ECONOMIC EVALUATION OF REFRIGERATION ALTERNATIVES FOR ALCOHOLIC FERMENTATION

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Abstract. *The use of sugarcane bagasse as fuel in cogeneration systems combined with thermal integration in sugar and ethanol production processes generates a surplus of electricity which Brazilian plants sell to local electricity distribution system. Thermal integration is dependent on the process steam distribution system which provides the thermal and mechanical energy for units operations. Thus, changes in the energetic balance of a plant have a direct effect on the economical performance of an industrial unit. To improve the ethanol production an absorption refrigeration system has been proposed to keep the reactor temperature at an optimum level. The aim of this study is to develop an economic analysis, based on a complete process simulation, of different alternatives to refrigerate the fermentation reactor: an single effect absorption chiller, double effect absorption chiller and a mechanical compression cycle. Also a hybrid system combining two cycles in series was analyzed. The economic evaluation is performed analyzing three operation modes and shows that cooling fermentation reactor is always profitable, but the small differences showed, between the alternatives analyzed, suggest that other parameters must been analyzed besides the economics.*

Keywords. *Economic evaluation, process simulation, alcoholic fermentation, sugar and ethanol plant.*

1 INTRODUCTION

Sugarcane represents one of the most important activities in terms of the Brazilian economy. Currently, there are more than 350 plants in operation including sugar production plants, ethanol distilleries and integrated sugar and ethanol plants. With the use of the residual biomass as a fuel in cogeneration systems and thermal integration processes, electricity has became a new product, since sugarcane plants sell surplus electricity.

Thermal integration is dependent on the steam distribution system, which provides the thermal and mechanical energy necessary for the different units operations involved in the production process. Therefore any change in the energetic balance of the industrial plant will affect its economic performance.

Alcoholic fermentation is the critical process in ethanol production, where the ethanol production rate is directly dependent on the reactor temperature (Atala et al. (2001)). As the process is exothermic it needs to be cooled in order to keep the temperature at a propitious temperature for yeast metabolic action. Currently, this operation is carried out by heat exchangers where the refrigerant is water from cooling towers. However, during the harvest season this system is insufficient to remove all of the heat released by the fermentation, reducing the ethanol production.

An absorption chiller to refrigerate the fermentation reactor has been proposed by Andrade (1999), using residual heat from the process. In this study, an economic evaluation was developed of four alternatives to refrigerate the fermentation reactor: full refrigeration by an single effect absorption chiller, double effect absorption chiller and a mechanical compression cycle. Also a hybrid model combining absorption and mechanical compression, was analyzed.

The economic evaluation was carried out considering the configuration of a real plant located in São Paulo state. The simulation model developed by Cardemil (2009), configured with plant information, was used to estimate the availability of waste heat to supply the absorption chiller and also to evaluate the performance of the refrigeration alternatives.

The results of simulation show two heat sources available to supply an absorption chiller. So three operation modes for the refrigeration alternatives were studied at this publication: under actual configuration (Base Case), buying bagasse from other plants and using a condensing steam turbine (CEST) at the cogeneration system. Also, in order to improve accuracy of the economic evaluation four settings for the sugarcane processing at the plant were defined, considering the sugarcane composition and flow variations during the harvest season.

2 PRODUCTION PROCESS

In the Brazilian sugarcane industry most of the sugar and ethanol are produced in integrated plants. These plants are configured according to the flow chart shown in Fig. 1. The industrial plants use sugarcane juice to produce sugar while the ethanol production is usually carried out using a mixture of sugarcane juice and molasses, a by-product of

Nomenclature

A1E	Single Effect Absorption Chiller	NPV	Net Present Value, US$
A2E	Double Effect Absorption Chiller	N	Project life, years
C_o	Initial Investment	P_E	Electricity Price
C_{mref}	Marginal Cost of Refrigeration	V1	Steam from evaporation 1st effect
CEST	Condensing Extraction Steam Turbine	V2	Steam from evaporation 2nd effect
COP	Coefficient of Performance	VE	Exhaust Steam
$EBCR$	Eckstein's Benefit-Cost Ratio	Y_z	Net annual cash flow ,US$
HS	Hybrid System	ΔE	Reduction of electricity surplus
i	Effective Discount Rate, %	ΔQ_{ref}	Refrigeration effect increment
MC	Mechanical Compression System	τ_{pb}	Payback time, years

sugar production. However, the study described in this paper considers a plant that uses only molasses, according to the reference plant configuration.

