

IMPROVEMENT OF PULP STRENGTH AND YIELD BY COMPUTER SIMULATION OF LO-SOLIDS KRAFT COOKING

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ABSTRACT

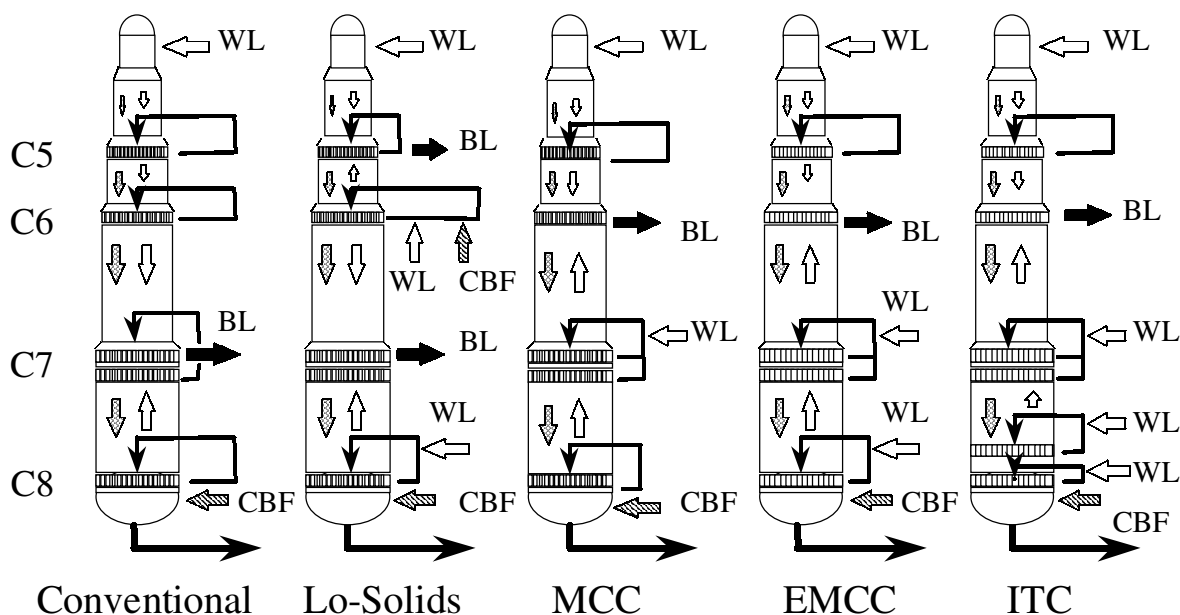
The computer model was developed to simulate the Lo-Solids Kamyr continuous digester. The model was a static model that simulated steady state processes for process optimization and retrofit feasibility studies. A new equation that described cellulose degradation by dissolved lignin was incorporated in the model, which made it possible to evaluate the effects of multiple black liquor extractions. One of the kraft digesters of Nippon Paper Industries was converted from the conventional cooking to Lo-Solids cooking. Comparing process data before and after the retrofit validated the computer model. Simulation shows that higher pulp viscosity and yield were obtained at a given kappa number by Lo-Solids cooking. The dissolution of the hemicelluloses was best understood by separately considering the two major types: glucomannan and xylan. Xylan remained relatively stable to alkali. The content of xylan in Lo-Solids cooked pulp was higher than that of the conventional pulp. As hardwood chips contain more xylan than softwood chips, it was predicted that the pulp yield increase in Lo-Solids cooking would be more pronounced in hardwood kraft pulp than in softwood kraft pulp. A low packing degree of chips was obtained in Lo-Solids cooking. The sulfide ion concentration was low in the initial phase and at the beginning of the bulk delignification phase in Lo-Solids cooking. There is a possibility for this drawback to be compensated for by making two kinds of white liquor with different sulfide ion concentrations and by feeding them to the different points of the digester.

INTRODUCTION

Kraft pulping is a complex mosaic of chemical and physical processes. The relationships between the rates of each process and the composition and temperature of the wood and liquor determine the behavior of the pulping system. An aspect of this behavior includes its ability to remove lignin without extensive attacks on the carbohydrate fraction of wood. In the ambition to force the kraft cook to low kappa numbers without losing pulp quality or yield, numerous studies have been carried out and a scheme for modified kraft cooking was developed. The basic conclusions were (10):

1. The alkali profile throughout the cook should be leveled out relative to conventional practice. In particular, a high alkali concentration at the beginning of the cook should be avoided and somewhat increased alkali concentration should be maintained towards the end of the cook.
2. The hydrogen sulfide ion concentration should be as high as possible in the initial phase and at the beginning of the bulk delignification phase. The concentration of free sulfide ion appears to be especially important. It has been shown that at the beginning of a conventional kraft cook there is typically a depletion of free hydrosulfide ion. This is probably the reason why lignin can sometimes undergo soda-type cooking reactions in the initial phase of the cook, which results in a residual lignin that is difficult to dissolve.
3. The concentration of dissolved lignin and sodium ion in the cooking liquor should be kept as low as possible, especially in the final phase of pulping, because the dissolved wood solids present in digester cooking liquor can cause:
 - a decrease in pulp viscosity and pulp tear strength
 - an increase in white liquor consumption
 - a decrease in pulp bleachability
 - a decrease in brown stock brightness
 - a decrease in final bleached pulp brightness ceiling
4. The temperature should be low in general, especially in the beginning and in the end of the cook.

The principles of modified cooking have been applied to continuous digesters by using split white liquor additions for alkali profiling and by using counter-current cooking methods to minimize lignin concentration at the end of the cook. The various types of modified cooking and a conventional cooking by single vessel steam/liquor phase digesters are presented in **Figure 1**. As compared with conventional kraft (CK) cooking, the cook starts at a reduced concentration of effective alkali and ends at a higher concentration of effective alkali. Modified Continuous Cooking (MCC) and Extended Modified Cooking (EMCC) are primarily designed to decrease the concentration of dissolved lignin in the final stage of cooking. In a single vessel MCC digester, cooking is divided into two zones: the co-current zone and the counter-current zone. The white liquor charge is split into three zones. EMCC was further developed in extended cooking. A third white liquor addition point is added to the wash circulation, and the temperature in the Hi-Heat zone is increased. Iso-thermal cooking (ITC) expands EMCC with an additional circulation loop. The ITC circulation operates at a very high volume and requires a special type of screen to handle the flow. The high circulation rates allow even more cooking in the counter-current zone. As compared with EMCC, the initial hydroxide is reduced. The temperature in the digester is lower and nearly uniform throughout the digester. The amount of cooking in the counter-current zone is increased. Lo-Solids aims at decreasing the concentration of all dissolved wood solids in both the bulk stage and the final stages of cooking. It derives its name from the multiple extraction flows, which lower dissolved lignin and sodium ion, thus improving selectivity. It is difficult to rank these systems in a particular order, since each has its own niche in particular applications. Wood species, pulp grades, system capacity, digester design, operating costs, capital costs, etc., all enter into the equation. It follows that the kinetics of kraft pulping is a subject of great importance. An understanding of pulping kinetics may be expected to provide a good basis for efforts to modify the existing process into a suitable cooking system.



WL = white liquor, BL = black liquor, CBF = cold black liquor filtrate

C5 = Upper heater circulation, C6=Lower heater circulation, C7 = Quench circulation, C8 = Wash circulation

Figure 1. Conventional and several types of modified cooking by single vessel steam/liquor phase digesters.

There was a need from the Production and Engineering Division of Nippon Paper Industries to our R&D Division to develop a computer theoretical model that describes the complex behavior of continuous kraft digesters. For a pulping model to be useful, it should predict the quantities that are measured in practice. The model should also be able to account for the main process variables in pulping. There are several mathematical models published in the literature. The characteristics of their models are summarized in **Table 1**. Models of No.1-9 can not simulate Lo-Solids cooking because they are unable to describe a decrease in pulp viscosity caused by the dissolved wood solids present in digester cooking liquor. In addition, they do not have the capability to independently simulate all of three important pulp qualities: lignin, carbohydrates and pulp viscosity. For example in Gustafson's model (6), the carbohydrate degradation rate is assumed to be proportional to the delignification rate.

The proposed model of No.10 in Table-1 is based on the mass and energy flow in the digester (14) and is an integration of the ideas behind the work of Smith, Williams (20), Christensen (4), Harkonen (7), Saltin (18), Agarwal (1,2), and Gustafson (3,6). In addition to their work, we created a new model to evaluate the effects of dissolved lignin in black liquor on cellulose peeling reactions to simulate Lo-Solids cooking. The model presented in this paper can simulate several types of Kamyr continuous digesters, estimate process variables, and predict pulp yield and viscosity at a given kappa number and at a given pulp production. The model is a static model that simulates steady state processes for process optimization and retrofit feasibility studies. The program is written using Microsoft FORTRAN Power Station version 4.0 on Windows95. This paper describes the following topics.

