ABSTRACT

Oxygen delignification and bleaching are integral parts of the closed mill concept. However, total closure can result in a chemical imbalance in the system as well as overloading of critical recovery equipment, such as evaporators, the recovery furnace, and the lime kiln. The chemical imbalance may be alleviated by purging some bleach effluent, by purging some recovery system streams (such as precipitator catch), or by using oxidized white liquor in the bleach plant. With the higher selectivity of oxygen vs. kraft chemicals, multiple oxygen stages may be used to reduce the impact on the recovery system load.

Pulp produced at a kraft kappa number of 40–50 followed by multiple oxygen stages and TCF bleaching, when compared to a conventional kraft (kappa no. 30) pulp followed by one oxygen stage and ECF bleaching, showed a similar or lower cost and a lower boiler throughput at constant production. Under constraints of limited digester capacity or limited boiler thermal capacity, production can be increased with a multiple oxygen/TCF closed system relative to a conventional kraft–oxygen system.

Application:

High-kappa kraft pulping coupled with multiple oxygen stages and TCF bleaching can be used to close a pulp mill without overloading the recovery system and with a production increase, under certain circumstances. This study also demonstrates the use of process simulation for determining optimal solutions to complex process problems.

Mill closure with high-kappa pulping and extended oxygen delignification

BRIJENDER PARSAD, ADRIANNA KIRKMAN, HASAN JAMEEL, JOSEF GRATZL, AND VINCE MAGNOTTA

IN THE INTEREST OF REDUCING THE environmental impact of pulp mill effluents, many companies are examining ways to modify effluent composition and reduce effluent volume. In the extreme, reduced effluent volume means a closed effluent-free pulp mill operation. With the chlorine-based bleaching processes currently in use, a closed mill is difficult to attain due to the corrosive behavior of the chlorine compounds when they are processed in the recovery system. In addition, for mills that are recovery limited, the closed-mill concept represents an unacceptable increase in the system load.

However, rapidly evolving experience with various nonchlorine bleaching chemicals, such as ozone and peroxide, is driving increased efforts to develop totally chlorine-free (TCF) bleaching. With TCF bleaching in a closed system, adsorbable organic halides (AOX) are eliminated, and the effluent with its attendant biological oxygen demand (BOD) can be recycled. The corrosion problem is also eliminated since no chlorine is present, but the recovery load issue must still be addressed. Other issues also arise with TCF bleaching, such as the potential for chemical (sodium/sulfur) imbalance in the recovery system *(1)*, and the impact of metal ions on the bleaching chemicals in a closed system. These process-related issues must be resolved before closure in the pulp mill can be fully implemented.

This study examines TCF bleaching and closure issues and economics through simulation of a kraft–oxygen mill with modification to the prebleaching and bleaching systems to incorporate TCF and closure. Optimum process configurations that allow a closed mill within various common mill constraints are determined. Metal ion buildup in the closed system will be examined in a later study.

PREVIOUS WORK

In previous work *(2)*, TCF bleaching systems were compared to a system of conventional kraft pulping to kappa no. 30, and kraft pulping to kappa no. 30 followed by oxygen delignification and bleaching [i.e., 30(O)]. Both of these "conventional" cases included elemental chlorine-free bleaching (ECF, sequence DEopD). The 30(O) case was also examined with TCF bleaching, sequence ZEopZP, with fresh NaOH used in the Eop and P stages. A summary of simulation results is shown in Table I *(3)*.

These results indicate that, with TCF bleaching, the load on the recovery system is increased by about 10% at key points, such as the evaporators, boiler, and lime kiln. In addition, the presence of sulfur makeup and negative saltcake makeup indicates the amount of the sodium imbalance in the system when the TCF inorganic chemicals are recycled to recovery.

