Master's thesis

NTNU
Norwegian University of Science and Technology
Faculty of Natural Sciences
Department of Chemical Engineering

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Control structures for consistent inventory control with moving bottleneck

Master's thesis in Chemical Engineering and Biotechnology
Supervisor: Sigurd Skogestad
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June 28, 2020

Abstract

There are three cases of inventory control that were considered in this report, each with their own objective: Maintaining production, where the objective is to keep mass flows between units constant despite temporary bottlenecks. Averaging level control, where smooth changes in flows is desired and levels can vary freely within given bounds. Tight level control, where the inventories must be controlled tightly (small deviations from desired value).

The aim of this study was to find a control structure that performed best at the case of maintaining production. The control structures tested where spilt range control (SRC), generalized split range control, controllers with different setpoints and model predictive control (MPC). The studied system consisted of three tanks in series, each representing the inventory of a real industrial unit (e.g. a tank, separator, reactor, distillation column or an evaporator). Matlab and Simulink was used for the modeling, together with CasADi for the implementation of MPC.

The simulations showed that the use of two controllers with different setpoints performed best at this, because it allowed the inventories to fill up during a temporary bottleneck. The three other control structures had only one setpoint for each tank, which they would always strive to keep. They also require some additional logic to handle switching between manipulated variables (SRC, generalized SRC) or a model of the plant (MPC) which can take a lot of time to design.

Sammendrag

Tre forskjellige variasjoner av regulering av innhold ble vurdert i denne rapporten, hver med sitt eget formål: Opprettholding av produksjon, hvor målet er å holde massestrømmene mellom prosessenheter konstant til tross for midlertidige flaskehalser. Utlikning av nivåer, hvor små endringer over tid i massestrømmene er ønsket og nivåene i tanker kan variere fritt innen gitte grenser. Stram nivåregulering, hvor regulering av innholdet i tanker er viktig (lite avvik fra ønsket verdi).

Målet med studied gjort her var å finne en reguleringsstruktur som var best til å opprettholde produksjonen gjennom prosessen. Regulering med bruk av split range (SRC), generalisert split range, regulatorer med forskjellig setpunkt og modell prediktiv regulering (MPC) har blitt testet. Prosessen som ble studert bestod av tre tanker i serie, som representerte innholdet i en virkelig industriell enhet (f. eks. en tank, separator, reaktor, destillasjonskolonne eller en fordamper). Matlab og Simulink ble brukt for modeleringen, samt CasADi for implementasjon av MPC.

Simuleringene viste at bruken med to regulatorer med forskjellig setpunkt var best til å opprettholde produksjonen, fordi det gjorde det mulig å fylle opp innholdet i tankene når en flaskehals inntraff midlertidig. De andre reguleringsstrukturene hadde kun et setpunkt for hver tank, som regulatorene alltid vil strebe etter å opprettholde. De krever også ytterligere logikk for å bytte mellom bruk av manipulerbare variabler (split range og generalisert split range) eller en model av prosessen (MPC) som kan ta lang tid å lage.

Preface

I have written this thesis as the final part of my Master of Science in Engineering degree at the Norwegian University of Science and Technology (NTNU).

I would like to thank my supervisor Sigurd Skogestad and co-supervisor Cristina Zotica for being able to work with them for this project. I would like to give an extra thanks to Cristina for the discussions, help and comments to my report both in my Master Thesis and feedback in my specialization project last semester.

Declaration of Compliance

I, David Alexander Lillevold Skaug, hereby declare that this is an independent work according to the exam regulations of the Norwegian Univeristy of Science and Technology.

Contents

List of Figures

List of Tables

Nomenclature

Acronyms

CV Controlled variable

LC Level controller

MPC Model predictive control

MV Manipulated variable

PID Proportional Integral Derivative

SP Setpoint

TPM Throughput manipulator

Symbols

τ_D Derivative time

τ_I Integral time

e Error

K_c Controller gain

u A generic manipulatd variable

y A generic controlled variable

Chapter 1

Introduction

In a chemical plant there may be numerous units (e.g. tanks, separators, reactors, distillation columns, evaporators) that require their inventory to be within a maximum and minimum limit, both so the contents does not spill out of the container (which can be both economical and environmentally harmful) nor that it runs empty so that the plant does not stop production. This is an important part in the process industry, and is called inventory control [3].

Three cases of inventory control and their objective will be discussed in this report (S. Skogestad 2020, personal communication, 7 April):

- Maintaining production, where the objective is to keep mass flows between units constant despite temporary bottlenecks.
- Averaging level control, where smooth changes in flows is desired and levels can vary freely within given bounds.
- Tight level control, where the inventories must be controlled tightly (small deviations from desired value).

In addition to inventory control in a plant there must also exist an degree of freedom where it is possible for plant personnel to set a production rate, which is often a flow rate set at either the plant inlet, outlet or inside the plant. This is called the throughput manipulator or TPM for short. The inventory control structure is designed around the location of the TPM, as shown in Figure 1.

This ensures that the steady-state mass balance is fulfilled, and the resulting inventory control structure is said to be consistent [3].

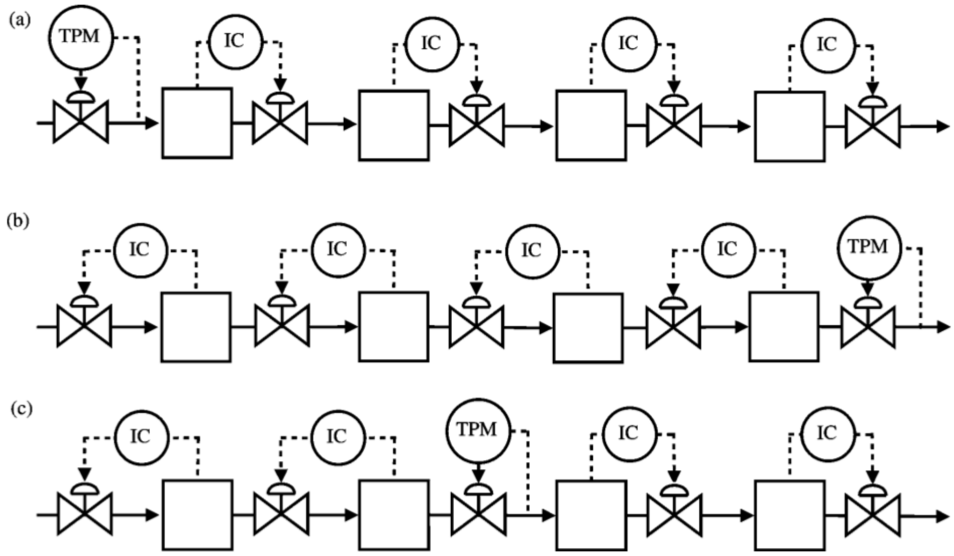


Figure 1: Consistent inventory control, to ensure that the steady-state mass balance is fulfilled. Figure retrieved from [3].

In terms of economics, it is usually so that optimal operation is obtained when the throughput or production rate is maximized, which means that the TPM should be placed at the bottleneck of the plant. However, the bottleneck is highly likely to move in a plant due to a change in active constraints, for example, some part of the plant is shut down for maintenance. This would require the inventory control structure to be rearranged to account for the new location of the TPM, which may be confusing to the plant personnel [4].

The work conducted in the specialization project was to design a supervisory control layer that could automatically move the TPM to the bottleneck of the plant, when different disturbances occurred. The system studied consisted of three tanks in series, where the levels where the controlled variables (CVs) and the flows rates between each tank the manipulated variables (MVs). The supervisory control layers where designed using split range control (SRC) and model predictive control (MPC). The control structure with SRC was designed such that each tank (with their own SRC) could manipulate the inlet and outlet flow of the tank, with use of min-selectors. The control structure with MPC were designed such that a desired throughput (flow rate) could be set in the cost function and the MPC would manipulate all flows simultaneously, settling at the desired flow rate. The conclusion was that MPC performed much better than SRC, due to large delays in the split range controllers when they switched from manipulating one MV to another [13].

A similar system to the one studied in the specialization project will be studied in this report, with three tanks in series but with valves instead of flows as MVs. This project will consider especially the case of ”maintaining production” for inventory control, where we want the flows between each unit or inventory to be constant during a temporary disturbance. As in the specialization project, several supervisory control layers will be designed using different control structures: split range control, generalized split range control, model predictive control, valve position control and multiple controllers with different setpoints.

This report is structured as follows. In Chapter 2 the theoretical background is presented, where the process control hierarchy for an entire chemical plant is described, along with each individual control structure and the process description for the studied system. In Chapter 3 the modeling of the system is presented together with the nominal operating point. Chapter 4 described the creation and implementation of each control structure. Chapter 5 shows the different control structures subjected to a number of disturbances and their responses. Chapter 6 discusses the results and findings and some concluding remarks and ideas for future work are given in Chapter 7.

Chapter 2

Theory background

2.1 Process control

The main purpose of process control is to keep a given process at the preferred operating conditions, while maintaining safety, environmental concerns and meeting the required quality of the product. For instance in an oil refinery there exists thousands of variables that all needs to be controlled for these requirements to be met, such as compositions, temperatures and pressures.

The controlled variables (CVs) are kept at or close to their desired value (setpoint) by changing other process variables know as manipulated variables (MVs). MVs can for instance be a valve position, a flow rate or the speed of a compressor. Other process variables that has an effect on the CV, but can not be manipulated, are called disturbance variables (DVs). The CVs are very often in industry controlled by the use of Proportional-Integral-Derivative (PID) controllers [12].

2.1.1 Proportional-Integral-Derivative controller

The output of a PID-controller consist of, as the name would suggest, three parts: a proportional-term (proportional to the error of the CV), integral-term (proportional to the error of the CV integrated over time) and derivative-term (proportional to the derivative of the error). The controller can be described by the equation

$$u(t) = K_c\left(e(t) + \frac{1}{\tau_I}\int_0^t e(\tau)d\tau + \tau_D\frac{de(t)}{dt}\right) \tag{1}$$

where u is the output from the controller and e is the error between the desired (setpoint) value and the measured value of the CV. The tuning parameters for the controller are proportional gain K_c, integral time τ_I and derivative time τ_D [5].

2.1.2 Controller tuning

Even though the PID-controller only has three tuning parameters, it is not easy to obtain good values for them without having a systematic procedure. Such a procedure is presented by Skogestad [15] (SIMC-PID tuning rules) and consist of two steps:

- Obtain a first- or second order transfer-function model
- Obtain tuning parameters from the model

A first-order model may be obtained by doing an open-loop step response of the process, practically meaning the controller we want to obtain the tuning parameters for is put in "manual" mode. This is done in Figure 2, where y is the controlled variable, u is the controller output, θ is the time delay and τ_1 is the open loop time constant, i.e. the time it takes for the output to reach 63 % of its total change.

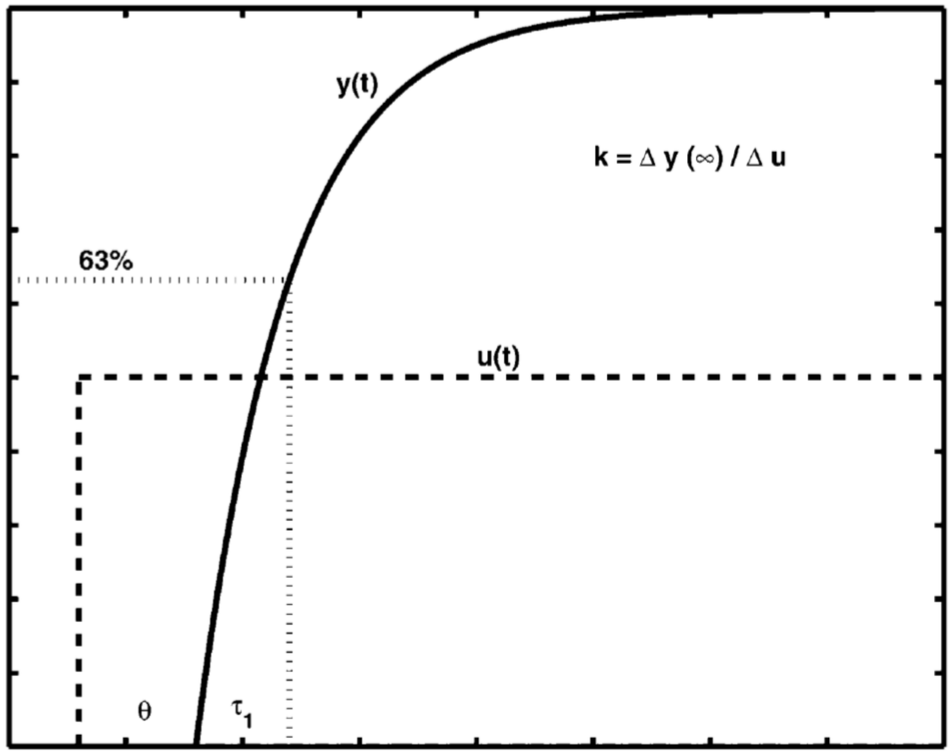


Figure 2: Open-loop step response of a first-order process with a time delay θ and time constant τ_1. u is the controller output and y is the controlled variable. Figure retrieved from [15].

From this step response, the first order transfer function model becomes

$$g(s) = \frac{ke^{-\theta s}}{\tau_1 s + 1} \tag{2}$$

where k is the steady-state process gain found by equation 3

$$k = \frac{\Delta y(\infty)}{\Delta u} \tag{3}$$

The SIMC-tuning rules then says that the tuning parameters for the PID-controller should be

$$K_c = \frac{1}{k}\frac{\tau_1}{\tau_c + \theta} \tag{4}$$

$$\tau_I = \min(\tau_1, 4(\tau_c + \theta)) \tag{5}$$

$$\tau_D = \tau_2 = 0 \tag{6}$$

which gives us a PI-controller (no derivative action for first-order processes) with tuning parameter τ_c.

