ISSN 0280-5316
ISRN LUTFD2/TFRT--5832--SE

Buffer Management Strategies for Improving Plant Availability

Anna Lindholm

Department of Automatic Control
Lund University
January 2009

Lund University
Department of Automatic Control
Box 118
SE-221 00 Lund Sweden

Document name
MASTER THESIS

Date of issue
January 2009

Document Number
ISRN LUTFD2/TFRT--5832--SE

Author(s)
Anna Lindholm

Supervisor
Krister Forsman Perstorp AB
Charlotta Johnsson Automatic Control, Lund
(Examiner)

Sponsoring organization

Title and subtitle
Buffer Management Strategies for improving Plant Availability(Bufferthanteringsstrategier för att förbättra tillgängligheten hos en produktionsanläggning)

Abstract
In an industrial plant availability is an important factor since increased availability often gives an increase of final production which in many cases means an increased profit for the company. It has been shown that good strategies for buffer management could increase the availability of an entire plant which motivates further research in this area. To be able to compute the optimal solution to a buffer management problem there are many constraints to be specified, among others the topology of the system, the disturbance model and the definition of flow variations. The optimal solution will depend on the choice of constraints and thereby the general buffer management problem becomes very complex. Today a few of the possible problem formulations have been solved but there are still many gaps in the research in this area. The purpose of this thesis is to sort out what has been done in the area of buffer management and to try to fill out some of the gaps. The developed methods and strategies are used for improving buffer management in two plants at Perstorp AB. These case studies reveal some buffer management problems that can be present within the process industry.

Keywords
Buffer management, buffer tanks, availability, bottleneck, averaging level control

Classification system and/or index terms (if any)

Supplementary bibliographical information

ISSN and key title
0280-5316

ISBN

Language
English

Number of pages
74

Recipient's notes

Security classification

Contents

Chapter 1

Introduction

An industrial plant often consists of a production line with buffer tanks in between some of the producing units. The purpose of the use of buffer tanks can be different and in many cases a buffer tank serves several purposes. The most common use of buffer tanks is either to separate production units from each other or to minimize flow variations in the in- or outflow of the buffer tank. Flow variations often cause poor behaviour or failure of sensitive units in the production line which motivates the use of buffer tanks in process sections where flow variations are present. A good buffer management strategy could thus in many cases increase the availability[1] of a plant. A buffer tank before the bottleneck unit[2] of the system could be used to ensure that this unit can run at its maximum speed even if a unit before the bottleneck suffers a shutdown, which in the long run will be a significant gain of production volume.

The purpose of this thesis is to investigate the topic of buffer management by looking at what has been done in the area and suggesting new methods for good buffer handling to maximize the production in a plant. In chapter 2 the complexities of buffer management are discussed and some of the achievements within the area are reviewed. The theory needed to understand the problem is presented in chapter 3.

In this thesis the approach is to begin from scratch and look at buffer management with no constraints on the involved variables and solve this simple buffer management problem theoretically. This is done in section 4.1. From the simple problem formulation the problem can be expanded in numerous ways to get a more realistic view of the problem. Here the effects when having level controlled buffer tanks is explored in section 4.2.

The master thesis is done at Perstorp AB and the developed methods and strategies for buffer management will be evaluated at real plants within the company. Information about the buffer management problems studied at Perstorp AB is available in chapter 5.

[1] The percentage of time the plant produces.
[2] The unit with the smallest capacity.

Chapter 2

Structuring the area of buffer management

The area of buffer management is very complex and before the optimal solution to a specific problem can be found four basic concepts has to be defined:

- Process model
- Flow roughness model
- Disturbance model
- Optimality

By process model it is here meant the transfer functions for all subprocesses as well as other process parameters that will affect the solution. The process that the optimization problem should be solved for can consist of one or several buffer tanks, here denoted as a local or global problem. The topology of a global problem could be a number of buffer tanks simply connected in series or possibly a setup with recycle flows from one tank to another. Depending on the contents of the recycle flow the solution to the buffer management problem could be affected in various ways. Different units in the process section could also have different start-up times, which could affect the optimal solution for the buffer management problem for the process section.

The flow roughness model gives constraints on the variations in the flow derivative, $\dot{u}$, or in the flow itself, u. A flow roughness model can for example give specifications on $\mathrm{Var}[\dot{u}(t)]$ or on $\max_t |\dot{u}(t)|$. The flow roughness model $\mathrm{Var}[\dot{u}(t)]$ punishes frequent variations whereas $\max_t |\dot{u}(t)|$ punishes large variations in the flow. Another possible flow roughness model is $\mathrm{Var}[u(t)]$, in which the flow itself is considered. There can also be maximum and minimum constraints on the flow.

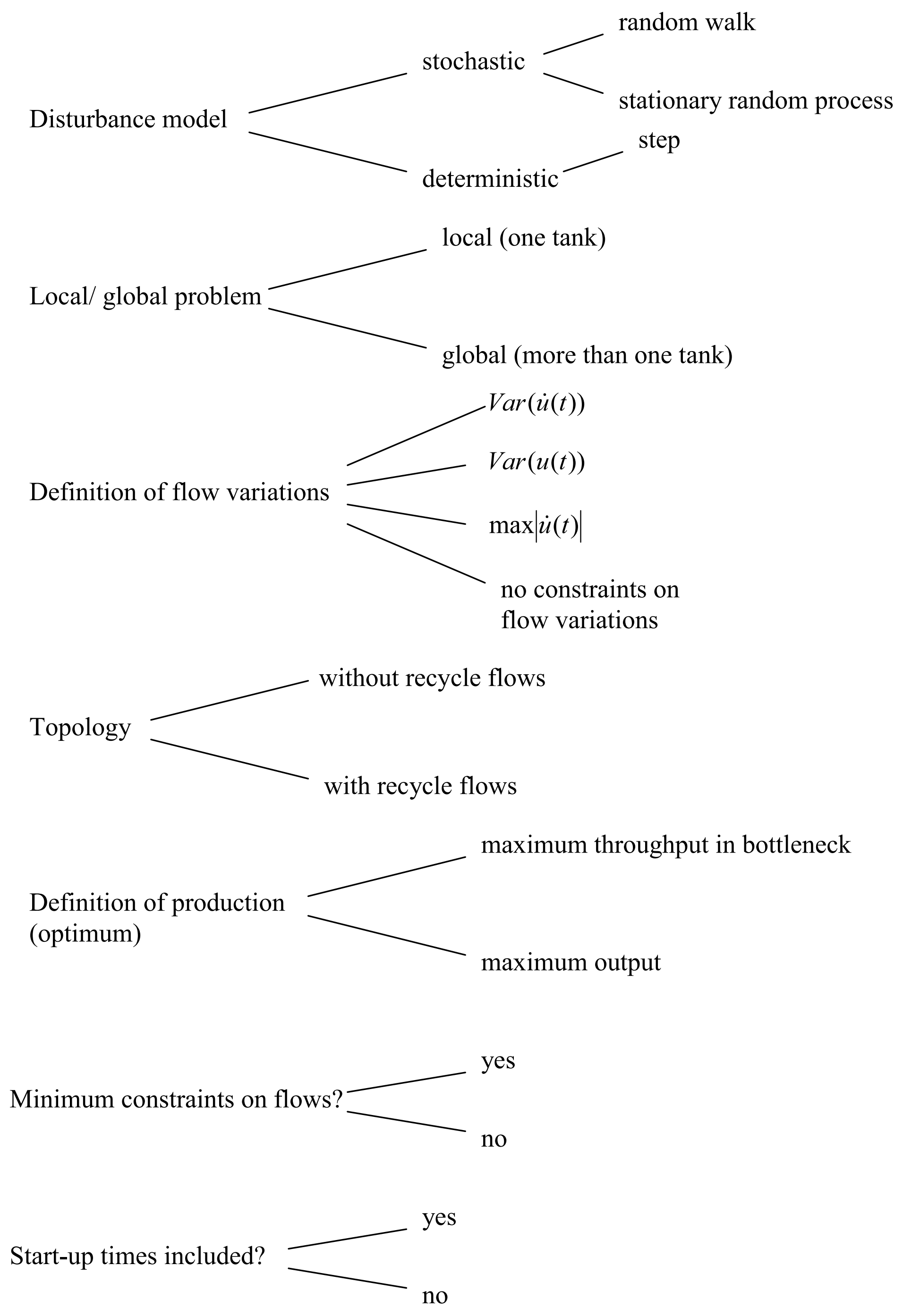

Figure 2.1: Map with some possible constraints for a buffer management problem.

To solve the buffer management problem a disturbance model must also be chosen. A stochastic model or a deterministic model can be chosen. Popular stochastic models for buffer management problems are a *random walk* or a *stationary random process*.

Finally, the optimization criteria for the problem must be specified. For a process with more than one buffer tank this could be equivalent with the definition of production. A common choice is here the final production (output) of the studied process section. However, in some cases it is preferable to define the criterion as throughput in the bottleneck unit. In the long run these two optimization criteria will be approximately equal since all units will be able to catch up with the bottleneck unit in the sense of production speed. If the difference between the capacity of the bottleneck unit and the unit with second slowest maximum production speed is small the time for the bottleneck to catch up will be very long though.

The four basic concepts and their related constraints are shown in Figure 2.1. To try to sort out for which of the constraints in Figure 2.1 the problem has been solved a map was made with indications of what has been done within the area of buffer management, see Table 2.1. The topics to be discussed in this thesis are also indicated at the bottom of the table.

	Disturbance model	**Local/ Global problem**	**Definition of flow variations**	**Topology**	**Definition of production**	**Minimum constraints on flows?**	**Startup-times included?**
Ogawa (1)	stochastic (random walk)	local	Var$[\dot{u}(t)]$	-	-	No	No
Ogawa (2)	stochastic (stationary random process)	local	Var$[\dot{u}(t)]$	-	-	No	No
Ogawa (3)	stochastic	local	Var$[u(t)]$	-	-	No	No
Shin	deterministic (step)	local	$\max_t \|\dot{u}(t)\|$	-	-	No	No
Åberg/ Brogårdh	stochastic	global	No limitations	with recycle flows	output	No	No
Buffer management with no constraints	deterministic	global	No limitations	without recycle flows	max throughput in bottleneck	No	No
Three-tank system with level control	deterministic (step)	global	$\max_t \|\dot{u}(t)\|$	without recycle flows	max throughput in bottleneck	No	No
Warrington	deterministic (step)	local	$\max_t \|\dot{u}(t)\|$	-	max throughput in bottleneck	Yes	Yes
Singapore	deterministic (step)	global	$\max_t \|\dot{u}(t)\|$	with recycle flows	max throughput in bottleneck	No	No

Table 2.1: Table showing what has been done in the area of buffer management and what will be handled in this thesis.

Ogawa, et al [1] and Ogawa [2] have solved the problem for a number of combinations of the above conditions but have only considered the local problem with one tank.

Åberg and Brogård [3] have considered the global problem with recycle flows, but the problem formulation considered in this internal report does not include a definition of flow variations.

In Shin, et al [4] the local problem considering one tank is solved for step disturbances with flow variations defined as the maximum rate of change of the outlet flow ($\max|\dot{u}(t)|$).

None of the above articles has considered minimum limitations on the flows or start-up times for the units.

A remark here is that there should possibly exist solutions to more problem definitions than those stated in Table 2.1, and it almost certainly does. The reason for the poorly filled table is that references for buffer management problems are hard to find and which keywords to search for is not evident. However, it can be concluded that because of the complexity of the problem there are still many problem formulations left to be solved. In this thesis the simplest possible case (section 4.1) will be considered and then the same case but expanded with level control of the buffer tanks (section 4.2) is explored. The conclusions from these two problem formulations will then be used to solve some buffer management problems at Perstorp AB (see chapter 5).

Chapter 3

Theory

3.1 Buffer tanks

The definition of a buffer tank is somewhat vague and a buffer tank can be used for several purposes. A common idea is to use a buffer tank to minimize flow variations to downstream processes. In many real plants flow variations are a large source of failure and poor behaviour of the plant which motivates the use of buffer tanks for this scenario. Another common use is to simply use the buffer tank to separate production units from each other and in that way increase the availability of the plant. In a real plant the purpose of a buffer tank is seldom put in black and white and is rather used for a combination of the reasons above. A buffer tank may or may not be level-controlled. In this chapter the dynamics of a tank will be described in order to understand the problems emerging while working with real plants. Only liquid tanks will be handled in this thesis. The following nomenclature is used.

Nomenclature	
V	Current volume of the tank
y	Current level of the tank
q_{in}	Inflow of tank [volume / time]
q_{out}	Outflow of tank [volume / time]
k_v	Process speed gain
A	Cross-sectional area of tank
r	Radius of the tank
L	Tank height or length
t	Time

3.1.1 The tank process

The cylindrical tank studied in this chapter is shown in Figure 3.1.

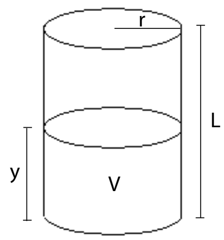

Figure 3.1: The cylindrical tank.

The relation between the volume and the inflow and outflow of a cylindrical tank is:

$$\frac{dV(t)}{dt} = q_{in}(t) - q_{out}(t) \tag{3.1}$$

Integrating both sides gives:

$$V(t) = V(0) + \int_0^t [q_{in}(\tau) - q_{out}(\tau)]d\tau \tag{3.2}$$

Laplace transformation gives:

$$V(s) = \frac{1}{s}[q_{in}(s) - q_{out}(s)] \tag{3.3}$$

which shows that a tank can be modelled as a simple integrator.

In reality the level is almost always measured instead of the volume. In that case it might be preferable to work with the level of the tank instead of the volume. We get:

$$y(t) = y(0) + \frac{1}{A}\int_0^t [q_{in}(\tau) - q_{out}(\tau)]d\tau \tag{3.4}$$

where A is the cross-sectional area of the tank.

To tune a controller for a specific tank process a mathematical model for the tank from control signal (often the inflow or outflow of the tank) to process value (level) is needed.

The model will be $\frac{k_v}{s}$, where the constant k_v specifies how fast the level of the tank changes depending on a change in the control signal. The constant has the unit $time^{-1}$. k_v can be determined either theoretically or by doing a simple step response and using sytem identification. Theoretically $k_v = \frac{1}{fill\ time}$. For more details on how to determine the process speed gain see [5].

3.1.2 Discussion

When working with real plants many problems emerge that require modification of the theoretical model of the tank. A common and time-demanding modification is to get all the scaling factors between the units correct. In a control system the volume of a tank is often indicated in % of the maximum volume or level. The in- and outflows could be mass-flows (in for example kg/h or tons/h) or in % of the maximum flow. Modelling of control systems thereby almost always requires scaling to get the model variables to agree with the measured variables available in the real plant. In this master thesis the choice has been made to handle volumes in % and flows in tons/h to agree with the measurements in the studied plants.

Several buffer tanks have a shape where the cross-sectional area varies with the level in the tank. The relationship between volume and level will then be non-linear which might yield problems when measuring the level and not the volume. However, in many cases the volume as a function of level is close to linear within the given boundaries for the tank and the error when using the linear model is small. To show this statement a lying cylindrical tank will be taken as an example. The volume of a lying cylindrical tank is given by:

$$V(y) = L[(y-r)\sqrt{r^2-(y-r)^2} + r^2\arcsin(\frac{y-r}{r}) + r^2\arcsin(1)] \tag{3.5}$$

If the working range is between 10 % and 90 % of the tank we get the relation between level and volume showed in Figure 3.2). The linear approximation between 10 % and 90 % is $V \approx 1.18y - 9.25$. (The lying cylindrical tank corresponds to D1017A/B in Warrington, see section 5.2.)

In a real plant there will be manipulating valves and sensors with delay and noise involved in the tank process. This might require modification of the tank model if the dynamics of these loops are extremely slow, but this is not the case in the typical physical process. Valves and sensors normally work in the time range 1 s which is much quicker than the level controller, which typically works in minutes or hours.

This discussion motivates that the integrator model of the tank is a sufficiently good model for working with the problems in this thesis. For information on how to design a buffer tank of appropriate size given the disturbance model [6] is recommended. The introduction in this paper also gives a good picture of the different purposes of installing a buffer tank.

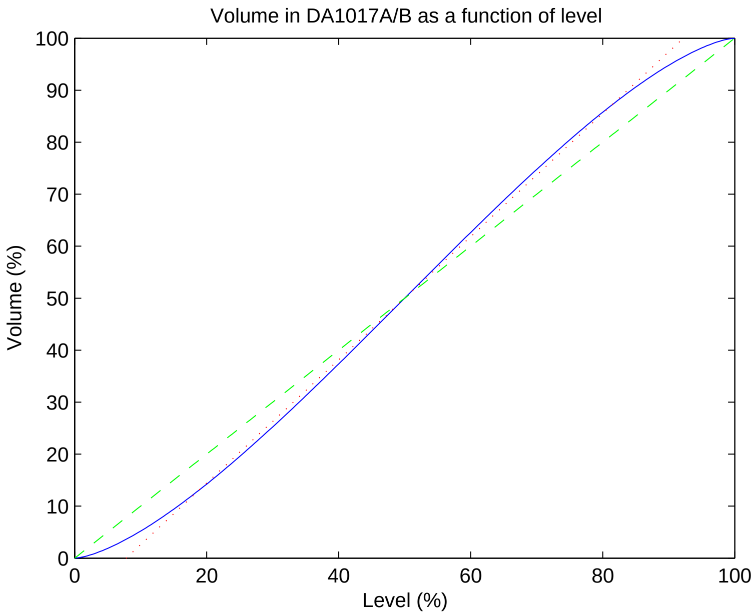

Figure 3.2: Relation between level and volume of a lying cylinder (blue solid line). The green dashed line marks the linear relation $V(t) = y(t)$ and the red dotted line the linear approximation in the interval 10 % to 90 %.

3.2 PID control

The PID-controller is today by far the most widely used controller in the process industry due to its relatively simple implementation and satisfactory performance. Here the PID-controller will be described briefly with focus on P- and PI-controllers for integrating processes. For a more thorough description [7] or [8] is recommended. In this chapter the following nomenclature is used.

Nomenclature

r, SP	reference value, set point
y, PV	output, process value
u, MV, OP	control signal, manipulated variable, output
e	control error
d	disturbance entering between controller and process
n	disturbance at the output of the process
K_c	proportional gain
T_i	integral time
T_d	derivative time
u_b	bias
T_t	tracking time constant
k_v	process speed gain
T_a	arrest time
L	dead-time

3.2.1 Basics

The PID-controller is a feedback controller operating on the control error, e, attempting to reduce the error and maintain the output from the process at its reference value, r. In the process industry the term output is exclusively used for the control signal and the output of the process is entitled the *process value* (PV). To avoid confusion the term output will not be used in this master thesis. The reference value for the controller is often denoted *set point* (SP) within the process industry. The block diagram of the typical feedback system is shown in Figure 3.3.

