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Coordination of Process Integration and Exergoeconomic Methodology for Analysis and Optimization of a Pulp and Paper Mill

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Abstract. *By simultaneously applying pinch technology and the exergoeconomic method to a complex process system, beneficial and energy-efficient measures are identified. The "three-link-model" exergoeconomic methodology optimizes the design and operability of a system. In this work, contrary to traditional exergoeconomic methods, a reversed method is used. The approach proposed for the optimization of such a complex system is to iteratively optimize subsystems. Since the reversed exergoeconomic method is used, assumptions considered by Tsatsaronis (based on four assumptions for calculating the cost-optimal exergetic efficiency and relative cost difference) are not applicable and new assumptions are to be considered. Unlike traditional exergoeconomic methods, the product exergetic specific cost is considered to be known and the objective will be to maximize the exergetic cost of the fuel. Heat flows costs are calculated with the assistance of a Pinch analysis. The strength of the combination of a Pinch analysis and the exergoeconomic method is elucidated in a case study applied to the Mazandaran wood and paper industry. Replacement of the pressure valve and Direct Cyclone Contact Evaporation (DCCE) is proposed, while by selection of the optimum decision variable and applying Pinch technology simultaneously, the recoverable black liquor could be increased by 7% and energy consumption decreased by 12%.*

Keywords: *Pinch technology; Exergoeconomic methodology; Process system; Optimization; Pulp and paper.*

INTRODUCTION

The pulp and paper industry is under increasing pressure to reduce its impact on the environment and its energy consumption. When modeling a system, it is very important to know where to introduce the system boundaries. Different results may be obtained, based on the boundaries chosen.

Pinch technology is a method to improve Heat Exchanger Networks (HEN), only improving the structure of a HEN under specific conditions. One key to good energy efficiency is the exchange of heat in the most effective way between components within the system, thereby cutting the need for additional heating or cooling [1]. The application of process integration tools and methods in the pulp and paper industry dates back to the 1990's when Pinch analyses of pulp and paper mills were first made. It was not until the late 1990's and early 2000's that process integration became more widely used as a tool for energy analysis and retrofit application in the pulp and paper industry [2].

Noel generalized the use of a Pinch analysis in pulp and paper mills in 1995 by applying the Pinch analysis to steam, water and liquor streams and, also, direct heat exchangers for the retrofitting of existing heat changer networks [3,4].

In 1999, Lombardo et al. [5] practiced Pinch and exergy analysis on a Kraft pulp and paper mill retrofitting. However, only indirect heat exchangers were considered and the exergy analysis was applied to each unit individually. The Department of Heat and Power Technology at Chalmers University of Technology in Sweden is a pioneer in the retrofitting of the pulp and paper industry and also in application of the Pinch analysis to pulp and paper mills [6-10].

Paris et al. have applied MILP (Mixed Integer Linear Programming) and Pinch technology to a pulp and paper mill and energy and exergy recovery opportunities have been examined to improve integration of

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the utility system. The MILP optimization targeting method has been applied to identify the best energy conversion options and to optimize production of combined heat and power [11].

Exergoeconomics combine exergy analysis and economic principles to provide the system designer with information not available through conventional energy analysis and economic evaluations, but crucial to the design and operation of a cost-effective system. A general methodology for this kind of analysis was presented by Tsatsaronis in 1985 [12] which was later called the exergoeconomic accounting technique [13]. In recent years, a different direction of developments in exergoeconomics has been taken and presented by Tsatsaronis et al. [14-18], Valero et al. [19-22], Fragopoulos [23] and Spakovsky [24]. Tsatsaronis introduces an iterative exergoeconomic optimization procedure based on exergoeconomic variables (relative cost difference, exergoeconomic factor and exergetic efficiency) and minimization of the product cost of each system; components are treated in thermoeconomic isolation. In the case of pulp and paper making processes, the problem, due to the high number of simultaneous products, is more complex compared to cogeneration systems producing only heat and power. Few studies have proceeded to systems which produce products other than heat and power [25-28]. No studies have applied exergoeconomics to pulp and paper processes (energy use and recovery processes).

APPLYING REVERSED EXERGY COSTING METHOD TO RECOVERY LINK

Exergy improvement potential is a measure for determining system performance, showing how much and how easy the system could be improved for optimization purposes [26].

$$\text{Pot} = Irr(1-\varepsilon) + Efl, \tag{1}$$

where:

$$Irr = \sum E_F - \sum E_P, \tag{2}$$

$$\varepsilon = \frac{E_{ntp}}{E_{nts}} = \frac{E_P}{E_F}, \tag{3}$$

$$Efl = \sum E_{\text{rejected to the environment}}. \tag{4}$$

It is formed by three contributions: Absolute potential (Irr), relative potential $(1-\varepsilon)$ and environmental potential (Efl). The relative potential $(1-\varepsilon)$ is a measure showing the system potential for improvement; if the effectiveness is very low, the relative potential approaches its maximum value, so in principle the system could be improved from inside itself. In order to optimize a system, it is necessary to preferably approach the blocks with the higher exergy improvement potential.

Exergoeconomic traditional sequential costing methods deal with energy systems; systems which convert one certain type of energy into another. Conversely, process systems produce certain products from certain raw materials.

Decomposition strategies have been proposed to reduce the complexity of complete systems. A 'three-link-model' energy/mass structure for process systems is proposed by Zhang et al. [25]; a rigorous quantitative mathematical energy/mass model which is suitable for any process system. This study is based on a 'reversed costing' strategy for computing recoverable energy/mass costs in recovery subsystems.

This model makes it possible to divide whole systems into main subsystems and recovery subsystems, optimizing each subsystem separately and improving the total system towards an optimum state. The 'energy/mass recovery' link recovers raw materials needed for main subsystems in addition to energy.

