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Design of Evaporation Systems and Heaters Networks in Sugar Cane Factories Using a Thermoeconomic Optimization Procedure*

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Abstract

Sugar cane production in Brazil is one of the most competitive segments of the national economy, producing sugar and ethanol for internal and external markets. Sugar production is done basically in several steps: juice extraction, juice clarification and evaporation, syrup treatment and sugar boiling, crystallization, centrifugation and drying. Much heat exchange equipment is used in this process.. An optimized design of the evaporation system with the correct distribution of the vapor bleed to attend other parts of the process may contribute to exhausted steam demand reduction. This paper presents a thermoeconomic optimization of the evaporation system and the heaters network of a sugar factory, aiming at minimum investments and operation costs. Data from Brazilian sugar factories were used to define the process parameters. The methodology proposed is used to evaluate the cost of the steam consumed by the factory and the optimized design of the equipment.

Keywords: Sugar cane, sugar process, thermoeconomic optimization, heat recovery, process integration.

1. Introduction

Sugar and ethanol production from sugar cane in Brazil is one of the most competitive sectors of the national economy. The bagasse generated by the productive process is used as fuel in cogeneration systems that offer thermal and electrical energy to the process. In the last few years many sugar cane factories have been producing a surplus of electricity that may be sold for the grid, becoming a new product.

Currently there are more than 300 cane factories operating all around the country (UNICA, 2006). These units crushed more than 394.4 MT of cane in the 2005/2006 harvest season, with a total production of more than 26.7 MT of sugar and 17.0 Mm^3 of ethanol (CONAB, 2006).

Sugar production is basically done in several steps: juice extraction, clarification and evaporation, followed by syrup treatment and sugar boiling, crystallization, centrifugation and drying. Heat requirements of the process occur mainly in the evaporation system and the sugar boiling step, but heaters of the extraction system, juice and syrup treatments also consume important amounts of heat.

The reduction of exhausted steam demand by the production process may increase the surplus of electricity generated by the cogeneration system, but its feasibility must be evaluated considering investment costs. From an economic point of view, the prices paid for the surplus of electricity generated and the investments necessary for new heat exchange equipment determine the feasibility of the exhausted steam demand reduction. A thermoeconomic optimization can indicate the most adequate investment.

The purpose of this paper is to perform a thermoeconomic optimization of the evaporation system and heaters network design, analyzing for the optimized distribution of the vapor bleed and the heat transfer area necessary for each piece of equipment.

Data of sugar process parameters used for the process simulation were obtained from sugar cane factories in Brazil.

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2. Sugar process description

The sugar production from sugar cane is basically done by the following steps depicted in *Figure 1*:

- *Juice Extraction System* (I): sugar cane bagasse and juice are separated. Traditional sugar factories use mills where juice is extracted by compression. Diffusers can also be used, extracting raw juice by a process of lixiviation, using for that imbibitions hot water and re-circulation of the juice extracted. Both systems require previous cane preparation, done using knives and shredders that operate with direct drive steam turbines. The plant studied in this paper operates with a diffuser that demands thermal energy for the heating of the re-circulating juice.

The sugar cane bagasse produced at the extraction is delivered to the cogeneration system, where it is used as fuel, producing the electricity and steam consumed by the process.