The gain for integrating processes (Fig. 2) or the steady-state gain for first-order processes that takes a very long time ($\tau_1 > 8\theta$) to settle at a new steady-state can be found from equation 7

$$k' = \frac{\Delta y}{\Delta t \Delta u} \tag{7}$$

The model of the process then becomes an integrating model with the transfer function

$$g(s) = \frac{k' e^{-\theta s}}{s} \tag{8}$$

and tuning parameters [16]

$$K_c = \frac{1}{k'} \frac{1}{(\tau_c + \theta)} \tag{9}$$

$$\tau_I = 4(\tau_c + \theta) \tag{10}$$

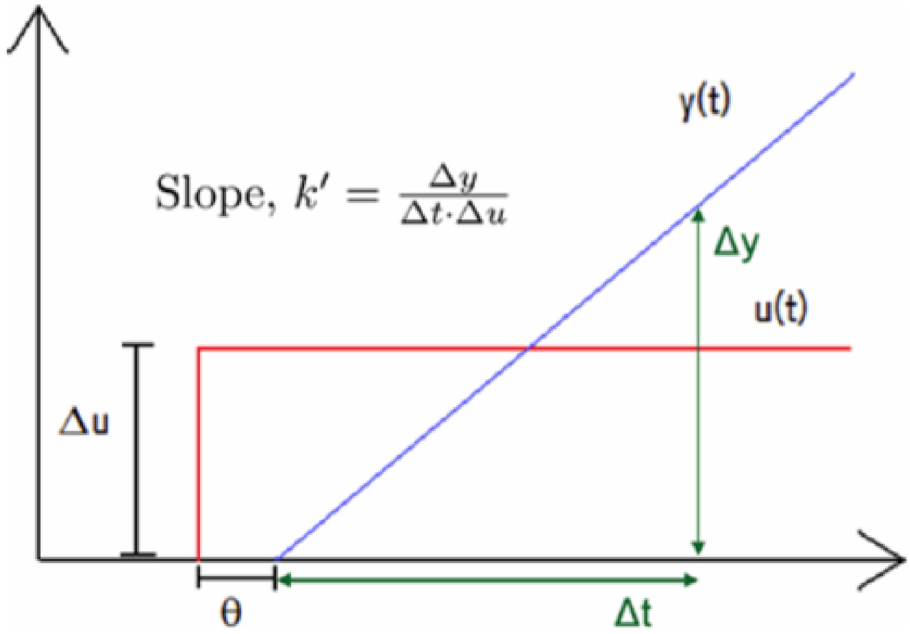


Figure 3: Open-loop step response for an integrating process with gain k', y is the CV, u is the MV and θ is the time delay. Figure retrieved from [16].

2.2 Control hierarchy

The control structure for a complete chemical plant may be separated by time scale into several layers, with the use of single-loop PID controllers at the bottom layer. Figure 4 show this hierarchical structure. The layers consist of scheduling (weeks), site wide optimization (daily), local optimization (hourly), supervisory control (minutes) and regulatory control (seconds). Each layer receives their setpoints by the layer above and implements them [14]. The scheduling, site wide optimization and local optimization is conducted based on an economic objective, while the supervisory and regulatory control layers are there to "stabilize" the plant [6].

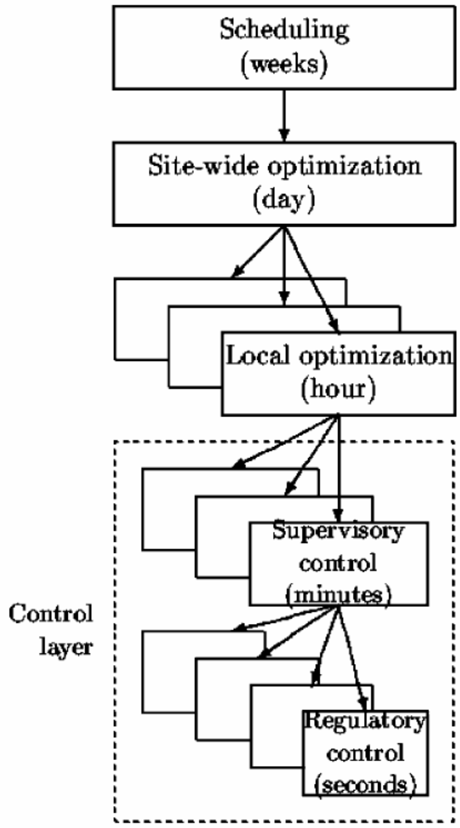


Figure 4: Control hierarchy in a chemical plant, separated into layers by different time scales. Figure retrieved from [14].

A procedure to design the overall control structure is given in [6] and consists of eight steps:

1. Defining the operational objective: Define operational constraints and a cost function J to minimize.

2. Manipulated variables and degrees of freedom: Find steady-state and dynamic degrees of freedom.

3. Primary controlled variables: Control active constraints. Control the remaining degrees of freedom at a setpoint which gives small economic loss when disturbances arise. An active constraint to control can be, for example, a concentration specification which are at a limit (found by minimizing J for different disturbances).

4. Production rate: Decide where the TPM should be placed. This choice determines how the inventory control will be arranged.

5. Regulatory control layer: Use of single-loop PID controllers to avoid that the plant drifts far way from the nominal operating point (stabilize the plant).

6. Supervisory control layer: Keep primary controlled variables at the best setpoints by manipulating the setpoints to the regulatory level and unused MVs.

7. Optimization layer: Identify active constraints and find the best setpoints for the supervisory control layer.

8. Validation: Simulate for different disturbances of critical parts of the plant.

The supervisory control layer can be designed using either single-loop PID-controllers, called decentralized control, or a multivariable (centralized) controller, for example model predictive control (MPC) [14].

2.3 Decentralized control

2.3.1 Split range control

When a controller uses more than one manipulated variable to control a controlled variable, it is called split range control (SRC). This is often implemented such that the controller only uses one MV at a time, while the others remain at their max or min values (saturated). A split range controller is depicted in Figure 6, where one CV (y) is controlled by the two MVs u_1 and u_2 [11].

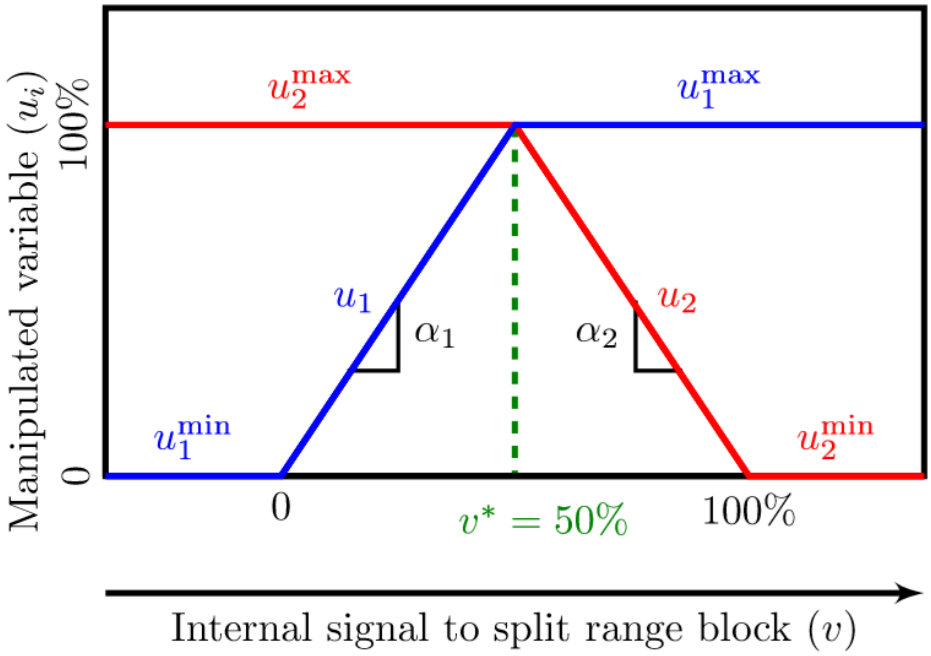


Figure 5: Split range block, with the split value v* at the midpoint (50%). Figure retrieved from [11].

The controller C in Figure 6 is usually a PI controller, and is only able to output one signal v. Since several MVs are used, we need a split range (SR) block that can give, in this case, two signals from v. An example of how a SR block can look is presented in Figure 5, where the α-values tells us something about how much effect the internal variable v has on each MV. The split value v* should be used as a tuning parameter to account for the dynamic behavior of each MV [11].

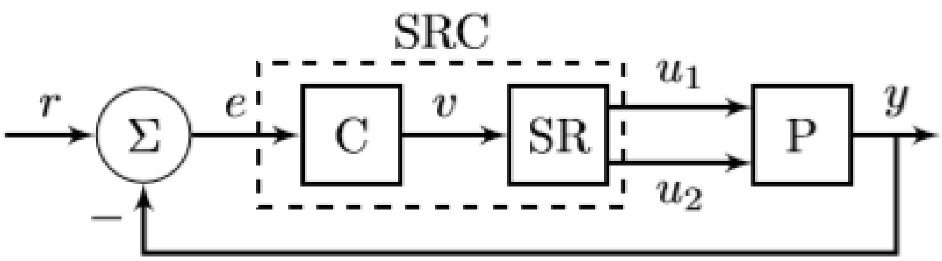


Figure 6: Split range controller. The CV y is controlled with the two MVs u_1 and u_2. Figure retrieved from [11].

A systematic procedure is given in [11] to design the SR blocks, and consists of the steps:

1. Decide the range of v
2. Define the limits of each MV (physical limitations)
3. Get the independent tuning parameters for each MV (as if they would have their own controller)
4. For a PI controller, the integral time for the SRC should be large for slow (integrating) processes and small for fast processes (compromise for the integral times found in step 3).
5. Choose in which order the MVs are used, based on economic reasoning.
6. Use equation 11 and 12 to find the α-values for a fast process, or equation 11 and 13 for a slow process. The proportional- and integral-gains $K_{C,i}$ and $K_{I,i}$ are found from the tunings in step 3, while K_C and K_I are for the SRC.
7. Lastly, use equation 14 to find the split values v* between all MVs.

$$v^{max} - v^{min} = \sum_{i=1}^{N} \frac{u_i^{max} - u_i^{min}}{|\alpha_i|} \tag{11}$$

$$K_{l,i} = \alpha_i K_l \tag{12}$$

$$K_{C,i} = \alpha_i K_C \tag{13}$$

$$\Delta v_i = v_i^* - v_{i-1}^* = \frac{u_i^{max} - u_i^{min}}{|\alpha_i|} \tag{14}$$

2.3.2 Generalized split range control

When several MVs are available for use to control a single CV, each MV may have very different dynamic effect on the CV. In the procedure to design the split range block in section 2.3.1, a compromise must be made for the chosen (common) integral time for all MVs. A way to overcome this limitation is given in [9] and is called "Generalized split range control".

In this control structure each MV has its own controller which can be tuned independently, but only one controller is active at a given time. The MVs that are not active are kept at a fixed value, e.g. minimum or maximum. This is handled by a "baton strategy logic", where it is compared to runners in a relay race [9].

This ensures that only one MV is used at a given time and the baton is passed among the controllers only when the active MV reaches its minimum or maximum value (saturates).

Before designing the logic, we must decide on the limits of each MV and the sequence in which they should be used, e.g. based on economic or operational aspects. When this is done, and with the active MV i the logic will then be:

1. Controller C_i computes MV_i'
2. If $MV_i^{min} < MV_i' < MV_i^{max}$, MV_i remains active with $MV_i = MV_i'$ and the other inactive MVs are fixed.
3. If $MV_i' \leq MV_i^{min}$ or $MV_i' \geq MV_i^{max}$, pass the baton to C_j and fix MV_i to the max or min value (depending on what limit caused the baton to be given away). Set $i = j$ and repeat from step 1.

2.3.3 Valve position control

Valve position control (VPC), also known as input resetting or midranging control [1], can be used when we have two MVs to control a single CV, where one MV has a much slower dynamic response than the other. Both MVs affects the CV, but one controls directly the CV and one brings the other MV to its desired setpoint. This control structure can be seen in Figure 7, where the MV u_2 controls the CV y and the MV u_1 controls u_2 to the desired setpoint u_2^{SP} [10].

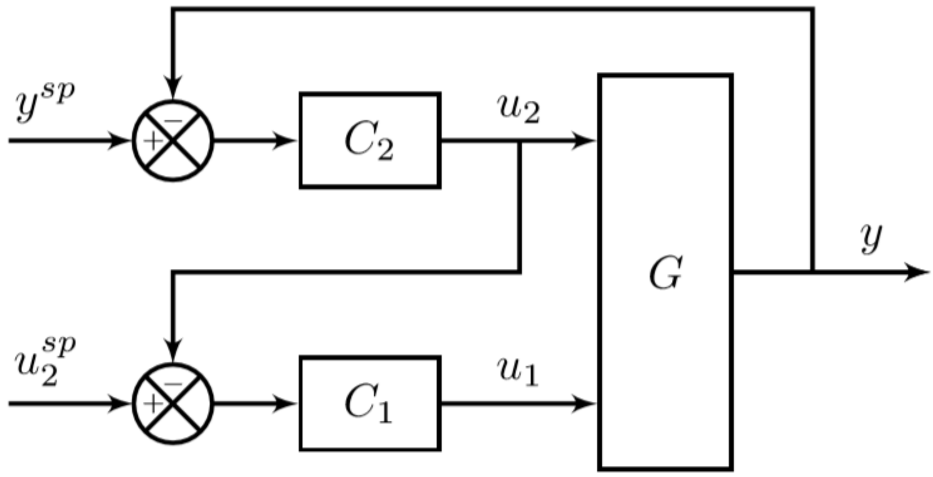


Figure 7: Valve position control structure where two MVs are used to control one CV. The MV u_2 controls the CV y, while the MV u_1 is used to control u_2 back to a desired setpoint. Figure retrieved from [9].

2.3.4 Controllers with different setpoints

In addition to SRC and VPC, controllers with different setpoints can also be used when several MVs are available for control of one CV. Each MV has their own controller with independent tuning. Figure 8 shows a block diagram of this structure with two MVs with a difference in the setpoints Δy^{SP}. This should be chosen large enough to let only one controller be active at a given point in time [9].

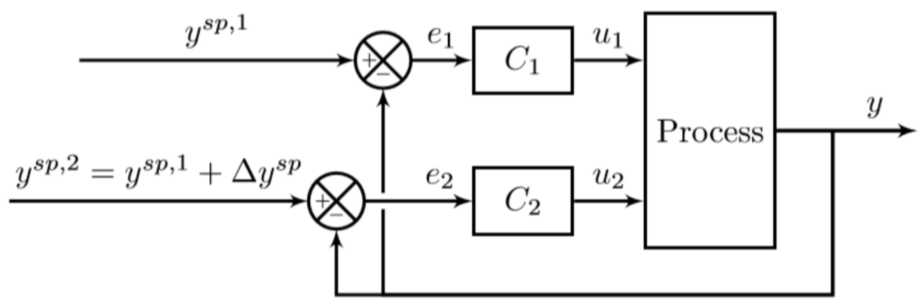


Figure 8: Controllers for different MVs with different setpoints for the same CV. A difference ΔY^{sp} is needed to let only one controller be active at a given point in time. Figure retrieved from [9].