In many systems, production costs are being reduced by the recovery of whole or parts of rejected streams. With the development of technology and changes in economical conditions and environment constraints, it has become possible to recover rejected energy/mass more and more effectively. The cost of recoverable exergy depends on the extent of recovery. Therefore, only the reversed exergy costing principle is available. The money balance equation for the recovery link is as follows [25]:

$$\sum_j c_{Oj}E_{Oj} + \sum_i c_{Dj}E_{Dj} + \sum_j Z_{Rj} = \sum_i c_{Ej}E_{Ej} + \sum_i c_{Ri}E_{Ri}. \tag{5}$$

The exergy-mass-economic model for the recovery link is shown in Figure 1.

Equation 5 differs from a conventional money balance. In this equation, the unknown term, c_{Oj} is not at the right side of the equation, but at the left side. First, the quantity of the recycled exergy, E_{Ri}, and the recovered export exergy, E_{Ei}, as well as their costs, c_{Ri} and c_{Ej}, respectively, should be determined.

'Reversed costing' means that the cost of the recoverable exergy E_O depends on the cost of the recovery driving exergy E_D and the exergies recovered from the energy-recovery subsystem, E_R and E_E. Factor c_D will be calculated from the energy-conversion link by using a sequential costing method.

The next step is determination of c_E and c_R. Two methods are suggested for calculating c_E:

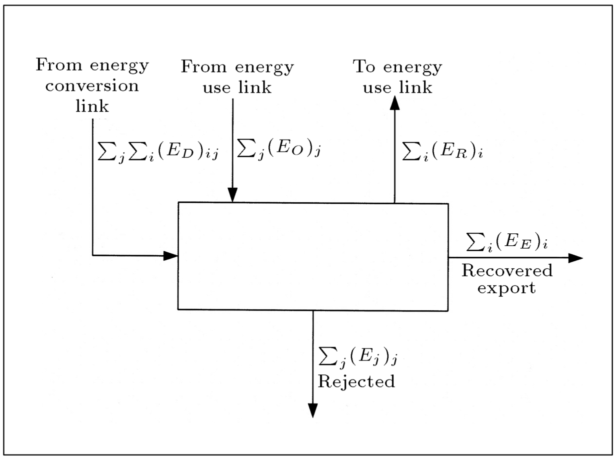

Figure 1. Recovery link exergy-mass-economic model.

1. If c_E is the specific cost of power, its cost is the same as the market cost of the same flow;
2. If c_E is the specific cost of heat, it can be calculated with the assistance of a Pinch analysis. c_R is the same as the market price of the same material.

These principles connect the energy use subsystem to the energy/mass recovery subsystem and have a great influence on the optimization results. The cost of the irreversible, effluent and total exergy losses of each unit and their exergoeconomic improvement potential is then calculated with the following expressions:

$$C_{irr,U} = Irr\ \bar{c}_{tte,U}, \tag{6}$$

$$C_{efl,U} = Efl\ \bar{c}_{tte,U}, \tag{7}$$

$$\mathrm{Potec}_U = C_{irr,U}(1-\varepsilon) + C_{efl,U}, \tag{8}$$

where:

$$\bar{c}_{tte,U} = \frac{\sum_{i=1}^{IN} E_{Ui} c_i}{E_{tte,U}}. \tag{9}$$

CALCULATING HEAT FLOW COST

For calculating c_E by Pinch technology, first energy-use link warm and cold streams are identified from mass and energy balance process simulations. The heat carrying streams are listed. Composite curves have to be drawn; minimum hot and cold utility requirements can be evaluated from composite curves and network capital cost has also to be calculated. In the next step, all warm and cold streams plus the "recovered stream" from the recovery link (the cost of which has to be calculated) are identified. Minimum hot and cold utility requirements and network capital cost will again be evaluated. The specific cost of a recovered stream is evaluated as follows:

$$c_E = \frac{(Q_{min,1} - Q_{min,2}) \times ce_D}{E_E} - \frac{(\text{network capital cost}_2 - \text{network capital cost}_1)}{E_E}. \tag{10}$$

COORDINATION OF EXERGOECONOMIC METHOD AND PINCH TECHNOLOGY

Bengtsson et al. have proposed a structure for combining Pinch technology and the MIND method [6]. The proposed procedure for combining Pinch technology and the MIND method is iterative. In this study, a similar structure is proposed for combining Pinch technology and the exergoeconomic method (see Figure 2).

The iterative procedure may be accomplished by using the exergoeconomic analysis to determine an investment opportunity for different alternatives.

OPTIMIZATION OF SYSTEM

The objective of this study is to minimize the total cost of mass and energy consumption, exergy destruction and the investment equipment for the whole system. Apart from investment and maintenance costs, the system's essential costs consist of electricity, steam and raw material costs. Therefore, the goal is to reduce their consumption. Exergoeconomic and Pinch methodology is applied for optimization, where Pinch analysis is applied to the whole system. In this study, since the energy/mass recovery subsystem is the most energy consuming subsystem, exergoeconomics are applied only to this subsystem. According to the reversed costing strategy proposed by Zhang et al. [25], the goal is to get the maximum cost of the recoverable mass, E_O, at the minimum driving exergy, E_D. The objective function of exergoeconomic optimization is:

$$\max c_O = \frac{(\sum_i c_{Ej}E_{Ej} + \sum_i c_{Ri}E_{Ri} - \sum_i c_{Dj}E_{Dj} - \sum_j Z_{Rj})}{\sum_J E_{Oj}}. \tag{11}$$

