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TAMPERE UNIVERSITY OF TECHNOLOGY

NIKO PUKKILA
UTILIZING MODEL PREDICTIVE CONTROL IN CONTINUOUS
COOKING APPLICATION
Master of Science Thesis

Examiner: Professor Risto Ritala
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ABSTRACT

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Continuous cooking is an important part of pulping process. High quality product is essential to preserve competitiveness of the mill. Modern automation increase profitability by improving product quality and reducing expenses.

The objective of this thesis was to familiarize myself with pulp process and to utilize model predictive control in continuous cooking application. As a result, performance and usability improves. A familiar MPC software was used for this complex process with long delays and strong interactions.

The work was done as a part of a pilot project. The created MPC controller was tested successfully with the MPC software. The information received from this projects probably increases the use of modern control technology in future projects.

TIIVISTELMÄ

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Jatkuvatoiminen keitto on tärkeä osa sellunvalmistusprosessia. Korkealaatuisen tuotteen valmistaminen on tärkeää kilpailukyvyn säilyttämiseksi. Modernit automaatioratkaisut lisäävät tuottavuutta parantamalla tuotteen laatua ja vähentämällä kustannuksia.

Tämän työn tavoitteena oli tutustua sellunvalmistusprosessiin ja soveltaa malliprediktivistä säätörakennetta jatkuvatoimiseen keittovaiheeseen, jolloin säädön suorituskyky ja käytettävyys paransi. Käytävissä oli valmis MPC-ohjelmisto, jonka toiminta oli todettu varmaksi. Monimuuttujaprosessi sisälsi pitkiä viiveitä ja voimakkaita ristikkäisvaikutuksia.

Työ suoritettiin osana pilottitehdasprojektia. Luotua MPC-säädintä testattiin työssä simuloimalla MPC-ohjelmistolla onnistuneesti. Projektin tuoma informaatio todennäköisesti lisää kehittyneiden säätömenetelmien käyttöä seuraavissa projekteissa.

PREFACE

I had a great opportunity to do this thesis for Metso Automation. Getting familiar with pulping process and modern control technologies was very interesting. The thesis was funded by Metso Corporation.

I would like to thank Professor Risto Ritala for your guidance. Special thanks for my supervisor Tapani Poranen and Johanna Newcomb. I would also like to thank Rami Rantanen, Manne Tervaskanto, Samuli Lehtonen, and Greg Fralic for your assistance.

I would also like to point out the importance of the support from my family and friends throughout my studies.

In Tampere, 24th March, 2015

Niko Pukkila

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LIST OF SYMBOLS AND ABBREVIATIONS

NaOH	Sodium Hydroxide (caustic soda)
Na ₂ CO ₃	Sodium Carbonate
Na ₂ S	Sodium Sulfide
OH ⁻	Hydroxide Ion
HS ⁻	Hydrosulfite Ion
HSO ³⁻	Bisulfite Ion (hydrogen sulfite)
H ₂ O ₂	Hydrogen Peroxide
MPC	Model Predictive Control
LPF	Low Pressure Feeder
HPF	High Pressure Feeder
MCC	Modified Continuous Cooking
EMCC	Extended Modified Continuous Cooking
ITC	Isothermal Cooking
BLI	Black Liquor Impregnation
Lo-Solids	Modified cooking method
EAPC	Enhanced Alkali Profile Cooking
EA	Effective Alkali
AA	Active Alkali
L/W	Liquor-to-Wood
A/W	Alkali-to-Wood
WL	White Liquor
BL	Black Liquor
DCS	Distributed Control System
MV	Manipulated Variable
CV	Controlled Variable
DV	Disturbance Variable
IV	Impregnation Vessel
FSR	Finite Step Response
FIR	Finite Impulse Response
ARX	Auto-Regressive With Exogenous Input
DMC	Dynamic Matrix Control
PRBS	Pseudo Random Binary Sequence
PID	A Proportional-Integral-Derivative Controller
APC	Advanced Process Control
Ad Hoc	A solution, based on assumptions, created for a specific problem
rpm	Rounds per minute
CMS	Chip Meter Speed

1 INTRODUCTION

Modern pulp industry automation has a significant role when aiming at high quality without disturbances in the process. The purpose of automation is to increase the capacity of production and to improve the quality of products while decreasing the cost of production in order to maintain competitiveness of the mill.

Pulping disintegrates constituent fibers from bulk structure of the fiber source, most commonly wood chips. Pulping can be mechanical, semi-mechanical or chemical. As a result, desired cellulose fibers can be used in papermaking industry.

An important part of the chemical pulping process is cooking. In chemical pulping, the aim of the cooking is to remove lignin that binds fibers from the wood chips with chemicals and heat. Continuous cooking process, which is discussed in this thesis, includes continuous feed of the chemicals and wood chips to a digester, and a continuous blow flow removing cooked fibers from the digester.

Controlling the cooking process is complex because of long delays and interacting control relationships. Therefore optimization of the process is challenging. One option to improve controllability of the process and to increase process performance is to utilize model predictive control (MPC). The objective of this thesis is to study the applicability of MPC in continuous cooking process.

MPC is based on a multi-input control algorithm that uses a dynamic process model and history of realized outputs. MPC provides optimal control actions for the process with respect to a given cost function over a given prediction time horizon.

The basic idea of cooking process is presented first, followed by the introduction of controlling the process. Then, the process of building a MPC controller for a pilot mill is described and the overall process is summarized in the final section.

2 COOKING AS A PART OF PULPING INDUSTRY

Pulping industry covers the process where wood raw material is refined into a pulp consisting of individual fibers which can be used, for example, in papermaking. This thesis focuses on continuous Kraft pulping. This chapter gives an overview of the structure of the pulping process and continuous cooking.

A way to measure effectiveness of pulping process is to define key performance indicators (KPI). Typical KPIs are introduced shortly and different cooking methods and digester types are described at the end of this chapter.

2.1 Structure of the pulping process

Pulp mill is a complex chemical processing plant consisting of several subprocesses. The structure of the pulping process is shown in Figure 1. The wood raw material comes to the mill as logs or sawdust, for example. Logs are treated to remove bark (debarking) and impurities and then cut into small chips. Before adding cooking chemicals, the chips are screened.

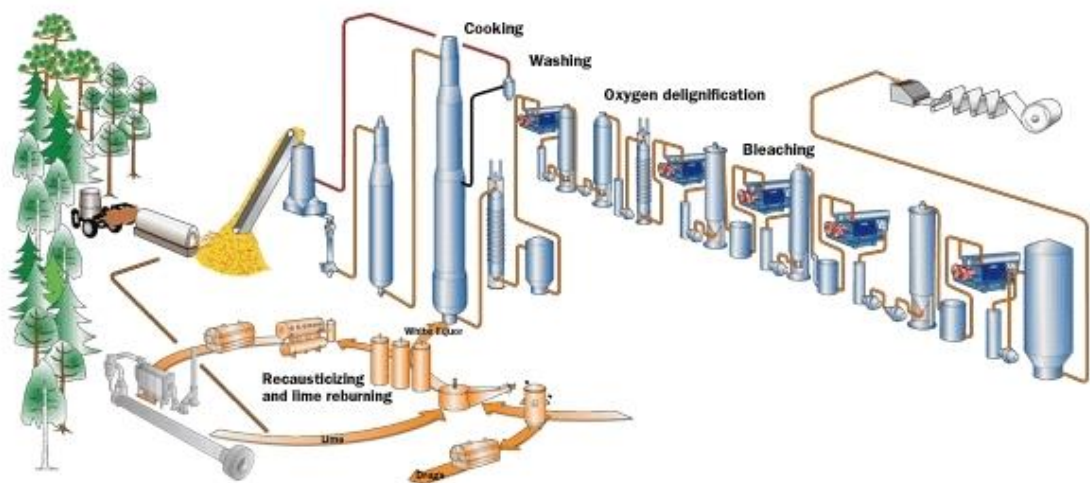


Figure 1: Structure of continuous Kraft pulp mill [1].

Chips are steamed to remove air so that chemicals could penetrate the chips composition more thoroughly. The chemicals remove fiber binding lignin from the chips to produce individual fibers in a fluid, pulp. The processing requires a high temperature and time. Therefore chips and chemicals are transferred into digester where the cooking takes place under these circumstances.

While removing lignin, also unwanted loss of cellulose, hemicellulose etc. happens. This affects to yield, but also to fiber properties such as strength. The challenge is to find the best compromise between Kappa (the amount of residual lignin), yield and other fiber properties. Different cooking modifications are designed to have optimal cooking conditions (temperature and chemical profiles, for example). It is “easy” to reach target Kappa, but how to maintain quality and yield!

After cooking, the pulp is washed to remove chemicals and impurities. Used cooking liquor, including lignin and chemicals, is transferred to a recovery line where chemicals are regenerated and organic material is burnt.

The recovery line consists of an evaporation phase for drying the black liquor and a recovery boiler for burning the organic material. Large amount of energy is released in the recovery boiler. Inorganic materials, such as sodium carbonate (Na_2CO_3) and sodium sulfide (Na_2S), from recovery boiler are fed to the causticizing circulation, where white liquor is recovered. The basic structure of the recovery line can be seen in Figure 2.

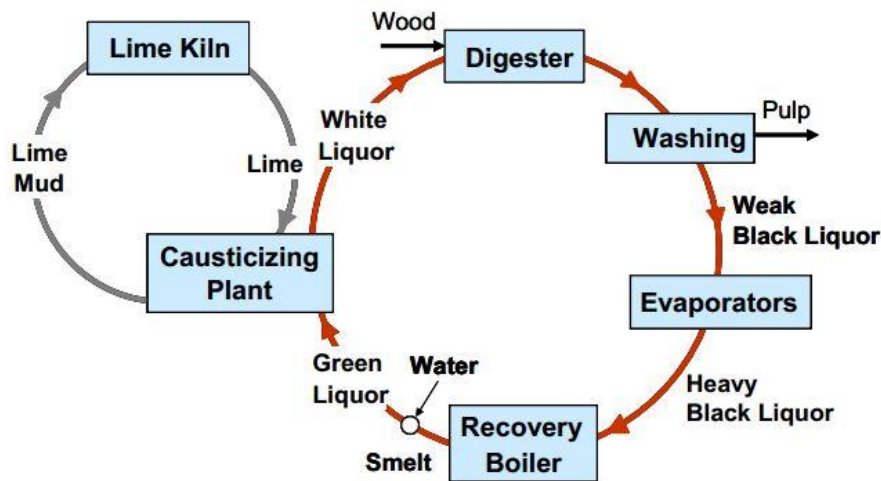


Figure 2: Structure of the Kraft Recovery Process [2].

After the washing stage, the pulp is fed to screening and bleaching. Screening removes impurities, such as knots and incompletely delignified chips, from the pulp to allow better further processing. The pulp flow that passes the screening is bleached by e.g. oxygen or hydrogen peroxide (H_2O_2), to achieve target brightness. On the other hand, unbleached pulp is also a desired raw material in some end product applications.

Pulp is dried and post-processed at the end of the pulping process. The pulp is ready for shipping or, in an integrated pulp and paper mill, can be fed without drying to paper machines. [3]

2.2 Key performance indicators of pulping process

The performance of the pulping process can be measured by a set of indicators. This section introduces some of the most important indicators such as Kappa number and H-factor.

Kappa number is one of the most important control parameters in the cooking process. It describes the residual lignin concentration of the mass and it is used as a target value in process. [4]

Yield represents the amount of pulp that is manufactured from raw wood material. Many variables affect to yield. For example, wood species, chip size distribution and pulping chemistry have to be taken into account.

Production is the rate at which final pulp is produced. It is usually indicated as air dried tons of pulp per day (ADT/day). Production can be estimated by calculation from digester blow flow or already from the chip feed to the digester.

Strength of the pulp can be specified in numerous ways. It depends on raw material, pulping process used and yield, the type of test it is submitted to, and the intended use of the product. Stress-strain curve of a solid material and tear-tensile curves are typical strength characteristics. Average fiber length, coarseness, wet compactability, intrinsic fiber strength, and cohesiveness are examples of strength parameters. [5]

Alkali consumption is calculated based on alkali dosage. These calculations use information about production rate as measured by the chip meter, strength of the white liquor (WL) measured on-line, and liquor flow measurements. Alkali dose is usually calculated as alkali-to-wood (A/W) ratio. Traditionally alkali is expressed as Na₂O and effective alkali (EA). Nowadays expression as NaOH and active alkali (AA) has become more common as the NaOH is the actual reacting chemical in cooking.

Setting the cooking temperature is based on the *H-factor*, which is a combination of cooking temperature and time as seen in Equation (1).

$$H = \int_0^t e^{(43,2 - \frac{16115}{T})} dt, \quad (1)$$

where t is time and T is temperature in Kelvin. [4]

Sulfidity describes the percentage of sodium sulfite in EA as follows:

$$S (\%) = \frac{Na_2}{Na_2S + NaOH} 100\% \quad (2)$$

Normal sulfidity is 25-35 %. [5]

2.3 Main functions of the cooking process

The cooking process digests wood chips in a cooking liquor in high temperature. When choosing the cooking time the objective is to remove enough lignin from the wood in order that the fibers are released from the chip structure without breaking up.

2.3.1 Structure of fibers

Wood chips consist of four main chemical components, which are lignin, hemicelluloses, cellulose and extractives. The middle lamella, between the fibers, is mostly lignin and acts as a glue binding fibers together. The actual fiber cell has multiple layers; primary wall and three-stage secondary wall. The distribution of lignin decreases from middle lamella towards the center of the cell. A cross section of wood structure is seen in Figure 3. Most of the lignin is removed first from the cell walls before middle lamella. Residual lignin in paper turns the color of paper yellow over time. For example, hardwood consists of 42-49 % cellulose, 23-34 % hemicelluloses, 20-26 % lignin, 3-8 % extractives and 0.2-0.8 % ash. [3]

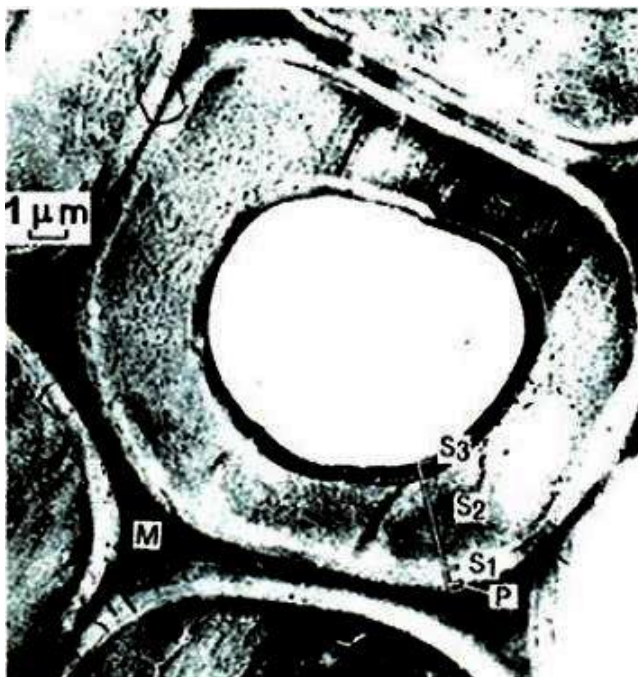


Figure 3: Cross section of a Pine latewood tracheid, where M is the middle lamella, P is the primary wall, S1, S2, and S3 are layers of the secondary wall [3].

Chips entering the chip bin have commonly been cut into length of 12-25 mm and 2-10 mm thick depending on the digester type. To avoid disturbance in the digesting process the wood chip dimensions should be homogenous. The length of the actual fibers depends on the wood species and how intact the fibers remain after the process. [3]

2.3.2 Cooking liquor

Cooking liquor consists of chemicals that can vary between different types of pulp mills and digesters. The most common process is the strongly alkaline kraft process which uses hydroxide (OH^-) ions and hydrosulfite (HS^-) ions. Sulfite or semi-chemical processes, which are neutral or acidic, use the bisulfite (HSO_3^-) ions. Other chemicals, such as alcohols can also be used to remove lignin from the wood. Choice of chemicals affects fiber properties. For example, the alkaline process produces strong but brown fibers [3]. The most important characteristics of the white liquor are: effective alkali (EA), active alkali (AA) and sulfidity. Effective alkali represents the concentration of the OH^- ions, active alkali is the sum of OH^- and HS^- ions and the sulfidity is the ratio of OH^- and HS^- ions as already seen in formula (2). [4]

2.3.3 Chip feed and pre-steaming

Wood chips are fed to the process through a chip bin. In some cases, chips are pre-steamed in atmospheric conditions in the chip bin. A chip metering device controls the amount of chips entering the process. This device is typically used to calculate the production rate. Low pressure feeder (LPF) transfers the chips into the steaming vessel. Chip meter and low pressure feeder are shown in Figure 4. [3]

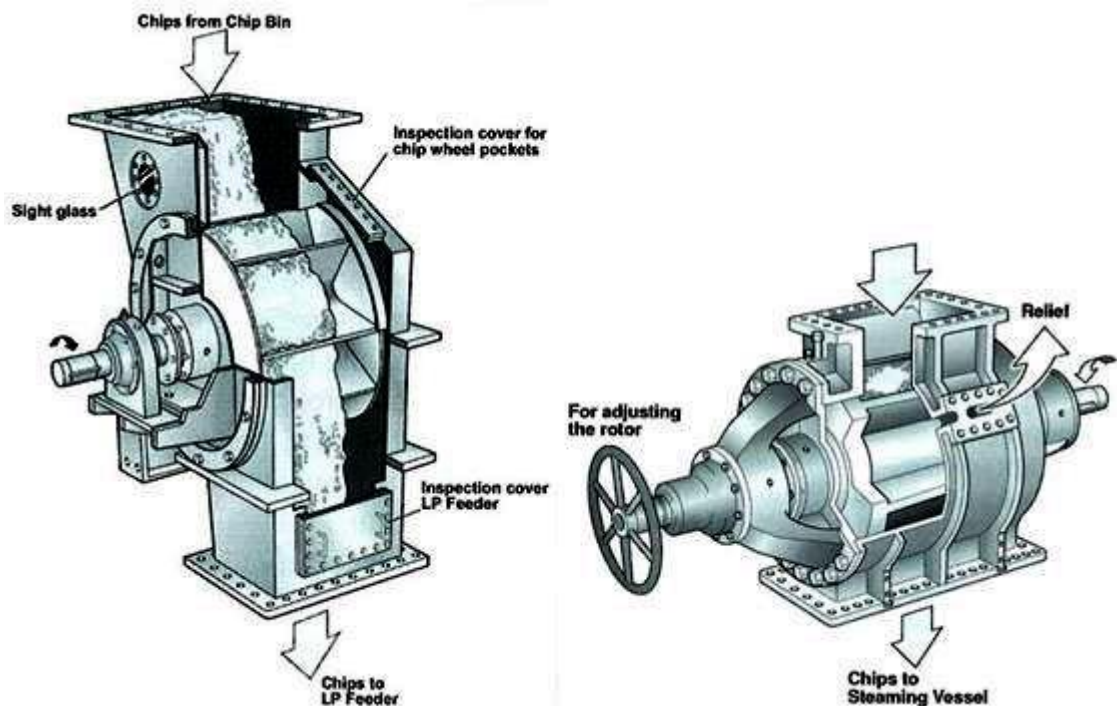


Figure 4: Chip metering device and low pressure feeder [3].

Steaming vessel, as seen in Figure 5, is a horizontal tube where chips are exposed to steam at a pressure of 100-150 kPa. Due to steaming, the air in the chips is mostly displaced with water vapor. This leads to a better penetration of the chemicals into the structures of the wood. High temperature (100°C - 120°C) improves the air removal due

to air expansion. Chips are transferred with a screw to a vertical chute which directs chips into a high pressure feeder.

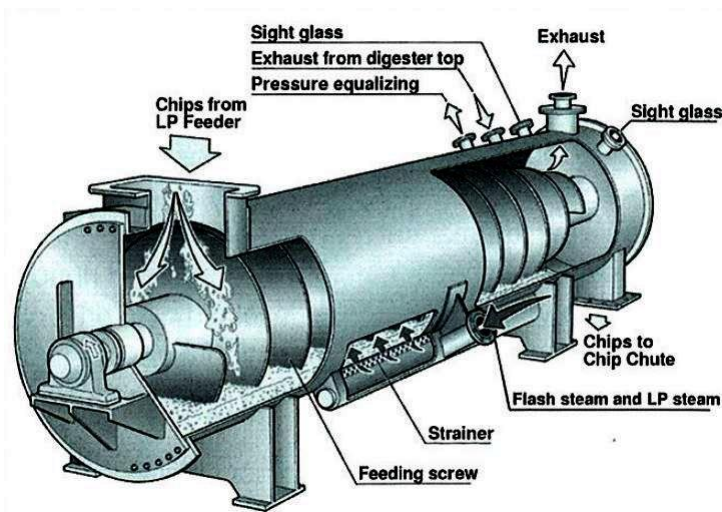


Figure 5: Steaming vessel [3].

Modern plants may have integrated steaming phase in the chip bin. In that case, chips heated to 90°C-100°C are transferred from the chip meter straight to a high pressure feeder (HPF). Feeder doses chips to top circulation, and further to the impregnation phase at cooking pressure. Top circulation system is shown in Figure 6.

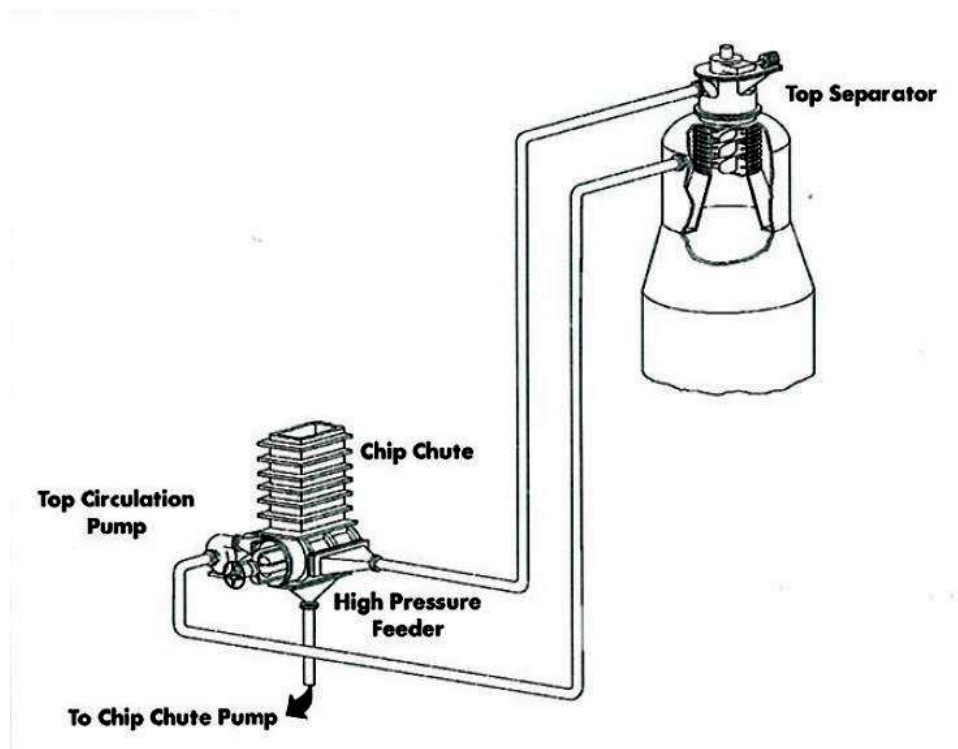


Figure 6: Chip transfer to impregnation vessel including Chip Chute, Top Circulation Pump, High Pressure Feeder and Top Separator [3].

2.3.4 Impregnation

In impregnation phase cooking liquor penetrates to wood chips. Chips are impregnated in a separate impregnation vessel (IV) or at the top of the digester depending on the type of the digester. Pressure is increased before the impregnation to the cooking pressure of over 1 MPa and temperatures are typically near 100°C. Liquids penetrate the air cavities of the wood filled with gas or steam due to either capillary force or pressure. [3]

In impregnation vessel, typical impregnation time is 30-60 min [4]. Too short time increases the amount of rejects due to incomplete chemical penetration. In contrast, too long impregnation time deteriorates the quality of the pulp, such as tensile strength [6]. After the impregnation phase, the chips are scraped with bottom scraper and, by the help of a sluice flow, discharged from the IV through a transfer circulation to the top of the actual digester.

2.3.5 Digester

Description of the digester is divided into flows, levels, cooking zone, temperatures, extractions and circulations, washing zone and blow flow. Continuous digesters can be divided into two main categories: hydraulic digesters and digesters with steam and liquor phase. Furthermore, both of these digester types are divided into single- and two-vessel models. This thesis concentrates on two-vessel digesters.