2.3.5 Selectors

In the previous sections 2.3.1 - 2.3.4 the case has been that we have more MVs than CVs. If we have more CVs than MVs, we can use selectors that chooses which CV(s) to control. There is three different kinds of selectors: max-, min- or mid-selectors, which outputs the maximum, minimum or middle value of their inputs, respectively [9].

Figure 9 shows the case where we have two CVs y_1 and y_2 that can be controlled by the MV u. Each CV has their own controller (C_1 and C_2) that computes the outputs u_1 and u_2, which is sent to the min- or max-selector. This can only be done if control of one CV can be given up or one CV is only constrained by a limit [9].

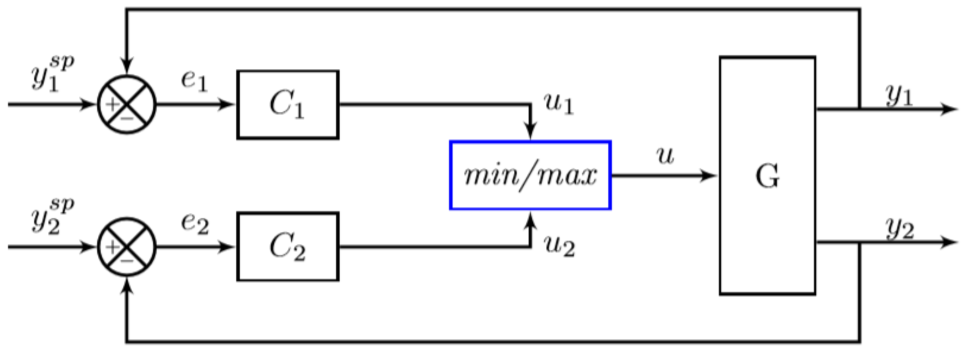


Figure 9: Min- or max-selector. Figure retrieved from [9].

Figure 10 shows the use of a mid-selector with one MV and one CV. The CV has a lower and upper limit, which is the setpoint in the controllers that computes a lower and upper value for the MV. We also have a desired value (u^{sp}) for the MV. The mid-selector then selects the middle value of the controllers' output and the desired value [9].

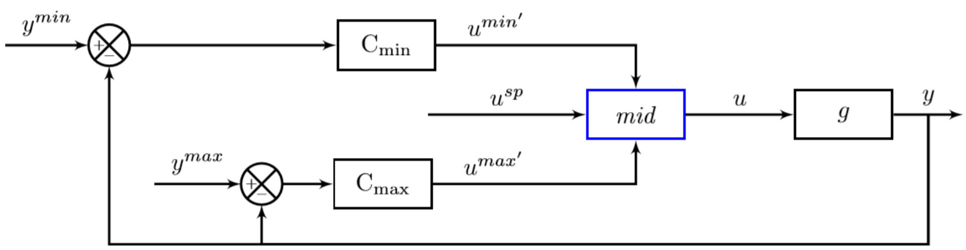


Figure 10: Mid-selector. Figure retrieved from [9].

2.3.6 Anti-windup

As long as there exists an error from the desired setpoint, a controller with integral action will continue to increase or decrease its output until the error reaches zero, as mentioned in section 2.1.1. If the error is sustained for a prolonged period of time, the integral-term will become quite large and surpass the physical limitations of the process equipment, e.g. a controller want to open a valve 150 %, which is of course not possible. This phenomenon is known as "windup" and is very common in the process industry [5].

There are several ways to avoid windup (implementing anti-windup), but only one option will be described here; namely the use of back-calculation [5].

The idea with back-calculation is that when the MV saturates (reaches a physical limitation) the integral-term in the controller is recomputed so it continues to give an output at the limit, and not crossing it. A controller where back-calculation is implemented is given in Figure 11. Here MV is the manipulated variable with inherent limitations, u_{real} is the real input to the plant and $u_{controller}$ is the value of the MV which the controller want with no physical limitations. At steady-state the error between u_{real} and $u_{Controller}$ (e_s) is zero, but when the MV saturates e_s in non-zero and is multiplied with $\frac{1}{\tau_t}$ and added to the integral-term. This resets the integrator dynamically [5].

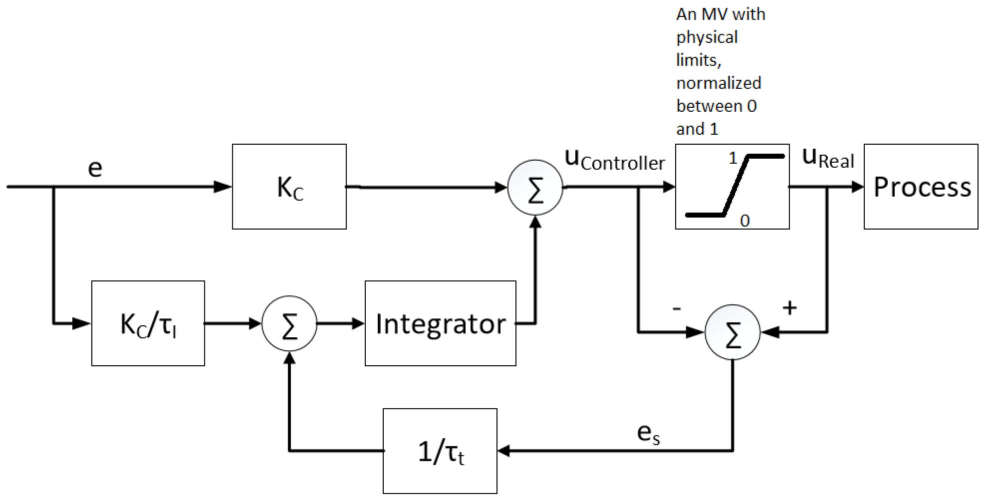


Figure 11: PI controller with back calculation. An MV has physical limits, but a controller does not. When there is a difference between the controller output and the value of the MV ($e_s \neq 0$) this value is multiplied with $1/\tau_t$ and added to the integrator, resetting the controller dynamically.

2.4 Centralized control

2.4.1 Model predictive control

The previously mentioned decentralized control structures can also be called feedback control, since a signal (measurement of the CV) is ”fed back” to the controller generating the error-signal.

Model predictive control (MPC) is the idea of combining feedback control and dynamic optimization. By dynamic optimization, it is meant that a discrete optimization problem is formulated from a model of the process we want to control for the time horizon t = 0 to t = N, and the optimized values of the MVs are implemented for all time steps. Another name for this is open loop optimization, sine the value for the MVs are purely based on a model (no feedback signal) [7].

This might not be a satisfactory approach to control the CVs, because the model does not take into account disturbances. This can however be solved by using closed loop optimization, where the optimization problem is solved at every time step t and the initial conditions are the measured values of the CVs. If the CVs can not be measured they can be estimated instead based on process data, but that will not be discussed further here. The solution gives a sequence of the optimal values of the MVs, but only the first value is implemented in the process from t to t+1 [7].

Figure 12 shows this closed loop optimization (MPC), where x is the state (CV) and u the MV. The bottom plot shows the historical data of the process for the CV and MV, while the top plot shows the predicted values of the CV and optimized MV values from the MPCs solution [7].

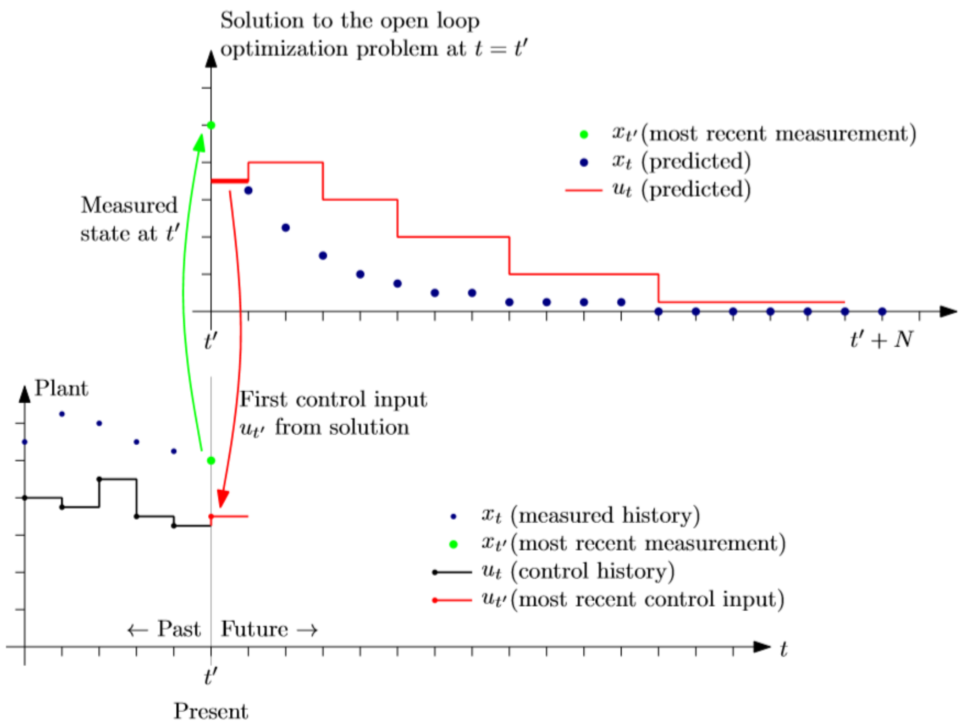


Figure 12: Principle of how an MPC works. x denotes the CV, while u denotes the MV. Figure retrieved from [7].

The objective function, also called a cost function, which the MPC will optimize can be given as

$$\min \sum_{t=0}^{N-1} (x_t - x_t^{sp})^T Q (x_t - x_t^{sp}) + \Delta u_t^T R \Delta u_t \tag{15}$$

subject to

$$x_{t+1} = g(x_t, u_t) \tag{16a}$$

$$x_0, u_{-1} = \text{given} \tag{16b}$$

$$x^{low} \leq x_t \leq x^{high} \tag{16c}$$

$$u^{low} \leq u_t \leq u^{high} \tag{16d}$$

$$-\Delta u^{high} \leq \Delta u_t \leq \Delta u^{high} \tag{16e}$$

where x are the CV(s), u the MV(s), g is the model and Q and R are the tuning parameters (usually diagonal matrices, which results in a quadratic cost function) [7]. The MPC inherently handles the operational and physical constraints, since they are given in equations 16c and 16d .

2.5 Process description

The process studied in this project is a system of three tanks in series presented in Figure 13, with valves before and after each tank. The mass flow through the system starts at the inlet of tank 1 and exits at the outlet of tank 3. The four valves, also called manipulated variables (MVs), can here be used for inventory control. We must also have a TPM, which will be set (manually by personnel) by using one of the valves. This gives us a multivariable system with four controlled variables (CVs, three levels plus a TPM) and four MVs.

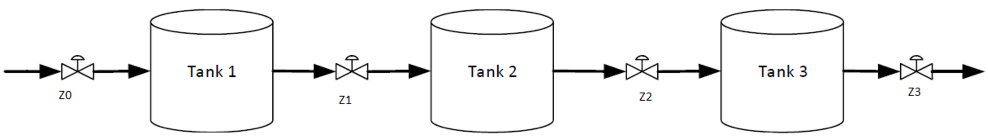


Figure 13: The process studied in this project.

Chapter 3

Modeling

A description of how the system was modeled and the assumptions are given here.

3.1 Assumptions

The system considered in this project consists of three cylindrical tanks in series with liquid inventories that needs to be controlled. The assumptions are:

- The liquid density is constant
- The tanks have a constant bottom area equal 5 m^2
- The measurement of the inventories (levels) are perfect (no error)

3.2 Modeling the system

Starting with the mass balance for tanks i=1,2,3 with the mass flow q having units kg/min

$$\frac{dm_i}{dt} = q_{in} - q_{out} \tag{17}$$

with the mass in tank i given as

$$m_i = V_i\rho = h_i A_i \rho \tag{18}$$

where h is the liquid height i m, A the bottom area in m^2 and ρ the liquid density in kg/m^3. Inserting equation 18 into 17 and rearranging we get

$$\frac{dh_i}{dt} = \frac{1}{A_i}(F_{in} - F_{out}) \tag{19}$$

with the liquid flow F in m^3/min

$$F = \frac{q}{\rho} \tag{20}$$

Since the variable that we have available for control is a valve and not a flow, we need a relationship that gives the flow F as a function of the valve position z. For this we can use a linear relationship for the flows j=0,1,2,3

$$F_j = Cv_j z_j \tag{21}$$

where Cv_j is the valve constant.

The nominal values for the MVs and CVs are given in Table 1. In this system we would like to have production set at the inlet of the process, meaning z_0 is the TPM and thus set manually to a certain position, with the remaining valves used for inventory control.

Table 1: Nominal operating conditions

Variable	Value	Unit
F_j^*	0.16	m^3/min
z_0^*	0.4	-
z_1^*	0.5	-
z_2^*	0.7	-
z_3^*	0.6	-
h_1^*	1.0	m
h_2^*	1.2	m
h_3^*	0.7	m

By using the values in Table 1 and equation 21 we can calculate the valve constants given in Table 2.

Table 2: Valve constants

Variable	Value	Unit
Cv_0	0.4000	m^3/min
Cv_1	0.3200	m^3/min
Cv_2	0.2286	m^3/min
Cv_3	0.2667	m^3/min

To have a more realistic case, some additional flows (disturbance variables, DVs) will be introduced to the model. The reasoning behind this is because in a real plant a given tank or inventory may be shared by multiple "sub-plants" or different groups of plant operators and thus flows may be "hidden" or unmeasured. This new case is presented in Figure 14 with additional inflows and outflows of the tanks. Note that this new addition came into the project at a later stage, so the model with MPC as the control structure was not updated for this case. This was collectively decided together with my supervisor.