The previous work demonstrated that the recovery load problem can be resolved by application of high-kappa pulping in combination with multiple stages of oxygen delignification. The process requires that kraft pulping be stopped at kappa no. 40–50 and be followed by two-stage oxygen delignification. Since the oxygen is more selective than the kraft liquor, the overall yield is increased, and less chips are used to

	30 ECF	30(O) ECF	30(O) TCF	40(OO) TCF	50(OO) TCF
Bleached pulp, o.d. tons/day	1100	1100	1100	1100	1100
Chip flow, o.d. tons/day	2762	2756	2853	2740	2592
Kraft yield, %	43.3	43.3	43.3	46.3	49.3
Post-O_2 yield, %	-	41.9	41.9	43.7	46.0
Overall yield, %	39.8	39.9	38.6	40.15	42.4
Recovery system loads					
Evap. solids, lb/o.d. ton	4134	4303	4700	4400	3908
Boiler solids, million lb/day	4.82	5.02	5.40	5.05	4.48
Kiln throughput, o.d. tons/day	318	336	361	337	288
White liquor makeups					
NaOH, lb/o.d. ton	41.1	36.5	0	0	0
Na_2SO_4, lb/o.d. ton	19.5	16	–46.8	–47.0	–49.0
Elem. sulfur, lb/o.d. ton	0	0	8.1	8.0	8.5

I. Effect of high-kappa pulping followed by multiple oxygen stages with ECF and TCF bleaching (3)

achieve the same bleached production.

Some of the results for two multiple-oxygen cases, 40(OO)-TCF and 50(OO)-TCF, are also shown in Table I. For the TCF sequences, the complete recycle of oxygen and bleach effluent is possible without increasing the recovery load relative to a kraft–oxygen system run in the conventional manner. Other major equipment in the recovery system is affected somewhat: Evaporator load and kiln throughput are both less for the 50(OO)-TCF case than for the 30(O)-ECF case. The 40(OO)-TCF case shows slight increases in the loads on the evaporators and kiln of 2.5 and 1.2%, respectively, relative to the 30(O)-ECF case.

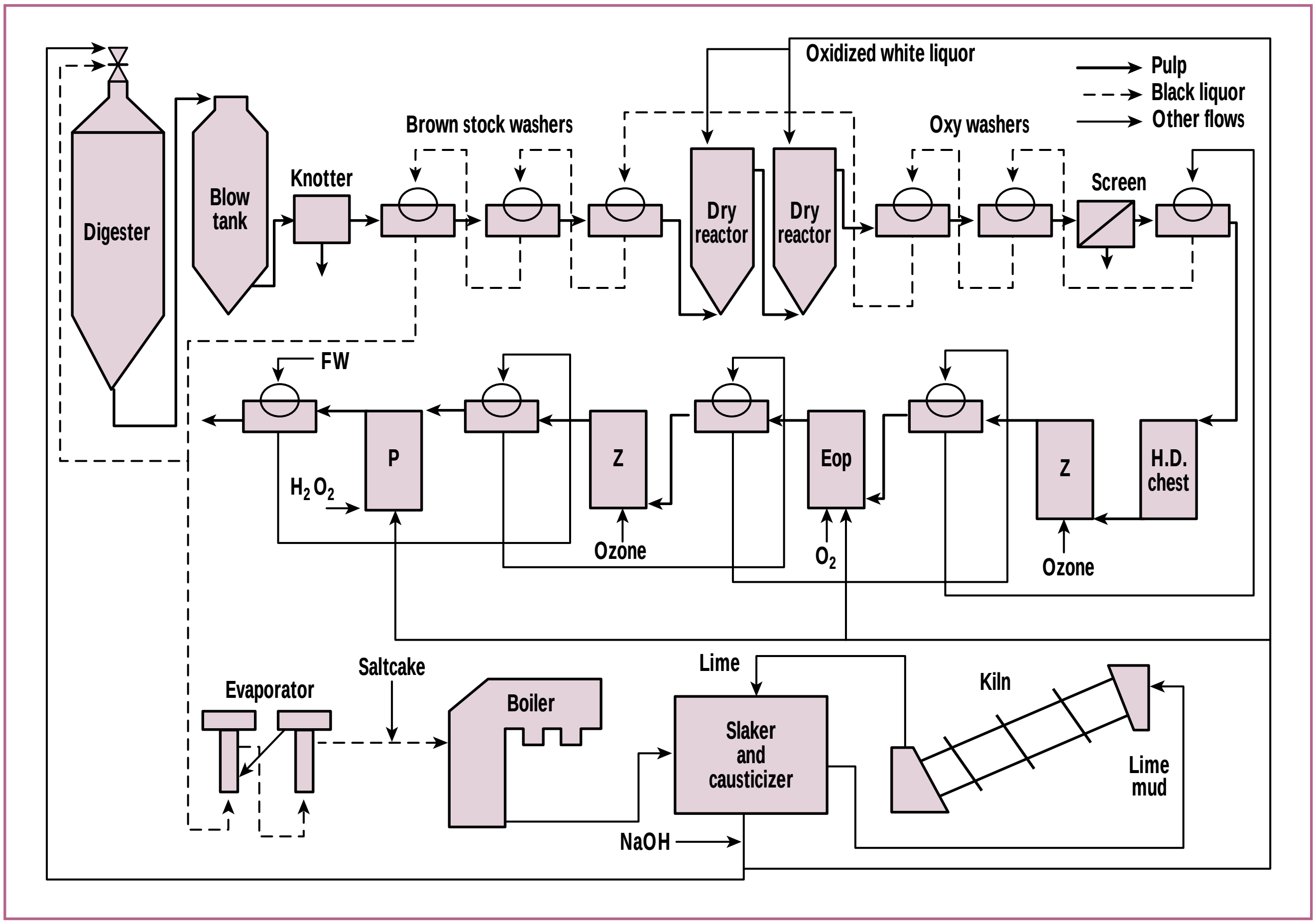
Thus, with multiple oxygen stages, it is possible to ameliorate some of the recovery system concerns that have hampered implementation of mill closure. However, the sodium/sulfur balance problem in the closed TCF systems must still be addressed. In addition, the previous study was restricted to constant production; this study examines mill closure under other constraints, such as limited digesters and thermally limited boilers.