Chip flow is fed from the impregnation phase to the top separator of the digester. There most of the transfer liquor goes back to the transfer circulation and the chips are separated to the top of the digester. In two-vessel digester with steam and liquor phase chips travel upwards on a screw conveyor and fall on top of the chip column in the steam zone of the digester. In a hydraulic vessel a screw conveyor moves the chips downwards into the digester. Both of these systems are shown in Figure 7 and Figure 8. [3]

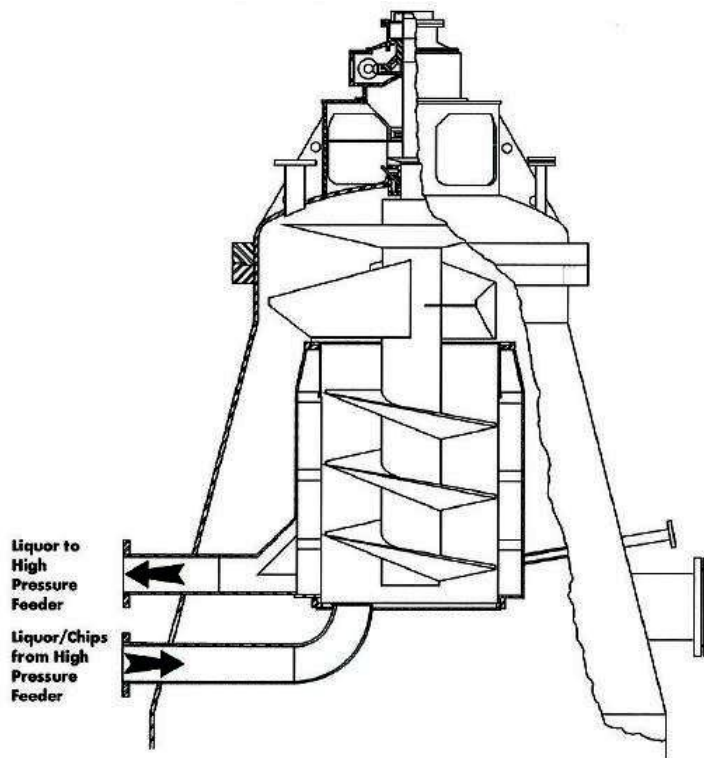


Figure 7: Top Separator of the steam and liquor phase digester [3].

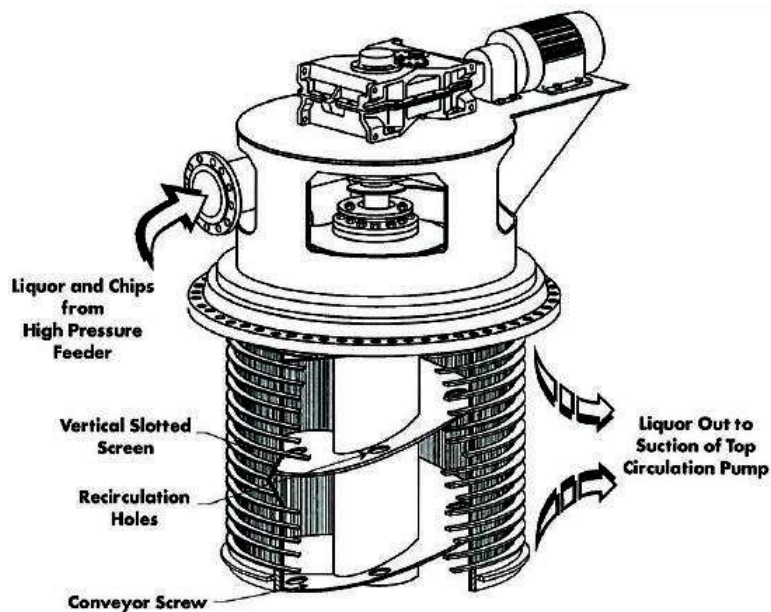


Figure 8: Top Separator of the hydraulic digester [3].

The liquor level of the steam-phased digester lies below the chip level. Both of these levels are measured and controlled by the chip feed, liquid additions and extractions, and blow flow. A hydraulic digester is always full of liquid, but the chip level is controlled.

The delignification process starts when the temperature of the liquor and chip mixture reaches 145-150°C. Diffusion of alkali into the chips, reaction of alkali with wood and diffusion of solubilized wood particles to the liquor occur simultaneously in

the cooking zone. It is important to maintain enough residual alkali in the process. Depleted alkali causes a drop of pH in the reaction mixture, which leads to poor pulp quality. Therefore, distributing liquor uniformly is important. Retention time in the cooking zone is typically 1.5-2.5 h depending on the furnish [3].

The methods to heat the contents in the digester to full cooking temperature vary depending on the type of the digester. In a two-vessel steam-phase digester the primary heating is done by direct steaming and secondary by indirect heating of the transfer liquor between the vessels with a heat exchanger [3]. The liquor circulation helps to achieve homogenous delignification by reducing the temperature and chemical concentration gradients.

The cooking process needs a temperature around 150°C -170°C [7] for chemicals to react with lignin in a specific time. Too high a temperature in cooking accelerates the unwanted cellulose depolymerization reaction, which decreases the strength of the fibers [5].

High temperature increases the rate of screening rejects of the chips because chemicals do not have time to diffuse deep enough into thicker chips [3]. Average cooking time is normally from less than one hour to three hours depending, for example, on the wood species, cooking temperatures, alkali dosages, desired fiber quality, and digester capacity.

Extractions allow selective removal of liquors from different points of the digester. At the cooking zone, extracted liquors are pumped and heated through an external heater. Liquors are then fed back to the same level in the digester via center pipes. Thus these extractions are heating circulations.

At a specific extraction zone, used cooking liquors are transferred to flash tanks where black liquor is cooled down by flashing. The steam produced is used in pressurized pre-steaming phase. There is typically more than one flash tank in series operating in lower pressures to decrease foaming effect (at 30-40 kPa, while the first tank operates at 130-150 kPa). The steam from lower pressure tank is used for atmospheric pre-steaming. The black liquor is transferred to the evaporator after fiber separation. This is illustrated in Figure 9. [3]

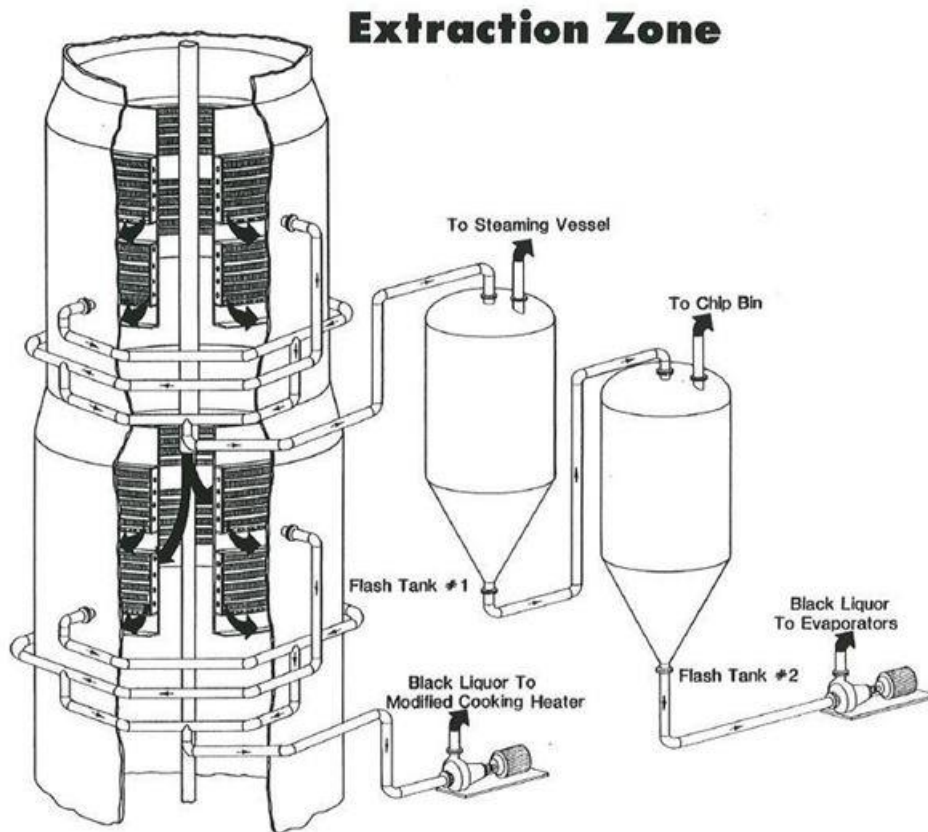


Figure 9: Schematic of an Extraction Zone [5].

The washing zone is near the bottom of the digester where cooled washing filtrate is pumped. A typical Hi-Heat washing zone washes the pulp counter-currently with circulated and heated washing filtrate. Pumped flow also cools down the chips just before blow flow. [3]

After travelling through the digester, cooked and washed chips are discharged from the bottom of the digester. Temperature is reduced to minimize fiber damage. Typical blow temperature is 85°C-90°C. The washing filtrate dilutes the pulp to 9-11 % consistency and keeps hydraulic digesters full of liquid. The blow process is where the fiberization mainly happens, because of sudden drop in pressure at the blow valve. [3]

2.4 Differences between continuous and batch cooking

Batch cookeries comprise many digester reactors, as seen in Figure 10, whereas in continuous systems there is only one digester. Continuous digesters are more efficient in reactor volume per unit retention time, because they do not have the filling and discharging phases needed in the batch process. The continuous digesters are larger vessels, but they still require less installation space than batch cookeries of same capacity.

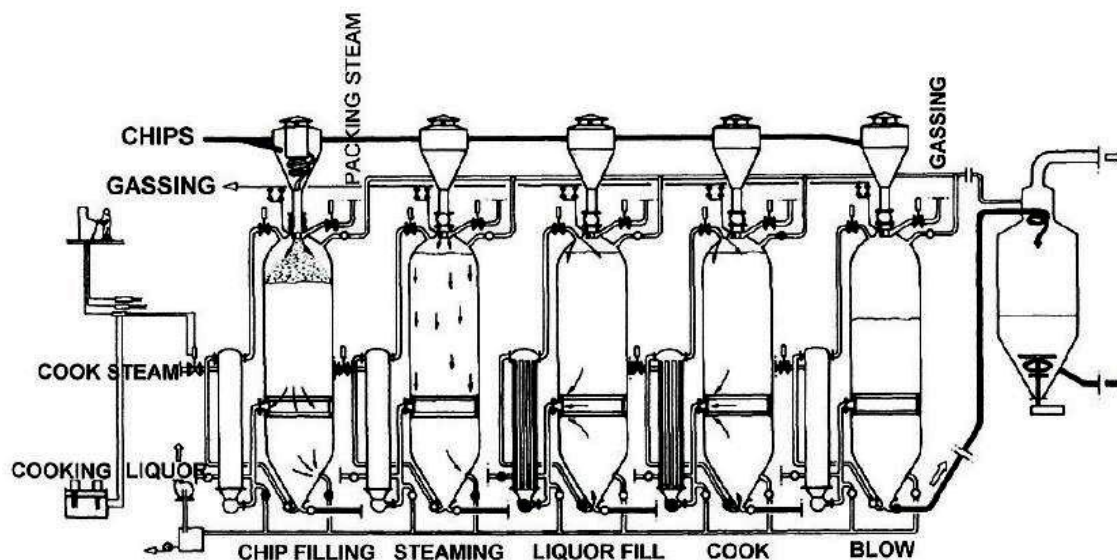


Figure 10: Schematic diagram of batch cooking [3].

Continuous digester installations require less powerful pumps and motors due to significantly lower and continuous flows of chips, chemicals and pulp. Batch systems have power consumption peaks during filling and discharging phases.

Continuous cooking applications have also other advantages over batch applications. These are, for example, energy efficiency due to steaming, easier control of environmental impacts, efficient in-reactor brownstock washing and more flexible process control. [3]

Batch cooking has also some advantages over continuous cooking. It is more flexible in grade changing and with different fiber sources. Production is also flexible as the amount of digesters can be increased. Down-times are shorter than in continuous digesters during maintenance issues which mean more production.

2.5 Cooking methods

The development of continuous cooking has led to improvements in fiber quality and more environment friendly operation. Conventional and some modified cooking modes are described in this section.

In *conventional cooking*, all the cooking chemicals are added in the beginning of the process with chips. After short impregnation time, temperature is increased to cooking temperature. In the cooking chips and chemicals travel co-currently and the cooking ends in a hi-heat counter-current zone. [3, 4]

Development of the digesters has aimed to lower Kappa and stronger pulp with lower cooking temperature and uniform alkali profile. The constructions of *modified cooking processes* are similar to conventional cooking, but modified systems include more screening zones and simplified feeding installations.

In *Modified Continuous Cooking* (MCC) white liquor is fed to a MCC circulation of the digester between extraction and washing screens. As a result, in the end of the cooking phase chips and chemicals flow counter-currently and the concentration of dissolved lignin is decreased towards the bottom of the digester.

MCC was modified to *Extended MCC* (EMCC) by adding heat and white liquor also in the washing circulation. Consequently, cooking is continued in the hi-heat zone simultaneously with washing and the overall cooking temperatures can be decreased.

The operation principle of *IsoThermal Cooking* (ITC) is similar to EMCC in that in both methods the washing zone is used for cooking. In addition, in ITC the temperature of the washing zone is chosen such that the temperature profile throughout the digester until the screens of the washing circulation is uniform. Liquor flows and temperature profiles of conventional, MCC, and ITC (or EMCC) cooking are compared in Figure 11.

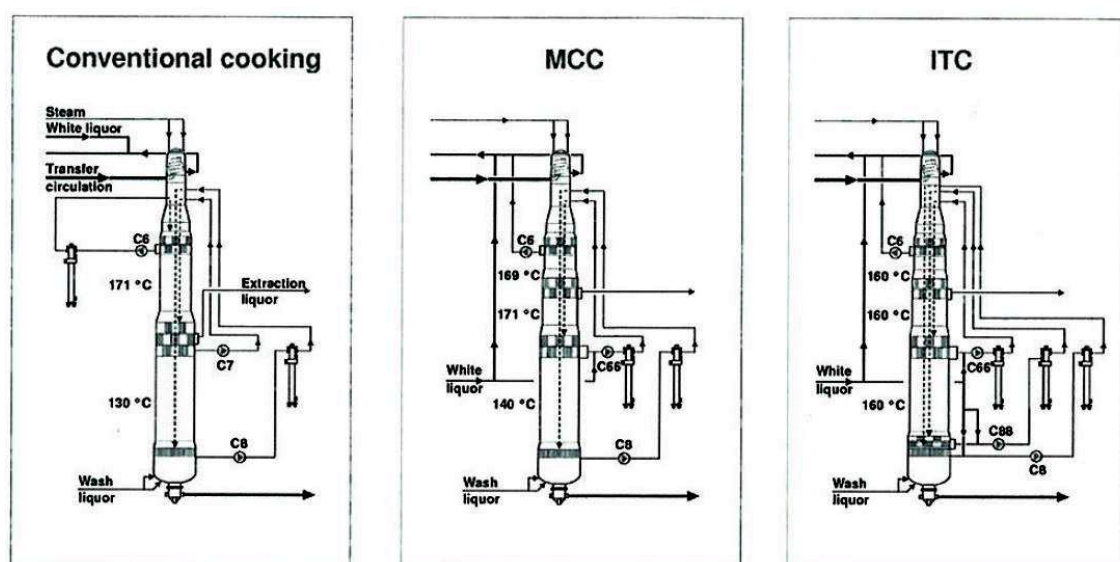


Figure 11: Comparison of different cooking methods applied to steam/liquor phase digester [1].

ITC cooking with Black Liquor Impregnation (BLI+ITC) differs from ITC cooking in that it includes an impregnation vessel for the black liquor impregnation through displacement of impregnation liquor with counter-current hot black liquor.

Lo-Solids cooking aims to lower concentration of dissolved wood solids in digester cooking liquors. This is achieved by using multiple extractions of spent liquors. Alkali and washing liquor are added to circulations. As a result, the concentration of dissolved organic materials is minimized, temperatures can be kept low and alkali profile flat.

Developed from Lo-Solids, *Enhanced Alkali Profile Cooking* (EAPC) aims to control the concentration of alkali at the beginning of cooking without increasing white liquor consumption. High residual alkali black liquor from lower extraction is used in impregnation. Therefore, alkali concentration at the end of cooking can be increased if necessary. The quality of pulp is enhanced and the sulfidity at impregnation phase is increased. [3, 4]

2.6 Special digester configurations

Chipping procedure always generates sawdust, which can also be used as a pulp raw material. The resulting pulp is not as strong as that from chips, because of greater amount of broken fibers. After separating it from chips, sawdust can be cooked in a specific digester.

Cooking process differs slightly from normal digesters, because of the fiber properties of sawdust. Sawdust dissolves almost entirely in the cooking liquor. Cooking circumstances should also be milder than in normal cooking; alkalinity, temperature and cook time should be lower to avoid material losses. Hence, cooking times are normally less than 20 min.

Due to fine structure of sawdust raw material, screening is difficult. Normal liquor circulation in cooking based on screening cannot be used for sawdust because of resulting operational problems. Therefore, digesters with screw or scraper conveyor, as seen in Figure 12, are typically used. [4]

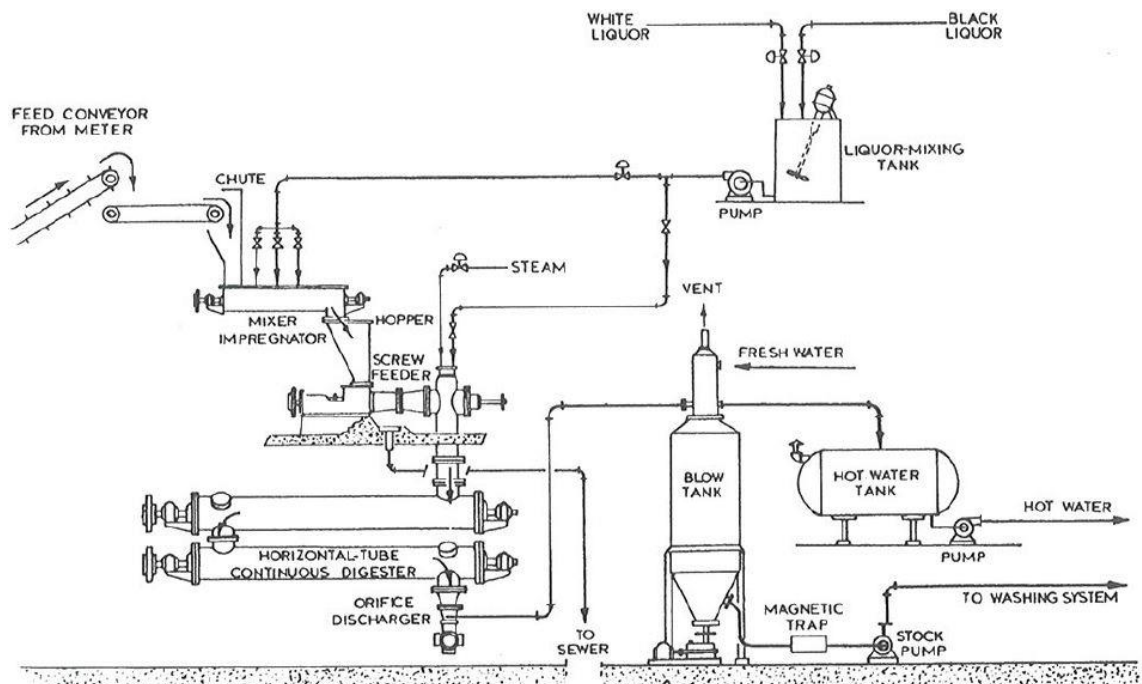


Figure 12: Sawdust digester with horizontal screw conveyor [3].

3 CONTROLLING THE COOKING PROCESS

This chapter deals with controlling a two-vessel digester. First, the basic controlled variables are introduced. Then, optimization principles are discussed. Challenging dynamics of the cooking process are noted before introducing the control methods in use.

3.1 The basic controlled variables

This section gives a description of the basic controlled variables in the cooking process. First, the controls related to chip movement are introduced. Then quality control is described.

3.1.1 Chip movement

It is important to make sure that chip flow stays uniform through the entire cooking process. This is managed by ensuring that chip feed is uniform. The first and very important phase in chip movement control is measuring the amount of chip feed. The production rate is calculated based on the rounds per minute (RPM) reading of a chip meter.

Chip properties and, for example, pre-steaming affects the packing degree of chip columns and therefore the chip flow. Variation in the chip dosing affects alkali dosing and chemical action (realized H-factor). As a result, even minor changes in chip flow result in fluctuations in the lignin content, the Kappa number. [4]

Controlling *the level of the chip bin* is necessary to supply chips to the digester in every situation. It is even more important in systems with no separate steaming vessel. Pre-steaming in the chip bin has to be done in right circumstances. Too low chip level causes steam to go through the chip layer, which results in incomplete pre-steaming. The level is difficult to control due to short residence time of 15-45 min, long feeding conveyor and high level in the bin. [4]

The first contact to liquor happens in the chip chute. The liquid level of the chute is kept constant. Chute circulation goes through the HPF, where chips are transferred to the feeding circulation. The flow in chute circulation has to be controlled to avoid chip plugs or chips bypassing the HPF. Mass flow to the top of the impregnation vessel in the feeding circulation is maximized. [4]

The setpoint for *chip level in the impregnation vessel* is maximized, but constrained through that the level is not allowed to reach the top screw or else the screw jams. The level is usually measured with mechanic sensors (Figure 13). The level is controlled by manipulating the chip screw speed and the bottom sluice flow to the transfer circulation. This is further done by controlling the speed of the bottom scraper and the amount of dilution liquid. Chip column reacts slowly to the changes. Thus the control should be cautious. Unsuccessful pre-steaming leads to problems with chip sinking in the impregnation vessel. As a result, the level control is disturbed. [4]

K1000 Level Indicator

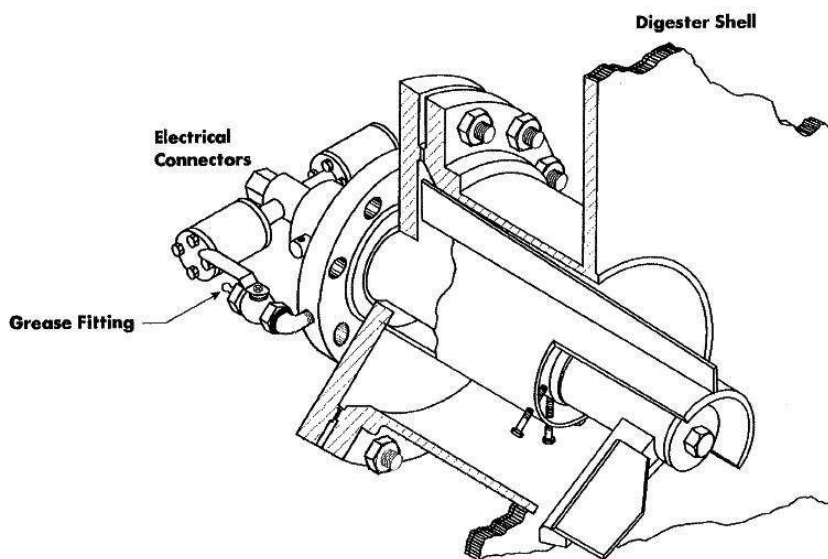


Figure 13: K1000 level indicator [5].

Chip flow from impregnation vessel to the digester is done via *transfer circulation*. The transfer flow is normally controlled by manipulating sluice flow and IV bottom scraper speed.

The chip level in the digester is normally managed by chip feed control or blow flow control. Blow flow is usually steadied to prevent changes in the delays of the digester and therefore stabilizing the residence time. IV sluice flow and bottom scraper can also be used to control digester levels.

Steam phase level can be measured with radiometric sensors but mechanical sensors are more common. Variation in chip quality disturbs the level measurements. Fines and too high packing degree can influence strainer jamming and flotation in the counter-current zone. Variation in chip size may affect chip packing and therefore disturb chip level control. The level control has to be set slow to minimize problems caused by disturbances. Variation in the packing degree can also affect Kappa number due to delay changes.

Packing degree can be influenced by the liquor-to-wood (L/W) ratio in the co-current zone. Higher L/W ratio leads to increase of the packing degree, because liquid flows faster than chips. Packing degree increases when the level difference of the liquor and chip levels increase. Too low or too high packing degree can cause problems in the process. [4]

3.1.2 Alkali dosing/charge (A/W ratio)

WL can be fed to the process in many phases. The most important WL flow to the process is before impregnation. Alkali profile control becomes easier to manage when there are multiple WL flows to the process in different phases.

Alkali doses can be fine-tuned based on residual measurements from different phases of the process. Typical measurement points are transfer circulation, trim circulation, cooking circulation and extraction liquor.

The basic principle of alkali charge control is presented in Figure 14. It includes all alkali flows and measurements of the process. Locations of specific instrumentation can vary in every mill. Therefore, the control design has to be built for every mill separately.

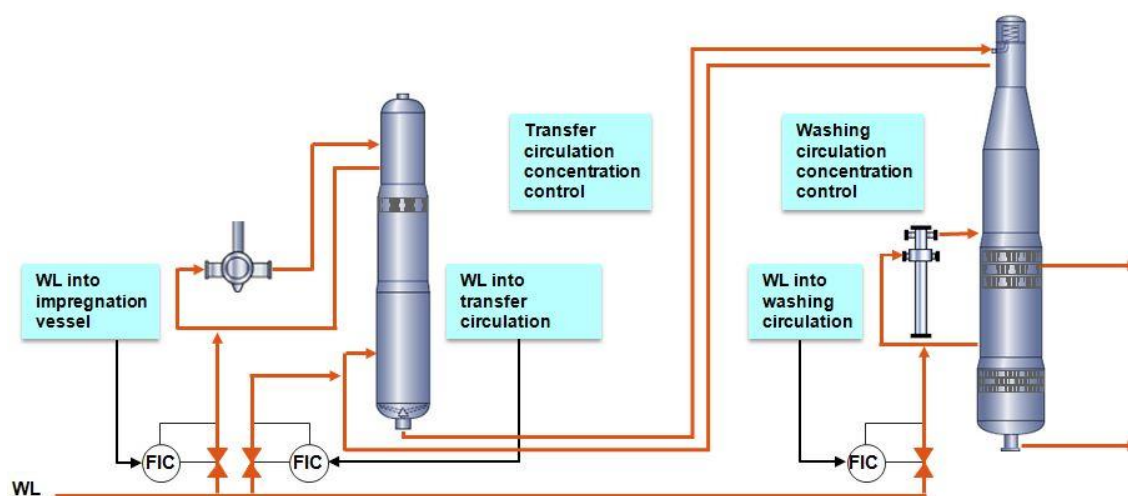


Figure 14: Basic Alkali Charge control.

