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ABSTRACT

The Stora Enso Skutskär mill has increased production and reduced effluent load by making some major changes in the recovery system. Some new, cost-effective technology has been implemented. Equipment installed includes a vertical air system in one recovery boiler, the burning of noncondensable gases in another recovery boiler, and a new evaporator line with a stripper column to remove COD. Along with other modifications, these changes have increased the mill's capacity by 25% and have reduced the effluent COD by 20%.

Application:

Practical experience in improving the performance of chemical recovery operations at Shutskär may benefit those in other mills in replacing similar kinds of equipment and upgrading the same production processes.

Recovery modernization at Stora Enso's Skutskär kraft mill

KAJ BÄCKMAN, HANS LINDBERG, AND HANS SJÖBERG

The Skutskär kraft mill is located on the east coast of Sweden, 180 km north of Stockholm. Since the merger of Stora and Enso in 1998, the former Stora Cell's Skutskär mill has been part of the Stora Enso Pulp Division.

Since 1995, the Skutskär mill has been undergoing an extensive modernization project called "S2000". The goal of S2000 is for the mill to be operating at higher production rates with better cost efficiency and minimized environmental impact by the year 2000. The projected production increase will take the mill from 430,000 to 550,000 a.d. metric tons per year.

To help meet the production and environmental goals for the project, we have upgraded or replaced several units in the mill with new equipment. Some new strategies and technologies have been implemented that are not yet fully proven in industry. The major projects so far are the replacement of the old fluff bleach plant with a new one, modernization of the woodyard, and a major rebuild of the recovery area. Our focus here is on the modernization of the recovery area.

THE SKUTSKÄR MILL AND S2000

The Stora Enso Skutskär mill produces bleached pulp on three fiber lines. Line 1 produces 450 tons/day of bleached softwood pulp for use in papermaking, Line 2 produces 650 tons/day of bleached fluff pulp from softwood, and Line 3 produces 440 tons/day of bleached hardwood pulp for use in papermaking. Each fiber line utilizes continuous cooking followed by oxygen delignification. The pulps from Line 1 and Line 3 are bleached with elemental chlorine-free technology, and the fluff pulp in Line 2 is currently bleached according to a total chlorine-free concept. The pulp is dried on four drying machines.

Production and COD targets for the project are outlined in **Table I**. The COD targets for 1997 and 1998 have been met, but the production target for 1998 was not reached, as a result of digester problems. For 1998, the total production was 445,000 a.d. metric tons, and the COD was 46 tons/day, as unfiltrated (about 41 tons/day as filtrated). The COD targets for 2001 and beyond are pending a new environmental permit and the installation of secondary treatment.

The design targets for the modernization of the recovery area are listed in **Table II**. **Figure 1** is a diagram of the recovery area in 1997.

BLACK LIQUOR EVAPORATION

For the production increase planned in S2000, the evaporation capacity of the mill has been increased from 500 tons/h to 670 tons/h. An old, five-effect, rising-film evaporator from 1966 was replaced with a new falling-film evaporator line from Ahlstrom. The new line, Line 6C, has a capacity of 350 tons/h, and it operates in parallel with an evaporation line built in 1977, Line 6AB.

Table III lists technical data for the two evaporation lines. The concentrators in Line 6AB have been disconnected, and all the liquor is concentrated in a new concentrator placed as the first unit in the new evaporator line. The evaporation plant operates today with six-effect economy and with a dry solids target of 72%, as diagrammed in **Fig. 2**.

The new evaporator line is designed with extended condensate segregation to improve the efficiency with which condensate is cleaned. The purpose was to reduce the COD emissions to meet a new, tighter effluent permit. The goal was set to 4 kg COD/a.d. metric ton in the combined condensates before reuse. An in-line condensate stripper with a capacity of 320 m^3/h and a new external stripper for foul condensates were therefore included in the plans.

The in-line stripper column is placed between Evaporation Units 3C and 4C, and it takes all the steam from Unit 3C. Units 1C, 2C, 3C, and

4C are designed with internal condensate segregation, as **Fig. 3** shows. (The numbering of the units is not fully conventional.) From Line 6AB, the condensate to the in-line stripper comes from Units 5 and 6 and from the precondensers. The condensate from Line 6C to the in-line stripper includes foul condensate from Units 1C, 2C, and 3C, the reflux from Unit 4C, and all condensate from Unit 5C. The clean condensate from an external stripper also goes to the in-line stripper. The accept stream from the in-line stripper is maintained at a level of 400–700 mg/L of COD.

The vapor leaving the in-line stripper is condensed on a pre-condensing surface and a secondary condensing surface in Unit 4C. The condensate from the pre-evaporation unit is returned to the in-line stripper. Noncondensable gases and the noncondensed vapor from Unit 4C are sent to a liquor heater for intermediate liquor heating and to a trim condenser for total condensation. The foul condensate is returned to the external stripper. Noncondensable gases are vented to the noncondensable gas collecting system.

To maintain a high availability in the condensate cleaning, we included the external stripper to take care of foul condensates from Line 6AB and from the digesters when Line 6C is not in use. Integrated with the external stripper is a methanol column with turpentine separation and a decanter. Turpentine is mixed with methanol and is burned in the RB6 recovery boiler. The main portion of turpentine is decanted from the digester flash condensates, but the operation is not always under full control, which results in disturbances in the methanol column.

