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Simulation and energy optimization of a pulp and paper mill – Evaporation plant and digester

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ABSTRACT

A detailed mathematical process integration model of a pulp and paper mill in the Northern Sweden has been developed. The main objective of this work has been set to describe the practical development of the model with particular emphasis on the development of the digester and evaporation plant sub-models. Actual plant measurements have been used to validate the model. By implementing the sub-models into the complete plant model, the influence of different operation parameters on the overall plant performance has been investigated. Furthermore, introductory studies with the main objective to minimize the plant energy cost have been carried out.

1. Introduction

The pulp and paper industry is an energy-intensive industrial sector that faces several challenges such as increased competition and rising feedstock and energy prices [1]. To address this, it is crucial for the industry to improve the material and energy efficiencies to the highest possible extent. Process integration methods provide useful tools for evaluating possible process alternatives. Several studies of paper and pulp mills have been carried out by using pinch analysis [2–8] and mathematical programming [9–15]. However, the scope of modelling has not been as complete and detailed as it is in many other process industries. More detailed modelling is required especially as large efforts are currently put on turning pulp mills into bio-refineries.

In our research group, mathematical process integration models with detailed material and energy balances have been developed for the steelmaking [16–20] as well as mining [21,22] industries based on mixed integer linear programming (MILP). The analysis was carried out using the reMIND software [23] in combination with the commercial optimization software CPLEX [24]. Recently, the research was extended to a pulp and paper mill located in the Northern Sweden [25] with the aim to produce results that were useful for practical plant decisions. Pre-studies indicated that the plant model should include very detailed sub-models, especially of the evaporation plant and digester. As the risk of errors and undiscovered bugs increase dramatically with increasing degree of complexity and detail level, a stepwise mode was applied, where the sub-models were firstly created and validated individually after each step, and then merged into a complete MILP model which then also was validated against plant measurements and experiences. In addition, a modular structure of the model was adopted to be able to study new process alternatives simply by adding new modules.

The main objectives of this work have been set to describe the practical development of this detailed model using a step-wise mode with a modular structure with particular emphasis on the development of the digester and evaporation plant sub-models. Furthermore, results from the modelling regarding optimization of various plant parameters are presented. Due to that the modelling approach is applied on a specific case, the results cannot be considered to be general. However, the approach itself is.

2. Process description and modelling

2.1. Process description

The mill can be divided into two major processing lines: the fiber line and the chemicals recovery. The fiber processing line extends from the digester to the pulp bleaching/paper making section. The main task of the fiber processing line is to remove the lignin from the wood to achieve bright pulp and paper. The chemical recovery cycle is necessary to make the process economically feasible. The by-product, black liquor, is concentrated in a multi-effect evaporation plant and burned in a recovery boiler, where the combustion of organics provides energy to produce high

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Nomenclature

c	price of an energy carrier (€/ton or €/MW h)
C_p	heat capacity (kJ/(kg °C))
f	flow rate (ton/h)
f_{in}	flow rate of light liquor to evaporation plant (ton/h)
$f_{liquorin}$	flow rate of liquor to effect (ton/h)
$f_{liquorout}$	flow rate of light out of effect (ton/h)
$f_{liquorsteamt}$	flow rate of steam out of effect (ton/h)
f_2	flow rate of liquor steam to stripping (ton/h)
h	enthalpy (kJ/kg)
m	energy carrier flow rate (ton/h)
p	absolute pressure (bar)
p_E	absolute pressure of the effect (bar)
q	electricity (el) flow rate (MW h/h, MW)
t	temperature (°C)
t_3	temperature of the liquor to $E3$ (°C)
t_{end1}	temperature of the digester after steaming (°C)
t_{end2}	temperature of the digester after heating (°C)
t_{ini1}	temperature of wood-chips to digester (°C)
t_{ini2}	temperature of white liquor before entering digester (°C)
t_{ini3}	temperature of black liquor before entering digester (°C)
TS	dry content, mass fraction
w	the amount of component (ton)
Δt	boiling point rise (°C)

pressure steam and to carry out the reduction reactions to recover cooking chemicals.

The inorganic product of the boiler recovery is used to generate the NaOH and Na_2S needed for pulping by passing through the causticizing plant. The high pressure steam (60 bar) produced in the recovery boiler (RB) and the bark boiler (BB) is expanded in a steam turbine producing 10 and 4 bar steam to be used in the process, and 30 bar steam is extracted from the turbine for soot-blowing in the RB. It should be noted that all pressures in this work are given in absolute pressure bar(a). Biomass fuels, in the form of bark and forest residues, are used in the BB. However, when the supply of biomass is not enough, oil as well as purchased biomass can be used, depending on the current market price. Oil is also used as the ignition fuel when starting up the RB.

A schematic process description of the studied pulp and paper mill is shown in Fig. 1.