Residual alkali must be kept on high enough level to ensure good pulp quality and bleachability. If the alkali runs out the lignin begins to precipitate and bleaching becomes more difficult and expensive. Too high alkali concentrations increase chemical expenses. [4]

3.1.3 Temperature, H-factor

H-factor is controlled by manipulating temperatures of cooking zones. H-factor calculation takes care of the changing delays due to changes in production. Realized H-factors are compared to H-factor targets.

Production speed changes can be executed by ramping zone temperatures with delays and making changes to chip feed. Increase of cooking time or temperature increases the H-factor. [4]

3.1.4 Kappa number

Kappa number was traditionally measured in laboratory, but nowadays it is normally measured on-line with special Kappa analyzers, as seen in Figure 15. These analyzers use optical measurement based on the UV-absorption of lignin. The measurement itself is done from the blow flow where the long delay of the digester causes problems. Kappa feedback control needs a good model with good management of disturbances to be efficient. Kappa number stays usually constant when digester runs well; chip movement and quality factors, like H-factor and residual alkalis, are constant. Disturbance related changes in Kappa number are very hard to correct. Kappa number is controlled by calculating the temperature needed to achieve desired H-factor target with corresponding delay of the production level.



Figure 15: Metso Kappa Q analyzer.

Mid-Kappa measurement from the side of the digester can also be used to reduce delay. This allows corrections to final Kappa number. The measurement point for mid-Kappa depends on the digester type. [4]

3.1.5 Liquor-to-wood ratio

L/W ratio is important to be controlled to ensure uniform packing and cooking of the chips. It is controlled by managing the liquor extractions on different circulations depending on the cooking and impregnation method. Calculations include chip moisture and condensates from steaming and direct steam heating. [4]

3.1.6 Counter-current washing (and dilution factor)

The amount of counter-current flow is adjusted by manipulating the amount of the liquor fed into the bottom of the digester. The control is usually based on dilution factor, which

is calculated by certain excess of liquid compared to liquid moving with blow flow. Too high-level counter-current can cause chip flow to stop. The structure of the counter-current washing is shown in Figure 16. [4]

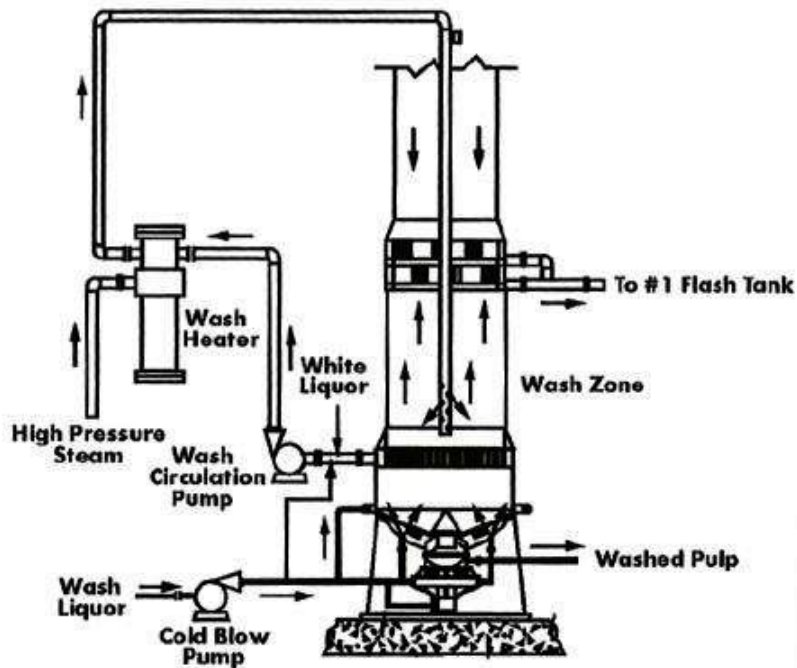


Figure 16: Counter-current wash zone [3].

3.2 Optimizing

Efficient optimizing requires well performing instrumentation and basic-level automation system (DCS). It helps operator to achieve high quality pulp with minimal costs and taking account of different cooking conditions, production and grade changes, and disturbances.

3.2.1 Quality control

Temperature profile throughout the digester is required to achieve the targeted Kappa number and thus good quality pulp. The profile is not controlled straightly but via H-factor control. [7]

Hatton's Kappa number model

$$Kappa = \alpha - \beta(\log(H) * EA^n), \quad (3)$$

is used for Kappa control. Parameters α , β and n are fitted to the process data [4, 7]. The new H-factor target is calculated based on Kappa analyses and old H-factor target. Kappa and yield calculations take into account time, temperature and A/W ratio.

As the Kappa number is the most important factor in the pulp quality, the use of Kappa analyzers is very useful. On-line measurements make it possible to react much quicker to disturbances than laboratory analyses.

3.2.2 Change management

Production rate and grade changes are process events that should be optimized to ensure uniform pulp quality. Automatic optimization schedules changeover starting times so that lost production due to grade change is minimized. An example of scheduling is shown in Figure 17.

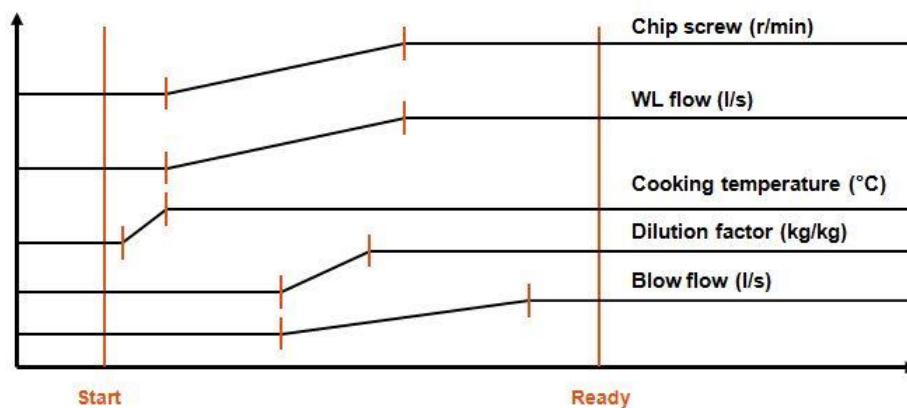


Figure 17: An example of change management scheduling.

Change scheduling covers production rate, alkali charge, liquor balance, and temperature profile. [4]

3.2.3 Disturbance management

Process disturbances can cause production losses and/or quality disturbances. Modern optimization software is able to give warnings about disturbances before they have any effect deteriorating performance. The software can give operator instructions how to make corrections or even make the corrections automatically.

Typical disturbances are related to packing degree and chip column movement. Control strategy can be based on, for example, fuzzy logic and it has to be configured specifically for every mill. [7]

3.3 Dynamics and current control methods

Vessels in pulping process have large volumes and thus long delays. Chip flow through the entire process can be many hours. For example changes in the white liquor addition before impregnation vessel can result in changes in the residual alkali measurements at the end of the digester vessel with several hours of delay.

Another characteristic feature of digesters are strong interactions between controls. For example, change in the IV sluice flow affects IV level, digester level, and therefore cooking time and temperature need to be changed.

Control of the cooking is based on efficient DCS which includes all the valve controls and measurements. Supervisory advanced process controls (APC) are built to optimize the process control. There are many approaches to how APC systems are implemented, e.g. fuzzy logic and - increasingly - model predictive control.

4 PILOT MILL DIGESTER DESIGN

The pilot mill studied in this thesis is a typical two-vessel continuous digester with separate impregnation vessel. The mill is also equipped with a separate steaming vessel. The digester vessel is steam and liquor phased. This chapter describes the process to level of detail needed for the MPC model identification.

The impregnation vessel is approximately 20 m high with a diameter of 3 m. The digester is approximately 70 m high. The diameter at the top is 3 m and 6 m at the bottom. The chip screw doses chips to the system approximately 100 kg/rev typically near 25 rpm.

The mill produces usually only one pulp grade. The typical production rate is over 1000 ADT/day. Total blow flow is typically around 350 m³/h. Typical WL addition flow is 110-120 m³/h. The mill produces pulp of Kappa number 88.

The level of the impregnation vessel is controlled by manipulating chip screw speed, IV bottom scraper speed, and sluice flow. The chip level of the digester is controlled by manipulating sluice flow and blow flow.

Material circulations in the pilot mill are shown in Figure 18. Feed circulation imports chips to the IV and transfer circulation C5 moves the impregnated chips to the digester. Cooking circulation C6 is normally not used. Cooking circulation C8, also used for washing, can be found at lower part of the digester.

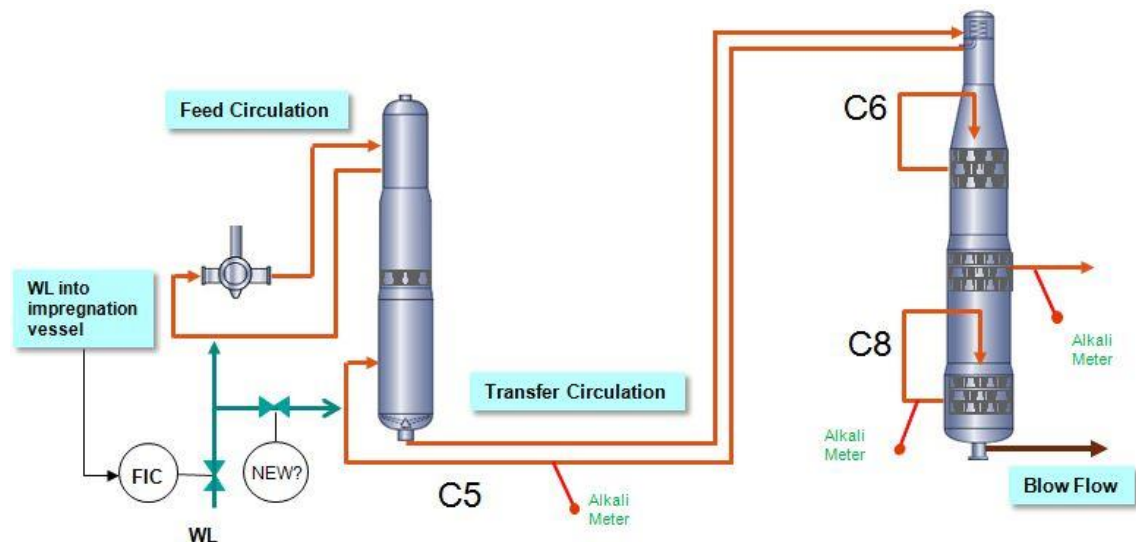


Figure 18: Circulations and Alkali Charge control of the pilot mill.

Screw speed information is used to calculate rates of dry wood and production. The pilot mill is exceptionally well instrumented with flow and temperature sensor and valves.

Chip level is estimated with three mechanical sensors which measure the pressure caused by the moving chip pillar. Each of the three sensors estimates the level and the final estimate is obtained by combining them.

The strategy for alkali charge control can be seen in Figure 18. At the moment only one WL addition point is employed, i.e. at in IV feed. However, a second addition point in the digester feed circulation can be employed at a later stage, and this has to be taken into account when designing the structure of the MPC controller. At present, as there is no alkali addition later in the process, the alkali profile cannot be controlled accurately.

WL addition is controlled with a control valve. A/W ratio is calculated in the alkali addition point based on dry wood calculation. Residual alkali is measured in transfer circulation, digester main extraction and washing circulation. Alkali addition and residual alkalis have their own targets. WL strength is measured with an analyzer, but laboratory analyzing is also possible.

H-factor value is calculated based on cooking times and temperatures according to Eq. (1). H-factor is controlled by manipulating the temperatures with PI-controller. Replacing this controller with MPC is one objective of this thesis.

Pulp quality is measured with a Kappa analyzer. The analyzer is placed after the blow flow. Prediction of Kappa is used to control the quality. Predicted Kappa is controlled via manipulation of cooking temperatures.

5 MODEL PREDICTIVE CONTROL

This chapter describes the basics of the MPC control method with advantages and disadvantages compared to other control methods. Suitability to the continuous cooking is discussed. In the end, the process of MPC implementation and the features of the MPC software used are discussed.

5.1 The basics of the MPC

MPC method is based on a model of the process that predicts the future evolution of the process given a set of control actions. MPC then optimizes the control signals [8]. A model is built based on process measurement data and realized control actions.

With the model, such actions are made which are to be expected of yielding the best long term performance. The control is with receding horizon update: an entire series of optimal actions are solved, but only the first action is implemented. At the next sampling instant the optimization is repeated.

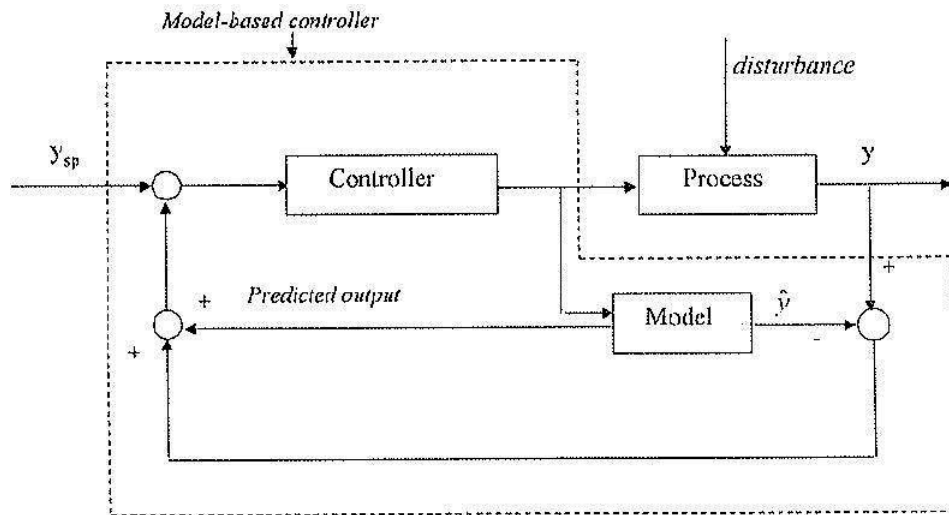
The control structure is based on a selected set of controlled variables (CV) and manipulated variables (MV). In some cases also disturbance variables (DV) are used. At time instant k the predictive control minimizes the sum of squares of future errors in controlled variables y over a finite time horizon

$$J = \sum_{i=1}^{N_2} \{y_{sp}(k+i) - \hat{y}(k+i)\}^2 \quad (4)$$

For stability the changes in manipulated variables u are limited with an additional term:

$$J = \sum_{i=1}^{N_2} \{y_{sp}(k+i) - \hat{y}(k+i)\}^2 + \sum_{i=1}^{N_u} \{\lambda \cdot \Delta u(k+i-1)\}^2, \quad (5)$$

where N_2 is the output prediction horizon, and N_u is the control horizon (denotes the number of future control moves ($N_u < N_2$)). Simple schematic of an MPC controller is shown in Figure 19. This is a quadratic programming (QP) problem with possibly some constraints [1, 11, 12]



$$(y - \hat{y}) = \text{disturbance} + \text{model-plant mismatch} + \text{noise}$$

Figure 19: Simple schematic of an MPC controller [3].

Process and quality constraints can be set as both high or low limits for variables instead of target setpoints. This is convenient when certain ranges of operating variables have to be maintained but the value itself is not important. Also maximum increments or decrements of control signals can be constrained. These constraints can be hard or soft, as in (5). Hard limits must always be satisfied, but soft ones are only undesirable as causing high values of the cost function.

The finite step response (FSR) model applied in this thesis as the process model is

$$y(k) = \sum_{j=1}^{N_2} S_j \Delta u(k-j) + z_o, \quad (6)$$

The basic operating idea and the control horizons are presented in Figure 20. [10]

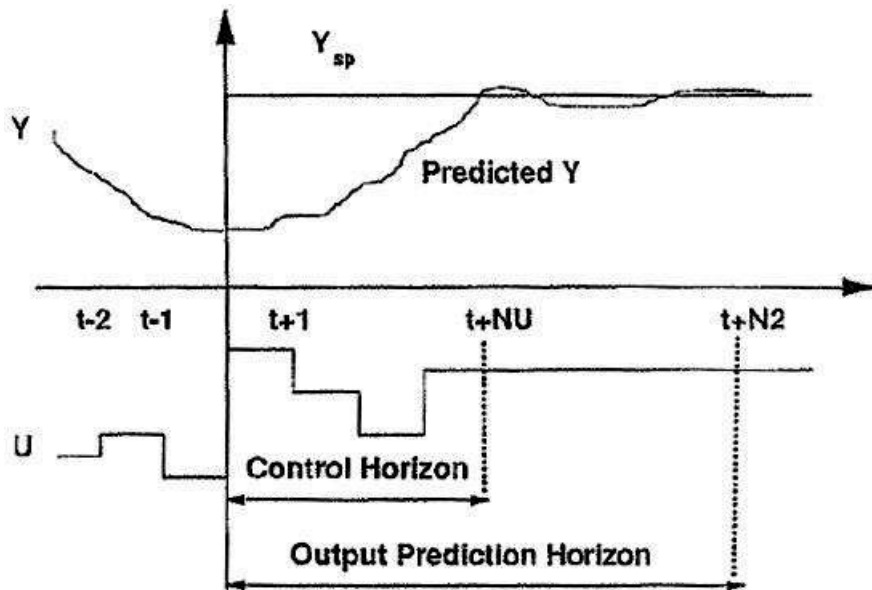


Figure 20: Graphical illustration of predictive control [10].

The model is typically a first principles one or obtained via system identification. In first principles modeling, the model is built theoretically based on knowledge of the process, for example, mass balances. Purely theoretical modeling is also called white box modeling.

System identification derives the mathematical model empirically based on experimental data. If no theoretical aspects are applied when setting up the model, it is called black box model. A combination of theoretical and empirical modeling is known as grey box modeling. The data for identification can be obtained by step response tests. Step response test is done by setting the controller in manual and changing the output of the controller and then observing the response in the process variable. Typical information of the process dynamics is a steady-state process gain, a dead time, and a time constant. [9]

The idea behind the operation of MPC controller is easy to understand. The structure is simple and it is easy to see which variables are used to achieve desired control results. MPC can handle complex MIMO processes that could be very difficult to figure out with traditional PID (Proportional-Integral-Derivative) controllers. [12]

The ability to handle constraints is an important advantage, because actuator limitations and economics can be considered. In addition, MPC can handle loops with long delays, which are typical in pulp process. The APC layer traditionally includes *ad hoc* solutions which are basically solutions to individual problems without taking into account the behavior of the entire process. MPC can be used to replace these *quick fix* solutions with a combination of linear dynamic models and constrained optimization. [10]

The ability to predict the behavior of the process results in decreased variation. As the variation decreases, targets can be set higher and closer to the system constraints while profitability increases. QP optimizing improves effectiveness.

The possibility to optimize tuning parameters with simulation studies is beneficial. The MPC software used in this thesis has the possibility to show predicted output of

the controller when the operation values are provided as inputs to the software. [13, 14, 15]

The most important disadvantage of MPC controller is that it needs a good model to be efficient. Poorly built model automatically leads to poor control. MPC controllers tend to be more complex and thus require excessive computing with short on-line response times. This is a challenge even though modern computers offer a lot of calculation capacity. One downside is the need of reconfiguration of the controller for every project.

A linear model cannot handle frequently changing operating conditions. In these circumstances a nonlinear model must be used. These are more complex to configure, solve, and understand. [13]

MPC has been implemented in the cooking process already in many pulp mills. Good results and a growing number of implementations encourage using this technology even more widely in the pulping process. MPC for level control, quality and alkali charge control has been used for example by Honeywell.

5.2 Using the MPC

Basic steps of building a MPC controller consists of system identification, model identification, data pretreatment, and tuning.

System identification can be done in multiple ways. Most of the MPC products use PRBS (Pseudo Random Binary Sequence) or multiple step response test signals. FSR, FIR (Finite Impulse Response), and ARX (auto-regressive with exogenous input) model structures are widely used. Other options are, for example, methods based on transfer functions or state space models. [14]

The model is built based on identified parameters of the system. It can be done automatically depending on the software. The accuracy of the model is critical for good results.

Usually the data retrieved from the process needs some *pretreatment*. This can be simple filtering, mean-centering or, for example, differencing. [10]

After building the controller, it needs to be *tuned*. Pre-tuning can be done before implementing the controller. Normally controllers may need tuning after on-line testing. This makes the controller more accurate.

The MPC design program applied in this thesis is a typical Dynamic Matrix Control (DMC) variant of MPC software based on parameters that can be obtained by step response tests. The model is built with an Excel tool. All the parameters are fed to the Excel and the dynamic matrix is generated automatically with macros.

Cost function optimization including variable weighting and constraining are used. Varying process models and complex dynamics can be taken into consideration. LP and QP optimizing can be implemented.

6 MPC FOR DIGESTER CONTROL

The MPC control structure is implemented in digester level control, alkali charge control and quality control. The implementation process of MPC is described in the following sections. First, the control problems are presented. Then the configuration of the dynamic model is introduced. Finally pre-tuning is dealt with.

6.1 Control tasks

MPC technology has already been used successfully in earlier pulp mill projects for production and level control. In the pilot mill production is controlled simply by manipulating chip screw speed. Level control requires integrating models and it is well suitable for MPC.

The control of the alkali charge should be easier to manage and tune with the new controller. Residual alkali profile should also be easier to control.

MVs and CVs for the alkali charge control were recognized to be as follows. A/W ratio targets were set as manipulated variables, and residual alkalis and rejects were set as controlled variables. WL flows are calculated by A/W ratio targets and these targets are manipulated to control residual alkali targets and rejects.

The controller needs to be tuned to minimize alkali consumption with some constraints. Residual alkali amount has to be maintained in certain range to avoid losing proper cooking conditions. Some variables can be weighted to react slowly to changes. This is needed because of long delays. Priorities between interacting variables can be managed by weighting.

The control of the alkali charge should be easier to manage and tune with the new controller. Residual alkali profile should also be easier to control.

The quality control includes H-factor calculations and Kappa control. Realized H-factor, predicted Kappa, and temperature difference in digester are set as CVs. And, digester cooking temperatures were logically selected as MVs.

6.2 Long delays

Lengths of horizons have to be long enough to incorporate the long delays in the process. Mass flow (from chip screw to Kappa analyzer) through the impregnation vessel and digester can take up to two hours depending on the production target. This travel time can be used for the length of prediction horizon. Quality horizon can be up to 15 hours, because Kappa analyzer is after the digester and the blow tank, which has delay up to 8 hours. The control horizon has to be smaller than output prediction horizon.

6.3 Building the Dynamic model

After entering MVs and CVs to an Excel file, the model matrix is formulated using built-in macros. Relationships between MVs and CVs are defined in the model matrix (Figure 21).

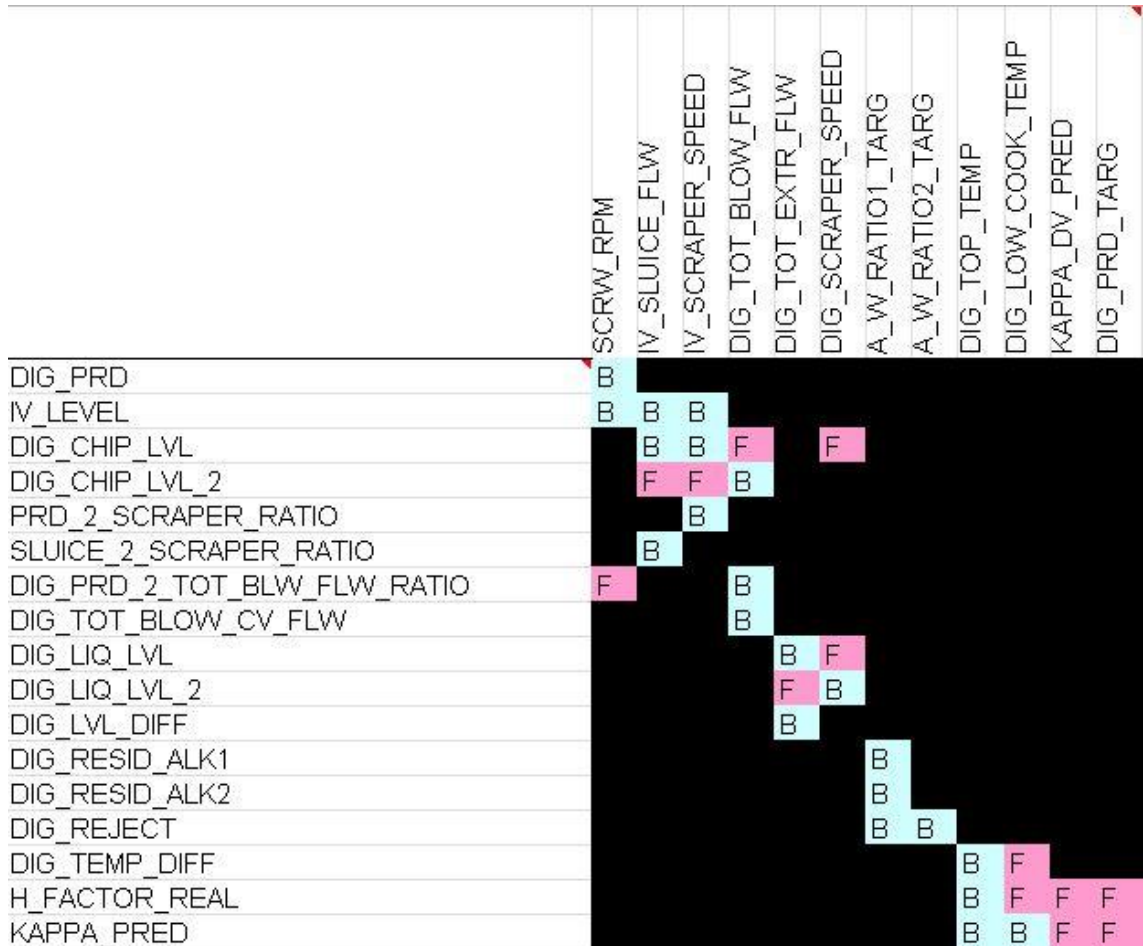


Figure 21: Structure of the MPC controller. CVs at left side and MVs at the top. B includes a model in both control and feedback, F includes only feedback model, and black area indicates no model between variables.

