

TKK Reports in Forest Products Technology
Series A6
Espoo 2008

MODELING OF CHIP BED PACKING IN A CONTINUOUS KRAFT COOKING DIGESTER

Sampsa Laakso

Dissertation for the degree of Doctor of Science in Technology to be presented with due permission of the Faculty of Chemistry and Material Sciences, for public examination and debate in PUU 2 Auditorium at Helsinki University of Technology (Espoo, Finland) on the 12th of December, 2008, at 12 noon.

**Helsinki University of Technology
Faculty of Chemistry and Material Sciences
Department of Forest Products Technology**

**Teknillinen korkeakoulu
Kemian ja materiaalitieteiden tiedekunta
Puunjalostustekniikan laitos**

Distribution:

Helsinki University of Technology
Faculty of Chemistry and Materials Sciences
Department of Forest Products Technology
P.O. Box 6300
FIN-02015 TKK

ISBN 978-951-22-9677-4 (printed)

ISSN 1797-4496

ISBN 978-951-22-9678-1 (PDF)

ISSN 1797-5093

URL: <http://lib.tkk.fi/Diss/>

Yliopistopaino

Helsinki 2008



ABSTRACT OF DOCTORAL DISSERTATION		HELSINKI UNIVERSITY OF TECHNOLOGY P.O. BOX 1000, FI-02015 TKK http://www.tkk.fi	
Author Sampsa Laakso			
Name of the dissertation Modeling of chip bed packing in a continuous kraft cooking digester			
Manuscript submitted January 20, 2008		Manuscript revised October 2, 2008	
Date of the defence 12.12.2008			
<input checked="" type="checkbox"/> Monograph		<input type="checkbox"/> Article dissertation (summary + original articles)	
Department Forest Products Technology Laboratory Chemical Pulping and Environmental Technology Field of research Chemical Pulping Opponent(s) Professor Richard Gustafson, University of Washington, USA Supervisor Professor Olli Dahl Instructor Dr. Panu Tikka			
<p>Abstract</p> <p>This work focused on modeling of the chip bed packing phenomena in a continuous kraft cooking digester. A better understanding of chip bed packing would make it possible to optimize chip flow conditions in the digester, thereby ensuring uniform fiber quality and production efficiency.</p> <p>Chips are fed continuously into the digester, with the chip flow forming a solid bed. As the solid chip bed moves slowly downwards, cooking reactions proceed, which leads to softening of the chips. The softening of chips and the process conditions, flows and gravitational, are major variables affecting the chip bed packing. A new computer model, TKK Packing Simulator, was created to study the chip bed packing in a continuous digester. The simulator was verified against mill-scale conditions by conducting measurements in a continuous cooking plant.</p> <p>The packing simulator was found to give an accurate picture of the process flows inside the continuous digester. As digesters have grown bigger, controlling the liquor flows in the chip bed has become increasingly important. A tracer test was used to verify the liquor flow, which can be far from ideal in a modern digester.</p> <p>It is important to have optimal packing and flow conditions inside the digester, because too high packing and channeling of the liquor flow leads to non-uniform cooking, production disturbances and impaired fiber quality. The simulator developed in the present research can be used to determine optimal packing and flow conditions.</p>			
Keywords Kraft cooking, Continuous digester, Modeling, Chip bed, Packing, Softwood, Liquor flow, Chip flow			
ISBN (printed) 978-951-22-9677-4		ISSN (printed) 1797-4496	
ISBN (pdf) 978-951-22-9678-1		ISSN (pdf) 1797-5093	
Language english		Number of pages 142	
Publisher Helsinki University of Technology, Department of Forest Products Technology			
Print distribution Helsinki University of Technology, Department of Forest Products Technology			
<input checked="" type="checkbox"/> The dissertation can be read at http://lib.tkk.fi/Diss/			



VÄITÖSKIRJAN TIIVISTELMÄ		TEKNILLINEN KORKEAKOULU PL 1000, 02015 TKK http://www.tkk.fi	
Tekijä Sampsa Laakso			
Väitöskirjan nimi Modeling of chip bed packing in a continuous kraft cooking digester			
Käsikirjoituksen päivämäärä 20.1.2008		Korjatun käsikirjoituksen päivämäärä 2.10.2008	
Väitöstilaisuuden ajankohta 12.12.2008			
<input checked="" type="checkbox"/> Monografia		<input type="checkbox"/> Yhdistelmäväitöskirja (yhteenveto + erillisartikkelit)	
Osasto	Puunjalostustekniikan osasto		
Laboratorio	Selluloosa- ja ympäristötekniikan laboratorio		
Tutkimusala	Selluloosatekniikka		
Vastaväittäjä(t)	Professori Richard Gustafson, University of Washington, Yhdysvallat		
Työn valvoja	Professori Olli Dahl		
Työn ohjaaja	Tekniikan tohtori Panu Tikka		
<p>Tiivistelmä</p> <p>Työn tavoitteena oli selvittää hakepedin pakkausilmiöitä jatkuvatoimisessa sulfaattikeitossa. Parempi ymmärrys hakepedin pakkautumisesta mahdollistaisi optimaalisten virtausolosuhteiden ylläpitämisen keittimessä. Tämä puolestaan takaisi tasaisen massan laadun ja korkean tuotantotehokkuuden massanvalmistuksessa.</p> <p>Jatkuva tasainen hakevirta virtaa keittimeen. Hakevirta muodostaa keittimeen hakepilarin eli hakepedin. Kun tämä yhtenäinen hakepilari liikkuu hitaasti alaspäin, keittoreaktiot edistyvät hakepedissä. Tämä johtaa hakepalasten pehmenemiseen. Hakepedin pakkautumiseen vaikuttavat hakkeiden pehmeneminen, keittimen prosessiolosuhteet sekä hakepilariin kohdistuvat voimat. Uusi tietokonemalli, TKK Pakkaussimulaattori, kehitettiin tutkimaan pakkausilmiötä jatkuvatoimisessa keittimessä. Simulaattorin tuloksia verrattiin tehdasmittakaavassa suoritettuihin mittauksiin.</p> <p>Pakkausmallin havaittiin antavan tarkan kuvan virtauksista jatkuvatoimisessa keittimessä. Virtauksien hallinta on noussut yhä tärkeämmäksi tekijäksi jatkuvatoimisten keittimien ohjauksessa, koska keittimien koko on kasvanut merkittävästi.</p> <p>On tärkeää pitää yllä optimaalisia pakkaus- ja virtausolosuhteita keittimessä, koska liian suuri hakepedin pakkautuminen ja virtauksen kanavoituminen johtaa keiton epätasaisuuteen, tuotantohäiriöihin ja massan laadun heikkenemiseen. Tässä työssä kehitettyä simulaattoria voidaan käyttää optimaalisten keitto-olosuhteiden selvittämiseen jatkuvatoimisessa keitossa.</p>			
Asiasanat Sulfaattikeitto, Jatkuvatoiminen keitto, Mallinnus, Hakepeti, Pakkautuminen, Havupuu, Nestevirtaus			
ISBN (painettu)	978-951-22-9677-4	ISSN (painettu)	1797-4496
ISBN (pdf)	978-951-22-9678-1	ISSN (pdf)	1797-5093
Kieli	englanti	Sivumäärä	142
Julkaisija Teknillinen Korkeakoulu, Puunjalostustekniikan laitos			
Painetun väitöskirjan jakelu Teknillinen Korkeakoulu, Puunjalostustekniikan laitos			
<input checked="" type="checkbox"/> Luettavissa verkossa osoitteessa http://lib.tkk.fi/Diss/			

Acknowledgements

I wish to express my deepest gratitude to all those who have helped me through this project. I would like to thank Docent Panu Tikka for his supervision and for being the driving force throughout the whole process. I would like to acknowledge Eric Enqvist for his support during the course of the project and tireless guidance during the finalization of the thesis.

I would like to thank Metsä-Botnia for giving me the possibility to work on the challenging thesis project and for funding the work. I thank Jukka Rantamäki and Juha Fiskari for their support and instructions during the project. Many thanks to Erkka Tahvanainen and Matti Toivonen for giving all the time and assistance needed during the process. Also I would like to thank Kari Kovasin, Esa Hassinen, Ismo Reilama, Risto Joronen and rest of the staff of Botnia for support and patience to see the project finalized. Special thanks to Lasse Hernesniemi for giving deeper knowledge about continuous kraft cooking.

Many warm thanks for the past and present colleagues in what was the Laboratory of Chemical Pulp and Environmental Science at TKK, who have given me inspiration, time and support during the project. Very special thanks to Marjo Määttä and Ville Laakso for the support and help during the project.

I wish to thank all of my family and friends; your encouragement has been most valuable. Finally, I wish to express my deepest gratitude to my dear Mia and children Oona and Konsta for endless love and understanding.

Otaniemi November, 2008

Sampsa Laakso

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LIST OF SYMBOLS AND NOTATIONS

Abbreviations and main symbols and notations used in the text:

AAS	Atomic Absorption Spectrophotometer
A_d	digester cross sectional area (m^2)
ADt	air dry metric tons
A_1	pre-exponential factor in Arrhenius equation
A_s	coefficient of pressure drop
B	reactant
BL	black liquor
B_s	coefficient of pressure drop
C	reactant
C_c	carbohydrate content (% on wood)
CFD	computational fluid dynamics
CSTR	continuous stirred tank reactor
D	diffusivity
$D_{[OH]}$	diffusion parameter (cm^2/min)
$D_{Digester}$	diameter of the digester
D_d	dispersion coefficient
dP	pressure drop trough packed bed (Pa)
E_a	activation energy of the reaction (J/mol)
$E(t)$	frequency function
EMCC	extended modified cooking
F	force
Filt.	filtrate from brown stock washing
F_p	reaction product
F_μ	friction force
HP	high pressure
HS^-	hydro sulfide ion
$[HS^-]$	sulfide concentration (mol/l)
ITC	iso thermal cooking
K	concentration (mol/l)
L_b	length of the packed bed (m)
L'	effective length of the chip (mm)
L_c	real length of the chip (mm)
L	lignin content (% on wood)
LD	chip and liquor level difference, m
Li	initial lignin content (% on wood)
LW	liquor-to-wood ratio (l/kg)
M	friction co-efficient
MCC	modified continuous cooking
OH^-	hydroxide-ion
$[OH^-]$	hydroxide concentration (mol/l)
P1, P2	pressure inside the digester (kPa)
PD	pressure difference (kPa)
P_{ext}	pressure at extraction line (kPa)
PDE	partial differential equation

PFR	plug flow reactor
R	universal gas constant. $R = 8.314 \text{ J/(K}\cdot\text{mol)}$
R_b	balls radius (mm)
R_k	chemical reaction rate change by the alkaline consumption
$R_{[OH]}$	reaction rate of the hydroxide
RTD	residence time distribution
R_1	coefficient of pressure drop correlation ($\text{Pa}\cdot\text{s/m}^2$)
R_2	coefficient of pressure drop correlation ($\text{Pa}\cdot\text{s/m}^2$)
T	the absolute temperature (K)
T_c	thickness of the chip (mm)
TKK	Teknillinen korkeakoulu
U	superficial velocity flow across the column (m/s)
V	volume (m^3)
V_c	chips free liquor volume
V_l	volume of the free liquor
V_{Total}	total volume of the cooking block
W_C	width of the chip (mm)
WL	white liquor

b	order of the reaction
c	order of the reaction
d_c	cylinder diameter
d_p	equivalent diameter of particles (m)
g	gravity coefficient (m/s^2)
k	rate constant for reaction
k'	pseudo rate constant for reaction
k_m	the mass transfer coefficient at the chip surface
k_0	empirical constant
k_1	empirical constant
k_2	empirical constant
k_3	empirical constant
l_c	cylinder length
l_c	element length
p_c	compacting pressure applied to chip bed (Pa)
p_l	liquor pressure (Pa)
p_{wall}	bed pressure near the wall
r	radial place in the ball(mm)
t	time (s)
u_i	superficial velocity of liquor (m/s)
v	rate of the reaction
v_l	volumetric flow of liquor (m^3/s)
w_j	weight fraction of the chips

Greek letters

ε	void volume inside the chip
ε_c	area taken by the chip in a cooking block
ε_l	area taken by the free liquor in a cooking block
μ_v	dynamic viscosity (Pa*s)
μ	friction coefficient between chips and digester wall
ρ_{wood}	density of solid wood material (kg/m ³)
ρ_{dc}	density of wood material (kg/m ³)
ρ_c	density of the chips (kg/m ³)
ρ_l	density of the cooking liquor (kg/m ³)
ρ_{bulk}	bulk density of the wood chips (kg/m ³)

1. Introduction

1.1. *General introduction*

Continuous cooking is the dominant method for kraft pulp production. Since the breakthrough of the continuous cooking process in the early 1960's, world pulp production has increased threefold. Over the years, the design of the continuous digester has been changed to meet the need for increased production and new kraft cooking modifications. This has led to growth in the size of continuous digesters. As digesters have grown bigger, controlling the flows in the chip bed has become more important in kraft pulp cooking.

Although continuous cooking is the dominant process for producing kraft pulp, our general knowledge and understanding about process conditions inside the digester is at a very low level. Process control systems include groups of measurements, which give information about conditions at the outer shell of the digester. These measurements mostly tell about cooking liquor inflows and outflows by flow rate and temperature. It is possible to measure pressure or pressure difference at screen section of the digester shell. Process conditions inside the digester are based on measurements at the shell of the digester and intelligent guesses, which are table data about chip residence times at different parts of the digester at different production rates. In reality, residence times along the digester are dependent on production rates and packing degree of the chip bed, which is not calculated by any means.

1.2. *Objectives*

Essential question for operating and controlling the continuous digester is to know the packing degree of the chip column in the digester. The main objective of this work was to develop a model, which describes 1) digester chip columns packing degree in vertical direction, 2) flows in chip bed and 3) delignification and lignin profile along the digester. New methods were used for accomplishing these aims, by combining porous bed flow and compaction model with wood chip delignification reaction-diffusion model. New approach was important, because both phenomenon influence on each other. Delignification of the chip bed changes compression characteristics and compression of the chip bed changes the residence times, which affects through reaction kinetics into delignification degree.

1.3. Starting hypothesis

The study is based on the following hypotheses:

1. A chip bed packing model can solve the true chip residence time and process conditions in the chip bed
2. Process conditions and mass transfer affect the delignification of the chip.
3. The chip bed packing degree is affected by chip softening caused by delignification
4. Packing can be modeled by a combined model using a reaction-diffusion model for delignification of chip and an Ergun equation for chip bed packing

1.4. Research methods

The research was based on simulations, which were executed with a new simulator created in this work. The idea of creating the new simulator was to combine two different models, namely a diffusion-kinetic based delignification model and a hydrodynamic porous bed model. The simulator was used for studying the packing of the continuous digester chip bed. The basic idea for this investigation was to determine how different kraft cooking parameters affect the runnability of the digester and how the digester production efficiency could be sustained without disturbances.

Some tracer tests were also performed for studying both liquor and chip flow in the digester. Liquor flow was followed by tracking the tracer liquor flow in digester at liquor outputs and chip flow was studied with monitoring the marked chip flow along the digester.

1.5. Outline of the thesis

The literature part of the thesis focuses on the historical development of continuous kraft cooking and developments in modeling and simulating the kraft cooking process. The outline of the thesis is shown in *Figure 1*. The history of pulping is discussed in *Chapter 2*. The development of the wood pulping process, with a focus on continuous kraft cooking, is examined in *Chapter 3*. *Chapter 4* reviews modeling in general. Cooking plants modeling at different process scales, from molecular to mill scale, are presented in *Chapters 5-8*. The chip bed packing model is presented in *Section 7.1*. Different kraft cooking kinetic models are compared in *Section 8.2*. Mass transfer and kinetic models are discussed in more detail in *Section 8.3*.

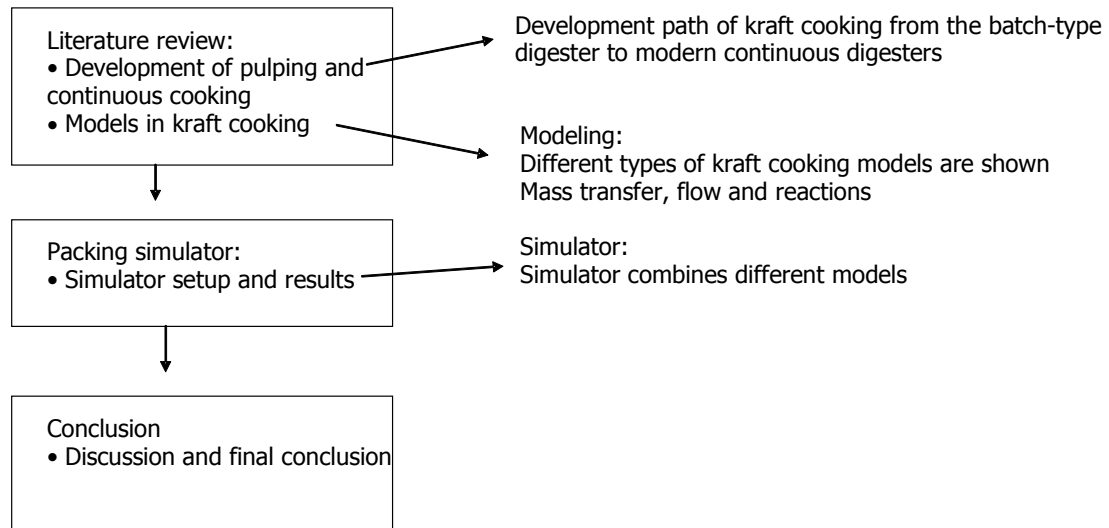


Figure 1 Structure of the thesis.

The new scientific contribution of the thesis is presented from *Chapter 10* onwards. *Chapter 10* presents the *TKK Packing Simulator*. Packing simulation results are discussed in *Chapter 11*. Simulation results and process measurements are compared in *Chapter 12*. An overall conclusion from the research is given in *Chapter 13*. Aspects of applications are given in *Chapter 14*.

2. Review of the evolution of early pulping digesters

The chapter is a brief review of digesters used from early batch reactors to present continuous digesters. The idea is to give the reader an idea of how digester evolution has advanced stepwise by solving one problem at a time. Most of the steps were not scientific, but were based on solving a runnability problem of the digester. Therefore, digester development has been evolutionary process based on solving practical problems rather than based on actual science.

Pulping of wood has gone through major technological changes from its invention in the mid 19th century to modern times. The industry itself has changed from using acidic sulfite cooking liquors to using alkaline sulfate cooking liquor. The sulfate process is also known as the kraft cooking process (cooking chemicals are NaOH and Na₂S), as it will be called in this work from now on. The kraft cooking process competed with the more popular sulfite process for a long time. However, as a result of the superior quality of kraft pulp and new market demand for better-quality fiber, the kraft cooking process gradually became the dominating process for wood pulp production. In addition, some major engineering problems were solved, allowing the kraft process to use a broader range of wood species.

In the following, the evolution of kraft pulp digesters is presented. The first digesters were of horizontal direct heated design, and did not become a big success. Because of the non-uniform heating of the digester and concentration differences inside the digester, the horizontal digesters produced poor-quality pulp [1]. The early digesters had no liquor circulation. Furthermore, the batch-type digester was difficult to empty. An example of a horizontal digester is shown in *Figure 2*.

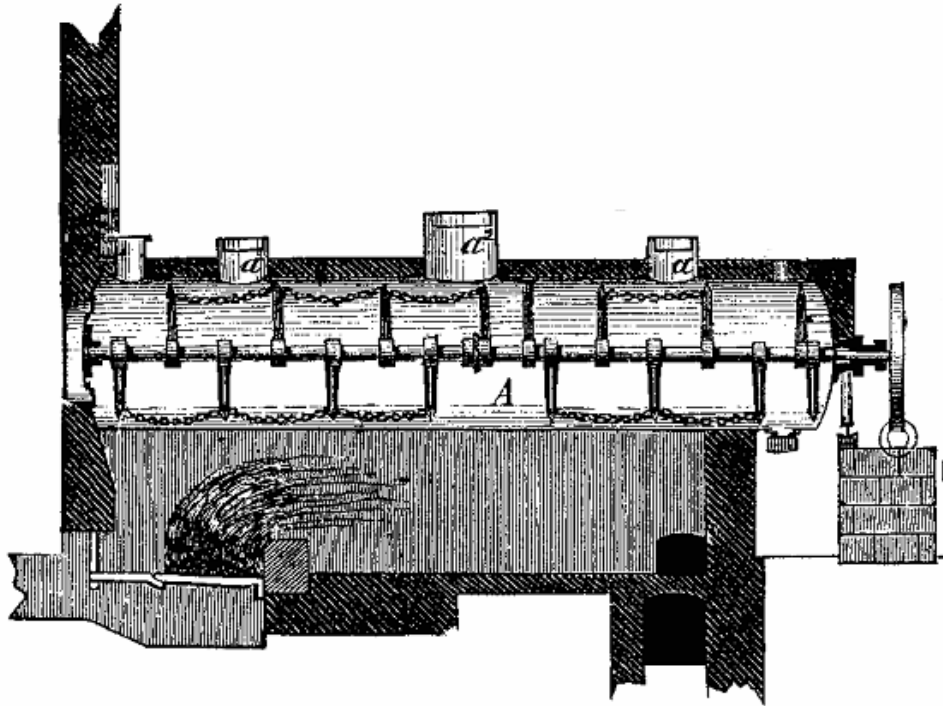


Figure 2 Direct heated horizontal digester for pulp production /1/.

Due to drawbacks of horizontal digester, soon vertical digesters replaced the horizontal digesters. The vertical digester was easier to empty. However, these vertical digesters were also directly heated with fire. Fire caused explosions, which damaged the digester shell. An example of a fire-heated vertical digester is shown in *Figure 3*.

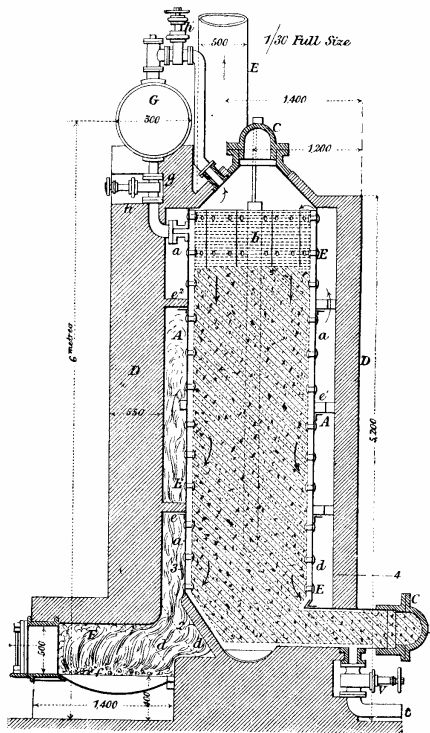


Figure 3 Fire-heated vertical digester /1/.

Due to the risk of explosions, fire heating was gradually replaced by steam heating. Steam heating was either direct or indirect /2/. In direct heating, the steam is introduced directly into the digester. In indirect heating, the cooking liquor is circulated through a steam-heated heat exchanger. Indirect heating was much slower than direct heating. Indirect heating produced better pulp quality because the temperature rise was slower and cooking temperature low. In addition, the cooking chemical concentration and chemical charge were lower than in direct heating.

At the same time with the vertical digester the rotating digester was introduced for pulping of wood (*Figure 4*). Rotating digesters, which had been earlier used for producing straw pulp, were common in kraft pulp production for some years. Demand for kraft pulp was small, so rotating digesters could meet the markets' need for kraft pulp. The maximum size of rotating digesters was about 40-50 m³, compared with 200-300 m³ for digesters with indirect heating /2/.

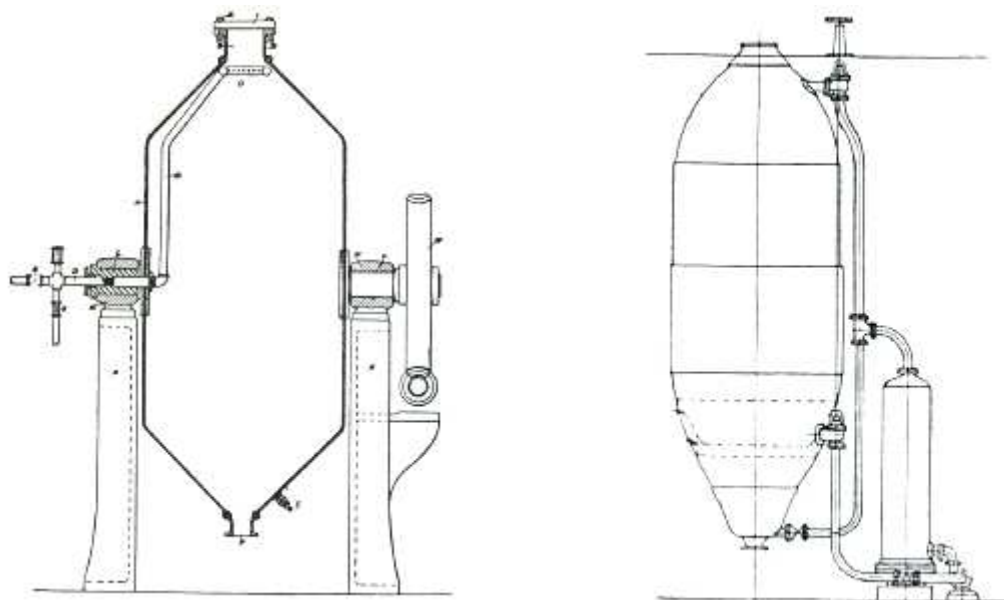


Figure 4 Rotating digester (left) and vertical digester with indirect heating /3/.

The rotating digester produced better-quality pulp than the vertical digester. It consumed less cooking liquor than the vertical digester, because the cooking liquor and wood chips were better mixed during the cook. However, there were many downsides to the rotating digester and it vanished from the industry in the early 20th century. For example, increasing the digester size was very expensive. The working procedure of the cook was complex and the steam piping suffered from malfunctions. Also, the rotating digester took a lot of space for a given production.

The development of multistage bleaching resulted in the development of continuously operating bleaching towers during the 1930's. Bleaching was further developed through the introduction of different bleaching chemicals,

such as chlorine dioxide and peroxide. At the same time, cooking chemicals recovery developed, which also increased the interest in the kraft cooking process. Chemicals recovery also made the kraft process competitive for making fully bleached pulps.

The advances in bleaching and chemical recovery made kraft pulping a cost competitive method for production of high strength pulp. Demand increased rapidly, leading to the need for higher capacity digesters. This demand led to the development of continuous digesters.

3. Continuous kraft cooking

Just as batch cooking, described in the previous chapter, continuous kraft cooking has evolved through a number of major incremental changes. This chapter gives detailed information about the development of the dominant production method, Kamyr continuous cooking.

Again, the idea is to highlight the evolutionary nature of the changes. This evolution through trial and error has again helped in building a complete picture of what happens inside the digester.

3.1. History of continuous cooking

The history of continuous kraft cooking dates back to the late 1920's and the early 1930's. There were early pioneers who were trying to design a continuous cooking process [4,5,6], but they failed to build a complete continuous mill-scale cooking process. Major obstacles were how to feed chips continuously into a pressurized vessel and how to discharge the chips from the vessel.

Major development of the continuous kraft cooking process has been conducted by only one company, Kamyr. Kamyr's name originates from two companies: AB Karlstads Mekaniska Werkstad of Sweden and Myrens Verkstad of Norway. These two companies merged in the 1920's. Finnish A. Ahlström Oy became a partner to this company in 1933. This new company, combining Finnish, Norwegian and Swedish interests, became known as Kamyr.

Eventually, the company's process setup and supremacy over patents concerning chip feeding and digester discharge made Kamyr the dominant continuous pulping method. This thesis is limited to discussing the continuous kraft cooking process developed by Kamyr. Impco, Bauer, Sunds Defibrator, Esco, Pandia and Rauma tried to develop a continuous digester, but their efforts were not successful, with the exception of special applications [7].

Kamyr established a North America-based affiliate company Kamyr Inc. in 1953. The company specialized in producing continuous digesters for the North American market. Kamyr AB continued its digester production in Sweden, Norway and Finland.

The two Kamyr companies differed in terms of technology. Kamyr AB was selling steam/liquor-phase digesters and Kamyr Inc. was delivering hydraulic digesters. In 1989, one of the three founding partners sold its shares to the other two companies [8]. This arrangement led to a split in the companies with Kamyr AB remaining based in Karlstad, Sweden and Kamyr Inc. being owned by Ahlström. After the split, Kamyr AB was renamed Kvaerner Pulping, whereas Kamyr Inc. started to use the name Ahlström Machinery. After their

renaming in 1993, the two companies started to compete with each other worldwide with both hydraulic and steam/liquor-phase digesters.

The mergers between pulp and paper machine suppliers continued. Soon after the Kamyr split, Ahlström Machinery was acquired by Andritz (2001) and Kvaerner Pulping became part of Metso (2006).

The first continuous digesters were started up in the early 1950's. It took almost 15 years for them to gain general acceptance. The main obstacle was the digester bottom and bottom conditions. In the early days, the digester was emptied at the process temperature. These conditions were too harsh for producing high-quality, strong pulp. Bottom cooling was invented and later Hi-heat washing, where the bottom part of the digester was washed with the washing liquor from brown stock washing. Hi-Heat washing strengthened Kamyr's position in the market. Hi-Heat washing was an industry standard for over 40 years. A new trend is that Hi-Heat washing is not included in digesters.

3.2. Development of continuous digester's production capacity

Production capacity and economy of scale have been the key driving forces in the capital-intensive pulp industry. This can be seen in *Figure 5*, which shows the nominal capacity of new continuous kraft digesters.

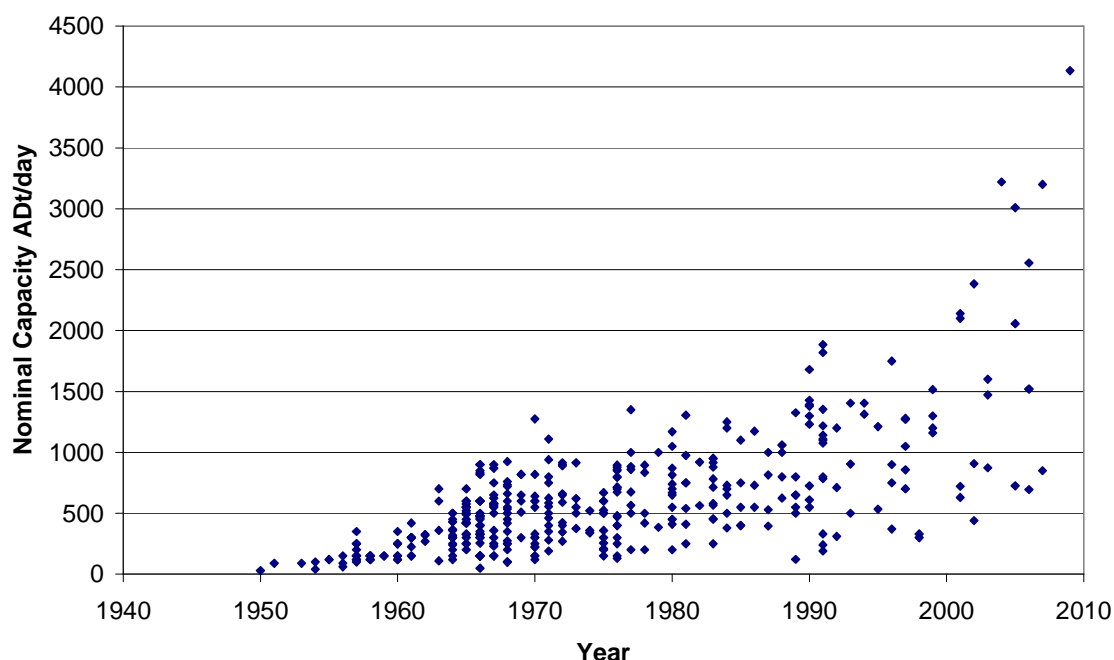


Figure 5 Nominal capacity of continuous kraft pulp digesters /9/.

The maximum capacity of new digesters follows a trend where capacity is doubled every 15 to 20 years. This is shown in *Table 1*.

Table 1 Maximum capacity of new continuous digesters in selected years.

Year	Production rate, ADt/day (approximately maximum)
<i>1961</i>	500
<i>1966</i>	1000
<i>1991</i>	2000
<i>2009</i>	4000

Table 1 and *Figure 5* show how the nominal capacity of continuous digesters has increased over the years. The digester production rate grew slowly until 1990 when more production was put through by increasing the digester's diameter. Within ten years, a production of 2 000 ADt/day became possible. The same development has continued, and the biggest digester so far, which will start up in 2009, will be producing over 4000 tons of pulp a day /10/.

In the early years of continuous cooking, new digester capacity was started up slowly. After 1965 the continuous digester made a commercial breakthrough (*Figure 6*). Average capacity has increased fast: in the 1960's 30 digesters were needed to produce 14 000 ADt/day, in the 1990's only 15 digesters. At present, five new digesters are able to produce the same amount as 20 digesters in the 1960's. This is illustrated by the fact that the new continuous digesters started up in 2005 worldwide had a nominal capacity equaling half of the total pulp production capacity in Finland /11/.

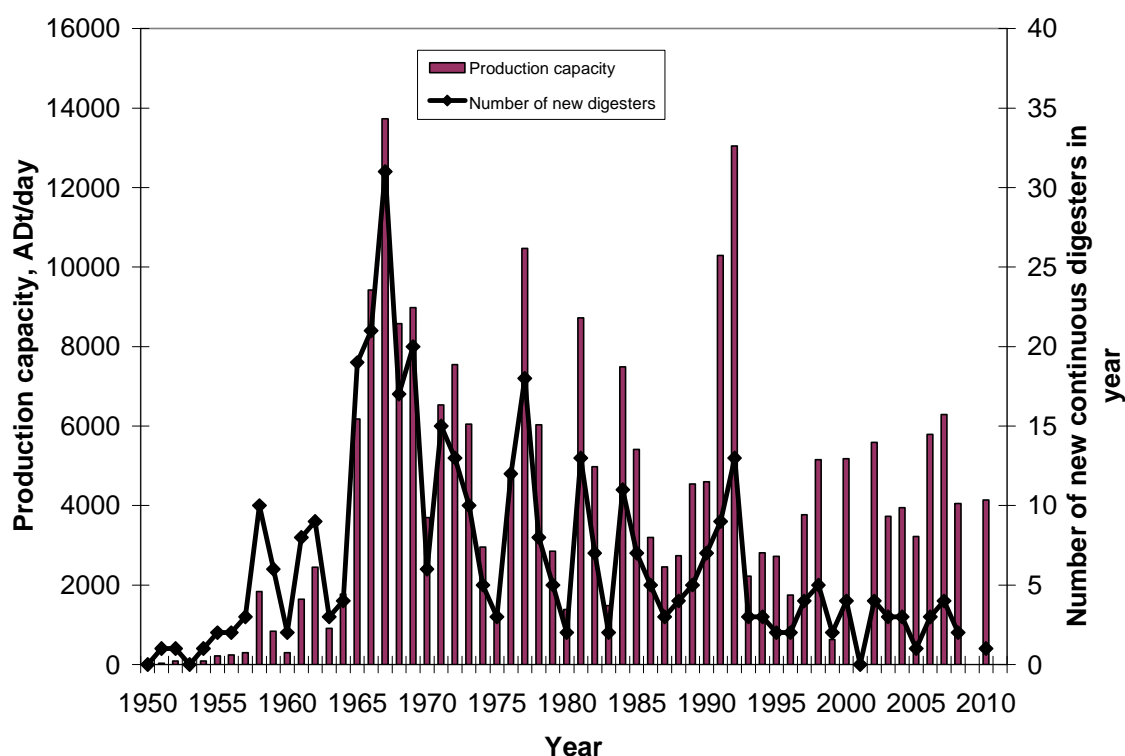


Figure 6 Total production capacity (ADt/day) of all new continuous digesters and number of new digester started up per year (1950-2007) /9/.

For the old digesters, the most important aspects of digester operation were runnability and steady movement of the high chip column. Currently, new digesters are facing different operational challenges. Uniformity and quality issues have become more and more important for modern digesters where different kinds of raw materials are put through the reactors.

3.3. Development of the Kamyr continuous digester

The idea for the continuous-cooking Kamyr digester was developed in 1938. The concept behind the continuous digester came from the continuous bleaching plant. Since bleaching could be done continuously, it should be possible to do the cooking in the same way /13/.

It took over ten years to solve mechanical and other problems involved in building a mill-scale digester /12/. The first commercial continuous digester unit started with a production of 30 ADt/day at Fengersfors in 1950 /13/. The digester was shut down after 20 years' operation, because it was too small to be economical. Figure 7 shows one of the earliest Kamyr continuous digesters, including a simplified presentation of it.

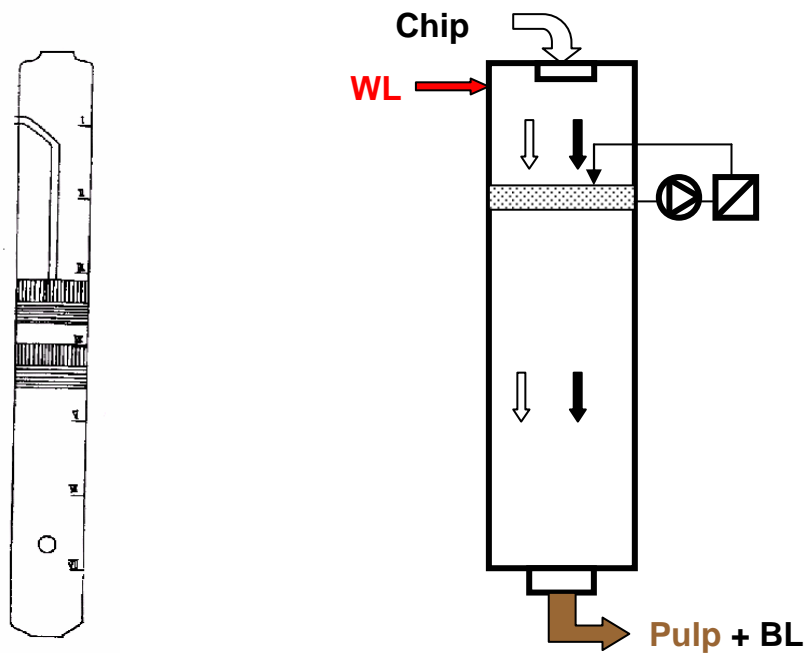


Figure 7 Typical Kamyr digester in the 1950's /13/ and a simplified presentation of the flows. Open arrow: wood chips flow direction, black arrow: liquor flow direction.

Chips and cooking liquor were fed into the digester top and there was a separate heating circulation to heat up the digester. The total volume of the digester was about 100-200 m³ /13/.

The key to continuous digester operation is to be able to feed the digester at high pressure. Kamyr solved this by constructing a HP (High Pressure) feeder. The basic design of an early digester is seen in *Figure 8*. The HP feeder is marked number 5 in the figure. HP-feeder is rotating pocket feeder, which feeds chips and liquor into digester.

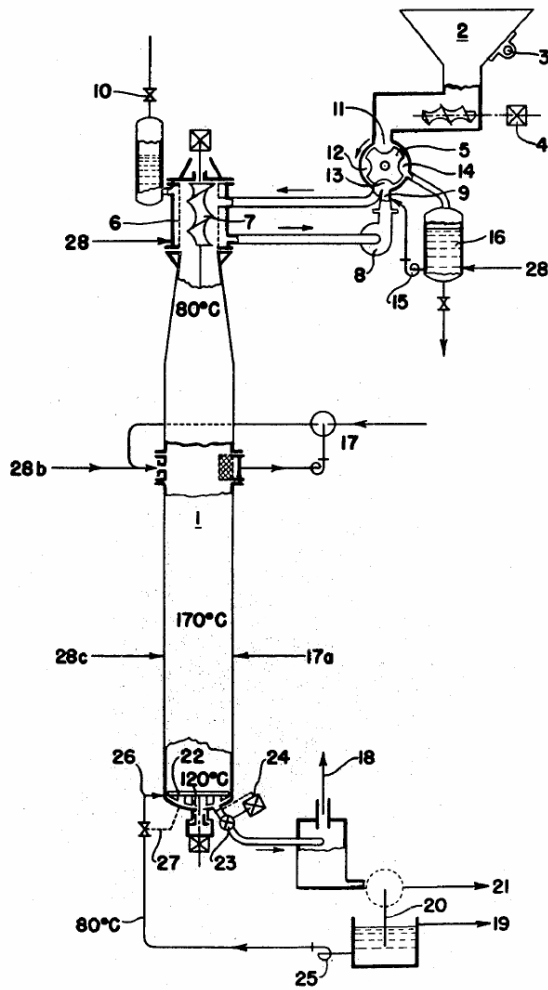


Figure 8 Hydraulic continuous digester with HP feeder /14/.

After a few digesters had been delivered it was noticed that the pulp quality did not fulfill expectations. There was a lack of fiber strength. One reason for this was that the pulp was discharged through the blow valve at too high temperature /13/. It was found that if the cooked chips were cooled to below 100 °C before blowing, the pulp quality was improved. The cooling can be accomplished by feeding cold washing liquor into the lower part of the digester. The idea of the cold blow was presented in 1957 (Figure 9) /13/.

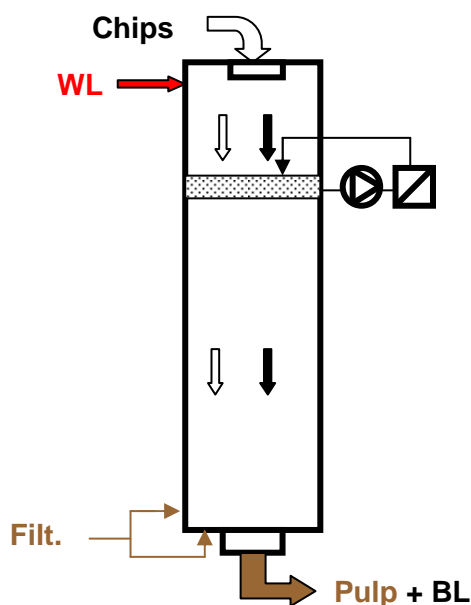


Figure 9 Cold blow schematic picture. Open arrow: wood chip flow direction, black arrow: liquor flow direction.