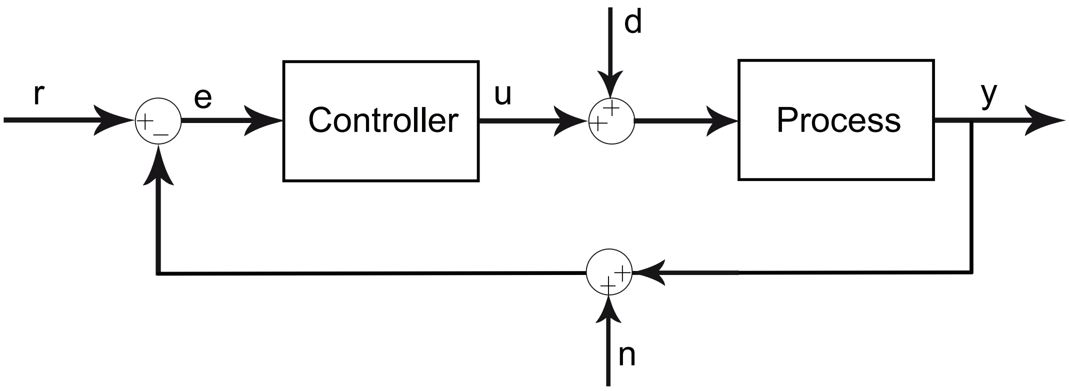

Figure 3.3: Block diagram of feedback control

The controller consists of three parts, the proportional part (P) acting on the current control error, the integral part (I) acting on previous errors and the derivative part (D) acting on the rate of change of the control error. The control signal is computed according to (3.6). In the process industry the control signal is often denoted the manipulated variable (MV) instead of u. The transfer function for the PID-controller is given in (3.7).

$$u(t) = K_c(e(t) + \frac{1}{T_i}\int_0^t e(\tau)d\tau \; + T_d\frac{de(t)}{dt}) \tag{3.6}$$

$$G_{PID}(s) = K_c(1 + \frac{1}{sT_i} + sT_d) \tag{3.7}$$

The derivative part is very sensitive to measurement noise and requires filtering to achieve a satisfactory result. This motivates the use of P- or PI-controllers if there are no certain circumstances that suggest otherwise. In this master thesis P- or PI-control will be used in all level control loops. These two types of controllers and their performance will be handled separately in the sections below. A comparison between the performance of a P-controller and a PI-controller with T_a-tuning will also be made in section 3.2.4.

3.2.2 P-control

With only P-control there will be a stationary error when a step disturbance enters before the process which is the main reason that this type of controller today is rarely used in the process industry. However, the P-controller has still shown to have good performance for buffer tank level control, where the importance is not to keep tight level control of the buffer tank but to keep the control signal smooth while keeping the level within high and low limits.

To avoid getting a steady-state error a bias term is included in the computation of the control signal. The bias term is the desired value of the control signal when the control error is equal to zero. The control signal becomes

$$u(t) = K_p e(t) + u_b \tag{3.8}$$

P-control for integrating processes

An advantage with the P-controller compared to the PI- and the PID-controller is its simplicity which makes the tuning of the controller very straightforward. When controlling an integrating process the bias term is needed to ensure that a correct steady-state level is attained. The following characteristics are obtained for an integrating process controlled with a P-controller.

- The system will have a pole in $-K_c k_v$ and no zeroes.
- The stationary error due to a step disturbance of C percent of the manipulated variable will be $\frac{C}{K_c}$. This is equal to the maximum control deviation due to the step disturbance, e_{max}.
- There will be no overshoot in the manipulated variable due to a step disturbance.
- The maximum derivative of the manipulated variable when a unity step disturbance occurs will be $\max|\dot{u}(t)| = K_c k_v$. The maximum occurs at the same moment as the step disturbance enters.
- When the set point makes a step change there will be no overshoot in the process variable.

The above stated characteristics are derived in the subsection below.

A difficulty when using these characteristics to tune a P-controller according to given specifications is the interpretation of C. C is the disturbance measured in percent of the manipulated variable whereas the measurable disturbances could be for example a step change in the inflow to a buffer tank. The step change in the inflow must then be translated to the range of the control signal if C should be achieved.

Derivation of P-control characteristics

Consider an integrating process with the transfer function $G_p(s) = \frac{k_v}{s}$ controlled with a P-controller $G_c(s) = K_c$.

The closed loop system will then have the transfer function

$$G_{ry}(s) = \frac{G_c(s)G_p(s)}{1+G_c(s)G_p(s)} = \frac{K_c\frac{k_v}{s}}{1+K_c\frac{k_v}{s}} = \frac{K_c k_v}{s+K_c k_v} \tag{3.9}$$

from set point to process value.

The characteristic equation is: $s+K_c k_v = 0$ which gives a pole in $s = -K_c k_v$.

When the set point makes a unity step change we get $Y(s) = G_{ry}(s)\frac{1}{s}$.

Inverse Laplace transform gives $y(t) = 1 - e^{-K_c k_v t}$

with the derivative $\dot{y}(t) = K_c k_v e^{-K_c k_v t}$.

The derivative is never equal to zero which proves that there will be no overshoot in the process variable due to a set point change. Since the transfer function from set point to process value is equal to the transfer function from a disturbance entering before the process to the control signal it will also hold that there will be no overshoot in the control signal due to a step disturbance.

The derivative of the control signal is $\dot{u}(t) = K_c k_v e^{-K_c k_v t}$

when the system suffers from a unity step disturbance entering before the process.

The maximum value of the derivative is obtained when $t = 0$, $\max|\dot{u}(t)| = |\dot{u}(0)| = K_c k_v$

The transfer function from a step disturbance d entering before the process to the control error e is:

$$G_{de}(s) = \frac{G_p(s)}{1+G_c(s)G_p(s)} = \frac{\frac{k_v}{s}}{1+K_c\frac{k_v}{s}} = \frac{k_v}{s+K_c k_v} \tag{3.10}$$

which gives $E(s) = G_{de}(s)\frac{1}{s}$ when the disturbance is a unity step.

The final value theorem gives that

$$\lim_{t\to\infty} e(t) = \lim_{s\to 0} sE(s) = \frac{1}{K_c} \tag{3.11}$$

The expression for the control error $e(t) = \frac{1}{K_c}(1-e^{-K_c k_v t})$ shows that the stationary error $\frac{1}{K_C}$ is also the maximum control deviation, e_{max}.

3.2.3 PI-control

The PI-controller is widely used within the process industry. In this section methods for anti-windup and PI-controller tuning for integrating processes will be handled.

Anti-windup

A complication that arises when including an integral part in the controller is integrator windup. This problem occurs when the control signal saturates and the integral part grows too large (winds up). As a result of this the control signal continues to saturate even when the control error becomes zero and large overshoots may be obtained. The phenomenon is illustrated in Figure 3.4.

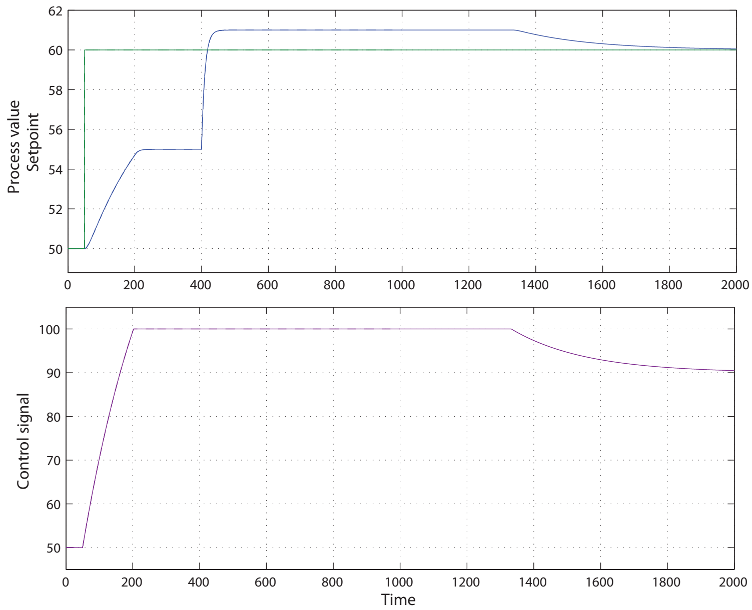

Figure 3.4: Without anti-windup. At time t=400 time units a step disturbance occurs but the control signal is not decreased until at approximately t=1350 time units.

To avoid this undesired behaviour of the controller there are several anti-windup methods available. Two of the most common approaches are conditional integration and back-calculation ("tracking"). The method used in this master thesis is back-calculation. This method is described briefly below.

The method of back-calculation or "tracking" suggests recomputation of the integral term when the control signal saturates by increasing the integral term with the difference between the saturated and unsaturated control signal. This means that if the controller output is greater than the upper limit of the actuator the integral part is decreased and if the controller output is less than the lower limit of the actuator the integral part is increased. The scheme is shown in Figure 3.5.

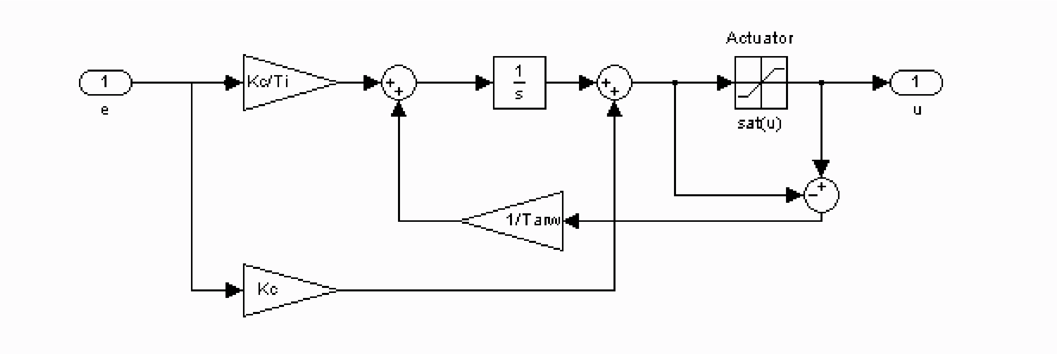

Figure 3.5: Anti-windup scheme with back-calculation.

The rate at which the integral part is reset is determined by the tracking time constant, T_t. A number of suggestions on how to choose this constant have been proposed. For a PID-controller it has been suggested that: $T_t = \sqrt{T_i T_d}$. However, this approach is not valid for PI-controllers where $T_d = 0$. For PI-controllers $T_t = T_i$ is suggested as an alternative, which is the choice to be used in this master thesis.

PI-controller tuning for integrating processes

Tuning of PID-controllers is a widely discussed topic within the area of automatic control. In this thesis it has been chosen to work with T_a-tuning of PI-controllers for integrating processes. The method and the characteristics of the controller is described in this subsection.

When having determined the process speed gain, k_v (see section 3.1) the tuning of a PI-controller for an integrating process can be achieved using the T_a-tuning rule[1]. The method has a single tuning parameter; the arrest time, T_a, which is the time elapsed before the process value curve starts to return to the set point after the occurrence of a step disturbance in the flow (see Figure 3.6). In other words, the tuning is based on the handling of disturbances and not on the performance during set point changes. A motivation for this choice of tuning specification for buffer management is that the set point in a level control loop is rarely changed.

[1] Sometimes also called the λ-tuning rule but the method is not consistent with λ-tuning for KLT processes, see [5].

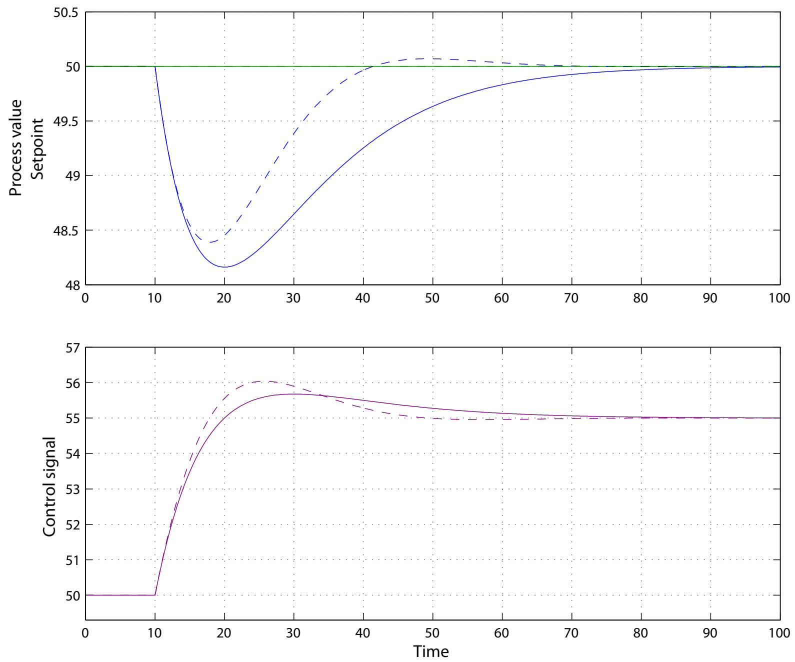

Figure 3.6: Simulation of a PI-controlled integrating process with arrest time T_a = 10. The solid line marks T_a-tuning with $T_i = 2T_a$ and the dashed line shows the behaviour of T_a-tuning with $T_i = T_a$

The characteristics of T_a-tuning for integrating processes are stated below.

- The system will have a double pole in $-\frac{1}{T_a}$ and a zero in $-\frac{1}{2T_a}$
- The maximum control deviation due to a step disturbance will be at time T_a (definition of arrest time). The maximum control deviation will be $e_{max} = 0.3679 k_v T_a C$ resulting from a disturbance of C percent of the manipulated variable.
- When handling a step disturbance the control signal will do an overshoot of 13.5 %.
- The maximum derivative of the manipulated variable when a unity step disturbance occurs will be $\max|\dot{u}(t)| = \frac{2}{T_a}$. The maximum occurs at the same moment as the step disturbance enters.
- The maximum overshoot when the set point makes a step change will occur at time $2T_a$. The amplitude of the overshoot will be 13.5 %.

The choice of controller parameters with T_a-tuning is

$$K_c = \frac{2}{k_v T_a}, \quad T_i = 2T_a \tag{3.12}$$

if there is no dead-time, and

$$K_c = \frac{T_i}{k_v(T_a + L)^2}, \quad T_i = 2T_a + L \tag{3.13}$$

if there is a significant dead-time L. The relationship between K_c and T_i in (3.12) and the characteristics of T_a-tuning stated above are derived in the subsection below.

A common choice is to choose the integral time to be half of that in the original T_a-tuning to get faster recovery from disturbances (see figure 3.6). This tuning, often denoted T_a-tuning with $T_i = T_a$, is in [4] proved to be the optimal PI-controller for averaging level control (see section 3.3) when the input disturbance is a step. The performance of a PI-controller with T_a-tuning with T_i = T_a compared to a PI-controller with T_i = $2T_a$ is shown in Figure 3.6

With T_a-tuning with $T_i = T_a$ the following characteristics are obtained:

- The system will have two complex conjugated poles in $-\frac{1}{T_a} \pm \frac{1}{T_a}i$ and a zero in $-\frac{1}{T_a}$.
- The maximum control deviation will be $e_{max} = 0.3224 k_v T_a C$ resulting from a disturbance of C percent of the manipulated variable.
- When handling a step disturbance the control signal will do an overshoot of 20.8 %.
- The maximum derivative of the manipulated variable when a unity step disturbance occurs will be $\max|\dot{u}(t)| = \frac{2}{T_a}$. The maximum occurs at the same moment as the step disturbance enters.
- The maximum overshoot in the process value when the set point makes a step change will be 20.8 %.

The characteristics stated above are derived in the subsection below.

Derivation of T_a-tuning characteristics

Consider an integrating process with the transfer function $G_p(s) = \frac{k_v}{s}$ controlled with a PI-controller $G_c(s) = K_c(1 + \frac{1}{sT_i}) = K_c\frac{1+sT_i}{sT_i}$.

The closed loop system will then have the transfer function

$$G_{ry}(s) = \frac{G_c(s)G_p(s)}{1 + G_c(s)G_p(s)} = \frac{\frac{k_v}{s}K_c\frac{sT_i+1}{sT_i}}{1 + \frac{k_v}{s}K_c\frac{sT_i+1}{sT_i}} = \frac{\frac{K_c k_v}{T_i}(sT_i + 1)}{s^2 + \frac{K_c k_v}{T_i}(sT_i + 1)} \tag{3.14}$$

from set point to process value.

The characteristic equation is: $s^2 + K_c k_v s + \frac{K_c k_v}{T_i} = 0$.

The poles of the system are given by $s = -\frac{K_c k_v}{2} \pm \sqrt{\frac{K_c^2 k_v^2}{4} - \frac{K_c k_v}{T_i}}$.

The system will have a double pole when $K_c = \frac{4}{k_v T_i}$, which gives the relationship between K_c and T_i with T_a-tuning:

$$K_c = \frac{2}{k_v T_a}, T_i = 2T_a. \tag{3.15}$$

These T_a-tuning parameters with $T_i = 2T_a$ give the following closed loop transfer function from r to y:

$$G_{ry}(s) = \frac{1 + 2sT_a}{(1 + sT_a)^2} \tag{3.16}$$

When the set point makes a unity step change we get $Y(s) = G_{ry}(s)\frac{1}{s}$.

Inverse Laplace transform gives $y(t) = 1 + (\frac{t}{T_a} - 1)e^{-\frac{1}{T_a}t}$

with the derivative $\dot{y}(t) = \frac{1}{T_a^2}(2T_a - t)e^{-\frac{1}{T_a}t}$.

The derivative is equal to zero iff $t = 2T_a$ which gives $y(2T_a) = 1 + e^{-2} \approx 1.1353$ corresponding to a maximum overshoot of 13.5 %. Since the transfer function from set point to process value is equal to the transfer function from a disturbance entering before the process to the control signal it will hold that the maximum overshoot in the control signal due to a step disturbance is also 13.5 %.

This also gives that $\dot{u}(t) = \frac{1}{T_a^2}(2T_a - t)e^{-\frac{1}{T_a}t}$

when the system suffers from a unity step disturbance entering before the process.

The second derivative of the control signal becomes $\ddot{u}(t) = \frac{t-3T_a}{T_a^3}e^{-\frac{t}{T_a}}$.

The second derivative is equal to zero iff $t = 3T_a$. $t = 3T_a$ gives $|\dot{u}(3T_a)| = \frac{1}{T_a}e^{-3}$. This is less than $|\dot{u}(0)| = \frac{2}{T_a}$ which proves that $\max|\dot{u}(t)| = |\dot{u}(0)| = \frac{2}{T_a}$ due to a unity step disturbance.

With T_a-tuning with $T_i = 2T_a$ the transfer function from a step disturbance d entering before the process to the control error e becomes:

$$G_{de}(s) = \frac{G_p(s)}{1 + G_c(s)G_p(s)} = \frac{k_v s}{s^2 + K_c k_v s + \frac{K_c k_v}{T_i}} = \frac{k_v s}{(s + \frac{1}{T_a})^2} \tag{3.17}$$

The step response is given by $e(t) = k_v t e^{-\frac{1}{T_a}t}$

with the derivative $\dot{e}(t) = k_v(1 - \frac{t}{T_a})e^{-\frac{1}{T_a}t}$.