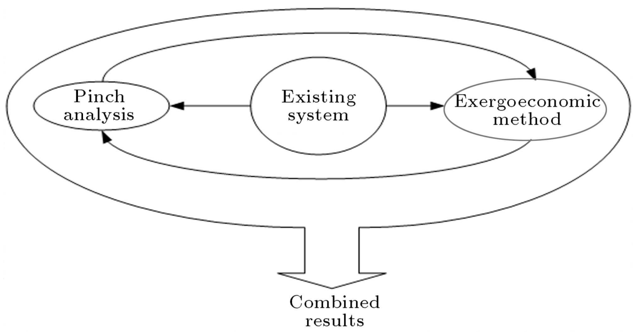

Figure 2. Procedure for combined method.

c_O is the coordinating variable of the energy-use link and mass/energy recovery link. Mass/energy recovery link optimization is done under the boundary constraint, E_O, which produces the optimization results of E_D and the new value of c_O. Iteratively, an optimization of the whole system and forward recovery link will be done. The feasible coordinating rules are:

$$(c_{Oj})_{k+1} = (c_{Oj})_k, \tag{12}$$

$$(E_{Oj})_{k+1} = (E_{Oj})_k. \tag{13}$$

The convergence judgment rules are:

$$|(c_{Oj})_{k+1} - (c_{Oj})_k| \leq \varepsilon. \tag{14}$$

Because of system complexity and the unavailability of input data and functions required, especially for black liquor, mathematical techniques cannot be used for optimization.

The usual approach to the optimization of such a complex system is to iteratively optimize subsystems. An alternative to this approach is an iterative exergoeconomic optimization technique proposed by Tsatsaronis, which consisting of seven steps [13].

Among the most important parameters in such an optimization are cost-optimal values for exergetic efficiency and the relative cost difference. Tsatsaronis [13] has proposed an approach, based on four assumptions, for calculating the cost-optimal exergetic efficiency and relative cost difference for a component isolated from the remaining system components. In this study, since a reversed exergoeconomic method is used, the assumptions considered by Tsatsaronis are no longer valid and new assumptions have to be considered. These assumptions are as follows:

1. The exergy flow rate of fuel, $\dot{E}_F$, and the unit cost of product, c_P, remain constant for the kth component:

$$\dot{E}_{F,k} = \text{constant}, \tag{15}$$

and:

$$c_{P,k} = \text{constatnt}. \tag{16}$$

2. Every system component investment cost increases with its capacity and exergetic efficiency. The total capital investment of the kth component can be represented by:

$$\text{TCI}_k = B_k \left(\frac{\varepsilon_k}{1-\varepsilon_k}\right)^{n_k} \dot{E}_{F,k}^{m_k}, \tag{17}$$

where:

$$\varepsilon_k = \frac{E_{P,k}}{E_{F,k}}.$$

Parameter B_k and exponents n_k and m_k are constant. These parameters are calculated by use of a purchased-equipment cost and least square method.

3. Annual levelized operating and maintenance costs attributed to the kth component are considered as:

$$Z_k^{\text{OM}} = \gamma_k(\text{TCI}_k) + \omega_k \tau \dot{E}_{F,k} + R_k. \tag{18}$$

The coefficient γ_k for such a system is considered as 0.1; ω_k is a constant that accounts for the variable operating and maintenance costs associated with the kth component and denotes the O & M cost per unit of the product exergy; τ is the average annual time of plant operation at the nominal load; and R_k includes all the remaining operating and maintenance costs that are independent of the total capital investment and the exergy of the product. The last two terms on the right side ($\omega_k \tau \dot{E}_{F,k}$ and R_k) may be neglected since they are small compared with the remaining term.

4. This study tries to optimize an established system, $Z_k^{\text{CI}} = 0$, by which Equations 19 to 21 are obtained:

$$\begin{aligned} Z_k &= Z_k^{\text{CI}} + Z_k^{\text{OM}} = Z_k^{\text{OM}} \\ &= \gamma_k(\text{TCI}_k) + \omega_k \tau \dot{E}_{P,k} + R_k. \end{aligned} \tag{19}$$

Thus, cost rate will be obtained as:

$$\dot{Z}_k = \frac{\gamma_k}{\tau}(\text{TCI}_k) + \omega_k \dot{E}_{F,k} + \frac{R_k}{\tau}. \tag{20}$$

By use of Equation 20:

$$\begin{aligned} \dot{Z}_k = {} & \frac{\gamma_k B_k}{\tau}\left(\frac{\varepsilon_k}{1-\varepsilon_k}\right)^{n_k} \dot{E}_{F,k}^{m_k} \\ & + \omega_k \dot{E}_{F,k} + \frac{R_k}{\tau}. \end{aligned} \tag{21}$$

The objective function to be maximized expresses the cost per exergy unit of product for the kth component. Accordingly:

$$\text{Maximize } c_{F,k} = \frac{c_{P,k}\dot{E}_{P,k} - \dot{Z}_k}{\dot{E}_{F,k}}. \tag{22}$$

So:

$$\begin{aligned} \text{Maximize } c_{F,k} = {} & c_{P,k}\varepsilon_k - \frac{\gamma_k B_k}{\tau E_{F,k}^{1-m_k}}\left(\frac{\varepsilon_k}{1-\varepsilon_k}\right)^{n_k} \\ & + \omega_k + \frac{R_k}{\tau \dot{E}_{F,k}}. \end{aligned} \tag{23}$$

During each iteration of optimization parameters $\beta, \gamma_k, \tau, \omega_k$ and R_k remain constant. ε_k is the only variable in this equation.