In 1960, a soda digester (cooking chemical NaOH), originally delivered to Australia in 1956, was equipped with counter-current displacement cooking and washing [15]. A slight counter-current flow was expected to help the column past the screens to compensate for the local increase in wall friction, thus reducing the possibility of hanging of the chip column. The new system was called Hi-Heat washing (Figure 10). This new cooking method became a standard in Kamyr continuous cooking systems in 1962. Hi-Heat improved brown stock washing and made it easier to run the digester. The Hi-Heat washing system is discussed in more detail in Section 3.5.

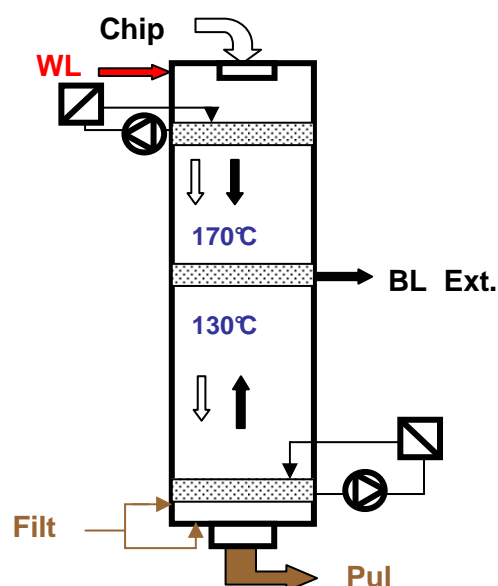


Figure 10 Hi-Heat washing system in continuous digester. Open arrow: wood chip flow direction, black arrow: liquor flow direction.

The first digesters were completely filled with liquor. This digester type was called a hydraulic digester (*Figure 12*). Chips and liquor were heated indirectly in heat exchangers. Later, a digester with a steam phase section was developed in order to be able to heat the chips using direct steam /16/. This digester type became known as the steam/liquor phase digester.

The digester size began to increase to accommodate greater production capacities. Liquor circulations increased along with increased production. It became more difficult to heat the increased liquor circulations. In addition, the chip bed loading increased as the digester height was increased. In response to all these needs, a separate impregnation vessel was invented in the 1960's /17/. The first two-vessel (impregnator and digester) steam/liquor phase cooking plant was delivered in the early 1970's. The first hydraulic two-vessel digester was delivered in 1978 (*Figure 13*). Today, these hydraulic and steam/liquor phase digesters are referred to as conventional continuous digesters.

The diameter of early continuous digesters was about 2-3 meters and the height about 20-30 meters. Soon, digester length was extended to increase the production. In the early 1970's the digester chip bed could be over 70 meters high. At this point, chip bed height was almost at the maximum. The production rate could not be increased much further because of the chip packing. As a result, some of these long digesters have been shortened. This is discussed in more detail in *Section 3.5*.

Same time as the digester became shorter, digester diameter has increased. This trend started in the 1990's. Chip bed height has settled at about 40 to 50 meters and the diameter of a new digester can be up to 12.5 meters /9/. Digester growth is illustrated in *Figure 11*.

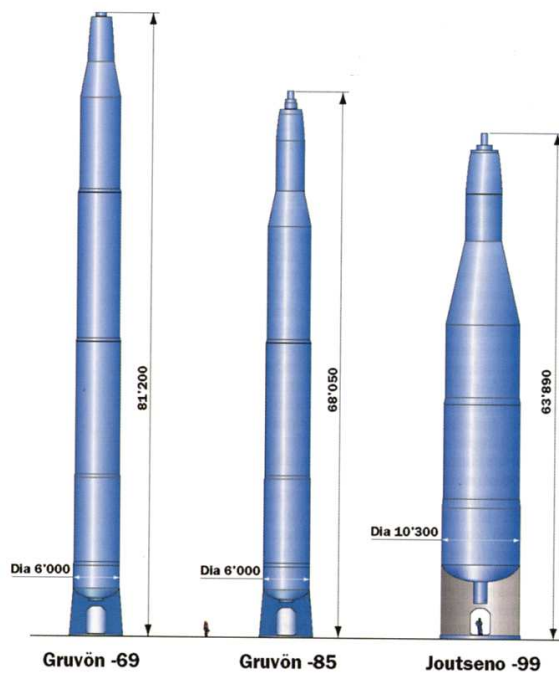


Figure 11 Gruvön's digester before and after rebuild and Joutseno's new digester /18/.

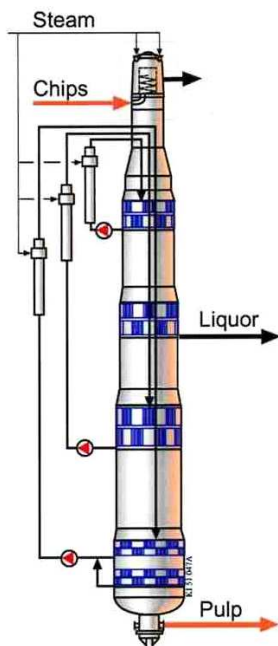


Figure 12 Single-vessel hydraulic digester /19/.

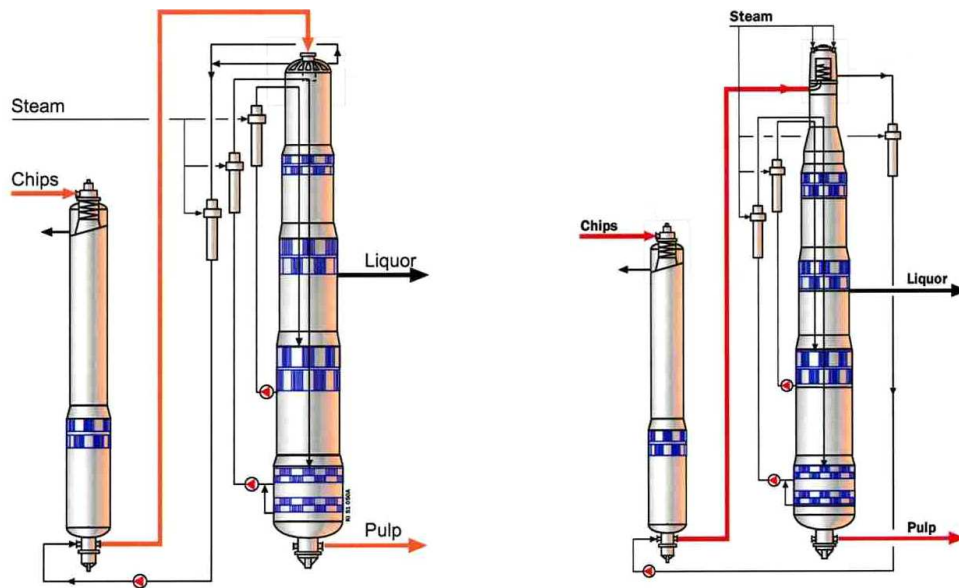


Figure 13 Two-vessel hydraulic and two vessel steam/liquor-phase digesters /19/.

Hydraulic and steam/liquor-phase digester tops are shown in *Figure 14*. Chips enter the digester trough the top separator. This unit has a cylindrical screen surrounding a screw conveyor. The slotted screen allows most of the liquor to be separated and reused in transfer circulation. Chips flow downwards in the hydraulic digester and upwards in the steam/liquor-phase digester top.

In steam/liquor-phase digesters chips overflow from the top of the screw conveyor. The chips form a chip column, which is above the liquor level. Direct steam is added to the top of the vessel in the vapor space. The chip and liquor level difference is controlled to affect the chip retention time in the steam phase.

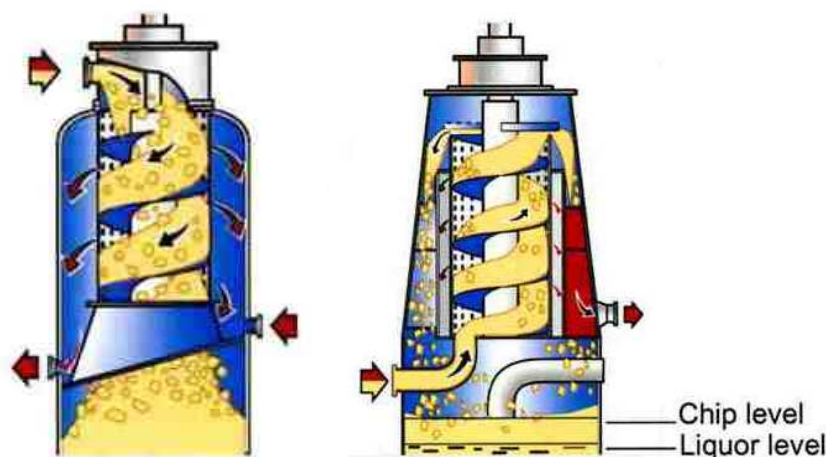


Figure 14 Hydraulic (left) and steam/liquor-phase (right) digester top /19/.

3.4. Development of cooking modifications after 1989

The development of the Kamyr continuous cooking process and product names is presented in *Figure 15*. The first major brand name of the Kamyr continuous digester was the MCC (Modified Continuous Cooking) digester. This name was used both in Europe and North America. After the company split in 1989, process development led to different digester set-ups. Ahlström's digesters were named EMCC (Extended Modified Continuous Cooking) and Lo-Solids. Kvaerner's cooking systems were called ITC (IsoThermal Cooking) and CC (Compact Cooking).

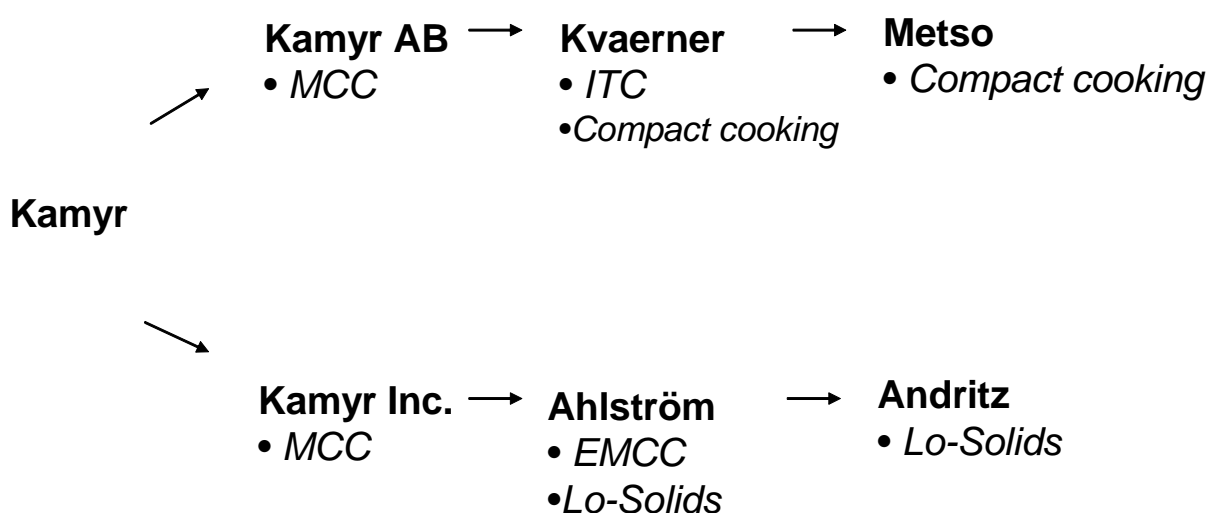


Figure 15 Development of the Kamyr continuous cooking process and different continuous kraft digester product names.

The market shares of different cooking processes are shown in Table 2. These figures refer to the situation at the end of 2006. Old digesters have been retrofitted with different cooking modifications whereas the number of new digesters sold has been more limited. This is also shown in *Figure 6*.

Table 2 Sales of modified continuous cooking systems reported by the machine vendors /20/.

Product (current manufacturer)	New	Retrofit	Year of first installation
MCC (Andritz)	13	26	1985
MCC (Metso)	17	0	
EMCC (Andritz)	0	5	1991
ITC (Metso)	12	13	1992
Lo-Solids (Andritz)	20	53	1993
CC (Metso)	15	7	1997
TOTAL	75	112	

Most of the continuous digesters sold at present are either Lo-Solids or Compact Cooking systems. The cooking chemical profile is uniform in the Lo-Solids digesters. Compact Cooking is based on black liquor impregnation, where part of the cooking liquor is extracted from the digester and re-circulated to the impregnation vessel. Most of the Lo-Solids and Compact Cooking digesters have been changed to run using downflow cooking. In downflow cooking the cooking liquor is flowing downwards in the bottom part of the digester, as opposed to the flow in Hi-Heat washing.

The common standard for Compact Cooking is at present a two-vessel system with a steam/liquor phase digester. The digester usually has two screen sections, where liquor is extracted. Currently, the standard for Lo-Solids cooking is a single-vessel hydraulic digester. A Lo-Solids digester has three to four screen sections. These comparisons refer to new digesters, because retrofits of digesters are always unique. For retrofits, the system design depends on the assembly of the digester and how many screens there are and whether there is an impregnation vessel or not.

3.5. Development of washing section of the continuous kraft cooking digester

The development of the Kamyr continuous digester has been a stepwise evolution. As an example, the development of the washing section is presented below.

The bottom washing area has been developed in stages, as there have been different problems to solve at different times. In the beginning, there was a need to ensure the pulp quality. Later on, the need was to have sufficient washing, which led to the invention of new washing techniques. During the 1980's and 1990's environmental concerns and the introduction of new bleaching chemicals led to lower kappa numbers. This was achieved by raising the washing section temperature. Currently, the bottom part of continuous digesters is commonly changed to downflow mode. Cooking liquor and chips are flowing in the same direction between the two lowest screen sections. This ensures stable chip column movement, while improving steam economy and counteracting channeling [21].

Initially, the digesters were designed so that pulp was discharged from the digester at cooking temperature, which was about 170 °C. Soon, mechanical action at high temperatures was found to impair pulp quality [22], for example strength properties. At this point, bottom cooling was introduced to cool down the discharge area to 30-40 °C below the cooking temperature. If the cooling liquor is introduced into the bottom of the digester, a new screen is needed to extract part of the liquor above the digester bottom. This ensures stable chip column movement in the digester. The bottom cooling system is presented in *Figure 16*.

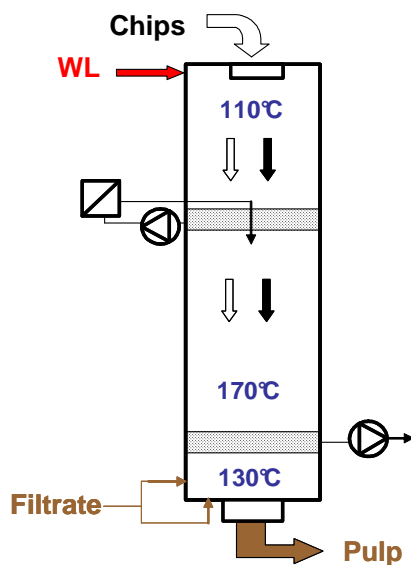


Figure 16 Bottom cooling of the continuous digester by washing filtrate. Open arrow: wood chip flow direction, black arrow: liquor flow direction.

Bottom cooling was the first step towards the digester washing. Soon after the invention of bottom cooling, Hi-Heat washing was introduced. This kind of digester set-up is shown in *Figure 10*. Hi-Heat washing made it possible to wash cooking chemicals from the wood chips counter-currently. This washing method reduced the need for brown stock washers [23]. This was one of the reasons for the peak in the sales of continuous Kamyr digesters in the 1960's (*Figure 6*). For mills thinking of investing in extra capacity, the options in the 1960's were to buy additional batch digesters or a single continuous digester. Additional batch digesters required more space and in the worst case the mill's brown stock washing system was not capable of handling the increased production. In this case, the mill would have to buy extra washing capacity while at the same time investing in extra production capacity. In this situation, a continuous digester equipped with good washing became an excellent option for pulping wood.

The sales peak in the 1960's cannot be fully explained by the superiority of the Hi-Heat washing system. Part of the peak is explained by the change in the market. Old sulfite mills were converting from sulfite to kraft pulping because of the need to improve strength properties, greater demand for fully bleached pulp and increased use of chlorine dioxide as a bleaching agent. The simplicity, economy and ease of maintenance of continuous cooking also contributed to making it a competitive option.

The Hi-Heat system proved to ensure excellent washing in its early applications. As mill sizes grew, digester capacity was increased by building taller digesters. While the old digesters were 20 to 30 meters high, the new digesters installed in the early 1970's were 70 to 80 meters high. For instance, Gruvön's digester (*Figure 11*) was over 80 meters high. This means that the chip bed inside the digester is about 65 to 70 meters high. This

increase in chip bed height led to problems caused by chip bed packing. The problems were seen as poor runnability and digester washing. Therefore, digester developers started looking for solutions that would reduce the chip bed packing and increase the continuous digester's production without sacrificing washing results. The options were to increase the digester diameter or to build a separate vessel for impregnation.

The impregnation vessel for continuous cooking was developed in the late 1960's. By using a separate impregnation vessel, the height of the digester could be reduced, thus decreasing the chip bed packing. For this reason, also the length of the Hi-Heat washing zone could be maintained at about 30 meters, as shown in *Figure 17*. A new trend in the late 1980's was to increase Hi-Heat-washing temperature to cooking temperature, which instead was reduced. In this way, the height of the Hi-Heat-washing section was decreased to about 10 meters. The importance of washing in new digesters decreased, as the brown stock washing system was improved and environmental factors increasingly favored cooking to lower kappa numbers. This is one reason for the temperature rise in Hi-Heat washing. As a result of better screen technology and the transformation of the washing section into the cooking section, the washing section was shortened further, as shown in *Figure 17*. For these reasons, new digesters are operating without Hi-Heat washing. This is the latest trend in continuous kraft cooking, which is called downflow cooking or co-current cooking, depending on whether it is an Andritz or Metso design. The term downflow cooking is used in the following.

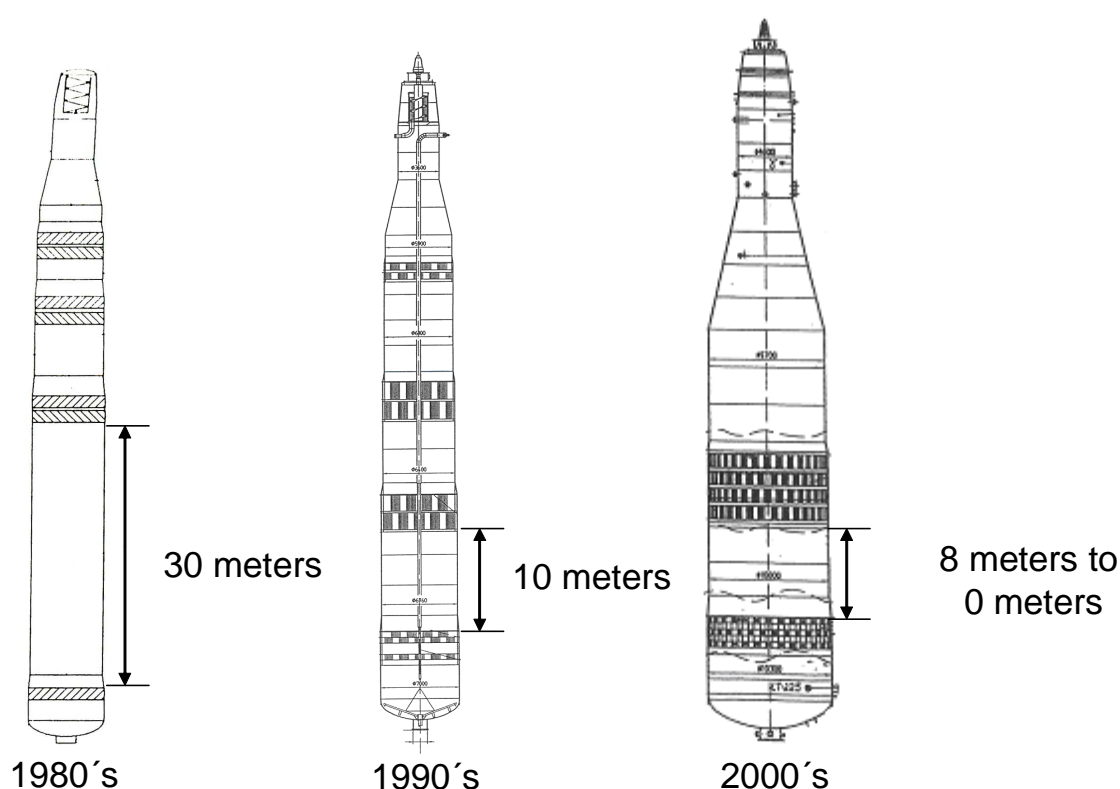
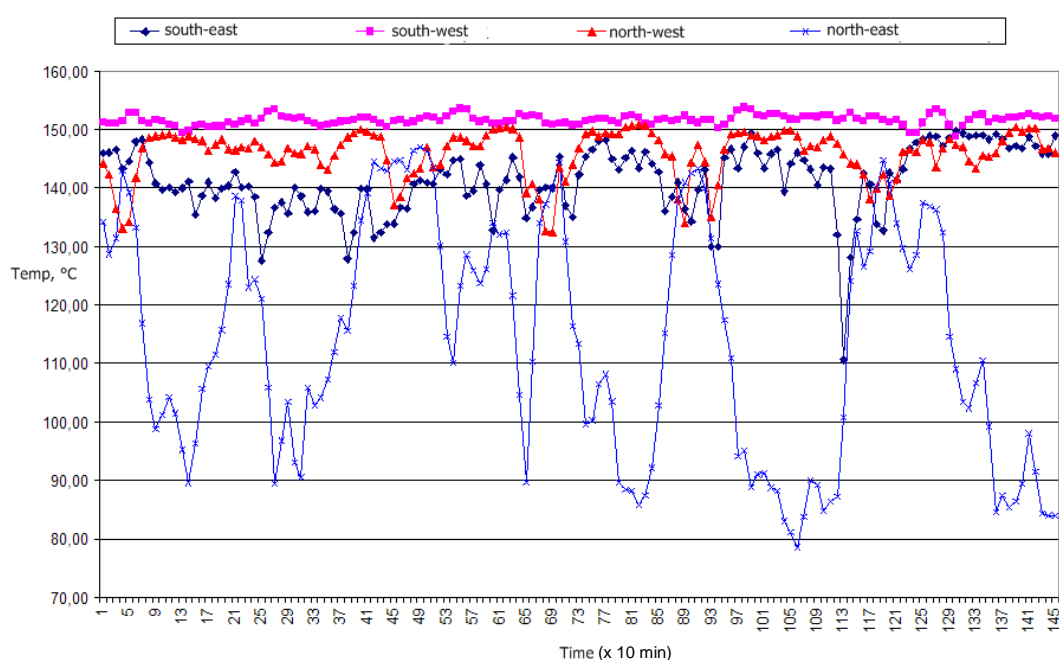


Figure 17 Hi-Heat washing zone length in different digesters.

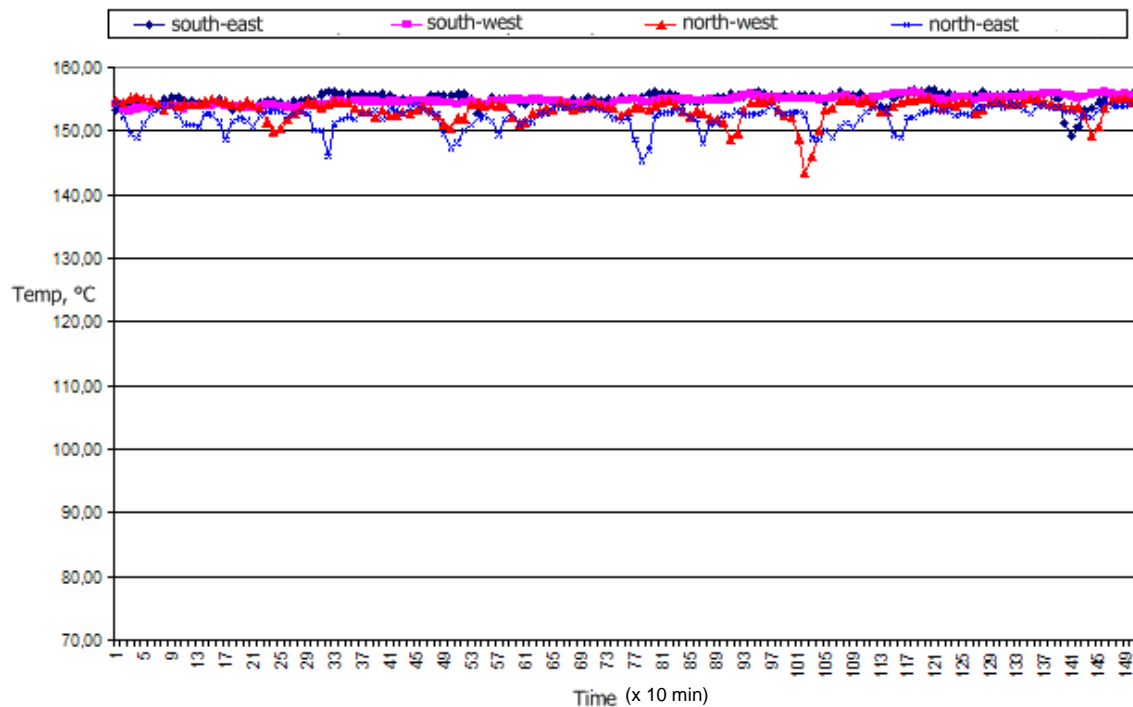
Downflow cooking occurs in the bottom part of the digester. The direction of the liquor in the Hi-Heat section is changed to move in the same direction as the chips are flowing. This is considered to stabilize the chip column movement /24/.

Over the years, it has been noticed that digesters with Hi-Heat washing have non-uniform flow inside the digester. Haas /25/ found in his research that the cooking liquor flow is subject to channeling in the bottom part of the digester.

The channeling in the bottom part is also seen in *Figure 18 and Figure 19*, which show washing screen temperature data before and after changing the digester to downflow cooking /26/. The flow in the bottom was monitored by temperature measurement. *Figure 18* shows huge temperature variations on one side of the screen. The temperature difference between the cool side and hot side is at worst over 60 °C. This temperature difference is caused by channeling of the bottom washing liquor. When the digester flow was changed to downflow cooking (*Figure 19*), screen temperatures evened out and channeling of the washing liquor was greatly reduced. Counter-current washing of the bottom of the digester is also said to increase the reject content /27/.



*Figure 18 Washing screen temperature on different sides of the digester in Hi-Heat washing (time*10 min) /26/.*



*Figure 19 Washing screen temperature on different sides of the digester in downflow cooking (time*10 min) /26/.*

The importance of the washing section has decreased. Now it is more important to increase the digester's production rate than to have good digester washing. One way to increase the throughput is to shorten the digester Hi-Heat washing zone, resulting in a smaller chip bed packing.

For example Kvaerner shortened the digester at the Steti mill in the Czech Republic. *Figure 20* shows the digester before and after the shortening. The washing effect of the digester was low before the shortening, because the consistency at the bottom of the digester was too low. The digester was shortened from 68 meters to 53 meters. Shortening of the digester reduced the pressure on the chips at the bottom and improved the washing /21/. At the same time, the mill was able to increase the digester's total production.

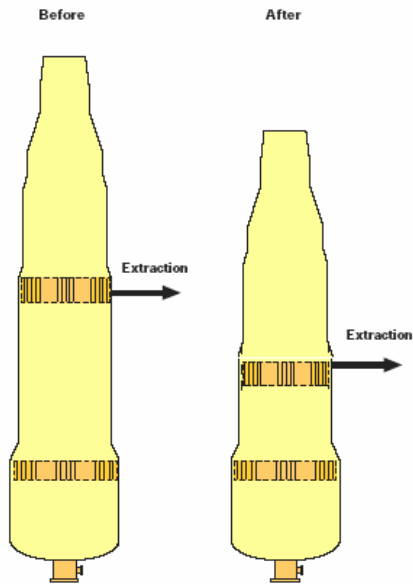


Figure 20 Digester vessel of Steti mill before and after shortening /21/.

Displacement washing technology has also been developed, making displacement washing more efficient and shortening the displacement height needed to achieve efficient liquor displacement. A new type of displacement washing technique is presented in *Figure 21 and Figure 22*. Wash liquor enters the chip column through the downward running central pipe and the upward pointing pipe from the discharge scraper. The liquor performs displacement radially through the digester. In this way the effective displacement length is reduced from 20 - 30 meters to 5 meters.

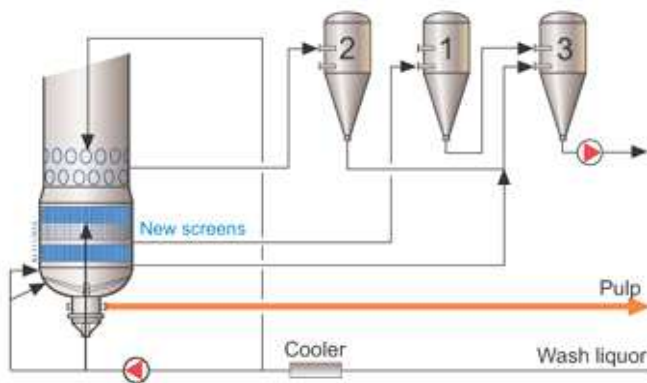


Figure 21 Liquor extraction from a rebuilt digester/28/.

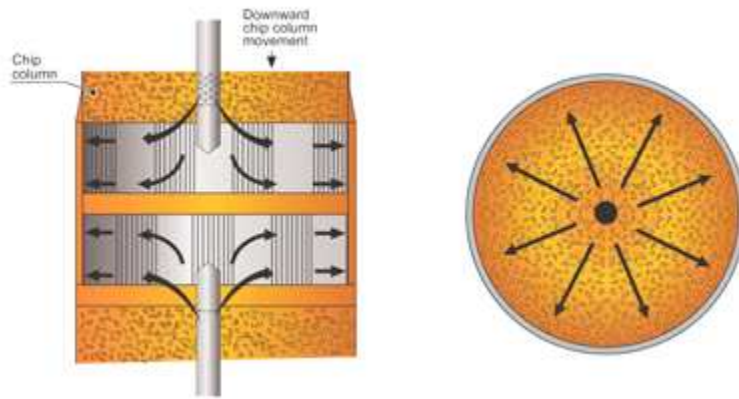


Figure 22 Principle of radial wash /28/.

Hi-Heat washing played an important role in chemical pulping until washing technology developed and digester length peaked in the early 1970's. Hi-Heat washing was useful as long as digester length did not limit the washing effect. When the machine suppliers were forced to reduce the digester length, the importance of Hi-Heat washing started to decline.

A new trend in digester development is to make the digester as simple as possible. The digester contains only from two to four screen sections, whereas older digesters had at least four screen sections. With fewer screen sections the digester is simpler and easier to operate, but at the same time there are fewer options how to run the digester. There are also other economic reasons affecting digester development and the modification of old digesters. For example, when the center pipe of the digester breaks down, it is sometimes not replaced; instead the whole digester is run in downflow cooking mode. Nevertheless, Hi-Heat washing has been found to improve the recovery of tall oil soap from the pulp /29/. Sometimes, Hi-Heat washing is wrongly claimed to cause channeling in the washing section, when the main reason for channeling is in fact poor digester bottom washing /30/.

It is important to understand the reasons why process changes were made in the Hi-Heat washing section and how these changes affected digester runnability, chip bed behavior and pulp quality. One way to understand these effects is to simulate process changes. The changes and their effect on chip bed packing is discussed in *Section 11.2.6*.

3.6. Conclusion concerning continuous digester development

Evolution of the kraft cooking digesters has been stepwise development. Digester development has gone through major changes from vertical digester to continuous digester. At the moment the continuous cooking digester is the dominant design of the pulping digesters. Most installed continuous digesters share a similar design for key process equipment. However, a variation on process configuration exists. Basically every continuous digester operated differently. This means that every continuous digester have different optimal operating conditions. To solve these conditions by measuring process conditions is difficult, because there are so few measurements in digester are. There is no measurement for measuring concentration profile of cooking liquor inside the chip bed. Models are needed to get better information about the conditions inside the digester. Next chapter discusses about kraft cooking models and how they give information about the process conditions inside the digester.

4. Kraft cooking modeling

4.1. Introduction

Although the changes in digester design over the years have largely been driven by trial and error some attempts have also been made to understand the fundamentals of kraft cooking. This has been done by the construction of fundamental models describing the chemistry and physics of kraft cooking. The following chapter starts with a brief review of how this type of models are constructed and goes on describing the most commonly used models at the moment.

One of the challenges in kraft cook modeling is the different scales of models needed for different phenomenon in the process. This chapter introduces the concept of different modeling scales in kraft cooking and shows what kind of models there are at different modeling scales.

4.1.1. Empirical versus theoretical models

Models can be divided into empirical (black box) and theoretical models. Empirical models are based on experimental data and the model does not describe the process behavior fully. An ideal theoretical model explains nature fully, and all the reactions and all different process conditions thoroughly. Usually, all the models used in chemical engineering include both some empirical results as well as some theoretical background of the different reactions.

Theoretical pulping models try to explain the complex behavior of kraft cooking. On the other hand, empirical models lack generality. Empirical models are based on a set of experimental conditions, which are unlikely to predict the outcome of the cooks outside of the experimental setup conditions. Empirical models cannot for example predict how changes in the raw materials affect the outcome of the cook. However, sometimes use of empirical models is recommended, for example, when simplicity is needed to solve the modeling problem.

Kraft pulping is all about chemical engineering. General models for chemical engineering consist of models for different unit operations. For example mixing, flow, heat transfer and so on. These models are used for modeling single unit operations. Different unit operations can be combined to model complete processes.

4.2. *Different modeling levels in kraft cooking*

Kraft cooking should be modeled at different reaction levels, because different reactions can be described at different levels of knowledge. Sometimes, it is enough that kraft cooking process is looked at one reaction level, but more often different reaction levels should be combined together. For example, lignin reactions are occurring at the molecular scale, but the mass transfer of the cooking chemicals is happening at chip scale.

As stated above, a model of kraft cooking should be able to deal with the fact that wood pulping is heterogeneous in many ways. This means that the model should also work at several levels of knowledge. The process should be examined at the level of a complete cooking plant. On the other hand, the system should also be understood at the reactor scale. The process can be examined one reactor at a time. An even more detailed way to look at the cooking plant is to examine the process at chip scale. Within an individual chip the process can be studied at fiber scale, and further, when taking one fiber at a time, the process can be understood at molecular scale. The different scales are presented in the list below and in Chapters 5-8. More detailed descriptions of different processes are provided later on.

This study concentrates on a kraft pulp cooking plant. The different levels of kraft cooking plant modeling are the following:

1. Mill scale
2. Reactor scale
3. Chip bed scale
4. Chip scale
5. Fiber scale
6. Molecular scale

A good model should be able combine different modeling scales. These models are used for predicting flows inside the reactor. These predictions are based on conditions from molecular scale to chip bed scale.

5. Mill scale

In this work, mill scale modeling is the most general level of describing the kraft cooking process. When modeling kraft pulping on the mill scale the objective is to get an idea of the consumptions of wood, chemicals and heat as a function of pulp production. Basically this means thinking of the cooking plant as a single unit (chip silo, impregnation vessel, digester). This kind of process thinking is illustrated in *Figure 23*.

At the cooking plant level, the process is viewed at production control level. This is the daily control level of the mill. There are crude models to calculate the production at this level. This is the mill manager level.

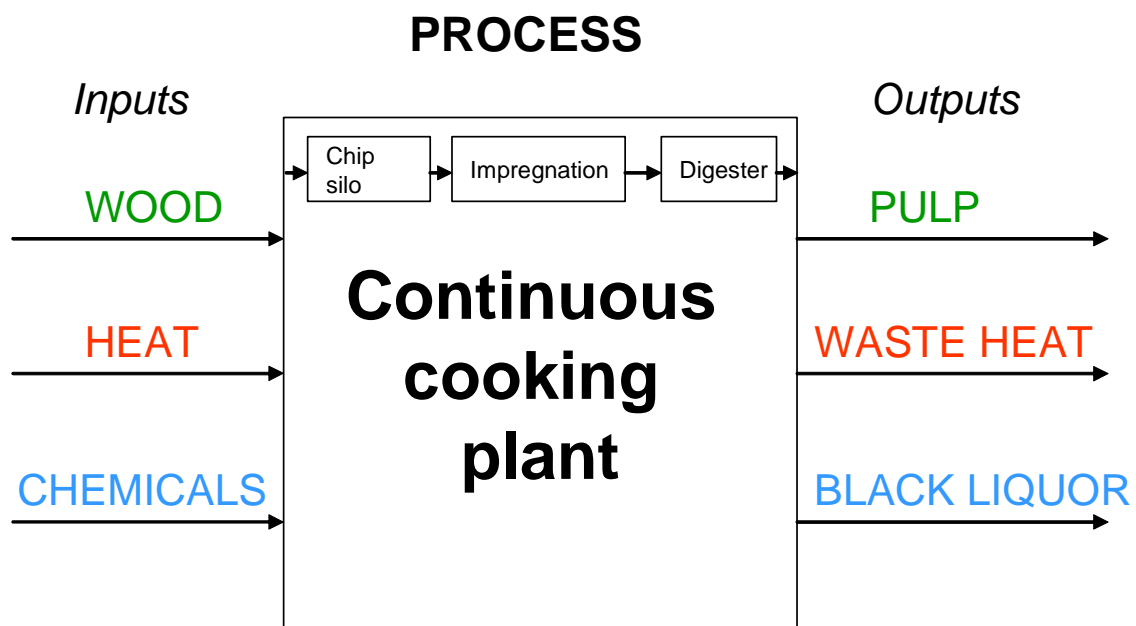


Figure 23 Mill-scale model of the kraft cooking process.

The following setup is needed to build an example of a mill-scale model:

The continuous cooking plant model combines different reactors within the cooking plant. Unit operations of this kind are for example the chip silo, impregnation vessel and digester. Commonly used variables for modeling a continuous cooking plant are:

- Wood
- Heat
- Chemicals

In addition, a simple mass conversion model is needed to transform wood, heat and chemicals into pulp, heat and black liquor.

Usually, the cooking plant is modeled either at verbal or diagram level. The verbal-level model gives a verbal idea of the cooking reactions. For example, in kraft cooking half of the wood is dissolved in the cooking plant.

If a mill-scale model is used on its own, it is generally to make economic evaluations. In this case, factors such as the cost of different inputs are added.

6. Reactor scale

A more detailed approach from mill scale is to look at the cooking plant at the reactor scale. Chip steaming, impregnation and cooking are done in separate unit operations. Each vessel can be simulated by combining appropriate unit operations. The cooking plant can therefore be modeled by a series of different unit operations. Mass and energy balances are then calculated for each unit operation. An example of a reactor-scale model of an impregnation vessel is shown in *Figure 24*. An impregnation vessel is used as an example because of its simplicity compared to a digester.

At the reactor level, different reactors are handled as unit operations. This is the process control level of the plant. Process control takes care of different parts of the plant. For example impregnation vessel temperature can be controlled at this level. The process control level is the operator level.

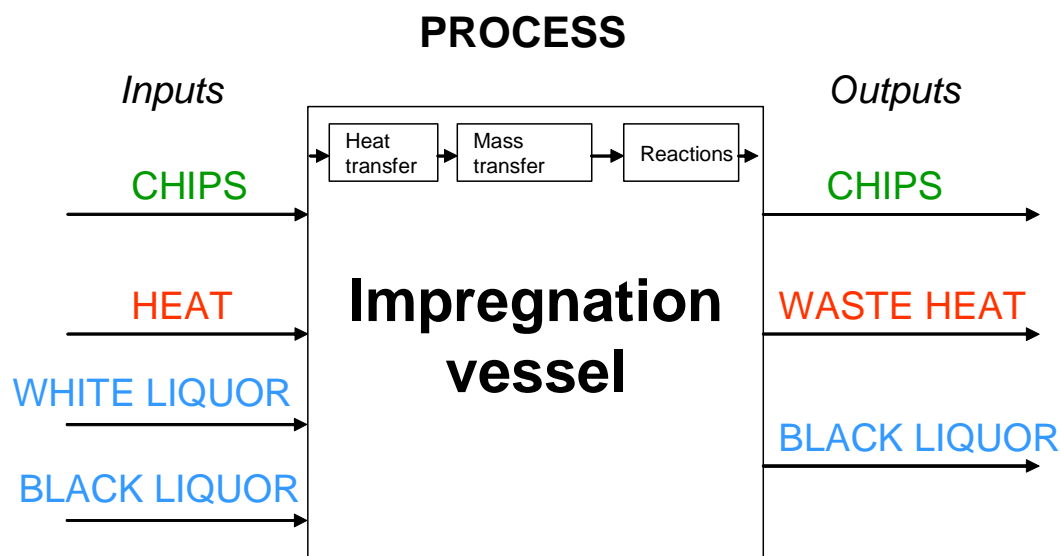


Figure 24 Reactor-scale model of kraft cooking process.

The following setup is needed to build an example of a reactor-scale model:

The reactor scale model combines different unit operations within the reactor. Commonly used variables for modeling the impregnation vessel are:

- Chip flow in
- Heat of the process
- White liquor flow in
- Black liquor flow in

In addition, models for heat and mass transfer and reactions are needed. Within each process, different process flows are mixed.

Models which describe the reactor-scale process are usually mathematical models. Both mass and heat transfer are solved with mathematical equations.