SIMULATIONS AND CALCULATIONS

The schematic of a full mill with two oxygen stages and a TCF bleaching sequence is shown in Fig. 1. The flowsheet is comprised of batch digesters, multiple oxygen stages, TCF bleaching (ZEopZP sequence), and a full recovery system. This system was compared to a conventional kraft–oxygen system (kraft pulping to kappa no. 30 followed by an oxygen stage, and either DEopD or TCF bleaching). The effluents from the oxygen stages and from the TCF bleach plant were recycled to recovery. The screening was done after the oxygen stages to maximize the yield of the system. Initially fresh alkali was used throughout the bleach plant. When oxidized white liquor was used in bleaching *(4)*, it was oxidized to thiosulfate.

For ECF bleaching, the chlorine dioxide charge was based on the kappa number entering the bleaching plant and the organic carryover in the pulp. By including both kappa number and organic carryover, the impacts of both pulping and washing conditions are accounted for. The kappa factor for all cases was kept constant at 0.22. For the constant-production and thermally limited boiler cases, the bleached pulp production and the energy produced in the boiler, respectively, were controlled by adjusting the amount of chip feed to the digester.

For the digester-limited case, the chip feed was held constant. The heating value of the black liquor was calculated in the simulation based on the ratio of oxygen and kraft organics recycled to recovery, and included a lowered heating value for oxygen organics. The total titratable alkali (TTA) concentration of the white liquor was controlled by adjusting the kiln scrubber shower flow. Sodium and sulfur makeups were automatically adjusted to reach the specified effective alkali charge and white liquor sulfidity. If there is an excess of sodium in the recovery system, the simulation calculates a negative saltcake makeup stream. Since sulfur is also removed in this process, elemental sulfur makeup is calculated to achieve the correct sulfidity. Adsorbable organic halides (AOX) were calculated by using the correlation given by Earl and Reeve *(4)*. No effluent treatment costs have been included for open systems (ECF cases). A variable operating cost was calculated for each case, based on chemicals, steam usage, fuel usage, and energy generated in the boiler.



1. Schematic of fiber line and recovery with two-stage oxygen and TCF bleaching sequence

The computer simulation package general energy and materials balance system (GEMS) *(5)* was used for simulating all cases. The unit operations block diagram for the simulation of the multiple-oxygen-stage system is shown in Fig. 2. The assumptions made for the simulation and cost analysis are listed in Tables II and III. The nomenclature and description of the cases are listed in Table IV.