2.2. Modelling strategy

The developed model is based on mathematical programming, i.e. mixed integer linear programming (MILP) to find the optimum operating conditions towards a lower energy cost. The integer function has not been used at this stage, and it will be used in the next step's model development. The software reMIND [23], which is a Java based equation editor, in combination with the commercial solver CPLEX [24] has been used. The model structure is represented as a network of nodes and branches, which represent process units and energy/material flows, respectively. The different nodes are connected depending on the input and output to/from each process unit. Each node contains linear equations to express the material and energy balances required in the process unit. Thus an entire energy system is created.

The dynamic of the energy system during the optimization period has been represented with a number of time steps, which should be chosen according to the variations in the energy system. Variations may, for example, occur in energy price or production process for different periods in a year.

The objective function has been set to minimize the energy cost in this work and can be expressed according to

$$\min f(x) = c_{oil} \cdot m_{oil} + c_{bark} \cdot m_{bark} + c_{el,purchased} \cdot q_{el,purchased} - c_{el,produced} \cdot q_{el,produced} \quad (1)$$

where c is the price of an energy carrier (oil, bark or electricity) in units of €/ton or €/MW h, m is flow rate in unit of ton/h for oil or bark, and q is electricity (el) flow rate in unit of MW h. The price used in this work is the same as that in [25] and listed in Table 1.

The first step of the modelling strategy has been to begin with creating rigorous Excel models of each sub-process together with plant engineers. Each of the sub-models has then been validated against operational data and thereafter translated to node equations in the reMIND software. This procedure has been followed

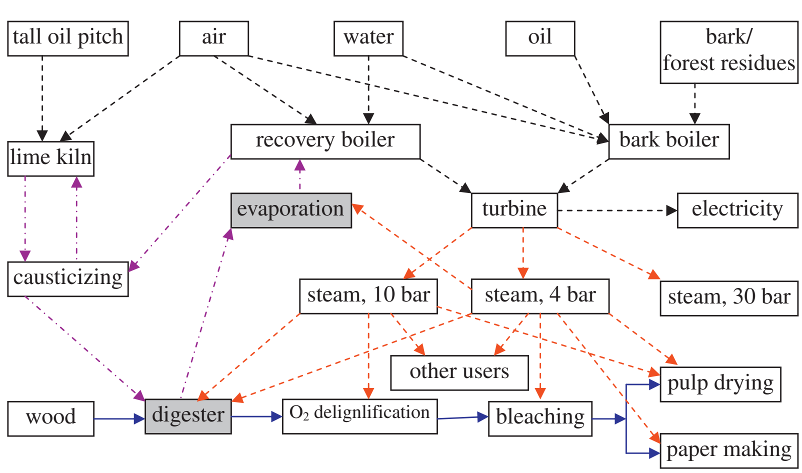

Fig. 1. Schematic representation of the pulp and paper mill. —, fiber-related streams; -·-·-, chemical recovery-related streams; - - - -, process steam streams; – – – –, streams including fuel, air, water, electricity, as well as high pressure steam.

Table 1
Price for fuels/electricity used in the model.

Fuel/electricity	Price*
Oil	350 €/ton
Bark (forest residues), purchase	87.4 €/ton, dry
Electricity, purchase	45.8 €/MW h
Electricity, sale	−31.6 €/MW h

* Recalculated from Swedish kronor (SEK) to Euro using the exchange rate 1€ = 9.50 SEK.

by a second validation step, where the complete reMIND model has been validated against process data.

There are two main options to express the material and energy balances for each unit, i.e. representing them theoretically (option 1), or obtaining linear equations from measurements under a certain set of conditions (option 2). In option 1, the energy balance of a unit is calculated from the properties of input and output flows with assumptions regarding the overall heat loss, the temperature-dependent heat capacities of the flows, etc. The material balance is based on matter conservation. Actual plant measurements are used to validate the assumptions in option 1. While in option 2, an empirical linear equation obtained from the plant measurements is used to represent the material and energy balances, respectively. As the plant measurements are directly linked to the operation conditions, the linear equations correspond to a specific set of operation conditions. For example, different linear equations representing the material and energy balances can be used for different seasons of the year.

One overall aim of this work has been set to develop a model that could be used for introductory studies of the current mill, but also for later investigations of alternative production routes (e.g., conversion to a bio-refinery) by adding sub-models of new processes. To accomplish that, a step-wise mode with a modular structure has been chosen to develop the process integration model, i.e. each unit has been modelled individually. Option 1 has been chosen for the detailed modelling of the most energy intensive process units, while option 2 has been used for less energy intense units. In this work, sub-models of two of the most energy intensive process units, the evaporation plant and digester, have been developed and then implemented into the process integration model of the complete mill.

2.3. Detailed modelling of the evaporation plant

A schematic layout of the evaporation plant is shown in Fig. 2. The liquor flows through effects (E) 4, 5, 6, 7, 3, 2 and 1 in which the temperature of liquor after $E7$ is increased before entering $E3$. The liquor is concentrated from 14.4% to 71.5% (mass). Live steam is only used in $E1$, and the generated liquor steam in an effect is used for the next effect. From $E1$, a certain amount of liquor comes out and is mixed with solid residue and then transferred back into $E1$ at a lower temperature. A part of the generated liquor steam by $E1$ is used for stripping.