Reactions are solved with a kinetic equation. The reactor-scale model is used by engineers for process study.

6.1. *Flows in reactors*

Any industrial chemical vessel, including continuous kraft pulp digesters, is a complex device in which heat transfer, mass transfer, diffusion and friction occur along with chemical reactions.

It is important to know process conditions inside the reactor, because the heat and flow conditions affect the outcome of the cooking. They also affect the runnability of the reactor and the quality of the product flowing inside it.

This section concentrates on theories concerning different flow models for reactors. Some earlier research on liquor and chip flows inside the digester is also covered.

6.1.1. Theory of flows in reactors

To predict the exact behavior of a vessel as a chemical reactor we need to know what is happening inside it. We need to know how different components are flowing in the reactor and how long it takes for liquor and solids to flow through the vessel. Only then can we know the process conditions inside the reactor.

By simplification, ideal reactor flow models are called either plug flow or mixing model. Usually the flow inside the reactor is something between the plug flow and mixing model. Though real reactors never fully satisfy these requirements, a large number of reactors can be approximated by these ideal models.

To be able to measure non-ideal processes, the residence time of the different flows must be known. Usually, residence time cannot be measured directly, so an experimental method is needed to solve the residence time. These methods are called stimulus-response techniques. In these techniques, something is done to the system and after a certain delay it is seen how the system reacts or responds to the stimulus. The desired information about the system is obtained by analyzing the response.

6.1.2. Ideal flow models

Real reactors may be simulated through a combination of ideal models. The most common ideal models are the following:

- In the Plug Flow Reactor (PFR) liquor is flowing as a plug, where no back mixing is assumed. Tracer added at the inlet will have the same concentration when exiting the process.
- In the Continuous Stirred Tank Reactor (CSTR) mixing is complete and instantaneous. If tracer is added at the inlet, the tracer concentration at the reactor outlet is initially equal to the total mass of tracer divided by the reactor volume, decaying at an exponential rate.
- The Dispersion model is based on Fick's diffusion law with an empirical dispersion coefficient D_d substituted for the diffusion coefficient. The flow through the reactor follows a plug pattern but will deviate from the ideal plug condition because of axial dispersion.

Different kinds of dispersion models and ideal plug flow and continuous stirred tank reactor models are shown in *Figure 25*. $E(t)$ is the response to the addition of tracer, which shows the fraction of tracer at any specific moment.

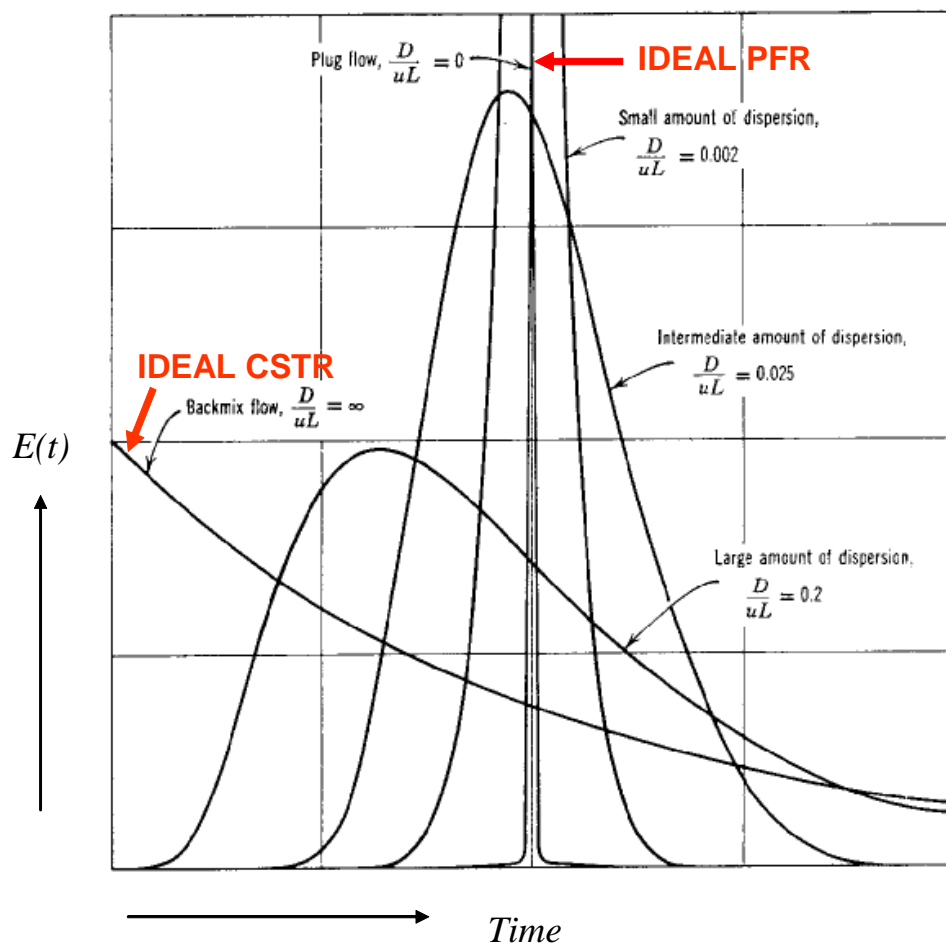


Figure 25 E curve response for variations of the dispersion model /31, p. 264/

Real reactors will not perform exactly like an ideal reactor or an ideal complete mix model and will therefore produce tracer outputs that deviate from the curves shown above. The flows inside a real digester are discussed in more detail in *Chapter 12*.

6.1.3. Residence time distribution

The distribution of residence times of reactants or tracers in a reactor, the RTD (Residence Time Distribution), is a key datum for determining reactor performance. In this section it is shown how a tracer test may be used to see how close the reactor is to the ideal flow. RTD gives information about the performance of the reactor and this information can be used to build non-ideal model reactor behavior.

Non-reactive (or radioactive) substances may be used for determining reactor performance. The progress of the concentration (or radioactivity) is followed along the reactor. The data is converted into residence time distribution, which tells the distribution of the residence time. RTD is often represented in the frequency function E , as shown in *Figure 26*. E is the exit age distribution of fluid leaving a reactor or the residence time distribution of fluid in a vessel /31, p. 245/.

Real reactor flows differ from the models available to describe the flow phenomena inside the reactor. Residence time distribution largely characterizes the reactor type.

The most typical fault flows are short-cut flow and dead zone volume (*Figure 26*). Short-cut means that part of the inflow material flows quickly out from the process. Dead zone refers to that part of the volume where mass transfer with the rest of the process volume is very slow. The short-cut and dead zone terms do in fact refer to the same phenomenon with extreme names, and the terms cannot be separated from each other. For example, 40 % dead zone can be interpreted to mean 60 % short-cutting /32/.

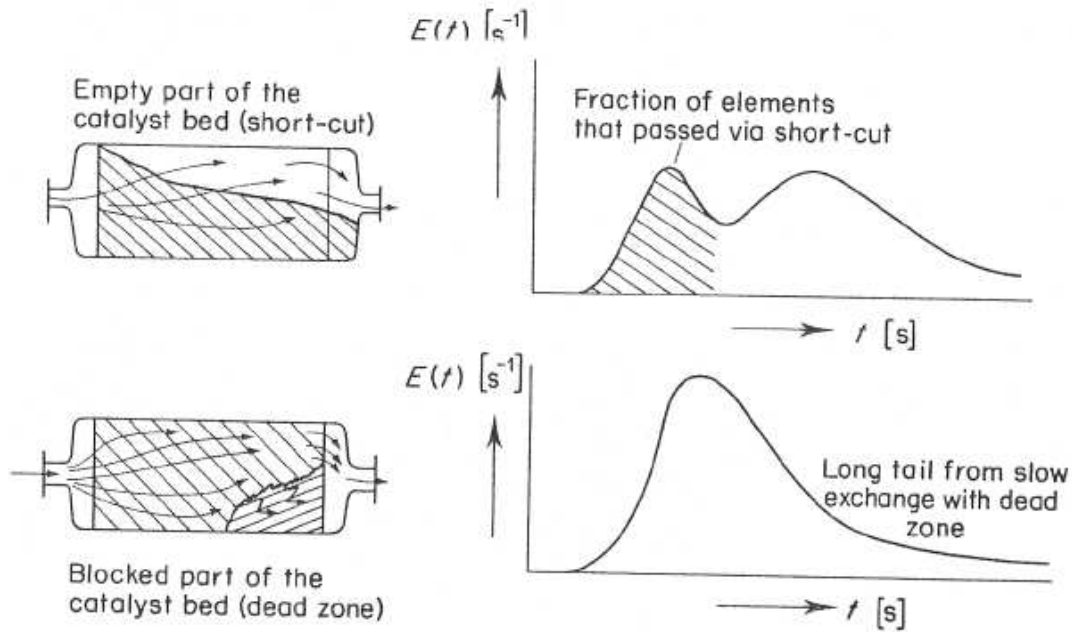


Figure 26 Examples of short-cutting and dead zones and their appearances in the $E(t)$ diagram /33/.

Different residence time distributions are shown in Figure 27 below. *Normal behavior* is an ideal flow pattern. *Bypassing* means that part of the flow bypasses the reactor or a particular flow region. When a part of the liquor leaving a flow region is circulated back into the reactor, the flow type may be *internal recirculation*. *Residence volume error* means that the total volume of the reactor has been measured incorrectly. *Parallel paths* means that there are two different paths which are measured.

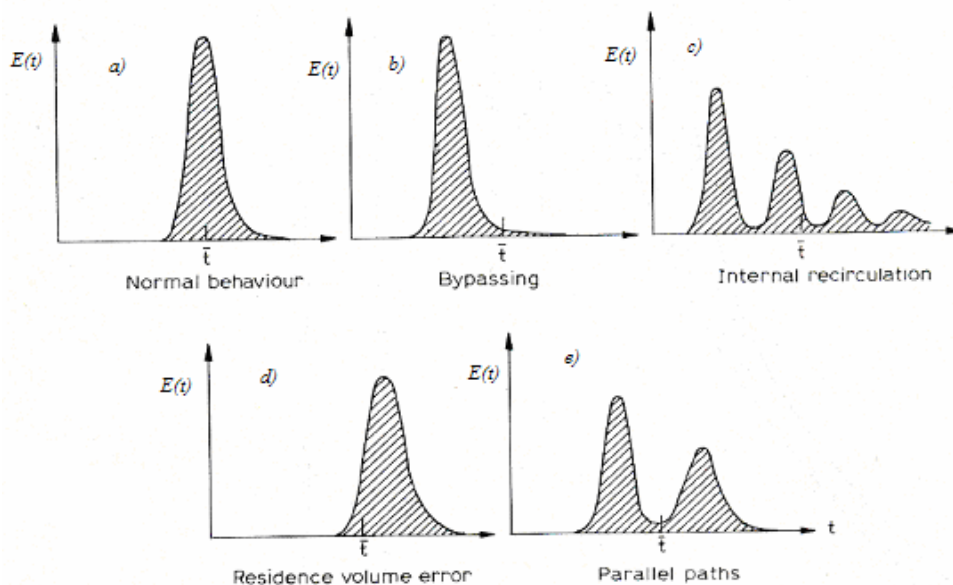


Figure 27 Different types of residence time distributions: a) Normal behavior, b) Bypassing, c) Internal recirculation, d) Residence time error and e) Parallel paths.

Many types of models can be used to characterize non-ideal flow patterns within vessels [31, p. 260]. Some draw an analogy between mixing in a flow and a diffusion process. These models are called dispersion models. Others visualize various flow regions connected in series or in parallel. These are called mixed models.

6.2. Flows in a continuous digester

It is important to know the real process flows inside the continuous digester and the interconnections between the cooking liquor and chips. In this way, we will know the process conditions and how they affect pulp quality inside the digester.

The liquor flow affects the concentration gradient, temperature profile and compaction of the chip bed. The concentration gradient and temperature profile affect the reactions of the chips. Delays in the chip flow affect the reaction times in different parts of the digester. If these delays are known in more detail, the digester can be more accurately controlled.

Direct liquor flow measurement is troublesome. For example, a step change in liquor temperature is difficult to detect because of strong mixing and it is often unclear how the wood affects the temperature. Using a non-reactive tracer is a practical way to study the liquor flows inside the digester. A commonly used tracer for a liquor-solid system is lithium chloride. One liquor flow study is discussed in *Section 6.2.2*.

Direct measurement of the chip flow is difficult. Only kraft pulp mills which alternately produce softwood and hardwood pulp have direct information on chip flow delays in the digester. Otherwise, the chip flow has to be measured indirectly. One possibility is to use a radioactive tracer.

6.2.1. Chip flow in digester

There are not very many studies of the chip flow in continuous digesters. One was made by Kamyr in 1957 [34]. Research was done by impregnating chips with radioactive material and the flow of radioactive chips was followed along the digester. The results are presented in *Figure 28*.

The height of the digester was 22 meters and its diameter about 2.5 meters. Five different radioactive chips were inserted into the digester and the chip flows were examined at different heights of the digester.

One chip (# 4) went near the digester wall. The other chips went through the inner part of the digester (*Figure 28*). Two of the chips were so close to the center part of the digester that they remained undetected until the blow line.

The two undetected chips moved faster than the three other chips. The velocity difference was mainly caused by the fact that in the center part of the digester the chips are moving faster at the bottom of the digester. If only the last one and a half meters are taken into account, tracer number 2 traveled it in 5 minutes, tracer number 5 in 10 minutes and tracer number 4 in 12 minutes.

The same phenomenon was noticed in a mill where the production was increased from 100 ADt/d to 130 ADt/d and the reject level rose to 3-4 % of pulp /35/. Chips in the middle of the digester passed faster through the digester, and the residence time in the middle was too short to achieve a low kappa number.

During the last 5 meters in the digester, chips move in a circular motion in the same direction as the bottom scraper is moving. When chips have reached the bottom scraper level, they have been moving 60° at cross-section level. At the bottom scraper, the chips turn full circle or even more before they go out from the digester. This circular motion with high temperature might affect fibers unfavorably, and this mechanistic treatment may lead to lower tear strength /36/.

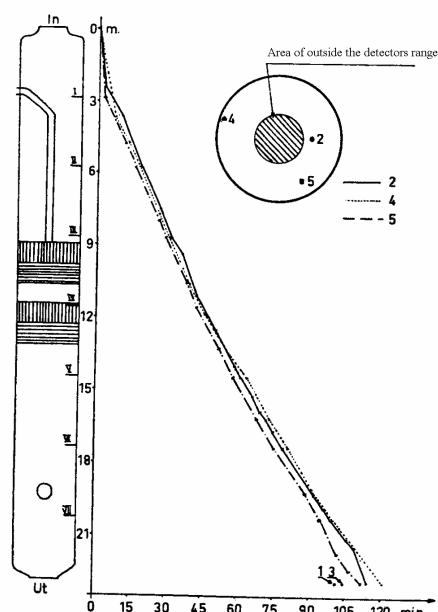


Figure 28. Diagram of chip flow along the digester and how chips are located in the cross-section (Joutseno, 100 ADt/d) /34/.

Another study to measure chip flow delays in the digester was made by Hamilton /37/. He noticed some rotation of the chips in the bottom part of the digester. In addition, he found that there was no evidence of channeling of the chip flow in the digester. This research was done by tracing the flow of an individual radioactive chip inside the digester. Proving the occurrence of channeling by measuring one chip at a time is difficult.

Based on Hamilton's /37/ results, measured chip flow in digester is 10 % slower compared theoretical values. This means that chip flow in digester is packing. Because of this, there is need for chip bed packing model.

6.2.2. Liquor flow in digester

Haas /25/ has studied the liquor flow inside the digester. The purpose of the study was to examine channeling in the bottom part of the digester. *Figure 29* shows how the experiment was conducted and the results. Lithium chloride was added into the MCC circulation. Samples were collected from the bottom part of the digester, in the blow line. Two experiments were made in MCC mode, where the liquor was flowing upwards in the MCC section (as shown in *Figure 29*). The flows were compared to conventional cooking where the liquor was flowing downwards in the MCC section.

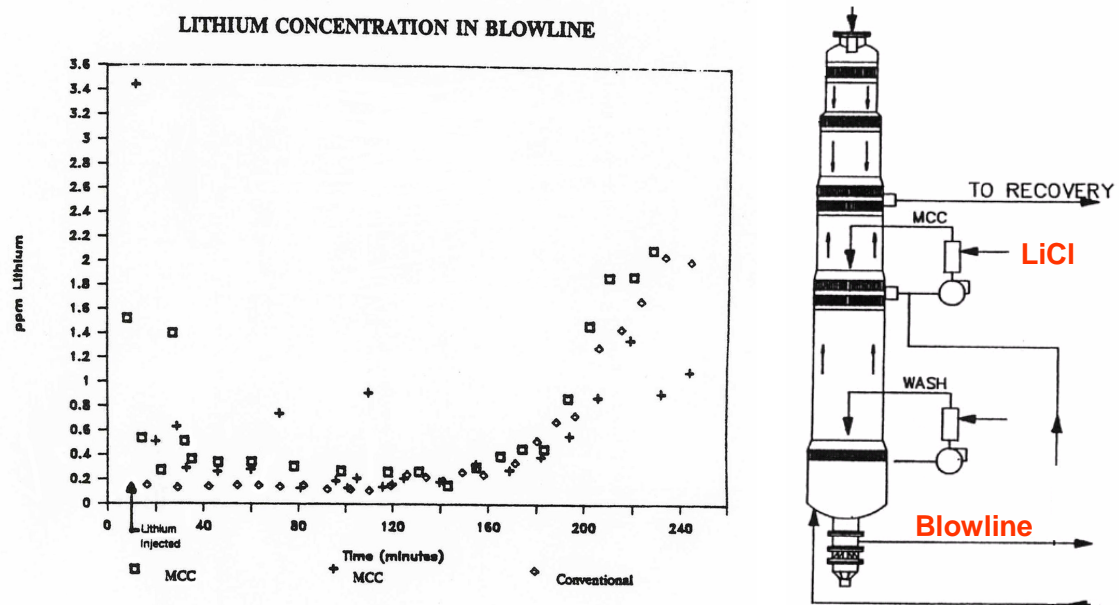


Figure 29. Lithium concentration in blowline liquor (left picture) and sample addition point in the MCC section of the digester and sample collected from the blow line (right picture) /25/. Arrows show the flow direction of the liquor in MCC cooking.

Haas noticed in his research that there was short-cutting of MCC cooking liquor, but not in conventional cooking. The cooking liquor added in the MCC section did not flow uniformly upwards. Part of the liquor added into the digester made a short-cut in the bottom part of the digester, which was seen as a higher lithium concentration in the blow line within minutes after adding the tracer. Liquor flowed non-uniformly in the digester. Because liquor flow inside the digester is non-ideal, there is need to model the liquor flow in chip bed scale.

7. Chip bed scale

If one wants to understand the behavior of the digester (or impregnation vessel) more deeply, chip conditions inside the reactor have to be modeled. In the digester the chips and cooking liquor form a chip column. This column is also called a chip bed, with the chips moving downwards. Liquor can flow in the same direction as the chips or it can flow against the chip flow. This is illustrated in *Figure 30*, where chips are moving downwards and cooking liquor upwards.

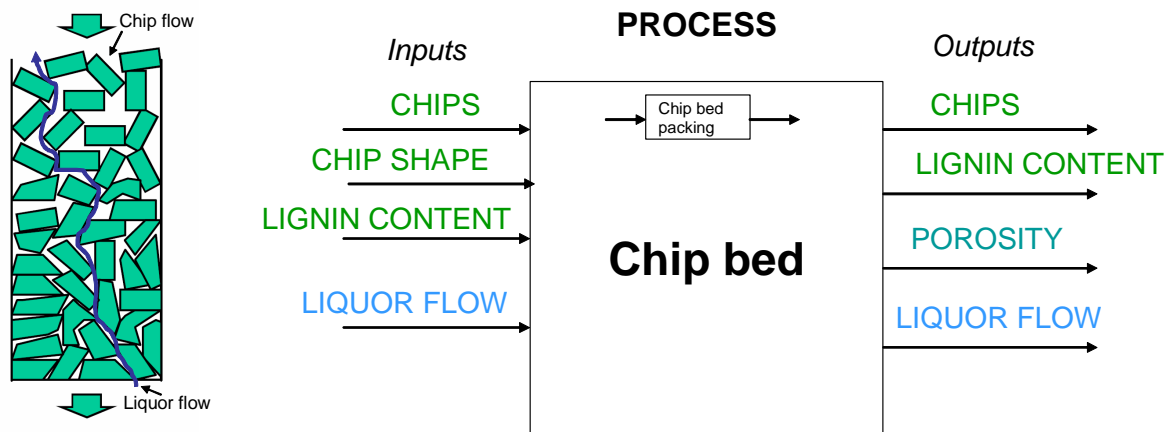


Figure 30 Chip bed-scale model of kraft cooking [38].

The following setup is needed to build an example of a chip bed-scale model:

The chip bed-scale model is based on a model which takes into account the softening of chips in the reactor. Chip softening leads to a more packed chip column. Commonly used variables for modeling chip bed packing are:

- Chip flow
- Shape of the chips
- Lignin content
- Liquor flow

Based on these input values, the chip bed packing can be modeled by using a model for chip bed softening. Chip and liquor flow affect the porosity of the bed. The shape of the chips affects the flow resistance of the liquor flow. Chip bed softening increases as the lignin content of the chips decreases along the reactor. The change in lignin content is solved using either a mathematical kinetic model or a diagram model, where the lignin content decreases in a linear fashion with the H-factor. Chip bed packing is solved as different forces act at in the bed. These packing forces are gravity, buoyancy, digester wall friction and flow friction of the cooking liquor.

The chip bed-scale model is used for studying the process flows and forces acting in the bed.

7.1. Chip bed packing

Again, when modeling the real phenomenon of chip bed packing different modeling scales is needed. Chip bed softening and packing is modeled at chip scale and flow phenomena are modeled in chip bed scale.

The chip bed is formed inside the continuous digester vessel. It is a long column, in which there can be over one billion chips. The chip column consists of many different kinds of chips. The chip thickness varies, the basic density of the wood varies, the sapwood content varies etc. A change in chip dimensions affects the chip bed packing. As the amount of the smaller chips increases, chip column takes less space, because smaller chips fill the open volume around bigger chips. Higher packing degree affects liquor flow inside the digester.

The chip bed is assumed to be elastic and compressible and the elasticity of the chip bed is "reversible". The flexibility of the chip bed considerably affects the void space of the bed /39/.

The bed pressure is used to define the contact forces over a reference surface /39/. The bed pressure affects the volume fraction of the solid particles and liquid in the reactor.

Different forces affect the total bed pressure and the chip bed compaction. These forces are presented in *Figure 31*. Gravitational force is pushing the chip bed downwards. If the chip bed is stationary, the packing degree increases towards the bottom of the bed. When the chip bed moves downwards, the digester wall and downward chip flow subject the chip bed to a frictional force. There is also flow friction between the liquor flow and chip column. The friction term is dependent on the directions and relative values of the flows of the chips and liquor inside the digester. The packing term is used as a common term to describe the conditions inside the digester. For example, when the pressure of the chip bed is high and the porosity of the chip bed low, the chip bed packing is high.

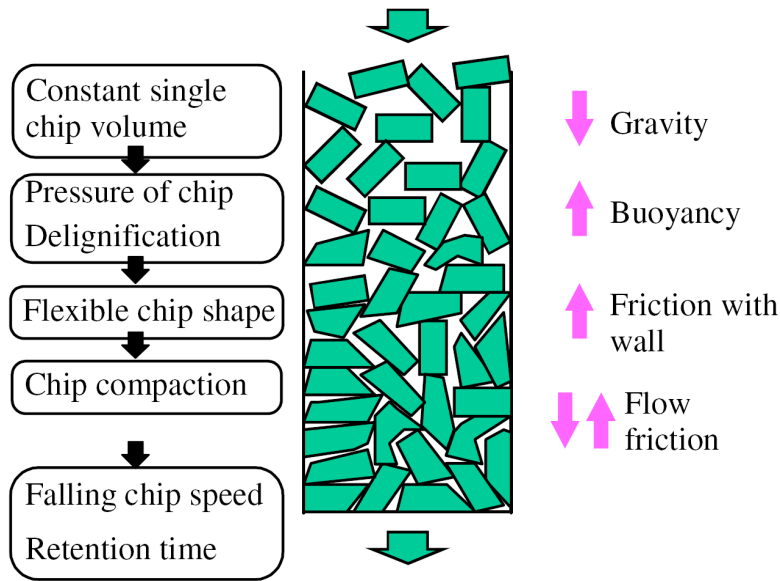


Figure 31 Chip compaction and compaction forces at chip scale /38/.

7.1.1. Volume fractions

The chip bed consists of chips and free space between the chips. This free space is called free liquor. When the chip bed is compressed, the free space between the chips decreases.

The chip volume (ϵ_c) and free liquor volume (ϵ_l) make up the total volume of the chip column. The chip column can be divided into different sections called blocks. In modeling, the total volume of one digester block is set to be 1. This is shown in

Figure 32 and in Equation 1. The volume difference between the free liquor and the chips is the basic idea behind the packing of the chip bed.

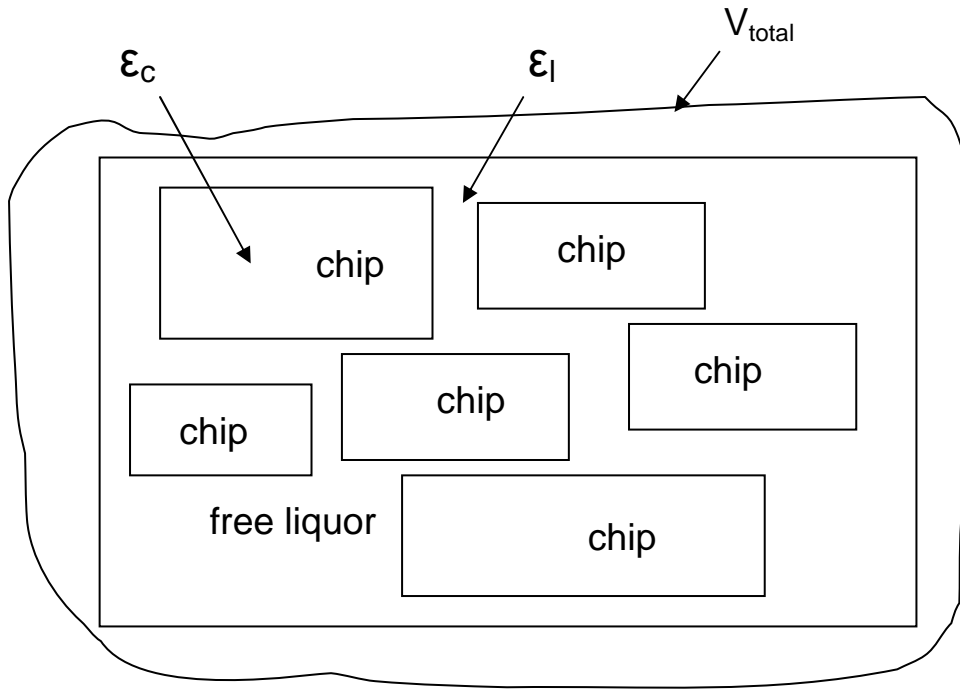


Figure 32 Chip and free liquor volume (ϵ_c is volume taken by the chips in the block, ϵ_l is volume taken by the free liquor and V_{total} is the total volume of the block).

$$\epsilon_c + \epsilon_l = 1$$

Equation 1

Where:

ϵ_c is the volume taken by the chips in a cooking block

ϵ_l is the volume taken by the free liquor in a cooking block

The compaction of the chip bed can be expressed as porosity, too. The porosity of the chip bed is defined as the volume fraction of the free liquor that is occupied in one digester block. Porosity (or free liquor volume fraction or void volume) is calculated in Equation 2.

$$\epsilon_l = \frac{V_l}{V_{total}}$$

Equation 2

Where:

V_l is the volume of the free liquor

V_{total} is the total volume of the cooking block

Lee /40/ studied the porosity (void fraction) of the chips. Chips were cooked to a certain kappa number. Chips were cooled and inserted in a separate container. In this separate container, the chips were compressed at a certain

pressure. After compression, the new volume occupied by the chips was measured. The volume change corresponded to the change in void fraction. The effect of compacting pressure and kappa number on the porosity of the chip bed is shown in *Figure 33*.

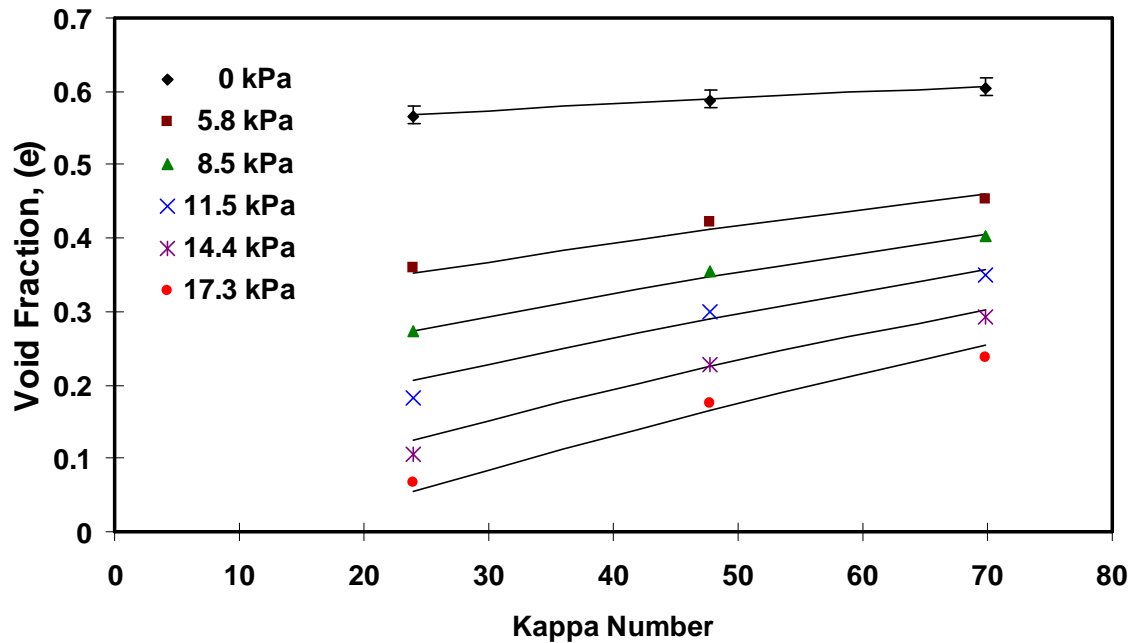


Figure 33 Effect of compacting pressure and kappa number on porosity (void fraction) of the chip bed /40/.

The porosity of the uncooked chip bed is typically 0.6-0.7, depending on the wood species and chip size distribution. During the course of the cook, chips are softened and become easier to compress into a smaller volume, so the porosity of the chip bed decreases. Porosity can decrease to less than 0.2 during the cook, if the chips are compressed with high compressing forces in laboratory studies, as shown by Lee /40/ in *Figure 33*.

7.1.2. Friction force

Modeling the distribution of the frictional force on the chip bed in detail would be impossible. Therefore, it is generally accepted that the bed of solid particles cannot carry shear forces. The shear force caused by the friction between the wall and the moving solid particles is therefore considered as a volume force /39/, i.e. the force acts on the whole volume of the chips.

Härkönen assumes the reactor vessel to be cylindrical /39/. The friction force between the wall and the bed is then

$$dF = \pi \mu p_{\text{wall}} d_c * dl_c \quad \text{Equation 3}$$

Where:

F is force

d_c is element diameter

l_c is element length

p_{wall} is bed pressure near the wall

μ is friction coefficient

The friction force as a volume force (*Equation 4*) is

$$F_{\mu} = \frac{dF}{dV} \quad \text{Equation 4}$$

Where:

F_{μ} is friction force

V is volume

The friction force between the wall and the bed is then

$$F_{\mu} = \frac{\mu p_c}{D_{Digester}} \quad \text{Equation 5}$$

Where:

p_c is pressure of the chip

$D_{Digester}$ is diameter of the digester

7.1.3. Flow resistance

The energy loss, characterized by the pressure drop of the moving process fluid, is an important operating parameter for packed columns. The Ergun equation is commonly used for calculating the pressure drop through the packed bed /41/:

$$\frac{dP}{dL_b} = A_s \frac{\mu_v}{d_p^2} \frac{(1 - \epsilon_l)^2}{\epsilon_l^3} U + B_s \frac{(1 - \epsilon_l)}{\epsilon_l^3} U^2 \quad \text{Equation 6}$$

Where:

dP is pressure drop trough packed bed (Pa)

L_b is length of the packed bed

A_s, B_s are coefficient of pressure drop

μ_v is dynamic viscosity (Pa*s)

d_p is equivalent diameter of particles (m)

U is superficial velocity flow across the column (m/s)

The Ergun correlation accounts for two energy loss terms: a viscous loss term and an inertial loss term. A_s and B_s are used to denote particle shape factors.

Wood chip shape and size vary. Consequently, it is difficult to obtain a good equivalent particle shape to use in *Equation 6*. This is why Härkönen /46/ rewrote the Ergun equation to describe the liquid flow resistance of a compressed chip column:

$$\frac{dP}{dL_b} = R_1 \left[\frac{(1 - \varepsilon_l)^2}{\varepsilon_l^3} \right] U + R_2 \left[\frac{(1 - \varepsilon_l)}{\varepsilon_l^3} \right] U^2 \quad \text{Equation 7}$$

Where:

R_1 , R_2 are coefficients of pressure drop correlation ($\text{Pa}\cdot\text{s}/\text{m}^2$)

The constant R_1 and R_2 depend on the wood species, chip shape and chip size distribution. Different values found in the literature are presented in *Table 3*.

Table 3 Chip flow resistance data in the literature.

Reference	R_1	R_2
Härkönen /46/, Pine (Pinus silvestris)	$0.046 \cdot 10^5$	$39 \cdot 10^5$
Lindqvist /47/, Pine, mix 1 (Pinus silvestris)	$0.28 \cdot 10^5$	$-1.3 \cdot 10^5$
Lindqvist /47/, Pine, mix 2 (Pinus silvestris)	$0.51 \cdot 10^5$	$-3.5 \cdot 10^5$
Lindqvist /47/, Pine, mix 3 (Pinus silvestris)	$0.055 \cdot 10^5$	$-0.057 \cdot 10^5$
Lammi /48/, Eucalyptus camaldulensis	$5.5 \cdot 10^5$	$7.5 \cdot 10^5$
Lammi /48/, Birch	$2.8 \cdot 10^5$	$-12 \cdot 10^5$
Lee /40/, White spruce	$0.4 \cdot 10^5$	$58 \cdot 10^5$
Wang and Gullichsen /42/, Pine (Pinus silvestris)	$0.52 \cdot 10^5$	$15 \cdot 10^5$
Wang and Gullichsen /42/, Pine (4 mm thick chips) (Pinus silvestris)	$0.82 \cdot 10^5$	$-1.1 \cdot 10^5$
Hernesniemi model /43/	$0.025 \cdot 10^5$	$21 \cdot 10^5$

The R_1 and R_2 values depend on the wood species used and chip geometry. Negative values for the shape factor are not physically possible, so there is something wrong with the laboratory experiments. Therefore, one should be careful when using packing model. There is variation in flow resistance parameters in the packing model. For example, softwood values are used for the flow resistance of hardwood chips /38, 44/.

Based on laboratory data, Härkönen derived an empirical equation for the volume fraction of the chips (*Equation 8*), which depends on the chip compacting pressure and the degree of delignification, i.e. kappa number. Empirical constants for *Equation 8* are collected in Table 4.

$$\varepsilon_l = k_0 + \left(\frac{p_c}{10^4} \right)^{k_1} (-k_2 + k_3 \ln(Kapppanumber)) \quad \text{Equation 8}$$

Where:

k_0, k_1, k_2, k_3 are empirical constants

p_c is compacting pressure applied to chip bed (Pa)

Table 4 Empirical constants for calculating volume fraction of chips.

Reference	k_0	k_1	k_2	k_3
Härkönen /46/, softwood	0.356	0.59	0.139	0.831
Lammi /48/, hardwood	0.37	0.64	0.151	0.697

The chip bed pressure is force acting on the chip column divided by the digester cross sectional area of the chip column. Bed pressure can be calculated using force balance on the chip column. The acting forces in a force balance are gravity, buoyancy, friction between the digester wall and the chip column and flow friction between the chip column and the liquor. The force balance for the chip column is given by Saltin /45/:

$$\frac{dp_c}{dL_b} = (\rho_c - \rho_l) \varepsilon_c g - \mu \frac{p_c}{D_{\text{digester}}} \pm \frac{dp_l}{dL_b} \quad \text{Equation 9}$$

Where:

ρ_c is density of the chips (kg/m³)

ρ_l is density of the cooking liquor (kg/m³)

g is gravity coefficient (m/s²)

p_c is chip pressure, contact force (Pa)

p_l is liquor pressure (Pa)

Static head combines the gravitational force and the buoyancy. Change in the volume fraction of the chips along the digester affects the buoyant force in the digester. In the flow friction term, “±”, is due to the possibility of counter-current and co-current liquor flow inside the digester.

7.2. Chip bed packing modeling

The continuous kraft cooking process takes place in a particle bed formed by chips. The chips provide the heat and mass transfer surface area and create the flow system for the liquids used in the cooking process. Chip properties change in the bed during the cooking process. Some researchers have attempted to model the behavior of the chip bed to understand the complex interactions between the chips, cooking liquor motion and cooking reactions.

Härkönen /39, 46/ measured the flow resistance trough a chip column and used the Ergun equation /41/ to describe the interaction between the liquor

and chip phase in a two-dimensional model of a continuous digester. This study was made with softwood (Nordic pine). The model has been used for modeling chip bed packing and flows of chips and cooking liquor.

Other researchers have tried to broaden their understanding of the chip bed packing, which affects the liquor pressure drop as a function of flow velocity. Lindqvist /47/ used different chip size distributions for softwood to examine chip bed packing and Lammi /48/ made the same kind of study for hardwood chips. More recently, Lee /49/ has built a model for the fluid pressure drop in a packed chip bed. Lee covers the progress of packing modeling in great detail in his research.

Chip bed modeling has become more widely used in recent years. Models have been used for a number of purposes: for dynamic studies of the digester with grade transition and model predictive control /44, 50/; for computational fluid dynamics study and for comparing different calculation methods to solve pressure distribution inside the digester /51/; for determining the kappa number profile inside the digester and for modeling the packing of a campaign digester producing both softwood and hardwood pulps /52, 53/. He et al. modeled the liquid flow through a packed bed and the pressure drop inside the digester, using simple empirical reaction models /54/. Fernandes et al. used an empirical packing model and a sophisticated kinetic model for the different wood component reactions as a basis for to build a mathematical model of a moving-bed reactor for kraft cooking /55/.

7.3. Liquor flow in chip bed

The main parameters of the chip bed are its porosity and compressibility. Porosity and compressibility are strongly inter-related: the more compressible the chip bed, the lower the porosity generated. The porosity of the chip bed means the volumetric share of the empty space between the chip particles.

The velocity of a fluid flowing through a tank containing a bed of particles is often described in terms of the superficial velocity, also called empty tower velocity. Average velocity is the velocity which also takes into account the changes in porosity in the tower. Different velocity terms are shown in *Figure 34*. The superficial velocity is defined as the velocity which would be achieved by the fluid if it flowed at the same volumetric flow rate through the tank when it is empty of particles.

v_l = liquor volumetric flow (m^3/s)

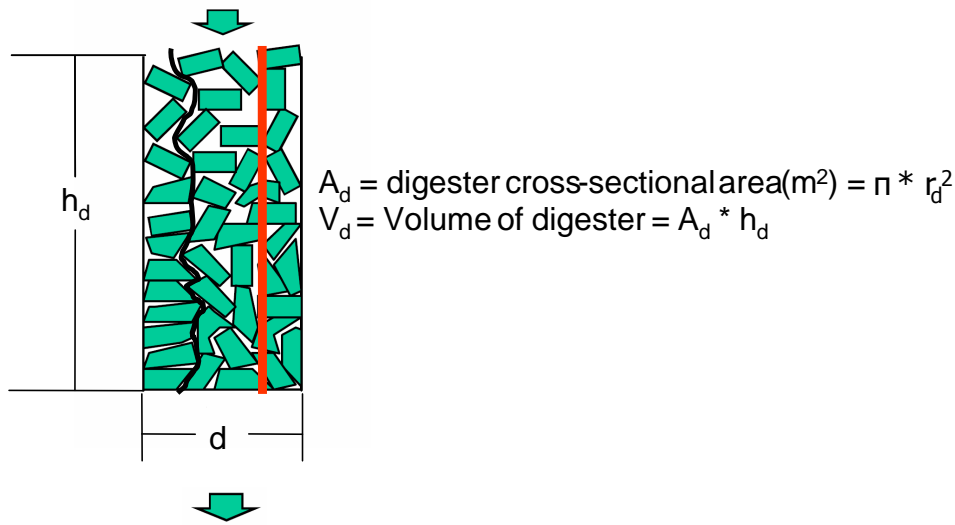


Figure 34 Flow speed in a chip bed. Liquor empty tower velocity (superficial velocity) is denoted by the red line and average liquor flow velocity by the black line.

Mathematically, the superficial velocity u_i is given by Equation 10:

$$u_i = \frac{v_l}{A_d} \quad \text{Equation 10}$$

Where:

v_l is volumetric flow of liquor (m^3/s)

A_d is digester cross sectional area (m^2)

u_i is superficial velocity of liquor (m/s)

In some cases, the average velocity of the liquor is calculated. This means that the free liquor porosity of the chip bed is taken into account in the calculation. The equation for mean velocity is given by Equation 11:

$$\bar{v} = \frac{v_l}{\epsilon_l A_d} \quad \text{Equation 11}$$

The liquor flow inside the digester is not uniform. Near the digester walls the porosity of the chip bed is higher compared to the average bed packing. The liquor flow near the wall is faster than in the middle of the bed. This effect is called wall effect. The wall effect of the digester wall is shown in Figure 35.

