ELECTRICITY-ETHANOL CO-PRODUCTION FROM SUGAR CANE: A TECHNICAL AND ECONOMIC ASSESSMENT

by

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"The process of making sugar is exceedingly interesting. First you heave your cane into the centrifugals and grind out the juice; then run it through the evaporating-pan to extract the fiber; then through the bone-filter to remove the alcohol; then through the clarifying-tanks to discharge the molasses; then through the granulating pipe to condense it; then through the vacuum-pan to extract the vacuum. It is now ready for market. I have jotted these particulars down from memory..."

Mark Twain
Life on the Mississippi

ABSTRACT

This thesis examines sugar cane based ethanol and electricity production, focusing on modern cogeneration technologies and of efficient steam use in autonomous ethanol distilleries and sugar factories. Three cogeneration technologies are considered: the condensing extraction steam turbine (CEST), the biomass integrated gasifier/ steam-injected gas turbine (BIG/STIG) and the biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG). CEST is the present state-of-the-art for sugar cane based cogeneration. Although not commercially availible at present, the BIG/(I)STIG cogneration systems offer particular promise, generating up to 20 times more power than is needed on-site.

When large amounts of excess electricity are cogenerated for sale, bagasse (the fibrous residue of cane milling) and perhaps barbojo (the cane tops and leaves) are regarded as valuable fuel resources. This valuation motivates the examination of steam economy in distilleries and sugar factories: steam which is consumed for process use cannot be used for power generation and sale. Additional motivation is provided by the electrically efficient BIG/(I)STIG cogeneration systems, which cannot meet the steam demands of sugar factories or distilleries without improvements in steam economy.

Numerous autonomous distillery designs are synthesized which reduce steam use by up to 70% over typical distilleries and by up to 50% over present "low steam use" distilleries. This is accomplished through the heat integration of the juice concentration section of the distillery and the distillation section of the distillery. (Heat integration is the arrangement of process equipment and heat exchangers to maximize waste heat use.) The pinch method of Linnhoff is used to determine the minimum steam use of a set of distillery equipment, and the mixed-integer linear programming technique of Papoulias and Grossmann is used to derive a heat exchanger network which achieves minimum steam use. The additional capital costs of the steam efficient systems are nearly always offset by their energy savings.

Within the pinch method, the concept of the "limiting process" is developed. By identifying which process in the system limits further steam savings, it provides the designer a valuble tool for integrating multiple processes into a single system.

Low steam use sugar factories are also derived using heat integration and employing steam efficient technologies found in other process industries. Through rigorous heat integration techniques, sugar factory steam use can be reduced 5%-15% below that achieved with informally heat-integrated designs.

An economic analysis of various co-production scenarios is also conducted by unbundling the cogeneration section from the distillery section, and, when appropriate, from the sugar factory section. The analysis shows that the use of BIG/(I)STIG cogeneration and treatment of bagasse (and perhaps barbojo) as valuable fuel resources strongly affects the economics of electricity-ethanol co-production. When bagasse is valued as a fuel resource and BIG/(I)STIG cogneration and low cost process equipment (such as is found in Brazil) are used, both electricity and ethanol (or sugar) can, in many circumstances, be produced competitively. One particularly attractive scenario is a facility which could produce either sugar or ethanol. Such a facility would be able to adapt its products to sugar and oil prices, and would be more profitable than either a sugar factory or an autonomous distillery.

Ethanol-electricity co-production is also shown to have the potential to make a large impact on energy supplies in many cane growing developing countries. Trends in cane production and gasoline consumption indicate that 25% of the gasoline demand in 2007 in cane growing developing countries could be met using ethanol (while still maintaining growth is sugar production), while almost half of the 1987 electricity demand in cane growing developing countries could be met using BIG/ISTIG cogenerated power.

ACKNOWLEDGEMENTS

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This thesis carries ‘1899-T’ in the records of the Department of Mechanical and Aerospace Engineering.

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Chapter 1: Introduction

1.0 Introduction

Biomass energy is already a significant global energy source, accounting annually for 48 EJ (10^{18} Joules) of energy in the developing world and 7 EJ in industrialized countries [1]. In developing countries, bioenergy comes in the form of fuelwood, crop residues and dung and accounts for approximately as much energy as from oil and coal combined [2]. It is, however, used very inefficiently, primarily by rural and urban poor for cooking, heating and light. For instance, energy use per capita for cooking using biomass fuels is three to five times higher than when modern fuel carriers (natural gas, propane, kerosene, LPG) are used [3]. Biomass energy could play a much more significant role in developing countries, if it is produced renewably and elevated to the level of a modern energy carrier by converting it into electricity and gaseous, liquid and processed solid fuels [4].

Sugar cane is a particularly attractive crop for bioenergy applications. Its cultivation is well established in over 80 developing nations. In 1987, over 900 million tonnes (metric) were harvested in developing countries [5]. Sugar cane is also a photosynthetically efficient crop; the global average cane productivity is around 58 tonnes cane per hectare per year,[a] which is equivalent in energy content to woody biomass production of 38 dry tonnes per hectare per year [2]. For comparison, yields of only 10 dry tonnes per hectare per year have been achieved in the US short rotation intensive culture demonstrations [6].

This thesis examines the potential for the efficient use of sugar cane as an energy source in the developing world.

2.0 Sugar Cane as an Energy Source

For every tonne of cane which is processed, approximately 11 GJ of energy can be extracted from the sugar cane plant (Figure 1.1). About 70 liters per tonne cane (1.64 GJ/tc) can be fermented into ethanol from the sugar in the cane, representing only 14% of the total energy content. Another 0.33 GJ of methane can be generated per tonne cane via the anaerobic digestion of the liquid by-products of fermentation (stillage)[7], boosting the sugar-juice's energy contribution up to 17%. Transforming the sugars into ethanol and methane incurs ~.3 GJ energy loss per tonne cane (about half of which cannot be avoided due to thermodynamics).

At 9.8 GigaJoules (GJ) per 50% moist tonne, bagasse, the sawdust-like by-product of milled sugar cane, has an energy content of over 2.85 GJ/tc (25% of the overall energy content). Bagasse, is usually burned in boilers which provide enough

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[a] Tonnes of moist cane milled. This includes only the cane stalks and does not include the cane tops and leaves (barbojo). References to "tonnes cane" (tc) refer to the stalk portion of the plant which is milled for sugar extraction.

steam and electricity to operate most sugar factories. The 1987 developing world sugar cane crop generated approximately 270 million tonnes of 50% moist bagasse (~300 kg/tc), totalling 2.5 EJ or 5% of the developing world bioenergy.

Bagasse, however, is not used efficiently. It is often seen as a waste product of sugar production, whose disposal can quickly become an issue if alternative uses cannot be found.[b] Therefore, sugar factories have been traditionally designed to use energy rather inefficiently in order to consume the bagasse at the same rate at which it is produced. Even in developed countries, excess bagasse is seen as a nuisance rather than a resource [8].

Barbojo, cane tops and leaves (Figure 1.2), represents the largest energy fraction of sugar cane (55%). Presently, it is usually either burned off of the fields prior to harvesting or left on the fields after harvest. Roughly 660 kg of barbojo (50% moist, 9.8 MJ/kg) are generated per tonne cane milled, with an energy content of 6.47 GJ/tc. Barbojo recovery has been attempted on modest scales in the Dominican Republic and the Philippines and is being investigated in Florida, Puerto Rico [9] and Thailand [10].

3.0 Motivations

Very few developing countries outside the Middle East have significant oil or natural gas resources, and, with the exception of China, very few have coal resources either. Importing petroleum often comes at a very high price for developing countries. In some cases, over 50% of their hard currency income is spent on oil imports (Brazil and Jamaica in 1981, for instance)[3]. Combining projected oil price increases and continued growth in energy demand, the balance of payments

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[b] Among other problems, large piles of bagasse in warm tropical climates tend to spontaneously combust.

problem can only become worse, and, in many cases hamper further development.

Ethanol fermented from sugar cane juice or from molasses, a by-product of sugar production, can make a significant impact on gasoline use in some developing countries. Brazil ferments by far the largest amount of ethanol from cane, over 11,700 million liters in 1988 [11], enough to operate its fleet of 4.2 million ethanol burning cars. Other countries have also investigated ethanol production from cane for energy purposes, but none has attempted it on as large a scale (Table 1.1).

Since 1980, electricity demand in the developing world has been growing at 7% annually [12]. Many countries already face electricity demands larger than their supply capacity. Pakistan, for example, is facing power shortages of over 25%, and India, shortages over 10% [11]. Power sector expansion is also proving to be very costly. It is not uncommon for more than 20% of a developing country's private sector capital budget to be spent on the power sector; foreign borrowing for the power sector is sometimes 40% of a developing country's total foreign debt [11]. Finding new, indigenous, low cost sources of power is becoming critical for continuing development.

The cogeneration of electricity from sugar cane residues in sugar factories and distilleries can make a major contribution in reducing this electricity supply crisis. For instance, in 1985 on the islands of Kauai, Hawaii and Maui (Hawaii), condensing, extraction steam turbine (CEST) cogeneration systems burning bagasse produced over 1100 million kWh of electricity [13]. (Admittedly, Hawaii is not a developing country, but it illustrates what is already being done.) Sugar factories supplied 8.4% of Mauritius's power generation in 1981 (30.6 million kWh)[9]. Using advanced gas turbine cogeneration technology, over 800 billion kWh could be generated annually using bagasse and barbojo in cane producing developing countries (Chapter 5). For reference, in 1987 1,630 billion kWh of electricity were generated in those same countries from all sources.

Cogenerated power from sugar cane residues would be located where it is most needed: in the countryside. Over 80% of developing nations's population lives in rural areas, much without electricity [12]. For example, in Guatemala, Honduras and El Salvador, only 10% of the rural population has access to electricity [12]. The difficulty with electrifying the countryside is the dispersed population; extending the power grid to rural areas is very costly because of the low electricity use intensity. By cogenerating electricity at sugar factories or distilleries at smaller scales (<100 MW), much of the rural electrification needs in cane growing regions could be met.

Expanding sugar cane cultivation for electricity and ethanol production might also provide rural employment opportunities and could help mitigate the exodus of rural workers into the cities. The direct capital investment required for creating new jobs in cane energy is 10% to 25% of that required in mining or petrochemicals [14]. Overall, almost a half million jobs (full time equivalent) were created between 1980 and 1985 by the alcohol fuel program in Brazil [15].

The cane sugar industry has ample reason to want to pursue energy from cane. The international sugar[c] price[d] tends to be very volatile (Figure 1.3). During low swings in the sugar market, all but the most efficient producers lose money. Growth in the sugar market is predicted to be slow in the next decade, with sugar demand rising at 1.5% per year [16], instead of at the historical rate of 3% per year. Cane sugar also faces challenges from oversupplies of beet sugar in Europe

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[c] Throughout this thesis, "sugar" refers to raw sugar, the product of sugar factories. Raw sugar must be further refined before it reaches the familiar white table sugar.

[d] Seventy percent of the world's sugar is consumed in the country in which it was produced (significant amounts of the sugar consumed in Europe and the US come from sugar beets). Of the 30% remaining, 40% is controlled by various import-export agreements between developing countries and the US or the EEC. The remaining sugar is sold on the international sugar market.

and high fructose corn syrups and artificial sweeteners in the US [13,16]. While at this time the international sugar market price is high, historic market volatility and continued competition from other sweeteners will keep export markets tight, providing incentive for the cane sugar industry to investigate ways to broaden its range of products.

The common practice of burning barbojo off of the fields prior to harvesting

Table 1.1: Annual Fuel Ethanol Production from Sugar Cane (Varying Years)[a]

	Country	Millions liters of Ethanol/yr.
South America	Brazil[b]	11,700
	Argentina[c]	380
	Paraguay	26
	Columbia	38
Central America	Costa Rica	31
	El Salvador	15
	Jamaica[d]	15
Africa	Kenya[e,f]	18
	Malawi[e]	11
	Zimbabwe[e,f]	42
	Mali	2
Asia and Oceania	Thailand	203
	Philippines	10
	New Zealand	15

Notes for Table 1.1:

[a] [2].

[b] [11].

[c] 1984 ethanol production capacity installed "Worldwide Review of Biomass Based Ethanol Activities," Meridian Corporation, 1985. Reference for all countries not otherwise noted.

[d] A 180,000 liter/day distillery was installed in 1988, with a planned production of 15 million liter/yr. To date, only one million liters have been produced. M.G. Hylton, Jamaican Sugar Industry Research Institute, Bernard Lodge, Jamaica, private communications, 1990.

[e] "Electricity and Ethanol Options in Southern Africa," USAID Office of Energy Report, September, 1988.

[f] Capacity installed as of 1984. "Power Alcohol in Kenya and Zimbabwe- A Case Study of the Transfer of a Renewable Energy Technology," UN Trade and Development Board Report, GE.84-55979, 1984.

is also quickly becoming an environmental issue in Brazil and other cane growing Latin American countries. Finding alternative uses for barbojo, such as a cogeneration fuel, can create productive alternatives to the environmentally damaging practice of field burning.

4.0 Electricity-Ethanol Co-Production

Electricity-ethanol co-production expands upon the idea of energy as a by-product of the sugar cane industry to include electricity as a co-product with either sugar or ethanol. By including electricity as a major focus of production, a number of significant changes must take place in the way in which the sugar resources are viewed. The primary change is the valuation of the agricultural residues-- the bagasse and barbojo. When these residues are seen as a source of fuel for cogenerated power, their value increases and their efficient use becomes an issue.

This thesis addresses this issue in several ways. First is the inclusion of efficient cogeneration equipment in evaluating the power potential of co-producing electricity and ethanol (or sugar). Three alternative cogeneration technologies are used in this analysis: the condensing-extraction steam turbine (CEST), the biomass integrated gasifier/steam injected gas turbine (BIG/STIG), and the biomass integrated gasifier/intercooled steam injected gas turbine (BIG/ISTIG). All three offer significant improvements over the turbo-alternators found in present day facilities, which generate just enough electricity to meet on-site needs (12-20 kWh/tc). CEST cogeneration is the present state-of-the-art in sugar factory cogeneration, generating approximately 100 kWh of electricity per tonne cane during the milling season while meeting the steam demands of the sugar factory or distillery. Although not commercially available at present, BIG/(I)STIG is estimated to be able to produce from 250 to 300 kWh per tonne cane during the milling season [2].

When the cogeneration facilities are operated during the off season as stand alone power plants using barbojo or some other off season fuel, average electricity production per tonne cane milled can exceed 800 kWh per tonne cane milled [2].

The second way in which the efficient use of bagasse is addressed is the efficient use of process steam in raw sugar and ethanol production. The more process steam consumed by the distillery and/or sugar factory, the less steam is available to drive the turbine and produce electricity. Without electricity export and sales, there has been little interest in steam economy in sugar factories or distilleries. Moreover, the more efficient BIG/(I)STIG systems produce modest amounts of steam and must be matched to steam efficient factories. However, when steam is valued at the opportunity cost of the electricity it might have generated, reductions in process steam demand become important. In this thesis, low steam use distilleries and sugar factories are formulated, and the additional cost of the energy efficiency is balanced against the revenue gained when the conserved steam is used for power production.

Steam conservation in sugar factories and distilleries is accomplished through two methods. The first method is the application of steam saving technologies found in other process industries. The primary steam consuming steps in sugar and alcohol production are characterized and technologies are identified which can perform the same functions more efficiently (Chapter 2).

The other way steam savings are achieved is through heat integration. Heat integration, the placement of heat exchangers to make optimal use of waste heat, is applied to distilleries and sugar factories in order to arrange the promising technologies identified into the most energy efficient configurations (Chapter 3). This is accomplished using "pinch" analysis to determine the minimum steam use of a set of components and linear programming modeling to derive the heat exchanger network which realizes minimum steam use.

Another major issue of electricity-ethanol co-production addressed here is the determination of the conditions under which co-production would be economically desirable (Chapter 4). The key parameters involved in co-production are identified as well as the market conditions required for co-production to compete.

Three co-production scenarios are evaluated. The first scenario is the autonomous distillery, which produces hydrous fuel ethanol[e] directly from cane. The second is the combination of a sugar factory and an annexed distillery, which ferments the sugar factory's molasses into anhydrous ethanol for blending with gasoline[f]. The third scenario investigated is the combination of a sugar factory and an annexed distillery with the capacity of an autonomous distillery. Such a setup would be able to switch production between sugar and ethanol, depending on the profitability of the two products.

Like fossil fuels, sugar cane energy is not the answer to all of the developing world's energy problems. It does, however, have the potential to make an impact on the future of the 80 cane growing developing countries. This thesis examines this potential from a technical perspective and evaluates the financial and economic conditions under which electricity-ethanol co-production would be a viable alternative to more conventional energy sources.

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[e] Hydrous ethanol contains approximately 5% water and is burned in Otto cycle engines specially designed for "neat" ethanol use. The water content of hydrous ethanol prohibits it from being blended with gasoline.

[f] There is no reason that an annexed distillery could not produce hydrous ethanol or that an autonomous distillery could not produce anhydrous ethanol. However, given that an autonomous distillery would produce ethanol on a large scale, it is not likely than an investment would be made in one unless large ethanol demands are assured, such as an existing or government supported market for neat ethanol cars. For instance, autonomous distilleries were not built in Brazil until after the decision was made by the government to introduce neat ethanol burning cars [15](annexed distilleries were already common). The small scale of the annexed distillery is appropriate for generating ethanol for blending with gasoline (~15%-20% ethanol).

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Figure 1.1a: Energy Content of Sugar Cane.

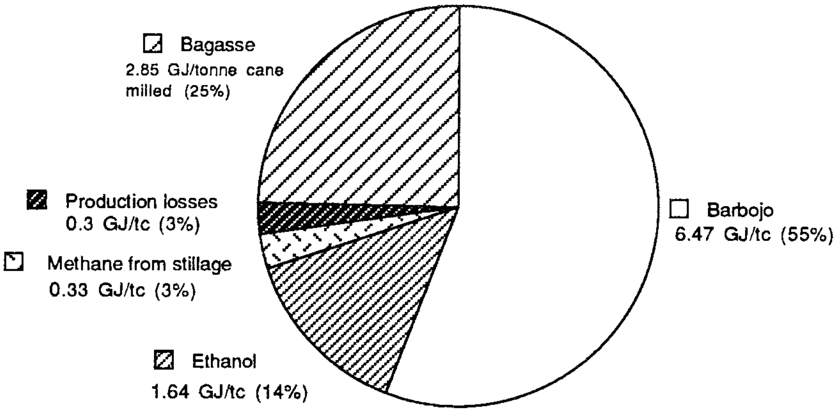


Figure 1.1b: Mass Breakdown of Sugar Cane.

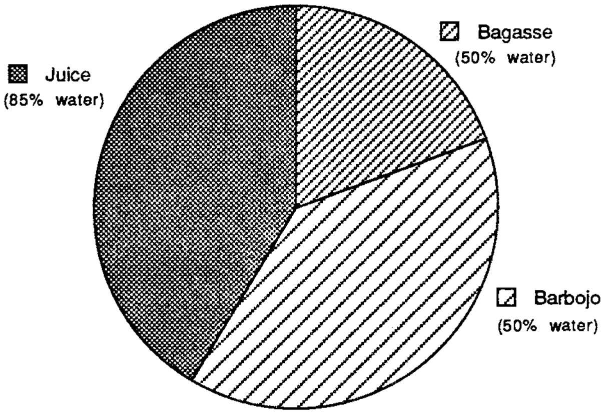
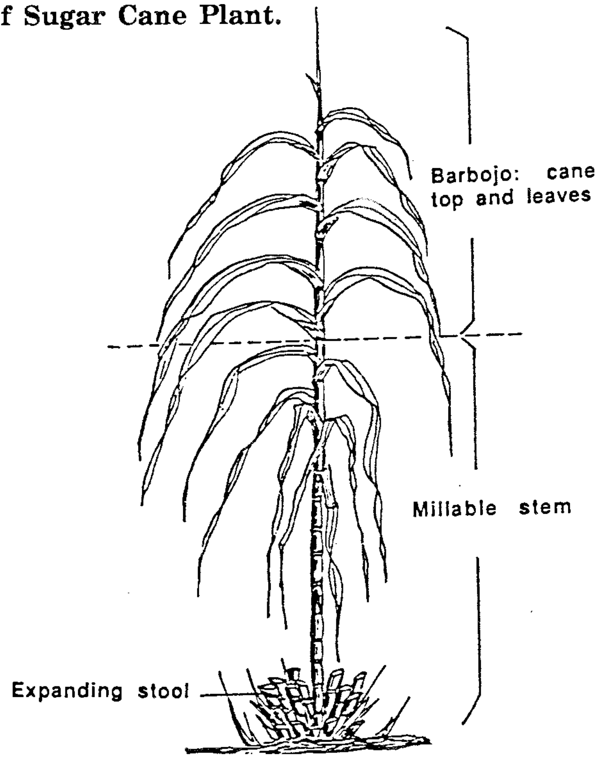
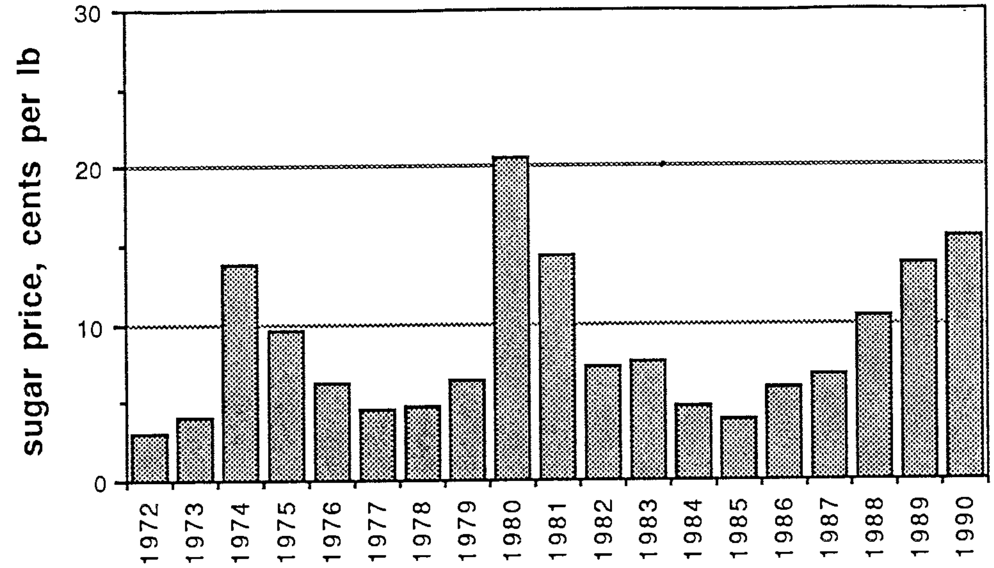


Figure 1.2: Diagram of Sugar Cane Plant.



Source: Alexander, A.G., *The Energy Cane Alternative*, Elsevier Science Publishing Co., New York, 1985, Fig 3-2.

Figure 1.3: Historical International Market Price for Raw Sugar, 1987 US$.



Source: *International Sugar Organization Year Book*, 1979 and 1987, ISO, London, and the *I.S.O. Statistical Bulletin*, January, 1988 through January, 1990. The 1990 sugar price is from *The Wall Street Journal*, May 1, 1990. Prices adjusted using gross national product deflator.

Chapter 2:
Survey of Technologies for Efficient Ethanol-Electricity(-Sugar) Co-Production

1.0 Introduction to Cane Sugar and Alcohol Production

This chapter examines the energy-consuming processes involved in ethanol and electricity production from sugar cane and presents energy-efficient alternatives which might be employed. Energy-conserving technologies are interesting in the context of ethanol-electricity co-production for two reasons. Obviously, the more the energy consumed in ethanol production or electricity generation, the less that is available for sale. Second, as will be shown in Section 2, the more promising cogeneration technologies are based on steam-injected gas turbines (STIGs), which have upper limits to the amount of steam which they can produce. Many of the technologies introduced in this chapter will be used in the quantitative steam economy analysis of Chapter 3.

1.1 Tours of Typical Sugar Factories and Alcohol Distilleries

Before examining the energy consuming processes in ethanol, sugar and electricity production, an introduction to the overall production procedures are in order. This section describes qualitatively the steps involved in converting sugar cane into ethanol, raw sugar and electricity using conventional technologies.

1.1.1 Overview of a Typical Cane Sugar Factory

Figure 2.1 shows the primary steps and material flows for a sugar factory with an annexed distillery. Once the cane has been delivered to the factory, the juice must be extracted from the fibrous cane. The long cane stalks are washed and broken into smaller pieces by cane knives and shredders in preparation for juice extraction. A typical sugar factory or autonomous distillery might have two or three cane knife/shredder stations placed before the mills.

A series of roller mills then squeezes the juice out of the cane bits. Often three to seven sets of "three roller mills" are used, depending on the toughness and fiber composition of the cane being milled. Water is sprayed onto the crushed cane ("imbibition") to enhance sugar extraction. From 90% to 95% of the sugar present in the cane will be removed during milling.

Two product streams leave the mills: raw sugar juice and wet bagasse. The wet bagasse (50% moisture content) is either fed directly to the boiler or stored in large outdoor piles, awaiting the boiler or some other use (e.g. fiber board manufacturing, pulp and paper, chemicals, etc.).

The raw juice is rather turbid and acidic and must be neutralized and cleaned. This is accomplished in the clarifier stage, where the juice is filtered, treated with milk of lime (CaO) to neutralize the naturally occurring acids and heated to near boiling so that the waxes, gums, etc., in the juice will coagulate and can be decanted off or filtered.

The clear juice is then heated and concentrated from about 15% dissolved solids (°Brix in sugar industry jargon) up to 60° to 70° Brix by boiling off the excess water. Concentration is performed in a multi-effect evaporator. A multi-effect evaporator consists of a series of separate evaporator vessels ("bodies" or "effects") operating at descending pressures so that the juice in each ensuing effect has a successively lower boiling temperature. This allows the vapors which have been evaporated from one effect to serve as the heat source of the following effect (the first effect is driven by steam), allowing more evaporation per kilogram of steam input. Generally, the evaporation occurs in three to five effects, descending from about 0.25 MPa (2.5 bar) to about 0.096 MPa (.96 bar or 68 cm water vacuum).

The syrup leaving the evaporators goes to the vacuum pan section, where the raw sugar is crystallized out of the solution. The syrup is placed into a vessel (vacuum pan) where it is evaporated further until the sucrose precipitates out of the syrup. The sugar-syrup slurry, known as massecuite, is centrifuged, the sugar crystals washed and the remaining syrup (now molasses) is reintroduced into another vacuum pan for further reduction. This process is repeated two to four times until the remaining molasses is too viscous to economically boil down further.

1.1.2 Overview of Typical Distilleries

Distilleries that derive their fermentation feedstock from sugar cane come in two varieties. The autonomous distillery ferments sugar cane juice directly into alcohol. The annexed distillery uses the molasses from an adjoining sugar factory as fermentation feedstock.

In the autonomous distillery, cane preparation, milling and clarification take place in the same manner as in a sugar factory (Figure 2.2). The clarified juice, however, is evaporated to only about 20°Brix. The slightly concentrated juice is sent to a fermenter (a large, sealed tank) and inoculated with yeast, which under anaerobic conditions ferments the sugars in the juice to ethanol and carbon dioxide.

It takes about 36 to 48 hours to completely ferment the juice, which is now known as "beer." The yeast is centrifuged from the beer and recycled. The beer is now approximately 8% to 12% ethanol by weight; most of the remaining ~90% of the beer is water, with small amounts of other organic compounds and salts.

The beer is concentrated in the distillation section. Because ethanol is more volatile than water, when the water-ethanol mixture is boiled the vapors given off are richer in ethanol. Distillation takes advantage of this fact by continually boiling and condensing the beer until the ethanol concentration reaches ~96% (by wt) At this point the volatility of ethanol and water are the same and further distillation will not increase ethanol purity. This constant boiling point mixture is an azeotrope. The product is now hydrous ethanol, which can be used as a motor fuel and is the assumed final product of an autonomous distillery (footnote f).

Rather than fermenting the partially concentrated cane juice, the annexed distillery uses the molasses by-product of raw sugar production as the fermentation feedstock. Before the molasses is fermented, it is diluted so that the final ethanol concentration is around 12%-15% ethanol. The beer is then distilled to ~95% alcohol in the same way as in the autonomous distillery.

At this point a third constituent is added in order to remove the last 5% of water and "break" the azeotrope. This is most commonly benzene, whose presence allows the ethanol to remain a liquid while the water and benzene are evaporated off. This final product is anhydrous ethanol, which can be used as an extending or octane enhancing additive to gasoline.

In both the annexed and autonomous distillery cases, about 10 to 12 times more "stillage" (the watery waste remaining from the beer once the ethanol has been removed) is generated as alcohol on a volume basis. Stillage is often disposed of into the local environment or sprayed onto the local cane fields as a weak fertilizer, and is often concentrated prior to disposal.

2.0 Cogeneration Technologies

Cogeneration has long been practiced in sugar factories. Burning bagasse in on-site boilers allows sugar factories to produce their own steam and electricity and is a convenient way to dispose of the by-product bagasse. A modest sugar factory grinding 1000 tonnes of cane per day can generate over 60,000 tonnes of bagasse in a 200 day milling season. Due to this bagasse disposal issue and the fact that there has been no market for bagasse or privately generated electricity, sugar factories have had little incentive to cogenerate more than on-site needs or to operate in an energy-efficient manner. If a market for independent electricity existed, then cogeneration at sugar factories and distilleries would become an option worth investigating. This section examines conventional cane based cogeneration technologies along with the new, more energy-efficient technologies.

2.1 Standard Technologies

Most sugar factories and alcohol distilleries use small, "medium pressure" (1.5-2.5 MPa) bagasse-fired boiler/steam turbine systems to provide just enough steam and electricity to meet the on-site needs, around 350 to 500 kgs[a] of steam per tonne cane milled (kg/tc) and 12 to 20 kilowatt-hours per tonne cane milled (kWh/tc) [11]. They are typically run inefficiently so that bagasse disposal does not become a problem. One exception to this is in southeast Brazil, where bagasse is sold as a boiler fuel and as a component of cattle feed, leading some factories to be

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[a] Unless otherwise noted kilograms steam refers the physical quantity of saturated steam at ~.2 MPa, with an energy content equal to approximately the H_{fg} of 2200kj/kg. In the heat integrated cases, another ~200kj/kg can be added due to sensible heating from the condensates ($\Delta T = 50°C$).

more energy efficient so as to free up surplus bagasse for these markets [1].

2.2 Advanced Technologies

The first step toward more efficient power generation in sugar factories is operating the boilers at a higher pressure so that more electricity and work can be obtained in the steam turbines. This is done in a few sugar factories and alcohol distilleries where condensing-extraction steam turbine systems (CEST) operated at 4.0 to 6.0 MPa have been installed. These systems produce enough steam to supply a typical sugar factory or distillery and export 50 to 100 kWh of electricity per tonne cane milled (kWh/tc) to the local grid (Figure 2.3)[2].

CEST systems are operating in Hawaii, Mauritius and Reunion, and are being considered elsewhere [2]. In 1985, over 58% of the power generated on the island of Kauai, Hawaii, was cogenerated at sugar factories [2]. Over 8% of the electricity generated in Mauritius is derived from bagasse [2].

Steam turbine technology, however, presents a number of drawbacks. First, there is a strong economy of scale, with a 3 MW CEST system costing almost twice as much per kW of capacity as a 30 MW system ($3010/kW vs. $1510/kW) [2]. The high thermal efficiencies seen in Rankine cycle central station power plants cannot be achieved at the modest scale indicative of sugar cane cogeneration[b,c][3]. Typical

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[b] Most sugar factories mill between 1000 and 3000 tonnes cane per day, enough to support between 6.3 and 18.9 MW of CEST cogeneration capacity.

[c] There are two fundamental reasons why smaller Rankine cycle systems have lower efficiencies. First, the "geometric similarity" between a small and large turbine cannot be maintained-- blade tip clearances and losses on small turbines are proportionally larger than on large ones. Second, economic considerations come into play. For a larger the system, the energy savings of more complex cycles (such as bleeding for feedwater heating) can be justified. Also, on smaller Rankine systems, cheaper materials are used in the boiler (to minimize the diseconomy of small scale), lowering the amount of steam superheat and reducing efficiency.[3]

performances and costs of CEST cogeneration are show in Tables 2.1 and 2.2.

One alternative to boiler-steam turbine cogeneration is the gas turbine. Recent studies have suggested that the biomass-integrated gasifier/steam-injected gas turbine (BIG/STIG) is an attractive technology for sugar cane cogeneration [4]. In a steam-injected gas turbine cycle, steam is generated using the turbine exhaust heat in a heat recovery steam generator (HRSG) and either used as process heat or injected into the combustor and turbine section of the engine to enhance power generation and efficiency (Table 2.1). Aircraft derived (aeroderivative) gas turbines are generally used for steam-injected cycles because their turbine sections are designed to accommodate mass flows much higher than their nominal rating [2].

Table 2.1: Performance of Biomass-Fired Cogeneration Systems.

	COGENERATION						POWER ONLY			
	Electricity,		Steam,		Fuel,	Cane,[c]	Electricity,		Fuel,	Cane,
	MW	%HHV	T/H	%HHV	T/H	T/H	MW	%HHV	T/H	T/H
CEST[a]										
Generic	17.5	13.0	65.6	35.9	50.8	169	27.0	20.3	50.2	167
Generic	6.1	11.4	26.4	36.4	20.2	67	10.0	17.8	21.2	71
Generic	1.8	10.1	9.0	37.2	6.73	22	3.0	15.7	7.22	24
BIG/STIG[b]										
LM-5000	38.8	31.3	47.7	30.0	27.6	157	53.0	35.6	33.0	188
LM-1600	15.0	29.8	21.8	33.8	11.2	65	20.0	33.0	13.2	75
GE-38	4.0	29.1	5.7	32.4	3.06	17	5.4	33.1	3.63	21
BIG/ISTIG[d]										
LM-8000	97	37.9	76.2	25.4	57.3	325	111.2	42.9	57.7	328

Notes for Table 2.1:

[a] Adapted for E.D. Larson and R.H. Williams, "Biomass-Fired Steam Injected Gas Turbine Cogeneration," Proceedings 1988 ASME Turbo-Cogen Symposium, Montreaux, Switzerland, Aug. 30-Sept. 1, 1988.

[b] Estimates for the LM-5000 are based on private communications from M. Erbes, General Electric Corp. to E.D. Larson. Other estimates are adapted from Table 3 in Chapter 4 [10], assuming that the gasification efficiency is the same for biomass as for coal (Table 2 in Chapter 4[10]).

[c] Assuming that the BIG/STIG and BIG/ISTIG use briquetted bagasse or barbojo with moisture content 15%, which has a higher heating value of 16,166 kJ/kg. CEST uses 50% wet bagasse having a higher heating value of 9350 kJ/kg. We further assume that 300 kg of 50% wet bagasse are produce per tonne of cane milled, or 176 kg of 15% wet briquetted bagasse are produced per tonne cane milled.

[d] Preliminary estimate of steam and electricity production, based on performance with coal. From J.M. Ogden and M.E. Fulmer, Assessment of New Technologies for Co-Production of Alcohol, Sugar and Electricity from Sugar Cane, PU/CEES report no. 250, May, 1990.

The aeroderivative STIG cycle has been industrially demonstrated by Allison Turbine Engine Company and General Electric [4,5] (among others) and can be found throughout the world.

The biomass gasifier adds some complexity to the system. Air from the compressor, a portion of the high pressure steam from the HRSG and the biomass fuel are fed into a pressurized gasifier. The resulting low energy density fuel gas is routed through a cyclone to remove particulate matter and sent to the combustor of the gas turbine and burned (Figure 2.4).

Both fixed and fluidized bed gasifiers appear attractive for biomass gasification. The fixed bed model would be the most likely candidate in the near term because of its simpler operation and fuel gas cleanup [6]. Its large disadvantage is that the bagasse or barbojo fuel would have to be densified, adding from ~$1 to $2 per GigaJoule to the cogeneration facility's operating costs [2].

Because of its ability to handle a wide variety of unprocessed biomass fuels and its projected lower unit capital cost, the fluidized bed gasifier is the more likely technology in the long term[6]. In either case, the gasifier would have to be pressurized (to avoid the cost of compressing the fuel gas to the combustor pressure) and be air blown (to avoid the scale sensitive O_2 plant required with an oxygen blown gasifier).

A biomass integrated gasifier/gas turbine with a heat recovery boiler/steam turbine (combined cycle) presents a more conventional approach to linking gas turbine cogeneration to biomass fuels. The drawback to these systems is similar to that of the straight steam turbine system: the cost of the steam turbine and associated condensers is too great at the small scales typical of biomass fired systems [6].[d]7 The steam bottoming cycle could be removed and the system operated as a

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[d] General Electric's is due to release the LM6000 in 1992, which in a combine cycle configuration is expected to generate ~55 MW at ~52% efficiency and a generating set (engine plus generator) cost of $250/kw [7]. This system is

Table 2.2: Capital and Operating Costs of Biomass-Fired Cogeneration Systems (1986 $).

	Capacity, MW	Installed Cost, $/kW	Maintenance Fixed, 1000$/Y	Variable, $/kWh	Labor, (1000$/Y)
CEST[a]					
Generic	27.0	1556	664	0.003	129.2
Generic	10.0	2096	246	0.003	97.2
Generic	3.0	3008	73.8	0.003	97.2
BIG/STIG[b]					
LM-5000	53.0	990	1304	0.001	297.0
LM-1600	20.0	1230	492	0.001	108.0
GE-38	5.4	1650	133	0.001	97.2
BIG/ISTIG[c]					
LM-8000	111.2	770	2736	0.001	405.0

Notes for Table 2.2:

[a] See note a, Table 2.1.

[b] E.D. Larson R.H. Williams, "Biomass-Fired Steam-Injected Gas Turbine Cogeneration," Proceedings of the 1988 ASME Cogen-Turbo Symposium, Montreux, Switzerland, Aug.30-Sept. 1, 1988.

[c] Based on personal communication with E.D. Larson.

simple cycle gas turbine, but the nagging problem of poor part-load performance encountered with simple cycle gas turbine cogeneration would be encountered[e] [2].

BIG/STIG cogeneration systems offer a number of advantages over CEST systems. Because of higher electrical efficiency, BIG/STIG systems could generate over twice as much power per tonne cane as CEST systems (up to 250 kWh/tc) and over 10 times as much as is generated in the turbo-alternators used in most sugar factories today. BIG/STIG systems also tend not to experience a strong economy of scale, allowing modest sized systems to cost around $1000/kW of capacity (Table 2.2 [2,4]).

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likely to prove viable for biomass-gasifier applications such as in sugar factories and alcohol distilleries.

[e] Simple cycle gas turbine cogeneration is largely restricted to applications with constant heat loads.

Although BIG/STIG systems are not commercially available at present, it is estimated that they could be commercialized in less than five years [8]. The research and development effort needed to bring BIG/STIG systems to market is believed to be rather modest, considering that much of the development work that has been done on gasified coal-gas turbines would be applicable to biomass versions.

The biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG) is similar to the BIG/STIG system, except that the gas turbine includes an intercooler between the two compressor stages. This reduces the required compressor work and allows operation at higher turbine inlet temperatures due to the improved blade cooling using the cooler air from the compressor. These changes result in a seven to eight percentage point increase in efficiency over BIG/STIG and an additional 50 kWh/tc electricity output. Installed costs are estimated to be around $770/kw for a 110 MW BIG/ISTIG system (Tables 2.1 and 2.2; [9,10]).

It is estimated that natural gas fired ISTIG technology would require a four or five year development effort [11]. If this development occurred parallel to that of BIG/STIG technology, then BIG/ISTIG could be available shortly thereafter.

Figure 2.5 shows the steam and electricity production for CEST, BIG/STIG, BIG/ISTIG and typical existing low pressure boilers with turbo-alternators operated in-season on bagasse. For each system, a range of operating values is possible, depending on the steam required for process use in the sugar factory or distillery. When the process steam demand is zero, the electricity output is maximized, as it would be during off-season operation.

The right hand end of each line shows the maximum steam output for each of the three systems. The more electrically efficient gas turbine cogeneration systems have much lower maximum steam outputs than the CEST system. With BIG/STIG, the maximum amount of 2.0 MPa steam (316°C) available is around 300 kg per tonne cane, while for BIG/ISTIG cogeneration it is only around 235 kg/tc. As has been discussed, the typical steam requirement of an autonomous distillery or a sugar factory can be well over 400 kg/tc. Clearly, if these efficient gas turbine cogeneration technologies are to be employed in the sugar cane industry, significant steam economy improvements must be made.

3.0 Steam Use and Economizing Technologies in Sugar Factories

Energy use in a typical sugar factory and autonomous distillery is shown in Table 2.3 while Figures 2.6 and 2.7 show schematics of the steam use in a typical sugar factory and autonomous distillery. Steam is produced in a medium pressure boiler (1.5 to 2.5 MPa), expanded through back-pressure turbines driving the cane mills and a turbo-alternator which generates enough electricity to meet the on-site needs. The exhaust steam from the mill turbine and turbo-alternator is at ~.2

Table 2.3a: Energy (Steam) Consumption In a Typical Cane Sugar Factory [12,13].

Component	Energy Use, MJ/tc	Kg(stm)/tc[a]	Source
Turbo-Alternator	35	180	Expand steam through backpressure turbine (2.→.2MPa)
Mill Turbines	50	200	Expand steam through backpressure turbine (2.→.2MPa)
Juice Heating	250	--[b]	Turbine exhaust steam and vapors bled from evaporator
Evaporation	620	320	Turbine exhaust steam and multi-effects.
Vacuum Pans	310	150	Turbine exhaust steam and vapors bled from evaporator
Misc. and Losses	100	40	Throughout system.

b: Energy (Steam) Consumption In a Typical Autonomous Distillery [12,13].

Component	Energy Use, MJ/tc	Kg(stm)/tc[a]	Source
Turbo-Alternator	35	180	Expand steam through backpressure turbine (2.→.2MPa)
Mill Turbines	50	200	Expand steam through backpressure turbine (2.→.2MPa)
Juice Heating	250	--[b]	Turbine exhaust steam and vapors bled from evaporator
Evaporation	150	360	Turbine exhaust steam and multi-effects.
Distillation	250	260	Turbine exhaust steam and vapors bled from evaporator
Misc. and Losses	100	40	Throughout system.

[a] Because of the varying steam pressure levels and steam reuse throughout the system (e.g. exhaust steam from mill turbines used for evaporation and vapors bled from the multi-effect evaporator used for juice heating), the steam use per tonne cane cannot be simply added to arrive at overall steam use.

[b] Because of the many combination of bleeding steam from the evaporators and using turbine exhaust steam, a "typical" steam use in juice heaters cannot be assessed.

MPa (120°C) and is used for juice heating and evaporation. In a sugar factory, the vacuum pans use a portion of the exhaust steam; the exhaust steam is also used in the distillation section of an annexed or autonomous distillery.

3.1 Cane Preparation and Milling

Cane knives and shredders consume around 1.5 to 3.0 kWh per tonne cane per station. Typically, a sugar factory or autonomous distillery would have two to three cane knife/shredder stations, with a total energy demand of 3.0 to 6.0 kWh per tonne cane, primarily powered by electric motors [12,13]. Small steam turbines are sometimes used in larger factories [12].

Backpressure steam turbines are used almost exclusively for mill power. Mill energy consumption depends on mill, turbine and gearing efficiencies and the amount of time spent in part load conditions. Operating the mill at throughput less than the design capacity can increase energy consumption dramatically. With typical efficiencies operating at design speed and capacity [14], mill energy consumption is approximately 50-70 MJ per tonne cane (~140-200 kg 2.0 MPa steam expanded down to 0.2 MPa per tonne cane) [13,14].

Electric motors also occasionally drive cane mills, and if steam economy is particularly important, provide a viable alternative to steam-driven mills. They offer better controllability, easier start-up and shutdown and lower operating and maintenance costs [13,15]. Electric mills, however, also consume 10% to 15% more fuel than steam-driven mills (about 10-15kWh of electricity per tonne cane) and require a larger initial investment [13].

One alternative to milling is "diffusion." For diffusion, the cane is first chopped and shredded and fed through a single mill which extracts approximately 70% of the sucrose. The remaining sugary bagasse is placed in a tank (or series of tanks) with water, which washes out most of the remaining sugar (there is very little true osmotic diffusion). The net sucrose recovery using diffusion can be over 98%. The 70% wet bagasse leaving the diffusion tanks must be dewatered in another set of roller mills or in presses down to 50% moisture [12,14]. Although there is no net energy advantage to diffusion, the process results in better sucrose extraction and requires lower initial costs [13]. The large disadvantage of cane diffusion is that it is more difficult to control [13].

3.2 Juice Heating

The cane juice must be heated on two separate occasions during processing. The raw juice must be heated to near boiling for clarification, and the clarified juice must be heated up to the temperature of the first evaporator effect (100°-120°C). This heat load is traditionally met by bleeding vapors from evaporator effects and/or using the mill or turbo-alternator exhaust steam [15]. Juice heating occurs in shell and tube heat exchangers, with the condensing vapors or steam on the shell side and the juice on the tube side.

The main opportunity for energy savings in juice heating is through the use of hot condensates from the evaporator. As will be shown in Chapter 3, up to 79 MJ/tc of the juice heating can be performed using evaporator hot condensates. This translates into a ~35 kg (0.2 MPa)/tc steam savings.

One warning is in order. The condensates from the low pressure steam are normally be returned to the boiler. Without other means of preheating this boiler feed stream (e.g. an economizer), then the additional steam economy gained by using it for juice heating would be offset by a higher fuel consumption in the boiler.

3.3 Evaporation

Most sugar factories use forward feed, four- or five-effect, short-tube rising film (STRF, or "Roberts") evaporators, with vapors bled from the first two or three effects for use in juice heating and the vacuum pans (Figure 2.8)[14]. The first one or two evaporator effects operate under slight pressure, while the latter effects operate under a slight vacuum.

There are a number of approaches which can be taken to improve steam economy of the evaporation section. The first is to add more effects. Rillieux's first principle[f] states that one kilogram of steam will evaporate as many kilograms of water as there are effects. Thus, a four-effect evaporator will evaporate one-third more water per kilogram of steam than a three-effect evaporator, a five-effect evaporator will evaporate one-fourth more water per kilogram of steam than a four-effect evaporator, etc.

More than five effects however, have not proven practical due to temperature constraints. When the cane juice is exposed to temperatures above ~120°C to 130°C, the sucrose caramelizes (inverts) rapidly, discoloring the crystals and reducing the sucrose yield [13]. Evaporation at temperatures less than 55°to 60°C is not practical for operational reasons (non-condensable gas entrainment, viscosity at low temperatures, controllability) [13]. Given the temperature constraints and the heat transfer coefficients in evaporator bodies, the steam savings achieved by adding a sixth effect cannot offset the additional costs of the sixth evaporator effect and of the increased evaporator heat transfer areas resulting from the reduced approach temperatures between effects.

The most promising technology for steam savings in the evaporator section is the use of long-tube, falling-film evaporators (LTFF, Figure 2.9). LTFF evaporators are common in the beet sugar and dairy industries, where cheap fuel such as bagasse is not available, and are being investigated in the cane sugar

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[f] Rillieux patented the first multi-effect evaporator in the 1840s, and developed a set of generalized rules for estimating evaporator performances [12].

industry [16,17]. The primary advantage of LTFF evaporators is the higher juice flow velocity (three to four times faster than in a Roberts STRF) which both enhances heat transfer and shortens juice residence times. This allows the first effect of the evaporator to operate at higher temperatures without the risk of caramelization [18]. With the first effect of a LTFF evaporator operating at 5°-20°C hotter than a STRF model and with the smaller temperature intervals between effects, hotter vapors can be bled from the third or fourth effect rather than from the first or second, allowing more evaporation per kilogram of steam. Evaporator condensates are also hotter, decreasing the vapors bled for juice heating.