The new evaporator plant was started up in April 1997. During the first year of operation, the liquor concentration varied between 65% and 72%, with an average of 68–69%. Problems with old black liquor pumps at the recovery boilers and with a low temperature for the weak black liquor had to be solved to maintain the dry solids at 72%. COD to the effluent from the brownstock area has decreased from about 20 tons/day to below 10 tons/day, as shown in **Fig. 4**. The reduction in COD is mainly the result of the improvement in condensate quality.

VERTICAL AIR

Recovery Boiler 7 was originally designed in 1977 for 1900 tons of dry solids per day. In practice, this load was also the upper limit for the maximum continuous rate (MCR). Above this level, the boiler started to plug, and the dust content after the electrostatic precipitator exceeded the permitted level.

	Pulp, tons/year	COD, tons/day
1997	406,000	65[a]
1998	500,000	50[a]
1999	525,000	55[b]
2000	550,000	55[b]

[a]Filtrated. [b]Unfiltrated.
COD targets for 1999 and beyond are pending a new environmental permit and installation of secondary treatment.

I. Pulp production and COD targets

The planned mill production of 550,000 a.d. metric tons/year meant that the capacity of RB7 had to be increased by 25% to 2400 tons of dry solids/day. A conventional expansion of the furnace was considered but was regarded as too expensive. The expansion of the furnace would have required at least three weeks of work, which corresponds to an extra

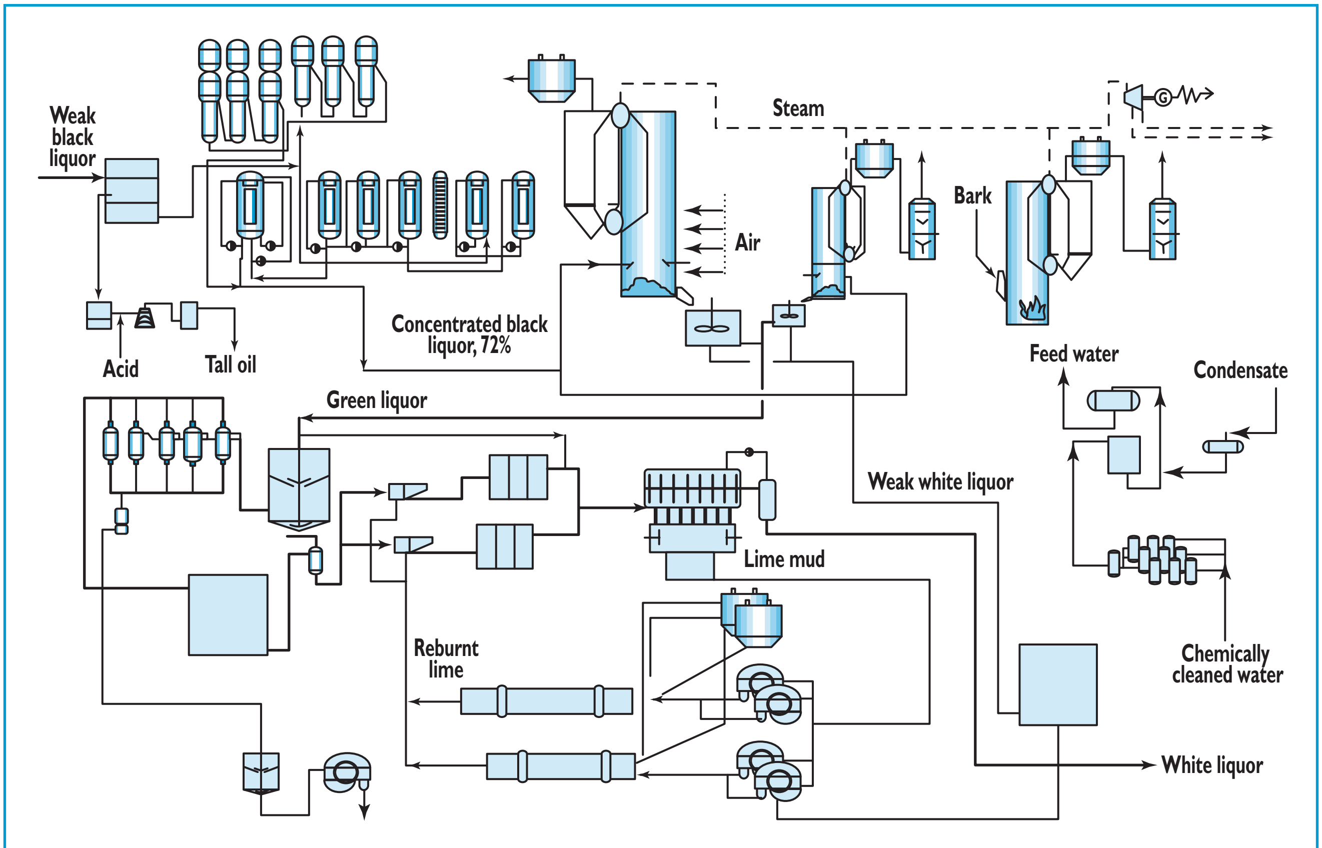
	Before	After
Annual pulp production, a.d. metric tons	430,000	550,000
Available days per year	355	355
Dry solids to the boilers, tons/day	2400	3100
White liquor production, m³/day	5500	7000
Evaporation, tons/hour	500	670
Lime reburning, tons/day	400	450

