

# Modelling and Control of an Evaporation Process

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# Abstract

Model-based design is in many ways seen as a potential instrument in process industries due to the close link between the process and the model. In Växjö, the process industry of Lantmännen Reppe AB produces syrup. The syrup goes through an evaporation process in order to raise the sugar concentration. If the concentration gets to high the syrup turns solid, which can make the process a bit tricky to control.

In this master's thesis the goal is to model this process in the MathWorks environment Simulink in order to gain understandings about different aspects of the process and eventually bring forward and evaluate different control strategies. Information about the real process has been obtained by visits at the factory and mail correspondence with the process engineer of the plant. Important advices along the project have been given by supervisors at the Department of Automatic Control and at Combine Control Systems AB.

The modelling is founded on approximations regarding no temperature or pressure dependencies, but only mass and energy balances. The model has been adjusted and tuned along the project in order to match the given process dynamics. The model verification has been conducted by comparisons of model simulations and process data. The model has in many ways proven to capture the fundamental behaviors of the process.

A number of different control strategies have been tested in the model and the results have been compared. It has been shown that the present feed forward controller improves the system control but also that new feed forward controllers and a new sensor can improve the system control furthermore. The greatest improvements have been seen when introducing an additional sensor in the model.



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# 1

## Introduction

*The project aims to model an evaporation process and investigate different control strategies. This chapter contains a presentation of the background of the project and the goals of the work. It also presents the method and the delimitations of the project. A description of all chapters and a list of nomenclature are also included.*

### 1.1 Background

Lantmännen Reppe AB has been producing glucose syrup at their factory in Växjö since 1886. The process is divided into a number of different parts. The last part, the final evaporation process, is the most energy consuming. This part consists of a preheating system and three evaporators. The system is controlled through several valves, regulating the steam flow to the heat exchangers in the evaporators. Lantmännen Reppe AB has been doing improvements as a part of their energy efficiency work and is aiming for further improvements of the system.

The evaporation process of Lantmännen Reppe AB is a relatively straight forward process, which basically boils a low concentrated glucose syrup solution to a higher concentration by leading off the vaporized water. The desired concentration of the resulting solution varies, but a dry substance percentage of 80 is often desired. The process control is, despite the simplicity of the process, very crucial. If the dry substance gets close to 85 % the sugar solution turns solid and will have to be removed from the system by hand, at a high cost for the company.

Sudden large overshoots increase the risk of the syrup turning solid. Overshoots may be caused by disturbances in the process inputs. To improve the process control Lantmännen Reppe AB asked Combine Control Systems AB in Lund for their expertise. Combine is working with model-based design and wish to further develop their knowledge in process control. They believe there is great potential for model-based design in process industries, since the model-based design provides well founded control systems due to the closeness between the process and the model. After contacting Combine the project eventually was made into a master's thesis project.

## 1.2 Goals

The overall goal is to evaluate the control of the process regarding disturbances of the process inputs. This goal has been divided into three different milestones of the project:

- Create a model of the real process
- Implement the existing control strategy in the model
- Compare and evaluate both existing and new control strategies

Each of these milestones will be including a number of smaller goals. There will be goals in understanding the different parts of the process – such as heat exchangers, evaporators, ejectors and valves – but also improve the understandings on how to use the model-based software Simulink in MATLAB.

## 1.3 Method

The method is essentially following the milestones of the project. The work was thus structured in the same way as the goals. The modelling and the control of the system were worked with in parallel during the project. The modelling and the model verification took most of the time.

Initially the focus was on understanding and modelling the process. This included reading about the different parts of the process as well as learning the software. During this step a first visit at the factory was conducted to get some initial information, see the process and get the opportunity to ask questions. As the work with the model progressed more visits at the factory were conducted in order to show the model, discuss the different features and eventually be able to verify the model itself. Unfortunately it was only possible to get plots of the measurement data from the process. The model verification was thus done by mimicking the input signals and studying how the model behaves compared to the real process.

Secondly the model was controlled. To begin with the present controllers were introduced as in the real process. The controllers were adjusted to fit the model.

Finally new control strategies were designed, implemented and evaluated to see if the present control of the system could be improved in any fashion. This was easily done with model-based design and the advantage was that new strategies could be evaluated without even touching the real process. When the different control strategies were implemented they were compared and evaluated to one another or together. To evaluate the different control strategies all the strategies were tuned in order to compare them equally.

## 1.4 Delimitations

Even though the principles of an evaporator system may seem as a simple system, it is quite complex. Having evaporators connected in series, a solution that changes its thermal and fluid characteristics along the system as well as different pressures and temperatures along the way create a lot of parameters and variables. A lot of sensors would be needed to be able to determine all these system values and material properties. To make the system more simplistic it was decided to eliminate all temperature and pressure dependencies from the model.

## 1.5 Disposition

### Chapter 1 – Introduction

This chapter covers background, goals, method and delimitations of the project. It also contains a chapter disposition and nomenclature.

### Chapter 2 – Background Theory

Here all background theory is explained. Throughout this chapter the theory about glucose syrup, evaporation techniques and basic control theory are covered. There is also a section about the used software last in this chapter.

### Chapter 3 – The Process at Lantmännen Reppe AB

This chapter contains a presentation of the process at Lantmännen Reppe AB and the present control system. Examples of the problems they have been having are also presented.

### Chapter 4 – Modelling

Here the mathematical modelling is presented together with a presentation of the Simulink model. It also contains a section about how the model has been adjusted to correspond to the real process.

### Chapter 5 – Control Strategies

The different control strategies used in the model and how they have been implemented are presented here. Both the present control strategy and new control strategies are explained.

### Chapter 6 – Results

This chapter covers the model verification with comparisons between the real process and the model. The different control strategies are also compared.

## Chapter 7 – Discussion

Discussion about the results regarding both the model verification and the comparisons between control strategies. There is also a section about future studies.

## Chapter 8 – Conclusions

Conclusions drawn from the discussion of the results. The conclusions are focused on how Lantmännen Reppe AB could continue to work on their control strategy.

### 1.6 Nomenclature

$A$	Heat transfer area in a heat exchanger
$c$	Concentration
$\dot{c}$	Time derivative of concentration
$c_0$	Concentration entering the system
$c_i$	Concentration out of effect $i$
$c_{in}$	Concentration of syrup entering an evaporator
$c_{out}$	Concentration of syrup exiting an evaporator
$D$	Derivative gain
$e$	Control error
$F_0$	Mass flow of syrup entering the system
$F_i$	Mass flow of syrup out of effect $i$
$F_{in}$	Mass flow of syrup entering an evaporator
$F_{out}$	Mass flow of syrup exiting an evaporator
$G$	First order transfer function
$G_c$	Transfer function of concentration in a perfectly stirred tank
$G_F$	Transfer function of mass flow through a tank, regulated to constant volume
$G_V$	Transfer function from steam to vapor
$h$	Enthalpy
$h_{F_{in}}$	Enthalpy of syrup entering an evaporator

$h_{F_{out}}$	Enthalpy of syrup exiting an evaporator
$h_{S_{in}}$	Enthalpy of steam entering an evaporator
$h_{S_{out}}$	Enthalpy of steam exiting an evaporator
$h_V$	Enthalpy of vapor exiting an evaporator
$I$	Integral gain
$K$	Proportional gain
$k$	General constant or gain of a transfer function
$K_b$	Back-calculation coefficient
$k_{ff}$	Feed forward gain
$k_{ffs}$	Steam feed forward gain
$L$	Latency, delay, dead time
$L_c$	Latency of dry substance
$L_F$	Latency of mass flow of syrup
$L_S$	Latency of mass flow of steam
$LMN$	Control signal
$P$	Proportional gain
$PV$	Process variable (measurement signal)
$q$	Volumetric flow rate
$r$	Reference signal
$S_0$	Mass flow of steam entering the system
$S_i$	Mass flow of steam entering effect $i$
$S_{in}$	Mass flow of steam entering an evaporator
$S_{out}$	Mass flow of steam exiting an evaporator
$SP$	Setpoint (reference signal)
$S_{pre}$	Mass flow of steam to preheater
$S_{TVR}$	Mass flow of steam entering the thermal vapor recompression system

## Chapter 1. Introduction

$T$	Time constant of a transfer function
$T_d$	Derivative time
$T_i$	Integral time
$T_{in}$	Temperature of steam entering a heat exchanger
$T_{out}$	Temperature of steam exiting a heat exchanger
$U$	Heat transfer coefficient between the two media in a heat exchanger
$u$	Control signal
$u_{61}$	Control signal from A061
$u_{66}$	Control signal from A066
$u_D$	Contribution from D part to control signal
$u_{ff}$	Output of feed forward system
$u_{ffs}$	Output of a steam feed forward system
$u_I$	Contribution from I part to control signal
$u_{out}$	Saturated control signal
$u_P$	Contribution from P part to control signal
$V$	Mass flow of vapor exiting an evaporator
$V_{tank}$	Tank volume
$W_{losses}$	Thermal losses of an evaporator
$W_{th}$	Transferred thermal energy of an evaporator
$y$	Measurement signal

# 2

## Background Theory

*The syrup manufacturing at Lantmännen Reppe AB in Växjö ends with an evaporation process to raise the concentration of sugar in the syrup. To model this process it is important to have an understanding of heat transition as well as mass and energy balances. This chapter presents a selection of background theory to give a basic understanding of the process and the control strategies. The following sections contain a brief overview of syrup manufacturing, evaporation processes, steam economy, control theory and a short description of model-based design and the software used.*

### 2.1 Glucose Syrup

Glucose syrup is a food syrup, which actually contains both glucose and maltose. The syrup is made from hydrolysis of starch from e.g. wheat, corn or potatoes. An addition of glucose syrup to a product prohibits crystallization of sucrose and increases the ability to obtain humidity from the air. It also increases the viscosity and decreases the sweetness. Because of these qualities glucose syrup is commonly used within ice cream and confectionery industry [Salomonsson, 2010].

Glucose syrup is produced by acid hydrolysis, enzyme hydrolysis or a combination of both. When using acid hydrolysis an acid is used to catalyze the cleavage of the chemical bonds in the starch molecules. With acid conversion the manufacturers cannot influence the saccharide distribution. In enzyme hydrolysis the enzymes  $\alpha$ -amylase,  $\beta$ -amylase and glucoamylase are added to a mixture of starch and water. The starch is broken into different oligosaccharides and glucose molecules. In this case it is possible to manipulate the glucose content [Schenck, 2012]. After the hydrolysis impurities are removed from the syrup. Finally the syrup is evaporated to raise the concentration of sugar to the requested level [Wikipedia, 2014b]. To get a measure of the sugar content in the syrup solution the term dry substance is used. This is simply a measure of the mass percentage of solids in the solution [Sugartech, 2014].

## 2.2 Evaporators

Solutions often consist of a solvent and a dissolved solid substance. By simple thermal separation the solvent can be vaporized leaving the solution with an increased concentration. The purpose of thermal separation can vary, but most often it is a matter of concentrating the solution, recovering the solvent or recovering the dissolved substance [Billet, 2012].

In the food industry a solution often contains more water than requested. When the foodstuff is a liquid the easiest way to remove the water is by evaporating it. As a part of a process plant the evaporator has two principal functions, to exchange heat energy for the solvent to absorb and to separate the generated vapor that is formed from the ingoing solution [Earle and Earle, 1983].

### The Plate Heat Exchanger

To raise the temperature and reach the boiling temperature of the incoming solution it is often convenient to use heat exchangers. A heat exchanger is basically an energy efficient piece of equipment for transferring heat from one medium to another and can be built in numerous ways [Thulukkanam, 2013].

In food industries plate heat exchangers are common due to the low viscosity of the solutions. The plate area creates a large contact surface, where heat can be transferred. Together with the large area, the thin layers make sure that a majority of the solution is in contact with the plates. The whole construction also leads to a turbulent flow, which improves the heat transfer capability. All the other aspects, such as the flow arrangement and the number of plates, are deliberately chosen to improve the heat transfer of the exchanger [Thulukkanam, 2013].

### The Flash Tank

The flash tank, sometimes known as a vapor-liquid separator, is the part of the system where the vapor gets separated from the solution that is still in liquid phase. The principle is very straight forward, using gravity in a vertical vessel causing the liquid to settle at the bottom and the vapor to rise to the top. The vapor and the liquid can be extracted through different outlets and the evaporation is complete. The resulting liquid will be left with a higher concentration than the entering solution [Wikipedia, 2014c].

### Climbing and Falling Film Plate Evaporator

The climbing and falling film plate evaporator is often used in process industries when a short residence time is desirable [Wikipedia, 2014a]. Reducing the residence time can be a critical factor when evaporating temperature sensitive materials, such as various foodstuff or pharmaceuticals. The residence time impacts important aspects considering color, texture and taste when it comes to evaporation in the food industry [Billet, 2012].

The design of a climbing and falling film plate evaporator does, as the name reveals, consist of two heating phases. Firstly comes the climbing phase, where the ingoing solution is heated by the steam flow as it climbs, or rises, through the plate [Wikipedia, 2014a]. Secondly comes the falling phase, where the momentum of the likewise downward flowing evaporated vapor assists the acceleration, reducing the residence time [Billet, 2012].

Vapor recycling is often taken into consideration to increase the energy efficiency of the evaporator system. The recycling strategy can vary, but common strategies are connecting a number of evaporators in series and utilization of thermal vapor recompression. [Billet, 2012].

### **Increasing the Steam Economy**

The low pressure vapor generated by the evaporator can be used in another evaporator. A single evaporator can be called an effect and a system where several effects are connected in series is thus called a multiple effect evaporator. The steam economy of a multiple effect evaporator can be increased by reusing the vapor from one effect by connecting it to the steam chest of another effect. The reuse of energy increases the coefficient of performance and thus the energy efficiency of the entire system [Earle and Earle, 1983].

If the vapor provided by an evaporator is going to boil off liquid in a following effect, the boiling point of the following effect must be lower than in the first effect. Therefore the following effect must be under lower pressure. Thus the pressure must be reduced after each effect. The case when the most concentrated solution will occur in the last effect is called forward feed. The alternate case is to reverse the flow of the solution so that the most concentrated solution occurs in the first effect. This is called backward feed [Earle and Earle, 1983].

Another way of reusing the steam produced by an evaporator is so called vapor recompression. This is done by recompressing the vapor and returning it to the steam chest of the evaporator that produced it. This can be done using a jet steam ejector. A jet steam ejector can recompress a portion of the vapor by using fresh highly pressurized steam. This increases the overall steam economy. The cost is however a pressure drop of the fresh steam. When using jet steam ejectors the method is called thermal vapor recompression [Earle and Earle, 1983].

One more aspect to be taken into account when considering the steam economy of an evaporator is the feed temperature. If the solution is not at its boiling point of the current pressure additional heat has to be used to raise the temperature before the evaporation begins. A separate preheater, or a series of preheaters, can be used to increase the solution temperature before entering the evaporator. Heat exchangers are commonly used as preheaters [Earle and Earle, 1983].

## Mass and Energy Balances of an Evaporator

The main purpose of an evaporator is vaporizing and eliminating an amount of the solvent from a solution. Hence the main equations of a single evaporator consists of the mass and energy balances, see equations (2.1), (2.2), (2.3) and (2.4).

The mass balance equations simply state that what enters the system must exit the system, regarding both the steam and the solution. Since the steam and the fluid do not mix, the mass equations contain the mass balances below. Equation (2.3) states that the amount of dissolved solid substance in the solution must be constant throughout the evaporator system [Ahlbeck et al., 2010]. The mass flow of syrup is denoted  $F$ , the mass flow of vapor  $V$  and the mass flow of steam  $S$ . The indices *in* and *out* denote the signals going into and out of the evaporator. The concentration of the dissolved substance is denoted  $c$ .

$$\begin{cases} F_{in} = V + F_{out} & (2.1) \\ S_{in} = S_{out} & (2.2) \\ F_{in} \cdot c_{in} = F_{out} \cdot c_{out} & (2.3) \end{cases}$$

The energy balance equation states that the energy of the entering masses must be equal to the energy of the exiting masses, including the thermal losses  $W_{losses}$  in the system. The energies of the masses are determined by the mass flows and the various enthalpies  $h$  [Ahlbeck et al., 2010].

$$S_{in} \cdot h_{S_{in}} + F_{in} \cdot h_{F_{in}} = S_{out} \cdot h_{S_{out}} + V \cdot h_V + F_{out} \cdot h_{F_{out}} + W_{losses} \quad (2.4)$$

An additional equation for the heat transfer can also be determined. The transferred thermal energy,  $W_{th}$ , between the steam and the fluid can be approximated by equation (2.5)

$$W_{th} = U \cdot A \cdot (T_{in} - T_{out}) \quad (2.5)$$

where  $U$  is the heat transfer coefficient and  $A$  is the heat transfer area.  $T_{in}$  and  $T_{out}$  are the temperature of the entering and exiting steam [von Böckh and Wetzel, 2012].

## 2.3 Control Theory

Utilization of relatively simple control theory and strategies are often sufficient even when controlling more complex processes. In this section the basic control principles of this project are covered. Initially the PID controller and its different parts are presented. Secondly the aspects of windup and the importance of anti-windup control are explained. Finally feed forward control, cascade control and mid-range control are briefly presented.

## The PID Controller

The PID controller is the most common controller and it is described by equation (2.6).

$$u(t) = K \left( e(t) + \frac{1}{T_i} \int_0^t e(\tau) d\tau + T_d \frac{d}{dt} e(t) \right) \quad (2.6)$$

The controller sends out a control signal,  $u$ , depending on the error,  $e$ . The error is defined as the difference between the measurement signal,  $y$ , and a reference signal,  $r$ , see equation (2.7).

$$e = r - y \quad (2.7)$$

As one can see in equation (2.6) the controller consists of three parts. The proportional term, or  $P$  part, has an output that is proportional to the current error. This output can be adjusted by tuning the proportional gain  $K$ . The advantage of an output proportional to the error as opposed to a controller that would give maximum output as soon as there is an error, is that it reduces the oscillations that would occur otherwise [Hägglund, 2011].

The next term is the integral term, or  $I$  part. This term is proportional to a weighted sum of all the previous errors.  $T_i$  is the integral time. This term makes it possible to eliminate the stationary error which could occur if there only was a proportional term [Hägglund, 2011].

The last term is the derivative term, or  $D$  part. This part predicts what will happen with the error in the future. It is proportional to the derivative of the error. This term makes the controller behave differently if the error is increasing than if it is decreasing even though the current error is the same size.  $T_d$  is the derivative time. Most real processes are controlled using a PI controller, i.e. a controller without a derivative term [Hägglund, 2011].

Since one most of the time refers to the parameters as  $P$ ,  $I$  and  $D$  equation (2.6) is sometimes written in an alternate form, according to equation (2.8) below.