Using mechanical vapor recompression (MVR), electricity can be substituted for a portion of the steam in evaporation. With MVR, vapor evaporated from the cane juice is compressed to a higher pressure and temperature and fed back into the same evaporator vessel on the hot side. If no vapor is bled from the system, the amount of vapor returned to the evaporator will equal approximately the amount of vapor boiled out of the cane juice, and only around 10% of the net steam demand would need to be met by steam. Tradeoffs include higher initial and operating costs for MVR for than steam-driven equipment and that electricity produced by the steam saved using MVR is in turn consumed in MVR.

3.4 Crystallization

In present practice, the raw sugar is crystallized out of the syrup in batch (discontinuous) vacuum pans. This is generally performed in two or three stages or "strikes." In each stage, the thick syrup is boiled down; as water is evaporated from the syrup, the solution becomes supersaturated with sucrose, which crystallizes out of the solution to form the molasses-sugar massecuite. When as much sugar has been crystallized out as is practical, the massecuite is removed from the pan and the sugar is centrifuged out of the molasses.

The first stage (or "A strike") removes about 68% of the sugar in the solution (the A sugar). The remaining solution (the A molasses) is diluted with clarified juice (for viscosity reasons) and the process is repeated in the B strike. The B strike crystallizes out ~21% of the sugar, leaving B molasses. The B molasses is diluted, and the process is repeated in the C strike. The C sugar is often of a lesser quality and is recycled back into the A and B strikes. The remaining C molasses is sold on commodity markets or used as feedstock for alcohol production.

Because steam is used for washing the pans after each batch and agitating the massecuite during boiling, and because the molasses is diluted between strikes, 1 kg of steam evaporates only 0.6 to 0.8 kg of water from the massecuite. Overall, vacuum pans consume from 120 kg/tc to 170 kg/tc [15], a portion of which is often bled from the first or second evaporator effect.

The greater difficulty with batch vacuum pans is not the high steam demand but rather the variability of the steam requirement. When a strike is first placed into the pan, the steam load is very high; as the strike progresses, the steam load diminishes to the point where it is negligible when the sugar and molasses are being "dropped" from the pan into a centrifuge. Operating a set of pans in a time-staggered manner minimizes this variability, but cannot eliminate it.

The continuous vacuum pan offers advantages both in the stability of the steam demand and the overall steam requirements. While there are a number of different schemes for continuous pans, they generally involve a horizontal cylindrical tank, partitioned into 10 to 20 sections. The massecuite flows from section to section, boiling off water until in the final section the desired level of crystallization has occurred. The massecuite is then centrifuged and the molasses diluted and returned to another continuous pan for more boiling.

Steam demand is 25% less with a continuous vacuum pan than with a batch pan (100-125 kg/tc, 0.1-0.25 MPa steam) [13]. Continuous pans also have increased capacity, reduced sugar losses and are easier to control [13].

4.0 Fermentation Technologies

Fermenting the cane juice or diluted molasses requires energy only for pumping the feed and centrifuging the yeast cells from the fermented broth. The fermentation itself is actually exothermic and requires cooling.

Fermentation can, however, have a direct effect on the heat requirements of distillation and stillage disposal. The more dilute the ethanol beer leaving the fermenters is, the more energy must be dedicated to distilling the beer and perhaps concentrating the stillage. Therefore, the basic chemistry of ethanol fermentation will be reviewed along with new and existing fermentation technologies.

Alcohol fermentation is a biochemical process by which microorganisms, generally yeasts or molds, partially oxidize sugars into alcohol and carbon dioxide. For the fermentation of ethanol, yeast has been found to be the most productive organism and is thus most common in large scale applications.

Prior to fermentation, the enzyme invertase hydrolyses sucrose into D-glucose and D-fructose (6 carbon or "invert" sugars):

$$\underset{\text{(sucrose)}}{C_{12}H_{22}O_{11}} + H_2O + \underset{\text{(enzyme)}}{\text{invertase}} \rightarrow \underset{\text{(D-glucose)}}{C_6H_{12}O_6} + \underset{\text{(D-fructose)}}{C_6H_{12}O_6}$$

Because invertase naturally occurs in sugar cane, this process can take place prior to fermentation. Very little of the sucrose is hydrolyzed prior to processing (the invert sugars do not crystalize), and in fact inversion and fermentation can severely degrade the quality of cane if too much time passes between cutting and milling.

In the absence of oxygen, yeast metabolizes the invert sugars into ethyl alcohol (primarily) and carbon dioxide, releasing energy in the process:

$$\underset{\text{(invert sugar)}}{C_6H_{12}O_6} \rightarrow \underset{\text{(ethyl alcohol)}}{2C_2H_5OH} + 2CO_2 + 31{,}200 \text{ cal/mol}$$

If this reaction were taking place in the presence of oxygen, the invert sugars would be completely oxidized into carbon dioxide and water. Approximately 3% of the energy content of the sucrose is unavoidably lost its conversion to ethanol.

Simple chemistry shows that if all of the invert sugars are converted to ethanol and carbon dioxide, the products would be 51.1% ethyl alcohol and 48.9% carbon dioxide (by wt). Typically, 90% of the theoretical yield is achieved [15].

A number of major practical considerations limit the rate and productivity of ethyl alcohol fermentation. Above a certain temperature yeast will not grow. Because the optimal temperature for ethanol fermentation is between 32°C and 38°C, and considering that the partial oxidation reaction is highly exothermic, cooling of the fermenting beer is always required.

Another practical consideration is alcohol inhibition. The higher the concentration of alcohol the fermenting beer contains, the slower the yeast growth and the slower the alcohol production. At alcohol concentrations above about 15%, yeast will not grow. For this reason, the solution to be fermented usually starts with 15-18% fermentable sugars so that the ethanol concentration in the final beer is no more than 8%-12%.

4.1 Present Fermentation Technologies

In virtually all distilleries, ethanol fermentation is a batch process, where the sugar-water-yeast mixture is placed in a large tank and allowed to react until the desired alcohol concentration is reached, usually taking about 36 to 48 hours. The tanks are kept at a constant temperature (around 32°C) using cooling coils and often agitated to promote mixing [19]. The yeast cells are frequently separated out of the beer prior to distillation and recycled. Typical productivity for batch fermentation is 1.8-2.5 grams ethanol per liter tank volume per hour (g/l·h)[20].

Continuous fermentation can dramatically increase productivity over batch processes, but very few continuous fermentation technologies have been demonstrated at an industrial scale [20]. The most common continuous method is cascade-coupled fermentation tanks, where the outflow of one fermentation tank is the input of the next, with each tank having successively higher alcohol concentrations. These systems use the same basic technologies as batch fermentation.

One variation on batch or cascade-coupled fermentation is the use of flocculent yeast. Although in principle it is no different from using standard yeast strains, flocculent yeast simplifies the separation procedure by replacing the centrifugal yeast separator with a simple settling tank, reducing both initial and operating costs [20]. It presents, however, serious drawbacks in yeast selection; other desirable yeast traits are sacrificed in a flocculent strain. Still, overall, the flocculent yeast allows higher yeast concentrations in the fermenter and thus increases productivity to 30-40 g/l·h [20].

4.2 Advanced Fermentation Technologies

Most of the advanced fermentation processes here increase the productivity of fermentation-- the amount of ethanol fermented per liter of fermenter volume per hour. High productivity reduces the volume capacity required for fermenters (and hence lowers costs), but does not affect the ethanol yield per gram sugar.

A more advanced application of flocculent yeast is in the tower fermenter. The cane juice or diluted molasses is fed into the bottom of the fermenter and slowly rises up the tower through the flocculent yeast. The ethanol concentration increases as the feed flows up the tower, reaching the desired concentration at the top. Because of the high concentration of yeast, productivity can be 80 times that of batch fermentation [20]. The tower fermenter is used in potable beer fermentation [20] and is being demonstrated in an annexed distillery in Australia [15].

Another method of maintaining high yeast cell density and high productivity is physically separating the yeast from the inhibiting ethanol and by-products by a filter or membrane. One proposed application of this method is the rotofermenter [21]. The rotofermenter consists of a cylindrical filter membrane through which the fermenting feed is drawn. The cylindrical filter is rotated at high speed to prevent clogging. While this technique has shown productivity of up to 27 g /l·h, it has not been demonstrated outside of the laboratory.

Another alternative is fermentation under vacuum (Vacuferm [22]). By maintaining the fermentation tank at a low pressure (30-35 mm Hg), the ethanol merely evaporates off as it is created. This allows high fermentable sugar and yeast concentrations, because at no point does the ethanol concentration approach inhibiting levels. The serious drawback to the Vacuferm process is the energy required to pump off the CO_2 to maintain the required low pressure. Although the method shows some promise, it has yet to be demonstrated at an industrial level.

One promising continuous fermentation method is the Biostil, produced by Alfa-Laval (Figure 2.10). Biostil consists of a fermentation tank from which the fermenting feed is continuously drawn. The yeast is centrifugally separated from the beer, which is then heated and introduced into a distillation column. A 30-40% ethanol solution is drawn off the top of the column and further distilled, while the column bottoms are used to pre-heat the beer entering the column before being returned to the fermentation vat. The primary advantage of the Biostil system is that the fermenter feed can have much higher sugar concentrations (because ethanol inhibition is not a factor), so that the stillage is highly concentrated. If the stillage must be concentrated by evaporation, it can greatly reduce the energy demands of stillage disposal. Biostil has been demonstrated at industrial levels in

Australia, Sweden and Brazil but is not presently in commercial use in Brazil [23].

5.0 Distillation and Other Alcohol Separation Technologies

Once fermented, the ethanol must be separated out of the fermented beer. Distillation is by far the most common separation process, but a number of alternative separation techniques have been proposed and are under development.

Distillation exploits the fact that a multi-component mixture has different component concentrations in the vapor and liquid phases. In the case of ethanol and water, ethanol, being the more volatile component, is more highly concentrated in the vapor phase of the solution than in the liquid phase. Therefore, by repeated vaporization and condensation, ethanol can be concentrated to the desired level.

This is practically accomplished by the use of a fractioning distillation column (Figure 2.11). Within a distillation column, the liquid phase flows downward to a reboiler, where heat is added to vaporize the mixture. The hot rising vapor exchanges heat with the cooler falling liquid, condensing the less volatile component out of the rising vapor and evaporating the more volatile component out of the falling liquid. The final vapor, rich in the more volatile component, is removed at the top of the column, condensed and a portion reintroduced into the column near the top as reflux for increased separation efficiency. The final liquid, rich in the less volatile component, is removed from the column at the reboiler.

The heat transfer between the hot vapors rising and the cooler condensate falling is enhanced by a series of perforated plates (bubble plates or sieve trays) along the length of the column. The more plates there are in a column, the better the separation, but also the greater the cost and complexity.

For ethanol distillation, the end product is the strongest factor determining the number of distillation columns required and the complexity of the distillation train. The final product can be fuel alcohol or low grade industrial alcohol (80%-90% ethanol), hydrous industrial alcohol, (96%-96.5%), fine alcohol (96%-96.5%) without organic impurities (used in pharmaceutics), or absolute or anhydrous alcohol (99.9%+), used for gasohol blending. The purer the final ethanol is, the more complex the distillation process.

5.1 Hydrous Alcohol Production

Fuel grade hydrous ethanol can generally be distilled in a single column[g]. If a higher quality hydrous ethanol is required, then the top product from the first column can be sent to an aldehyde column, where the high boiling organic impurities, mostly aldehydes, are drawn off the top. If further distillation is required, the ethanol-water mixture is fed into a rectifying column, where the low boiling organic impurities (fusel oil) any remaining water are removed, resulting in a product up to 96.5% ethanol (hydrous ethanol).

Standard two column distillation consumes from 4730-7700 kJ per liter hydrous ethanol (2.1 to 4.5 kg .2MPa steam)[24,25]. Steam and energy consumption can vary considerably due to tradeoffs in design simplicity and capital cost versus steam use. It is not uncommon to inject steam directly into the bottoms in the reboiler, adding water (which will have to be removed) along with heat.

Heat integration is a particularly effective way to significantly improve steam

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[g] Often this single column is broken up into two sections: a stripping section and a rectifying section. The system behaves as if the rectifying section is mere sitting on top of the stripping section: the vapors from the top of the stripping section feed into the bottom of the rectifying section, and the liquid from the bottom of the rectifying section is pumped to the top of the stripping section. The system still contains only one reboiler and one condenser.

economy in hydrous ethanol distillation. Heat integration is the technique where heat from streams which need to be cooled (such as distillation column bottoms) are used as the energy source for streams which need heating. This approach is applied extensively in Chapter 3.

Using two columns operating at pressures such that the condenser of the higher pressure column can serve as the reboiler of the lower pressure column

Table 2.4: Energy Use for Ethanol Separation From Water: Hydrous (Azeotropic) Ethanol From Dilute Solutions.[a]

Ethanol Concentration (% wt) initial	final	Process	Energy Use kJ/liter ethanol	Process Steam Use (kg/l)[b]	Status[c]
8-10	95	Conventional 2 column distillation in typical cane alcohol distillery	6600-7700	3.0-3.5	Com[1]
8-10	95	Heat integrated distillation in innovative cane alcohol distillery	3300-4400	1.5-2.0	Com[1]
6-10	95	Conventional 2 column distillation	4730-5850	2.1-2.7	Com[24]
6-10	95	Conventional distillation with vapor re-use (multi-effect)	1950-3340	0.9-1.5	Com[24]
10	95	Conventional distillation with vapor recompression[d]	1610-1780		Com[24]
10	95	Three column distillation with vapor re-use[e]	4730-5830	2.1-2.7	Com[24]
10	95	Four column distillation	8080	3.7	Com[24]
10	95	Three effect vacuum distillation	2010	0.9	Lab[24]

[a] Based on Table 15b, reference 10.

[b] Process steam is assumed to be saturated at 120°C, with enthalpy of vaporization of 2202 kJ/kg.

[c] Status: **Com**mercially available, **Lab**oratory scale only.

[d] For vapor recompression, it is assumed that heat is converted into electricity at 33% efficiency.

[e] Three and four column distillation results in a higher grade of ethanol (without the organic impurities), therefore the energy per liter should not be directly compared with the two column values.

along with using standard heat integration, hydrous ethanol distilleries have been designed which require as little as 1.0 to 1.2 kg of steam per liter (1950kJ/liter) [24,26]. Employing this multi-effect two column distillation with heat pumps and heat integration can reduce the energy demand still further to 1610 to 1780 kJ per liter [24]. The energy consumptions of the hydrous ethanol distillation techniques discussed here and others are summarized in Table 2.4.

5.2 Anhydrous Alcohol Separation Techniques

5.2.1 Distillation Techniques

There are numerous methods in which the 96.5% ethanol-water azeotrope can be broken and the ethanol further purified. The most common method is to add a third constituent to the mixture to suppress the volatility of either the water or the ethanol. If the ethanol boiling point is to be suppressed, trichlorethylene, ethyl ether, pentane or most commonly benzene can be added [15].

A standard anhydrous ethanol distillery using benzene distillation (without heat integration) consumes from 7630 kJ to 11,000 kJ per liter ethanol produced (3.5-5.0 kg steam/l, Table 2.5). Heat integration, including operating columns at different pressure levels in a "multi-effect" configuration can dramatically reduce energy use. For instance, Raphael Katzen Associates markets a number of heat integrated systems for motor grade anhydrous ethanol distillation which consume only 5000 kJ/liter (2.3 kg/l) [26]. Other proposed (and patented) heat integrated azeotropic distillation systems claim to require only 4210 kJ per liter (1.9 kg/l) [27].

An alternative to azeotropic distillation is extractive distillation using a dehydrating agent such as ethylene glycol. The dehydrating agent suppresses the volatility of water and allows the ethanol to be boiled off the top of a distillation column. The entraining solvent and water are separated in another distillation column and the solvent is recycled. An optimized, non-heat integrated ethylene glycol extractive distillation process consumes 6050 kJ per liter (2.7 kg/liter). Heat integrated distillation sequences based on ethylene glycol extraction, it is claimed, can reduce energy use to 2150 to 1475 kJ per liter ethanol (0.9 to 0.7 kg/liter) [28,29]. While ethylene glycol extractive distillation is a demonstrated technology, the ultra-low energy use systems exist only on paper.

5.2.2 Non-Distillation Separation Techniques

A number of non-distillation separation techniques have been proposed for breaking the azeotrope and producing anhydrous ethanol. The simplest of these systems is the use of molecular sieves which selectively adsorb the water, leaving anhydrous ethanol. The only additional energy required for molecular sieves, beyond that used to distill to hydrous ethanol, is the heat to regenerate the adsorbent, around 1400 kJ per liter ethanol beyond that for distillation [24].

Two substances are used commercially as the adsorbent. Union Carbide, among others, produces zeolite molecular sieves for ethanol dehydration. A more surprising adsorbent is common corn meal or corn grits. Corn meal and corn grits appear to be as effective as the zeolite in dehydrating ethanol [30] and are easily obtained and produced in developing nations.

Dramatic energy savings are theoretically possible if non-distillation separation techniques are used. Batelle Pacific Northwest Laboratories has demonstrated in the laboratory a technique in which a solvent is brought into contact with the ethanol-water mix and extracts the ethanol. The pressure of the ethanol-solvent solution is then reduced, flashing off the solvent and leaving pure ethanol, at less than 15% of the energy needed for standard distillation [31]. Other methods include contacting the water-ethanol mixture with liquid carbon dioxide to extract the ethanol and flashing off the CO2 (67% energy savings), or blending the

Table 2.5: Energy Use for Ethanol Separation From Water: Anhydrous Ethanol from Dilute Solutions.[a]

Ethanol Concentration (% wt)			Energy Use kJ/liter ethanol	Process Steam Use (kg/l)[b]	Status[c]
initial	final	Process			
8-10	99.9	Conventional 2 column distillation in typical cane alcohol distillery + azeotropic distillation w/ benzene	9900-11,000	4.5-5.0	Com[1]
8-10	99.9	Heat integrated distillation in innovative cane alcohol distillery + azeotropic distillation w/ benzene	6600-7700	3.0-3.5	Com[1]
6.4-10	99.9	Conventional 2 column distillation + azeotropic distillation w/ benzene	7630-9650	3.5-4.4	Com[24]
10	99.9	Conventional distillation with extractive distillation with Ethylene glycol.	6050	2.7	Pro[e][28]
6.3-10	99.9	Conventional distillation + azeotropic distillation w/ benzene with vapor re-use	5000	2.3	Com[24]
10	99.9	Conventional distillation + azeotropic distillation w/ benzene with vapor recompression and re-use[d]	4230		Com[24]
10	99.9	Conventional distillation with water adsorption in molecular sieve	3340	1.5	Com[24]
10	99.9	Extractive distillation with Ethylene glycol and vapor re-use	2150		Pro[28]
10	99.9	IHOSR distillation with extractive distillation with KAc salts	1700	0.8	Lab[31]
10	99.9	Extraction with CO_2	2232-2791		Lab[31]
10	99.9	Extraction with "ideal" solvent	1005		Theo[28]
10	99.9	Vacuum Distillation	10,330		Lab[31]

[a] Based on Table 15a, reference 9.

[b] Process steam is assumed to be saturated at 120°C, with enthalpy of vaporization of 2202 kJ/kg.

[c] Status: **Com**mercially available, **Lab**oratory scale only, **Pro**posed on paper, **Theo**retical.

[d] For vapor recompression, it is assumed that heat is converted into electricity at 33% efficiency.

[e] Ethylene glycol distillation is a demonstrated technology, although this particular configuration has only been proposed and is not commercially available.[28].

mixture with gasoline, and freezing out the water, leaving gasohol (23% energy savings) [31]. These and other similar methods show promise, but have not been demonstrated at anything beyond the laboratory scale. The energy use of several of these experimental separation processes are included in Table 2.5.

5.2.3 Stillage Disposal and Use

For every liter of ethanol produced, 10 to 13 liters of stillage are generated. Stillage, the liquid waste remaining after the ethanol has been removed from the fermented beer, can have chemical oxidation demands[h] as high as 70 kg/m^3, making it as organically potent as sewage and a serious threat to the local environment.

Traditionally, stillage is either dumped into the local environment or returned to the cane fields as a fertilizer. (In addition to its organic content, stillage contains phosphorus and potassium, two important fertilizer components) When used as fertilizer, stillage is either piped or trucked directly to the cane fields dilute or concentrated at the distillery prior to shipment and re-diluted at the fields. Alternatively, the stillage can be concentrated and incinerated in a boiler, recovering a portion of the extra energy required for concentration.

The most promising alternative method of stillage disposal is anaerobic digestion. Anaerobic digestion is a process whereby micro-organisms metabolize the organic matter in a waste stream, producing methane, carbon dioxide, more microbes and a waste stream free of organic pollutants. Anaerobic digestion can produce 4.5 MJ of methane per liter of ethanol, about 1/5 as much energy as is in the ethanol itself. Even so, anaerobic digestion is more typically viewed as a waste treatment method rather than an energy production technique [33].

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[h] Chemical Oxygen Demand (COD) expresses the amount of oxygen required to oxidize all of the organic matter present in a substrate. COD is a standard method of expressing the amount of biodegradable organic matter present in a waste stream.

Anaerobic digesters are in use in a number of distilleries around the world, including at the Baccardi rum distillery in San Juan, Puerto Rico [32] and at the Sao Joao autonomous distillery in Sao Joao da Boa Vista, Sao Paulo, Brazil [33]. The Brazilian digester has a capacity of 6500 normal cubic meters of methane per day, which replaces 300,000 liters of diesel fuel and 1,200,000 liters of fuel ethanol per season. This is enough methane to run the distillery's 42 truck fleet [33].

6.0 Prior Investigation of Steam Efficient Sugar Factories and Distilleries

Because higher pressure, condensing-extraction steam turbine technology is employed in the cane sugar industry, and because alternative uses of bagasse are applied in some locations, steam economy in sugar factories and to a lesser degree autonomous distilleries has been delved into by previous researchers. The highest "low steam use" sugar factory found in the literature consumed 400 kg steam (0.1 MPa) per tonne cane [34]. This setup used quadruple effect evaporation (evaporator type not stated), with vapor for juice heating being bled from two evaporator effects. Antoine investigating cogeneration in sugar factories showed that 50 kWh/tc could be cogenerated, with process steam use at 330 kg (.2MPa) per tonne cane [35]. (The factory configuration was not described.) Paturau has designed a sugar factory which consumed only ~300 kg steam per tonne cane [36]. Research carried out at the Center for Energy and Environmental Studies has shown that sugar factory steam demands can be reduced to 240 kg steam (.2 MPa) per tonne cane [14]. Such a system would employ quintuple-effect, falling-film evaporators, continuous vacuum pans and would use hot condensates for juice heating.

Less has been written on the steam consumption of autonomous distilleries. Brazilian researchers have shown that steam demands in autonomous distilleries can be as low as 258 kg steam (.25 MPa) per tonne cane milled [1]. Such systems use five-effect evaporation with vapor bled from every effect for juice heating and a distillation section which consume 1.5 kg steam (.25 MPa) per liter ethanol.

While these steam demands are significant improvements over conventional practice, if electricity is to be a full co-product with sugar or alcohol, steam use in sugar factories and distilleries should be investigated further and more rigorous techniques applied to the problem of steam use in sugar cane processing.

7.0 Summary

Because historically there has been little reason to conserve bagasse and there has been few markets for cogenerated electricity, typical sugar factories and distilleries operate rather inefficiently, producing only enough steam and electricity to meet on-site demands. If opportunities arise to sell cogenerated electricity, a number of cogeneration technologies appear particularly attractive for improving the power output of cane-based cogeneration. These include biomass gasifiers coupled with steam-injected gas turbines (BIG/STIG and BIG/ISTIG), which could generate over ten times more electricity per tonne cane milled than is generated today.

There are a number of technologies in use today in other process industries and to a limited degree within the cane sugar industry which consume significantly less steam than systems in conventional sugar factories. These include the use of hot condensates for pre-heating the cane juice, long-tube, falling-film evaporators for juice concentration, and continuous vacuum pans for sugar crystallization.

Heat integration is a promising technique (investigated in the next chapter) for reducing energy and steam demands in distillation. One particularly attractive application of this technique is operating distillation columns at different pressures such that the condenser of a column can serve as the heat source in the reboiler of a lower pressure one.

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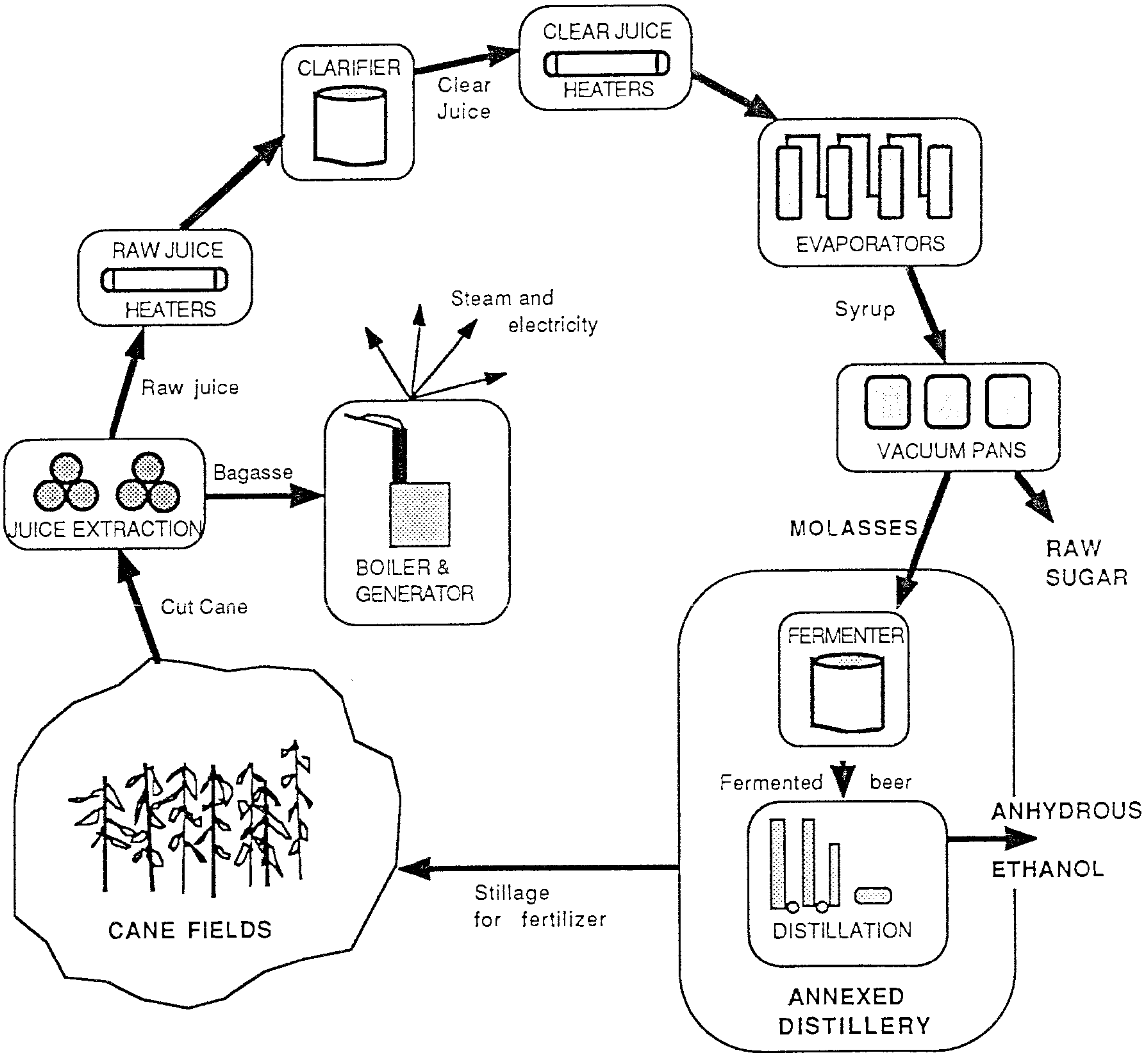
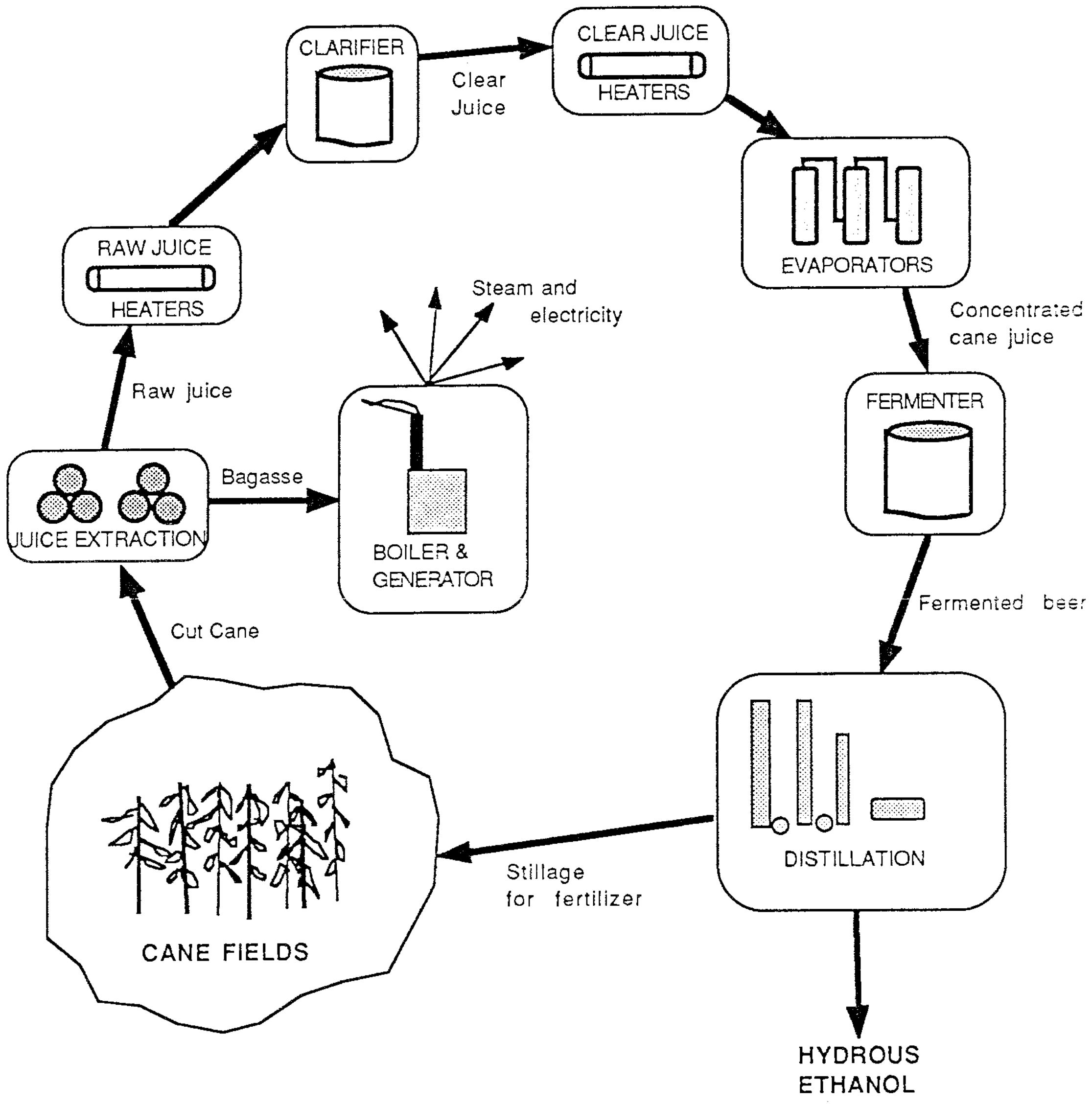
Figure 2.1: Basic Processes and Material Flows of a Typical Sugar Factory with an Annexed Distillery.



After the cane is cut and delivered to the factory, the sugary cane juice is extracted via crushing in roller mills. The cane juice is heated and clarified, the clear juice concentrated in evaporators to a thick syrup, and the sugar crystallized out of the syrup in vacuum pans. The by-product molasses from the vacuum pans is sent to the annexed distillery, where it is fermented and distilled into anhydrous ethanol. The stillage waste from the annexed distillery is used as fertilizer on the cane fields.

Figure 2.2: Basic Processes and Material Flows of a Typical Autonomous Distillery.



After the cane is cut and delivered to the distillery, the sugary cane juice is extracted via crushing in roller mills, the cane juice heated and clarified and the clear juice concentrated in evaporators. The concentrated juice is fermented and distilled into hydrous ethanol. The stillage waste from the annexed distillery is used as fertilizer on the cane fields.

Figure 2.3: Schematic of Possible Cane Sugar Application of a High Pressure Condensing-Extraction Steam Turbine (CEST) Cogeneration System.

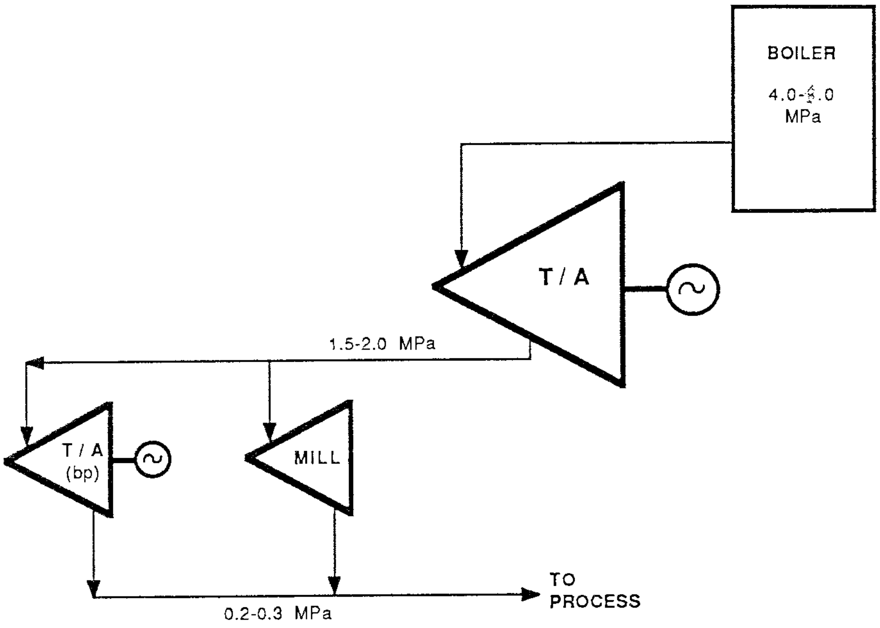


Figure 2.4: Schematic of Possible Cane Sugar Application of a Biomass Integrated Gasifier/Steam Injected Gas Turbine (BIG/STIG) Cogeneration.

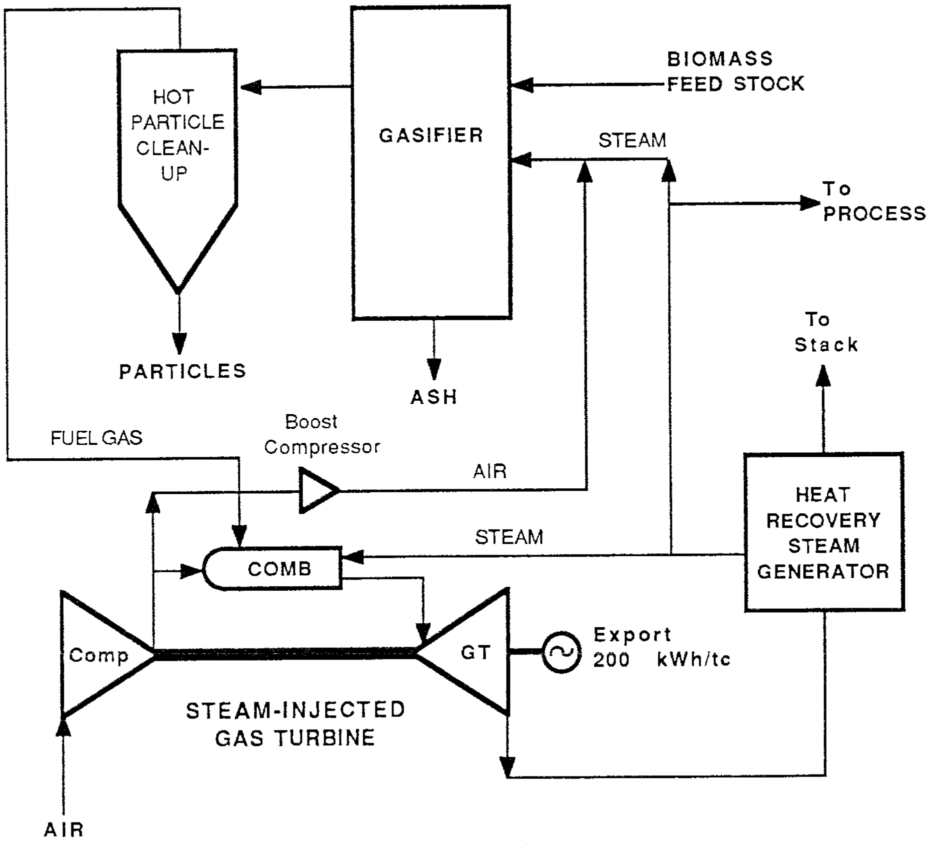


Figure 2.5: Electricity and Steam Production with Bagasse Fired Cogeneration Systems.

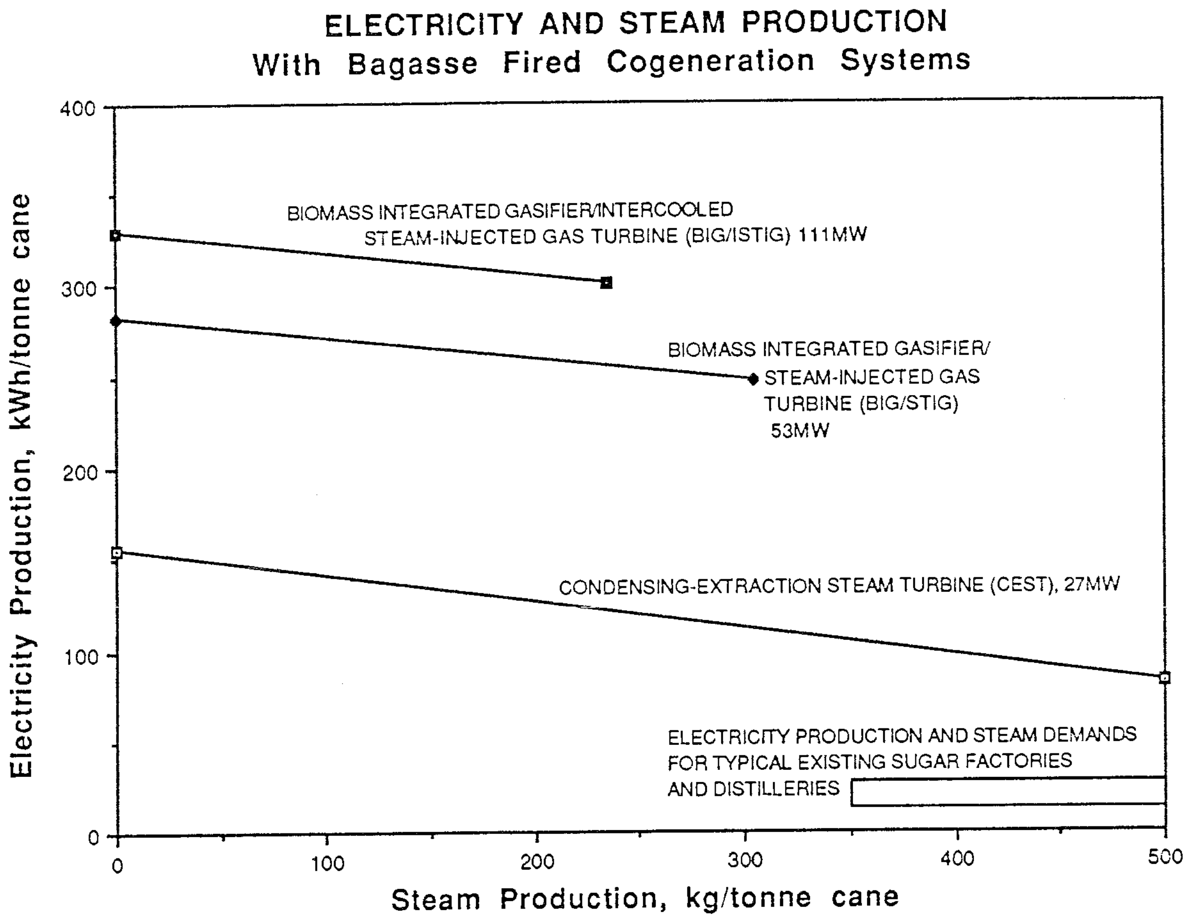


Figure 2.5 shows the in-season electricity and steam production (~2.0 MPa) per tonne cane for alternative bagasse fired cogeneration systems: CEST, BIG/STIG and BIG/ISTIG operating during the milling season at either a sugar factory or autonomous distillery. Also included is the steam and electricity production form a typical sugar factory or autonomous distillery.

The right end of each line shows the maximum steam output for each of the systems; the left hand end shows then maximum amount of electricity possible with each system, as would be the case for off-season power generation.

Figure 2.6 Steam and Material Flow for a Typical Sugar Factory Milling 125 Tonnes Cane per Hour.

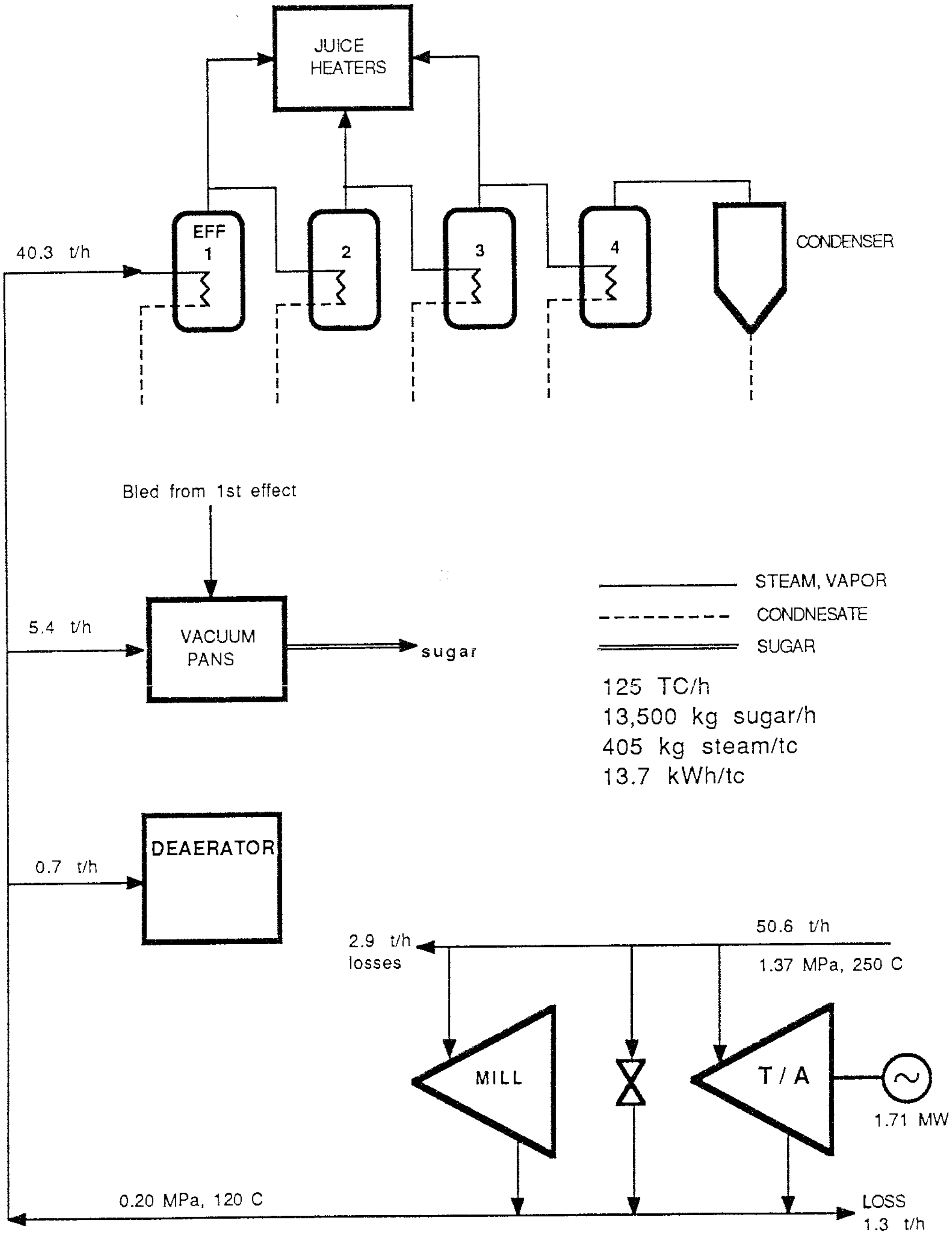


Figure 2.6 shows the steam and product flows of a typical sugar factory, milling 125 tonnes cane per hour and producing 13,500 kg sugar per hour. Conventional technology is used, requiring 406 kg/tc. The electricity demand is estimated to be 15 kWh/tc. Based on reference [12].

Figure 2.7: Steam and Material Flow for a Typical Brazilian Autonomous Distillery Milling 125 Tonnes Cane per Hour.