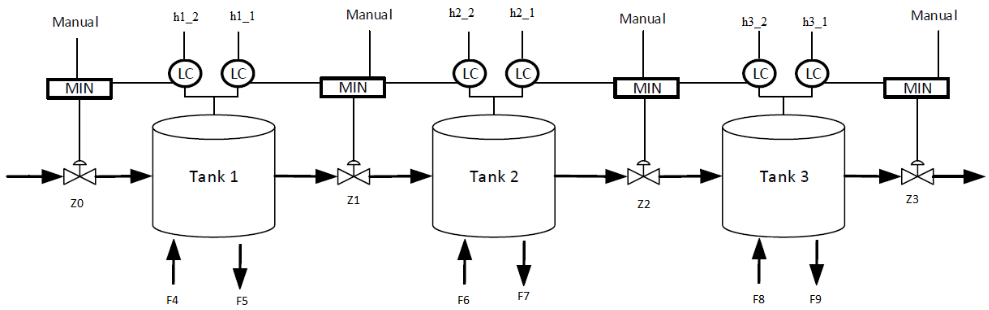


Figure 14: Control structure with additional disturbances (inflows and outflows).

Chapter 4

Implementation

A description of how the modeled process and the control structures was implemented is given here. A bit more practical information about the simulation files are given in Appendix A.

Matlab and Simulink (ver. 2018a) was used to simulate the modeled process with SRC, generalized SRC and controllers with different setpoints. This was done by creating a Matlab function with equation 19 and 21, which gives us the differentials $\frac{dh_i}{dt}$ as a function of the valve positions z_j. This function was then brought into Simulink by use of the Interpreted MATLAB Function block [8]. The solver used was *ode15s* with a relative tolerance of 10^{-5}.

For the control structure with MPC Matlab and CasADi [2] was used. The optimization problem was formualted by use of the CasADi-syntax in Matlab and solved using the solver IPOPT.

We also need to consider some operational constraints or limits, since all physical tanks has a limited volume. These limits are given in Table 3. Valves have limits being fully closed or fully open, represented by the values 0 and 1, respectively.

Table 3: Operational limits for the process

	Lower limit	Upper limit	Unit
Tank 1	0	2.0	m
Tank 2	0	2.4	m
Tank 3	0	1.4	m

The model is based on a linear relationship between the valve position z and the flow F, as described earlier. By doing an open-loop step test on any of the valves, we then observe that the level in the tanks can be represented by an integrating process on the form of equation 8. These are shown in Table 4. Since there is no time delay, the term $e^{-\theta s}$ have a value of 1.

Table 4: Transfer functions in the Laplace-domain for the different valves and tanks.

	z_0	z_1	z_2	z_3	Unit
Tank 1	$\frac{0.0801}{s}$	$\frac{-0.0640}{s}$			$\frac{m}{min}$
Tank 2		$\frac{0.0640}{s}$	$\frac{-0.0458}{s}$		$\frac{m}{min}$
Tank 3			$\frac{0.0458}{s}$	$\frac{-0.0533}{s}$	$\frac{m}{min}$

4.1 Split range control

Figure 15 shows the structure with split range control applied, where the inlet and outlet valve are the manipulated variables for a given tank. Here we need to use min-selectors, because two controllers can use the same valve and it should be possible to set the TPM manually by use of any of the four valves. Since we only have one single controller for a given tank, we must have setpoints so that the system can handle disturbances ”in both directions”, meaning disturbances that would either drain a tank or fill a tank. The setpoints are then chosen to be at 50% of the tanks maximum allowed volume from Table 3, given in Table 5.

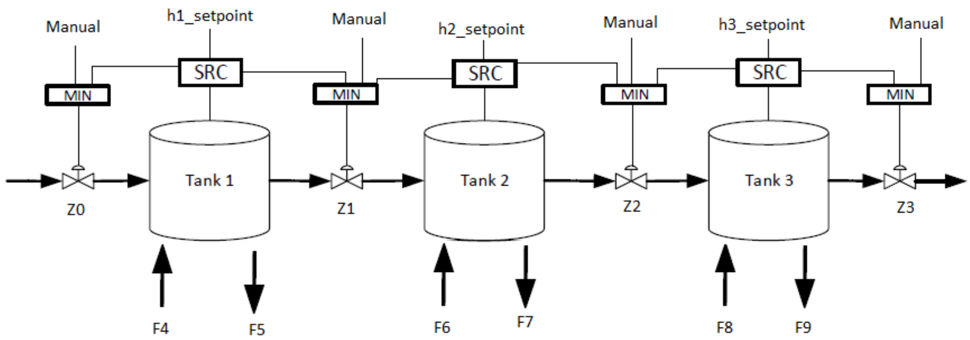


Figure 15: Control structure for the modeled process with split range control.

Table 5: Setpoints for the level controllers in the control structure with split range control.

Setpoint	Value	Unit
h_1	1.0	m
h_2	1.2	m
h_3	0.7	m

The control objective here is to maintain production during a temporary bottleneck, meaning that we are not interested in tight control of the levels. For this reason we want to have a large value for τ_C. In the design of a SRC given in section 2.3.1, we need to define independent tuning parameters for each MV (as if they would have their own controller) and then use these to design the common split range controller. For the independent tunings, we then choose to have $\tau_c = 30$ minutes, and with the transfer functions given in Table 4 the resulting SIMC tuning parameters are given in Table 6.

Table 6: Ideal tuning parameters for the control structure with split range control used in the design of the common controller for all manipulated variables.

	Valve	K_C	Unit	τ_I	Unit
Tank 1	z_0	0.4161	m^{-1}	120	min
	z_1	-0.5208	m^{-1}	120	min
Tank 2	z_1	0.5208	m^{-1}	120	min
	z_2	-0.7282	m^{-1}	120	min
Tank 3	z_2	0.7282	m^{-1}	120	min
	z_3	-0.6254	m^{-1}	120	min

To design the split range controllers, we follow the seven-step procedure in section 2.3.1:

1. Range of v: Between 0 and 100.
2. Limits of each MV: These are physically valves, so the limits are fully closed or fully open, here represented by the values 0 and 1, respectively.
3. Independent tuning for each MV: Given in Table 6.
4. Integral time: The integral times are equal for every independent controller, so it is set to 120 minutes for all split range controllers.
5. Order of MV usage: The valve z_0 is nominally the TPM, meaning the outlet valves of each tank should be used first followed by the inlet valve so the control structure remains consistent.
6. The processes we are dealing with are all integrating processes, so equation 11 and 13 are used to find the α-values in the split-range blocks.
7. Equation 14 are used to find the split value in the blocks. Note that here for any of the three SRC we only have two MVs, so there is only one split value in each block.

The values for the three resulting split-range blocks are given in Table 7.

Table 7: Controller tunings and split-range block parameters for the three split range controllers.

	Gain	Integral time	Split value	Valve	α-value
Tank 1	-23.13	120	44.41	z_0	-0.01799
				z_1	0.02252
Tank 2	-30.36	120	41.71	z_1	-0.01716
				z_2	0.02398
Tank 3	-33.64	120	53.78	z_2	-0.02164
				z_3	0.01859

4.1.1 Update on internal variable v

By only implementing the split range controllers and selectors as mentioned above for the studied system, we will observe that when a bottleneck occurs the level in a tank will become uncontrolled.

As an example, consider what happens if a bottleneck were to occur at the outlet of tank 1 (z_1) when the system is initially at the nominal point (see Table 1). If an operator gives a manual input on z_1 of 0.3, the valve position will immediately go from 0.5 (from controller of tank 1) down to 0.3. This will make the level in tank 1 increase and the controller would like to increase the opening of z_1, but nothing happens because of the min-selector. The controller then needs to "unwind" or increase its output v until a value of 0.4 is reached for the second MV, namely z_0 which is also set manually as it is the nominal TPM. This will leave the level of tank 1 uncontrolled (open-loop) during the unwinding.

A fix for this unwanted time delay for switching between the two MVs for a given tank, was to include the use of a function that could update the internal value v. This function was written as a Matlab-function script and introduced in Simulink by use of the Interpreted MATLAB function block [8]. The algorithm for this function is shown in Algorithm 1.

The main part of the algorithm are two IF statements, checking if an update should be done for the switch between using the outlet valve to start using the inlet valve, or the reverse. The conditions to trigger an update is:

- The error signal e_s between the physical valve position and controller output is non-zero, which means that the controller is not active.
- Check if the value of v is above or below the split value. This statement confirms if an update is necessary.
- Check if an update has happened in the last 25 minutes (25 minutes was chosen based on experience during the programming, where updates happened when they didn't need to). With slow dynamics and disturbances expected to last at least an hour, updates are expected to happen on an hourly scale.

If all conditions are met, then the internal variable v is updated by a value Update_v calculated by

$$Update\_v = \frac{(Z_{inlet} - Z_{inlet}^{max})}{\alpha_{inlet\_valve}} + v_{split} + v0 \tag{22}$$

if switching from using the outlet valve to inlet valve, or

$$Update\_v = \frac{(Z_{outlet} - Z_{outlet}^{max})}{\alpha_{outlet\_valve}} + v_{split} + v0 \tag{23}$$

if switching from manipulating the inlet valve to manipulating the outlet valve. Z_{inlet} and Z_{outlet} are the physical valve positions, Z_{outlet}^{max} and Z_{outlet}^{max} the maximum valve openings (value of 1), α_{outlet_valve} and α_{inlet_valve} the slopes of each MV in the SR block (see Table 7), v_{split} is the split value for the SR block for each individual tank and $v0$ is the un-updated v. Because of this updating of v, anti-windup was not implemented for SRC.

Algorithm 1: Updating of the internal value v for a SRC

if $e_{s,inletvalve} \neq 0$ ***AND*** $v < v_{split}$ ***AND*** $time \geq time\_before\_next\_update\_is\_allowed$ **then**

  Update_v = $\frac{(Z_{inlet} - Z_{inlet}^{max})}{\alpha_{inlet\_valve}} + v_{split} + v0$;

  time_before_next_update_is_allowed = current_time + 25 minutes ;

else if $e_{s,outletvalve} \neq 0$ ***AND*** $v > v_{split}$ ***AND*** $time \geq time\_before\_next\_update\_is\_allowed$ **then**

  Update_v = $\frac{(Z_{outlet} - Z_{outlet}^{max})}{\alpha_{outlet\_valve}} + v_{split} + v0$;

  time_before_next_update_is_allowed = current_time + 25 minutes ;

Result: Updated_v = v + Update_v ;

4.2 Generalized split range control

The generalized split range control structure for the studied process are presented in Figure 16. Independent controllers are here used for each MV, again with the inlet and outlet valves for a given tank with use of min-selectors. The structure resembles the one in section 4.3, but here the level controllers have the same setpoint. This is possible because of some additional "switching logic" mentioned in section 2.3.2, which is not depicted in Figure 16. The ideal tunings found when designing the split range blocks and the setpoints in the previous section are used here (Table 5 and Table 6).

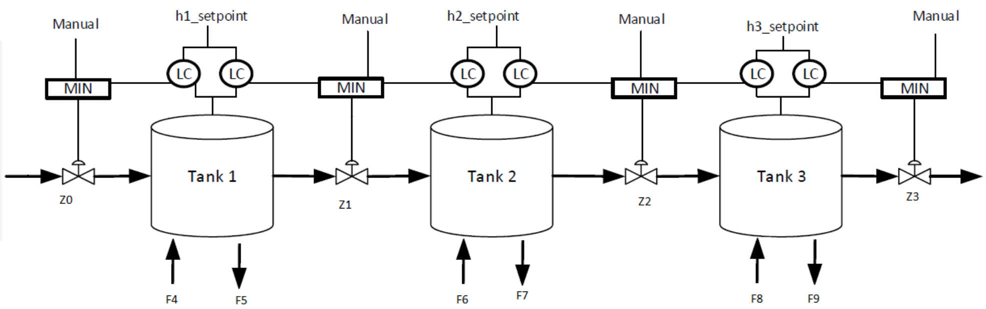


Figure 16: Control structure with generalized split range control. The two level controllers are split up with additional logic (not shown) to prevent unwanted switching, as they have the same setpoint.

The switching logic is implemented in Simulink as a Interpreted MATLAB function, with the logic given in Algorithm 2. Some additions had to be made to the logic applied here than what was described in section 2.3.2, because we here have multiple controllers able to manipulate the same valve, which was not accounted for earlier. Instead of the switching only being triggered by a MV saturating, the switching must in this case also be triggered if the active controller for a given tank looses control over a valve. To give an example; z_0 is nominally the TPM, meaning for tank 1 the controller for z_1 is active (has the baton). If production is suddenly reduced at z_1 manually, the controller for z_1 would no longer be the active input to the valve z_1 because of the min-selector. A time delay of six seconds was also implemented, to avoid very fast switching back and fourth. A given controller is here made inactive by stopping its integral-action.

Algorithm 2: Switching logic for a given tank

Initialization: Controller for outlet valve has the baton;
if *The outlet valve saturates at max opening or another controller starts to manipulate it and no switching has occurred in the last six seconds* **then**
    Give baton to the controller for inlet valve;
else if *The inlet valve saturates at max opening or another controller starts to manipulate it and no switching has occurred in the last six seconds* **then**
    Give baton to the controller for outlet valve;
else
    Do nothing;

Anti-windup was also here implemented by use of back-calculation with $\tau_t = 1$, which also results in the inactive controllers tracking the actual valve position instead of being fixed at either max or min values.

4.3 Controllers with different setpoints

The control structure with two controllers for the studied process can be seen in Figure 17. We also here need to use min-selectors for the same reason as before. The setpoints for the controllers are given in Table 8.

Table 8: Setpoints for the level controllers in the control structure with two controllers.

Setpoint	Value	Unit
$h_{1,1}$	0.3	m
$h_{1,2}$	1.7	m
$h_{2,1}$	0.3	m
$h_{2,2}$	2.0	m
$h_{3,1}$	0.3	m
$h_{3,2}$	1.1	m

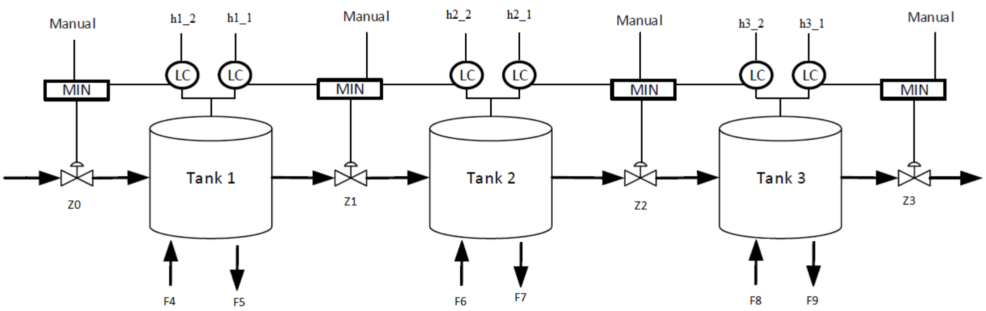


Figure 17: Control structure for the modeled process with two controllers with different setpoints.