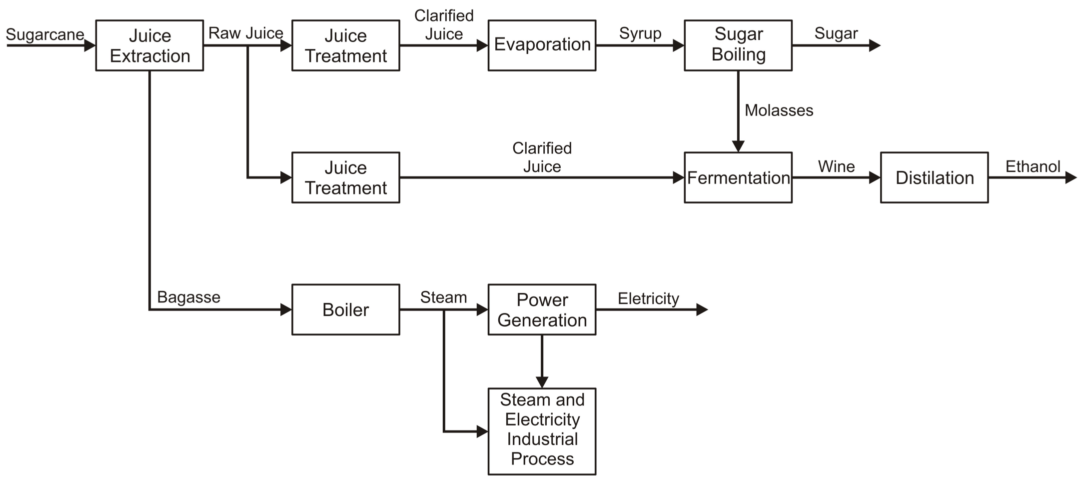


Figure 1: Scheme of sugar and ethanol production process

The basic process steps shown in Fig. 1 and the reference plant specific configuration are described as follows:

i Juice Extraction: Sugarcane is washed before entering the extraction system in order to remove excessive amounts of soil. Also, before extraction, sugarcane is broken apart into small pieces, by rotating knives and shredders, in order to be fed into the extraction system. Mills separate the juice from the bagasse by compressing the sugarcane, then the juice produced is pumped to the treatment system. Reference plant has a extraction system composed of five mills driven by an arrangement of steam and electric engines.

ii Juice Treatment: Impurities present in the sugarcane juice are removed by different treatment steps. Firstly the insoluble solids are removed by a sieve, then a chemical treatment is carry out by addition of substances such as sulphur and lime, among others. In order to complete the chemical reactions the juice is heated to 105^{o}C before passing through a flash tank, where non-condensable gases are eliminated. This chemical treatment is applied to coagulate and precipitate the soluble impurities, which are separated by sedimentation. Juice treatment for sugar and ethanol production are basically the same, differing only with regard to the sulphur addition, which is used only for refined sugar production. The juice treatment at plant consists only in lime addition and heating, since no refined sugar is produced.

iii Juice Evaporation: Juice for sugar production undergoes a concentration process by removing the water contained in it. The first stage of concentration is carried out in a multiple-effect evaporator. This system normally consists of four or five tanks, connected in series so that the juice undergoes progressive concentration in each tank. To make this possible exhaust steam is used as thermal energy source at the first tank, since evaporated water heats the juice in the subsequent tanks. The system works with decreasing pressure between the effects, due to vacuum conditions imposed in the last tank, which produces the necessary temperature difference between each effect. Before entering the evaporator the juice has a concentration of 14–16 °Brix, reaching 60–65 °Brix at the last effect, when it is called syrup. The steam generated during the first and second effects is usually bled to attend the heat requirements of other processes like the juice treatment heating system and sugar boiling system. At reference plant the multiple-effect evaporator consists of five effects, where steam is bled only from the first and second effects.

iv Sugar Boiling: The syrup is boiled in vacuum tanks for sucrose crystal formation and then the crystallization is completed in crystallizers increasing the crystal size. After the crystallizers, the cooked paste is centrifuged to obtain the raw sugar and the saturated solution is called molasses. After centrifugation, raw sugar passes to a dryer where the moisture content is reduced. The plant analyzed in this study use a two-boiling scheme.