Types of digesters

- Conventional and Lo-Solids
- Single vessel with liquid/vapor phase using softwood chips

Estimation of process variables

- Residual effective alkali and dissolved solids in black liquor
- Temperature profile
- Chip compaction and retention time

Prediction of pulp quality and production

- Kappa number
- Pulp yield
- Intrinsic viscosity

The Kamyr kraft continuous digester of softwood species was rebuilt from conventional cooking to Lo-Solids cooking in a mill of Nippon Paper Industries. Using process data before and after the retrofit validates the computer model and provides a good understanding of the modified process that is achieved.

No.	Year	Authors	Scope of simulation						Type of digesters			
			Lignin	Carbo-hydrate	Yield	Vis-cosity	Com-paction	Rejects	CK	MCC	Lo-Solids	Wood species
1	1974	Smith & Williams (20) (Purdue Univ.)	Yes	Yes	Yes	No	No	No	Yes	No	No	Softwood Hardwood
2	1982	Christensen & Williams (4) (Purdue Univ.)	Yes	Yes	Yes	No	No	No	Yes	No	No	Softwood Hardwood
3	1983	Gustafson (6) (Univ. of Washington)	Yes	Middle	Middle	No	No	Yes	Yes	Yes	No	Softwood
4	1984	Sandstorm & Teder (17) (STFI)	Yes	No	No	No	No	No	Yes	Yes	No	Softwood
5	1987	Harkonen (7) (Kamyr AB.)	No	No	No	No	Yes	No	Yes	No	No	Softwood
6	1992	Saltin (18) (Univ. of Idaho)	Yes	Yes	Yes	No	Yes	Middle	Yes	No	No	Softwood
7	1992	Agarwal (1,2) (Univ. of Washington)	Yes	Middle	Middle	Yes	Yes	Yes	Yes	Yes	No	Softwood
8	1993	Michelson & Foss (12,13) (Univ. of Trondheim, Norway)	Yes	Middle	Middle	No	Yes	No	Yes	No	No	Softwood
9	1997	Krishnagopalan (21) (Auburn Univ.)	Yes	Middle	Middle	No	No	No	No	No	No	Softwood
10	1999	Miyaniishi & Shimada (14) (Nippon Paper Industries)	Yes	Yes	Yes	Yes	Yes	No	Yes	Yes	Yes	Softwood Hardwood

Yes: Possible to simulate, no: Impossible to simulate,
 Middle: Carbohydrate, viscosity or reject are the function of delignification and are not calculated independently.

Table 1. Mathematical models for kraft continuous digesters in the literature.

MODELING

Kraft Cook Kinetics

The kraft delignification process is known to proceed in three distinct phases: initial, bulk and residual delignification. During the initial phase, which accounts for 20-25% of the delignification, low molecular weight lignin and carbohydrates are dissolved. The main delignification takes place during the bulk delignification phase, which starts at about 150°C, during which lignin is degraded and dissolved by reactions with hydroxide and hydrogen sulfide ions. Cleavage reactions occur between the alkyl group and the aryl group in a variety of alkyl-aryl ether linkages, which hold together the majority of the phenyl propane units of lignin. Cleavage of non-phenolic β -aryl ether linkages seems to be the rate-determining step during the bulk delignification. The slow residual phase, which begins after about 92-94% delignification, is characterized by an increase in the amount of carbohydrates being dissolved with the lignin.

Several models for the reaction kinetics of the kraft cook can be found in the literature. The first attempt to model the kinetics is made by Vroom (22) when he developed the H-factor tables. Vroom finds that each H-factor corresponds to a specific lignin and carbohydrate content of the wood, independent of the temperature profile that had been used. However, the H-factor correlation only provides an accurate fit for data with similar cooking chemical charges and wood compositions. Hatton (9) describes the delignification by H-factor and effective alkali charge. The next level of reaction kinetic models includes the cooking chemical concentration. Several different models of this type can be found in the literature (4,6,11,15,16,20,21). They have the same general form.

$$-dW/dt = \{k_1[OH]^{-a} + k_2[OH]^{-b}[HS]^{-c}\}(W - W_\infty) \quad [1]$$

Where

W	= wood component concentration	kg/kg wood
W_∞	= wood component concentration after infinite time	kg/kg wood
k_1, k_2	= Arrhenius constant	$m^3/kg \cdot s$
a, b, c	= exponents in pulping kinetics	
t	= time	s

Gustafson's model (6) assumes that the rate of carbohydrate removal is proportional to the rate of lignin removal, which results in the same yield at any given kappa number. However, the carbohydrate removal changes with the cooking conditions. As this paper aims to evaluate pulp yield and pulp strength independently at a given kappa number, the kinetic model in this work is based on the work done by Smith, Williams (20) and Christensen (4). Namely a chip is assumed to have six components: high and low reactivity lignin, cellulose, galactoglucomannan, xylan and extractives. High reactivity lignin contains γ ether and low reactivity lignin contains β ether. The main xylan components of softwood and hardwood species are arabinoxylan and glucuronoxylan respectively. Extractives are assumed to dissolve immediately at the beginning of a cook. The removal of the other five components is modeled by five differential equations of the type shown in Equation [1], one for each component. These equations are used in three phases: initial, bulk and residual delignification.

Transport Between Free Liquor and Wood Chips

The development of rate expressions for kraft pulping is complicated by the heterogeneous nature of the system. One consequence of this is that the process may be considered to occur in a stepwise manner, as presented in **Figure 2**.

1. Transport of the ions from the bulk liquor to the exterior surface of the chip.
2. Diffusion of the ions to the interior of the chip.
3. Chemical reaction between the ions and the wood components.
4. Diffusion of the reaction products to the chip exterior.
5. Transport of the reaction products into the bulk liquor.

Under normal conditions, the transport steps 1 and 5 are not important but the diffusion steps 2 and 4 play a significant role unless the effective thickness of the chips is less than about 3 mm. The following assumptions are used in a model:

1. Ideal plug flow of wood chips in a digester.
2. Dividing the total height into 50 horizontal sections provides a method of predicting conversions and compositions at any point in the digester. This digester is divided into basically 50 continuous stirred differential reactions. With this assumption, there is a step change in conditions from one section to the next.
3. Each section is assumed to consist of three phases: free liquor, entrapped liquor inside a chip and a chip. Entrapped liquor moves with a chip and only free liquor is transported by a circulation pump and is extracted from a screen.
4. In each section, the temperature is assumed to be uniform in these three phases: free liquor, entrapped liquor and a chip.

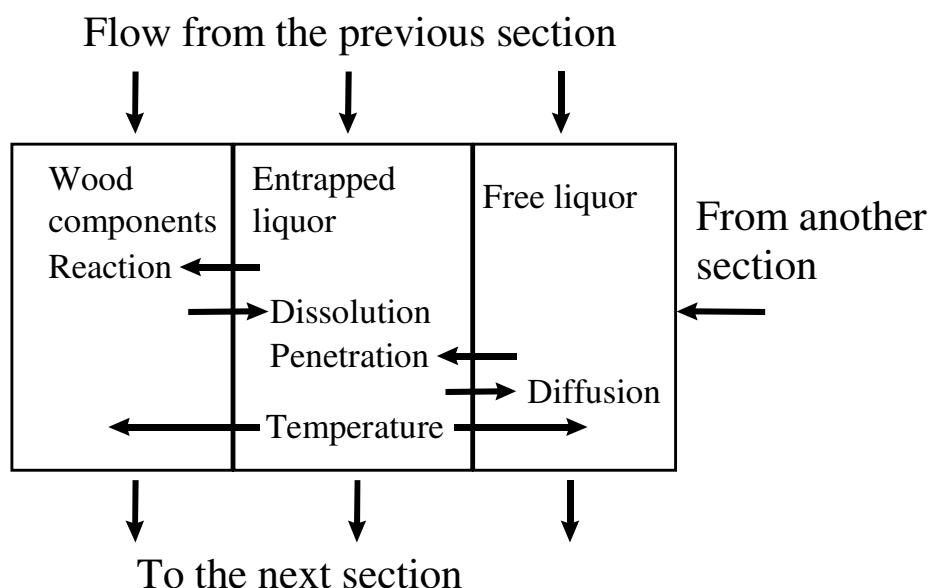


Figure 2. *Three-phase model of a section.*

In the beginning of the cook, cooking free liquor penetrates the wood chips. The penetration gives an even distribution of the cooking chemicals between the chip phase and the liquid phase. As soon as the temperature is high enough, the cooking begins and cooking chemicals are consumed. Wood components are transferred from the chips to the penetrated entrapped liquor. This leads to a concentration gradient between the entrapped liquor and the free liquor. This concentration gradient leads to the next stage in the transport process, the diffusion of chemicals and dissolved wood components between free liquor and entrapped liquor. This continues until the cook is finished. Previously, the diffusion has been modeled with two different approaches. The first approach by Christensen (4) assumes that the concentration in the chips is constant. This gives a simplified mass transport expression, Equation [2].

$$dX_{\text{liquid}}/dt = k_{\text{mass}}(X_{\text{chips}} - X_{\text{liquid}})m_{\text{chips}}A_{\text{chips}} \quad [2]$$

Where

X_{liquid}	= concentration in free liquid	kg/m^3
X_{chips}	= concentration in penetrated liquid	kg/m^3
k_{mass}	= mass transfer coefficient	
m_{chips}	= chip weight in segment	kg
A_{chips}	= specific surface are of chips	m^2/kg
t	= time	s

In the second approach, Gustafson (6) examines the concentration gradient inside the chips. Their model is supposed to give more information about the quality of the pulp. Gustafson shows that it is possible to calculate the screening rejects from the lignin profile. Gustafson's model gives more complex mass transfer calculations and it needs chip thickness. Our model is based upon the first assumption that the concentration in the chips is constant, because mill chips have a wide range of chip thickness and a number of the concentration gradient measurements inside chips for verification of Gustafson's model is difficult to perform.