Because of the setpoints being close to the operational limits, we need to control the levels tightly. This is achieved by having a small value for τ_c when using the SIMC tuning rules, in this case chosen to be 8 minutes for all six controllers. The resulting tuning parameters for the controllers are given in Table 9 for integrating processes using the SIMC-rules. The controllers were implemented in Simulink by use of gain blocks and transfer-function blocks. Anti-windup was handled by use of back-calculation with $\tau_t = 0.1$ for each controller.

Table 9: Tuning parameters for the control structure with two controllers.

	Valve	K_C	Unit	τ_I	Unit
Tank 1	z_0	1.56	m^{-1}	32	min
	z_1	-1.95	m^{-1}	32	min
Tank 2	z_1	1.95	m^{-1}	32	min
	z_2	-2.73	m^{-1}	32	min
Tank 3	z_2	2.73	m^{-1}	32	min
	z_3	-2.35	m^{-1}	32	min

4.4 Model predictive control

The control structure with MPC is depicted in Figure 18 for the process studied here. The cost function the MPC will minimize is given by equation 24 for i = 1,2,3 and j = 0,1,2,3

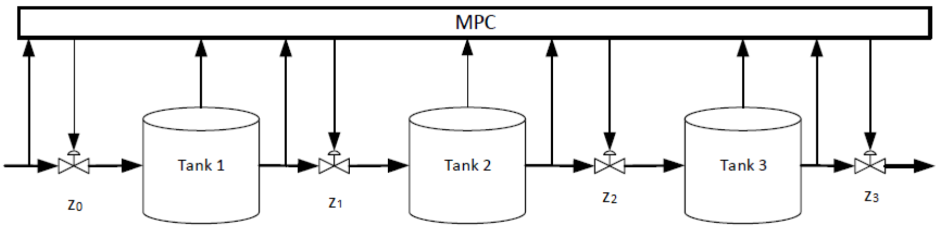


Figure 18: Control structure with MPC.

$$\min \sum_{t=0}^{N-1} Q(h_{i,t} - h_{i,t}^{sp})^2 + R\Delta z_{j,t}^2 + W(z_{TPM} - z_{TPM}^{Desired})^2 \tag{24}$$

subject to

$$\frac{dh_i}{dt} = \frac{1}{A_i}(Cv_{i-1}z_{i-1} - Cv_i z_i) \tag{25a}$$

$$h_{i,0}, z_{j,0} = \text{given} \tag{25b}$$

$$z_{TPM} = \text{given} \qquad [-] \tag{25c}$$

$$h_1^{SP} = 1 \qquad [m] \tag{25d}$$

$$h_2^{SP} = 1.2 \qquad [m] \tag{25e}$$

$$h_3^{SP} = 0.7 \qquad [m] \tag{25f}$$

$$0 \leq h_{1,t} \leq 2 \qquad [m] \tag{25g}$$

$$0 \leq h_{2,t} \leq 2.4 \qquad [m] \tag{25h}$$

$$0 \leq h_{3,t} \leq 1.4 \qquad [m] \tag{25i}$$

$$0 \leq z_{j,t} \leq 1 \qquad [-] \tag{25j}$$

$$\Delta z_{j,t+1} = z_{j,t+1} - z_{j,t} \tag{25k}$$

with tuning parameters Q = 30, R = 20 and W = 0.01. The MPC was set to predict 60 minutes into the future, with sampling time every 10 seconds resulting in N = 360. The control structure with MPC does not use selectors for the valve positions, as the other three structures do. The TPM is then here implemented in the cost function as $z_{TPM}^{Desired}$, which include information about which valve is the TPM and with what desired value. The tuning parameters Q, R and W was found by trial-and-error so the MPC does not change the valve positions much faster or slower than the control structures with SRC, because they have the same setpoints. It was especially noted that the relationship between R and W had a large impact on the use of the MVs.

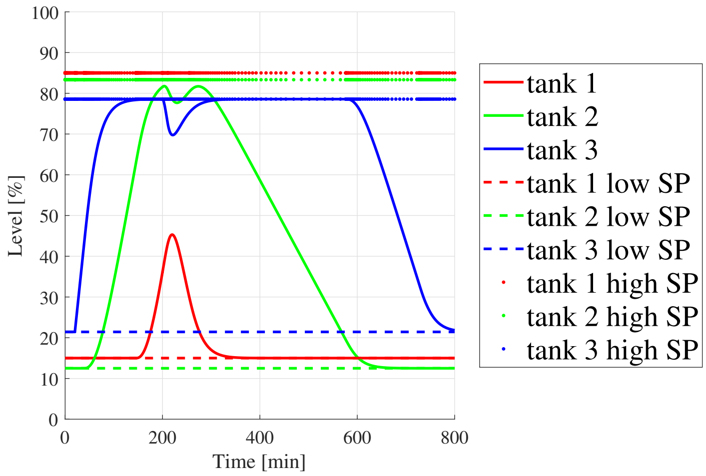
Chapter 5

Simulations

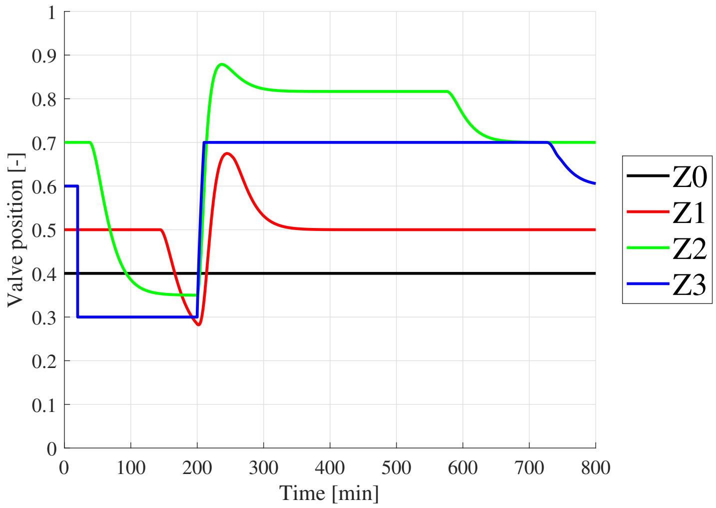
A number of simulations will be shown here for the control structures subjected to different disturbances. We would especially like to test the control structures for how good they are to "maintain production". With this it is meant that for a temporary disturbance, we would like to maintain the (nominal) flow or valve positions at every point in the process for as long as possible without violating safety or operational constraints (e.g. tank overflowing). To have the results on the same scale, the levels are normalized between 0 and 100 %, representing the lower and upper operational limit given in Table 3.

5.1 Temporary bottleneck at the outlet

A bottleneck on the outlet of the process (z_3) is conducted here for three hours, because production must be reduced by half for maintenance. The manual input for $z3$ is reduced from 1 (inactive) down to 0.3 at t = 20 minutes becoming the active input, and at t = 200 minutes raised to 0.7 where it becomes a temporary bottleneck before the system settles back to the nominal point.

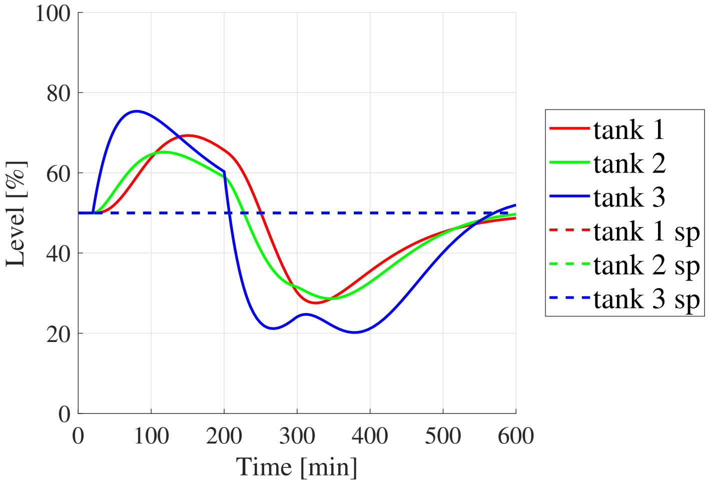


(a) Tank levels

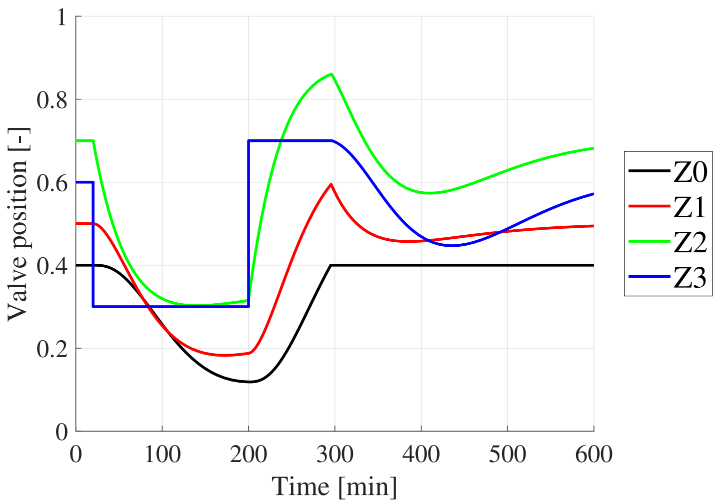


(b) Valve positions

Figure 19: Disturbance on z_3 for control structure with two controllers with different setpoints. The manual in-put for z3 is reduced from 1 (inactive) down to 0.3 at t = 20 minutes becoming the active input, and at t = 200 minutes raised to 0.7 where it becomes a temporary bottleneck before the system settles back to the nominal point.

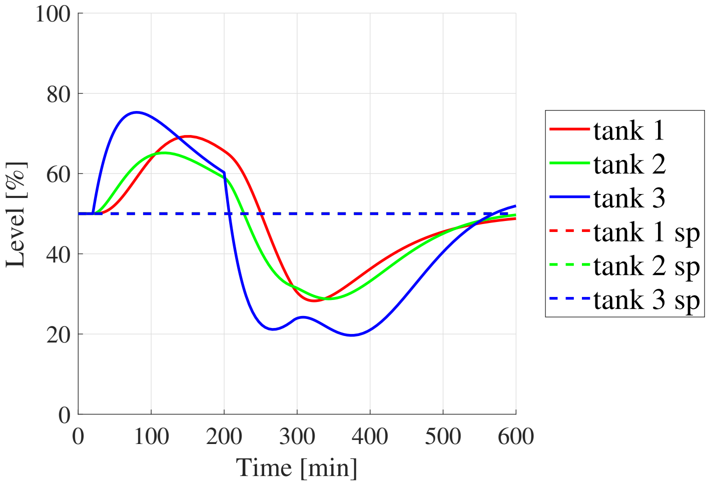


(a) Tank levels

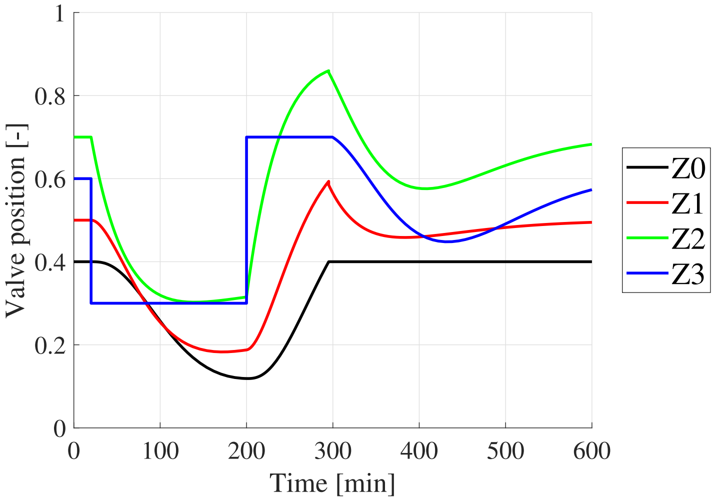


(b) Valve positions

Figure 20: Disturbance on the outlet (z_3) with split range control. The manual input for z3 is reduced from 1 (inactive) down to 0.3 at t = 20 minutes becoming the active input, and at t = 200 minutes raised to 0.7 where it becomes a temporary bottleneck before the system settles back to the nominal point.

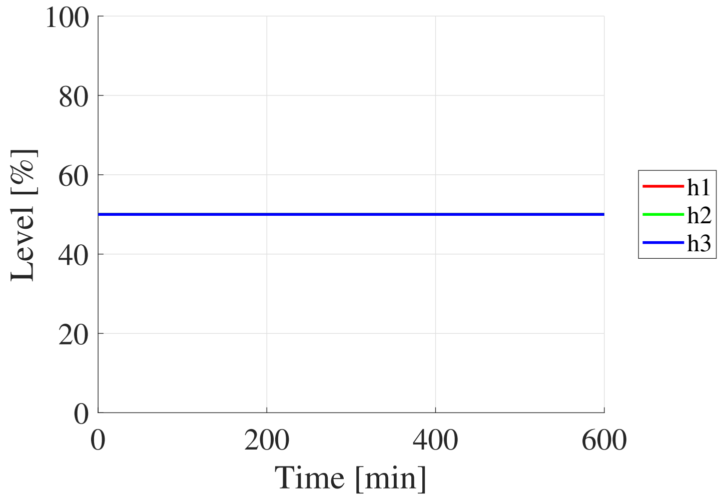


(a) Tank levels

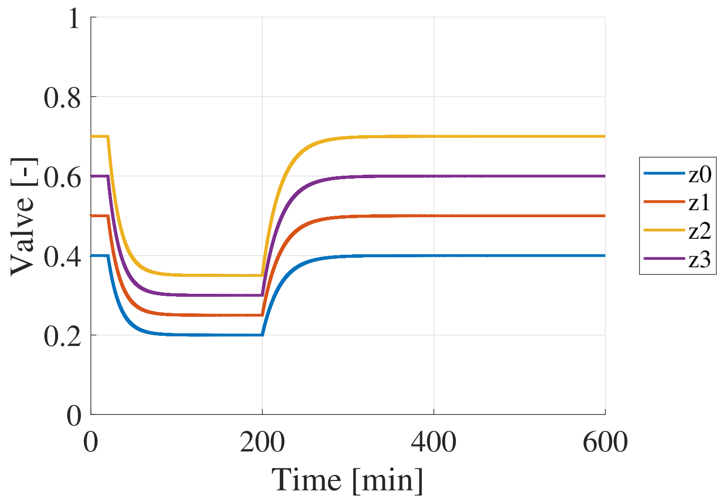


(b) Valve positions

Figure 21: Disturbance on z_3 with generalized split range control. The manual input for z3 is reduced from 1 (inactive) down to 0.3 at t = 20 minutes becoming the active input, and at t = 200 minutes raised to 0.7 where it becomes a temporary bottleneck before the system settles back to the nominal point.