v Fermentation: As previously mentioned most sugar and ethanol plants use a mixture of molasses and sugarcane juice to prepare the mash to be fermented. Thus, the process begins with the adjustment of the solid concentration of the mash to reach the optimum level, in order to facilitate the yeast action. The fermentation process is then carried out in fed-batch reactors for approximately eight hours. Fermented liquor, called wine, has around 8% ethanol concentration (mass basis). This wine is transferred to the storage tank from where it is pumped to the distillation system. The real system considered in this work consist of five reactors operating independently under fed-batch mode and using only molasses.

vi Distillation: The ethanol contained in the wine is recovered by distillation. Usually the distillation system used in sugar and ethanol plants consists of three distillation columns for hydrous ethanol production by stripping and rectification stages. To obtain anhydrous ethanol two more columns are required for the dehydration process. Reference plant distillation system. At reference plant only anhydrous ethanol is produced by a dehydration process consisting of azeotropic distillation by ethylene glycol. At this plant the by-product of distillation called vinasse is concentrated in vacuum evaporators, this procedure allows to reduce transport expenses to farm where is used as fertilizer.

vii Cogeneration: The bagasse produced in the juice extraction, which contains the fibers present in the sugarcane with approximately 50% of moisture content, is fed to a steam boiler where high pressure steam is produced. This steam serves as thermal and mechanical energy for different production processes, also part of this steam is conducted to a steam turbine where it generates the electricity required for the production process. Currently, with process thermal integration it is possible to generate a surplus of electricity and sell it to the local distribution system. Reference plant has two steam boilers and two backpressure turbines.

viii Steam distribution: For the processes described above the steam distribution system which supply thermal requirement for all unit operations is shown in Fig. 2. The V1 and V2 steams flows observed in Fig. 2 are the steam bled at the first and second evaporation effects, respectively. At this plant configuration all V3 steam is used at fourth evaporation effect.

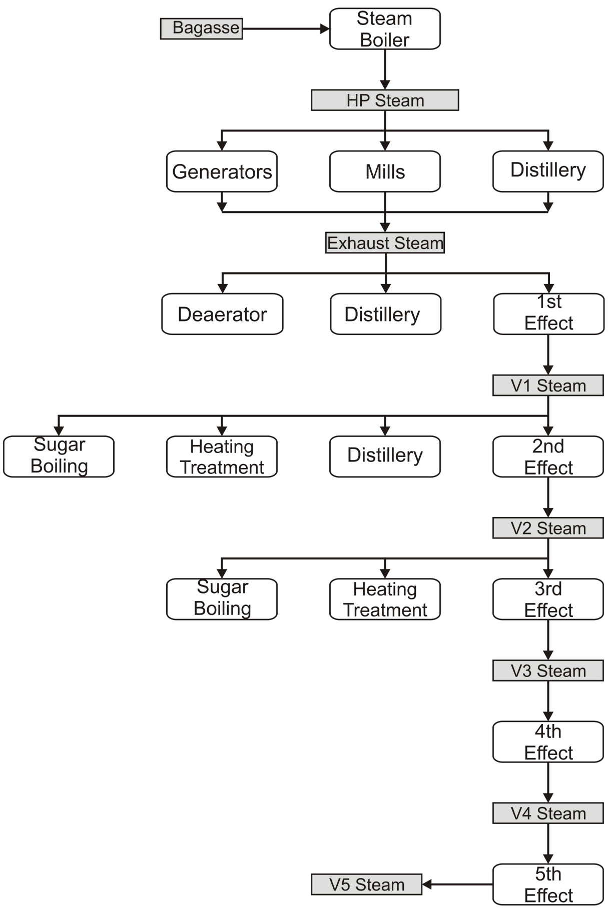


Figure 2: Standard steam distribution system

3 HEAT AVAILABILITY

In order to estimate the availability of heat sources to provide absorption cycles, a process simulation was developed according with the methodology presented by Cardemil (2009). The simulation was built in ASPEN Plus (Aspentech (2008)) software, using the NRTL (Non-Random Two Liquids) model and the NREL database (Wooley and Putsche (1996)) for physical properties calculation. The process simulation represents the reference plant characteristics described before and the specific parameters listed at Tab. 1.