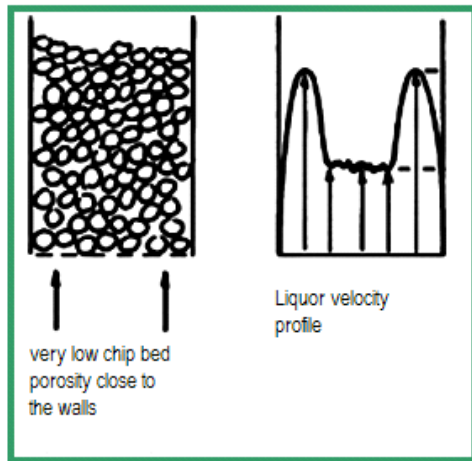


Figure 35 Wall effect phenomena inside the digester.

Chip bed packing and its effect on the uniformity of the cook in continuous cooking are not very well known. Joutsimo /56/ studied the packing degree and uniformity in a SuperBatch digester. An increase in the initial chip bed packing from 160 to 210 kg/m³ (porosity from 0.6 to 0.5) decreased the kappa variation somewhat, as shown in Figure 36. Too low chip bed packing leads to non-uniform liquor flow inside the digester. This leads to greater kappa variation of the pulp after cooking. With normal chip bed packing, the kappa standard deviation was 1.5 to 3.6. As the packing degree was increased, the kappa standard deviation decreased to less than 1.

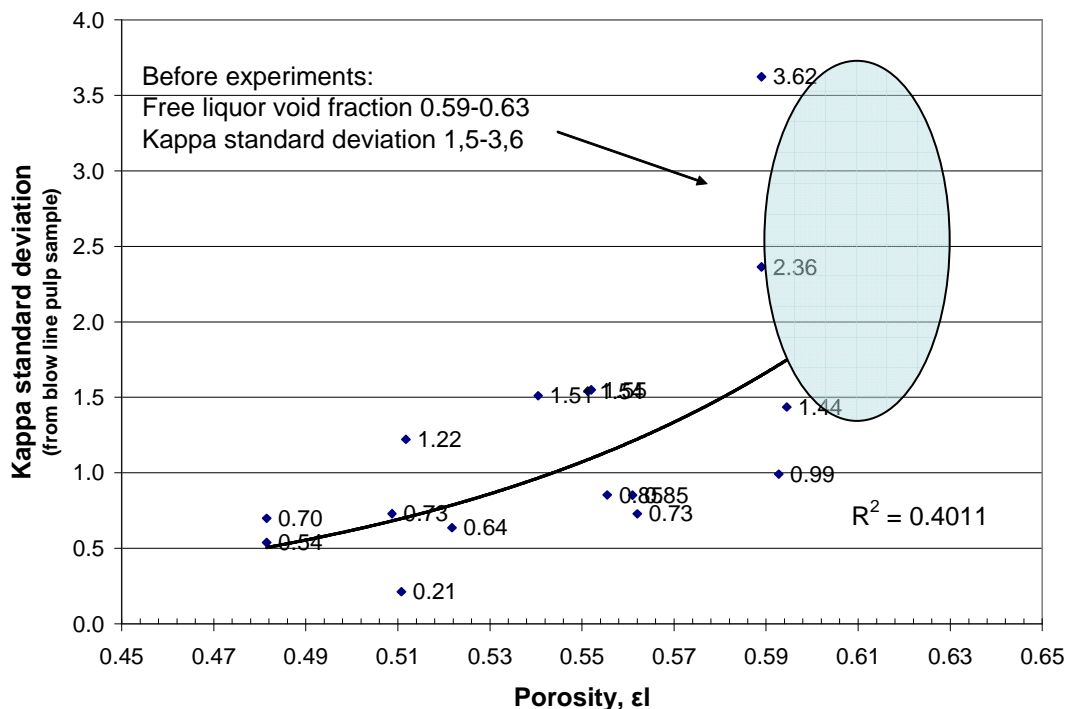


Figure 36 Kappa standard deviation at different porosities in the beginning of a batch cook /56/.

8. Chip scale

If one wants to have an accurate value for the chip lignin content in a given part of the chip bed, a chip-scale model is needed. The chip-scale model combines fiber-scale and molecular-scale models.

In the chip-scale model, cooking liquor is transported into the chip through mass transfer (diffusion and penetration). As the reactions proceed, the reaction products are transported from the chip with diffusion. Lignin and carbohydrates react at reaction sites inside the chip (*Figure 37*).

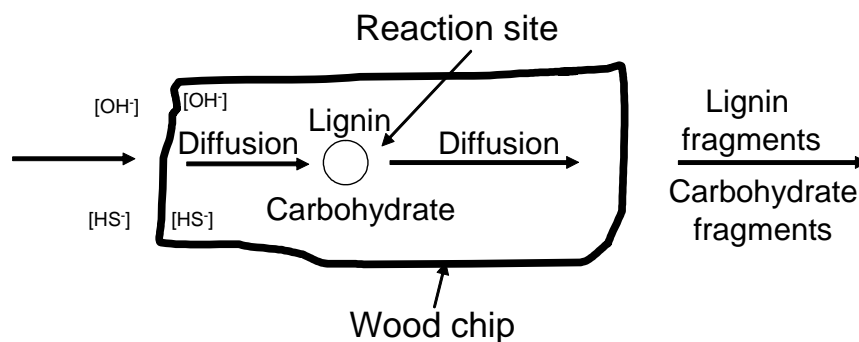


Figure 37 Chip-scale model of kraft cooking.

The following setup is needed to build an example of a chip-scale model:

The chip-scale model is based on chip-scale mass transfer and molecular-scale reactions. Mass transfer is modeled with a diffusion model and reactions are modeled with a kinetic model, where each carbohydrate (cellulose, glucomannan and xylan) has its own model or all carbohydrates are treated together.

Commonly used variables are:

- Carbohydrate content on wood
- Lignin content on wood
- Hydroxide concentration of chip [OH concentration]
- Hydrosulfide concentration of chip [HS concentration]
- Temperature of the chip

Based on these input values, the single chip can be modeled. The reacting components of a wood chip are carbohydrates and lignin. Hydroxide and hydrosulfide are diffused into the chip. These chemicals react with different wood components. The temperature of the wood chip affects the reaction rates of the wood components. The unit operations of the chip-scale model are diffusion and reactions.

8.1. Rate of delignification reactions

The most common way to model kraft cooking is to examine the lignin reactions. Lignin reactions are modeled at either molecular scale or chip scale.

Much of the early research into kraft pulping concentrated on the temperature dependency of lignin reactions. The Arrhenius equation can be used for solving the temperature effect on the lignin reaction rate. The equation was first proposed by the Dutch chemist van 't Hoff /57/. Arrhenius /58/ provided a physical justification and interpretation for it. This reaction rate can be expressed by the following equation (*Equation 12*):

$$k = A_1 * \exp \frac{-E_a}{RT} \quad \text{Equation 12}$$

Where:

T is the absolute temperature

A₁ is the pre-exponential factor

R is universal gas constant. R = 8.314 J/(K*mol)

E_a is called the activation energy of the reaction (J/mol)

The justification deals with understanding the reaction rate as the sum of molecular collisions ending the reactions. A₁ is a factor, which expresses the probability that the molecules contain a favorable orientation and will be able to proceed in a collision. For first-order reactions the unit is 1/s. E_a is the energy, needed for the reaction to take place.

The following expression can be used for the lignin reaction rate:

$$-dL/dt = k * L \quad \text{Equation 13}$$

Where:

L is lignin (% on wood)

Kraft kinetic models are good for predicting the characteristics of cooking, for example kappa number, yield, rejects of the wood and so on. These models can be used for controlling, training and optimization studies. The models can also be used for designing the digester.

8.2. Kraft cooking kinetic models

Kraft and soda cooking modeling has in the past concentrated on lignin reactions. The reason for this was that the analysis methods were not sufficiently advanced for more detailed modeling. This is also seen as a lack of understanding of what really takes place inside the digester.

One of the first kinetic models for pulping of wood, put forward in the 1920's, was built just by looking the end point of the cook and the level of un-reacted material in soda cooking. At the time, wood was divided into cellulose and incrust. Over the years, the methods for analyzing wood components gradually improved, and the kinetic model became more detailed. In the 1930's, it was assumed that lignin reactions consist of three consecutive reactions. After carbohydrate analysis was developed for analyzing wood, the kinetic models became more accurate for other wood components. The most recent models also take into account the carbohydrate kinetics.

At the moment, there are three common approaches to kinetic modeling for lignin and carbohydrates in kraft cooking. The first approach is called the Purdue model, which was developed at the University of Purdue in the 1970's. The second approach is called the Gustafson model, which was developed by Gustafson et al. at the University of Washington in the 1980's. The third model, developed in 2003, is called the Andersson model. The basic idea of all of these kinetic models is the same. All the models are based on the Arrhenius kinetic model, where different wood components have different activation energies. Otherwise, there are differences between the models. The main difference is that these kinetic equations are either parallel or consecutive. The Gustafson model is consecutive and the Andersson and Purdue models are parallel. The difference between lignin reactions with different models is also discussed in the following.

8.2.1. History of kraft cooking kinetic modeling

According to Arrhenius, the experiments with pulping of sawdust using different cooking liquors and cooking temperatures, conducted by Bruun /59/ in 1923, gave good results that can be interpreted using the laws of monomolecular reactions. Arrhenius /60/ calculated reaction velocity for the cellulose and non-cellulose part (lignin and hemicellulose), referred to as incrust by Arrhenius, in different temperatures by using the reaction rate law. The incrust reaction rate is presented in Figure 38 below.

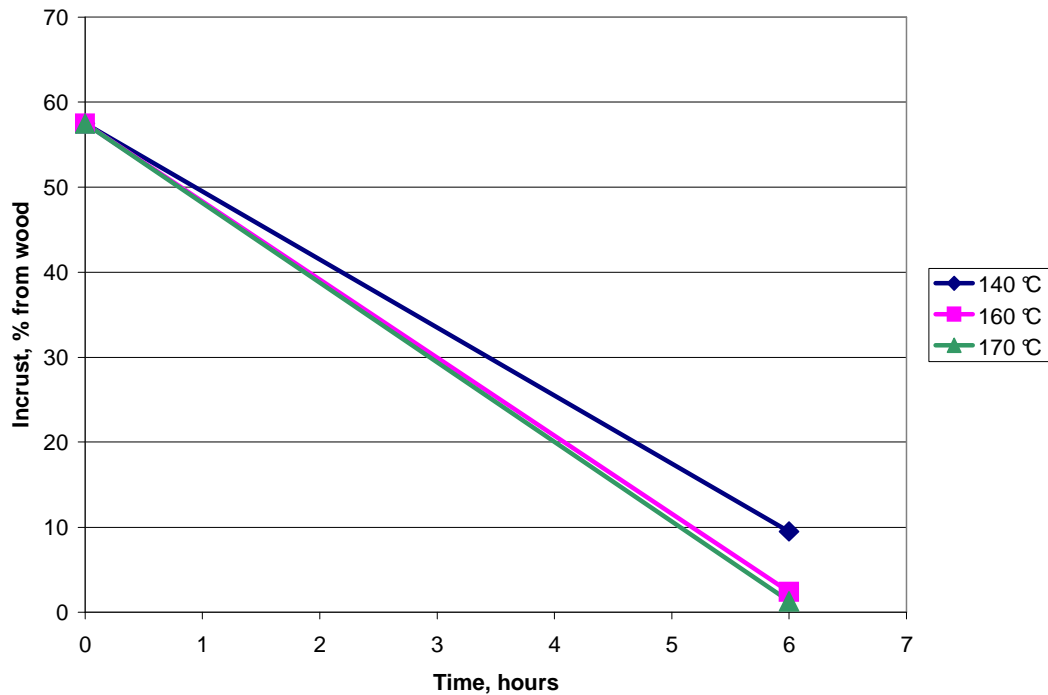


Figure 38 In crust reaction rate as interpreted by Arrhenius from the experiments of Bruun /59/.

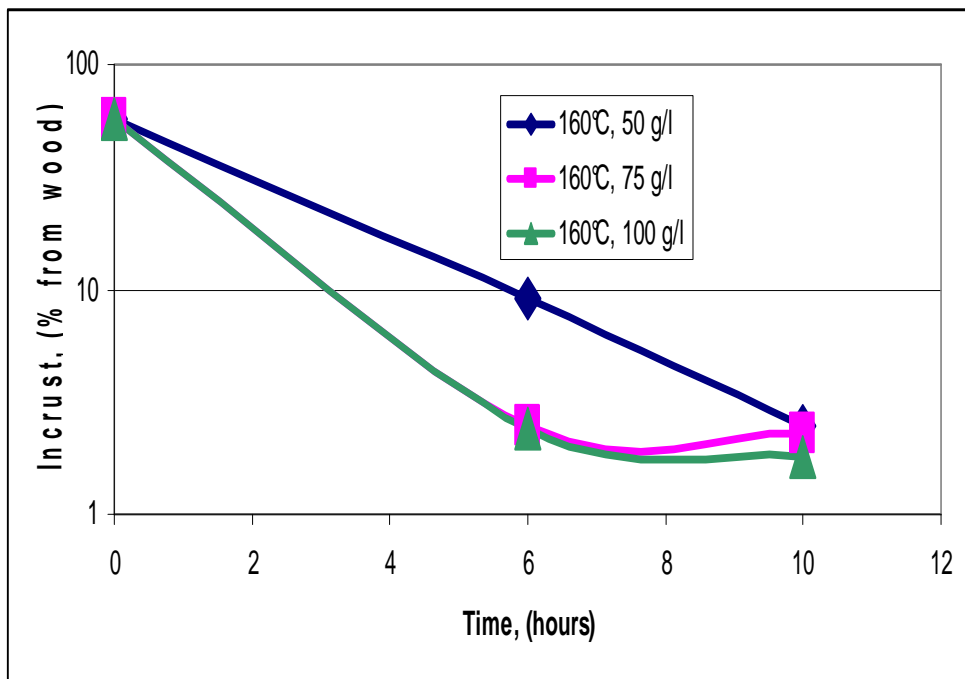


Figure 39 In crust reaction rate as interpreted by Arrhenius from the experiments of Bruun (for more results). In crust reactions are not linear /59/.

Based on Bruun's cooking data /59/, the activation energy of the removal of in crust in soda cooking is about 38 kJ per gram of reacting wood. In crust was the part of wood which dissolves during the cook. The complement of in crust

was cellulose, which the rest of the wood consists of. For example in Bruun's data, the wood incrust content was 57.5 % on wood. At the end of the cook the incrust yield was 10 % on wood compared with a total yield of 50 % on wood.

Schmidt-Nielsen, Bruun's superior, /61/ criticized Arrhenius based on the following: a) cooking reactions are not monomolecular, b) the cooking chemical concentration decreases during cooking and c) diffusion varies during cooking.

Also, Arrhenius did not take into account the preliminary period of cooking where 12-15 % of wood is removed. Cooking data was gathered at the end of the cook, when the total yield of the pulp is less than 55 % on wood (*Figure 38*).

Mitchell and Yorston /62/ were among the first to divide the kraft cooking reactions into three parts. These three parts are called the initial, bulk and residual stage. However, it seems doubtful that the rate equations which they brought forward have any real theoretical background /63/.

Laroque and Maass /64/ showed that the rate of delignification is the same, regardless of the wood particle size (from wood meal to 10 mm wood chip thickness). In another investigation, Laroque and Maass /65/ studied the value for sodium hydroxide (soda delignification) activation energy. They found that the activation energy for the soda cook is about 134 kJ per gram of lignin. Experiments were done at three different temperatures and three different concentrations (120, 140 and 160 °C and 2, 9, 14 moles of NaOH). At the high cooking liquor concentrations, particle size may not affect the reaction rate, but the concentrations are on the high side for commercial pulp production.

Goldfinger /66/ briefly reviewed the definitions of kinetics, examining the findings of Maass and collaborators /67, 68/ for the kraft pulping process ($T \sim 50-130$ °C). According to Goldfinger, the activation energy of lignin dissolution at the beginning is lower than shown by Laroque and Maass /65/ (about 67 kJ/mol) but starts to increase when about 40 % delignification is reached. Finally it reaches the value of 92 kJ/mol. The energy of non-lignin removal is considered as a constant, being about 88 kJ/mol.

The kinetic studies by Laroque and Maass /65/ were used as a reference when Vroom /69/ established H-factor model for kraft cooking. In Vroom's model, the same value for activation energy as the one determined by Laroque and Maass /65/ for soda cooking, is used for kraft cooking.

Nolan et al. /70,71/ criticized the work done, stating that kraft cooking reactions do not follow the first-order reaction law and that kraft cooking reactions should only be considered as an apparent first-order reaction.

According to Nolan, the complexity of the process makes a true first-order reaction highly improbable.

Yan and Johnson commented that rate-determining reactions are unknown for kraft cooking lignin reactions /63/. Nevertheless, the temperature dependence of the delignification reaction using the Arrhenius equation can be applied, provided that the reaction does not change over the temperature range studied /72/.

Still, the activation energy is most commonly used for building kraft kinetic models, which are used for controlling the cooking operation. Sixta /73/ has collected the values from most recent studies into his book. The activation energy varies widely between different studies. The variation is partly explained by the use of different wood species and by differences in how the laboratory experiments are conducted, but the values still vary within the same wood species.

8.2.2. Purdue model

The Purdue model was set up by Smith and Williams /74/. It was one of the first kinetic models set up both for softwood and hardwood. Figure 40 illustrates the normal lignin reaction rate during the kraft cook. The kinetic model for lignin reactions is divided into three parts. There is high- and low-reactivity lignin and un-reactive lignin. In this way, the model gives quite reasonable results at the end of the cook. The model is set up in a way that whatever changes are made in the process, the outcome of the lignin reaction is about the same /75/. These different lignin reaction models are set up experimentally.

There is not much theoretical background on how different wood components are set in the model. For instance, Wisniewski /76/ in his study of the Purdue model, sets the cellulose content in the beginning at 50 % on wood for softwood. In addition, the glucomannan content in the beginning is 2 % on wood. Sjöström shows that the cellulose content of softwood varies from 33 % to 42 % on wood and the glucomannan content from 14 % to 20 % on wood /77/.

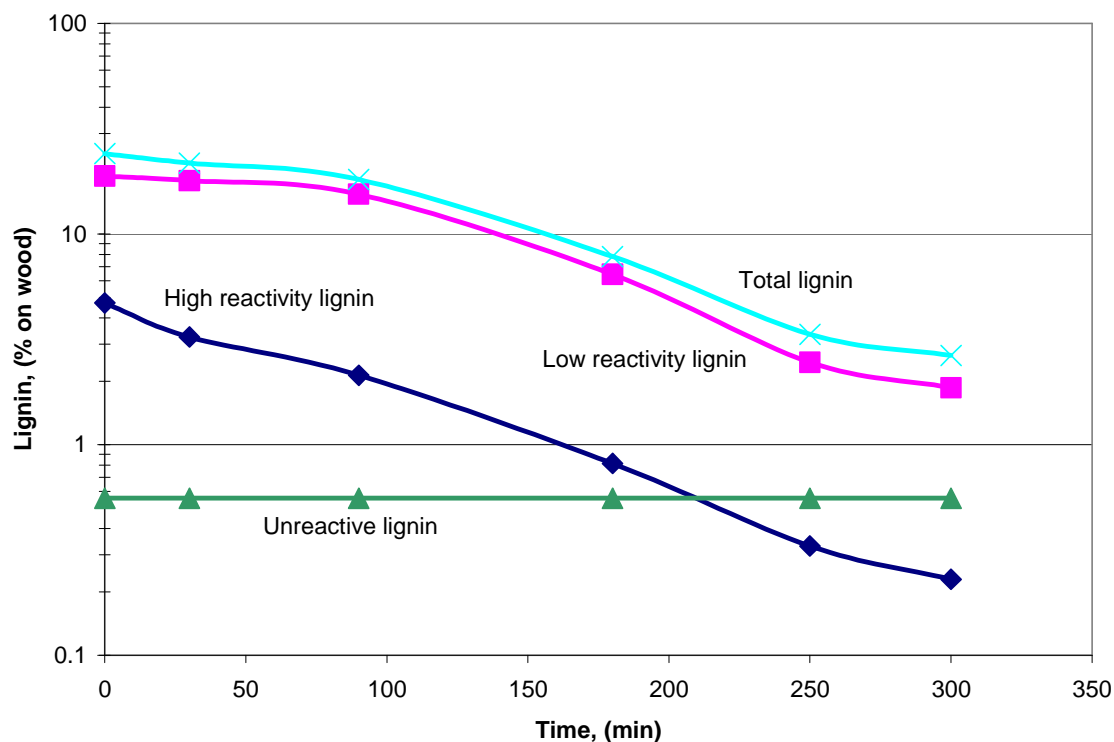


Figure 40 Schematic trend for lignin reactions according to Purdue model (Total lignin = Un-reactive lignin + High-reactivity lignin + Low-reactivity lignin). Equations adapted from /78/.

8.2.3. Gustafson model

The Gustafson model's /79/ lignin reactions are presented in Figure 41. Lignin reactions are divided into three parts. A more detailed presentation of the lignin delignification of the Gustafson model is given below, starting on page 78. The early part of delignification is called the initial stage. In laboratory experiments the initial stage is the heating-up period of the cook. During that stage a major portion of the cooking chemicals are consumed. About 20 % of the total lignin is removed and the total yield on wood can drop by about 20 %.

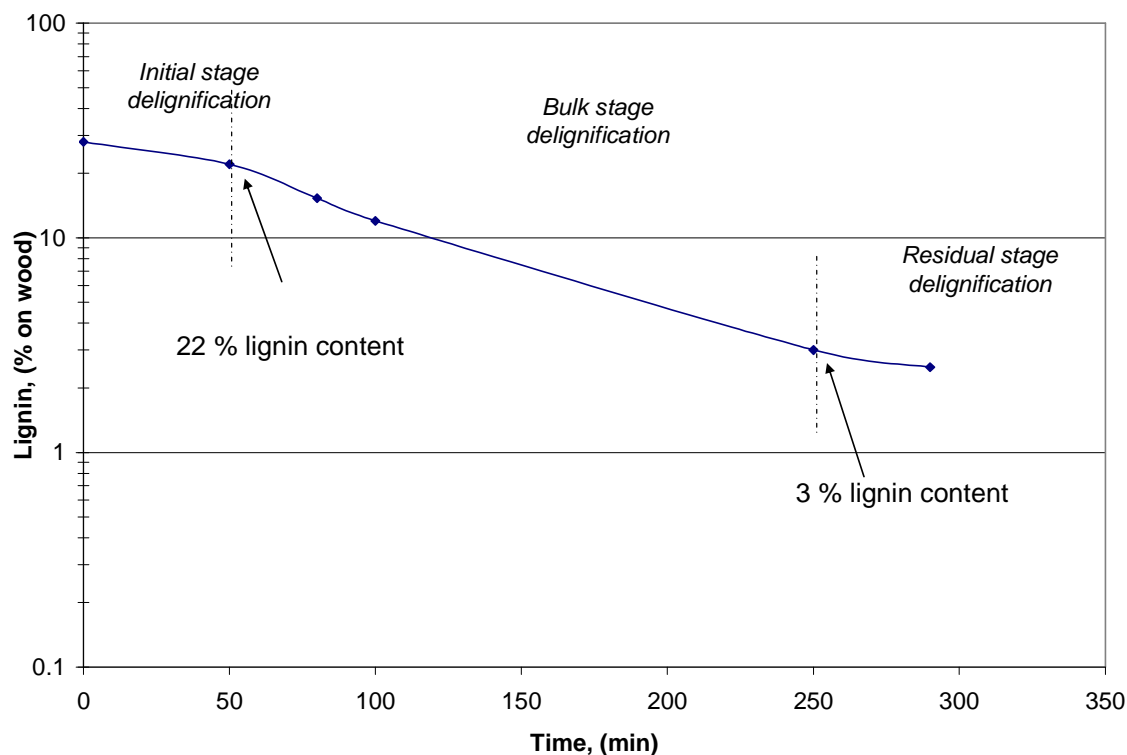


Figure 41 Schematic trend for the Gustafson model's lignin reactions.

The bulk stage can be said to start when the lignin content achieves a level of about 22 % on wood, as presented in Gustafson's model. More generally, the bulk stage is said to start when the cooking temperature reaches 140 °C in a laboratory cook /80, 81/. In laboratory kinetic studies, it usually takes one to two hours to reach this temperature.

When the lignin content of the pulp is decreased by 90 % of the original lignin content of the wood (from 28 % to about 2-3 % on wood), the lignin reactions start to slow down. From this point, the cooking reaction is said to be at a residual stage. The lignin content of the pulp at the transition point varies from 2.5 % on wood /82/ to 1.1 % on wood /83/. The transition from the bulk stage to the residual stage is shown in Figure 42.

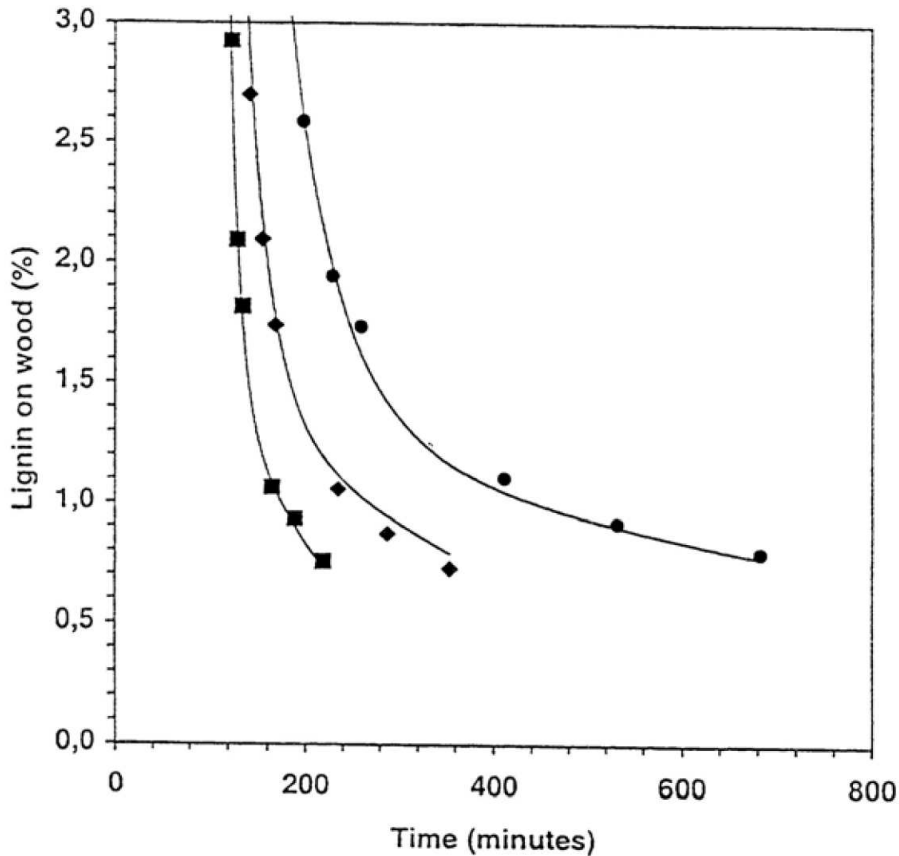


Figure 42 Remaining lignin, on wood, as a function of time for "normal" cooks at 160 °C (●), 170 °C (◆) and 180 °C (■) for spruce /84/. Initial cooking: 60 min at 135 °C, $[OH^-] = 0, 5 \text{ mol/l}$ and $[HS^-] = 0, 3 \text{ mol/l}$, liquor-to-wood ratio 31:1. Heat-up 1 °C per min. Cooking stage: $[OH^-] = 0, 44 \text{ mol/l}$ and $[HS^-] = 0, 28 \text{ mol/l}$, liquor-to-wood ratio 41:1.

8.2.4. Andersson model

The Andersson model /85/ is based on three different parallel lignin kinetic equations, as shown in Figure 43. This kind of approach has been previously presented by Lindgren and Lindström /86/. Andersson model is fitted from Lindgren and Lindström data and was validated by using autoclave and circulation cooks.

The starting value of each type of lignin has to be determined over again for new process conditions. These different lignin contents are dependent on the concentration of OH and HS ions and cooking temperatures. Different lignin components have about the same activation energy for delignification as in the Gustafson model. A major difference is that in the Andersson model, the lignin kinetic reactions are divided into three different parallel equations, while the Gustafson model is based on three different cooking stages.

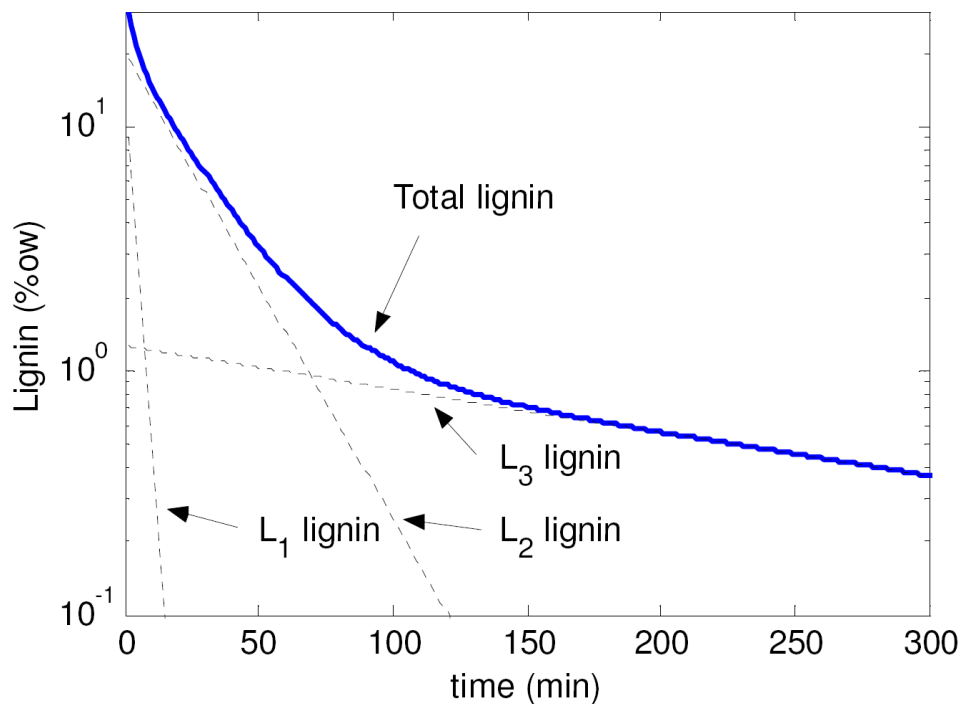


Figure 43 Schematic trends for total lignin (solid line) and L1 (initial lignin), L2 (bulk lignin) and L3 (residual lignin) (dashed lines) for a constant concentration according to the Andersson model /85/.

8.2.5. Comparison of kinetic models

This section discusses differences between different kraft cooking models. Different delignification models are shown in Figure 44. Figure is based on models developed by the Andersson /87/, Gustafson /79/ and Purdue /78/. These results are illustrated in Figure 44. The Andersson model simulates an ITC digester (Iso Thermal Cooking), the Gustafson model a CC digester (Compact Cooking) and the Purdue model an EMCC digester (Extended Modified Continuous Cooking). The lignin simulations of the Gustafson and Purdue models follow each other quite closely. The lower initial lignin content in the Gustafson and Andersson model is explained by separate impregnation before the cooking stage.

Dissolution of the lignin proceeds much faster in the Andersson model. Cooking temperature in Andersson and Gustafson model are 160 °C. Still the reactions are fast in Andersson, because after 100 minutes of cooking, the lignin level is less than 5 % from wood and it takes about 200 minutes to achieve that level in other models. Also the minimum level of the lignin is much smaller in Andersson model.

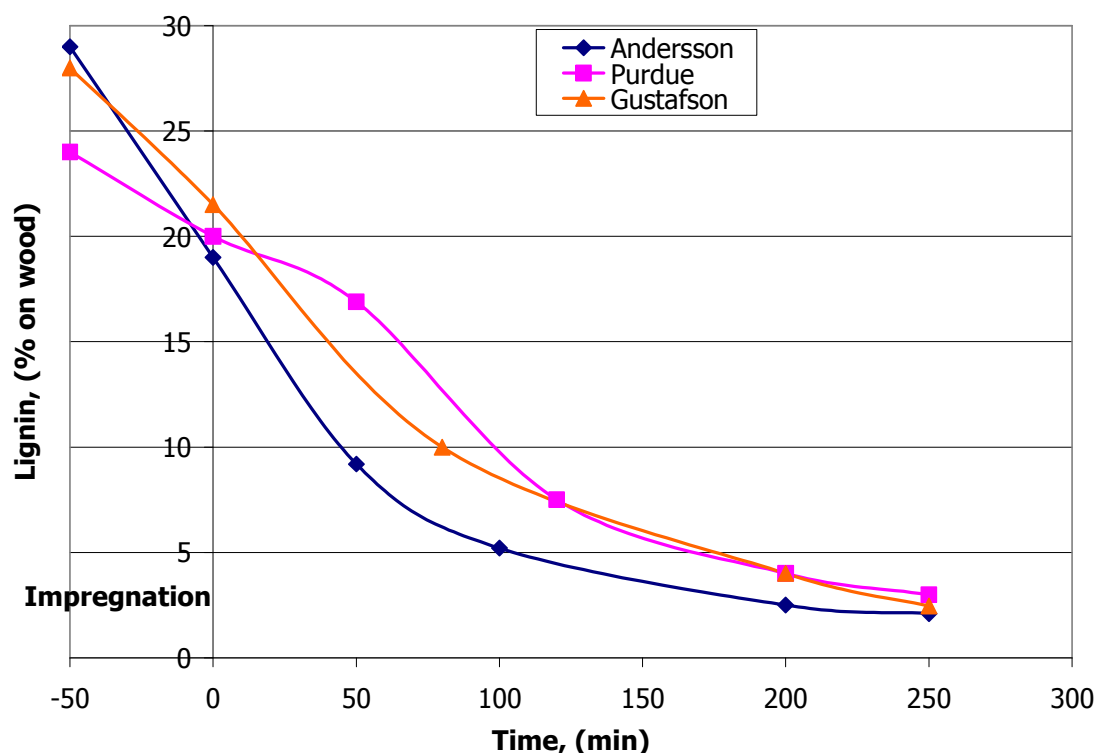


Figure 44 Lignin reactions during different kraft cooking models. The Andersson model's results are reproduced from presentation /87/.

Modeling lignin content range

Andersson states in his thesis /85/ that the delignification models in which lignin reactions are consecutive were found not to be suitable for extended delignification, so called residual stage delignification. When simulating the real kraft cooking process, the residual stage delignification is unimportant for modeling modern digester, because lignin content is so low in residual stage to continue the cooking. Andersson discusses modeling the lignin levels less than 1 % on wood, when making comparison between cooking models. Lignin levels for modern commercial pulps are left at about 3-4 % lignin contents on wood. It is of minor interest to discuss what kinds of results models are giving at the residual stage.

Consecutive versus simultaneous model

Andersson /85/ criticizes the Gustafson model for dividing the kraft cook into three stages. It is true that the cooking process does not follow these three stages as such, so this kind of division into different cooking stages is in some cases artificial. Mostly, the different stages can be distinguished because of the laboratory experimental setup.

Verification of models

Laboratory kraft cooking experiments can cause some artificial changes in the reaction rate for lignin. The influence of cooking time and temperature on lignin yield is illustrated in *Figure 45*. The initial part of the cook is quite straightforward. The temperature is low and the cooking chemicals diffuse inside wood and react rapidly with acetyl groups and with carbohydrates. These reactions consume a lot of chemicals. As long as the temperature is kept under 140 °C, the lignin reactions remain slow.

When temperature rises over 140 °C, lignin reactions speed up. The cooking liquor's alkali concentration changes very little during this stage. When the lignin level continues to decrease under 5 % on wood, the alkali concentration starts to decrease.

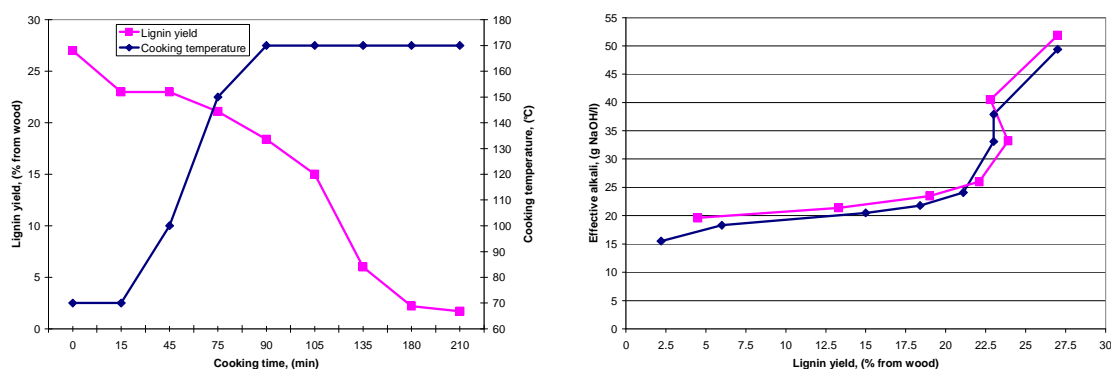


Figure 45 Influence of cooking time and temperature on lignin yield. Effective alkali 23 % NaOH, sulphidity 38%, impregnation time 15 min, heat-up time 90 min, liquor-to-wood ratio 3.8:1 /88/. Figures reproduced from Rekunen's thesis.

In the early part of the cook, the lignin reactions are slow. The lignin reaction rate increases as the temperature rises to about 140 °C

A general laboratory cooking temperature profile is shown in Figure 46, in which the lowest line is "air bath autoclave", which is usually used for laboratory cooking. A typical temperature profile in laboratory cooking is as follows: At first temperature is raised by one degree Celsius per minute. Eventually, the set-point temperature is achieved and cooking is continued at this temperature until a certain reaction level of lignin is accomplished.

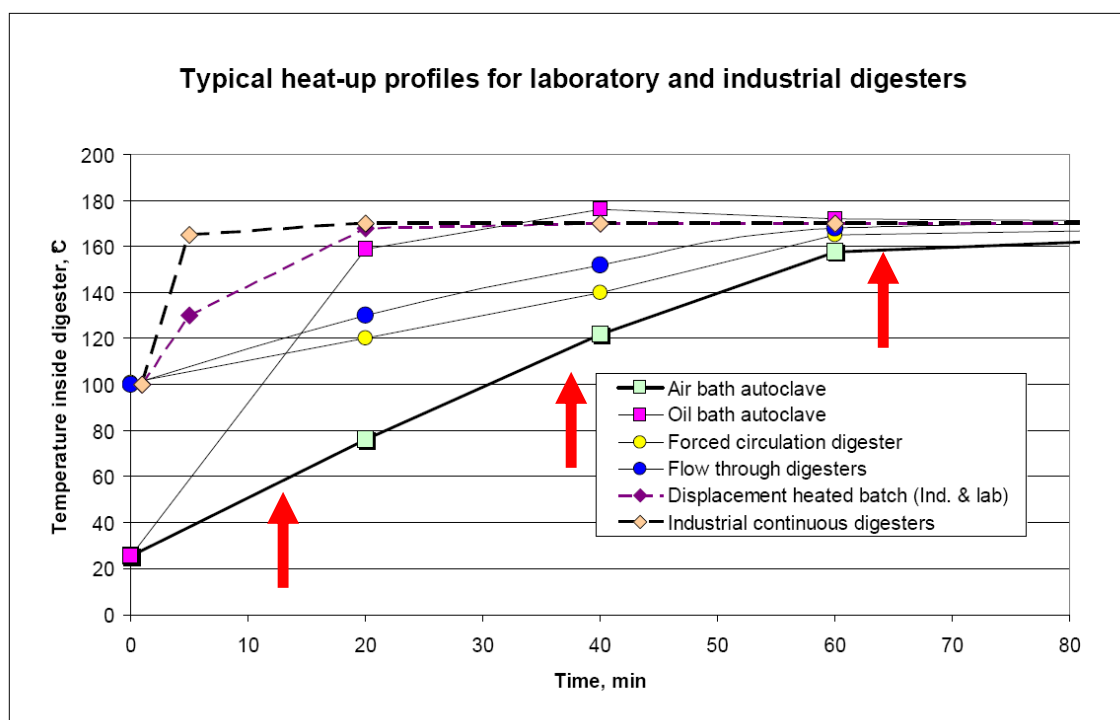


Figure 46 Typical heat-up profiles for laboratory and industrial digesters. Autoclave cooking typically starts at room temperature as there is no pre-steaming. In other cooking methods, the chips are generally heated to approximately 100°C by pre-steaming of chips [89]. Typical laboratory air bath experiment temperature profile indicated by arrows.

During the heat-up period, some lignin reactions occur. Eventually, the temperature is high enough for the lignin reactions to start to accelerate. The transition point from initial stage to bulk stage is said to occur at a lignin level of 22 % on wood. This is the point where the bulk stage of lignin reactions starts.

As a result of the experimental arrangement, the transition point between the initial and bulk delignification stage usually occurs at 140 °C, as shown in Figure 45. If the temperature is maintained at less than 140°C, say 120 °C for more than 5 hours, the lignin content decreases to less than 20 % on wood, but still the lignin reactions are slow. The so-called initial stage continues because the temperature is kept low. The lignin reaction speed in the initial stage is more dependent on temperature than on lignin content.