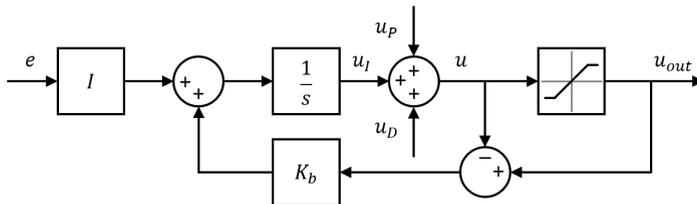
$$u(t) = P \left( e(t) + I \int_0^t e(\tau) d\tau + D \frac{d}{dt} e(t) \right) \quad (2.8)$$

## Anti-Windup Control

The integral term in the PID controller makes the controller unstable. A feedback loop is crucial to stabilize an unstable process. This means that the controller needs to be stabilized by a feedback loop. If the output of the controller is saturated the feedback is not working and problems might occur. This since the output of the controller is not the same signal as the signal that is affecting the process [Hägglund, 2011].

When the control signal is saturated at an upper limit and not large enough to eliminate the error the integral of the error will increase linearly. This also causes the output to increase but the signal affecting the process still lies at a constant level. If the setpoint is decreased to a level where the control signal is high enough to eliminate the error the sign of the error will change. The integral term and the output will then decrease. Since the output of the controller is at a higher level than the upper limit the signal affecting the process will be constant during some time before it starts to decrease. This problem is called windup [Hägglund, 2011].

This can be avoided by implementing an anti-windup circuit in the controller. One option is to use the back-calculation method. This method uses a feedback loop by measuring the output of the saturation and form an error as the difference between the output of the controller and the output of the saturation. This error is then sent through a gain,  $K_b$ , and back to the integrator, see figure 2.1. The integral term is reset and the output of the controller will be the same size as the signal affecting the process. When the control signal is not limited this error will be zero and will not affect the controller [Åström and Wittenmark, 2011].



**Figure 2.1** A block scheme of the integral part of a PID controller with an anti-windup system using the back-calculation method. The contributions from the different parts of the PID controller are denoted  $u_p$ ,  $u_I$  and  $u_D$ . The signal affecting the process is denoted  $u_{out}$ .

## Feed Forward Control

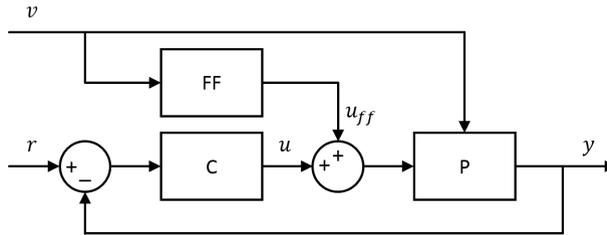
In most systems there are more signals entering the process than just control signals. The classic feedback loop is often enough but comes with the drawback of not being able to detect disturbances or changes of the ingoing signals until they have affected the measurement signal. Adding feed forward control can then be a good way of improving the system control, as long as the disturbances or the changes of the ingoing signals can be measured [Hägglund, 2011].

In figure 2.2 a simplistic scheme over how feed forward control can be added to a feedback control system is seen. Disturbances or other signals entering the process are called  $v$  and the feed forward control signal is called  $u_{ff}$ .

It is crucial when the signal  $v$  can be measured. If the signal can be measured long before affecting the process, the feed forward will be able to compensate in an effective way. If the signal has already started to affect the process, the feed forward

might be useless. Therefore processes with long dead times are well suited for this type of control [Hägglund, 2011].

The implementation of a feed forward control system can be done in different ways. The feed forward transfer function needs to be sufficiently chosen in order to get the right properties of how the signal  $v$  will affect the process. Sometimes a proportional feed forward is sufficient and sometimes a more advanced high order transfer function must be found.

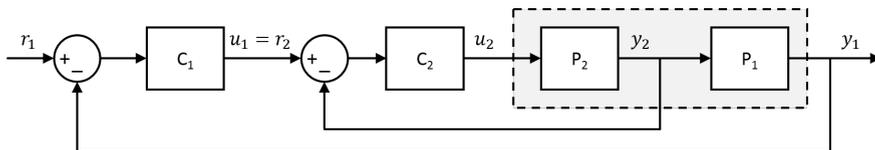


**Figure 2.2** A block scheme of the principles of feed forward control.

## Cascade Control

Usually more than one process variable is measured and to use as much information as possible is generally a good way to improve the control. Cascade control does exactly this by adding a secondary controller, supporting the primary controller. A simple scheme of a cascade control system can be seen in figure 2.3, where  $C_1$  is the primary controller and  $C_2$  the secondary controller [Hägglund, 2011].

The cascade control aims to control  $y_1$  with the primary controller. This could be done without the secondary controller, but with the implementation of the secondary loop a more efficient control is attained. It also suppresses disturbances more quickly due to the secondary feedback loop. As can be seen in figure 2.3 the control signal of  $C_1$  works as the reference signal to  $C_2$ , while the control signal of  $C_2$  is the actual control signal going into the process. In order to achieve good cascade control a significantly faster secondary loop is required [Hägglund, 2011].



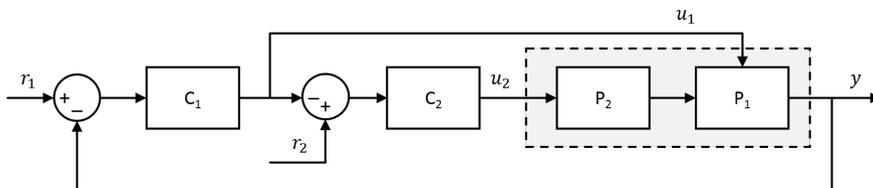
**Figure 2.3** A block scheme of the principles of cascade control.

## Mid-Range Control

When having two actuators controlling the same system variable, letting them work in different ranges might improve the control efficiency. The range difference can sometimes occur due to physical aspects of the system, making one controller faster and possibly more accurate. Since control signals are more or less saturated it is important to make sure that they do not work too close to their boundaries in order to have a robust control [Sörnmo et al., 2013].

Mid-ranging control is a control strategy where the slower controller is using the control signal of the faster one as a measurement signal. By doing so the slower controller aims to keep the faster controller in the middle of its range, while the faster controller takes care of the main control of the system variable [Sörnmo et al., 2013].

One way to illustrate a mid-range control system, added to a feedback loop, can be seen in figure 2.4. Here the first controller,  $C_1$ , is the faster controller and the second controller,  $C_2$ , is the slower controller. The control signal of  $C_2$  is controlling the part of the process,  $P_2$ , which is connected in series with another part of the process,  $P_1$ . The measurement signal from  $P_1$  is directly controlled by  $C_1$  and only indirectly by  $C_2$  since it is aiming to keep the control signal of  $C_1$  at its mid-range value.



**Figure 2.4** A block scheme of the principles of mid-range control for a process with two subprocesses in series.

## 2.4 Model-Based Design

Model-based design is a mathematical and visual method for development of control systems but it can be used in other areas as well, for example signal processing. Developers create models to test if algorithms will work before implementing them in a real system. The procedure starts by modelling of a process. Then a controller can be modelled as well and a simulation can be run. This makes it possible to understand whether a control strategy will work before the embedded code is written. In this way costs of testing something in a real process can be reduced [MathWorks, 2014].

## **The Simulink Software**

Simulink, created by MathWorks, is a block diagram environment for simulation and model-based design. It contains a graphical editor, a block library that can be customized and solvers for modelling and simulating dynamic systems. Since it is integrated with MATLAB it is possible to integrate MATLAB algorithms into models and use the simulation results for further analysis in MATLAB. Simulink blocks represent basic mathematical operations. When blocks are connected the resulting diagram is equivalent to the mathematical model of the system [MathWorks, 2015b].

# 3

## The Process at Lantmännen Reppe AB

*All knowledge about the evaporation process at Lantmännen Reppe AB in Växjö is presented in this chapter. The set-up of this specific evaporation process is described together with gains, dead times and time constants. The properties of the process and all parameters are given by Lantmännen Reppe AB. In the past they have conducted several tests and estimated different system parameters.*

### 3.1 The Evaporation Process

At Lantmännen Reppe AB in Växjö the final evaporation of the syrup is done in several steps. A scheme of the entire process can be seen in figure 3.1, where orange signals represent the pipes containing steam or vapor, blue signals condensed water and gray signals the syrup. The syrup is initially extracted from a storage tank and pumped into the preheater, consisting of three heat exchangers. In the figure heat exchangers are represented by blue rectangles. The preheater is followed by three evaporator effects where the evaporation is done. Each of the effects consists of a heat exchanger and a flash tank, where the separation of syrup and vapor is taking place. The flash tanks are represented by the three gray blocks.

All steam used in the process comes from the same source which also supplies other parts of the factory with steam. The steam that enters the process is extracted from the steam feed through the three green valves in the upper part of the figure. In the heat exchangers both primary steam from the steam feed and recompressed steam from a thermal vapor recompression system, called TVR, are used. The first heat exchanger in the preheater uses condensed steam from the first evaporator effect while the second heat exchanger uses primary steam and recompressed steam given from the first effect through a small jet steam ejector. The third and last preheating heat exchanger uses only primary steam since it also pasteurizes the syrup. Effect 1, which is the first effect, uses primary as well as recompressed steam from the effect itself, by using another larger jet steam ejector. Effect 2 uses secondary



steam from effect 1 by a forward feed and the last evaporator, effect 3, only uses primary steam. The green circles are the sensors measuring the ingoing dry substance and mass flow, the pressure of the steam feed and the resulting dry substance and mass flow. These are not the only sensors in the systems, but essentially the most important ones along with the sensors of the controllers.

It is also possible to return the resulting syrup to the storage tank. This may occur if the dry substance is too low compared with the setpoint value and the product then needs to go through the process again. If the level in the storage tank is low, returning the higher concentrated syrup will induce a sudden raise of dry substance of the ingoing syrup.

The process is quite slow and the transport delays are one of the reasons why the process can be hard to control. Since the only sensors, measuring different properties of the syrup, are located before the preheater and after effect 3 it is hard to analyse the dynamics of different parts within the process. By manually inducing a step in the steam flow to effect 3 and analysing its influence on the dry substance the dynamics of the third effect have been estimated. From the step responses Lantmännen Reppe AB has been able to determine that there is a transport delay of 165 seconds until the dry substance is measured. There is also a gain of 0.12 and a time constant of about 80 seconds that characterize the third effect. The gain is calculated as the ratio of the relative change in the step response and the control signal. Since the three effects are similar, although not identical, it can be assumed that their dynamics are similar as well.

Apart from the dynamics of the third effect there are known properties about the system as a whole. A variation in the ingoing dry substance is noticed before the third effect after about 400 - 500 seconds, while a change in the ingoing mass flow is noticed after about 35 seconds. All known parameters of the real process are listed in appendix A. It is worth noticing that since the operation point – mass flow rates, ingoing dry substance, levels in the flash tanks, etc. – is not fixed, these values will vary from time to time. This means that the parameters will vary too. The values mentioned above are however seen as good approximations.

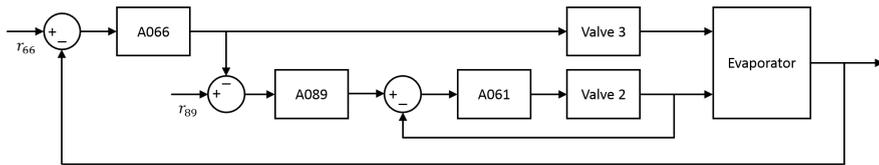
As seen in figure 3.1, the thermal vapor recompression is implemented using two jet steam ejectors. These ejectors approximately take as much vapor from the first effect as they are fed primary steam from the valve. The ejectors are not of the same size. The one recompressing vapor to effect 1 is larger than the one used for the preheater. This also means that more recompressed steam is used in effect 1 than in the preheater. The primary steam fed to these ejectors are split so that the large ejector uses four fifths of the primary steam and the small ejector one fifth. Thus the large ejector uses four times as much recompressed vapor as the small ejector.

The incoming mass flow of syrup is either controlled by a manually set reference value or by a level control of the tank that contains the syrup before it is pumped to the final evaporation process. This level control aims to keep a high volume of syrup to avoid sudden changes of dry substance of the syrup going into the evaporator. It also makes sure that the evaporator always is utilized to a high degree.

## 3.2 Control System

The dry substance of the syrup is controlled using three controllers. The control goals of these controllers are to reach the requested dry substance of the syrup and to move the main evaporation from the last effect to the first two effects. This is due to the superior steam economy of these effects compared to the last effect.

One controller, called *A066*, regulates the input of steam to effect 3, see figures 3.1 and 3.2. This is a PID controller, using the dry substance of the resulting syrup as measurement signal. The control signal sets the level of a valve. The requested dry substance is used as reference signal. The valve regulated by this controller has some characteristics that are important to be aware of. If the control signal sent to the valve is less than 20 % no steam passes through the valve. It has also been discovered that if the valve opens too much it takes a long time to decrease the mass flow of steam through it. To keep the valve from opening too much the control signal from *A066* is limited to 0.37. This means that the valve is open when it gets a control signal between 0.2 and 0.37, resulting in a small operation range for the controller.



**Figure 3.2** A block scheme of the present control system used in the process.

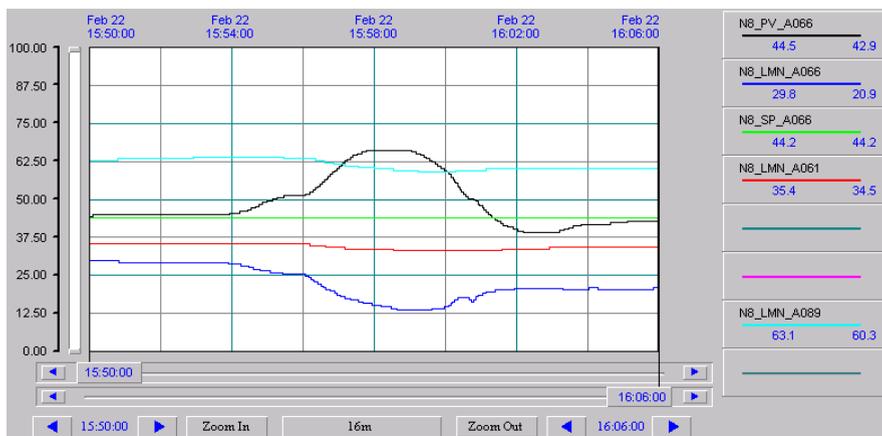
The other part of the control strategy aims to move the focus of the evaporation from effect 3 to the two first effects. This is done by using mid-range control and two different controllers in a cascade control system. The first one, *A089*, uses the control signal from *A066* as measurement signal. Reference value for this controller is manually set and represents a desired level for the valve controlled by *A066*. The control signal of *A089* sets the reference value of the third controller, *A061*. The control signal from *A061* controls another valve that regulates the pressure of the primary steam flowing into the jet steam ejectors and thereby the flow of steam into effect 1. *A061* uses the pressure of the primary steam of the jet steam ejector as measurement signal. These controllers are both PI controllers. The parameters of the controllers can be seen in table 3.1. All controllers have anti-windup systems.

Figure 3.3 shows what the operators at Lantmännen Reppe AB call a trend, which basically is a plot of up to eight of the different sensor measurements in the system. This particular trend shows how the controllers work together and how they react to an increase in the dry substance of the resulting syrup. In the figure signals of *A066*, *A089* and *A061* are plotted. The measurement signal is called process variable, *PV*, the reference signal setpoint, *SP*, and the control signal *LMN*.

Controller	P	I	$T_i$	D	$u_{min}$	$u_{max}$
A066	0.40	0.0050	200	20	0	0.37
A089	-0.30	0.0025	400	-	0	0.70
A061	0.50	0.1439	7	-	0	0.50

**Table 3.1** The PID parameters of the controllers regulating the mass flow of steam into the evaporation process. The integral times,  $T_i$ , are given in seconds.

To be able to plot different signals in the same trend all the signals are scaled to a percentage value between 0 and 100. The resulting dry substance is scaled so that a value of 70 % sugar corresponds to 0 % and 91.5 % to 100 %. The same scale is used for the setpoint of A066. The control signal of A066 is the valve position in percentage and does thus not need to be changed. When it comes to the signals for A061 the setpoint is a pressure in the range 0 - 15 bar and the same goes for the measurement. The control signal of A061 is also a valve position, just like the control signal of A066. As previously mentioned, the setpoint of A061 is the same signal as the control signal of A089. Other interesting signals not shown in this plot are the mass flow and the dry substance of the incoming syrup as well as the pressure of the steam feed. They are denoted A058, A067 and A069 respectively. The scaling intervals and a short description of all signals occurring in this report are listed in appendix A.



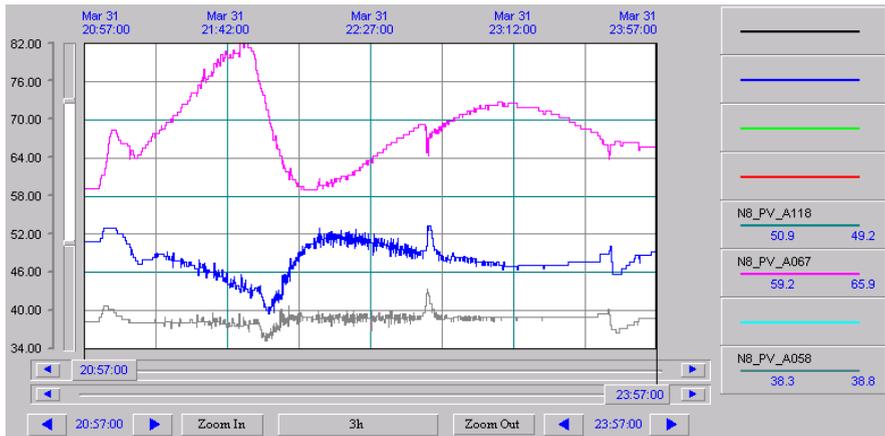
**Figure 3.3** A trend showing how the controllers work together and how they react to an increase in the dry substance of the resulting syrup. A066 lowers its control signal to decrease the steam flow into the last effect. This causes A089 to lower its control signal to keep A066 at the requested level. Since A089 decreases the reference value for A061 the control signal for A061 decreases too. When the dry substance reaches the required level again the control signals stabilize. Information about the signals can be found in appendix A.