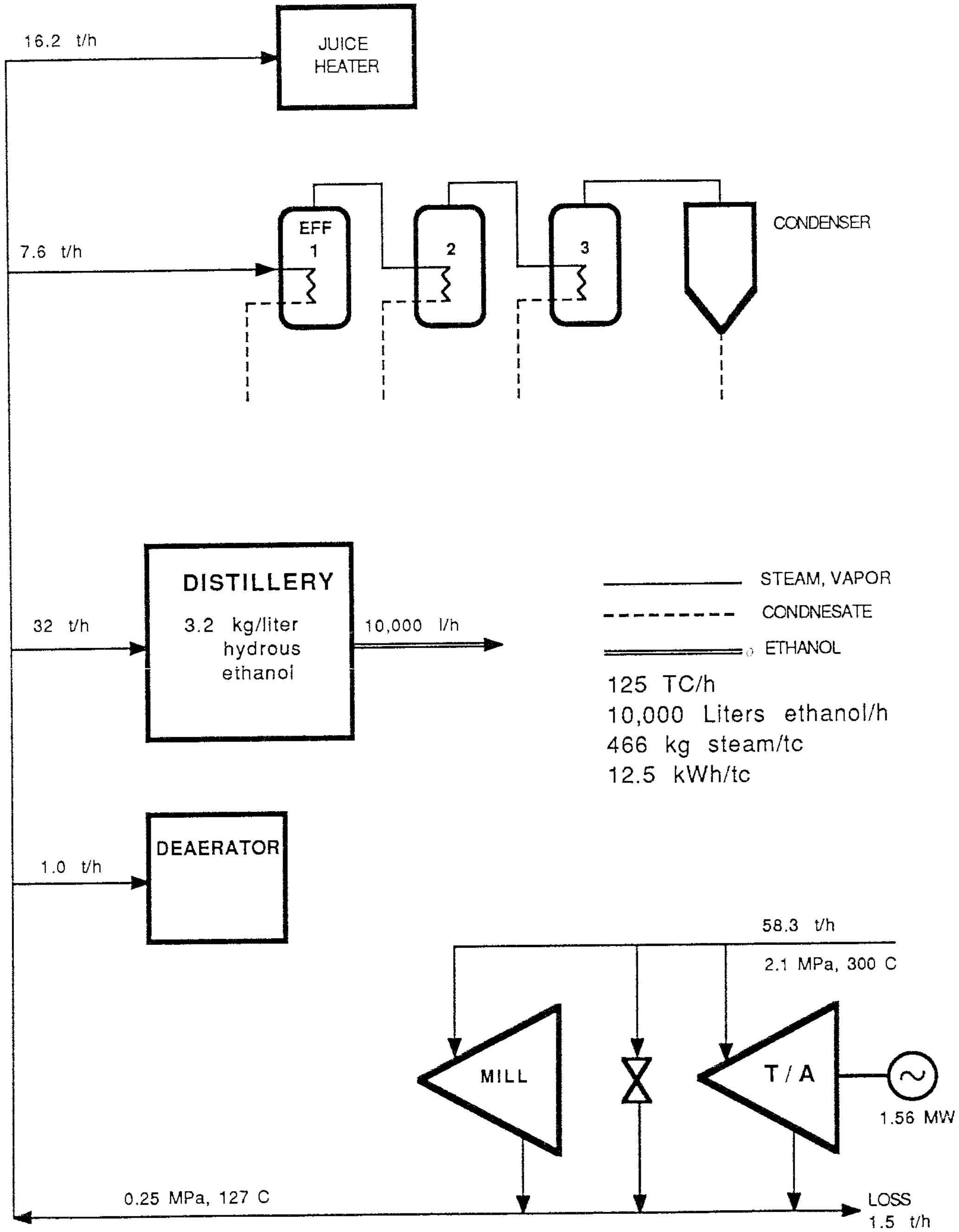
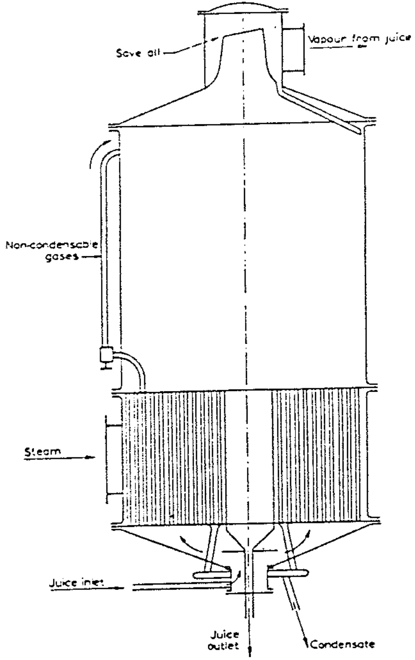


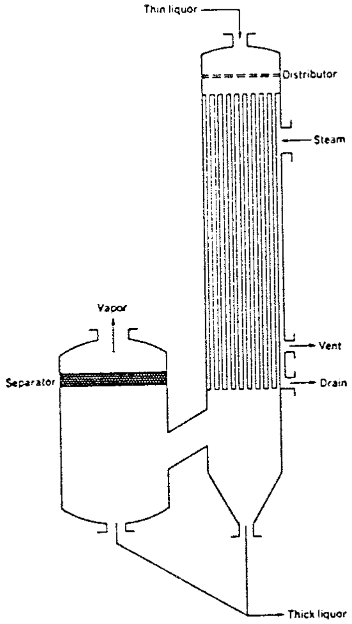
Figure 2.7 shows the steam and product flows of a typical Brazilian autonomous distillery, milling 125 tonnes cane per hour and producing 10,000 liters hydrous ethanol per hour. A conventional distillation system is used, requiring 3.3 kg steam per liter ethanol. Overall factory steam demand is 466 kg/tc. The electricity demand is estimated to be 12.5 kWh/tc. Based on reference [1].

Figure 2.8: Short Tube Rising Film Evaporator (STRF).



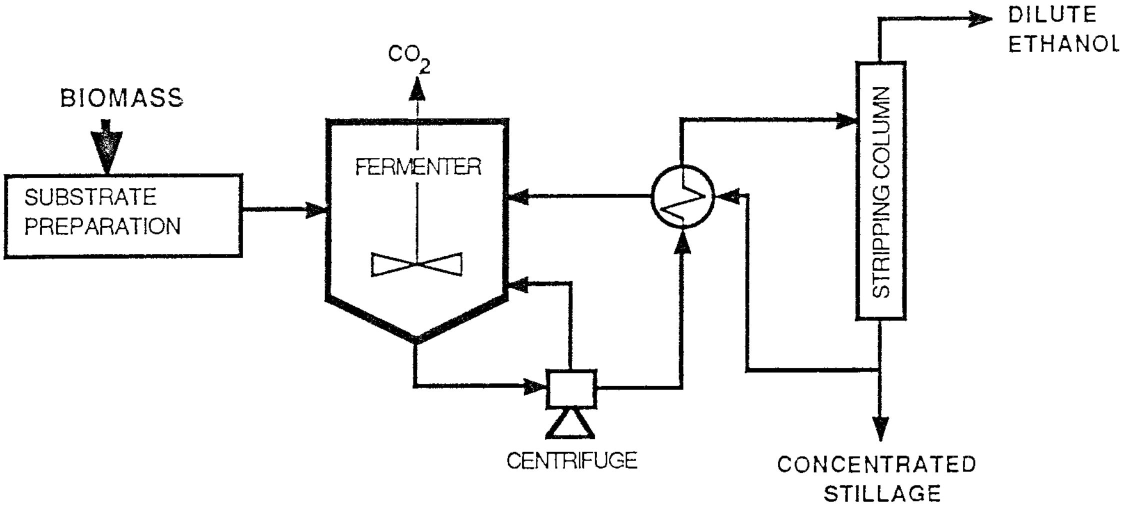
Short Tube Rising Film Evaporator (STRF) is typical of that used in sugar factories today. From reference [13].

Figure 2.9: Long Tube Falling Film Evaporator (LTFF).



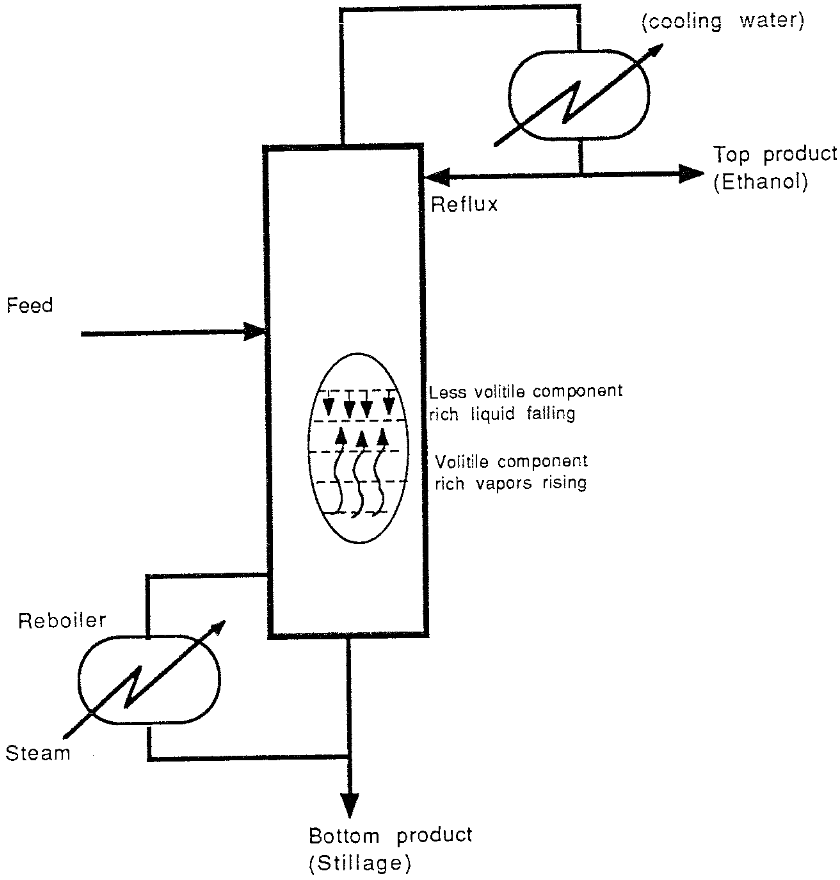
Long Tube Falling Film Evaporator (LTFF) offers increased juice velocities and heat transfer coefficients over the STRF evaporator. From reference [13].

Figure 2.10: Diagram of Biostil Fermentation Process.



Source: [18]

Figure 2.11: Diagram of a Distillation Column.



Chapter 3:
Synthesizing Low Steam Use Autonomous Distilleries and Sugar Factories/Annexed Distilleries

1.0 Introduction and Background

The co-production of electricity with ethanol or sugar changes the way in which bagasse and barbojo, the residues of sugar cane processing, are viewed. When large amounts of cogenerated electricity can be sold, bagasse and barbojo become valuable and identifying strategies for their efficient use is of interest.

In particular, strategies for making more efficient use of the cogenerated steam in the sugar factory or distillery. Steam consumed in either ethanol or sugar production cannot be used to generate electricity. Also, when electrically efficient BIG/(I)STIG technologies are employed, limits on process steam use in sugar factories and distilleries come into consideration[a]. If these cogeneration technologies are to be available for electricity-ethanol/sugar co-production, the process steam use

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[a] With BIG/STIG cogeneration, only 305 kilograms steam per tonne cane (kg/tc) is available for process use; with BIG/ISTIG only 235 kg/tc (Figure 2.5).

must be reduced from the present industry practice of ~350-500 kg/tc.

In this chapter, low steam use autonomous distilleries and sugar factories are synthesized, and the value of the steam savings is balanced against the additional costs of implementing the steam-conserving configurations. The economics of the most promising factory configurations in association with co-production are then analyzed in the following chapter.

Two steam economy improvements are considered. The first source is the use of the energy efficient process equipment discussed at length in the previous chapter. The second involves "heat integration": the configuring of heat exchangers to efficiently recover "waste" heat. As will be shown, sugar factory steam demand can be reduced below that of typical facilities by 100-150 kg/tc using heat integration alone (~70% of the overall steam savings), without introducing any low steam use technologies. When improved process equipment and integration are combined, steam use can be reduced 5% to 15% below that of "low steam use" sugar factories found in the literature.

For the autonomous distillery case, all of the steam economy improvements derived come from heat integration. The critical new aspect in the distillery analysis is the integration of the juice concentration section with the distillation section. The low steam use distilleries synthesized here consume 120 kg/tc to 235 kg/tc. This is 300-350 kg/tc less steam (75% reduction) than typical autonomous distilleries and some cases almost half as much as existing "low steam use" distilleries.

In most cases, the distillery and sugar factory configurations can achieve the steam economy constraints set by BIG/STIG and BIG/ISTIG cogeneration systems without increasing capital costs above standard sugar factories or distilleries.

In this analysis, heat integration is accomplished by first determining the minimum energy required to operate the distillery or sugar factory ("energy targeting") and then synthesizing a heat exchanger network which meets the minimum energy target. The "pinch" method of B.D. Linnhoff [1] is used to derive the minimum energy requirements, while the mixed integer linear programming method of Papoulias and Grossmann [2] is used to synthesize the heat exchanger network. The combination of these two techniques provides both physical insight into the integration process and a rigorous analytical framework.

2.0 Optimization Methodology

2.1 The Pinch Point Method for Energy Targeting

The pinch method is used to calculate the minimum energy required to perform a set of thermo-chemical processes by optimally matching the heating requirements of streams which need to be heated with the cooling requirements of streams which need to be cooled. The beauty of the pinch method is its analytical simplicity. Unlike mathematical programming methods for predicting minimum energy use, it does not require specialized software, or for that matter, even a computer. It presents a clear global picture of what is happening (energetically) in a set of thermo-chemical processes.

Appreciating the results in Section 3.0 requires a basic understanding of how a pinch analysis is performed. The rest of this section presents a brief overview of pinch analysis; a more detailed discussion is found in Appendix I.

Composite Curves

The first step in performing a pinch analysis is identifying and characterizing the streams involved in the processes being investigated: their source (initial) temperatures, target (final) temperatures and specific enthalpies (C_pxMass flow=CP). All the hot streams (streams from which heat can be extracted) and all the cold streams (streams which require heating) are combined, respectively, into one composite hot stream and one composite cold stream (Figure 3.1). The composite hot and cold streams reflect all of the heating and cooling which must be performed and the temperatures at which they occur.

The two composite curves are plotted with temperature as a function of total enthalpy (Figure 3.2). The hot stream composite curve begins (arbitrarily) with zero total enthalpy at the lowest temperature to which any stream must be cooled. It rises at rates determined by the sums of the specific enthalpies and temperature ranges of the hot streams involved. (This construction is opposite to what is physically happening, where the flows are releasing heat down to the lowest temperature rather than beginning there.)

The position of the cold composite curve on the plot depends on the minimum approach temperature specified in any heat exchanger. The two composite curves must never be any closer together (vertically) than a minimum approach temperature, ΔT_{MIN}. This can be conceptually constructed by starting with the cold composite stream far to the right and sliding it to the left until the point where the two composite curves differ by the prescribed ΔT_{MIN}. The location where the curves are at their closest is the "pinch point."

A number of observations can be made about this plot of the composite curves. Where one curve is above the other, heat can be exchanged between streams and heat integration is possible; otherwise, external heating or cooling is required. The difference in total enthalpies (horizontal distance) between the right end points of the composite curves is the minimum amount of heat which must be added from external sources. The difference in total enthalpies between the left end points of the composite curves is the required external cooling.

One can now easily see the effect of changing ΔT_{MIN}. As the minimum approach temperature increases, the two curves must be pulled further apart, increasing both the required heating and cooling. Conversely, as the minimum temperature difference decreases, the curves can be brought closer together, reducing both heating and cooling loads.

The pinch divides the problem into two regions. The region above the pinch acts as a heat sink: not all of the heat load can be met using the hot streams being cooled, thus external heat must be added. The region below the pinch acts as a heat source: not all of the cooling required can be met by heating the cold streams, thus a heat sink must be introduced. If more heat than is required is added to the region above the pinch, it would be transferred across the pinch into the lower region, which would then reject it into the external heat sink. The net effect would be the transfer of heat from the heat source to the cold sink! Therefore, a hot stream above the pinch should not exchange heat with a cold stream below the pinch. This is known as transferring heat "across" the pinch and implies additional energy being both added and rejected. While this rule has no bearing on the minimum heating and cooling requirements, it is an element often used in the formulation of the heat exchanger network.

Grand Composite Curve

Another way of looking at a system is the "grand composite curve." The grand composite curve is in essence the horizontal (enthalpy) difference between the cold and hot composite curves (Figure 3.3). It is expressed with temperature on the y-axis and ΔTotal Enthalpy on the x-axis. Segments of the curve with a positive slope indicate net heating is required in that temperature interval (the hot composite curve is steeper that the cold composite curve); segments with a negative slope indicate heat must be removed in that temperature interval (the cold composite curve is steeper than the hot). Where a negatively sloped section vertically is above a positively sloped section, the potential for heat integration exists. If a positively sloped section has no negatively sloped section over it, then its heat demand must be met by a hot utility (e.g. the top of the curve). Negatively sloped sections of the grand composite curve under which no heat requiring segments exist must reject heat to a cold utility (the bottom of the curve). Pinch point(s) occur where the curve touches the *y*-axis.

In order for the grand composite curve to work out correctly, the minimum approach temperature must be accounted for when the difference in the two composite curves is taken. This is accomplished in this analysis by adding the minimum approach temperature to the hot composite curve. One can visualize this by adding the ΔT_{MIN}'s to the hot composite curve on plot of the two composite curves. The two composite curves would then just touch wherever there was a pinch point, corresponding to the grand composite curve having ΔTotal Enthalpy equalling zero. (See Appendix II for details on grand composite curves.)

2.2 Example Pinch Point Analysis and Design

The pinch process is illustrated with the following examples. Table 3.1a shows the stream data for a sugar factory with quadruple effect falling-film evaporators, continuous vacuum pans and condensate juice heating. This factory design assumes that the area in each effect should be similar. Table 3.1b presents the stream data for a sugar factory with equipment operating at the same temperature conditions, except that the effects are sized so that steam use is minimized.

In both tables, each stream is characterized by mass flow, specific heat, source and target temperatures, both actual and adjusted. For phase change entries, an artificial temperature difference is assumed: for evaporating streams, the temperature difference is assumed to be between T_{actual}– 0.01°C and T_{actual}; condensation is assumed to occur between T_{actual}+ 0.01°C and T_{actual}. While assuming that both condensation and evaporation occur in the same temperature interval would be closer to correct, this method creates a more meaningful grand composite curve. If the condensing vapors in an evaporator effect were assumed to occur in the same temperature interval as the evaporating liquid, then the net result would be the two streams cancelling each other out. The magnitude of the heats involved in these phase changes relative to the sensible heating would be completely lost. By forcing the evaporation and condensation occurring in the same evaporator effect into different temperature intervals, the phase change "spikes" in the grand composite curves the problem table are emphasized (Figure 3.4).

The "effect" and "effect vapor" entries in Tables 3.1a and b illustrate the way in which the latent heat is accounted. The "C_P" entries are actually the latent heat multiplied by ten. Thus when this "C_P" is multiplied by the imposed 0.1°C ΔT, the correct latent heat results.

Figures 3.4a and 3.4b show the grand composite curves for the two sugar factories. Figure 3.4a shows clearly a single pinch, occurring between the vacuum pans and the fourth evaporator effect. In this case, steam must be used in both the first effect and the vacuum pans. In the grand composite curve of the low energy case, pinches occur following the first effect, second effect, vacuum pans and fourth effect. The effects in the evaporator are sized such that the vacuum pans can be operated using vapors evaporated in the second effect rather than steam. *By maximizing the number of pinches, the total energy requirement is minimized.* While this increases the heat transfer area and the steam use in the first effect, it is more than compensated for by the reduced steam load of the vacuum pans.

The composite curves for the two cases are shown in Figures 3.5a and 3.5b. These curves can supply little new information which has not already been gleaned from the grand composite curve, but a few features should be noted. The most obvious feature is the "vacuum pan well": the drop in the cold composite curve corresponding to the heat load of the vacuum pans. This well is required because of the large heat exchanger approach temperature in the vacuum pans. If the vacuum pan portion of the composite curve is placed such that the composite curve

Table 3.1a: Stream Data for Sugar Factory with Quad. Effect, Falling Film Evaporators, Continuous Vacuum Pans and Condensate Juice Heating. All evaporator effects have similar heat transfer areas.

QUAD EFFECT
FALLING FILM EVAPORATORS
VAC PANS AT 100 KG/TC

TOTAL STEAM USE:
319.7241193 KG/tc

steam temp: 135

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	HEAT FLOW WATTS	EVAPORATOR AREA, SQ M
RAW CANE JUICE	13.6	157,000	3.84	34	102	41,013,639		
CLEAR JUICE	13.6	157,000	3.84	90	125	21,109,961		
EFFECT 1	17.50	35,000	21,900	124.9	125	76,650,000	21,291,667	665
EFFECT 2	23.21	30,000	22,160	114.9	115	66,480,000	18,466,667	616
EFFECT 3	34.44	30,000	22,710	96.9	97	68,130,000	18,925,000	478
EFFECT 4	62	27,561	23,210	74.9	75	63,969,755	17,769,376	538
VACUUM PAN		MCp = 387625000		84.9	85	38,762,500		
EFFECT 1 VAPORS		35,000	21,900	125.1	125	-76,650,000		
EFFECT 2 VAPORS		30,000	22,160	115.1	115	-66,480,000		
EFFECT 3 VAPORS		30,000	22,710	97.1	97	-68,130,000		
EFFECT 4 VAPORS		27,561	0	75.1	75	0		
SUMMED CONDENSATE		150,952	4.18	121.15	40	-51,206,729		

Tables 3.1b: Stream Data for Sugar Factory with Quad. Effect, Falling Film Evaporators, Continuous Vacuum Pans and Condensate Juice heating. Evaporator effects sized so to minimize steam consumption.

QUAD EFFECT
FALLING FILM EVAPORATORS
VAC PANS AT 100 KG/TC

TOTAL STEAM USE:
260.5720728 KG

steam temp: 135

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	HEAT FLOW WATTS	EVAPORATOR AREA, SQ M
RAW CANE JUICE	13.6	157,000	3.84	34	102	41,013,639		
CLEAR JUICE	13.6	157,000	3.84	90	125	21,109,961		
EFFECT 1	18.60	42,200	21,900	124.9	125	92,418,000	25,671,667	802
EFFECT 2	28.97	41,100	22,160	114.9	115	91,077,600	25,299,333	937
EFFECT 3	39.65	19,844	22,710	96.9	97	45,066,632	12,518,509	290
EFFECT 4	62	19,417	23,210	74.9	75	45,066,602	12,518,501	474
VACUUM PAN		MCp = 387625000		84.9	85	38,762,500		
EFFECT 1 VAPORS		42,200	21,900	125.1	125	-92,418,000		
EFFECT 2 VAPORS		41,100	22,160	115.1	115	-91,077,600		
EFFECT 3 VAPORS		19,844	22,710	97.1	97	-45,066,632		
EFFECT 4 VAPORS		19,417	0	75.1	75	0		
SUMMED CONDENSATE		148,745	4.18	121.57	40	-50,714,442		

*The Vacuum pan entry shows a total M*CP rather than delineating the mass flow and heat of vaporization. This is done because more steam is needed in a vacuum pan than is actual required to vaporize the water content (See chapter 2, Section 3.4). The total energy expressed here reflects the typical steam use in a batch vacuum pan, 150 kg steam ΔH_{fg} of ~2200 kJ/kg.*

increases monotonically, the hot streams directly above it to which it would be matched for heat exchange would be at a temperature less than the required ΔT above the vacuum pan temperature. When all of the temperatures of the components are adjusted for their respective minimum approach temperatures (as in the grand composite curve), everything increases monotonically.

Along the same line of thinking is the different approach temperatures required by each effect of the evaporator. The ΔT_{MIN}'s increase as one moves from the early effects to the latter effects. The temperature between the first effect vapors and the second effect evaporation is only one third the temperature difference between the third effect vapors and the fourth effect evaporation. As will be discussed in Section 3, this is due to the decrease in heat transfer coefficient with increasing sugar concentration.

2.3 Synthesizing Minimum-Unit Heat Exchanger Networks

The pinch method only determines the minimum heating and cooling loads. The mixed-integer linear programming (MILP) trans-shipment model of Papoulias and Grossmann [2] is used to derive the heat exchanger network with the minimum number of heat exchangers which achieves these minimum heating and cooling loads. Such a solution derives the simplest heat exchanger network, but does not guarantee the minimum investment cost. Since the cost of a heat exchanger network goes with both heat exchanger area and number of units, a mixed integer nonlinear programming technique would have to be employed to minimize overall investment cost. While such techniques are being developed, they are beyond the scope of this thesis. Therefore, the MILP model determining the minimum number of heat exchanger units is used, with the implied assumption that the overall cost of such a network would be near the minimum.

2.4 Other Techniques for Energy and Cost Minimization

Numerous techniques have be proposed and developed over the past ten years to derive low energy heat exchanger networks [1,2,3,4,5], ranging from heuristic algorithms, relying on rules of thumb derived from engineering experience up through rigorous mathematical programming techniques.

Energy targeting such as is accomplished by the pinch analysis can be accomplished through trans-shipment linear programming (LP) models [2]. Such models offer an advantage over the pinch method in their ability to handle multiple heat sources and cold sinks at varying costs. However, since the distillery and sugar factory cases analyzed required only one level of hot and cold sinks, this advantage becomes moot.

The pinch method has an advantage over the LP models because of its simplicity of implementation and ability to immediately access the results graphically. The physical insight gained by examining the composite and grand composite curves allows the user a degree of judgement in designing a system. The "feel" of a system acquired by manipulating the pinches, effect loadings, and distillation column balances cannot be experienced using programming methods.

Techniques beyond the MILP trans-shipment model used here exist for explicitly deriving the heat exchanger network and minimizing the overall investment cost. Floudas, Cirac and Grossmann [5] derived a non-linear programming (NLP) technique which uses the minimum utility cost and minimum number of matches predicted using the LP and MILP mentioned earlier and synthesizes the heat exchanger network which minimizes the overall investment cost. By linking the minimum utility LP model to the MILP minimum matches model to the NLP minimum investment cost heat exchanger network model through driver programs (such as Floudas, et.al.'s MAGNETS program), the user need only input the stream definition and heat transfer data and the program will calculate the heat exchanger network achieving the minimum utility and investment costs.

Similar programming algorithms exist to synthesize networks which contain variable and uncertain flows and temperatures [6], optimal retrofits [7] and overall (lifecycle) cost minimization (in contrast to first minimizing utility costs and then minimizing investment cost)[8]. These techniques involve increasingly complex programming techniques (mixed integer non-linear programming, for instance), require more precise cost and heat transfer data than are available for sugar factories/distilleries and involve extensive computational time and effort which were not appropriate for the analyses addressed in this thesis.

3.0 Autonomous Distillery Design

With the tools outlined in Section 2, the problem of designing a minimum steam use autonomous distillery is addressed. This section describes the approaches and assumptions made in performing the analysis along with the results of a pinch point and heat exchanger network configuration study.

An autonomous distillery can be broken down into three major steam consuming sections: the distillation section, where the dilute aqueous ethanol solution is concentrated; the cane juice pre-treatment section, where the juice is heated to approximately 100°C and clarified (to remove the cane waxes and other particulate impurities and to kill all of the wild, non-ethanol producing bacteria and yeasts which might infect the fermenters and drastically reduce yields); and the evaporation/concentration section, where the treated juice is concentrated up to the level required for economical fermentation. Assuming standard, non-steam conserving practices, the energy use for each of these tasks is [9]:

3.3 kg(.25 MPa steam)/liter, distillation
1.6 kg/liter, juice heating/pasteurizing
0.8 kg/liter, juice concentrating/evaporation

As will be shown, large reductions in energy use can be achieved at minimal increased capital investment by integrating the three sections, carefully choosing their operating conditions and optimally sizing the components.

3.1 Pinch Analysis: Approach and Implementation

The autonomous distillery analysis is approached by creating a database of components (evaporator effects and distillation columns) and selecting from these components when developing factory designs using pinch analysis. The database consists of six evaporator effects operating at a range of temperatures and three distillation columns operating at different pressures. Component particulars are discussed in Section 3.1.

All of the components are integrated into one spreadsheet, including the operating conditions of each component, the thermodynamic and mass characteristics of each possible distillery stream (including steam and condensates), a "problem table"[b] containing all possible temperature intervals and a section which composes the hot and cold composite curves. A particular factory configuration is implemented by setting the mass flows of the unused components to zero. The mass flows of the remaining components are then adjusted to meet the desired steam demand and component balances. This often means maximizing the number of pinch points and thereby minimizing overall steam demand.

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[b] The problem table is the numerical version of the grand composite curve (See Appendix II for details).

3.1.1 Equipment and Operating Assumptions

The conditions of the feedstock, final and intermediate products are based on typical values seen in operating distilleries and sugar factories. 1000 kilograms of raw juice per tonne cane at 13.5°Brix, 30°C is assumed to come from the mills. 144 kilograms of fermentable sugar is assumed per tonne cane, 84% of which is recovered [10]. 91% of the theoretical Guy-Lassac ethanol yield is achieved in fermentation [13]. Distillation recovers 99.9% of alcohol in feed [11]. The final product is 94% ethanol, 6% water by weight (hydrous alcohol). The net alcohol yield is 70.45 liters of hydrous alcohol per tonne cane milled, a typical to conservative ethanol yield.

Evaporator Effects Operating Conditions

A typical sugar factory evaporation scheme consists of three to five effects, concentrating the cane juice from approximately 13°Brix up to 65°Brix [12]. Although the autonomous distillery only requires concentration up to 20°Brix, 45% of the water must still be evaporated and therefore a multi-effect evaporation system is required for distilleries to achieve low steam use.

This analysis assumes up to six evaporator effects, providing the range of temperatures found in sugar evaporation systems [11]. The highest pressure effect operates at 120°C (juice temperature), with an effect operating at ten degree intervals down to 80°C. The sixth effect operates at 68°C. The sixth effect operates at 68°C rather than 70°C so that the condenser from the atmospheric distillation column could serve as a condensing vapor source. With an assumed ten degree minimum heat exchanger approach temperature, evaporation in each effect can be driven by vapors from the previous effect. A summary of evaporator and distillation equipment and operating conditions can be found in Table 3.2 and Figure 3.7.

The approach temperature in each effect is assumed to be constant at ten degrees. In sugar factories, approach temperature increases with increasing Brix to account for the marked decrease in heat transfer coefficients with increased solids: the heat transfer coefficient drops from approximately 3000 $W/m^{2\circ}C$ in the first effect down to less than 1000 $W/m^{2\circ}C$ by the last effect (Figure 3.6) [12]. Because in an autonomous distillery the juice is concentrated only to 20°Brix, the evaporator heat transfer coefficient does not drop significantly, and thereby does not require the increased approach temperatures normally found in lower temperature sugar factory evaporator effects.

This relatively small increase in Brix also affects the choice of type of evaporator used in an autonomous distillery. Earlier CEES research suggested that the use of falling-film evaporators in sugar factories would conserve steam by their increased heat transfer coefficients and their ability to operate at higher temperatures than the standard Roberts' short tube rising film evaporators [13]. The concentration required in the autonomous distillery can be performed equally as well by either of the systems. At low Brix (less than 20°), both types of evaporators have nearly the same heat transfer coefficient; the falling-film evaporator's advantage only comes into play above 30°Brix.

The falling-film evaporator's ability to operate at higher temperatures is due to the juice being exposed to higher temperatures for much shorter periods of time. The sucrose in sugar solutions will invert and caramelize if exposed to high temperatures (above about 120°C) for any significant time, darkening the color and decreasing its value. This color change is the limiting factor in sugar evaporation design [12]. However, if the sugars in the cane juice are to be fermented, sugar inversion and coloration is no longer an issue. Hence, the longer residence times of the rising film evaporators are no longer a liability at the higher temperatures.[c]

Distillation Operation Conditions

Three operating pressures for the distillation columns are used for the analysis: 6.0 atm, 4.25 atm and 1.0 atm. The pressures were chosen for the convenient temperatures of their reboilers and condensers. The condenser of the high pressure column operates at 132°C, allowing it to serve as the heat source for the highest evaporator effect. The condenser of the medium pressure column operates at 110°C, allowing it to serve as the heat source of the low pressure column reboiler or the third effect of the evaporator, all of which operate at 100°C. The condenser of the low pressure column operates at 78°C, which can either preheat feeds or drive the sixth evaporator effect.

Ethanol distillation columns generally do not operate at pressures higher than 8 bars (0.8 MPa) or lower than 1 bar (0.1 MPa) [14,15]. As the system pressure increases, so does the required strength and cost of the distillation equipment; above this regime it no longer becomes economical, even at high energy costs, to run columns at higher pressure [15]. Azeotropes also tend to be pressure sensitive. As pressure increases, the water content of the azeotropic mixture increases, lowering the ethanol content of the hydrous ethanol product.

For distillation under a vacuum, more constraints come into play. As the pressure drops, the specific volume of the vapor in the column increases, thereby requiring larger column volumes and more material. Also, when a column is operated at lower pressures the condenser temperature is reduced (<50°C), potentially requiring a refrigeration cycle to provide a low temperature sink [15].

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[c] Higher temperatures than those being considered can lead to excessive tube fouling.

Other column pressures could have been chosen, but no more were necessary. Given the operating pressure constraints, a three effect distillation sequence could not be achieved (i.e. a column whose condenser could drive the reboiler of the 4.25 atm column); consequently, a higher pressure column could not be justified. More importantly, except in all the very low energy configurations, the steam use is limited by evaporation (Section 3.1.3). Columns could be hypothesized whose condensers could drive, perhaps, the second or fourth effect, but these would offer no advantage in steam economy and merely present nearly redundant configurations.

3.1.2 Pinch Point Results

The results of nine pinch point analyses are presented in Table 3.3. Schematics of the distillation and evaporation equipment for each of the nine cases

Table 3.2: Evaporator and Distillation Column Operating Temperatures.

Component	Section	Operating Temperature
Effect 1		120°C
Effect 2		110°C
Effect 3		100°C
Effect 4		90°C
Effect 5		80°C
Effect 6		68°C
HP distillation Column		
	Reboiler	158°C
	Feed	150°C
	Condenser	132°C
MP distillation Column		
	Reboiler	135°C
	Feed	125°C
	Condenser	110°C
LP distillation Column		
	Reboiler	100°C
	Feed	90°C
	Condenser	78°C

are shown in Figure 3.7. The composite and grand composite curves for these designs are seen in Figures 3.8 through 3.17 and the problem tables and flow data are in Appendix I. These cases encompass the best scenarios at each steam pressure level along with some cases where the minimum steam use criterion was relaxed. The relaxed cases were selected to observe the effect of increasing steam use on heat exchanger configuration complexity, overall surface area and cost.

The steam savings in the autonomous distillery designs come exclusively from heat integration. The use of multi-effect style distillation sequences and the use of hot condensate for juice heating are considered forms of heat integration (making the best use of heat liberated in the condenser and sensible waste heat in from the evaporator condensates).

The conventional low steam use case where the evaporation section is separately integrated from the distillation section is "MP-LP + 2345" (see note a Table 3.3). This case assumes thermally integrated medium pressure and low pressure distillation columns with a four effect evaporation section, starting with "effect 2" (110°C) and going down to "effect 5" (80°C). The total steam use (without parasitic loads and losses) for this case is 216 kg steam per tonne cane milled, over 200 kg steam per tonne cane more efficient than a typical autonomous distillery.

The steam use of this case is similar to the steam conserving designs found in some Brazilian autonomous distilleries. The differences between this design and the reference one are that the Brazilian case uses a quintuple effect evaporator for juice concentration and that its distillation section is not operated at multiple pressures. The quintuple effect evaporation of the Brazilian case uses slgihtly less steam than the quadruple effect evaporation of the MP-LP + 2345 case, while the distillation section of the Brazilian case uses ~38 kg/tc more steam than the distillation section of the MP-LP + 2345 case.

Table 3.3: Steam Demands of Various Factory Configurations as Predicted Using Pinch Analyses.

Configuration[a]	Steam demand kg steam/TC[b]	Steam Condition[c]	Limiting Process[d]
New Configurations:			
MP-LP	119	HP	n/a
HP-12345-LP-6	136	HP	Evaporation
HP-12345-LP-6[e]	140	HP	Evaporation
HP-1234	155	HP	Distillation
MP-345-LP-6	157	MP	Evaporation
MP-3456	161	MP	Distillation
MP-345	174	MP	Evaporation
MP-LP + 2345	216	MP + LP	n/a
2-LP-6	218	LP	Distillation
1-2-LP	236	LP	Distillation
Existing Configurations[f]			
234 + LP (Typical existing)	465	~LP	n/a
12345 + LP (Existing Brazilian steam conserving)	258	~LP	n/a

Notes for Table 3.3:

[a] Configuration nomenclature:
HP: Distillation column operating at **H**igh **P**ressure (6.25 atm).
MP: Distillation column operating at **M**edium **P**ressure (4.25 atm)
LP: Distillation column operating at **L**ow **P**ressure (1.00 atm)
1 : Evaporation effect **1**, operating at 120°C
2 : Evaporation effect **2**, operating at 110°C
3 : Evaporation effect **3**, operating at 100°C
4 : Evaporation effect **4**, operating at 90°C
5 : Evaporation effect **5**, operating at 80°C
6 : Evaporation effect **6**, operating at 68°C

[b] No losses, no steam to deaerator.

[c] hp = 0.60 MPa, 178°C, mp = 0.425 MPa, 155°C,lp = 0.1 MPa, 130°C.

[d] Process which forbids further steam economy. For instance, if "Evaporation" appears in this column, adding another distillation column wound not reduce steam demand, but adding another evaporator effect would.

[e] The load in the evaporation section of this configuration has been adjusted to simplify the heat exchanger network.

[f] [9]

Two fully heat integrated cases use more steam than the MP-LP + 2345 case. Both of these cases involve using a single low pressure column with two evaporation effects. The "2-LP-6" configuration requires 218 kg steam per tonne cane milled while the 1-2-LP case has the highest steam consumption of any of the heat integrated configurations analyzed at 236 kg/tc.

Two cases using the medium pressure distillation column were analyzed: MP-345 and MP-3456. These cases correspond to the condenser of the 4.25 atmosphere column driving the 110°C evaporator effect, which would be followed by either two or three more evaporation effects. The steam demand for these two cases were 174 kg/tc for the MP-345 case and 161 kg/tc for the MP-3456 case. Adding the atmospheric (LP) column to the MP-3456 case does little to improve the steam economy, dropping the steam use to only 157 kg/tc.

The simplest system considered using the HP column was the HP-1234 case. This case uses 155 kg/tc, which is less than any of the MP column cases. Because the pinch occurs at the reboiler of the distillation column and none of the evaporator effects are near pinching, the evaporator effects could be sized for simplification of the heat exchanger network. The second and third evaporator effects are sized such that vapors from those effects exactly match the heating loads of the third and fourth effects, respectively. This implies that vapor is not bled from either of these effects for juice heating, thereby simplifying the heat exchanger network (Figure 3.21). If the HP column did not already define the minimum steam requirement for the system, these simplifications would come at the price of increased steam demand. While this purposely moves away from the maximum number of pinches, because the steam use is set elsewhere in the system, these conveniences can be implemented with no penalty in steam economy.

Two cases were considered using both the HP and the LP distillation columns and all six evaporator effects. HP-12345-LP-6(a) maximizes the number of pinches (see grand composite curve Figure 3.9b), and therefore minimizes the total steam demand. The second case, HP-12345-LP-6(b), assumes the vapors from the third and fourth effects exactly meet the evaporation loads of the fourth and fifth effects, respectively (Figure 3.10b). Unlike the HP-1234 case where such manipulation of the evaporator effect loads has no effect on the overall steam demand, in this case a price is paid for the network simplification. However, this price is not too great, with the maximum pinch case (a) using 136 kg/tc and the simplified network case (b) using only 140 kg/tc.

Lastly, the most steam efficient autonomous distillery case analyzed is the one where the feed is sent to the fermenter dilute, without any pre-concentration. In this case, the MP and LP columns are assumed (MP-LP case). While there are a number of operating and capital cost tradeoffs associated with the non-preconcentrated fermenter feed setup (particularly the greatly increased capital costs of the fermentation section, discussed in Section 3.3), the very low steam use of 119 kg/tc is over 12.5% lower than the next lowest steam use configuration. Without the juice concentration section, its heat exchanger network is also the simplest.

The Limiting Process

The idea of a "limiting process" can be used to understand most of the trends observed in the preceding section and in Table 3.4. The limiting process is the one (either evaporation or distillation in our case) which excludes the possibility of further steam economy through heat integration.

This is best understood by looking at cases MP-345 and MP-3456. When only three evaporator effects were used, pinches occurred at the three evaporator effects, while the distillation column "spike" at 120°C is still around 30 MJ/tc away from the *y*-axis on the grand composite curve (Figure 3.14b). This case is evaporator limited, because the portion of the grand composite curve involving the distillation column is not near pinching. Neither changing the design of the distillation column nor adding another distillation column operating at another pressure would reduce steam demand.

When the additional evaporator effect is added, the pinch transfers to the distillation column condenser (Figure 3.13b), and distillation becomes the limiting process. Adding another evaporation effect would pull the evaporators further away from pinching, but would not reduce steam use. The only way which steam use can be reduced is if an LP column is added so that some of the load from the MP distillation column could be shifted and the reboiler "spike" reduced.

With the exception of the case where the juice was not pre-concentrated (MP-LP), the higher the steam pressure used, the lower the overall steam demand. This trend is due in part to the ability to incorporate more evaporator effects, but more important is the operating temperature of the limiting process. This can be seen by comparing the grand composite curves of a high pressure case (HP-1234, Figure 3.11b) and a low pressure case (2-LP-6, Figure 3.16b). In both cases, distillation is the limiting process; the critical difference is the temperature at which this limiting process occurs relative to the rest of the system. In the low pressure case, sensible heating occurs within the same temperature regime as the limiting process. In the grand composite curve for this case, net sensible heating can be seen occurring between 110°C and 120°C, between the evaporation in the first effect and the condensation of those same vapors (grand composite curves use adjusted temperatures, and evaporation and condensation of the same vapors occur 10°C apart). This sensible heating must be provided by steam, as there is no negative sloping section of the grand composite curve from which heat can be drawn.

More net sensible heating occurs between the condensation of the first effect vapors and the condenser of the LP column at 78°C, which must be met by the condensing vapors out of the second effect. This sensible heating immediately prior to the limiting process prevents the reboiler of the LP column from approaching the y axis more closely. If less vapor from effect 2 is required for juice heating, the load of that effect could be reduced up to the point that all three points on that grand composite curve were pinches (effect 2, LP distillation, effect 6), lowering the overall steam requirement. More effects could be added, but they could not decrease total steam demand because the pinch of the limiting process LP column condenser prevents further economy.

This case is contrasted with Figure 3.11b, the grand composite curve for HP-1234. Because no sensible heating occurs at temperatures higher than the HP reboiler, all of the steam can go towards evaporation. More importantly, the limiting process pinch can occur at the reboiler. Because the limiting process, distillation, occurs at temperatures above the net sensible heating load, the overall steam demand of this configuration is not a function of how the juice heating is accomplished but rather by the design of the limiting process. The limit in steam economy consequently is set by the heat demand of the limiting process (reboiler heat load in our example) rather than by the heat load of the limiting process plus some amount of sensible juice heating.

In cases where evaporation is the limiting process, there is no way of avoiding the interaction of the sensible heating load with the evaporator effects. The juice must be heated up to the temperature of the highest pressure effect, which is occurring in the same temperature range as all of the evaporator effects.

Evaporation is the limiting process in all cases with multiple distillation columns. In these cases, the bulk of the distillation load is carried by the higher pressure column, with the lower pressure column handling anywhere from less than 2% of the distillation load (MP-345-LP-6) up to 10% of the distillation load (HP-12345-LP-6). Thus, there is only marginal benefit from switching from one higher pressure column (MP or HP) to multiple columns.

The most energy efficient system using juice concentration, HP-12345-LP-6, is evaporation limited even with six effects. If the number of effects were not limited, steam use could ideally drop down to 81 kg/tc, the level of the optimized case of distillation integration with the concentrated feed (the MP-LP portion of the MP-LP + 2345 case).

In this light, it is clear why the lowest steam use case is the one where no fermenter feed concentration occurs (MP-LP). The lack of evaporation capacity no longer stands in the way of further steam economy. The increase in distillation steam requirements is more than offset by the elimination of evaporation heat needs, however the more dilute feed prevents it from reaching the concentrated feed minimum of 81 kg/tc. Nevertheless, this case is by far the most steam efficient considered in this study.

3.2 Configuring the Heat Exchanger Network

The pinch analysis provides upper bounds for the steam economy possible in autonomous distilleries. Translating those predicted steam uses into actual heat exchanger networks will be addressed in the following section.

Four autonomous distillery heat exchanger networks were derived using the MILP modeling techniques discussed in Appendix II. They include (using the same nomenclature): MP-LP (dilute feed to fermenters), HP-12345-LP-6(b), HP-1234, and MP-LP + 2345. The first three include configurations with the lowest overall steam demands. The higher steam use case for HP-12345-LP-6 was used for the simplification of modeling and heat exchanger network synthesizing. The MP-LP + 2345 case corresponds to where the evaporation with juice heating and distillation sections were optimized separately. This case is included as an upper limit of steam economy when a more standard approach is taken. The pinch analysis of the distillation section of this case predicted a steam demand of 1.3 kg per liter of ethanol, which falls within the range of other published values for heat-integrated hydrous distillation sequences [16]. The basic heat exchanger configuration and duty-split between the two columns also closely resemble published (and patented) heat integrated hydrous alcohol distillation sequences [17].

Sample GAMS input files for implementing the MILP model can be found in Appendix III. The flow diagrams derived from these outputs are shown in Figures 3.18 through 3.21.

3.3 Analysis and Comparisons of Derived Configurations

Tables 3.4 and 3.5 present rough technical and economic comparisons among the four systems configured using the MILP modeling. Heat exchanger areas were calcualated assuming clean tubes and optimal operation. Costs were calculated per unit of heat transfer area using conservative (high) price estimates.

The MP-LP case not only consumes the least steam but also has the simplest heat exchanger network (11 units). Its tradeoffs include large heat exchanger areas for the sensible feed heating, up to three times other configurations. The large heat exchanger area is due primarily to its reliance on liquid-liquid heat exchange, occurring mostly near the minimum approach temperature. The other cases rely on a significant amount of juice heating from condensing vapors, which have comparable heat transfer coefficients but much higher temperature differences.

The larger cost tradeoff is the increased fermentation and distillation capacity required with the dilute feed. The additional costs were calculated using a power scaling law:

Table 3.4: Comparison of Heat Transfer Areas.

Configuration(a)	Number of Units[b]	Evaporator/ Reboiler/Condenser Area[c]	Shell & Tube Heating Area[d]	Plate and Gasket Heating Area[e]
MP-LP	11	700m^2	2200m^2	1400m^2
HP-12345-LP-6	22	1700	2600	950
HP-1234	13	1700	1300	480
MP-LP + 2345	20	2100	1800	750

Notes for Table 3.4:

[a] See note a, Table 3.1 for configuration nomenclature.
[b] Total number of heat exchangers, evaporators, reboilers, condensers and distillation columns.
[c] Combined areas of heat exchanger equipment involving phase changes on both sides: evaporators, condensers cooled by evaporation and reboilers. See Appendix III and Section 4.
[d] Combined area of sensible heat exchangers, assuming all are shell and tube.
[e] Combined area of sensible heat exchangers, assuming all are plate and gasket.

$$\text{Cost}_{\text{SCALED}} = \text{Cost}_{\text{BASE}}\ (\text{Capacity}_{\text{SCALED}}/\text{Capacity}_{\text{BASE}})^{\alpha}$$

where α is an empirical scaling factor, generally around 0.6 [18]. The vast majority of fermenter costs are tanks and pumps [19], with an average α of 0.55 [18,20]. For distillation columns, the α is approximately 0.6 to 0.65 [18,20]. The nearly 50% increase in feed volume increases the capital cost of the fermentation equipment by approximately $700,000 and the distillation columns by approximately $300,000 over the cases where the feed is concentrated. With a discount rate of 12% and a life of 20 years, these additional installed costs translate into an additional $133,000 annual capital recovery.

As seen in Table 3.5, the least cost configuration depends upon the oppor-

Table 3.5: Comparison of Annualized Costs of Heat Transfer Equipment and Annual Energy Costs.

Configuration	Annual Cost of Heat Transfer Equipment[a] w/ P & G	w/ S & T	Annual Energy Costs[c] for Electricity at: 3.0¢/kwh	5.0¢/kwh	Σ(S & T. + 3¢/kWh Elect.) Annual costs	Σ(S & T. + 5¢/kWh Elect.) Annual costs
MP-LP	$209,000[b]	$210,000	$267,000	$444,000	$478,000	$655,000
HP-12345-LP-6	$132,000	$148,000	$314,000	$523,000	$462,000	$671,000
HP-1234	$124,000	$131,000	$347,000	$579,000	$479,000	$710,000
MP-LP + 2345	$159,000	$169,000	$484,000	$807,000	$653,000	$976,000

Notes for Table 3.5

[a] The costs include heat exchangers, evaporators, reboilers, condensers and distillation columns **only**. The "w/ P & G" column is the case were only plate and gasket exchangers were used; the "w/ S & T" column is the case were only shell and tube heat exchangers were used. The balance of the distillery is comparable for all three cases and therefore not addressed. Assumed costs were:

Evaporator cost:	$500/m^2
Shell and tube heat exchanger cost:	$100/m^2
Plate and gasket heat exchanger cost:	$150/m^2

A discount rate of 12%, along with a plant lifetime of 20 years was assumed.