Table 1: Parameters adopted for the process simulation

Process step	Parameter	Value
Juice Extraction	Nominal milling capacity	$600\,\mathrm{t/h}$
Juice Treatment	Heating temperature	$105\,°\mathrm{C}$
Evaporation	Syrup solid concentration	$65\,°\mathrm{Brix}$
Fermentation	Heat released at reaction (Williams (1982))	$678.262\,\mathrm{kJ/kg_{glu}}$
Distillation	Stripping column pressure	$149\,\mathrm{kPa}$
	Rectification column pressure	$140\,\mathrm{kPa}$
Boiler 1	Pressure	$5001.4\,\mathrm{kPa}$
	Temperature	$400\,°\mathrm{C}$
	Capacity	$200\,\mathrm{t/h}$
Boiler 2	Pressure	$2157.5\,\mathrm{kPa}$
	Temperature	$350\,°\mathrm{C}$
	Capacity	$150\,\mathrm{t/h}$
Cogeneration	Exhaust steam pressure	$147.1\,\mathrm{kPa}$
	Power of generator 1	$5\,\mathrm{MW}$
	Power of generator 2	$30\,\mathrm{MW}$
Average process energy consumption (Ensinas et al. (2007))		$12\,\mathrm{kWh/t_{cane}}$

The fermentation reaction was simulated as suggested by Pascal et al. (1995), as follows:

$$Sucrose + H_2O \rightarrow 2Glucose \qquad (1)$$

$$Glucose + 0.032H_2O \rightarrow 1.9EtOH + 1.9CO_2 + 0.075Glycose + 0.0105Succinic\ Acid + \Delta H \qquad (2)$$

where ΔH is the heat released per kmol of glucose fermented.

The harvest season lasts eight months and during this time some variations occur, for example the amount of sugarcane processed. Regarding these variations four production settings, with 60 days of duration, were considered. The first variation at the industrial plant is the flow of sugarcane processed. At the beginning of the harvest season this is approximately the nominal capacity of the plant, but at the end it decreases with the increasing rainfall. This study is based on a plant configuration where the nominal capacity for sugarcane processing is $600\,\mathrm{t/h}$. Thus for the first two production settings a flow of sugarcane equal to nominal capacity was defined, and for the next two settings the sugarcane flows were 573 and $512\,\mathrm{t/h}$ respectively.

Another variation observed at the industrial plant was the sugarcane composition, according to the cane varieties, maturation level and therefore. Thus, the four production settings defined before have the sugarcane compositions, in mass basis, shown in Tab. 2.

Table 2: Sugarcane composition for each production setting

	I	II	III	IV
Water	71.5 %	69.0 %	68.5 %	68.0 %
Sugars	13.0 %	15.5 %	15.0 %	13.0 %
Fiber	12.5 %	13.5 %	14.5 %	17.0 %
Impurities	3.0 %	2.0 %	2.0 %	2.0 %

In order to improve the accuracy of the simulation model, the non-conventional components (mainly impurities) present in sugar and ethanol production process were modeled regarding the parameters suggested by Wooley and Putsche (1996).

3.1 Cooling fermentation process

As previously mentioned there is a consensus in the literature regarding the benefits of cooling alcoholic fermentation. In an earlier theoretical analysis it was found that cooling fermentation reactors generates an increase of 5% in fermentation productivity. In order to perform a conservative analysis, a 2% increase was considered i.e. 0.4% of ethanol concentration in the wine (mass basis).

Currently fermentation reactor are cooled by cooling towers, but as was proved in field it is insufficient. Through the process simulation it is possible to estimate the refrigeration requirement of the fermentation reactors, at the four production settings considered. Regarding the maximum mean ambient temperature registered at plant location, the mean temperature of cooling tower outlet is calculated using the methodology presented by Braun et al. (1989), thus refrigeration deficit is estimated (Tab. 3).

Table 3: Refrigeration requirement for each production setting

Setting	Ambient Temperature		Temperature Outlet Cooling Tower		Heat Release	Refrigeration
	Mean	Max.	Mean	Max.	Fermentation	Requirement
I	21.3 °C	27.2 °C	30.4 °C	29.94 °C	4620 kW	3280.14 kW
II	19.9 °C	26.6 °C	30.2 °C	29.47 °C	4368 kW	2203.65 kW
III	23.4 °C	30.7 °C	30.3 °C	29.60 °C	4860 kW	2378.54 kW
IV	25.3 °C	39.8 °C	30.6 °C	30.13 °C	3804 kW	2973.18 kW

The simulation results also provide the waste heat available to supply an absorption chiller. In this study, mainly two heat sources were considered: condensate from the second and third effect steams and the unused exhaust steam. The distillation by-product vinasse is mentioned in literature as a possible heat source for an absorption cycle, but at reference plant it is concentrated in vacuum evaporators, so its outlet temperature is below 70 °C. Thus, in order to keep the same process configuration, others heat sources were not considered since they are all ready used in another operations. Hence, Tab. 4 shows the amount of condensate and exhaust steam available for each production setting.