Flow Characteristic of a Continuous Digester

A model of chip compaction in a digester is illustrated in **Figure 3**. The flow of the chips through the digester determines how long the chips stay in each section of the digester and is therefore important to the pulp quality. The downward velocity of the chips changes as the chips travel through the digester. The liquid flow, the degree of delignification, and the height of the chip column and the liquid level in the digester can cause this velocity change. The volume of a single chip is maintained practically unchanged as it moves through the digester. The reason for the varying compaction is that the shape of the chips changes as they move through the digester. Harkonen (7) derives a complex model of the two-phase flow inside a continuous digester. This model predicts the chip packing, flow of liquid, flow of chips and liquid pressure. The model involves elaborate solutions of two-dimensional equations and would be difficult to use in conjunction with the pulping reaction models. Saltin (18), using Harkonen's data and theory, derives a simplified one-dimensional model. The new model calculates the void fraction in the chip column, the liquid pressure, pressure on the chip column, the flow of chips and the flow of liquid. The liquid pressure drop as a function of the liquid flow through a packed bed of chips is described in Equation [3].

$$dP_{\text{liquid}}/dh = 4600\varepsilon^2/(1-\varepsilon)^3 \times v + 3.96 \times 10^6 \varepsilon^2/(1-\varepsilon)^3 \times v^2 \quad [3]$$

Void fraction of the chips in the chip column changed according to Equation [4].

$$\varepsilon = 0.356 - (P_{\text{chips}}/10^4)^{0.59} \times \{0.139 \ln(KN) - 0.831\} \quad [4]$$

The chip pressure, P_{chips} , is the contact force acting on the chip column divided by the cross sectional area of the chip column. This pressure is calculated using a force balance on the chip column. Three acting forces are gravity forces on the chip column, the friction between the digester wall and the chip column and the flow friction between the chip column and the liquid. The force balance is given by Equation [5].

$$dP_{\text{chips}}/dh = \{\rho_{\text{liquid}}(1-\varepsilon) + \rho_{\text{chips}}\varepsilon\} g - \mu P_{\text{chips}}/D_{\text{digester}} \pm dP_{\text{liquid}}/dh \quad [5]$$

Where

D_{digester}	= digester diameter	m
g	= gravity coefficient	m/s^2
h	= digester height	m
KN	= Kappa number	
P_{chips}	= chip pressure	Pa
P_{liquid}	= liquid pressure	Pa
ε	= void fraction of chips	
v	= superficial velocity in packed bed	m/s
ρ_{liquid}	= density of free liquid	kg/m^3
ρ_{chips}	= density of chips	kg/m^3
μ	= frictional coefficient between chips and digester wall	

The “ $\pm$ ” in the flow friction term is due to the possibility of counter-current and co-current flow inside the digester.

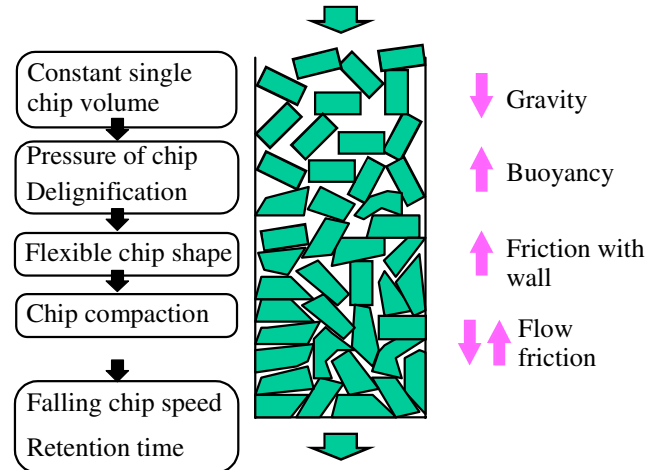


Figure 3. A model of chip compaction in a digester.

Cellulose Degradation Kinetics

Carbohydrate degradation and dissolution in kraft pulping reduces pulp yield and the strength potential of pulp. To improve the selectivity of delignification, it is necessary to better understand and quantify the kinetics of depolymerization reactions of wood carbohydrates. Agarwal (1,2) shows that a theoretical model for depolymerization is applicable to the degradation of cellulose in alkaline pulping. Changes in viscosity during pulping are mainly due to the secondary peeling of cellulose. The following equation is valid for secondary peeling of wood carbohydrates,

$$dDP_n/dt = -k_s DP_n^2 \quad [6]$$

$$k_s = A \exp(-E/RT)(OH) \quad [7]$$

Equation [6] can be derived from a theoretical treatment of secondary peeling of cellulose. The treatment assumes depolymerization is initiated by random chain-cleavage. Sjoblom (19) reports that alkalinity increases with an increasing concentration of dissolved lignin even at the same temperature. It is due to the changes of the acid-base equilibrium. Sjoblom reports that in a solution without lignin the alkali concentration at 25°C has to be 31% higher than that in a solution containing 44g/dm³ lignin in order to give the same alkalinity at 170°C. From his data, 0.31/44 = 0.007 a new model that describes cellulose degradation by dissolved lignin is proposed as Equation [8] in this paper.

$$(OH)' = (0.007 \times \text{lignin concentration} + \alpha) \times (OH) \quad [8]$$

Evans and Wallis (5) present the following relationship for cellulose viscosity and polymerization degree.

$$(DP_v)^{0.9} = 1.65 \times [\eta] \quad [9]$$

$$DP_v = 1.954 \times DP_n \quad [10]$$

Tappi viscosity, η_{sp} , is calculated from the following equation.

$$\log \eta_{sp} = \log(\eta/\eta_0 - 1) = \log(c[\eta]) + 0.13c[\eta] \quad [11]$$

Where

- c = solution concentration
- DP_n = number-average degree of polymerization
- DP_v = viscosity-average degree of polymerization
- k_s = rate constant for cellulose chain scission
- (OH) = alkalinity
- (OH)' = modified alkalinity
- α = constant
- η = intrinsic viscosity of solution
- η₀ = viscosity of CED (cupriethylenediamine)
- η_{sp} = Tappi viscosity

SIMULATION

Process Description

The softwood conventional cooking digester was converted to Lo-Solids cooking in a mill. The single vessel digester has the steam/liquor phase at the top. One to two meters of the chip column is above the liquor level. Having part of the chip column above the liquor level enables the packing degree to be controlled by adjusting the distance between the top of the chip column and the liquor level. The liquor level is normally held constant by a level controller regulating extraction. The added weight of the free chip column ensures that there is enough driving force to press the chip column to flow downward against the liquor flow. Before the retrofit, semi bleached softwood kraft pulp was produced for newsprint furnishes by the single vessel steam/liquor phase conventional digester. The pulp was bleached to 63% Hunter brightness by O-H bleaching sequence. But due to a decrease of newsprint basis weight and an increased usage of deinked pulp, the demand of semi-bleached kraft pulp for newsprint furnishes has decreased. To produce the market pulp, the semi-bleaching sequence O-H was modified to a full bleaching sequence O-D-Eop-D. This particular sequence has become the first ECF in Japan. It copes with the environmental restriction of AOX and chloroform. Two months prior to the start of the new bleaching sequence, the conventional digester was rebuilt to Lo-Solids. The objectives of the retrofit were:

- To decrease bleaching chemicals.
- To increase pulp strength.
- To stabilize digester operation.

Input Variables and Simulation Outputs

Process data taken before and after the retrofit was used for model validation and provided a better understanding of the new cooking process. **Table 2** shows the information that a user inputs into the model. After the retrofit, white liquor was split into three points: feed, lower heater circulation (C6) and wash circulation (C8). The flow rate of lower heater circulation (C6) and wash circulation (C8) was increased. The multiple extraction points resulted in an increased overall extraction flow rate from 118m³/h to 141m³/h, which improved washing in the Lo-Solids digester. Generally speaking, when continuous digesters are pushed beyond their design capacity, the filtrate applied in the wash zone begins to impede the downward movement of the chip column. Typically, a portion of the washer filtrate is diverted directly to evaporation in these digesters, which is particularly true of overloaded digesters with little or no upflow. The by-pass flow was avoided in Lo-Solids due to the low chip compaction. Temperature set points at the top and lower heater circulation (C6) were decreased and temperature set point of wash circulation (C8) was increased. The outputs of simulation are summarized in **Table 3**.