The derivative is equal to zero iff $t = T_a$ which gives the maximum control deviation $e(T_a) = k_v T_a e^{-1} \approx 0.3679 k_v T_a$.

The same calculations as above can be done for T_a-tuning with $T_i = T_a$. The closed loop transfer function from r to y then becomes

$$G_{ry}(s) = \frac{\frac{2}{T_a^2}(T_a s + 1)}{s^2 + \frac{2}{T_a}s + \frac{2}{T_a^2}} \tag{3.18}$$

When the set point makes a unity step change we get $Y(s) = G_{ry}(s)\frac{1}{s}$.

Inverse Laplace transform gives $y(t) = 1 + e^{-\frac{1}{T_a}t}(\sin(\frac{1}{T_a}t) - \cos(\frac{1}{T_a}t))$

with the derivative $\dot{y}(t) = \frac{2}{T_a}e^{-\frac{1}{T_a}t}\cos(\frac{1}{T_a}t)$.

The derivative is equal to zero iff $t = (\frac{\pi}{2} + n\pi)T_a$ where n is an integer. The expression for $y(t)$ is largest for $n = 0$ which gives $|y(\frac{\pi}{2}T_a)| = 1 + e^{-\frac{\pi}{2}} \approx 1.208$ corresponding to a maximum overshoot of 20.8 %. Since the transfer function from set point to process value is equal to the transfer function from a disturbance entering before the process to the control signal it will hold that the maximum overshoot in the control signal due to a step disturbance is also 20.8 %.

This also gives that $\dot{u}(t) = \frac{2}{T_a}e^{-\frac{1}{T_a}t}\cos(\frac{1}{T_a}t)$.

when the system suffers from a unity step disturbance entering before the process.

The second derivative of the control signal becomes $\ddot{u}(t) = -\frac{2}{T_a^2}e^{-\frac{1}{T_a}t}(\cos(\frac{1}{T_a}t) + \sin(\frac{1}{T_a}t))$.

The second derivative is equal to zero iff $t = (\frac{3\pi}{4} + n\pi)T_a$ where n is an integer. The expression for $\dot{u}(t)$ is largest for $n = 0$ which gives $|\dot{u}(\frac{3\pi}{4}T_a)| = \frac{2}{\sqrt{2}T_a}e^{-\frac{3\pi}{4}}$. This is less than $|\dot{u}(0)| = \frac{2}{T_a}$ which proves that $\max|\dot{u}(t)| = |\dot{u}(0)| = \frac{2}{T_a}$ due to a unity step disturbance.

With T_a-tuning with $T_i = T_a$ the transfer function from a step disturbance d entering before the process to the control error e becomes:

$$G_{de}(s) = \frac{G_p(s)}{1 + G_c(s)G_p(s)} = \frac{k_v s}{s^2 + K_c k_v s + \frac{K_c k_v}{T_i}} = \frac{k_v s}{(s^2 + \frac{2}{T_a}s + \frac{2}{T_a^2})} \tag{3.19}$$

The step response is given by $e(t) = k_v T_a e^{-\frac{1}{T_a}t}\sin(\frac{1}{T_a}t)$

with the derivative $\dot{e}(t) = k_v e^{-\frac{1}{T_a}t}(\cos(\frac{1}{T_a}) - \sin(\frac{1}{T_a}))$.

The derivative is equal to zero iff $t = (\frac{\pi}{4} + n\pi)T_a$ where n is an integer.

The maximum control deviation is obtained for $n = 0$ and has the value $e(\frac{\pi}{4}T_a) = \frac{k_v T_a}{\sqrt{2}}e^{-\frac{\pi}{4}} \approx 0.3224 k_v T_a$.

More derivations of characteristics for PI-controllers with T_a-tuning can be viewed in [9].

3.2.4 Comparison of the performance of the P- and the PI-controller for integrating processes

The characteristics of the P-controller and the PI-controller with T_a-tuning with $T_i = 2T_a$ and $T_i = T_a$ have been concluded in Table 3.1. e_{max} and $\max|\dot{u}(t)|$ is the maximum control deviation and the maximum derivative of the control signal due to a unity step disturbance respectively. The term "Overshoot" estimates the overshoot in the level when the set point makes a step change or the overshoot in the control signal due to a step disturbance.

	Poles	**Zeroes**	$\mathbf{e_{max}}$	$\max\lvert\dot{\mathbf{u}}(\mathbf{t})\rvert$	**Overshoot**
P	$-K_c k_v$	—	$\frac{C}{K_c}$	$K_c k_v$	—
PI $\mathbf{T_i = 2T_a}$	$-\frac{1}{T_a}$ (double) $= -\frac{K_c k_v}{2}$ (double)	$-\frac{1}{2T_a}$ $= -\frac{K_c k_v}{4}$	$0.3679 k_v T_a C$ $= 0.7358 \frac{C}{K_c}$	$\frac{2}{T_a}$ $= K_c k_v$	13.5 %
PI $\mathbf{T_i = T_a}$	$-\frac{1}{T_a} \pm \frac{1}{T_a} i$ $= -\frac{K_c k_v}{2} \pm \frac{K_c k_v}{2} i$	$-\frac{1}{T_a}$ $= -\frac{K_c k_v}{2}$	$0.3224 k_v T_a C$ $= 0.6448 \frac{C}{K_c}$	$\frac{2}{T_a}$ $= K_c k_v$	20.8 %

Table 3.1: Characteristics of the P-controller and the PI-controller with T_a-tuning with $T_i = 2T_a$ and $T_i = T_a$.

Here it can be seen that the maximum control deviation due to a step disturbance is greater for the P-controller than for the PI-controllers but with PI-control there will be an overshoot in the level due to set point changes and in the control signal due to step disturbances. The overshoot increases with the integral part, i.e. decreases with increasing integral time, T_i.

3.3 Averaging level control

Averaging level control is a control strategy aiming to keep the outlet flow of a tank smooth while keeping the level within high and low limits. Many suggestions on how to achieve this have been documented and the optimal controller for different specific cases has been derived. For a more thorough description of averaging level control and deriving optimal controllers for different scenarios [2] is recommended.

Before the optimal controller for averaging level control can be derived these three concepts have to be defined, compare the discussion in chapter 2:

- Process model
- Flow roughness model
- Disturbance model

Once these concepts are defined the optimal controller for this specific case can be derived by minimizing some performance index subject to a constraint condition. The optimization problem can be solved using either a state space method with a noise free observer (solving the Ricatti equation) or the Wiener-Hopf method (transfer function approach). In [2] a few different cases are considered. The results of the calculations of the optimal *linear* controllers made in [2] are summarized in Table 3.2. For derivation see [2].

Flow roughness / **Disturbance model**	Var[$\mathbf{u}(\mathbf{t})$]	Var[$\dot{\mathbf{u}}(\mathbf{t})$]
Random walk	PD controller $C_{PD} = c_0 + c_1 s$	PI controller $C_{PI} = K_c \frac{s+b}{s}$
Stationary random process	PD controller $C_{PD} = c_0 + c_1 s$	Phase lag network $C_L = K_c \frac{s+b}{s+a}$

Table 3.2: Optimal linear controllers for a few buffer management problem formulations.

In many articles averaging level control problems constrained to a certain type of controller or control strategy are considered, for example the optimal P-controller for averaging level control is derived in [10] and the optimal PI-controller is derived in [4].

The definition of the flow roughness model and the disturbance model will affect the properties of the optimal controller. The most appropriate choice of definition of the flow roughness model and the disturbance model depends entirely on the characteristics of the real process and the desired behaviour of the optimal controller.

3.4 Bidirectional inventory control

The strategy of bidirectional inventory control is handled in [11] and [12]. A brief description of the strategy will be given here.

Consider a production line with three level-controlled tanks where the master flow is the inflow to the first tank of the system. Given this master flow the three tanks should be controlled by manipulating the outflow as in Figure 3.7.

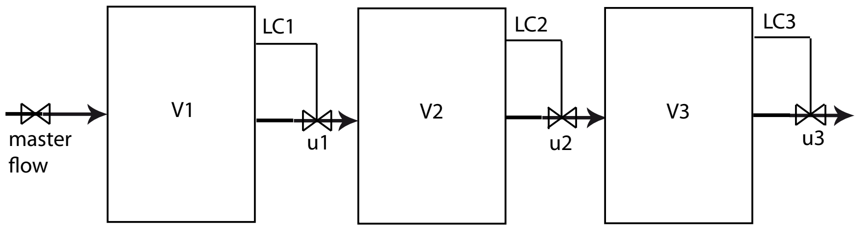

Figure 3.7: Three level-controlled tanks manipulating the outflow.

If a shutdown occurs after the third tank in the production line the manipulated variables should be changed for the three controllers so that the inflow of the tanks are manipulated, see Figure 3.8. The outflow from the third tank has now become the new master flow of the production line.

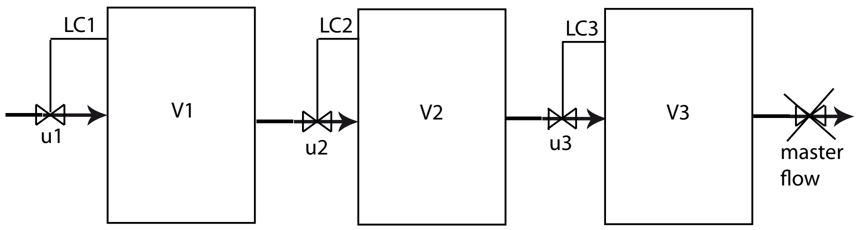

Figure 3.8: Bidirectional inventory control of the three level-controlled tanks. When a shutdown occurs after the third tank the controllers change the manipulated variable so that the inflows of the tanks are controlled.

In practice the change of manipulated variable is done automatically, for example using a high-level controller that takes over manipulation of the inflow to the tank when a high level limit is reached. An example of this type of bidirectional flow control can be viewed in the section about the case study in Singapore, section 5.3.

Chapter 4

Buffer management

4.1 Buffer management without constraints

A plant often has a very complex structure and consists of several buffer tanks and it is not elementary to determine how the set points for the buffer tank levels should be chosen to maximize the final production when one of the units suffers a shutdown or reduced production rate. In this section a very simplified buffer management problem is studied, with the intention of getting a sense of how the choices of buffer tank levels affect the final production.

4.1.1 Problem definition

Consider a system with i producing units, P_i (Figure 4.1) connected in series. Each unit is followed by a buffer tank with a certain current volume, x_i, and a total volume capacity, V_i. The producing units have different capacities and the bottleneck unit is defined as the unit with the smallest capacity. The bottleneck unit's capacity is limiting for the entire system and thus the total production for the system can be defined as the output flow of the bottleneck unit. In this simplified approach it is assumed that the output flows from the units can be changed momentarily.

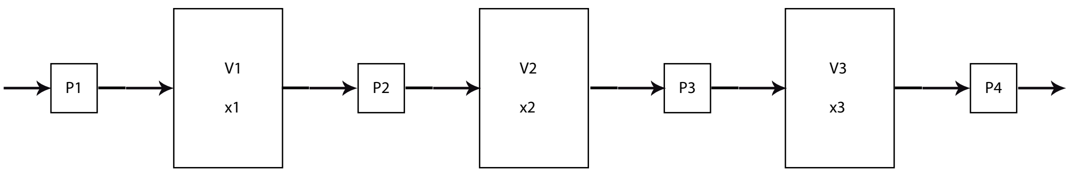

Figure 4.1: The simple system considered in this chapter.

If for some reason one of the producing units fails and is forced to run at reduced speed, the failure can be considered as a loss of production volume, f, during the time that the failing unit is forced to run at reduced speed, see Figure 4.2. So, notice that by failure it is not neccesarily meant a complete stop of a unit. The total production will only be affected by the failure if the bottleneck unit is constrained to slow down its production speed, since all remaining units produce faster and thereby in the long run have the ability to catch up. If the buffer levels are chosen inappropriately there will be scenarios where the bottleneck has to slow down due to empty buffers before the unit or full buffers after the unit and an unnecessary loss of production volume will be obtained. How large the total production loss will be depends on the levels in the buffer tanks between the failing unit and the bottleneck unit at the time of the failure. The approach is to run every unit at maximum speed for as long as possible.

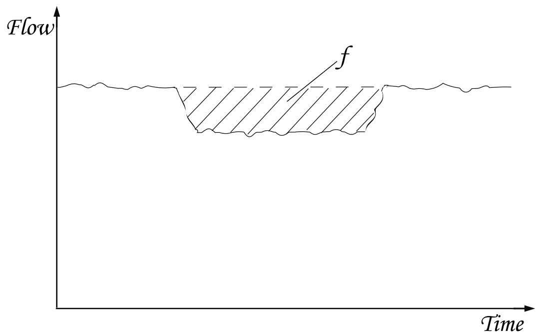

Figure 4.2: Volume loss due to failure.

4.1.2 Solution

The problem can be divided into two cases. The first scenario is when the bottleneck unit is located before the failing unit in the production line. To allow the bottleneck unit to run at full speed for as long as possible we have to have sufficiently low levels in the buffer tanks between the bottleneck unit and the failing unit. If the total ullage[1] for all buffer tanks between these units are larger than the volume lost in the failing unit during the failure there will be no loss of total production, since all of the units have the possibility to catch up with the bottleneck unit. The second scenario is when the bottleneck unit is located after the failing unit in the production line. In this case the volume in the buffer tanks between the failing unit and the bottleneck has to be sufficiently large to guarantee that the bottleneck unit can produce at full speed during the failure. The total volume in these buffer tanks has to be at least as large as the volume lost during the failure to ensure that there will be no loss of production due to the failure. Two figures illustrating the statements above are shown in Figure 4.3 and 4.4.

[1] The non-filled volume of the tank.

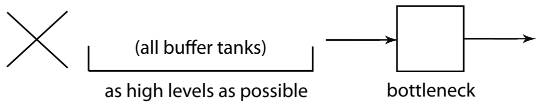

Figure 4.3: Choice of buffer levels when the breaking unit is located *before* the bottleneck.

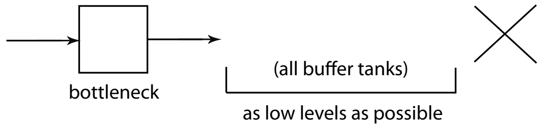

Figure 4.4: Choice of buffer levels when the breaking unit is located *after* the bottleneck.

The constraints of the buffer tank levels are summarized mathematically in equation 4.1 and 4.2 below.

Failing unit: P_j, Bottleneck unit: P_i

If f_j is the volume loss for unit j due to the failure the total loss of production is given by:

1. If P_j is *before* P_i in the production line

$$\max(0, f_j - \sum_{k=j}^{i-1} x_k) \tag{4.1}$$

2. If P_j is *after* P_i in the production line

$$\max(0, f_j + \sum_{k=i}^{j} (x_k - V_k)) \tag{4.2}$$

where f_j is the loss of production volume due to the failure in unit j, x_i is the current volume of tank i as the failure occurs and V_i is the total volume of tank i.

4.1.3 Conclusions

The buffer management problem without constraints discussed in this section illustrates the importance of considering the location of the bottleneck unit in relation to the failing unit when choosing set points for the buffer tank levels. In this problem it has been shown that in the long run the only parameter affecting the total production is the production speed of the bottleneck compared to its maximum production speed. If the difference in maximum capacities between the bottleneck and the unit with the second slowest maximum capacity is small it will however take long time for this unit to catch up after a failure.

This simple approach is not directly applicable on a real plant since a lot of non-realistic assumptions have been made. The strategy to run all units at maximum speed for as long as possible is often not an option because of the impossibility of changing the production speed momentarily. Even if it is possible, it will introduce large and rapid variations of the flows and thereby possibly cause units to trip. We have not at all considered the control strategy or constraints on the flow smoothness and on the minimum obtainable flow, which are important aspects when working with real processes. When introducing control the case with no volume loss does not exist since the controllers will start working immediately as the failure takes place.

In a real process it is mostly not evident where the bottleneck unit and the unit with highest probability of failing are located. If asking the process operators where the bottleneck unit is, the answer would probably be different depending on who you ask and when you ask. It is also not evident how the bottleneck unit should be defined; should maximum speed or average speed be considered? Regarding failing units it is unlikely that only one unit in the production line fails during a certain time period. This suggests a switch from the deterministic model to a statistical approach. The failing unit can for example be considered as the unit with most frequent failures over a certain period of time. However, the definition of a suitable statistical model is not always apparent and requires some consideration.

Finally, the structure with producing units and buffer tanks connected only in series is quite unrealistic. In a real plant the topology is more complex, often with recycle flows and several branches which make the modelling more complicated.

4.2 System with three level-controlled tanks

If the simple problem described in section 4.1 is extended with automatic level control of each of the tanks the problem will become more complex. To start with, there will be a choice whether to control the level of the tanks by manipulating the inflow or the outflow of the tanks. For information about how to choose which variable to manipulate, see [13]. A strategy that can be used when a unit in the production line fails is to change manipulated variables for the affected controllers in a certain way. This strategy is called bidirectional inventory control and was described briefly in section 3.4. According to this approach the three-tank system when manipulating the inflow can be seen as a strategy to handle a failure in the unit located after the third tank (see Figure 4.5). The outflow of the third tank is then the master flow. Similarly the system when manipulating the outflow can be seen as a strategy to handle a failure of a unit located before the first tank in the production line (Figure 4.6). Here the master flow is thus the inflow to the first tank.

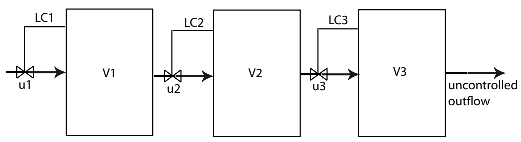

Figure 4.5: Three level-controlled tanks manipulating the inflow.

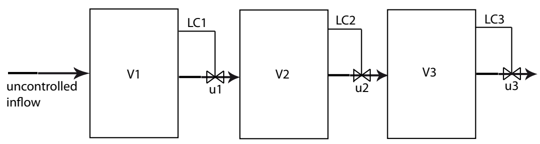

Figure 4.6: Three level-controlled tanks manipulating the outflow.

Another choice to be made is which type of controller to use and how to tune the controller. In this chapter the dynamics of a system with three level controlled tanks using P- and PI-control manipulating the outflow will be observed.

The scenario studied is when a unit located before the first tank is forced to run at reduced speed and the bottleneck unit is the unit located after the third tank. The shutdown can be seen as a step disturbance in the inflow to the first tank and the objective is here to see how the bottleneck unit is affected by the disturbance. In the simulations in this chapter it has been assumed that the breaking unit has to reduce its working speed from 50 % to 40 % of its maximum working speed and that the levels have to be kept between 10 % and 90 % of the tanks.

The Simulink model used in the simulations is shown in Figure 4.7.