Table 4: Heat sources availability

Setting	Condensate		Exhaust Steam		
	Mass Flow	Temperature	Mass Flow	Temperature	Pressure
I	275 t/h	95 °C	0–7.5 t/h	127 °C	248.42 kPa
II	272 t/h	95 °C	18.70 t/h	127 °C	248.42 kPa
III	248 t/h	95 °C	23.17 t/h	127 °C	248.42 kPa
IV	189 t/h	95 °C	44.13 t/h	127 °C	248.42 kPa

The low fiber content at the beginning of harvest season is the cause of the unavailability of exhaust steam at setting I, however this situation can be corrected buying bagasse from plants of the neighborhood. Therefore, a surplus of exhaust steam can be generated at setting I, this explain the mass flow range shown in Tab.4.

The following economic evaluation compares the installation of a single and double-effect lithium bromide absorption chiller and a mechanical (centrifugal) compression chiller, using R123 as refrigerant. In order to evaluate the performance of these refrigeration cycles under the reference plant conditions, simulations were carried out in ASPEN Plus to determine the coefficient of performance (COP) of each cycle. The results of these simulations are 0.81 and 1.13 for the single and double-effect absorption chiller, respectively. The mechanical chiller presents a COP of 6.09.

The hybrid system considered in the following analysis consists of two cycles in series, one single-effect absorption chiller which uses only low quality heat sources (condensate) and a mechanical compression chiller which complements the refrigeration effect defined in Tab.3. According to the COP calculations described above, the refrigeration capacity of the absorption chiller using only condensate is shown in Tab. 5. The remaining refrigeration effect for the hybrid case has to be produced by the mechanical compression system.

The use of the different refrigeration systems affects the energetic balance of the industrial plant. In the case where the fermentation reactors were refrigerated by a single-absorption system, the low quality waste heat available is not sufficient to generate the refrigeration capacity needed, so another heat source must be supplied. The exhaust steam can be used to provide the remaining heat, however, this source is not available in all the production settings, unless bagasse is bought. Furthermore if high pressure steam is supplied, the electricity surplus will decrease. The full mechanical compression cycle has the benefits of lower capital cost and higher COP than the absorption cycle, but it also decreases the quantity of electricity sold by the plant, because of the energy consumed by the compressor. Therefore, the hybrid system appears to be a good option, because it can combine the benefits of the other systems.

Table 5: Absorption chiller refrigeration capacity using only condensates

Setting	Refrigeration Effect		Refrigeration Deficit	
I	2511.9 kW	714.42 ton	768.24 kW	218.49 ton
II	2484.0 kW	706.48 ton	0.00 kW	0.00 ton
III	2265.3 kW	644.28 ton	113.24 kW	32.22 ton
IV	1726.2 kW	490.96 ton	1246.98 kW	354.66 ton

4 ECONOMIC ANALYSIS

The analysis methodology performed in this paper is based on an evaluation of project assessments as described by Helfert (2004). Through this procedure the project cash flow can be estimated in order to calculate the Net Present Value (NPV) of the investment and other economic indicator described in this section.

The economic parameters adopted in the further analysis are listed below:

- Project life time: 10 years.
- Harvest season: 240 days.
- Ethanol price: $400\,\mathrm{US\$/m^3}$ (UNICA (2009)).
- Electricity price: $66.7\,\mathrm{US\$/MWh}$, as the normative value for biomass generation (ANEEL (2009))
- Refrigeration system capital cost as listed in RSMeans.Co.Inc. (2009): $160\,\mathrm{US\$/kW}$ for single effect absorption chiller, $279\,\mathrm{US\$/kW}$ for double effect absorption chiller and $116\,\mathrm{US\$/kW}$ for mechanical compression chiller (centrifugal).
- CEST capital cost: $600\,\mathrm{US\$/kW}$ (Neto and Ramon (2002)).
- Depreciation: linear.
- Tax: 15% of profits.
- Discount rate: 15%, as the rate normally used at this type industrial projects.

Using the parameters described above, the cash flow of the project was constructed. Table 6 presents a summary of the annual cash flow including ethanol and electricity gains, maintenance expenses, depreciation and taxes. Therefore the exercise result (ER)by cooling fermentation reactors is calculated.