- *Juice Clarification* (II): some non-sugar impurities are separated by the addition of some chemical reactants as sulfur, lime, among others. Juice heating is necessary for the purification reactions. After heated, the juice passes through a flash tank, before entering the clarifier.
- *Juice Evaporation* (III): juice is concentrated in a multiple-effect evaporator. Exhausted steam from the cogeneration system is used as thermal energy source in the first evaporation effect, which separates an amount of the water presented in the juice, and so producing, the heating steam for the next evaporation effect. The system works with decreasing pressure due to a vacuum imposed in the last effect, to produce the difference of temperature between each effect. The vapor generated in each effect may be used to attend other heat requirements of the process.
- *Syrup Treatment* (IV): syrup (concentrated juice) from the evaporation system is purified to improve the quality of the final product. Firstly the syrup is heated and then a flotation of the impurities is done with addition of some chemical reactants.
- *Sugar Boiling, Crystallization and Centrifugal Separation* (V): syrup is boiled in vacuum pans for crystal formation and then directed to crystallizers to complete crystal enlargement. After that, the sugar crystals formed are separated from molasses in centrifugals.
- *Sugar Drying* (VI): sugar is dried to reduce its moisture content in order to be stored.
- *Cogeneration System* (VII): Steam and electricity are produced to attend process demand. Usually sugar cane bagasse is used as fuel for boilers that produce high pressure steam to move back-pressure or extraction-condensation turbines in a steam cycle.

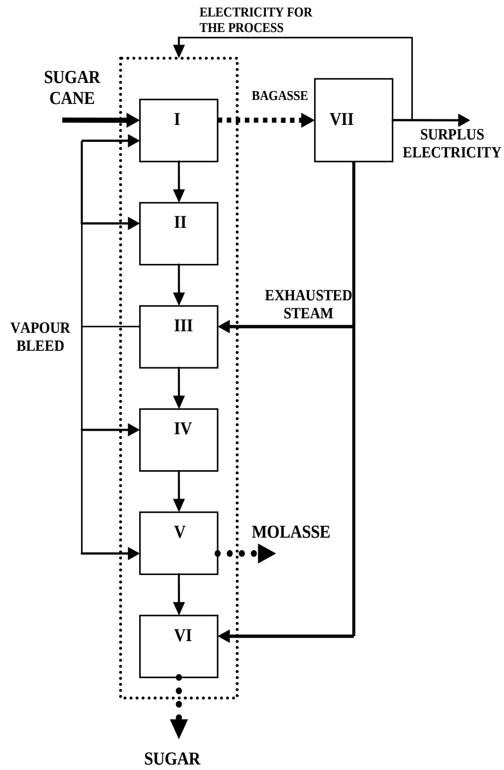

Figure 1. Scheme of a sugar factory with the cogeneration system.

3. Thermoeconomic optimization procedure

The thermoeconomic optimization procedure was performed using the EES software (EES, 2006), aiming at the optimum design of the evaporation system and the heaters network with a minimum total cost including operation costs (heating steam cost) and investments costs (equipment). The final cost results were based in a sugar factory that crushes 10,000 t cane/year. The reference environment presented by Szargut (1988) was used for the determination of the exergy of sugar cane, bagasse, steam and condensates.

A base case was defined that represents the current design found in Brazilian sugar factories, and used for the comparison of the results obtained after the optimization procedure.

The optimization was performed, dividing the plant into sub-systems that could be optimized separately in an iterative procedure with satisfactory results (Lozano et al., 1996). Four sub-systems listed below were optimized:

A. **Extraction system**
B. **Juice clarification**
C. **Syrup treatment**
D. **Evaporation system**

For the heaters that were designed in sub-systems A, B and C, the decision variable was the juice/syrup outlet temperature in each heater ($t_{j,out}$), and for the evaporation system (D) the saturation temperature of the steam generated at the evaporation effects ($t_{w,sat}$).

Figure 2 shows the adopted steps of the optimization procedure.