		Conventional	Lo-Solids
Chip meter	rpm	15	←
Wood density	BD.kg/m ³	394	←
Wood bulk density	BD.kg/m ³	140	←
Production	UKP.ADMT/D	340	←
Chip components			
Lignin	%	28.3	←
Cellulose	%	43.9	←
Galactoglucomannan	%	17.5	←
Arabinoxylan	%	6.6	←
Extractives	%	3.7	←
White liquor strength	AA as Na ₂ O g/L	114	←
Sulfidity	%	28	←
AA charge	%/BD chip	19	←
WL splits ratio			
Feed	%	100	54
Lower heater circ.(C6)	%	0	27
Wash circ. (C8)	%	0	19
Circulation flow rate			
Upper heater circ.(C5)	m ³ /hr	20	20
Lower heater circ.(C6)	m ³ /hr	130	195
Quench circ. (C7)	m ³ /hr	50	
Wash circ. (C8)	m ³ /hr	50	100
Extraction flow rate			
Upper heater circ. (C5)	m ³ /hr		60
Main extraction	m ³ /hr	118	81
Addition of CBF to			
Lower heater circ.(C6)	m ³ /hr		28
Bottom	m ³ /hr	166	166
Temperature set point			
Top	°C	148	132
Lower heater (C6)	°C	165.8	155
Wash circ. heater (C8)	°C	115	160

Chip components is given from the literature of Douglas Fir.

Temperature of C6 and C8 is that of heater outlet.

Table 2. User input for simulation of softwood kraft cooking.

		Conventional		Lo-Solids	
		Mill data	Simulation	Mill data	Simulation
Temperature					
Top	°C	148~152	148.0	132	132.0
Upper heater outlet (C5)	°C	158	154.2	135~140	136.7
Lower heater outlet (C6)	°C	164	165.8	153~159	155.0
Main extraction	°C	163	157.0	165~167	166.6
Wash circ. heater inlet (C8)	°C		94.0	130~145	136.2
Wash circ. heater outlet (C8)	°C	115	115.0	160~164	160.0
Blow line	°C	80	73.8	75~85	75.4
Effective Alkali as Na ₂ O					
Upper heater outlet (C5)	g/L	14	20.5	3~5	5.6
Lower heater outlet (C6)	g/L	12	16.9	15~19	19.6
Main extraction	g/L	8	9.2	8~11	8.6
Wash circ. heater inlet(C8)	g/L	3	3.1	12~14	14.6
Wash circ. heater outlet (C8)	g/L	3	3.1	18~22	21.0
Blow pulp					
Kappa Number		28~30	28.0	25	25.2
Yield	%		45.7		46.1
Intrinsic viscosity	dm ³ /kg		1008.0		1033.0
TAPPI viscosity	cP	27	27.2	29	28.9
Retention time	hr		6.7		6.1
Pulp components					
Lignin	%		2.0		1.8
Cellulose	%		36.8		36.9
Glucomannan	%		4.4		4.4
Arabinoxylan	%		2.6		3.1

Table 3. Output of simulation of softwood kraft cooking.

RESULTS AND DISCUSSION

Simulation results are given in Figures 4-14. Each zone is named by conventional cooking: the first zone is the upper impregnation zone, the second zone is the lower impregnation zone, the third zone is the cooking zone and the fourth zone is the Hi-Heat washing zone. Strainers separate these zones and section numbers shows the position of strainers in the horizontal axis of a graph. Addition and extraction of liquors (WL, BL) are made at the position of strainers.

Figure 4 shows the OH concentration profiles in free liquor, which are calculated for conventional and Lo-Solids cooking. Free liquor movement, multiple additional and extraction points of white liquor (WL) and black filtrate (BL) are also illustrated. Traditionally, the lower part of a continuous digester has been used for washing. It is referred to the Hi-Heat zone. In the upper part of the conventional digester down to the point where liquor is extracted, the flow is co-current with the liquor flow i.e. both are directed downwards. From the extraction zone, the

chip flow continues downwards but the liquor surrounding the chip moves upwards. This counter-current zone is used for Hi-Heat washing in conventional cooking.

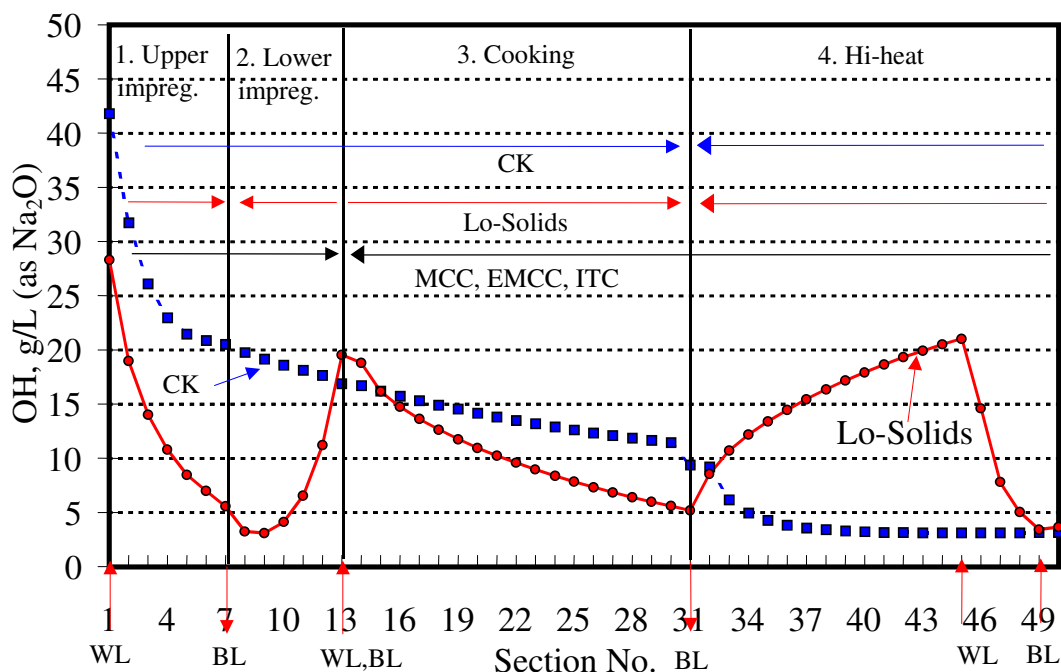


Figure 4. Simulation profile of OH concentration in free liquor.

Using this part of the digester for cooking without impairing the wash efficiency is the basic concept of a modified cooking. In Lo-Solids digester, multiple white liquor additions, multiple filtrate additions and multiple extractions are used. There are two points of liquor extraction. Extraction occurs at an upflow downflow interface, which is represented by the No.7 and No.31 sections in Figure 4. The flow above an extraction screen is co-current and the flow below the screen is counter-current. A heating circulation is not used at this point of extraction. In Figure 4, the No. 45 section represents make-up white liquor filtrate flows that are added to the digester in a downstream wash circulation. Preheated white liquor and filtrates are introduced in the center of the digester through the central pipe discharge. Part of the combined make-up, flows counter-currently upward and the balance flows co-currently downward. Two benefits gained by adding white liquor to the counter-current zone are that the initial hydroxide concentration is lowered, and some cooking is performed in the counter-current zone, where the dissolved solids are lower. Since there is a limit to the amount of filtrate and white liquor which can be added to the digester, the extraction liquor at each of the extractions should be as low in cooking residual alkali and as high in dissolved solids as possible. Minimizing the white liquor application to the feed accomplishes both of these objectives.

The free liquor effective alkali concentration at the two extraction points should be carefully monitored and be maintained above 3g/L-5g/L. Otherwise, serious cellulose degradation would occur when increasing the wood chip temperature without sufficient effective alkali in a chip. Kappa number is indicated by the effective alkali concentration at the first extraction point, which is continuously monitored by the online instrument. By adjusting the amount of white liquor added to the top of the digester, the effective alkali concentration at the upper heater circulation (C5) can remain constant. The blown pulp kappa number is monitored with an online

instrument and changing the white liquor charge at the wash circulation (C8) makes the final control of kappa number.

Figure 5 shows that the sulfide ion concentration is low in the initial phase and at the beginning of the bulk delignification phase in Lo-Solids cooking. This is the drawback of Lo-Solids cooking, and would be compensated for by a new chemical recovery process to make two kinds of white liquor with different sulfide ion concentration and to feed them to the different points of the digester. White liquor, which contains high sulfide ion, gets fed to the top of the digester, while the other white liquor, which is of low sulfide ion concentration, gets pumped into the initial cooking zone and the bottom of the digester.