## Feed Forward

Because of the long dead times in the system it also takes a long time for the controllers to make adjustments due to a change of the mass flow or dry substance of the incoming syrup. To give the controllers a head start and improve the performance of the control system a feed forward system has been implemented. It is rather simple and gives an addition to the output of *A066*. This addition,  $u_{ff}$ , is proportional to the amount of water in the incoming syrup, see equation (3.1). By using the amount of water both changes in the mass flow and dry substance are taken into account.

$$u_{ff}(t) = k_{ff} [100 - c_0(t - L_c)] F_0(t - L_F) \quad (3.1)$$

The feed forward gain  $k_{ff}$  has been adjusted to give an appropriate contribution to the control signal of *A066*. Out of precautionary reasons the feed forward signal can be limited. The properties of the incoming syrup are measured before the syrup enters the preheater. Since the feed forward system affects the steam flow into the last heat exchanger it must be taken into account how much time the syrup spends in the preheater and the first two effects. Due to the latencies mentioned in section 3.1, a delay of  $L_F = 35$  seconds has been applied to  $F_0$  and a delay of  $L_c = 400$  seconds has been applied to  $c_0$ . How the output of the feed forward system, denoted *A118*, is varying with  $F_0$  and  $c_0$  can be seen in figure 3.4. When the mass flow of syrup increases or decreases the feed forward signal follows. When the dry substance varies the feed forward signal varies in the opposite direction.

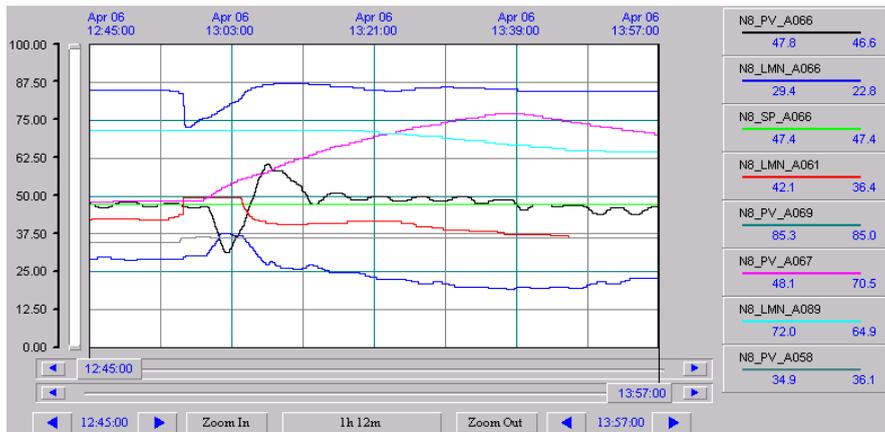


**Figure 3.4** An example of how the output of the feed forward system varies with  $F_0$  and  $c_0$ . When the mass flow of syrup increases the feed forward increases as well. When the dry substance increases the feed forward signal decreases. Information about the signals can be found in appendix A.

### 3.3 Problems

When the input signals vary it is hard to keep the resulting syrup at a stable level of dry substance. The dry substance and the mass flow of the ingoing syrup as well as the pressure of the steam feed are the main varying input signals of concern. Too large variations of the dry substance can have costly consequences. If the dry substance of the syrup rises to around 85 % the solution turns solid and needs to be manually removed.

Lantmännen Reppe AB experiences an improvement since the implementation of the feed forward system, but there are still situations when the control system is struggling to keep the dry substance at the requested level. One situation when the feed forward system does not improve the performance is when there are variations in the pressure of the steam feed. An example of this can be seen in figure 3.5. The figure shows a situation when the pressure suddenly drops. This decreases the vaporization of water in the effect. When the controllers notice that the dry substance of the resulting syrup starts to decrease they open the valves more to increase the vaporization and the dry substance again. When the pressure begins to increase again the valves are opened a lot and the dry substance rises too much before the controllers get a chance to start closing the valves again. As mentioned, the issue in this situation is that if too much water is vaporized the syrup becomes a solid mass and the process needs to be shut down and cleaned.



**Figure 3.5** An example of when a variation of the steam feed pressure results in a decrease in dry substance and a following overshoot. When the pressure drops the controllers open the valves more. When the pressure rises again the valves are opened a lot and the dry substance level increases too much. It is worth noticing that a change of the ingoing dry substance happens to occur at the same time. Information about the signals can be found in appendix A.

# 4

## Modelling

*The modelling of the process is based on the mass and energy balances mentioned in section 2.2. To model the entire system these equations have been extended to be adequate for a multiple effect evaporator with thermal vapor recompression. Together with the dynamics of heat transfer and mixing of a solution in a tank these balances are the foundation of the Simulink model. To adjust the model to behave as the real process parameters have been tuned to fit the description in Chapter 3.*

### 4.1 Mathematical Modelling

By combining equations (2.1), (2.2) and (2.4) an expression for the vapor exiting the evaporator can be determined according to equation (4.1) below.

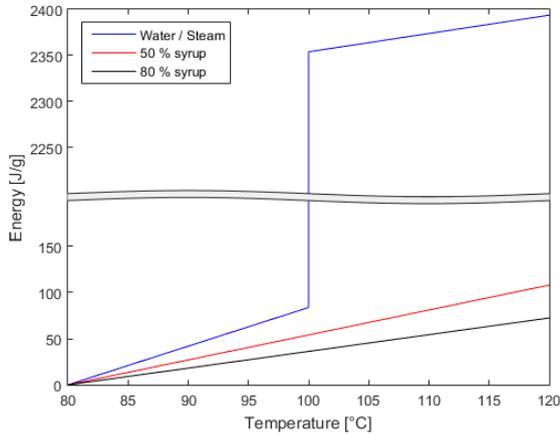
$$V = \frac{h_{S_{in}} - h_{S_{out}}}{h_V - h_{F_{out}}} \cdot S_{in} + \frac{h_{F_{in}} - h_{F_{out}}}{h_V - h_{F_{out}}} \cdot F_{in} - \frac{W_{losses}}{h_V - h_{F_{in}}} \quad (4.1)$$

Approximating no or at least very low thermal losses the last term in equation (4.1) can be eliminated. When looking at the second term one can see that the numerator will be much smaller than the denominator. This due to the enthalpy of the vapor being much larger than the enthalpy of the syrup. This can be seen in figure 4.1 and the second term can thus also be eliminated by approximation.

With these approximations the equation results in the following relation in equation (4.2), where the mass flow of vapor is proportional to the mass flow of steam. This proportion is an important part in the model, since this approximation eliminates the temperature and pressure dependencies.

$$V \propto S_{in} \quad (4.2)$$

As shown in figure 4.1 it takes much more energy to vaporize the water in the syrup than it takes to raise the temperature of the syrup to the boiling point. This together with equation (4.2) makes it possible to assume that almost all energy that leaves the steam is contained in the vapor after the effect. Equation (4.2) can then be written



**Figure 4.1** How the enthalpy increases with the temperature for water and syrup with 50 % and 80 % dry substance. At 100 °C the water vaporizes into steam. Due to the high enthalpy of vaporization for water the enthalpy of the steam is much higher than the enthalpy of the liquid syrup solutions.

as equation (4.3) where  $k$  is a constant close to but less than one. This constant can also account for heat losses.

$$V = kS_{in} \quad (4.3)$$

In the process the mass flow of steam going into the evaporators, after the preheater, is not measured. This value can be calculated when the system is at steady state with known mass flows and dry substances of the incoming and resulting syrup. Rewriting equations (2.1), (2.3) and (4.2) gives equation (4.4) which is the mass flow of steam needed to get a requested dry substance with specific inputs to an effect.

$$S_{in} = \frac{F_{in}(c_{out} - c_{in})}{kc_{out}} \quad (4.4)$$

The heat exchanger system is known to be a delayed first order process [MathWorks, 2015a]. This can be seen in equation (2.5). Supposing a linear relation between heat transfer energy and mass flow of steam, the transfer function from mass flow of steam to mass flow of vapor can be written as equation (4.5).

$$G_V = e^{-Ls} \frac{k}{sT + 1} \quad (4.5)$$

where  $L$  is the dead time,  $T$  the time constant and  $k$  is the transfer function gain. Since this is the transfer function from the mass flow of steam to the mass flow of vapor the transfer function gain will be the same as  $k$  in equation (4.3).

After leaving the heat exchanger the syrup enters the flash tank. This is modeled as a perfectly stirred tank with an inlet and an outlet. As the concentration changes it firstly has to pass through the tank, raising the total tank concentration, before the change is noticed throughout the system. This is described by the differential equation (4.6), which can be written as a transfer function, according to equation (4.7) below.

$$\dot{c} = \frac{q}{V_{tank}}(c_i - c) \quad (4.6)$$

$$G_c = \frac{1}{\frac{V_{tank}}{q}s + 1} \quad (4.7)$$

where  $V_{tank}$  is the tank volume and  $q$  the volumetric flow rate into the tank. There will also be similar dynamics for the mass flow through the tank. This is also approximated as a first order system, see equation (4.8).

$$G_F = \frac{1}{sT + 1} \quad (4.8)$$

To use the model of the whole evaporation process with three effects in series and thermal vapor recompression, an appropriate mass flow of steam needs to be used. To calculate this mass flow an extension of equation (4.4) that applies to the whole process is needed. This is given by equation (4.9), where  $u_{66}$  and  $u_{61}$  are the control signals from the controllers *A066* and *A061* respectively. The derivation of this equation can be found in appendix B.

$$S_0 = \frac{F_0(c_3 - c_0)}{c_3k \left( \left( \frac{8}{5}k + \frac{3}{5} \right) u_{66} + u_{61} \right)} \quad (4.9)$$

The modelling of the valves is kept as simple as possible and the control signals decide how many percents of the total steam feed that is fed into each effect. The characteristics of the valves discussed in section 3.2 are not a part of the model. The mass flows of steam entering the thermal vapor recompression system and effect 3 are given by equations (4.10) and (4.11) respectively. Since both *A066* and *A061* have upper limits to their control values and the sum of these will never be greater than one, the controllers can never send in more steam in the effects than what exists in the steam feed.

$$S_{TVR} = u_{61}S_0, \quad 0 \leq u_{61} \leq 0.50 \quad (4.10)$$

$$S_3 = u_{66}S_0, \quad 0 \leq u_{66} \leq 0.37 \quad (4.11)$$

The modelling of the thermal vapor recompression is based on the approximations in section 3.1 – the jet steam ejectors take as much vapor as they are feed with primary steam and that they have a 4:1 size ratio. The resulting steam flows are then given by the following equations.

$$S_{pre} = \frac{1}{5}(\min(V, S_{TVR}) + S_{TVR}) \quad (4.12)$$

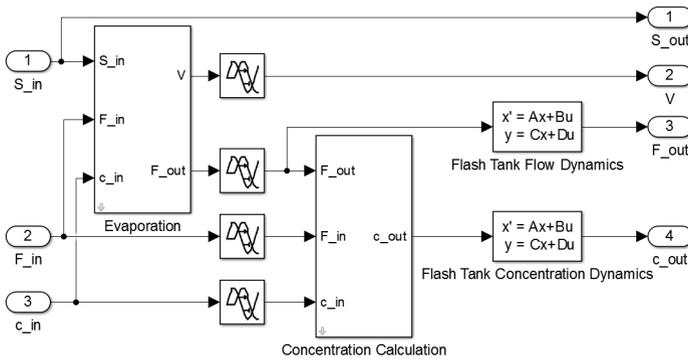
$$S_1 = \frac{4}{5}(\min(V, S_{TVR}) + S_{TVR}) \quad (4.13)$$

$$S_2 = V - \min(V, S_{TVR}) = \max(0, V - S_{TVR}) \quad (4.14)$$

## 4.2 Simulink Model

A model of an evaporator effect has been created in Simulink. The model is based on mass flows and the assumption that the syrup is at its boiling point when it reaches each evaporator. This means that the model is not depending on temperature or pressure.

To model the complete multiple effect system, a model representing a single effect is needed and it can be seen in figure 4.2. The input signals are the mass flow of steam,  $S_{in}$ , the mass flow of syrup,  $F_{in}$ , and the dry substance of the syrup,  $c_{in}$ . The first output signal is the mass flow of condensed steam,  $S_{out}$ , that leaves the evaporator. This value is simply the same as the mass flow of steam going into the evaporator, just as in equation (2.2). The other output signals are the mass



**Figure 4.2** The Simulink model of an evaporator effect. The effect consists of three parts – a block representing the dynamics of evaporation, a block calculating the new dry substance and two blocks representing the dynamics of the flash tank. All signals are also delayed to model the time the syrup spends in the effect.

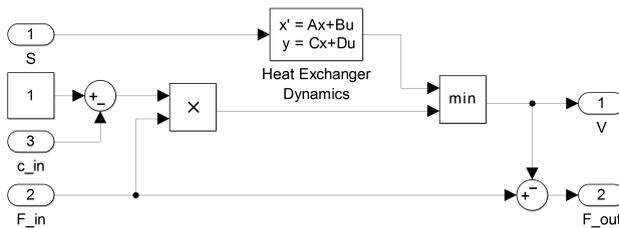
flow of vapor,  $V$ , that is vaporized from the syrup, the resulting mass flow of syrup,  $F_{out}$ , and the resulting dry substance of syrup,  $c_{out}$ .

The block *Flash Tank Concentration Dynamics* represents the mixing of syrup in the flash tank and the block was set according to equation (4.7). The change of the mass flow of syrup through the flash tank is also described by a state space block. This block is called *Flash Tank Flow Dynamics* and is based on equation (4.8). To calculate the dry substance of the syrup when the steam has vaporized some of the water the block *Concentration Calculation* is used. Here the dry substance is computed according to equation (2.3). All signals, except the steam, are passed through delay blocks that represent the time the syrup and vapor spends in the evaporator.

The *Evaporation* block contains the dynamics of the heat exchanger and can be seen in figure 4.3. The inputs to this block are the mass flow of steam,  $S_{in}$ , the mass flow of syrup,  $F_{in}$ , and the dry substance,  $c_{in}$ . The *Heat Exchanger Dynamics* block was set according to equation (4.5). The delay factor in equation (4.5) is represented by the aforementioned delay blocks in figure 4.2. To prevent the dry substance value from exceeding one which would mean that there is more than 100 % sugar in the syrup the mass flow of water in the syrup is calculated. If the mass flow of steam is greater than the mass flow of water that can be vaporized, simply all existing water is vaporized.

To model the total multiple effect evaporator process three effect models shown in figure 4.2 were connected as shown in figure 4.4. Two of the input signals are the steam flows through the valves controlled by *A061* and *A066*,  $S_{TVR}$  and  $S_3$ . The other two inputs entering the system are the mass flow of syrup,  $F_{in}$ , and the dry substance of the syrup,  $c_{in}$ . The outputs are the mass flow of the resulting syrup,  $F_{out}$ , and the resulting dry substance,  $c_{out}$ . The preheating heat exchangers seen in figure 3.1 are not part of the model more than by two delays. These delays, on the left side of *Effect 1*, represent the time it takes for the syrup to pass through the preheater. The delay after each effect represents the time it takes for the syrup to flow through the flash tank and the pipes to the following effect.

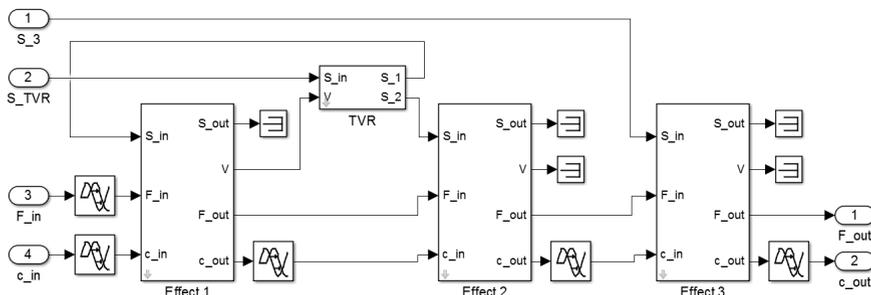
The *TVR* block handles the thermal vapor recompression and contains models of the two jet steam ejectors. The block is shown in figure 4.5. Here  $S_{in}$  is the mass



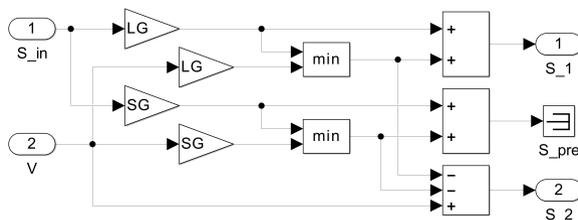
**Figure 4.3** The evaporation block containing the dynamics of the heat exchanger. There is also a system that makes sure that the upper limit of how much water that can be vaporized equals the water content of the ingoing syrup.

flow of steam that flows into the two jet steam ejectors as seen in figure 3.1.  $V$  is the mass flow of vapor from *Effect 1*.  $S_1$  and  $S_2$  are the mass flows of steam that flows into *Effect 1* and *Effect 2* respectively. The gains  $LG$  and  $SG$  represents how much steam the two jet steam ejectors use in regard to each other. As mentioned in section 3.1 the larger ejector takes four fifths of the steam and the smaller one fifth. Also mentioned in section 3.1 the ejectors take as much vapor from the first effect as they are fed primary steam from the valve. This is done as long as there is enough vapor. If there is not enough vapor the larger ejector takes four fifths of  $V$  and the smaller one fifth. In this case there is no vapor left for the second effect. The recompressed steam going to the preheater is included in the block diagram, but since the preheater is excluded from the model this signal is terminated and does not continue anywhere in the model.

The MATLAB scripts where all parameters are defined and from where the simulation is run can be found in appendix D.



**Figure 4.4** The multiple effect evaporator. Three effects and a block representing the thermal vapor recompression, of the vapor exiting the first effect, are connected. In between the effects there are sometimes transport delays and the unused signals are terminated.



**Figure 4.5** The thermal vapor recompression block. Both ejectors takes as much vapor as they are fed primary steam, but they are different in size. Therefore the different gains. Since the preheater is not part of the model the signal representing the mass flow of steam to the preheater is terminated.

### 4.3 Adjustments of the Model

To get the model to behave like the real process, as well as possible, all the knowledge of the process needs to be transferred into the model. The adaption of the model was considered with respect to a number of known behaviors, such as transport delays, gains and time constants throughout the system as well as simulating and comparing real process behaviors in the model. Everything that is known about the process is covered in appendix A.

Due to the lack of measurements the transfer function gain,  $k$  in equation (4.5), of the model of the heat exchanger was set to one. This is not realistic, but since the thermal efficiency, heat losses and other aspects of the effects are unknown, it is as good of an approximation as anything. In general, all three effects are modelled with the same properties due to the difficulties of measuring their individual dynamics.

#### Mass Flow of Steam

The model is based on the mass flows of syrup and steam. The preheater is not a part of the model but in the real process it uses some of the steam from the steam feed, see figure 3.1. It is not known exactly how much steam that flows into the effects, but the pressure of the steam feed is known. To get a translation between the steam feed pressure and the steam flow four steady state cases were analyzed. In table 4.1 the steady state values of the four cases are listed. Using equation (4.9) the steam flow needed to get the correct resulting dry substance in the model was calculated. The mean steam flow value from the four cases was then considered as the normal steam flow value and the result was  $S_0 = 0.54$  kg/s.