Because of the internal cycle of the liquor (part of the liquor out of $E2$ is supplied to $E4$), the boiling point rise, the counter-current/parallel flow of the liquor and the steam and the heat capacity dependence on the dry content of the liquor, two-layer iterations are needed to obtain the steam consumption of the evaporation plant. The iteration procedure is shown in Fig. 3.

There are mainly six parameters affecting the live steam consumption, i.e. (1) the dry content of the liquor to the evaporation plant ($TS\%_{\mathrm{in}}$), (2) the temperature of the liquor to the evaporation plant (t_{in}), (3) the flow rate of the liquor to the evaporation plant (f_{in}), (4) the dry content of the liquor out of the evaporation plant ($TS\%_{\mathrm{out}}$), (5) the liquor steam leaving from $E1$ for stripping (f_2), and (6) the temperature of liquor to $E3$ after heat exchanger (t_3). These six key parameters have been used as variables in the expression representing the steam consumption of the evaporation plant.

The Excel model has been established on the basis of material and energy balances for each effect. To illustrate the material and energy balances in detail, $E1$ illustrated in Fig. 4 is used as an example.

The material balances for each effect i can be expressed according to

$$f_{liquorin,i} = f_{liquorout,i} + f_{liquorsteam,i} \quad (i = 1, 7) \tag{2}$$

$$(f_{liquorin} \cdot TS_{in})_i = (f_{liquorout} \cdot TS_{out})_i \quad (i = 1, 7) \tag{3}$$

$$f_{water,i} = f_{steam,i} \quad (i = 1, 7) \tag{4}$$

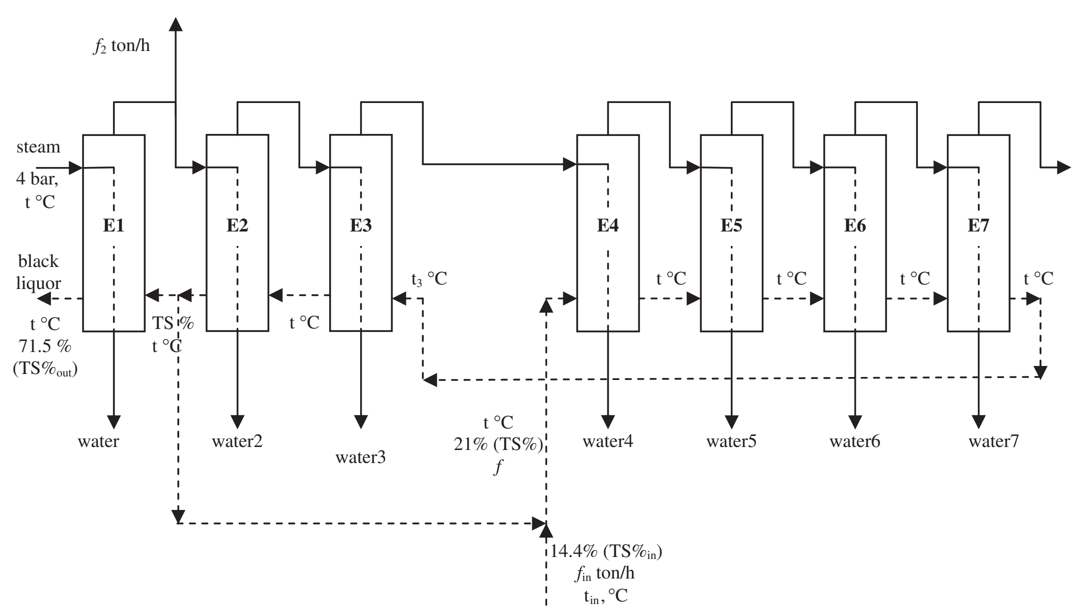

Fig. 2. Schematic description of the evaporation plant.