The maximum cost per exergy unit of fuel is obtained by differentiating the above equation and setting the derivative to zero:

$$\frac{dc_{F,k}}{d\varepsilon_k} = 0. \quad (24)$$

Finally, Equation 25 is achieved, which has to be solved by numerical methods (contrary to direct exergoeconomic methods) for $\varepsilon_k^{\text{OPT}}$:

$$\frac{\gamma_k B_k n_k}{\tau \dot{E}_{F,k}^{1-m_k} c_{P,k}} = \frac{(1-\varepsilon_k^{\text{OPT}})^{n_k+1}}{(\varepsilon_k^{\text{OPT}})^{n_k-1}}. \quad (25)$$

Another useful and important variable for optimization is the relative cost difference:

$$r_k = \frac{c_{P,k} - c_{F,k}}{c_{F,k}}, \quad (26)$$

or:

$$r_k = \frac{c_{P,k}\dot{E}_{P,k} - c_{F,k}\dot{E}_{P,k}}{c_{F,k}\dot{E}_{P,k}}. \quad (27)$$

Equations 28 and 29 are used for eliminating $\dot{E}_{P,k}$:

$$\dot{E}_{P,k} = \dot{E}_{F,k} + \dot{E}_{D,k} + \dot{E}_{L,k}, \quad (28)$$

$$c_{P,k}\dot{E}_{P,k} = c_F \dot{E}_{F,k} + \dot{Z}_k. \quad (29)$$

So:

$$c_{P,k}\dot{E}_{F,k} = c_{F,k}\dot{E}_{F,k} + c_{P,k}\dot{E}_{L,k} + c_{P,k}\dot{E}_{D,k} + \dot{Z}. \quad (30)$$

Finally, the following relation is achieved:

$$r_k^{\text{OPT}} = \frac{1}{\frac{c_{P,k}\dot{E}_{F,k}}{c_{P,k}(\dot{E}_{D,k}+\dot{E}_{L,k})+\dot{Z}^{\text{OPT}}} - 1}. \quad (31)$$

Tsatsaronis' priority for optimization of components in the iterative exergoeconomic optimization technique is $(\dot{Z}_k + C_{D,k})$ [13]. The optimization proposed in this study specifies its priority based on exergoeconomic and exergetic improvement potentials. Components with small exergoeconomic improvement potential, but acceptable exergetic improvement potential, have to be replaced. Components with high exergoeconomic and exergetic improvement potential are the best components for improvement and modification. Components which have very low exergoeconomic improvement potential are not worth any capital investment. Components with low exergoeconomic and exergetic improvement potential and high exergetic efficiency are considered optimum.

CASE STUDY

Mill Description

The paper mill under consideration is an Iranian pulp and paper mill with a capacity of 600 tons/day. This mill is a Chemical Thermal Mechanical Pulp (CTMP) integrated mill. The effluent that is to be evaporated is from the CMTP plant, amounting to 137.9 tons/h, with a Dry Solid (DS) content of 5%, which is evaporated to a final content of 63% DS before incineration in the recovery boiler.

Evaporation of the effluent means that the it is divided into two fractions, clean condensates and residue with a high content of dry substance, containing both organic and inorganic material. The residue can be combusted in the recovery boiler and condensates can be reused in the process, thereby increasing the efficiency of the mill. The recovery link is shown in Figure 3.

The following assumptions have been considered during a simulation of the recovery link:

- System operates under steady state conditions.
- Ideal-gas mixture principles are applied to air and combustion products.
- Combustion in the recovery boiler chamber and conversion tower is complete.
- Except for the recovery boiler, all components operate without heat loss.

Applying Pinch to Heat Exchanger Network

The main aim is to make a survey of the excess heat in the system, with unchanged live steam consumption, therefore some parts of the system are excluded. These parts are steam consumers such as paper machines, digesters and cold streams whose heat demands, due to process conditions, have to be met with steam. The heat supplied from incinerating the effluent residue is neglected because the heating value is poor. Table 1 summarizes the cold and hot streams chosen for the formation of cold and hot composite curves.

For streams containing fiber, water and/or liquor, c_p, is calculated by use of the following relations:

$$c_p = (0.005 + 1.092^* \left(\frac{T_{\text{in}} + T_{\text{out}}}{2}\right)$$
$$^*\text{consistency}/100 + (1 - \text{consistency})^* 4.178, \quad (32)$$

where consistency is defined as [29]:

$$\text{consistency} = \frac{\dot{m}_{\text{fiber}}}{\dot{m}_{\text{fiber}} + \dot{m}_{\text{water}}}. \quad (33)$$

After computing streams $T_{\text{in}}, T_{\text{out}}, c_p$ and $\dot{m}$, the hot and cold composite curves are drawn by specifying a