The residual stage of the cook is not important for kraft cooking modeling. The lignin content is so low that modern digesters do not cook to so low lignin contents. Besides, modern digesters alter the "transition point" from bulk stage to residual stage to lower the lignin content [84].

In modern digesters temperature changes are rapid. Chips can be heated in the impregnation vessels to a temperature of 120 °C and a cooking temperature of 150 °C within minutes. Modern digesters differ a lot from laboratory digesters, in which heat-up can take one to two hours.

Decision

Thus, a new approach in modeling is needed to solve rapid heat-up of modern digesters. The delignification model can be divided into two stages, a low-temperature stage and a cooking stage, where the major cooking reactions take place. If the cooking plant is a two-vessel system, the model can be easily divided into an impregnation stage and a cooking stage in different vessels. In this way the model produces reproducible results. Hence, the Gustafson model becomes a valid choice to model the lignin reactions in continuous kraft cooking.

When building a chip bed packing model, accurate lignin dissolution model is needed, because change in lignin content correlates with the compaction of chip bed. Gustafson's lignin modeling is very well validated and kinetic equations are fitted to Nordic softwood species, mainly pine (*Pinus silvestris*) and spruce (*Picea abies*). At this point, the Andersson model is not suitable for simulating the lignin profile all the way from impregnation to blow line. It is suitable for modeling kraft cooks for low lignin contents. That is one of the reasons, why Gustafson model is more suitable for chip bed packing compared to Andersson model

8.3. *Mass transfer and kraft cooking kinetics*

The Gustafson model's mass transfer and cooking kinetics are discussed in the following sections. The equations are based on two articles: cooking kinetics are fully presented in an article by Gustafson et al. /79/ and the approach to modeling and mass transfer in Agarwal's thesis /90/. The equations, which are used in the TTK Packing Simulator are discussed detail in Sections 8.3-7.1.

8.3.1. Diffusion-reaction phenomena

To get a detailed picture of kraft pulping, mass transfer and cooking reactions have to be known. Most important is to understand the lignin reactions. Lignin keeps the fibers together, so removal of lignin is the main task in kraft cooking. In addition, to be able to determine the lignin reaction rate in different parts of the chip, the concentrations of cooking chemicals inside wood chip need to be known. To find out these concentrations, mass transfer must be modeled accordingly. Carbohydrates, mostly glucomannan and xylan, in wood consume most of the cooking chemicals during kraft cooking. This is why carbohydrates also have to be considered in the cooking model. For this reason, it is important to have good models for lignin and carbohydrate kinetics and for mass transfer when building a kraft cooking model.

The reaction front is visualized in five steps, which are the following:

1. Diffusion of reactant A through a film surrounding the particle to the surface of the solid
2. Penetration and diffusion of A through the moving reaction front into the reaction site
3. Chemical reaction of reactant A with solid
4. Diffusion of the reaction products through the particle to the surface of the particle
5. Diffusion of the reaction products through the film into the surrounding liquid

The following section explains how the chip is transformed into spherical coordinates and how the mass transfer is done according to Agarwal's thesis, in which the Gustafson model is modified /90/.

8.3.2. Changing chip dimensions into spherical coordinates

Mass transfer into chip was simplified by changing the three-dimensional slab model can be into a spherical ball model. This spherical model makes the mass transfer model easier to solve mathematically. Chip thickness is the smallest of the three chip dimensions. In the Gustafson pulping model it is assumed that chip thickness is the most critical dimension /79/. Wood chips are treated in the model as a sphere. The radius of the sphere is chosen so that the sphere has the same surface-area-to-volume ratio as the actual chip /90/. In this way, the three-dimensional chip is transformed into a one-dimensional sphere. A sphere is mathematically easier to solve than a three-dimensional chip.

The three dimensions of chips, i.e., thickness (T_C), length (L_C) and width (W_C) (*Figure 47*), are changed into spherical coordinates. The coordinates are calculated from the following *Equation 14*:

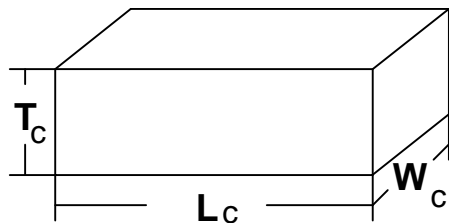


Figure 47 Chip dimensions in terms of length, thickness and width.

$$\frac{R_b}{3} = \frac{LW_C T_C}{2(LW_C + W_C T_C + L T_C)} \quad \text{Equation 14}$$

Where:

R_b is ball radius (mm)

L' is effective length ($L' = L_c / \sqrt{3}$) because diffusion is three times greater in length direction than in thickness direction /91/.

W_c is width of the chip (mm)

T_c is thickness of the chip (mm)

While the assumption of a spherical chip may not be reasonable from a geometrical perspective, a sphere approximates the wood chip quite well from a diffusional standpoint /90/. This has been shown earlier by Neretnieks /91/ in his study on washing of chips.

8.3.3. Diffusion

Alkaline diffusion into wood is dependent on the alkali concentration and lignin content:

$$D_{[OH^-]} = 0,33T^{0,5}e^{-2456/T} \left[-0,02L + 0,13[OH^-]^{0,55} + 0,58 \right] \quad \text{Equation 15}$$

$D_{[OH^-]}$ is diffusion coefficient (cm²/min)

The diffusion coefficient increases at the same time as the wood material dissolves during cooking. It is assumed that lignin and carbohydrates are dissolving proportionally. The change in lignin content takes this variable into account. The swelling of wood is taken into account by the hydroxide ion concentration.

Unsteady state reaction-diffusion in a spherical body is represented by the following partial differential equation (PDE) (Equation 16):

$$\frac{\partial K}{\partial t} = \frac{1}{r^2} \frac{\partial}{\partial r} \left\| Dr^2 \frac{\partial K}{\partial r} \right\| - R_k(K) \quad \text{Equation 16}$$

Where

K is concentration of reactive species (mol/l)

t is time (s)

r is radial position of sphere (mm)

D is diffusion co-efficient

R_k is change due chemical rate

Boundary conditions are solved in the following manner:

At the center of the chip no slope condition is applied:

$$\left. \frac{\partial K}{\partial r} \right|_{r=0} = 0 \quad \text{Equation 17}$$

Concentration flux at the chip surface is balanced by mass transfer across the liquor film

$$D \left. \frac{\partial K}{\partial r} \right|_{r=R} = k_m (K_{liq} - K_R) \quad \text{Equation 18}$$

Where

k_m is mass transfer coefficient

The change in cooking liquor concentration rate inside the chip is presented in *Equation 19*. More detailed derivation of the equation is shown in Ming's dissertation /92/. For pulping liquor outside the chips, the mass balance between the chips' surface and free liquor gives the following equation:

$$\frac{dK_{liq}}{dt} = -D \left. \frac{\partial K}{\partial r} \right|_{r=R} \left[\frac{3V_c}{RV_l} \right] \quad \text{Equation 19}$$

Where

V_c is chips' free liquor volume

V_l is the volume of the surrounding liquor outside the chips

The amount of surrounding cooking liquor is calculated by subtracting the absorbed amount of liquor from the total amount of liquor. For the alkali concentration, the following equation is obtained:

$$\frac{d[OH^-]_{liq}}{dt} = -D_{[OH]} \left. \frac{\partial [OH^-]}{\partial r} \right|_{r=R} \left[\frac{3V_c}{R_k V_{liq}} \right] \quad \text{Equation 20}$$

8.3.4. Lignin reaction rate

The lignin reaction rate depends on the lignin content of the wood. In Gustafson's model, the lignin reactions are divided into three stages. The initial stage at the beginning of the cook is followed by the bulk stage and, at the lower level of lignin, by the residual stage. The different stages are presented in *Figure 48*.

The initial delignification stage is based on *Equation 21*. When the lignin content reaches a certain level (transition point 22.5 % lignin on wood /79/), the bulk delignification equation, *Equation 22*, is used. At a lignin content less than 2.2 % on wood, the residual stage *Equation 23* is used.

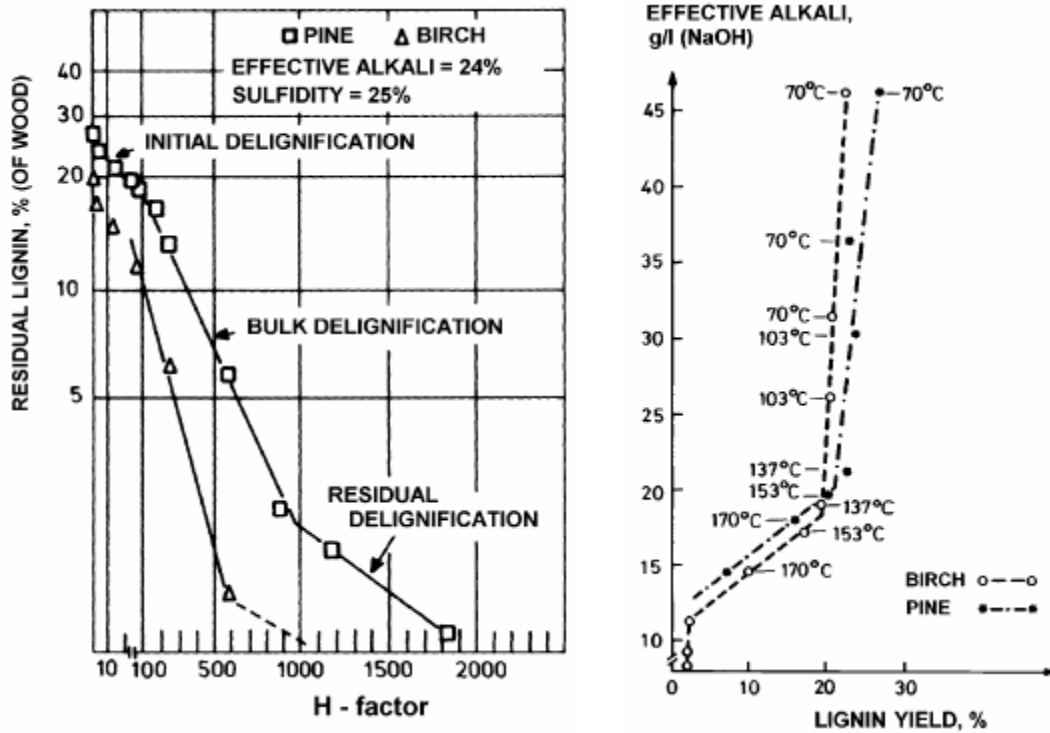


Figure 48 Dissolution of lignin as a proportion of H-factor in pine and birch kraft cooking (left) and alkali consumption with lignin yield (right) /80/.

The reaction rate for delignification at different stages of the cook can be calculated from the following equations:

Initial stage ($L > 22.5\%$):

$$-\frac{dL}{dt} = 36,2T^{0,5}e^{\frac{-4807,69}{T}}L \quad \text{Equation 21}$$

Bulk stage ($2.2\% < L < 22.5\%$):

$$-\frac{dL}{dt} = \left[e^{\frac{35,19 - 17200}{T}}[OH^-] + e^{\frac{29,33 - 14400}{T}}[OH^-]^{0,5}[HS^-]^{0,4} \right] L \quad \text{Equation 22}$$

Residual stage ($L < 2.2\%$):

$$-\frac{dL}{dt} = e^{\frac{19,64 - 10804}{T}}[OH^-]^{0,7}L \quad \text{Equation 23}$$

8.3.5. Carbohydrate reaction rates

In the Gustafson model, the carbohydrate reaction in each delignification stage is a linear function of the lignin reaction rate:

Initial stage:

$$-\frac{dC}{dt} = 2,53 \frac{dL}{dt} [OH^-]^{0,11} \quad \text{Equation 24}$$

Bulk stage:

$$-\frac{dC}{dt} = 0,47 \frac{dL}{dt} \quad \text{Equation 25}$$

Residual stage:

$$-\frac{dC}{dt} = 2,19 \frac{dL}{dt} \quad \text{Equation 26}$$

Lignin and carbohydrate reactions are dependent on the alkali concentration and temperature through the reaction-diffusion process.

8.3.6. Alkali reaction rate

The rate of alkali consumption depends on the lignin and carbohydrate reaction rates:

$$R_{[OH^-]} = \left[0,045 \frac{dL}{dt} + 0,09 \frac{dC_c}{dt} \right] \frac{\rho_{dc}}{\epsilon} \quad \text{Equation 27}$$

Where

ρ_{dc} is density of wood (kg/m³)

ϵ is volume inside the chip

L is lignin content (% on wood)

C_c is carbohydrate content (% on wood)

Hydroxide ions which react during the cooking reactions, dissolves organic matter and becomes water. During the impregnation stage, wood extractives and acetyl groups consume alkali rapidly, and the alkali consumption is higher in the early stage of cooking. Therefore, the alkali consumption is 5-fold at the beginning of the cooking stage, when the first percentage of lignin is

reacted. This alkali consumption can be understood as the initial consumption of alkali.

The free liquor volume is calculated from the assumption that the density of the wood material inside the wood matrix is 1.5 kg/m³.

$$\varepsilon_l = 1 - \frac{\rho_{dc}}{1,5} \quad \text{Equation 28}$$

9. Conclusion of literature review

The literature part begins with a short description of the development of digesters from 1880's to modern continuous digesters. Continuous cooking became the dominant design for new cooking plants in the late 1960's and early 1970's, because the process was easier to operate, investment costs were lower and digester washing was better compared to batch cooking. The importance of digester washing in new continuous digesters has decreased, because new digesters are delivered mainly for greenfield investments, where increased washing capacity is taken into account in the brown stock area.

Continuous digester production capacity has been increasing over the years, because economy of scale favors large production units especially in capital intensive industries. The capacity of the digester has increased by increasing the digester diameter. This sets new challenges for controlling the flows in the digester to get uniform quality. To get a more accurate picture about the process conditions inside the digester models should be used, because digester process measurements do not give sufficient amount of information about packing and flow conditions. Typically only liquor flow, cooking chemical charge and temperature are parameters, which are measured and controlled in pulp digester.

This thesis gives information about chip bed packing in continuous digester. To gain better understanding about chip bed packing, packing has to be modeled on chip bed scale. In order to model the packing phenomenon on chip bed scale, chip softening has to be known. Chip softening depends on delignification of the chip. This has to be modeled at chip scale. To get accurate picture about delignification reactions on chip scale, mass transfer and cooking kinetics has to be known.

Several kraft cooking kinetic models are covered in *Sections 8.2.2-8.2.5*. Based on existing information kinetic model was chosen. This is discussed in more detail *Section 8.2.5*. Only one suitable packing model was found in the literature, which was chosen. Based on these decisions, a packing simulator was built, which is covered in *Chapters* later on.

10. The TKK Packing Simulator

The packing simulator for modeling chip bed was developed to achieve the general understanding about conditions inside the continuous digester. In the following sections, the simulator is described in detail.

The packing simulator was built to simulate the chip bed reactions in continuous kraft pulp digesters. The simulator combines the chip scale Gustafson and chip bed scale Härkönen models. The models are presented in more detail in Sections 8.3 and 7.1. The Gustafson model handles the mass transfer and kinetic equations of the simulator and the Härkönen model is used for packing simulation. The foundation of the model is presented in *Figure 49*.

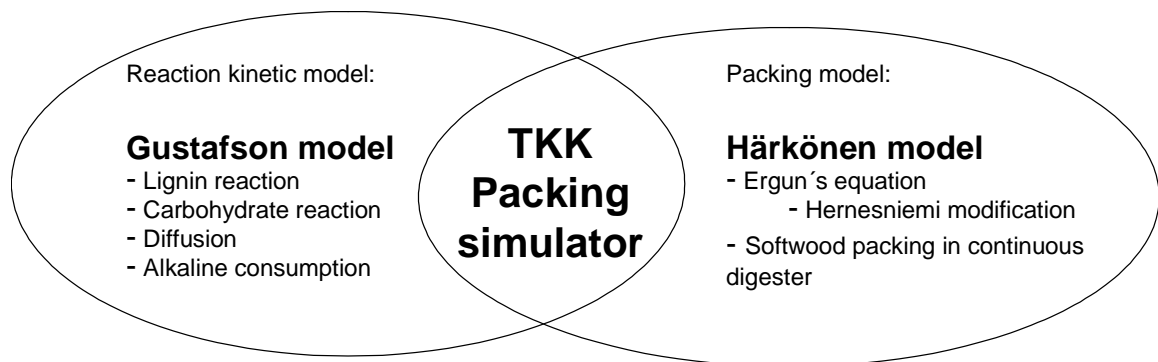


Figure 49 Foundation of the TKK Packing simulator.

The reaction kinetics model consists of cooking reactions and mass transfer. The packing model is based on a chip bed packing model, which Hernesniemi /43/ has modified. Hernesniemi's modifications concern the Ergun equation, where the R1 and R2 values were changed from the original values set by Härkönen /39/.

The *TKK Packing Simulator* was designed using the MATLAB software. Differential equations in the simulator were solved using the ODE15s solver, which is used for solving stiff problems for ordinary differential equations.

10.1. Model assumptions

Modeling of a chip bed inside the continuous digester is a difficult task. This study concentrates on studying the chip beds only in the digester. The chip bed is less compacted in the impregnation vessel, because cooking reactions have not been advanced much. Radial and vertical liquor flows affect the chip bed packing. Vertical flow is more dominant for chip bed packing. Chip heating takes place through liquor circulation and heat of reaction. Heat of the reaction is taken account by increasing the temperature in certain digester blocks, where this phenomenon should be take account.

The following assumptions have been made:

1. The chip bed is modeled only in vertical direction. The packing simulator does not simulate the digester in radial direction. The vertical direction is the most important direction in chip bed packing.
2. The heat of the reaction is not included in the simulator. At each simulation block, the user sets the value for the temperature.
3. The kraft cooking kinetic model is used only for modeling the softwood reactions.
4. The diffusion of reaction products is left out from the model. Reaction products do not affect the packing model, so at this stage of model development, this phenomenon is not included.
5. The hydrosulfide concentration in the chip bed is constant. In other words, there is no consumption of HS⁻-ions. The real concentration of HS⁻ ions is difficult to measure and there is no general kinetic model for calculating HS⁻-ion reactions. At this stage, the concentration is kept constant.
6. Extractives are removed before cooking. Gustafson /79/ assumes that all extractives are removed before the wood chips enter the digester. The same assumption is made in the simulator.
7. Chips are fully penetrated with liquor and no air is left inside the chips before cooking. Mass transfer of chemicals into chips occurs by diffusion only.
8. Mass transfer and chemical kinetics are the rate-determining steps of this model. Mass transfer and reaction kinetics affect the physical properties of the chip bed.
9. The porosity of the chip bed in the simulator is limited, so it cannot go under 0.1. As a result, the calculations become more stable. Also, it is physically impossible to force out all the liquor from the wood chips.
10. The maximum chip bed pressure is about 20-25 kPa. If the simulator results exceed this value, the packing conditions are too severe to be realistic. The chip bed cannot bear that kind of load, and pressures at this level are impossible in real life.

10.2. Structure of the simulator

10.2.1. General structure

The Packing Simulator is divided into different levels of information. On the reactor scale, the simulator is divided into different parts, called blocks. There are nine different types of blocks available in the simulator. These blocks are used for setting liquor flows in different parts of the digester. Blocks are used for simulating liquor additions and removals in combinations with different liquor flows. In the simulator the chips are always flowing downwards. The blocks are also used for dividing the digester into different diameters and heights. These different blocks are shown below in *Figure 50*.

Block type *Flow direction*

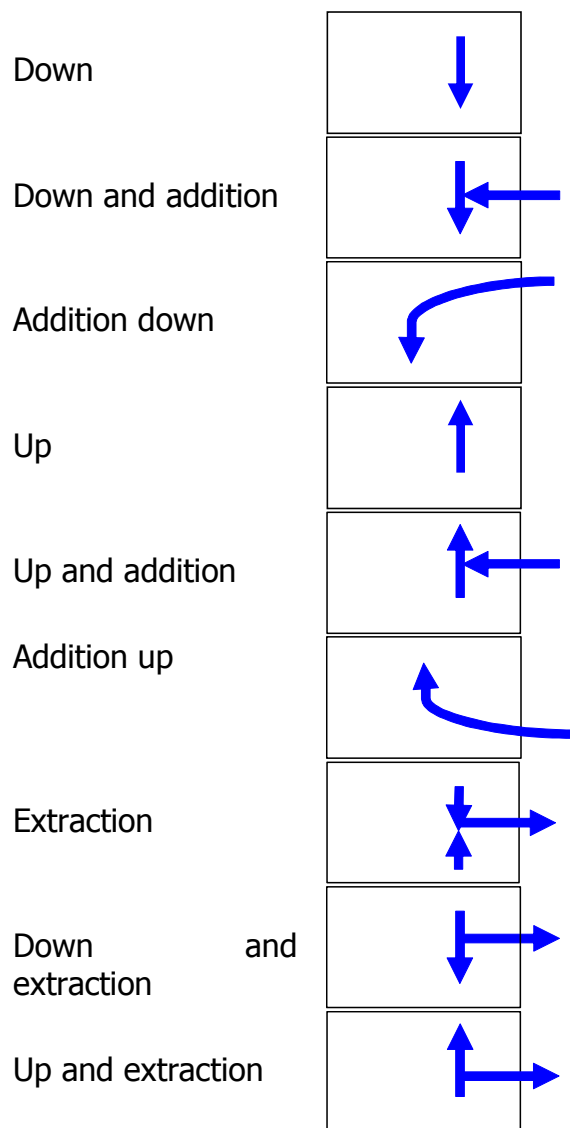


Figure 50 Different types of liquor flows in the blocks of the TKK Packing Simulator.

Examples of the digester and simulation blocks are given in *Figure 51*. The blocks are used to simulate the mass balance of the chip bed.

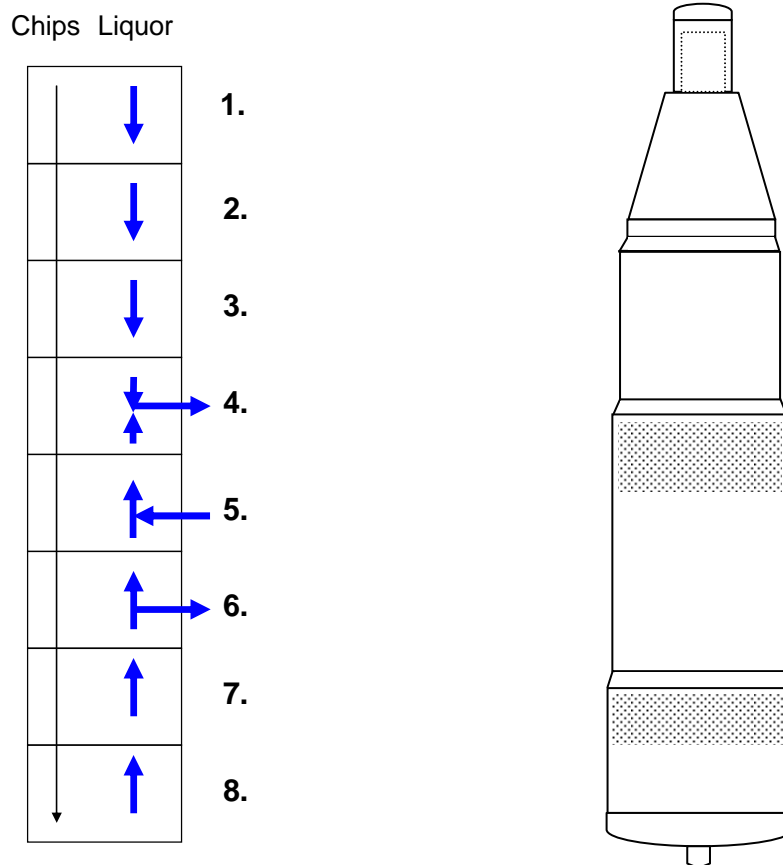


Figure 51 Block diagram of the digester (left figure) and schematic picture of the digester.

The digester is divided into 8 blocks in *Figure 51*. The figure shows an upflow-type (Counter-Current) Compact Cooking digester. Chips and liquor are added in *block 1* and the liquor is extracted in *block 4*. This first extraction position is called the extraction screen. After the extraction screen, the liquor flows in opposite direction relative to the chips. Washing liquor is added from the bottom of the digester (*block 8*). Part of the liquor is extracted in *block 6*. This second extraction screen is called the ITC screen (IsoThermalCooking). Part of the extracted liquor (from *block 6*) is recirculated back into *block 5*. Liquors flowing in opposite directions are extracted in *block 4*.

10.3. Simulator setup

A simplified simulation calculation order is shown in *Figure 52*. First, the length of the *Time step* is defined via the initial packing of the chip bed. The length of the time step is the retention time in one simulation step as defined in Section 10.4.6. Then the *Mass transfer* equations are solved for the given *Time step*. Chemical concentrations are thus derived for the different parts of the wood chips and surrounding liquor. This information is used for solving *Reactions* of wood components and cooking chemicals. After the reaction rate

is solved, *Packing* is solved. Based on the packing calculation the length of the time step in the next simulation step is calculated.

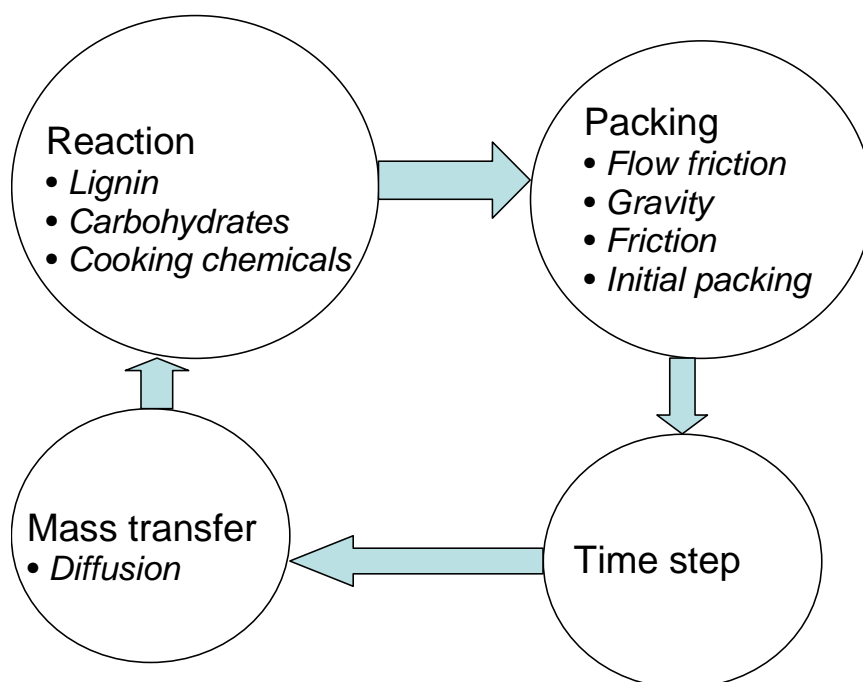


Figure 52 TTK Packing Simulator calculation order.

10.3.1. Impregnation

Chip bed packing can be simulated with or without impregnation. Impregnation means that the chips are impregnated, i.e. the voids of the fibers are filled with cooking liquor in a separate impregnation vessel before the digester.

Before the digester simulation studies, the initial conditions when chips enter the digester are modeled. In this way, the initial conditions of the digester, i.e. the lignin and carbohydrate contents, are solved. These values are used in chip bed packing simulation.

An impregnation block can be used for impregnation studies. The impregnation model uses the Gustafson kinetic and mass transfer equations. The only difference compared to the digester packing simulation is that Härkönen's packing equations are not used. Chip bed packing is not an issue in the impregnation vessel and chip softening is very small.

All simulations in this thesis were done with impregnation, even though the impregnation results are not presented here in detail. The focus is on packing phenomena inside the digester, so a more detailed presentation of chip packing inside the impregnation vessel is irrelevant.

Impregnation is used throughout the simulations in this work. Conditions are the following: impregnation time 50 minutes, impregnation temperature 130 °C, $[\text{OH}^-]$ is 0.5 mol/l, $[\text{HS}^-]$ is 0.2 mol/l, lignin content 28 % on wood and carbohydrate content 68 % on wood.

10.4. Simulator parameters

The initial parameters for setting up the simulation are shown in *Figure 53*. These values are used for the chip bed's initial conditions. Underneath, there is a short explanation of each value.

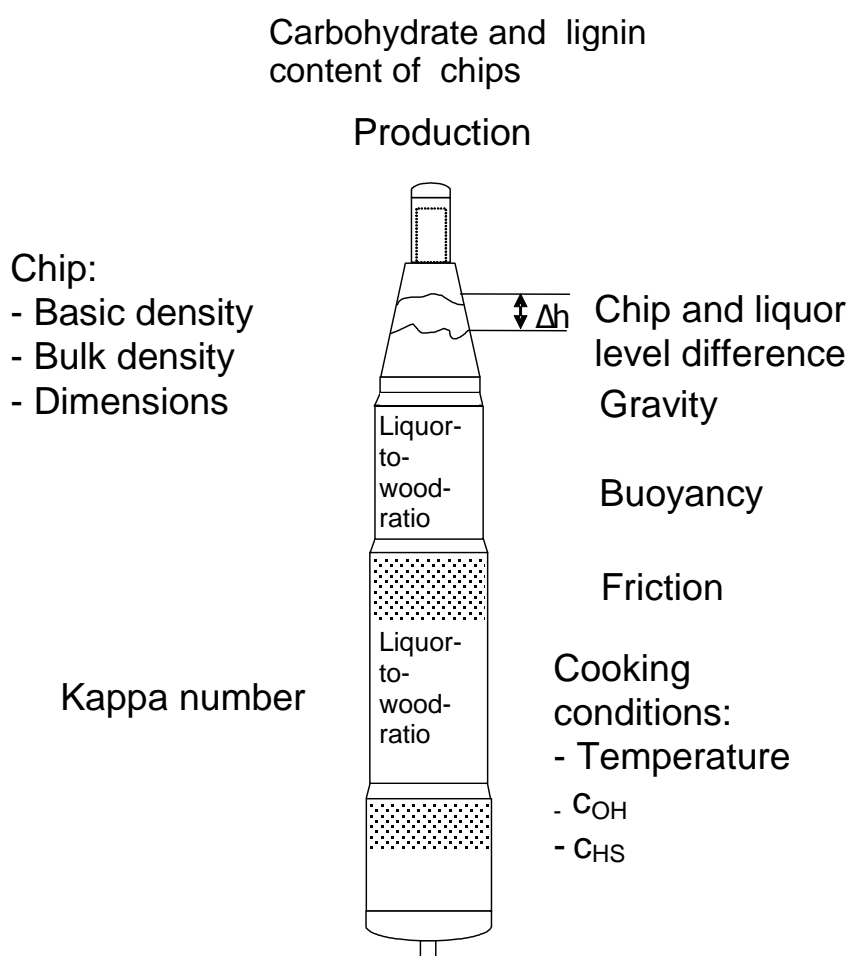


Figure 53 Initial parameters for modeling chip bed packing.

10.4.1. Chip properties

The chip material properties used in the simulation are: basic density, chip bulk density and chip dimensions.

Basic density (ρ_c) is the density of wood material. It varies for softwood from 350 to 500 kg/m³.

Chip bulk density (ρ_{bulk}) is the free density of the pile of chips. The bulk density of dry chips varies from 130 to 185 kg/m³.

Chip dimensions are **length** (L_C), **width** (W_C) and -- the most important dimension -- **thickness** (T_C). A more detailed presentation of chip thickness is given in *Chapter 8.3.2*. Chip density affects the packing of the chip bed and chip dimensions affect the mass transfer during cooking. The further the cooking advances inside the digester, the softer the chips become. This increases the packing of the chip bed.

10.4.2. Chip and liquor level difference

The **chip and liquor level difference** is important in steam-liquor-phase digesters. The level difference affects the initial packing of the chip bed. This level difference acts as a force pushing the chip bed downwards, as the chip level above the liquor level is increased. The level difference is used to heat up the chips in steam-liquor-phase digesters. In the top part of the digester, chips are heated from impregnation temperature to cooking temperature in a short time, approximately one to two minutes. A certain level difference should be maintained to guarantee effective heating in the digester, but if the heating zone grows the downward force in the chip bed and, accordingly, chip bed packing, will increase.

10.4.3. Forces affecting the chip bed

Buoyancy and **flow friction** are forces acting in the same direction in chip bed packing when liquor is flowing counter-currently against the chip flow. When liquor is flowing in the same direction as the chips, these forces act as counter-forces. Buoyancy provides uplift for the chip bed. The liquor flow either pushes the chip bed downwards or provides uplift for the bed. The packing in the chip bed is usually densest where two opposite liquor flows meet.

In *Figure 51*, the highest packing is in block 4. Liquor flows downwards from block 3. This flow produces a downward force. Liquor flows from block 5 and this flow produces an upward force. These forces meet in block 4, where the packing is densest. At the same time, the lignin content in the chips is decreased, making the chips softer. This phenomenon increases the packing effect. For detailed modeling, the liquor flows are more important than chip softening through lignin removal.

Gravity acts as a downward force in the chip bed. In a stable, non-moving chip bed, the packing degree is highest in the bottom of the chip bed. But when the chips and liquor start to move and the cooking degree is taken into

account, the packing might be densest in the area where two of the biggest flows meet.

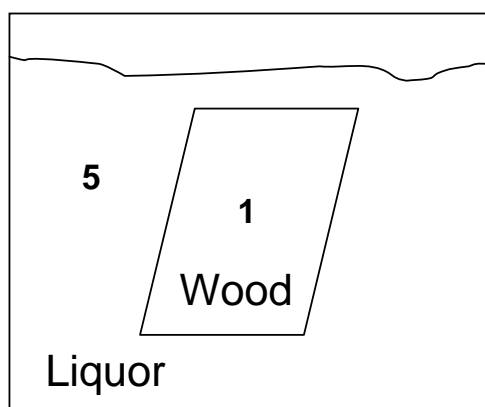
Friction affects the chip flow inside the digester. The real friction inside the digester is unknown, but according to Härkönen, in a digester wall section the coefficient of friction can be assumed to be 0.01 and at the screen section 0.02 /39/. Lacking better knowledge of the real phenomena determining the friction force inside the digester, the same values for friction are used in the present work.

10.4.4. Liquor-to-wood ratio

The **liquor-to-wood ratio (LW)** affects the liquor amount inside the digester. The liquor-to-wood ratio is a common factor used to control the liquor flow in a cooking plant.

The liquor-to-wood ratio is not unambiguous term. In literature, the LW term is discussed at least two different ways. Those terms are packing LW and feed LW. Packing LW is used in used commonly at batch cooking and feed LW is used at continuous cooking. These two methods are illustrated in *Figure 54*. In both examples, LW is 5:1. This means that there are 5 volume units of liquor in every 1 weight unit of wood (l/kg or dm³/kg). In packing LW, liquor and wood are at the same volume all the time. In feed LW, liquor flow in is 5 times the wood flow in. In this study, the liquor-to-wood ratio is understood as a feed LW.

Packing liquor-to-wood ratio



Feed liquor-to-wood ratio

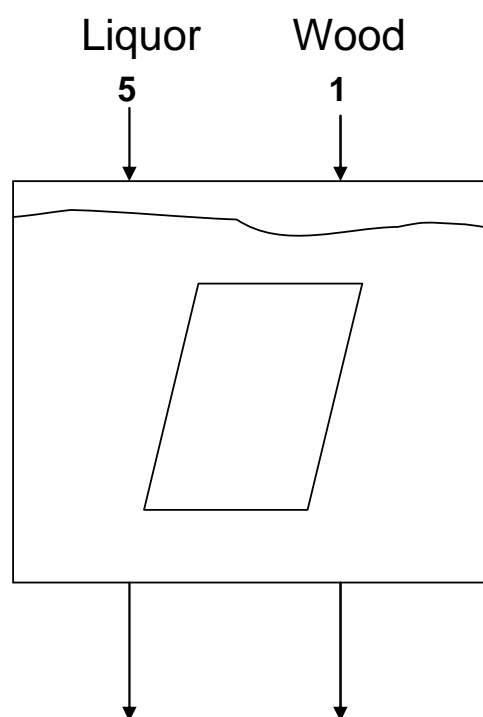


Figure 54 Packing and feed liquor-to-wood ratio.

10.4.5. Cooking conditions

Cooking conditions, such as cooking temperature and the concentration of cooking liquor affect the kraft cooking reaction rate, and consequently also the lignin content of the wood chips. The change in lignin content causes softening of the chips, which subsequently affects the packing of the digester.

10.4.6. Simulator step size

The chip bed height is divided into simulation steps of even length. The number of simulation steps affects the accuracy of the simulator. Usually about 500 simulation steps are needed to get accurate simulation results with the TKK Packing Simulator. This step size is also used in the following simulations. The modeling accuracy at different simulation steps is shown in *Figure 55*.

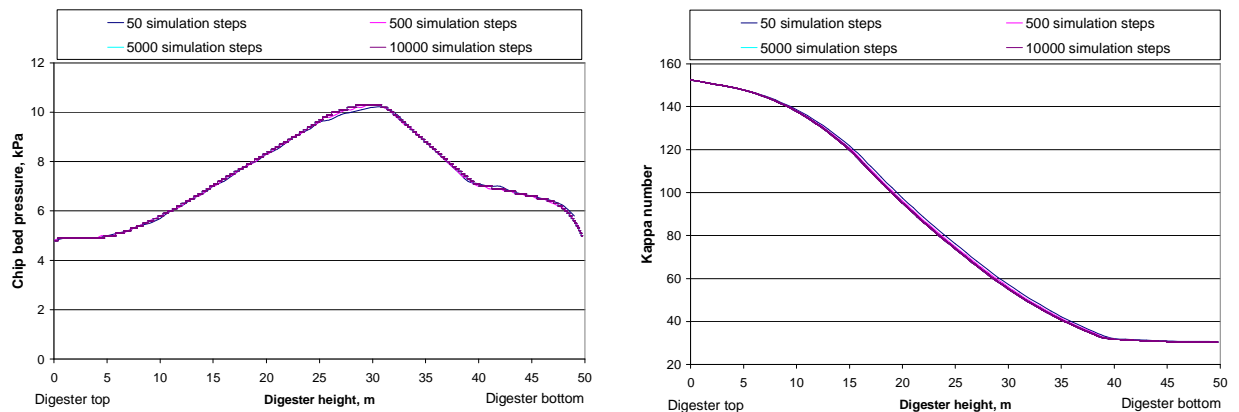


Figure 55 Accuracy of the TKK Packing Simulator in simulating the packing of the chip bed with normal packing conditions. Liquor-to-wood ratio is 5 and chip and liquor level difference is 1 m.

If the digester height is 50 meters and the simulation step size is 0.1 m, there are 500 calculation rounds within the simulation. This has often been found to be the optimal step size for solving simulation problems, when taking into account the simulation time and accuracy of the simulation result.

10.4.7. Simulation inputs

In the examples in this thesis, counter-current cooking of softwood chips is simulated. The input parameters for these basic "default" cases are found in *Table 5*. These are the values used, unless otherwise specified.

Table 5 Default input parameters for digester simulation.

Input	Value	Units
Basic density of chip	400	kg/m ³
Bulk density of chip	160	kg/m ³
Density of wood	1500	kg/m ³
Production	2000	ADt/day
Feed liquor-to-wood ratio, digester top	5	l/kg
Gravity	9.81	m/s ²
Chip and liquor level difference	1	m
Cooking liquor density	1100	kg/m ³
Washing liquor density	1080	kg/m ³
Dilution factor	2.4	m ³ /ton of pulp
Blow flow consistency	10	%
Friction coefficient (wall)	0.01	
Friction coefficient (screen section)	0.02	
Ergun equation, parameter R1	2500	
Ergun equation, parameter R2	2.125*10 ⁶	
OH ⁻ -concentration in the beginning of cook, mol/l	1.2	
HS ⁻ -concentration in the beginning of cook, mol/l	0.2	
Chip length	25	mm
Chip width	15	mm
Chip thickness	5	mm
Simulator step size	0,1	m

The cooking temperature profile for the “default” case is shown in *Figure 56*. Temperature rises gradually because there is some heat from the reaction. After the first extraction screen there is a counter-current cooking zone. In this section of the digester liquor extracted from the second extraction screen is heated in a heat exchanger and returned to the digester through the center pipe above the second extraction screen. Cooking liquor is heated to the same temperature that it has at the first extraction screen. These temperature measurements are based on real temperature measurements in the digester.

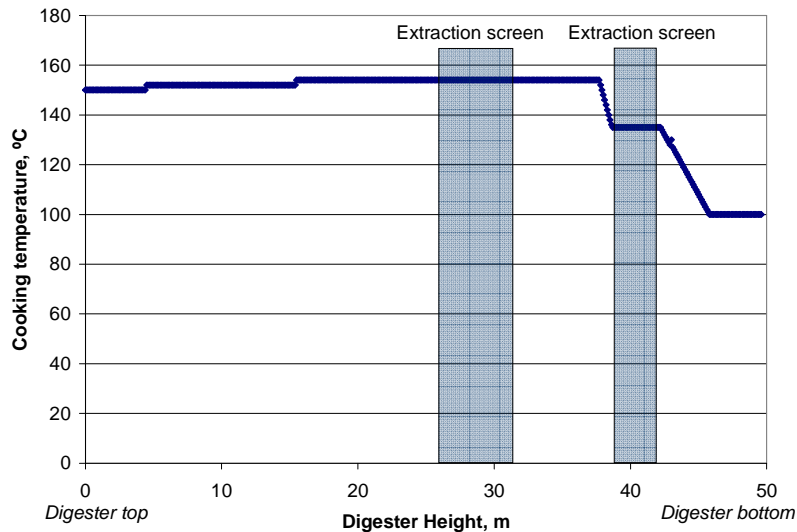


Figure 56 Cooking temperature in normal counter-current cooking simulation.