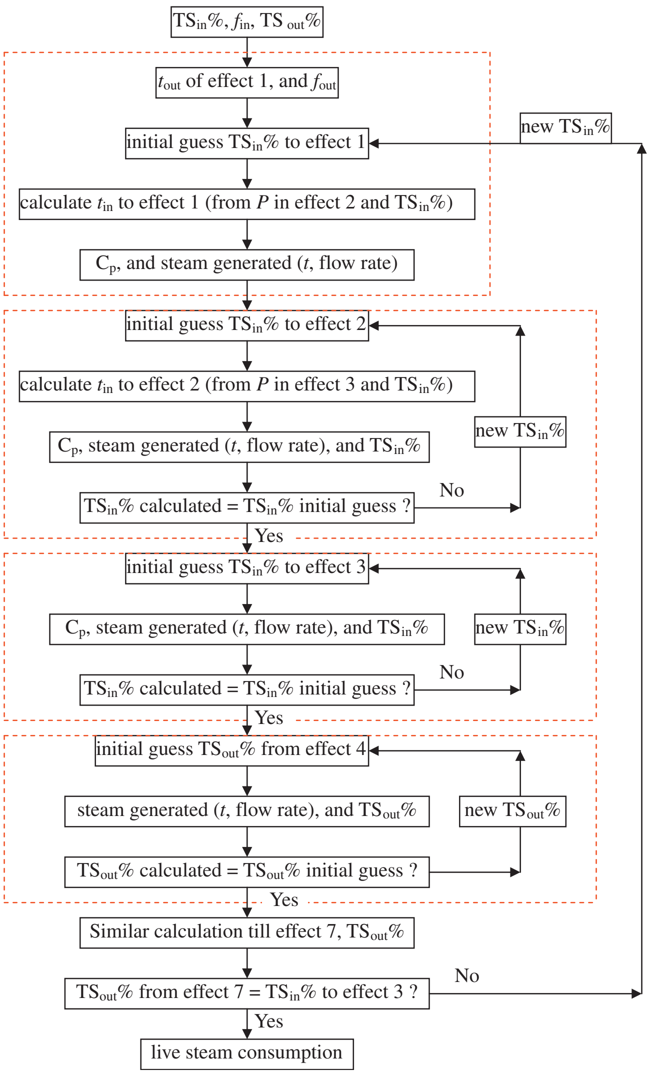

Fig. 3. Schematic illustration of the Excel model for the evaporation plant.

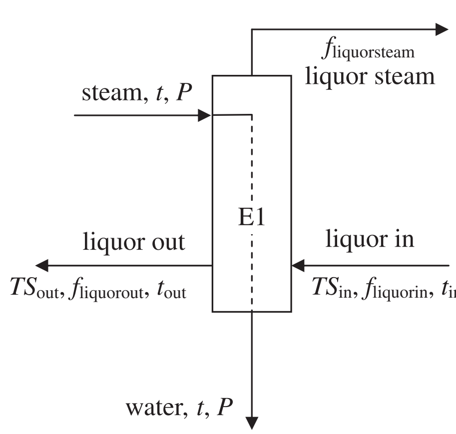

Fig. 4. Schematic representation of the $E1$ in the evaporation plant.

The energy balance for $E1$–$E7$ can be expressed according to

$$f_{water} \cdot (h_{steam} - h_{water}) = f_{liquorin} \cdot C_p \cdot (t_{out} - t_{in}) + f_{liquorsteam}(h_{steam} - h_{water})|_{t_{out},p_E} \quad (5)$$

where h has been calculated with steam-table [26], p_E is the pressure of the effect, and C_p depends on temperature as well as the dry content of the liquor. C_p has been calculated according to Eq. (6) [27]

$$C_p = 4.216 \cdot (1 - TS) + (1.675 + 3.31 \cdot t/1000) \cdot TS + (4.87 - 20t/1000) \cdot (1 - TS) \cdot TS^3 \quad (6)$$

The boiling temperature of the liquor increases because of the existence of solid substances. In this work, the boiling point rise (Δt) has been calculated according to Eq. (7) for $E1$ [28] and Eq. (8) for other effects [27], respectively

$$\Delta t = 1.3 \times 10^{-6} TS^4 - 0.00013 \cdot TS^3 + 0.0046 \cdot TS^2 - 0.011 \cdot TS + 5.2 \quad (7)$$

$$\Delta t = (6.173 \cdot TS - 7.48 \cdot TS^{1.5} + 32.747 \cdot TS^2) \cdot [1 + 0.006(T_{sat,p} - 373.16)] \quad (8)$$

The temperature t_{out} has been calculated according to

$$t_{out} = t_{sat}(P) + \Delta t \quad (9)$$

where t_{sat} is the saturation temperature of pure water at pressure P and it has been calculated with steam-table [26].

In $E1$, the temperature and pressure of the condensate were taken from mills, and live steam was used. While in $E2$–$E7$, the condensate has been assumed to be saturated water at pressure p_E, and the liquor steam was used as shown in Fig. 2.

The Excel model has been developed based on the principle shown in Fig. 3. By executing the Excel model, the linear expression representing the steam consumption has been obtained as shown in Eq. (10) in which the range of each parameter is based on the suggestions from the mill

$$f_{steam} = -0.6897TS\%_{in} - 0.552t_{in} + 0.8655f_2 - 0.4288t_3 + 0.2445TS\%_{out} + 0.1182f_{in} \quad (10)$$

It should be mentioned that the Excel model has been validated very carefully together with plant engineers by comparing model calculations both to plant measured values and plant experience.

2.4. Detailed modelling of the digester

The digester is operated batch-wise in several steps, i.e. wood-chips filling, steaming, liquor filling, heating, cooking, displacement, and emptying (blowing). 4 bar steam is used for wood-chips filling, 4 bar as well as 10 bar steam is used for the wood-chips steaming up to a temperature of 108 °C (t_{end1}), the white liquor is preheated with 10 bar steam from 90 to 122 °C and added into digester together with the black liquor (16%), and 10 bar steam is used for the heating step up to 168 °C (t_{end2}). The displacement liquor (black liquor, 11%) is preheated from 125 to 168 °C using 10 bar steam.