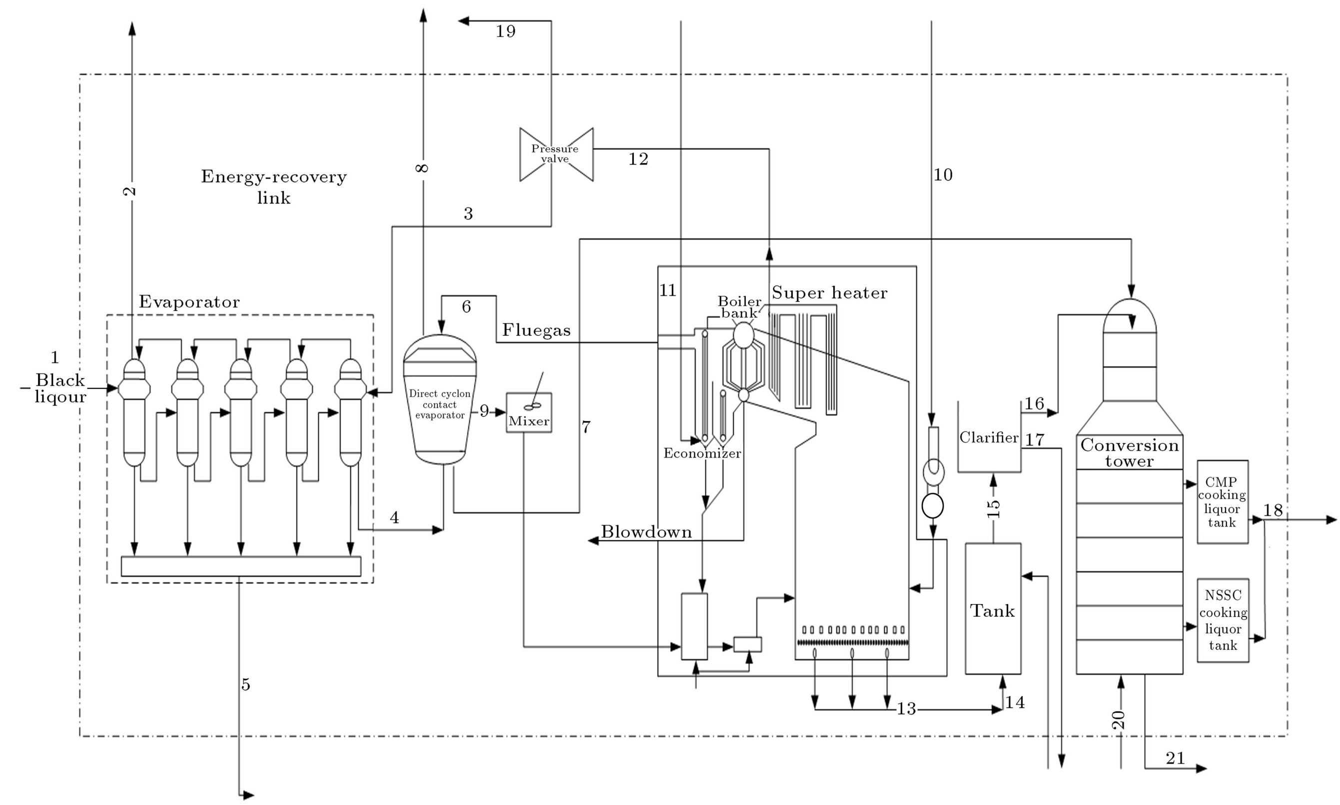

Figure 3. Mill's recovery link.

Table 1. Summary of cold and hot streams.

Cold Streams	Hot Streams
Inlet chips	
Black liquor	
White liquor	
Fresh water	Effluent
Process water	Blow steam
Evaporator inlet	Evaporator condensate
Boiler make up water	
Process pulp*	

* Only pulps whose temperature has to increase considerably.

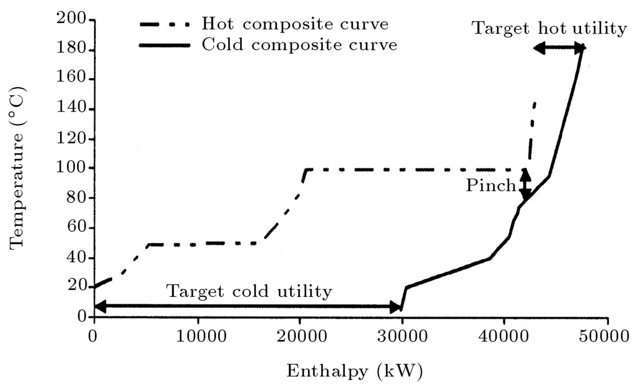

Figure 4. Mill's cold and hot composite curve.

minimum temperature driving force for heat transfer from hot to cold streams ($\Delta T_{\min}$) of 25°C, selected according to the process constraints and investment criteria of the mill. Figure 4 shows the Mazandaran mill's cold and hot composite curve.

APPLYING EXERGOECONOMIC METHOD TO THE MILL

An exergetic and exergoeconomic system analysis is performed to determine all mass, exergy, exergy destruction, exergetic cost, exergy destruction cost, component investment cost flow rates, component exergetic efficiencies and system total cost.

Selection of the independent variables for characterizing design options is very important. All important variables that affect the performance and cost effectiveness of the system must be included. Independent variables whose values are amenable have to be distinguished; only decision variables may be varied. Decision variables for cogeneration systems have been presented and introduced, but since the system under consideration is a process system, selection of its decision variables is not the same as energy systems. Choosing decision variables is difficult, especially because some variables affect the product cost without influencing the effectiveness of the system fine details, thus, variables of minor importance should not be included. Compounded methodology is used for the identification of decision variables that affect the cost and exergetic efficiency of the system. This methodology combines recent available exergoeconomic techniques with new qualitative and quantitative criteria [30].

Table 2 shows decision variables and their limits for the system under consideration.

OPTIMIZATION

Tsatsaronis has proposed an iterative optimization procedure for complex systems, consisting of the following steps [13]:

- For components, particularly the ones having a relatively high value of the sum $(\dot{Z}_k + \dot{C}_{D,k})$, $\Delta\varepsilon_k$ and Δr_k, the calculation for one decision variable changes, while other decision variables are kept constant, where:

$$\Delta\varepsilon_k = \frac{\varepsilon_k - \varepsilon_k^{\mathrm{OPT}}}{\varepsilon_k^{\mathrm{OPT}}} \times 100, \tag{34}$$

$$\Delta r_k = \frac{r_k - r_k^{\mathrm{OPT}}}{r_k^{\mathrm{OPT}}} \times 100. \tag{35}$$

- The effects of the decision variables are examined on the objective function. If, in comparison with the previous design, this value has been improved, it may be decided to proceed with another iteration. However, if the value of the objective function is not better in the new design than in the previous one, some design changes may either be revised or a previous step be repeated.
- Repetition of the above steps for the other decision variables.