(a) Tank levels

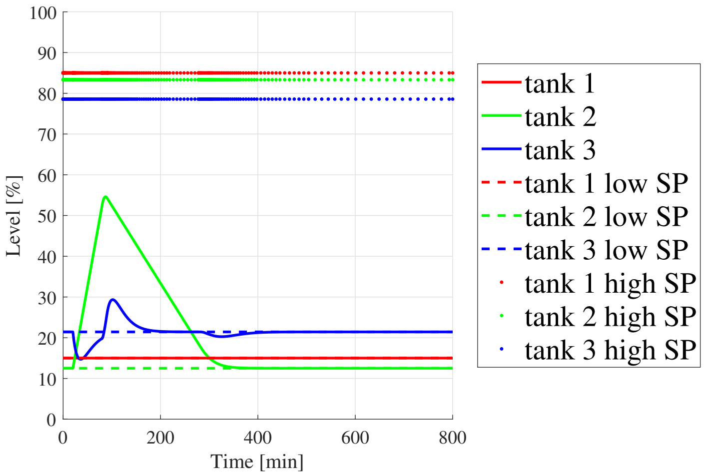


(b) Valve positions

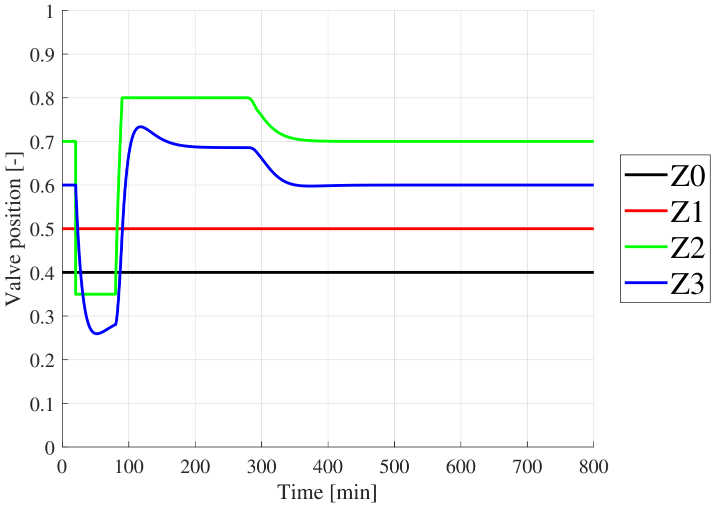
Figure 22: Disturbance on z_3 with MPC. The MPC is given the information that from t = 20 minutes z3 is the desired TPM with a desired value of 0.3, until at t = 200 where z0 is the TPM again with a desired value of 0.4.

5.2 Temporary bottleneck inside

A disturbance on z_2 is conducted here for one hour, because production must be reduced by half for maintenance. The manual input for z_2 is reduced from 1 (inactive) down to 0.35 at t = 20 minutes becoming the active input, and later raised to 0.8 where it becomes a temporary bottleneck before the system settles back to the nominal point.

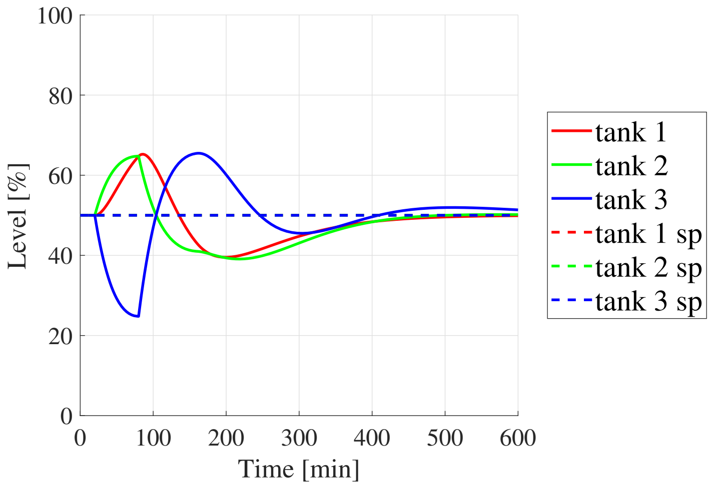


(a) Tank levels

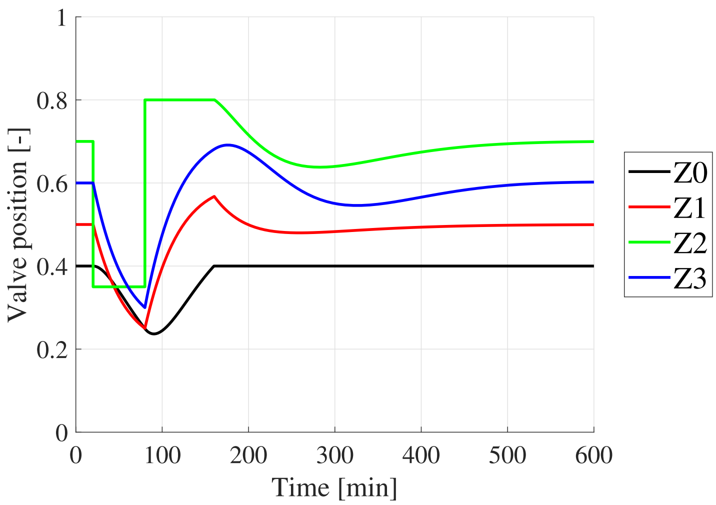


(b) Valve positions

Figure 23: Disturbance on z_2 for control structure with two controllers with different setpoints. The manual input for z_2 is reduced from 1 (inactive) down to 0.35 at t = 20 minutes becoming the active input, and later raised to 0.8 where it becomes a temporary bottleneck before the system settles back to the nominal point.

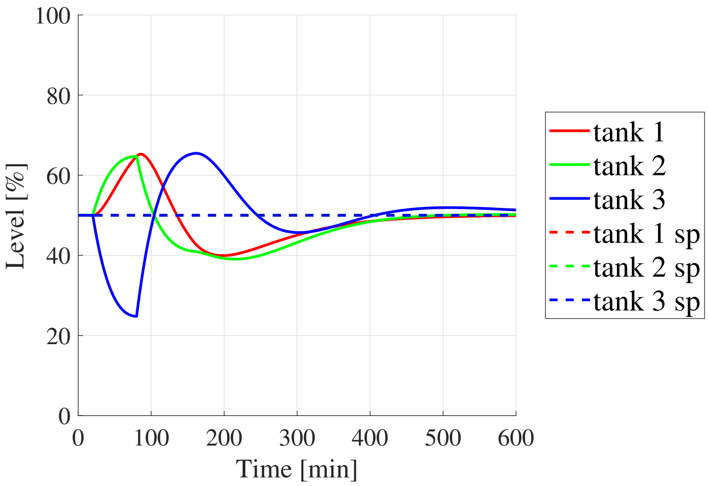


(a) Tankj levels

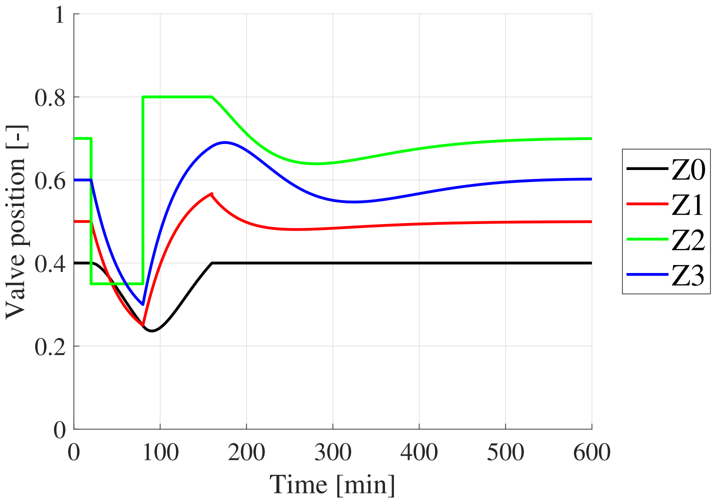


(b) Valve positions

Figure 24: Disturbance on z_2 with split range control. The manual input for z_2 is reduced from 1 (inactive) down to 0.35 at t = 20 minutes becoming the active input, and later raised to 0.8 where it becomes a temporary bottleneck before the system settles back to the nominal point.

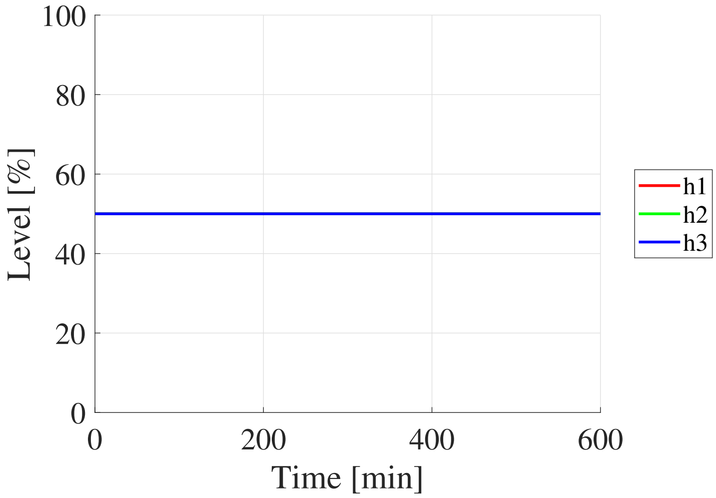


(a) Tank levels

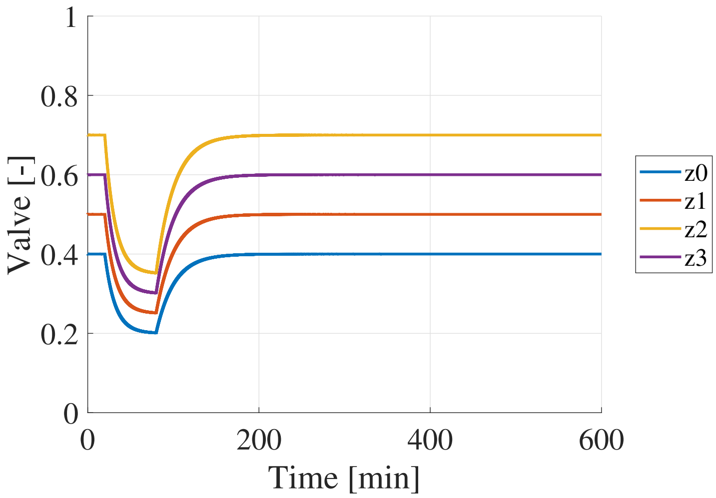


(b) Valve positions

Figure 25: Disturbance on z_2 with generalized split range control. The manual input for z_2 is reduced from 1 (inactive) down to 0.35 at t = 20 minutes becoming the active input, and later raised to 0.8 where it becomes a temporary bottleneck before the system settles back to the nominal point.



(a) Tank levels

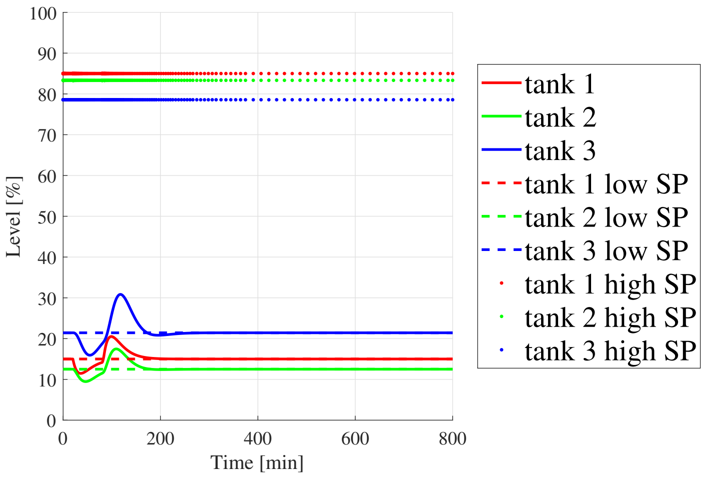


(b) Valve positions

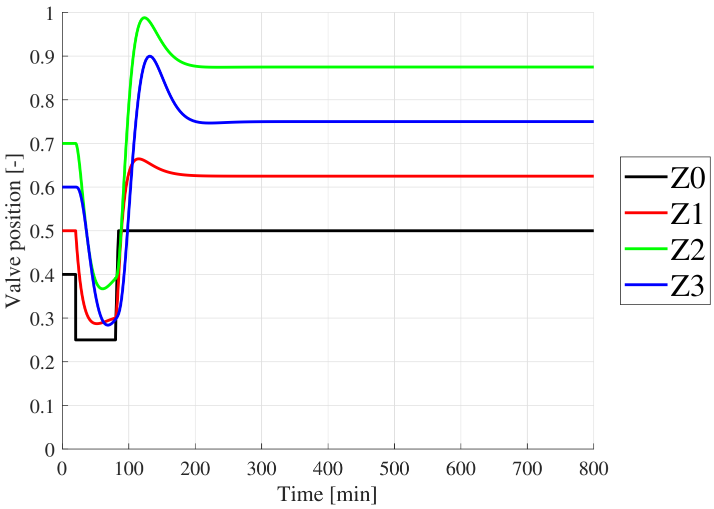
Figure 26: Disturbance on z_2 with MPC. The MPC is given the information that from t = 20 minutes z2 is the desired TPM with a desired value of 0.35, until at t = 200 where z0 is the TPM again with a desired value of 0.4.

5.3 Temporary bottleneck at the inlet

A disturbance on z_0 is conducted here for one hour, because production must be reduced by half for maintenance upstream the process. The manual input for z_0 is reduced from 1 (inactive) down to 0.25 at t = 20 minutes becoming the active input, and later raised to 0.5.