Material and energy balances for the steps of filling and steaming as illustrated in Fig. 5a have been calculated according to

$$w_{chips,dry} = w_{chips,wet} \cdot TS \quad (11)$$

$$w_{water} = w_{chips,wet} \cdot (1 - TS) + w_{4bar,steam} + w_{10bar,steam}, \quad (12)$$

$$w_{chips,dry}C_{p,chips,dry}(t_{end1} - t_{ini1}) + w_{chips,wet}(1 - TS)C_{p,water}(t_{end1} - t_{ini1}) + h_{loss} = w_{4bar,steam} \cdot (h_{4bar,steam} - h_{water}|_{t_{end1}}) + w_{10bar,steam} \cdot (h_{10bar,steam} - h_{water}|_{t_{end1}}), \quad (13)$$

where w is the amount (mass) of wood-chips, water or steam supplied into the digester. It should be mentioned that the input temperature of the wood-chips is generally higher than 0 °C. In the modelling, only the sensible heat has been considered for the

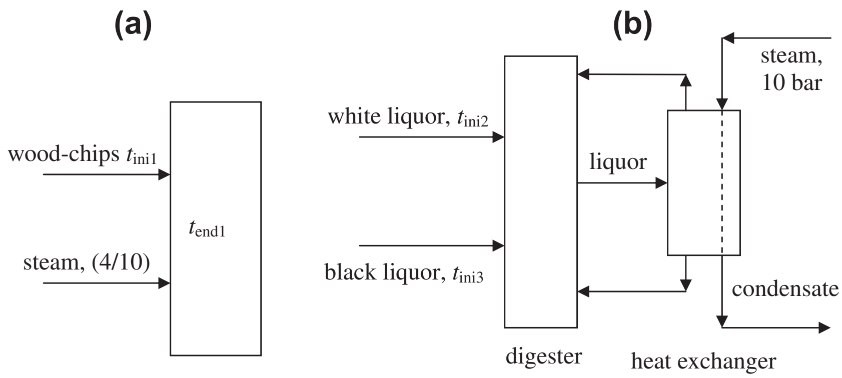

Fig. 5. Schematic representation of the steps of filling and steaming (a) and steps of liquor filling and heating (b) for the digester.

wood-chips warming up. h_{loss} is the estimated heat loss, and h is the enthalpy of steam or water.

The energy balance for the steps of liquor filling and heating as shown in Fig. 5b is established according to

$$\begin{aligned} & w_{chips,dry}C_{p,chips,dry}(t_{end2} - t_{end1}) + w_{water}C_{p,water}(t_{end2} - t_{end1}) \\ & + w_{whiteliquor}C_{p,whiteliquor}(t_{end2} - t_{ini2,whiteliquor}) \\ & + w_{blackliquor}C_{blackliquor}(t_{end2} - t_{ini3,balckliquor}) + h_{loss} \\ & = w_{10bar,steam}(h_{10bar,steam} - h_{condensate}) + h_{heatrelease,reaction} \end{aligned} \quad (14)$$

where $h_{heatrelease,reaction}$ is the heat release due to the chemical reaction before cooking. The chemical reaction heat has been measured experimentally. The heat release will however be started after the liquor filling, and only a part of the heat is released till the end of the heating (before cooking). It is estimated that the temperature of the digester may increase by 2–3 °C before cooking, corresponding to 1/3 of the total heat release. This estimation has been used in this work to account for the heat release due to the chemical reactions before cooking.

The energy balance for the preheating of the white liquor or the displacement liquor can be expressed according to

$$wC_p(t_{out} - t_{in}) = w_{10bar,steam}(h_{10bar,steam} - h_{condensate}) \quad (15)$$

where $h_{condensate}$ is the enthalpy of the condensate that has been assumed to be the saturated water.

In Excel model, the enthalpy of steam or water has been calculated with steam-table [26] directly. While in the reMIND model, it has been calculated in another way. The enthalpy of steam or water has been primarily calculated from the NIST online database [29] and then fitted to an expression as shown in

$$h = (2400.1 + 2.1856t) + (2.910 \times 10^{-2}t - 11.00)(p - 9) \quad \text{(steam around 10 bar)} \quad (16)$$

$$h = (2443.9 + 2.1157t) + (8.06 \times 10^{-2}t - 20.5)(p - 3) \quad \text{(steam around 4 bar)} \quad (17)$$

$$h = 4.2570t - 4.4463 \quad \text{(water)} \quad (18)$$

In addition, because of the batch operation, the steam consumption is time-dependent. However, in the process modelling, the average steam consumption of a certain period for a fixed consumption of dry wood-chips has been used. The heat capacity used in this work is listed in Table 2.

Table 2
Heat capacities (kJ/kg K).