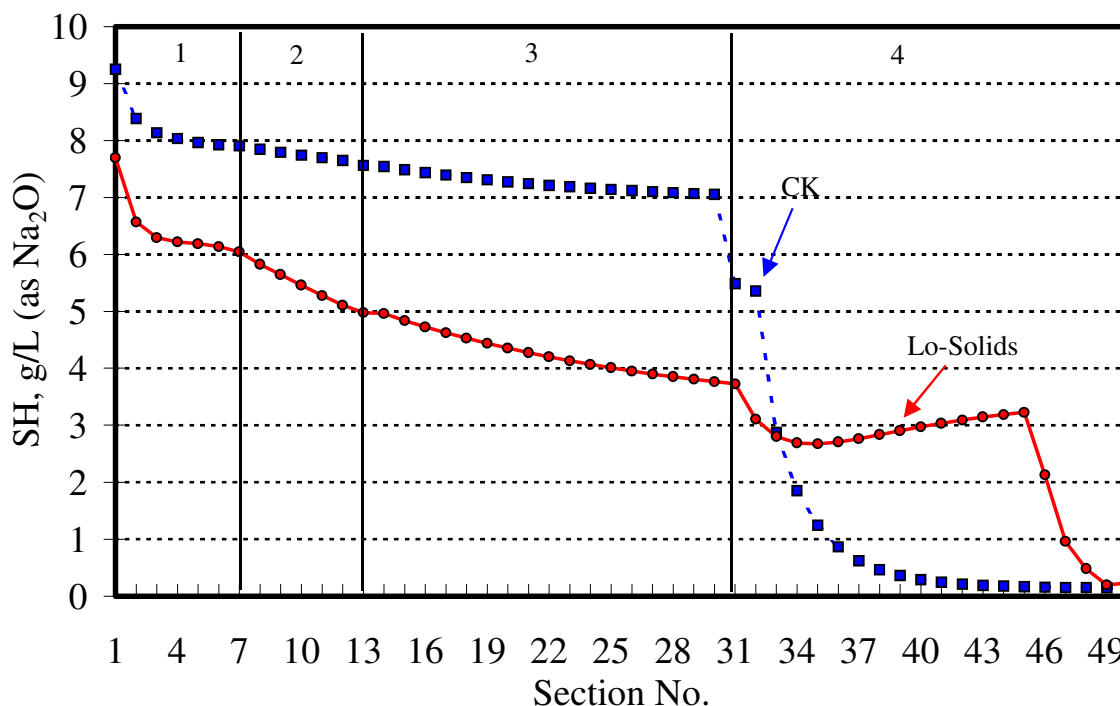


Figure 5. Simulation profile of SH concentration in free liquor.

Figure 6 shows that dissolved solids in free liquor are low in the cooking zone of Lo-Solids digester. In general, both the dissolved wood solids generated and the alkali consumed during impregnation are independent of the initial white liquor application, so long as there is sufficient amounts of white liquor present. Excess white liquor to the feed will increase the impregnation zone's liquor to wood ratio and therefore will decrease the concentration of dissolved wood solids in the spent impregnation liquor. Any white liquor applied to the feed which is beyond the amount consumed in impregnation will remain as residual. Shifting white liquor away from the feed, results in increased concentrations of dissolved wood solids and decreased alkali residuals for spent impregnation liquor. The amount of dissolved wood solids removed from the system for a given post impregnation extraction flow rate will therefore increase. At the same time, the amount of alkali removed from the system will decrease.

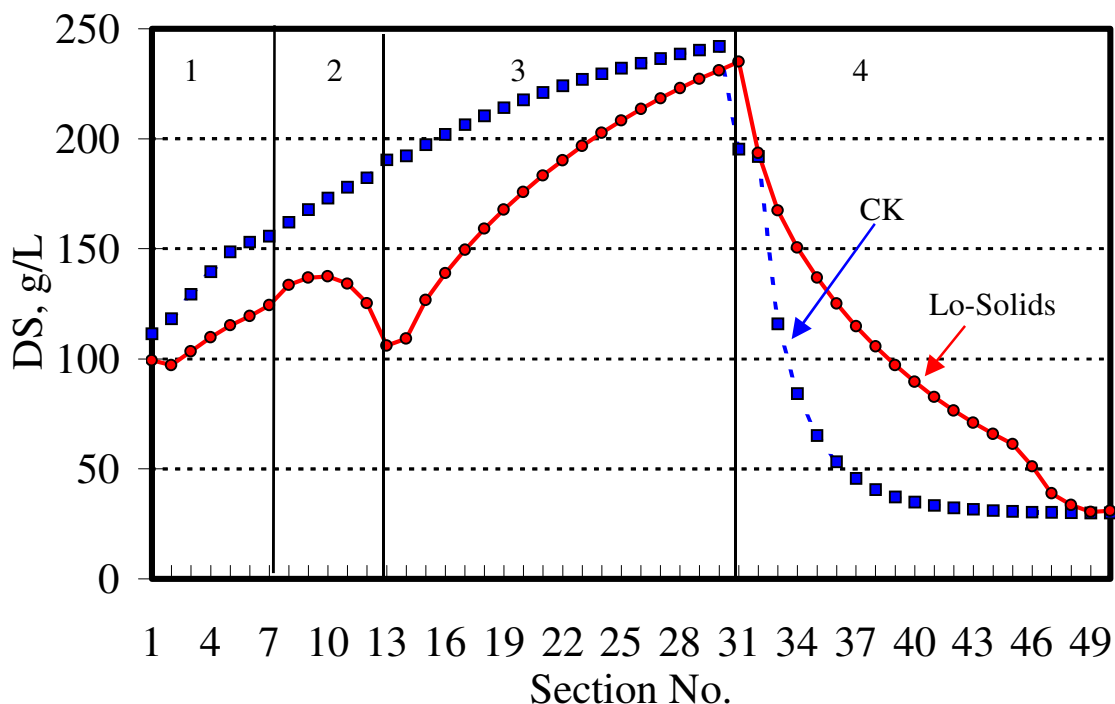


Figure 6. *Simulation profile of dissolved solids concentration in free liquor.*

Figure 7 shows that the top temperature is very high in the conventional liquid/vapor phase digester. After the rebuild into Lo-Solids, the top temperature decreases to a large extent. The temperature of the Lo-Solids digester gradually increases and at the beginning of the second zone, ultimately reaching the same temperature with that of the top of the conventional digester. During the first, second and third zones, the temperature is always lower in Lo-Solids than in conventional. As the top temperature is high in the conventional cooking, the delignification is fast in the upper impregnation zone. **Figure 8** shows that the initial delignification phase is not observed in the conventional cooking. After the rebuild to Lo-Solids, the top temperature is much lower than that of conventional cooking and the removal of lignin is slow at first. The independence of delignification rate and alkali concentration suggests that lignin removal during this part of the cook is not chemically controlled, but rather is diffusion-limited. As the temperature increases, delignification accelerates and an alkali concentration effect becomes apparent. Subsequently, a marked effect of alkali concentration is seen, suggesting that the chemical reaction becomes the rate-controlling factor at intermediate cooking times. Later in the cook, the lignin content levels off quite sharply and appears to approach a limiting value, which is lower at higher alkali concentrations. In the mill, the low initial temperature of the Lo-Solids digester made for better impregnation of the alkali solution into wood chips, and the prolonged cooking time resulted in a more uniform cook. Rejects of the blown pulp decreased by half. Higher kappa numbers at two extraction points reduced plugging problems at the screens, increased extraction flow, and achieved stable chip level control.