II. Design targets for the modernization of the recovery area

	Line 6AB	Line 6C
Manufacturer	Rosenblad	Ahlstrom
Year	1977	1997
Capacity, tons evaporated/h	320	350
Economy, stages	6	6
Dry solids out, %	56	72
Steam, bar at 144°C	3	3
Strippers		
Internal, m³ of condensate/h	n.a.	320
External, m³ of condensate/h	n.a.	85
Methanol column, kg of MeOH/h	n.a.	800

Atm. = bar × 0.9869.

III. Technical data for the evaporation lines



1. The recovery area at the Skutskär Mill in 1997.

production loss of more than two weeks, compared to a normal maintenance stop. A solution based on increasing the black liquor dry solids content, combined with a progressive air distribution system, was found to be the most attractive solution.

The theory and the computer simulation of "Vertical Air," a patented process (*1*), were convincing, even though no full-scale references were available. The vertical air approach offered improved horizontal mixing while providing more uniform vertical gas velocities in the furnace. Increased capacity, reduced emissions, and lower costs were the main reasons for considering the vertical air concept.

Air ports were installed according to the vertical air concept in the front and rear walls of the furnace. The arrangement is made up of three columns in the front wall and two in the rear wall in a so-called "five finger" arrangement. There were eight air port levels in the front wall and nine in the rear wall. The vertical arrangement is shown in **Fig. 5**.

To reduce installation costs, the new ports were located in areas with easy access. Secondary air ducts and the original liquor gun arrangement precluded the possibility of a full five-finger arrangement on the new upper secondary air level. Also, only one level of air ports was placed between buck stays.

The boiler was equipped with four new liquor guns in 1995, two on each side wall about 1.5 m higher than the original two liquor guns. One of the original two is in the front wall, and one is in the rear wall. Firing with four or more liquor guns became successful when we installed a liquor heater in 1996 that enabled us to control the liquor temperature.

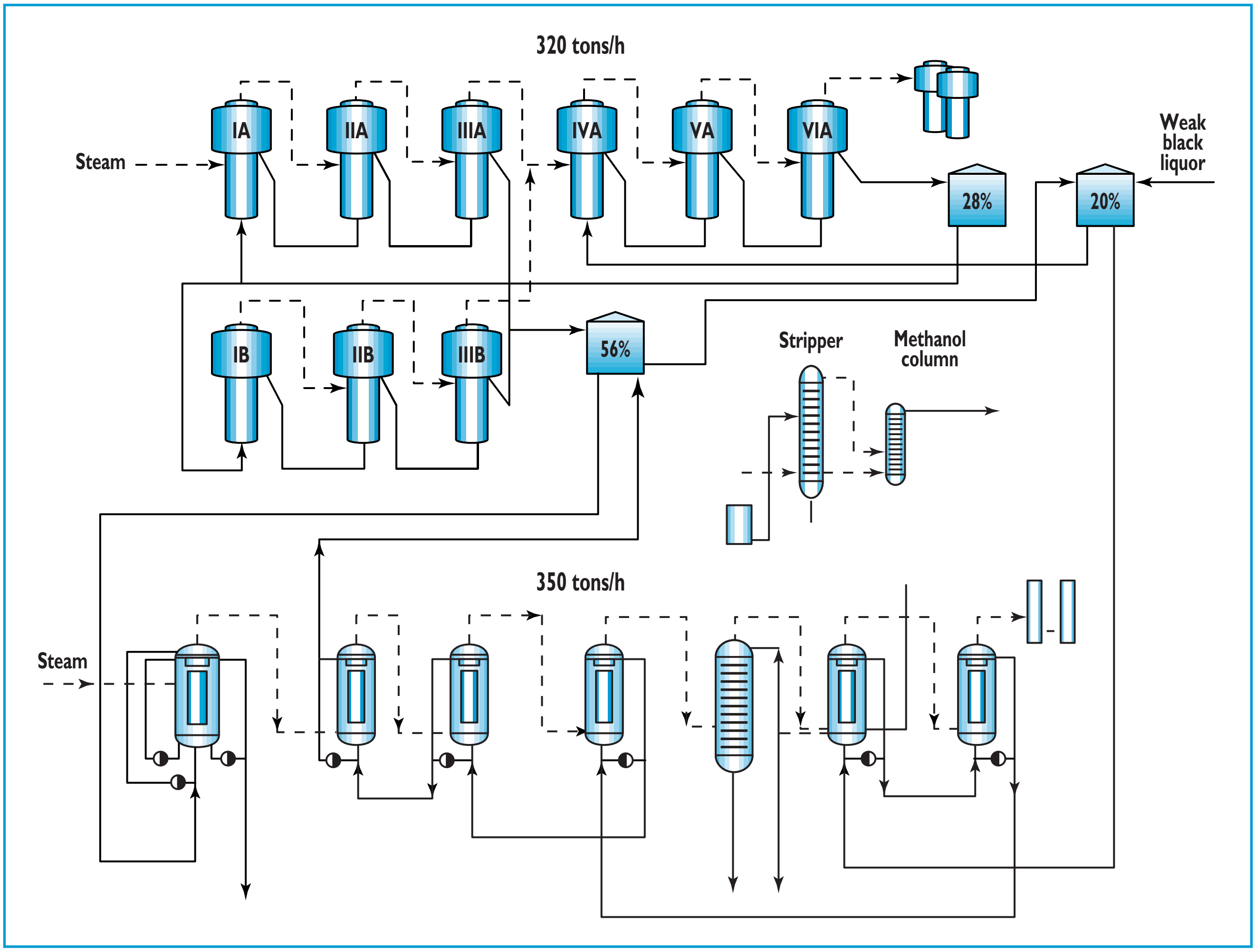
Since June 1996, the new air system has been in continuous operation. The black liquor concentration was increased in April 1997, when the new evaporation line was put into operation. During extended periods, the load has been at approximately the new maximum continuous rate of 2400 tons of dry solids per day, which corresponds to a heat load of 2450 kW/m² in cross section.