[b] Includes extra cost to upscale fermentation and distillation capacity to handle the more dilute feed. (Appendix III).

[c] Steam valued at opportunity cost of electricity not generated by that steam; steam-electricity equivalence assumed to be .129 kwh/kg steam, the average value for the three cogeneration systems. Distillery rated at 4000 tonne cane per day, 160 days per year. (While actual steam opportunity costs vary from .115 (BIG/ISTIG) to .148 (CEST), they do not effect the overall ranking.)

tunity cost of the steam being sent to the distillery[d]. If the steam could have produced electricity which could be sold at 3¢/kWh, then the least cost configuration is HP-12345-LP-6, with the HP-2345 and MP-LP cases about the same. When the opportunity cost of steam is 5¢/kWh, then the steam efficient MP-LP case becomes

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[d] The annual costs in Table 3.6 and in this section refer to the steam cost and capital costs of the heat transfer and distillation equipment. Because the balance of the distillery is the same for our cases, these costs were not addressed.

the most desirable. HP-12345-LP-6 and MP-LP have about the same annual costs when electricity is valued at 4¢/kWh. If higher discount rates and/or shorter lifetimes are assumed, the tradeoff electricity value would shift higher, favoring the cases where the fermenter feed is pre-concentrated.

If pre-concentration is required, the HP-12345-LP-6 and HP-1234 cases are comparable in terms of annual costs, with the HP-1234 case's reduced capital costs offsetting the HP-12335-LP-6 case's better steam economy.[e] The simplicity of the HP-1234 case, however, gives it a strong edge. Particularly in a developing country setting, straight forward operation and design simplicity should be stressed.

The more standard case for pre-concentrated feed, MP-LP + 2345, has by far the highest heat transfer equipment capital costs and energy costs. Separating the distillation from the evaporators is an operational advantage, but one not likely to outweigh its cost and energy consumption disadvantages.

In all of the cases with steam demands below approximately 200 kg/tc, the limiting factor in steam economy could become the steam turbines driving the mills. Although in a new factory these mills could be designed to operate at higher steam pressures, the larger mill design problem is the variability of the load. The cane tends to be fed into the mill in large bunches, resulting in periods of high steam demand when a clump of cane is being fed through and rather low requirements the rest of the time. This problem could be addressed by, for instance, installing variable speed electric mills at some or all of the mill stations, at a price of additional capital costs. The general problem of milling and mill turbine design is not addressed in this thesis, but would have to be considered if very low steam use configurations, such as MP-LP, are to be used.

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[e] Other than possible economic reasons, preconcentrating the fermenter feed is sometimes done in practice to allow the fermentation feedstock to be stored so that the distillation section can operate while the milling section is idle. This is done in some Brazilian distilleries, where cane is not delivered on weekends [9].

4.0 Applications in Annexed Distillery/Sugar Factory Design

The annexed distillery's steam use per liter of ethanol is not significantly different than that of the autonomous distillery, from 1.3 to 4.0 kg steam per liter. However, because most of the sucrose in the cane juice goes toward sugar production, only about 9 liters of hydrous ethanol can be produced in an annexed distillery per tonne cane milled (C molasses as the feedstock).

This makes the sugar factory by far the more energy intensive portion of the annexed distillery/sugar factory system. A typical sugar factory uses ~350 to 500 kg of medium pressure steam (.2-.3 Mpa) per tonne cane [21]; a heat integrated factory with steam conserving components (as will be shown) can operate on less than 220 kg/tc. A typical annexed distillery operates at 36 kg/tc (4 kg/l), while an efficient one can run on less than 12 kg/tc (1.3 kg/l) [16], approximately 5% of the overall steam demand in annexed distillery/sugar factory. Because the sugar factory is the over-whelming steam consumer in the annexed distillery/sugar factory system, it is the focus of the analysis; the distillery portion of the system is treated as a steam consuming black box on the side.

There are other reasons that this approach was taken. The interest in gasohol in the US during the 1970's spurred considerable research and development into low energy methods of producing anhydrous ethanol involving a number of very different approaches (see Chapter 2, Section 2). Rather than attempt to integrate the annexed distillery portion of the sugar factory/distillery, existing low steam use heat integrated anhydrous alcohol distillery designs are considered in conjunction with the low steam use sugar factories addressed in the analysis.

Another reason that the sugar factory alone was the focus of the analysis is that complete integration of a distillery/sugar factory would not yield significantly improved steam demands beyond the low steam use cases which were addressed.

The lowest steam demand of a separately optimized sugar factory and distillery would be approximately 223 kg/tc[f]. Assuming that evaporation in the sugar factory is the limiting process, the minimum steam use for the completely integrated system would be 211 kg per tonne cane. When compared to the 277 kg/tc improvement made over the standard sugar factory/annexed distillery, the extra 12 kg/tc improvement gained by complete integration comes at a rather high price in terms of computational effort and overall system flexibility.

Complete integration of a flexible product distillery (being able to stop sugar crystallization at some point prior to complete sucrose extraction and devote the remaining uncrystallized sugar to ethanol production) is yet another significant problem in itself. A distillery using A molasses as the fermentation feedstock would use almost three times the steam as one fermenting C molasses with a similar drop in vacuum pan steam use. This potentially large variation in steam demand in different sections of the sugar factory-annexed distillery system creates large difficulties in deriving an optimal heat exchanger network robust enough to handle the variations.

When the two sections are integrated separately, the operating load of one section does not affect the operation of the other. Thus, flexible product operation is feasible, with the only costs being the extra capacity of the annexed distillery and the potential (small) loss in steam economy by operating the vacuum pans at a less reduced load.

This is not to say that synthesizing a flexible annexed distillery/sugar factory is impossible (programming algorithms exist to do just that) [6], but rather that it is not appropriate given the scope of this thesis. If the economics of such a system

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[f] Assuming quintuple effect, falling-film evaporators with continuous vacuum pans and condensate juice heating in the sugar factory, the steam demand is approximately 211 kg/tc. Assuming also that the alcohol is produced at 1.4 kg/liter, with a yield of 9 liters per tonne cane.

can be justified, it would present a challenging task for some future researcher.

4.1 Pinch Analysis: Approach and Implementation

The approach taken in the pinch analysis of the sugar factory is somewhat different from that taken for the autonomous distillery. The database of potential components used in the autonomous analysis is not appropriate for the sugar factory analysis. This is because (a) all of the heat demand is for either juice heating or evaporation; and (b) the approach temperatures in the effects evaporating highly concentrated sugar syrup must be higher [13], limiting the number of potential evaporator effects. Therefore, specific evaporator configurations are proposed, with the energy conserving components and strategies implemented incrementally.

4.1.1 Sugar Factory Equipment

Selected combinations of the following standard and energy saving components and techniques are considered:

Standard sugar factory equipment:

- Quadruple effect evaporation
- Short tube rising film (Roberts) evaporators
- Batch vacuum pans

Low steam use equipment/techniques:

- Quintuple effect evaporation
- Falling-film evaporators
- Continuous vacuum pans
- Vapor condensates from the evaporator effects used for juice heating.

Details of each of these processes can be found in Chapter 2.

Of course, not all possible combinations are addressed. Generally, those with the simplest implementation and the greatest steam savings are implemented first,

Table 3.6: Evaporator Operating Temperatures and Heat Transfer Coefficients.

	EFFECT 1		EFFECT 2		EFFECT 3		EFFECT 4		EFFECT 5	
evaporator	T	U[a]	T	U	T	U	T	U	T	U
4 effect, RF	110°C	3150[b]	99.8	2400	82.9	1200	60.1	500	n/a	
4 effect, FF	125	3200	115	2800	97	2300	75	1000	n/a	
5 effect, FF	130	3200	123	2950	115	2500	102	1950	90	1000
Vacuum pans	85°C, 30°C minimum approach temperature.									

Notes for Table 3.6:

[a] Watts/m^{2}°C.

[b] No heat exchanger analysis was performed using rising film evaporators, however the values are included for comparison.

adding more marginal improvements later.

The approach temperatures chosen for the factory designs are based on existing sugar factories[g]. Due to the operating characteristics of sugar factory evaporation systems, optimizing the approach temperatures and effect areas cannot be done with much precision. The problem revolves around the extreme variability of the heat transfer coefficients, which depends on Brix, amount of tube scaling and the juice loading in a particular evaporator body. The heat transfer coefficients used in this study come from reference 13 and are for clean, optimally operated effects. While actual areas might be as much as twice that calculated here, they provide a means of comparison between the different configurations. The operating

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[g] The sugar factory size and approach temperatures of the rising film, quadruple effect cases were those used in the Monymusk Sugar factory, Jamaica. The quintuple effect, falling film approach temperatures are based on a test evaporator train at the Sugar Industry Research Institute, Bernard Lodge, Jamaica. The falling film, quadruple effect case was a synthesis of the two prior cases [13].

temperatures and heat transfer coefficients are summarized in Table 3.6 and presented graphically in Figure 3.6.

Distillery Systems Considered

Three different anhydrous alcohol distillation/separation systems are annexed to the sugar factory: heat integrated benzene distillation, using 5017 kJ/liter (2.42 kg steam/liter at .7 MPa) [22]; ethylene glycol extractive distillation, using 3286 kJ/liter (1.60 kg/liter) [14]; and distillation with molecular sieve drying at 3350 kJ/liter (1.62 kg/liter) [15]. Flow diagrams/schematics of the three systems can be seen in Figures 3.22 through 3.24.

Because all three systems being considered are heat integrated and optimized, the streams leaving each are at relatively low temperatures. Because there is already an abundance of low quality heat in all sugar factory designs, these streams were not considered for further integration with the sugar factory.

4.1.2 Pinch Analysis: Results

Pinch point analyses were performed on the following sugar factory configurations:

- 4 effect, rising film evaporators with batch vacuum pans and no condensate heating (standard sugar factory).
- 4 effect, rising film evaporators with batch pans and condensate heating;
- 4 effect, falling-film evaporators with batch pans and condensate heating;
- 4 effect, falling-film evaporators with continuous pans and condensate heating;
- 5 effect, falling-film evaporators with continuous pans and condensate heating.

The composite curves and grand composite curves for these cases are presented in Figures 3.25 through 3.31. Table 3.7 presents the net steam use of the different factory configurations. The results of the earlier CEES steam economy analysis [13] are included as a reference for low steam use sugar factories using the same evaporation effect operating conditions.

The cases shown in Table 3.7 start at the most common, least efficient sugar factory design and add the various steam economizing features as one goes down the table. The largest improvement in steam economy is the addition of the fifth effect and accompanying approach temperature reductions, which decreases steam use by 50 kg/tc (Figure 3.32). This is seen in the transition from case 5 to case 6. The next most effective steam reducing strategy is the shift from rising film to falling-film evaporators (38 kg/tc savings, case 2 to 3), which allows the evaporation section to operate at higher temperatures, resulting in hotter bled vapors and condensates. Using hot condensates for juice heating decreases the steam use by 36kg/tc (case 1 to 2), while switching from batch to continuous vacuum pans improves the steam economy by 20 kg/tc (case 3 to 4). Although the continuous vacuum pans provide the smallest improvement in steam consumption, their larger asset is their stability of steam demand. The steam consumption values assumed for batch vacuum pans are average values over many strikes, with the actual steam demand coming in large peaks and valleys.

Two analyses are performed on the last two configurations (cases 4 to 7): one minimizing steam demand, including bleeding vapor from every effect and treating condensate from each effect as separate streams (cases 4 and 6) and the other assuming vapor bleeding from only two or three effects and combining condensates from all of the effects aggregated into a single stream (cases 5 and 7). It is more efficient to treat the condensates from the evaporator effects as separate streams because the high temperature condensates are degraded when mixed with the lower temperature condensates. This increase in efficiency comes at a high price of system complexity. The simplifications add only two to three kilograms of steam per tonne cane more than the optimized versions, however they significantly reduce the complexity of the heat exchanger network.

4.2 Sugar Factory/Annexed Distillery Configurations

Full heat exchanger networks are derived using MILP modeling for the two factories with quadruple effect falling-film evaporators and continuous vacuum pans

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Table 3.7: Steam Requirements for Various Sugar Factory Configurations.

	Configuration:				Steam Use, Kg Steam per Tonne Cane[a]:		
Case	Effects	Type[b]	Condensate Heating	Vacuum Pans	Using Pinch Point	From Ogden et. al.[c]	Percent Difference
0	4	RF	No	Batch	350-500 (Standard sugar factory)		
1	4	RF	No	Batch	353	381	7.9%
2	4	RF	Yes	Batch	317	337	6.3%
3	4	FF	Yes	Batch	279	322	15.4%
4	4	FF	Yes	Continuous	259	———	N/A
5	4	FF	Yes[d]	Continuous	261	———	N/A
6	5	FF	Yes	Continuous	211	231	9.5%
7	5	FF	Yes[d]	Continuous	214	231	7.9%

Notes for Table 3.7:

[a] The steam use numbers assume no losses and no steam to deaerator. See Appendix I for more details on assumptions.

[b] RF = Rising film evaporator, presently the industry standard.
FF = Falling-film evaporator, commonly used in beet sugar processing.

[c] J.M. Ogden, et.al, "Steam Economy and Cogeneration in Cane Sugar Factories," Proceedings of the Meeting of West Indies Cane Sugar Technologists, Bridgetown, Barbados. 1988. [13]

[d] For all other cases where condensates are used, it is assumed that the condensates are not mixed prior to their use in juice heating. In these cases, along with all of those in reference 13 it is assumed the condensates from all the effects are combined before being used for juice heating.

(cases 4 and 5) and the simplified quintuple effect case (case 7). The sugar factory designs developed are shown in Figures 3.33 through 3.35, with a summary of steam use, total heat transfer areas and number of units presented in Table 3.8. A sample GAMS input files are included in Appendix III.

Figure 3.33 shows the heat exchanger network for the quadruple effect case where the condensate streams from all of the effects were kept track of independently. This case requires a sugar factory with 9 juice heaters, 23 different streams and a Rube Goldberg-like heat exchanger network.

This network contrasts with the simpler factory design using quadruple effect evaporation (Figure 3.34). In this case, all of the condensates are combined into a single stream before being used for juice heating, costing two kilograms of steam per tonne cane over the configuration in Figure 3.33, but also accomplishing the task with four fewer heat exchangers and nine fewer streams. If the raw juice need only be heated to 100°C rather than the assumed 102°C, one more heat exchanger could be eliminated.

The heat exchanger network for the combined condensate version of the quintuple effect sugar factory is shown in Figure 3.35. With five juice heaters (one of which is doubling as the condenser), the network is actually simpler than either of the quadruple effect cases. This is due to the fact that the quadruple effect evaporators were loaded such that vapor could be bled from each effect, decreasing steam use, but adding heat exchanger area and complexity.

4.3 Analysis and Comparisons of Derived Configurations

Comparisons of the evaporator effect areas, juice heating areas and estimated annual steam and capital recovery costs are shown in Tables 3.8 and 3.9. The two quadruple effect cases are similar in both heat transfer area and energy costs. The

Table 3.8: Comparison of Heat Transfer Areas for Sugar Factory Evaporation and Juice Heating Sections.

Configuration	Evaporator area[a]	Shell & Tube heating area[b]	Plate and Gasket heating area[c]
4 effect, falling-film evap., separate condensates	2700m^2	1600m^2	700m^2
4 effect, falling-film evap., combined condensates	2800	1500	670
5 effect, falling-film evap., combined condensates	3400	1200	460

Notes for Table 3.8

[a] Areas calculated using estimates in Ogden, et.al 1988 [13]. These are for non-scaled evaporators; while actual areas could be up to twice these values, they provide a means of comparison.

[b] Combined area of sensible heat exchangers, assuming all are shell and tube.

[c] Combined area of sensible heat exchangers, assuming all are plate and gasket.

sum of the energy costs and heat exchanger costs of the more complex of the two cases (separate condensate accounting) is about $13,000 less per year (~0.03¢/kg sugar). It should be considered, however, that the costs are calculated per square meter of heat transfer surface area, and that the fixed costs per heat exchanger unit and increased operation and maintenance costs of the more complex network not included in this rough analysis might easily overshadow this difference, particularly in a developing country setting.

The quintuple effect case requires 16% more evaporator area than the quadruple effect cases, but also 25% to 30% less heat transfer area for juice heating. This decrease in the area required for juice heating is due the higher temperature of the bled vapors, which increases approach temperatures and decreases heating

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Table 3.9: Comparison of Capita Recover Costs of Heat Transfer Equipment and Annual Energy Costs.

Configuration	Annual Cost of Heat Transfer Equipment[a]		Annual Energy Costs[c] for Electricity at:		Σ(S & T + 3¢/kWh Elect.) Annual Costs[d]	Σ(S & T + 5¢/kWh Elect.) Annual Costs
	S & T	P & G	3.0¢/kWh	5.0¢/kWh		
4 eff, falling-film, separate condensate	$205,000	$197,000	$591,000	$985,000	$796,000	$1,190,000
4 eff, falling-film, combined condensate	$209,000	$203,000	$595,000	$992,000	$804,000	$1,201,000
5 eff, falling-film, combined condensate	$242,000	$235,000	$488,000	$814,000	$730,000	$1,055,000

Notes for Table 3.9:

[a] The costs include heat exchangers, evaporators, reboilers, and condensers **only**. The balance of the distillery is comparable for all three cases and not addressed. Because the heat transfer area values were optimistic, higher price estimates for capital costs were assumed [13]. See Appendix III. A discount rate of 12% and a plant life of 20 years were assumed.

[c] Steam valued at opportunity cost of electricity not generated by that steam; steam-electricity equivalence assumed to be .129kwh/kg steam, the average value for the three cogeneration systems.

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areas. These two effects tend to cancel each other out, with the net capital cost of the heat transfer equipment of the quintuple effect case about 11% higher than the quadruple effect cases.

This higher capital cost, though, is greatly overshadowed by the decrease in steam costs. Even with steam valued at a low opportunity cost of the electricity not generated by that steam (electricity at $0.03/kWh), the quintuple effect case would save over $70,000 per year over the more efficient quadruple effect case, and over $530,000 per year over a standard sugar factory[h]. A matrix of the steam consumption of the combined sugar factories/annexed the steam use in each case, with little relative difference between the three alcohol distilleries is shown in Table

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[h] Assuming a 4000 tonne cane per day factory, operating at 90% capacity factor, .132 kWh per kg steam, 160 day milling season. The standard sugar factory uses 450 kg per tonne cane.

Table 3.10: Comparison of Overall Steam Demands for Sugar Factory/Annexed Distillery Combinations

Steam use, kg/tonne cane	Benzene Distillation	Ethylene Glycol Distillation	Mol. Sieve Extraction
4 effect, falling-film evap. separate condensates	281	274	275
4 effect, falling-film evap., combined condensates	283	276	277
5 effect, falling-film evap., combined condensates	236	226	227

3.10. As expected, the sugar factory dominates separation technologies. The lowest overall steam demand of 226 kg/tc comes with the combination of the quintuple effect evaporation case with ethylene glycol extractive distillation. (For more a more detailed discussion of the tradeoffs between these and other separation technologies, see Chapter 2.)

5.0 Summary and Conclusions

Large reductions on steam requirements can be made for both autonomous distilleries and sugar factories. The integration of the evaporation section with the distillation section of an autonomous distillery can dramatically reduce steam demands relative to both conventional autonomous distilleries and those with the two sections integration separately. When evaporation and distillation are integrated separately, the overall steam demand is around 220 kg(.27 MPa steam)/tc; when the evaporation and distillation sections are integrated together, the steam demand can drop to below 140 kg/tc. Higher energy efficiency still can be achieved when the juice is not concentrated at all before fermentation and distillation (~120 kg/tc).

The idea of the limiting process developed here can be a useful tool in the total integration of an autonomous distillery. By identifying which of the processes in the system (either distillation or evaporation) limits further steam economy, the designer can concentrate his effort on the appropriate section of the configuration. The limiting process concept could probably be adapted to other pinch point analyses where two or more separate processes are being integrated.

The use of efficient process technologies and heat integration can reduce sugar factory steam use to below 220 kg/tc. Meeting the steam economy levels dictated by BIG/ISTIG cogeneration requires a heat-integrated sugar factory with quintuple effect, falling-film evaporation and a heat-integrated annexed distillery.

When rigorous energy targeting and heat integration techniques are applied to the sugar factory, steam demand can be reduced by 5% to 10% below the level that results when heuristic, informal heat integration techniques are used.

In all but one of the cases analyzed, when steam is valued at the opportunity cost of the electricity it might have produced (3¢/kWh minimum), the savings associated with high steam efficiency are greater than the additional annual capital costs of efficient technology and heat exchanger networks. The exception to this is the autonomous distillery where the feed is not concentrated (case MP-LP), where electricity must be valued above ~4¢/kWh before the energy savings can overcome the additional capital costs.

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Figure 3.1: Sample Construction of a Composite Curve[1]:

Data:

Stream	M^*C_P	Source Temp.	Target Temp.
1	A	T2	T5
2	B	T1	T3
3	C	T2	T4

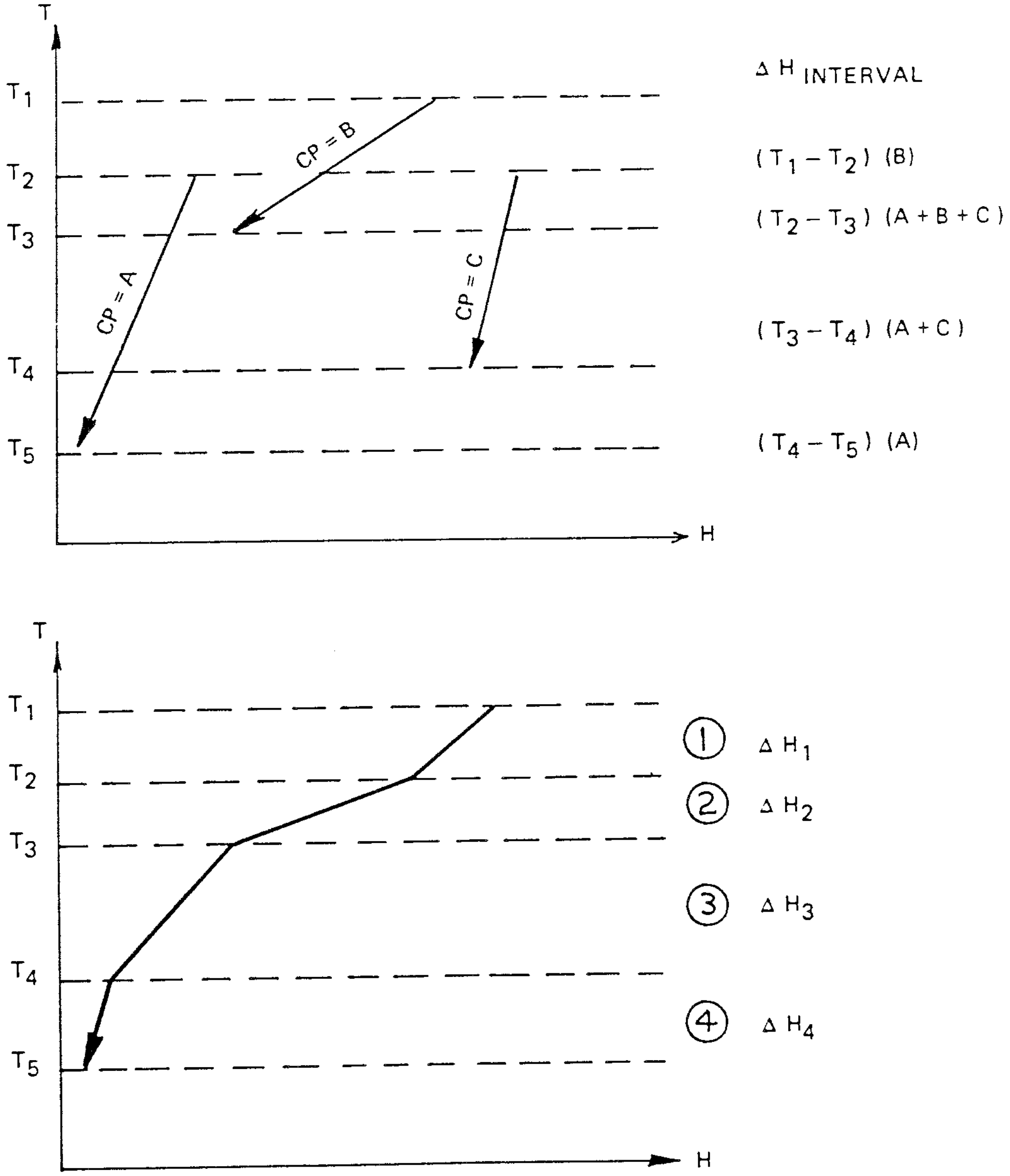


***Figure 3.1a** shows three sample streams 1, 2, 3, (hot streams being cooled) with temperature drops as a function of enthalpy. In interval T2-T3 all three streams exist, while in interval T3-T4 only streams 1 and 3 exist, etc. The far left column represents the composite ΔH of the streams in each temperature interval, showing the summing of the specific enthalpies over each interval.*

***Figure 3.1b** shows the composite curve of the three streams; at each source or target temperature the slope of the composite curve changes, reflecting the different specific enthalpy of the composite.*

Figure 3.2: Composite Hot and Cold Streams.

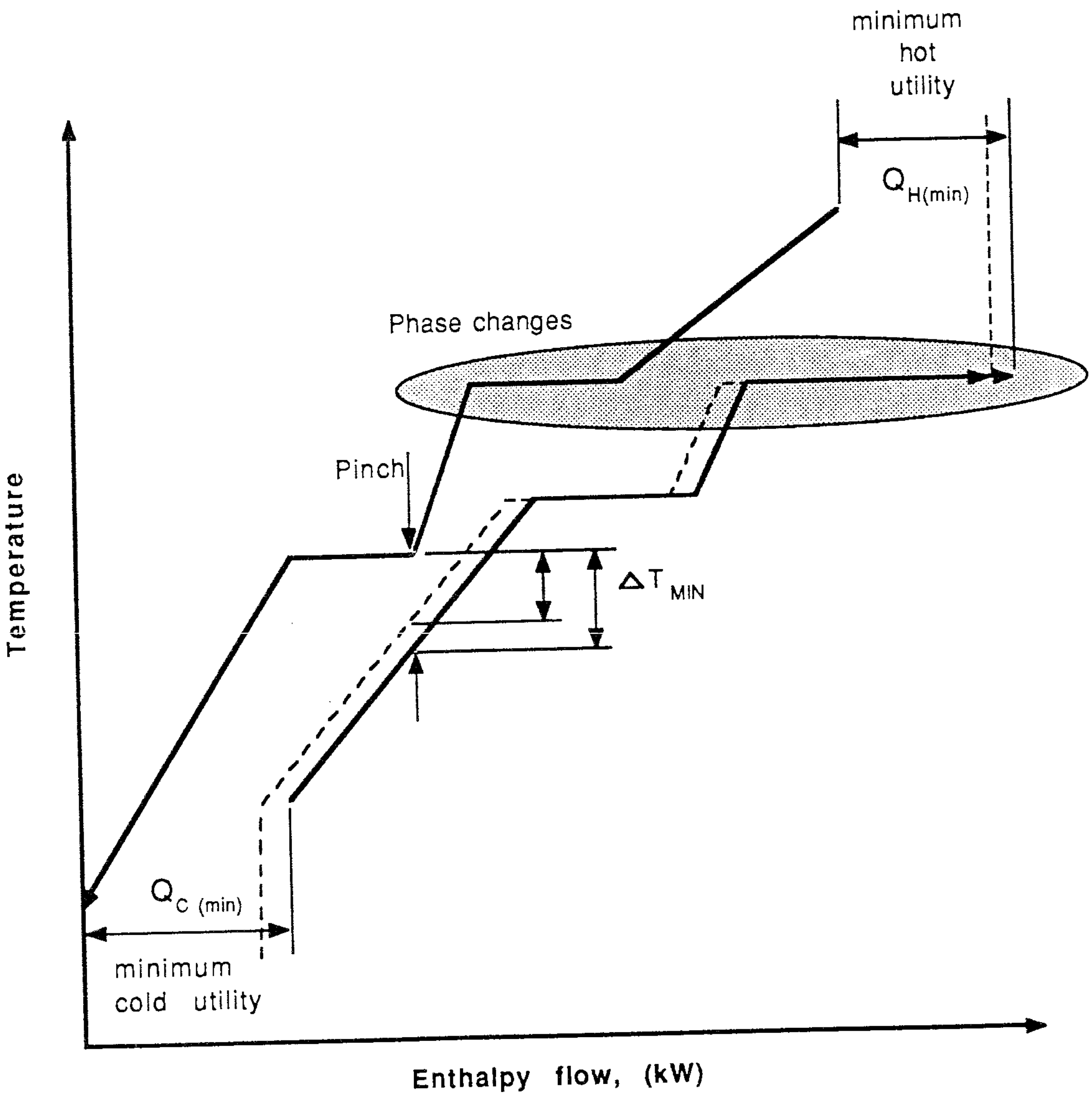


Figure 3.2 shows a plot of composite hot and cold streams for a hypothetical process. The dashed line represents the change in location of the cold composite stream if the minimum approach temperature, ΔT_{MIN}*, was reduced. The flat portions of the curve are where phase changes are occurring.*

Figure 3.3: Grand Composite Curve.

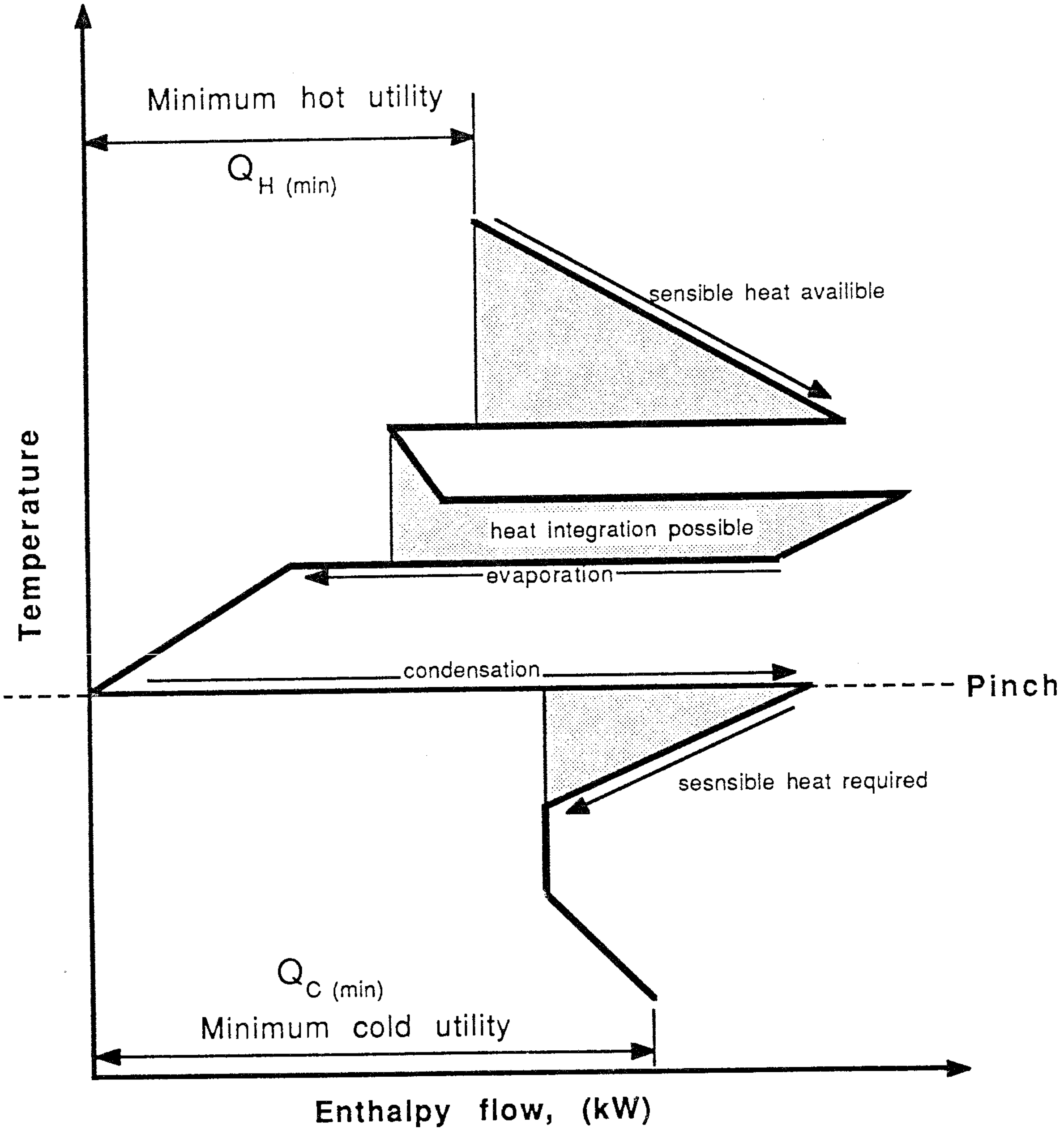


Figure 3.3 *shows a grand composite curve of the same hypothetical process. The pinch is indicated by a dashed line and occurs where the grand composite curve touches the y axis. The shaded areas indicate where process heat integration is feasible. The flat portions of the curve indicate phase changes.*

The minimum hot and cold utilities are given by the horizontal distance from the ends of the curve to the y axis. They also show at what temperature the heat must be added or rejected. For instance, most of the heat is rejected during the condensation occurs at the pinch temperature; therefore, that portion of the cold utility load might be met using a warmer cooling water than is required by the last segment of cooling.

Figure 3.4a and b: Grand Composite Curves for Sugar Factories Using Balanced and Unbalanced Multi-Effect Evaporators.

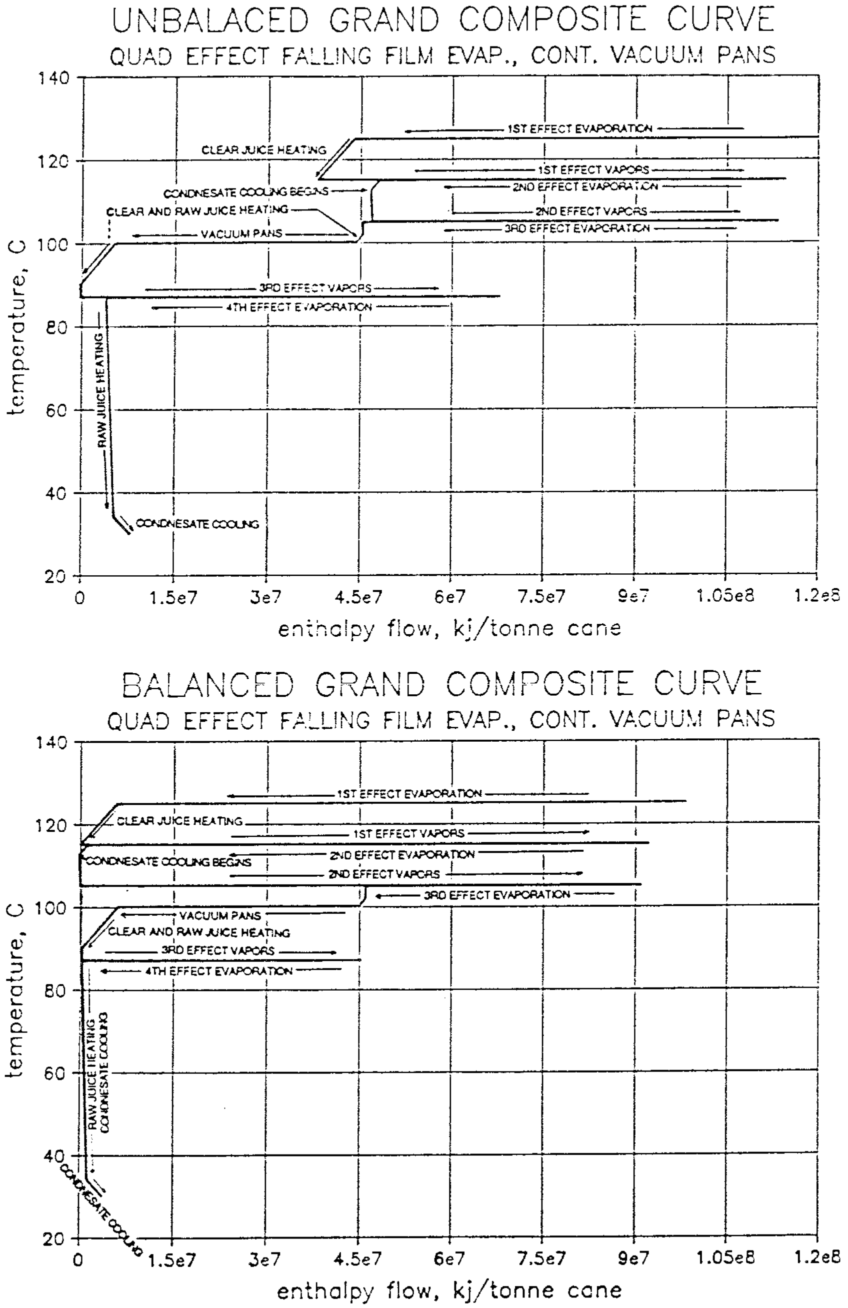
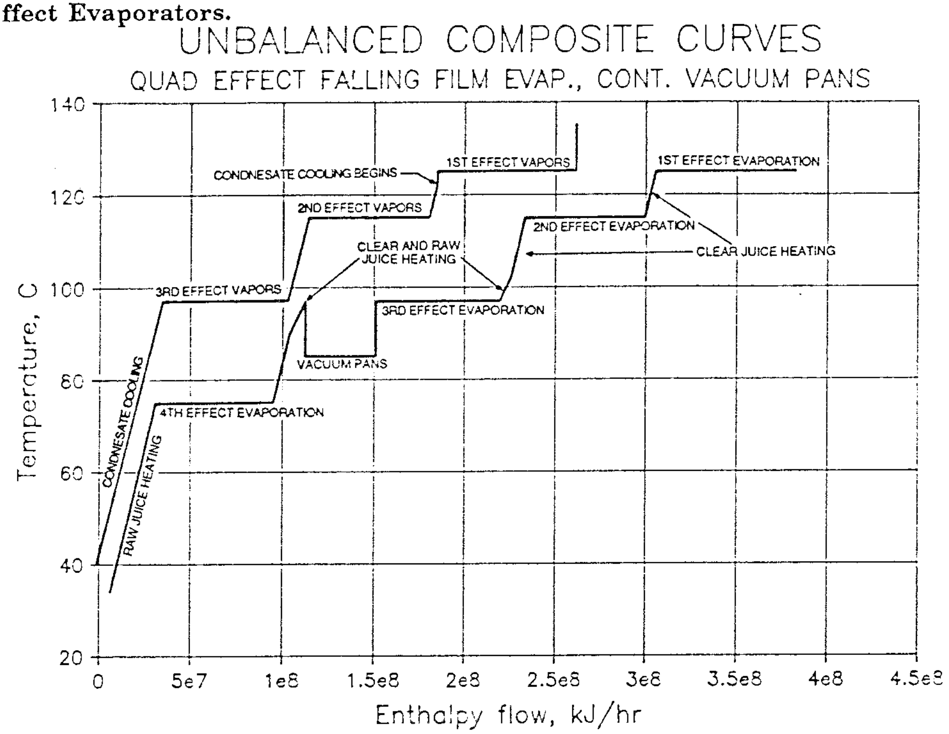
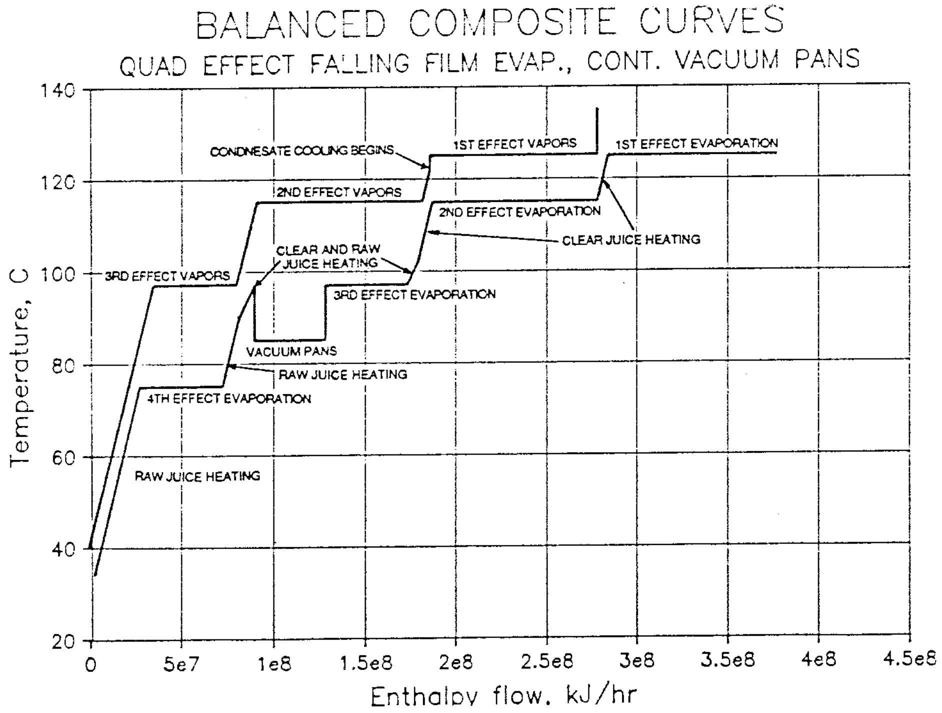


Figure 3.4a *Shows the grand composite curve of a sugar factory with an "unbalanced" multi-effect evaporator. In this case, the pinch occurs only at the fourth effect, with the bulk of the curve a long distance from the y axis. By looking at this curve one can see that steam will have to be used not only in the first effect, but also in the vacuum pans.*
Figure 3.4b *Shows the grand composite curve of a sugar factory with a "balanced" multi-effect evaporation. Notice that a pinch occurs at each effect, bringing the curve as close as possible to the y axis, and thereby minimizing the hot utility load. "Balanced" in these cases mean that each effect is involved in a pinch rather than each effect having the same evaporation duty.*

Figure 3.5a and b: Composite Curves for Sugar Factories Using Balanced and Unbalanced Multi-Effect Evaporators.

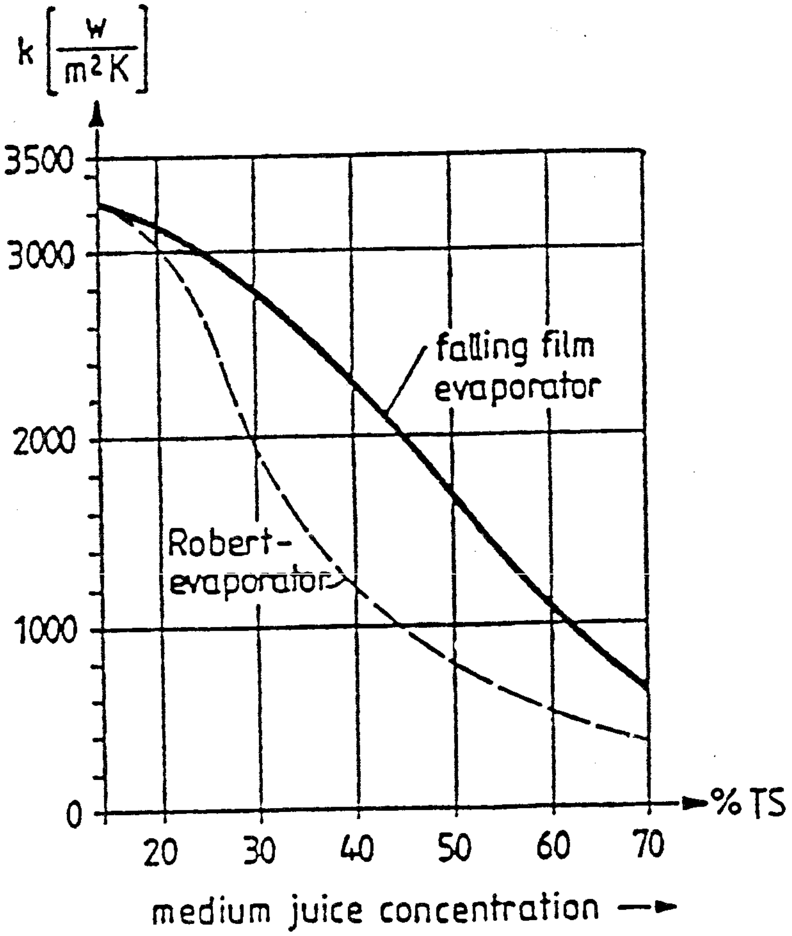




***Figures 3.5a and b** show the composite curves of sugar factories with "balanced" and "unbalanced" multi-effect evaporation. For the balanced case, the hot and cold composite curves remain much closer together throughout the length of the curve.*

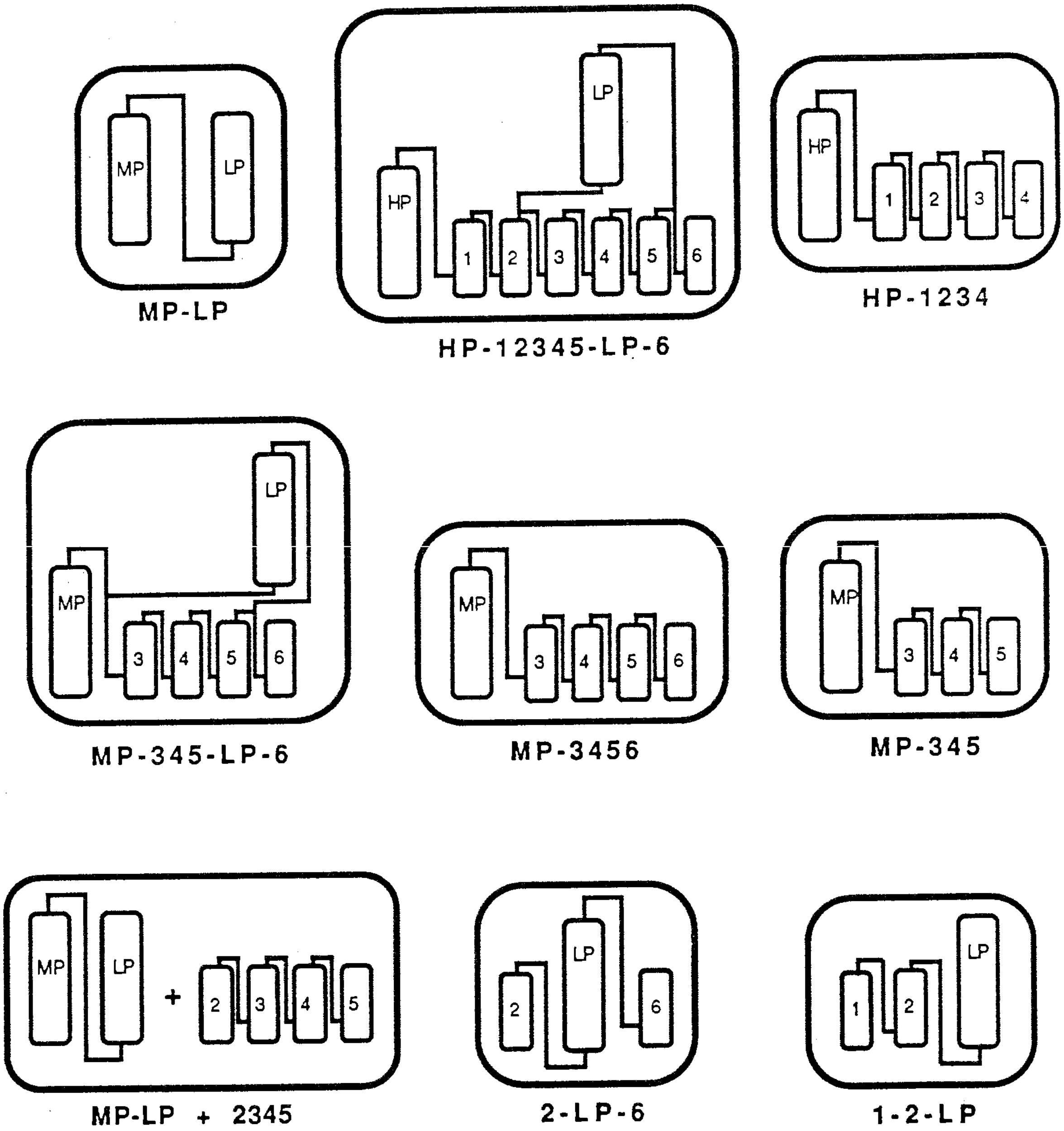
Also clearly evident in both figures is the "vacuum pan well". This dip in the cold composite curve is required so that higher approach temperature is maintained in the vacuum pans. When corrected temperatures are used in the grand composite curve, the vacuum pan well disappears.