Material	Heat capacity	Material	Heat capacity
Wood-chips, dry	1.47	Black liquor (16%)	3.74
Water	4.20	Displacement liquor (11%)	3.74
White liquor	3.81		

Again, the digester model has been validated carefully by comparing to the plant measured values and plant experience.

3. Model validation

A model of a sub-process can be validated individually and/or when implemented into a process integration model of the complete mill. In this case an individual valuation of the sub models had been carried out when they were developed as Excel models. Since the sub-processes will affect each other, an additional validation has been carried out for the process integration model. Actual plant measurements have been compared to values calculated for evaporation plant and digester for the same case.

The modelling results of the evaporation plant and digester have been validated against actual plant measurements during full production (900 ton pulp per day during spring/fall season). The steam consumption of the evaporation plant has been calculated to 72.9 ton/h assuming a heat loss of 2%. According to plant measurements the steam consumption amounted to 72.5 ton/h under prevailing operational conditions. For the digester, the 10 bar steam consumption has been calculated to 41.9 ton/h, to be compared with the operational data of 42.0 ton/h. It may be concluded that the calculated and measured data show good agreement for the reference case.

4. Results and discussion

Introductory studies with the main objective to minimize the plant energy cost have been carried out in the process integration model and the influence of different operation parameters of the evaporation plant and digester on the overall plant performance has been investigated. For the evaporation plant, the temperatures of light liquor and liquor to $E3$, the dry content of light liquor, and the utilization of liquor steam from $E1$ were investigated; while for the digester, the case of preheating wood-chips as well as four cases on preheating of the white liquor were addressed.

It should be pointed out that all the results (all points in Figs. 6–8 and cases in Table 3) are results of a cost optimization. For example, in Fig. 6a the point corresponding to a value of 95 on the horizontal axis shows the consumptions of biomass fuel (bark) and steam and the electricity generation if the light liquor to $E4$ is fixed at 95 °C and the other plant parameters are optimized to give the lowest energy cost.

4.1. Parameter study of the evaporation plant

The effect of temperature of the light liquor supplied into $E4$ is illustrated in Fig. 6a. If the temperature increases from 85 to

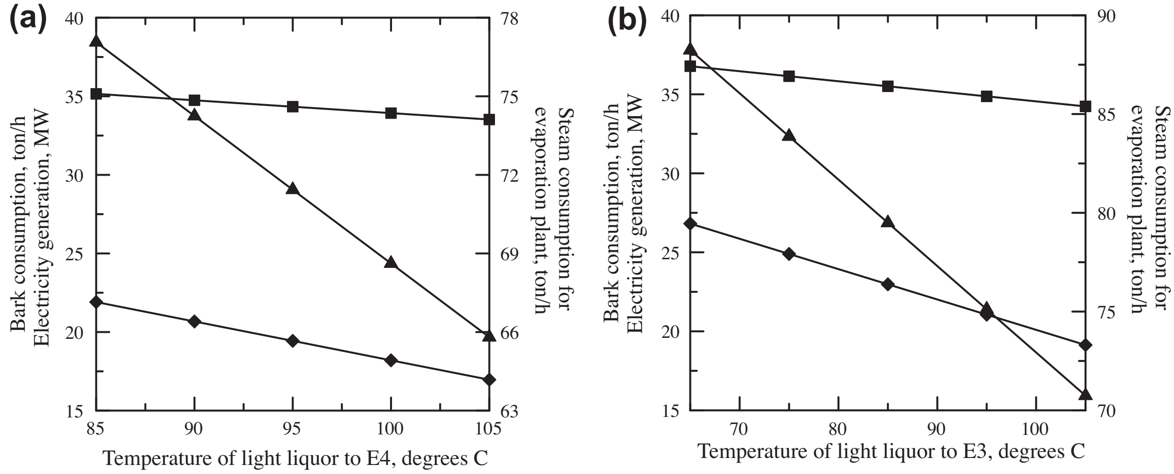

Fig. 6. The effect of the liquor temperature to *E*4 (a) /the liquor temperature to *E*3 (b) on the steam consumption of the evaporation plant and the performance of BB. —◆—, bark consumption; —■—, electricity generation; —▲—, steam consumption of evaporation plant.

