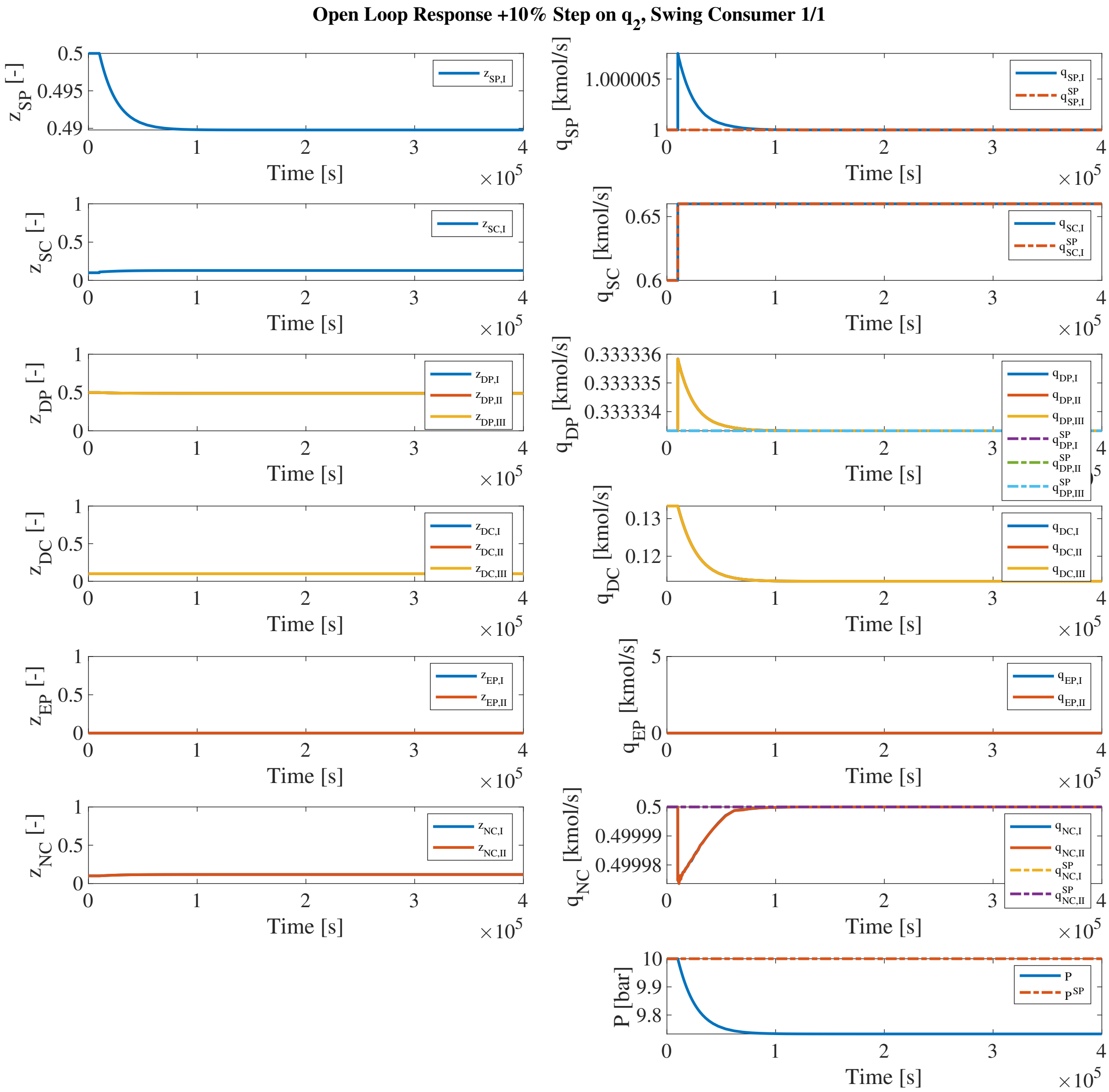

Figure 7: Closed loop system response for a +10% step on $q^s_{SC,I}$, that is the flow setpoint for swing consumer I at $t = 1000$s.