Table 6: Project annual cash flow

	Base Case				Buying Bagasse				Buying Bagasse - CEST			
Parameter	A1E	A2E	MC	HS	A1E	A2E	MC	HS	A1E	A2E	MC	HS
Production												
Eth. gain ($\mathrm{Mm^3}$)	4.19	4.19	4.19	4.19	4.19	4.19	4.19	4.19	4.19	4.19	4.19	4.19
Elec. gain (GWh)	(1.15)	(5.51)	(2.56)	(0.50)	3.06	3.06	0.56	2.61	6.12	5.69	3.77	5.83
Incomes												
Eth. (MMUS$)	1.67	1.67	1.67	1.67	1.67	1.67	1.67	1.67	1.67	1.67	1.67	1.67
Elec. (MMUS$)	(0.08)	(0.37)	(0.18)	(0.03)	0.20	0.20	0.04	0.17	0.41	0.38	0.25	0.39
Expenses												
Main. (MMUS$)	(0.03)	(0.05)	(0.02)	(0.04)	(0.03)	(0.05)	(0.02)	(0.04)	(0.04)	(0.05)	(0.03)	(0.05)
Others												
Dep. (MMUS$)	0.07	0.11	0.06	0.07	0.07	0.11	0.06	0.07	0.13	0.17	0.12	0.13
Tax (MMUS$)	(0.23)	(0.17)	(0.21)	(0.23)	(0.24)	(0.23)	(0.22)	(0.23)	(0.26)	(0.24)	(0.24)	(0.25)
ER (MMUS$)	1.28	0.97	1.21	1.31	1.40	1.39	1.27	1.38	1.59	1.55	1.46	1.57

4.1 Evaluation Parameters

In order to compare the economic performance of the different refrigeration alternatives three evaluation parameters were considered: Net Present Value, Benefit-Cost Ratio and Payback Period(Helfert (2004) and Bejan et al. (1996)).

4.1.1 Net Present Value

A rational criteria to evaluate projects must consider the time value of money during the life of the project. Hence, a widely used method is the Net Present Value (or Net Present Worth), defined by the following formula:

$$NPV = \sum_{z=0}^{N} Y_z(1+i)^{(-z)} \tag{3}$$

where N is the project life, Y_z is net cash flow at the end of zth time period and i is the effective discount rate.

4.1.2 Eckstein's Benefit-Cost Ratio

The Eckstein's benefit-cost ratio is used to evaluate the profitability of an investment, giving a criteria based on the net present value per dollar outlay. That is

$$EBCR = \frac{b_1 + b_2}{C_o + b_2} \tag{4}$$

where C_o is the initial investment and b_1 and b_2 are defined as follows:

$$b_1 = NPV + C_o \tag{5}$$

$$b_2 = A_{exp}\frac{(1+i)^n - 1}{i(1+i)^n} \tag{6}$$

where A_{exp} represents the constant annual cash expenses.

4.1.3 Payback Period

The payback period (τ_{pb}) is defined as the length of time required for the cash inflow recover the initial investment. Thus,

$$\sum_{z=0}^{\tau_{pb}} Y_z = 0 \tag{7}$$

4.2 Results

With the above definitions the NPV is estimated for the four alternatives studied, in each mode of operation. The results are observed in Fig 3.

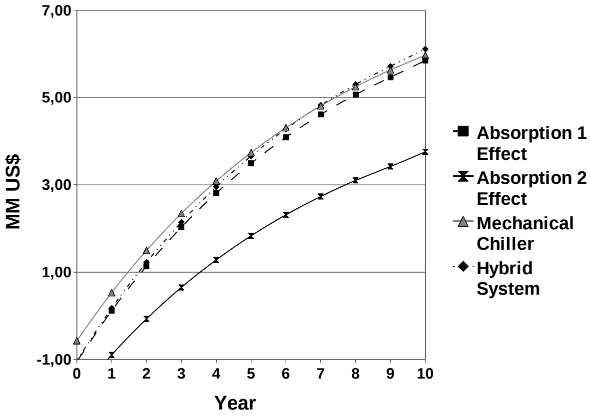
Table 7 shows a summary with the economic parameters calculated for the refrigeration alternatives under the three modes of operation.