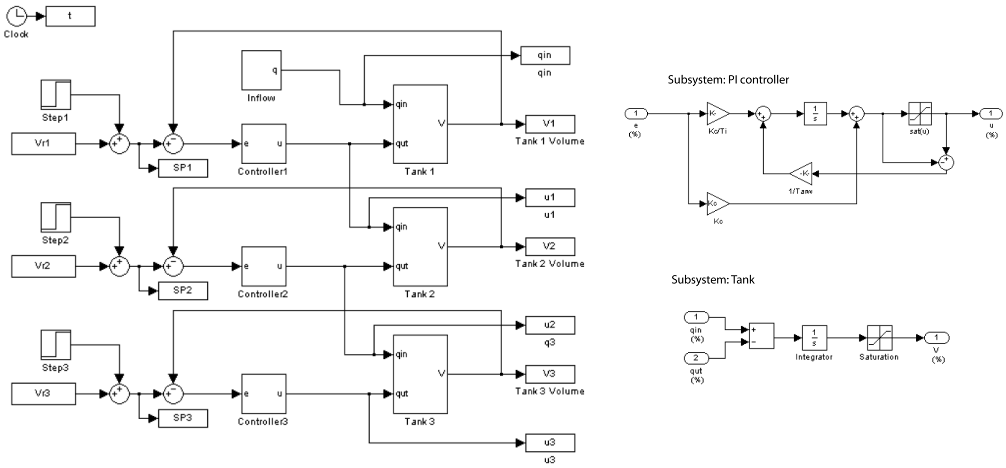

Figure 4.7: Simulink model of the three-tank system with subsystems.

Since the purpose of this chapter is to get a sense of the dynamics of a production line with controlled buffer tanks and not to solve a specific problem all scaling factors in the Simulink model have been excluded. A comment is that if the unit after the third tank really is bottlenecking the system, it would probably run at its full speed i.e. 100 % of the control signal and not 50 % as in the simulations.

In the chapter the following holds for all of the simulation plots: The blue solid line marks the flow or the level of the tank, the green dashed line marks the initial flow or the set point for the controller and the red dash-dotted lines mark the critical limits for the level. An example plot is shown in Figure 4.8.

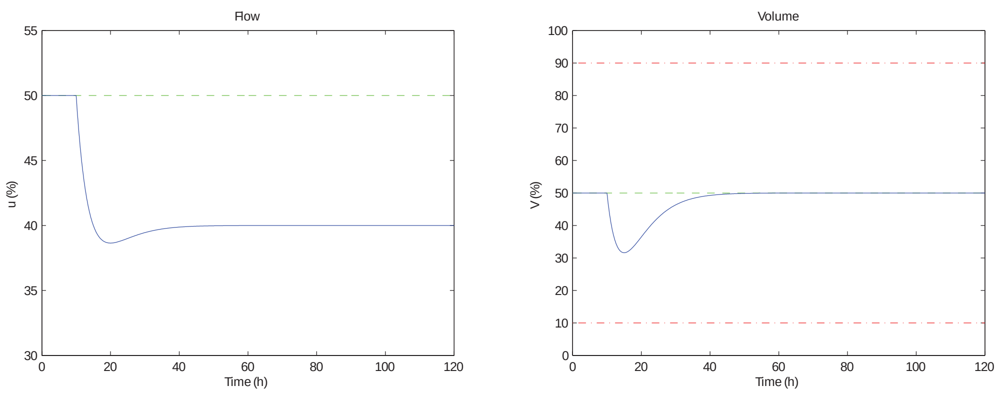

Figure 4.8: Example plot.

4.2.1 PI-control of the three-tank system

Tuning

To get a sense of the dynamics of the system when using PI-control the three controllers were equally tuned with the slowest T_a-tuning for the first controller to handle the step disturbance in the uncontrolled inflow of 10 % of the manipulated variable without letting the level of this tank go outside the limits 10 % and 90 %. The tuning to achieve this is derived in section 3.2.3. The set point is selected as 50 % which ensures that equally large disturbances upwards and downwards in the inflow are handled. The simulation results are showed in Figure 4.9.

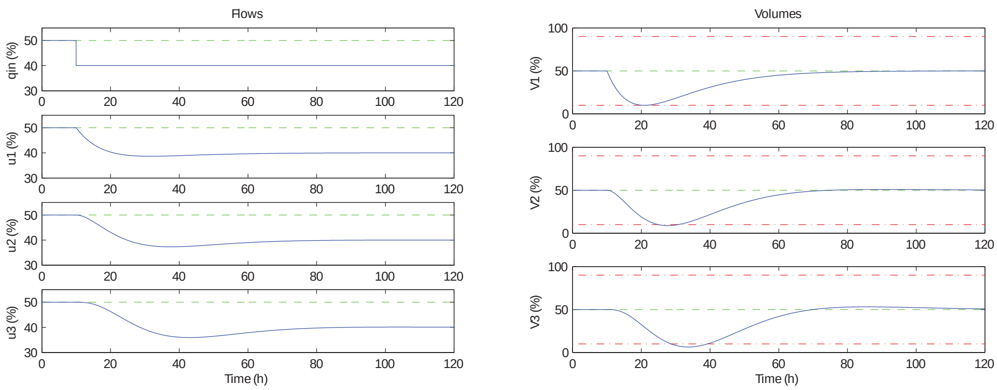

Figure 4.9: PI-control of the three-tank system with set point 50 % and $T_a^1 = T_a^2 = T_a^3 = 10.87$ h. The first unit in the production line starts running at reduced speed at time $t = 10$ h.

Here it can be seen that the second and third PI-controller in the system cannot handle the disturbance well enough and their volumes drift outside the limit 10 %. This can be explained by the change of character of the disturbance while propagating through the system. The second controller sees a smoother disturbance than the first controller but the amplitude is larger since the control signal of the first controller (which is also the inflow to the second tank process) makes an overshoot when handling the disturbance in the uncontrolled inflow. The controllers are tuned to handle a step disturbance of 10 % but here the second and third controller see a non-steplike disturbance with an amplitude larger than 10 %, which explains that the specifications are not met. A simple approach to solve the problem for this scenario is to tune the second and third controller to handle a disturbance which is as large as the overshoot of the previous control signal in addition to the initial inflow disturbance. The allowed values of the arrest time for the three controllers with this approach is shown in Figure 4.10, where also the possible lower bound of arrest time due to constraints on the derivative of the control signal has been illustrated. Here the maximum allowed derivative of the control signal when handling step disturbances has been set to $\max |\dot{u}(t)| = 2.5$.

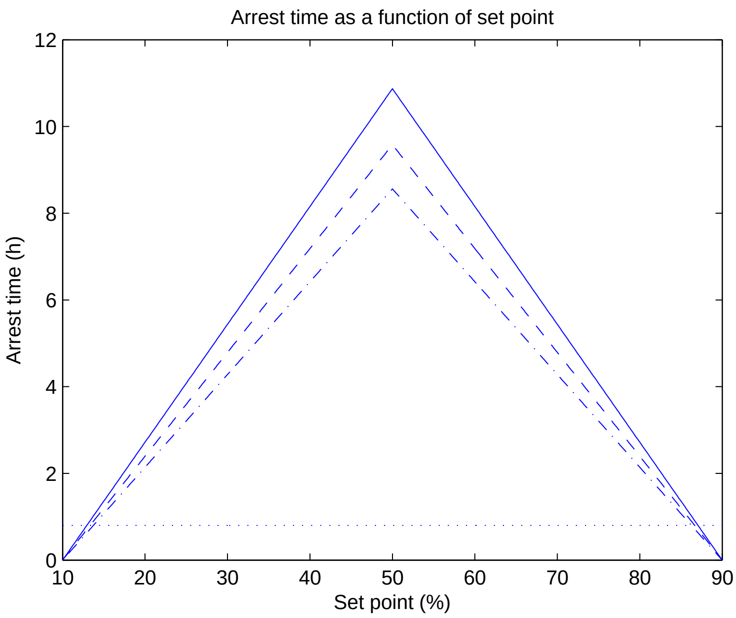

Figure 4.10: Allowed arrest times for the three controllers as a function of set point when the controllers are tuned to handle step disturbances of different amplitudes. The allowed combinations of set point and arrest time are located inside the triangles. The solid line belongs to the first controller, the dashed line to the second and the dash-dotted line to the third. The dotted horizontal line indicates the constraints on the derivative of the control signal.

When simulating the system with the maximum allowed value of the arrest time for the set point 50 % given in Figure 4.10 we get result showed in Figure 4.11.

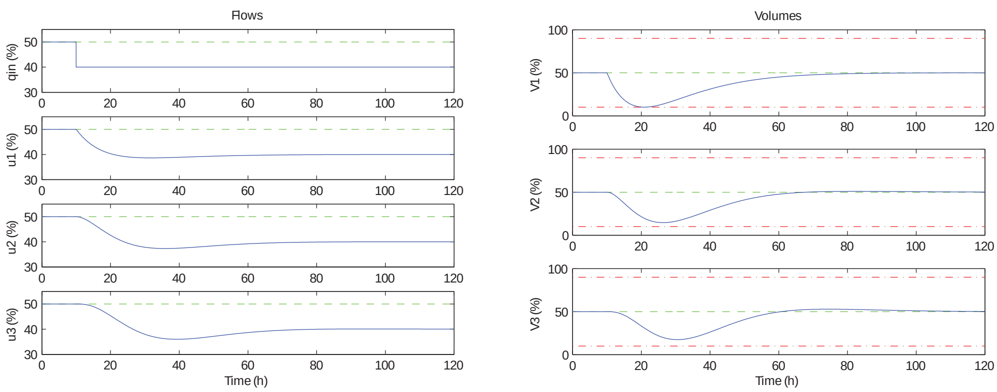

Figure 4.11: PI-control of the three-tank system with set point 50 % and T_a^1 = 10.87 h, T_a^2 = 9.58 h, T_a^3 = 8.56 h. The first unit in the production line starts running at reduced speed at time t = 10 h.

In Figure 4.11 it can be seen that with this tuning we still have a margin for the levels of tank two and three, depending on the change of character of the disturbance. Since the second and third controller do not see a step disturbance but a smoother disturbance these controllers could be slightly detuned without filling or emptying their respective tanks. If the controllers are detuned to get the levels to stay exactly within the limits it will look as in Figure 4.12.

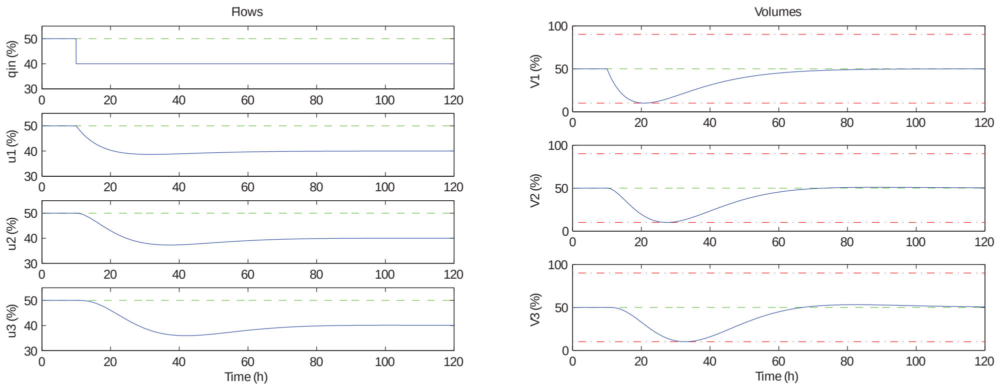

Figure 4.12: PI-control of the three-tank system with set point 50 % and T_a^1 = 10.87 h, T_a^2 = 10.60 h, T_a^3 = 10.12 h. The first unit in the production line starts running at reduced speed at time t = 10 h.

The tuning to achieve the behaviour in Figure 4.12 can be derived using transfer function analysis. The transfer function from the inflow disturbance to the control errors of the three tanks can be computed and by specifying the maximum control deviation for each of the tanks the tunings can be computed, starting with the tuning of the controller closest to the disturbance. For i tanks connected in series the Laplace transform of the control error of tank i, $E_i(s)$, can be computed as

$$E_i(s) = \frac{s\prod_{k=1}^{i-1}(K_c^k s + \frac{K_c^k}{T_i^k})}{\prod_{k=1}^{i}(s^2 + K_c^k + \frac{K_c^k}{T_i^k})} D(s) \tag{4.3}$$

when K_c^i and T_i^i are the control parameters for controller number i and $D(s)$ is the Laplace transform of the inflow disturbance to the first tank.

Unfortunately the inverse Laplace transform of this expression becomes quite complicated for the control errors in tanks far away from the original disturbance and are not easily solved by hand calculation or even by Maple. Even for tank number two in this case the computations get tricky.

Disturbance charachteristics

The problem gets even more interesting if looking at the start-up of the unit that was running at reduced speed. The length of the failure in relation to the arrest time of the PI-controllers will then affect the behaviour of the system. If the stop time is much larger than the arrest time the level of the tank will have time to almost return to the set point after the step downwards before the step upwards occurs. This means that the level will make an undershoot due to the first step and an equally large overshoot due to the second step. If the maximum arrest time for each of the controllers to keep the level within high and low limits is used the set point of the three tanks will have to be 50 % since there will be an equally large step upwards and downwards in the inflow to the first tank. The simulation result with the set points 50 % and the maximum arrest times (as in Figure 4.12) is showed in Figure 4.13. If the step upwards occurs when the level has started to go back towards the set point but not yet reached the set point there will even be scenarios when the specifications are not met because of the overshoot in the levels.

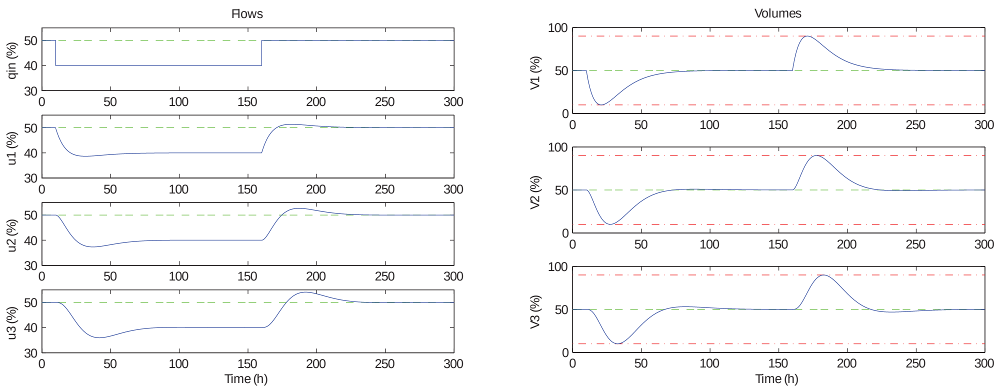

Figure 4.13: PI-control of the three-tank system with set point 50 % and $T_a^1 = 10.87$ h $T_a^2 = 10.60$ h $T_a^3 = 10.12$ h. The first unit in the production line runs at reduced speed for 150 hours starting at time t = 10 h.

If the stop time is less than the arrest time the levels will not have time to drop down to the lower limits during the stop. In this case the controllers could be slightly detuned. However, it is not trivial how to detune the three controllers in order to make the level of the first tank stay exactly within its limits. With the detuned controller a longer stop than expected will cause the level to drop below the lower limit. The behaviour of the system during a short stop is illustrated in Figure 4.14. The tuning and set point used is the same as in Figure 4.13.

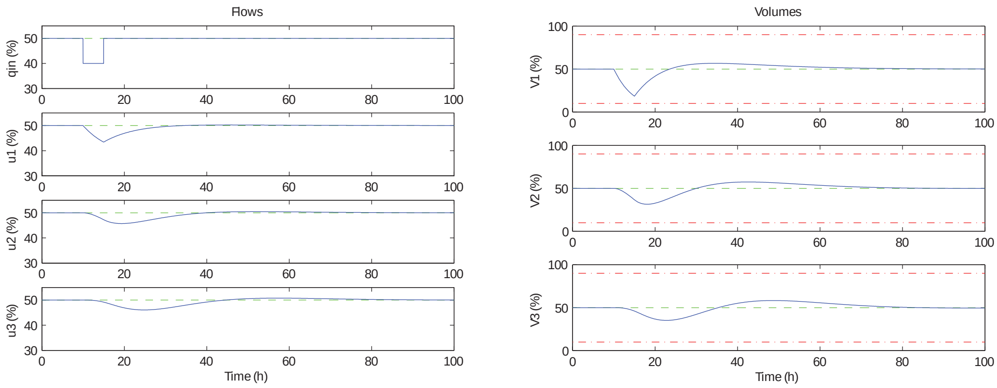

Figure 4.14: PI-control of the three-tank system with set point 50 % and $T_a^1 = 10.87$ h $T_a^2 = 10.60$ h $T_a^3 = 10.12$ h. The first unit in the production line runs at reduced speed for five hours starting at time t = 10 h.

Loss of production volume

In section 4.1 the loss of production volume due to the bottleneck being forced running at reduced speed was computed as a function of the levels in the buffer tanks. When controlling the levels with PI control the bottleneck will begin reducing its working speed immediately as the shutdown occurs and a loss of production volume will always be obtained independently of buffer tank levels. How large the loss will be will depend on the tuning of the controllers and on the duration of the shutdown. The loss of production volume is here the volume lost in the bottleneck during the time of the shutdown compared with the initial running speed of the bottleneck unit. If the shutdown is infinitely long, the levels of the tanks will have time to return to their set points and thus the loss of production volume will always be 100 % of the volume loss in the unit suffering from the shutdown, independently of the tuning of the controllers. For shorter shutdowns the production volume loss could be less than 100 % and the loss will depend on the tuning of the controllers. The theory is that the tuning for the three tanks that gives the least loss of volume for a certain duration of the shutdown is the tuning that utilizes as much of the volume in the buffer tanks as possible, i.e. the slowest allowed tuning that keeps the level between its limits. This tuning gives the smoothest possible flow response and thereby probably the least loss of production volume in the bottleneck.

4.2.2 P-control of the three-tank system

Tuning

When using P-control and tuning the three controllers to handle a step disturbance in the inflow of 10 % the process will behave as in Figure 4.15. The choice of proportional gain for the controllers to make the levels stay exactly within their limits is derived in section 3.2.2. The P-controller will give a stationary error and the level will stay at the lower limit 10 %.

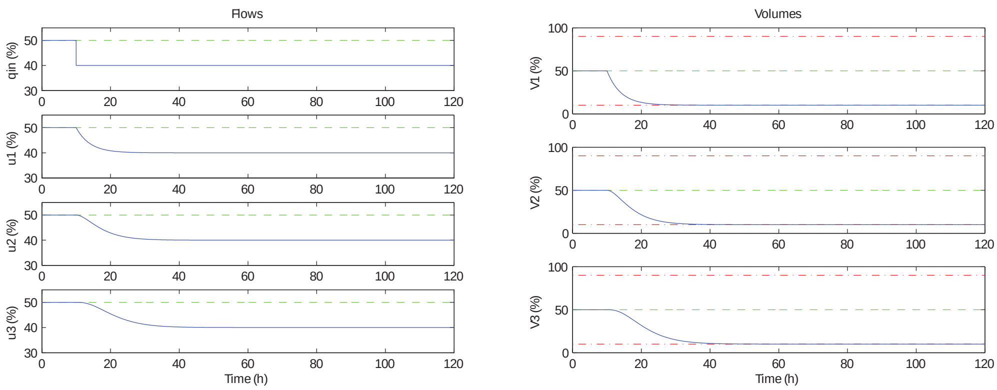

Figure 4.15: P-control of the three-tank system with set point 50 % and K_c = 0.25. The first unit in the production line starts running at reduced speed at time t = 10 h.