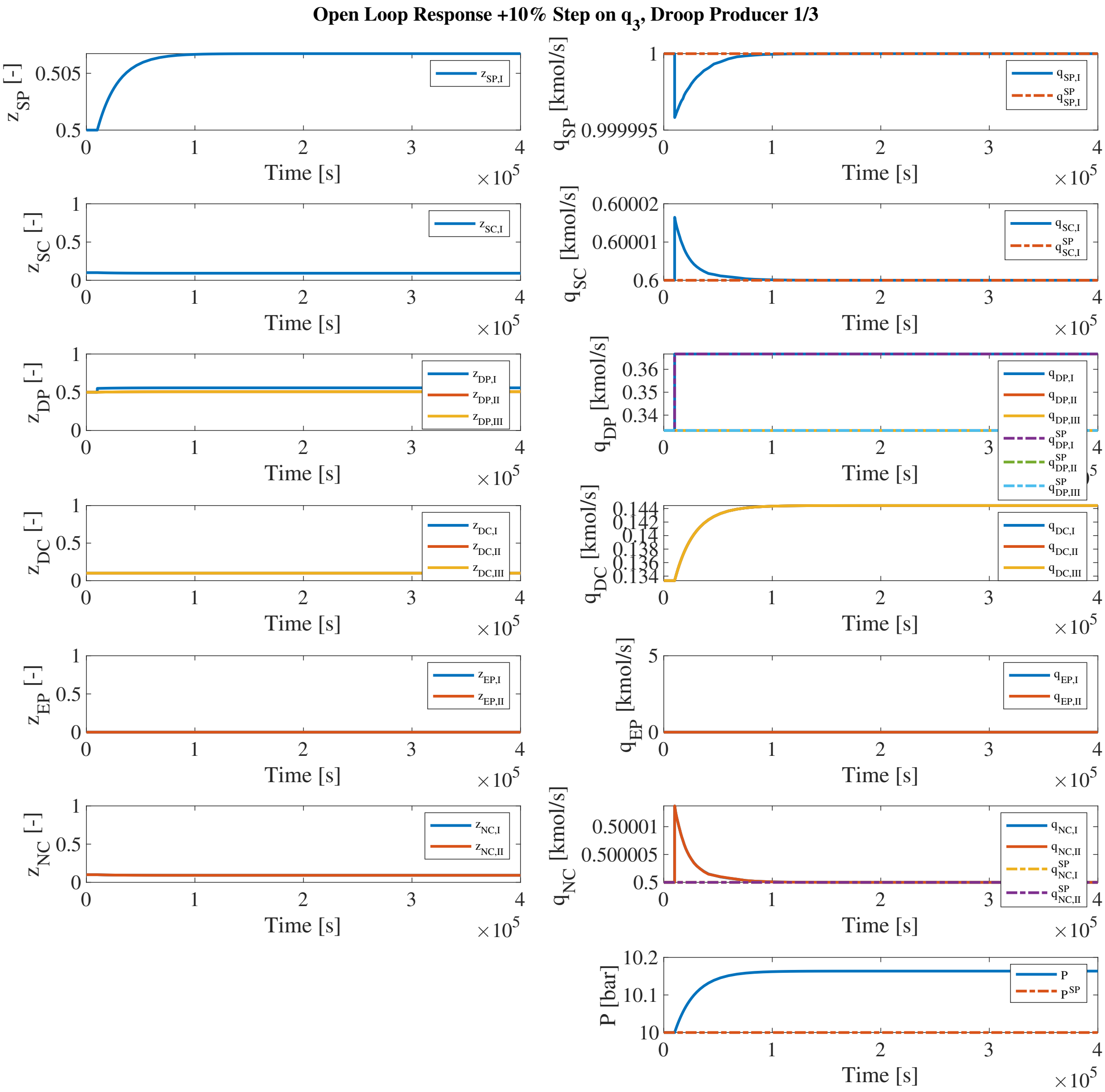

Figure 8: Open loop system response for a +10% step on $q^s_{DP,I}$, that is the flow setpoint for droop producer I at $t = 1000$s.