Figure 3.6: Heat Transfer Coefficients in Long Tube, Falling Film Evaporators and Short Tube, Rising Film Evaporators.



Source: [12].

Figure 3.7: Schematic Diagrams of the Autonomous Distillery Configurations Considered in the Pinch Analyses.



MP-LP

HP-12345-LP-6

HP-1234

MP-345-LP-6

MP-3456

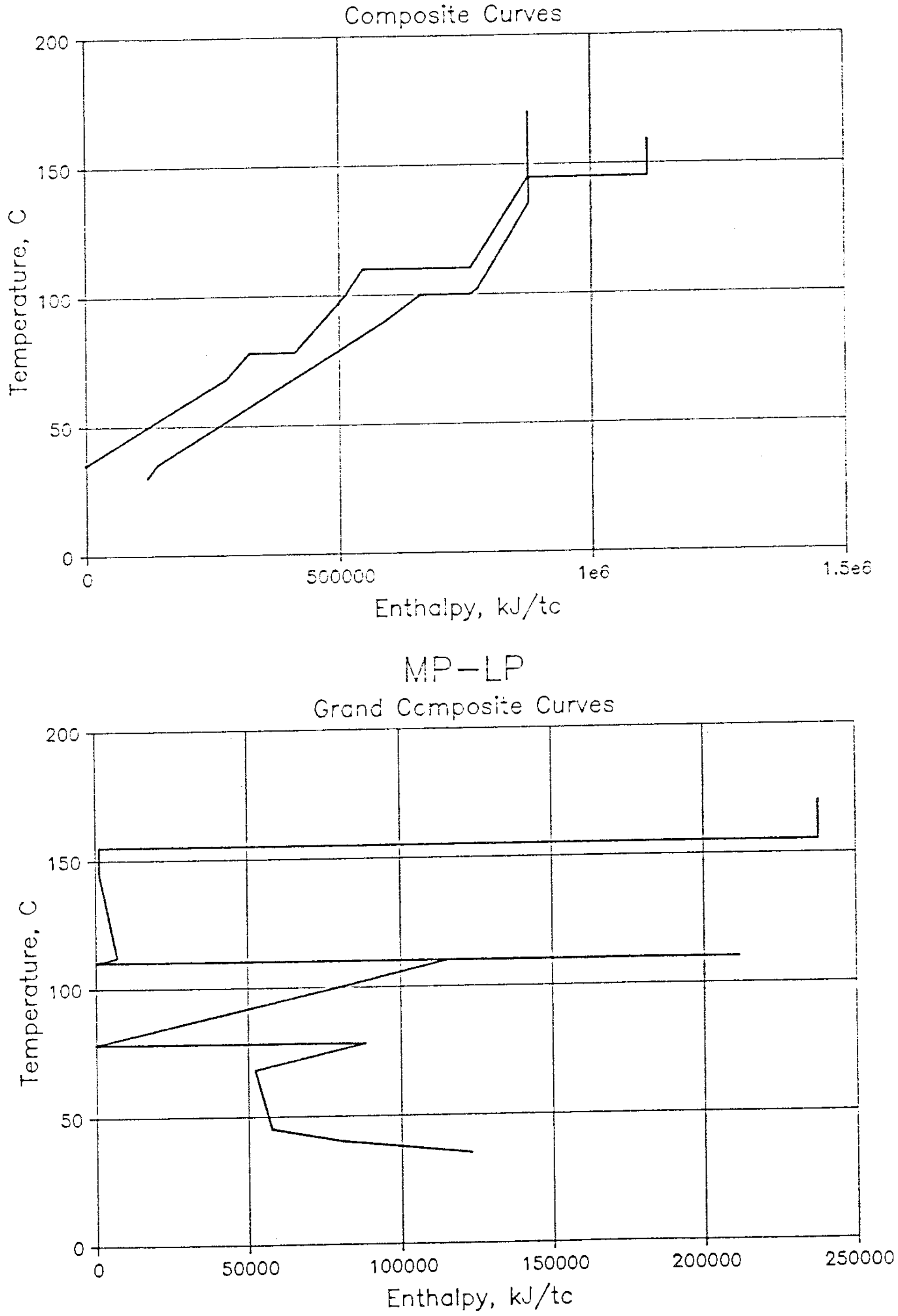
MP-345

MP-LP + 2345

2-LP-6

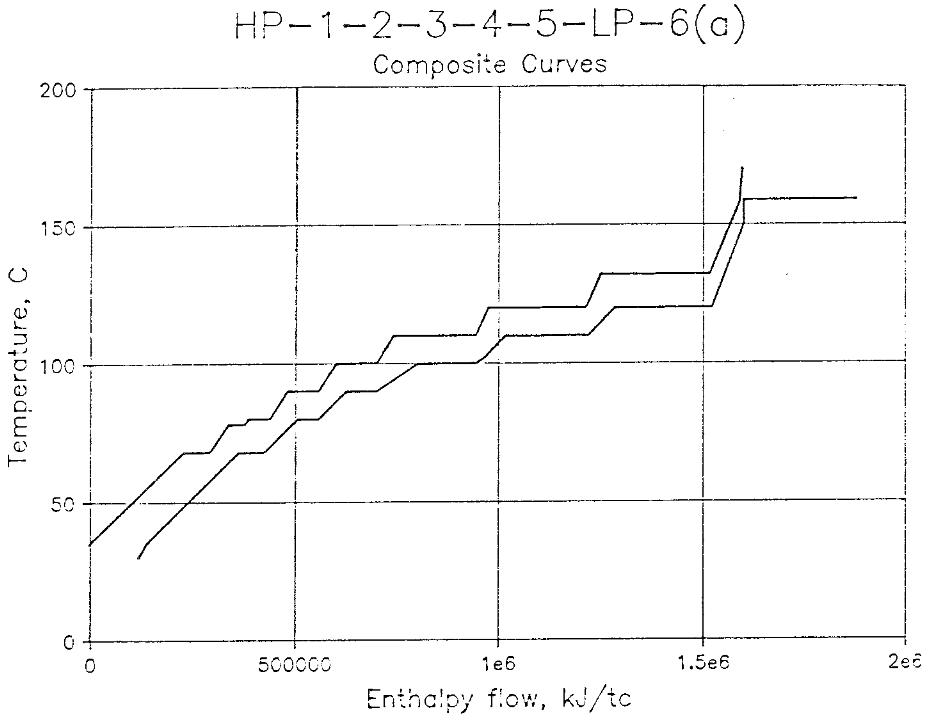
1-2-LP

Figure 3.8a and b: Composite and Grand Composite Curves for Autonomous Distillery Case MP-LP.



The apparent intersection of the hot and cold composite curves in ***Figure 3.8*** *is actually due the combination of finitely thick lines on the composite curves and the fact that no heating is occurring between 132°C and 148°C; the cold composite is vertical and* ΔT_{MIN} *is still being maintained.*

Figure 3.9a and b: Composite and Grand Composite Curves for Autonomous Distillery Case HP-1-2-3-4-5-LP-6(a).



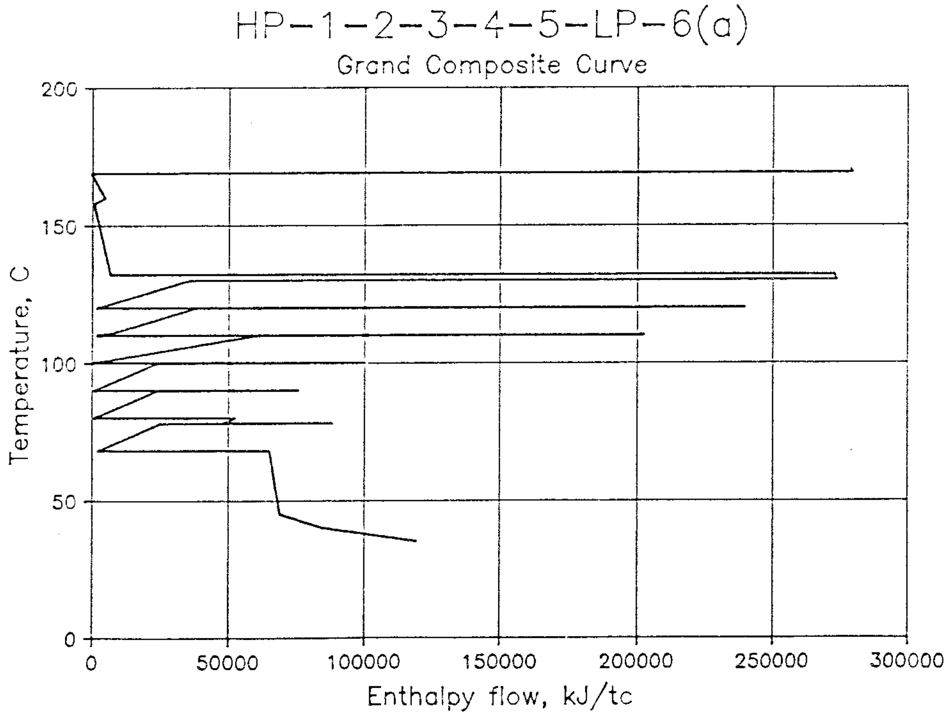
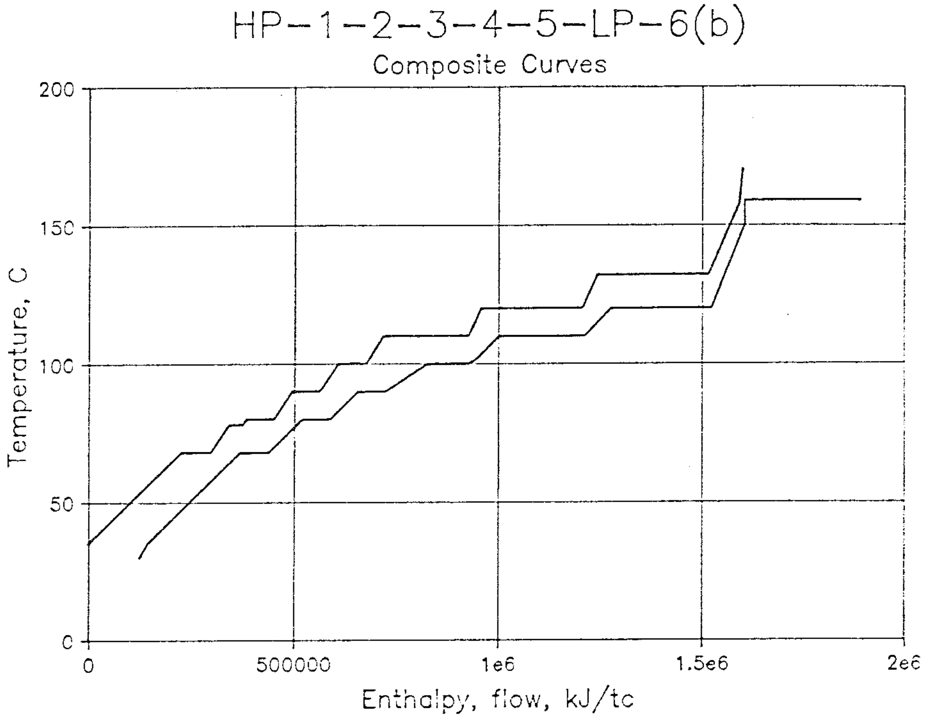


Figure 3.9: *The number of pinches are maximized, thus minimizing the overall steam requirement for this set of equipment.*

Figure 3.10a and b: Composite and Grand Composite Curves for Autonomous Distillery Case HP-1-2-3-4-5-LP-6(b).



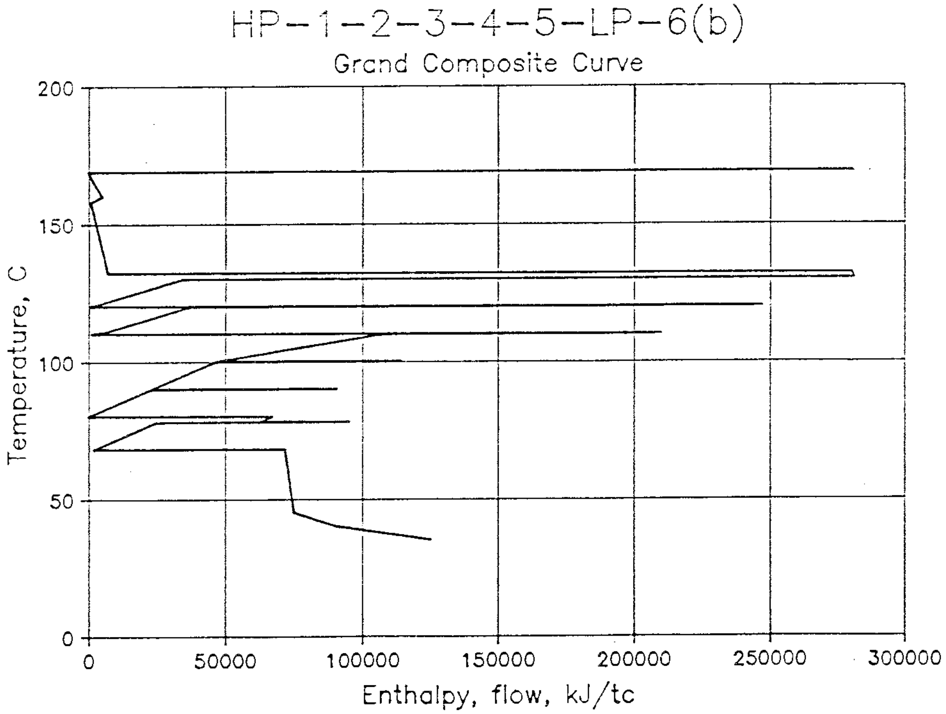


Figure 3.10: *The third and fourth evaporator effect "spikes" in the grand composite curve mesh into the next lower evaporator effect spikes. This implies that the heat from the vapors of the previous effect exactly meet the heat requirements of that effect. This simplification costs approximately 4 kg/tc steam, but greatly simplifies the heat exchanger network.*

Figure 3.11a and b: Composite and Grand Composite Curves for Autonomous Distillery Case HP-1-2-3-4.

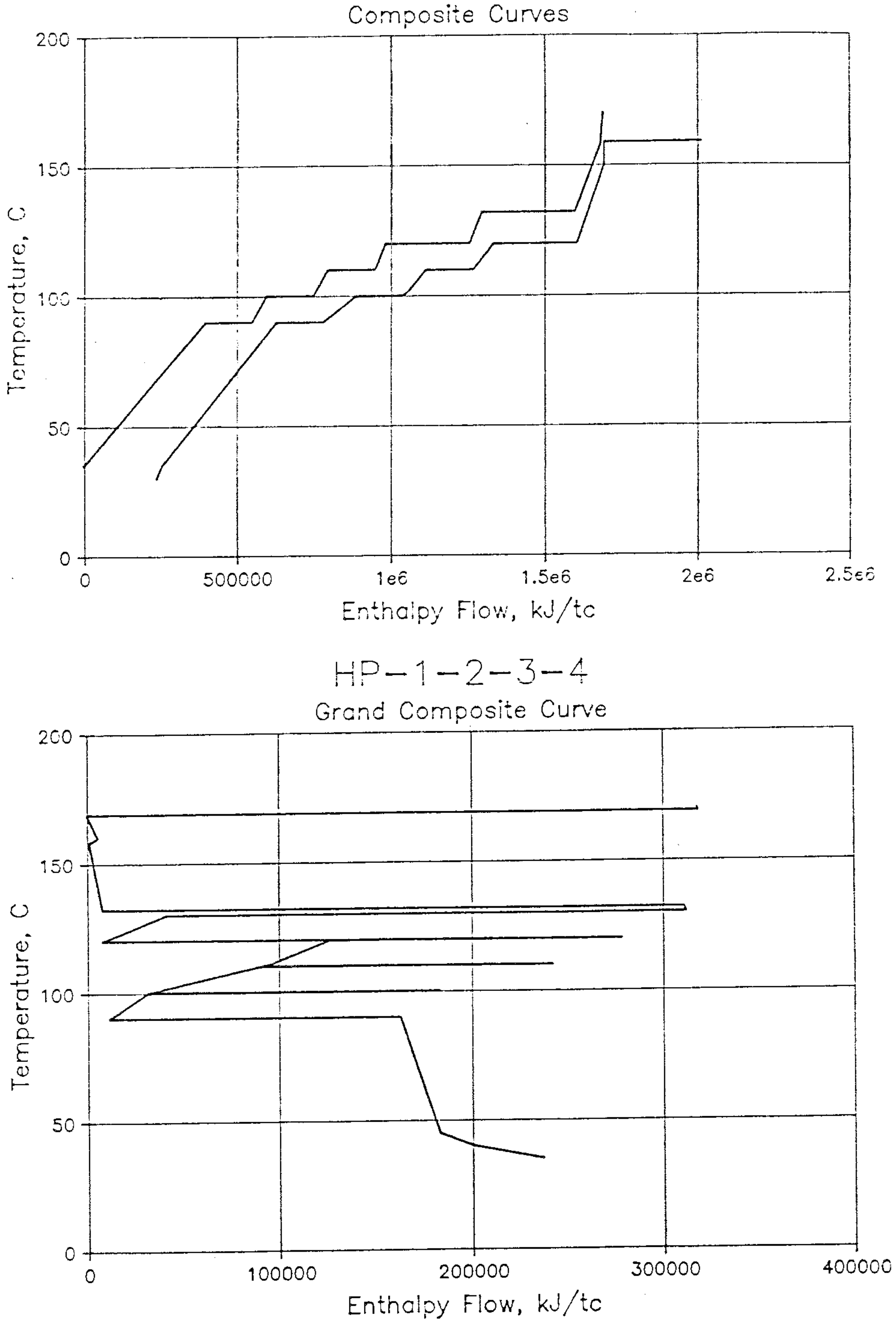


Figure 3.11: *Note that the uppermost "spike" (more like a "bar" in this case) is the distillation column. The smaller spikes below are the four evaporation effects. Like the previous figure, the second and third evaporator effects were sized such that the vapors out of then exactly met the heat requirements of the third and fourth effects. No steam penalty is incurred because distillation is the limiting process.*

Figure 3.12a and b: Composite and Grand Composite Curves for Autonomous Distillery Case MP-3-4-5-LP-6.

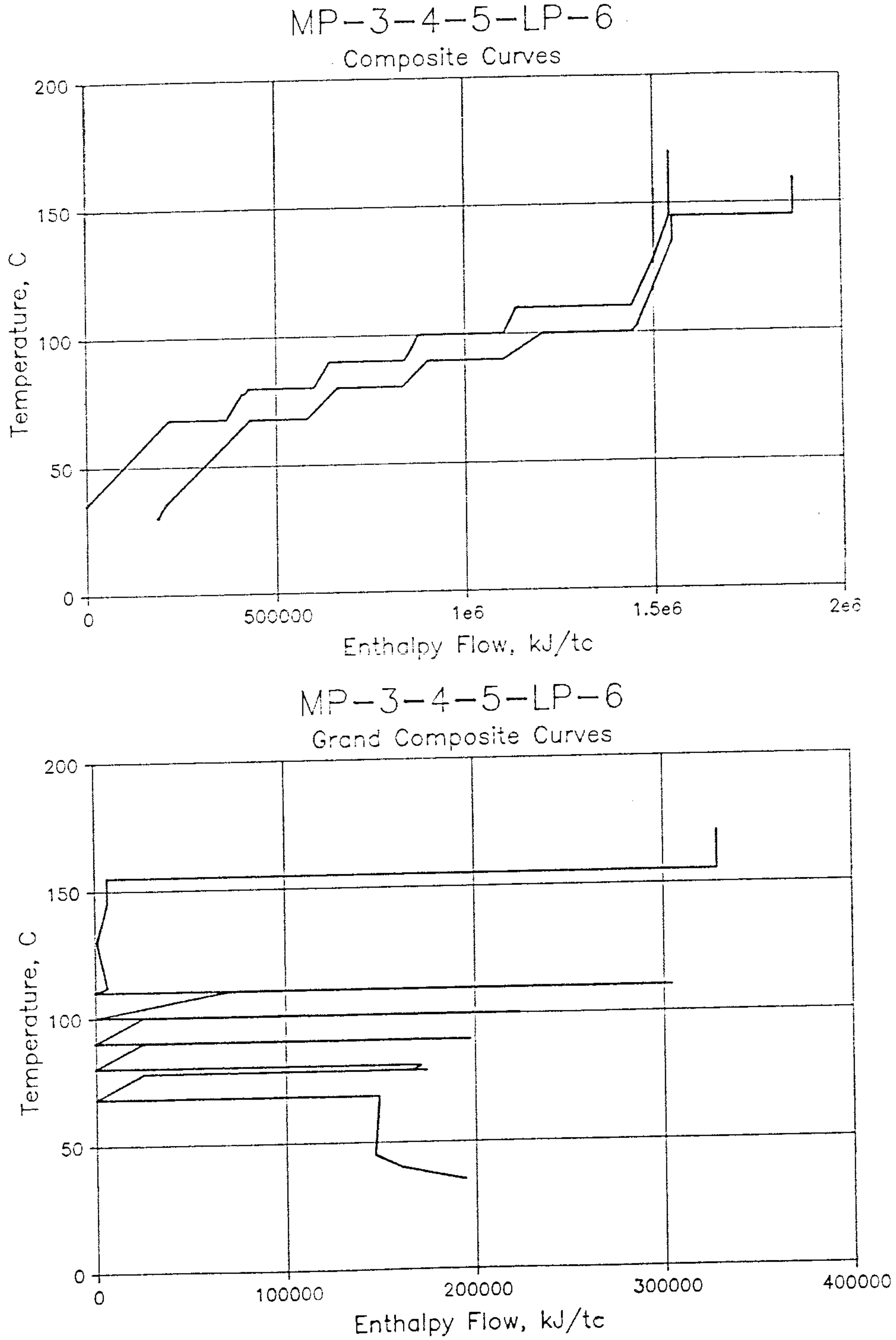


Figure 3.16: The very small load of the low pressure distillation column is noted on both the composite and grand composite curves.

Figure 3.13a and b: Composite and Grand Composite Curves for Autonomous Distillery Case MP-3-4-5-6.

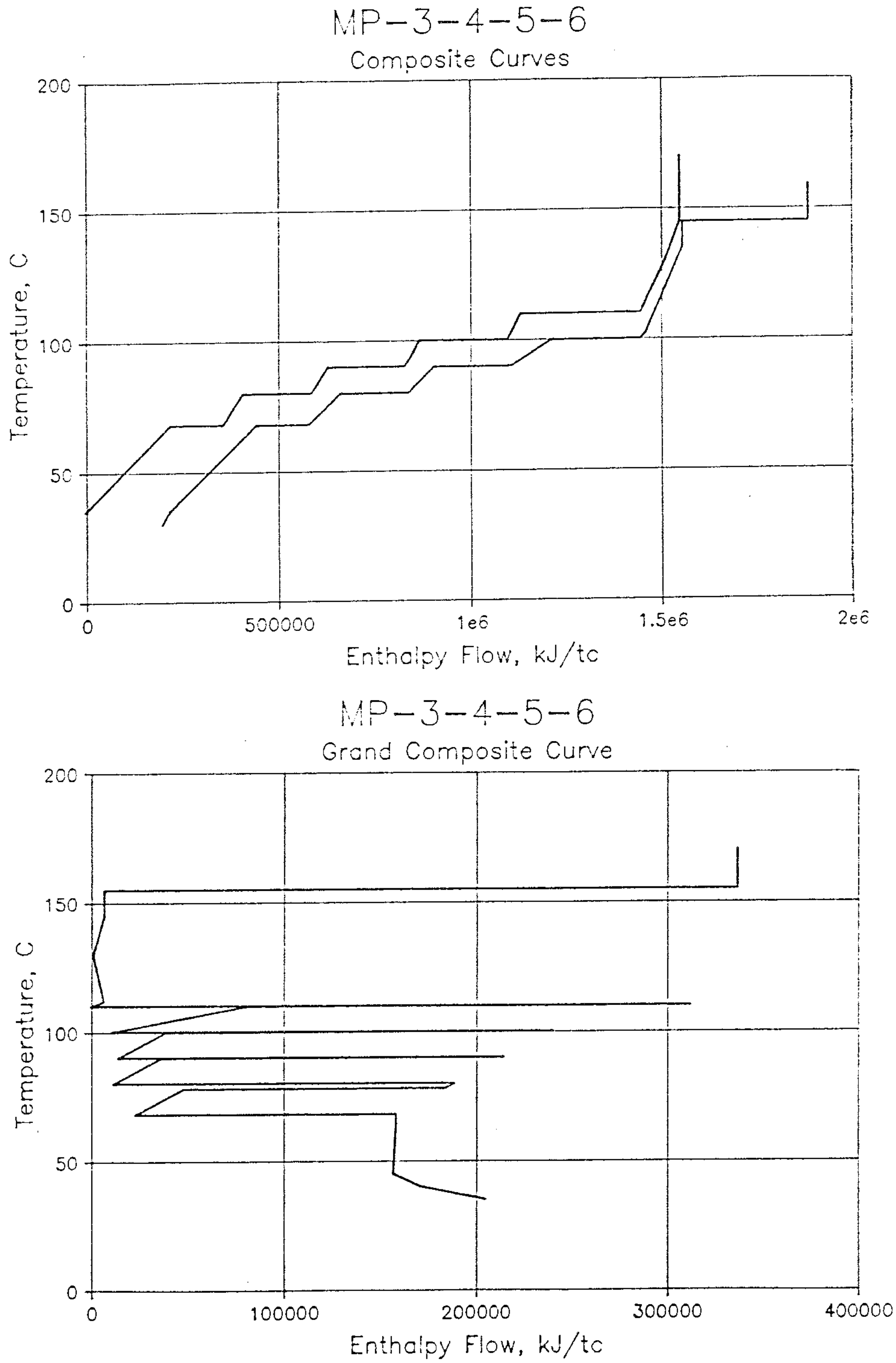


Figure 3.13: *The case is distillation limited, with the four evaporator spikes on the grand composite curve not approaching the y axis.*

Figure 3.14a and b: Composite and Grand Composite Curves for Autonomous Distillery Case MP-3-4-5.

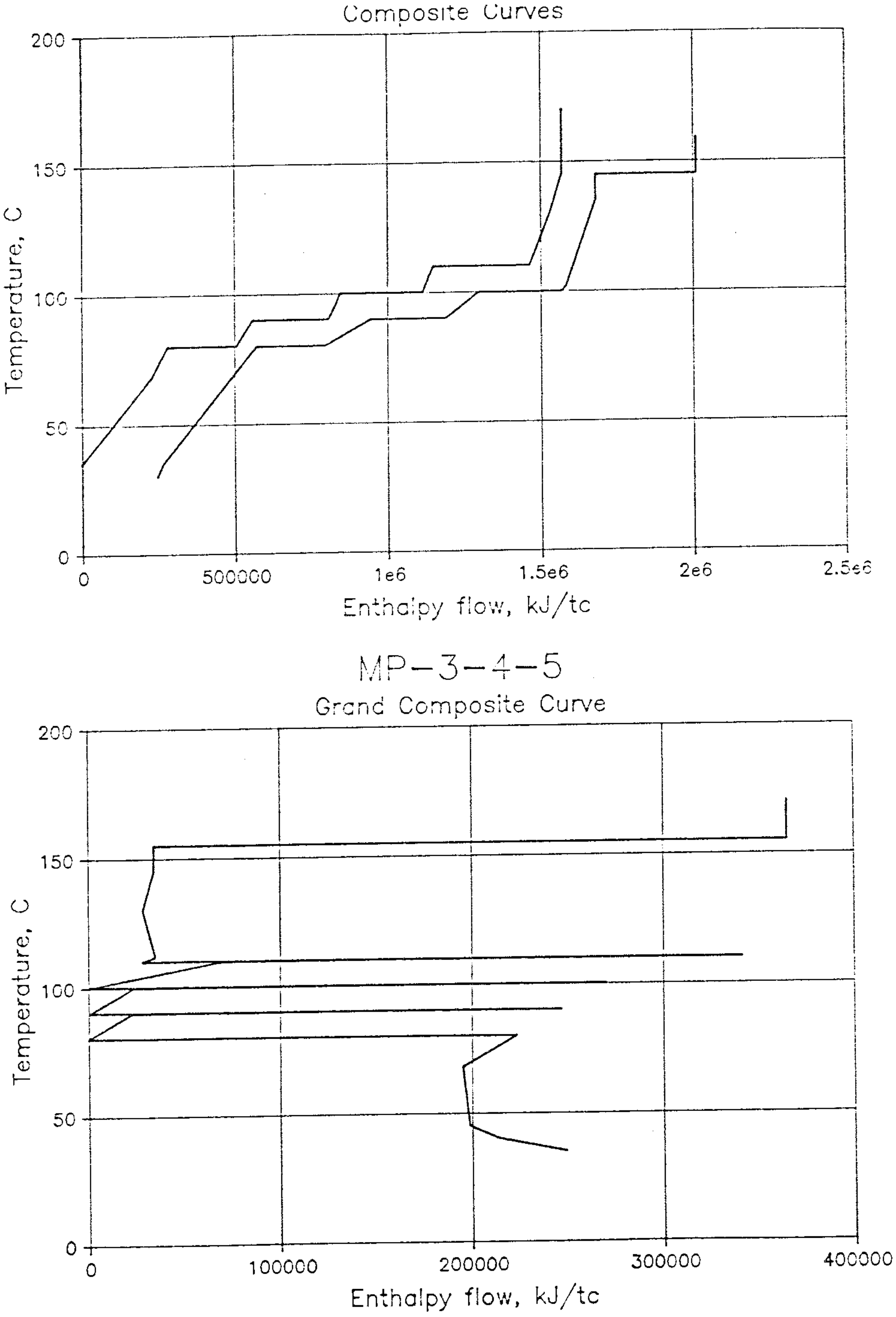


Figure 3.14: *With only three evaporator effects, the limiting process is evaporation. This is seen in the grand composite curve, where the three lower evaporator effect spikes form pinch points while the distillation column section remains away from the y axis.*

Figure 3.15a and b: Composite and Grand Composite Curves for the Distillation Section (2-3-4-5) of Autonomous Distillery Case MP-LP + 2-3-4-5.

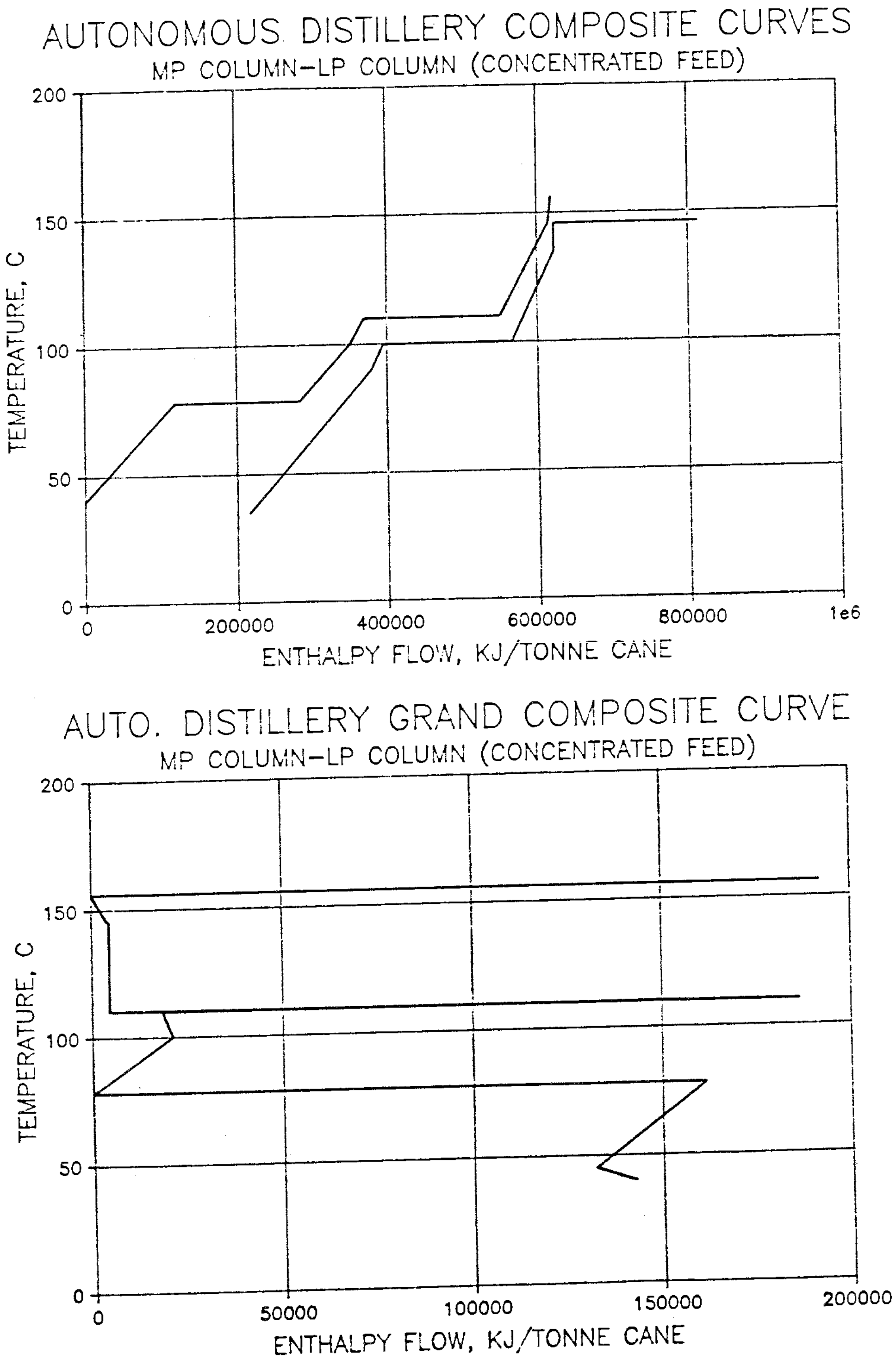
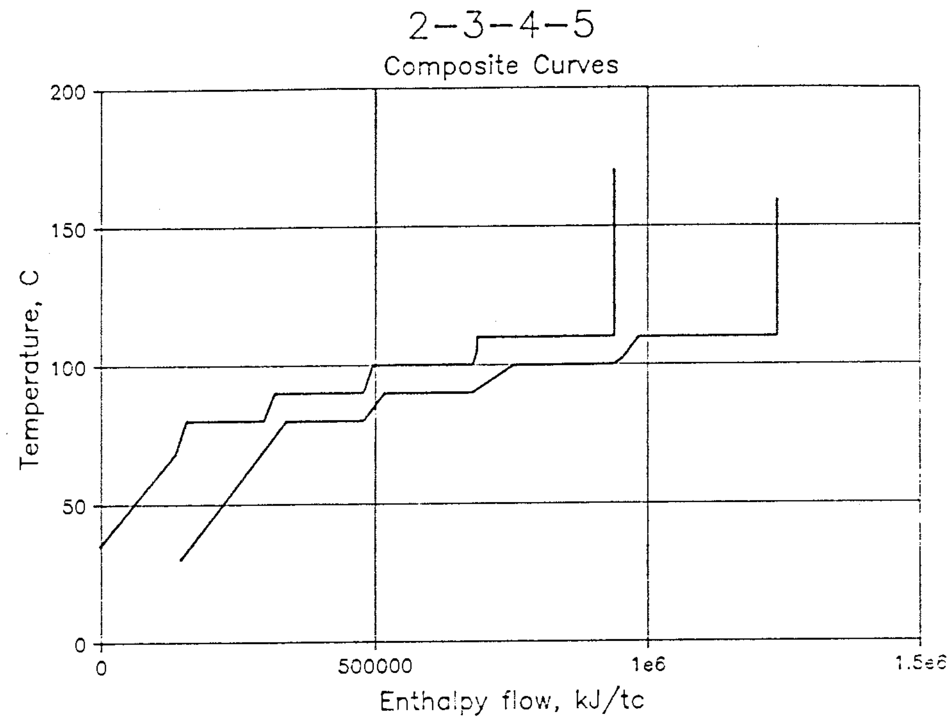
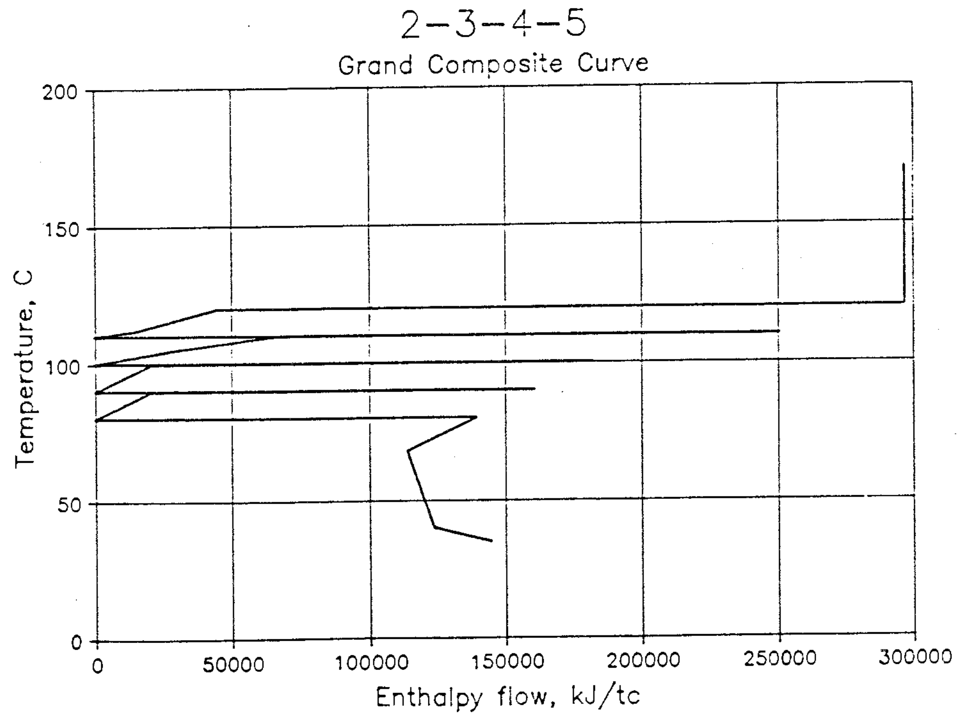


Figure 3.15c and d: Composite and Grand Composite Curves for the Evaporation Section (MP-LP) of Autonomous Distillery Case MP-LP + 2-3-4-5.


2-3-4-5

Composite Curves


2-3-4-5

Grand Composite Curve

Figure 3.16a and b: Composite and Grand Composite Curves for Autonomous Distillery Case 2-LP-6.

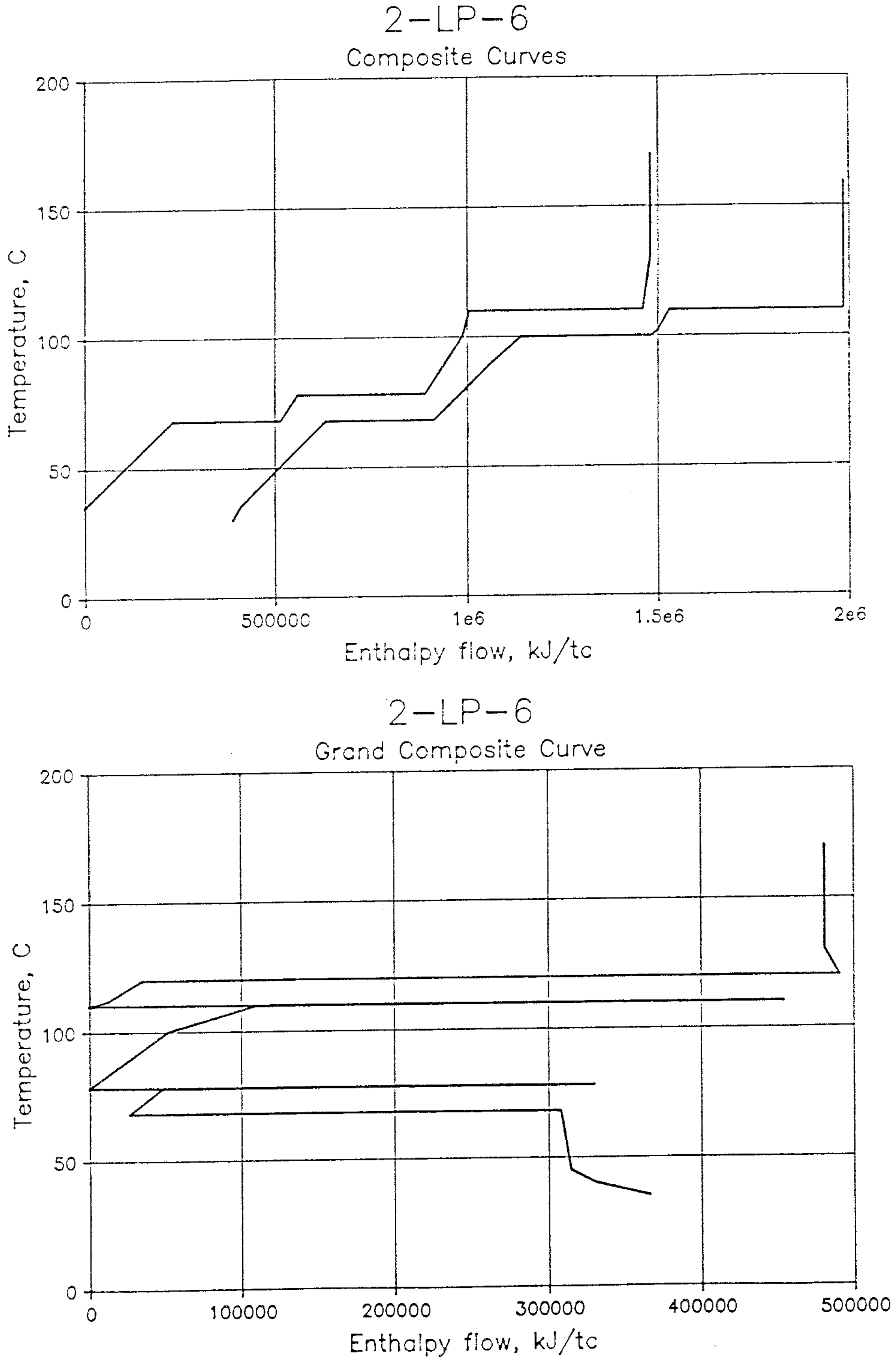


Figure 3.16: The case is distillation limited, with the low temperature evaporator effect (effect 6) spike on the grand composite curve not reaching the y axis.

Figure 3.17a and b: Composite and Grand Composite Curves for Autonomous Distillery Case 1-2-LP.

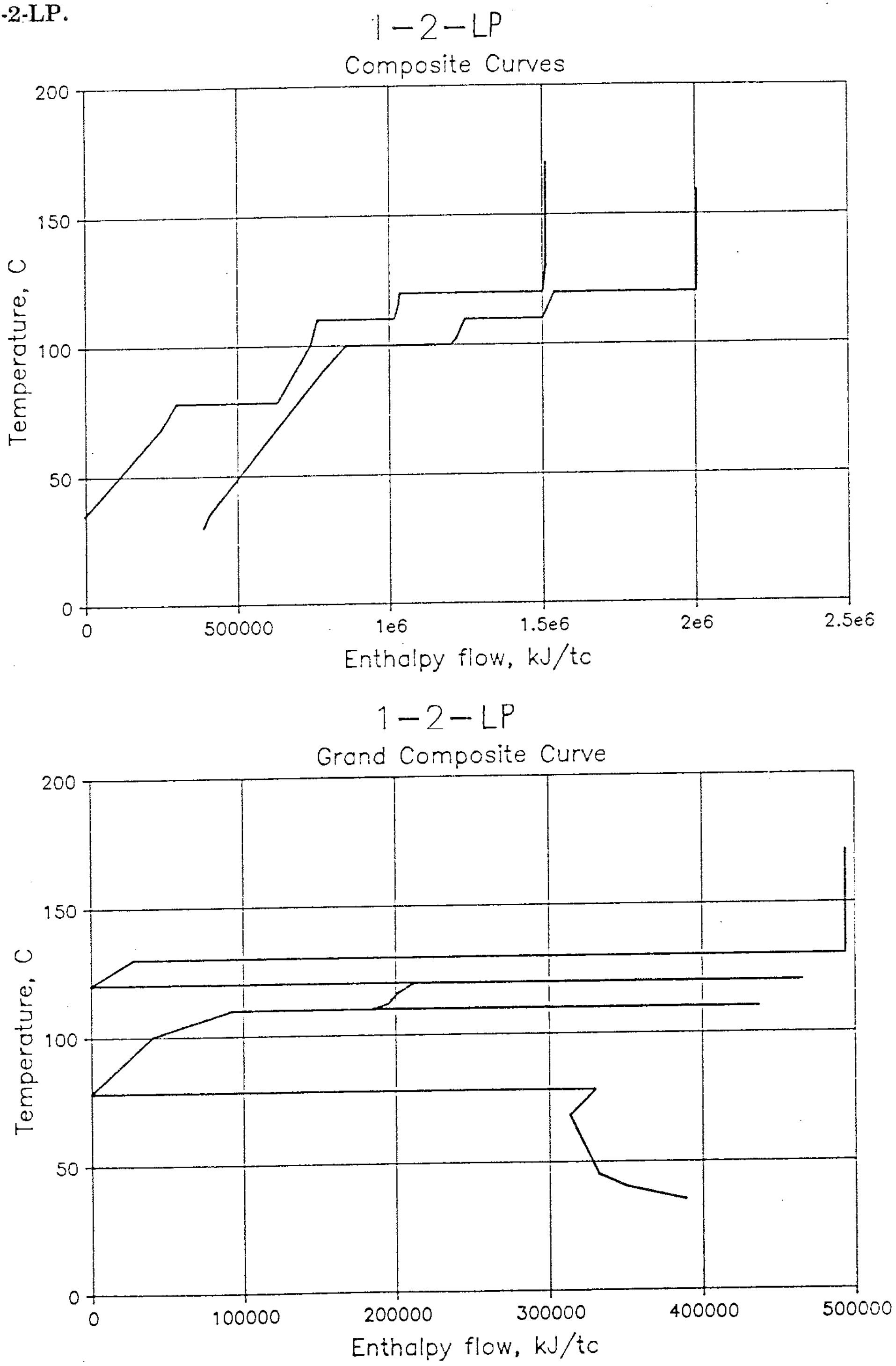


Figure 3.17: The case is distillation limited, with the lower temperature evaporator effect (effect 2) spike on the grand composite curve not reaching the y axis.