Since the P-controller will not give an overshoot in the control signal when handling a step disturbance the chosen tuning with the same proportional gain for each of the three controllers will make all the levels stay within their limits.

Disturbance charachteristics

If taking into account the start-up of the failing unit the system will behave as in Figure 4.16 during a long stop. The tuning used is the tuning to make the levels stay exactly within their limits.

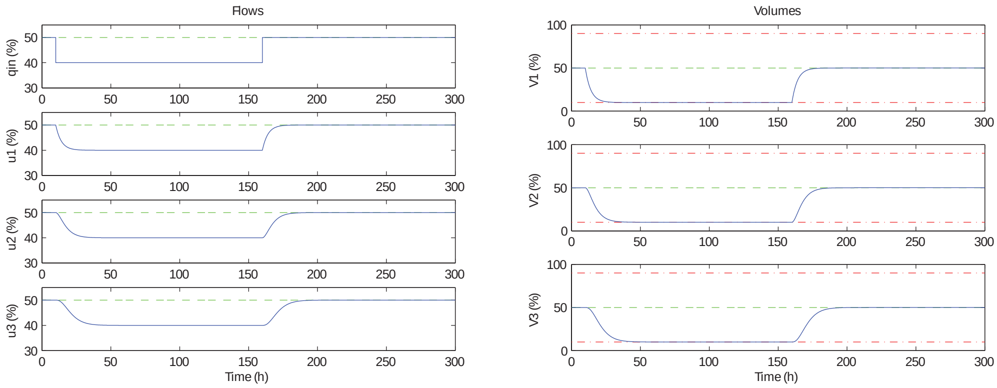

Figure 4.16: P-control of the three-tank system with set point 50 % and $K_c^1 = K_c^2 = K_c^3 = 0.25$. The first unit in the production line runs at reduced speed for 150 hours starting at time $t = 10$ h.

Here it can be seen that there will be no stationary error in the levels after the disturbance if the step disturbance upwards and downwards in the flow are equally large, which eliminates the main drawback with P-control in this case.

The P-controller will not give an overshoot in the level due to a step disturbance (shown in section 3.2.2) and thus the set point can be changed in order to be able to decrease the proportional gain and achieve a smoother control signal. If the set point is increased to 90 % as much tank volume as possible will be used. The simulation result for this scenario is shown in Figure 4.17.

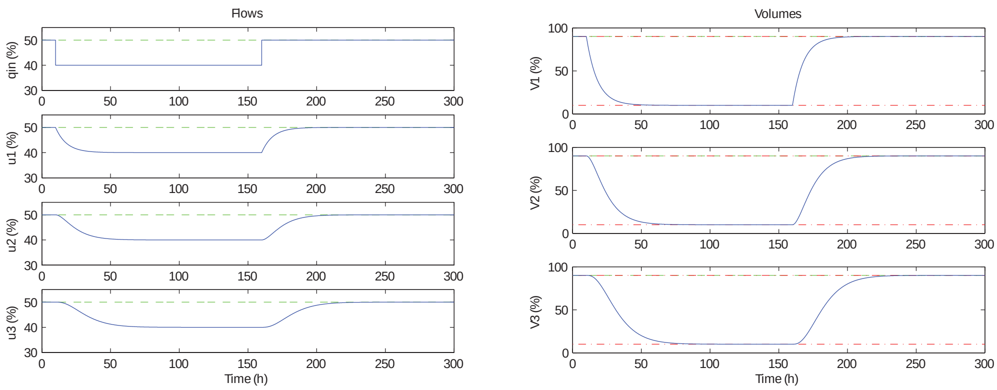

Figure 4.17: P-control of the three-tank system with set point 90 % and $K_c = 0.125$. The first unit in the production line runs at reduced speed for 150 hours starting at time $t = 10$ h.

If the stop is short the levels might not have time to drop down to the lower limits before the step upwards occurs. In this case the controllers can be slightly detuned. The drawback with detuning the controller is that a longer stop than expected will make the levels drop below their lower limits. Simulation of the system during a short stop with the same parameters as in Figure 4.17 is shown in Figure 4.18.

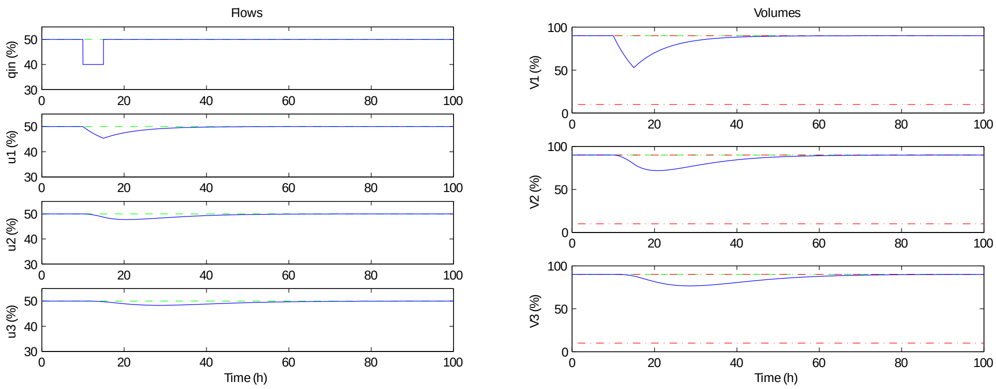

Figure 4.18: P-control of the three-tank system with set point 90 % and $K_c = 0.125$. The first unit in the production line runs at reduced speed for five hours starting at time $t = 10$ h.

Loss of production volume

In subsection 4.2.1 it was concluded that a loss of production volume due to reducing the working speed of the bottleneck is obtained immediately as the shutdown begins independently of the buffer tank levels in the system. This is true also for P-controllers. The fundamental difference between P- and PI-control is that if P-control is used a stationary error will be obtained. When using PI-control the loss of production volume will always be 100 % after an infinitely long stop since the levels will have returned to their set points. This is not the case when using P-control, where the levels never will return to the set point and the loss of production volume will be less than 100 %. The loss of production volume will increase slowly towards 100 % with the time of the shutdown.

4.2.3 Comparison of P- and PI-control of the system

In the previous subsection on P-and PI-control it has been showed that PI-control will give an overshoot due to the a step disturbance in the inflow where the flow returns to its initial value and that P-control will give no overshoot. If the objective is to have as smooth a flow as possible (averaging level control, see chapter 3.3) the tuning of the controllers should be chosen to utilize the entire volume of the tanks. The question is if P- or PI-control gives the smoothest outflow for the discussed disturbance model. To begin with, P-control and PI-control with the same set point are compared in Figure 4.19 where the disturbance characteristics is as in Figure 4.13 and 4.16.

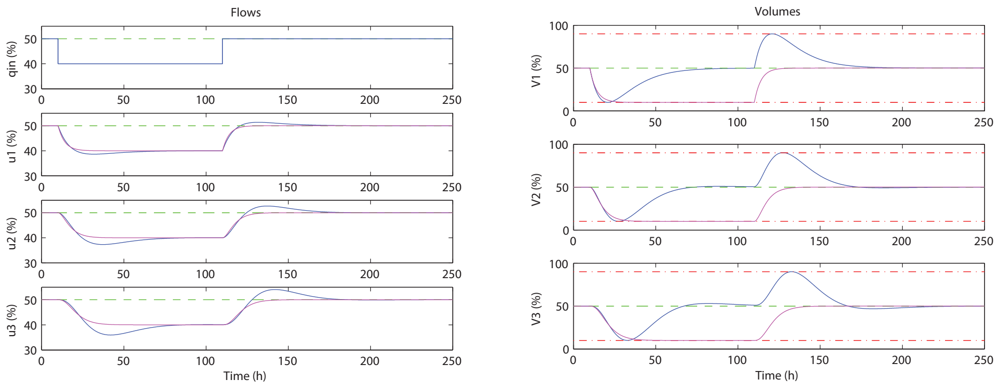

Figure 4.19: Comparison between P-control with $K_c^1 = K_c^2 = K_c^3 = 0.25$ (magenta solid lines) with the set point 50 % and PI-control with $T_a^1 = 10.87$, $T_a^2 = 10.60$, $T_a^3 = 10.12$ (blue solid lines) with the set point 50 %.

The maximum derivative of the control signal due to a step disturbance can be computed given the tuning, see chapter 3.2. The maximum derivative of the control signal for the first controller is thus $\max|\dot{u}_1(t)| = K_c k_v = 0.25$ when using P-control with the set point 50 % and $\max|\dot{u}_1(t)| = \frac{2}{T_a} = 0.18$ when using PI-control with the set point 50 % with the tuning that keeps the level exactly within the limits. The third and second controller do not suffer from a step response and since it has been showed that transfer function analysis for the three-tank system with PI-control leads to heavy computations the maximum derivatives of the control signals for the second and third controller have not been computed. It can though be seen in Figure 4.19 that with P-control the control signal seems to get smoother the further away from the original disturbance the buffer tank is located. This does not seem to be the case with PI-control.

It has been seen that the set point for the P-controllers can be increased since there will be no overshoot in the level. A comparison of P-control with the set point 90 % and PI-control with set point 50 % can be viewed in Figure 4.20. The maximum flow derivatives for the first controller is in this case $\max|\dot{u}_1(t)| = K_c k_v = 0.125$ using P-control and $\max|\dot{u}_1(t)| = \frac{2}{T_a} = 0.18$ using PI-control.

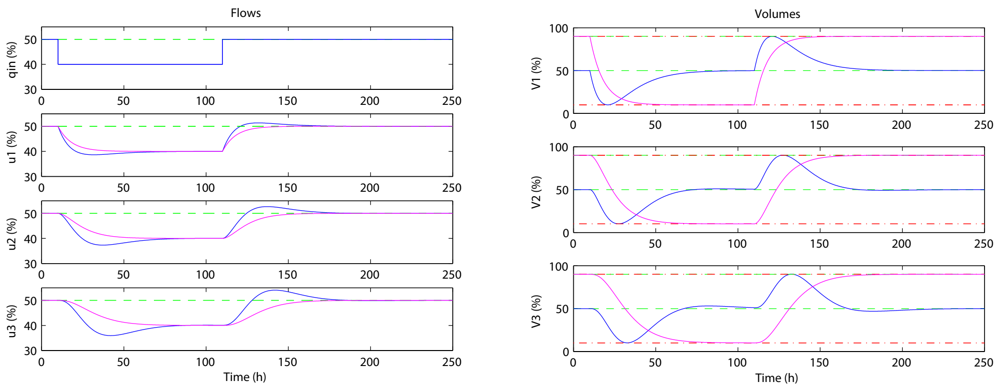

Figure 4.20: Comparison between P-control with $K_c^1 = K_c^2 = K_c^3 = 0.125$ (magenta solid lines) with the set point 90 % and PI-control with $T_a^1 = 10.87$, $T_a^2 = 10.6$, $T_a^3 = 10.12$ (blue solid lines) with the set point 50 %.

4.2.4 Conclusions

The dynamics of the simple system described in chapter 4.1 are changed when introducing automatic control and there are more parameters to consider than only the buffer tank levels. The tuning of the controllers, the set points in the tanks and duration of the stop in relation to the tuning will also affect the loss of production volume due to a shutdown. The set point is connected with the tuning and should be chosen considering probable disturbances. After choosing a suitable set point the optimal tuning in the sense of using the entire tank volume when handling a specified disturbance could be computed.

The tuning of a controller will affect downstream processes since the output from the controller will be the inflow to the next process. The character of the disturbance will depend on the tuning of the preceding controller in the system. If using PI-control the amplitude of the disturbance will increase when propagating through the system due to the overshoot in the control signal of the preceding controller, but the disturbance will also become smoother. Because of this it is not trivial to determine the tuning of the PI-controllers that keep the levels exactly within the limits; transfer function analysis leads to heavy computations even for only a few tanks. With P-control there will be no overshoot in the control signal and the same choice of K_c for all of the three controllers will guarantee that the levels are kept within the limits.

Another advantage with P-control is that the level of a P-controlled tank will never make an overshoot when a step disturbance in the flow occurs. The set points and tuning of the controllers can thus be adjusted to utilize the entire volume of the tanks yielding smooth outlet flows of the tanks. When working with real processes probable disturbances must be considered. Increasing or decreasing the set point should be done so that probable disturbances upwards and downwards are handled. If the disturbance characteristics are such that equally large disturbances upwards and downwards from the initial flow should be handled, the set point will have to be 50 % in order to meet the specifications. If on the other hand only smaller disturbances upwards should be handled the set point could be increased as in Figure 4.17.

It has been showed that when handling step disturbances the maximum derivative of the control signal is less for PI-control than for P-control if the set point is equally chosen to 50 % but if the set point for the P-controller could be increased the maximum derivative might be smaller using P-control. Furthermore, P-control seems to affect downstream processes in a more preferable way than PI-control, yielding smoother flows downstream. If the average of disturbances upwards and downwards in the flow are equally large there will be no stationary error during normal operation when using P-control. When the first controller suffers from a step disturbance where the flow returns to its initial value after a certain time and PI-control control is used the length of the disturbance in relation to the tuning has to be considered since the overshoot in the level because of the integral part might make the level go outside the specified limits.

Finally it has to be mentioned that the optimal tuning of the controllers in a system with level-controlled tanks connected in series depends on the purpose of the buffer tank. In this chapter it has mainly been discussed how to minimize flow variations in a such system and P-control has here shown to have good performance. However, if the objective is to keep rather tight level control even if large disturbances occurs it might probably be preferable with PI-control to avoid the risk of ending up with a stationary error in the level.

Chapter 5

Case studies

This chapter contains information about Perstorp AB as a company (section 5.1) and the case studies made at Perstorp AB. Two real cases are studied; the first one is found in Warrington in the United Kingdom (section 5.2) and the second in Singapore (section 5.3).

The simulation figures in this chapter will look as in Figure 5.1. The blue solid line marks the simulated parameters, the magenta solid line data from real operation, the green dashed line the set point for the controller and red dash-dotted lines mark the critical limits between which the level should be kept.

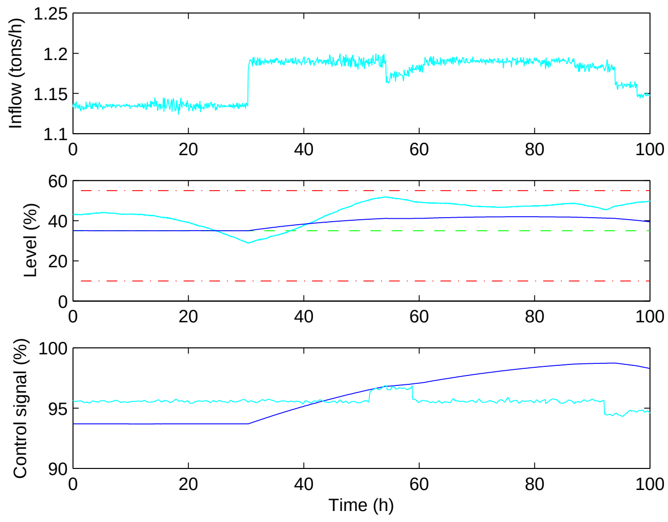

Figure 5.1: Example figure for the simulations in the chapter.

5.1 Perstorp AB – the company

The history of Perstorp AB begun in 1881 when young engineer Wilhelm Wendt decided to start producing acetic acid, and charcoal under the company name Stensmölla Kemiska Tekniska Industri. The acetic acid production was a success and the company was soon renamed Skånska Ättiksfabriken (”Scania acetic acid factory”). As a result of the success the company also started producing its own bottles for the acetic acid. Today only bottling of the acetic acid is carried out in Perstorp. Already in the beginning of the company history Perstorp’s policy to maximize refinement and utilization of raw materials and minimizing the quantity of waste products was founded. Wilhelm Wendt managed to make the production more efficient and in the process he discovered that there was a possibility to reuse some of the waste products to develop new products.

A major breakthrough in the company’s history came in 1907 when the company managed to produce methanol from beech wood and refine it to formaldehyde (formalin). Formaldehyde turned out to be a useful raw material for many other processes and is still an important product at Perstorp AB. As the years went by the range of formaldehyde-based chemicals produced by Skånska Ättiksfabriken increased rapidly and in 1917 the company also starts producing plastics, which makes them the first company in Scandinavia entering this industry.

The next milestone in the history of the company was reached when Skånska Ättiksfabriken started manufacturing laminate. After the Second World War there was a demand for modernisation and renewal in Sweden and the laminate ”Perstorps-Plattan” became an astonishing success. The company continued to expand rapidly and in 1955 the first international laminate production began in Brazil. At this time the production of polyalcohols from formaldehyde was developed and pentaerytritol and trimethylprophane became important products for the paint industry. These chemicals are today still produced in large amounts at Perstorp AB. A few years later the company changed its name to Perstorp AB and was listed on the Stockholm Stock Exchange. This was followed by an extensive local and international expansion in the 1970’s and 1980’s and Perstorp became a well-known company within the chemical industry. The laminate flooring Pergo was introduced and soon became a global success.

The wide range of products and the rapid expansion eventually made it necessary to concentrate the production to fewer areas to maintain the good quality and cost-effectiveness. Pergo and the plastic division were sold and the focus was concentrated on the specialty chemicals. Today the main products are alcohols, polyalcohols and acids. Perstorp AB is now controlled by the French company PAI Partners and has about 2700 employees and production in 12 countries all over the world, see Figure 5.2.

In this thesis process sections of the plants in Warrington in the United Kingdom and in Singapore are studied. In Warrington caprolactones (Capa) are produced, with its main applications within the plastics industry. Capa thermoplastics have also become a useful

Figure 5.2: Perstorp sites around the world.

component in the shoe industry. In Figure 5.3 the Capa plant can be viewed. Because of high corrosivity of some substances in the plant some process sections in the plant are entirely made of glass. A picture from such a process section is showed in Figure 5.4.

The information about Perstorp AB in this chapter is collected from [14] where also further information about the company and their products is available.

Figure 5.3: The Capa plant in Warrington.