Together with the approximated steam flow a steam scale factor can be determined. The steam scale factor is simply a linear approximation of how much one

Case	1	2	3	4	Average	Unit
A066_PV	82.04	74.73	78.04	80.17	-	%*
A066_SP	82.00	74.71	78.00	80.19	-	%*
A066_LMN	24.50	28.20	27.40	35.00	-	%*
A061_LMN	42.40	29.30	40.40	45.80	-	%*
A069_PV	12.78	12.53	12.48	12.50	12.57	bar
A067_PV	46.35	49.24	51.32	49.24	49.04	%*
A058_PV	1.442	1.500	1.875	1.850	1.667	kg/s
Calc. steam flow	0.532	0.552	0.552	0.526	0.541	kg/s
Steam scale factor	0.042	0.044	0.044	0.042	0.043	(kg/s)/bar

\* The values are not scaled

**Table 4.1** Four operation cases when the process was running in steady state. The steam flow for each case has been calculated together with a scale factor between steam pressure and mass flow of steam.

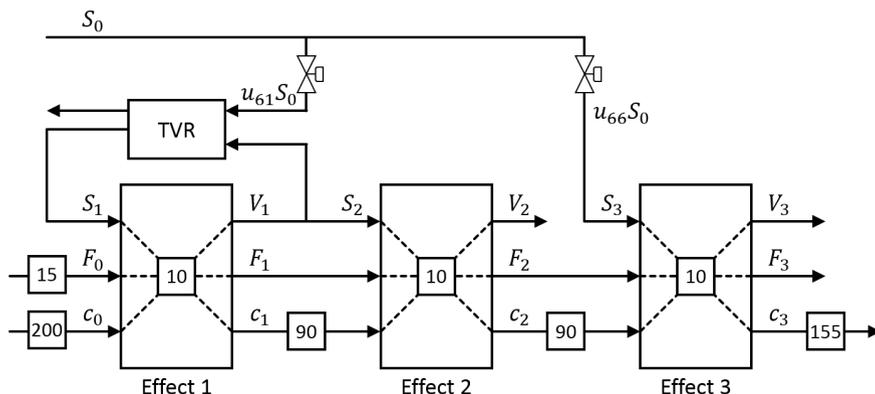
bar of steam pressure will give in steam flow. The average steam scale factor of the four cases was determined to 0.043 (kg/s)/bar, which also can be seen in table 4.1. When looking at trends of other cases the steam feed pressure has been translated into mass flow with this scale factor.

### Transport Delays

A change in mass flow of the ingoing syrup is known to spread much more rapidly throughout the system than changes in dry substance, 35 seconds compared with 400 - 500 seconds until the change has reached the third effect. In order to get these system properties right, firstly the focus was put on the shorter transport delay. The delay of the preheater is set to 15 seconds while the delay of each effect is set to 10 seconds, resulting in a 35 second delay until the ingoing mass flow of syrup affects the syrup going into the last effect. The different transport delays in the model are shown in figure 4.6 and can easily be added to get a hold of the total delay for different signals.

For a change in dry substance it has been chosen to set the total delay time to 400 seconds, of which 200 are assumed to take place in the preheater and 90 between each of the effects. This together with the 10 seconds in each effect results in a total delay of 400 seconds from the ingoing syrup to the beginning of the last effect.

From section 3.2 it is known that it takes 165 seconds for a change in the incoming steam flow to the last effect to affect the dry substance of the outgoing syrup. Therefore the delay after this effect is 155 seconds, due to the 10 second delay within the effect itself. Since there are no known transport delays within the steam feed or the steam flow from the feed to the effects, these flows are assumed to flow without any delays.



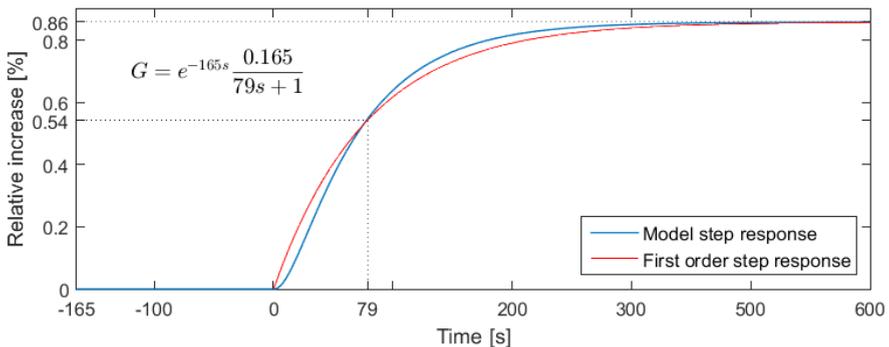
**Figure 4.6** A scheme of the model with all delays presented. The delays have been adjusted to fit the description in section 3.1.

## System Gains

As mentioned in section 3.1 there is a gain of 0.12 in the dynamics between the mass flow of steam into effect 3 and the resulting dry substance. When adjusting the model of the third effect to have the same gain, the model needs to be at the same operation point as when Lantmännen Reppe AB did their step response. It is however not clear at exact what operation point the step response was done. Even if not knowing the exact dry substance nor the mass flow of the ingoing syrup at the time, the step response itself can be tested. With the average values from the four cases in table 4.1 and the control signal of *A061* at 0.37 the same step response done in the model gives a gain of 0.165. Even though this is not the same gain as Lantmännen Reppe AB calculated from their step response, it is approximately of the same size.

## Time Constants

From the same step response as mentioned in the previous section a time constant of about 80 seconds was found. In each effect there are three transfer functions that had to be adjusted to give the same total time constant. The two blocks representing the dynamics of the flash tank were set to have the same parameters, both for the mass flow and the concentration. These time constants were set to 15 seconds since the volume in the tanks are quite low in the real process. The time constant of the evaporation dynamics was adjusted to 60 seconds. These time constants give a total time constant of approximately 79 seconds. This can be seen when comparing the model step response with a first order step response, see figure 4.7. Since the three effects are similar and there is no information about the time constants of the first two effects all three effects in the model were adjusted with the same parameters.



**Figure 4.7** A comparison of a model step response and a first order step response. The first order transfer function,  $G$ , has a delay of 165 seconds, a gain of 0.165 and a time constant of 79 seconds. The dotted lines show the time constant with its corresponding value and the end value.

# 5

## Control Strategies

*The present control strategy has been implemented in the model together with the feed forward system used in the real process. To be able to compare different control strategies all new strategies have been implemented in the same model. These strategies include feed forward systems using various signals and affecting the different controllers, but also a new sensor. The new sensor is measuring mass flow and dry substance of the syrup after the second effect and gives new alternative control strategies which also have been implemented.*

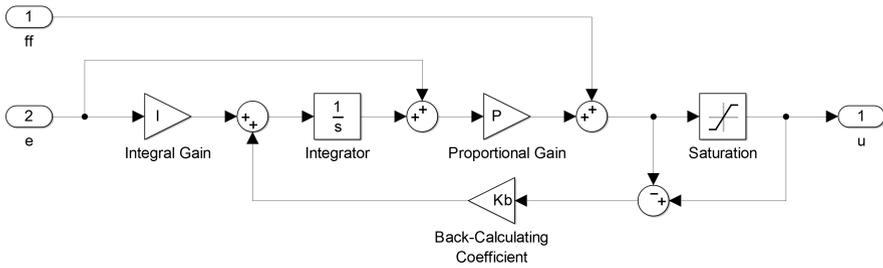
### 5.1 Introduction

Throughout this chapter the different control strategies are presented. The present control system described in section 3.2 is naturally one of them, but there are also new feed forward strategies and an implementation of a completely new strategy using a sensor, which does not exist in the present system.

All strategies are presented in the following sections. In order to minimize confusion and the risk of mix ups, each strategy has been assigned a letter and a color. The given letters and colors are attributes to the control strategies throughout the rest of the report. A summary of the different strategies, with these attributes, can be found in the end of this chapter, in section 5.5. The Simulink model containing all strategies and the MATLAB scripts where all control parameters are defined can be found in appendices C and D.

### Controller Implementation

The PI controllers used in the model have an anti-windup system activated since all control signals have limitations. When a feed forward signal is added to a controller it has to be included within the limitations of the control signal. Therefore a PI controller, where the feed forward signal is added before the saturation and anti-windup circuit, has been created, see figure 5.1. The signal denoted  $ff$  is the signal from the feed forward system.  $Kb$  is the back-calculation gain and was set to one.



**Figure 5.1** The Simulink model of a PI controller with an anti-windup circuit and a feed forward signal included.

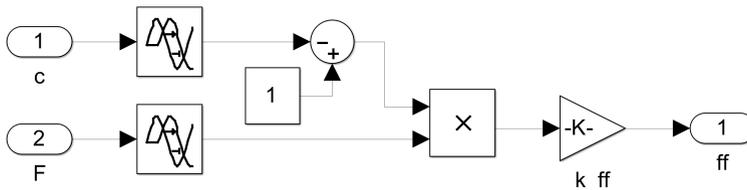
When there is no feed forward system added to a controller the signal  $ff$  is not included in the controller.

There are two different feed forward control systems among the strategies and both of them are proportional feed forward controllers. The first one is the present feed forward, which has the dry substance and the mass flow of the ingoing syrup as inputs. From these inputs the mass flow of water is calculated. The output of the feed forward system is calculated according to equation (3.1) and the block of the controller can be seen in figure 5.2. The signals  $c$  and  $F$  are the mass flow and dry substance of syrup and  $k_{ff}$  is the feed forward gain. Due to system latencies the signals are connected to transport delays.

The second feed forward system only uses the mass flow of steam as an input. The output is calculated according to equation (5.1). The general structure of this feed forward controller is the same as the previous, but has a negative gain since a rise in steam flow implies shutting the valves and vice versa. The block of this feed forward controller is shown in figure 5.3.

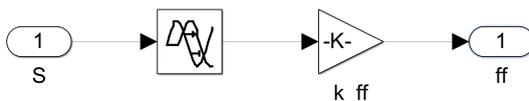
$$u_{ffs}(t) = k_{ffs}S_0(t - L_S) \quad (5.1)$$

A drawback with a proportional feed forward controller is however that the second overshoot, or undershoot, tends to increase in amplitude. Since the first of the two different feed forward control systems contains two inputs the gain of the con-



**Figure 5.2** The Simulink model of the present feed forward system. The mass flow of water is calculated using the mass flow and dry substance of syrup. The feed forward gain adjusts the control signal to be at the right size.

troller can be tuned either by changes in the dry substance or changes in the mass flow. Therefore two different gains have been determined for each of the following control strategies implementing this feed forward controller with two inputs. To be able to compare the different control strategies the feed forward gain of the present system was adjusted to reduce the first impact of a disturbance in either dry substance or mass flow by 30 %. All other control strategies are then compared with the present feed forward system by tuning their gains to match the undershoot of that system.



**Figure 5.3** The Simulink model of the steam feed forward system. The controller uses the mass flow of steam in the steam feed as input. The feed forward gain is negative for this controller.

## 5.2 Present System

As mentioned in section 3.2, one of the goals with the present control strategy is to focus more of the evaporation process to the first two effects due to their preferable steam economy. This has been accomplished by introducing the controller *A089*, creating a mid-range control system. *A089* is controlling the mass flow of steam into the thermal vapor recompression system through a cascade control loop together with the controller *A061*. By letting the control signal from *A066* form the measurement signal for *A089*, the first two effects can be controlled through *A061* to reduce the utilization of the last effect to a suitable or desirable level.

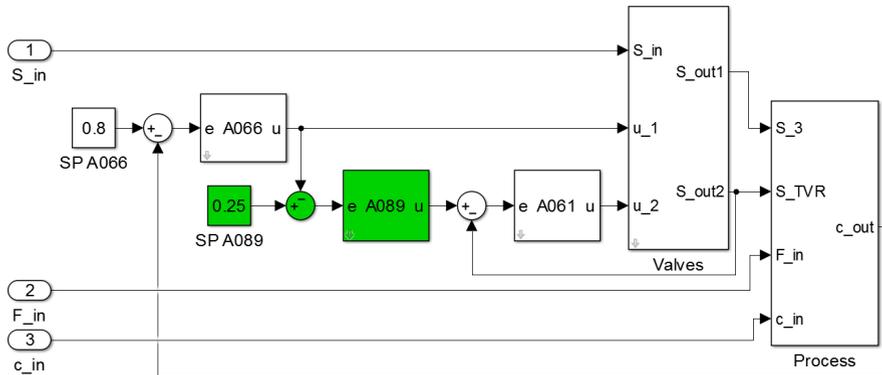
Almost the entire process can be seen as a black box. Basically, the only sensors used, when regulating the dry substance of the syrup, are the measurements of mass flow and dry substance of the ingoing and resulting syrup, as well as the pressure of the steam feed. Of course the different signals in connection with the controllers are logged as well. This implies a lot of unknown dynamics, which together with the long transport delays sometimes implicate a seemingly unreliable process.

To further improve the present control system a feed forward system has been implemented. This feed forward system is taking the dry substance and mass flow of the ingoing syrup into account, giving the system a heads up of changes that sometimes occur.

## A: Present Mid-Range Control

The Simulink model of the present control strategy, without the feed forward, is shown in figure 5.4. This figure can be compared with the previous figure 3.2.

The *Valves* block is sending the steam flows to the effects and these are computed according to equations (4.10) and (4.11). The limitations discussed in section 3.2 were set in the controllers so they have the same space to work in as the controllers in the real process. In the controllers there are anti-windup circuits activated. The parameters and limits of the controllers can be seen in table 5.1. These are based on the parameters used in the real process but tuned a bit to better fit the model. The controller *A066* is a PID controller in the real process but in the model it is a PI controller.



**Figure 5.4** The Simulink model of the present control strategy with the controllers *A066*, *A089* and *A061*. The *Valves* block represents the valves in the steam feed and the *Process* block represents the multiple effect evaporator.

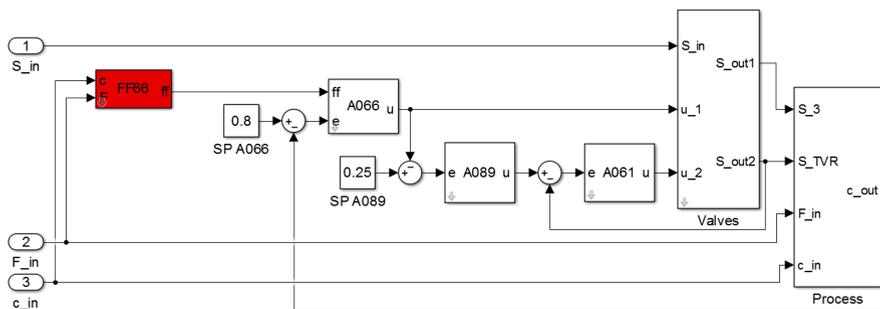
Controller	P	I	$T_i$	$u_{min}$	$u_{max}$
<i>A066</i>	1.3	0.0050	200	0	0.37
<i>A089</i>	-0.1	0.0025	400	0	0.70
<i>A061</i>	0.5	0.1000	10	0	0.50

**Table 5.1** The PI parameters of the controllers in the Simulink model of the present control strategy. The integral times,  $T_i$ , are given in seconds.

## B: Present Feed Forward Control to A066

To improve the control of this black box a feed forward system was integrated to the control strategy. The feed forward system uses the measurements of mass flow and dry substance of the ingoing syrup. In this way variations in these two signals can be compensated for when they enter the process. The output of the feed forward helps controlling the mass flow of steam fed to the last effect, see figure 5.5.

The feed forward signal affects the evaporation in effect 3 and therefore this signal needs to be delayed by the time the syrup spends in the first two effects. These delays were set to 35 seconds for the mass flow and 400 seconds for the dry substance. The results of the two different tunings, according to section 5.1, were to set the feed forward gains to  $k_{ff} = 0.78$  or  $k_{ff} = 0.41$ .



**Figure 5.5** The Simulink model of the present mid-range control strategy with a feed forward system using the mass flow of water in the ingoing syrup. The feed forward controller is connected to A066.

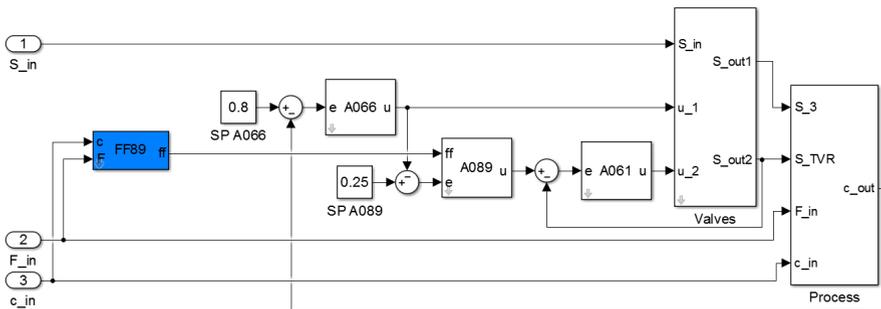
## 5.3 New Feed Forward Control

One way to improve the control of the process could be to use the signals, that are presently measured, in new ways. It is possible to implement a new feed forward system that affects  $A089$  in the same way as the present feed forward system affects  $A066$ . A possibility would then be an implementation where both these feed forward systems are used. Another option is to implement a feed forward system that uses the measurement of the steam feed to affect the control signal of  $A066$ . This measurement is not used in the present feed forward control strategy.

### C: Feed Forward Control to A089

To improve the robustness against changes in the incoming syrup a feed forward system can be added to the controller *A089* too, see figure 5.6. This would make it possible to compensate for changes of the inputs in an earlier stage of the process than in the strategy above.

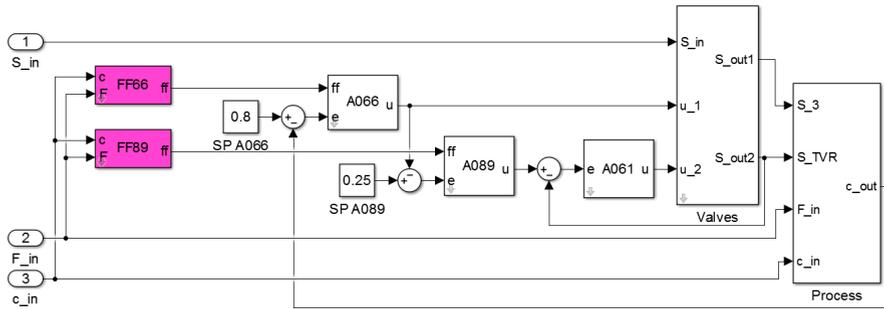
This feed forward system is also based on equation (3.1) and gives an offset to the control signal of *A089*. The signals has to be delayed in this system instead. The delays here represents the time it takes for the syrup to go from the sensor to the first effect. These were set to 15 seconds for the mass flow and 200 seconds for the dry substance. As for the present feed forward to *A066*, two gains were used. The gains for this strategy were adjusted to result in an undershoot equal to the one given by *A066*. This was done without having the feed forward system to *A066* activated. The gains were set to 0.54 or 0.15.