At this phase of creating the model, a dummy controller was built. It has all the MVs and CVs without any tuning parameters. The dynamic matrix of the dummy MPC controller is shown in Figure 22. Responses to control actions can be seen in this matrix after the model is tuned.

	WLIQ_FLW1	WLIQ_FLW2	A_W_RATIO1	A_W_RATIO2
A_W_RATIO1_CV		0	0	0
A_W_RATIO2_CV			0	0
RESID_ALK1	0	0		0
RESID_ALK2	0	0		0
RESID_ALK3	0	0		
REJECT	0	0		
TOT_A_W_RATIO_CV	0	0		

Figure 22: Dummy version of the dynamic matrix of an early stage version of the alkali charge MPC controller.

To obtain the dynamics of the controller, some information about the process is needed. Lag, gain and dead time between each MVs and CVs are entered to the Excel. These parameters are calculated either based on step response tests or physical dimensions and knowledge of the digester.

Model identification process started with step response tests at the pilot mill. At the second identification phase another identification method was used also based on pilot mill tests. The last identification method is based on physical model of the process.

6.3.1 Model identification based on first step response tests

Step response tests are done at the pilot mill. The testing process is a typical open loop test and it is done by varying one MV with specific step sizes with certain schedule and record changes in corresponding CVs. This is done with every MV.

The recorded data is analyzed to find out what are the lags, gains and dead times of the variables. The identification was executed in three steps: data cleaning, manual identification and identification using model identification tools.

Data cleaning was done after receiving the data package from the pilot mill. Collected data consisted of several process states with different setpoint values. Corresponding data sections to the step tests made were separated from the data. Basically MV and CV related data from step test time intervals were gathered.

Manual identification was done in Excel and responses of CVs to each MV were plotted based on the model matrix and the dynamics of the system were inspected. Gain and dead time averages were calculated based on fitted slopes. An example of the process is seen in Figure 23.



Figure 23: Fitting slopes to step response data. Red curve indicates sluice flow process value, green curve indicates sluice flow setpoint value, and magenta curve indicates IV level process value.

After having the first set of parameters, the Excel model identification tool was used. It includes a sheet with macros that calculate predicted behavior of the model based on the parameters used. An example of the model test is shown in Figure 24. The tool calculates how well the model prediction correlates with actual measurements. It is also possible to find out better parameters for model using the solver function.

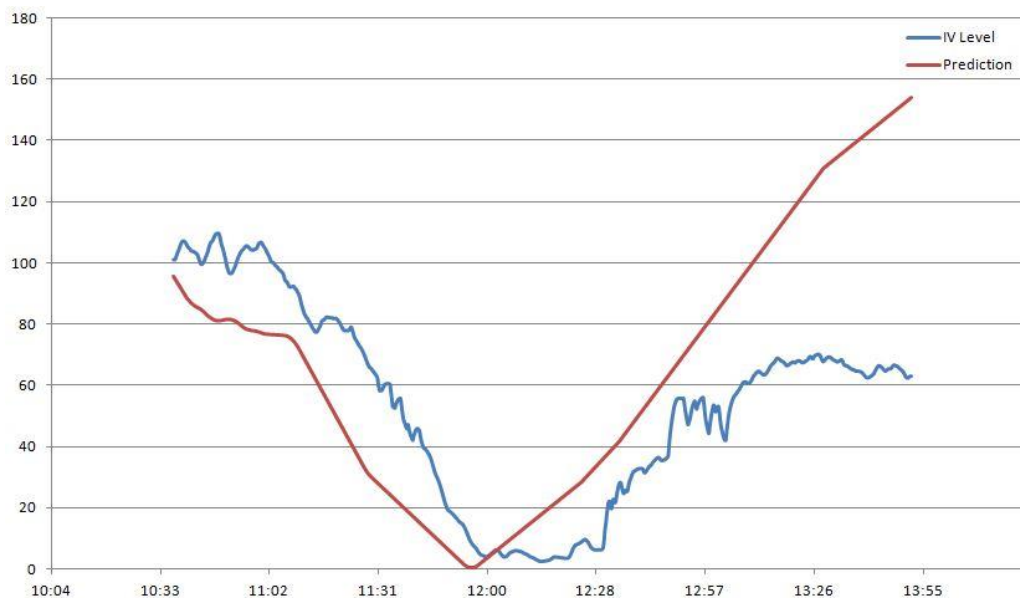


Figure 24: Model test for IV level based on sluice flow and chip screw speed.

At this model identification phase only few model parameters were obtained. Basically production, IV level and liquid level of the digester were included. Alkali charge models could not be identified and needed more testing. Chip level of the digester needed also a few more step tests and temperature response tests will be done in a later phase.

6.3.2 Model identification based on second pilot mill tests

Second pilot mill testing was done with Zhu Yucai's Tai-Ji software comprising automated testing signals and model identification. This method has been used widely in petrochemical and refining industry with significant benefits. However, the suitability to cooking process was not clear and models had to be critically evaluated. [16, 17]

In a test the signals were fed simultaneously to every manipulated variable. Test signal consisted of a large number of steps of certain size. An example of a test signal is shown in Figure 25. The step sizes and step durations can be selected so that process will not be disturbed. The tests were made under closed loops: the controls were left in automatic mode.

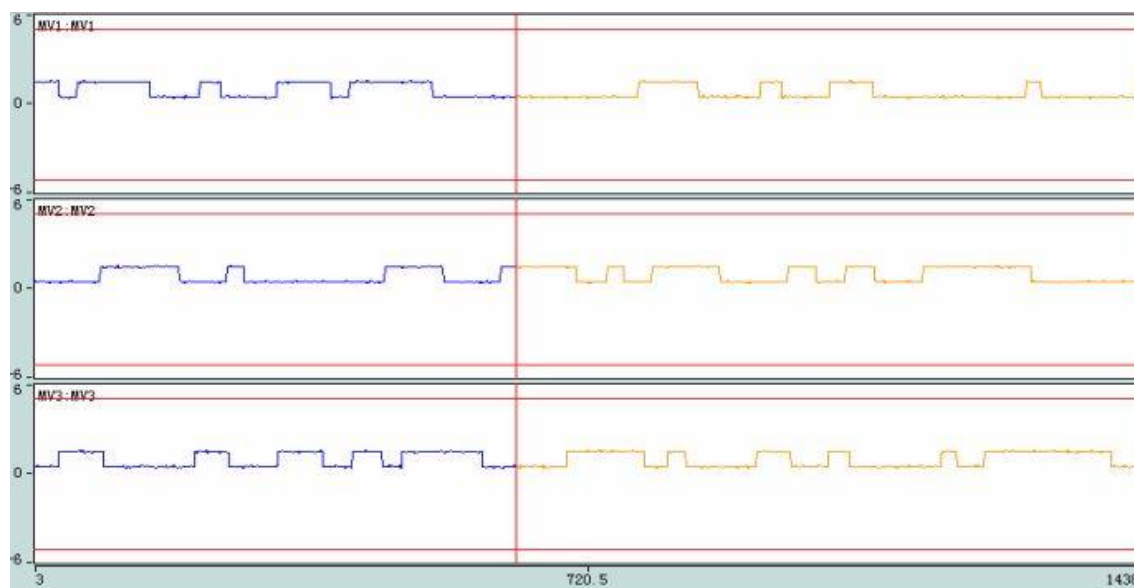


Figure 25: Step signal example [21].

The testing program gave gains for each MV and CV pairs with a rating of goodness (from A to D) based on the frequency response errors. An example of these results is presented in Figure 26. The gains appeared to be somewhat different from the earlier results. There were some problems interpreting the results.

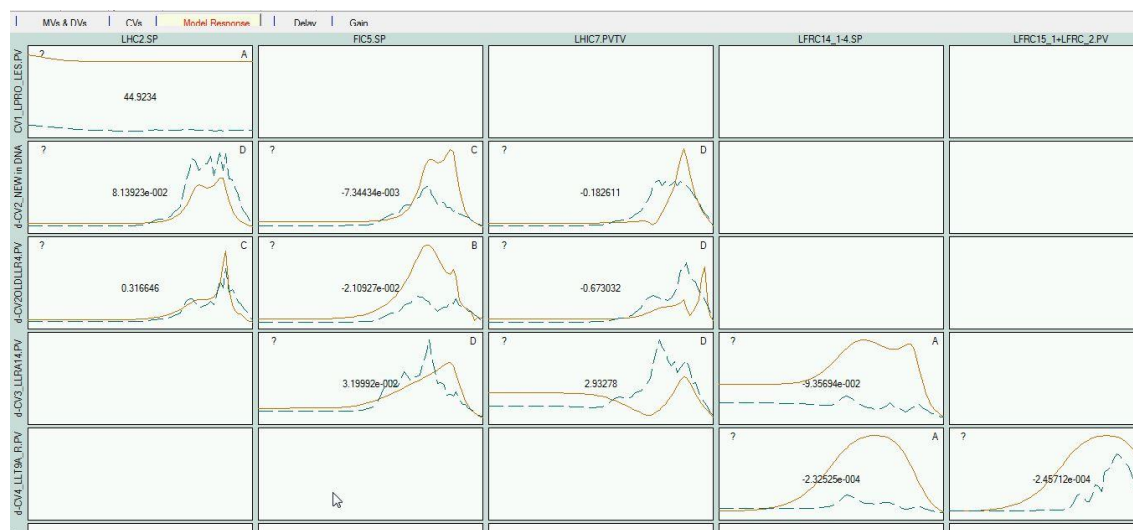


Figure 26: An example of the results based on second pilot mill tests.

Some of the gains appeared to have wrong sign. For example, the gain from blow flow to digester liquid level should be negative; liquid level should be decreasing when blow flow from the digester is increased. Also, temperature increase in lower circulation C8 should have negative response to Kappa number and positive response to H-factor.

6.3.3 Model identification based on mass balance calculations

Third identification phase included comparing earlier models with models based on the physical dimensions of the digester. Changes in the level were calculated based on mass flow in and out of the vessel and the dimensions of the vessel.

IV level and digester chip level models were calculated. Model for the IV level included chip meter speed and sluice flow. Chip level model included sluice flow and blow flow.

First, some information about the vessels had to be discovered. Calculation required volume information: the level measurements and diameters of the vessels. Volumes of measuring areas were calculated based on this information. Then the mass flow into the IV had to be calculated. Chip meter speed provides data as rounds per minute (rpm). Mass of one rpm was known but the volume of one revolution had to be calculated based on chip data like density, moisture, and yield.

Level change for IV and digester needed two operating point values for CMS, sluice flow and blow flow. Level change (%/min) was calculated based on this information and, as a result, gains could be calculated. Dilution factor and packing degree had to be taken into account.

6.3.4 Model identification based on third pilot mill tests

The second pilot mill test was repeated. The same Tai-Ji identification software was used with a larger step size and longer step durations. An example of the results is shown in Figure 27. The model gains were of wrong sign from sluice flow to digester chip level and in almost every gain related to Kappa number.

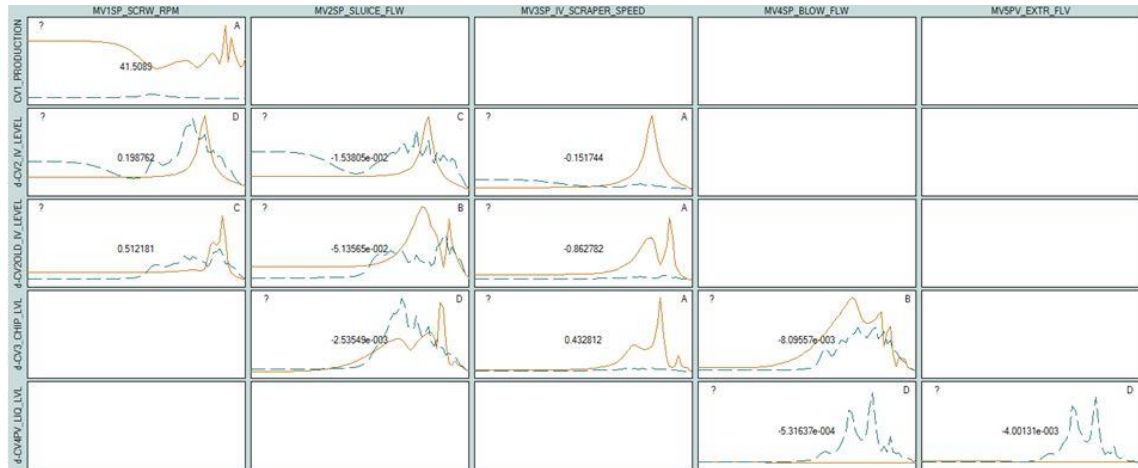


Figure 27: Tai-Ji test results from third pilot mill test.

6.3.5 Delay identification

Delays can be identified from pilot mill tests. Tai-Ji software also calculates delays. However, very long delays, characteristic to pulping process, appeared to be problematic. The delays not obtained from step response tests were set based on process knowledge. Some of the delays calculated from step response tests are presented in Table 1.

Table 1: Delays calculated based on step response tests.

DELAYS, s		MV1	MV2	MV3	MV4	MV5
		Chip Meter	Sluice	IV SCRAPER	Blow Flow	Extraction
CV1	Production	0				
CV2	IV Level	480	630	300		
CV3	DIG Level		0	0	180	
CV4	LIQ Level				0	150

6.3.6 Choosing the best models

The model parameters were set based on model comparison and process knowledge. Gain parameters chosen for simulation test are given in Figure 28. An example of the dynamic matrix with selected model parameters are presented in Figure 29 and Figure 30.

	SCRW_RPM	IV_SLUICE_FLW	IV_SCRAPER_SPT	DIG_TOT_BLOW_F	DIG_TOT_EXTR_F	DIG_SCRAPER_S	A_WJ_RATIO1_TAF	A_WJ_RATIO2_TAF	DIG_TOP_TEMP	DIG_LOW_COOK	KAPPA_DV_PREC	DIG_PRD_TARG
DIG_PRD	44						0	0	0	0	0	0
IV_LEVEL	41.2	-1.804	-1.1205				0	0	0	0	0	0
DIG_CHIP_LVL		12.3	0.83	-17.01		0	0	0	0	0	0	0
DIG_CHIP_LVL_2		12.3	0.83	-17.01		0	0	0	0	0	0	0
PRD_2_SCRAPER_RATIO	0.005						0	0	0	0	0	0
SLUICE_2_SCRAPER_RATIO		0.2	-33.33				0	0	0	0	0	0
DIG_PRD_2_TOT_BLW_FLW_RATIO	0			-0.007			0	0	0	0	0	0
DIG_TOT_BLOW_CV_FLW				1			0	0	0	0	0	0
DIG_LIQ_LVL												
DIG_LIQ_LVL_2					-0.0836	0	0	0	0	0	0	0
DIG_LVL_DIFF					-0.0836	0	0	0	0	0	0	0
DIG_RESID_ALK1	0	0	0	0	0	0	2		0	0	0	0
DIG_RESID_ALK2	0	0	0	0	0	0	1.4		0	0	0	0
DIG_REJECT	0	0	0	0	0	0	0	0	0	0	0	0.00
DIG_TEMP_DIFF	0	0	0	0	0	0	0	0	0	0	0	0
H_FACTOR_REAL	0	0	0	0	0	0	0	0	0	0	0	0
KAPPA_PRED	0	0	0	0	0	0	0	-3.49	-0.7	0.49861	0	0

Figure 28: Gain parameters chosen based on identification tests and process knowledge.

	SCRW_RPM	IV_SLUICE_FLW	IV_SCRAPER_SPEED	DIG_TOT_BLOW_FLW	DIG_TOT_EXTR_FLW
DIG_PRD		0	0	0	0
IV_LEVEL				0	0
DIG_CHIP_LVL	0			0	0
DIG_CHIP_LVL_2	0	0	0		0
PRD_2_SCRAPER_RATIO	0	0		0	0
SLUICE_2_SCRAPER_RATIO	0		0	0	0
DIG_PRD_2_TOT_BLW_FLW_RATIO	0	0	0		0
DIG_TOT_BLOW_CV_FLW	0	0	0		0
DIG_LIQ_LVL	0	0	0	0	

Figure 29: An example of the dynamic matrix with the chosen model parameters. Integrating models can be seen as increasing or decreasing lines.


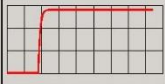

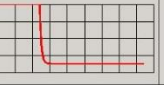
	A_W_RATIO1_TARG	A_W_RATIO2_TARG	DIG_TDP_TEMP	DIG_LOW_COOK_TEMP	KAPPA_DV_PRED	DIG_FRD_TARG
DIG_RESID_ALK1		0	0	0	0	0
DIG_RESID_ALK2		0	0	0	0	0
DIG_REJECT	0	0	0	0	0	0
DIG_TEMP_DIFF	0	0	0	0	0	0
H_FACTOR_REAL	0	0	0	0	0	0
KAPPA_FRD	0	0			0	0

Figure 30: An example of the dynamic matrix with the chosen model parameters. Delay can be noticed.

6.4 Pre-tuning with the simulator

Pre-tuning is done in the MPC software simulator. The controller is tuned by making modifications to the variable parameters. Changes in the constraints and weightings can easily be simulated. Also horizon length changes can be simulated.

7 SIMULATION RESULTS AND ANALYSIS

The implemented MPC controller was tested both by simulation with the MPC software and, based on simulation results, real life tests at the pilot mill were allowed after this thesis work. The simulation results are represented and analyzed in this chapter.

7.1 Simulation plan

A testing environment was built on a testing server. The MPC software has a built in simulation feature. After entering normal operation process values to the simulator, abnormal conditions can be simulated. For example, level of digester can be set low and the controller actions can be simulated. The simulation plan consisted of three main cases: level simulations, alkali simulations, and Kappa simulations.

7.2 Simulation of IV and Digester levels

Most of the simulations are level scenarios. The level of impregnation vessel and digester are important to be kept under control in different situations. Level simulation plan consisted of situations in which one of the levels is clearly below or above its setpoint. Under normal operation conditions IV level was taken to be 50%, digester chip level 55%, production 1145 tons, and total blow flow 370 m³/h. All the normal state values are presented in Figure 31 and Figure 32.

Tuning [?] [X]

Controlled | Manipulated | FeedForwards | Optimization | Advanced | Options

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	50	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	55.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	55.00	20.00	80.00	150.00	150.00	1.00	0.00	ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00	250.00	1.00	1.00	1.00	0.00	OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00	55.00	3.00	3.00	1.00	0.00	OFF	1
7	DIG_PRD_2_TOT_BLOW_FLW_RATIO	2.70	2.60	2.80	500.00	500.00	1.00	0.00	ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00	370.00	500.00	500.00	1.00	0.00	ON	1
9	DIG_LIQ_LVL	60.00	55.00	65.00	10.00	10.00	1.00	0.00	ON	1
10	DIG_LIQ_LVL_2	60.00	20.00	80.00	10.00	10.00	1.00	0.00	ON	1
11	DIG_LVL_DIFF	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	1
12	DIG_RESID_ALK1	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	2
13	DIG_RESID_ALK2	7.00	6.00	8.00	1.00	1.00	1.00	0.00	ON	2
14	DIG_REJECT	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	2
15	DIG_TEMP_DIFF	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
16	H_FACTOR_REAL	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
17	KAPPA_PRED	88.00	87.50	88.50	20.00	20.00	1.00	0.00	ON	3

Refresh from Online | Update | Cancel

Figure 31: CV values in normal operation conditions.

Tuning [?] [X]

Controlled | **Manipulated** | FeedForwards | Optimization | Advanced | Options

ID	Name	Value	Min	Max	Step Low	Step Hi	MovSupp	On/Off	Grp
1	SCRW_RPM	25.00	22.00	27.00	0.1	0.1	20	ON	1
2	IV_SLUICE_FLW	195.00	150.00	220.00	1.5	1.5	50	ON	1
3	IV_SCRAPER_SPEED	5.00	3.00	10.00	0.01	0.01	50	OFF	1
4	DIG_TOT_BLOW_FLW	370.00	350.00	400.00	0.2	0.2	200	ON	1
5	DIG_TOT_EXTR_FLW	470.00	350.00	500.00	0.1	0.1	20	ON	1
6	DIG_SCRAPER_SPEED	7.00	5.00	12.00	0.01	0.01	20	ON	1
7	A_W_RATIO1_TARG	17.00	15.00	19.00	0.05	0.05	10	ON	2
8	A_W_RATIO2_TARG	0.00	0.00	0.00	0	0	0	OFF	2
9	DIG_TOP_TEMP	157.00	156.00	159.00	0.02	0.02	40	OFF	3
10	DIG_LOW_COOK_TEMP	116.00	114.00	118.00	0.01	0.01	10	ON	3

Refresh from Online | Update | Cancel

Figure 32: MV values in normal operation conditions.

Level simulation scenarios fall into four phases. In the first phase IV level was abnormal. IV level was first set to 0% in *simulation 1* and to 100% in *simulation 2*. In the second phase the chip level was abnormal in *simulations 3* and *4*. In third phase both levels were changed from their normal values in *simulations 5-8*. In the last phase, the screw control for IV level control was disabled, *simulation 9*.

7.2.1 Level simulation 1

Figure 33 shows the settings where IV level was simulated from initial value of 0%. This simulates a situation where the level suddenly drops. The predicted IV level change and IV sluice flow manipulations can be seen in Figure 34 and predicted digester chip level changes and screw speed manipulations can be seen in Figure 35.

The screenshot shows the 'Tuning' software interface. It features a table of controlled variables and a 'Simulation Time' dialog box. The table has columns for ID, Name, Value, Min, Max, Min Wt, Max Wt, NoiseFilter, Rotation, On/Off, and Grp. The 'IV_LEVEL' row is highlighted in red, indicating a value outside the allowed boundaries. The 'Simulation Time' dialog box shows a 'Simulation Intervals' field set to 600 and a 'Run' button.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	0.00	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	55.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	55.00	20.00						ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00						OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00						OFF	1
7	DIG_PRD_2_TOT_BLW_FLW_RATIO	2.70	2.60						ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00						ON	1
9	DIG_LIQ_LVL	60.00	55.00						ON	1
10	DIG_LIQ_LVL_2	60.00	20.00						ON	1
11	DIG_LVL_DIFF	0.00	0.00						OFF	1
12	DIG_RESID_ALK1	0.00	0.00						OFF	2
13	DIG_RESID_ALK2	7.00	6.00						ON	2
14	DIG_REJECT	0.00	0.00						OFF	2
15	DIG_TEMP_DIFF	0.00	0.00						OFF	3
16	H_FACTOR_REAL	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
17	KAPPA_PRED	88.00	87.50	88.50	20.00	20.00	1.00	0.00	ON	3

Figure 33: IV level was set to 0% for simulation. Simulation time was set to 600 minutes. IV level minimum is red because the current value is not inside the allowed boundaries.

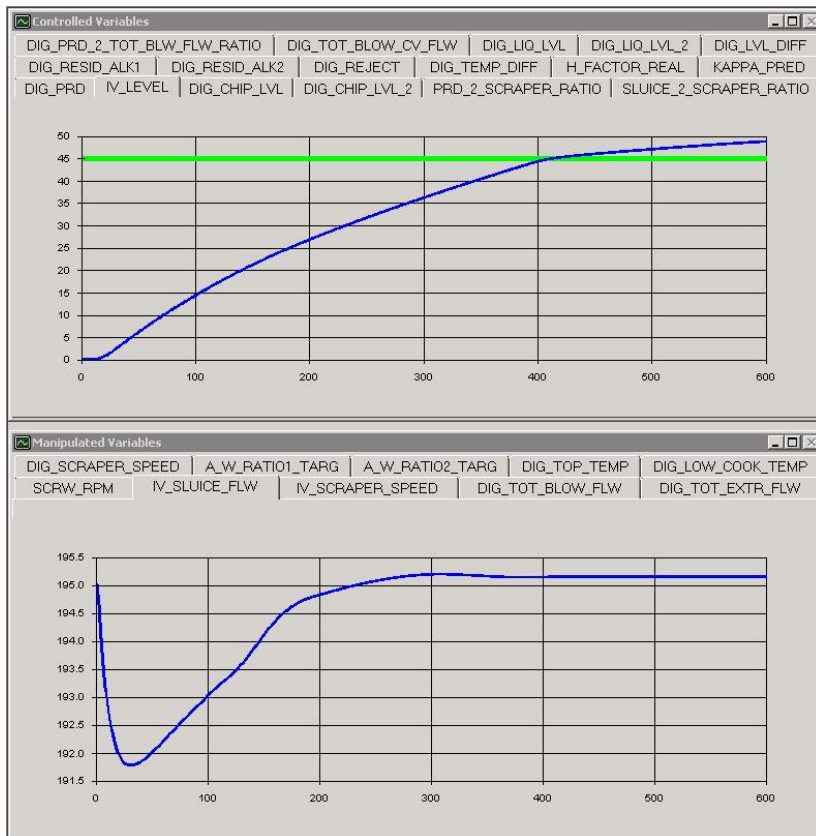


Figure 34: Simulation results for IV level and IV sluice flow when IV level was set to 0%. The green line represents IV level minimum value.