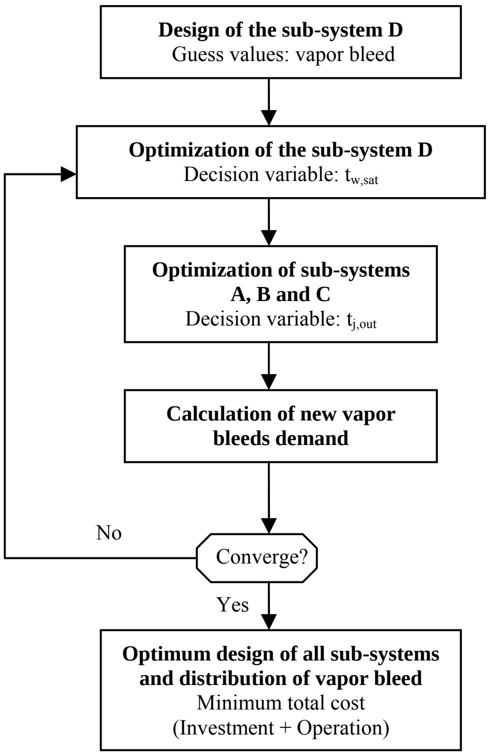

Figure 2. Iterative optimization procedure steps.

Other parameters listed below were calculated, defining an optimized design of the equipment with the minimum total cost.

For the heaters network the following parameters were defined:

- Number of heating stages;
- Heating requirements;
- Logarithmic mean temperature difference;
- Heat transfer area;
- Investment cost;
- Monetary cost of heating steam consumed.

And for the evaporator system the following parameters were defined:

- Operation pressure of evaporators;
- Juice boiling point elevation;
- Temperature of boiling juice;
- Intermediate juice concentration;
- Heat transfer area;
- Investment cost;
- Monetary cost of heating steam consumed;
- Monetary cost of vapor bleed and condensates produced.

The objective function, for the evaporation system and the heaters network, were defined by Equations 1 and 2 respectively.

Equation 1 minimizes the total cost of the evaporation system including investment cost of the heat transfer area in each evaporator (Z_e). The operation cost considers and the heating steam cost in the first effect of evaporation (C_s) and as products, which reduce the total cost of this sub-system, the vapor bleeds (C_v) and useful condensates costs (C_c).

For the heaters network, the objective function, defined by Equation 2, includes the investment cost of the heat transfer area in each heater and operation cost which considers the cost heating steam (C_s) and the useful condensate (C_c).

$$MinC_{evap} = \sum_e Z_e + \sum_s C_s - \sum_v C_v - \sum_c C_c \quad (1)$$

$$MinC_{he} = \sum_e Z_e + \sum_s C_s - \sum_c C_c \quad (2)$$

The total cost of the plant is defined by Equation 3, which considers the sum of minimum cost of all sub-systems obtained after the procedure of optimization.

$$C_{tot} = \sum_n C_n \quad (3)$$

3. 1. Economic model

3.1.1. Determination of the steam cost

The cost of each stream of steam demanded by the process was estimated using the theory of exergetic cost (Lozano and Valero, 1993).

Firstly, exergy of the bagasse and sugar cane was calculated. For the determination of bagasse exergy, a methodology presented by Sosa-Arnao and Nebra (2005) was adopted. The referred methodology is a variation of one proposed by Szargut et al. (1988) for wood, with the necessary changes in the composition and low heat value of the fuel. For the bagasse at the reference environment conditions, its total exergy is equal to its chemical exergy. The following composition of the bagasse in mass and dry base was assumed: C(47.0%), H(6.5%), O(44.0%) and Ash(2.5%) (Baloh and Wittner, 1990). The exergy of the sugar cane was obtained with the sum of the bagasse exergy and the juice exergy calculated following procedures for sucrose-water solutions presented by Nebra and Fernández-Parra (2005).

Thus, for a determined production cost of the sugar cane ready to be processed, the monetary cost per unit of exergy (c) of the sugar cane could be calculated as follows:

$$c_{cane} = \frac{C_{cane}}{m_{cane} ex_{cane}} \quad (4)$$

The "c" of the bagasse used as fuel at the cogeneration system was assumed to be the same as the sugar cane that enters the factory at the extraction system.

$$c_{cane} = c_{bag} \quad (5)$$

So, the live steam produced by the boiler at the cogeneration system had its "c" obtained using Equation 6.

$$c_s = \frac{(c_{bag} \dot{m}_{bag} ex_{bag}) + Z_{bl}}{\dot{m}_s (ex_s - ex_w)} \quad (6)$$

The exhausted steam from the back pressure steam turbine of the cogeneration and the vapor generated in the evaporators were considered alternatives of heating sources to the process. So, to perform the optimization, it was assumed as a hypothesis, that they have the same "c" of the live steam. The "c" of condensates generated after the steam condensation at the heat exchangers are also the same.