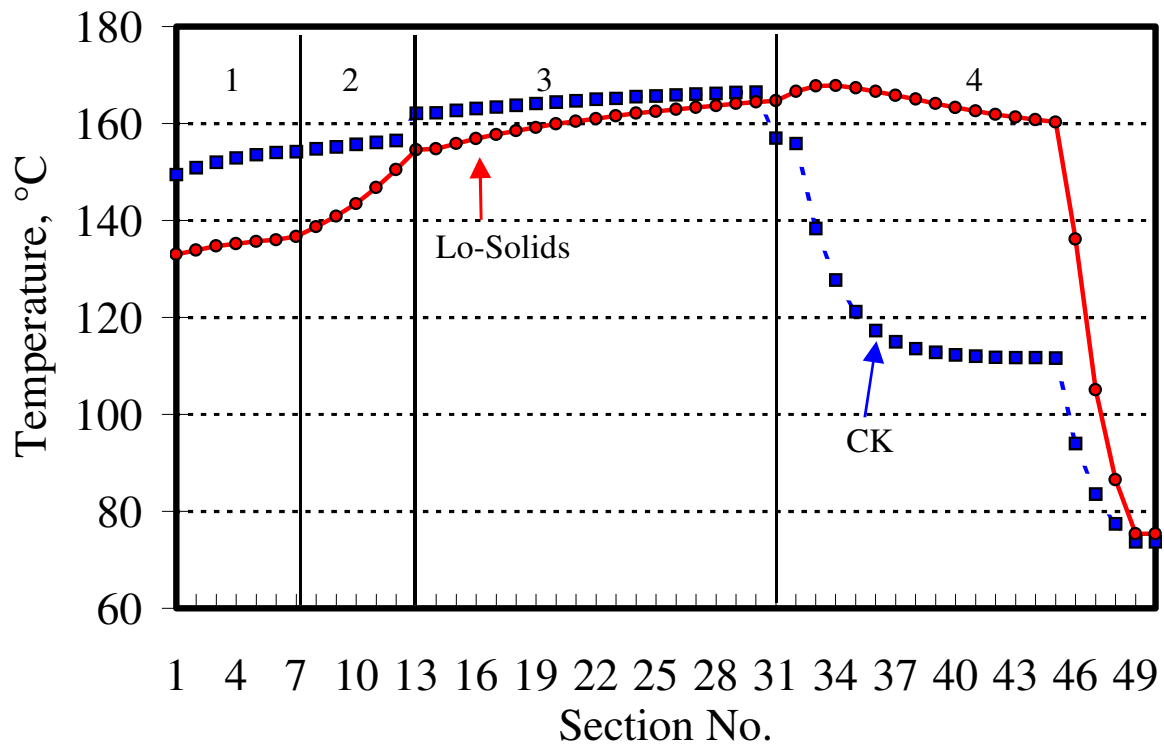


Figure 7. Simulation profile of temperature.

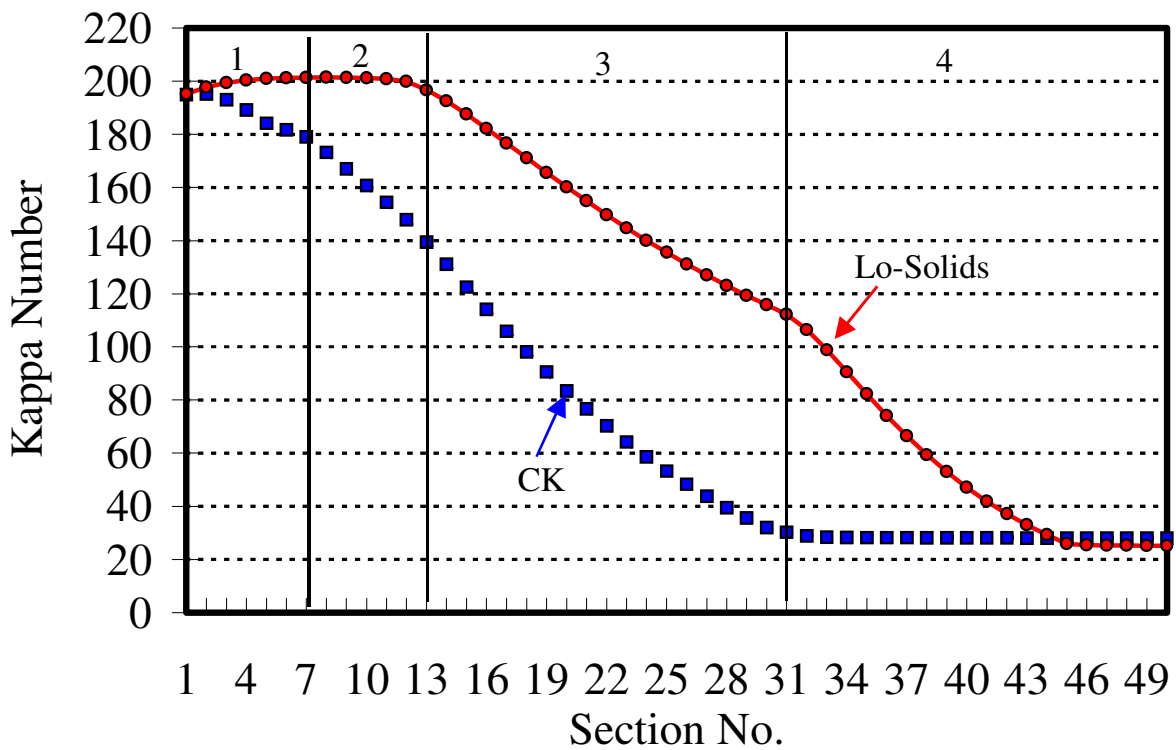


Figure 8. Simulation profile of kappa number.

The dissolution of the hemicelluloses is best understood by separately considering the two major type -glucomannan and xylan. **Figure 10** shows that glucomannan is removed at a relatively rapid rate at the onset of the cook, and this rate becomes even greater as the temperature exceeds 100°C. It is due to the dissolution of the soluble fraction. When about 70% of the glucomannan originally present has been removed, the rate drops off sharply. It may be inferred that this acceleration is due to peeling reactions. No difference is observed on the amount of glucomannan in blown pulp between conventional and Lo-Solids cooking. **Figure 11** shows that the pattern of xylan removal is similar to that of lignin. It is comprised by an initial slow period, and is accelerated as the maximum temperature is approached and subsequently levels off. Dissolution, not peeling, is the dominant mechanism, since xylan is relatively stable to alkali. A stable substituent on the C-2 of a terminal aldose would prevent the peeling reaction by blocking the formation of ions. The content of xylan in Lo-Solids pulp is higher than that of conventional pulp. As more xylan is present in hardwood chips than in softwood chips, it is predicted that the pulp yield increase in Lo-Solids cooking would be more pronounced in hardwood kraft pulp than in softwood kraft pulp.

Cellulose removal starts at 120-130°C and levels off when the maximum temperature is reached, which is probably due to peeling reaction, initiating at higher temperatures by alkaline hydrolysis of glucosidic bonds. An important reaction that does not lead directly to the removal of any material from the wood is hydrolytic cleavage of cellulose. This results in a drop in the degree of polymerization or molecular weight of the cellulose polymer. It is manifested as a viscosity reduction of the pulp solution in a solvent such as cupriethylenediamine. If it proceeds to a great extent, it will also appear as a loss in pulp strength, especially after bleaching. **Figure 9** shows that Lo-Solids cooking produces pulp with higher TAPPI viscosity.

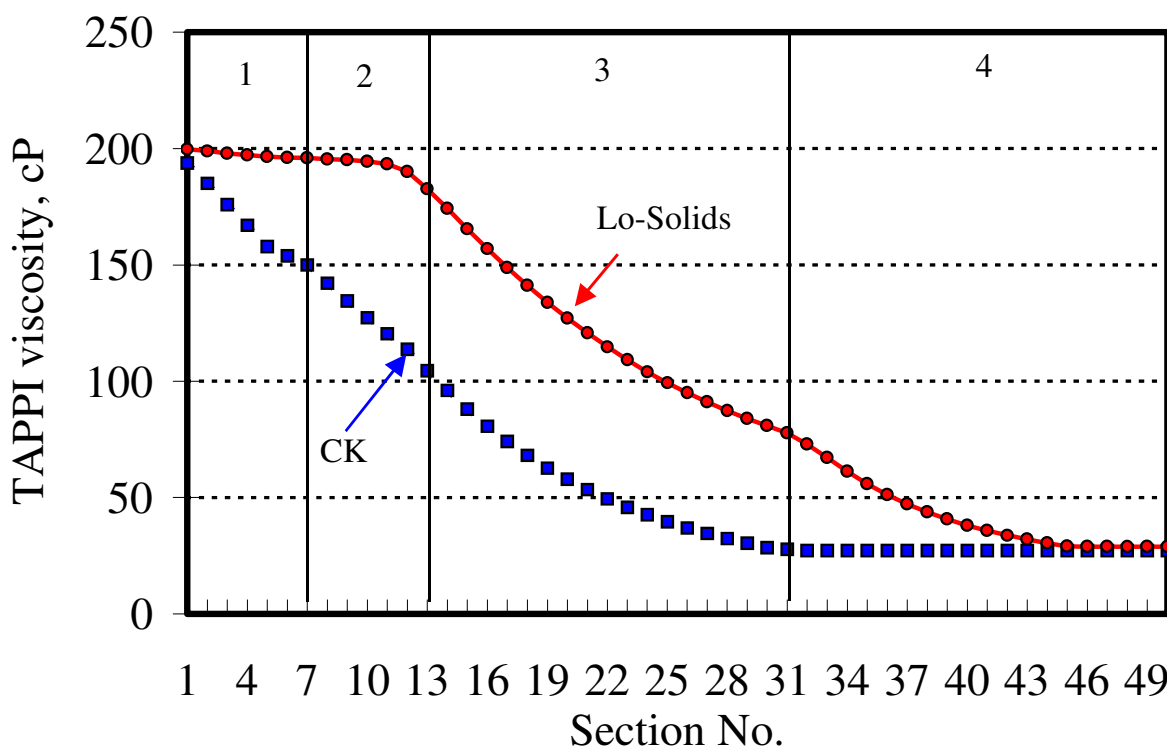


Figure 9. Simulation profile of TAPPI viscosity.