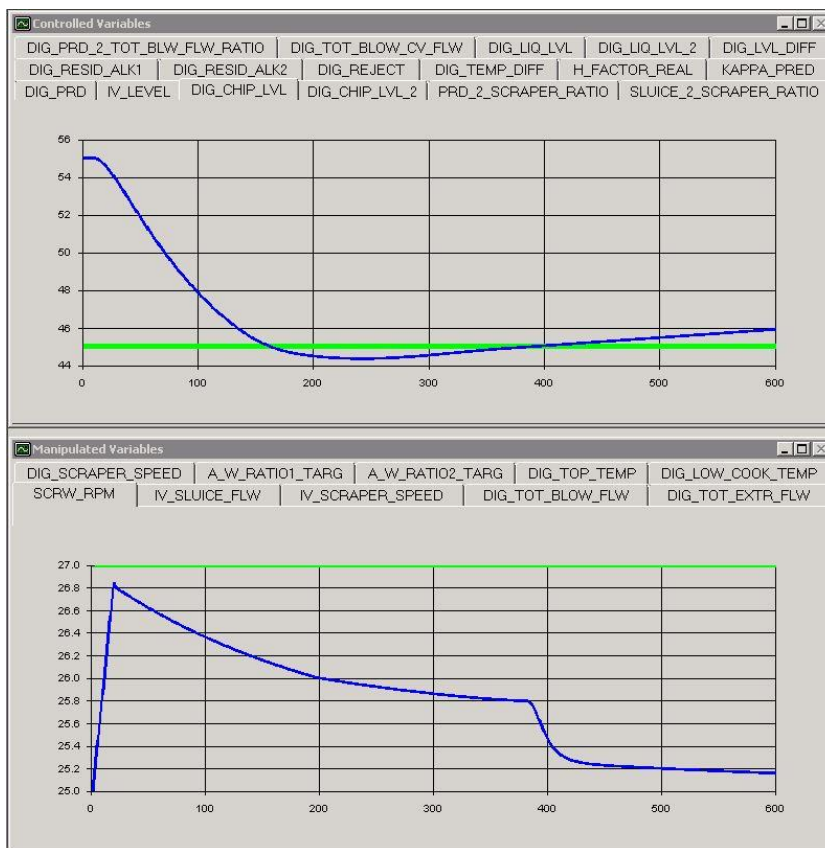


Figure 35: Simulation results for digester chip level and screw speed when IV level was set to 0%. The upper green line represents digester level minimum value. The lower green line is screw speed maximum limit.

Based on results, the controller reacts to the sudden drop of IV level by manipulating IV sluice flow and chip screw speed as wanted. Sluice flow decreases significantly and screw speed increases very fast to recover the minimum IV level. After 400 minutes, the minimum IV level is reached. Digester chip level drops slightly below minimum value 45% for 200 minutes.

7.2.2 Level simulation 2

Figure 36 shows the settings where IV level was simulated from initial value of 100%. The predicted IV level change and screw speed manipulations can be seen in Figure 37 and predicted digester chip level changes and IV sluice flow manipulations can be seen in Figure 38.

The screenshot shows the 'Tuning' window with a table of controlled variables. A 'Simulation Time' dialog box is open, showing 'Simulation Intervals' set to 600 and a value of 7.79. The dialog has 'Run' and 'Done' buttons.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	100.00	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	55.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	55.00	20.00	80.00	150.00	150.00	1.00	0.00	ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00						00 OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00						00 OFF	1
7	DIG_PRD_2_TOT_BLOW_FLW_RATIO	2.70	2.60						00 ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00						00 ON	1
9	DIG_LIQ_LVL	60.00	55.00						00 ON	1
10	DIG_LIQ_LVL_2	60.00	20.00						00 ON	1
11	DIG_LVL_DIFF	0.00	0.00						00 OFF	1
12	DIG_RESID_ALK1	0.00	0.00						00 OFF	2
13	DIG_RESID_ALK2	7.00	6.00						00 ON	2
14	DIG_REJECT	0.00	0.00						00 OFF	2
15	DIG_TEMP_DIFF	0.00	0.00						00 OFF	3
16	H_FACTOR_REAL	0.00	0.00						00 OFF	3
17	KAPPA_PRED	88.00	87.50	88.50	20.00	20.00	1.00	0.00	ON	3

Figure 36: IV level was set to 100% for simulation. Simulation time was set to 600 minutes.

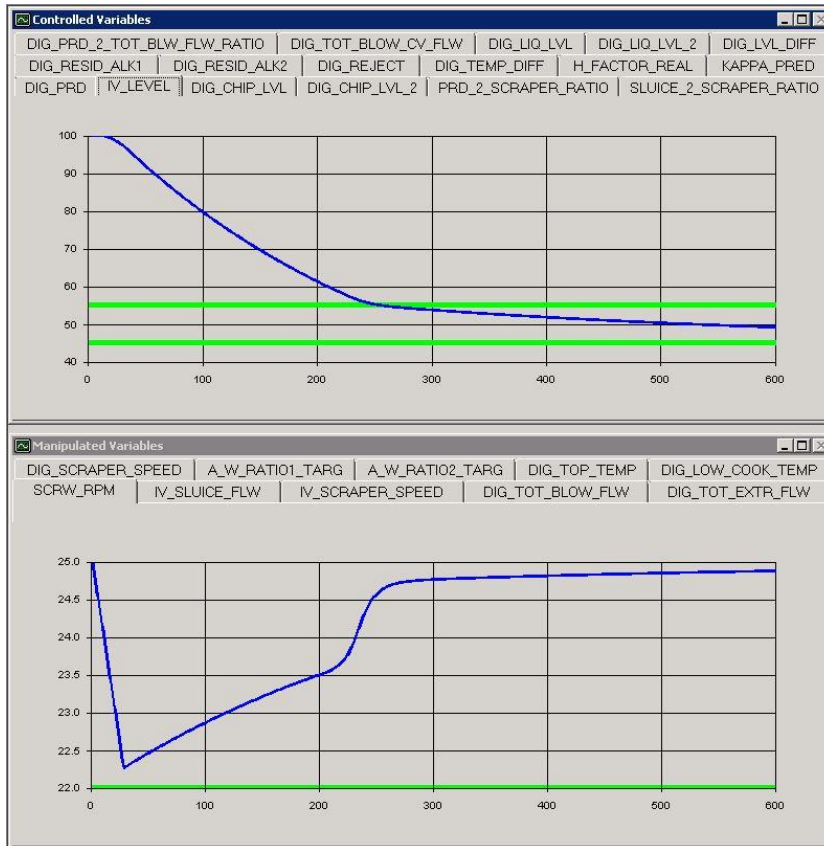


Figure 37: Simulation results for IV level and screw speed when IV level was set to 100%. The green lines represent minimum and maximum values.

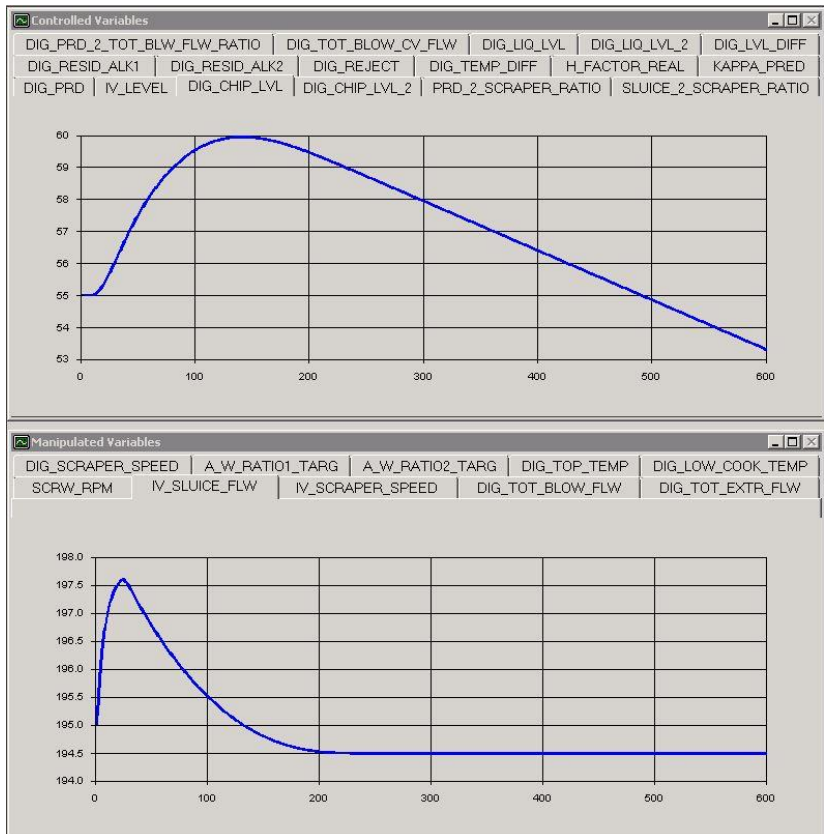


Figure 38: Simulation results for digester chip level and IV sluice flow when IV level was set to 100%.

Simulation 2 indicates, that the controller reacts to the suddenly full IV by manipulating chip screw speed and IV sluice flow again as wanted. Screw speed decreases significantly and sluice flow increases very fast to recover the desired IV level. After 250 minutes, the desired IV level range is reached. Digester chip level increases at first, but continues to drop after 600 min simulation time.

7.2.3 Level simulation 3

Figure 39 shows the settings where digester chip level was simulated from initial value of 0%. The predicted digester chip level change and IV sluice flow manipulations can be seen in Figure 40, predicted digester chip level 2 changes and digester total blow flow manipulations can be seen in Figure 41, and predicted IV level changes and screw speed manipulations can be seen in Figure 42.

The screenshot shows the 'Tuning' software interface. A table lists 17 controlled variables with their current values and ranges. A 'Simulation Time' dialog box is open, showing a simulation interval of 600 minutes and a value of 7.49. The dialog has 'Run' and 'Done' buttons. Below the table are 'Refresh from Online', 'Update', and 'Cancel' buttons.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	50.00	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	0.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	0.00	20.00	80.00	150.00	150.00	1.00	0.00	ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00	250.00	1.00	1.00	1.00	0.00	OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00	45.00	1.00	1.00	1.00	0.00	OFF	1
7	DIG_PRD_2_TOT_BLOW_FLW_RATIO	2.70	2.60	2.80	1.00	1.00	1.00	0.00	ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00	450.00	1.00	1.00	1.00	0.00	ON	1
9	DIG_LIQ_LVL	60.00	55.00	65.00	1.00	1.00	1.00	0.00	ON	1
10	DIG_LIQ_LVL_2	60.00	20.00	80.00	1.00	1.00	1.00	0.00	ON	1
11	DIG_LVL_DIFF	0.00	0.00	0.00	1.00	1.00	1.00	0.00	OFF	1
12	DIG_RESID_ALK1	0.00	0.00	0.00	1.00	1.00	1.00	0.00	OFF	2
13	DIG_RESID_ALK2	7.00	6.00	8.00	1.00	1.00	1.00	0.00	ON	2
14	DIG_REJECT	0.00	0.00	0.00	1.00	1.00	1.00	0.00	OFF	2
15	DIG_TEMP_DIFF	0.00	0.00	0.00	1.00	1.00	1.00	0.00	OFF	3
16	H_FACTOR_REAL	0.00	0.00	0.00	1.00	1.00	1.00	0.00	OFF	3
17	KAPPA_PRED	88.00	87.50	88.50	20.00	20.00	1.00	0.00	ON	3

Figure 39: Digester chip level was set to 0% for simulation. Simulation time was set to 600 minutes.

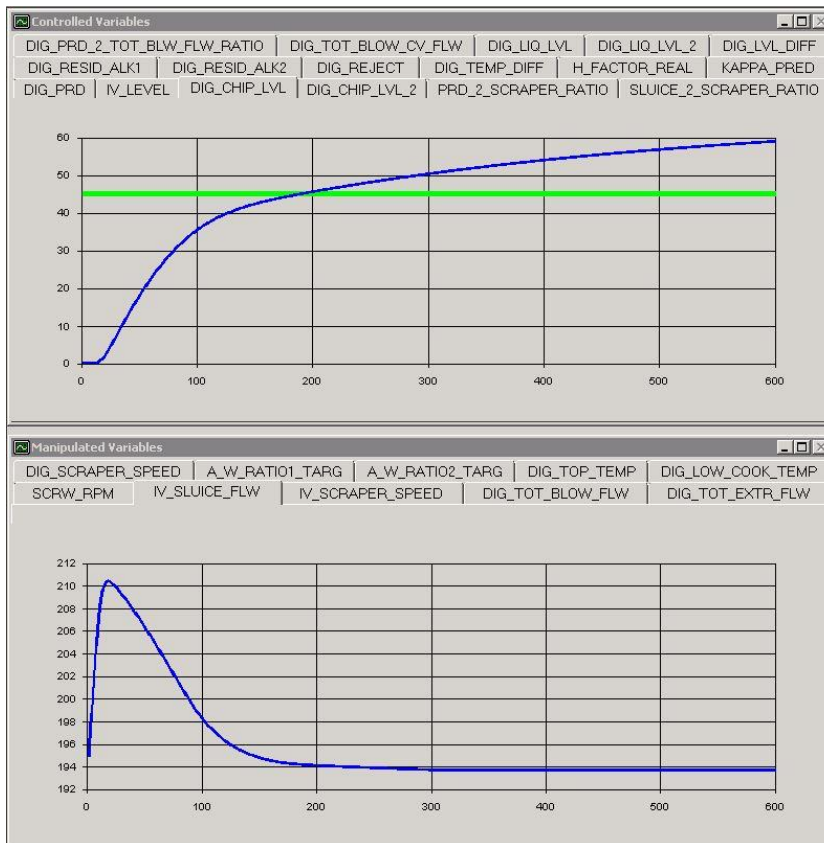


Figure 40: Simulation results for digester chip level and IV sluice flow when digester chip level was set to 0%.

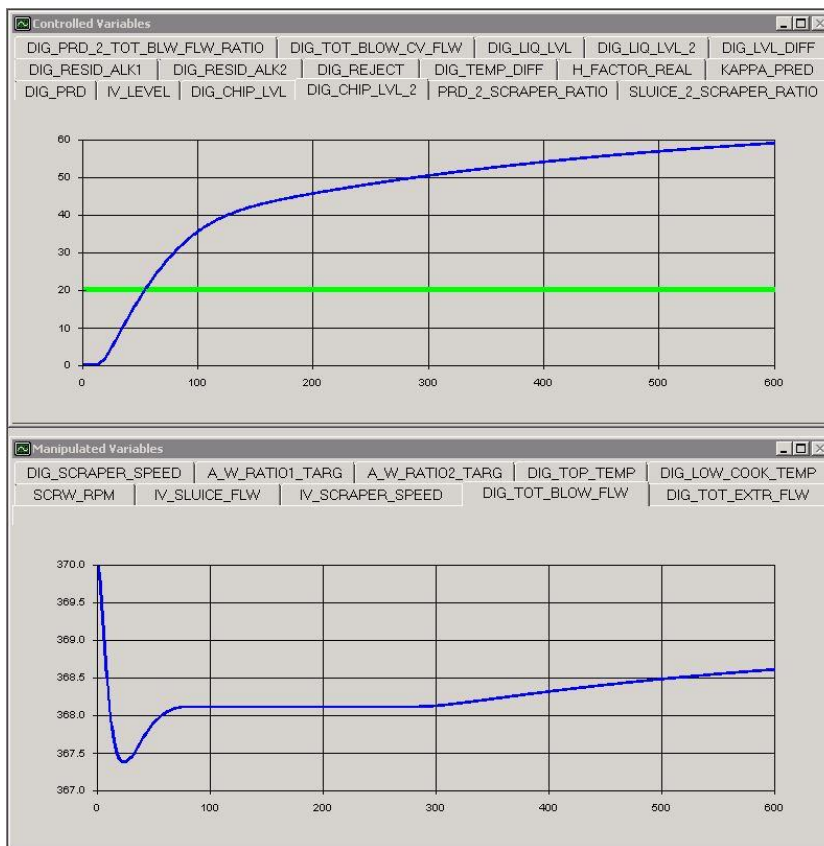


Figure 41: Simulation results for digester chip level 2 and digester total blow flow when digester chip level was set to 0%.

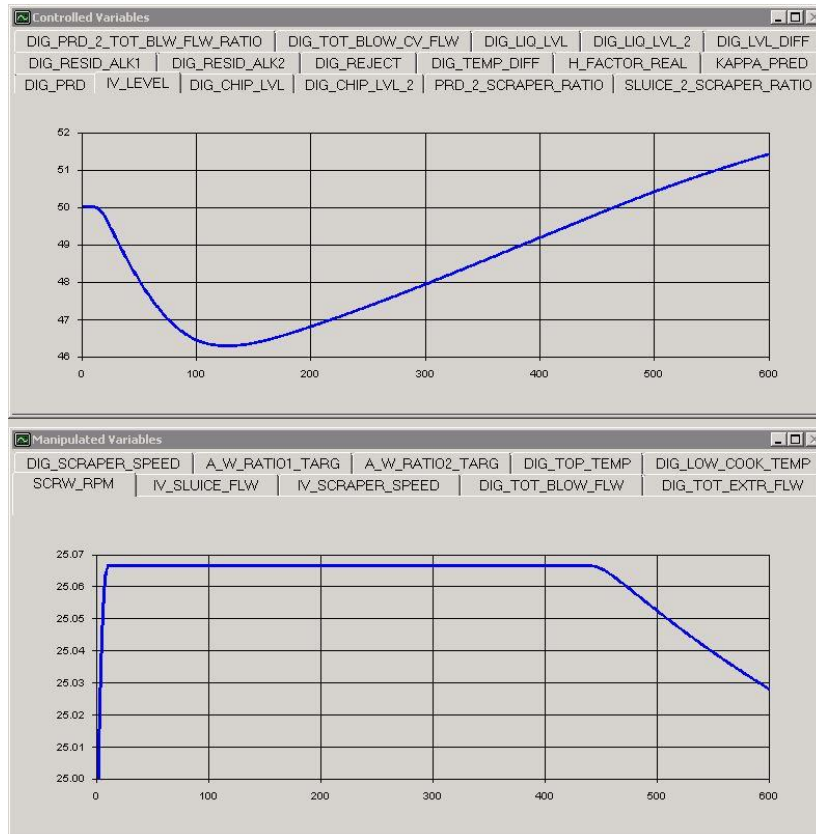


Figure 42: Simulation results for IV level and screw speed when digester chip level was set to 0%.

The results from level simulation 3 show, that controller reacts to sudden drop in digester chip level as required. IV sluice flow increases instantly to recover the level of digester. This decreases the level of impregnation vessel. Figure 41 represents a secondary digester chip level minimum, which is lower than primary digester level minimum. When this level is exceeded, the controller uses blow flow manipulation to recover the level of the digester faster. Digester chip level 2 minimum is reached in 50 min and primary minimum level in 200 min. IV level stays in acceptable range and continues to increase after 600 min.

7.2.4 Level simulation 4

Figure 43 shows the settings where digester chip level was simulated from initial value of 100%. The predicted digester chip level change and IV sluice flow manipulations can be seen in Figure 44, predicted digester chip level 2 changes and digester total blow flow manipulations can be seen in Figure 45, and predicted IV level changes and screw speed manipulations can be seen in Figure 46.

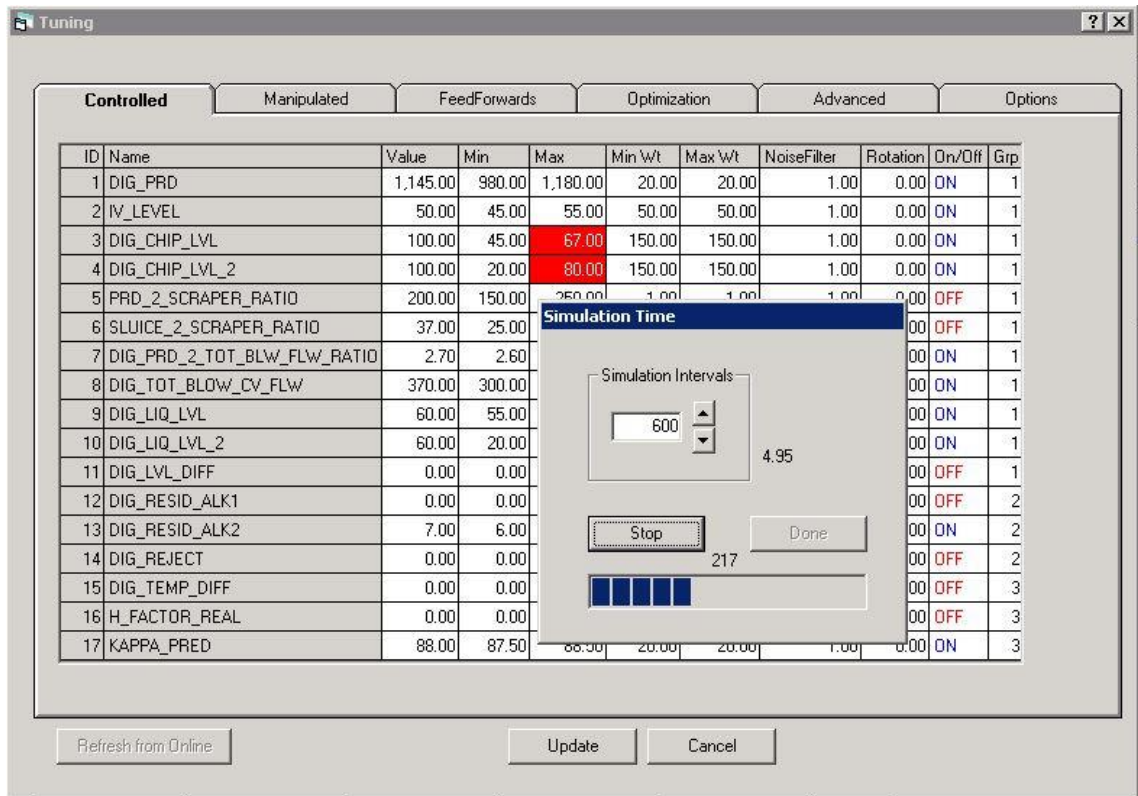


Figure 43: Digester chip level was set to 100% for simulation. Simulation time was set to 600 minutes.

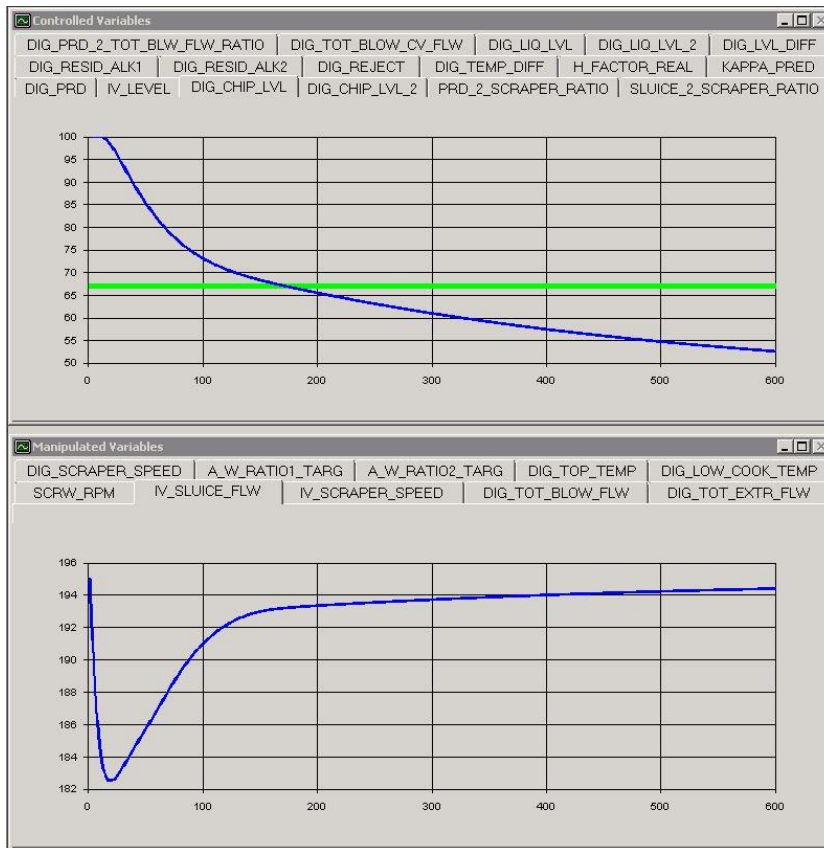


Figure 44: Simulation results for digester chip level and IV sludge flow when digester chip level was set to 100%.