RESULTS AND DISCUSSION

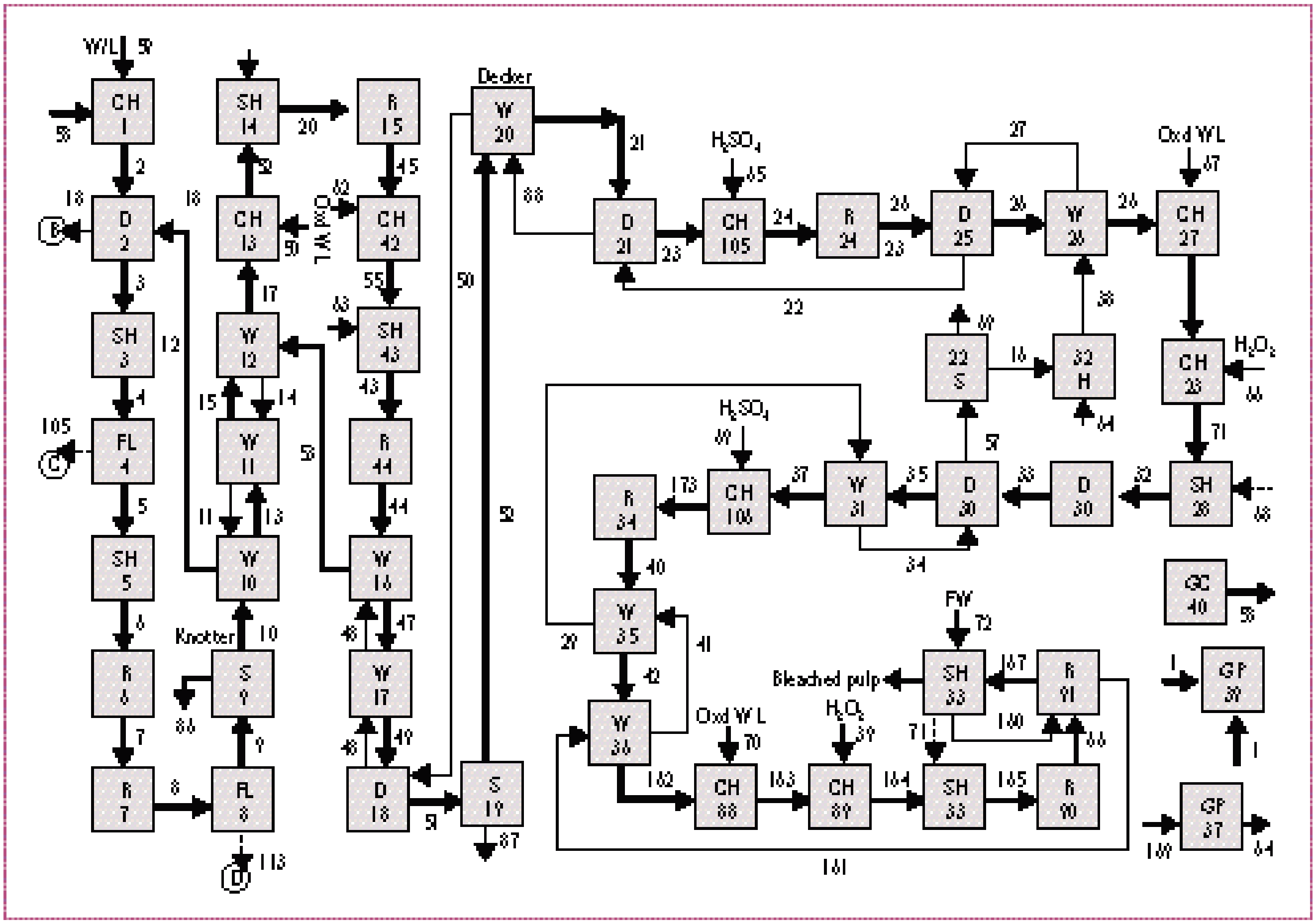
Sodium/sulfur balance

Three options were examined to eliminate the sodium-to-sulfur imbalance when the TCF bleach effluent is recycled:

- Reduce the amount of precipitator catch recycled to the boiler
- Purge some of the TCF effluent from the system
- Use oxidized white liquor in the bleach plant *(6, 7)*.

For the first two options, splits were taken on either the recycled precipitator catch or on the first ozone stage effluent in amounts that would result in zero or slightly positive makeups. The precipitator catch removed may be used elsewhere in the mill or could be used to produce NaOH and H_2SO_4 by electrolysis. If not used elsewhere, this material must be landfilled. The purging of a stream from the system may have the benefit of removing some of the metals that tend to build up in a totally closed system and that interfere with the bleaching process. However, either landfilling precipitator catch or purging bleach effluent results in a stream that must be discharged and thus represents a waste treatment cost to the mill. The results for these two options are shown in Table V.