**Figure 5.6** The Simulink model of the present mid-range control strategy with a feed forward system using the mass flow of water in the ingoing syrup. The feed forward controller is connected to *A089*.

### D: Feed Forward Control to A066 and A089

It is possible to use the previously mentioned feed forward systems at the same time. In this way both *A066* and *A089* get a feed forward controller and the system gets two chances to compensate for the variations in the inputs before the output is affected, see figure 5.7. When using both feed forward systems at the same time the gains have to be decreased. The same ratio between the gains of strategy B and C was maintained for the combination of the strategies. The gains were for this strategy set to 0.52 or 0.30 for *A066* while 0.36 or 0.11 for *A089*.



**Figure 5.7** The Simulink model of the present mid-range control strategy with a feed forward system using the mass flow of water in the ingoing syrup. Feed forward controllers, with reduced gains, are connected to both *A066* and *A089*.

### E: Steam Feed Forward Control to *A066*

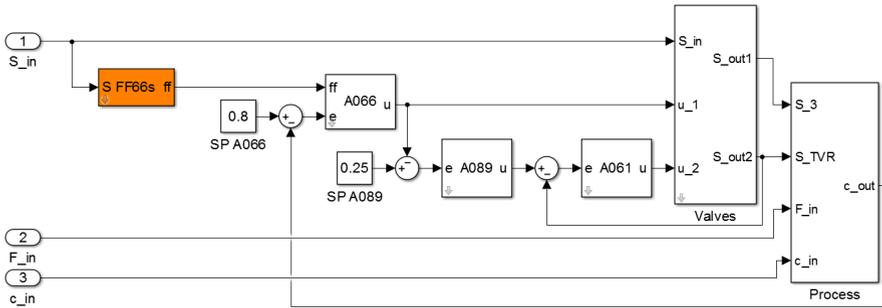
One problem with the process is when the pressure in the steam feed drops and then increases again. This is mentioned in section 3.3 and shown in figure 3.5. The problem occurs because if there are variations in the steam feed the controller *A066* does not compensate for it until a variation in the resulting dry substance has been measured. A way to compensate faster is to implement a feed forward system of the measured signal of the steam pressure, see figure 5.8.

This is a different feed forward system than the ones in the previous strategies, but is implemented in a similar fashion, see figure 5.3. It gives an offset to the control signal of *A066*. This offset increases the control signal when the pressure drops and decreases it when the pressure rises. Since the model does not include pressure this signal is proportional to the mass flow of the steam feed, see equation (5.1).

This control strategy would however work in a similar way in the real process where it could be a signal proportional to the steam pressure. Since no delays between the sensor in the steam feed and the evaporator are known there are no delays in the model either. This means that this feed forward system can work without delaying the signals. It would however be simple to add delays both in the model and in the feed forward system. The gain in this model was set to  $k_{ff_s} = -0.30$ .

## 5.4 New Sensor

Since the dry substance of the syrup is only measured at the end of the process it takes a long time before the controllers notice a change in any of the inputs. To improve the control of the process an option to consider is to install additional sensors in the process. One alternative is to install a sensor that measures the dry substance of the syrup that comes out of the second effect. With this sensor it is possible to get measurements from inside the black box and hopefully this information will be helpful when controlling the process. This new measurement signal can be used in different strategies.

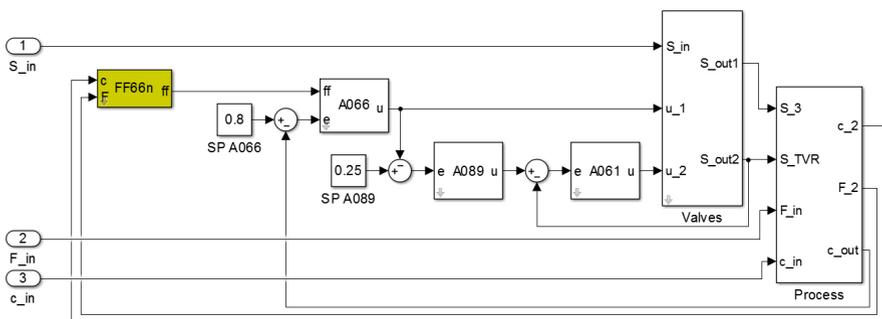


**Figure 5.8** The Simulink model of the present mid-range control strategy with a feed forward system using the mass flow of steam in the steam feed. The feed forward controller is connected to *A066*.

## F: New Feed Forward Control to A066

To keep the present control strategy intact the new sensor can be implemented into another feed forward system to *A066*. With the present feed forward system *A066* gets an offset proportional to the mass flow of water into the first effect. This new sensor allows the controller to compensate more correctly to changes of the inputs to the last effect, since the exact variations now are known.

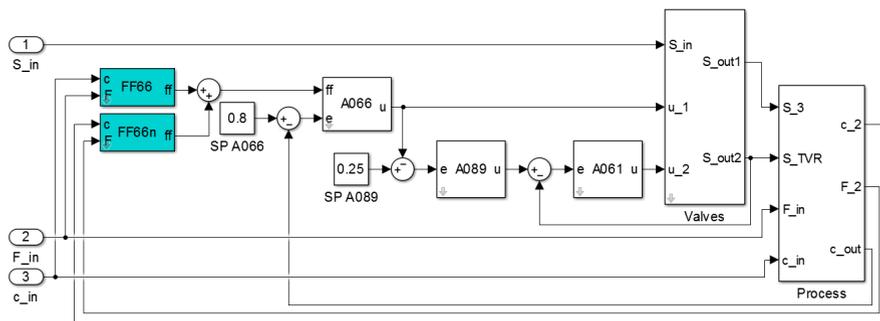
The model of this control strategy is shown in figure 5.9. In the model there is a delay of 90 seconds for dry substance between the last two effects. There is however no delay for the mass flow. Therefore a delay of 90 seconds was implemented for the dry substance in this feed forward system but there is no delay for the mass flow. Just as in strategy B, C and D there are two different gains for this strategy. The gains were for this strategy set to 2.49 or 1.69.



**Figure 5.9** The Simulink model of the present mid-range control strategy with a feed forward system using the measurements from a new sensor, located between the second and the third effect. The feed forward controller is connected to *A066*.

### G: Present and New Feed Forward Control to A066

The feed forward system in the previous section can be used together with the present feed forward system. As in strategy D, the gains have been decreased to some extent since two feed forward controllers now are working together. The gains were set to 0.55 or 0.31 for the present feed forward controller to A066 while 1.77 or 1.29 for the new feed forward controller. The gain ratio of the standalone strategies B and F are preserved when setting the gains of strategy G. The model of this strategy can be seen in figure 5.10.



**Figure 5.10** The Simulink model of the present mid-range control strategy with a feed forward system using both the present feed forward controller and the measurements from a new sensor, located between the second and the third effect. The two feed forward controllers are connected to A066.

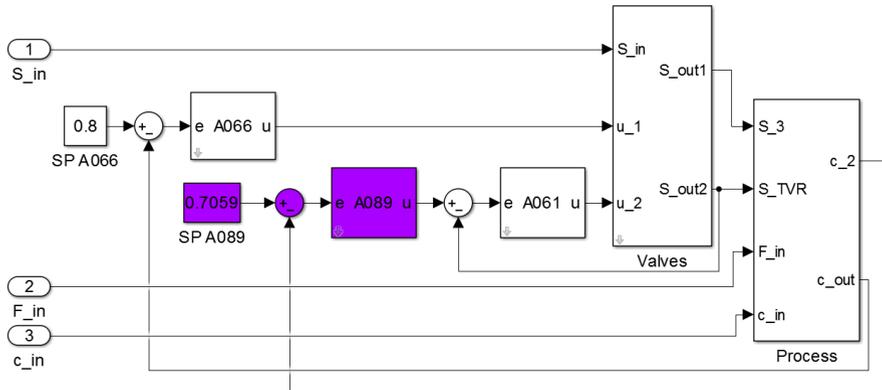
### H: New Feedback to A089

Another way to benefit from the new sensor would be to divide the process into two parts. The first part consisting of the first two effects, controlled by A089, and the second part the last effect, controlled by A066. This strategy would mean that the connection between the to valves is removed. Losing this connection will however not have to imply losing the goal to focus the evaporation to the first two effects. With the new sensor it will be possible to control the dry substance level of the syrup going into effect 3. The 25 % valve position to the last effect can, if desired, still be obtained, by choosing the setpoint of the new A089. With the mean values from the steady state cases in table 4.1 loaded as inputs and the present control strategy implemented, the dry substance before the last effect is  $c_2 = 70.6 \%$  at steady state. Using these mean values as inputs with the new strategy and 70.6 % as setpoint to A089, A066 sends a control signal of 25 %. With similar inputs the control signal would still be a value close to 25 %. The PID parameters of A089 in this case can be seen in table 5.2.

With this control strategy the present feed forward system to A066 is still useful, but also the new feed forward systems discussed in the previous sections. In figure 5.11 the model of this strategy is shown.

Controller	P	I	$T_i$	$u_{min}$	$u_{max}$
A089	0.09	0.007	143	0	0.70

**Table 5.2** The PI parameters of A089 with the dry substance after the second effect as measurement signal. The integral times,  $T_i$ , are given in seconds.



**Figure 5.11** The Simulink model of the new control strategy using the resulting dry substance from effect 2 as a measurement signal to A089. The new strategy breaks the present mid-range control, creating two different feedback loops.

## X: All New Strategies Included

One final strategy implementing all new strategies mentioned in previous sections has also been implemented. This strategy X includes the present feed forward system, the new feed forward systems and the new sensor, utilizing the sensor signals in a feed forward system as well as creating a new feedback loop. In context a combination of strategy D, E and G together with the feedback in H is used. The model where it is possible to choose a combination of all these systems can be seen in appendix C .

For this strategy, only tuning according to a step in the mass flow of syrup was conducted. This since it gives the best behavior. The gains of the different feed forward controllers are as in previous combined strategies modified to some extent. The gains of the feed forward controllers in strategy D were reduced by 50 %, the gain of the new sensor feed forward controller in strategy G remained the same, while the gain of the feed forward system in strategy E was adjusted from -0.3 to -0.35.

## 5.5 Summary

Since the different control strategies presented in this chapter might be fairly hard to grasp a summarizing table has been put together, see table 5.3. The table gives a condensed overview of the control strategies regarding their given attributes – color and letter – as well as the controllers included in each strategy.

Table 5.3 is also useful in the following chapters since the strategies are referred to according to their given letters and graphs are plotted in their corresponding colors.

	Present control		New feed forward		New sensor	
	Present mid-range control	Feed forward from $c_0$ and $F_0$ to A066	Feed forward from $c_0$ and $F_0$ to A089	Feed forward from $S_0$ to A066	Feed forward from $c_2$ and $F_2$ to A066	New feed-back from $c_2$ to A089
A	x					
B	x	x				
C	x		x			
D	x	x	x			
E	x			x		
F	x				x	
G	x	x			x	
H						x
X		x	x	x	x	x

**Table 5.3** A summarizing table, listing all the different control strategies with their attributed colors and letters as well as the controllers included in the strategy.

# 6

## Results

*The results of the project have been divided into two sections – model verification and a comparison of the control strategies, both present and new. The verification of the model is done by comparing a number of trends with plots from the model. The inputs of the model are adjusted to correspond to the course of events in the trend. To compare the control strategies they are simulated at the same operation point while meeting conditions concerning overshoots, undershoots and settling time.*

### 6.1 Model Verification

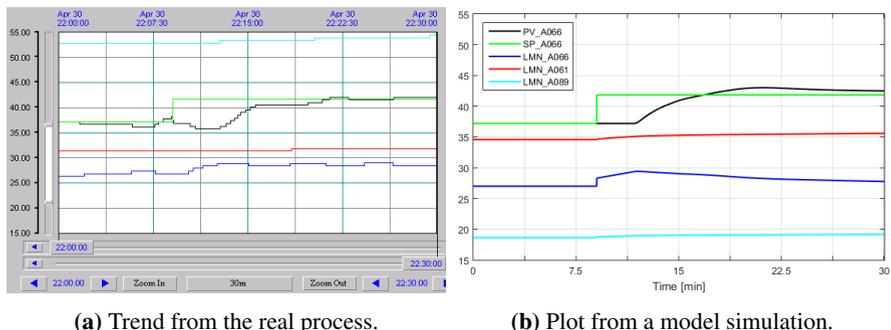
To verify if the model behaves similar to the real process several cases with known input signals were analyzed. There is no measurement data available from the real process and therefore the inputs have been imitated as well as possible. In these cases the present control strategy was used to make it possible to compare the performance of the model to the real process. Since it is not known if the feed forward system was used in the real process it was not used in the model verification either. The resulting dry substance was then plotted together with the control signals and other system inputs. These graphs were then compared to the trends from the real process.

For the model verification the interesting signal variations are changes in the different input signals – dry substance and mass flow of the ingoing syrup as well as changes in the steam feed – but also step responses when changing the setpoint value of the resulting product.

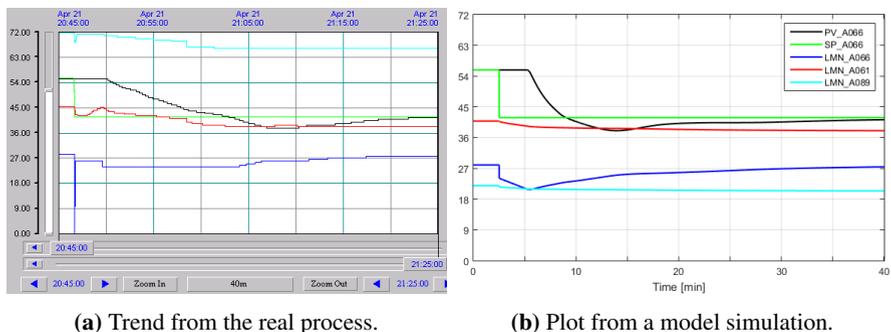
Notice that the axes of the plots from the model do not necessarily match the axes of the trends from the real process. All y axis values are in percent with the scales used by Lantmännen Reppe AB, see appendix A.

### Change of Setpoint

The results from two set point changes can be seen in figure 6.1 and figure 6.2. In figure 6.1 a positive step response is compared, while a negative step response is compared in figure 6.2.



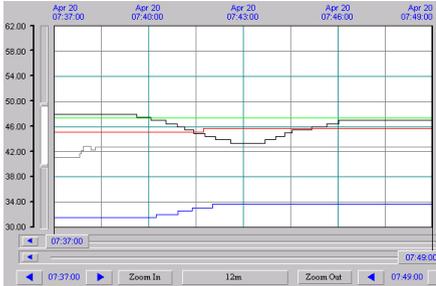
**Figure 6.1** An increase of the setpoint from 37.2 % to 41.9 %, which corresponds to 78 % and 79 % in real dry substance values. All signals with descriptions and corresponding scale intervals are listed in appendix A.



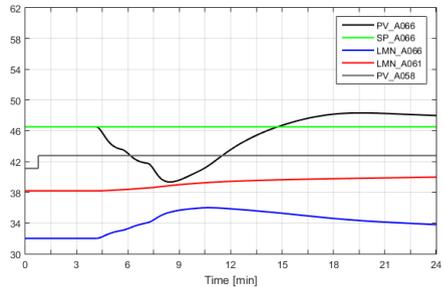
**Figure 6.2** A decrease of the setpoint from 55.8 % to 41.9 %, which corresponds to 82 % and 79 % in real dry substance values. All signals with descriptions and corresponding scale intervals are listed in appendix A.

## Variations of the Ingoing Syrup

The properties of the ingoing syrup can be adjusted by changing one of two variables, the mass flow and the dry substance of the syrup. In figure 6.3 a positive step response of the mass flow is shown, while in figure 6.4 a decrease of the dry substance is shown.

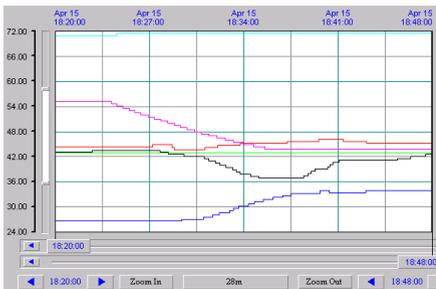


(a) Trend from the real process.

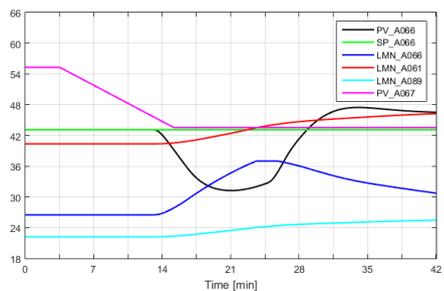


(b) Plot from a model simulation.

**Figure 6.3** An increase of the mass flow of syrup from 41.1 % to 42.8 %, which corresponds to 6165 kg/h and 6420 kg/h in real mass flow values. All signals with descriptions and corresponding scale intervals are listed in appendix A.



(a) Trend from the real process.

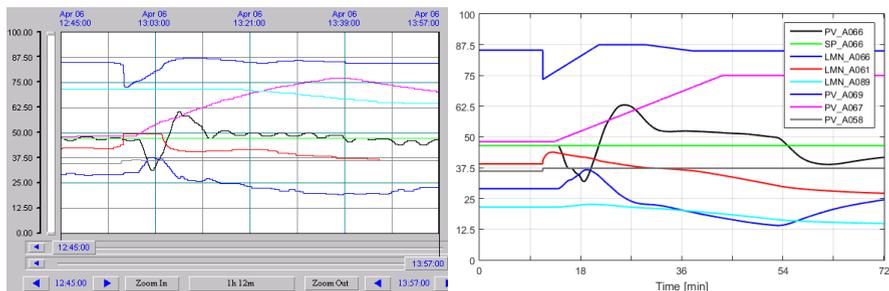


(b) Plot from a model simulation.

**Figure 6.4** A decrease of the dry substance of syrup from 55.3 % to 43.5 %, which corresponds to 48.8 % and 44.8 % in real dry substance values. All signals with descriptions and corresponding scale intervals are listed in appendix A.