Among design guidelines, some are very useful when considering a recovery link. These are:

- Maximizing the use of cogeneration of power and process steam, this can be interpreted as the use of a turbine instead of a pressure valve.
- Minimizing the mixing of streams with different temperatures, DCCE (Direct Cyclone Contact Evaporation) can be replaced by an indirect heat exchanger.
- Minimizing the use of combustion or preheating the reactants and minimizing the use of excess air.

The use of these guidelines can reduce the total number of iterations required.

In this study, exergoeconomic and exergetic improvement potential are used instead of $(\dot{Z}_k + \dot{C}_{D,K})$, specifying the order of component modification. Figures 5 and 6, respectively, show the exergetic and exergoeconomic improvement potential of the components and exergetic efficiency at initial decision variable values.

These figures give ideas about conditions when encountering different components. For example, the evaporator is optimum, however decision variable changes can improve its function. DCCE and pressure valve replacement is proposed, while the recovery boiler and conversion tower can be modified. It should be noted that the conversion tower low efficiency and

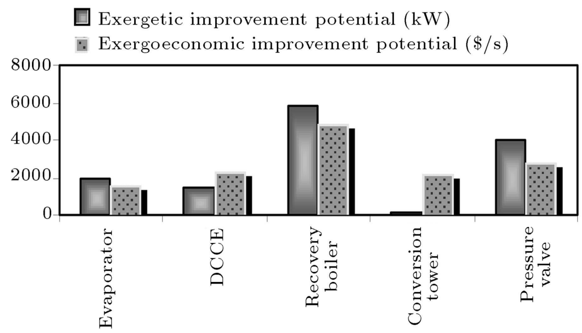

Figure 5. Exergetic and exergoeconomic improvement potential of components at initial decision variable values.

Table 2. Decision variables of all components and their limit.

Component	Input Variable	Value	
		Minimum	Maximum
Evaporator	Number of effect	4	6
	Backward-forward central		
	$T_{\mathrm{steam,in}}$(°C)	90	400
	$\dot{m}_{\mathrm{steam,in}}$ (kg/s)	3	4.5
Recovery boiler	AFR	1	1.6
	T_{fluegas} (°C)	250	1200
	$T_{\mathrm{air,in}}$ (°C)	25	200
	$T_{\mathrm{water,in}}$ (°C)	25	200

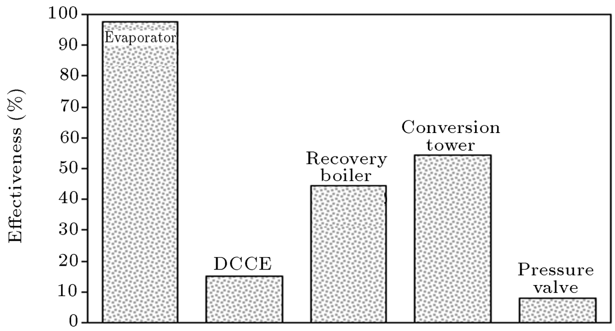

Figure 6. Exergetic efficiency of components at initial decision variable values.

exergetic and exergoeconomic improvement potential are related to the combustion nature.

After defining component priorities, based on exergetic and exergoeconomic improvement potentials, the relative deviations of the actual values from the cost optimal values for the exergetic efficiency and relative cost differences are calculated [13]. The design is modified to reduce the values of $\Delta\varepsilon_k$ and Δr_k. If conflicting design changes are suggested from the evaluation of different components, the design changes with higher exergoeconomic improvement potential values prevail.

Subsequent figures show the influence of some decision variables on different components, $\Delta\varepsilon_k$ and Δr_k. For better perception, absolute values of $\Delta\varepsilon_k$ are used in theses figures. It should be noted that some variables have a great influence on some components like $\Delta\varepsilon_k$ and/or Δr_k, but little or even no influence on other components. Figure 7 shows an impression of the recovery boiler's flue gas temperature on the recovery boiler, DCCE and conversion tower, $\Delta\varepsilon_k$.

Since a pressure valve performance is not dependent upon decision variables under consideration, it is not included in these figures.

Since DCCE's delta exergetic efficiency is noticeably higher than other components, its replacement is suggested. Besides, an evaporator has the highest exergetic efficiency and the lowest improvement potential. Therefore, the selection of decision variable values is based on the recovery boiler and conversion tower, $\Delta\varepsilon_k$ and Δr_k.

As seen in Figures 7 to 13, the recovery boiler and conversion tower optimum decision variable values are close, leading to an easier selection of decision variables. Figures 12 and 13 show the relative cost difference versus decision variables in some components. The decision variables, influence on DCCE and the recovery boiler's $\Delta\varepsilon_k$ is also shown in Figures 7 to 11.

It is deduced from Figure 7 that probably the best flue gas temperature is at the point where the recovery boiler curve cuts the conversion tower curve. Based on professional research [31], the best flue gas outlet temperature for recovery boilers in pulp and paper mills is about the same. It should be noted that delta exergetic efficiency and the delta relative

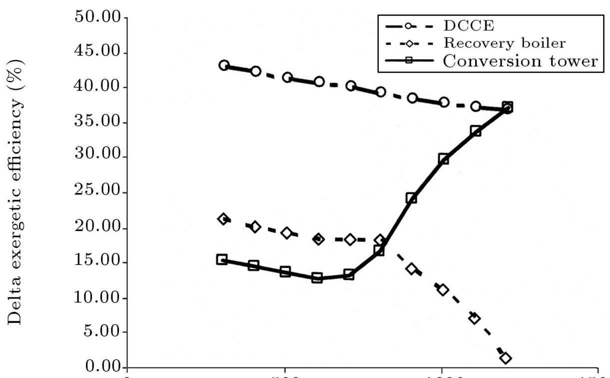

Figure 7. Recovery boiler's flue gas temperature influence on recovery boiler, DCCE and conversion tower, $\Delta\varepsilon_k$.