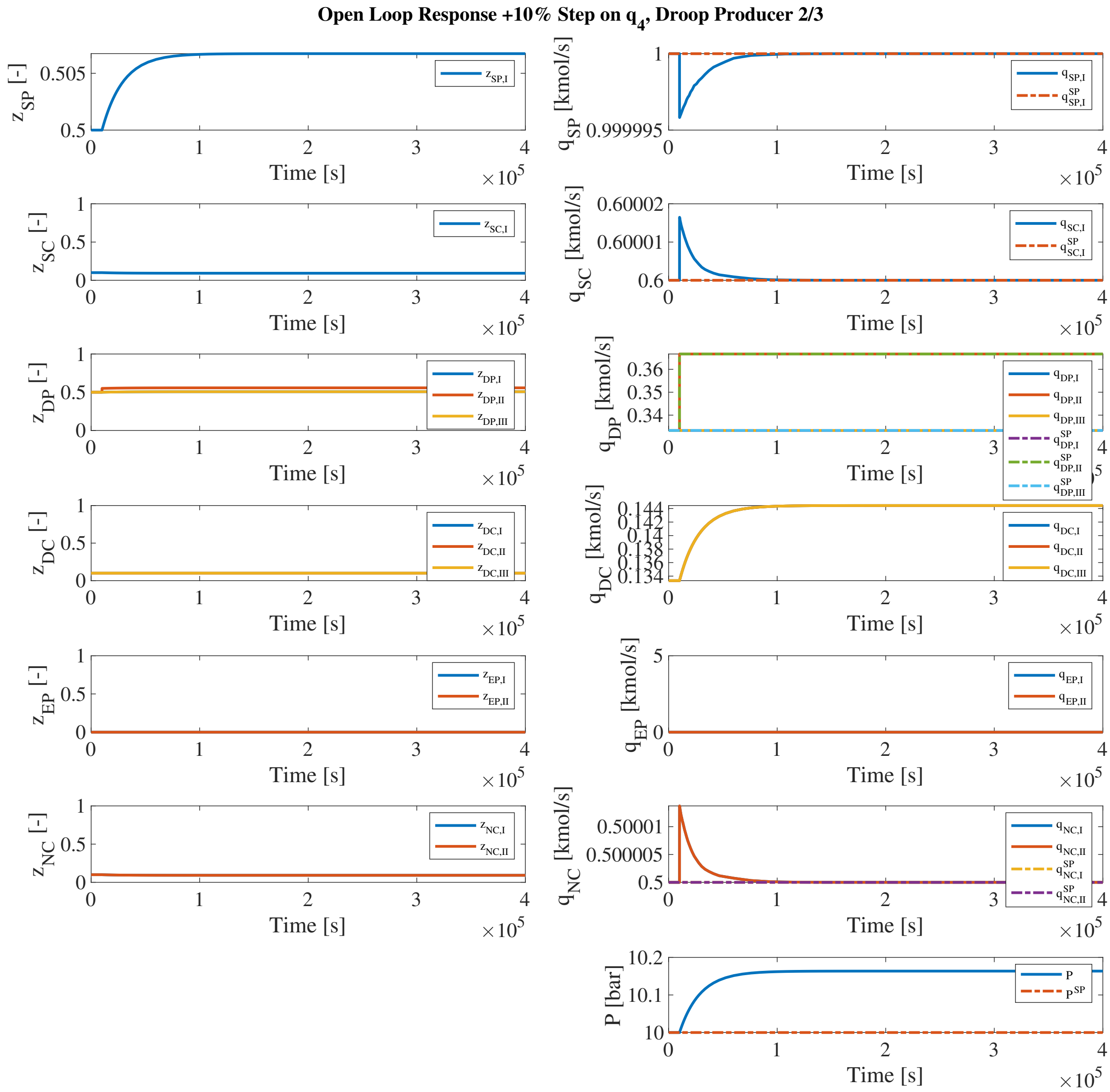

Figure 9: Closed loop system response for a +10% step on $q^s_{DP,II}$, that is the flow setpoint for droop producer II at $t = 1000$s.