So, the monetary cost (C) of each steam stream could be calculated multiplying its "c" by its total exergy (Equation 7). TABLE I shows the parameters adopted for the determination of the steam monetary cost.

$$C_s = c_s \dot{m}_s ex_s \quad (7)$$

TABLE I. DATA FOR DETERMINATION OF STEAM COST.

Parameter	Value
Boiler capital cost (10^6US$)[1]	12
Sugarcane production cost (US$/t)[2]	14
Available bagasse (kg/t cane)[3]	252
Live steam pressure (bar)	63
Live steam temperature (ºC)	480
Boiler feed water temperature (ºC)	122
Boiler efficiency (%)[4]	85

[1] cost of boiler with following characteristics: 63 bar, 480ºC, 200 t of steam/h including costs of installation and instrumentation (Dedini, 2006).
[2] cost of sugar cane ready to be processed at the State of Sao Paulo, Brazil in 2006 (Usina Santa Isabel, 2006).
[3] wet base (50% of moisture)
[4] LHV base

3.1.2. Determination of investment costs

The investment cost of evaporators and heaters could be calculated using Equations 8 to 10. Scaling exponent is used to correct the reference equipment purchase cost for the optimized heat transfer area (Equation 9) (Bejan et al., 1996). Data used are shown in TABLE II.

$$\dot{Z}_e = E_e \xi \quad (8)$$

where:

$$E_e = E_r \left(\frac{A_e}{A_r} \right)^{\alpha} \quad (9)$$

$$\xi = \frac{\dfrac{i(1+i)^j}{(1+i)^j - 1}}{3600\tau} \quad (10)$$

TABLE II. DATA FOR DETERMINATION OF EQUIPMENT COST.

Parameter	Value
Evaporator purchase cost (10^3 US$)[1]	476
Heater purchase cost (10^3 US$)[2]	43
Evaporator scaling factor[3]	0.7
Heater scaling factor[3]	0.5
Annual interest rate (%)	10
Equipment useful life (years)	15
Factory operation hours per year	4000

[1] cost of installed evaporator Robert type with 4000m² of area (Usina Santa Isabel, 2006).
[2] cost of installed carbon steel shell and tube juice heater with 300m² of area (Usina Santa Isabel, 2006).
[3] Source: Chauvel et al., 2001.

A maximum heat transfer area was adopted for evaporators or heaters to represent a realistic design of the equipment. The maximum size of an evaporator was admitted as 5000m² and for heaters this limit was 1000m². If the optimization indicates that an equipment size is bigger than these values, the procedure divides the total area, showing some equipment in parallel which attends the limits imposed.

3.2. Physical model

3.2.1. Evaporation system

The developed evaporation system model uses Robert type five-effect evaporators, which operate with a vacuum at the last effect, producing the difference of temperature between each effect. Some restrictions are imposed for the optimization:

- Juice enters at 15% of solid content and leaves at 65%.
- Maximum temperature of 115ºC for juice boiling at the first effect to avoid juice sucrose loss and coloration (Baloh and Wittner, 1990).
- Minimal pressure of 0.16 bar at the last effect (Hugot, 1986).
- 5% of heat loss (Hugot, 1986).

A heat demand of sugar boiling system was estimated as 98 kWh/t cane, and it was assumed that it is attended by the vapor produced in the first effect of evaporation. Sugar drying heat was estimated as 5 kWh/t cane is provided by the exhausted steam from the cogeneration system.