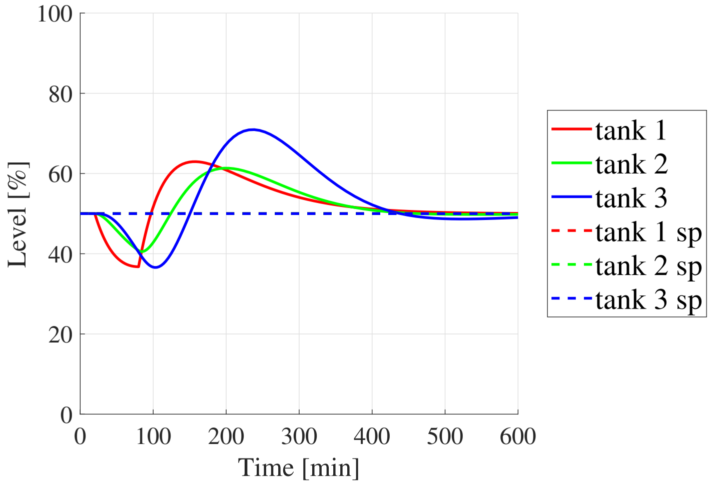


(a) Tank levels

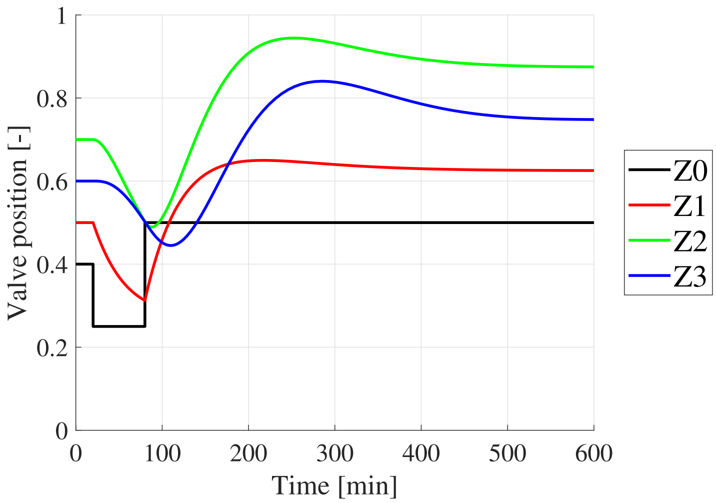


(b) Valve positions

Figure 27: Disturbance on z_0 for control structure with two controllers with different setpoints. The manual input for z_0 is reduced from 1 (inactive) down to 0.25 at t = 20 minutes becoming the active input, and later raised to 0.5

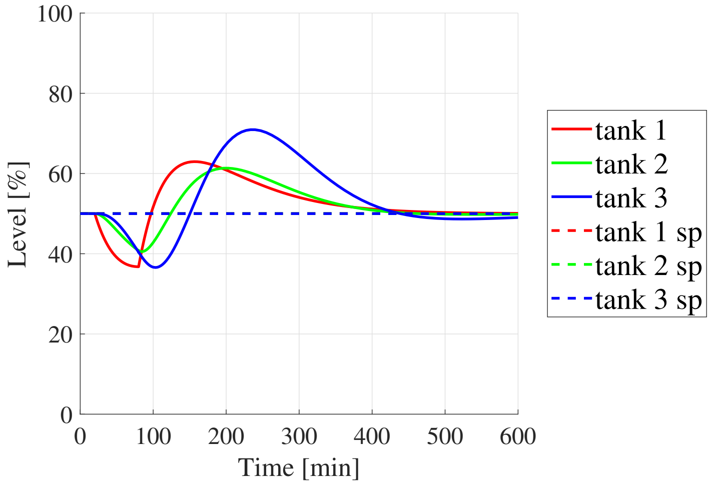


(a) Tank levels

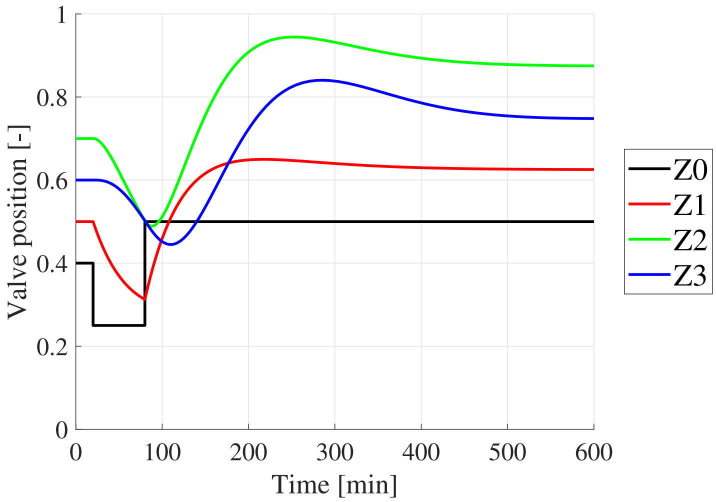


(b) Valve positions

Figure 28: Disturbance on z_0 with split range control. The manual input for z_0 is reduced from 1 (inactive) down to 0.25 at t = 20 minutes becoming the active input, and later raised to 0.5

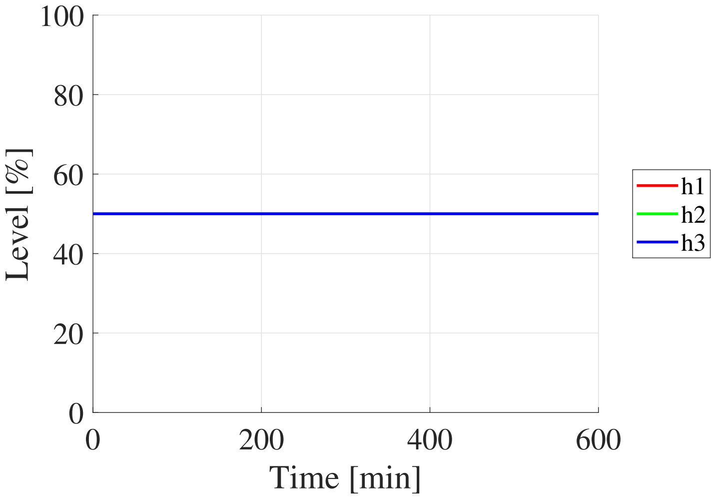


(a) Tank levels

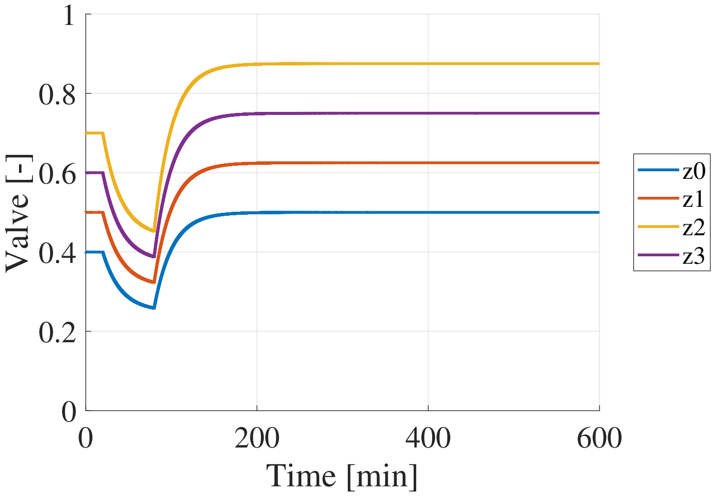


(b) Valve positions

Figure 29: Disturbance on z_0 with generalized split range control. The manual input for z_0 is reduced from 1 (inactive) down to 0.25 at t = 20 minutes becoming the active input, and later raised to 0.5



(a) Tank levels

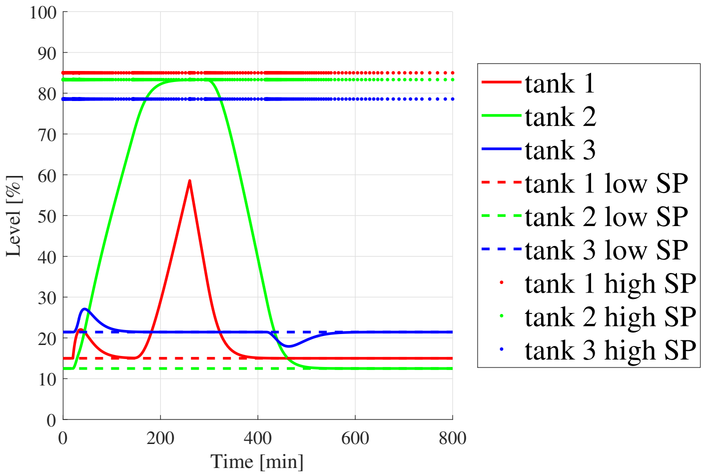


(b) Valve positions

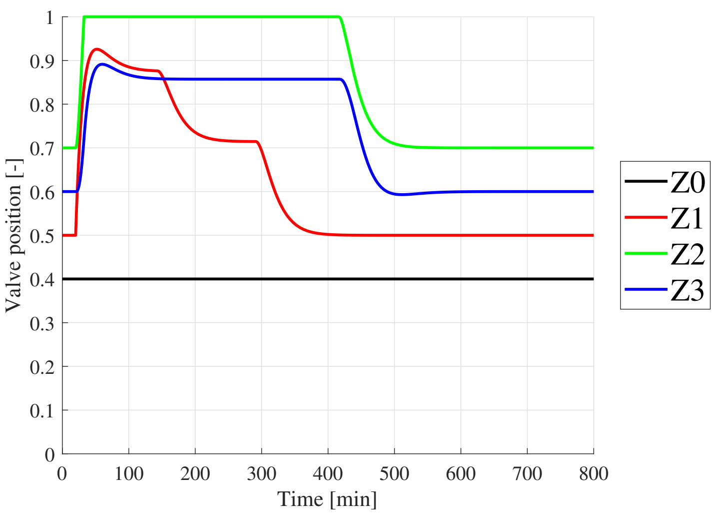
Figure 30: Disturbance on z_0 with MPC. The MPC is given the information that from t = 20 minutes the desired value of z0 is 0.25, and after t = 200 the desired value is 0.5.

5.4 Filling of tank 1

Here tank 1 is filled with an outside stream F_4 with a rate of 0.12 m^3/min from t = 20 to t = 260 minutes. As a reminder the nominal flow rate through the system is 0.16 m^3/min.

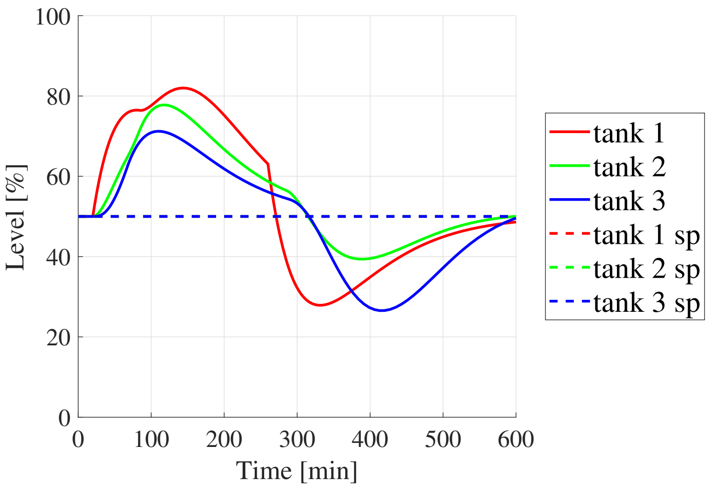


(a) Tank levels

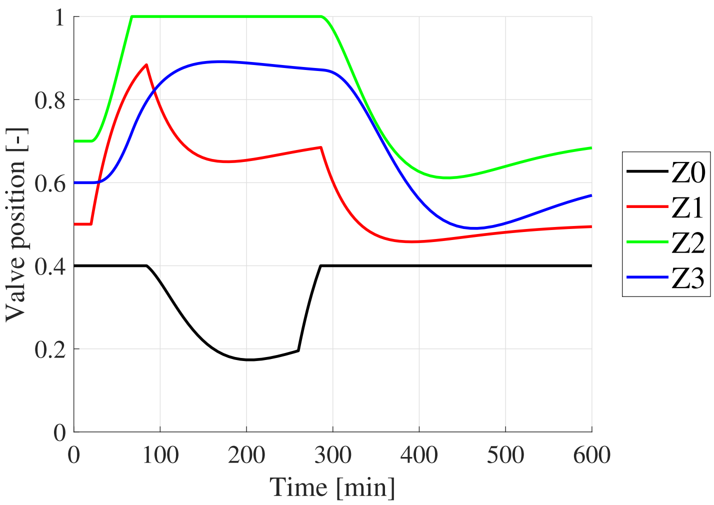


(b) Valve positions

Figure 31: External filling of tank 1 with a flow (F4) of 0.12 m^3/min from t = 20 to t = 260 minutes with two controllers for each tank.