Figure 3.18: Process Flow Diagram for Autonomous Distillery Case MP-LP.

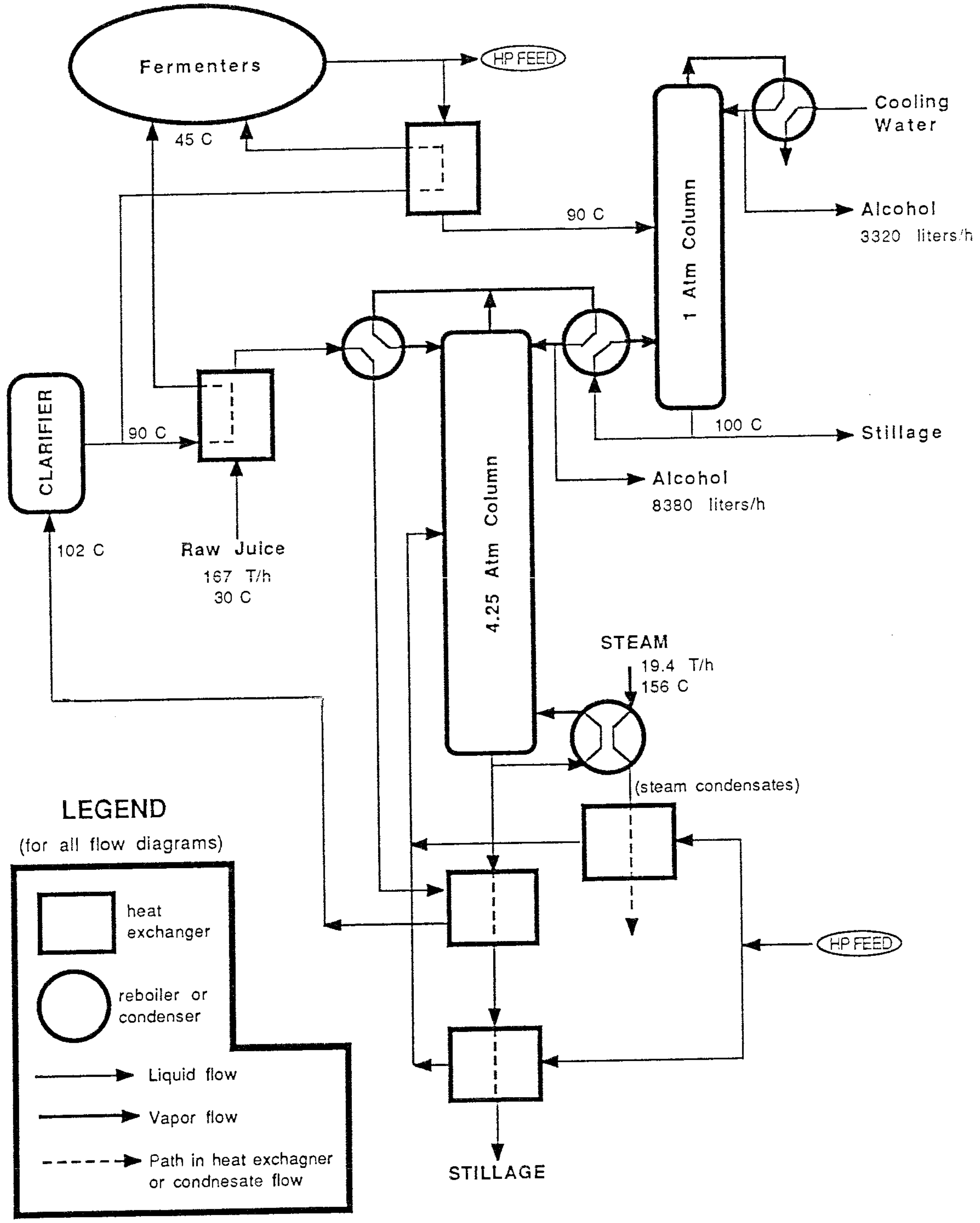


Figure 3.19: Process Flow Diagram for Autonomous Distillery Case HP-1-2-3-4-5-LP-6.

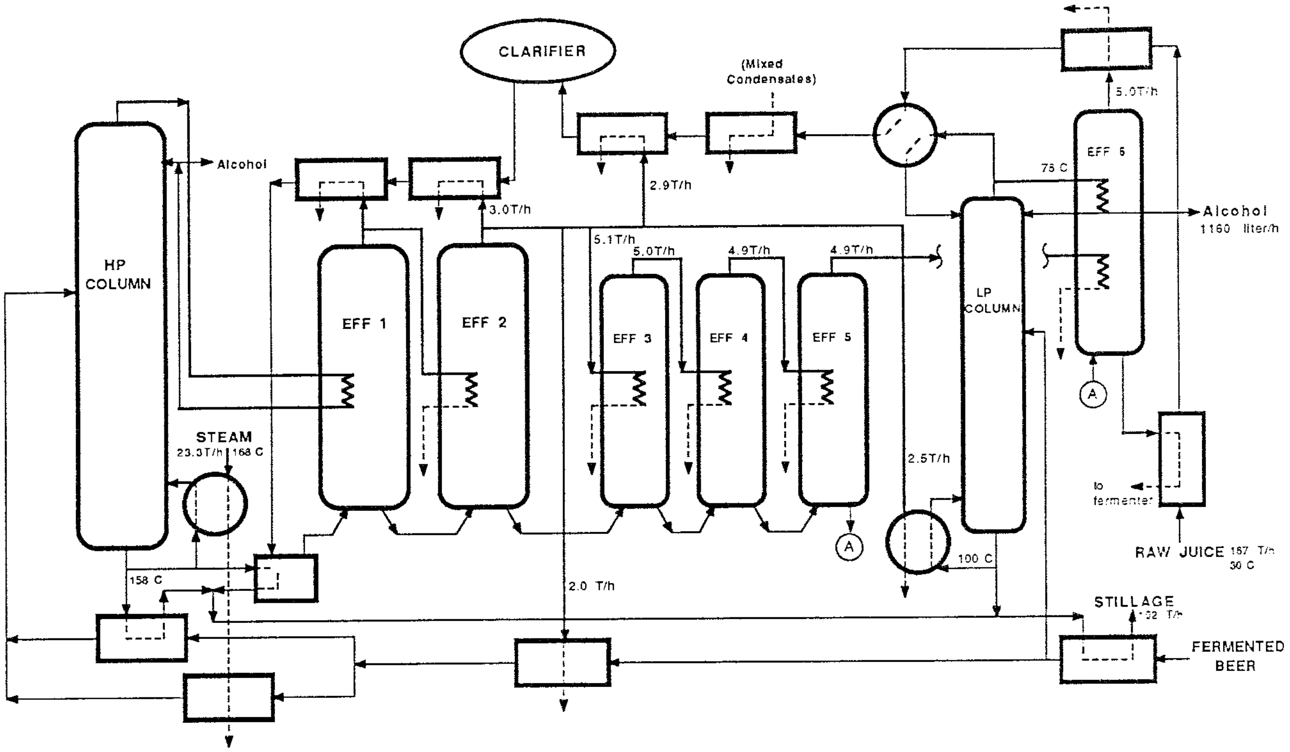


Figure 3.20: Process Flow Diagram for Autonomous Distillery Case HP-1-2-3-4.

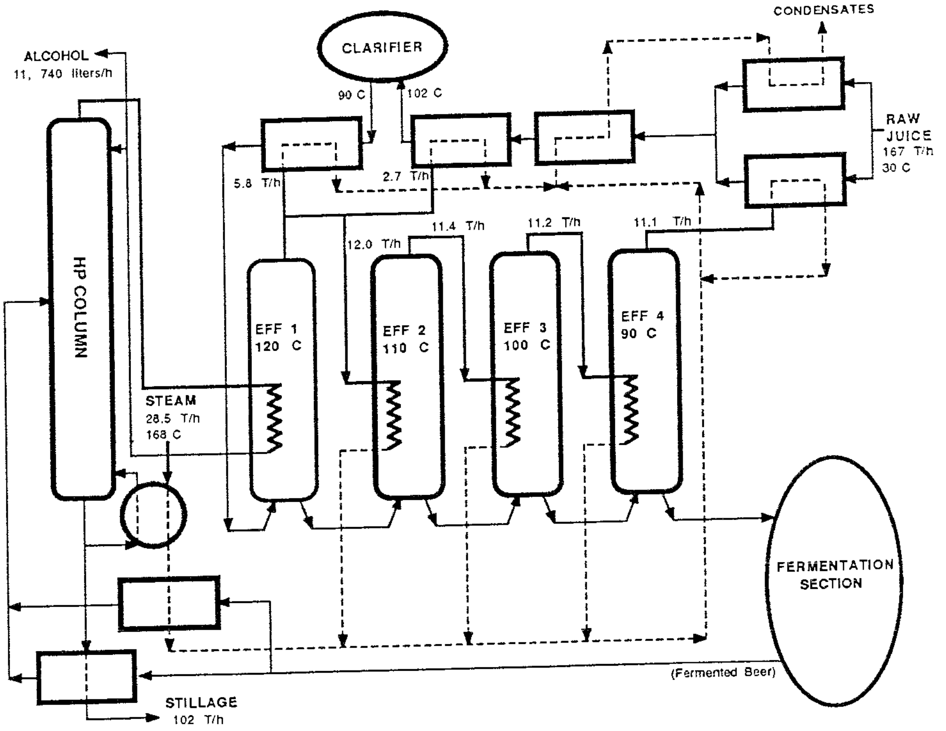


Figure 3.21a Flow Diagram for Evaporation Section of Autonomous Distillery Case MP-LP + 2-3-4-5.

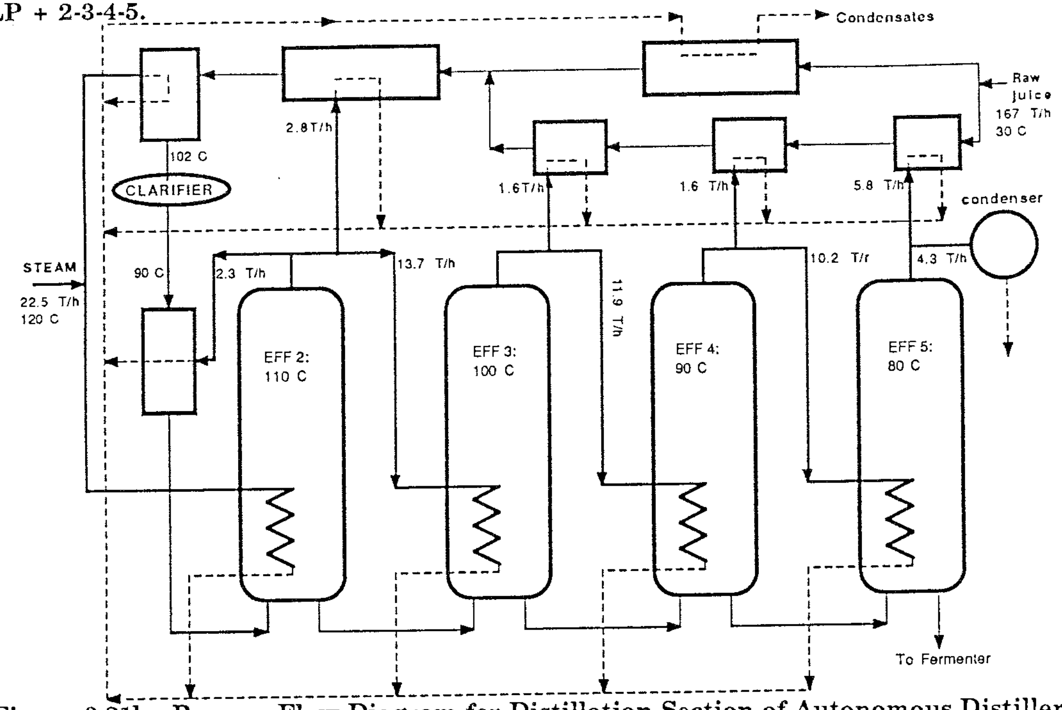


Figure 3.21b: Process Flow Diagram for Distillation Section of Autonomous Distillery Case MP-LP + 2-3-4-5.

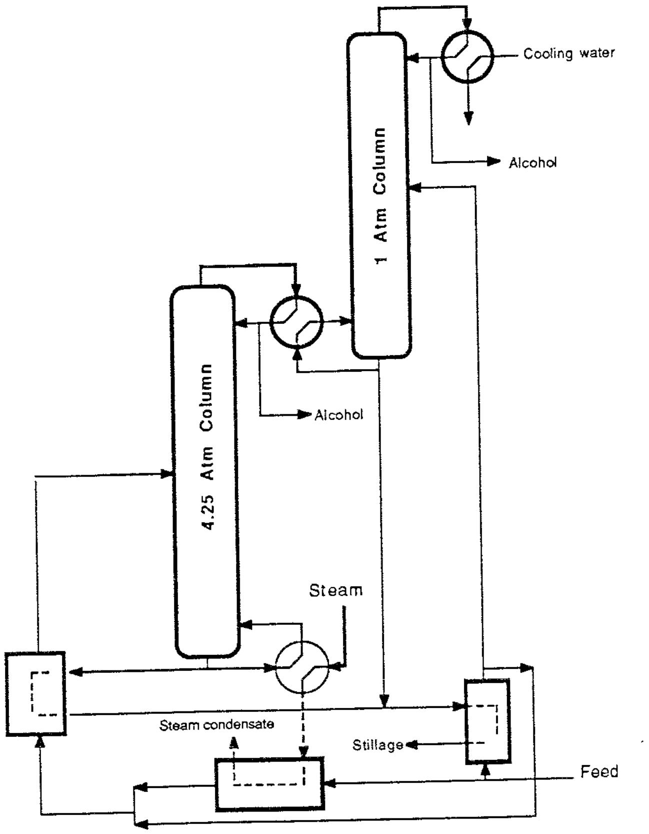


Figure 3.22: Process Flow Diagram for Benzene Azeoptropic Distillation of Anhydrous Ethanol. From reference 19.

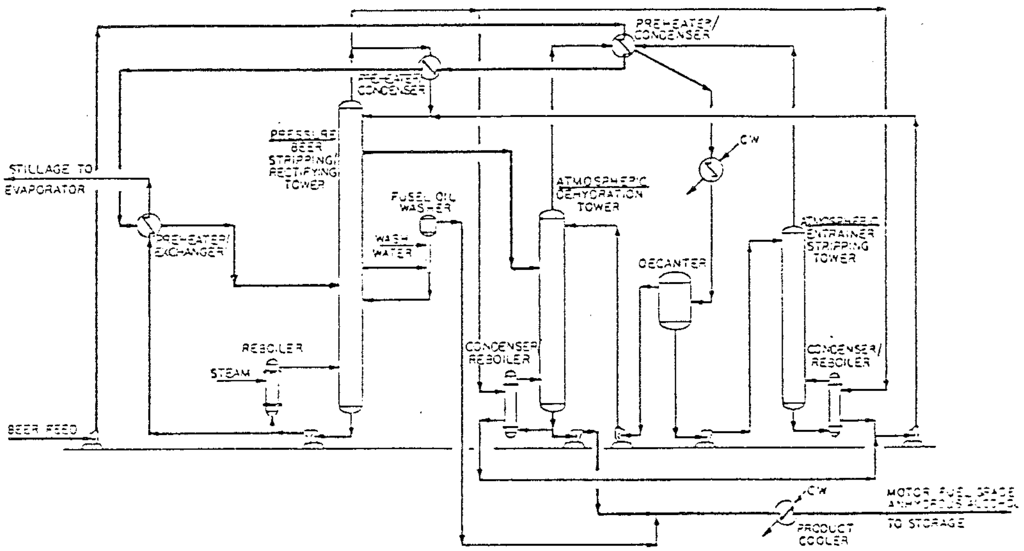


Figure 3.23: Process Flow Diagram for Ethylene Glycol Extractive Distillation of Anhydrous Ethanol. [13]

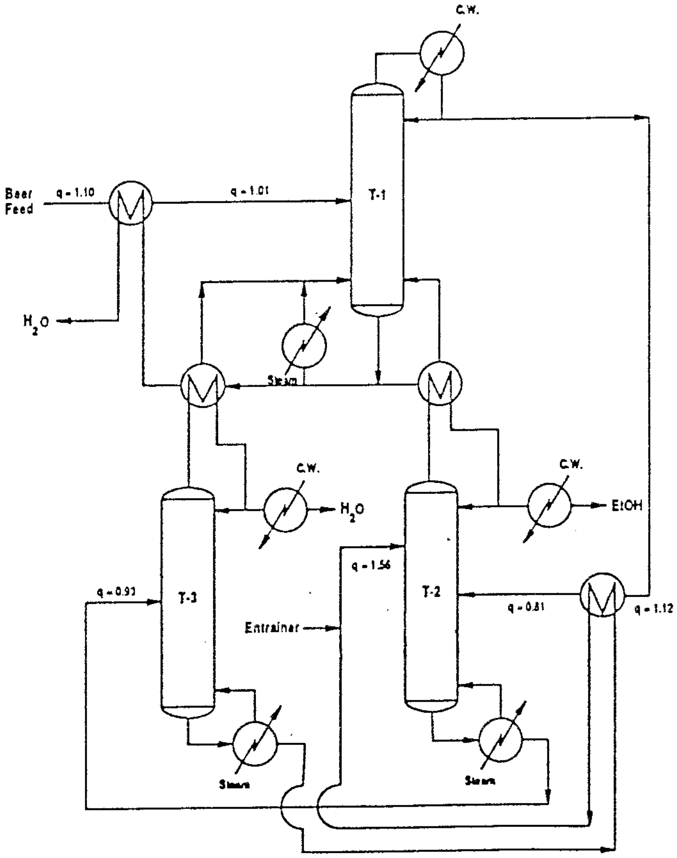


Figure 3.24: Process Flow Diagram for Ethanol Dehydration Using Molecular Sieves.

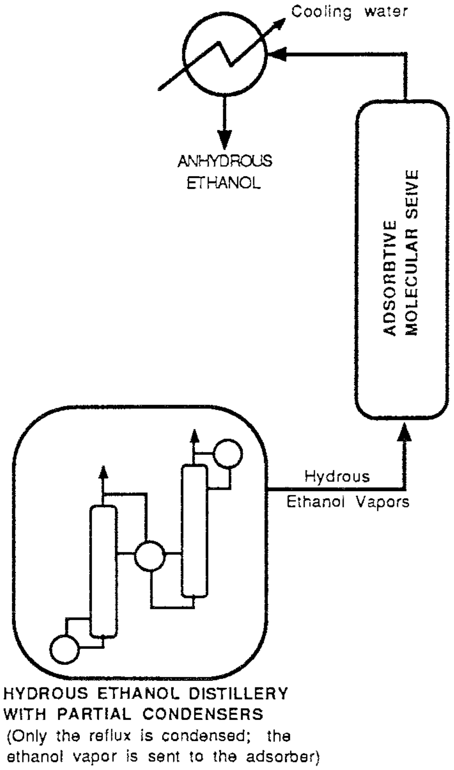
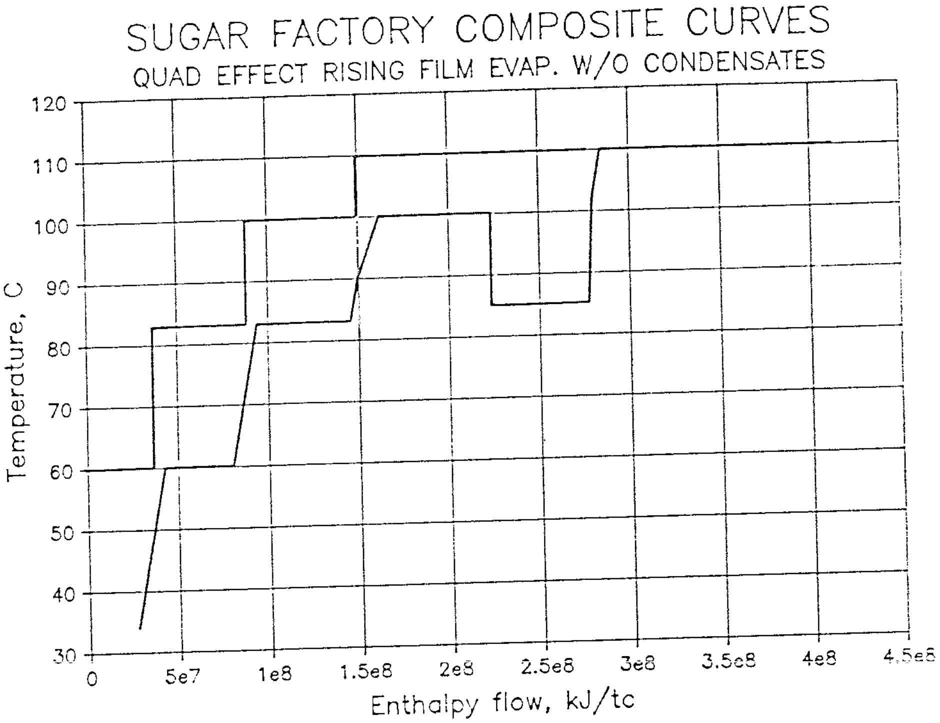


Figure 3.25a and b: Composite and Grand Composite Curves for Sugar Factory Using 4 Effect, Rising Film Evaporators, Batch Vacuum Pans and Without Condensate Juice Heating (Case 1).



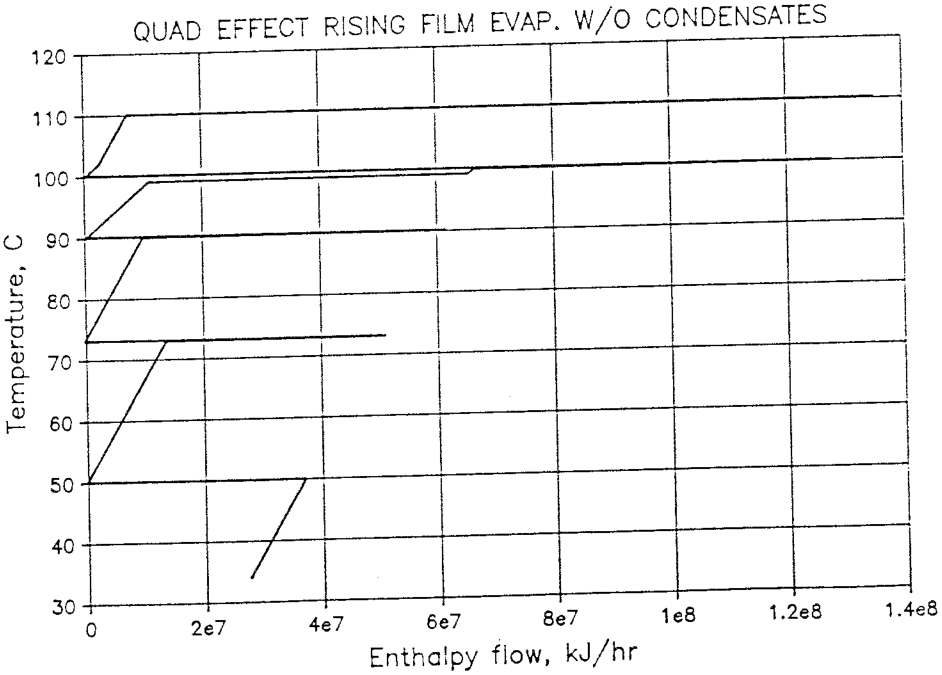
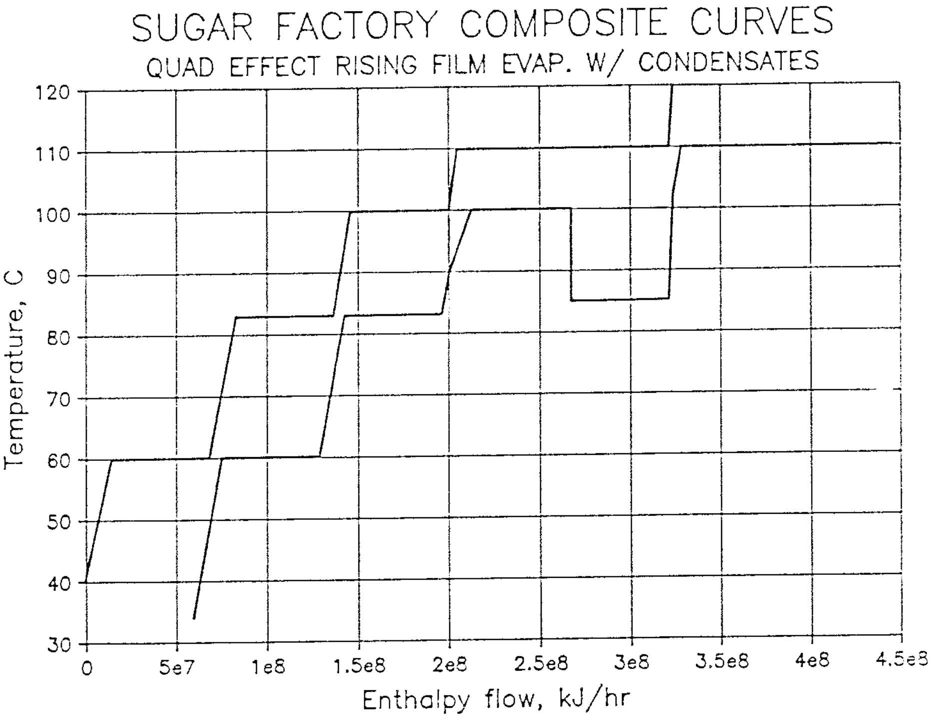


Figure 3.26a and b: Composite and Grand Composite Curves for Sugar Factory Using 4 Effect, Rising Film Evaporators, Batch Vacuum Pans With Condensate Juice Heating (Case 2).



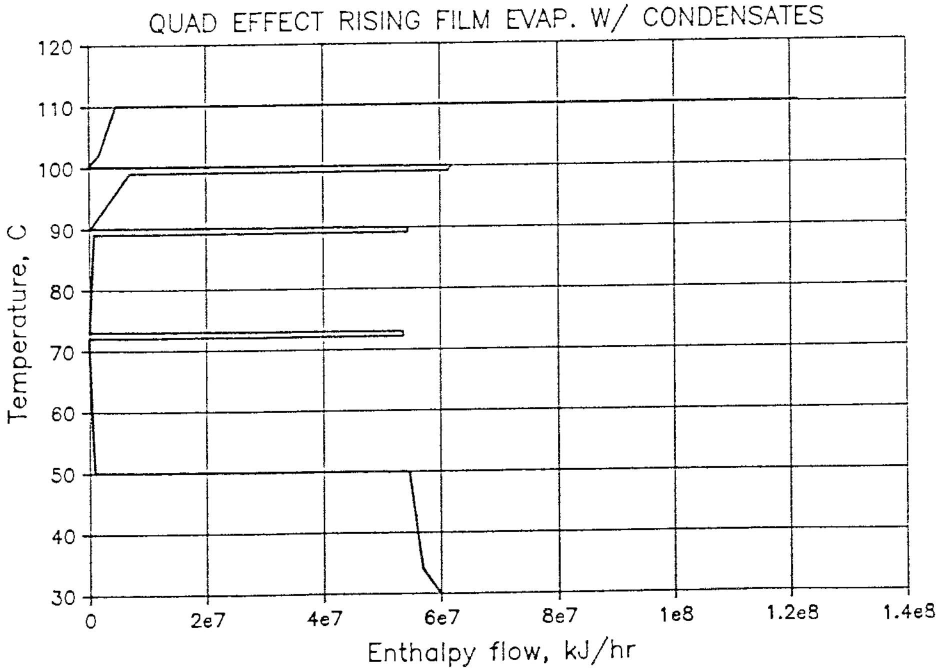
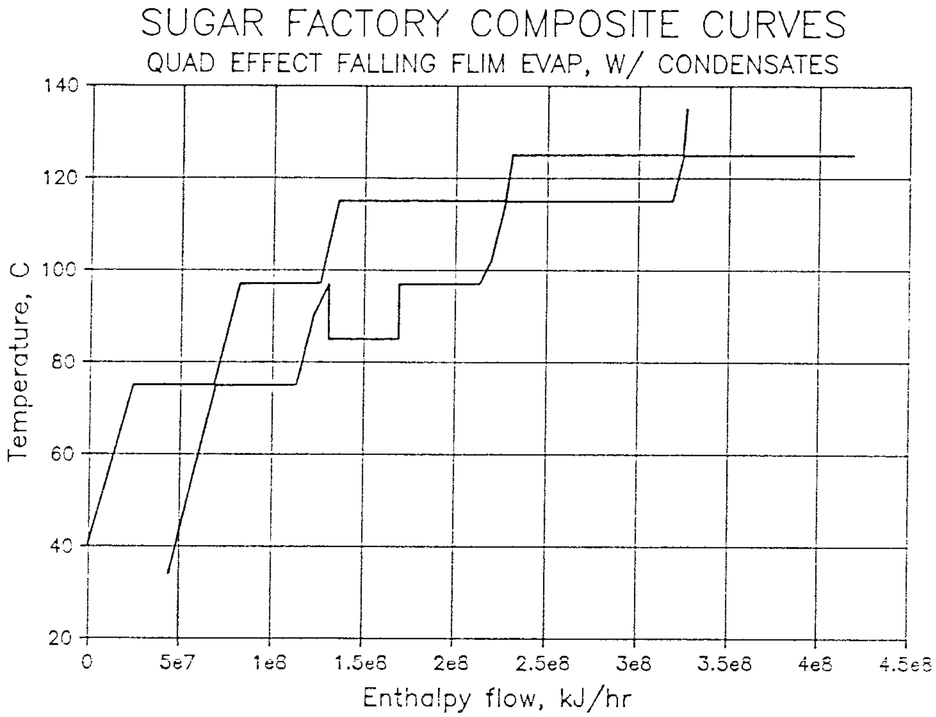


Figure 3.27a and b: Composite and Grand Composite Curves for Sugar Factory Using 4 Effect, Falling Film Evaporators, Batch Vacuum Pans With Condensate Juice Heating (Case 3).



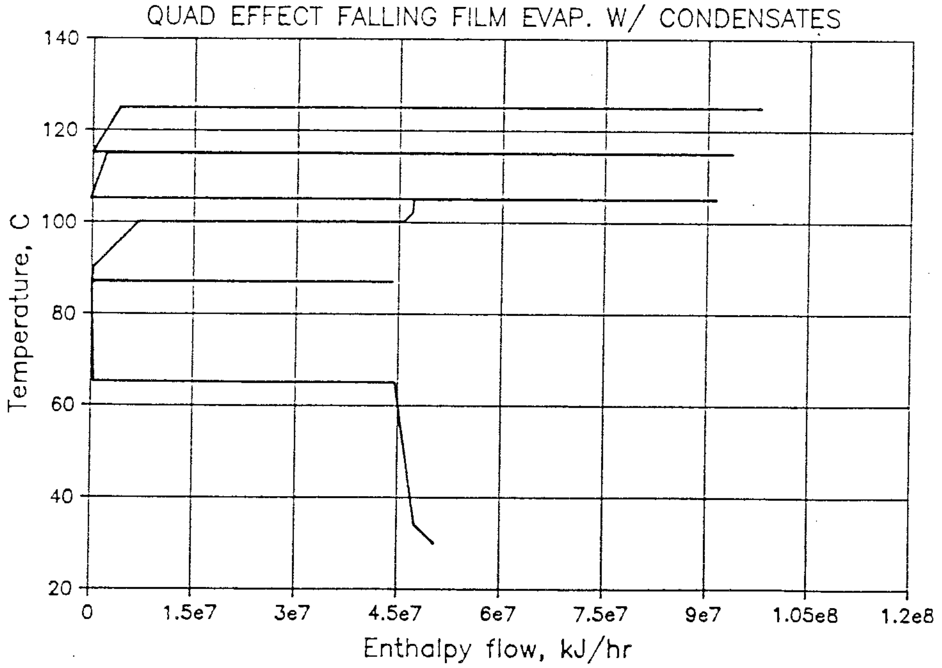
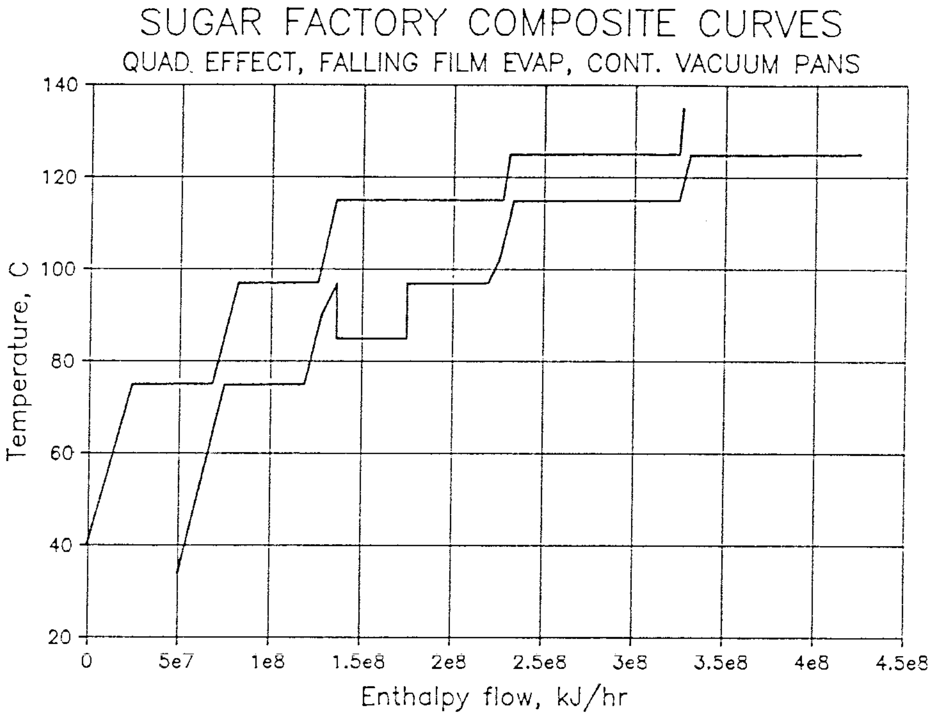


Figure 3.28a and b: Composite and Grand Composite Curves for Sugar Factory Using 4 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Condensate Juice Heating (Case 4).



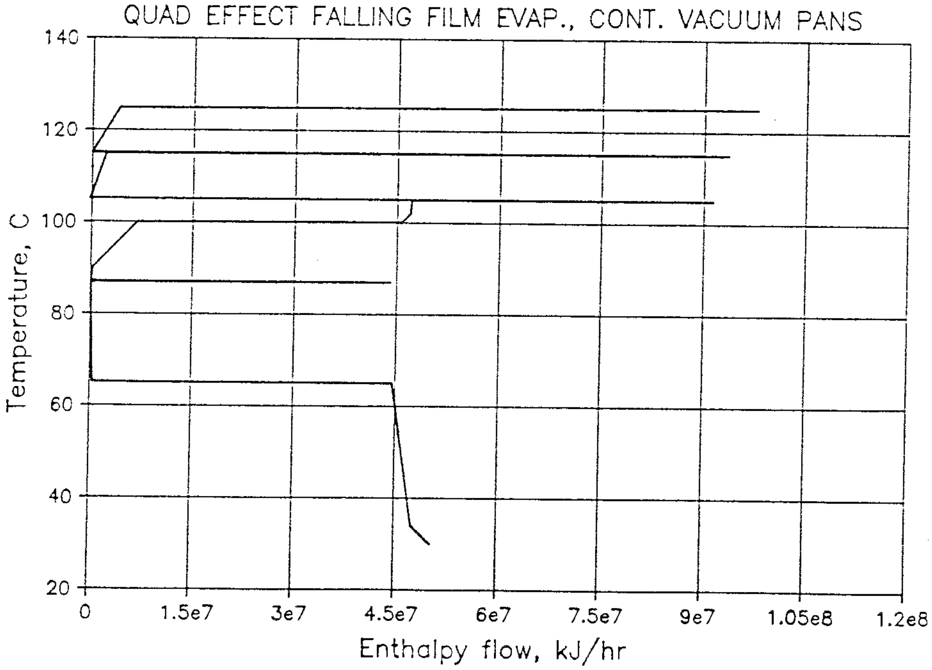


Figure 3.29a and b: Composite and Grand Composite Curves for Sugar Factory Using 4 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Aggregated Condensate Stream Juice Heating (Case 5).

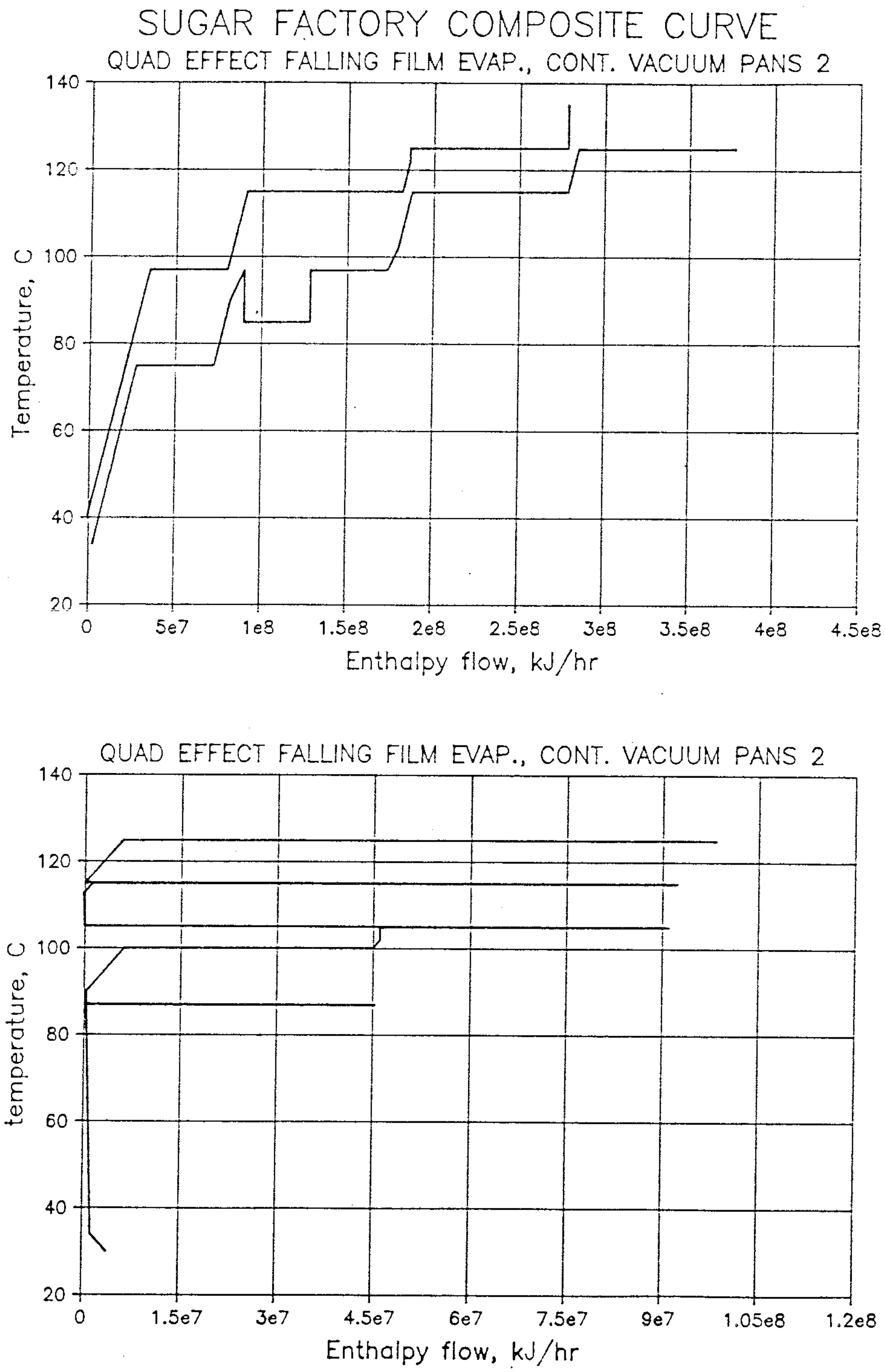


Figure 3.30a and b: Composite and Grand Composite Curves for Sugar Factory Using 5 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Condensate Juice Heating (Case 6).

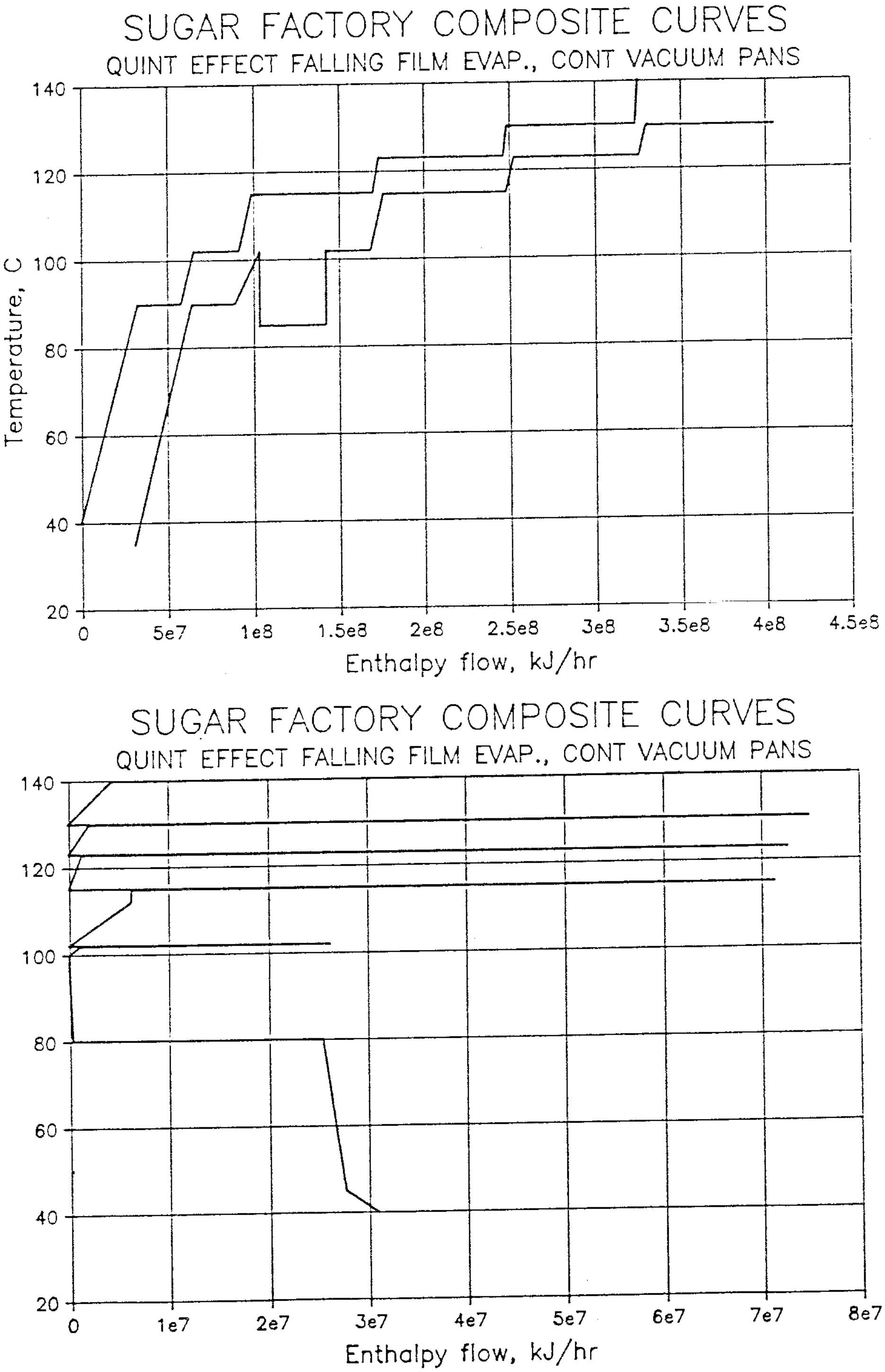


Figure 3.31a and b: Composite and Grand Composite Curves for Sugar Factory Using 5 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Aggregated Condensate Stream Juice Heating (Case 7).

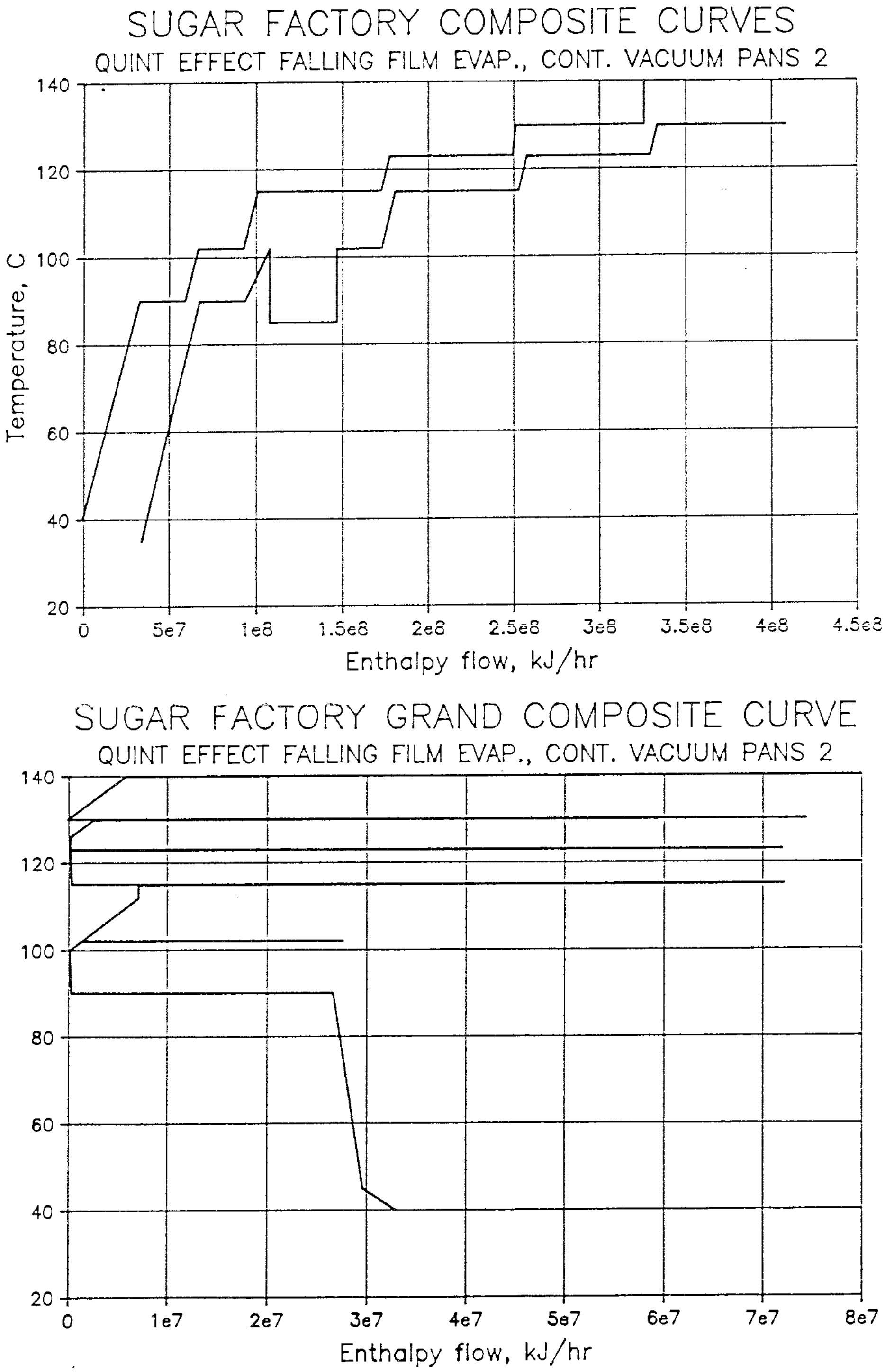
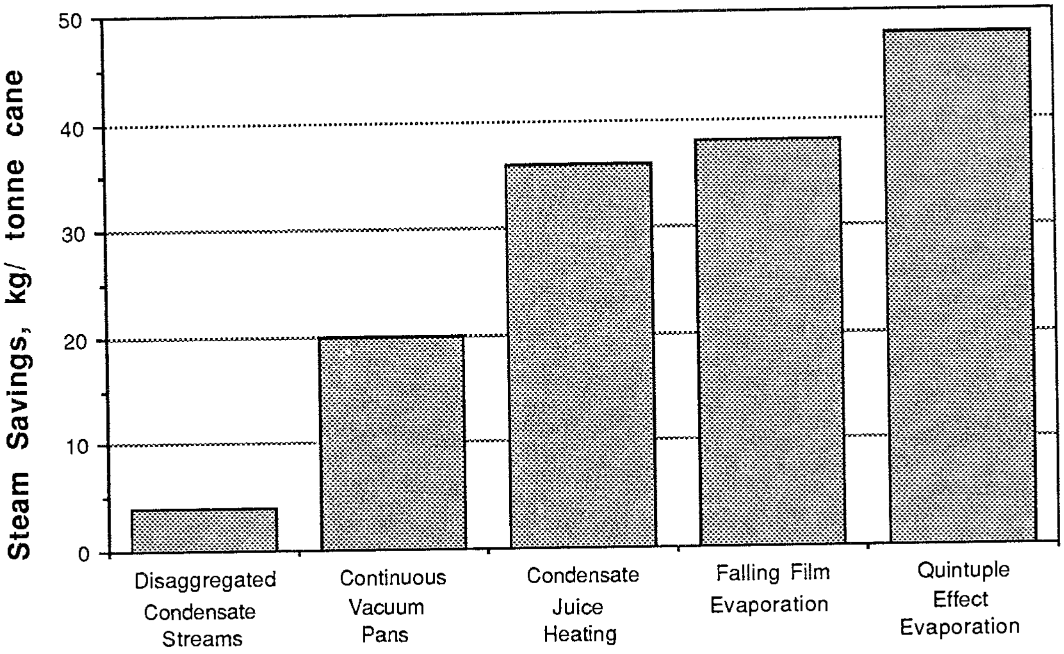


Figure 3.32: Amount of Steam Saved by Various Implementing Various Steam Economizing Measures.