10.5. Overview of the TKK packing simulator

The TKK packing simulator was developed based on the Gustafson reaction-kinetic model and Härkönen packing model, which has been modified by Hernesniemi. The equations, which the model is based on, have been shown in the previous sections. Matlab 7.4 has been used for programming the simulator. The Matlab code was developed by Juha Päivärinta.

The simulator is run using an interface, which communicates with Matlab. The main interface is shown in *Figure 57*. Input data of process variables is set up by this interface. The input variables are discussed in more detail in *Section 10.4.7*.

Parameter	Value
Basic density	400
Bulk density	160
Wood density	1.52
Production	2000
Yield t=0	80
Yield	48
H-factor	1200
Init kappa #	150
Liquor-to-wood	5
Gravity	9.81
Chip level	1
BL density	1100
Wash I. density	1080
Screen temp	150
Yield at screen:	55
Temp. Displ.	2.5
Dil. Fact.:	2.4
Blow cons.	10
Ergun 1	2500
Ergun 2	2125000
Tot. Alkali	19
WL conc	120
WL to impr.	6
WL to cook	11.5
No. of blocks	8
Block	Block1

Buttons: Close, Der. vals, Simulate

Figure 57 TTK packing simulator interface.

The simulator has two flow examples of digester, upflow Compact Cooking and downflow Compact Cooking. The default example is upflow CC-digester. Comparison between these two cooking methods is discussed in *Section 11.2.6*. After each simulation, the results are visualized. The window for visualization is shown in *Figure 58*. Based on selection in visualization, graphs of the results are drawn. That is shown in *Figure 59*.

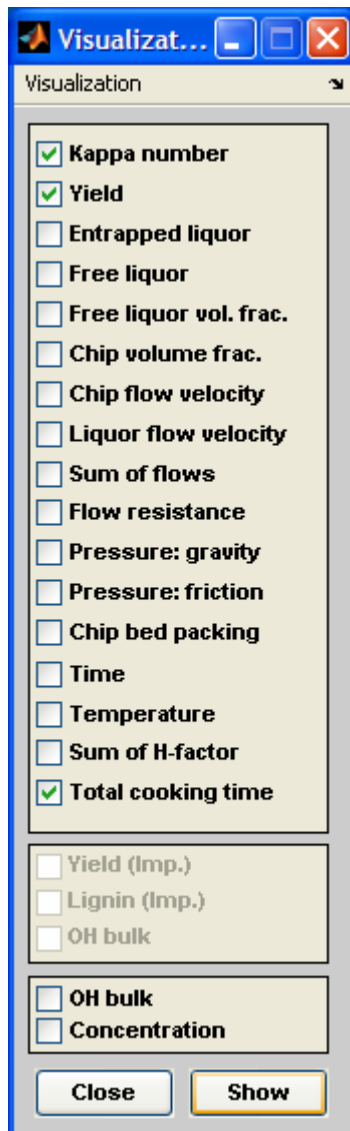


Figure 58 Window for selecting different parameters for visualization.

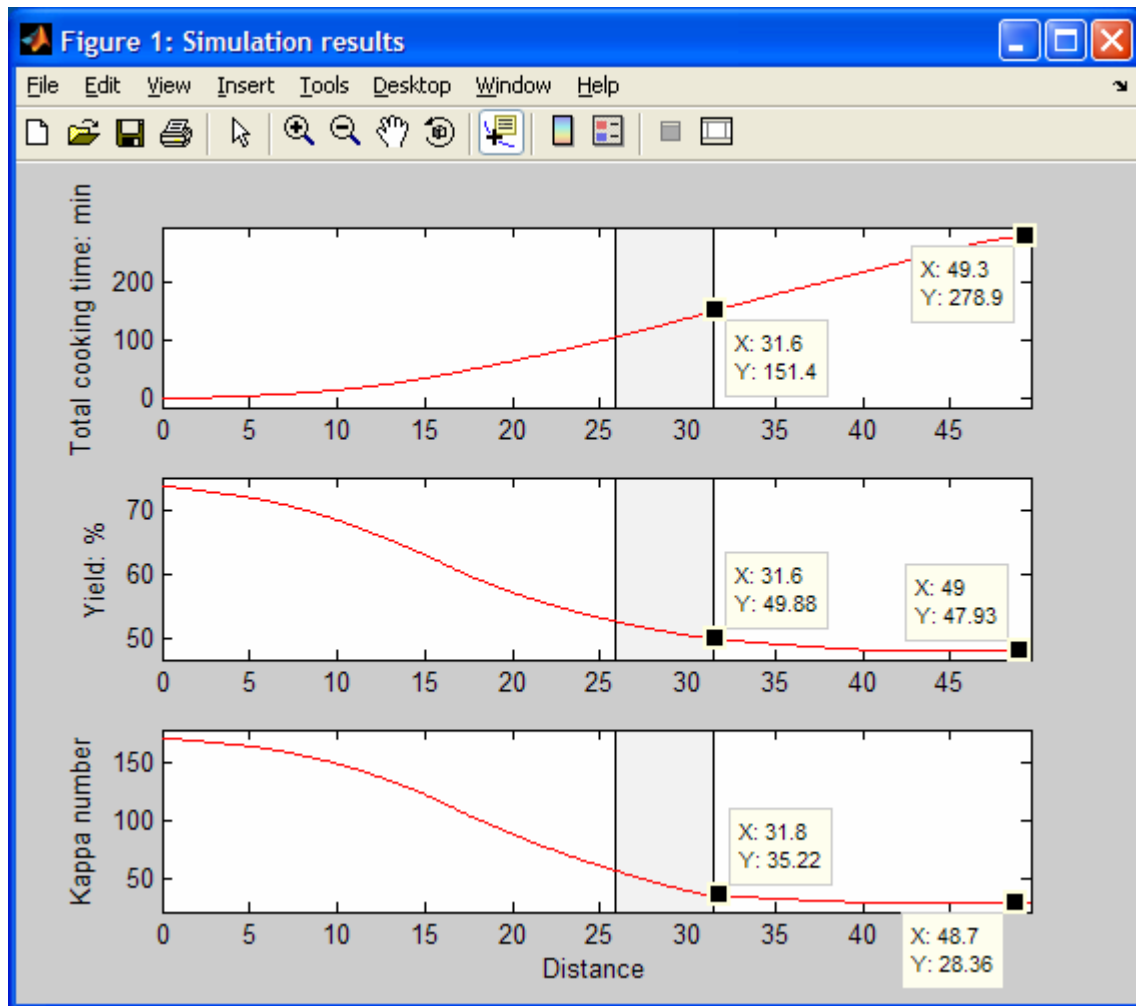


Figure 59 Simulation results for TKK Packing simulator. The shaded area is extraction screen area of the digester. The distance scale starts from digester top.

Simulation results may be examined in Matlab, but simulation results can be exported from Matlab to Excel, where further analyzes may be done. Next chapter concentrates on further discussion about the simulator and simulation results.

11. Simulation results

The following chapter is divided into two sections. The first is a general presentation of the simulation results. The second section shows how different variables affect the chip bed packing and other output parameters, assuming realistic conditions. The default simulator input values are shown in *Table 5*.

11.1. Presentation of output values of the simulator

The TTK Packing Simulator can be used to simulate continuous digester cooking conditions. The following parameters can be solved from the simulation of the Packing Simulator:

- Lignin profile in digester
- Lignin profile in chip
 - Uniformity of cook
- Total yield
- Chip and free liquor volume fraction (porosity)
- Chip and liquor velocity in vertical direction inside the digester
- Chip bed packing
- Chips delay inside the digester
- Cooking chemical concentration in surrounding liquor and inside the chip

Lignin and total yield describe how cooking proceeds in the chip bed. Porosity and packing force illustrate the chip bed's hydrodynamic properties and the quality of the pulp produced. Quality wise, there is no clear upper limit for packing force which should not be exceed, because the chips tolerate different loadings in different parts of the digester. Packing force and compressibility depend on the kappa number of the chips, i.e. their lignin content. The best way to examine the packing conditions in the chip bed is to combine information concerning the packing and porosity of the chip bed.

General results of packing simulations are shown in *Figure 60-Figure 64*, below. Lignin yield and cooking yield are the most common results to solve from the packing simulation. Lignin yield can be compared to real mill data to see how accurate the delignification model is. *Figure 60* shows the lignin yield from a normal counter-current cooking simulation.

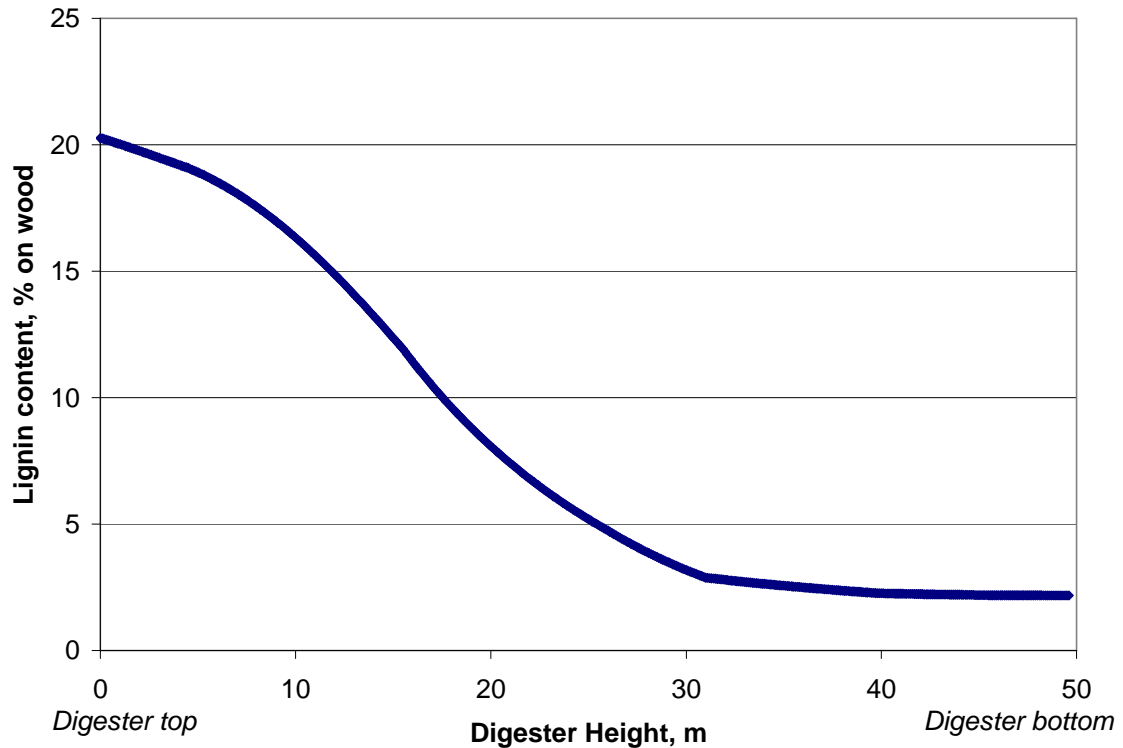


Figure 60 Lignin content in default counter-current continuous digester simulation.

Chip bed pressure in a default counter-current cooking simulation is shown in *Figure 61*. The pressure of the chip bed is highest in the first extraction screen. The pressure is at its highest, because down-flowing cooking liquor and counter-currently flowing cooking liquor meet in the extraction screen.

There is already some pressure in the bed at the top of the digester, because chip liquor level difference is causing force downwards. In the simulation, top of the digester means the zero level of the cooking liquor. That is the reason why in top of digester the chip bed pressure is already 5 kPa.

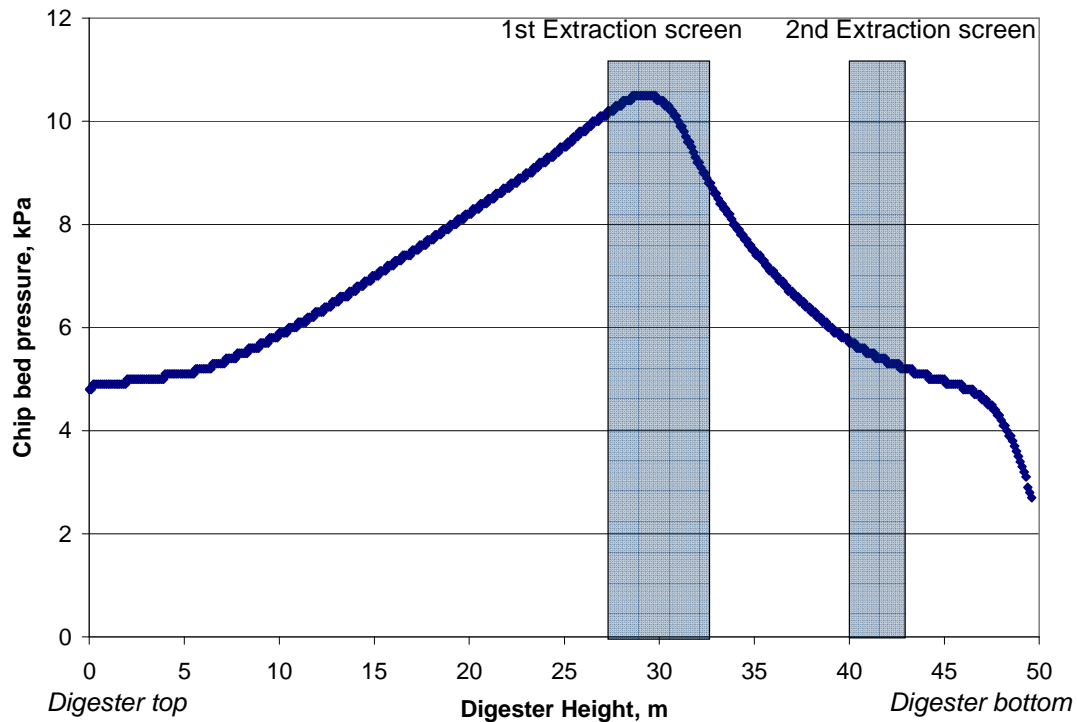


Figure 61 Chip bed pressure in default counter-current digester simulation. Shaded areas are the screen sections.

The porosity of the chip bed is shown in *Figure 62*. At the beginning, the porosity is about 0.6 with and decreases towards minimum of 0.3. Porosity changes affect the compaction of the wood material. When the chip bed's porosity decreases under 0.2, the pulp's uniformity and quality will suffer from the increased compaction. However, Joutsimo /56/ showed that a decrease in initial porosity decreased the blow-line standard deviation of the kappa number. A reduction in initial porosity in a continuous digester affects the chip bed in the same way, but if the initial porosity is too small, this will lead to too dense packing of the chip bed during cooking. Initial porosity of chip bed should be closer to 0.5 to minimize the short cutting of the cooking liquor flow.

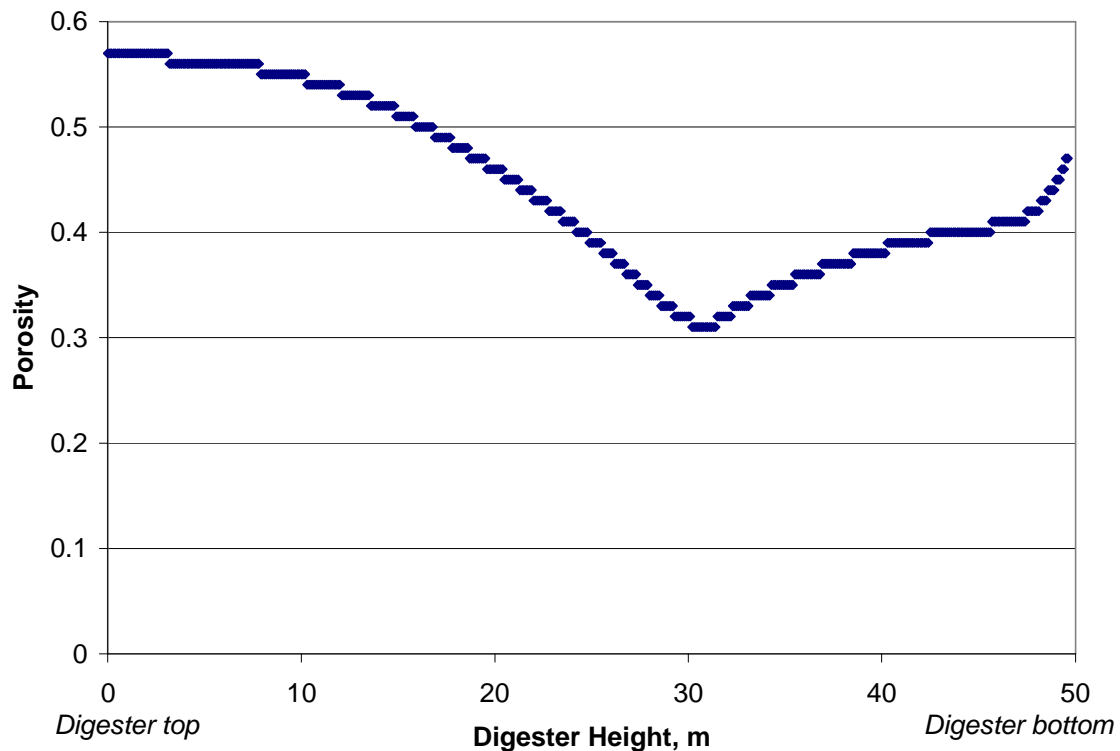


Figure 62 Porosity (free liquor volume fraction) in default counter-current digester simulation.

Chip and cooking liquor superficial velocities (see section 7.3) are shown in *Figure 63*. Chip and cooking liquor are flowing at about the same speed at the top of the digester. After 15 meters, the liquor velocity increases as the chip velocity decreases. At this point, the porosity of the chip bed starts gradually to decrease due to chip bed packing. The volume occupied by the cooking liquor decreases, which leads to a shorter retention time for liquor at a certain level of the chip bed. Negative values in cooking liquor flow mean that the cooking liquor is flowing upwards relative to the digester. Near the bottom of the digester, the flows start to increase again. The reason for this speed change is that the digester diameter decreases and both liquor and chips must flow faster due to this volumetric change.

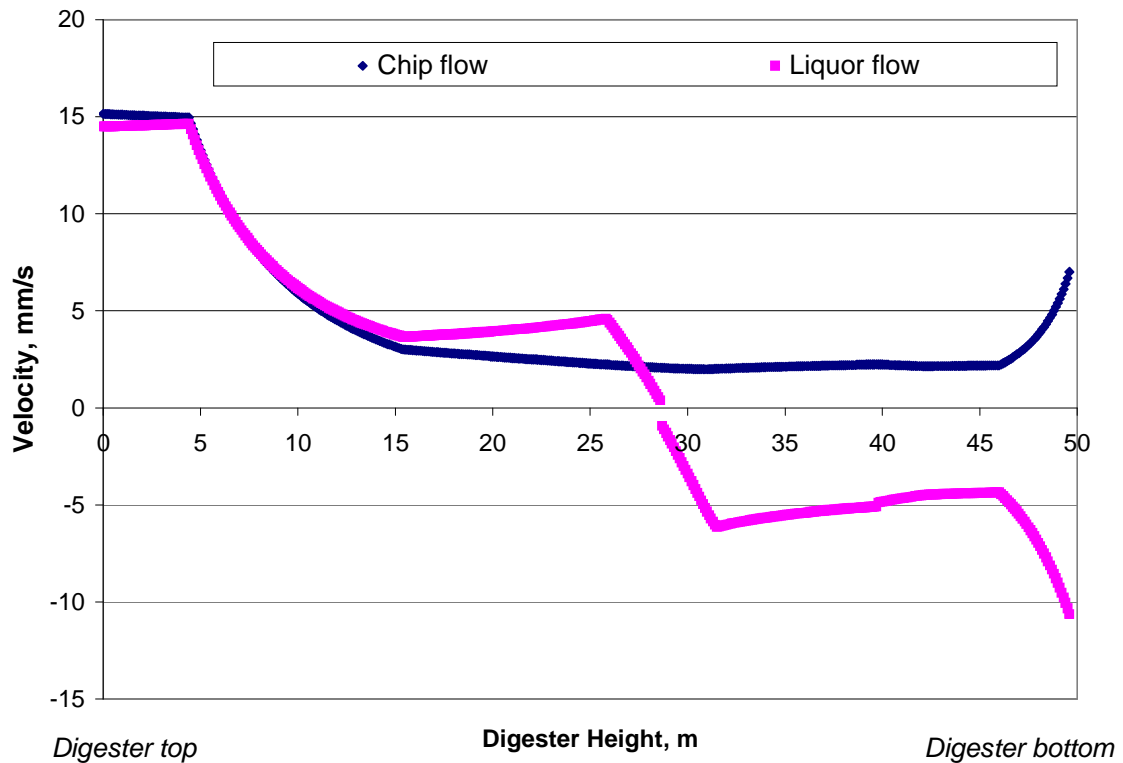


Figure 63 Chip and cooking liquor superficial flow velocity in default counter current digester simulation.

The total cooking time of the chips is illustrated in *Figure 64*. The total cooking time is about 280 minutes. The chip residence time in a mill digester with similar process conditions is about 290 to 300 minutes for counter-current cooking.

The total liquor flow time is a more complicated issue, because liquor is flowing from the top and bottom to the center of the digester. Liquor is extracted from the first extraction screen. For cooking liquor it takes about 80 minutes to reach the top of the extraction screen and about 100 minutes to reach the center part of the extraction screen. It takes 75 minutes for the liquor to reach from the bottom of the digester to the middle of the extraction screen.

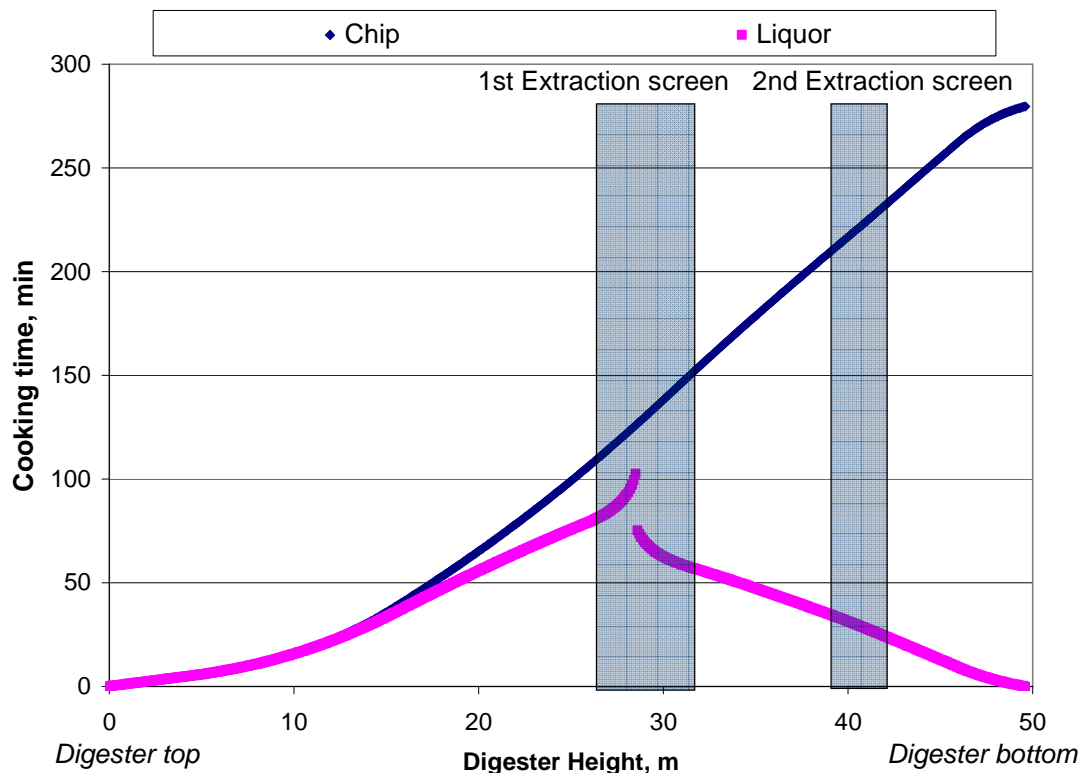


Figure 64 Total cooking time for chips and cooking liquor in normal counter-current digester simulation. Shade areas are the screen sections. NOTE: Liquor is flowing downwards from the digester top. At the middle of the 1st screen to liquor flow direction changes to upflow. This is the reason why time starts from zero at the digester bottom.

11.2. Impact of selected parameters on simulation results

11.2.1. Basic density of the chip

The production rate in continuous kraft cooking is based on the volumetric flow rate of the feeding chip screw. This flow volume is measured as the filling degree of the screw.

The basic density of wood chips affects the weight of a certain volume flow of wood chips. If the volume flow is kept constant and the basic density of the wood changes, the total amount of wood going to the digester will change. Default value for basic density is 400 kg/m³.

Table 6 contains a selection of different basic chip densities. The values are based on mill data. The basic density of the chip depends for example on the wood species and the growth site of the wood.

Table 6 Basic density of different raw materials based on mill measurements.

	Basic density, kg/m ³
Mill debarking, pine	400
Pine	385
Sawmill pine	450

The TKK Packing Simulator was used for determining how basic density affects the kappa number and cooking time. Volumetric chip flow rate was assumed to be constant. The volume of the flow changes as the basic density of the wood changes. The results are shown in *Table 7*.

When the basic density increases, the total mass flow of the wood chips is increased. This increase of the mass flow affects the total residence time in the digester, because packing of the chip bed is increased. As the packing is increased, chips kappa number decreases, because chip residence time is increased. Higher volume flow causes longer residence time for chip flow.

Table 7 Simulation results for different basic densities. Values shown in boldface are typical mill values; the other values have been selected to illustrate realistic margins of variation.

Basic density, kg/m ³	Kappa number	Cooking time, Min
350	31	240
400	28	280
430	27	310

11.2.2. Feed liquor-to-wood ratio

The feed liquor-to-wood ratio shows the proportion of cooking liquor in the cooking process compared to wood. When the LW is 2, there is 2 cubic meters of cooking liquor per 1 ton of oven-dry wood. In a continuous digester, the liquor-to-wood ratio affects the total flow inside the digester. As the liquor-to-wood ratio is increased, more liquor is flowing in the process.

The liquor-to-wood ratio affects the total liquor flow in continuous cooking. Different simulation results are shown in *Table 8*. The liquor-to-wood ratio indicates the amount of liquor entering in the upper part of the digester. The simulation solves how extra liquor flow affects chip bed packing and total cooking time. Default value for liquor-to-wood ratio is 5.

When the liquor-to-wood ratio is increased from 4 to 5, with a wood flow of 50 kg/s, the liquor flow in the top part of the digester increases from 200 l/s to 250 l/s. The change in the amount of liquor reduces the volume occupied by the wood chips and increases the liquor flow speed. And when the liquor-to-wood ratio is 6.5, the total liquor flow is 325 l/s. This further increases the chip bed packing. Increased packing can be seen by longer cooking time and higher maximum packing value.

Table 8 Simulation results for different liquor-to-wood ratios. The values shown in boldface are typical mill values; the other values have been selected to illustrate realistic margins of variation.

Liquor-to-wood	Max. packing value in simulation kPa	Cooking time, min
4	10	270
5	11	280
6,5	13	290

11.2.3. Production rate change

The following simulation was done to see how well the model simulates a change in the production rate. As a general rule, when the production rate changes by ± 100 ADt/day, the temperature is changed by ± 0.4 - 0.6 °C in the steam phase of the digester top in this type of mill. The nominal production rate is 2000 ADt/day.

In the simulation study, the production rate was decreased from 2000 ADt/day to 1600 ADt/day. Production rate was decreased in steps of 200 ADt/day. At the same time, the cooking temperature dropped by 1 °C for each 200 ADt/day change in the production rate. The results are compared to a case without a temperature change in the digester.

The results are shown in Table 9. The change in production rate mainly affects the cooking plant by causing an increase in residence time in the cooking process. The cooking time increases as the volume flow of chips into the digester decreases.

Table 9 Simulation results for different production rates.

Production, ADt/day	Kappa number	Kappa number (without temperature correction)
2000	28	28
1800	27	26
1600	26	23

Based on the simulation results, the kappa number decreases when the production rate is decreased. If the production rate changes with an amount close to the nominal production rate, the change in digester top temperature should be close to 0.4 °C for each 100 ADt/day change in production. If there is a bigger change in the production rate, the corresponding change in digester top temperature should be close to 0.6 °C.

The drop in kappa number is more drastic when there is no temperature correction in the digester. This is of course not common in mill operation, but it is still useful to know how changes in mill operation affect pulp quality and how quality can be improved.

11.2.4. Chip and liquor level differences effect on chip bed

This section discusses how initial packing in steam-liquor phase digesters affects on bed packing. The most important parameter affecting initial packing is the chip and liquor level difference (LD). LD measures, how much of the chip pile is above the liquor level. LD is discussed in more detail in *Section 10.4.2*. Level differences unit is meters.

Packing conditions in chip bed were simulated by changing chip and liquor levels. Process conditions for different simulations are shown in *Table 10*. Simulated cases are based on default digester parameters with counter-current washing section.

Table 10 Input values for simulation of chip and liquor level difference on chip bed packing. The values shown in boldface are typical mill value.

	Chip and liquor level difference (m)
Case 1	0.5
Case 2	1
Case 3	1.5

First is discussed how chip and liquor level difference affects the lignin content. Varying LD does not affect lignin content of the chip along the digester very much. This is shown in *Figure 65*. Lignin content of the chip is un-affected, because alkali charge and cooking temperature are kept constant.

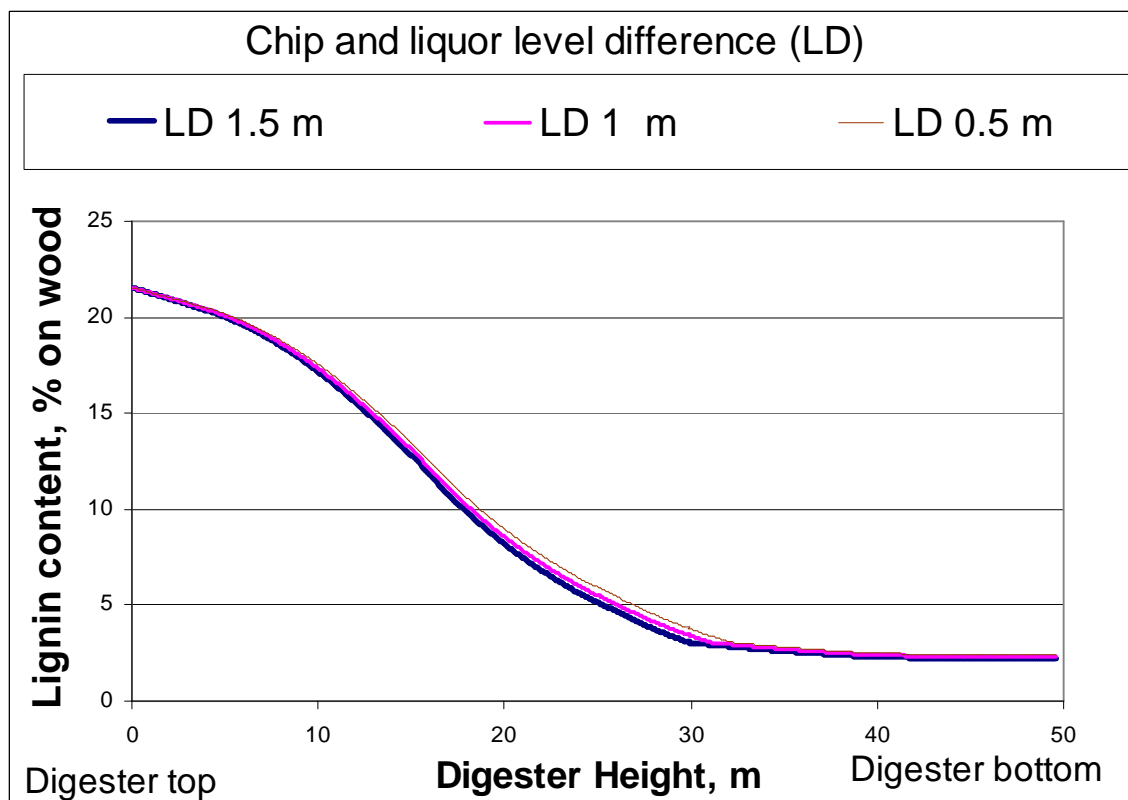


Figure 65 Lignin content along the digester height in default counter-current digester simulation.

Next is shown how chip and liquor level affects the porosity. When LD is increased, initial compaction of the chip bed is increased. There are more chips above the liquor level, which is causing downward force. Higher LD decreases the initial porosity (*Figure 66*). Porosity decreases and achieves its minimum value at the extraction screen about 30 meters from the digester top. At the screen section, downward cooking liquor and upward washing liquor flow are taken out from the digester. These contacting liquors are causing the maximum packing in bed.

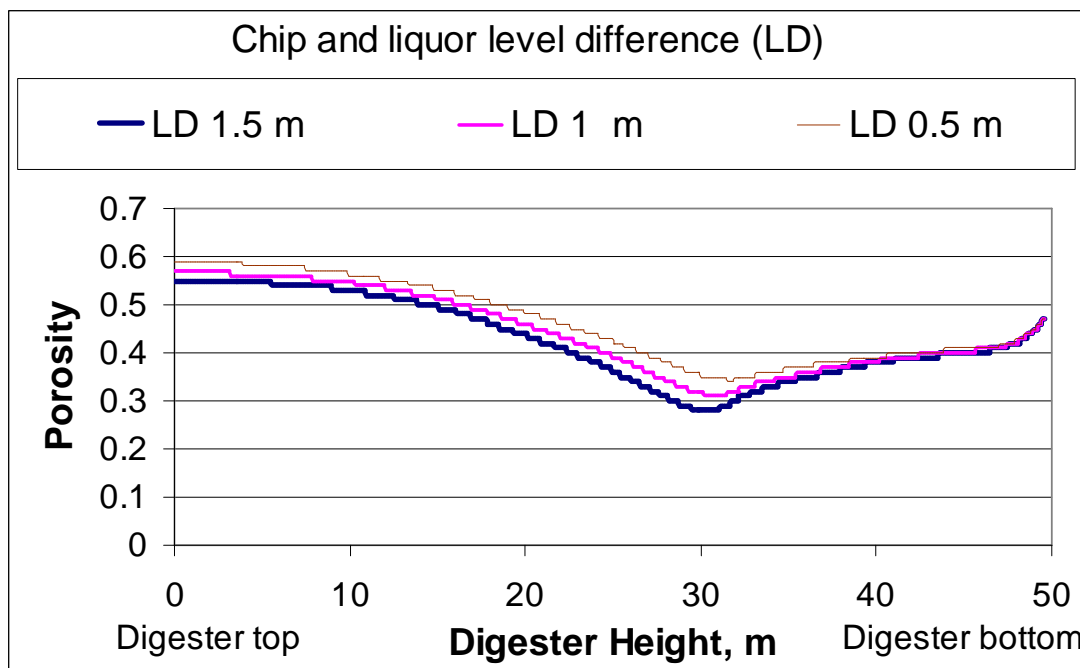


Figure 66 Porosity along the digester height in default counter-current digester simulation.

At last is discussed chip liquor level affect on chip bed pressure. The chip and liquor level difference affects the initial chip bed pressure. This is illustrated in Figure 67. Maximum chip bed packing pressure varies from 12 kPa to 9 kPa, when the LD is varied from 1.5 to 0.5 meters. Higher the initial pressure, higher the maximum pressure is.

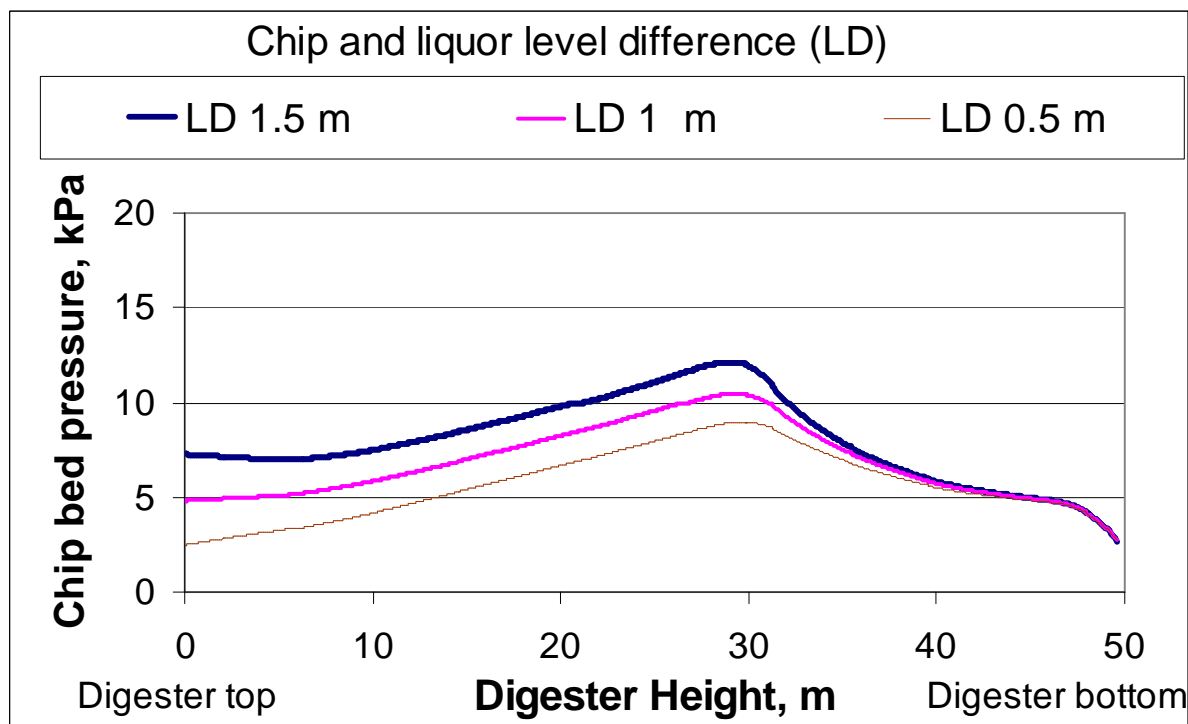


Figure 67 Chip bed pressure along the digester height in default counter-current digester simulation.

11.2.5. Chip softening effect of chip bed packing

Chip bed packing can be reduced by changing the lignin profile inside the digester. Lignin content of the chip affects the compaction of the chip bed. Higher the lignin content, less compacted the chip bed is.

In this simulation case lignin reactions are slowed down by decreasing the cooking temperatures. Following cooking examples are compared:

- lower cooking temperature in upper part of the digester, H-factor 900
- higher cooking temperature in the upper part of digester, H-factor 900

Counter-current cooking method is used. Chip and liquor level difference at top of the digester was 2 meters is used in both cases. Higher LD is used for showing the effect of the temperature in delignification. Base case cooking temperatures are the same as in the default case. Second case cooking temperature is adjusted in by decreasing the cooking temperature in top part of the digester by 4 °C. Temperature profiles of the different cases are shown in *Figure 68*. The lower part digester temperature was adjusted to meet the same H-factor demand as in the default case.

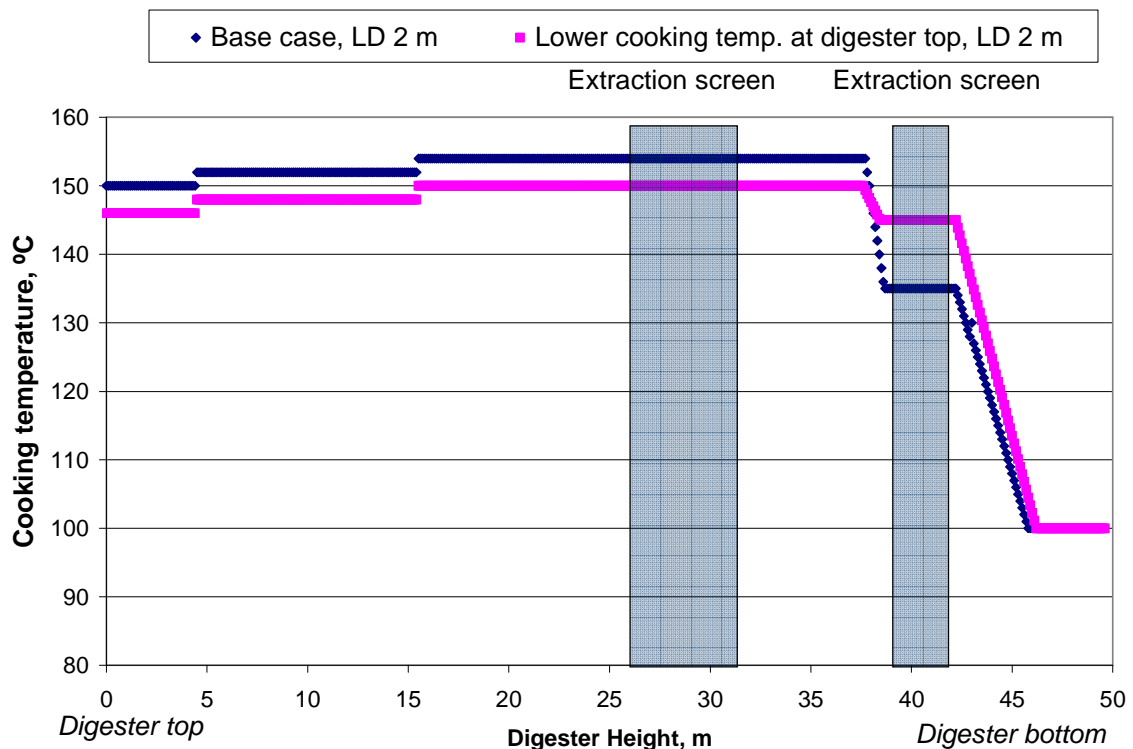


Figure 68 Cooking temperature in normal counter-current cooking (base case) and with lower cooking temperature at the digester top simulation.