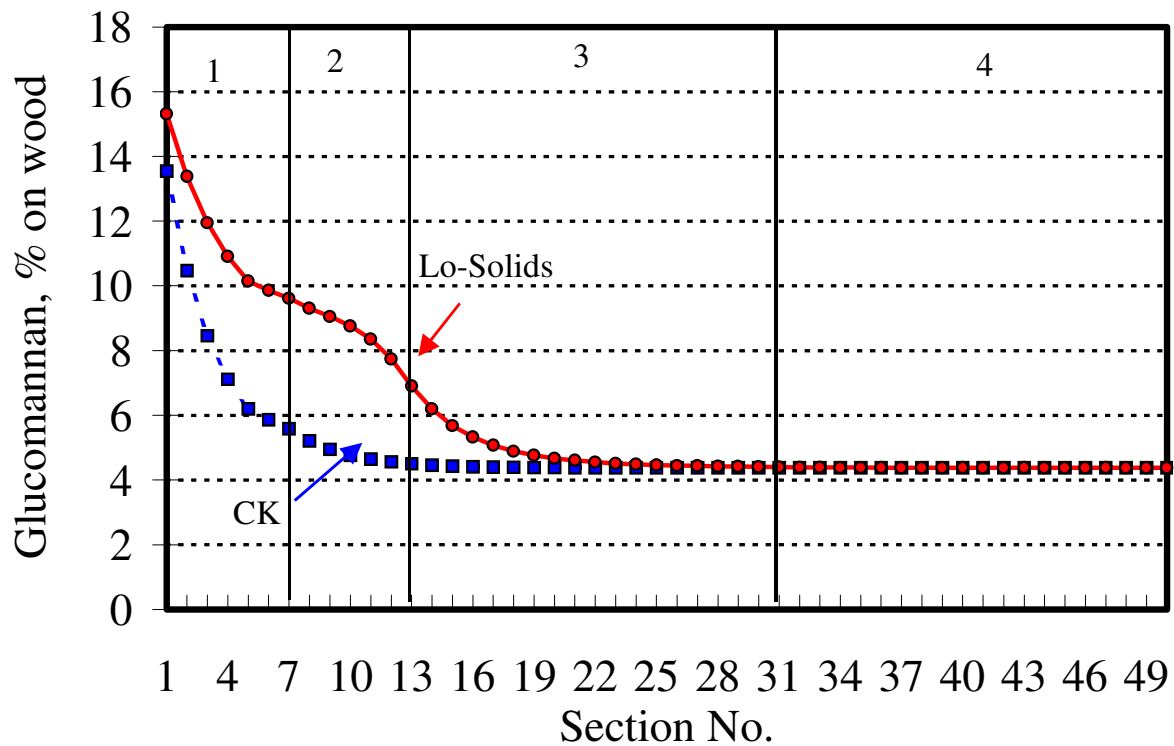


Figure 10. Simulation profile of glucomannan.

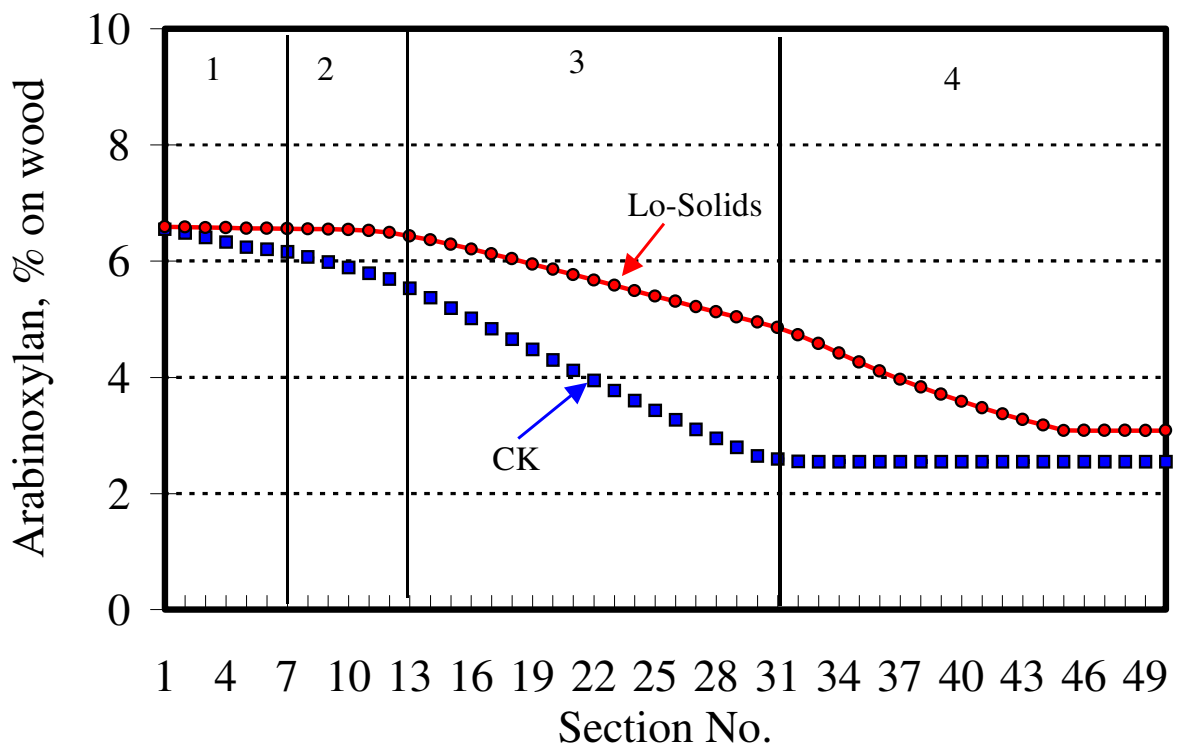


Figure 11. Simulation profile of arabinoxylan.

As previously shown in Figure 1, the third zone between the lower heater circulation (C6) and the main extraction screen is the largest zone where most of the cooking proceeds. The liquor flows in this zone of Lo-Solids are different from MCC, EMCC and ITC. Only Lo-Solids has the

co-current liquid flow in this zone and the downward movement of the chip column is not hindered by liquor upward flow, with chips falling smoothly in the digester. This is one of the reasons for the stable chip level at the top of the Lo-Solids digester and a stable kappa number after the retrofit was observed in the mill. MCC, EMCC and ITC have the counter-current flow in this zone. This free liquor flow is against the smooth downward movement of chip flow and has a tendency to result in “hanging” and unstable chip level. If the packaging degree becomes too high, a very compact chip column will form and movement of free liquor will be impeded. **Figure 12** shows a low package degree in Lo-Solids cooking. It will result in a soft chip column, giving room for the finer fractions of the chips to move smoothly. Table-3 shows that the retention time of Lo-Solids is shorter than that of conventional. It comes from the lower chip compaction.

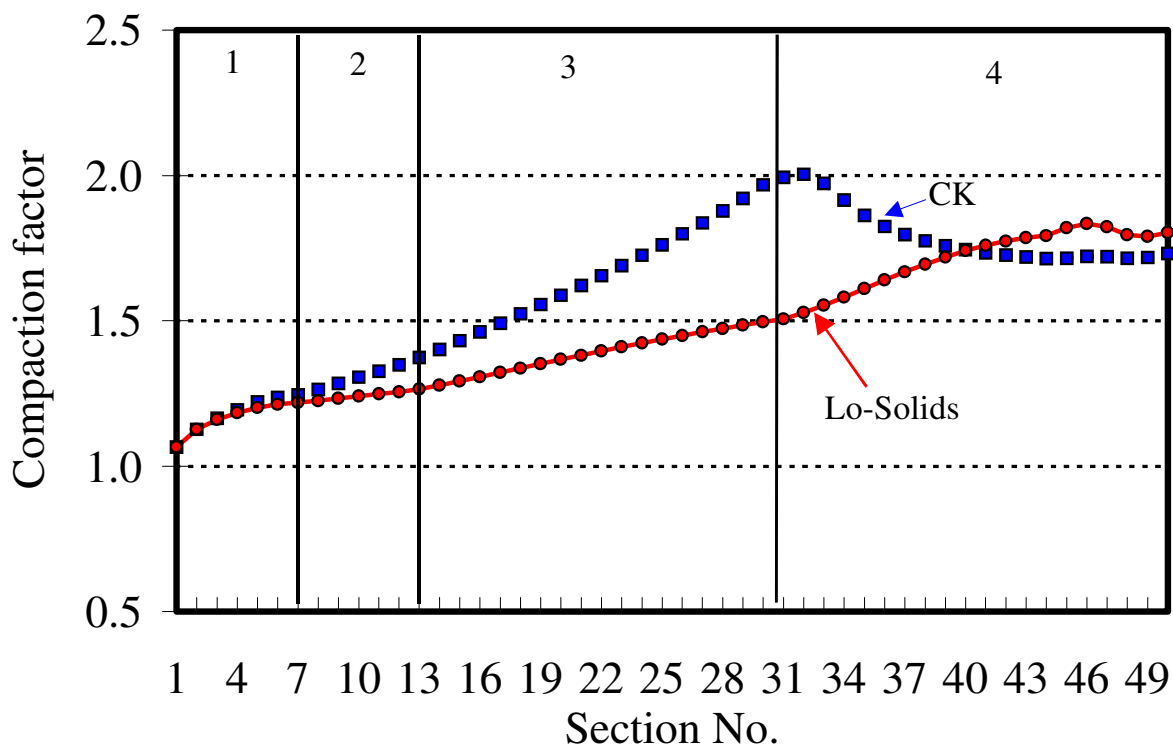


Figure 12. Simulation profile of chip compaction.