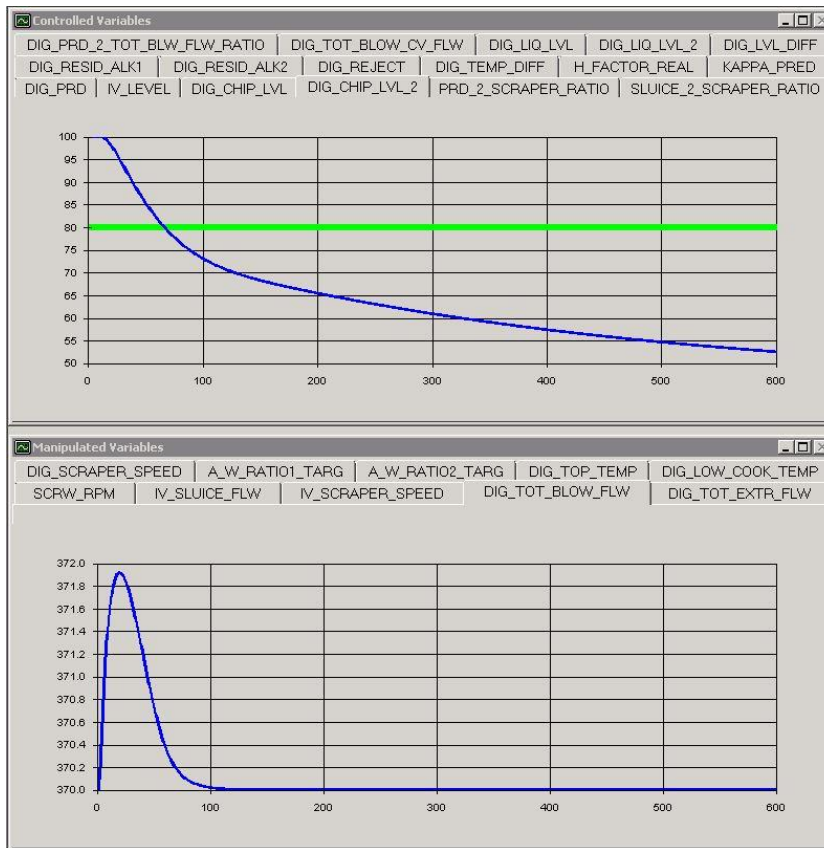


Figure 45: Simulation results for digester chip level 2 and digester total blow flow when digester chip level was set to 100%.

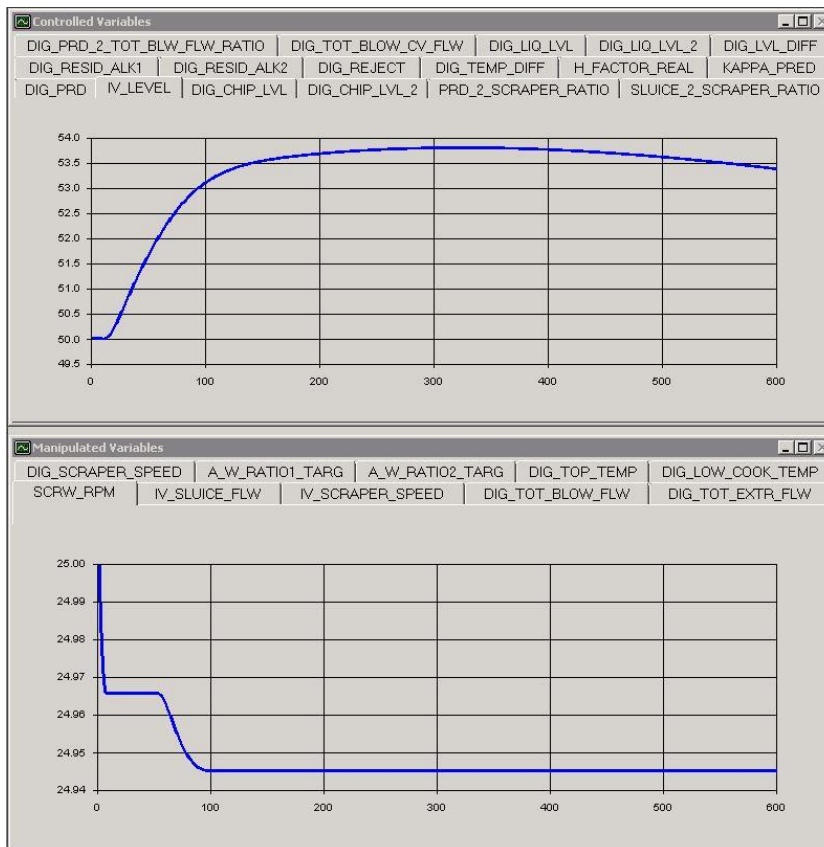


Figure 46: Simulation results for IV level and screw speed when digester chip level was set to 100%.

Level simulation 4 results indicate, that controller starts immediately to manipulate IV sluice flow and total blow flow to decrease the level of the digester. Sluice flow drops from 195 to 183. The upper maximum limit of the digester (digester chip level 2 in Figure 45) results in more aggressive changes to the level by manipulating the blow flow. Screw speed drops to a lower level and IV increases, but still remains in optimal range.

7.2.5 Level simulation 5

Figure 47 shows the settings where IV and digester chip levels were simulated from initial value of 0%. The predicted IV level change and chip screw speed manipulations can be seen in Figure 48, predicted digester chip level changes and IV sluice flow manipulations in Figure 49, and digester total blow flow manipulations are presented in Figure 50.

The screenshot shows the 'Tuning' software interface. The main window displays a table of controlled variables under the 'Controlled' tab. The table has columns for ID, Name, Value, Min, Max, Min Wt, Max Wt, NoiseFilter, Rotation, On/Off, and Grp. The 'Value' column for rows 2, 3, and 4 is highlighted in red, indicating they are set to 0.00. A 'Simulation Time' dialog box is open over the table, showing 'Simulation Intervals' set to 600 and a value of 8.93. The dialog has 'Run' and 'Done' buttons. At the bottom of the main window are 'Refresh from Online', 'Update', and 'Cancel' buttons.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	0.00	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	0.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	0.00	20.00	80.00	150.00	150.00	1.00	0.00	ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00	250.00	1.00	1.00	1.00	0.00	OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00						OFF	1
7	DIG_PRD_2_TOT_BLW_FLW_RATIO	2.70	2.60						ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00						ON	1
9	DIG_LIQ_LVL	60.00	55.00						ON	1
10	DIG_LIQ_LVL_2	60.00	20.00						ON	1
11	DIG_LVL_DIFF	0.00	0.00						OFF	1
12	DIG_RESID_ALK1	0.00	0.00						OFF	2
13	DIG_RESID_ALK2	7.00	6.00						ON	2
14	DIG_REJECT	0.00	0.00						OFF	2
15	DIG_TEMP_DIFF	0.00	0.00						OFF	3
16	H_FACTOR_REAL	0.00	0.00						OFF	3
17	KAPPA_PRED	88.00	87.50						ON	3

Figure 47: IV level and digester levels set to 0% for simulation.

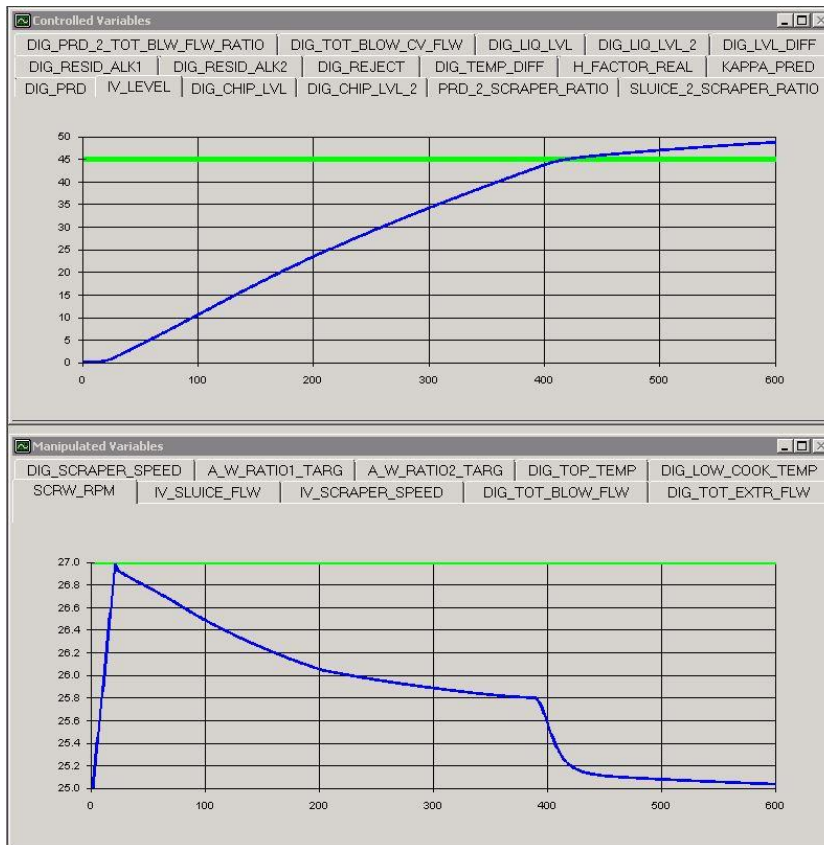


Figure 48: Simulation results for IV level and screw speed when IV and digester chip level were set to 0%.

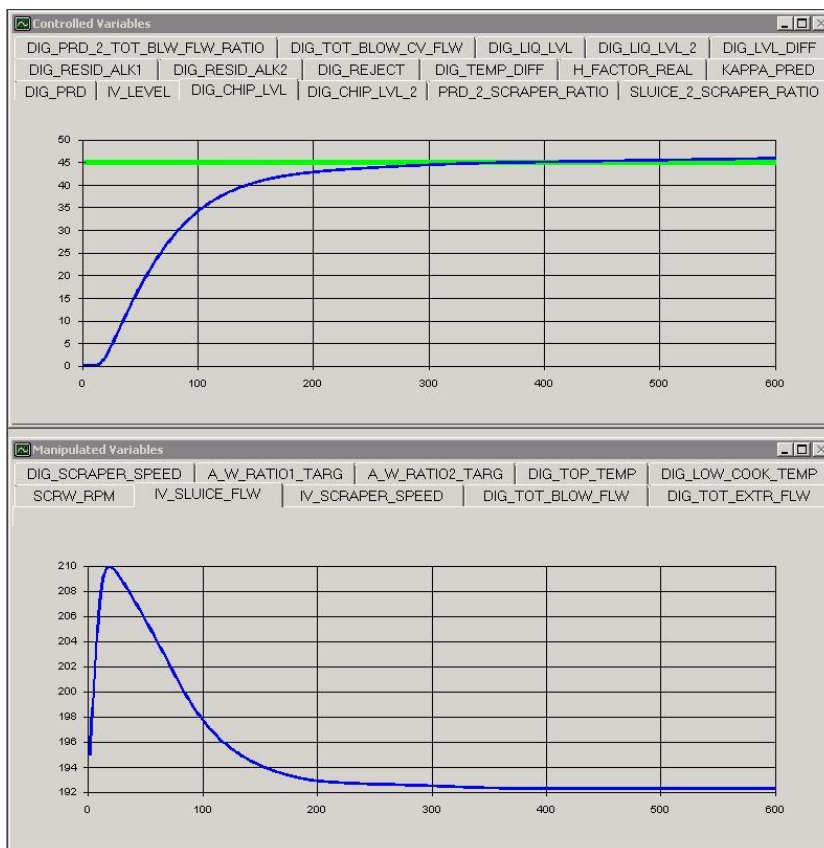


Figure 49: Simulation results for digester chip level and IV sluice flow when IV and digester chip level were set to 0%.

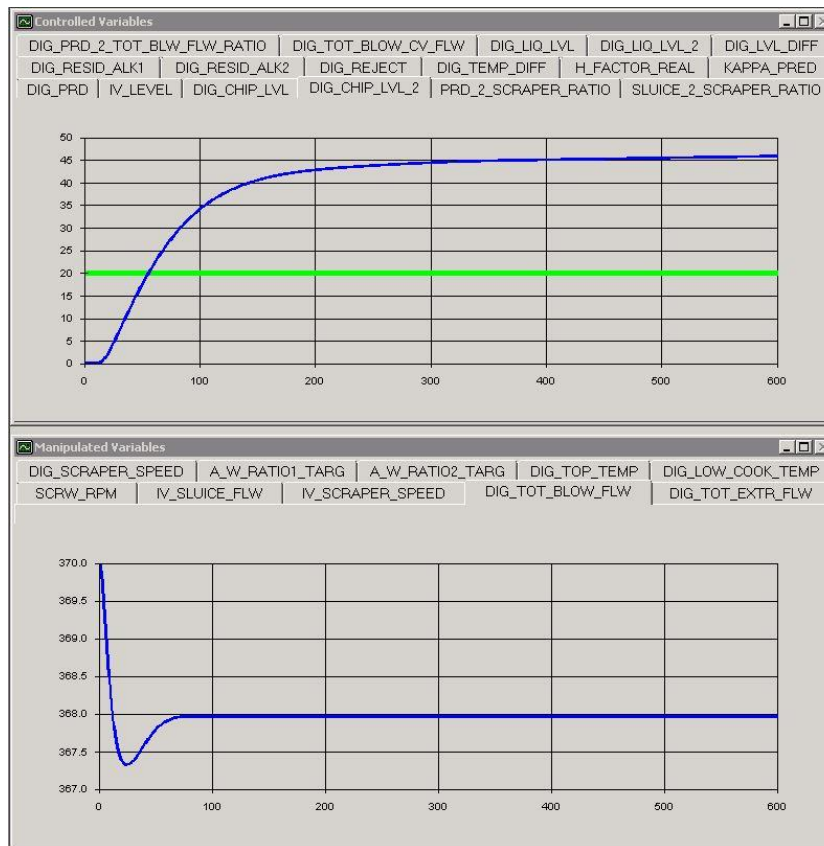


Figure 50: Simulation results for digester chip level 2 and digester total blow flow when IV and digester chip level were set to 0%.

Results show that controller starts to manipulate chip screw speed, IV sluice flow and digester blow flow. Blow flow decreases to help recovering the digester chip level and chip screw speed increases to help recovering the IV level. The IV sluice flow increases for a short time and then decreases. Blow flow manipulation is active only when digester chip level 2 is out of range. IV level reaches the minimum range in 400 minutes and digester chip level in 300 minutes.

7.2.6 Level simulation 6

Figure 51 shows the settings where IV and digester chip levels were simulated from initial value of 100%. The predicted IV level change and chip screw speed manipulations can be seen in Figure 52, predicted digester chip level changes and IV sluice flow manipulations in Figure 53, and digester total blow flow manipulations are presented in Figure 54.

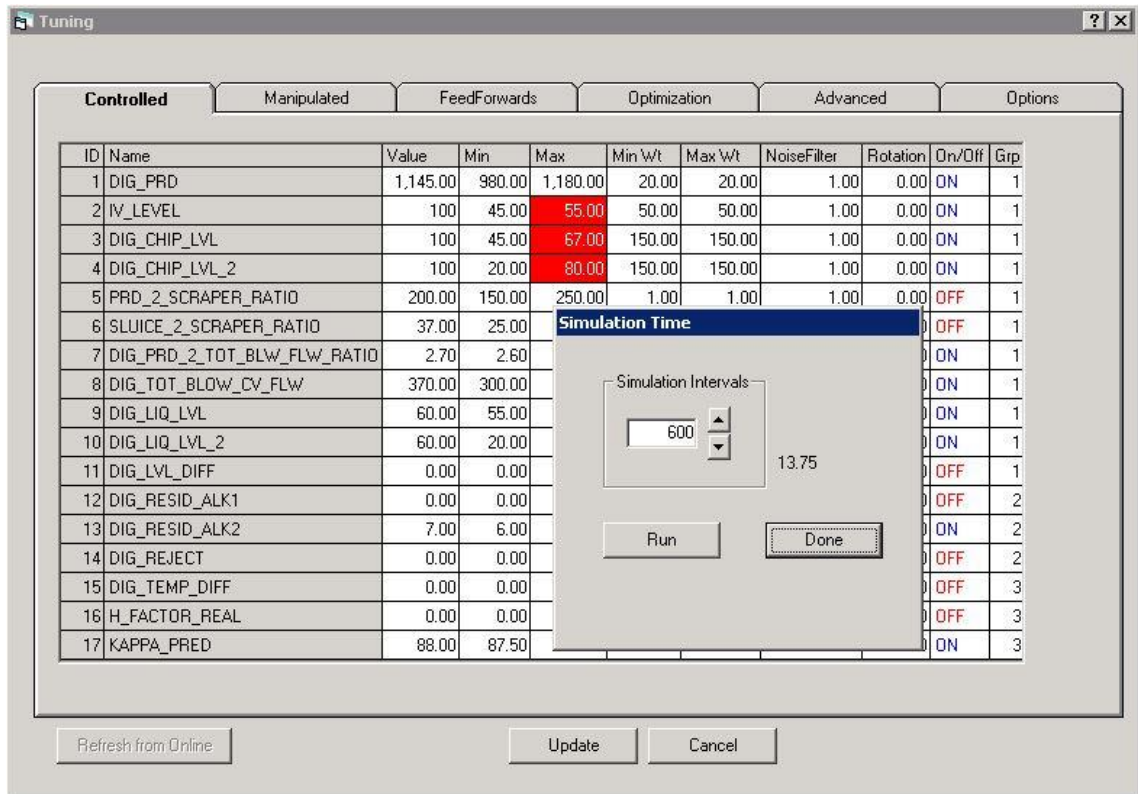


Figure 51: IV level and digester levels set to 100% for simulation.

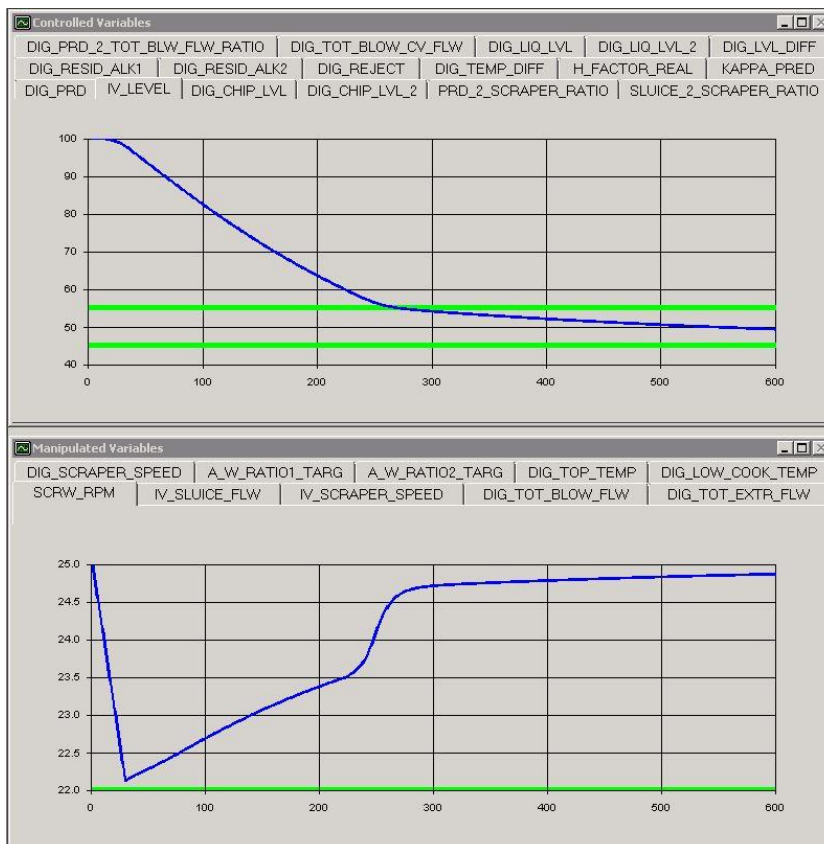


Figure 52: Simulation results for IV level and screw speed when IV and digester chip level were set to 100%.

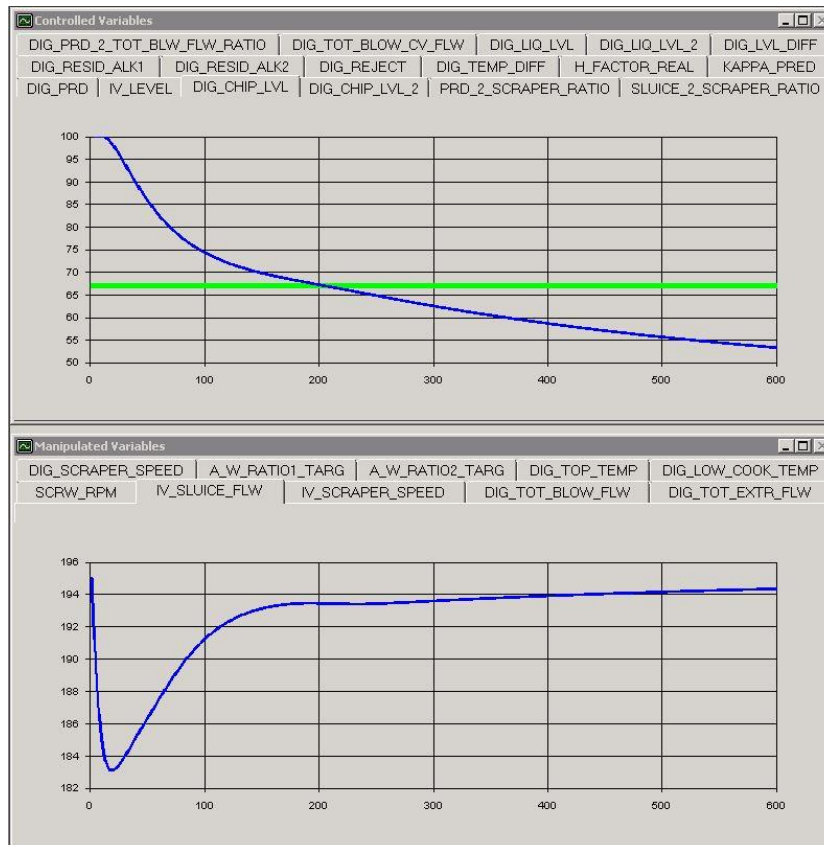


Figure 53: Simulation results for digester chip level and IV sluice flow when IV and digester chip level were set to 100%.

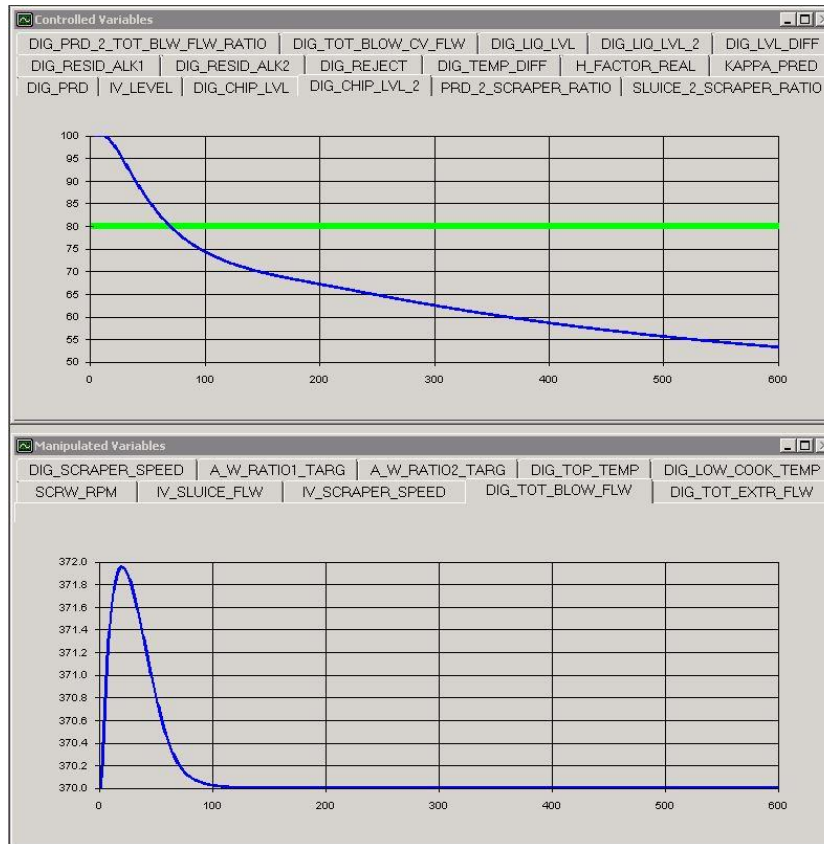


Figure 54: Simulation results for digester chip level 2 and digester total blow flow when IV and digester chip level were set to 100%.

Simulation 6 indicates similar behavior as simulation 5, with opposite manipulations. Screw speed, IV sluice flow and blow flow are manipulated to recover both vessel levels. IV level maximum range is reached now faster in 250 minutes. Digester chip level range is reached in 200 minutes.

7.2.7 Level simulation 7

Figure 55 shows the settings where IV level was simulated from initial value of 0% and digester chip level was simulated from initial value of 100%. The predicted IV level change and chip screw speed manipulations can be seen in Figure 56, predicted digester chip level changes and IV sluice flow manipulations in Figure 57, and digester total blow flow manipulations are presented in Figure 58.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	0	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	100.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	100.00	20.00	80.00	150.00	150.00	1.00	0.00	ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00	250.00	1.00	1.00	1.00	0.00	OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00						OFF	1
7	DIG_PRD_2_TOT_BLOW_FLW_RATIO	2.70	2.60						ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00						ON	1
9	DIG_LIQ_LVL	60.00	55.00						ON	1
10	DIG_LIQ_LVL_2	60.00	20.00						ON	1
11	DIG_LVL_DIFF	0.00	0.00						OFF	1
12	DIG_RESID_ALK1	0.00	0.00						OFF	2
13	DIG_RESID_ALK2	7.00	6.00						ON	2
14	DIG_REJECT	0.00	0.00						OFF	2
15	DIG_TEMP_DIFF	0.00	0.00						OFF	3
16	H_FACTOR_REAL	0.00	0.00						OFF	3
17	KAPPA_PRED	88.00	87.50						ON	3

Simulation Time dialog box:

Simulation Intervals: 600 (up/down arrows), 18.22

Buttons: Run, Done

Main window buttons: Refresh from Online, Update, Cancel

Figure 55: IV level set to 0% and digester level set to 100% for simulation.