The lignin profiles for both cases are shown in *Figure 69*. Lignin content decreases in the slow delignification case. In the middle of the extraction screen the lignin content is about 5 % of wood, which corresponds to kappa number 60. In the default case the lignin content is 3.5 % of wood, which corresponds to kappa number 40.

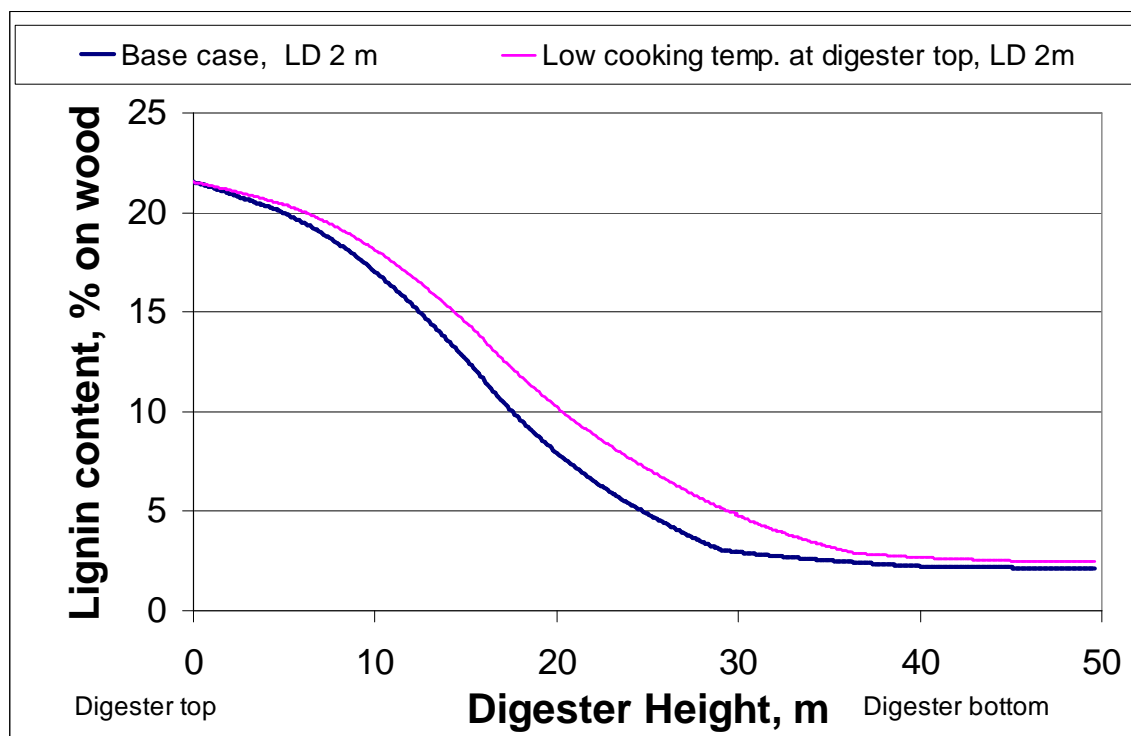


Figure 69 Lignin content along the digester height in default counter-current digester. Level difference is 2 meters. Other case is simulated with lower cooking temperature at digester top.

Porosity is used for calculating flow conditions in the digester. As the porosity decreases, liquor flow becomes more uniform. Porosity for two cases is shown in *Figure 70*. Default case porosity decreases below 0.2, which means that liquor flow becomes very restricted. This might lead channeling of liquor flow in chip bed or chip plug movement stops. When the lignin content is higher, as in the case with slow delignification, chip bed porosity is much higher. Packing of the chip bed is reduced by keeping the lignin content much higher level compared to default case.

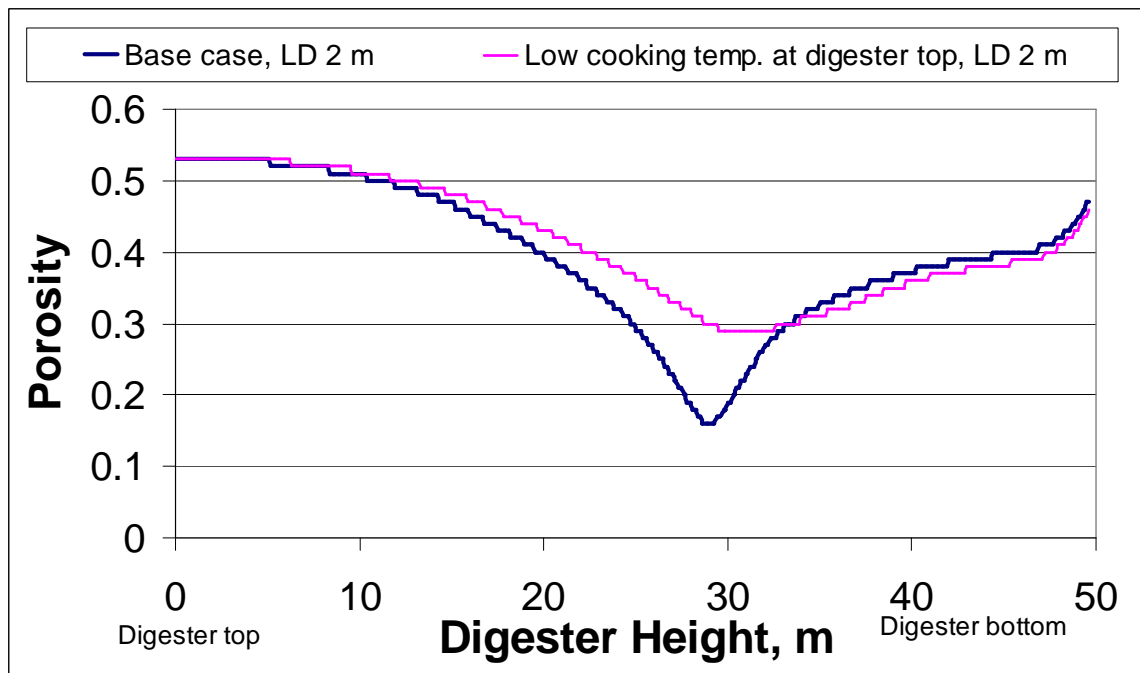


Figure 70 Porosity along the digester height in default counter-current digester simulation. Level difference is 2 meters. Other case is simulated with lower cooking temperature at digester top.

Chip bed pressure is shown in *Figure 71*. Chip bed pressure in both cases is high. Lee shows /40/, that compacting pressure over 14 kPa caused high pressure drop in bed even with high kappa numbers (kappa number 69). Still, maximum chip bed pressure can be reduced from 19 kPa to 15 kPa by just increasing the kappa number from 40 to 60 at the middle of the extraction screen. Controlling the lignin profile inside the continuous digester is one way to reduce the chip bed packing.

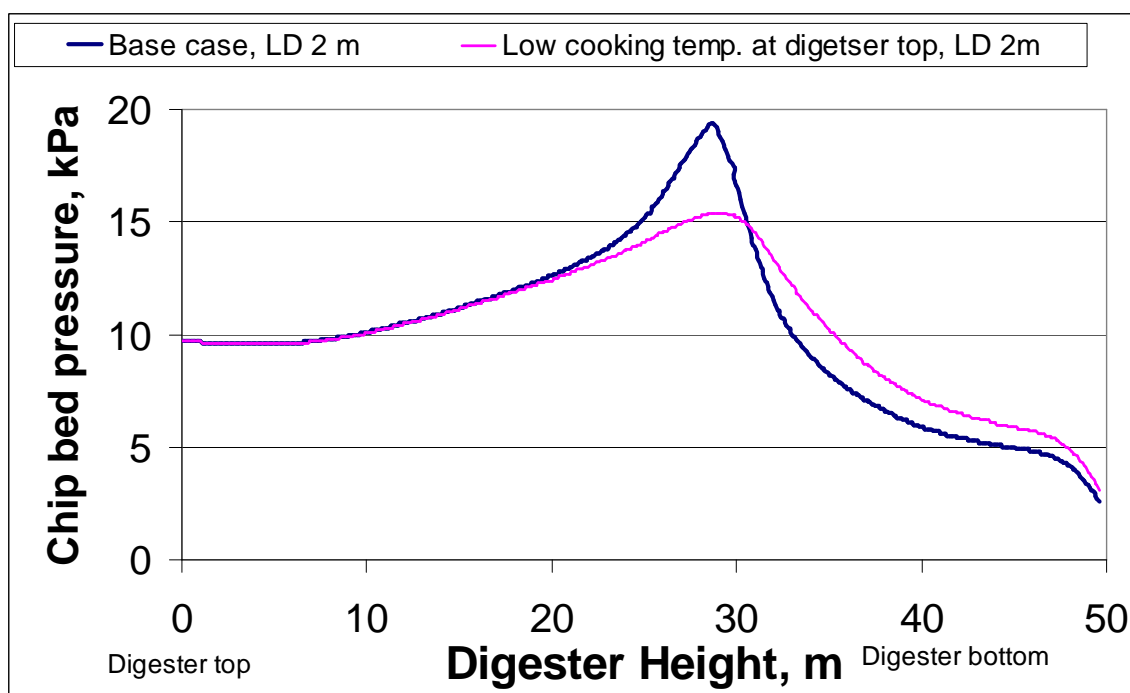


Figure 71 Chip bed pressure along the digester height in default counter-current digester simulation. Level difference is 2 meters. Other case is simulated with lower cooking temperature at digester top.

11.2.6. Liquor flows effect on chip bed packing

The following section discusses how changes in liquor flow directions affect chip bed pressure and packing. A change in the liquor flow direction means that the flow direction of the cooking liquor between two extraction screens is changed. As a “normal” case, cooking liquor is flowing upwards between the screens as the chips are flowing downwards, as shown in *Figure 72*. Cooking liquor flows are shown in the same figure.

This upflow is also called “Hi-Heat-washing”. In some digesters the liquor flow direction has been changed from upflow to downflow. Downflow cooking reportedly leads to better heat economy, less liquor channeling and improved chip column movement [26, 93].

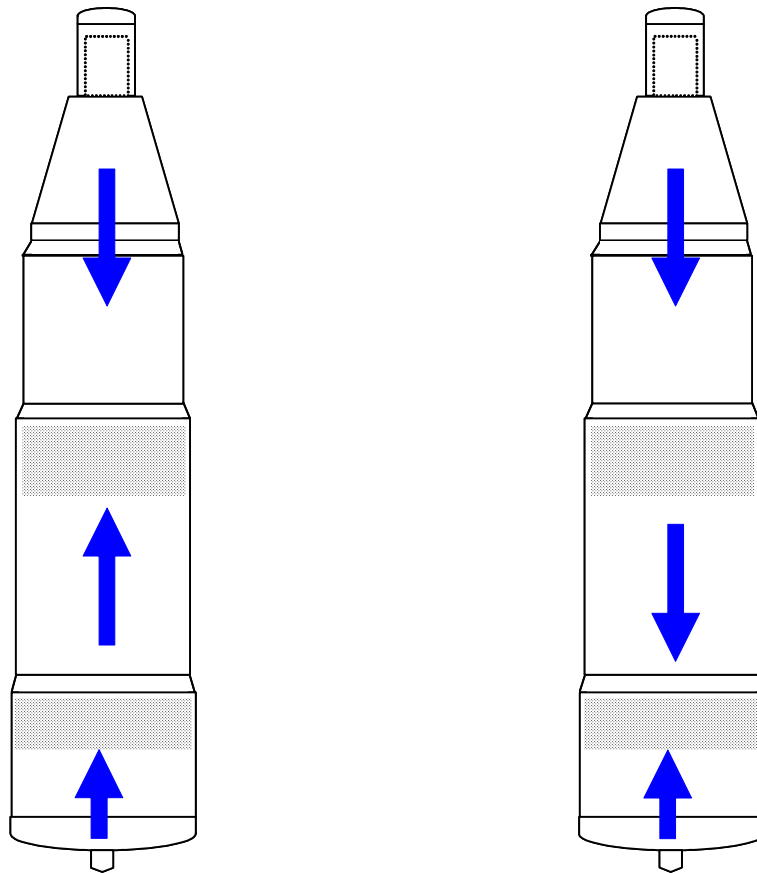


Figure 72 Upflow (left) and downflow digester.

Process conditions are kept more or less the same in both downflow and upflow cooking. This means that temperatures and alkali concentration in the beginning are the same. Flows are varied in such a way that the downflow cooking liquor is flowing in the same direction as the chips between two extraction screens. Also, the amount of extraction between the screens differs between the two cooking methods.

A change in liquor flow direction affects the chip bed pressure and packing. A change in liquor flow direction affects the volume where the chip bed pressure is at its maximum, as shown in *Figure 73*. In downflow cooking, the liquor and chips move in the same direction all the way from the top of the digester to the second extraction screen. At the second extraction screen, the chip flow encounters the upflowing cooking liquor that has been added at the bottom of the digester. At this point, the chip bed pressure reaches its maximum.

The maximum chip bed pressure is higher in a downflow digester. As the chips and liquor are flowing in the same direction over 40 meters, compared to a little less than 30 meters in the upflow digester, the force of the chip bed is bigger at that point. Furthermore, the chip bed is softer, due to lower kappa number, which increases the packing effect.

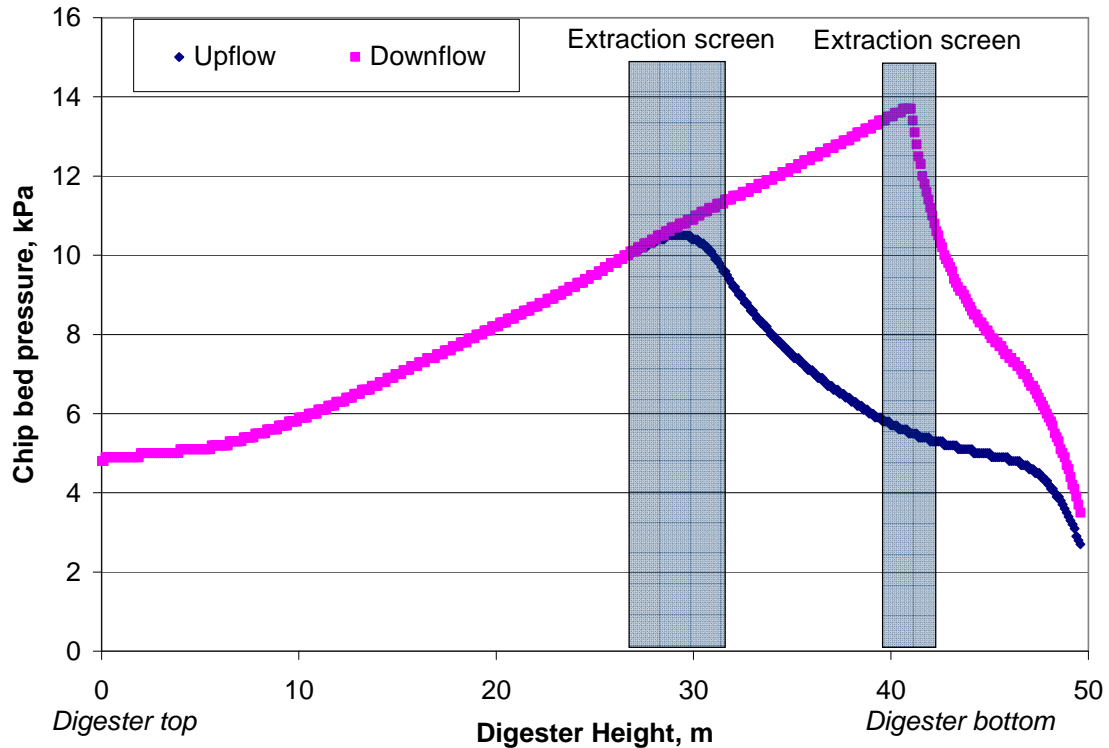


Figure 73 Chip bed pressure in upflow and downflow digester simulation.

The porosity of the chip bed in downflow cooking is very similar to that in upflow cooking (*Figure 74*) until the first extraction screen. In upflow cooking, the porosity starts to rise at the first extraction screen, whereas in the downflow cooking the porosity continues to decrease until the second extraction screen. This is mainly because the pressure of the chip bed increases and chips are compressed more intensively. The volume occupied by the wood chips increases and the porosity of the chip bed decreases. As the chip bed pressure decreases after the second extraction screen, the porosity starts to increase.

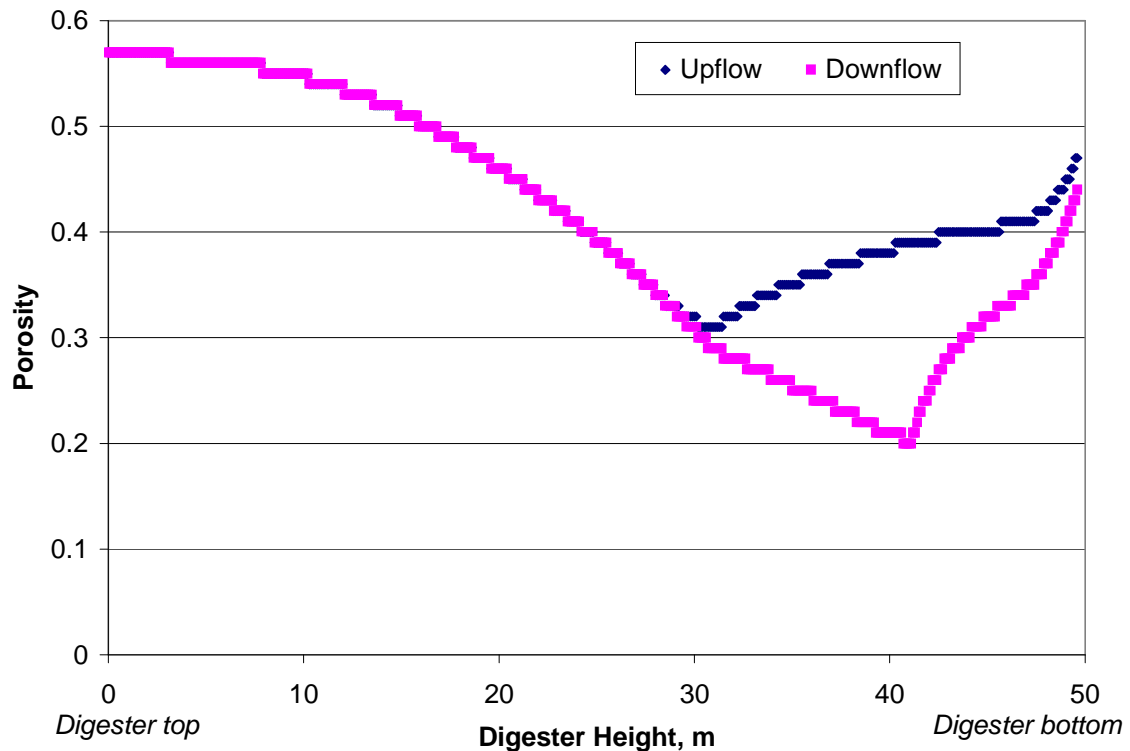


Figure 74 Porosity of chip bed in upflow and downflow digester simulation.

Chip and liquor velocities in downflow and upflow digester simulations are shown in *Figure 75*. Chip velocities are about the same in both downflow and upflow cooking. It is only between the extraction screens that the chips are moving slower in downflow cooking, because the compaction of the chip bed is greater in downflow cooking.

Liquor flow velocities follow each other closely until the first extraction screen. At that point, the flow direction in upflow cooking changes. The velocity of the downflow liquor decreases somewhat at the extraction screen. This is because part of the free liquor is extracted from the screen and the volume occupied by liquor is reduced. The liquor flow direction in downflow cooking changes at the second extraction screen. This explains the discontinuity in the liquor flow at about 41 meters.

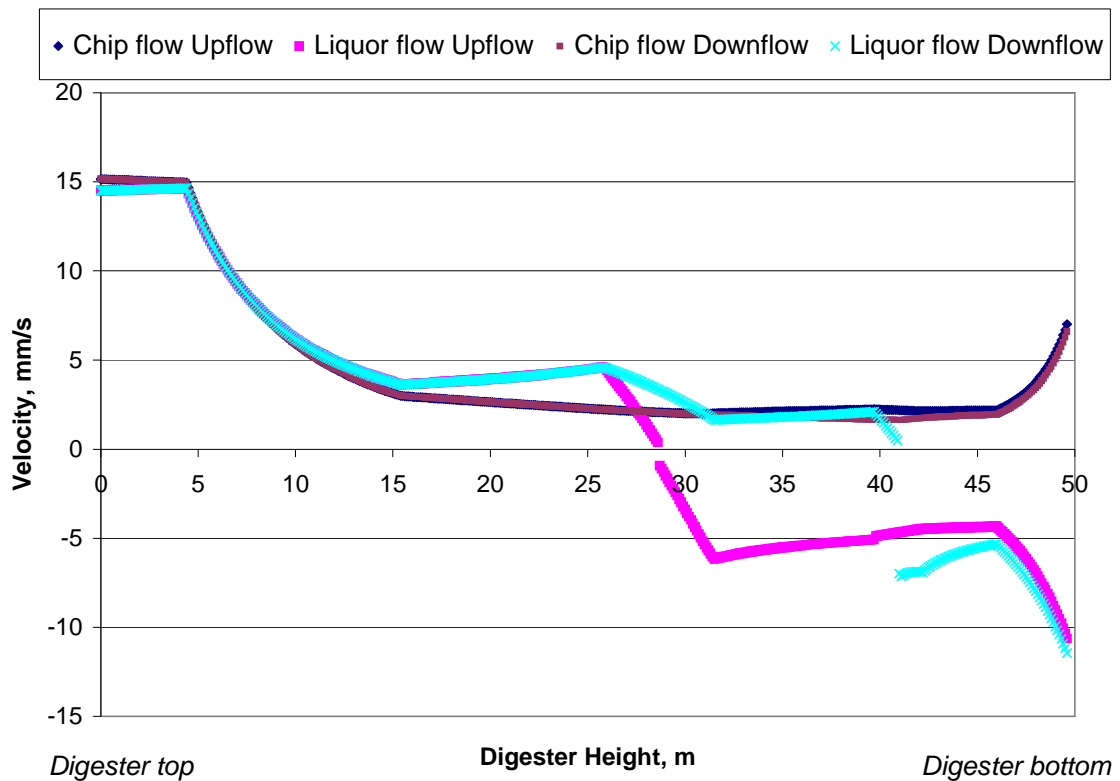


Figure 75 Chip and cooking liquor superficial flow velocities in default upflow and downflow digester simulation.

Chip and liquor flow times are shown in *Figure 76*. Liquor flow times are very similar up to the first extraction screen. At the middle of the first extraction screen, the upflow liquor residence time is a little greater than that of the downflow liquor, because the velocity of the upflow liquor decreases at the extraction screen. This is due to packing of the chip bed and higher flow resistance in the screen area. The total flow time for the downflow cooking liquor to the beginning of the second extraction screen is about 180 minutes.

The cooking method does not affect the residence time in the chip flow until close to the second extraction screen. The packing of the chip bed increases the porosity of the bed and thus reduces the chip residence time. In the end, the difference in residence times is about 20 minutes.

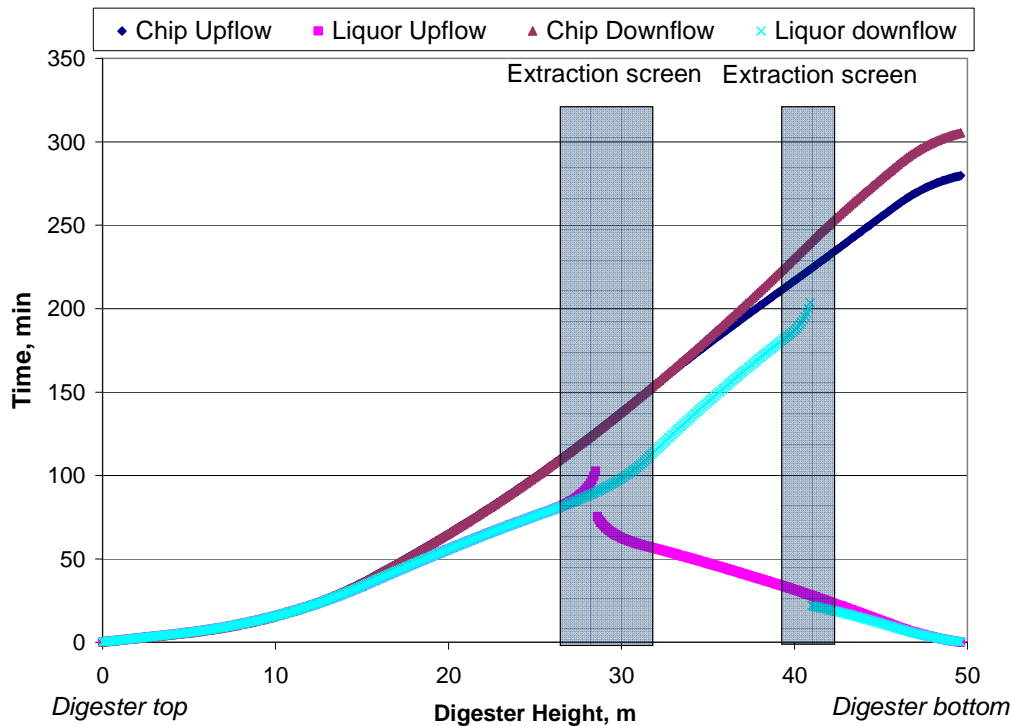


Figure 76 Total cooking time for chips and cooking liquor in default upflow and downflow digester simulation. NOTE: Liquor is flowing downwards from the digester top. At the middle of the 1st screen to liquor flow direction changes to upflow. This is the reason why time starts from zero at the digester bottom.

11.3. Summary of simulated results

Simulator results of chip bed packing have been presented about this *Chapter*. The chapter shows how different parameters affect the chip bed compaction. The effect of the parameters on the chip bed pressure and effects are given in detail below.

The simulator describes the packing phenomenon of the continuous digester trough the concept of chip bed pressure. The chip bed pressure is the pressure affecting the chips at a given height of the digester (constant in horizontal direction).

In the simulator *Equation 9* is used to calculate the chip bed pressure. An increase in chip bed pressure leads to higher compaction and thereby longer cooking time. Longer cooking time means lower kappa number and softer chips, which leads more compacted chip bed. The equation can be seen containing following terms in calculation (*Figure 77*):

- the *gravity force* on the chip column
- the *wall friction* between the digester wall and the chip column
- the *flow friction* between the chip column and the liquor

$$\frac{dp_c}{dL_b} = \underbrace{(\rho_c - \rho_l)\epsilon_c g}_{\text{Gravity force}} - \underbrace{\mu \frac{p_c}{D_{\text{digester}}}}_{\text{Wall friction}} \pm \underbrace{\frac{dp_l}{dL_b}}_{\text{Flow friction}}$$

Figure 77 Force terms affecting the chip bed packing: gravity force, wall friction and flow friction.

The gravity force terms calculates the net effect of gravity and buoyancy i.e. the weight of the chip column above the studied level. The cooking variables affect the term following way:

Chip basic density – denser chips mean that the column above a given level weighs more. This in turn leads to higher compaction.

Production rate – volume flow of the chips is increased. As the volume of chip flow is increased, the chip bed is less compacted.

Chip and liquor level difference – the part of the chip level that is not submerged is only affected by gravity and not by buoyancy. The more of the column that is above the liquor level the greater the compaction.

Chip softening – lignin content has effect on how individual chip can bear a loading. As the lignin content decreases, softer the chips become. Also the volume occupied by the chips is decreased, as the lignin content decreases. If the lignin removal in upper part of digester is increased, more compacted the chip bed becomes.

The wall friction terms calculate the force of the chip bed above the studied level divided by the circumference of the digester. The cooking variables affect the same way into the friction term as the gravity force term, except the force direction is opposite to the gravity force terms. The friction coefficient at digester wall section is 0.01 and at the screen section wall friction is 0.02.

Flow friction terms calculate the liquor pressure drop along digester. This term is based on Ergun equation (*Equation 6*).

Feed liquor-to-wood ratio – increase in ratio leads to bigger liquor flow in the upper part of the digester. Increase in the flow leads to higher flow friction, which in turn leads to higher chip bed compaction.

Liquor flow direction - liquor flow direction has an effect on chip bed compaction. When liquor and chips are moving in the same direction, the liquor flow increases the chip bed packing.

12. Simulated and measured mill results

During the course of this study chip bed conditions were evaluated to see how chip and liquor are flowing inside the digester. It was suspected that liquor flows are not uniform inside the digester. At the same time, the packing model was compared to real mill flow conditions.

In the beginning simulator results are compared to mill measurement. The aim was to verify that the simulator results are close to measured mill results. The pressure difference in digester screen is compared to simulated chip bed pressure. In addition, real chip and liquor flows are compared to simulated flows.

12.1. Chip bed packing at the first extraction screen

There are very few direct ways to measure compaction directly in a continuous digester. Indirect measurements are therefore needed. One such measure is the pressure difference measurement (PD) over certain screen section in the digester.

The pressure difference gives an indication of the chip bed pressure at a given height of the digester (*Figure 78*). The PD is the difference between the pressure inside the digester at the certain height (P_1 , P_2 , ..., P_N) and pressure of the extraction line (P_{ext}). These PD measurements give relative information about compaction of the chip bed at different heights.

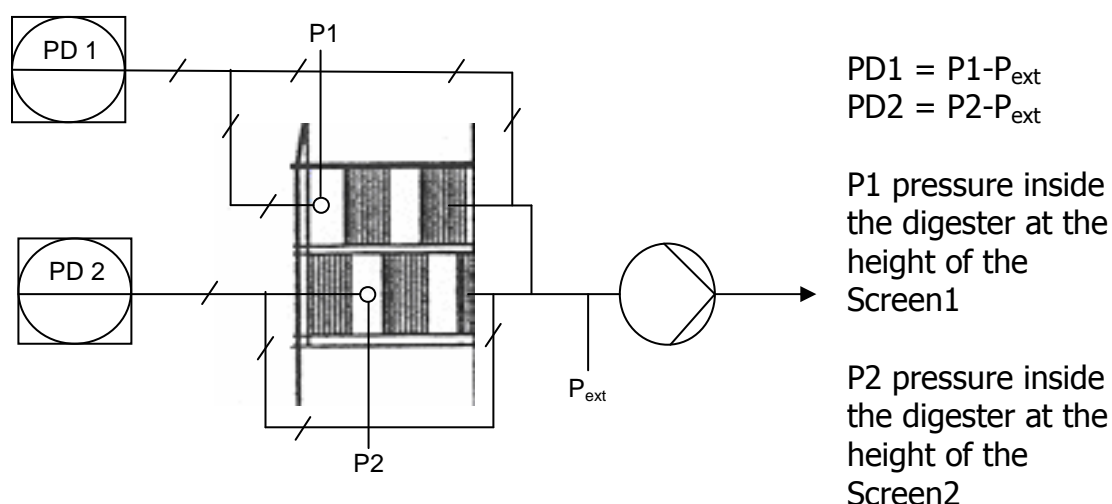


Figure 78 Pressure difference (PD) measurement in continuous digester.

The digester has four screens in the first extraction section, with continuous measurement of the pressure difference between the screens. Based on simulation of upflow cooking, maximum packing was expected at the upper half of the extraction screen (*Figure 61, page 98*). At the mill, the highest pressure difference was measured also at the upper of screens (PD 1 and

PD2). This is shown in *Figure 79*. Based on this comparison, it can be said that the simulator gives a realistic picture of the packing conditions inside the digester.

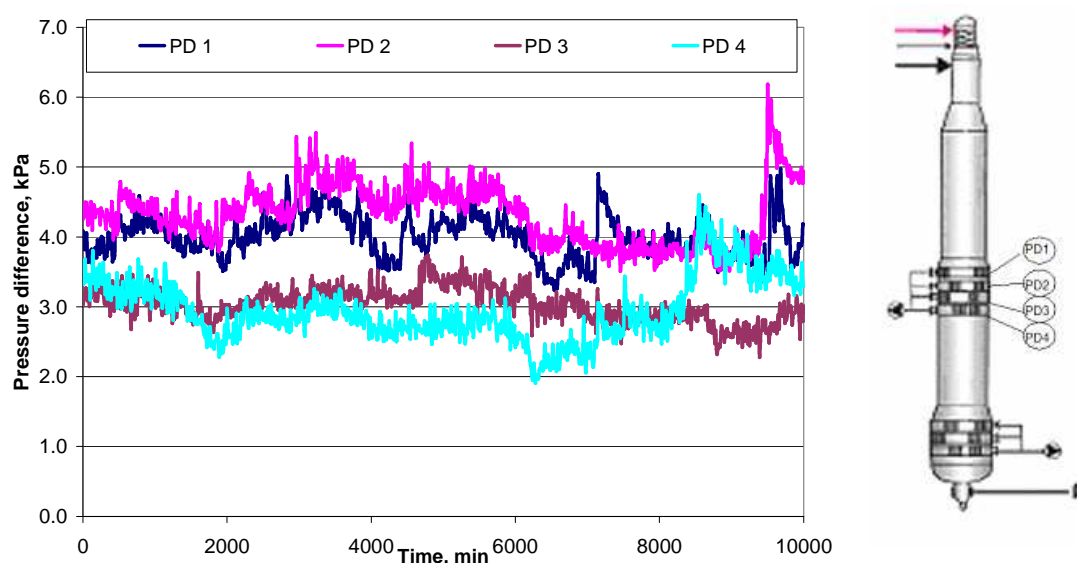


Figure 79 Pressure difference measurements (PD) in the extraction screens (at four different levels) in counter-current cooking mode in mill conditions (10 minutes average value).

12.2. Simulation of chip residence time and chip flow measurement in digester

The chip flow inside the Compact Cooking digester was studied by using a radioactive tracer. Measured chip flow was compared to simulated chip flow. Chip flow measurement was done in co-operation with Indmeas Oy. Radioactive golden staples were stapled to wood chips and the flows of these radioactive chips were followed throughout the cooking plant. The experiments were done twice. The digester was running in upflow cooking mode. The cooking method is explained in detail in *Section 11.2.6*.

Radioactive chips (200 chips) were added to the chip silo in a paper bag. The chips flowed as a plug from the chip silo. After the impregnation vessel, the chip flow into the digester can be said to be uniform. The digester detector setup for the experiments is shown in *Figure 80*.

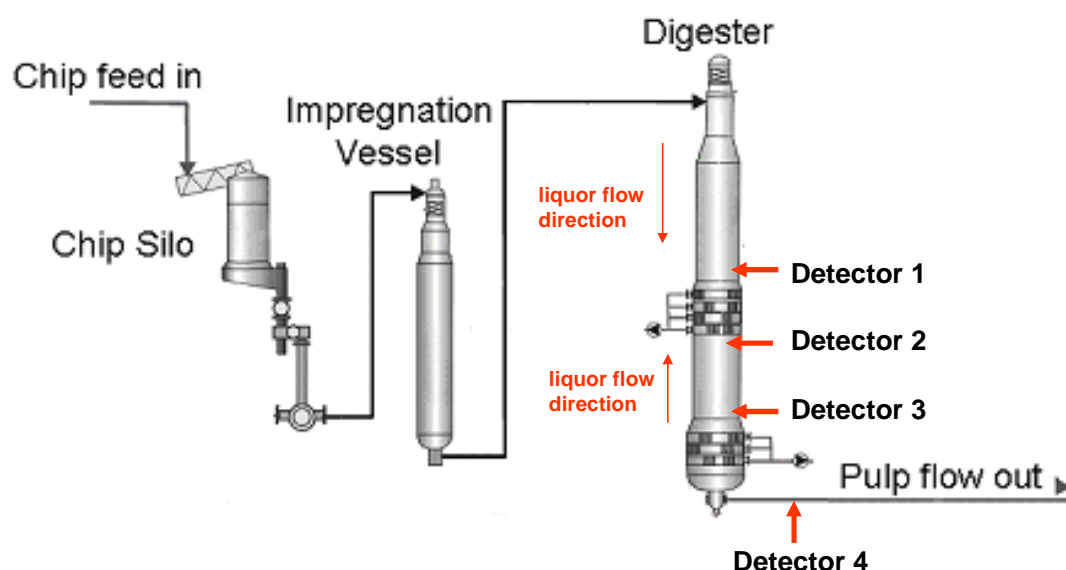


Figure 80 Detector setup in radioactive tracer test for chip flow in continuous digester.

Tracer test were done during one day, when the production rate at the digester was maintained at 1700 ADt/day. Digester liquor-to-wood ratio was 5.2. Cooking temperature was 150 °C. Mid-point kappa number was 34 and blowline kappa number was 30. Mid-point kappa sample is taken out just below the extraction screen. Raw material was 45 % spruce and 55 % pine.

The total residence time and standard deviation for both experiments are shown in *Table 11*.

Table 11 Total residence time in digester in radioactive tracer tests in continuous digester.

	Experiment 1	Experiment 2
Production, ADt/day	1700	1700
Measured average residence time, min	407	410
Standard deviation, min	7	8

The chip flow in the digester is uniform. Based on the shape of the response curve (*Figure 81*), the chip flow is almost of plug-flow type.

During the test, the detectors were fitted along the digester wall. In the first tracer test, detectors did not detect any of these radioactive chips. During the second tracer test, the detectors detected the chip flow along the digester. The tracer test results are compared to simulated results in *Table 12*.

Table 12 Measured and simulated tracer test results for chip flow in the continuous digester.

		Digester Time, min	Simulator Time, min	Digester Porosity	Simulator Porosity average	Digester Velocity, mm/s	Simulator Velocity, mm/s
Detector 1	Digester top part	160	150	-	-	2.7	2.9
Detector 2	Upper screen area	90	90	-	-	1.5	1.5
Detector 3	Upflow cooking	70	50	-	-	1.2	1.7
Detector 4	Digester bottom part	90	100	-	-	1.8	1.7
	TOTAL	410	390	0.36	0.39	2.0	2.1

When comparing the simulated results to real digester data, it is evident that the simulator calculated cooking conditions quite accurately. The simulator predicted only 5 % shorter total cooking time than indicated by the tracer test results. Total average porosity was about the same in both cases and there was only a minor difference in average chip velocities.

Closer comparison of the cooking time data shows that the digester bottom part was not so accurate. Both the upflow zone and bottom part calculations vary a lot from the measured results. The measured chip flow was much slower than the simulator calculation in upflow cooking.

The average porosity, i.e. how much open space there is on average in a certain part of the digester, is more complicated. The simulator takes into account the yield changes which affect the porosity of the bed. When the yield decreases, more open space becomes available for free liquor. It is difficult to estimate the yield change along the different parts of the digester in real digester. That is why only the whole digester porosity is shown *Table 12*.

The flow speed of chips was almost the same throughout the digester, both in the simulator and in the tracer test. The only difference was in the upflow cooking section, where chips are moving slower compared to simulated results. The packing effect in the upflow section seems to be greater than indicated by the simulator calculations.

12.2.1. Chip flow measurement in digester

A radioactive tracer was used to study the chip flow in the digester. The method proved to be a good method for studying the chip flow. However, there are still some problems with using radioactive staples for the flow measurement. The detector's range to detect the staples is limited. Because of the thick steel wall and insulation, the detection range in the digester area was only 1.5 meters. This limits the detection to the outer layer of the digester. As the digester examined had a diameter of 10 meters, the stapled chip flow in the middle of the digester remained undetected.

Furthermore, the pulse response of an individual stapled chip depends on how closely the chip passes the detector. Because of this, the pulse response of radioactive chips varies in different parts of the digester. For example, if there is some tailing in the tracer, like in *Figure 81*, it is uncertain how many chips or how big a portion of the flow that are causing the tailing phenomena. The figure refers to the second tracer test.

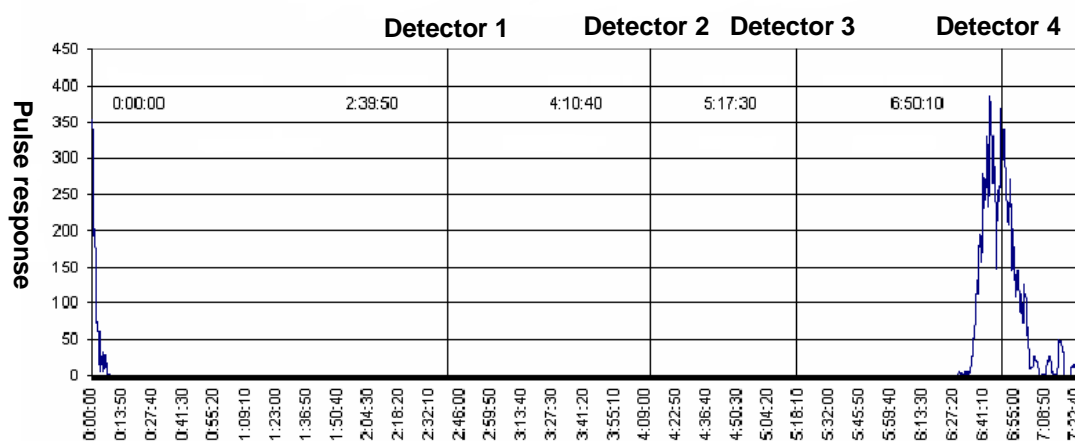


Figure 81 Residence time distributions of radioactive chips in continuous digester (2nd tracer test experiment).

When the test results were compared with simulator results, the simulator was found to give a fairly realistic picture of the average flow of chips in the digester. The simulated values were about 5 % lower than the measured values for the total delay. Mills upper level control system calculated the total residence time in digester to be 20 % lower. It can be said that simulator worked properly for studying the chip flow in continuous digester.

12.3. Simulation of digester liquor flow and liquor flow measurement in digester

It was assumed that there was some liquor channeling in the bottom part of the digester, because the temperature measurements showed variations in the second screen section in the mill. This phenomenon is illustrated in *Figure 18* and *Figure 19*. It was decided to examine the liquor flow in more detail with a tracer test. Lithium chloride was selected as the tracer.