Each column represents the incremental steam savings achieved with the implementation of each particular technology or technique. As the chart shows, the greatest amount of steam savings occurs by adding the fifth evaporator effect. This implies that a heat integrated sugar factory with four effects (falling film) would consume 48 kilograms more steam per tonne cane than a heat integrated sugar factory wit 5 effects (falling film).

Figure 3.33: Flow diagram of Sugar Factory Using 4 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Disaggregated Condensate Stream Juice Heating (Case 4).

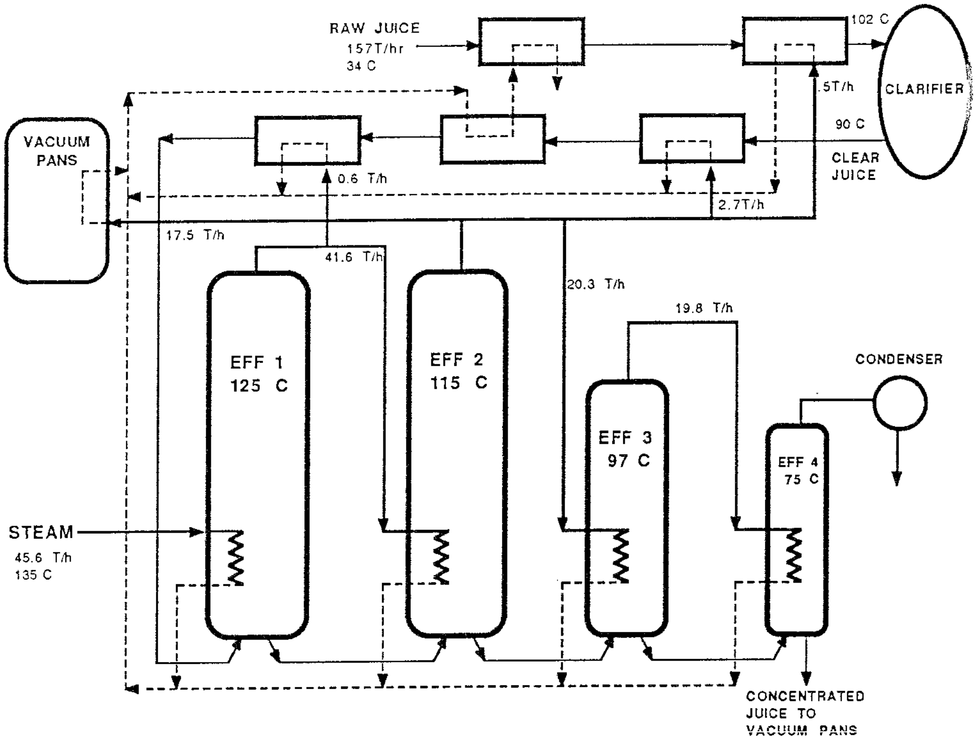


Figure 3.34: Flow Diagram of Sugar Factory Using 4 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Aggregated Condensate Stream Juice Heating (Case 5).

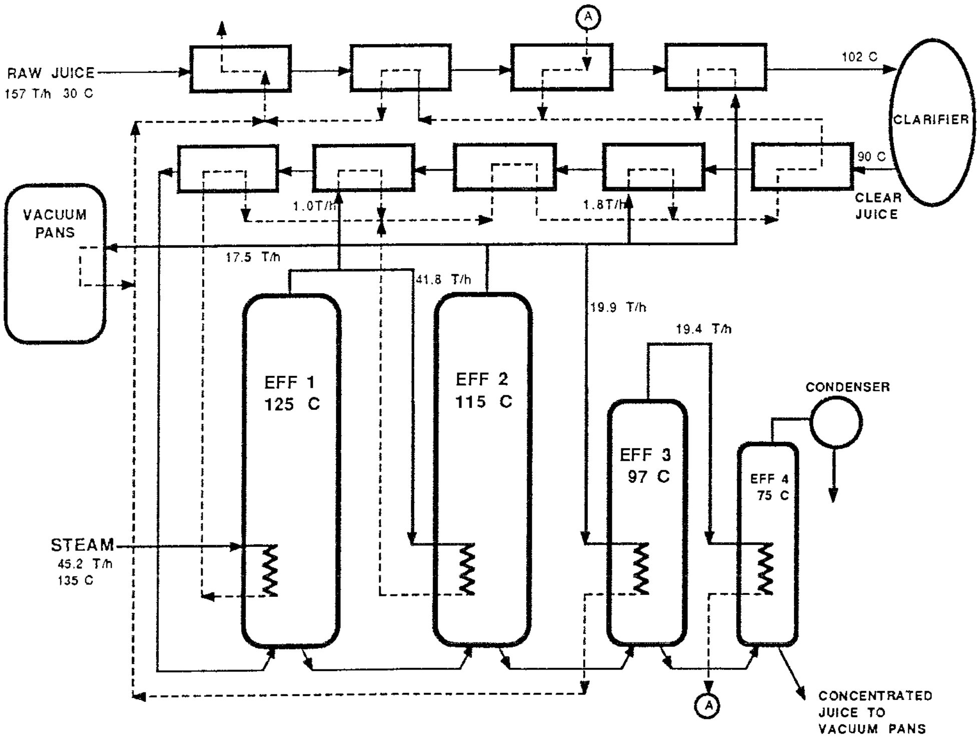
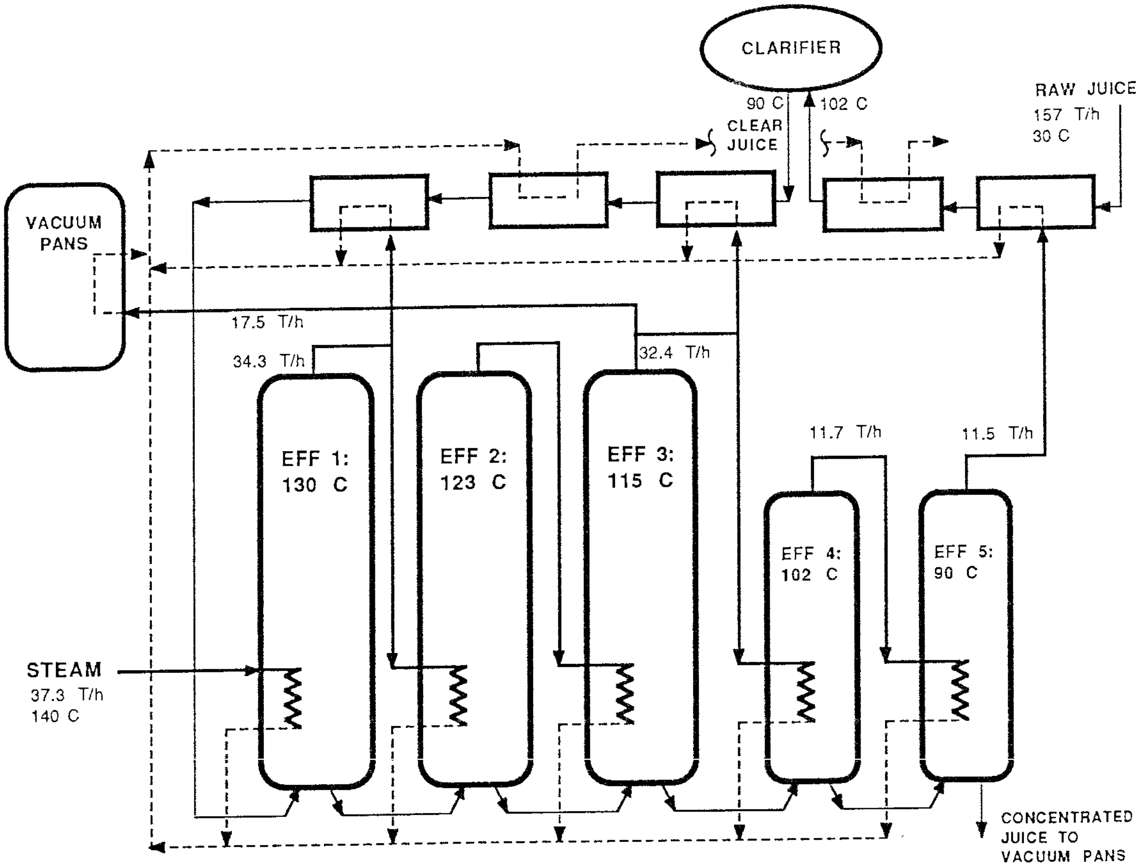


Figure 3.35: Flow diagram of Sugar Factory Using 5 Effect, Falling Film Evaporators, Continuous Vacuum Pans With Aggregated Condensate Stream Juice Heating (Case 7).



Chapter 4:
Economic Analysis of Electricity-Ethanol Co-Production in Autonomous and Annexed Distilleries

1.0 Introduction and Approach

In this chapter, the co-production strategies described in the preceding chapters are evaluated using basic microeconomic principles. The critical differences between this analysis and more traditional ones are the inclusion of high efficiency cogeneration systems producing large amounts of electricity for sale and the treatment of bagasse and barbojo as resources with economic value, specifically as fuels for cogeneration. Without the incentive of cogenerated electricity sales, there has been little reason to value bagasse as anything other than a cheap fuel to burn in the sugar factory and to see barbojo as nothing more than a nuisance when harvesting. When the possibility of exporting electricity from the sugar factory or distillery is considered, the values of these waste-products become evident.

In order to account for the value of the biomass fuels, electricity production is unbundled (i.e. separated for economic analysis) from alcohol production, and in the case of annexed distilleries, from sugar production. To the extent possible, each section of the facility is treated as an independent entity, purchasing the raw materials, labor and capital required for the production of its specific product. Because we seek the optimal use of resources, including capital, labor and biomass, each unbundled section must be able to stand on its own relative to other uses of each particular resource. Market imperfections will not be addressed directly in this chapter, but will be discussed in terms of potential government interventions and roles in Chapter 5.

1.1 The Cogeneration Section

The cogeneration section is treated as a stand alone power plant. As long as the sugar factory/distillery can provide biomass fuel at or below the market price, the cogeneration facility will purchase the fuel from the distillery/sugar factory. It is assumed that the cogenerated electricity is sold to a local utility. Alternatively, the cogeneration facility might also be owned by a local utility, so that the production costs are made internal to the existing structure. Steam sold to the distillery or sugar factory is valued at the opportunity cost of the electricity which could have been generated with that steam. Therefore, the cogeneration facility is indifferent between selling electricity to the grid and selling steam to the neighboring distillery/sugar factory.

1.2 The Autonomous Distillery Section

Many new transactions are introduced by unbundling the distillery (Figure 4.1). First, steam and electricity must be purchased. It is assumed that the neighboring cogeneration facility provides these. Bagasse and barbojo are viewed as a sellable resource, whose value depends on the local circumstances. Bagasse has been used in the past as a feedstock from chemical production, pulp and paper production, or as a boiler fuel for other local industries [1]. The distillery should of course sell the bagasse to whomever values it the most and offers the highest price. The highest value the cogenerator would be willing to pay for the biomass is the market cost of the next cheapest available fuel. The cogenerator is the assumed buyer throughout this analysis.

Because of the price dependence on local conditions, no single bagasse price is assumed. Instead, it is allowed to conform to local conditions, varying between the value of the steam and electricity used in the distillery (the distillery trades the biomass fuel for steam and electricity enough to operate) and $3.00/GJ, the estimated costs for chipped, delivered plantation fuel wood in Brazil [2].

One difficulty which quickly arises when trying to determine profitability of an autonomous distillery is finding the appropriate market price for the alcohol. The value of ethanol depends on its use and competitors. Anhydrous ethanol can be used as an octane enhancing additive to gasoline (gasohol), as which its value on a volume basis is about 16% higher than gasoline [3]. As a motor fuel itself, it has been estimated that hydrous ethanol should be valued on a volume basis at ~80% of gasoline [4]. Throughout this analysis, the present (1989) wholesale gasoline price and the projected gasoline prices in 2000 and 2010 are used as references.

1.3 The Sugar Factory/Annexed Distillery Sections

The sugar factory and annexed distillery are handled very similarly to the autonomous distillery, however additional interactions between the two sections must be introduced (Figure 4.2). Steam and electricity are still purchased from the cogeneration facility. Bagasse and perhaps barbojo are purchased by the cogeneration facility. The sugar factory sells molasses at the market price to anyone who will purchase it. If the adjacent annexed distillery can produce ethanol competitively at the market price of molasses then it will buy the molasses. If it cannot produce ethanol competitively at the market molasses price, then the molasses is sold on the market and no ethanol is produced.

An alternative way of looking at this is when the same party owns the sugar factory and distillery. The only decision the owner faces concerning ethanol production is whether it is more profitable to sell the molasses directly or to ferment it into ethanol and sell the ethanol. This decision is independent of both sugar and bagasse/barbojo production and revenue.

A minor complication arises because of the differing steam needs of the sugar factory and annexed distillery. The sugar factory purchases "high" pressure steam (2 MPa) to run the cane mills. It then uses a portion of the low pressure exhaust steam from the mills to supply the evaporators, vacuum pans, etc. The distillery requires only the low pressure exhaust steam, which must be purchased from the sugar factory. It is assumed that low pressure steams is assigned the same dollars per kilogram as the higer pressure steam. (This simplifying assumption does not effect outcome). An alternate way of looking at this is splitting the overall steam costs between the sugar factory and distillery per their respective low pressure steam use.

1.4 Sugar Factory/Annexed Distillery with Autonomous Distillery Capacity

This case addresses the possibility of being able to produce either all sugar (plus molasses) or all ethanol, depending on the market circumstances. This is accomplished by having a sugar factory with an annexed distillery capable of fermenting and distilling all of the cane juice produced in the mills. The transactions between the cogeneration facility and the sugar factory/high capacity annexed distillery are the same as before. The critical difference in this analysis in determining under what conditions it is desirable to produce which products (sugar, ethanol, molasses), and the effect of the added capital costs for the large capacity annexed distillery on the profitability of all of the products.

2.0 Autonomous Distillery Analysis

2.1 Assumptions and Approach

Rather than performing a specific case study on an autonomous distillery, a more general approach is taken. An envelope of possible production costs for both alcohol and electricity is created by assuming high and low values of costs and operating parameters (Table 4.1). The sensitivities of alcohol and electricity production costs to each of parameters are also identified, which creates a set of screening criteria which can quickly identify locations and circumstances under which co-production would be desirable.

The edges of the envelope of operating and cost parameters come from numerous sources. The high costs are based on a study by F.C. Schaffer and Associates, Baton Rouge, for a hypothetical sugar factory/distillery in Costa Rica [5]. The low cost edge is based on actual Brazilian conditions [6], and was corroborated

Table 4.1: Major Assumptions for Economic Analysis.

	Autonomous Distillery	Sugar Factory		Cogeneration
Product	Hyrous ethanol	Sugar		Electricity
Capacity	300,000 l/day	430,000 kg/day	↔	27 MW CEST, 45 MW BIG/STIG,
	600,000 l/day	860,000 kg/day	↔	110 MW BIG/ISTIG
Capital cost,	18-48 million	44 million	↔	$1500/kW, CEST, $990/kW BIG/STIG,
	36-96 million	88 million	↔	$770/kW, BIG/ISTIG
Discount rate, life	12%, 20 yrs.	12%, 20 yrs.		12%, 30 yrs.
Feedstock costs	$8-18/tonne cane	$8-18/tonne cane		up to $3/GJ total fuel cost.
Operating costs/yr	$1.1-3.1 million	$3.4 million	↔	$1.6 mil. CEST, $2.3 mil. BIG/STIG
	2.2-6.2 million	$6.8 million	↔	$4.3 million, BIG/ISTIG
Historical: Product prices	14-20¢/l neat	8-50¢/kg		4-8¢/kWh
Price trends	increasing	cyclical		increasing, eventually stabilizing

by American and Brazilian sugar industry researchers and manufacturers [7,8,9].

2.1.1 Physical Operating Parameters

Cogeneration Operating Parameters

The three cogeneration technologies discussed in Chapter 2 were considered: the condensing, extraction steam turbine, (CEST), the biomass integrated gasifier /steam injected gas turbine (BIG/STIG), and the biomass integrated gasifier /inter-cooled steam injected gas turbine (BIG/ISTIG). A summary of the estimated performances of these three cogeneration systems at different scales is shown in Table 4.2 [10]. The "cogeneration mode" columns show the electrical output, efficiency and fuel consumption with the maximum amount of steam going for process use. The "power only mode" columns show the same parameters (save steam production) when no process steam is required and all of the steam is consumed in power production.

Because the distillery steam demand of 165 kg of steam per tonne of cane milled (kg/tc) assumed in the analysis is much less than the steam available in full cogeneration mode, the performance of the cogeneration systems in the cogeneration mode is scaled from the nominal values in Table 4.2. The percentage of the fuel converted into electricity (apparent electrical efficiency) in this reduced cogeneration mode is scaled between full power and maximum cogeneration modes by the ratio of steam being exported (per tonne cane) by the maximum amount of steam produced in full cogeneration mode:

Table 4.2: Performance of Biomass-Fired Cogeneration Systems.

	COGENERATION						POWER ONLY			
	Electricity,		Steam,		Fuel,	Cane,[c]	Electricity,		Fuel,	Cane,
	MW	%HHV	T/H	%HHV	T/H	T/H	MW	%HHV	T/H	T/H
CEST[a]										
Generic	17.5	13.0	65.6	35.9	50.8	169	27.0	20.3	50.2	167
Generic	6.1	11.4	26.4	36.4	20.2	67	10.0	17.8	21.2	71
Generic	1.8	10.1	9.0	37.2	6.73	22	3.0	15.7	7.22	24
BIG/STIG[b]										
LM-5000	38.8	31.3	47.7	30.0	27.6	157	53.0	35.6	33.0	188
LM-1600	15.0	29.8	21.8	33.8	11.2	65	20.0	33.0	13.2	75
GE-38	4.0	29.1	5.7	32.4	3.06	17	5.4	33.1	3.63	21
BIG/ISTIG[d]										
LM-8000	97	37.9	76.2	25.4	57.3	325	111.2	42.9	57.7	328

Notes for Table 4.2:

[a] Adapted for E.D. Larson and R.H. Williams, "Biomass-Fired Steam Injected Gas Turbine Cogeneration," Proceedings 1988 ASME Turbo-Cogen Symposium, Montreaux, Switzerland, Aug. 30-Sept. 1, 1988.

[b] Estimates for the LM-5000 are based on private communications from M. Erbes, General Electric Corp. to E.D. Larson. Other estimates are adapted from Table 3 in [10], assuming that the gasification efficiency is the same for biomass as for coal (Table 2 in [10]).

[c] Assuming that the BIG/STIG and BIG/ISTIG use briquetted bagasse or barbojo with moisture content 15%, which has a higher heating value of 16,166 kJ/kg. CEST uses 50% wet bagasse having a higher heating value of 9350 kJ/kg. We further assume that 300 kg of 50% wet bagasse are produce per tonne of cane milled, or 176 kg of 15% wet briquetted bagasse are produced per tonne cane milled.

[d] Preliminary estimate of steam and electricity production, based on performance with coal. From J.M. Ogden and M.E. Fulmer, Assessment of New Technologies for Co-Production of Alcohol, Sugar and Electricity from Sugar Cane, PU/CEES report no. 250, May, 1990.

$$\%HHV_{\text{elect, cogen}} = \%HHV_{\text{max cogen}} + \left\{ \frac{\text{Steam}_{\text{actual}}(\%HHV_{\text{full pwr}} - \%HHV_{\text{max cogen}})}{\text{Steam}_{\text{max cogen}}} \right\}$$

The scaling required for the 165 kg/tc process steam demand assumed in the analysis raises the electrical generating efficiency for the three cogeneration systems up to 17.2%, 33.2%, 39.4% for CEST, BIG/STIG and BIG/ISTIG, respectively. The power output in this reduced cogeneration mode is scaled in the same manner. The power output during the milling season is assumed to be 101.5MW, 45.4MW and 23.0MW for BIG/ISTIG, BIG/STIG and CEST cogeneration, respectively.

Distillery Operating Parameters

The steam use of the autonomous distillery is based on the HP-1234 case presented in Chapter 3, Section 3.1.2. This case assumes a distillation column operating at .60 MPa, whose condenser serves as the heat source for the highest temperature effect (120°C) of a quadruple effect evaporator. The more steam efficient case where the fermenter feed was not preheated was not used because of the uncertainties in milling steam requirements and the increased fermenter and mill capital costs (Chapter 3, Section 3.3). The minimum low pressure steam demand of such a system was estimated to be 155 kg steam per tonne cane (kg/tc, Table 3.3). Ten kg/tc of steam is added to account for losses.

The electricity use for the autonomous distillery is assumed to be 13 kWh per tonne cane milled, a typical value for sugar factories and distilleries[a]. There is no doubt that using improved control techniques and electricity conserving technologies could reduce this value further, however, because the overall cost of

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[a] The typical sugar factory or distillery actually consumes around 20kWh/tc, however, ~7kWh/tc is consumed in the powerhouse, which is now assumed to be part of the cogeneration section.

electricity is orders of magnitude less than other ethanol production costs these improvements would have minimal affect on the overall cost of ethanol production.

The bagasse and barbojo yields are assumed to be 300 kg of 50% moist bagasse and 660 kg of 50% moist barbojo per tonne cane milled. Ethanol yields are taken to be 70 liters per tonne cane for the high cost scenario [5] and 73 liters per tonne cane for the low cost scenario [6]. While values as high as 80 liters/tc and as low as 65 liters/tc can be found in the literature [11,12] and are used in the sensitivity studies, the yields used in the base analyses can reasonably be expected in a modern, autonomous distillery.

The distillery capacity is assumed to be approximately 4000 tonnes of cane milled per day (tc/day; approximately 290,000 liters/day) for distilleries with CEST or BIG/STIG cogeneration and approximately 8000 tonnes cane per day (~580,000 liters/day) for a distillery matched to BIG/ISTIG[b]. The 4000 tonne cane per day size is typical for a moderate size autonomous distillery and was the capacity assumed in the Schaffer study from which the high cost scenario values are taken[c]. The distillery is assumed to operate on average at 90% of the rated capacity.

2.1.2 Costs

Cogeneration Costs

Table 4.3 presents a summary of capital, operation and maintenance costs

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[b] The actual capacities were adjusted so that the bagasse met the in-season fuel requirement of the cogeneration system. When BIG/ISTIG was used, the actual grinding capacity was 3980 tc/day, for CEST 4050 tc/day and for BIG/ISTIG, 7780 tc/day.

[c] While a majority of sugar factories worldwide mill between 1000 and 4000 tc/day, over 280 factories have grinding capacities greater than 4000 tc/day and more than 75 are rated over 8000 tc/day [20].

Table 4.3: Capital and Operating Costs of Biomass-Fired Cogeneration Systems (1986 $).

	Capacity, MW	Installed Cost, $/kW	Maintenance		Labor, (1000$/Y)
			Fixed, 1000$/Y	Variable, $/kWh	
CEST[a]					
Generic	27.0	1556	664	0.003	129.2
Generic	10.0	2096	246	0.003	97.2
Generic	3.0	3008	73.8	0.003	97.2
BIG/STIG[b]					
LM-5000	53.0	990	1304	0.001	297.0
LM-1600	20.0	1230	492	0.001	108.0
GE-38	5.4	1650	133	0.001	97.2
BIG/ISTIG[c]					
LM-8000	111.2	770	2736	0.001	405.0

Notes for Table 4.3:

[a] See note a, Table 4.2.

[b] E.D. Larson R.H. Williams, "Biomass-Fired Steam-Injected Gas Turbine Cogeneration," Proceedings of the 1988 ASME Cogen-Turbo Symposium, Montreux, Switzerland, Aug.30-Sept. 1, 1988.

[c] Based on personal communication with E.D. Larson.

for the three cogeneration systems at the same sizes shown in Table 4.2. The 27 MW generic CEST, LM 5000 BIG/STIG and LM8000 BIG/ISTIG were assumed throughout the analyses.[d] The BIG/(I)STIG costs are based on estimates for coal integrated gasifier STIG systems [10]. (See Appendix IV for disaggregated BIG/STIG and BIG/ISTIG capital costs.)

A 12% discount rate and a 30 year facility life are assumed for all three cogen-eration technologies in the base case scenarios. In the sensitivity studies, the discount rate is allowed to vary from 10% to 20%, while the plant life varies from 20 to 40 years. The labor cost estimates are based on typical man-hours per kilo-

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[d] Unless otherwise noted, all costs and revenues are in 1987 US$.

watt for fossil fuel steam-electric plants with Jamaican wage conditions applied [14]. Fixed operating costs include annual insurance costs of 0.5% of installed capital costs in addition to operating and maintenance costs from Table 4.2.

Distillery Costs

The overall capital cost for the high cost scenario was assumed to be $48 million (Table 4.4). The capital cost for the low cost scenario was $18 million, and was based on a 1644 tonne cane per day distillery and scaled up linearly to 4000 tonnes cane per day (Table 4.5). This scaled value fell within the range quoted by Brazilian manufacturers for distilleries of this size (Figure 4.3). Both the high and low capital costs were scaled linearly up to $35.31 million and $94.85 million, respectively, for 7780 tonne cane per day distillery matched to the BIG/ISTIG cogeneration system.

The higher cost case assumes that US equipment and engineering is used, while the lower cost case assumes Brazilian technology. It is believed that the Brazilian costs are so much lower because of manufacturing experience gained from the Brazilian National Alcohol Fuels Program (production economies of scale, equipment standardization and experience) and the different engineering and materials standards found in Brazil [22]. In fact, Brazil is now a prominent exporter of distillation equipment, implying that the low costs used here could reasonably be duplicated elsewhere.

The high and low estimates for operation and maintenance (O & M) costs came from the same sources as the capital costs. The fixed costs were scaled in the same manner as the capital costs. It should be noted that the two sources for the O & M costs did not always treat the fixed and variable costs in the same manner. For instance, the low cost Brazilian estimates treated the bulk of the maintenance costs as fixed, while the higher cost Schaffer study treated most of the maintenace costs as variable. Thus, the relatively small difference in the fixed

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TABLE 4.4: High capital and operating cost estimates for 4000 tonne cane per day milled (290,000 liters per day) autonomous distillery (not including power house equipment; i.e. generator, boiler, turbines, etc.).[a]

CAPITAL COSTS:

ITEM	COST, $
Cane handling	661,000
Milling	5,939,000
Juice Processing	2,750,000
Water/Chemical treatment	634,000
Fermentation/Prefermentation	2,887,000
Distillation	1,700,000
Warehouse, shop, office facil.	1,680,000
Plantwide piping	4,000,000
General Services	2,694,000
Misc.	466,000
Subtotal, industrial facilities	23,411,000
Labor to erect	15,222,000
Contractor fee, overheads & profits	4,674,000
Shipping, ocean freight	1,869,000
Engineering fee @ 7.5% of above	3,388,000
Total Erected Cost	$48,564,000

OPERATING COSTS:

ITEM	ANNUAL FIXED COSTS, $	VARIABLE COSTS, $/tc
Wages & Salaries	509,000	.00
Repair Materials & Parts	53,000	1.15
Chemicals & Lubricants	7,000	1.16
Equipment rental	10,000	0.02
Vehicles (incl. maintenance)	50,000	0.19
Cane costs	0	18.87
Other/ Misc.	37,000	0.23
Subtotal	666,000	22.13
Administrative Costs	500,000	0.10
Total Operating Cost	$1,166,000	$21.62/tc

TABLE 4.5: Low capital and operating cost estimates for 4000 tonne cane per day milled (290,000 liters per day) autonomous distillery (not including power house equipment; i.e. generator, boiler, turbines, etc.).[b]

TOTAL CAPITAL COST:	$18,083,000

FIXED OPERATING COSTS:

ITEM	COST, $
Labor	560,000
Maintenance	362,000
Supply	36,000
Insurance	90,000
Total Fixed Cost	$1,048,000

VARIABLE OPERATING COSTS:

ITEM	COST, $/TC
Products	0.084
Cane	8.09
Chemicals/Misc.	0.092
Total Variable Cost	$8.226/TC

Notes for Tables 4.4 and 4.5.

[a] F.C. Schaffer and Associates, "Evaluation and Technical and Financial Assessment, Central Azucerera Tempisque, S.A. (CASTA)," report to CODESA/USAID, Vol 1, July, 1987.

[b] Goldemberg, J., J.R. Moreira, P.U.M. Dos Santos, and G.E. Serra, "Ethanol Fuel: A Use of Biomass Energy in Brazil", *Ambio*, Vol. 14, No. 4-5, 1985. Adjusted to 1987 US$ using Gross National Product Deflator (Economic Indicators, March, 1988) and scaled up linearly from 1764.7 tonnes cane per day (120,000 liter per day) autonomous distillery.

O & M costs ($1,047,000 versus $1,393,000) and the particularly large difference in the variable O & M costs ($0.15/tc versus $2.85/tc) is chiefly due to accounting differences rather than any more intrinsic factors.

The cost of cane delivered to the factory is assumed to be $18.87 per tonne in the high cost scenario and $8.07 per tonne in the low cost scenario (from the same sources the capital costs). These values represent the range of cane costs which can be expected throughout the developing world[e]. Cane costs vary with soil fertility, crop yield, level of mechanization in agricultural practice, local wage rates and transportation costs, to name a few. The cost to grow and harvest cane can be lowered to some extent by improved agricultural practices, but most of the factors tend to be specific to the location. Therefore, the low Brazilian cane costs are not as likely to be duplicated as the low Brazilian capital costs.

2.1.3 Analysis Method

The economic analysis is performed using a spreadsheet model constructed using Supercalc 4. The spreadsheet allows the user to input all of the pertinent performance and cost information for both the distillery and cogeneration sections, and calculates the overall and average production costs of both products along with numerous intermediate variables. A sample spreadsheet is shown in Figure 4.4. Details of all the calculations performed in the spreadsheets appear along with the spreadsheets in Appendix VI.

2.2 Results

2.2.1 Baseline Results

Figure 4.5 shows ethanol production cost versus electricity production cost for each of the three cogeneration technologies being considered with high and low

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[e] Cane costs in Louisiana and Florida can approach $20/tc.

distillery cost assumptions. It is assumed that the distiller sells bagasse during the milling season to the cogenerator, from whom he purchases electricity and steam. The cogeneration facility is also assumed to operate during the off-season as an independent power producer.

For each of the six cases, two lines are shown. The upper line is the case where the distiller receives revenue from the sales of bagasse only; the cogeneration facility still operates off season on either barbojo or some other fuel of equal cost to the cogenerator such as plantation fuel wood, coal or oil. The lower line in each pair represents the case where the distiller receives revenue from the sale of both bagasse and barbojo. This occurs in vertical integration, where the same party owns the cane fields and the distillery.

The upper left end point of each line represents the case where no net money is transacted: the value of the biomass fuel sold to the cogenerator equals the value of the steam and electricity sold to the distiller. This point is approximately equivalent to a standard distillery without cogeneration. The overall fuel costs at this endpoint (including all fuel transportation and processing [14]) for the cogeneration facilities are ~\$1.50/GJ for BIG/STIG and BIG/ISTIG and \$0.85/GJ for CEST cogeneration.

Increasing fuel costs to the cogenerator (and revenues to the distiller) are expressed by moving to the right along each line. As the fuel price increases, average electricity production costs rise while average ethanol production costs fall. (It is assumed that the economic rents accrued by the distiller from bagasse and barbojo sales are credited towards alcohol production.) The right hand end point of each line represents the case where the cogenerator pays \$3.00/GJ for his fuel (including processing costs), the estimated cost of dried, chipped and delivered plantation fuel wood in Brazil [2]. For locations where alternative fuels are higher than \$3.00/GJ, the lines can be extended to the right beyond these points.

There are significant differences in electricity production costs among the three cogeneration systems and with the different fuel costs. When no fuel revenues are credited to alcohol production (left end points), the electricity busbar cost ranges from 3.0¢/kWh for BIG/ISTIG to 3.8¢/kWh for BIG/STIG to 4.8¢/kWh for CEST cogeneration. When the biomass fuels are valued at $3.00/GJ, busbar costs rise to 4.2¢/kWh for BIG/ISTIG, 5.2¢/kWh for BIG/STIG and 8.5¢/kWh for CEST.

There is also large spread in ethanol production costs. For the cases where no net fuel revenue is earned (left hand end point of each line), estimated ethanol production cost is about 20¢/liter (using low production cost estimates); when the higher costs estimates are used, the average ethanol production cost is around 50¢/liter. A breakdown of these cost differences is discussed in the sensitivity analysis in Section 2.2.2.

For co-production to be economically viable, both ethanol and electricity must be competitive in their respective markets. Figure 4.6 represents the low ethanol production costs estimate scenarios with each cogeneration system with a number of competing reference lines. Line E1 (3.9 cents/kWh) shows the fuel and operating costs of an existing heavy oil burning steam-electric generating plant characterized by parameters typical for a developing nation.[f] Line E2 (4.25 cents/kWh) represents the cost of hydroelectricity in Brazil[g], while line E3 (5.3 cents/kWh) represents the cost of electricity from a new, coal burning power plant.[h]

Line A1 shows the value of ethanol as a neat transportation fuel (24¢/liter ethanol, ~29.6¢/l gasoline) with wholesale gasoline at the US DOE estimated 2010 wholesale gasoline price [16]. Line A2 shows the value of neat ethanol at the

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[f] Heat rate of 13,120kJ/kWh, fuel cost of $2.63/GJ.

[g] $1500 per installed kilowatt, including capital cost of generating equipment and transmission from Amazon to population centers in southeast Brazil. Assumes discount rate of 10%.[15]

[h] $1400/installed kw and 10% discount rate with coal at $1.70/GJ.

estimated 2000 wholesale gasoline price (19.5¢/l ethanol, ~24¢/l gasoline), while line A3 shows the value of neat ethanol at 1989 US wholesale price (13.8¢/liter ethanol, ~15.2¢/l gasoline)[16].[i] While on an energy basis, one liter ethanol is worth less than .7 liters of gasoline, when the higher octane of ethanol and the improved performance of ethanol burning engines are factored in, the ethanol/gasoline ratio increases to 0.81 [3].

A number of regions are created on the plot by these reference lines. For instance, the area to the left of E1 and below A1 shows average production cost combinations where both ethanol and electricity could be produced competitively with electricity generated from an existing oil fired power plant (operating costs only) at 1989 oil (an gasoline) prices. The only co-production line which falls within this region is when low ethanol production costs are achieved in conjunction with BIG/ISTIG cogeneration. In Brazil (where hydro-electricity is the alternative electricity source) with 2000 projected oil prices, both ethanol and electricity could be produced competitively using either gas turbine technology with either bagasse or bagasse and barbojo revenues accrued by the distillery.

Another viable scenario is where the cogeneration facilities operate only during the milling season. This might be the case in southeast Brazil, where the dry season causes a trough in the hydroelectric supply, but also corresponds to the sugar cane milling season (Figure 4.7) [17]. By filling in the supply trough, the marginal value of the cogenerated power far exceeds the long run marginal cost of new "firm" hydroelectricity. This is because existing rainy season hydroelectric supply exceeds firm power (power available year round).

The value of the cogenerated electricity can be calculated from:

$$V_{FH} = LRMC_{hydro} = \alpha V_{cogen} + (1-\alpha)SRMC_{hydro}$$

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[i] As a rule of thumb, wholesale gasoline prices can be related to world crude oil prices by: GAS ($/gal) = 0.0273 OIL($/bbl) + 0.05.

where:

V_{FH} = Value of firm hydroelectric power.
$LRMC_{hydro}$ = Long run marginal cost of Amazon hydro, 4.25¢/kWh [14]. This is the production cost of hydroelectric power (reference line E2).
α = Fraction of the new firm power actually provided by cogeneration, .25.(shaded region of Figure 4.7)
V_{cogen} = Value of the cogenerated power, to be solved for.
$SRMC_{hydro}$ = Short run marginal cost of Amazon hydro, .34¢/kWh (operating costs) [14].

In other words, the value of firm electricity from a new source (assumed hydroelectric) equals the fraction of the new supply provided by cogeneration times its value plus the fraction of the new capacity provided by existing "non-firm" hydroelectric power times the cost of providing that electricity (hydro operating cost). In this way, the *value* of the in-season cogenerated electricity can be found. Solving for the V_{cogen}, the in-season cogenerated power is worth almost 16¢/kWh! Thus, a utility would be indifferent between building new hydroelectricity facility (producing power at 4.25¢/kWh) or paying the cogenerator 16¢/kWh for the dry season cogenerated power to the extent that the "hydro-trough" can be filled.

Figure 4.8 shows the production cost of electricity and ethanol assuming only in-season power generation. The figure clearly shows that even with CEST cogeneration, production costs are far below the 16¢/kWh at which the additional power would be valued. If the milling season, however, extends beyond the dry season trough, the value of the electricity averaged over the milling season would have to take into account these periods when the alternative electricity sources is the existing, non-firm hydroelectricity, whose marginal cost is quite low. Therefore, the electricity production during the dry season should be maximized by perhaps adding supplementary firing in the boiler or HSRG or adjusting the milling season and distillery and cogeneration capacity so that the plant need operate only during the hydro-trough.

2.2.2 Sensitivity Analysis

Autonomous Distillery Sensitivities

As was shown in the previous section, the envelope encompassing possible ethanol production costs varies from economically undesirable under almost all circumstances to feasible under present day conditions. This section identifies the critical parameters and quantifies the sensitivity of the overall ethanol production cost to those critical parameters.

A summary of the sensitivities of autonomous distillery ethanol production costs are shown in Table 4.6. For each parameter, the high and low values are shown, along with the nominal difference in average ethanol production cost between the high and low estimates of the parameter. The last column shows the change in ethanol cost with a change in the parameter equal approximately to 10% of the range from the assumed high value to low value.

One of the most critical ethanol production parameters was discussed in the previous section: fuel credits. As was shown, ethanol production costs could range from marginally attractive levels (20¢/liter) down to less than 10 cents per liter merely by attaching a reasonable market value to the bagasse and barbojo.

Another important parameter is the cost of the sugar cane. Given nominal season length and ethanol yield, each dollar increase in cane costs increases ethanol production costs by 1.37¢/liter. Over half the 31¢/liter difference between the high and low costs scenarios in Section 2.2.1 (16.5¢/liter) can be accounted for by the difference in cane costs.

Overall capital cost is the next most important parameter effecting ethanol production costs. Assuming a discount rate of 12%, each additional $2.5 million dollars of capital expenses implies over three-fourths of a cent per liter increase in

Table 4.6: Sensitivity of Ethanol Production Costs of Ethanol Produced in an Autonomous Distillery.

Parameter	Low Value	High Value	ΔEtOH cost w/ change from low to high value of param (cents/liter)	$\frac{\partial(\text{EtOH cost})}{\partial(\text{param.})}$	Units
Cane cost	$8/tc	$20/tc	16.44	1.37	(¢/liter)/($/tc)
Economic Rents on Fuel	$0.00/GJ	$2.45/GJ*			
Bagasse only			-6.05	-0.62	(¢/liter)/($.25/GJ)
Bagasse and barbojo			-16.79	-1.71	(¢/liter)/($.25/GJ)
Capital Cost	$18 million	$48 million	7.73	0.64	(¢/liter)/$2.5 million
Disc. Rate	10%	25%			
High Capital Costs			15.46	0.95	(¢/liter)/(1% IRR)
Low Capital Costs			5.80	0.36	(¢/liter)/(1% IRR)
Season Length	120 days	220 days	-4.65	-.55	(¢/liter)/(10 days)
Yield	65 liter/tc	80 liter/tc	-4.07	-.27	(¢/liter)/(liter/tc)
Variable Costs	$0.15/tc	$2.85/tc	3.70	0.34	(¢/liter)/($0.25/tc)
Fixed Costs	$1,047,000	$1,393,000	0.82	0.19	(¢/liter)/($100,000)
Distillery Life	15 years	30 years			
High Capital Costs			-2.59	-.22	(¢liter)/(year life)
Low Capital Costs			-0.97	-.08	(¢/liter)/(year life)

* Actual price paid to the cogenerator for the biomass fuel minus the value of the steam and electricity which the distillery consumes.

ethanol production cost. The $20 million difference in capital costs between the high and low cost scenarios of the previous section make up 9.35¢/liter of the 31¢/liter difference (28%).

The effect of the milling season on ethanol production cost is tied to the fuel credits received by the distillery. With small credits, lengthening the milling season (same daily cane throughput) spreads the fixed costs over more liters of ethanol, lowering average production costs (Table 4.6). If no economic rents are accrued from fuel sales, every additional 10 days of milling lowers the average ethanol production cost by 0.57¢/liter. However, when the net fuel costs to the cogeneration facility are $3/GJ (assuming BIG/ISTIG cogeneration and both bagasse and barbojo sales), the ethanol production cost *increases* 0.40¢/liter per ten day increase in season length. This is because the revenue from the reduced off season barbojo sales more than compensates for the higher capacity factor on the distillery (Figure 4.9). While ethanol production costs can be significantly affected either positively or negatively by milling season length, the electricity production costs are only slightly affected.

One advantage of the shorter milling season which is not shown in the figure is the efficient use of the barbojo resource. Using BIG/ISTIG cogeneration, during the 205 day off-season of a 160 day milling season, only about 60% of the barbojo available is used. If the season length is shortened to 133 days per year, then 80% of the barbojo resource is used. This raises the electricity generated per tonne cane milled up from 730 kWh/tc (160 day milling season) to nearly 900 kWh/tc (Figure 4.10). While the costs of barbojo retrieval, its agricultural value and the environmental benefits of not burning the fields would have to first be carefully calculated before extensive barbojo retrieval is undertaken, barbojo is a resource which should not be overlooked.

Sensitivities to yield, fixed and variable operating and maintenances costs and distillery life (over which the capital costs are recovered) are included in Table 4.6 but are not significant relative to the parameters already addressed.

Cogeneration Sensitivities

Although high and low cogeneration cost scenarios were not developed as they were for the autonomous distillery, the sensitivity of cogenerated electricity to the key economic parameters is important. The sensitivity analysis of the cogeneration section was performed the same way as the autonomous distillery. For a given parameter, a range was defined which would encompass high and low estimates for that parameter. The change in average electricity production cost with the high to low range of the parameter in question is determined, along with the derivative (taken at a typical value) of the average electricity production cost with respect to a 10% change in the parameter being analyzed.

These results are shown in Table 4.7. The parameter having the largest effect on the average production cost of electricity is the fuel price. This is not unexpected seeing fuel costs make up from one to two thirds of the overall production cost of electricity. The sensitivity to fuel cost runs inversely with cogeneration efficiency, with the average electricity cost of the very efficient BIG/ISTIG system varying 0.9 cents per $1.00/GJ change in fuel cost and a 1.9 cent variation per $1.00/GJ change in fuel cost for the much less efficient CEST cogeneration system. The assumed discount rate also has a large effect on electricity production cost. Assuming that the capital recovery takes place over the 30 year life of the facility, a change of 1% in the assumed discount rate results in a .1 to .2 cent per kWh change in average electricity production cost.

Effect of Variation of Critical Parameters on Co-Production Competitiveness

In the previous sections, variation in cane cost and capital costs were identified to be the most important parameters affecting autonomous distillery ethanol production, and fuel cost was identified to the critical parameter in electricity production. In this section, a number of intermediate combinations of these parameters are assumed in order to identify some of the cut-offs between competitive and non-competitive ethanol and electricity co-production.

This is shown in Figures 4.11 through 4.14. As in the earlier electricity-ethanol production cost figures, each line shows the change in production cost with fuel price, ranging from a direct barter of fuel for steam and electricity up to a net fuel cost (including processing) of \$3.00/GJ. In Figures 4.11 and 4.12, families of curves show varying distillery capital cost and cane costs with the distillery receiving revenues from both bagasse and barbojo. Figures 4.13 and 4.14 are the same except that the distillery receives revenues only from bagasse sales. The same reference lines which appear in Figure 4.6 are included in all of the figures.

Table 4.7: Sensitivity of Average Electricity Production Costs for Electricity Produced in Cogeneration Facility.