## Variations in Steam Pressure

A pressure drop in the steam feed is shown in figure 6.5. The drop comes suddenly and then slowly recovers back to the original pressure level. In the meantime the dry substance of the ingoing syrup increases.



(a) Trend from the real process.

(b) Plot from a model simulation.

**Figure 6.5** A decrease in the steam feed pressure from 85.3 % to 73.4 %, which corresponds to 12.8 bar and 11.0 bar in real steam pressure values. An overshoot at 87.5 % (13.1 bar) occurs when the steam flow is recovering. All signals with descriptions and corresponding scale intervals are listed in appendix A.

## 6.2 Comparison of Control Strategies

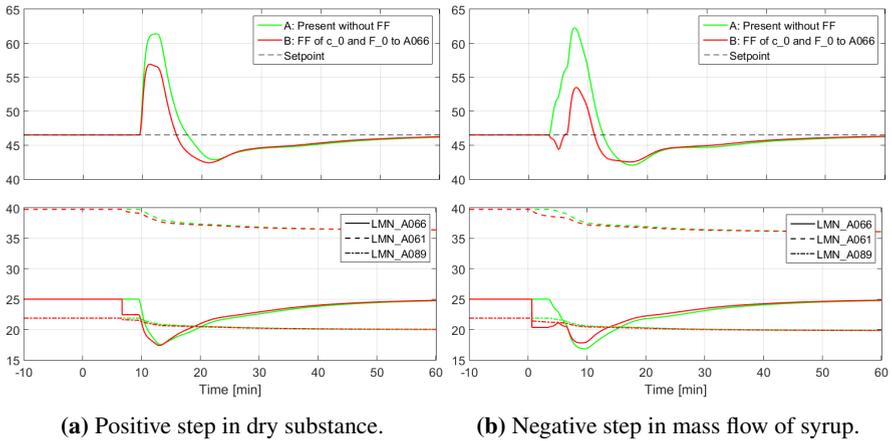
When doing the comparisons of the different control strategies, they are compared in the same plots in order to easily see what differences they make. The strategies are compared two or three at a time.

For all of the following plots, the average case input signals in table 4.1 have been used in order to achieve an equal comparison between the strategies. The upper plots in each figure contain the step responses of the measurement signal. This signal is always the dry substance of the resulting syrup in percent. These are plotted with the scale that Lantmännen Reppe AB uses in their trends. 80 % dry substance corresponds to 46.5 % in the plots. The lower plots contain the control signals, in percent, from all three controllers of each case. The signal from *A066* is represented with a solid line, the signal from *A089* with a dash-dot line and the signal from *A061* with a dashed line. When the strategies are compared each strategy is plotted with the same color as it was assigned in chapter 5.

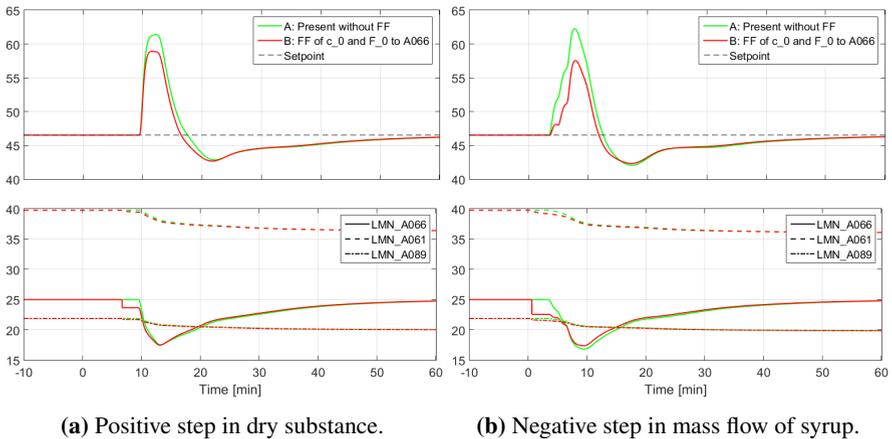
The steps of the input signals have the same sizes in all cases. A step in dry substance starts at 50 % and ends at 52 %, in real dry substance values. The mass flow of syrup is decreased from 5940 kg/h to 5508 kg/h. These have been chosen to affect the resulting dry substance equally. A step in the mass flow of steam goes from 1980 kg/h to 2340 kg/h. The steps are occurring at time zero in the plots.

### A-B: Present Control Strategies

A comparison of the present mid-range control strategy, with and without the present feed forward system. The green graphs are from simulations with strategy A, while the red are from strategy B.



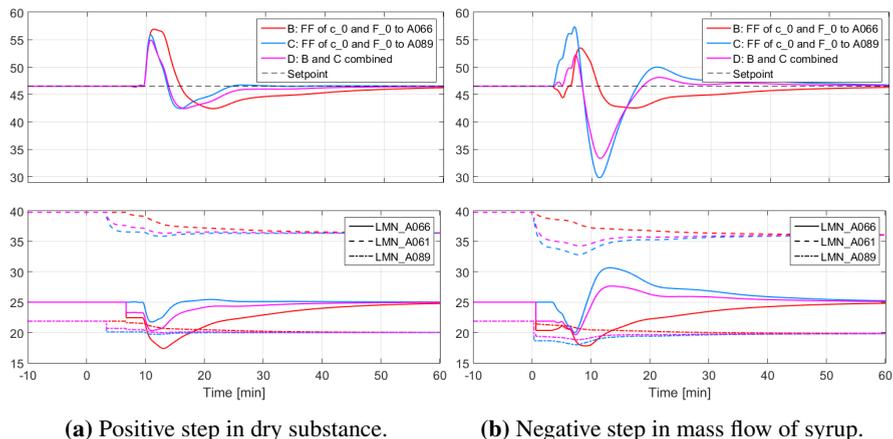
**Figure 6.6** Present mid-range control strategy, with and without the present feed forward system. In this case the feed forward gain has been tuned to reduce the impact of the step in dry substance by 30 %.



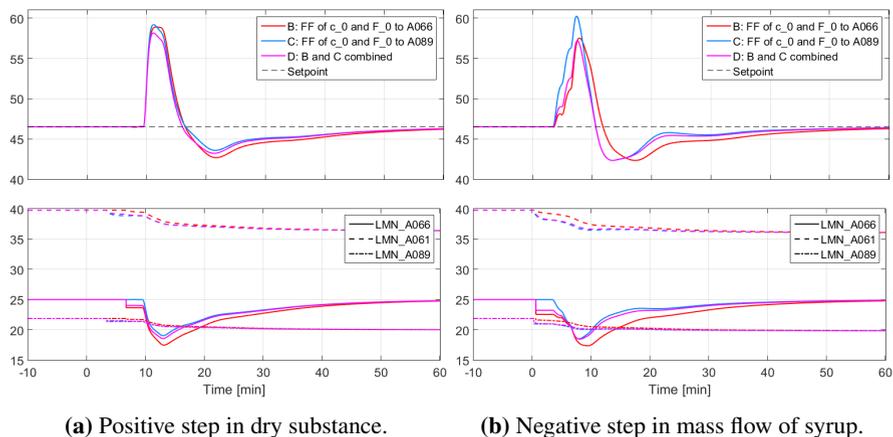
**Figure 6.7** Present mid-range control strategy, with and without the present feed forward system. In this case the feed forward gain has been tuned to reduce the impact of the step in mass flow of syrup by 30 %.

### B-C-D: Feed Forward Control to A066 and A089

A comparison of the present mid-range control strategy with feed forward control to A066, A089 or to both. The red graphs are from simulations with strategy B, the blue are from strategy C, while the magenta are from strategy D.



**Figure 6.8** Feed forward control to A066, A089 or to both using the measurements of the ingoing syrup. The feed forward to A066 has been tuned to reduce the impact of the step in dry substance by 30 %.

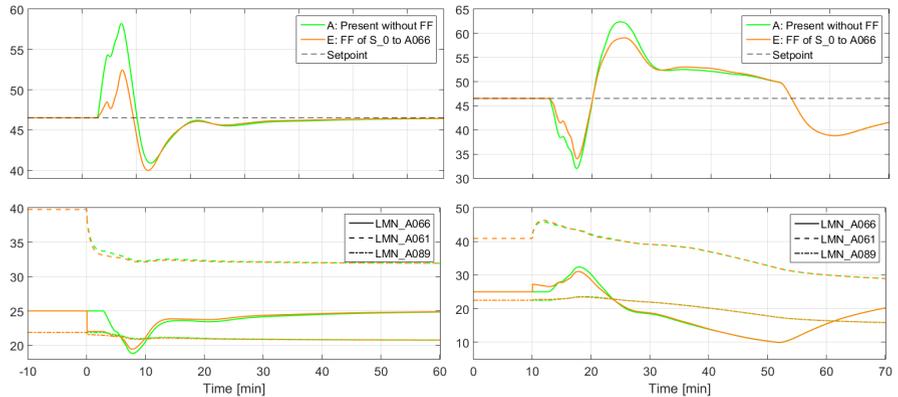


**Figure 6.9** Feed forward control to A066, A089 or to both using the measurements of the ingoing syrup. The feed forward to A066 has been tuned to reduce the impact of the step in mass flow of syrup by 30 %.

### A-E: Steam Feed Forward Control to A066

A comparison of the present mid-range control strategy with and without steam feed forward control to A066. The green graphs are from simulations with strategy A, while the orange are from strategy E. The left graphs are the step response from a positive step in mass flow of steam. The right graphs are the response to a simulation of the drop and slow increase of steam pressure in figure 6.5.

As mentioned in section 3.3 the present control does not take changes of the steam feed into account. This means that strategy B would give the same step response as strategy A.



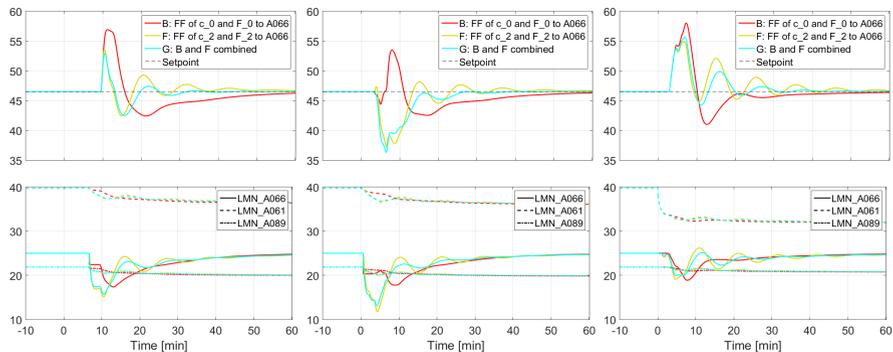
(a) Positive step in mass flow of steam.

(b) Drop and slow rise of steam pressure.

**Figure 6.10** Positive step response in mass flow of steam with and without a feed forward of the signal to A066 as well as the drop and slow rise of the mass flow of steam from figure 6.5 with and without a feed forward signal to A066.

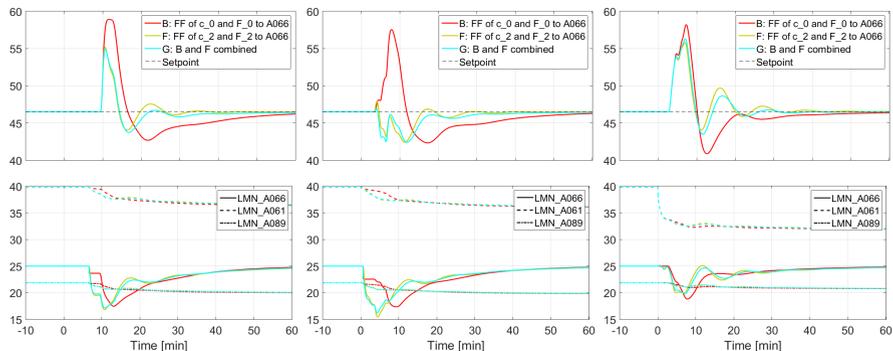
### B-F-G: New Feed Forward Control to A066

A comparison of the present mid-range control strategy with feed forward control to A066, new feed forward control to A066 or a combination. The red graphs are from simulations with strategy B, the yellow are from strategy F, while the cyan are from strategy G. The new sensor is implemented in strategy F and G.



(a) Positive step in dry substance. (b) Negative step in mass flow of syrup. (c) Positive step in mass flow of steam.

**Figure 6.11** Feed forward control to A066 using both the measurements of the ingoing syrup and from the new sensor. The feed forward to A066 has been tuned to reduce the impact of the step in dry substance by 30 %.

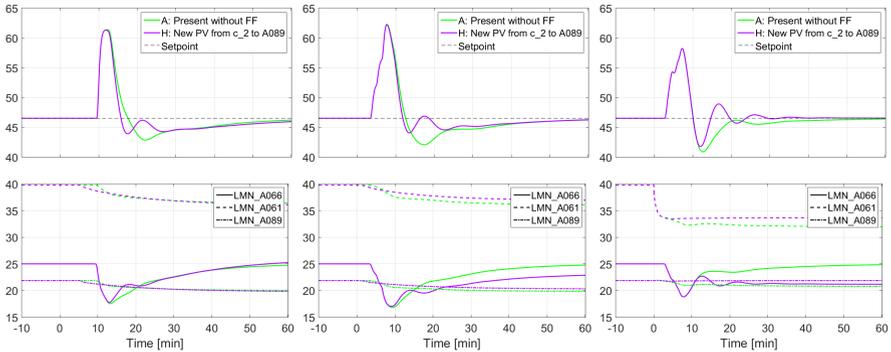


(a) Positive step in dry substance. (b) Negative step in mass flow of syrup. (c) Positive step in mass flow of steam.

**Figure 6.12** Feed forward control to A066 using both the measurements of the ingoing syrup and from the new sensor. The feed forward to A066 has been tuned to reduce the impact of the step in mass flow of syrup by 30 %.

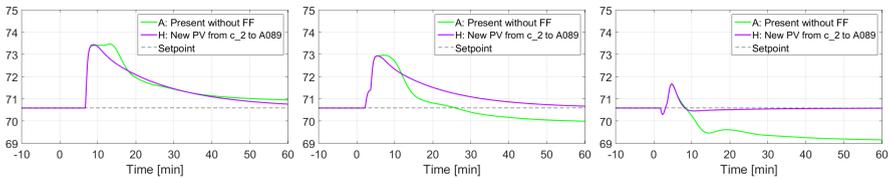
### A-H: New Feedback to A089

A comparison of the present mid-range control strategy and a new strategy that regulates the dry substance of the syrup coming out of the second effect. The green graphs are from simulations with strategy A, while the purple are from strategy H. The new sensor is implemented in strategy H. In figure 6.14 the dry substance after the second effect is also shown.



(a) Positive step in dry substance. (b) Negative step in mass flow of syrup. (c) Positive step in mass flow of steam.

**Figure 6.13** Present mid-range control strategy without feed forward control together with the new strategy regulating the dry substance of the syrup coming out of the third effect.

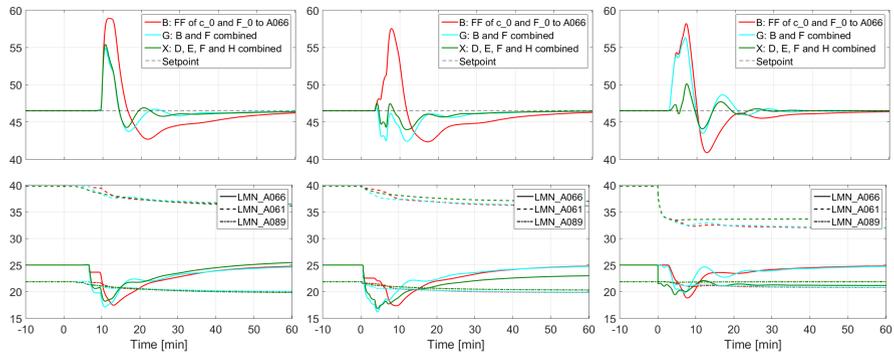


(a) Positive step in dry substance. (b) Negative step in mass flow of syrup. (c) Positive step in mass flow of steam.

**Figure 6.14** Present mid-range control strategy without feed forward control together with the new strategy regulating the dry substance of the syrup coming out of the second effect. This plot shows the dry substance out of the second effect, which now can be measured due to the new sensor. The dry substances are real and not scaled values.

### B-G-X: All New Strategies Included

A comparison of strategy G and strategy X, which uses all new control systems, has also been conducted. Strategy B has been included in order to compare it with strategy X as well. The red graphs are from simulations with strategy B, the cyan are from strategy G, while the dark green are from strategy X.



(a) Positive step in dry substance. (b) Negative step in mass flow of syrup. (c) Positive step in mass flow of steam.

**Figure 6.15** Present mid-range control strategy with feed forward control, the strategy using both feed forward systems to A066 and the strategy using all new control systems.

# 7

## Discussion

*This chapter initially follows the same structure as the results chapter. The discussion therefore begins with two sections covering the model verification and the comparison of the control strategies. How to proceed after this master's thesis and what further investigations or implementations that are reasonable are eventually also discussed in this chapter.*

### 7.1 Model Verification

The comparison of the plots from the model and the trends from the real process gives an understanding of the model behavior. There are both cases where the model gives results similar to the process and where the model differs more. Visual differences – latencies, time constants, gains, overshoots and undershoots – are discussed below. Both measurement signals and control signals are considered.

Considering the latencies, the process and the model simulations are about the same size. The delay times of the model are always constant even though the delay times of the real process vary with especially the mass flow of the ingoing syrup. It can be seen, when comparing the trends and the simulation plots, that the delay times only differ with approximately  $\pm 100$  seconds. This can be considered fairly small and shows that the estimations for normal operation done by Lantmännen Reppe AB are sufficient even during more general operation.

By approximating the process as a first order system it is possible to consider what time constants that would correspond to the different cases. The setpoint changes in figures 6.1 and 6.2 show that the model seems to have a smaller time constant than the real process. The input variations in figures 6.3 and 6.4 however show that the model is slower than the real process. This could perhaps be improved by reducing the time constants of the transfer functions for the flash tanks, see equations (4.7) and (4.8), and increasing the time constants of the transfer functions for the heat exchangers, see equation (4.5). These are tuned to fit the time constant of 80 seconds that has been estimated for a step response from A066 to the dry substance. This could be achieved with other time constants as well. A decrease of the time

constant of the flash tank dynamics would however correspond to having a very low volume in the tank. One option could also be to not have exactly the same dynamics in all three evaporators. This was not done since only the dynamics of the last effect have been estimated. It is hard, if even possible, to estimate the dynamics of the first two effects with the measurements that are currently available.