For this load, the typical air distribution is:

- 22% above the liquor gun level
- 42% between the liquor gun level and the primary air level
- 36% primary air.

The 22% above the liquor gun level is the maximum load for the tertiary air fan. The 42% between the liquor gun level and the primary air level is the maximum load for the secondary air fan. The direct obstacle for further load increases is the tertiary air fan.

The vertical air system involves a redistribution of the air supply to the furnace. The vertical air distribution has preconditions for reducing NO_x,



2. Steam and liquor flows for the rebuilt evaporation plant.

and the system does decrease NO_x. The NO_x reduction depends also on the boiler load and on the dry solids content of the black liquor. Results from test runs are shown in **Fig. 6**, where the reduction is given as a percentage from a base level of 110 ppm as NO.

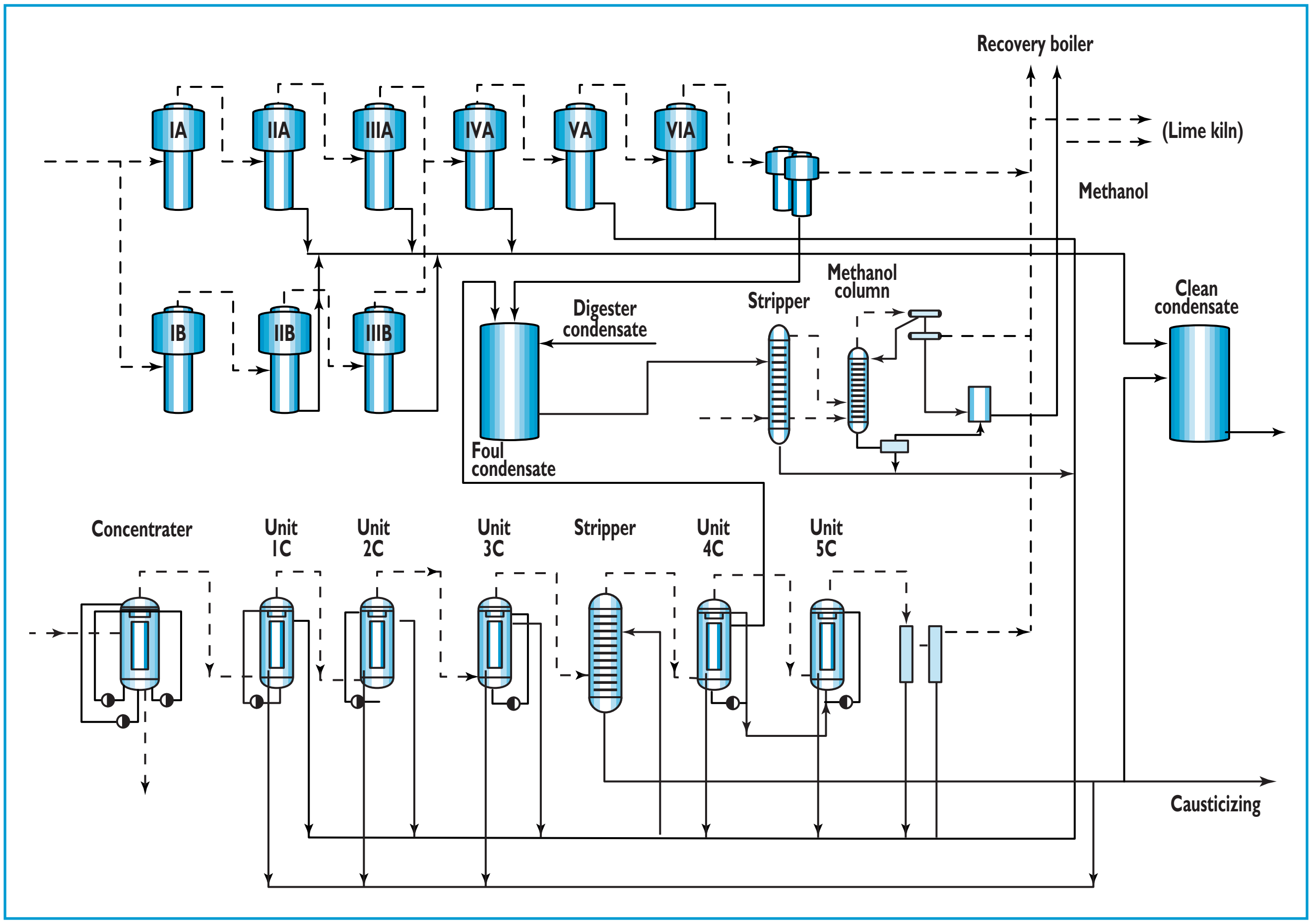
At a partial load of 60% MCR and a low dry solids content of 64%, the NO_x level has been reduced by more than 60%. At a load close to MCR with a dry solids content of 70% or more, the corresponding decrease in NO_x has been only 10–25%. In this case of low removal, a tertiary fan shortage decreases the percentage of air distributed to the upper furnace. Less than stoichiometric air supply to the area below the liquor guns is one important factor in reducing NO_x emissions.