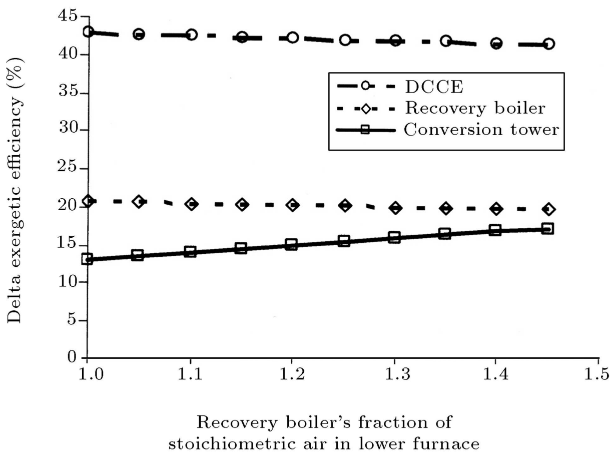

Figure 8. Recovery boiler's fraction of stoichiometric air in the lower influence on recovery boiler, DCCE and conversion tower, $\Delta\varepsilon_k$.

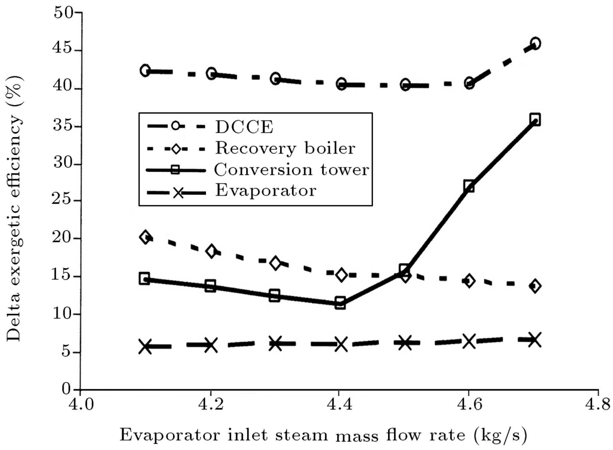

Figure 9. Evaporator's inlet steam mass flow rate influence on recovery boiler, DCCE and conversion tower, $\Delta\varepsilon_k$.

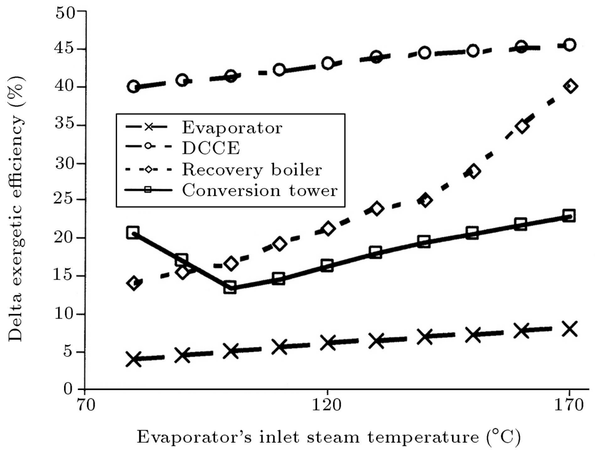

Figure 10. Evaporator's inlet steam temperature influence on recovery boiler, DCCE and conversion tower, $\Delta\varepsilon_k$.

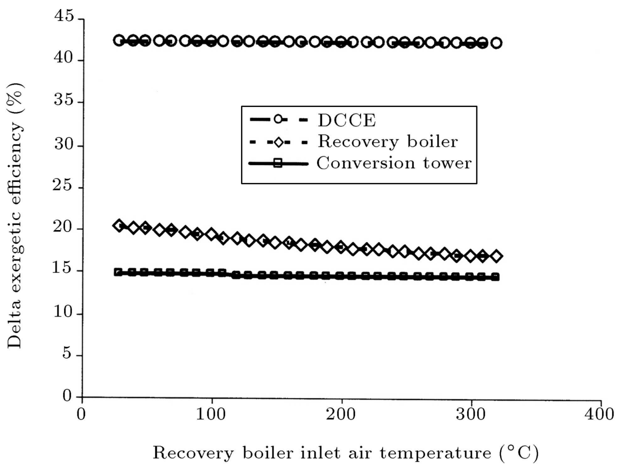

Figure 11. Recovery boiler's inlet air temperature influence on recovery boiler, DCCE and conversion tower, $\Delta\varepsilon_k$.

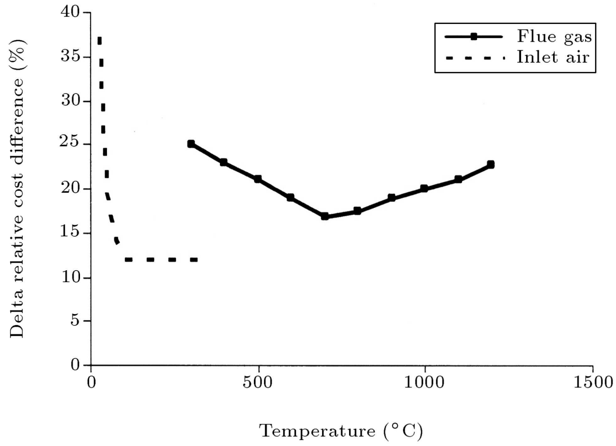

Figure 12. Recovery boiler's flue gas temperature influence on DCCE's delta relative cost difference and recovery boiler's inlet air temperature influence on recovery boiler's delta relative cost difference.