From 18 to 24% (25.0–26.8 o.d. tons/day) of the precipitator catch must be purged to eliminate the excess sodium in the recovery system. Similarly, the TCF-effluent purge ranged from 23 to 27% of the flow—511 to 600 gal/min. (Note: The amount of flow removed was replaced at the washer with water to maintain the same hydraulics in the washing system.)



2. Flow diagram of the fiber line with two-stage oxygen and TCF bleaching

The 30(O)-TCF case under both scenarios has higher loads throughout the recovery system than the 30(O)-ECF system. The 40(OO) case has a higher evaporator load but is similar in terms of boiler and kiln loads for the precipitator catch purge case. This is to be expected, since the purge is taken after the evaporators. When TCF effluent is purged, both 40(OO) and 50(OO) cases have lower loads on the evaporators, boiler, and kiln since less flow is recycled. The energy available for steam generation is indicative of the thermal load on the boiler and is less in both multiple-oxygen systems under either discharge condition. It is interesting (but not unexpected) to note that the makeup composition ranges from mostly elemental sulfur if precipitator catch is discharged to mostly caustic if the effluent is discharged.

The third option requires the use of oxidized white liquor in the Eop and P stages of the bleach plant. Since oxidized white liquor (OWL) is being used in the oxygen stages, an oxidizing unit would already be in place, and this option becomes an obvious solution to the sodium problem. Oxidizer capacity may be a factor if the bleach plant demand for OWL is very high. In this scenario, there is no fresh alkali added in the bleach plant, and thus no excess of sodium is added relative to the kraft–oxygen pulping system requirements. Alkali makeup is added to the white liquor and is a function of both the pulping and bleaching needs. Table VI shows the results of using OWL for the caustic in the bleach plant.

The caustic makeup in the OWL cases ranges from 69 to 71 lb/o.d. ton vs. the zero value when fresh alkali is used. The saltcake becomes a positive number at a value of 1.6–1.9 lb/o.d. ton, and there is no elemental sulfur makeup in the OWL cases. The magnitude of the makeup relative to the 30(O)-ECF system is significantly larger, since the bleach plant demand is now included in the total system makeup requirement. The sulfuric acid used in the ozone stages adds a small amount of sulfur to the system, so the saltcake makeup is relatively low compared to an ECF system.

The OWL cases have slightly higher evaporator, boiler, and kiln loads than the fresh alkali cases, as would be expected, since more white liquor must be processed. However, the values for the 50(OO)-TCF case are actually lower than the 30(O)-ECF case, 4.57 vs. 5.02 million lb/day, and the 40(OO)-TCF case is only 2% higher (5.14 million lb/day) than the base case.

By using OWL, the variable operating cost drops by about US$ 2/o.d. ton, primarily as a function of the reduced alkali costs and in spite of the increased oxygen demand at the oxidizer. The increased cost relative to the 30(O) case is mostly a result of the more expensive TCF bleaching chemicals used (ozone and peroxide). Relative to the other methods of controlling the sodium balance, the OWL method is the least expensive, because the spent chemicals are retained in the system and because some of the makeup requirement for sodium is met with the cheaper saltcake in the recovery system instead of fresh NaOH in the bleach plant. The cost differences between the three options, in the range of US$ 3–5/o.d. ton, do not take into account the cost of landfill or effluent treatment of any purge streams. Costs associated with metals buildup when there is no purging of bleaching effluent were also not included.

Closure with a thermally limited boiler

In the constant-production cases previously discussed, it was found that the load of solids on the boiler was decreased, since the yield improved and thus the chip feed could be decreased. Many mills, however, are thermally limited in the boiler. The impact of the multiple oxygen stages, TCF bleaching, and full closure is shown in Table VII. The energy production was held to a target of 858 million Btu/h.

To implement a single oxygen stage and TCF bleaching with closure at the same energy for steaming, the production must drop by 72 o.d. tons/day, a 6% decrease in production with a 2% increase in furnace solids load relative to a 30(O)-ECF system. The solids increase is primarily the result of recycling the bleaching solids containing both organics and inorganics. Since the bleaching yield in TCF is lower [38.6% than ECF (39.9%)], more organics are dissolved in bleaching