Table 7: Economic indicators

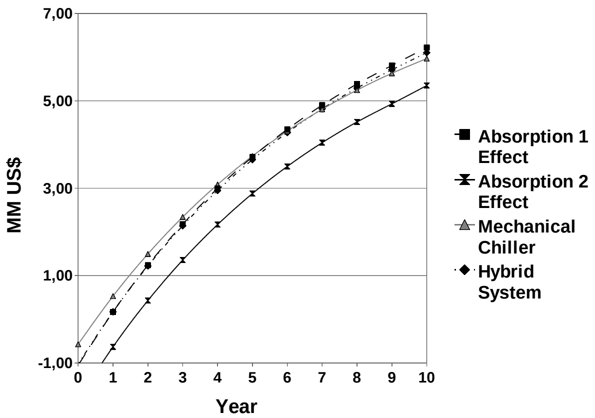
	Base Case				Buying Bagasse				Buying Bagasse - CEST			
Parameter	A1E	A2E	MC	HS	A1E	A2E	MC	HS	A1E	A2E	MC	HS
NPV	5.86	3.76	5.99	6.12	6.22	5.35	5.97	6.10	6.59	5.60	6.34	6.53
EBCR	5.79	2.76	9.74	6.04	3.55	2.59	4.13	3.51	3.16	5.41	3.53	3.15
τ_{pb}	0.89	1.70	0.49	0.85	0.86	1.52	0.52	0.85	1.20	1.80	0.93	1.19

Results in Tab. 7 show that Buying Bagasse - CEST operation mode is the most profitable, where absorption chiller present the highest NPV. This option also present interesting result in terms of Benefit-Cost Ratio, nevertheless the payback time grows due the higher initial investment.

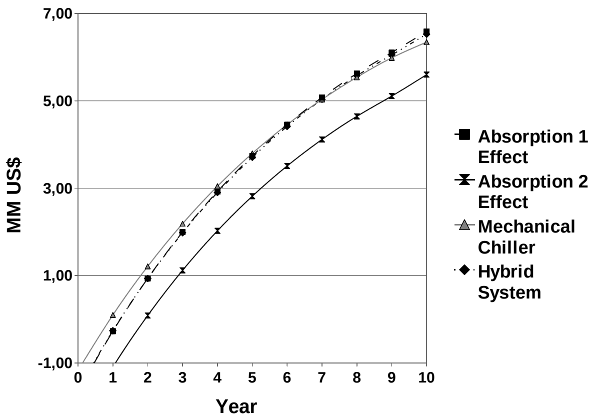
Also, a break-even point analysis was carried out in order to evaluate the level of fermentation productivity increase, due to refrigeration of the fermentation reactors, which merits the investment in a cooling system. Figure 4 shows this analysis, where the break-even point is a gain of 0.66% in the fermentation productivity.



(a) Base Case



(b) Buying Bagasse



(c) Buying Bagasse - CEST

Figure 3: Evolution of NPV for the base case

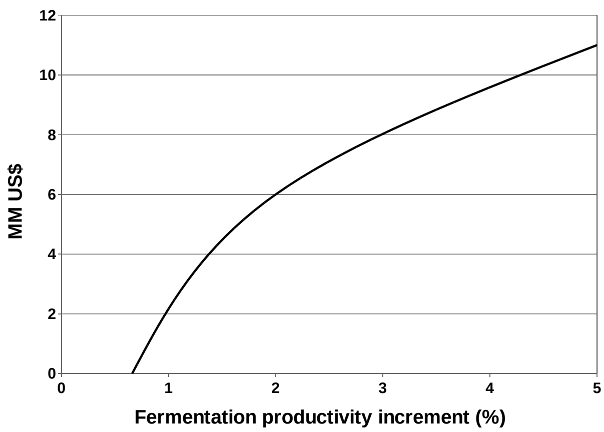
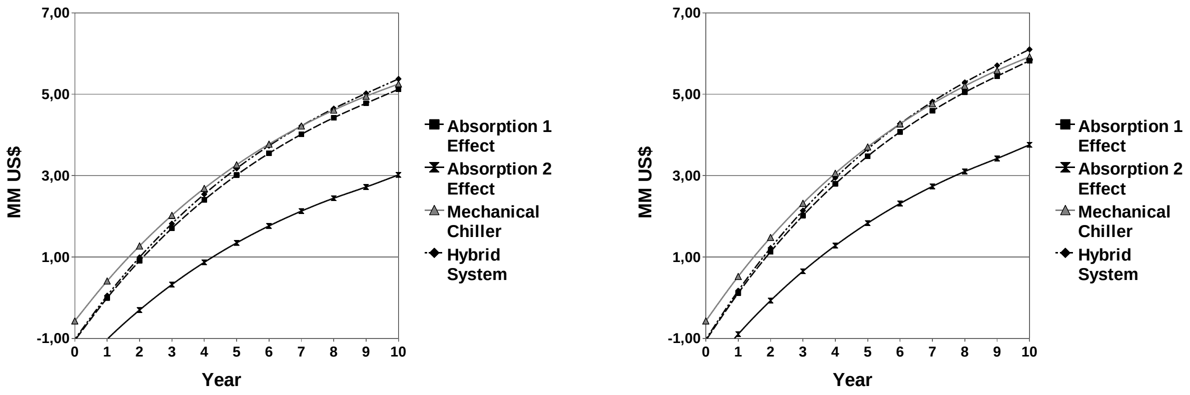