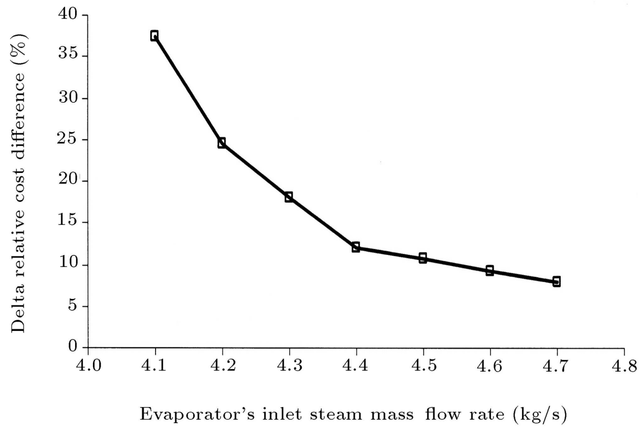

Figure 13. Influence of evaporator's inlet steam mass flow rate on recovery boiler delta relative cost difference.

cost difference minimum point is not necessarily the optimum point. However, Figure 8 shows very small change in the recovery boiler, DCCE and conversion tower delta exergetic efficiency. The proposed value for recovery boilers' LFA is about 1.2 to 1.4 [31]. This figure indicates that optimization iteration is in very good agreement with professional research.

Figure 9 shows that the best value is proportional to the black liquor mass flow rate. The results in Figure 10 are very acceptable, since the best evaporator inlet temperature is 20 to 30 degrees above the inlet stream, 90-100°C.

A slight change in the recovery boiler, DCCE and conversion tower delta exergetic efficiency versus the recovery boiler inlet air temperature is shown in Figure 11. However, from a graphical optimization method (which it is not discussed in this paper) it can be deduced that the best value for the inlet temperature is about 200°C.

This contradiction is due to the heat exchanger network and the impression of the inlet stream cost on the objective function. This result coincides with the consequences in Figure 12.

After choosing the best value in each iteration, the objective function is calculated. If, in comparison with the previous iteration, this value has been improved, it will proceed to the next iteration, otherwise changes should be revised and the iteration repeated. As explained before, the objective value is the value of C_{fuel} for the recovery link for which the black liquor cost has to be maximized.

In the next step, Pinch technology is applied to the heat transfer network, and the design is optimum when the annual levelized costs associated with the network are minimum and the recovery link objective values are maximum. With 3 to 4 iterations, the design optimum is achieved.

Figure 14 shows the objective value in some iterations.

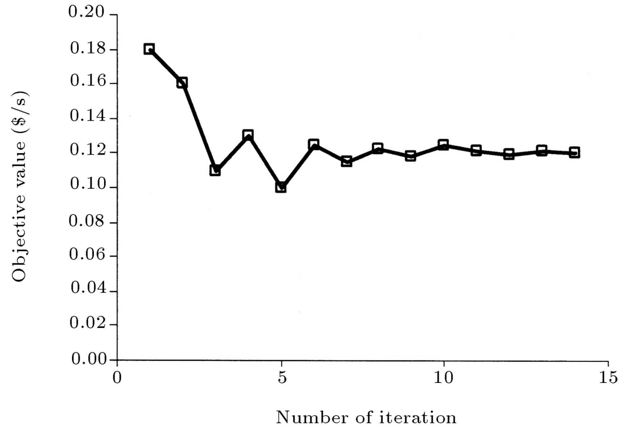

Figure 14. Variation of objective values with number of iterations.

CONCULUSION

The present study shows that the existing design consumes 23% more energy than the gross root design. By retrofitting the existing design, about 15% reduction in the energy consumption is possible. However, optimization of the heat exchanger network and recovery link together will lead only to 12% reduction in energy consumption. From components, exergetic efficiency, exergetic improvement potential and exergoeconomic improvement potential figures, it is concluded that DCCE and the pressure valve should be replaced by more efficient components. The best values for other decision variables are as indicated in Table 3. These values are achieved after 14 iterations. By applying these values, the mass flow rate of recoverable black liquor increases by almost 7%.

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Table 3. Optimized decision variables.

Component	Input Variable	Value
Evaporator	Number of effect	5
	Backward-forward central	Central
	$T_{\text{steam,in}}$ (°C)	97
	$\dot{m}_{\text{steam,in}}$ (kg/s)	4.15
Recovery boiler	AFR	1.2
	T_{fluegas} (°C)	459
	$T_{\text{air,in}}$ (°C)	112
	$T_{\text{water,in}}$ (°C)	73

NOMENCLATURE

B	constant
c	exergy specific cost (US $/kW)
C	cost of a stream (US $)
c_p	heat capacity (kJ/kgK)
D	exergy destruction (kW)
e	specific exergy (kW/kg)
E	exergy (kW)
EFL	effluent exergy losses (kW)
Irr	irreversible exergy losses (kW)
m	constant
$\dot{m}$	mass flow rate (kg/s)
n	constant
Pot	exergy improvement potential (kW)
Potec	exergoeconomic improvement potential ($/s)
$Q_{\min}$	minimum hot and cold utility requirement (kW)
r	relative cost difference
R	remaining operating and maintenance costs ($)
T	temperature (K) or (C)
TCI	total capital cost ($)
Z	equipment cost (US $/s)

Greek Symbols

γ	coefficient
ω	constant
τ	average annual operating time of plant (s)
ε	exergetic efficiency

Subscripts

1	pinch analysis without stream from recovery link
2	pinch analysis with stream from recovery link
D	recovery driving, destructed
E	recovered export
efl	effluent exergy losses
F	fuel
i	exergy form
irr	irreversible exergy losses
j	unit number
J	rejected
k	component and number of iterations
L	losses

ntp	net produced
nts	net supplied
O	recoverable
P	product
R	recycled
tte	total input
U	unit

Superscripts

CI	capital investment
OM	operating and maintenance
OPT	optimum

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