	30 (O)	30 (OO)	40 (OO)	50
AA charge, % as Na_2O	19.84	19.84	17.81	15.79
Sulfidity, %	25	25	25	25
Yields, %				
Kraft	43.3	43.3	46.3	49.3
Post-O_2	-	41.9	43.7	45.9
Overall-ECF	39.8	39.9	41.8	43.6
Overall-TCF	39.8	38.6	40.2	42.4
Kraft kappa no.	30	30	40	50
Washing shower, gal/min	2091	2091	2091	2091
Oxygen stage(s)				
NaOH charge, % (fresh/oxid. wh. liquor)	-	3	3/3	3/3
O_2 kappa no.	-	15.5	13.0	17.0
Oxygen, %	-	2.6	4.6	5.6
Bleaching				
Kappa factor	0.22	0.22	0.22	0.22
ECF–DEopD				
ClO_2 charge D0, %	2.77	1.35	1.16	1.51
Caustic charge, %	3.96	2.04	1.72	2.24
H_2O_2 charge, %	0.25	0.25	0.25	0.25
Oxygen, %	0.5	0.5	0.5	0.5
ClO_2 charge DI, %	1	1	1	1
TCF–ZEopZP				
Ozone charge Z1, %	-	0.75	0.75	0.75
Ozone charge Z2, %	-	0.52	0.52	0.52
H_2SO_4 charge, %	-	1	1	1
Alkali charge, E/P, %	-	3/1.34	3/1.34	3/1.34
H_2O_2 charge, %	-	1	1	1
Oxygen, %	-	0.5	0.5	0.5

II. Assumptions for simulation cases. Product: bleached softwood pulp.

Material	Unit cost	Material	Unit cost
Chips, US$/kg	0.062	Oxygen, US$/lb	0.040
NaOH, US$/lb	0.141	$MgSO_4$, US$/lb	0.223
Na_2SO_4, US$/lb	0.050	H_2SO_4, US$/lb	0.032
Sulfur, US$/lb	0.070	Kiln fuel, US$/bbl	20.00
Lime, US$/lb	0.032	Steam, US$/kg	0.006
H_2O_2, US$/lb	0.573	BOD, US$/kW·h	0.025
Ozone, US$/lb	1.000	Pulp, US$/ton	285
ClO_2, US$/lb	0.300		

III. Cost assumptions for variable operating cost calculation

Case	Kraft kappa no.	No. of O_2 stages	Bleach kappa no.	Bleach sequence	Case name
Base	30	0	30.0	DEopD	30-ECF
Case 1	30	1	15.5	DEopD	30(O)-ECF
Case 2	30	1	15.5	ZEopZP	30(O)-TCF
Case 3	40	2	13.0	ZEopZP	40(OO)-TCF
Case 4	50	2	17.0	ZEopZP	50(OO)-TCF

IV. Description and nomenclature of simulation cases

	ESP CATCH PURGE				TCF EFFLUENT PURGE		
	30 (O) ECF	30 (O) TCF	40 (OO) TCF	50 (OO) TCF	30 (O) TCF	40 (OO) TCF	50 (OO) TCF
Bleached pulp, o.d. tons/day	1100	1100	1100	1100	1100	1100	1100
Chip flow, o.d. tons/day	2756	2853	2740	2592	2853	2740	2592
Precipitator catch purge, %	-	18.1	19.6	24.2	-	-	-
TCF effluent purge, %	-	-	-	-	23	24	27
Recovery system loads							
Evaporator solids, lb/o.d. ton	4304	4700	4400	3910	4598	4293	3802
Boiler solids, million lb/day	5.02	5.40	5.05	4.48	5.33	4.98	4.40
Kiln throughput, o.d. tons/day	336	361	337	288	359	334	286
Energy prod., million Btu/h	806	865	792	716	851	778	702
White liquor makeups							
NaOH, lb/o.d. ton	36.5	0	0	0	5.3	6.7	6.6
Na_2SO_4, lb/o.d. ton	16.0	0.9	0.96	0.8	0.2	0.03	0.4
Elemental sulfur, lb/o.d. ton	0	6.0	5.9	6.4	0	0	0
NaOH in bleaching, lb/o.d. ton	42.2	90.8	90.8	90.6	90.8	90.8	90.6
Net variable op. cost, US$/o.d. ton[a]	193.2[b]	225.5	227.5	221.3	227.8	230.2	224.0

[a] Costs for landfilling precipitator catch and TCF effluent treatment not included
[b] Effluent for 30(O)-ECF: 9 million gal/year, AOX=2.55 lb/o.d. ton, BOD=4355 lb/h

V. Adjustment of sodium/sulfur balance by purging precipitator catch or TCF effluent

per unit of pulp. With the lower 30(O)-TCF chip feed, however, a 2.2% decrease in organics is expected to be generated. By recycling the bleaching stream, the overall organics flow to recovery increases. The overall inorganics increase due to the increased demand for oxidized white liquor in TCF bleaching. Thus, in comparing 30(O)-TCF to 30(O)-ECF bleaching, the net changes in solids composition to the boiler to achieve the target steam production are 1.9% increase in organics and 4.8% increase in inorganics.