Figures 13 and **14** show that higher pulp yield and viscosity are obtained at a given kappa number by Lo-Solids cooking. Figure 14 shows that pulp viscosity simulation fits well with measurement data. As there is not an established method to accurately measure pulp yield in a mill digester, comparison can not be made on pulp yield between simulation and measurement.

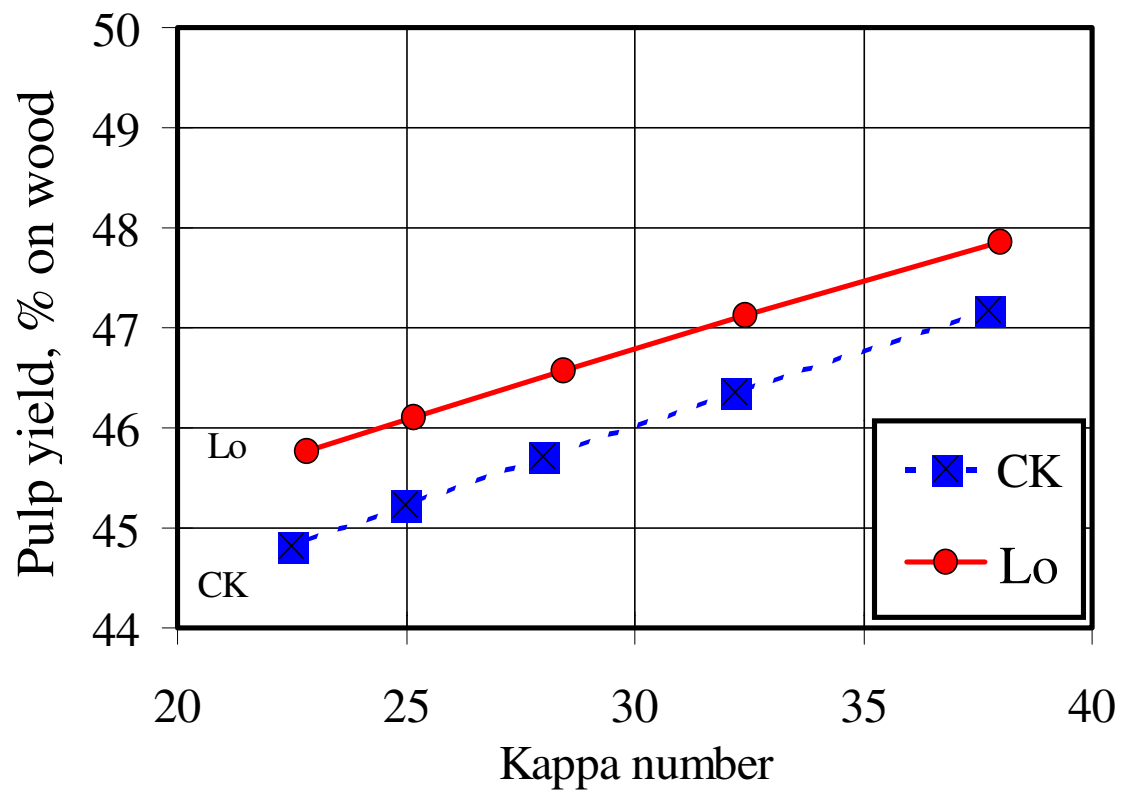


Figure 13. Simulation of kappa number vs. yield.

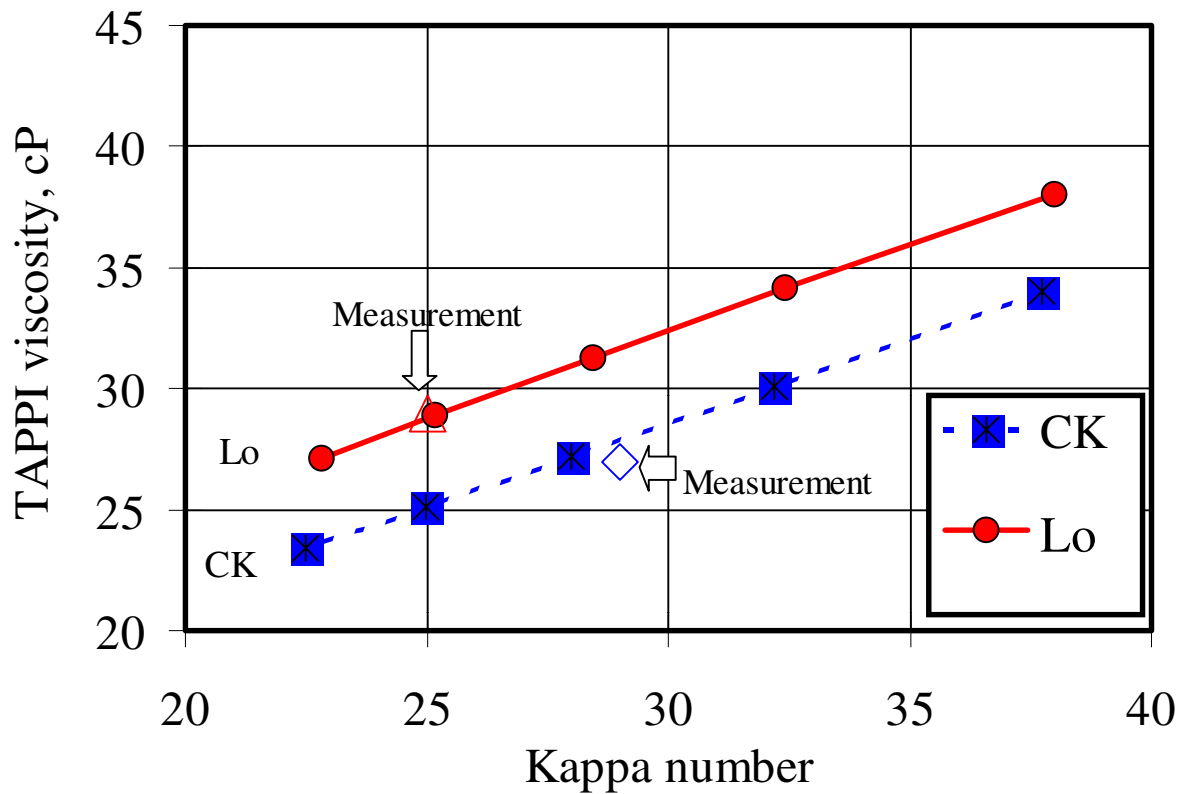


Figure 14. Simulation and measurement of kappa number vs. viscosity.

CONCLUSION

The computer model was developed to simulate the Lo-Solids Kamyr continuous digester and to make comparison of its performance with the conventional digester before and after the retrofit. The Hi-Heat washing zone was used for cooking without impairing washing efficiency. Simulation outputs fit very well with measured process data. Simulation shows that temperature decreases in the top, impregnation and cooking zones were observed in Lo-Solids cooking. Higher pulp viscosity and yield were obtained at a given kappa number by Lo-Solids cooking.

Multiple white liquor additions, multiple filtrate additions, and multiple extractions leveled out the alkali profile throughout the cook. The sulfide ion concentration was low in the initial phase and at the beginning of the bulk delignification phase in Lo-Solids cooking. This was the only drawback of Lo-Solids cooking, and could be compensated for by making two kinds of white liquor with different sulfide ion concentrations and by feeding them to the different points of the digester. Multiple filtrate additions and multiple extractions decreased the dissolved solids concentration in the free liquor in the digester.

In the mill digester, effective alkali concentrations at the two extraction points were a good estimate of kappa number and were continuously monitored by online instruments. In order to control kappa number, effective alkali concentrations were maintained constant by adjusting the multiple white liquor flows. The cooking temperature primarily controlled kappa number. Additional control of the kappa number was achieved by changing the temperature and the white liquor charge at the wash circulation (C8).

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