Looking at figures 6.3 and 6.4, the resulting dry substance gets affected more in the model than in the real process, which gives rise to a longer settling time as well. These results show that there are room for improvements, even though the basic behaviors are initially corresponding correctly. One option could be to adjust the gains of the transfer functions representing the flash tanks, see equations (4.7) and (4.8). These are set to one in the model to satisfy the mass balances at steady state, but perhaps a lower gain would result in a better corresponding model. In figures 6.3 and 6.4 it is also clear that the model gives overshoots that are not seen in the trends. This probably occurs as a result of the mid-ranging control, when it aims to return the control signal of *A066* to the setpoint of *A089*. A deeper understanding of the implementation of the real mid-range control is desirable to be able to model the process control even better.

Another general understanding from the comparison of trends and simulations is that the control signals from *A066* and *A061* in the model are behaving mostly the same way as in the real process. A couple of differences are however noticeable. Firstly the control signal of *A066* gives a rapid negative spike response in the trend when the setpoint is decreased, which is not seen in the simulation plot. This could be because of the implementation of *A066* in the real process. The negative spike might occur in the trend due to the derivative part being connected to the error. If so, the derivative part should be connected to the measurement signal instead, since the rapid changes in control signal will imply large valve wear [Hägglund, 2011]. Secondly the control signal of *A066* also seems to be unaffected of the mid-ranging control. This can be seen in many of the trends since the control signal of *A066* does not return to the same steady state value as before. Whether the mid-ranging control, for some reason, might not be set or if the control of the real process is much slower than the control of the model is uncertain. The control parameters in the model are fairly equal to the ones in the real process and should not be causing these distinct differences. The control signal from *A089* is on the other hand always settling at a lower value in the simulations than in the trends. This might correspond to the lack of valve dynamics. The model uses only valve positions and no pressure, which probably gives rise to missing dynamics and therefore imperfections in the amplitude of the control signal. In the model, the controller *A061* is more or less without any influence due to the missing pressure and valve dynamics.

The final part of the model verification is perhaps the most satisfying. In figure 6.5 the overall behavior of the model corresponds very well to the trend. The alternations in the resulting dry substance are very similar. Some oscillatory behavior is not captured and the model also gives a decreased resulting dry substance after the ramp of the ingoing dry substance has flattened out. The control signals of both

*A066* and *A061* are saturated in the trend but not in the simulation, but except from this they look much like the control signals in the trend.

## 7.2 Comparison of Control Strategies

The different control strategies are discussed according to the figures of the results. Here each section directly corresponds to the same section in the results chapter. The final and general conclusions concerning the control strategies can be found in chapter 8.

### A-B: Present Control Strategies

When comparing the present system with and without the feed forward system connected it is clear that by adding the feed forward the results are significantly improved. The improvement can be seen in the step responses for both the incoming dry substance and mass flow. By looking at the plots of the control signals it is clear how the feed forward signal contributes with a step in the control signal of *A066*. Both *A089* and *A061* are affected secondarily but not as much.

Regarding the cut off of the peaks, the feed forward affects a change in mass flow more than a change in dry substance. In other words, the impact of the feed forward due to changes in mass flow is larger than the impact due to changes in dry substance. This is expected since the actual value that is fed forward is the amount of water in the syrup. A problem with this distorted impact is that a gain resulting in a better control at changes in the dry substance might be too large regarding changes in the mass flow. As can be seen in figure 6.6, the chosen gain results in a negative step response initially. It might be problematic if this negative response gets too significant since it in the same way creates a positive response for a positive step in syrup mass flow. To be able to have more equal impacts at the different step responses it is possible to weight the signals when calculating the amount of water. It is also possible to completely separate the signals, creating two different feed forward controllers and therefore have two separated gains.

One disadvantage with a proportional feed forward system is that the undershoot sometimes gets deeper with the feed forward than without. In figure 6.6 this happens for a change in dry substance but not for a change in mass flow. In this case the difference between the undershoots is fairly small, but it is still an important aspect to keep in mind when introducing a proportional feed forward. If the step would be in the opposite direction the undershoot would instead be an overshoot and large overshoots are exactly what is unwanted.

In figure 6.7 a reduced feed forward gain is used and instead 30% of the mass flow step response is cut off. It is clear that the unwanted effects are gone, but the price to pay is a reduced effect. Both figures 6.6 and 6.7 have their strengths and weaknesses. In order to determine which one is the better all aspects have to be considered and evaluated.

### **B-C-D: Feed Forward Control to A066 and A089**

In this case the strategy with a feed forward signal to A066, B, is compared to having a feed forward signal to A089, C, instead. The third alternative is to have a combination of these feed forward systems, D. In figure 6.8, where the systems have been tuned for a step in dry substance it is clear that the gain makes strategy C better than B in the case of a step in dry substance. This however gives a much bigger undershoot after a step in mass flow, without decreasing the first impact compared to strategy B. The time it takes to reach the setpoint after the steps have improved in both strategy C and D. Strategy D, the combination, is more similar to C than to B. It decreases the impact of the step in dry substance but is a little slower than strategy C. It also decreases the impact of the step in mass flow, has a smaller undershoot than strategy C and is the fastest of the three. The control signals correspond to the measurement signals. When both controllers have feed forward systems they share the burden of compensating for the disturbances of the inputs. This could be positive regarding valve wear.

In figure 6.9 it is seen that strategy C is not being able to cut off the first impact of the mass flow step response without obtaining larger undershoot than B. Strategy C is however a little faster than B. Strategy D seems to be a good alternative to decrease both the impact, the undershoot and the settling time. From the control signals the same conclusion as above can be drawn. The differences are smaller in this case since all feed forward gains are smaller.

A summarized conclusion from comparing strategy B, C and D is that a combination of feed forward systems to A066 and A089 is a promising alternative. In this way both controllers can compensate for disturbances when it reaches the effects respectively. Since the impact from a step in mass flow affects the feed forward signals the most it seems to be safer to tune the gain with this in mind. It also seems that the feed forward system to A089 tends to be harder to tune to get a decrease of the impact and at the same time not get too big undershoots. This sensitivity might be occurring due to A089 controlling two effects and A066 only one. These results show that with a combination of the feed forward systems it is good to let the one to A066 have a bigger impact than the one to A089.

### **A-E: Steam Feed Forward Control to A066**

Since the present feed forward system only takes properties of the ingoing syrup into account it does not compensate for changes in the steam feed pressure, which is one of the mentioned problems in section 3.3.

One solution to this problem is to add similar feed forward control as the present but to use the steam pressure to affect A066. By doing so the present system would be able to adapt faster considering changes of all the inputs of the system. It is worth noticing that even though the feed forward of the steam pressure is disconnected, the control signal of A061 still gets affected. The reason for this is that A061 controls its valve regarding steam pressure rather than just the position of the valve, which is

the case of *A066*. As a matter of fact, *A066* was previously in a cascade control like the one with *A089* and *A061*. *A066* was then acting as *A089* is today and another controller, called *A062*, was controlling the valve by regulating the pressure. The issues with changes in the steam feed might have worsened after removing the previous pressure controller *A062*. By reintroducing this controller the system might be more robust when it comes to changes of the pressure in the steam feed. A feed forward from the steam feed might nevertheless further improve the control.

Just as when comparing strategy A and B, the impact of the feed forward can clearly be seen in the plots of the control signals and that the undershoot in the measurement signal gets deeper with the feed forward connected.

In figure 6.10b the same scenario as in figure 6.5 is shown. Here the response of strategy E is certainly better than strategy A. It can be seen that strategy E reduces both the first undershoot and the first overshoot. Strategy E has got a slightly higher dry substance value between about 30 and 50 minutes in the plot, which has to do with what has been discussed about the undershoot in figure 6.10a. This is however almost irrelevant due to the tiny difference between the strategies.

Another finding when looking at figure 6.10b is that the final undershoot has nothing to do with the change in steam feed pressure. This is certain since both strategies come together and thus the undershoot must be originating from changes in another signal. As mentioned in section 6.1 the undershoot occurs due to a ramp change of the dry substance of the ingoing syrup, which can be seen in the more detailed figure 6.5.

### **B-F-G: New Feed Forward Control to A066**

Installing a new sensor between the second and last effect is based on the lack of information about the changes in the properties of the syrup along the process. Since the present control is dividing the process into two parts in series, where the first part consists of the first two effects and the second part the last effect, it seems natural to place the sensor in between the two parts. The new sensor would provide details about the dry substance as well as the mass flow of the syrup and thus a good check up along the way. Using these new measurements in a feed forward to *A066* would implicate a better feed forward, F, since it actually forwards the exact syrup properties that goes into the last effect.

When comparing the different strategies in figure 6.11 it shows to be mainly as expected. Strategy F gives a larger cut off than strategy B, without resulting in a larger undershoot. Strategy F also speeds up the control and thus finds the setpoint faster. It does create an oscillatory behavior, which can be seen as a disadvantage. When comparing figures 6.11 and 6.12 it is however clear that the oscillations can be reduced significantly by reducing the feed forward gain. The difference in cut off is very small compared to the more non-oscillatory behavior. The smaller gain is actually more desirable when comparing figure 6.11b with figure 6.12b, due to the larger impact of a change in mass flow, as discussed in previous sections.

Since the steps of the properties of the ingoing syrup has been passing through the first two effects, the dynamics of the effects have transformed the step into a more smooth change. This can be seen when looking at the control signals of figures 6.11 and 6.12. The impact of strategy B on the control signal of A066 is affecting the signal as the original step going into the process, while strategy F has the softer characteristics of the syrup leaving the second effect. It is also clear that strategy F has got an impact on changes in the pressure of the steam feed, in figures 6.11c and 6.12c, which strategy B does not provide.

Strategy G is using the combined feed forward controllers in strategy B and F. One could perhaps find it a bit strange to have both these feed forward controllers connected at the same time since the two controllers essentially have the same purpose. But by taking advantage of the rapid change of strategy B and the more soft change of strategy F, the resulting strategy G actually has some interesting characteristics. Even though very similar, the general differences between strategy F and G are that strategy G is somewhat slower but slightly more damped. It is hard to say which one is more favorable just by looking at figures 6.11 and 6.12.

To install a new sensor might be a good solution to improve the control of the process. The installation does however implicate both direct and indirect costs. The sensor itself will cost, but perhaps more significant are the costs considering stopping, restructuring and restarting the production and the system. Depending on both the economic and the control beneficial aspects it might be more or less motivated to install a new sensor. One possibility could be to add dynamics to the feed forward system that is using the inputs to the first effect. These dynamics could be a model of the first two effects. This would result in a feed forward system corresponding better to the inputs of the last effect. However this would demand some system identification of the first two effects and if some part of the system would change or be replaced, the model might not correspond as well anymore.

### **A-H: New Feedback to A089**

Another alternative with adding a new sensor is to control the dry substance of the syrup after the second effect. The comparison of the present mid-range control, A, and this new strategy, H, can be seen in figure 6.13. Both step responses from dry substance and mass flow give similar differences between the strategies. Strategy H tends to give smaller undershoots than A. This is since the disturbances are noticed earlier in the process. In strategy A the effects of disturbances are noticed at the end of the process and that is when A066 reacts to them. This then causes A089 to react as well. With the new sensor, A089 notices the changes and compensates for them when they are measured after effect 2. The settling time is however not affected that much.

When it comes to disturbances in the steam feed there are bigger differences. Strategy H gives smaller undershoots since the disturbance is noticed earlier in the process in this case too and both controllers can react to it.

There are some differences in the control signals as well. The first difference is that *A089* reacts faster when a disturbance enters the system. For the step in dry substance it can be seen that the control signal of *A066* stabilizes at 25 % for both strategies. This is since the inputs to the third effect are the same after the step response as it was before. On the other hand, when one of the mass flows have changed the inputs to the third effect have changed even though the same level of dry substance is sent in. This makes *A066* work at different levels. Therefore it is harder to keep the control signal at the requested level. An option could be to have a dynamic setpoint of *A089* and adjust it when changes in the mass flow occur. This would however mean that one more algorithm needs to be implemented.

In figure 6.14 the dry substance of the syrup coming out of the second effect is shown. In the present mid-range control strategy this dry substance is not regulated to reach a specific setpoint. In the new strategy it is however regulated to be 70.6 %. Here it is obvious that there is a difference using the mid-range control and the new strategy. The present control indirectly adjusts the dry substance going into the last effect to keep the valve position at a constant level. The new strategy splits the process in two parts and this dry substance is controlled without respect to the last effect. This makes the control of the first part a little smoother since it makes changes directly instead of waiting for *A066* to notice the change.

### **B-G-X: All New Strategies Included**

From the comparison in figure 6.15 it is clear that strategy X performs a lot better than strategy G when it comes to the steam disturbances. This has to do with the new feedback to *A089* and the feed forward from the new sensor, but mostly due to the steam feed forward system. Strategy X is generally faster, but also reduces both overshoots and undershoots to some more extent than strategy G.

Comparing strategy X with strategy B the differences are distinct. By including all control systems the control has been greatly improved, no matter what of the inputs are disturbed. The least difference is seen for a step in dry substance but it is still improved a lot regarding both the cut off and the settling time.

## **7.3 Further Studies**

The project has in many ways shown how improvements of the control of the process can be done, by presenting a number of different control strategies and comparing them to one another. It was shown that it is possible to create a model similar to the real process, despite all approximations. To approximate the process with only mass and energy balances is actually giving a similarly behaving model even though there are a lot of temperatures, pressures and other properties involved in a process like this.

The model is nevertheless not completely waterproof and it is of course possible to improve it further. There are some characteristics in the dynamics that could

be further studied. The evaporators in the process are approximated to be identical, which is not the case. There are no latencies or dynamics in the valves and steam feed of the model. The transport delays and transfer function time constants are constant throughout the entire model, but are probably related to especially the different mass flows in the system. There are also other parts of the system, such as the preheater, that have not been considered in the model.

Perhaps a better model could have been achieved using another software. Simulink was originally chosen due to the previous contacts and collaborations between Combine and MathWorks. Other software might include packages that are easier or better when it comes to modelling of an evaporator system. It is hard to know whether this is the case or not since this project did not compare different software during the pre-studies.

Since only trends of the data were available a more solid model verification has not been conducted. It would be very interesting to get real process data and use it as an input in the model. This would in a more detailed way show the strengths and weaknesses of the model and thus what could be further improved.

Hopefully this project has resulted in useful input to Lantmännen Reppe AB on how to continue their work on the control strategy and their efficiency work. The project will hopefully lead to a deeper understanding of the process and to some extent advise Lantmännen Reppe AB of what the next control implementation could be. It would be very satisfactory if this master's thesis would turn out to be an important tool in their future work.

# 8

## Conclusions

*This final chapter is a condensed summary of the results and the discussion. The conclusions are focused on how Lantmännen Reppe AB could continue to work on their control strategy according to the results of this master's thesis. This chapter therefore includes suggestions and recommendations of future improvements to their control strategy.*

The model is capturing the overall behavior of the process. There are differences since the dynamics depend on several aspects that are not included in the model. The model can be used for conceptual analysis of different control strategies, but because of the differences parameters cannot be taken directly from the model and implemented in the real process.

The simulations show that there are a number of ways to improve the control of the process. This can also be seen in the summarizing table 8.1. The implementation of the present feed forward system improves the control for disturbances in the ingoing syrup, both in the real process and the model. A combination of feed forward systems to both *A066* and *A089* would imply that *A089* can start to compensate directly and not wait for *A066* to get the feed forward signal. This could be done by using similar feed forward systems as the one that is already implemented, but with different gains and delays. However this strategy only improves the robustness to some extent when there are varying properties of the ingoing syrup, as can be seen in table 8.1.

These feed forward systems do not affect the process when there are disturbances in the steam feed. An implementation of a feed forward system that uses the steam pressure seems to be the easiest solution to improve the control for disturbances in the steam feed. This would probably be a fairly simple addition to the present control strategy. It can be based on the existing feed forward system but only use one input and different gain and delay. This helps to compensate the valve position directly when the pressure is varying.

Only having a sensor measuring the resulting dry substance makes the process harder to control, especially since there are delays in the system. A sensor between

		Cut off			Over/undershoot			Settling time		
		$c_0$	$F_0$	$S_0$	$c_0$	$F_0$	$S_0$	$c_0$	$F_0$	$S_0$
C	C		-		+	*				
D	D	+			+	*				
E	E			+			-			
F	F	+	+		+	*	+	+	+	+
G	G	+	+		+	*	+	+	+	+
H	H				+	+	+			+
X	X	+	+	+	+	+	+	+	+	+

\* Zero impact due to tuning

**Table 8.1** A summarizing table, listing all the different control strategies and what impact they have on different aspects of the dry substance when a disturbance inflicts one of the inputs –  $c_0$ ,  $F_0$  and  $S_0$ . They are all evaluated with regard to strategy B. It grades the strategy according to the aspects cut off, overshoot or undershoot and settling time. A plus (+) implicates that the strategy improves the control, a minus (-) that it worsens and a blank cell that it has no effect. Strategy C, D, F and G are all compared according to the gain tuned for a step in the mass flow of syrup.

the last two effects gives the opportunity to implement new strategies as well as providing additional information about the process. A new sensor might however be an expensive alternative.

The new feedback reacts faster to disturbances in the steam feed. By switching to this strategy, the present mid-range control will be lost, but the new strategy may provide the same effects by having a dynamic setpoint.

To break the mid-range control may not necessarily be the best option. However, to use the information to adjust the control signal of A066 seems to be a good idea. To be able to use a feed forward system depending on exactly what goes into the last effect is a clear advantage. The drawback of the existing feed forward system – that it uses a signal that is quite different from what really is going into the last effect – can in this way be reduced. The new sensor would also give the process the advantage to earlier detect disturbances caused by the steam feed.

Finally strategy X needs to be mentioned. The strategy is undoubtedly the most complex strategy to implement, since it contains a large number of controllers. Most of these controllers are new, but also the present controllers would need to be readjusted in order to implement the entire control strategy. Regardless of this the strategy seems to give a distinct improvement to the control system, no matter which one of the inputs that are disturbed. The implementation of strategy X could be seen as a long term goal, which could be achieved by introducing one of the included control systems at a time. As a first step the steam feed forward system is recommended. This since the present system does not include any compensation for pressure disturbances in the steam feed. Secondly the new sensor could be installed and eventually used to add new control strategies to the control system.



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# Appendices

## A Information from Lantmännen Reppe AB

All the given information about the process is listed in this appendix. Firstly descriptions and scaling intervals of the signals seen in the trend diagrams are summarized and secondly the different process properties.