The enthalpy of the juice was calculated using Equation 11 (Kadlec, 1981):

$$h_j = (4.1868 - 0.0297x_j + 4.6\mathrm{E}^{-5} x_j Pu_j)t_j + 3.75\mathrm{E}^{-5} x_j t_j^{\,2} \quad (11)$$

The temperature of evaporation (Equation 12) in each effect is defined as the sum of the temperature of saturation of pure water at the vapor space for a given operation pressure and the boiling point elevation due the concentration of the juice (Peacock (1995) (Equation 13). The effect of the boiling point elevation due the hydrostatic effect of liquid column was neglected.

$$t_{evap} = t_{w,sat} + \Delta t_{\,bpe} \quad (12)$$

$$\Delta t_{bpe} = 6{,}064\mathrm{E}^{-5}\left[\frac{(273 + t_{w,sat})^2 x_{j,out}^{\;2}}{(374{,}3 - t_{w,sat})^{0{,}38}}\right]\left(5{,}84\mathrm{E}^{-7}(x_{j,out} - 40)^2 + 0{,}00072\right) \quad (13)$$

3.2.2. Heaters network

The heaters network includes sub-systems A, B and C previously indicated.

Data of each sub-systems are presented in TABLE III. The purity of the heated flow is considered constant at 85% and 5% of heat loss is considered for each heater (Hugot, 1986).

TABLE III. DATA OF SUB-SYSTEMS A, B AND C.

	Sub-System	$\dot{m}_j$ (kg/s)	x_j (%)	$t_{j,in}$ (ºC)	$t_{j,out}$ (ºC)
A	Extraction System[1]	124.8	15	80.0	92.0
B	Juice Clarification	124.8	15	62.0	105.0
C	Syrup Treatment	28.8	65	Evaporator outlet	80.0

[1] this sub-system is composed by a diffuser with re-circulation of the raw juice in 3 heating stages (Usina Cruz Alta, 2005).

3.2.3. Heat transfer area

The heat transfer area defines the investment cost of evaporators and heaters, and can be calculated by Equation 14.

$$A = \frac{Q}{U\Delta t} \quad (14)$$

Equations 15 (Van der Poel et al., 1998) and 16 (Hugot, 1986) were used to calculate the heat exchange coefficients of evaporators and heaters respectively. The juice velocity circulation at the heaters was assumed constant at 1.5 m/s.

$$U_{evap} = \frac{465 t_{evap}}{x_{j,out}} \quad (15)$$

$$U_{he} = 6.978 t_{s,in}\left(\frac{v_j}{1.8}\right)^{0.8} \quad (16)$$

For the determination of the difference of temperature (Δt) in the evaporators and heaters, Equations 17 (Hugot, 1986) and 18 were used respectively.

$$\Delta t_{evap} = t_{s,in} - t_{evap} \quad (17)$$

$$\Delta t_{he} = \frac{(t_s - t_{j,in}) - (t_s - t_{j,out})}{\ln\left(\frac{t_s - t_{j,in}}{t_s - t_{j,out}}\right)} \quad (18)$$

3.3. Base case

A base case is assumed to compare and validate the results of the optimization procedure. The evaporation system is composed by a Robert type five-effect evaporator working at the following pressure in each effect: 1.69, 1.07, 0.76, 0.46, 0.16 bar of absolute pressure (Usina Cruz Alta, 2005)

As it occurs in many sugar factories in brazil, for this base case, there is not a thermal integration of the process and all the juice and syrup heaters of the factory consume vapor from the first effect of evaporation that attends the heat demand of sugar boiling system too. So, sub-systems a, b and c have only one heater as shown in *Figure 3*