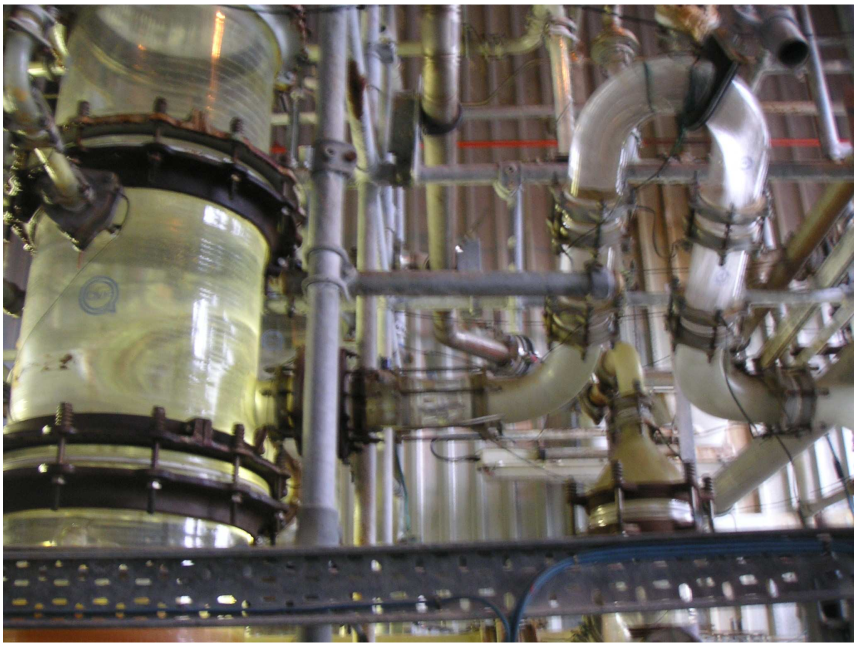

Figure 5.4: Picture from a process section made of glass.

5.2 Warrington

The process section studied in the Warrington plant produces peracetic acid (PAC) from 70 % hydrogen peroxide (H_2O_2). The main objective is to increase the average production rate of the plant by increasing availability of the presumed bottleneck of the plant, alternatively to determine where the bottle neck really is located. This is done by introducing level control in one of the buffer tanks to prevent the level from going outside specified limits.

5.2.1 System description and problem formulation

The simplified model of the process section of the plant in Warrington that has been studied is shown in Figure 5.5. None of the tanks in the process section where level controlled as this study began.

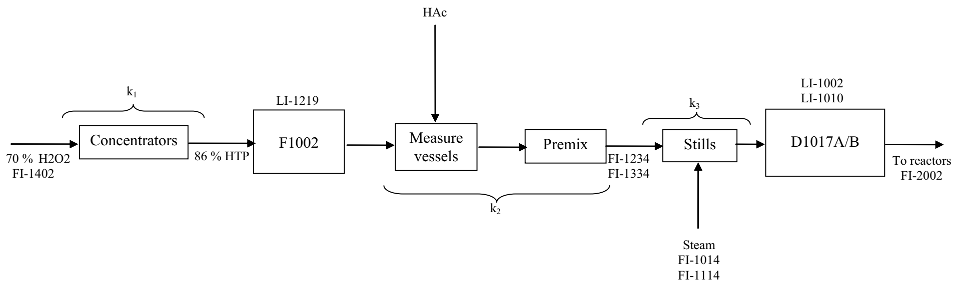

Figure 5.5: A simple model of a the studied process section of the plant in Warrington

The section includes two buffer tanks, F-1002 and D1017A/B, containing High Test Peroxide (HTP) and peracetic acid (PAC) respectively. Every process unit in between these two tanks has been approximated with a constant ratio to the incoming flow based on data from normal operation. For the concentrators no measurements of the outflow are available and thus the factor k_1 has been computed analytically considering the amount of water removed in the concentration process. A table was made with ratios from one substance to another (table 5.1) when working with mass flows. This table could be useful to get a sense of the size of the buffer tanks compared to their contents. The volume of F-1002 is 10 m^3 and the volume of D1017A/B is 17 m^3.

	$\mathbf{H_2O_2}$	**HTP**	**Steam**	**PAC**
$\mathbf{H_2O_2}$	1	0.65	1.94	3.19
HTP	1.53	1	2.97	4.90
Steam	0.52	0.34	1	1.65
PAC	0.31	0.20	0.61	1

Table 5.1: Conversion table for the studied process section of the plant in Warrington.

The master flow of this section of the plant is the incoming flow of 70 % hydrogen peroxide which is then concentrated to 86 % High Test Peroxide (HTP). HTP at these high concentrations is extremely explosive and therefore a lot of safety switches control the operation of the concentrators. One of them redirects the flow of 86 % hydrogen peroxide if the level of F-1002 rises above 55 %. If it is true that the concentrators are bottlenecking the process section, a redirect is of course a loss of production of the entire plant. The suggestion to minimize these redirect periods is to introduce level control in F-1002 by manipulating the steam flow to the stills (FI-1014 and FI-1114) to ensure that the level stays within 10 % to 55 % of the tank volume. This level control is possible since a certain amount of steam to the stills will give a certain outflow of F-1002 (see table 5.1). A larger steam flow will cause the stills to work faster and thus increase the outflow of F-1002. A picture of F-1002 is showed in Figure 5.6 and the radar level sensor of the tank in Figure 5.7.

Figure 5.6: The tank to be controlled in Warrington.

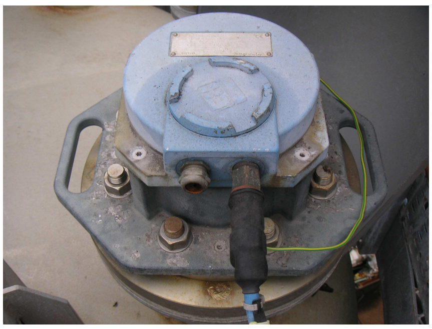

Figure 5.7: The radar level sensor of the tank in Warrington.

To get a suitable tuning of the level controller probable disturbances and constraints on the control signal must be defined. By studying data from the operation during the year of 2008 it has been found that a possible disturbance is a set point change in the feed to the concentrators. By evaluating the size of these changes upwards and downwards during normal operation and translating the disturbance from feed to concentrators to percent of the manipulated variable it has been found that a step disturbance of 2.97 % of the manipulated variable upwards and of 5.70 % downwards should be considered. Another possible disturbance is variations in the concentration of the incoming hydrogen peroxide. However, a variation of 1 % of the concentration only gives a disturbance of 1.14 % of the manipulated variable, which means that if set point changes in the feed to the concentrators are handled by the controller, so will the variations in the concentration. The disturbances due to redirect periods are not considered since a good controller will ensure that the level of F-1002 never exceeds its specified limits and consequently no redirects will occur.

The maximum allowed amount of steam to the stills gives a constraint on the control signal. There could also be specifications on the derivative of the control signal. During the year of 2008 the steam flow has not been changed with a higher rate than 2.65 h^{-1} and thus the constraints on the derivative of the control signal is suggested to be specified as $\max|\dot{u}(t)| =$ 2.65 h^{-1}. The rate has been computed as $\max|\dot{u}(t)| = \frac{\Delta u/\Delta t}{u_{max}}$ which explains the unit [h^{-1}].

5.2.2 Tuning

To ensure that the level never goes outside its specified limits and that the controller never gives a control signal with higher derivative than allowed if a step disturbance occurs a PI-controller can be tuned using T_a-tuning. In section 3.2.3 it is derived that the maximum deviation due to a inlet flow step disturbance of C % is $e_{max} = 0.3679 k_v T_a C$. In this case the maximum tolerated deviation depends on the set point. If for example the set point is 30 % we can tolerate a positive deviation, C_{up}, of 25 % and a negative deviation, C_{down}, of 20 %. This can generally be expressed as:

$$e_{max}^{up} = 55 - SP \tag{5.1}$$

$$e_{max}^{down} = SP - 10 \tag{5.2}$$

The allowed values of T_a is then

$$T_a^{up} \leq 2.718 \frac{e_{max}^{up}}{C^{up} k_v} \tag{5.3}$$

$$T_a^{down} \leq 2.718 \frac{e_{max}^{down}}{C^{down} k_v} \tag{5.4}$$

It is also derived in section 3.2.3 that the maximum allowed rate of change of the manipulated variable will give a lower bound on T_a according to:

$$T_a \geq \frac{2}{\max|\dot{u}(t)|} \tag{5.5}$$

In Figure 5.8 the arrest time as a function of set point is illustrated. The allowed values of the combinations of arrest time and set point are located inside the triangle given by the three lines. The optimal choice of arrest time and set point among these allowed combinations depends on the probability of even larger or more frequent disturbances than the specified versus the wear on the equipment and the effect of downstream processes when having rapid changes in the control signal. In this thesis a set point of 35 % and an arrest time of T_a = 55 h with $T_i = T_a$ is suggested. The choice is marked with a green cross in Figure 5.8.

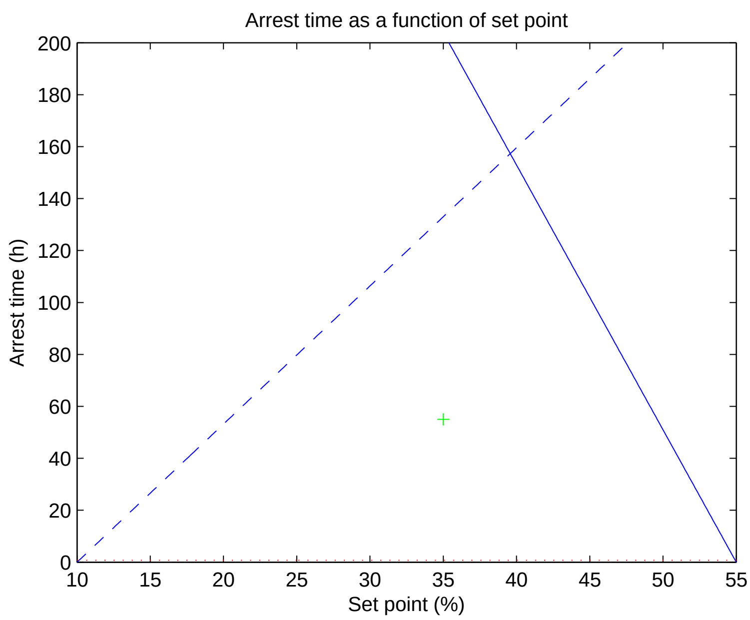

Figure 5.8: Allowed combinations of arrest time and set point for the controller of F-1002. The blue solid line is given by the specifications on disturbances upwards, the blue dashed line on disturbances downwards. The hardly visible red dotted line at the bottom of the triangle illustrates the specifications on the derivative of the control signal and the green cross marks the suggested choice of arrest time and set point.

5.2.3 Simulation

The process section under study was built up in Simulink (see Figure 5.9) and simulated using constants or data from normal operation as input to the first tank and as output of the second tank. A possibility to simulate a step disturbance in the inflow has been included.

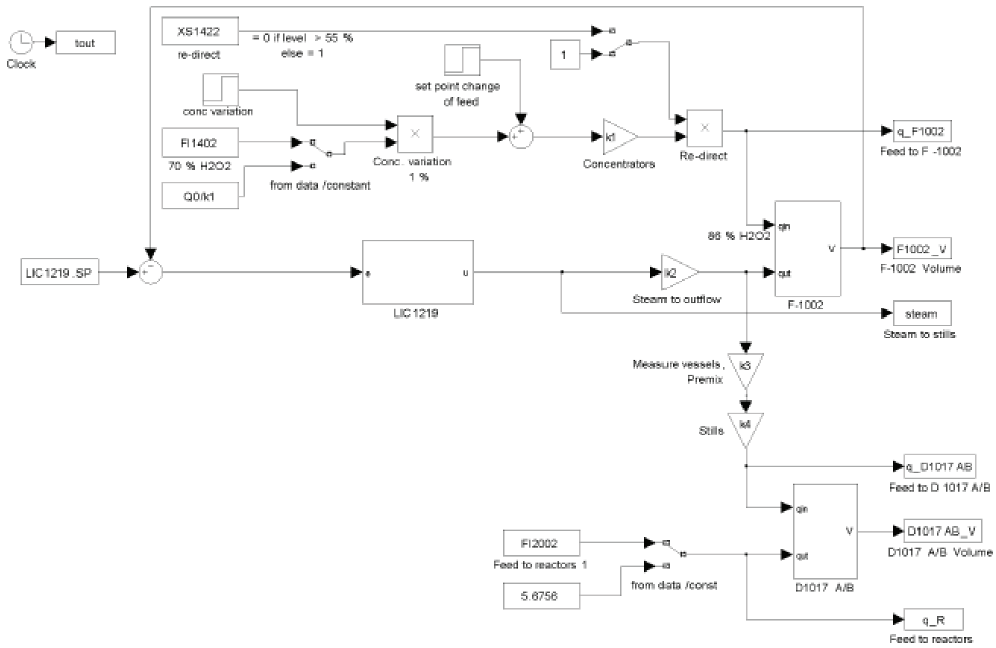

Figure 5.9: Simulink model of the studied process section in Warrington.

At first a verification that the computed tuning handles the specified disturbances was made using the maximum allowed arrest time for the set point 15 %, T_a = 26.6 h. The simulation was made using data from real operation during a time period with no major disturbances as input to get a sense of how the measurement noise affects the controller. Furthermore, a step of the specified size downwards in the inflow was simulated at time t = 20 h. The result of the simulation is shown in Figure 5.10.

If looking closely at Figure 5.10 it can be seen that the level drops slightly below the lower critical limit. Except for this the controller behaves as expected. The fact that the specifications are not entirely met can be explained by fact that the controller is tuned to handle *step disturbances* of the specified size. The small and rapid variations (noise) in the inflow will make the level drop slightly below the critical limit. This is a motive for choosing a somewhat faster tuning than the slowest allowed tuning for the chosen set point. To get a sense of the different dynamics when using the slowest allowed tuning and the suggested tuning T_a = 55 h with $T_i = T_a$ for the set point 35 % a simulation was made with stepwise constant inflow to the tank. The result is showed in Figure 5.11 and 5.12.

When simulating the system using input data from a period with real disturbances in the inflow the suggested controller behaves as can be seen in Figure 5.13. Here data from the

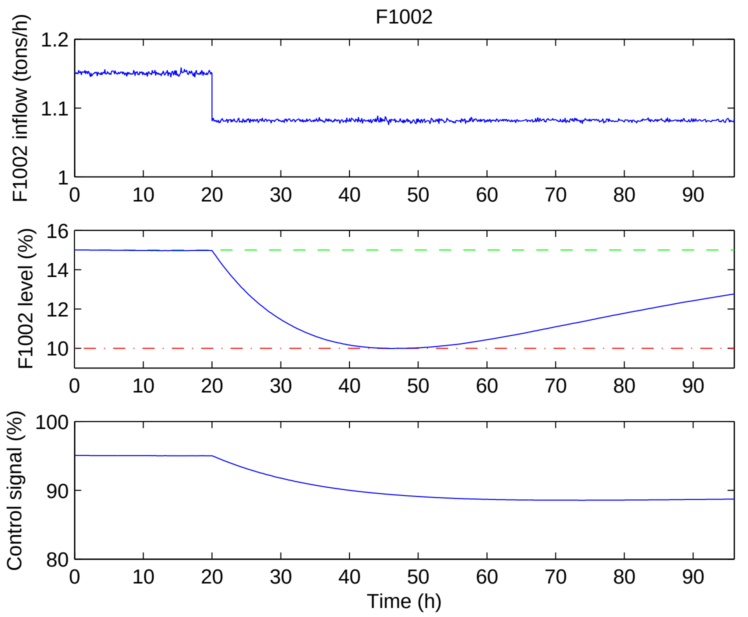

Figure 5.10: Inflow, level and control signal for level control of F-1002 with the slowest allowed tuning for set point 15 %: T_a = 26.6 h. Inflow simulated with data from a time period with no major disturbances with a simulated step at t = 20 h.

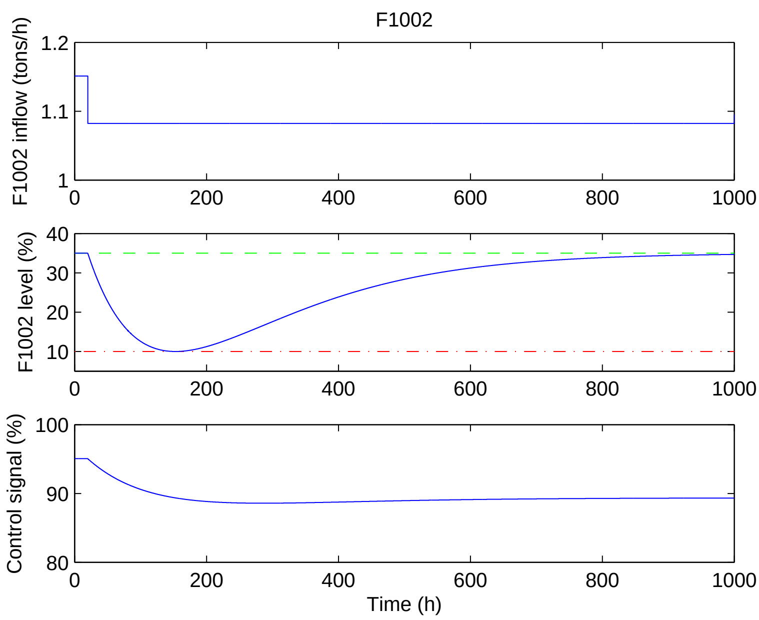

Figure 5.11: Inflow, level and control signal for level control of F-1002 with set point 35 % with the slowest allowed tuning, T_a = 133 h. Simulated step at time t = 20.

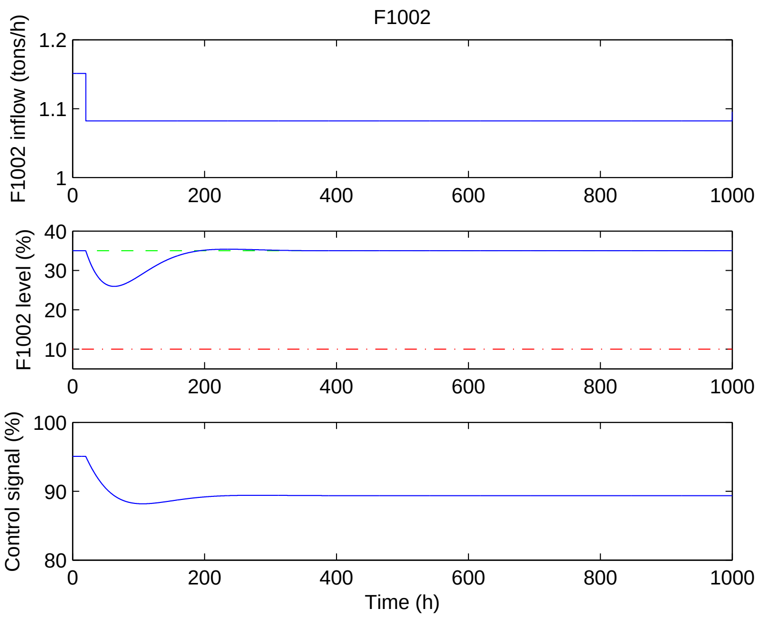

Figure 5.12: Inflow, level and control signal for level control of F-1002 with set point 35 % and the suggested tuning T_a = 55 h with $T_i = T_a$. Simulated step at time t = 20.

manual operation during this period has been included for comparison.