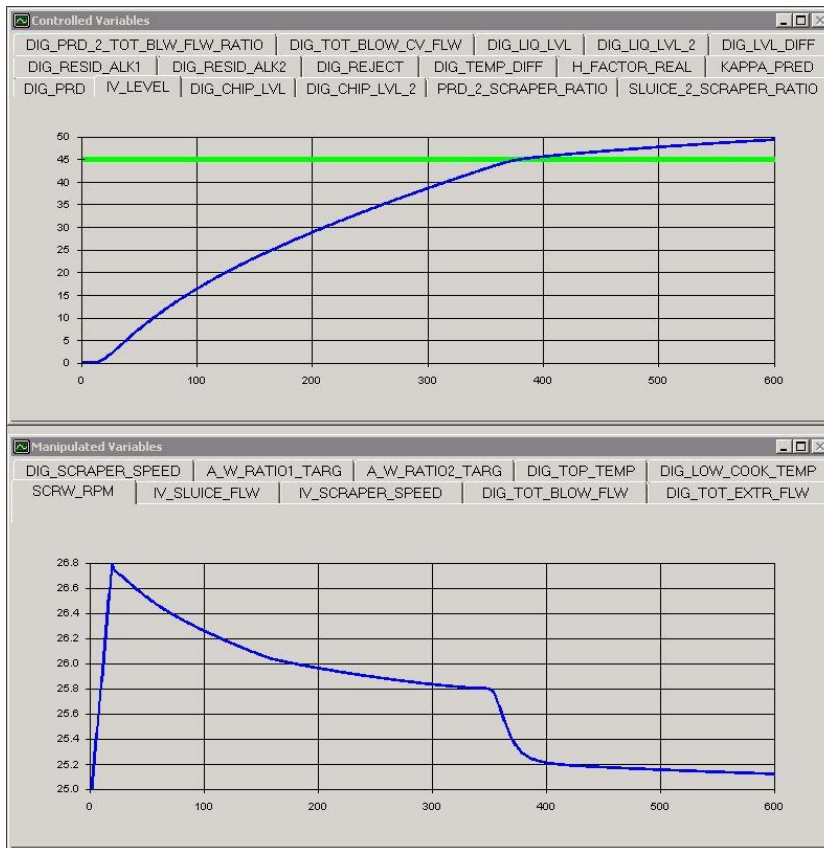


Figure 56: Simulation results for IV level and screw speed when IV level was set to 0% and digester chip level was set to 100%.

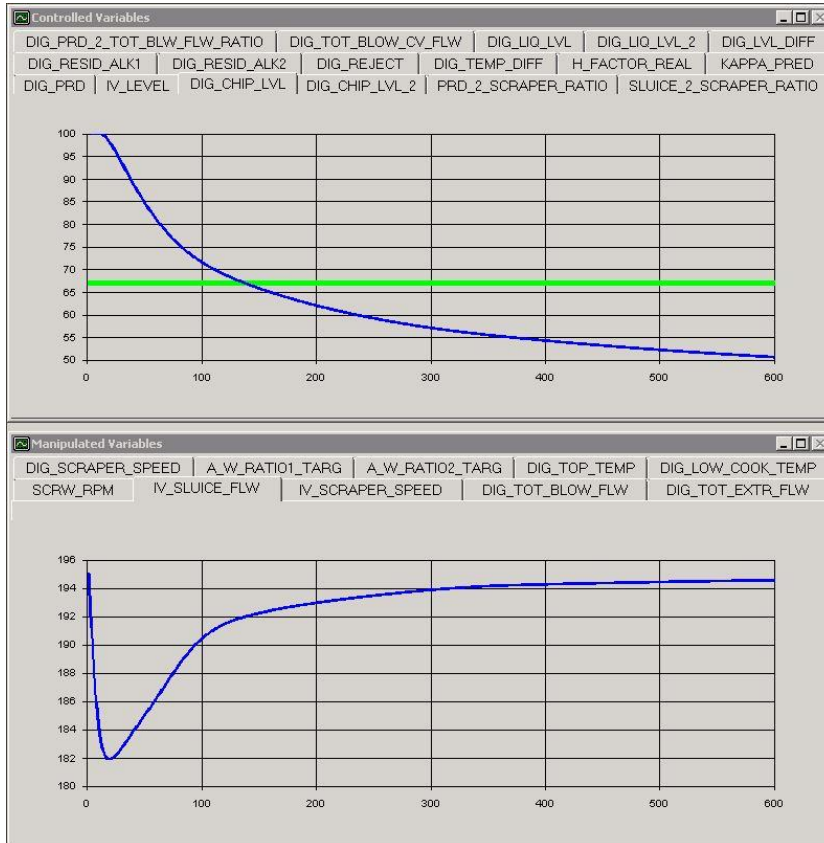


Figure 57: Simulation results for digester chip level and IV sluice flow when IV level was set to 0% and digester chip level was set to 100%.

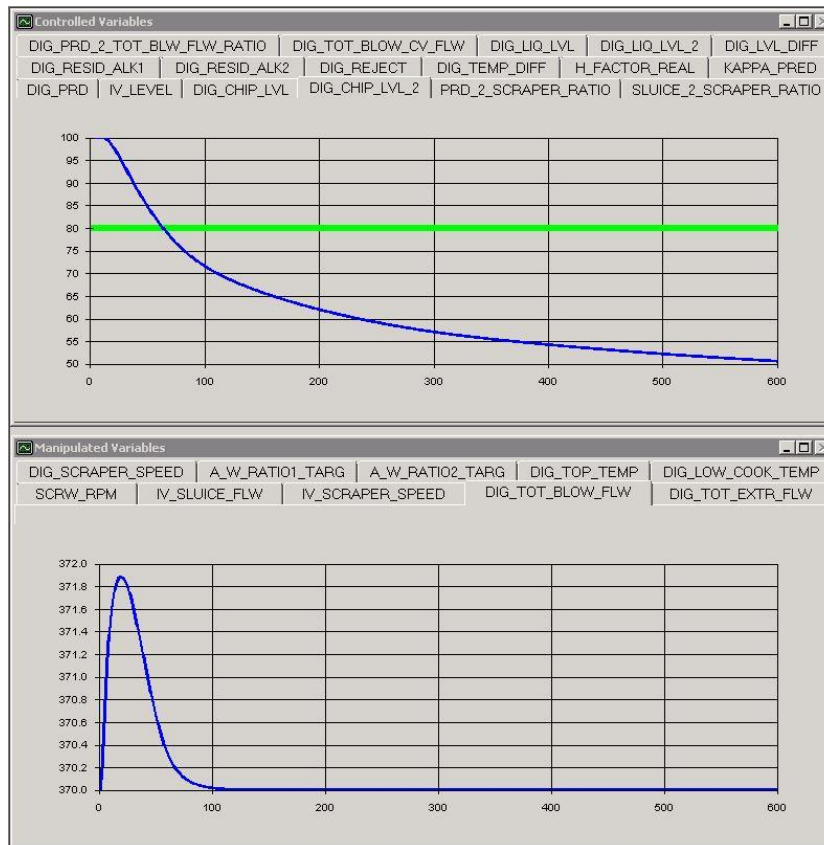


Figure 58: Simulation results for digester chip level 2 and digester total blow flow when IV level was set to 0% and digester chip level to 100%.

This simulation shows how the controller works when IV level is low and digester level is high. Screw speed increases and blow flow decreases. IV sluice flow is manipulated almost identically as in the simulation 6. Sluice flow is now primarily used to decrease digester level. IV level minimum is reached at 370 minutes and digester chip level at 150 minutes.

7.2.8 Level simulation 8

Figure 59 shows the settings where IV level was simulated from initial value of 100% and digester chip level was simulated from initial value of 0%. The predicted IV level change and chip screw speed manipulations can be seen in Figure 60, predicted digester chip level changes and IV sluice flow manipulations in Figure 61, and digester total blow flow manipulations are presented in Figure 62.

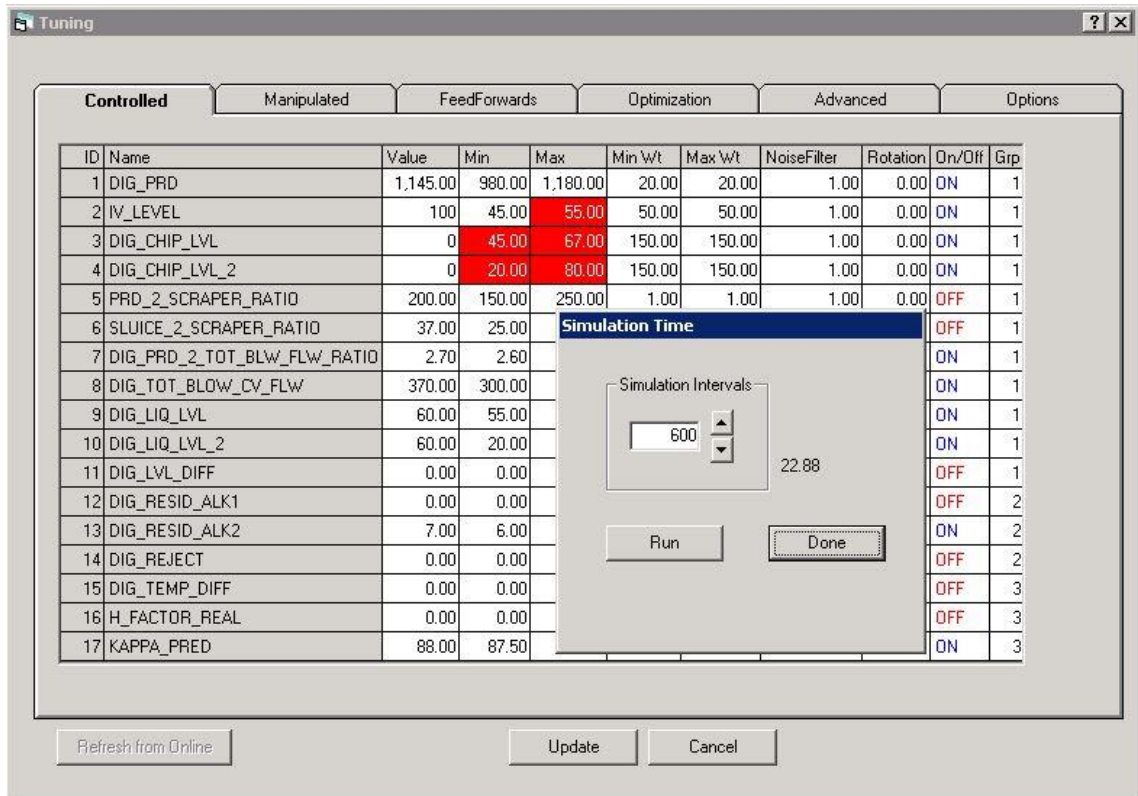


Figure 59: IV level set to 100% and digester level set to 0% for simulation.

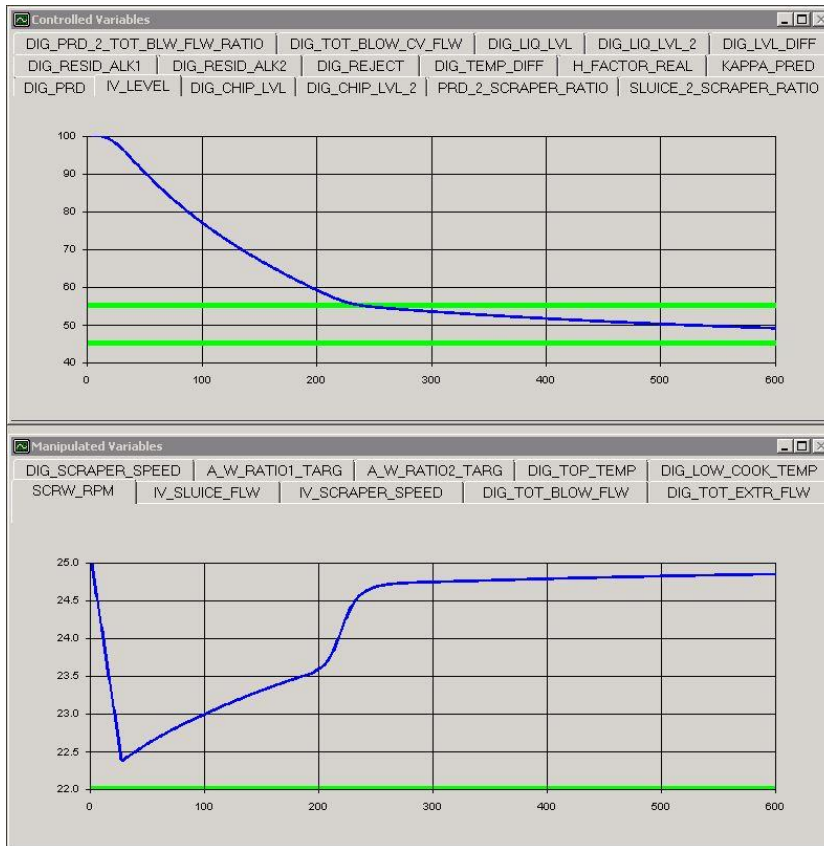


Figure 60: Simulation results for IV level and screw speed when IV level was set to 100% and digester chip level was set to 0%.

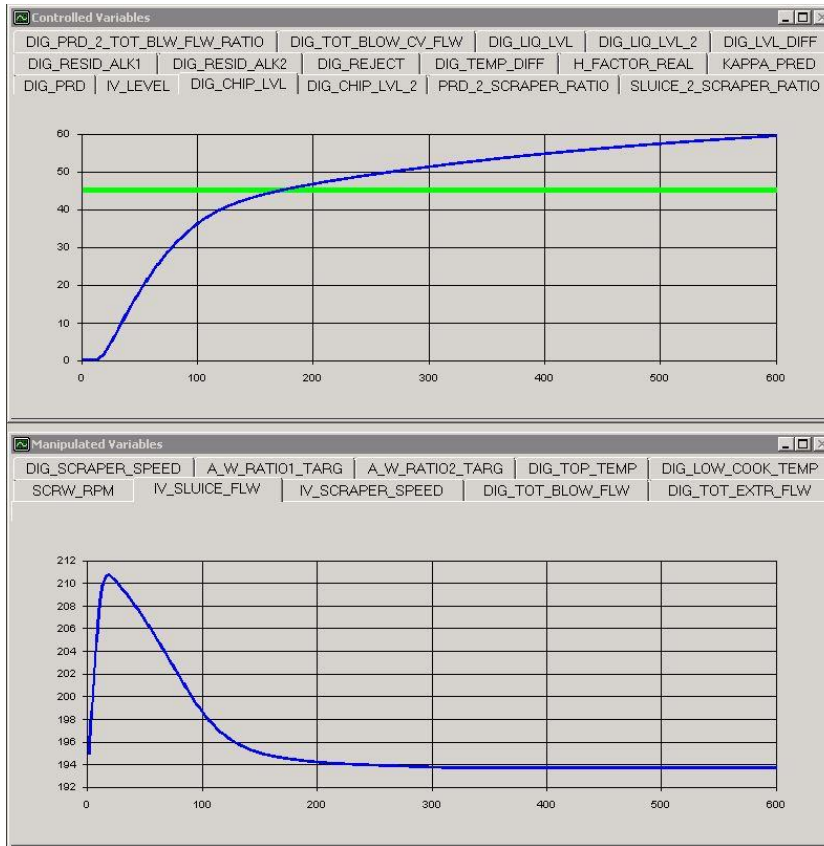


Figure 61: Simulation results for digester chip level and IV sluice flow when IV level was set to 100% and digester chip level was set to 0%.

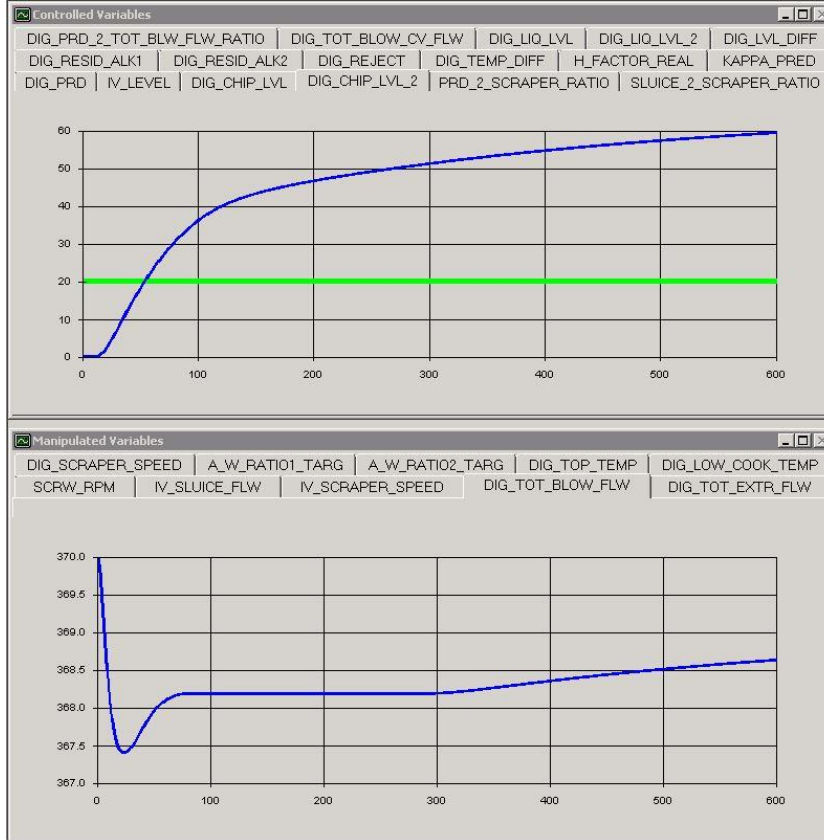


Figure 62: Simulation results for digester chip level 2 and digester total blow flow when IV level was set to 100% and digester chip level to 0%.

Simulation 8 shows similar operation from the controller as in simulation 7. When IV level is high, the controller decreases chip screw speed. Very low level of the digester is controlled by manipulation of the blow flow. IV sluice flow is increased to help regain the digester level. This time IV level range is reached in 225 minutes and digester chip level in 175 min.

7.2.9 Level simulation 9

Figure 63 shows the settings where IV level was simulated from initial value of 25% (low) and screw speed manipulation is offline. The predicted IV level change and IV sluice flow manipulations can be seen in Figure 64 and predicted digester chip level changes is presented in Figure 65.

The screenshot shows the 'Tuning' window with a table of controlled parameters. The 'IV_LEVEL' parameter is highlighted in red, indicating its current value of 45.00. A 'Simulation Time' dialog box is open, showing a simulation interval of 600 and a value of 9.15. The 'Run' button is visible, and the 'Done' button is highlighted.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	25.00	45.00	55.00	50.00	50.00	1.00	0.00	ON	1
3	DIG_CHIP_LVL	55.00	45.00	67.00	150.00	150.00	1.00	0.00	ON	1
4	DIG_CHIP_LVL_2	55.00	20.00	80.00	150.00	150.00	1.00	0.00	ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00	250.00	1.00	1.00	1.00	0.00	OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00						OFF	1
7	DIG_PRD_2_TOT_BLW_FLW_RATIO	2.70	2.60						ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00						ON	1
9	DIG_LIQ_LVL	60.00	55.00						ON	1
10	DIG_LIQ_LVL_2	60.00	20.00						ON	1
11	DIG_LVL_DIFF	0.00	0.00						OFF	1
12	DIG_RESID_ALK1	0.00	0.00						OFF	2
13	DIG_RESID_ALK2	7.00	6.00						ON	2
14	DIG_REJECT	0.00	0.00						OFF	2
15	DIG_TEMP_DIFF	0.00	0.00						OFF	3
16	H_FACTOR_REAL	0.00	0.00						OFF	3
17	KAPPA_PRED	88.00	87.50						ON	3

Figure 63: IV level set to 25% (low) and chip screw speed manipulation is offline for simulation.

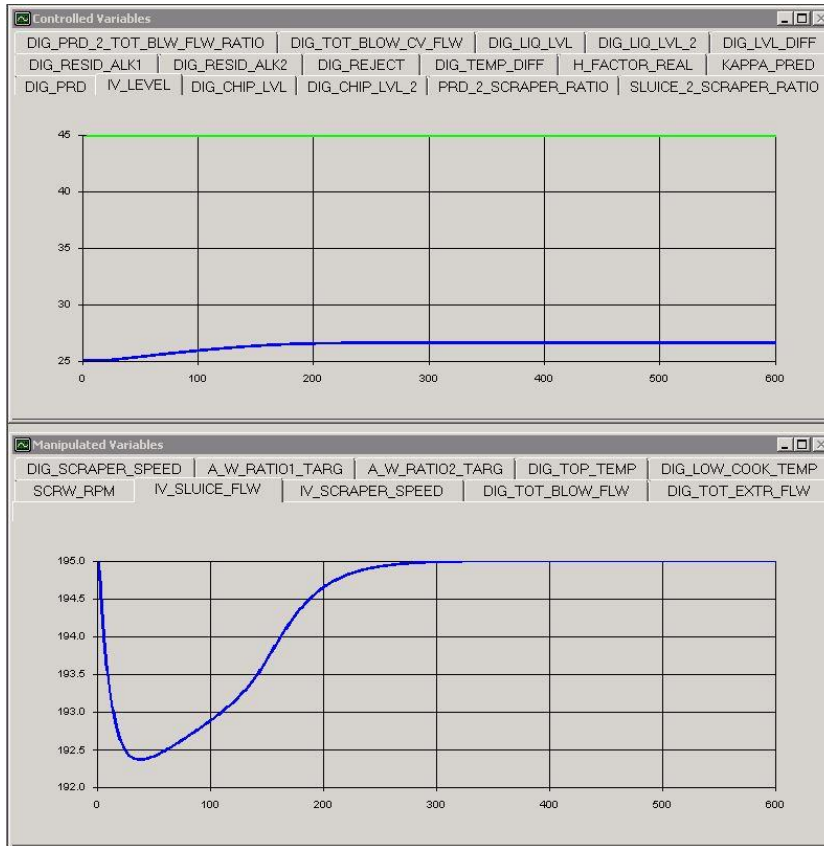


Figure 64: Simulation results for IV level and IV sluice flow when IV level was set to 25% and screw speed manipulation is offline.

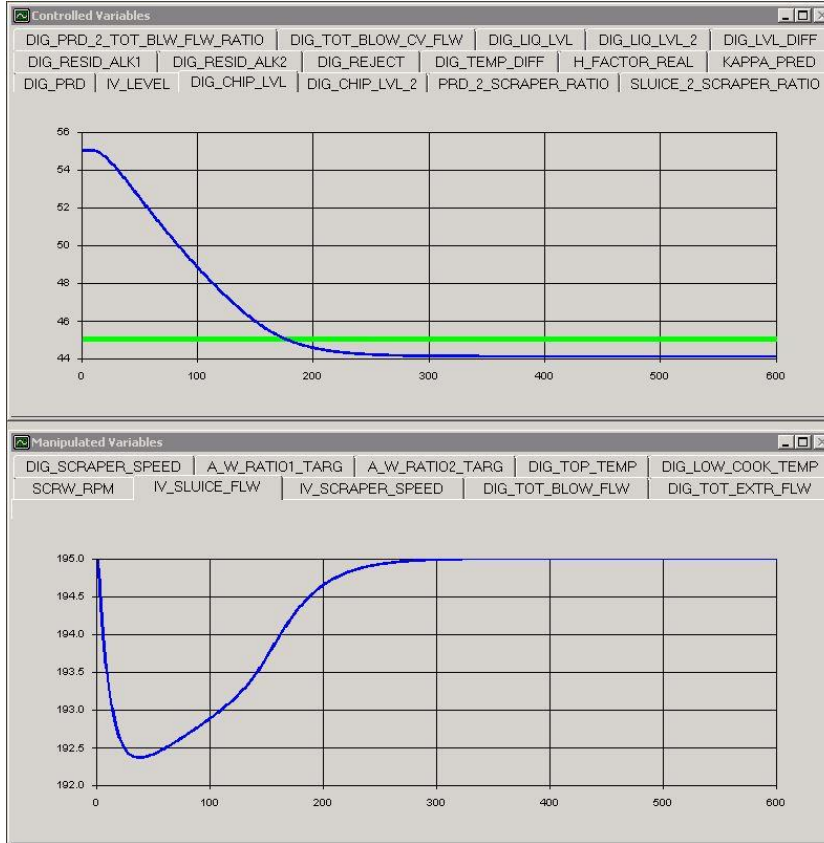


Figure 65: Simulation results for digester chip level and IV sluice flow when IV level was set to 25% and screw speed manipulation is offline.

The last level simulation shows that the process cannot be controlled correctly when the chip screw speed is not manipulated. IV level is only controlled by manipulating the IV sluice flow. Digester chip level drops below minimum limits and IV level does not reach the minimum limit 45 and stays very low at 26-27. If this situation is typical for the process, controller must be tuned to function properly.

7.3 Simulation of Alkali control

Alkali simulation cases consist of residual alkali abnormalities. Simulation starting points were set to low, very low, high, and very high compared to normal operating conditions. As the residual alkali level is controlled via manipulated A/W ratio target the results can be pointed out straightforward.