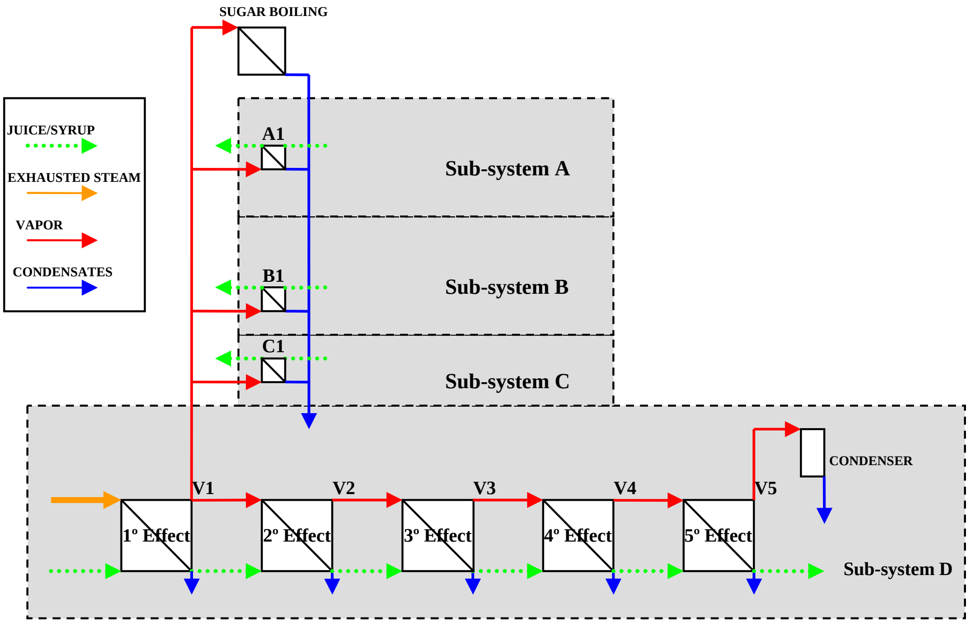

Figure 3. Lay-out of the base case.

4. Results

Using the optimization procedure described above, the optimum design of the evaporation system and heaters network of the sugar factory could be determined. TABLE IV shows the monetary costs per unit of exergy calculated for the sugar cane and live steam that were used for the calculation of the operation costs, once determines the costs of heating steam and condensates as previously explained.

TABLE IV. MONETARY COSTS PER UNIT OF EXERGY.

	Monetary cost per unit of exergy (10^{-6}US$/kJ)
Sugar Cane	2.788
Steam	10.864

Figure 4 shows the lay-out of the equipment optimized at the factory. In this figure the proposed sub-systems can be seen, including also the distribution of the vapor bleeding from the evaporator. The detailed parameters of each heater are shown in TABLE V.

TABLE V. OPTIMIZED HEATERS NETWORK.

	Area (m^2)	Heating Steam Consumed[1]	Heating Steam Flow Hate (kg/s)	$T_{j,in}$ (ºC)	$T_{j,out}$ (ºC)
A1	1000	V4	1.76	80.0	87.9
A2	1000	V4	1.76	80.0	87.9
A3	1000	V4	1.76	80.0	87.9
A4	265	V3	0.94	87.9	92.0
A5	265	V3	0.94	87.9	92.0
A6	265	V3	0.94	87.9	92.0
B1	1000	V4	5.92	62.0	88.5
B2	1000	V4	5.92	62.0	88.5
B3	994	V3	2.30	88.5	98.6
B4	749	V2	1.48	98.6	105.0
C1	162	V4	0.82	58.3	80.0

[1] "V" denotes vapor bleed and the "number", the evaporation effect it was produced.