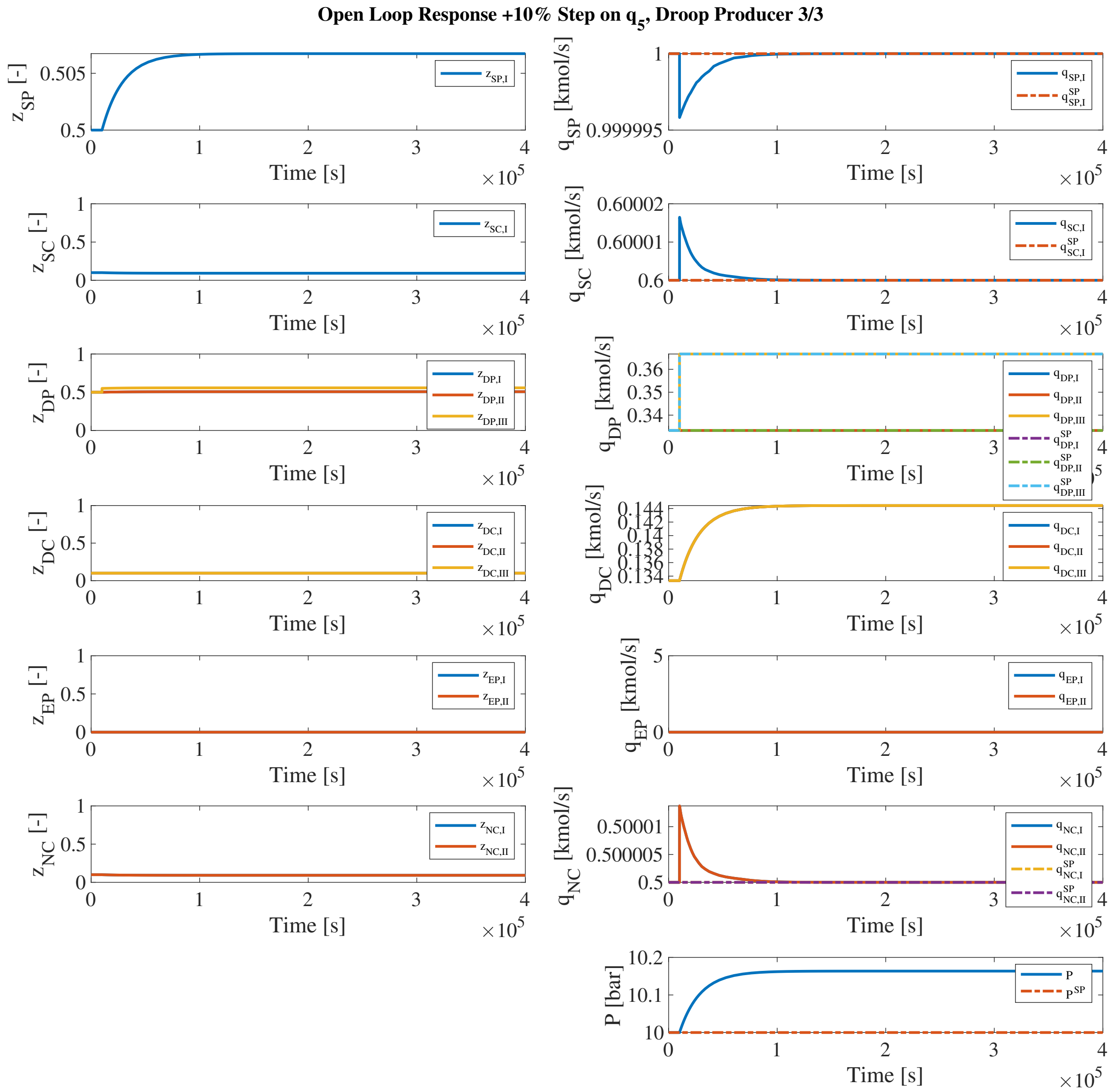

Figure 10: Closed loop system response for a +10% step on $q^s_{DP,III}$, that is the flow setpoint for droop producer II at $t = 1000$s.