5.2.4 Results and conclusions

When beginning the implementation of a controller in a plant, new facts about the process often emerge. During the visit to the plant in Warrington it was discovered that the volume 10 m^3 corresponding to 100 % of F-1002 that was used in the simulations was really only 2 m^3, since 0–100 % in the measurements only corresponds to part of the physical tank. This fact requires a faster tuning of the level controller and even questions whether or not the tank actually can be regarded as a buffer tank. Another issue that was discussed was the actual concentration of the incoming hydrogen peroxide. Lab results available at the plant confirmed that this concentration varies around approximately 70.5–72.5 %, i.e. the concentration is almost never below 70 % but rather a few percent above 70 %. According to the operators at the plant it would even be difficult running the plant at the present speed today if the concentration was actually 70 %. The effect when comparing the simulations with the real operation of the plant will mainly be that the ratio k_1 will change over time. This fact can actually be seen in Figure 5.13 where it seems like there is an "offset" in the simulated control signal compared to the corresponding data from the period. The ratio k_1 was computed assuming that the incoming hydrogen peroxide concentration was 70 % whereas the actual concentration during this period was higher, giving an offset between the

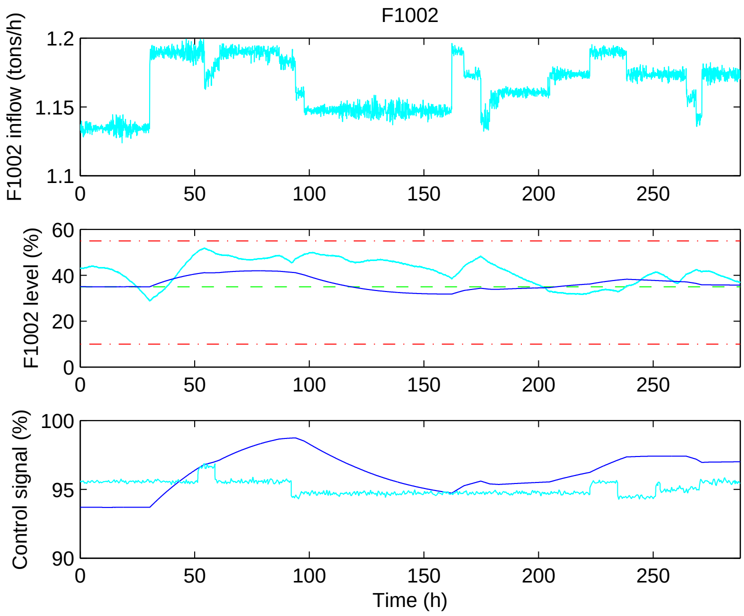

Figure 5.13: Inflow, level and control signal for level control of F-1002 with set point 35 % and the suggested tuning T_a = 55 h with $T_i = T_a$. Inflow simulated with data from a time period with real disturbances in the inflow.

simulated and measured control signal in steady state.

A level controller for F-1002 was implemented in Warrington on the 28th of October 2008. Data from the plant after the implementation can be viewed in Figure 5.14.

As can be seen in Figure 5.14 the set point is chosen as high as 45 %. The reason for this is that the operators want to have as much contents of F-1002 as possible to be prepared for a shutdown in the concentrators. If a concentrator shutdown occurs the reactors and the stills should be able to run at full speed for as long as possible before having to shut down the units due to empty buffer tanks before these units. If considering the flow through the bottleneck as the definition of production this might seem strange when comparing to the conclusions for the simple system discussed in section 4.1. If following the suggestions from this section it would be preferable to have a low level in F-1002 so that the concentrators can run at full speed for as long as possible if a shutdown occurs after this unit.

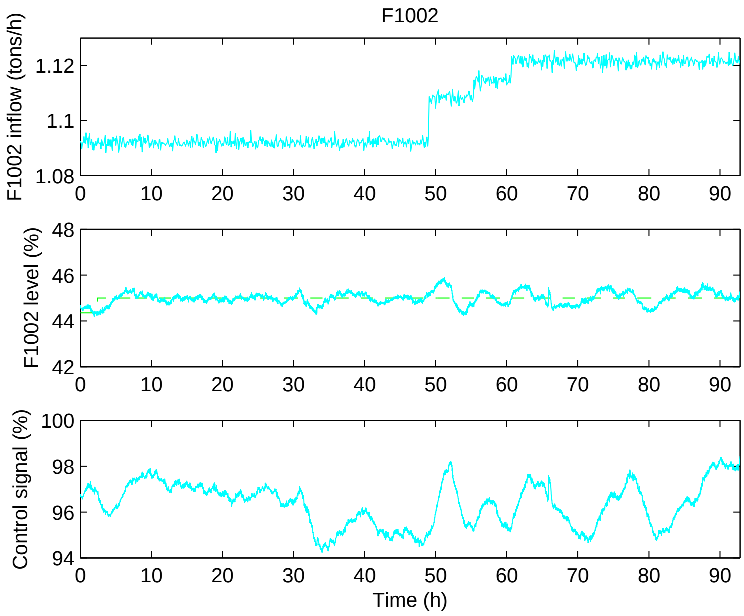

Figure 5.14: Data for the level control of F-1002 in November 2008.

However, if there are units with long start-up times (here the stills and the reactors) the optimal choice of buffer tank levels will not be as trivial as in the simple case. Long start-up times for the stills and the reactors will require a high level of F-1002 and the bottleneck analysis will require low level of the tank. The optimal level will be a weighting of the above factors. If the capacities of the bottleneck unit and the other units in the production line do not differ much this must also be taken into consideration. A first step in determining the optimal level of F-1002 is to evaluate the effects of shutdowns in the different units. With the ratios stated in table 5.1 the tolerated stop time before having to shut down a unit due to empty buffer tanks can be computed given the buffer levels at the time for the shutdown, the minimum allowed buffer tank levels and the desired rate at which the working units should be run. Alternatively the speed at which the plant can be run without having to shut down a unit can be computed given the planned duration of the shutdown. This has been presented as an Excel sheet considering shutdown of the reactors and the concentrators. An example from the Excel sheet is showed in table 5.2.

The high set point of the level controller used today will demand a faster tuning to keep the level below its high limit 55 %. Today the level is very tightly controlled with K_c = 1.3 and T_i = 55 min compared with the suggested controller in this thesis with K_c = 0.41 and T_i = 55 h. A motivation for keeping the level control tight in the first implementation is to build up a trust for the controller among the operators. If the level varies much when disturbances occur the operators will probably regard the controller as hazardous and run the level control manually which gives no use of the level controller. The large difference between the suggested control parameters in this thesis and the parameters for the controller

Concentrator trip

Compute maximum PAC feed given the down-time of the concentrators					
Estimated down-tim e (h)	Current F-1002 Level (%)	Curren t D-1017A/ B Level (%)	Minimum allowed level F-1002 (%)	Minimum allowed level D-1017A/B (%)	Maximum PAC fee d (tons/h)
1	45	50	20	10	10,92

Compute the tolerated down-time of the concentrators given the desired PAC fee d					
Desire d PAC feed (tons/h)	Current F-1002 Level (%)	Curren t D-1017A/ B Level (%)	Minimum allowed level F-1002 (%)	Minimum allowed level D-1017A/B (%)	Tolerated down-tim e (h)
5,2	45	50	20	10	2,10

Table 5.2: Example from Excel sheet where allowed stop times for different units are considered. In the left column the estimated down-time or the desired PAC feed could be entered. The shadowed boxes contain parameters to be read from the data collecting system and the computed maximum PAC feed or the maximum tolerated down-time is given in the right column.

implemented in Warrington also to a certain extent depends on different methods used for determining the process speed gain, k_v (see section 3.1). In this thesis k_v was computed as $k_v = \frac{1}{filltime}$ using the volume 10 m^3 of F-1002. When implementing the controller a step response was made and an identification of the tank process parameter k_v performed.

However, the implemented level controller behaves well so far. Further improvements could easily be made when the controller has been in use for some time and the behaviour of the controller when suffering from disturbances can be evaluated. The largest gain from the level controller might be that it takes us one step closer to determining the actual bottleneck of the process section.

5.3 Singapore

This is a censored version and thereby details on the process are not included.

The plant in Singapore studied in this thesis is producing a Polycarboxylic acid. The plant has a few typical buffer management problems which will be discussed in this chapter. A control strategy is developed for averaging level control in one of the buffer tanks and some further suggestions for better buffer management are made.

5.3.1 Problem formulation

A simple sketch of the system can be viewed in figure 5.15.

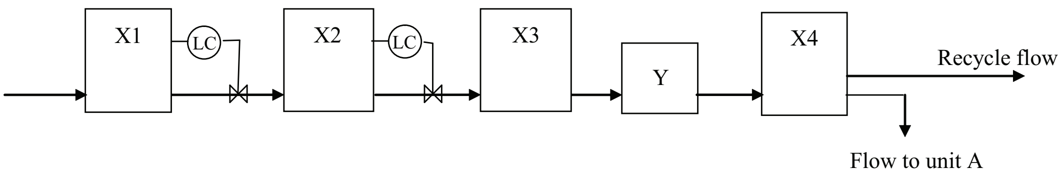

Figure 5.15: Model of process section to be studied.

The unit here called unit Y has to be shut down for maintenance frequently which causes the main problems in the production line. During a shutdown of Y the flow to this unit is bypassed back to the tank before unit Y, X3, and the level in tank X4 will drop rapidly due to the large disturbance in the inflow of the tank. If the level in this tank drops down to the lower limit of the tank, the outlet flow has to be reduced to be less or equal to the incoming flow in that moment which could lead to large variations in the outlet flow of the tank. When controlling the level of X4 there will be a choice whether to have tight level control of the tank or to make the outflow of the tank smooth. Today the controller is run in manual mode during a shutdown of unit Y and the outflow is decreased drastically immediately as the shutdown begins, resulting in a minor level drop with the price of poor smoothness of the outlet flow.

From X4 there is a recycle flow to another tank in the production line. The recycle flow consists mainly of a certain chemical, here called C, to be reused in the reaction. The ratio of C and a certain substance, here denoted X, is important to keep constant to maximize the yield of the chemical reaction in the tank where the recycle flow is directed. If there are variations in the recycle flow from X4 the amount of new C to the tank where the reaction takes place is manipulated to keep the ratio constant. Large and rapid variations in the recycle flow will lead to an incorrect X/C-ratio and thereby poor yield of the reaction, which motivates the definition of a buffer management strategy to minimize flow variations from X4. When decreasing the recycle flow in a controlled way it would also be possible to predict the amount of new C needed where the reaction takes place. Another reason for keeping the recycle flow smooth is to prevent the disturbance from propagating through the system (see chapter 4.2).

As can be seen in Figure 5.15 there are two outflows from X4. One of them is a recycle flow, the other a flow to a production unit which is very sensitive to flow variations, here called unit A. The flow to unit A is even more important than the recycle flow to keep smooth and today the disturbance when a shutdown of unit Y occurs is mainly handled by the recycle flow.

Except for the smoothness of the outlet flows from X4 there are other buffer management problems to be solved in the plant. One of them concerns the buffer management in the tanks before unit Y during a shutdown of this unit. When a shutdown of unit Y occurs, the flow to unit Y is redirected back to the last tank before unit Y, X3. If this tank becomes full a choice must be made of where to direct the flow. Today the set point of the second tank before unit Y, X2, is often slightly increased by the operators to take care of a fraction of the flow and the remaining flow goes directly to X4. Unit Y will consequently be bypassed and the quality of the end product could be detoriorated.

In this thesis level control of X4 will be handled in detail in the simulation subsection. For the buffer management problem in the tanks before unit Y a suggestion is made in section 5.3.3. This suggestion is not evaluated by simulations in the thesis.

5.3.2 Simulation

A model was built up in Simulink for the simulation of the level control of X4. The level of the tank is constrained to stay within 30 % to 90 % of the tank.

The main disturbance the process section suffers from are the shutdowns of unit Y for maintenance. This gives a disturbance of approximately 30.6 % of the manipulated variable. Other present disturbances are very small in comparison to the disturbances caused by these shutdowns and consequently only the disturbances due to shutdowns of unit Y will be considered. A shutdown of unit Y results in a step downwards in the feed to the unit from the current value to approximately 2.5 tons/h. The shutdown takes approximately 25 minutes and after this time there will be a step upwards in the feed to unit Y, back to the value of the flow before the shutdown.

If the controller is tuned using T_a-tuning to handle the disturbance in the inflow due to a shutdown of unit Y without letting the level go outside its specified limits we get the allowed values of the combination of arrest time and set point under the line in Figure 5.16.

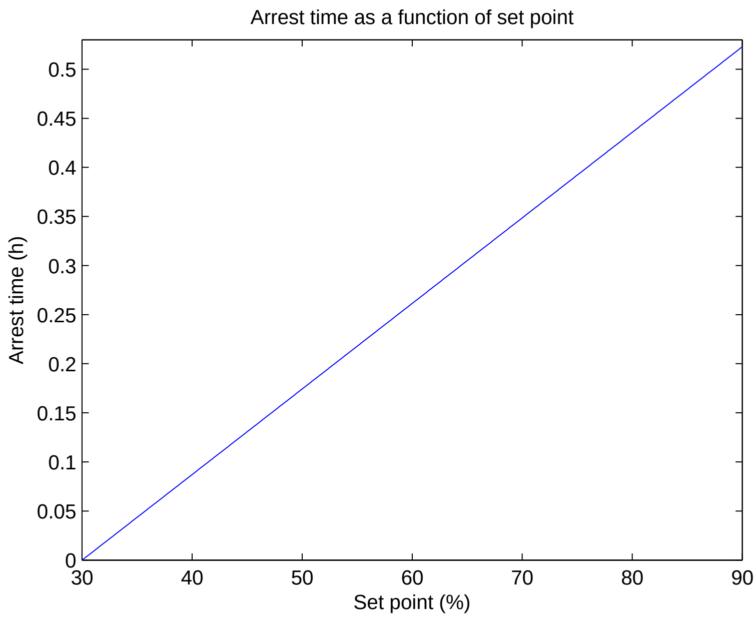

Figure 5.16: Allowed combinations of arrest time and set point.

A problem with using PI-control to solve this buffer management problem is that the specified disturbance here actually is two disturbances; at first one downwards in the inflow to X4 and after 25 minutes a similar disturbance upwards in the inflow. Assume the PI-controller is tuned to handle the first step disturbance using the entire volume of the tank (consequently with the approach in Warrington, section 5.2). When the step upwards occurs the integrator of the controller still works on bringing the level back from its minimum value to the set point, resulting in an overshoot in the level of the tank. The larger the integral part is, i.e. the faster the integral time is, the larger overshoot is obtained. This means that the overshoot will be larger for T_a-tuning with $T_i = T_a$ than for $T_i = 2T_a$. If PI-control is used we will thus get a constraint on how high the set point can be chosen if the level should stay inside the specified limits. With the set point 70 % and the maximum tuning that makes the level stay within the lower limit we get the behaviour shown in Figure 5.17. An observation here is that if the shutdown of unit Y takes longer time than expected the overshoot in the level might go over the high limit of the tank. To ensure that this never happens independently of how long time the shutdown takes the set point must be decreased to 50 % and the arrest time of the controller also must be decreased. For a more thorough description of this scenario, see section 4.2.1.

If we want to eliminate the overshoot in the level we can diminish the integral part or simply use P-control instead of PI-control. With P-control the level will never make an overshoot and we can raise the set point. For more information about P- and PI-control for these kind of disturbances see section 4.2. In Figure 5.18 the system is simulated with P-control with the set point 85 % and the maximum allowed gain (for derivation see section 3.2.2). If the

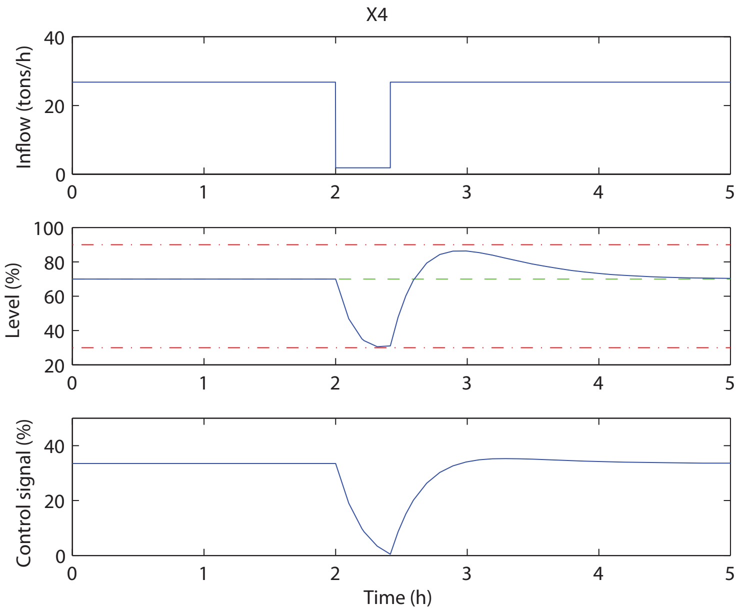

Figure 5.17: Simulation of the level control in X4 during a shutdown of unit Y with PI-control with T_a = 0.35 h and $T_i = 2T_a$.

controller is tuned in this way the level will never go below or above the specified limits independently of how long the duration of the shutdown is. In Figure 5.18 it can also be seen for this duration of the shutdown the level never has time to drop down to the lower limit of the tank. The gain of the controller can thus be decreased slightly to utilize the entire volume of the tank. However, this is not recommended since there will then be no margin for longer durations of the shutdowns than 25 minutes.

Figure 5.19 and 5.20 show the simulation of the system with P-control using real inflow data. A comparison to the way the tank is controlled today is also made. Here the outflow of the tank has been divided into its two components, the recycle flow and the flow to unit A. Since unit A is most sensitive to flow variations the approach that has been chosen is to let the recycle flow take care of as much as possible of the flow variations. Not until the recycle flow is decreased to zero the flow to unit A will be used.

Since the shutdowns of unit Y give two equally large disturbances, one upwards and one downwards, no stationary error will build up. However, if for some reason the average of disturbances upwards and downwards are not equally large there will be a stationary error in the level of the tank. A strategy to solve this could be either to introduce an integral part between the shutdowns or to use gain scheduling and increase the gain of the proportional controller when the level goes over or under certain limits. With gain scheduling the controller can be tuned to handle the normal operation including the shutdowns of unit Y using the volume 30 - 90 % of the tank and if an unusual disturbance occurs and the level goes

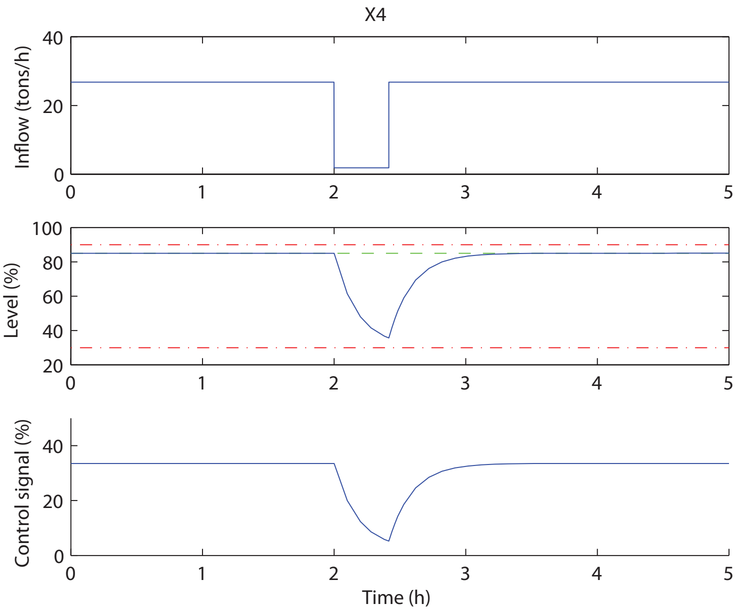

Figure 5.18: Simulation of the level control in X4 during a shutdown of unit Y with P-control with $K_c = 0.57$.

outside these limits the controller will have a higher gain to prevent the tank from being emptied or filled. The controller gain as a function of the level in the tank could look as in Figure 5.21.