Experience has shown that the air system can be further improved through an increase in the number of horizontal air levels. This option will be implemented, along with a new tertiary air fan, during the maintenance stop in the fall of 1999.

The dust content of flue gases downstream of the electrostatic precipitator seems to depend strongly on the stability of the burning process. Before the modernization of the air distribution system, periods with high dust concentration were common. Now only extreme disturbances cause increases in dust content.

Besides meeting the load increase and the reduction in emissions, the vertical air distribution has also provided a stabilized burning process. Beginning from very small partial loads, the surface temperature of the char bed has been uniform, without any dark areas. The height of the char bed has also been uniform, steady but low at all loads. Stabilized burning has been obtained easily without auxiliary fuel at loads that are 25–30% lower than load levels previously observed.

Low O_2 content in the flue gas at the economizer outlet indicates good mixing of the gases in the furnace. Now the furnace can operate with only 1.3–2.5% O_2 at a CO level below 300 ppm at the economizer outlet, compared to 3.5–4.5% earlier. The degree of reduction of the green liquor is normally exceeding 92% as calculated on total sulfur. Because of less carryover, we have not had deposit problems in the superheater or in the boiler bank.



3. Steam and condensate flows for the rebuilt evaporation plant.

The capacity of the old secondary and tertiary air fans still limits the load to about 2450 tons of dry solids/day. Also, the drum pressure is close to its limit at the steaming rate at this load, especially at a high liquor dry solids content of 72%.

MODERNIZATION OF THE CAUSTICIZING PLANT

To meet future production levels, we increased the capacity of the causticizing plant from 5500 m³/day of white liquor to 7000 m³/day. The main changes completed include green liquor filtration, a rebuilt causticizing line, a new disc filter for white liquor filtration, and a new dregs filter. The modifications were based on proven technology except for the capacity of the white liquor filter. The capacity is the largest capacity in one filter in the world. (Aracruz started up a filter of the same capacity in late 1997.)

Existing tubular pressurized filters used for white liquor filtration have been modified for green liquor filtration. To improve the filtration properties, we mix lime mud, corresponding to 1–2% of the lime charge, with the green liquor prior to filtration. Full production is reached with four of five existing filters in use.

Green liquor dregs, including lime mud, is discharged on a new dregs precoat filter with a surface area of 42 m².

The density and temperature of the green liquor are adjusted before the liquor is charged to the slaker. Temperature control is considered important for proper slaker control. To control the temperature, we use pressure-controlled expansion of the green liquor. In other words, the green liquor is flashed to control temperature.

The factors that control the lime charge to the slaker are the temperature difference over the slaker and the degree of causticization from after the slaker and from after the last causticizer. The degree of causticization is measured by a new Procheck alkali analyzer. This method of controlling the lime charge to the slaker is so far the most reliable we have tested.

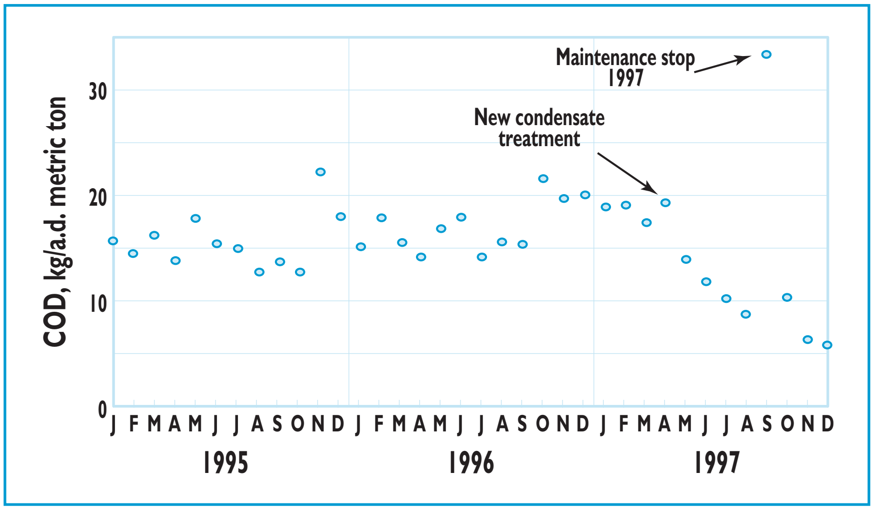
The lime mud is separated from the white liquor on a new pressure disc filter with a diameter of 3 m with 12 discs. The capacity of the filter is 7000 m³/day. The operation of the filter has been stable at all production levels experienced. The peak production levels have been about 330 m³/h (8000 m³/day). The capacity flexibility in the filter makes it possible to rapidly pick up the white liquor storage even at high production levels after disturbances in the white liquor production. Acid washing of the filter takes 2–3 h and has to be done about once a month.