Over the course of one year period, a number of attempts (> 15) were made to measure the liquor flows using LiCl. In the following results from the three most successful attempts are presented. The results given in this section have been collected on three separate dates. The production varied from 2000 to 2100 ADt/day. The digester was operating in downflow cooking mode. Process conditions of different experiments are given in *Table 13*. Mid-point kappa is kappa number of pulp sample just below the extraction screen.

Table 13 Process conditions in different liquor tracer test experiments.

Production rate	LD	LW	Cooking temperature (Digester top)	Mid-point Kappa	Blowline kappa	Wood in Spruce	Wood in Pine
ADt/day	m		°C			%	%
2000	1	5	151	46	27	0	100
2050	1.2	4.8	151	39	24	53	47
2100	1.4	5	150	48	25	61	39

The experimental setup is shown in *Figure 82*. Pure LiCl-powder (5 kg) was dissolved in water (8 l). Solution was put in a pressure vessel. The vessel was pressurized above the process pressure. The tracer (LiCl-solution) was introduced into the white liquor line. As the chips were fed into the digester, the white liquor with the tracer contacted the chips. Chips and liquor flowed downwards. Part of the liquor was taken out from the digester at *Screen 1*, which was also the first sample collection point. Liquor and chips flowed downwards after the first screen. At *Screen 2*, part of the liquor was taken out from the digester. This was the second sample point. From these collected liquor samples the lithium content was analyzed with the Atomic Absorption Spectrophotometer (AAS). According to the laboratory performing the AAS measurements, the uncertainty of the measurement was estimated to be $\pm 20\%$. As the aim of the LiCl-tracer experiments was only to measure residence times for cooking liquors, it is felt that the level of accuracy is sufficient.

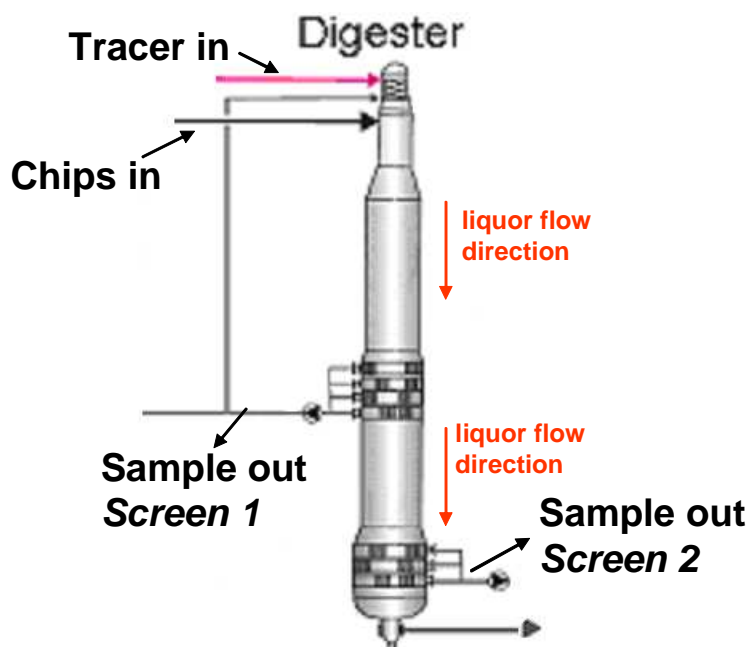


Figure 82 Lithium chloride tracer test setup for liquor flow in continuous digester.

Part of the liquor removed from *Screen 1* is re-circulated to the digester top. This recirculation flow is approximately 15 % of the total liquor flow at the

digester top. This flow will cause internal recirculation, which is seen in the tracer test.

Tracer was added into the process as a single pulse. The pulse was one minute long. At the same time as the tracer was added into the cooking system, liquor samples were taken from the screens. From these samples the lithium content was measured. This level is called the floor-level content of lithium.

12.3.1. Liquor flow measurement in digester

An example of the tracer test results is shown in *Figure 83*. The production rate during this test was 2050 ADt/day. Samples from *Screen 1* were collected at every 8 minutes and from *Screen 2* every 20 minutes or every 10 minutes.

The lithium reaches *Screen 1* in about 30 minutes and the whole pulse passes through in about 110 minutes. After the tracer pulse has passed *Screen 1*, the floor-level lithium content is about 20 % higher than the pre-test level. This is probably caused by the recirculation of liquor from *Screen 1* to the digester top.

The first traces of the tracer start to show up in *Screen 2* in about 150 minutes. Sample collection was stopped too early to see the whole pulse go through the screen.

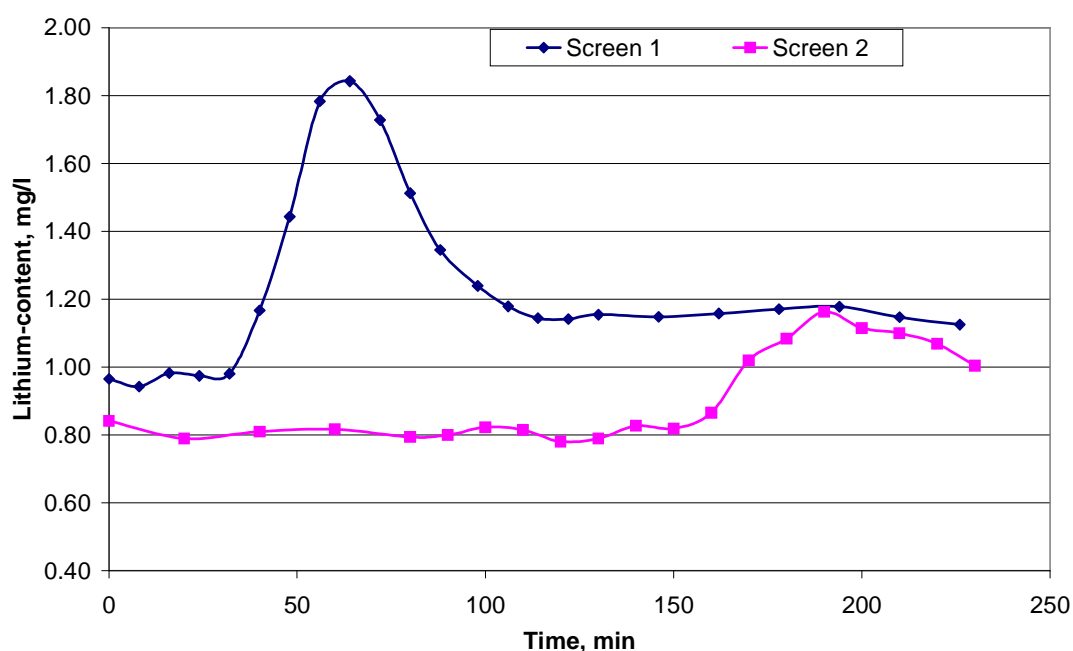


Figure 83 Tracer content at the first and the second extraction screen. Production rate is 2050 ADt/day.

Based on these first results, there was no sign of channeling or short-cutting in the flow. There might be some tailing, but recirculation obscures that

effect. Half of the pulse is detected in *Screen 1* within 60 minutes from the start of the test. An approximation for half of the pulse to go through *Screen 2* is 200 minutes. The average superficial velocity for *Screen 1* is 8 mm/s and for *Screen 2* it is 4 mm/s.

It is hard to calculate the characteristics of the reactor's hydraulic performance, because only part of the tracer is recovered in the screens. Only about 10 % of the tracer is recovered in *Screen 1* and about 5 % in *Screen 2*. The rest of the tracer leaves the digester with the pulp.

12.3.2. Liquor flow at different production rates

Following two tracer tests were conducted to find out if there is any short-cutting of the liquor flow. The tracer flows at different production rates are shown in *Figure 84*. To confirm this phenomenon, liquor sample collection was started right after tracer addition. Both experiments showed a rise in the lithium concentration shortly after tracer addition. In the experiment with a higher production rate, more liquor samples were collected during the first ten minutes. In this experiment, the lithium content reached its maximum in two minutes. With the lower production rate, short-cutting lasted about 10 minutes and with the higher production rate the lithium content remained higher than the floor-level content until the main flow of tracer reached the screen.

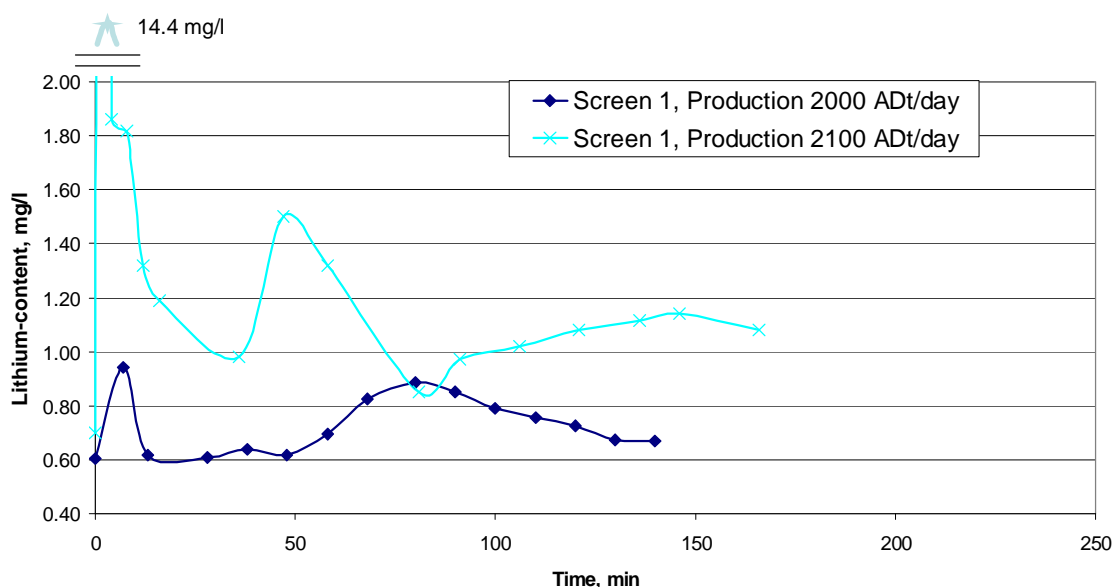


Figure 84 Tracer content at the first screen with two different production rates.

The shapes of the tracer curves are different. This can clearly be seen when comparing the shape of the major response curves. The tracer content starts to rise again after 40 minutes with the higher production rate and after 50 minutes with the lower production rate. Half of the pulse has passed the

screen within 50 minutes with the higher production rate and within 82 minutes with the lower production rate. With the higher production rate, most of the tracer has passed *Screen 1*, whereas with the lower production rate only half of the tracer has passed the screen. The average superficial velocity with the higher production rate is 10 mm/s and with the lower production rate 6 mm/s. The liquor flow with the lower production rate is more uniform, and after the short-cut flow the tracer content returns to the floor level.

The measured liquor flow results were compared with simulated results as shown in *Table 14*. Process conditions vary a little at different production rates, because the experiments were done on different days. For example, with the production rate 2050 ADt/day the liquor-to-wood ratio was 4.8, whereas in the two other measurements it was 5.0. This explains the variation between different simulation results.

Table 14 Measured and simulated liquor flow times in different parts of the digester. Process parameters vary slightly between different measurements.

		2000 ADt/day		2050 ADt/day		2100 ADt/day	
		Digester Time, min	Simulator Time, min	Digester Time, min	Simulator Time, min	Digester Time, min	Simulator Time, min
Screen 1	Digester top part	82	80	60	83	50	76
Screen 2	Digester middle part	-	-	200	215	-	-

The simulated results differ somewhat from the measured liquor flow results. At a production rate of 2000 ADt/day, the liquor flow results are close to each other and at the highest production rate, the simulated flow time is 50 % greater than the measured value. It seems that the tracer test flows are not ideal. This is discussed more detail in next section.

12.3.3. Non-ideal liquor flow in digester

The liquor flow inside the digester was studied by adding lithium to the liquor flow. The lithium test was found to be an efficient way to study the liquor flow in the continuous digester, making it possible to get a more realistic picture of the flows inside the digester. As the case was with chip flow, the simulator assumes an ideal plug flow model. The information provided by the simulator was compared against real measurements.

The lithium tracer test was an easy way to evaluate the liquor flows inside reactor, in spite of some drawbacks. Due to the nature of the test, only residence time aspects of the flow can be measured that way. This type of tracer test is not very useful for mass-balance purpose. One reason is that part of the lithium interacts by ion-exchange with the chips. This causes that part of the lithium to move along with the chips and not with the liquor flow. This explains partly, why most of the tracer does not follow the liquor flow but goes with the chips.

The lithium test revealed some recirculation and short-cutting in the liquor flows. It was revised that there was no leakages in the process to cause any short-cutting (etc. valves, pipes or pumps). Also the experiments were repeated several times during the course of approximate one year with a major maintenance shut down in the middle of the year, where process equipment was repaired. The standard deviation of the tracer test was estimated to be $\pm 20\%$. The error itself cannot either explain all the short cutting effect, which happened during the first 10 minutes after the tracer addition.

The real causes of recirculation were difficult to determine, because some of the extracted liquor from Screen 1 was re-circulated to the digester top. Furthermore, sample collection is a slow method for determining short-cutting. Even though the process delays are really long for the continuous cooking process, liquor samples should be collected with short intervals, to make it possible to detect any short-cutting of the liquor flow. This means a very large number of samples need to be analyzed.

Figure 85 shows the results of the last tracer test at *Screen 1*. As can be seen from the figure, there was some recirculation in the tracer flow after 100 minutes. Recirculation means that part of the tracer is re-circulated to the process. In these tests, part of the cooking liquor was re-circulated from *Screen 1* to the digester top. The recirculation is reflected as a rise in the floor level of the lithium after the major pulse has passed *Screen 1*. The lithium concentration is over 10 % higher than the pre-test level in every experiment.

Both the recirculation and a possible dead zone cause the floor level of the lithium to rise. In these experiments, the recirculation overshadows the dead-zone effect. Based on these results, it is difficult to say anything definite about the proportion of the dead zone in the digester, because of the high recirculation flow in the digester.

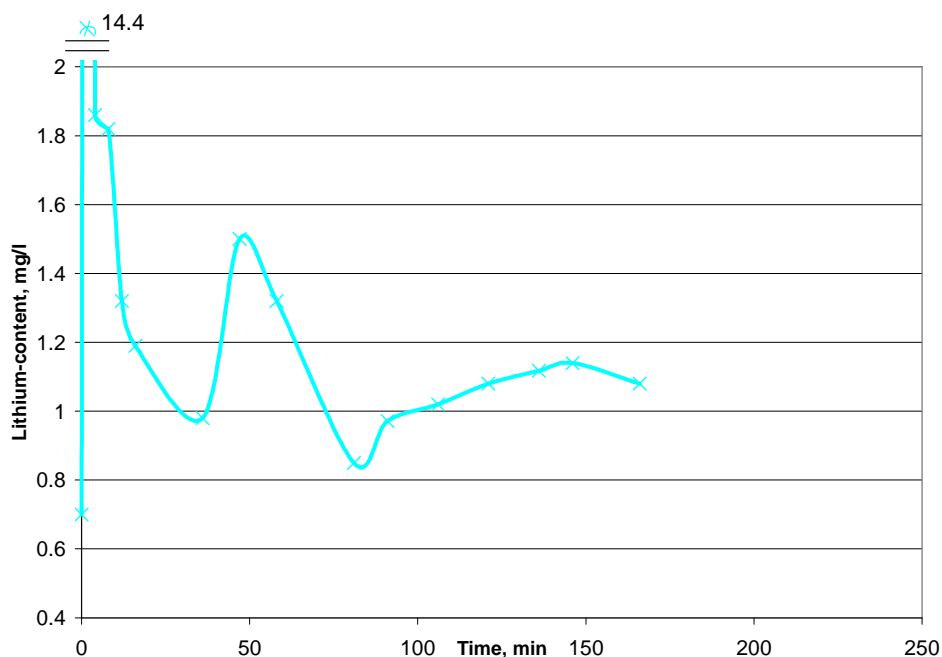


Figure 85 Tracer content in tracer test in the first extraction screen (Production 2100 ADt/day).

It is evident that there was some short-cutting in the digester (Figure 86). Because of the limitations of the sample collection method, the total amount of short-cut flow from the total flow is difficult to define.

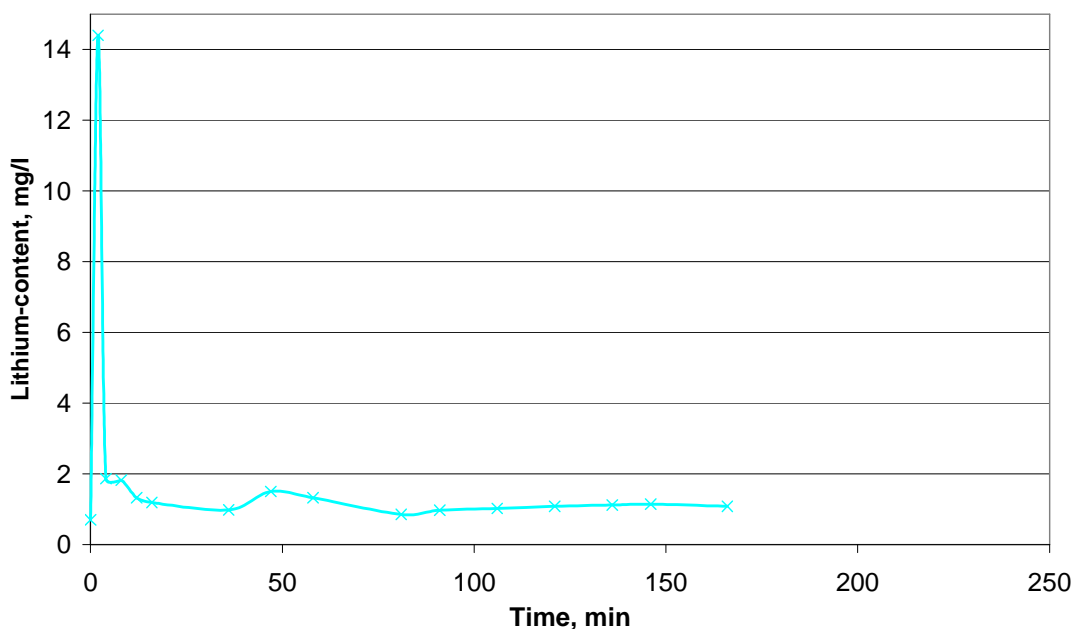


Figure 86 Tracer content in tracer test in the first extraction screen (Production 2100 ADt/day).

To gain a better understanding of the different flows inside the reactor, a simple mass balance was constructed. An example of the tracer test mass balance is shown in Figure 87. The amount of added tracer was 5000 grams. 20 % of the added tracer was found in *Screen 1* and 5 % of the tracer was

detected in *Screen 2*. It is likely that rest of the tracer leaves the digester with the pulp. Pulp samples were not analyzed within the tracer test.

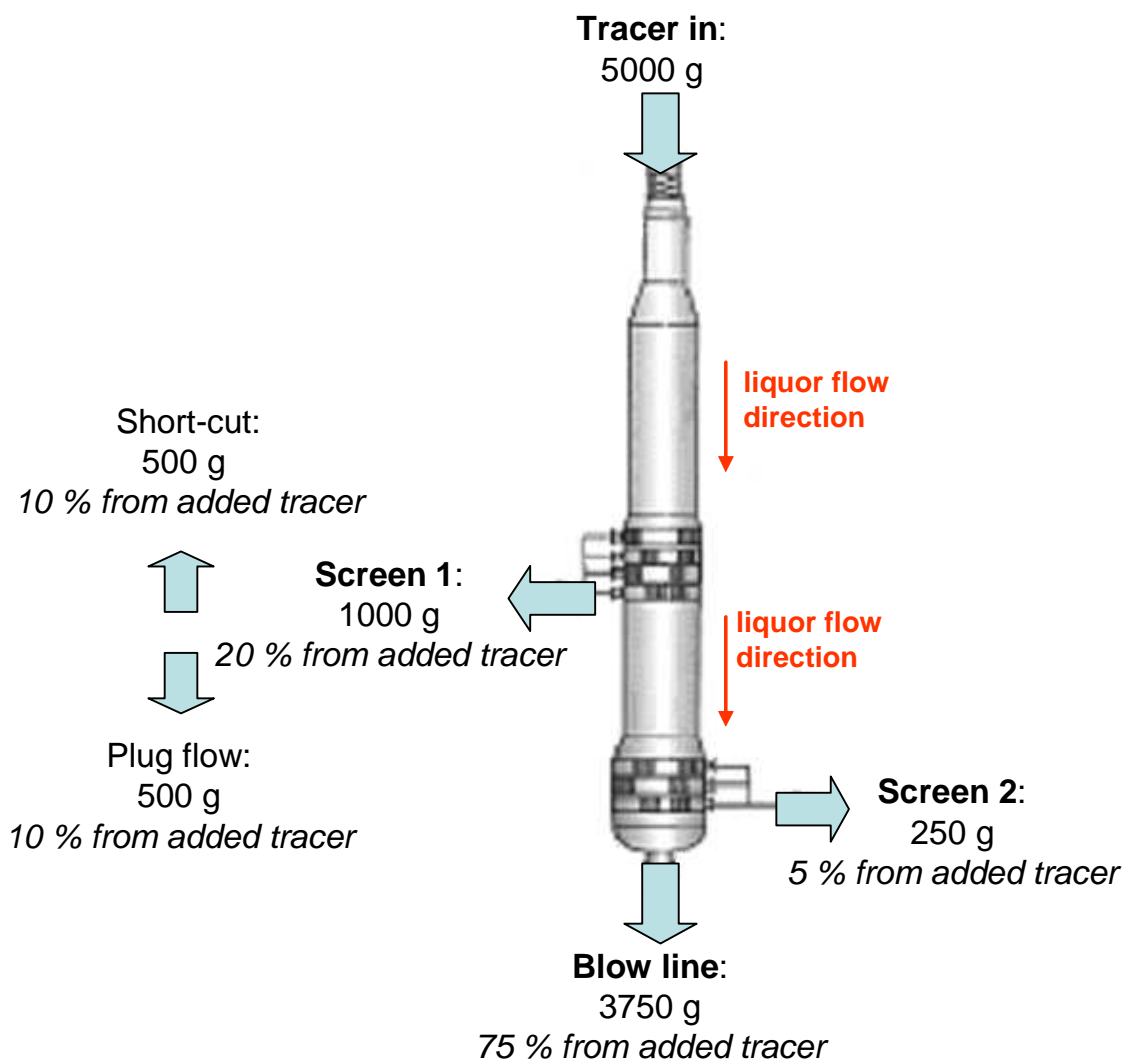


Figure 87 Hypothetical tracer flow scheme in digester.

Of the tracer detected at *Screen 1*, half is detected within 15 minutes and other half of the tracer was detected from 40 minutes to 90 minutes. In other words, half of flow was short-cutting and half of flow was plug flow type.

A general liquor flow balance of the tracer test is shown in *Figure 88*. Balance is based on mills real flows. The liquor flow into the digester is about 12 m³/ADt. The amount of liquor extracted from *Screen 1* is about 9 m³/ADt and that from *Screen 2* about 5 m³/ADt. The amount of washing liquor added to the bottom of the digester is about 9 m³/ADt and the blow line flow is about 7 m³/ADt.

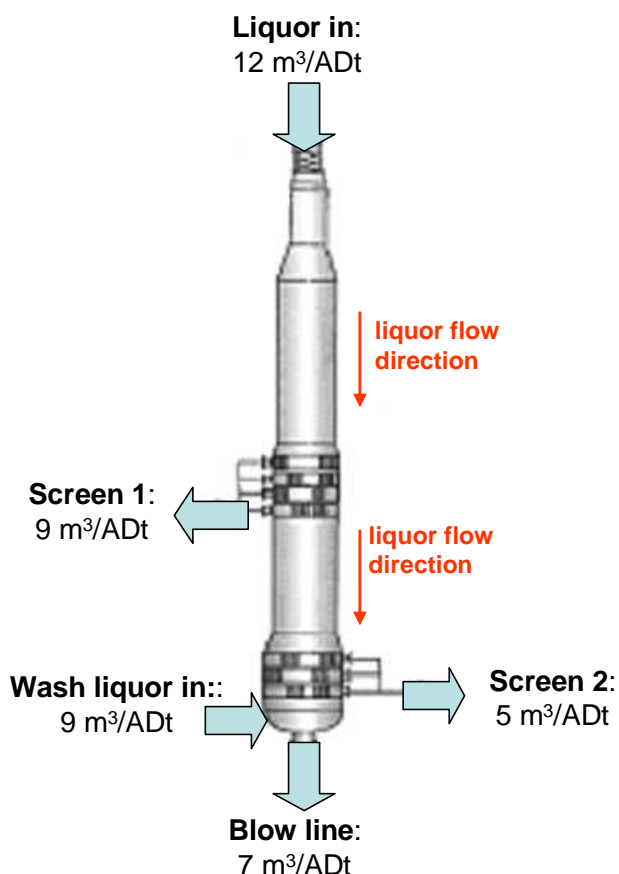


Figure 88 Flow balance of tracer test in digester.

When making a balance over Screen 1, 75 % of the liquor flowing with chips is taken out at Screen 1, but only 20 % of the added lithium is recovered at Screen 1. Half of the lithium observed at Screen 1 was found within ten minutes from the start of the experiment. In other words, the liquor flow in the digester is not very uniform.

It seems that most of the lithium flows with the chips, since only 25 % of the lithium is recovered from the screens during sample collection. Still, the flow path of the tracer cannot be determined in detail, because the tracer flow in the blow line was not examined.

12.4. Comparison of Härkönen's and Hernesniemi's parameters in TKK Packing Simulator

The packing model is based on the Ergun equation, which is commonly used for calculating the pressure drop in packed beds. Effect of the Ergun equation coefficients on chip bed packing is discussed in this section. Härkönen /39/ modified the equation by setting up two empirical coefficients to describe the shape factor of the chips (R_1 and R_2). Chip size affects the R_1 value: the smaller the chips, the bigger the R_1 value. This leads to a bigger pressure loss in the chip bed. R_2 is defined as multiplying the R_1 with a constant. Härkönen

developed these coefficients in laboratory scale digester. Hernesniemi /43/ fitted Ergun equation empirical parameters to better describe chip flow in modern continuous digester. These parameters are shown in *Table 15*.

Table 15 Ergun equation coefficients in Härkönen and Hernesniemi models.

	R₁	R₂
Härkönen	4600	3,900,000
Hernesniemi	2500	2,100,000

When comparing the values used in the Ergun equation, the original values given by Härkönen give smaller packing values than those given Hernesniemi. This is shown in *Figure 89*.

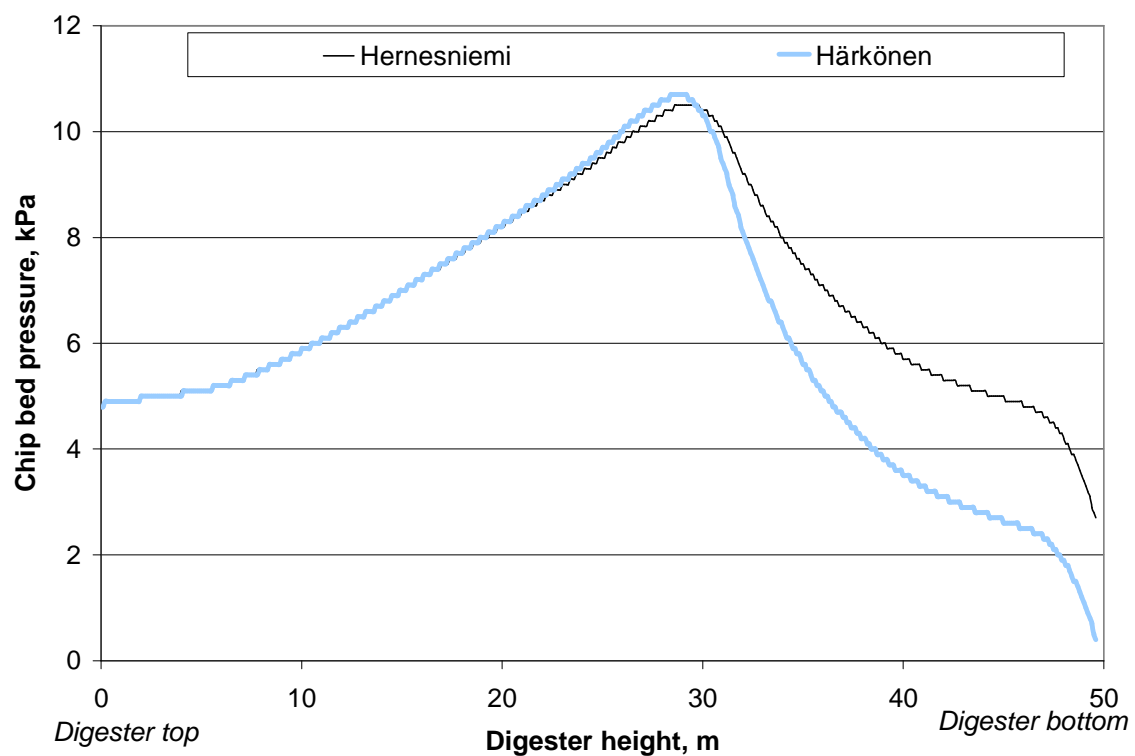


Figure 89 Chip bed packing using Hernesniemi's or Härkönen's values in Ergun equation.

The effect of the flow resistance is smaller, which is seen more strongly in the counter-current section of the chip bed. Accordingly, the packing effect and total cooking time (*Figure 90*) are smaller with Härkönen's values.

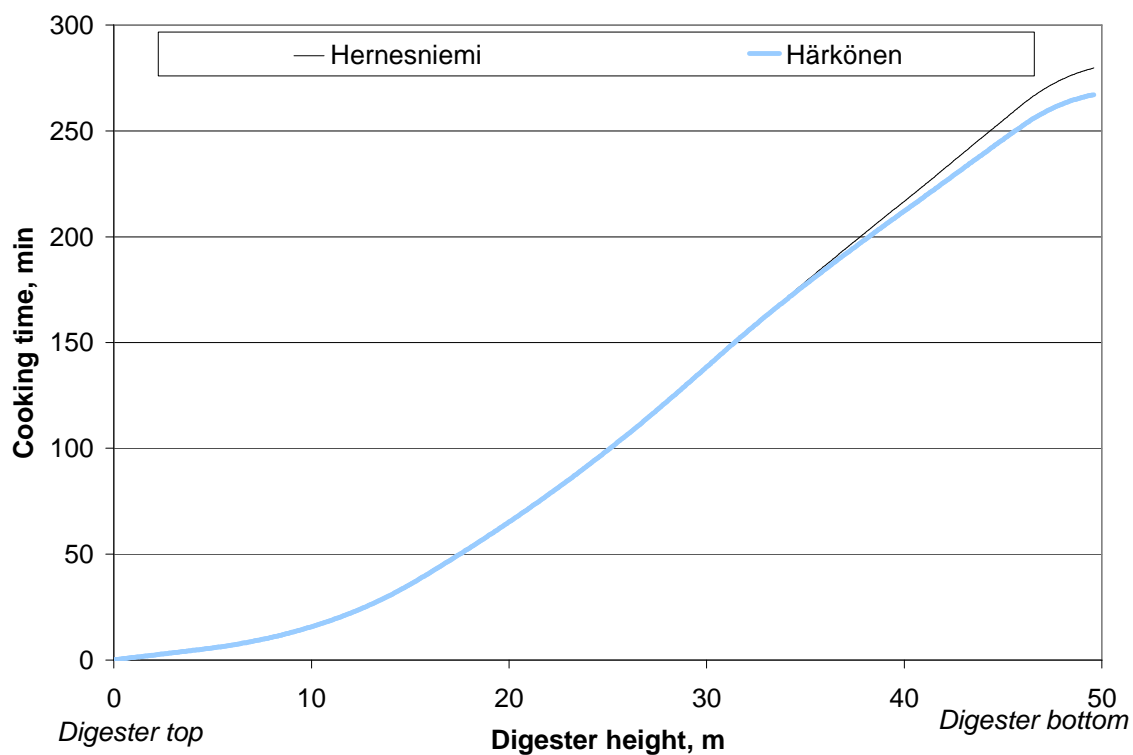


Figure 90 Total cooking time using Hernesniemi's or Härkönen's values in Ergun equation.

Hernesniemi's values give better accuracy when chip residence times are compared to real mill data, as shown in previous *Section 12.2*. Therefore, it is justified to use Hernesniemi's values in the Ergun equation.

13. Conclusions

The objective of the present thesis was to develop a model, which describes the packing conditions in a continuous digester chip column, flows in the chip bed and delignification along the digester. The purpose was to increase understanding and thereby control of the cooking process. The present new model was used for solving the vertical packing degree of the chip column along the vessel of continuous digester.

The present research work was based on the following hypotheses:

- 1) A chip bed packing model can solve the true chip residence time and process conditions in the chip bed.
- 2) Process conditions and mass transfer affect the delignification of the chip.
- 3) The chip bed packing degree is affected by chip softening caused by delignification.
- 4) Packing can be modeled by a combined model using a reaction-diffusion model and the Ergun equation for chip bed packing.

The developed TKK Packing Simulator combines a reaction-diffusion model for delignification and a chip bed packing model. The so called Gustafson model was chosen the reaction-diffusion model, because the model is very well validated and kinetic equations are fitted to Nordic softwood species. The so called Härkönen model was chosen as the chip bed packing model, because the model accurately describes packing phenomenon in chip bed.

The simulator was adjusted to describe full-scale industrial digester cooking Scandinavian softwood and the effects of various operational parameters were studied. The simulated packing and residence time was compared to the industrial operation by radioactive tracers. The simulator proved to give an accurate picture of the digester chip plug. Thereby, the set four hypotheses were shown to provide a good basis for the work and the objectives were met. This work shows for the first time that known theoretical models can be combined to form a practical industrial scale process simulator.

14. Aspects of applications

Basic research question about this work takes form in a question of the chip bed packing and modeling of these phenomena. This chapter discusses more detail about simulating the packing conditions in the continuous digester and how they affect on digester operation. General guidelines are given for operating the continuous digester. In addition, utilization of the simulator is discussed in more detail. Also more general discussion about conducting tracer tests in continuous digester given. Finally there is discussion about recommendations for future work.

14.1. Practical guidelines for running continuous digester

The following findings were made from simulation of the chip bed packing. The chip bed packing is affected by the process conditions and the chip properties. Most important process parameters are cooking conditions, chip and liquor level difference and feed liquor-to-wood ratio. The effects of selected initial parameters are shown in *Table 16*.

Table 16 Effect of selected cooking parameters on chip residence time in the digester and maximum chip bed packing. (When one parameter is increased, + means an increasing effect and – a decreasing effect on the variable.)

	Chip delay in digester	Maximum packing in digester
<i>Production</i>	-	(-)
<i>Feed liquor-to-wood ratio</i>	+	+
<i>Chip and liquor level difference</i>	+	+

An increased production rate decreases the total cooking time, because the volumetric flow increases inside the digester as the chips have to flow faster through the digester. The effect of the production rate on chip packing is more complicated.

If the kappa number profile (cooking degree at different parts of the digester) remains about the same at different production rates and temperature control works properly, there should not be any effect on packing. This needs good control and understanding the chip flow dynamics at different production rates.

If the production rate is decreased markedly from the nominal production rate, the packing is increased. This is because the digester flows are far from optimal and chips are flowing slower in the digester. As the lignin content of chip decreases, chip becomes softer. Chip bed becomes more compacted. For example, if the production rate is decreased from 2000 to 1400 ADt/day, the total cooking time increases by about 50 %, whereas the total production rate

decreases by 30 %. For this reason, all pulp mills with a continuous digester should define the most sensible production limits for the digester. The minimum value is more important, because the packing is more severe at lower production rates. In this way, digester control can be improved and severe packing conditions avoided.

The feed liquor-to-wood ratio affects the flow in the top part of the digester. As the liquor flow is increased, the total packing is increased, and as the packing is increased, the total cooking time is increased, too.

The chip and liquor level difference is important for steam/liquor-phase digesters. The chip and liquor level difference works as an initial packing force in the chip bed, forcing the chips to sink under the cooking liquor. The greater the level difference, the higher the initial packing.

It is difficult to define the optimum chip and liquor level difference (initial packing) in a steam-liquor phase digester, because every digester is unique. When comparing the findings of Joutsimo's study /56/ (on page 60) with the simulated results of the present study (see page 115), some differences in initial porosity can be noted. Joutsimo showed that in batch digesters the porosity of the chip bed should be less than 0.5 to minimize the kappa standard deviation. The initial porosity in the simulated base case was about 0.57. The chip and liquor level difference was 1 meter.

When the initial packing is increased in the continuous digester, cooking becomes more uniform. This is probably caused by the more uniform liquor flow in the digester, while channeling inside the digester is decreased. The initial packing also affects the maximum packing of the digester. As the initial porosity of the chip bed is decreased to 0.5, the maximum packing of the chip bed is increased somewhat. At the same time, the minimum value of the simulated porosity is decreased to 0.2, which means high packing in a certain part of the digester. High initial packing might lead to dense packing of the chip bed in the lower parts of the digester. The optimum chip and liquor level difference for the digester studied here is about 0.9-1.4 meters.

Simulator and mill data were used for building a general understanding of how to run the continuous digester under optimum conditions as a means to minimize the maximum packing of the digester. Based on the simulation results and mill measurements, the following guidelines can be given:

1. The porosity of the chip bed is an important factor for chip bed packing. The porosity of the chip bed should be less than 0.5 at the top of the digester. The porosity of the chip bed along the digester should not go below 0.2.
2. If the packing force exceeds 15 kPa, the chip bed might be plugged by the packing force. The effect of the packing forces on the chip bed is more drastic when the kappa number of chips is less than 60.

3. To minimize chip bed packing, the kappa number should be kept as high as possible before the blow line. For the digester studied here, better movement of the chip column is achieved if the kappa number at the Screen 1 is raised above 60.

These results can be generalized for modern steam/liquor-phase digesters running on Nordic softwood.

14.2. Utilization of the simulator

The simulator can be used to increase the knowledge level of the operating personnel in a pulp mill, and also used for process development for new and old processes. Different possibilities for utilizing simulator data are listed in the following:

1. Teaching and training of mill personnel
2. Preparation of operating manuals and searching for realistic operating conditions
3. Trouble-shooting and mill research
4. Planning new configurations in an old mill
5. Examining new investments, comparing different digester configurations
6. Planning of automation and process control
7. On-line and off-line monitoring and alarm systems for process operation
8. Development of pulp quality, optimizing fiber quality and minimizing fiber damage

Teaching and training is one possibility for wider use of the simulator. At present, the TKK Packing Simulator is being used in a pulping course in the Pulping Laboratory of the Helsinki University of Technology. The simulator is used to examine how different process parameters affect chip bed packing, for example to determine how high the chip bed can be in a conventional digester until the packing of the chip bed is increased too much. This increases the awareness of how different phenomena and situations affect digester runnability.

On the other hand, the simulator can be used as a part of a mill's operating manual and for searching for realistic operation conditions, making the users more familiar with the process conditions in the mill. The simulator can be used to determine the critical limits for different operating parameters, for example the upper and lower limits for the chip and liquor level difference, as a means guarantee smooth digester operation.

The simulator can be used as a planning tool for retrofits as well as for examining new investments. Critical process parts can be modeled with the

simulator, for example, to determine the loading of different screens and how different cooking methods differ from each other. At the same time, the level of knowledge about different cooking methods increases. In addition, information about the limitations and possibilities of different cooking methods becomes more readily available.

Measurement data, both mill data and simulator data, can be combined to produce quality data. The results can be used for steering the cooking process to produce better pulp quality. In this way, information can be combined for example to see how chip bed pressure affects pulp quality.

14.3. Further discussion about flow studies in continuous digesters

This section concludes the main findings about the flow studies in this thesis and gives further information about ways to investigate flows in continuous digesters. Chip and liquor flow in digesters was studied in this thesis. Radioactive tracer was used for studying the chip flow and LiCl-was used for studying the liquor flow. Underneath advantages and drawbacks of the measurements are listed. Also some instructions are given for flow studies.

Chip flow measurements using radioactive tracers:

- + Gives information along the digester
- + Method is useful for studying the channeling of the chip flow
- + Real time measurement
- Working with radioactive material needs a special caution and equipment
- Radioactive tracer measurement for chip flow studies is expensive
- Detection range of radioactive material is limited
 - Sometimes no information gained along the digester

Liquor flow measurements using tracers:

- + Quick and easy way to measure liquor flow in commercial digester, if there are sampling points for liquor in digester area
- + Lithium analyzes are inexpensive
- + Lithium does not affect the process
 - o Same lithium already in process
 - o Floor level of lithium in studied mill was 0.5-0.8 mg/l
 - o Lithium tracer pulse has to be big enough to be detect in analyzes
 - 5 kg of lithium was used in tracer test
- Sample taking takes a minimum of 3 minutes
- Back-mixing of liquor flows distracts the measurement

14.4. Recommendations for future work

The packing simulator presented in this work is a good tool for studying continuous digesters. Important future work can be related to hardwood modeling and simulator setup.

Recommendations for future work are given below:

1. Hardwood packing models should be included in the simulator. Both a packing model and a kinetic reaction model for different hardwood species are needed.
2. Radial flow in the digester should be modeled. Radial flow modeling would give more information about liquor flows in the digester.
3. Digester bottom washing area should be modeled in more detail. Computational fluid dynamics (CFD) could be used for modeling.
4. Chips should be modeled in 2-D or 3-D. At present, chips are modeled using a one-dimensional spherical model (1-D). It is accurate for chip softening at the moment, but when the packing model is developed further, it might be useful to model the chips in 2-D or even 3-D.
5. Heat from the reaction is not included. At the moment, the temperature change along the digester is estimated empirically based on real digester temperature measurements.

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