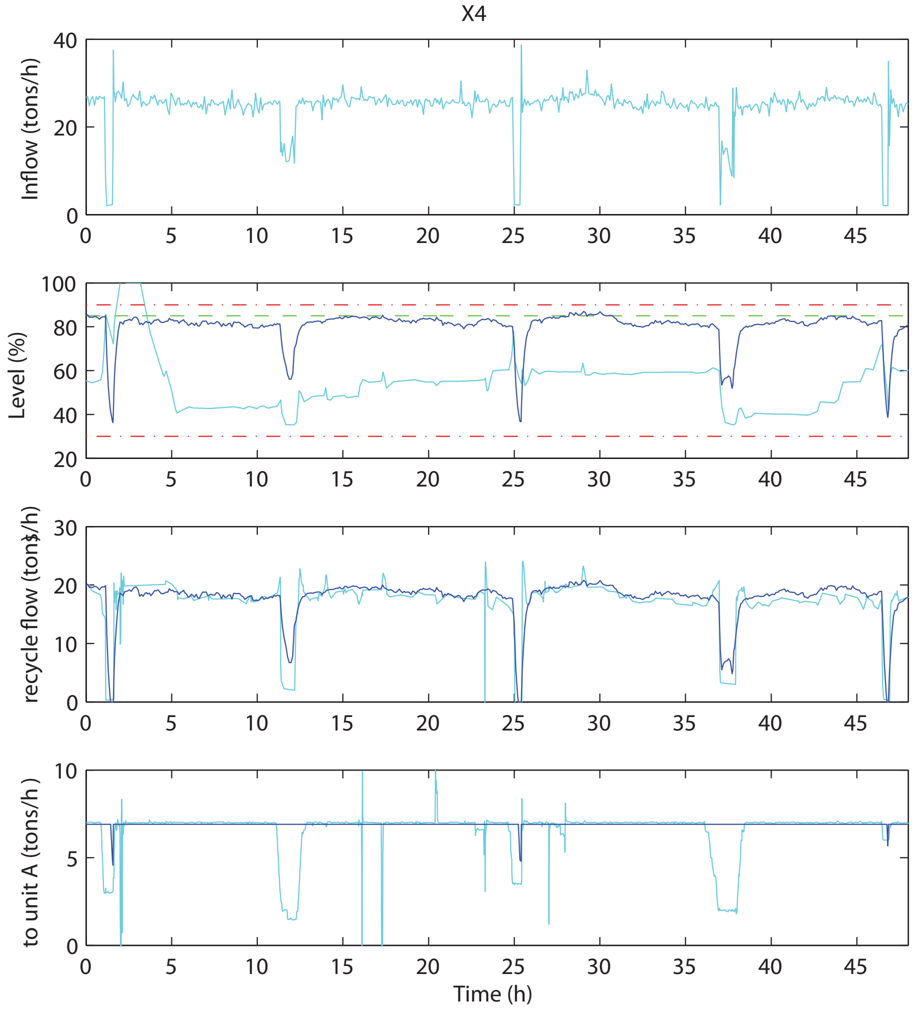

Figure 5.19: Simulation of the level control in X4 during two days with the suggested P-controller with $K_c = 0.57$ and the set point 85 %.

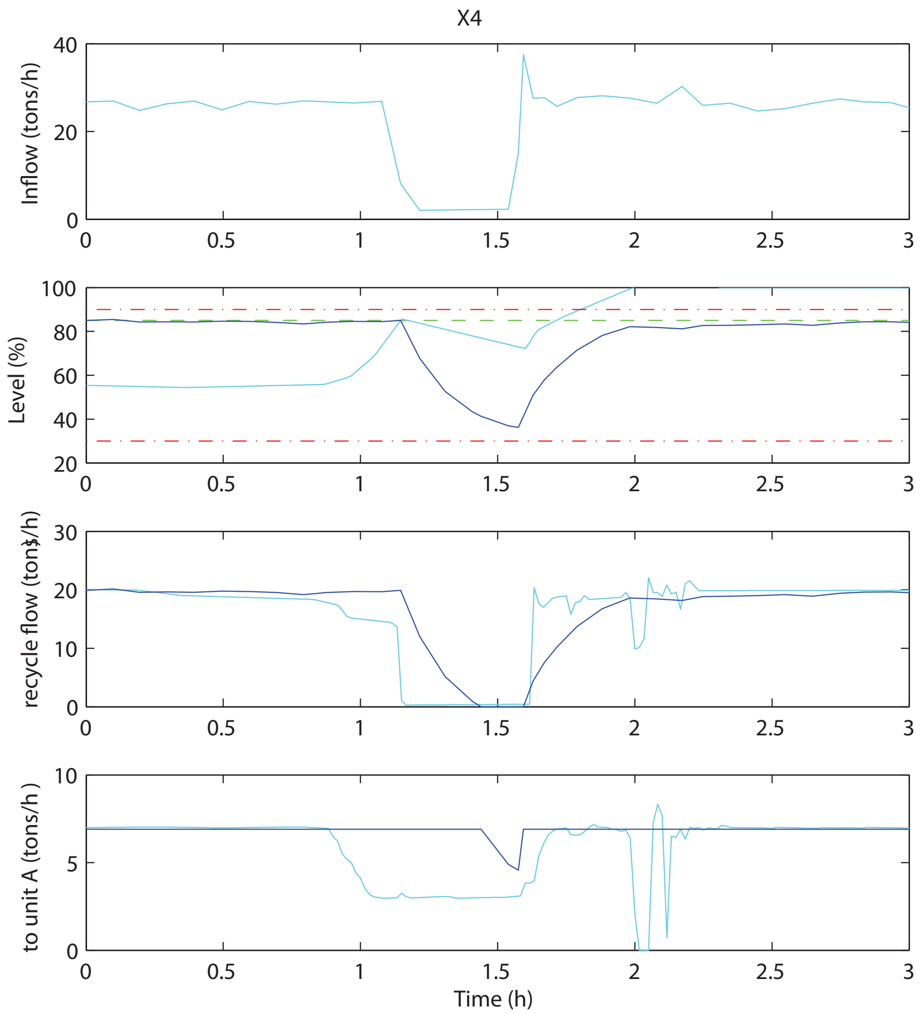

Figure 5.20: Simulation of the level control in X4 during a shutdown of unit Y with the suggested P-controller with $K_c = 0.57$ and the set point 85 %.

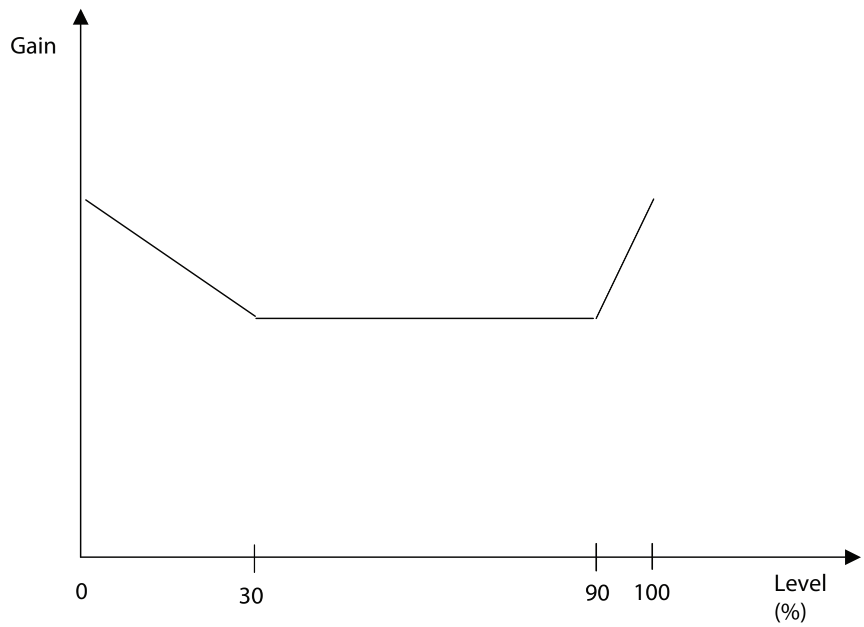

Figure 5.21: Example of gain scheduling for level control of X4.

5.3.3 Conclusions

The suggested P-controller for X4 seems to handle shutdowns of unit Y much better than today's control strategy if the objective is to have small and not too rapid variations in the outflows from the tank. However, the main reason for today's strategy of reducing the outflow drastically immediately as the shutdown begins is to be able to increase the ratio of new C to the tank where the reaction takes place accordingly and in this way keep the X/C-ratio constant. The X/C-ratio is today computed as the amount of X over the amount of new C, where the amount of new C is varied to keep the ratio constant. The recycle flow of C is thus not included and variations in this flow will give an incorrect X/C-ratio. It is most likely easier to compensate for a step disturbance in the outflow, caused by controlling the tank manually as is done today, than from the variations caused by the suggested control strategy for X4. However, if the strategy for computing the X/C-ratio is modified a correct ratio could be achieved independently of the characteristics of the recycle flow. A suggestion is to include the recycle flow of C in the computations of the X/C-ratio.

Another comment on the level control of X4 is that it has been designed to have good behaviour when running the plant at a certain production speed. If the production speed is changed the optimal tuning of the controller will be different since the size of the disturbance to be handled will be changed. A flexible controller which would automatically adjust to differences in production speed could be achieved by re-tuning the controller before every shutdown of unit Y. It is known that the inflow to X4 will go down to a certain minimum value during a shutdown of this unit and thus the maximum control error to be handled by the controller will be given by the inflow to X4 before the shutdown subtracted by this minimum value.

In the simulations the approach for handling the flow to unit A might not be optimal. The proposal is to keep the flow constant for as long as possible and only if the recycle flow has to go down to zero the flow to unit A will be decreased. If this happens the flow will have to be decreased rapidly in order to keep the level within the low limit, which could have disastrous effects on unit A. A better strategy is probably to begin to decrease the flow to unit A earlier but smoother.

For the buffer management problem in the tanks before unit Y in the production line the bidirectional flow control strategy described in section 3.4 is suggested. In this case we have two master flows, both the flow to the first of the three tanks and the feed to unit Y, which makes the problem slightly more complicated. At normal operation the level in X3 will not be controlled and X2 and X1 will be level controlled by manipulating the outflow of the tanks (see Figure 5.22). When a shutdown of unit Y occurs the tanks will still be controlled in this way until a given maximum limit of the level in X3 is reached. When this limit is reached the level controller of X3[1] will take over manipulation of the outflow of X2, which is also the inflow to X3 (see Figure 5.23). When a given maximum limit of X2 is reached the level controller of X2 takes over manipulation of the outflow from X1 (5.24). If X1 becomes full the flow to the first of the three tanks has to be reduced. However, this is not likely to happen if the shutdowns are of normal durations and the levels in the tanks are not too high when a shutdown begins. The result of using this strategy for the buffer management in the three tanks before unit Y is that they are filled one by one starting with the tank closest to unit Y. A parallel can be drawn to the simple problem in section 4.1 where it is desirable to have high buffer tank levels before the bottleneck unit and low levels after the bottleneck unit to ensure that this unit can run at full speed for as long as possible. In this case the unit before the first of the three tanks can be seen as the bottleneck and if this flow should be able to be chosen arbitrarily it would be desirable to fill up the buffer tanks furthest away from this unit first, accordingly with the strategy described.

[1] This level controller does not exist today.

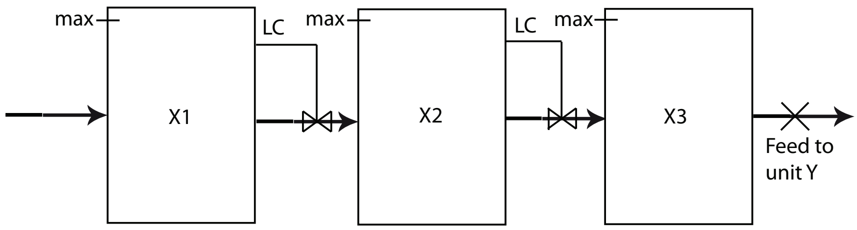

Figure 5.22: Bidirectional flow control of the tanks before unit Y. Step one: A shutdown of unit Y occurs. X1 and X2 are level controlled manipulating the outflow.

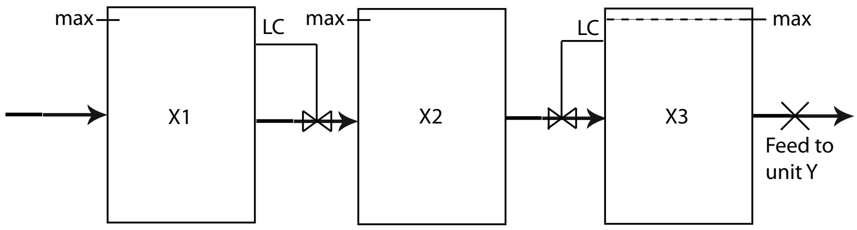

Figure 5.23: Bidirectional flow control of the tanks before unit Y. Step two: The maximum level in X3 is reached. The level controller of X3 takes over manipulation of the outflow of X2.

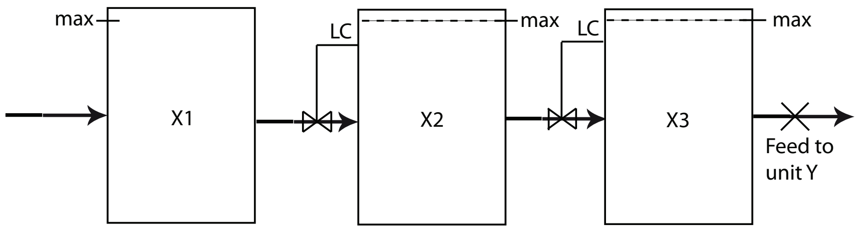

Figure 5.24: Bidirectional flow control of the tanks before unit Y. Step three: The maximum level in X2 is reached. The level controller of X2 takes over manipulation of the outflow of X1.

Chapter 6

Conclusions

Buffer management has shown to be an important factor within the process industry and good strategies for buffer handling can possibly increase both the availability of the plant and the quality of the end product.

The solution of a buffer management problem could be of different character, for example the solution could be a choice of set point for the level controller of a buffer tank, the choice of which variable to manipulate or the tuning of the level controller. The possible solutions mentioned will not be independent of one another and if the problem should be solved all possible combinations of the solution parameters should to be considered. What has been done today is to solve the optimization problem for some certain choices of constraints.

In this master thesis the buffer management problem without constraints is considered and then the problem formulation is extended to include level control of the buffer tanks. It is shown that the solution will change when adding constraints such as constraints on flow derivatives or start-up times to the problem formulation. This can be seen clearly in the case studies made at Perstorp AB. In the plant in Warrington PI-control showed to yield good result for the level control of the buffer tank and the start-up times on certain units in the production line showed to affect the optimal choice of set point and tuning of the controller. In Singapore P-control of a buffer tank was discovered to yield better behaviour because of the certain characteristics of the disturbances. A fundamental difference between the two case studies is the purpose of the level controlled buffer tank. In Warrington the objective of introducing level control in the buffer tank is to prevent the level from going outside the specified limits whereas in Singapore the purpose of the buffer tank is mainly to minimize flow variations in the outflow of the buffer tank.

Because of the complexity of the area of buffer management there are still numerous problem formulations left to be solved. This master thesis contributes to structuring the multi-dimensional map of buffer management problems with different constraints and filling in a few of the gaps within the domain.

Acknowledgements

This thesis is a partial fulfilment of the requirements for the degree of Master of Science in Technical Physics and is developed in collaboration with the Department of Automatic Control in Lund and Perstorp Specialty Chemicals AB. The work corresponds to 20 weeks full time studies and was carried out during the autumn of 2008.

First I would like to thank my supervisors, Krister Forsman at Perstorp AB and Charlotta Johnsson at Automatic Control in Lund for their guidance and for giving me the opportunity to perform such an interesting project as my master thesis work. I would also like to thank them and Tore Hägglund at Automatic Control for their positive attitude, for always being helpful and inspiring and for encouraging me to a further academic career within automatic control. Alf Isaksson at Automatic Control in Linköping has also been of great help during my master thesis work.

My other colleagues at Perstorp AB, Nils-Petter Nytzén, Jesper Jönsson and Lena Persson also deserve a big thank you for their assisting guidance and for nice company during my work at Perstorp AB.

Finally I would like to thank my fiancé, my family and my friends for supporting me at all times and for always being available for a phone call or a visit.

Bibliography

[1] S. Ogawa, B. Allison, G. Dumont, M. Davies: *A new approach to averaging level control*, Pulp & Paper Centre, In: Process Control Systems, Stockholm, Sweden (2002)

[2] S. Ogawa: *Stochastic Averaging Level Control and Its Application to Broke Management in Paper Machines*, PhD Thesis, University of British Colombia (2003)

[3] A. Åberg, T. Brogårdh: *Optimization of the Buffer Storage Levels in a Pulp Mill controlled by the ASEA PMPC System*, Unpublished report, ASEA (1980)

[4] J. Shin, J. Lee, S. Park, K. Koo, M. Lee: *Analytical design of a proportional-integral controller for constrained optimal regulatory control of inventory loop*, Control Engineering Practice (16) 1391 - 1397 (2008)

[5] K. Forsman: *Reglerteknik för processindustrin*, Studentlitteratur (2005)

[6] A. Faanes, S. Skogestad: *Buffer Tank Design for Acceptable Control Performance*, Industrial & Engineering Chemistry Research (42) 2198 - 2208 (2003)

[7] K.J. Åström, T. Hägglund: *PID controllers: Theory, design and tuning*, Instrument Society of America (1995)

[8] A. Visioli: *Practical PID control*, Springer London (2006)

[9] K. Forsman: *Some useful formulas in automatic control*, Unpublished report, Perstorp (2008)

[10] F.G. Shinskey: *Special rules for tuning level controllers (Part-1)*, ControlGlobal.com (2005)

[11] W.L. Luyben, B.D. Tyreus, M.L. Luyben: *Plantwide Process Control*, McGraw-Hill (1998)

[12] F.G. Shinskey: *Controlling Multvariable Processes*, Instrument Society of America p. 46-47 (1981)

[13] S. Skogestad: *Control structure design for complete chemical plants*, Computers & Chemical Engineering (28) 219 - 234 (2004)

[14] R. Rahmberg: *The good circles and the bad*, Perstorp, AM-tryck & reklam (2006)