Suspended solids are today maintained below 50 mg/L in the filtrated green liquor and between 25 mg/L and 50 mg/L in the white liquor. The dry solids content in the cake from the dregs filter is above 50%.

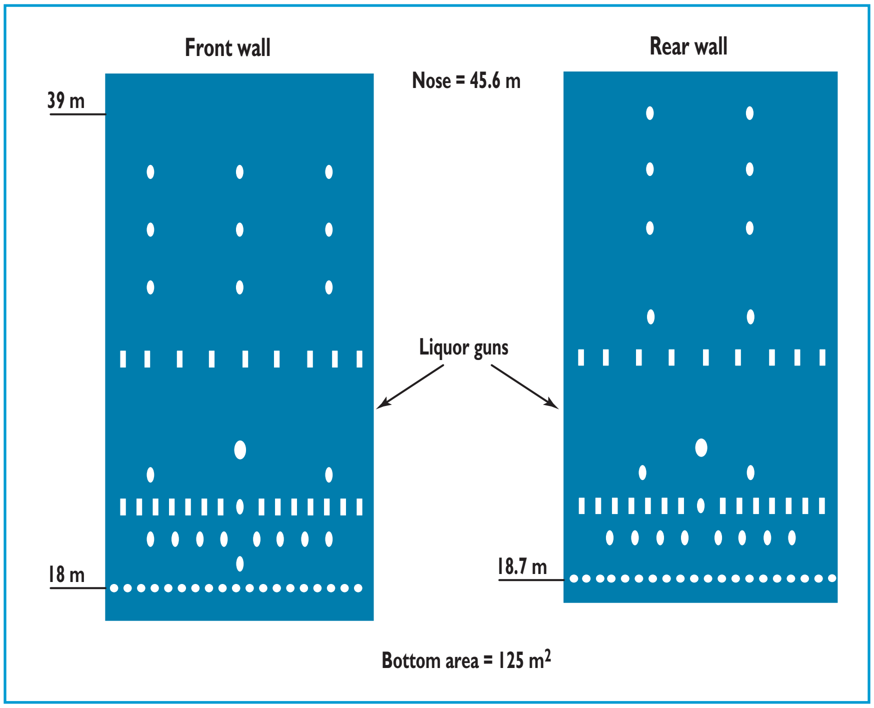
The modernization of the causticizing plant also included a total modernization of the electrical system and a new distributed control system. Kværner supplied the recausticization rebuild, and ABB supplied the control system.

The start-up and operation of the modernized causticizing plant has been relatively trouble free. The capacity of the white liquor filter meets our expectations, but the concentration of suspended solids in the white liquor was above the guaranteed level at first. Lime mud consumption has been unacceptably high as a result of large losses of lime mud in the green liquor filtration. The filtration process is being adjusted to lower the lime mud charge. For the moment, we do not see any possibility of operating the green liquor filters without lime mud.

Besides the main purposes of increasing capacity and improving quality, the goals included the improvement of the overall internal working environment. For example, we chose closed conveyors for lime charges in order to minimize the dusting. A new control room has also been built for the operators in the recausticizing and lime reburning area.



4. COD measured in the recovery area effluent from Jan. 1995 to Dec. 1997.



5. Arrangement of air ports according to the vertical air concept in Recovery Boiler 7.

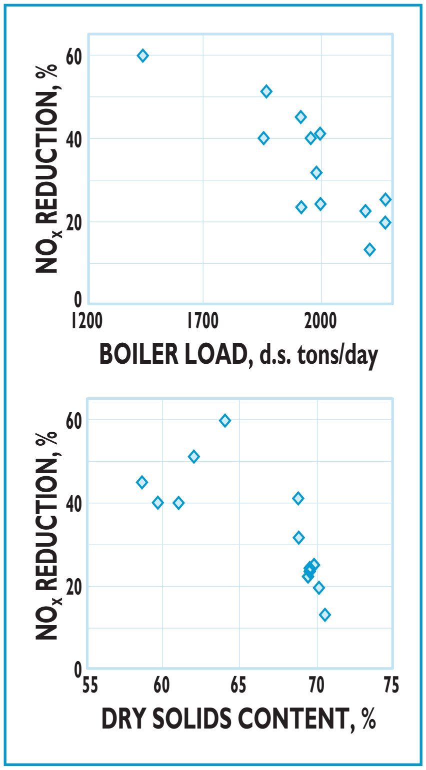
BURNING OF LVHC AND METHANOL IN RB 6

Burning of digester blow gases, evaporation vent gases (noncondensable gases), and stripper gases in the lime kilns caused frequent ring formation and plugging. Refractory damages related to the burning of noncondensable and stripper gases were also too frequent.