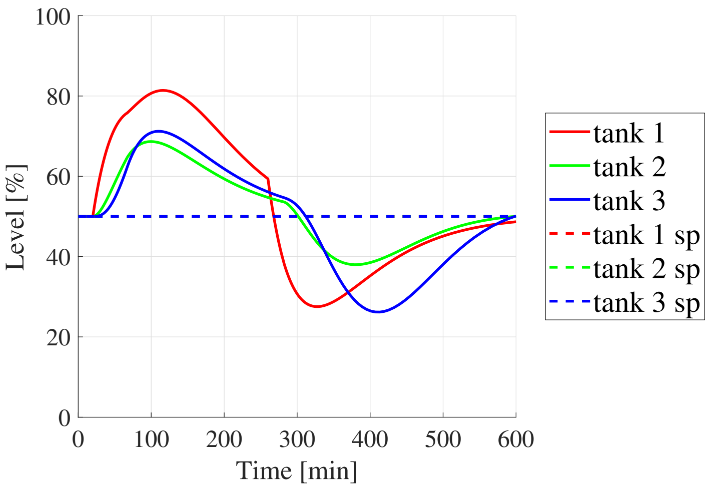


(a) Tank levels

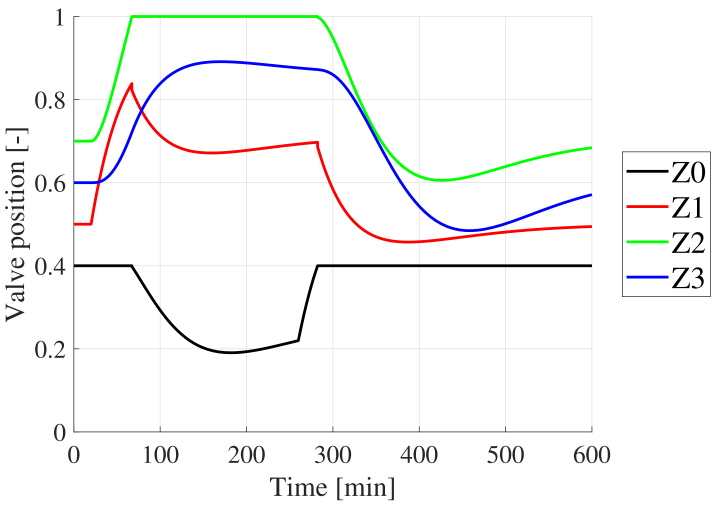


(b) Valve positions

Figure 32: External filling of tank 1 with a flow (F4) of 0.12 m^3/min from t = 20 to t = 260 minutes with split range control



(a) Tank levels

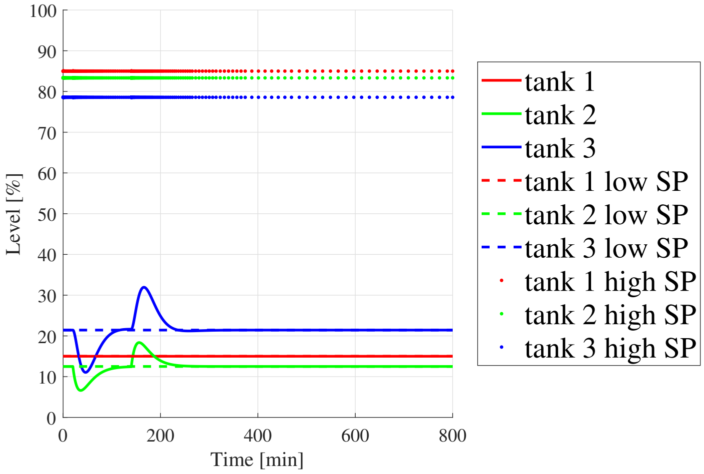


(b) valve positions

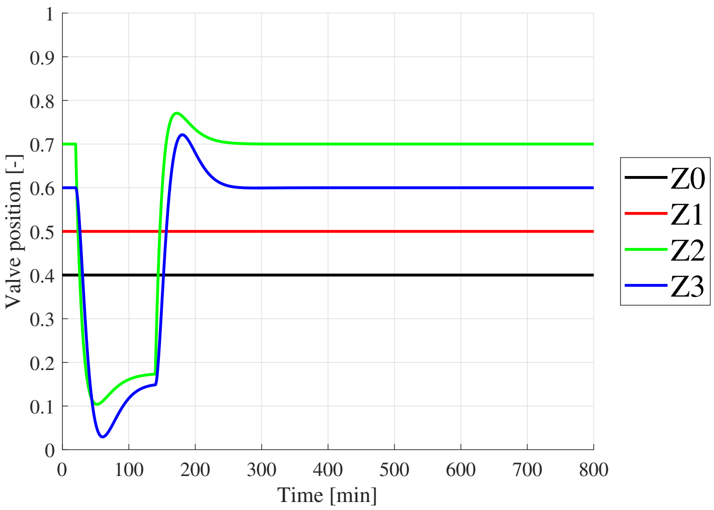
Figure 33: External filling of tank 1 with a flow (F4) of 0.12 m^3/min from t = 20 to t = 260 minutes with generalized split range control

5.5 Draining of tank 2

Here tank 2 is drained with an outside stream F_7 with a rate of 0.12 m^3/min from t = 20 to t = 140. As a reminder the nominal flow rate through the system is 0.16 m^3/min.

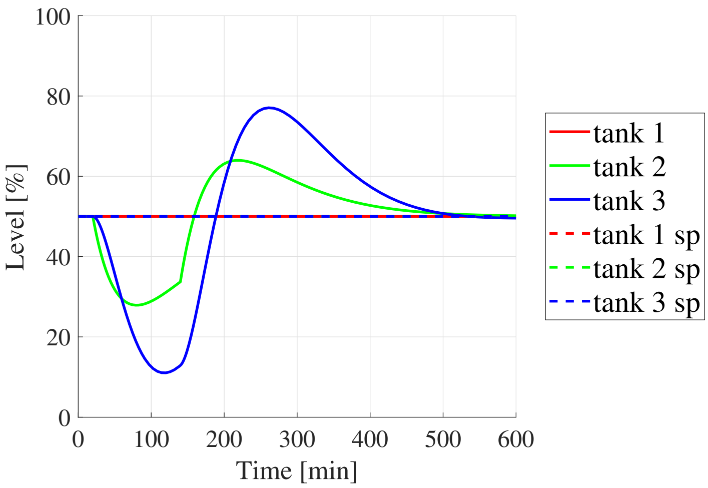


(a) Tank levels

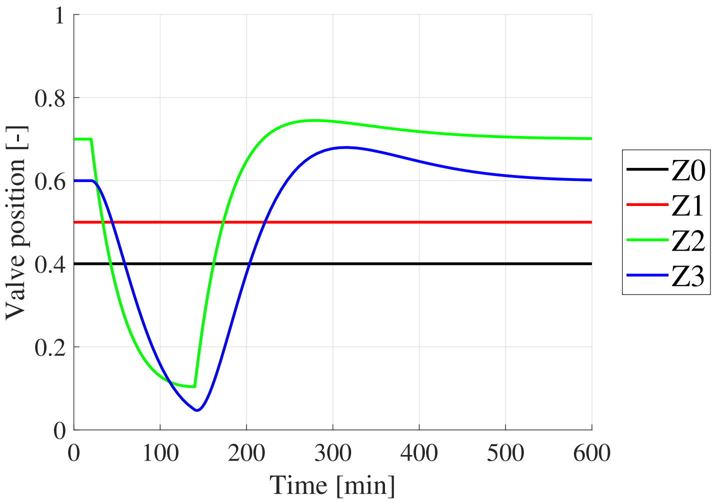


(b) Valve positions

Figure 34: External draining of tank 2 with a flow (F7) of 0.12 m^3/min from t = 20 to t = 140 minutes with two controllers for each tank.

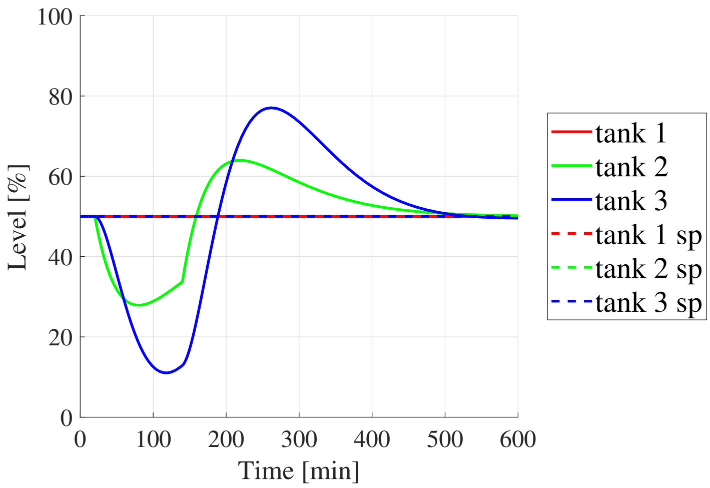


(a) Tank levels

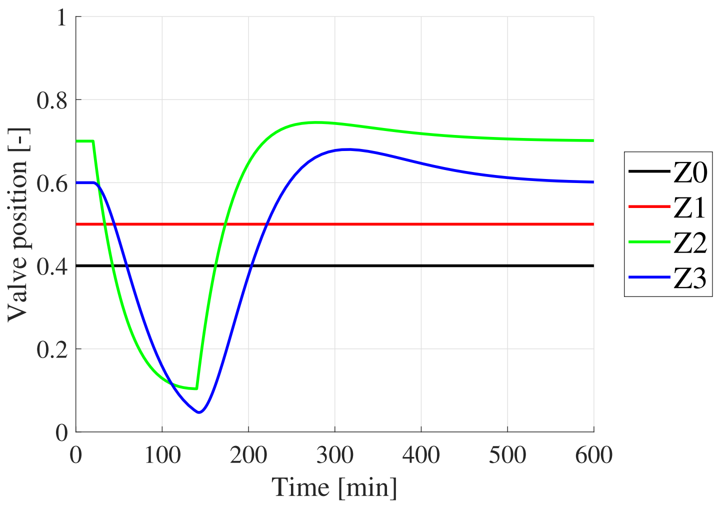


(b) Valve positions

Figure 35: External draining of tank 2 with a flow (F7) of 0.12 m^3/min from t = 20 to t = 140 minutes with split range control.



(a) Tank



(b)

Figure 36: External draining of tank 2 with a flow (F7) of 0.12 m^3/min from t = 20 to t = 140 minutes with generalized split range control.

Chapter 6

Discussion

The objective of this project was to find a control structure that is best suitable for inventory control when we want to maintain production during temporary disturbances. By "maintaining production" it is meant that the objective is to keep flows between units constant despite temporary bottlenecks. The modeled system, consisting of three tanks coupled in series with liquid inventories, were tested for different disturbances with four different control structures: Split range control, generalized split range control, model predictive control and use of two controllers with different setpoints.

Valve position control can be used when we have two MVs to control a CV, where one MV has much slower dynamics than the other. For a given tank here we do indeed have two MVs, but the valves are very equal in size, as can be seen from the nominal point given in Table 1. We also want the TPM to move in the process, meaning that the setpoint u_2^{SP} in Figure 7 could not have been constant. Both of these arguments makes it so that VPC does not fit as a control structure for the studied process, and will not be discussed further here.

A few observations can be made based on the simulations conducted here:

- The responses of the control structures for split range control and generalized split range control are very similar, if not a perfect match comparing, e.g., Figure 28 and 29. This is not surprising, as the only benefit of using generalized SRC compared to "standard" SRC is to avoid the compromise of a common integral time. Here the controllers are tuned with the same integral time, since all MVs have a similar dynamic behavior.

- The MPC handles the disturbances perfectly in terms of level control since there is no model error (perfect level measurement and prediction) by comparing Figure 22, 26 and 30. This makes it however bad at "maintaining production", since when a disturbance happens it immediately reduces production at all points in the process. It could be however good for inventory control with the objective of "averaging level control", because it can manipulate all flows simultaneously and thus smoothing out the flows. Another way of implementing it could have possibly made it better to maintain production, but such an implementation will not be considered further here.

For a temporary bottleneck at the outlet of the plant it is observed that only z_2 is being manipulated by a controller with use of two controllers with different setpoints (Figure 19). The remaining valves z_0 and z_1 remains constant. For SRC, generalized SRC and MPC all valves are manipulated at the instant the disturbance occur (Figures 20, 21 and 22, respectively). The same happens for a temporary disturbance inside the plant, with a manual reduction in z_2. With the use of two controllers for each tank in Figure 23 only z_3 is manipulated by a controller while z_0 and z_1 is kept at a constant value. For SRC, generalized SRC and MPC all valves are again being manipulated (Figures 24, 25 and 26, respectively).

With a temporary reduction of the inlet flow to the process (z_0) (Figures 27 to 30), all flows are manipulated for all control structures tested. This is also expected, because z_0 is the nominal TPM which the control structures are designed around to have consistent inventory control.

When there is an external filling of a tank Figure 31 clearly shows that the control structure with two controllers are able to take advantage of this and fill up the inventory when a "natural" bottleneck is reached in the plant, here given by z_2 saturating fully open. SRC and generalized SRC again reduces the flow through the system when this natural bottleneck is reached, as seen in Figures 32 and 33 respectively.

Chapter 7

Conclusion and future work

The work conducted in this report was to find a control structure that performed best at "maintaining production", where the objective is to keep mass flows between units constant despite temporary bottlenecks. Control structures using split range control (SRC), generalized split range control, model predictive control (MPC) and multiple controllers with different setpoints where designed and compared against each other for different disturbances with this objective in mind. The results showed that the control structure with use of two controllers, one with a low setpoint and one with a high setpoint where the best at maintaining production for a temporary bottleneck. This is because during the temporary bottleneck, the different setpoints makes it possible to automatically fill the inventories. For SRC, generalized SRC and MPC we have only one setpoint for the level in a given tank, which the controllers will always strive to keep the levels at. They also require some additional logic to handle switching between manipulated variables (SRC, generalized SRC) or a model of the plant (MPC) which can take a lot of time to design. The way the control structures SRC, generalized SRC and MPC is formulated here, they seem to be better at the case of "Averaging level control", because of the smoother variations in the use of MVs.

For future work it might be of interest to find a way to implement MPC which can let the levels vary more freely between the operational limits, but to the best of the knowledge of the author this would go beyond the "standard" MPC-formulation described in this report.

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Appendix A

Comments to MATLAB files

These are some comments to my Matlab and Simulink files that will be delivered as appendixes to my report in Inspera.

The control structures SRC, generalized SRC and controllers with different setpoints are each implemented in their own Matlab-script and Simulink-file. The scripts to run the simulation are respectively named Project_3_tanks_SRC, Project_3_tanks_GSRC and Project_3_tanks_2_controllers. The Simulink files ahve the names Project_three_tanks_SRC, Project_three_tanks_GSRC and Project_three_tanks_2_controllers. The control structure with MPC are written in the file "MPC" with no corresponding Simulink-diagram, since the simulation are run in a FOR-loop in that file.

The model of the plant are given in "SysThreeTanks" for SRC, Generalized SRC (GSRC) and controllers with different setpoints, which as mentioned in the report are introduced in the Simulink files by use of the Intrepreted MATLAB function block. The mathematical model is the same for the MPC structure, but given in the file "Plant_three_tanks" instead.

The split range blocks are written in the files "SRC_tank1", "SRC_tank2" and "SRC_tank3" for controllers of each tank, also implemented in the Simulink file by using the Intrepreted MATLAB function block. The functions to update the internal variable v are written in "update_v1", "update_v2" and "update_v3" for tank 1, 2 and 3 respectively. The switching logic for the generalized SRC structure are given in the files "Switching_logic_tank_1", "Switching_logic_tank_2" and "Switching_logic_tank_3".

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