Tertiary Controller Tunings

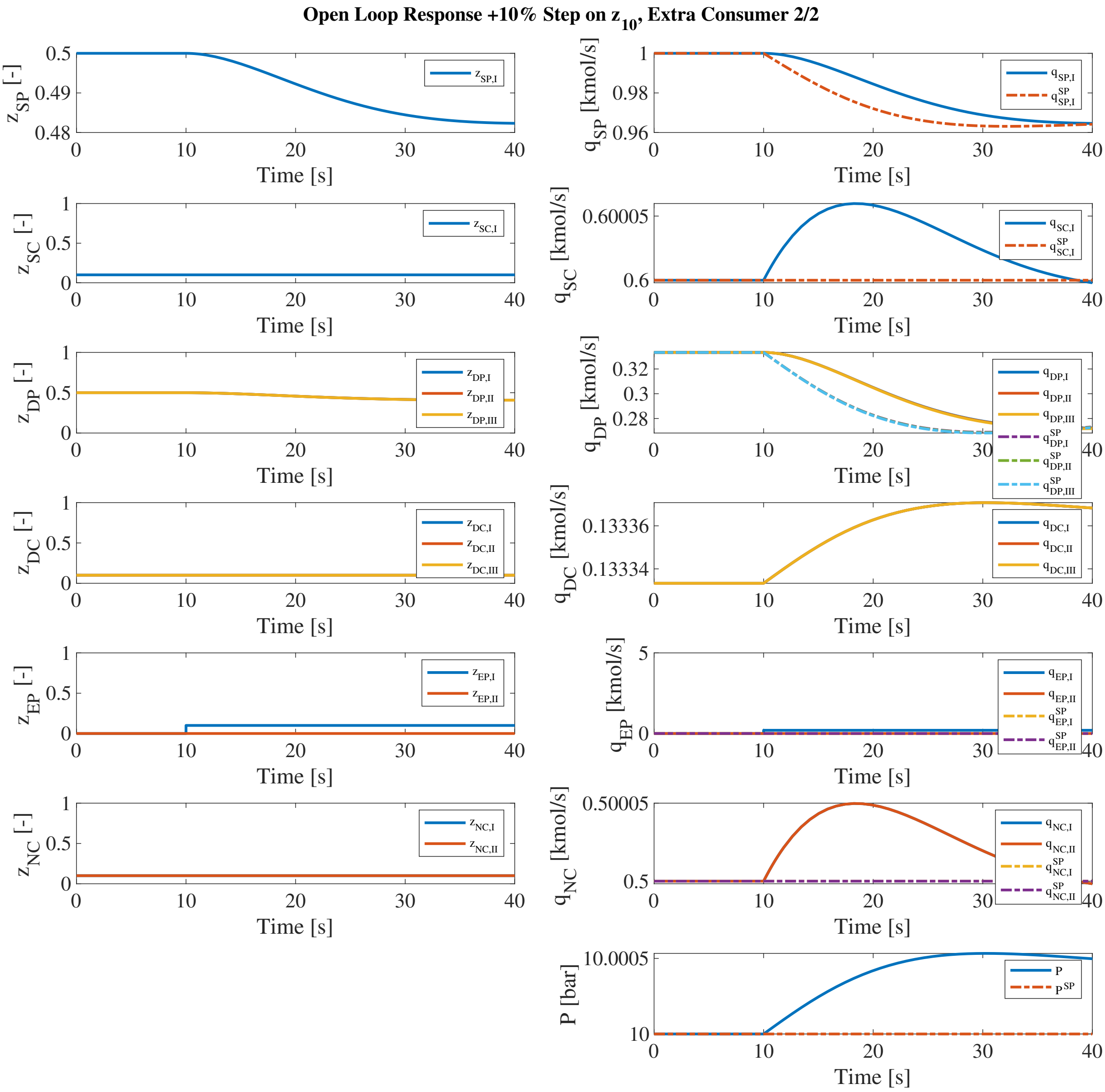

Figure 11: Closed loop system response for a +10% step on $z_{EP,II}$, that is valve position for extra producer II at $t = 10$s.