Lime kiln plugging reduced the operating capacity. The downtime for cleaning the kilns and for refractory repairs reduced the overall lime reburning capacity.

It was obvious that an alternative location for burning noncondensable gases and methanol had to be found to ensure kiln availability. The modernizations included a rebuild of the existing blow-gas collecting system from an open to a closed system.

The alternatives discussed for burning noncondensable gases (low-volatility hydrocarbons, or LVHC) and methanol were: (a) a separate incinerator, including heat recovery and SO_2 scrubbing and (b) burning them in the recovery boiler. Data from two Finnish mills indicated that burning methanol and noncondensable gases in the recovery boiler would lower the total NO_x



6. Reduction of NO$_x$ as a function of boiler load (upper) and as a function of dry solids content before the mixing tank (lower).

emissions, compared to burning them in a separate incinerator. Several Japanese mills also use the recovery boiler as the main place to burn noncondensable gases. These reports of good results and the cost effectiveness of this technique made us select this option of using the recovery boiler.

Since the lower furnace of Recovery Boiler 6 was to be replaced with a composite-tube furnace and since RB6, unlike RB7, is equipped with a flue gas scrubber, we selected Recovery Boiler 6 for burning noncondensable gases. The only drawback was that it would increase the sulfur load proportionally more than it would for the considerably larger RB7. An increased sulfur load has been the case, and it initially resulted in severe plugging problems during some periods.

When firing intensity decreases, the concentration of SO_2 increases rapidly in the flue gas, and the pH of ash decreases. At an ash pH below 4.0, ash rapidly plugs the boiler bank and the inlet to the electrostatic precipitator. A decrease in the dry solids content of the black liquor rapidly shows up as sticky ash. Also, the ash scrapers in the precipitator stop at times as a result of the sticky ash. The electrostatic precipitator is placed before the economizer, and the inlet flue gas temperature is above 350°C.

Fortunately, we have overcome these problems by establishing firing liquor concentrations of greater than 70% and by increasing the tertiary air. Purging precipitator ash from the RB6 recovery boiler helps to control the sulfidity level in the mill, which also raises the burning intensity in the furnace and the ash pH.

The burning of noncondensable gases in recovery boilers was not recommended by the Swedish Recovery Boiler Committee. Consequently, the installation of the burner for noncondensable gases and methanol at Skutskär, the first such installation in Sweden, initiated a discussion about this subject. After consideration, the committee has published a recommendation on system design and safety aspects related to the burning of noncondensable gases (*2*).

LIME REBURNING

Lime Kiln 1 was originally built for 170 tons of lime per day, and Lime Kiln 2 was originally built for 250 tons of lime per day. To meet the future pulp production, the lime reburning capacity has to be increased to 450 tons/day. That load corresponds to 170–200 tons/day for Kiln 1 and 250–280 tons/day for Kiln 2. This capacity increase can be met with minor process improvements and tuning.

The burning of noncondensable gases and stripper gases in the lime kilns results in reduced capacity as a result of ring formation and long downtimes for cleaning and refractory repair. To solve this problem, methanol and noncondensable gases have been burnt primarily in Recovery Boiler 6 since May 1997. The lime kilns are used today as back-up only for LVHC gases and methanol.

The chain sections in both kilns have been rebuilt. The original burners, which were pressure-atomized, have been replaced with steam-atomized burners. The noncondensable gas burner is concentrically placed in relation to the oil and methanol guns.

The front end of Kiln 2 has been rebuilt, and it includes a new Compax cooler. The new cooler is designed for a capacity of 325 tons/day and is thus prepared to meet a future capacity increase in case the kiln is equipped later with a flash dryer. The replacement of the kiln front end included 15 m of the kiln shell, the first live ring, and the roller set.