As seen in TABLE V, after the thermoeconomic optimization a certain distribution of the vapor produced by the evaporation system was obtained. The new lay-out promotes a better use of each evaporation effect vapor bleed, contributing for a higher thermal integration of the factory, with a minimal cost

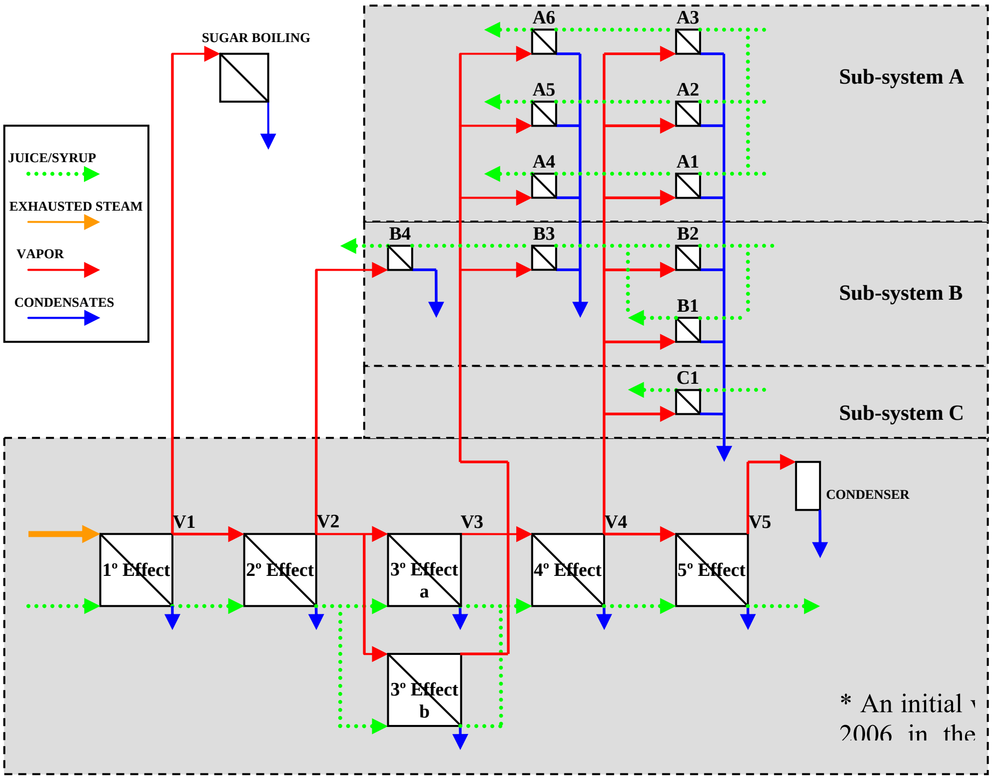

Figure 4. Lay-out of the optimized case.

The higher cost of the exhausted steam from the cogeneration system limits its use as a heating source for the heaters in an optimized design. The available vapor bleeds have adequate temperatures for the heating requirements, reducing the total cost of these sub-systems.

The consumption of exhausted steam from the cogeneration system decreased when compared with the base case (TABLE VI). As expected, the use of vapor bleeds from the last evaporation effects, substituting vapor bleed from the first one, promoted the reduction of steam requirements of the evaporation system, that could evaporate the same amount of water using 16% less energy.

TABLE VI. EXHAUSTED STEAM DEMAND.

	Exhausted Steam Demand[1] (kg/t cane)
Base Case	490
Optimized Case	412

[1]saturated at 2.5 bar of pressure

The evaporation system was designed to have the minimum cost at the optimized case, as its parameters are those presented in TABLE VII.

TABLE VII. OPTIMIZED EVAPORATION SYSTEM DESIGN.

Effect	Area (m^2)	p (bar)	$t_{w;sat}$ (ºC)	Δt_{bpe} (ºC)	t_{evap} (ºC)
1º	3324	1.69	115.0	0.5	115.5
2º	4808	1.38	109.0	0.7	109.7
3ºa	3615	1.12	102.9	1.3	104.2
3ºb	3615	1.12	102.9	1.3	104.2
4º	5000	0.75	91.9	3.1	95.0
5º	246	0.16	55.0	3.3	58.3

The final results for base and optimized cases in each sub-system can be seen in TABLES VIII and IX respectively.

TABLE VIII. COSTS FOR THE BASE CASE.