Figure 4: NPV of the project versus fermentation productivity increase

4.3 Sensitivity analysis

To evaluate how changes in the economical parameters will affect the NPV, a sensitivity analysis was performed. Figures 5(a) and 5(b) show the evolution of the NPV for the base case, considering a price of ethanol 10% lower and a price of electricity 10% higher. These prices represent a situation less favorable to ethanol market, attending to the common instabilities, and a possible sell of electricity in private auction.



(a) NPV for ethanol price 10% lower

(b) NPV for electricity price 10% higher

Figure 5: Sensitivity analysis

Figure 5 shows, as expected, higher sensitivity of NPV front ethanol price variations. Others sensitivity analysis were carried out to evaluate how the NPV changes with market variations, however, even when the value of the NPV increases or decreases, the best alternative in each mode of operation remains the same.

4.4 Marginal Cost

The economic indicators calculated in section 4.2, did not present significant differences between the alternatives analyzed. Hence, in order to complement the economic evaluation other parameters must be considered, as the system flexibility. Heat released during a feed-batch fermentation is not constant, therefore the system needs flexibility to manage variations on refrigeration demand. A first approach to evaluate this parameter is to define the marginal cost of producing an additional kW of refrigeration effect, in terms of electricity surplus reduction, as follows:

$$C_{mref} = \frac{\partial(E\ P_E)}{\partial Q_{ref}} \tag{8}$$

where E is the electricity surplus, P_E is the electricity price and Q_{ref} is the refrigeration effect increment.

Thus, regarding only the most profitable mode of operation (Buying Bagasse - CEST), the marginal cost for the refrigeration alternatives are calculated obtaining the values shown in Tab. 8.

Table 8: Marginal Cost of Refrigeration

Refrigeration Alternative	Marginal Cost
Absorption Single Effect	US\$ 4.7×10^{-3}
Absorption Double Effect	US\$ 3.3×10^{-3}
Mechanical Compression	US\$ 10.7×10^{-3}
Hybrid System	US\$ 10.7×10^{-3}

The small difference between the NPV of the hybrid model and the single effect absorption chiller contrast with the difference presented in terms of marginal cost of refrigeration. This situation is explained when the heat source supplied to the absorption chiller has latent heat to exchange. Otherwise generating electricity and providing it to a mechanical chiller will be preferable.

The low marginal cost presented by the absorption chiller strengthens it as investment option, attending refrigeration demand variations with low cost, therefore, high economic flexibility.

5 CONCLUSIONS

The analysis performed in this paper showed that the refrigeration of the fermentation reactors is always a profitable option, independent of the refrigeration alternative chosen. The break-even point analysis for fermentation refrigeration, based on the productivity increase, was calculated and found to be lower than the theoretical increase predicted in the literature. It explains why the option of refrigerated fermentation reactors is always profitable. The analysis performed for the reference plant configuration is consistent with result obtained by Andrade (1999) for different market conditions. In this study the break-even point was an increase of $0.62\,°\mathrm{GL}$ on wine ethanol concentration for a period of analysis of five years.

The small difference observed in the NPV for the systems analyzed shows that choosing one alternative above other must consider other parameters as system flexibility. The marginal cost of refrigeration were defined like an additional parameter to evaluate system performance. A comparison between the economic indicators of the refrigeration alternatives shows that the single effect absorption chiller and the hybrid system are the best investment alternatives, however the small differences presented suggest a detailed study for each particular plant.

Besides the refrigeration flexibility, the determination of overall plant's flexibility is researched in order to improve the system configuration to attend common changes in the volatile sugar and ethanol market.

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