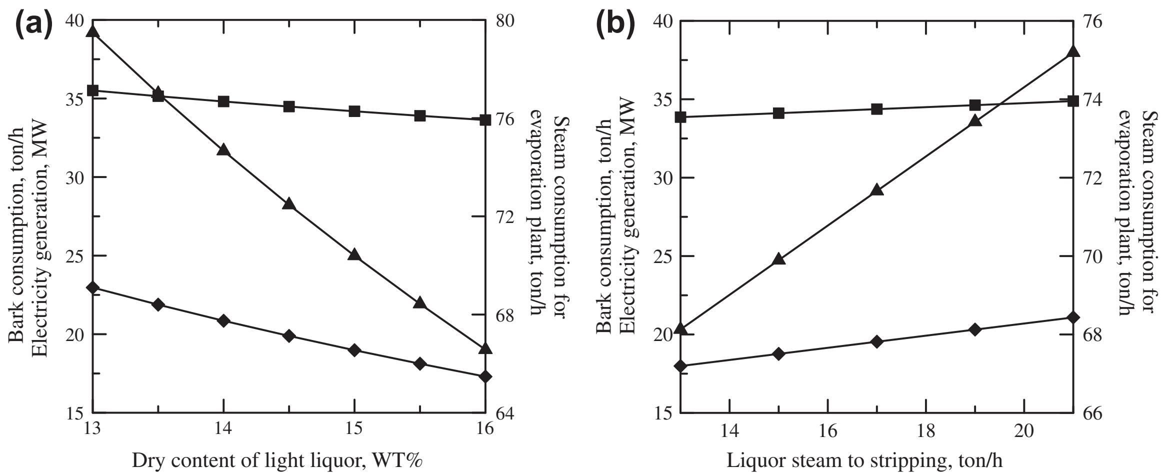

Fig. 7. The effect of the dry content of the light liquor (a) /the steam liquor to stripping (b) on the steam consumption of the evaporation plant and the performance of BB. —◆—, bark consumption; —■—, electricity generation; —▲—, steam consumption of evaporation plant.

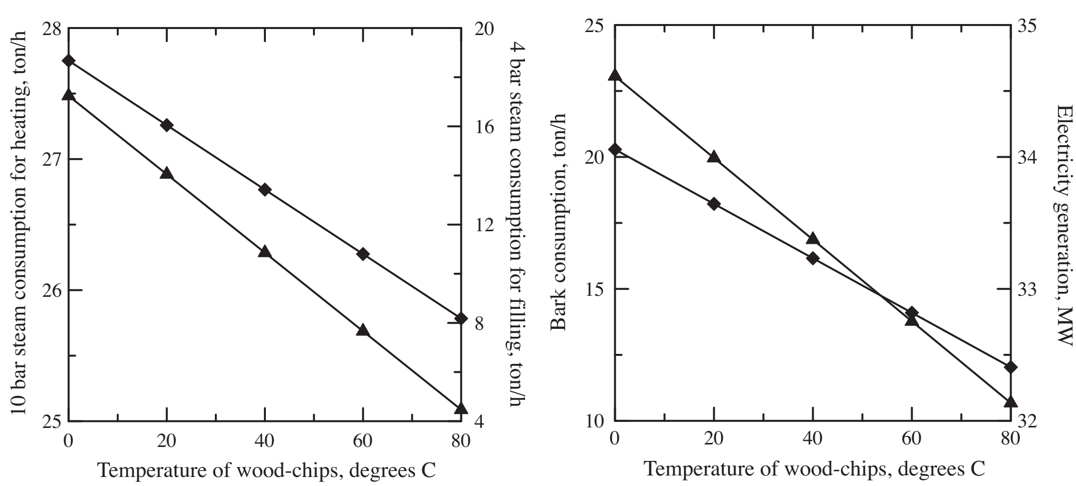

Fig. 8. The effect of the temperature of wood-chips on the steam consumption of the digester and the performance of BB. (a): —◆—, 10 bar steam consumption for heating; —▲—, 4 bar steam consumption for filling; (b), —◆—,bark consumption; —▲—, electricity generation.

105 °C, the steam consumption of the evaporation plant decreases from 77 to 66 ton/h, the biomass fuel (bark) consumption decreases from 22 to 17 ton/h at the same time as the electricity generation decreases from 35 to 34 MW h/h (=MW, the same in the

Table 3
Energy utilization in the digester for four cases regarding the preheating of the white liquor.

Substance	Case 1	Case 2	Case 3	Case 4
Biomass fuel (bark) consumption, ton/h	20.08	20.23	20.36	19.68
Steam generation from BB, ton/h	46.43	46.78	47.08	45.50
Electricity from turbine, MW h/h (MW)	34.55	34.95	34.62	34.46
10 bar Steam consumption for preheating white liquor, ton/h	8.25	–	–	20.11
4 bar Steam consumption for preheating white liquor, ton/h	0	8.59	–	–
10 bar Steam consumption for heating, ton/h	27.70	27.70	36.59	14.93

flowing text and in all figures). The temperature of the light liquor depends on the operating conditions at the washing plant, and an improved temperature control at the washing plant can reduce the steam and fuel consumption.

The temperature of the liquor from *E*7 is normally in the range of 60–70 °C depending on the dry content of liquor. To use less live steam for the evaporation plant it would be desirable if the liquor to *E*3 would be around 100 °C (close to the boiling point). One way to accomplish that could be to use a heat exchanger to pre-heat the liquor before it enters to *E*3, for example, using the condensate from effects to preheat liquor. The influence of inlet liquor temperature to *E*3 on the steam consumption has been studied to find out how much live steam that can be saved by installing a heat exchanger. The results are shown in Fig. 6b. If the temperature of liquor to *E*3 increases from 65 to 105 °C, the steam consumption of the evaporation plant decreases from 88 to 70 ton/h leading to a reduced biomass fuel consumption from 27 to 19 ton/h. At the same time the electricity generation decreases from 37 to 34 MW.