Sub-system	Investment (US$/h)	Operation[1] (US$/h)	Total (US$/h)
Extraction System	4.3	167.1	171.4
Juice Clarification	4.7	199.4	204.1
Syrup Treatment	0.5	17.0	17.5
Evaporator	62.9	334.3	397.1
TOTAL	72.3	717.7	790.1

[1] Operation cost considers the cost of heating steam and the reduction cost due the production of vapor and/or condensates that are used in other parts of the factory.

As can be seen in TABLES VIII and IX, after the thermoeconomic optimization was performed, a reduction of the total cost was obtained in each sub-system when compared with the base case. The investment in equipment increased, as heating steam of a lower quality is used at the heaters, requiring bigger surfaces of heat exchange. On the other hand, the reduction of consumed steam cost compensates its investment in new heaters and the total cost decreases for each sub-system.

The use of vapor from the 1^{st}, 2^{nd}, 3^{rd} and 4^{th} effect of evaporation reduced the total cost of all sub-systems considerably, even increasing more than 50% the investment in new equipment.

The clarification, contributes with 13.5% of the total cost reduction. The extraction system reduced 10.9% of the total cost substituting the use of V1 in the base case for V3 and V4 in the optimized one. Syrup treatment contributed with 1.7% of the total reduction using V4.

TABLE IX. COSTS FOR THE OPTIMIZED CASE.

Sub-system	Investment (US$/h)	Operation[1] (US$/h)	Total (US$/h)
Extraction System	13.7	138.0	151.7
Juice Clarification	12.5	167.2	179.7
Syrup Treatment	0.9	13.4	14.3
Evaporator	82.6	181.2	263.8
TOTAL	109.6	499.8	609. 5

[1] Operation cost considers the cost of heating steam and the reduction cost due the production of vapor and/or condensates that are used in other parts of the factory.

The higher reduction of cost was obtained at the evaporation system with 33.6% of savings in this sub-system. This represents 73.8% of the total cost reduction obtained, showing the importance of the optimized design of this equipment for the sugar process cost reduction.

5. Conclusions

The thermoeconomic optimization presented in this paper showed to be very useful in analyzing the cost generation when designing a heaters network and evaporation system of a sugar factory, aiming at minimum investment and operation costs, and so choosing an alternative for thermal integration of the factory.

The evaporation system represents the largest part of the total cost of the factory in thermal energy consumption. Its investment cost is substantially higher than the heaters, showing the importance of its optimized design. Moreover it produces the heating source for the other sub-systems, influencing their designs too.

The optimization of the thermal energy consumption in sugar factories can also be important for evaluating the cost of exhausted steam demand reduction. The decision of the best technology to be implemented in the cogeneration system depends on the quantity of steam consumed by the process. The analysis of both systems must be made together aiming at the best alternative for the factory as a whole.

Acknowledgements

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Nomenclature

A	Heat transfer area (m^2)
ex	Specific exergy (kJ/kg)
c	Monetary cost per unit of exergy (US$/kJ)
C	Monetary cost (US$/s)
E	Equipment purchase cost (US$)
h	Specific enthalpy (kJ/kg)
i	Annual interest rate (%)
j	Equipment useful life (years)
$\dot{m}$	Mass flow rate (kg/s)
p	Pressure (bar)
Pu	Purity (%)
Q	Heat power (kW)
t	Temperature (ºC)
U	Heat exchange coefficient (kW/m^2K)
x	Solid content (%)
Z	Equipment cost (US$/s)

Greek Letters

α	Scaling exponent
ν	Velocity (m/s)
ξ	Amortization factor (s^{-1})
τ	Operation Hours (hours/year)

Subscripts

bag	bagasse
bl	boiler
bpe	boiling point elevation
c	useful condensate
cane	sugar cane
e	equipment
evap	evaporator
he	heater
in	inlet flow
j	juice/syrup
out	outlet flow
r	reference equipment
s	heating steam
sat	saturation

tot total
v vapor bleed
w water

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