As the kraft kappa number is increased and a second oxygen stage is added, the production can be increased compared to both of the 30(O) cases. Incremental production values above the 30(O)-TCF case are 109 o.d. tons/day and 271 o.d. tons/day for the 40(OO)-TCF and 50(OO)-TCF cases, respectively. These increases are the result of the lower heating value of the oxygen/bleaching liquors, which allow more solids to be burned. In addition, the lower wood solids dissolved in the more selective oxygen stages, compared to kraft pulping, further reduces the flow of organics to the boiler. Increasing production results in increased total solids flow: more inorganics to pulp and bleach

	FRESH ALKALI				OXIDIZED WHITE LIQUOR		
	30 (O) ECF	30 (O) TCF	40 (OO) TCF	50 (OO) TCF	30 (O) TCF	40 (OO) TCF	50 (OO) TCF
Bleached pulp, o.d. tons/day	1100	1100	1100	1100	1100	1100	1100
Chip flow, o.d. tons/day	2756	2853	2740	2592	2853	2740	2592
Recovery system loads							
Evaporator solids, lb/o.d. ton	4304	4700	4400	3908	4721	4422	3933
Boiler solids, million lb/day	5.02	5.40	5.05	4.48	5.49	5.14	4.57
Kiln throughput, o.d. tons/day	336	361	337	288	364	339	291
Energy prod., million Btu/h	806	865	792	716	858	785	709
White liquor makeups							
NaOH, lb/o.d. ton	36.5	0	0	0	70.9	71.0	69.0
Na_2SO_4, lb/o.d. ton	16.0	–46.8	–47.0	–49.0	1.9	1.6	1.9
Elemental sulfur	0	8.1	8.0	8.5	0	0	0
NaOH in bleaching, lb/o.d. ton	42.2	90.8	90.8	90.6	0	0	0
Net variable op. cost, US$/o.d. ton	193.2	225.5	227.6	221.4	223.9	225.9	219.5

VI. Comparison of alkali source for TCF bleaching in a closed system

	30(O) ECF	30(O) TCF	40(OO) TCF	50(OO) TCF
Bleached pulp, o.d. tons/day	1172	1100	1209	1371
Chip flow, o.d. tons/day	2937	2853	3014	3231
Kraft yield, %	43.3	43.3	46.3	49.3
Post-O_2 yield, %	41.9	41.9	43.7	46.0
Overall yield, %	39.9	38.6	40.2	42.4
Recovery system loads				
Evap. solids, million lb/day	5.03	5.19	5.33	5.32
Evap. flow, gal/min	2044	2178	2015	1694
Boiler solids, million lb/day	5.34	5.49	5.62	5.49
Kiln throughput, o.d. tons/day	358	364	371	347
Energy prod., million Btu/h	858	858	858	858
Net variable op. cost, US$/o.d. ton[a]	-	223.9	199.7	166.8
Net variable op. cost, US$/o.d. ton[b]	182.8	242.6	217.9	186.8

[a]Incremental production credit relative to 30(O)-TCF
[b]Incremental production credit relative to 30(O)-ECF

VII. Comparison of TCF cases at constant energy for steam production in the boiler

more wood, and increased organics to provide sufficient energy to offset the increased inorganic load while still producing a constant energy for steam production. This increased solids load is seen in all TCF cases at the evaporators and at the boiler. The double-oxygen cases, however, have lower evaporator volumetric flow rates and higher percent solids than the 30(O)-ECF case. A lower steam requirement results in those two cases.