The dry content of the light liquor depends on the degree of washing. Meanwhile, it also affects the steam consumption of the evaporation plant considerably as shown in Fig. 7a. If the dry content of the light liquor increases from 13% to 16%, the steam consumption decreases from 79.5 to 66.6 ton/h, the biomass fuel (bark) consumption decreases from 23 to 17 ton/h, while the electricity generation decreases slightly from 35.6 to 33.6 MW. However, exactly how the overall operation of the washing changes with a modified dry content has not yet been investigated, but the optimization potential has been illustrated.

One part of steam generated by *E*1 is used for stripping and the residual part is used in *E*2. The steam distribution between the stripping and *E*2 affects the performance of the mill as shown in Fig. 7b. If the utilization of the liquor steam to the stripping increases from 13 to 21 ton/h, the steam consumption of the evaporation plant increases from 68 to 75 ton/h, the biomass fuel consumption increases from 18 to 21 ton/h, and the electricity generation increases from 33.9 to 34.9 MW.

4.2. Parameter study of the digester

The temperature of the wood-chips supplied to the digester depends on the outdoor temperature, and in winter, the wood-chips can be frozen. To investigate the effect of the initial temperature of the wood-chips on the steam consumption and the performance of BB, the process integration model has been run at different initial temperatures of the wood-chips but with the same ratio of 4 and 10 bar steam for steaming. The results are shown in Fig. 8. With increasing initial temperature of the wood-chips, the steam consumption decreases considerably. Because the amount of steam for filling will affect the water content in the digester, and the steam consumption for the later step (steaming and heating) decreases with increasing initial temperature of the wood-chips. Subsequently, the biomass fuel consumption and the electricity generation are reduced as shown in Fig. 8. If low quality residual heat (such as condensates as well as flue gas from boilers) can be used for preheating of the wood-chips, it may be a fairly simple way to save fuel.

To investigate the influence of the temperature of the white liquor supplied to the digester on the steam consumption, four case studies have been carried out. As a reference case (Case 1), the white liquor is, as presently, preheated using 10 bar steam before supplied to the digester. From the energy point of view, it would be better to use 4 bar steam to preheat the white liquor to the same temperature as in case 1, which represents Case 2. In Case 3, the white liquor is not preheated at all. In Case 4, the white liquor is preheated using 10 bar steam to the final temperature of digester, 168 °C. The results are shown in Table 3.

As the case study shows, using 4 bar steam for preheating (Case 2) increases the biomass fuel supply slightly compared to the reference case (Case 1) due to the higher mass-flow of steam supplied into the digester. This also leads to an increased electricity generation. From a practical point of view, it is more convenient to use 10 bar steam for the preheating. As the temperature of the white liquor to the digester will affect the steam consumption for the subsequent steps, it is better to preheat the white liquor to a higher temperature, for example, to the final temperature of the digester (Case 4). As shown in Table 3, the biomass fuel consumption in case 4 (19.68 ton/h) is lower than that in Case 1 (20.08 ton/h) and consequently the electricity generation becomes slightly reduced. In Case 3, the biomass fuel consumption becomes higher, leading to an increased electricity generation.

The calculations show that the energy utilization is lowest in Case 4 and highest in Case 3, and the energy utilization in Case 2 is lower than that in Case 1. This implies that it is better to preheat the white liquor to a higher temperature if 10 bar steam is used for preheating, while if the white liquor is preheated to the temperature at current status, it is better to use 4 bar steam instead of 10 bar steam. However, investments in new or modified existing heat exchangers have not been taken into consideration. The best option depends on the price of the fuel, the price of the electricity, the investment and the operational cost of the heat exchangers.

5. Conclusions

A mathematical process integration model of a pulp and paper mill in the Northern Sweden has been developed. This paper focused on describing the modelling approach regarding material and energy balances of the evaporation plant and digester of a mill. The models of these two sub-processes have been implemented into a complete plant model and validated with the actual plant measurements. The approach may be considered general and may be applied to more or less any industrial process. It provides the possibility to carry out parameter studies of a single subprocess in which the total system optimum is found at prevailing parameter value.

Case studies regarding how various operational parameters influence the steam consumption, the electricity production and the biomass fuel consumption have been carried out. As a result, several options to save steam and fuel have been identified. For example, if the wood-chips supplied to the digester is preheated from a temperature of 0 °C to say 60 °C by the use of low grade residual heat, approximately 5 ton/h of biomass can theoretically be saved. Another case study shows that if the inlet liquor temperature to *E*4 of the evaporation plant increases from 85 to 105 °C, the steam consumption of the evaporation plant decreases from 77 to 66 ton/h and then the biomass fuel consumption decreases from 22 to 17 ton/h.

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