Normal operating value for residual alkali was set to 7.0. Simulated values were set to 5 for low starting point, 3 for very low, 9 for high, and 11 for very high. Simulation settings for low residual alkali can be seen in Figure 66, and simulation results are presented in Figure 67, Figure 68, Figure 69, and Figure 70.

The screenshot shows the 'Tuning' software interface with the 'Controlled' tab selected. A table lists 17 process variables with their current values, minimums, and maximums. A 'Simulation Time' dialog box is overlaid on the table, showing a simulation interval of 300 minutes. The 'Value' column in the table shows a value of 5.00 for DIG_RESID_ALK2, which is highlighted in red.

ID	Name	Value	Min	Max	Disturb	Disturb	Noise Filter	Disturb	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00						1
2	IV_LEVEL	50.00	45.00	55.00						1
3	DIG_CHIP_LVL	55.00	45.00	67.00						1
4	DIG_CHIP_LVL_2	55.00	20.00	80.00						1
5	PRD_2_SCRAPER_RATIO	200.00	150.00	250.00					F	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00	55.00					F	1
7	DIG_PRD_2_TOT_BLOW_FLW_RATIO	2.70	2.60	2.80						1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00	370.00						1
9	DIG_LIQ_LVL	60.00	55.00	65.00						1
10	DIG_LIQ_LVL_2	60.00	20.00	80.00						1
11	DIG_LVL_DIFF	0.00	0.00	0.00					F	1
12	DIG_RESID_ALK1	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	2
13	DIG_RESID_ALK2	5.00	6.00	8.00	1.00	1.00	1.00	0.00	ON	2
14	DIG_REJECT	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	2
15	DIG_TEMP_DIFF	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
16	H_FACTOR_REAL	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
17	KAPPA_PRED	88.00	87.50	88.50	20.00	20.00	1.00	0.00	ON	3

Figure 66: Simulation settings for low residual alkali conditions. Simulation time was set to 300min.

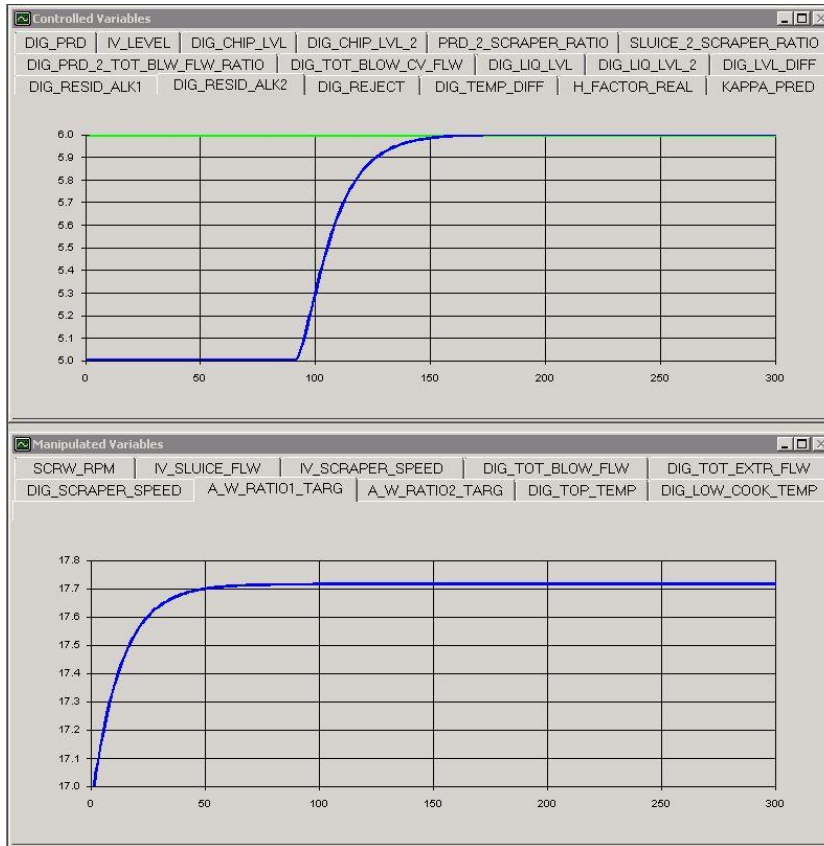


Figure 67: Simulation result for low residual alkali conditions.

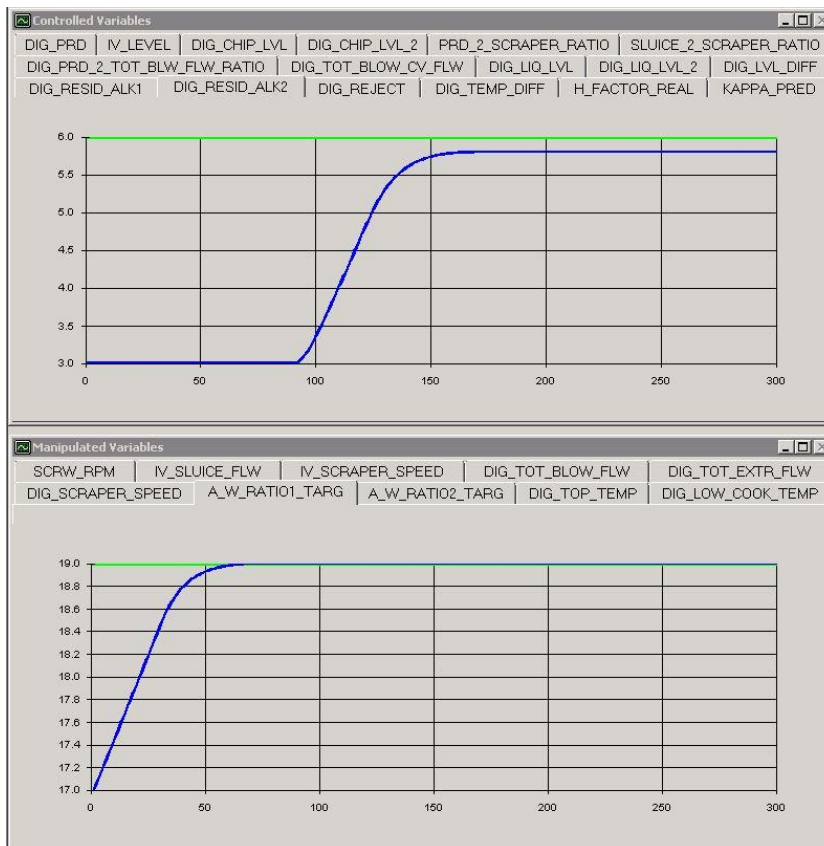


Figure 68: Simulation result for very low residual alkali conditions.

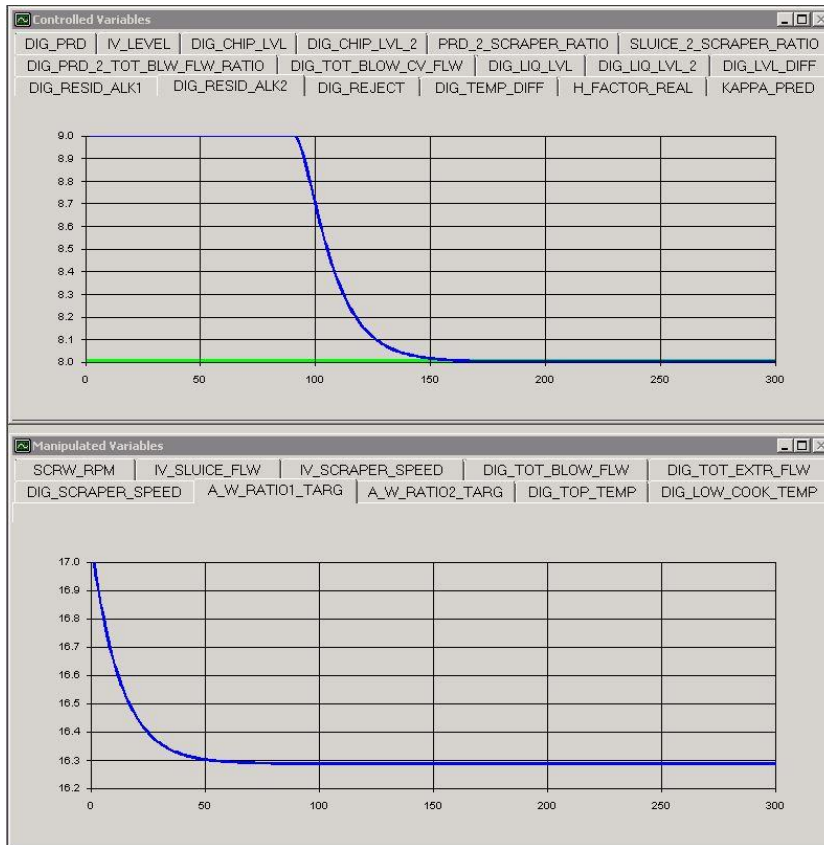


Figure 69: Simulation result for high residual alkali conditions.

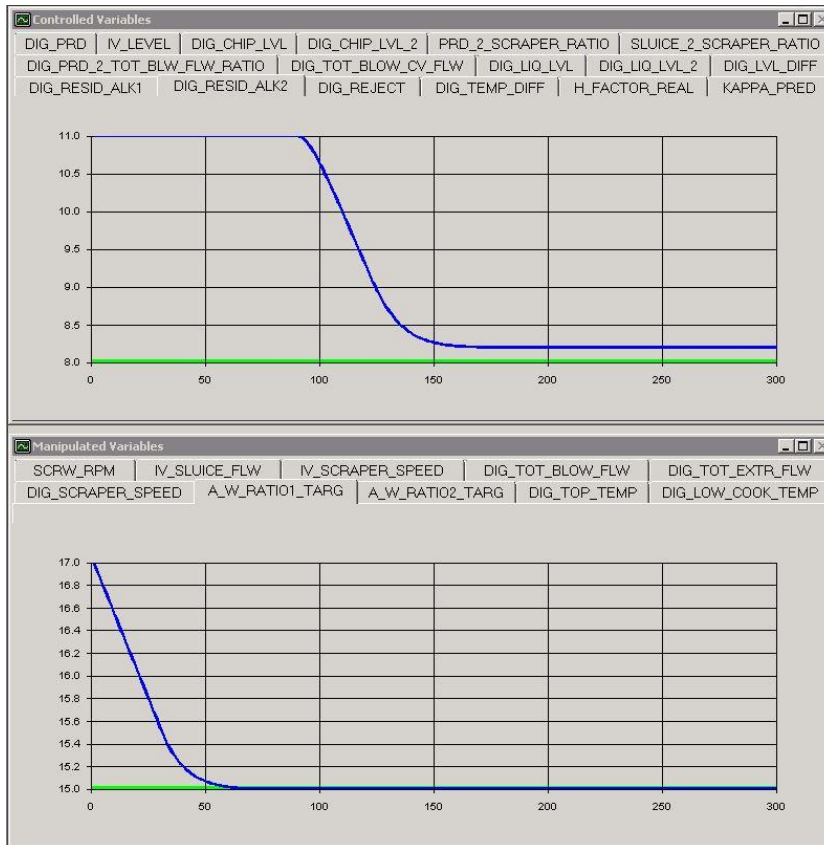


Figure 70: Simulation result for very high residual alkali conditions.

Simulation results show that the residual alkali of the digester is controlled by manipulation of the A/W ratio target. If the residual alkali value is below the operating range, A/W ratio target is increased. Respectively, when the value is high, A/W ratio target decreased. Manipulation effects can be seen in the residual alkali values after a delay of 90 minutes. If the residual value is very low (Figure 68) or very high (Figure 70), the A/W ratio target cannot be manipulated sufficiently to regain the normal operating range of the residual alkali. A/W ratio target range limits or weighting may need further adjustments.

7.4 Simulation of Kappa control

Kappa simulations consisted of low and high Kappa number starting values. Normal operation value for Kappa is 88 and low point was decided to be 85 and 91 for high. Kappa value is controlled via manipulated digester lower cooking temperature and digester top temperature.

Figure 71 shows the simulation settings for low Kappa starting value. Results for both simulations are presented in Figure 72, Figure 73, Figure 74, and Figure 75

The screenshot shows the 'Tuning' software interface. It features a table of controlled variables and a 'Simulation Time' dialog box.

ID	Name	Value	Min	Max	Min Wt	Max Wt	NoiseFilter	Rotation	On/Off	Grp
1	DIG_PRD	1,145.00	980.00	1,180.00	20.00	20.00	1.00	0.00	ON	1
2	IV_LEVEL	50.00	45.00						ON	1
3	DIG_CHIP_LVL	55.00	45.00						ON	1
4	DIG_CHIP_LVL_2	55.00	20.00						ON	1
5	PRD_2_SCRAPER_RATIO	200.00	150.00						OFF	1
6	SLUICE_2_SCRAPER_RATIO	37.00	25.00						OFF	1
7	DIG_PRD_2_TOT_BLW_FLW_RATIO	2.70	2.60						ON	1
8	DIG_TOT_BLOW_CV_FLW	370.00	300.00						ON	1
9	DIG_LIQ_LVL	60.00	55.00						ON	1
10	DIG_LIQ_LVL_2	60.00	20.00						ON	1
11	DIG_LVL_DIFF	0.00	0.00						OFF	1
12	DIG_RESID_ALK1	0.00	0.00						OFF	2
13	DIG_RESID_ALK2	7.00	6.00						ON	2
14	DIG_REJECT	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	2
15	DIG_TEMP_DIFF	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
16	H_FACTOR_REAL	0.00	0.00	0.00	0.00	0.00	1.00	0.00	OFF	3
17	KAPPA_PRED	85	87.50	88.50	20.00	20.00	1.00	0.00	ON	3

The 'Simulation Time' dialog box is open, showing 'Simulation Intervals' set to 600 and a value of 9.15. It includes 'Run' and 'Done' buttons.

Figure 71: Simulation settings for low Kappa starting point.

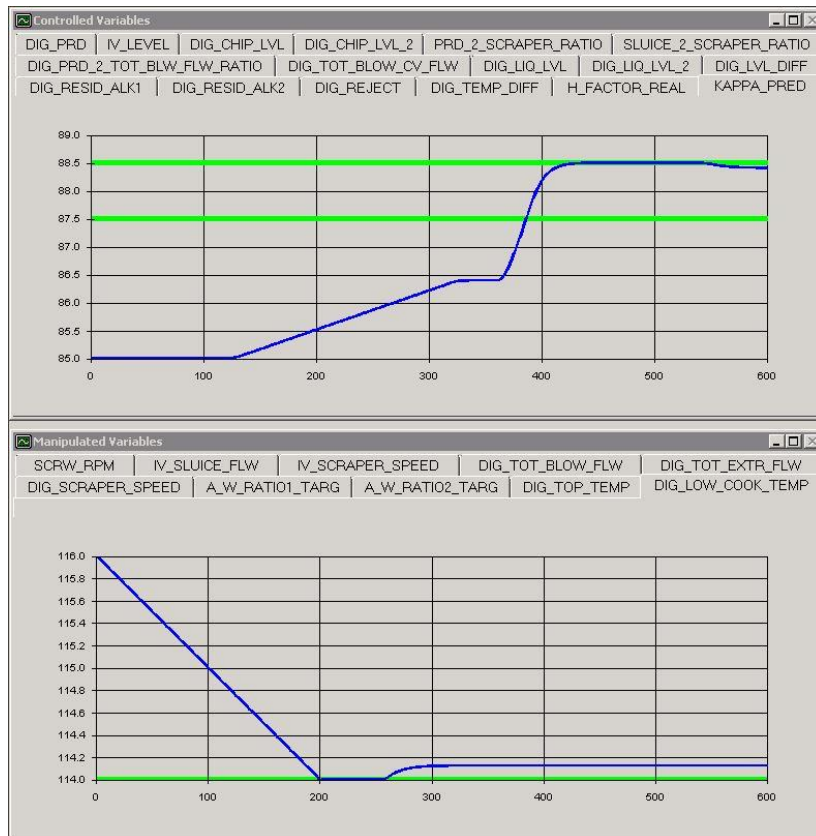


Figure 72: Simulation result for low Kappa conditions. Predicted Kappa and manipulated digester low cooking temperature.

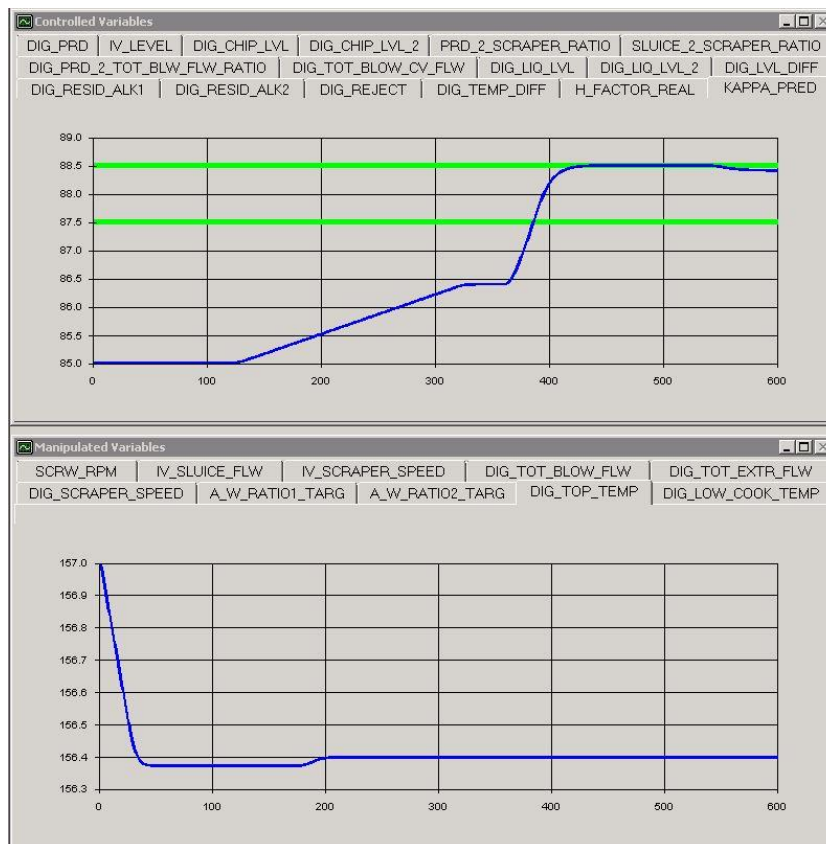


Figure 73: Simulation result for low Kappa conditions. Predicted Kappa and manipulated digester top temperature.

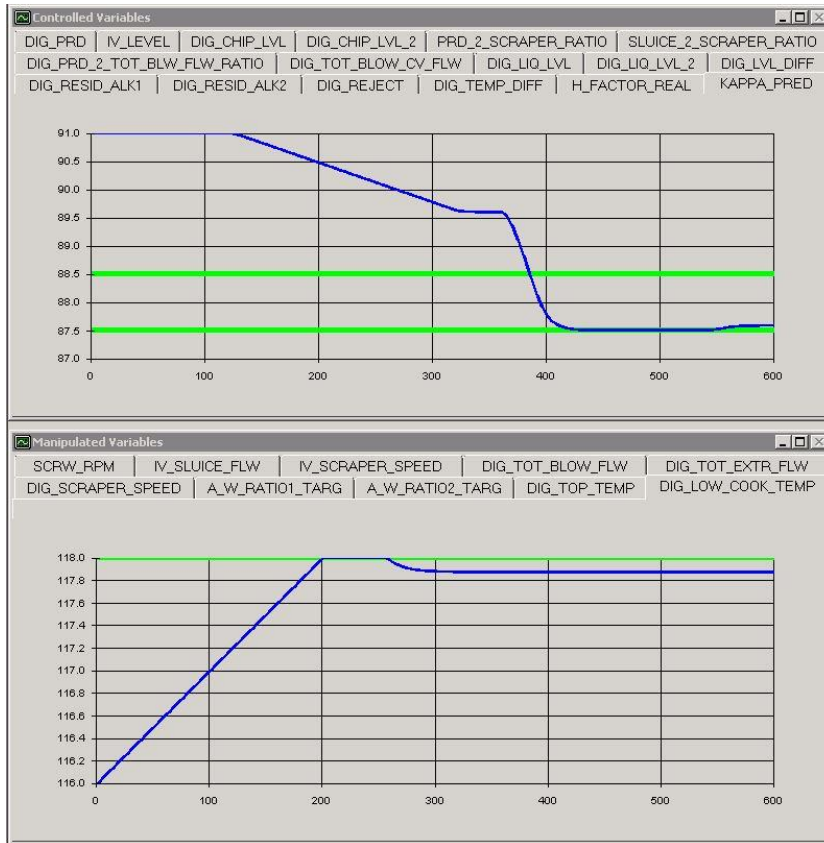


Figure 74: Simulation result for high Kappa conditions. Predicted Kappa and manipulated digester low cooking temperature.

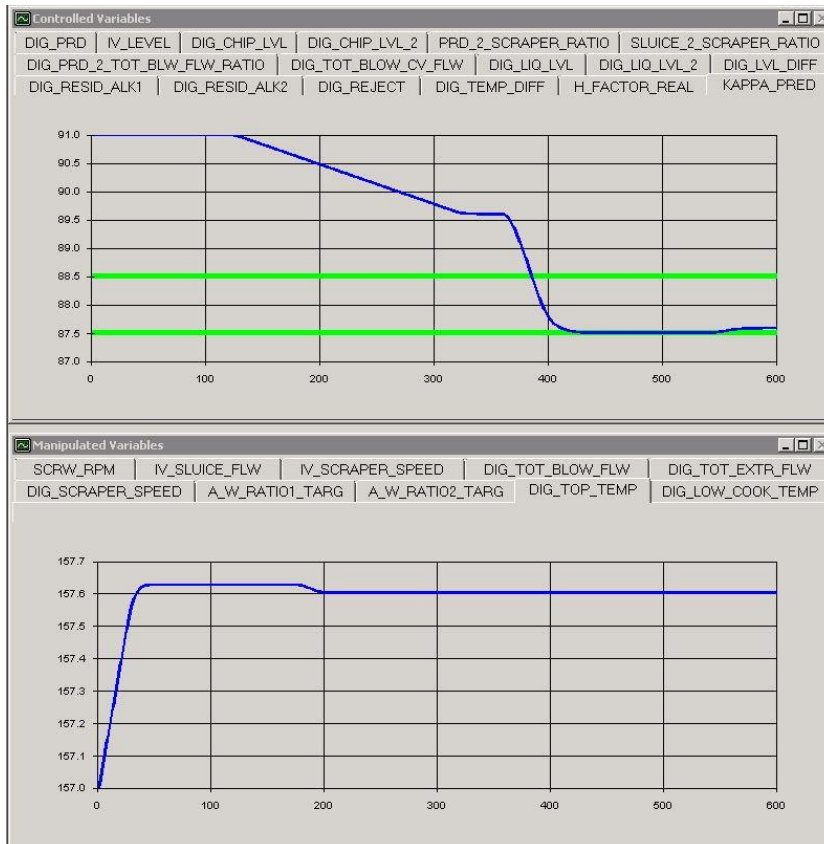


Figure 75: Simulation result for high Kappa conditions. Predicted Kappa and manipulated digester top temperature.

Kappa simulations indicate that the controller is functioning properly. When predicted Kappa is too low, digester low cooking temperature and top temperature are decreased. Respectively, when the predicted Kappa is too high, digester temperatures are increased. Controlling the Kappa is very slow. It takes up to 400 minutes to reach the normal operation conditions.

7.5 Discussion of simulation result

Simulation results show that the MPC provides mostly satisfying performance. The controller reacts to problems very well. Level simulations show that the controller is able to recover the normal levels in a variety of abnormalities. The level control without the chip screw speed control has to be reconsidered, as the level range was not achieved. Alkali simulations were simple and successful. However, extremely low or high residual alkali conditions were problematic. Kappa number control was successful. The overall conclusion is that MPC controller appears completely suitable in this control problem.

Obviously, simulations could have been more diverse, as proper process simulations are missing. However, the main problematic situations were tested with the software. Without adjustments made to the controller, these simulations should be sufficient for moving into mill trials. Thanks to simple structure of the controller, further modifications to the controller can be done in the mill testing phase.

8 CONCLUSIONS

The objective of this thesis was to implement model predictive control to the continuous cooking application. Cooking application is only one small part of pulping processes in which advanced control can be implemented. Complex and widely interactive processes can be controlled quite easily with these modern controllers. The ability to take process and quality constraints into consideration is a huge advantage compared to old control methods.

The MPC controller is as good as its model. Complex process needs in-depth system identification. The cooking application proved to be theoretically easy to model, although the process is highly interacting. However, the identification phase of this thesis took most of the time. Test response tests were done at the pilot mill and the model identification was done based on that knowledge. The effort was worthwhile, as the software simulations showed acceptable results.

The results from the simulations indicated that the controller is working as planned. Impregnation vessel level and digester chip level problems were mostly successfully controlled. In addition, residual alkali control and Kappa control were also simulated successfully. Some modifications could be done to the controller to get even better simulation results, but it is more important to move towards mill trials.

The controller is not ready for use in production only based on simulation results of this thesis. Actual process simulation was not available, so the next testing is done at the mill. Then, the true behavior of the controller is revealed. The controller will most likely need some modifications and fine tuning to work properly. The accuracy of the model will then improve significantly.

This thesis demonstrates how straightforward the implementation of a modern controller can be. The number of advanced controllers will increase in the future, as old technologies still widely exist and the benefits of modern controllers are noticed.

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