### Signals in Trend Diagrams

Signal	Description	SP	PV	LMN
<i>A066</i>	The PID controller regulating the steam valve to effect 3	70 - 91.5 %	70 - 91.5 %	0 - 100 %
<i>A061</i>	The PI controller regulating the steam valve to the TVR	0 - 15 bar	0 - 15 bar	0 - 100 %
<i>A089</i>	The PI controller creating the mid-range control system	0 - 100 %	0 - 100 %	0 - 15 bar
<i>A058</i>	The mass flow of the incoming syrup	-	0 - 15 tonne/h	-
<i>A067</i>	The dry substance of the incoming syrup	-	30 - 64 %	-
<i>A069</i>	The pressure of the steam feed	-	0 - 15 bar	-
<i>A118</i>	The feed forward signal	-	0 - 30 %	-

**Table A.1** The signal names with descriptions and scaling.

## Process Properties

All given process properties are gathered in this section. The different properties are listed in the tables below.

Process parameter	Value
Gain	0.12
Time constant (s)	83
Dead time (s)	165
Latency from $c_0$ to effect 3 (s)	400-500
Latency from $F_0$ to effect 3 (s)	35

**Table A.2** The estimated parameters of the dynamics of the evaporator effects. The gain, the time constant and the dead time are all estimated from a step response of the resulting dry substance when a step of the valve position, controlled by *A066*, is induced.

TVR property	Ratio
Primary to secondary steam	1:1
Large to small ejector	4:1

**Table A.3** The estimated ratios of the jet steam ejectors used in the thermal vapor recompression system (TVR).

Controller	P	I	$T_i$	D	$u_{min}$	$u_{max}$
<i>A066</i>	0.40	0.0050	200	20	0	0.37
<i>A089</i>	-0.30	0.0025	400	-	0	0.70
<i>A061</i>	0.50	0.1439	7	-	0	0.50

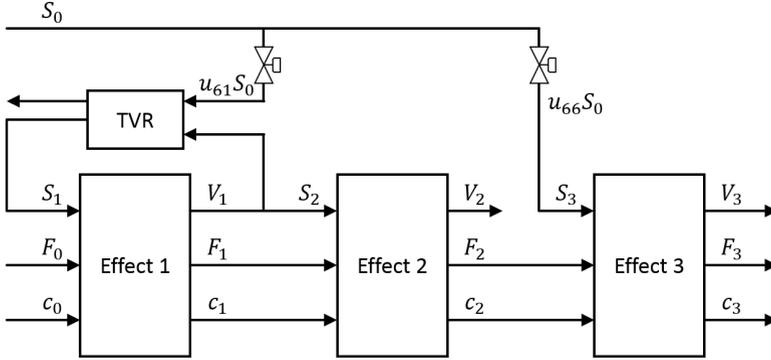
**Table A.4** The PID parameters of the controllers regulating the mass flow of steam into the evaporation process. The integral times,  $T_i$ , are given in seconds.

Feed forward parameter	Value
Gain, $k_{ff}$	0.025
Latency of dry substance, $L_c$ (s)	400
Latency of mass flow, $L_F$ (s)	35

**Table A.5** The parameters of the present feed forward controller.

## B Derivation of the Mass Flow of Steam

At steady state all mass flows and dry substances can be calculated using the inputs to the system. For an explanation of where each signal enters or exits the effects see figure B.1.



**Figure B.1** The mass flows and dry substances in the entire process.

### Outputs of Effect 1

$$V_1 = kS_1 = k2\frac{4}{5}u_{61}S_0 = \frac{8}{5}ku_{61}S_0 \quad (\text{B.1})$$

$$F_1 = F_0 - V_1 = F_0 - \frac{8}{5}ku_{61}S_0 \quad (\text{B.2})$$

$$c_1 = \frac{F_0c_0}{F_1} = \frac{F_0c_0}{F_0 - \frac{8}{5}ku_{61}S_0} \quad (\text{B.3})$$

### Outputs of Effect 2

$$V_2 = kS_2 = k(V_1 - u_{61}S_0) = \frac{8}{5}k^2u_{61}S_0 - ku_{61}S_0 = k\left(\frac{8}{5}k - 1\right)u_{61}S_0 \quad (\text{B.4})$$

$$\begin{aligned} F_2 &= F_1 - V_2 = F_0 - \frac{8}{5}ku_{61}S_0 - k\left(\frac{8}{5}k - 1\right)u_{61}S_0 \\ &= F_0 - k\left(\frac{8}{5}k + \frac{3}{5}\right)u_{61}S_0 \end{aligned} \quad (\text{B.5})$$

$$c_2 = \frac{F_0c_0}{F_2} = \frac{F_0c_0}{F_0 - k\left(\frac{8}{5}k + \frac{3}{5}\right)u_{61}S_0} \quad (\text{B.6})$$

**Outputs of Effect 3**

$$V_3 = kS_3 = ku_{66}S_0 \quad (\text{B.7})$$

$$\begin{aligned} F_3 &= F_2 - V_3 = F_0 - k \left( \frac{8}{5}k + \frac{3}{5} \right) u_{61}S_0 - ku_{66}S_0 \\ &= F_0 - k \left( \left( \frac{8}{5}k + \frac{3}{5} \right) u_{61} + u_{66} \right) S_0 \end{aligned} \quad (\text{B.8})$$

$$c_3 = \frac{F_0c_0}{F_3} = \frac{F_0c_0}{F_0 - k \left( \left( \frac{8}{5}k + \frac{3}{5} \right) u_{61} + u_{66} \right) S_0} \quad (\text{B.9})$$

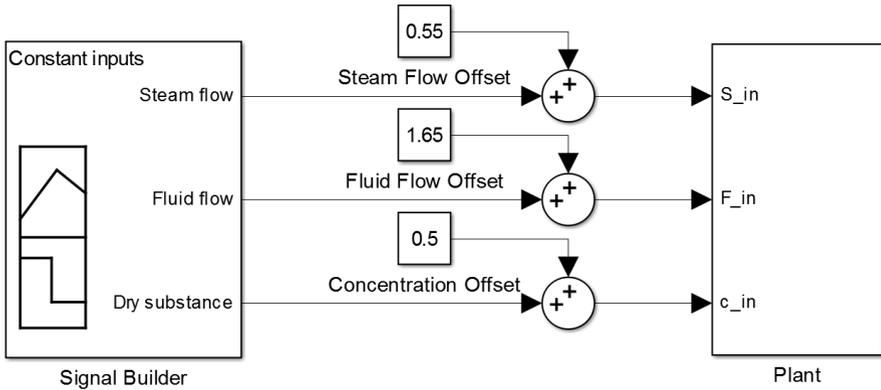
**Mass Flow of Steam Feed**

Rewriting equation (B.9) gives the mass flow of steam entering the process.

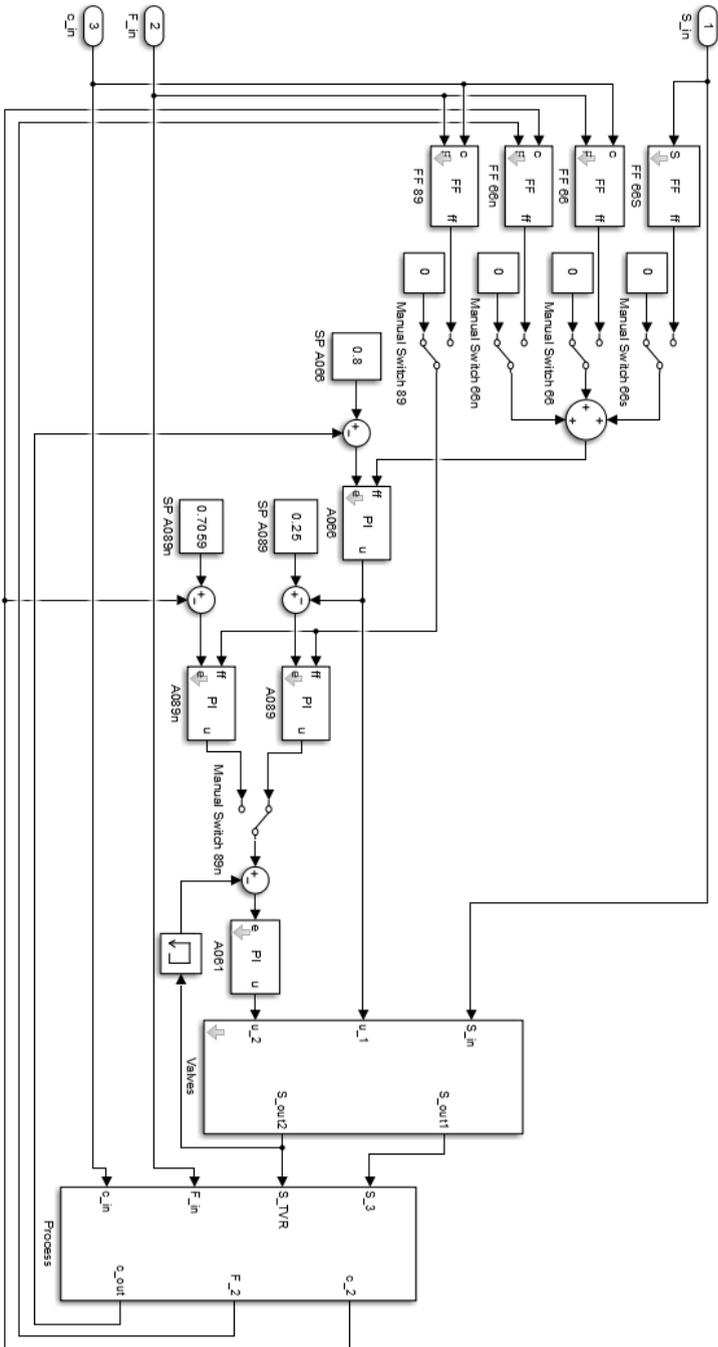
$$S_0 = \frac{F_0(c_3 - c_0)}{c_3k \left( \left( \frac{8}{5}k + \frac{3}{5} \right) u_{61} + u_{66} \right)} \quad (\text{B.10})$$

### C Simulink Model

The inputs to the model are sent from a *Signal Builder* block. This can be seen in figure C.1. The *Plant* block contains the model of the process including the control strategies, seen in figure C.2. Different control strategies can be tested by using the switches. The scripts from where the simulation model is run can be seen in appendix D.



**Figure C.1** Mass flow of steam, mass flow of syrup and dry substance are the inputs to the model. They are sent from a *Signal Builder* block in Simulink. This block decides the form of the signals, in this case it sends out constant signals. *Signal Builder* sends out signals around zero. The actual size of the inputs are then set by the constants that are added to each signal. The signals are sent to the evaporation process in the *Plant* block.



**Figure C.2** The Simulink model used for testing control strategies and combinations of different strategies. This can be done by using the switches.

## D MATLAB Scripts

The MATLAB scripts are gathered in this appendix. Three of the scripts contain the simulation, evaporator and control properties used in the blocks of the Simulink model, which can be seen in appendix C. The properties are stored in three different mat-files. One main script is then used to load the mat-files and run the model.

### Simulation Properties

```
clearvars
% ----- Simulation Properties -----
% Initial values of the system inputs
c_init = 0.50;
F_init = 1.65;
S_init = 0.55;

% Setpoints
SP_A066 = 0.80; % Resulting dry substance
SP_A089 = 0.25; % Valve position
SP_A089n = 0.71; % Dry substance out of the second effect

% Start and stop time of the simulation
start_time = 0;
stop_time = 70000;

save('simulation_properties')
```

### Evaporator Properties

```
clearvars -except c_init F_init S_init
% ----- The Dynamics of the Evaporators -----
% Heat exchanger
T11 = 60; T21 = T11; T31 = T11; % Time constants
k11 = 1; k21 = k11; k31 = k11; % Gains
x011 = 0; x021 = 0; x031 = 0; % Initial conditions

% Flash tank conc.
T12 = 15; T22 = T12; T32 = T12;
k12 = 1; k22 = k12; k32 = k12;
x012 = c_init*T12;
x022 = c_init*T22;
x032 = c_init*T32;

% Flash tank flow
T13 = 15; T23 = T13; T33 = T13;
k13 = 1; k23 = k13; k33 = k13;
x013 = F_init*T13;
x023 = F_init*T23;
x033 = F_init*T33;
```

```

% ----- The First Effect -----
% Heat exchanger
num11 = k11; % Transfer function numerator
den11 = [T11 1]; % Transfer function denominator
[A11, B11, C11, D11] = tf2ss(num11, den11); % State space matrices

% Flash tank conc.
num12 = k12; den12 = [T12 1];
[A12, B12, C12, D12] = tf2ss(num12, den12);

% Flash tank flow
num13 = k13; den13 = [T13 1];
[A13, B13, C13, D13] = tf2ss(num13, den13);

% ----- The Second Effect -----
% Heat exchanger
num21 = k21; den21 = [T21 1];
[A21, B21, C21, D21] = tf2ss(num21, den21);

% Flash tank conc.
num22 = k22; den22 = [T22 1];
[A22, B22, C22, D22] = tf2ss(num22, den22);

% Flash tank flow
num23 = k23; den23 = [T23 1];
[A23, B23, C23, D23] = tf2ss(num23, den23);

% ----- The Third Effect -----
% Heat exchanger
num31 = k31; den31 = [T31 1];
[A31, B31, C31, D31] = tf2ss(num31, den31);

% Flash tank conc.
num32 = k32; den32 = [T32 1];
[A32, B32, C32, D32] = tf2ss(num32, den32);

% Flash tank flow
num33 = k33; den33 = [T33 1];
[A33, B33, C33, D33] = tf2ss(num33, den33);

% ----- Thermal Vapor Recompression -----
tvr_lg = 4/5; % Large ejector
tvr_sg = 1 - tvr_lg; % Small ejector

% ----- The Delays and their Initial Values -----
% Delays of the system
F0_delay = 15;
E1_delay = 10;
E2_delay = E1_delay;
E3_delay = E1_delay;
c0_delay = 200;
c1_delay = 90;
c2_delay = 90;
c3_delay = 165 - E3_delay;

```

## Appendices

```
% Initial values of the delays
F0_init = F_init;
F1_init = F_init;
F2_init = F_init;
F3_init = F_init;
V1_init = 0;
c0_init = c_init;
c1_init = c_init;
c2_init = c_init;
c3_init = c_init;

save('evaporator_properties')
```

### Control Properties

```
clearvars -except c_init F_init S_init
% ----- Present Mid-Range Control to A066 -----
% A066
P66      = 1.3;
I66      = 0.005;
u66_max  = 0.37;

% A089
P89      = -0.1;
I89      = 0.0025;
u89_max  = 0.7;

% A061
P61      = 0.5;
I61      = 0.1;
u61_max  = 0.5;

% ----- Present Feed Forward to A066 -----
ff_gain_66      = 0.7768;
ff_c_delay_66   = 400;
ff_F_delay_66   = 35;

% Initial values in delays (used in all feed forward controllers)
ff_c_init = c_init;
ff_F_init = F_init;

% Manual switch on/off
sw66 = '0';

% ----- New Feed Forward to A089 -----
ff_gain_89      = 0.5404;
ff_c_delay_89   = 200;
ff_F_delay_89   = 15;

% Manual switch on/off
sw89 = '0';
```

```

% ----- Steam Feed Forward to A066 -----
% A066
ff_gain_66s = 0.3;
ff_delay_66s = 0;
ff_S_init   = S_init;

% Manual switch on/off
sw66s = '0';

% ----- New Feed Forward to A066 -----
% A066
ff_gain_66n   = 1.4386;
ff_c_delay_66n = 90;
ff_F_delay_66n = 0;

% Manual switch on/off
sw66n = '0';

% ----- New Feedback to A089 -----
% A089n
P89n   = 0.09;
I89n   = 0.007;
u89n_max = 0.7;

% Manual switch on/off
sw89n = '0';

save('control_properties')

```

## Main

```

simulation_properties, evaporator_properties, control_properties
clearvars -except model signal_group sw66 sw89 sw66s sw66n sw89n
% ----- Run the Simulation -----
% Model
model = 'Final_evaporator';

% Signal group in signal builder
signal_group = 1; % Constant inputs
% signal_group = 2; % Concentration step
% signal_group = 3; % Fluid flow step
% signal_group = 4; % Steam flow step
% signal_group = 5; % Pulse inputs

% Manual switches on/off
sw66 = '0'; % Present feed forward to A066
sw89 = '0'; % New feed forward to A089
sw66s = '0'; % New steam feed forward to A066
sw66n = '0'; % New feed forward to A066
sw89n = '0'; % New feedback to A089

```

## Appendices

```
% Load model
if bdIsLoaded(model) ~= 1
    open_system(model)
elseif strcmp(get_param(model, 'shown'), 'off') == 1
    open_system(model)
end

% Load signal group in signal builder
sgnbldr = strjoin({model, '/Signal Builder'}, '');
signalbuilder(sgnbldr, 'activegroup', signal_group);

% Turn on/off switches
try
    main_path = strjoin({model, '/Plant'}, '');

    path66 = strjoin({main_path, '/Manual Switch 66'}, '');
    path89 = strjoin({main_path, '/Manual Switch 89'}, '');
    path66s = strjoin({main_path, '/Manual Switch 66s'}, '');
    path66n = strjoin({main_path, '/Manual Switch 66n'}, '');
    path89n = strjoin({main_path, '/Manual Switch 89n'}, '');

    if strcmp(sw89n, '1')
        sw89n = '0';
    else
        sw89n = '1';
    end

    set_param(path66, 'sw', sw66)
    set_param(path89, 'sw', sw89)
    set_param(path66s, 'sw', sw66s)
    set_param(path66n, 'sw', sw66n)
    set_param(path89n, 'sw', sw89n)

catch
    disp('Something went wrong with the switches');
end

sim(model) % Run simulation
```

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<i>Title and subtitle</i> <b>Modelling and Control of an Evaporation Process</b>			
<i>Abstract</i> <p>Model-based design is in many ways seen as a potential instrument in process industries due to the close link between the process and the model. In Växjö, the process industry of Lantmännen Reppe AB produces syrup. The syrup goes through an evaporation process in order to raise the sugar concentration. If the concentration gets to high the syrup turns solid, which can make the process a bit tricky to control.</p> <p>In this master's thesis the goal is to model this process in the MathWorks environment Simulink in order to gain understandings about different aspects of the process and eventually bring forward and evaluate different control strategies. Information about the real process has been obtained by visits at the factory and mail correspondence with the process engineer of the plant. Important advices along the project have been given by supervisors at the Department of Automatic Control and at Combine Control Systems AB.</p> <p>The modelling is founded on approximations regarding no temperature or pressure dependencies, but only mass and energy balances. The model has been adjusted and tuned along the project in order to match the given process dynamics. The model verification has been conducted by comparisons of model simulations and process data. The model has in many ways proven to capture the fundamental behaviors of the process.</p> <p>A number of different control strategies have been tested in the model and the results have been compared. It has been shown that the present feed forward controller improves the system control but also that new feed forward controllers and a new sensor can improve the system control furthermore. The greatest improvements have been seen when introducing an additional sensor in the model.</p>			
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