For the 50(OO)-TCF case, there is also a lower lime kiln load than in

Property	30(O) ECF	30(O) TCF	40(OO) TCF	50(OO) TCF
Bleached pulp, o.d. tons/day	1139	1100	1145	1210
Chip flow, o.d. tons/day	2853	2853	2853	2853
Recovery system loads				
Evap. solids, million lb/day	4.89	5.19	5.05	4.70
Evap. flow, gal/min	2041	2177	2052	1817
Boiler solids, million lb/day	5.19	5.49	5.34	4.98
Kiln throughput, o.d. tons/day	348	364	353	318
Energy prod., million Btu/h	834	858	815	774
Net variable op. cost, US$/o.d. ton*	226.2	233.9	224.1	202.6

*Incremental production credit relative to 30(O)-ECF

VIII. Comparison of TCF cases at constant chip flow to digester

40(OO)-TCF. This results from higher NaOH makeup added after the white liquor clarifier due to higher soda losses, particularly with the pulp, at the higher 50(OO)-TCF production rate. These evaporator and kiln conditions are the result of maintaining a constant wash shower flow rather than a constant dilution factor in the simulation.

As production increases, a credit can be taken for increased sales revenue, which offsets the variable operating costs. Obviously, the credit taken depends on the current selling price of the bleached pulp; US$ 285/o.d. ton of incremental production was used as the incremental profit for comparison purposes. Relative to the 30(O)-TCF system, the costs drop from US$ 224/o.d. ton to US$ 200/o.d. ton and US$ 167/o.d. ton for 40(OO)-TCF and 50(OO)-TCF, respectively. Relative to 30(O)-ECF, the TCF systems tend to be more expensive because of the bleaching chemical costs. However, the 50(OO)-TCF case has a comparable cost at US$ 187/o.d. ton vs. US$ 183/o.d. ton for the 30(O)-ECF system.

Closure in a digester-limited system

For a digester-limited mill, the chip feed must remain constant. The results for a TCF/closed mill, for a constant chip feed of 2853 o.d. tons/day, are shown in Table VIII. By implementing TCF bleaching and closure in a kappa no. 30 kraft–oxygen system, production drops from 1139 to 1100 o.d. tons/day because of the lower bleaching yield in the 30(O)-TCF system compared to the 30(O)-ECF. By using multiple oxygen stages, production increases by 0.5 and 6% for 40(OO)-TCF and 50(OO)-TCF, respectively. The net cost relative to 30(O)-ECF at US$ 226/o.d. ton is slightly lower for 40(OO)-TCF at US$ 224/o.d. ton and significantly lower for 50(OO)-TCF at US$ 203/o.d. ton.

Relative to the 30(O)-ECF base case, in all other measures, such as boiler load, evaporator load, evaporator steam requirement, and kiln throughput, the 30(O)-TCF case is significantly higher, 40(OO)-TCF is comparable or slightly higher, and 50(OO)-TCF is significantly lower. Thus, under the constraint of a limited digester, a kappa no. 30 kraft–oxygen mill might opt for 40(OO)-TCF if it can handle a 2.5% increase in boiler solids load, a 2% decrease in steam production, a 0.5% increase in evaporator liquor flow rate and a 1.4% increase in kiln throughput. Otherwise, 50(OO)-TCF would be the best choice, with the major concern being the 7% drop in steam production.

CONCLUSIONS

The following conclusions can be drawn:

- If fresh alkali is used in a TCF bleaching sequence with complete recycle of effluent, the amount of sodium returned to recovery exceeds the requirements for the white liquor.
- Using oxidized white liquor in TCF bleaching to maintain the sodium/sulfur balance in the liquor system results in the lowest operating cost compared to either purging precipitator catch or purging some of the bleach plant effluent.
- Boiler loads are higher for sequences using oxidized white liquor than for the same sequences using fresh alkali, or for systems that purge either saltcake or effluent. However, the boiler load for 50(OO)-TCF is still lower than conventional 30-ECF, 30(O)-ECF and 30(O)-TCF under all scenarios considered.
- For a system with a thermally limited boiler, TCF bleaching and full closure can be implemented with a 50(OO)-TCF configuration. The product cost is US$ 4/o.d. ton higher than a 30(O)-ECF system. Solids loading at the boiler is higher, but other throughputs in the recovery system are similar or lower than the 30(O)-ECF system.
- For digester-limited systems, TCF bleaching and full closure can be achieved with a production increase. Both 40(OO)-TCF and 50(OO)-TCF provide lower product cost than 30(O)-ECF. Slightly higher boiler load occurs with 40(OO)-TCF, and significantly lower steam production occurs with 50(OO)-TCF. TJ

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