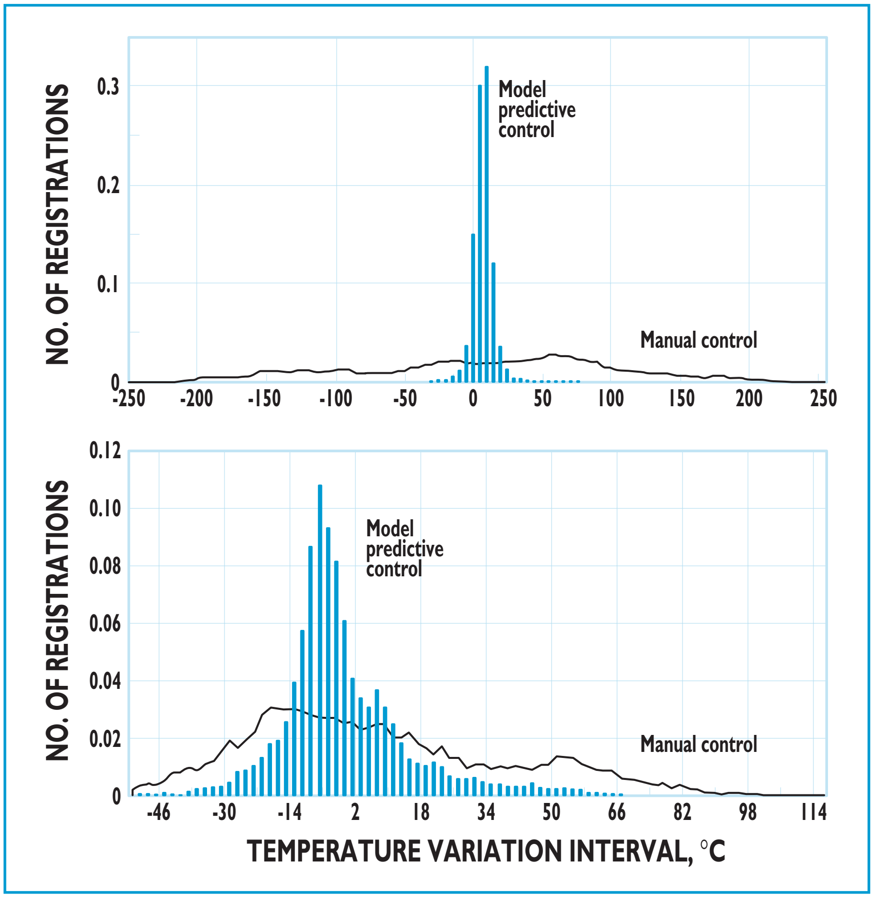
The mechanical design of the 104 m long kiln on three supports has been tight from the beginning. With the refractory change to higher densities and with an increase in the dam height, the front end of the kiln became heavily overloaded. Loads were exceeding the highest acceptable design load. Reoccurring refractory damages have also been a problem. The pattern of the damages indicate that overheating of the front end not only wore out the refractory but also increased the ovality of the shell. This damage promoted the collapse of the refractory in the burning zone. These factors initiated the decision to rebuild the kiln.

Low availability of the kilns was by far the most serious problem that needed to be straightened out. The manually operated kilns varied unacceptably in performance. Disturbances in operation were caused by the lime mud filters or by the switching of gases from one kiln to the other. The disturbances resulted in large variations in the front-end and back-end temperatures, with the front-end temperature often exceeding acceptable levels. Temperature variation causes the lime quality to vary and also leads to ring formation.

We had to improve the production stability and the lime quality. To reduce disturbances caused by filter blowing, we equipped the lime mud filters for Kiln 2 with high pressure jets to continuously renew the precoat. To improve the availability and operability of the kiln, we initially installed a control system. Promising results from some American mills using model predictive control for kilns helped us to select this strategy. Kiln 2 has been controlled by a model predictive controller (*3*), supplied by Symetric Control, since August 1996. The controller is an optimizing controller that uses dynamic models to predict and control the process. The kiln control problem is solved by the simultaneous optimization of a number of variables.

To stabilize the kiln operation, our initial strategy was to control the oxygen and back-end temperature to specific targets with high and low limits on the front-end temperature. When the front-end constraints were active, the controller would give the highest priority to satisfying those limits. It would then simultaneously minimize the deviation of back-end temperature and oxygen, along with reducing front-end variability. However, the front-end temperature variations were still regarded as high. We felt it was important to tighten the temperature variation to maximize the lifetime of the refractory and to improve lime quality.

The control configuration implemented has the front-end temperature as its primary control variable. The back-end temperature target is a function of the kiln load and is dynamically minimized until a low oxygen constraint limit is achieved. This control strategy gives stable front-end conditions while ensuring minimum energy consumption at all production rates.



7. Front-end (upper) and back-end (lower) temperature distributions for the kiln. The y axis represents the number of registrations within each interval of 5°C for the front end and within each interval of 2°C for the back end.

The graphs in **Fig. 7** show that the model predictive controller has made performance less variable than it was under manual operation. In each graph, the distribution curve for the reference case is generated by minute values collected for a two-month period, during which time operation was under manual control. The curve of the model predictive controller is drawn from data generated by minute values collected during the first four months of operation.

We have achieved a more stable operation in the kiln and have avoided continual refractory damages. Results show that the controller is doing an excellent job. The controller is the kiln's autopilot, and the project was paid for in less than six months. The controller is tuned to be active in order to reduce the variations as quickly as possible. To have an active controller is something new to the operators and has not been without its problems. However, the implementation of this controller has led to a better understanding of the lime reburning process among the operators.

CONCLUSIONS

The modernized recovery plant meets expectations in regard to capacity improvements. In addition, the modifications have reduced the effluent load. Extended condensate segregation together with extensive condensate stripping reduces the COD load in the mill effluents by more then 10 kg/a.d. metric ton.

The vertical air system implemented on RB7 has made it possible to increase the burning capacity of the boiler by 25%, to more than 2400 tons of dry solids/day. This increased load corresponds to a cross-sectional heat load of 2450 kW/m². NO_x reductions of up to 50% are achieved at lower boiler loads and liquor dry solids concentrations below 65%.

The burning of noncondensable gases in Recovery Boiler 6 is a competitive alternative to burning these gases in a separate incinerator. To avoid sticky ash, it is important to keep the burning intensity high in the boiler furnace.

Controlling Kiln 2 with model predictive control has given us excellent results in stabilizing the front-end and back-end temperatures of the kiln, improving and stabilizing the quality of lime. ***TJ***

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