Parameter	Low Value	High Value	ΔElect. cost w/ change from low to high value of param (cents/kWh)	$\frac{\partial(\text{EtOH cost})}{\partial(\text{param.})}$	Units
Fuel Cost	\$1.00/GJ	\$3.00/GJ			
BIG/ISTIG			1.74	0.09	(¢/kWh)/(\$0.10/GJ)
BIG/STIG			2.07	0.10	(¢/kWh)/(\$0.10/GJ)
CEST			3.80	0.19	(¢/kWh)/(\$0.10/GJ)
IRR	10%	25%			
BIG/ISTIG			1.46	0.09	(¢kWh)/(1%IRR)
BIG/STIG			1.93	0.12	(¢/kWh)/(1%IRR)
CEST			3.05	0.19	(¢/kWh)/(1%IRR)
Facility Life	20 years	40 years			
BIG/ISTIG			0.13	0.01	(¢/kWh)/(year life)
BIG/STIG			0.17	0.01	(¢/kWh)/(year life)
CEST			0.27	0.01	(¢/kWh)/(year life)
Fixed O&M	\$200,000 maximum variation				
BIG/ISTIG			0.02	neglig.	(¢/kWh)/(\$10,000)
BIG/STIG			0.05	neglig.	(¢/kWh)/(\$10,000)
CEST			0.10	neglig.	(¢/kWh)/(\$10,000)
Season Length	120 days	220 days			
BIG/ISTIG					
BIG/STIG			0.05	neglig.	(¢/kWh)/(10 days)
CEST					

Many competitive scenarios can be devised when BIG/(I)STIG cogeneration is employed and the sale of bagasse (and perhaps barbojo) can be credited to ethanol production. Even under the most stringent conditions considered (electricity from new cogeneration facilities competing with existing oil fired power plants and ethanol competing with wholesale gasoline at the 1989 price), cane cost can be up to $11 per tonne or capital costs as high as $28 million and still remain competitive using BIG/ISTIG cogeneration. Scenarios using CEST cogeneration (either bagasse or both bagasse and barbojo) are competitive only with gasoline prices at the 2000 or 2010 levels and new coal fired electricity, and then only when low capital and cane costs are employed.

Competitiveness is enhanced when both bagasse and barbojo sales are credited to the distillery. For example, using BIG/STIG cogeneration and accruing only bagasse revenues, ethanol can be produced competitively at present gasoline prices only if electricity is competing against new coal fired generating facilities.

This does not mean that the competitiveness at present oil prices depends upon the presence of barbojo revenues; in many developing countries the marginal cost of electricity is well over 5¢/kWh. Rather, it highlights the fact that barbojo sales enhance the prospects of competitive co-production.

As mentioned in Section 2.1.2, the low capital costs of Brazilian equipment are likely to be duplicated in other developing countries, while the cane costs are more liable to be less flexible. This underscores the importance of Figures 4.11 and 4.13: competitive scenarios can be constructed with the higher cane costs apt to be encountered, as long as the lower cost, Brazilian equipment is used.

3.0 Annexed Distillery/Sugar Factory Analysis

3.1 Introduction and Approach

The annexed distillery/sugar factory analysis is approached in as general a way as possible, with a baseline case first addressed followed by quantifying the sensitivities of ethanol and sugar production costs to all of the major production parameters. Basic criteria are developed to quickly identify locations and situations in which co-production of electricity, sugar and ethanol is attractive.

The major difference between this analysis and the analysis of the autonomous distillery is the possibility of "flexible production": the ability to produce varying combinations of ethanol and sugar. While the most common feedstock for ethanol production is C molasses, the sugar recovery process can be stopped at nearly any stage, reducing the sugar yield but increasing the molasses or ethanol yields. When an higher ethanol yield is desired, the sugar crystallization is most routinely stopped after the A strike. The desirability of such actions depends on the relative prices of sugar and ethanol and is analyzed in Section 3.3.

3.1.1 Physical Parameters

Cogeneration Facilities

The same three cogeneration systems considered in the autonomous distillery case are used in the sugar factory and annexed distillery analysis. The summary of the performances of these systems are shown in Table 4.2.

The electrical efficiency and power generated in the cogeneration mode are also scaled to the steam demands of the sugar factory and annexed distillery, 235 kilograms per tonne cane.[j] Therefore, the BIG/ISTIG system is already functioning in full cogeneration mode, with an in-season power output of 97.4MW operating at 37.9% of the fuel's HHV. During the milling season, a BIG/STIG cogeneration system would generate 42.1MW at 31.3% of the fuel's HHV, while a CEST system would generate 21.2MW at 20.3%.

Sugar Factory

The high pressure (2.0 MPa) steam use in the sugar factory is assumed to be 235 kg per tonne cane. This corresponds to the case in Chapter 3 of quintuple effect, falling-film evaporators with continuous vacuum pans and condensate juice heating. While the sugar factory itself requires only 214 kg steam per tonne cane, the additional mill turbine exhaust steam is required by the annexed distillery and to account for losses and parasitic loads. The electricity use for the sugar factory is assumed to be 13 kWh per tonne cane milled.

The sugar yield is assumed to be 108 kg per tonne cane milled with a C molasses yield of 30 kg per tonne cane. These values are calculated using typical cane sugar contents, extraction and crystallization efficiencies [18]. Sugar yields ranging from 100 kg per tonne cane up to 120 kg per tonne cane yields are assumed in the sensitivity studies. When sugar crystallization is stopped following the A strike, 73.5 kg of sugar is recovered with 55.3 kg of A molasses.

The sugar factory capacity is assumed to be approximately 4000 tonnes cane milled per day (~432,000 kg raw sugar per day) for factories matched with CEST or BIG/STIG cogeneration systems and approximately 8000 tc/day (841,000 kg raw sugar per day) when BIG/ISTIG cogeneration is used. The actual cane grinding

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[j] See Section 2.1.1.1 for details on scaling procedure.

rates of the sugar factories are scaled slightly such that the bagasse output meet the in-season fuel requirements of the cogeneration system.[k] The sugar factory is assumed to operate at 90% capacity during the milling season.

Annexed Distillery

With 235 kg of high pressure steam going to the mill turbine and 214 kg of low pressure exhaust steam being used in the sugar factory, 20 kg per tonne cane low pressure steam remains for losses and the annexed distillery. This low steam availability can be matched to a distillery using either ethylene glycol extractive distillation (14.5 kg/tc) or molecular sieve ethanol dehydration (14.7 kg/tc).[l] This allows for 5 kg steam per tonne cane for losses. Electricity consumption for the annexed distillery is assumed to be .05 kWh per liter [18].

Ethanol yields are assumed to be .303 liters ethanol per kg of C molasses, or 9.09 liters anhydrous ethanol per tonne cane. Yield increases to .443 liters of ethanol per kg molasses when A molasses is used as the fermentation feed stock, resulting in an 24.5 liters ethanol per tonne cane milled. These values are derived using typical cane composition, extraction, fermentation and distillation efficiencies. In the sensitivity studies, the ethanol yield (using C molasses as the feed stock) ranges from 8 to 10 liter/tc.

Annexed distillery capacities are assumed to be 36,000 liters per day in conjunction with the 4000 tonne cane per day sugar factory and 72,000 liters per day when used with the 8000 tonne cane per day sugar factory. The distillery is assumed to operate during the milling season at 90% capacity.

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[k] Actual capacities are: 4053 tc/day with CEST cogeneration, 3948 with BIG/STIG and 7806 with BIG/ISTIG.

[l] See Chapter 2, Sections 4.1.1.2 and 4.3.

3.1.2 Costs

The capital, operating and maintenance costs of the cogeneration systems considered with the sugar factory and annexed distillery are the same as in the autonomous distillery analysis. All sugar factory capital, operating and maintenance costs are based on estimates by F.C. Schaffer and Associates [5], except for cane costs, which are assumed to be the low Brazilian value of \$8.07 per tonne. Because data from this study are used as the high cost scenario for the autonomous distillery cost case, these estimates provide a conservative base line for nominal sugar production costs. The capital cost for the sugar factory is assumed to be \$44.0 million, with a grinding capacity of approximately 4000 tonnes of cane per day (Table 4.8). When BIG/ISTIG cogeneration is applied, the capital costs are \$85.64 million for a factory with ~8000 tonne cane per day milling capacity. Labor, operation and maintenance costs for the sugar factory are assumed to be \$1.02 million per year for the 4000 tc/day factory and \$1.99 million per year for the 8000 tc/day factory. Variable costs excluding cane costs are assumed to be \$4.17 per tonne cane milled, independent of capacity (Table 4.8).

The annexed distillery capital, operating and maintenance costs are based on estimates by F.C. Schaffer and Associates [5] along with the data shown in Figure 4.15. The capital cost for the annexed distillery is assumed to be \$3.0 million, with anhydrous ethanol production capacity of 36,000 liters per day. (Schaffer annexed distillery estimates are similar to those of Brazilian manufacturers.) When BIG/ISTIG cogeneration is assumed, capital costs are scaled linearly up to \$5.84 million for a distillery with 71,000 liters anhydrous ethanol per day capacity.

Labor, operation and maintenance costs for the annexed distillery are based on the F.C. Schaffer study [5] and assumed to be \$135,000 per year for the 36,000 liter per day distillery and \$262,700 per year for the 71,000 liter per day facility.

When flexible sugar factory/distillery scenarios are considered in Section 3.3, the fixed labor, operating and maintenance costs are scaled up to $216,000 per year. Variable costs (excluding molasses) are $0.38 per liter.

TABLE 4.8 Sugar Factory Capital and Operating Costs for 4000 Tonnes Cane Per Day Sugar Factory [5].

CAPITAL COSTS

ITEM	CAPITAL COST, $
Site Preparation	213,000
Cane Handling	661,000
Milling Equip.	5,939,000
Juice Treatment, incl. clarifiers evaporators, & Crystallization Equip.	5,570,000
Sugar Handling	1,141,000
Molasses Handling	165,000
Water & Chemical Treatment	539,000
Warehouse	1,289,000
Factory wide Piping	3,885,000
Plantwide Services	1,676,000
Office, Lab and Shops	578,000
Storage & Shipment	184,000
Subtotal	21,840,000
Contractor Fee, Overhead @ 20%	4,368,000
Labor to Erect @ 50% of above	13,104,000
Shipping and Freight @ 8%	1,747,000
Engineering Fee @ 7.5% of above	3,079,000
TOTAL ERECTED COST	44,138,000

OPERATING COSTS

ITEM	FIXED COST, $	VARIABLE COST, $/tc
Wages & Salaries	234,000	1.15
Repair Materials & Parts	68,000	1.80
Chemicals & lubricants	6,000	0.25
Equipment Rental	8,000	0.02
Vehicles & vehicle maint.	43,000	0.20
Miscellaneous	46,000	0.30
Administrative Costs	616,000	0.45
TOTAL OPERATING COSTS:	$1,021,000	$4.17/tc

A discount rate of 12% and a factory life of 20 years are assumed for both the annexed distillery and sugar factory analyses. Values from 10% to 25% discount rate and factory lifetimes from 15 to 30 years are investigated in the sensitivity studies.

3.1.3 Analysis Methods and Implementation

The economic analysis was performed using a SuperCalc 4 spreadsheet similar to the one used for the autonomous distillery analysis. The spreadsheet allows the user to input all of the pertinent performance and cost information for the cogeneration facility, sugar factory and annexed distillery. It calculates the annual production, overall and average production costs of electricity, sugar and ethanol. A sample spreadsheet is shown in Figure 4.16, with more details in Appendix V.

3.2 "C" Molasses as Feedstock: Results

The baseline analysis of the annexed distillery producing ethanol from C molasses assumes that all of the economic rents associated with the sale of bagasse and barbojo are credited to sugar production. Any fuel credits assigned to ethanol could just as reasonably be assigned to the molasses. Ethanol production costs are independent of both cane costs and fuel credits. The annexed distillery receives only enough fuel credits to cover its steam and electricity bills; therefore, average alcohol costs are a function of molasses price, capital and operating costs only.

3.2.1 Baseline Results

The relationship between sugar and electricity production costs are derived in the same manner as Section 2.1.3. Figure 4.17 is analogous to Figure 4.5 and shows sugar production cost as a function of average electricity production cost for each of the three cogeneration systems. It is assumed that the cogeneration facility operates year round.

For each of the three cogeneration systems, a range of sugar and electricity production costs are shown. The upper left hand end of each line is where no net money is exchanged; the value of the steam and electricity sold to the sugar factory and annexed distillery equals the value of the biomass fuel sold to the cogeneration facility. The upper of the two lines for each cogeneration system is the case where the sugar factory receives revenue from the sale of bagasse; the off-season fuel is purchased from an outside agent. The lower line is where the sugar factory receives revenue from the sale of both bagasse and barbojo.

As in Figure 4.5, increasing fuel prices/revenues are expressed by moving downward and to the right along each of the lines. As fuel prices/revenues increase, the electricity production costs rise while sugar production costs fall. The right hand end of each line is where the average fuel cost to the cogeneration facility is $3.00/GJ, including processing costs.

A number of reference lines are included on Figure 4.17. Again, electricity production costs for existing oil fired power (3.9¢/kwh), new hydro-power (4.2¢/kWh) and new coal-fired power (5.3¢/kWh) are shown as vertical dashed lines. Line S1 shows sugar at 10¢/lb (22¢/kg), and line S1 shows sugar at 5¢/lb (11¢/kg). Over the past 20 years, the world sugar price has been below S2 approximately one-third of the time and above S1 one-third of the time. The sugar market tends to cycle without any long term average increase or decrease.

Analogous to the autonomous distillery analysis, regions of competitiveness are created by the reference lines. For sugar to be produced competitively at the S2 level (competitive over two-thirds of the market cycle), CEST cogeneration would have to be employed, along with both bagasse and barbojo sales by the sugar factory and cogeneration fuel costs nearly $3.00/GJ. At these fuel costs, however, CEST could not produce electricity competitively with any of the references.

Note that the capital and operating costs assumed in this analysis correspond to the high estimates in the autonomous distillery analysis. If low Brazilian costs are used, then sugar production would become much more competitive (Section 4.0).

The only tradeoff to be considered in this baseline analysis involving ethanol production is between selling the molasses directly or using it to produce ethanol. Figure 4.18 shows the production costs of sugar and ethanol (no economic fuel rents in either case) as a function of molasses price. One can see, for example, that for ethanol to be competitive as an octane enhancing additive[m] with wholesale gasoline at $0.64/gallon (1989)[n], molasses would have to sell for less than $17 per tonne. While this is significantly below the world market price for molasses (Figure 4.19), it is not at all unreasonable for distilleries where molasses transportation costs are particularly high, such as in southern Africa [21].

Also evident in Figure 4.18 is the large disparity in the sensitivity of sugar and annexed distillery ethanol production to molasses price. An $80 per tonne change in molasses price would decrease average sugar production costs by 2.4¢/kg. That same change in molasses price would change the average production cost of ethanol by over 26¢/liter.

3.2.2 Sensitivities

The parametric analysis of ethanol from an annexed distillery and sugar are

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[m] The most likely use of annexed distillery anhydrous ethanol would be for ethanol blending. See Chapter 1.

[n] On a performance basis, ethanol as an octane enhancing additive is worth 1.16 times gasoline. [3].

performed in the same manner as in the autonomous distillery analysis. A range of high and low costs or operating parameters are identified, and the change of production cost of either sugar or ethanol with that difference from the assumed high to low costs is derived. Also, the effect of moving along 10% of the identified range of each parameter is noted.

Annexed Distillery Sensitivities

Similar to the autonomous distillery, the average ethanol production cost in the annexed distillery is the most sensitive to the price of the feedstock. In the annexed distillery case, a $5 per tonne increase in C molasses price increases the average ethanol production cost by 1.67 cents per liter (assuming 160 days/year season, 9 liters/tc). This strong dependence on molasses price makes all the more important the large variation seen in molasses prices, both historically and locally.

As in the autonomous distillery cases, the capital costs and the cost of capital (assumed discount rate) also are particularly sensitive parameters. Each $0.25 million increase in annexed distillery costs increases average ethanol production costs by over 0.5 cents. A 1% increase in discount rate results in a .79 cent per liter increase in average ethanol production costs when high capital costs are assumed and .32 cents per liter when low capital costs are assumed. This difference in discount rate sensitivity between the high and low capital costs points out the double importance of achieving low capital costs.

The average annual capacity factor (as expressed in the length of the milling season) also has a moderate effect on the average production cost of ethanol from an annexed distillery. Every ten days of additional operation (at the same daily production rate) decreases the average production cost by .8 cents per liter. Because fuel revenues are not credited against ethanol production, this difference

Table 4.9: Sensitivity of Ethanol Production Costs of Ethanol Produced in an Autonomous Distillery.

Parameter	Low Value	High Value	ΔEtOH cost w/ change from low to high value of param (¢/liter)	$\frac{\partial(\text{EtOH cost})}{\partial(\text{param.})}$[a]	Units
Molasses cost	$10/tonne	$70/tonne	20.60	1.67	(¢/liter)/($/tonne)
Capital cost	$2 million	$5 million	7.75	0.52	(¢/liter)/$.25 million
Disc. Rate	10%	25%			
high Capital costs			12.95	0.79	(¢/liter)/(1% IRR)
Low Capital costs			5.17	0.32	(¢/liter)/(1% IRR)
Season Length	120 days	220 days	-15.08	-.80	(¢/liter)/(10 days)
Yield	65 liter/tc	80 liter/tc	-5.67	-.28	(¢/liter)/(liter/tc)
Variable costs	$0.15/tc		--	0.11	(¢/liter)/($0.01/liter)
Fixed costs	$100,000		--	0.15	(¢/liter)/($10,000)
Distillery life	15 years	30 years			
high Capital costs			-2.17	-.18	(¢/liter)/(year life)
low Capital costs			-0.87	-.07	(¢/liter)/(year life)

Notes:

[a] Partial derivatives are taken at an ethanol yield of 9 liter per tonne cane and 160 day per year milling season.

is independent of the fuel price being charged for the bagasse and possibly barbojo. The remaining parameters (yield, fixed and variable O & M costs, distillery life over which the capital costs are recovered) affect ethanol production costs much less.

Sugar Factory Sensitivities

Like the cost of ethanol produced in an autonomous distillery, sugar production costs are particularly sensitive to sugar cane cost. When sugar cane prices vary from the assumed low price ($8/tonne) to the assume high price

Table 4.10: Sensitivity of Sugar Production Costs.

Parameter	Low Value	High Value	Δsugar cost w/ change from low to high value of param (¢/kg)	$\frac{\partial(\text{EtOH cost})}{\partial(\text{param.})}$	Units
Cane Costs	$8/tonne	$20/tonne	10.91	0.91	(¢/kg)/($/tonne)
Capital Cost	$40 million	$50 million	2.11	0.17	(¢/kg)/$.25 million
Fuel Rents	$0.00/GJ	$2.45/GJ			
Bagasse Only			-6.76	-.69	(¢/kg)/($.25/GJ)
Bagasse and Barbojo			-3.03	-.31	(¢/kg)/($.25/GJ)
Disc. Rate	10%	25%			
High Capital Costs			12.95	0.79	(¢/kg)/(1% IRR)
Low Capital Costs			5.17	0.32	(¢/kg)/(1% IRR)
Season Length	120 days	220 days	-15.08	-.80	(¢/kg)/(10 days)
Yield	100 liter/tc	120 liter/tc	-4.42	-.23	(¢/kg)/(liter/tc)
Fixed Costs	$1,000,000		--	0.13	(¢/kg)/($10,000)
Factory Life	15 years	30 years			
High Capital Costs			-2.17	-.18	(¢/kg)/(year life)
Low Capital Costs			-0.87	-.07	(¢/kg)/(year life)

Notes:

[a] Partial derivatives are taken at an sugar yield of 110 kg per tonne cane and 160 day per year milling season. For second order sensitivities to these parameters, see Appendix VI.

($20/tonne), sugar prices change almost 11¢/kg: each dollar change in sugar cane price increases average sugar production costs 0.9¢/kg. Sugar production cost sensitivities to other parameters are found in Table 4.10.

3.3 Flexible Production of Sugar and Ethanol From Molasses

3.3.1 Introduction

The annexed distillery analysis so far has assumed that the fermentation feed stock has been C molasses, the standard by-product of sugar production. There is nothing inherent that sugar boiling stops exactly at this point; C molasses is not completely devoid of sucrose. Rather, stopping after three strikes has been found to be the point of diminishing returns.

The possibility of stopping the sugar crystallization earlier and using the sugar rich ("sweeter") A molasses for ethanol production rather than C molasses is considered in this section. If ethanol can be produced more cheaply (higher profits) from A molasses or some intermediate level prior to C molasses than C molasses, then flexible production should be seriously considered.

The sweeter A molasses is valued at the value of the C molasses which it contains plus the opportunity cost of the sugar which is not crystallized out of the molasses. The sugar factory is indifferent between selling the higher value sweeter molasses and the standard C molasses. Consequently, the value of the sweeter molasses is depends on the market price of both molasses and sugar.

3.3.2 Ethanol Production Using "A" Molasses

The first case to be considered in the flexible distillery analysis is the use of A molasses as the fermentation feedstock. Specifically, the set of all possible sugar and ethanol price pairs can be divided into two regions: one in which it is more profitable to make ethanol from A molasses (low sugar prices and high ethanol prices) and one in which it more profitable to make ethanol out of C molasses only (high sugar prices and low ethanol prices). The "indifference" line separating these two regions is identified analytically from the ethanol and sugar production cost functions.

Deriving the Ethanol from A-C Molasses Indifference Line

What will be shown is that line of indifference between producing ethanol from either A or C molasses is a linear function of the price of sugar. This will provide an easy reference to ascertain whether fermentation of the sweeter molasses is financially desirable. To find the indifference line, the profit using A molasses for fermentation is set equal to the profit using C molasses:

$$\Pi_{A\ mol \rightarrow EtOH} = \Pi_{C\ mol \rightarrow EtOH}$$

$$P_{EtOH}{*}E_{A\ mol} - \{FC + VC{*}E_{A\ mol} + P_s(S_C - S_A)\} = P_{EtOH}{*}E_{C\ mol} - \{FC + VC{*}E_{C\ mol}\}$$

Where:

$\Pi_{A,C\ mol \rightarrow EtOH}$ = Profit from producing ethanol from either A or C molasses.
P_{EtOH} = Market price of anhydrous ethanol, \$/liter.
$E_{A\ mol}$ = Liters of ethanol produced fermenting A molasses.
$E_{C\ mol}$ = Liters of ethanol produced fermenting C molasses.
FC = Fixed costs of ethanol production, \$.
VC = Variable costs of ethanol production \$/liter.
P_S = Sugar price, \$/kg.
S_C, S_A = Sugar produced when crystallization is stopped following the C or A strikes, kg.

Rearranging the expression to solve for P_{EtOH} as a function of P_S yields:

$$P_{EtOH} = VC + \frac{(S_C - S_A)}{(E_{A\ mol} - E_{C\ mol})} * P_S$$

This expression is general enough so that any feasible production pair of ethanol and sugar can be compared. For the case at hand of A and C molasses and the assumed variable costs, the indifference expression becomes:

$$P_{EtOH} = 3.8¢/l + 2.25 kg/l {*} (P_S[¢/kg])$$

This is plotted in Figure 4.20. For ethanol-sugar price pairs above the indifference line, it is more profitable to ferment A molasses than to ferment C molasses and make thee additional sugar. For price pairs below the line, it is more profitable to ferment C molasses and sell the extra sugar.

3.3.3 Flexible Production: Stopping Sugar Production Between A and C Molasses

The question of producing ethanol from either A or C molasses is actually just a step towards the bigger question of production flexibility: Given a sugar price, what is the optimal sugar content of the molasses to be fermented? To approach this less constrained problem, one must first quantify the relationship between sugar and ethanol.

A quadratic function, S(E), can be used to fit the three standard vacuum pan outputs (A sugar/molasses, B sugar/molasses, C sugar/molasses):

$$S(E) = 0.07329E^2 - 4.715E + 145$$

Where S(E) is the kilograms of sugar produced per tonne cane and E is the liters of ethanol produced per tonne cane. One must not extrapolate beyond either ethanol from A molasses or ethanol from C molasses. More sugar and less molasses than the left end of the line (C sugar and molasses) is not possible using standard sugar boiling techniques. Less sugar and more ethanol than the right end of the line (A sugar/molasses) is technically possible, but because a quadratic fit was chosen for the ethanol-sugar relationship any extrapolations would not accurately reflect the relationship.

This result is inserted into the indifference line equation derived in the previous section. Because the *y*-intercept is constant, increasing ethanol yields with sweeter molasses only affects the slope of the indifference line. If the indifference line is assumed to be between using C molasses for ethanol fermentation and using any other molasses (up to A molasses), the slope of the line is:

$$\text{slope} = \frac{108.15 - 0.7329E^2 + 4.715E - 145}{E - 9.09}$$

For practical purposes, the slope of the indifference line between ethanol from C molasses and ethanol from any sweeter molasses (up to A molasses) will be steeper than for the A-C indifference line. Therefore, for a given sugar price, the lowest indifference line will be for the tradeoff between A and C molasses. For the fermentation of any intermediate molasses (between C and A) to be profitable, the ethanol price would have to be higher than that required for profitable fermentation of A molasses. This implies that the only production decision required (assuming that it is not more profitable to sell the molasses directly) is between using the A or C molasses for ethanol fermentation.

4.0 Sugar Factory/Annexed Distillery with Autonomous Capacity Analysis

4.1 Introduction

This section discusses the option of a sugar factory with an annexed distillery which has the capacity to ferment all of the cane juice. Such a sugar factory/distillery facility would have the ability to adapt to local market conditions. When the sugar market is down, it could produce all ethanol; when the sugar market is up but molasses prices are down, then it could produce sugar, ferment the molasses and sell ethanol. When the both sugar and molasses prices are up, then the facility could perform like a standard sugar factory and produce sugar and molasses.

The first portion of the section develops an algorithm which determines the optimum output of sugar, alcohol and molasses given a set of prices for the products. The remainder of the section investigates the production costs of the flexible facility and overall financial returns when the cogeneration section is added.

4.2 Determining the Optimum Product Mix

The combined alcohol/sugar facility is assumed to produce one of the following combination of products: (1) anhydrous ethanol, fermented directly from cane juice; (2) sugar and C molasses; (3) sugar and anhydrous ethanol, fermented from C molasses; (4) A sugar only and ethanol fermented from A molasses. The possibility of sending a portion of the cane juice to the sugar factory and a portion to the distillery is not considered; the design of a factory which would be robust enough to handle efficiently the full spectrum of sugar-ethanol production is not within the scope of this thesis.

The following assumptions are made in the optimum product-mix algorithm:

1. The sugar factory/distillery operates 160 days per year with a grinding capacity of 4000 tonnes cane per day. It operates at 90% capacity during the milling season.

2. The costs of operating the facility as a sugar factory are $90,000 per year more than operating it as an autonomous distillery. This assumes that the percent difference in operating costs between the high cost autonomous distillery and the sugar factory in the sugar factory/annexed distillery analysis [5], 7.5%, is the same when the low cost Brazilian operating conditions are assumed. When the distillery ferments molasses, the annual operating costs are $320,000.

3. When the distillery capital costs are not assumed to be sunk, the annexed distillery with sufficient capacity to ferment all of the cane juice (~300,000 l/day) costs $8 million. A discount rate of 12% and a distillery life of 20 years are also assumed. Sugar factory costs are considered sunk throughout the product mix analysis, but are

assumed in the production cost analysis of Section 4.3.

4. Yields for the various potential products are:

	Ethanol from <u>C Molasses</u>	Ethanol From <u>A molasses</u>	Ethanol from <u>Cane Juice</u>
Sugar:	108 kg/tc	73.5 kg/tc	0.0 kg/tc
Ethanol:	9.09 l/tc	24.5 l/tc	73 l/tc
	0.303 l/kg mol	0.443 l/kg mol	n/a
Molasses:	30 kg C mol/tc	55 kg A mol/tc	n/a

These assumptions are based on low cost Brazilian conditions and good product yields. Site specific data could just as easily be used, and the algorithm tailored to any particular location or facility.

<u>Method</u>

The optimum product mix algorithm is a set of simple profitability decisions. Each potential decision is presented as an indifference line, on one side of which it is more profitable to produce one product (or set of products) while on the other side it is more profitable to produce another product (or set of products).

The flow chart in Figure 4.21 shows the basic decisions which must be made to determine the most profitable product mix for the combination sugar/ethanol facility. The flow chart begins by assuming that sugar is being produced and first determines the best use of the molasses: ferment the C molasses into ethanol or sell the C molasses directly. Quantitatively, this is asking if higher profits could be achieved by selling the molasses or producing and selling ethanol:

$$P_{MOL}{*}M <^{?}> P_{etOH}{*}E - C_{MOL\rightarrow EtOH},$$

where:

P_{MOL} = the price paid to the distillery for C molasses, \$/tonne,

M = the annual production of C molasses, tonnes,
P_{etOH} = the price paid to the distillery for anhydrous ethanol, \$/liter,
E = the annual production of ethanol from C molasses, liters,
$C_{MOL \rightarrow EtOH}$ = the additional cost of producing ethanol for C molasses over that of just selling the molasses directly, \$.

If the distillery has already been built, then $C_{MOL \rightarrow EtOH}$ includes only the additional operating costs. If the distillery is only being considered and the capital costs have yet to be sunk, the $C_{MOL \rightarrow EtOH}$ term would also include the distillery capital recovery.

As in the A versus C molasses fermentation question of the Section 3.3.2, an indifference line is derived from the profitability inequality. The indifference line for the first decision is:

$$P_{EtOH} = 0.0033*P_{MOL} + \begin{cases} 0.061 \leftrightarrow \text{distillery capital already sunk.} \\ 0.266 \leftrightarrow \text{distillery capital not sunk.} \end{cases}$$

These lines are plotted in Figure 4.22.

Assuming for the moment that it is more profitable to produce ethanol from the C molasses, the next question becomes whether it is more profitable to divert sucrose in the form of A molasses for ethanol production. This question was addressed in Section 3.3.2 and 3.3.3 of this chapter. The indifference line derived there (Figure 4.20) is:

$$P_{EtOH} = 2.25*P_{Sugar} + 0.038$$

The next decision is whether one should produce ethanol directly from cane juice or produce sugar plus either molasses or ethanol from molasses. When the molasses is sold directly, the indifference line is:

$$P_{EtOH} = 1.445*P_{Sugar} + 0.0004*P_{MOL} \begin{cases} - 0.002 \leftrightarrow \text{distillery capital already sunk.} \\ + 0.024 \leftrightarrow \text{distillery capital not sunk.} \end{cases}$$

The indifference line between producing all ethanol and sugar plus ethanol from C molasses is:

$$P_{EtOH} = 1.651*P_{Sugar} - 0.01$$

An indifference line between making all ethanol and making A sugar and fermenting A molasses is not needed. The ethanol from cane juice versus sugar indifference line is below the indifference line between making sugar from either A or C molasses. For a given sugar price, the ethanol price at which it would be desirable to produce all ethanol is lower than the price at which it is desirable to ferment the A molasses. Under all conditions for which it is desirable to ferment A molasses, it is more desirable to ferment the cane juice directly and not produce any sugar or molasses.

The indifference lines between producing ethanol from cane juice and sugar plus either molasses or ethanol from molasses are shown in Figures 4.23a (distillery capital sunk) and 4.23b (distillery capital not sunk). At very low ethanol prices, the decision is always between producing ethanol from cane juice or producing sugar and molasses. At ethanol prices below ~6¢/liter (capital costs sunk), the additional operating costs of producing ethanol cannot be recouped, even with worthless molasses. At high ethanol prices, the tradeoff is between producing ethanol from cane juice and producing sugar and fermenting the molasses for ethanol. At intermediate ethanol prices, the appropriate indifference line depends upon the price of molasses.

One can see from the sugar price reference lines in Figure 4.23a that during approximately one-third of the sugar price cycle (sugar less than 5¢/lb) it would be more profitable to produce ethanol at today's gasoline prices. For gasoline at the estimated 2010 prices, it would be more profitable for the combined sugar/ethanol facility to produce ethanol during almost two-thirds of the sugar price cycle and sugar only one-third. When the distillery costs are not sunk, the indifference line shifts upwards 2.5¢/liter, which only marginally lessens the amount of time in which the dual facility would be producing all ethanol rather than sugar.

Following this determination, the most profitable combination of products is known. The last issue which must be decided is whether or not the whole operation is profitable. The revenues from the sale of bagasse and possibly barbojo are added in, or, if the cogeneration facility is owned by the same party as the sugar/alcohol facility, the revenues and costs of the cogeneration facility are accounted for. The total revenues less the total operating costs are compared to the sunk costs:

$$\text{sunk costs} <^{?}> \text{total revenues} - \text{total operating costs}$$

If less money is lost by remaining idle and paying off the sunk costs than by operating, then the facility should consider sitting out a season.

This is, in essence, the same decision implied in the graphs of ethanol (or sugar) production cost versus electricity production cost (Figures 4.6 and 4.17). When a production cost line falls below both the ethanol (or sugar) market price reference line and to the left of an electricity price reference line, then co-production is profitable. If the production cost line is always above and to the right of the market reference lines for ethanol and electricity, then co-production is not profitable, and one should consider remaining idle until markets change such that profitability is achieved.

4.3 Production Costs of a Sugar Factory/Autonomous Capacity Annexed Distillery

Because the capital cost of such a facility is larger than the capital costs of either a single sugar factory or autonomous distillery, the production cost lines derived earlier must be reevaluated to determine if the additional capital costs paid for the flexibility would harm the competitiveness of either ethanol or sugar. The overall capital cost of a combined sugar/alcohol facility grinding 4000 tonnes cane per day is assumed to be $26 million-- $18 million for the sugar factory and $8 million for the 290,000 liter per day annexed distillery. Both of these values correspond to the low cost Brazilian conditions used in the autonomous distillery analysis [5]. When the facility is operated as an autonomous distillery, the Brazilian condition operating costs are assumed. When the facility is operated as a sugar factory, equivalent Brazilian low cost conditions are assumed (Table 4.11).

Figures 4.24 and 4.25 show the electricity production costs for the three cogeneration systems versus either ethanol production costs or sugar production costs for the dual sugar/ethanol facility. Even with the additional capital costs, either sugar or ethanol production is profitable. Anhydrous ethanol can be produced below 2000 gasoline prices even without selling any fuel to the cogeneration

TABLE 4.11: Low capital and operating cost estimates for 4000 tonne cane per day milled sugar factory (not including power house equipment; i.e. generator, boiler, turbines, etc.).[b]

TOTAL CAPITAL COST:	$18,083,000
FIXED OPERATING COSTS:	$765,000/yr
VARIABLE OPERATING COSTS:	
Maintenace & Operating	0.32
Cane	8.09
Total Variable Cost	$8.41/tc

[a] Based on Goldemberg, J., J.R. Moreira, P.U.M. Dos Santos, and G.E. Serra, "Ethanol Fuel: A Use of Biomass Energy in Brazil", *Ambio*, Vol. 14, No. 4-5, 1985, which presented costs for a Brazilian autonomous alcohol distillery. The distillery values were scaled according to:

$$\text{Costs}_{\text{Sugar, low}}=\text{Costs}_{\text{distillery, low}}*(\text{Costs}_{\text{Sugar, high}}/\text{Costs}_{\text{distillery, high}}).$$

Adjusted to 1987 US$ using Gross National Product Deflator (Economic Indicators, March, 1988) and scaled up linearly from 1764.7 tonnes cane per day (120,000 liter per day) autonomous distillery.

facilities. Assuming 1989 gasoline prices (A3), existing oil fired electricity (E1) and BIG/ISTIG cogeneration, ethanol and electricity can be co-produced competitively with either bagasse and barbojo sales or just bagasse sales. Many other competitive co-production combinations can also be derived.

Sugar and electricity can be co-produced just as competitively. Even with only bagasse revenues accrued to the sugar factory, sugar could be produced at or below the world market price during two-thirds of the price cycle (line S2). With full bagasse and barbojo revenues, sugar could be produced profitably even with sugar prices lower than 5¢/kg (2.7¢/lb).

When Brazilian cost conditions are assumed, the dual sugar/ethanol facility would be competitive as a sugar factory or as an autonomous distillery producing anhydrous ethanol; when it is used as a flexible product factility, it is more profitable than either a sugar factory or autonomous distillery°. When the sensitivities derived for the autonomous distillery and sugar factory in Sections 2 and 3 are applied to the dual sugar/ethanol facility, the competitiveness remains robust. Ethanol and BIG/ISTIG cogenerated electricity can be competitive with 1989 gasoline prices and existing oil fired power for capital costs up to ~$45 million or cane costs up to ~$15/tonne. Sugar can be produced competitively during two-

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° The operating costs of a sugar factory without a distillery (low cost assumptions, 160 day season length, 4000 tc/day capacity, no fuel credits) is 11.7¢/kg; with the average cost of sugar over the past 19 years at 18.9¢/kg, the average annual profit would be ~$4,420,000. An anhydrous ethanol autonomous distillery would produce ethanol at 19.5¢/l (same cost, season and capacity assumptions; no fuel credits); with 2000 gasoline prices (28¢/l ethanol) the annual profit would be $3,617,000). With 2000 gasoline prices and distillery costs not sunk, the facility would operate 11/19 of the time as a distillery and 8/19 of the time as a sugar factory. The average sugar price during the years in which it operates as a sugar factory is 29.1¢/kg. Annual production costs of ethanol and sugar for the expanded distillery (no fuel credits) are 24¢/l and 13.4¢/kg, resulting in an average total profit of $5,037,000. While capital costs, cane costs and fuel credits would affect the bottom lines for each of these three scenarios, they would effect them equally and not change their relative profitabilities.

thirds of the sugar price cycle and BIG/ISTIG power competitive with existing oil fired power for capital costs up to $42 million or cane costs up to almost $12/tonne.

4.4 Financial Return on Flexible Sugar Factory/Distillery with Cogeneration

The sugar factory/autonomous capacity annexed distillery analysis so far has centered on the sugar and alcohol products of co-production. This section examines the internal rates of return (IRR)[p] for the dual sugar-ethanol facility when cogeneration is included.

The IRR analysis assumes that the 20 year life of the sugar factory/distillery is divided into six three year cost periods and one two year cost period. During the first three cost periods, ethanol sells at 1989 gasoline price equivalent (19¢/l); during the next two periods at the 2000 gasoline price equivalent (27¢/l), and during the last two periods at the 2010 gasoline price equivalent (34¢/l).

Sugar prices are assumed to follow the same trends as the they have in the past twenty years. The historic range of sugar prices is broken into three segments: the lowest one-third of the cycle (average: 4.2¢/lb, 9.4¢/kg), the middle one-third of the cycle (average: 7.8 ¢/lb, 17.2¢/kg) and the highest one-third of the cycle (average: 14.8¢/lb, 32.6¢/kg). The sugar price in each period is assigned one of these averages, alternating ...high-medium-low-medium-high... to simulate the cyclical nature of the sugar market. The first period is assigned either the high sugar price average or the low sugar price average. The indifference lines derived in the previous section are used to determined the more profitable product in each period. Low Brazilian operating and capital costs are assumed.

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[p] Lifetime internal rate of return is calculated by:

$$\sum_{i=1\text{ year}}^{20\text{ years}} \{(\text{revenues}_i - \text{operating costs}_i)(1/(1+\delta)^i)\} - \text{capital costs} = 0$$

δ, the internal rate of return, is determined via iteration.

The IRR for each scenario is plotted against the electricity price paid to the facility for the cogenerated electricity, assuming that the off season fuel is either purchased from an outside agent at a net cost of $3.00/GJ (Figure 4.26) or is barbojo, available to the facility at the cost of gathering and processing ($1.35 STIGs, $0.97/GJ, CEST, Figure 4.27). For each cogeneration technology, two lines are shown. The upper line assumes that the first period experiences high sugar prices, while the lower line assumes that the first period occurs during low sugar price. These two lines provide a window in which the IRR can be expected to fall.

The largest possible return on a sugar-ethanol-electricity facility occurs using BIG/ISTIG cogeneration and varies from 17.5% (electricity at 4¢/kWh, low sugar prices in the first period and purchasing off-season fuel) to over 35% (electricity at 5.5¢/kWh, high sugar prices in first period, barbojo used as off season fuel and provided internally). The IRR of a facility with BIG/STIG cogeneration falls between ~12.5% and 22.5% (same extreme conditions), while the IRR of a facility with CEST cogeneration ranges from 7.5% to 21% (same extreme conditions).

Figures 4.28 and 4.29 show the same life cycle IRR versus electricity cost for a low cost Brazilian autonomous distillery and a low cost Brazilian sugar factory. When the first period sugar prices are high, the sugar factory alone retrieve better returns than the flexible facility. This is because the initial, minimally discounted revenues are the same for the two scenarios while the flexible facility costs $8 million more. The same is true when high return BIG/ISTIG cogeneration is added; the same revenue is accrued, but with the sugar factory alone it is spread over fewer investment dollars. When the first period sugar market is in a price trough, the additional revenue received from selling ethanol rather than cheap sugar counterbalances the higher capital costs of the flexible facility so that the returns are about the same. The autonomous distillery alone consistently yields the lowest investment return of the three options.

5.0 Summary and Conclusions

The use of efficient steam injected gas turbine cogeneration equipment and the treatment of bagasse and perhaps barbojo as valuable fuel resources allow electricity-ethanol co-production to be economically viable under many circumstances. Using CEST technology cogeneration (the present state of the art in sugar factory cogeneration) in an autonomous distillery with free bagasse and barbojo available at only their processing and retrieval costs, the most optimistic co-production costs achievable are around 20¢/liter for ethanol and 4.75¢/kWh for electricity. While electricity produced for less than 5 cents per kWh is competitive in many developing world settings, the wholesale gasoline price would have to be greater than 90¢/gallon before neat ethanol would become competitive.

The only way in which autonomous distillery produced neat ethanol costs can be lowered is to value the agricultural by-product bagasse (and barbojo) as marketable fuel resources. When bagasse and barbojo are valued at as little as \$1.00/GJ[q], ethanol costs can be reduced to below 14¢/liter (Brazilian capital and production costs), the level required to match 1989 wholesale gasoline prices.

However, even when these small values are attached to the bagasse and barbojo, the additional fuel costs raises the average production cost of CEST cogenerated power to beyond most competitive levels. In order for both electricity and ethanol to be co-produced competitively, a lower cost, more electrically efficient cogeneration system such as the BIG/STIG or BIG/ISTIG is required.

When BIG/STIG cogeneration is employed and only bagasse sales are credited toward alcohol production, ethanol can be produced competitively at 1989 gasoline prices while electricity production costs are ~5.3¢/kWh, the estimated pro-

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[q] The price paid to the autonomous distillery beyond the value of the steam and electricity the cogenerator provides.

duction cost of a new coal fired power plant. When BIG/ISTIG cogeneration is employed and both bagasse and barbojo revenues are credited to the autonomous distillery, the electricity production costs are lowered to 3.5¢/kWh while still producing ethanol competitively with 1989 gasoline prices. This electricity cost is lower than the *operating* costs of many developing world oil fired power plants.

The analysis of a sugar factory with cogeneration yield similar results. A low cost sugar factory operating under Brazilian cane cost conditions with a BIG/ISTIG cogeneration facility can co-produce sugar during the lowest swings in the sugar market and sell electricity at 4.0¢/kWh and still remain profitable.

The key parameter in ethanol production in a conventional annexed distillery (C molasses is fermented) is the opportunity cost of the molasses being fermented. Anhydrous ethanol for blending with gasoline can be produced competitively with 1989 gasoline prices only if molasses sells for $17/tonne.

A particularly interesting scenario is gas turbine cogeneration with a sugar factory having an annexed distillery capable of fermenting all of the cane juice extracted in the mills. Such a facility would be able to produce anhydrous ethanol and forgo sugar when the sugar market is down and switch to all sugar when the market is up. Given the historical range of the world sugar price and the projected gasoline prices for the next 20 years, such a facility could maximize profits by operating as an autonomous distillery approximately half of the years and as a sugar factory the other half of the years.

The results outlined here show a great deal of robustness with respect to increased capital and feed stock costs. Even when the distillery capital costs are increased over 50% from the low Brazilian level or cane costs are increased 30%, ethanol and electricity can be co-produced competitively with existing oil fired power plants and gasoline at 1989 prices (BIG/ISTIG cogeneration, bagasse and barbojo revenues credited to alcohol production).

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Figure 4.1: Material and Energy Transactions Between an Autonomous Distillery and a Cogeneration Facility.

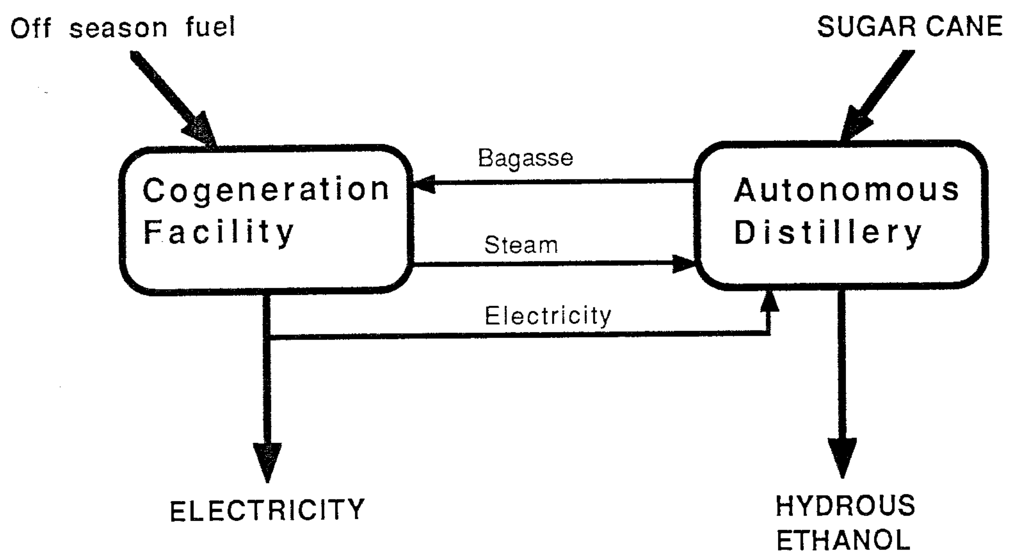


Figure 4.2: Material and Energy Transactions Between a Sugar Factory, an Annexed Distillery and a Cogeneration Facility.

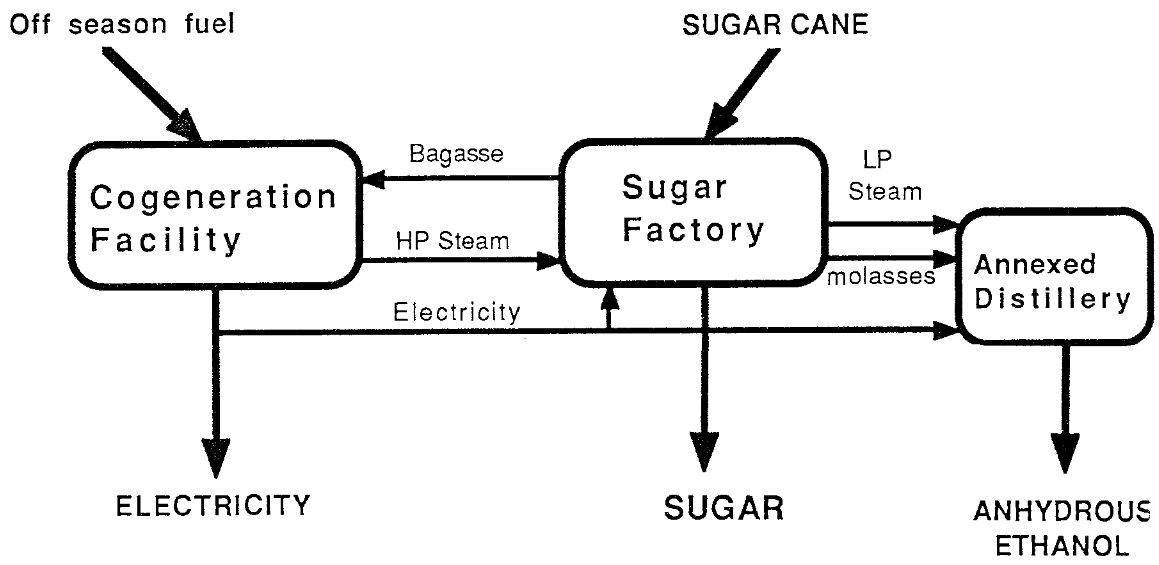
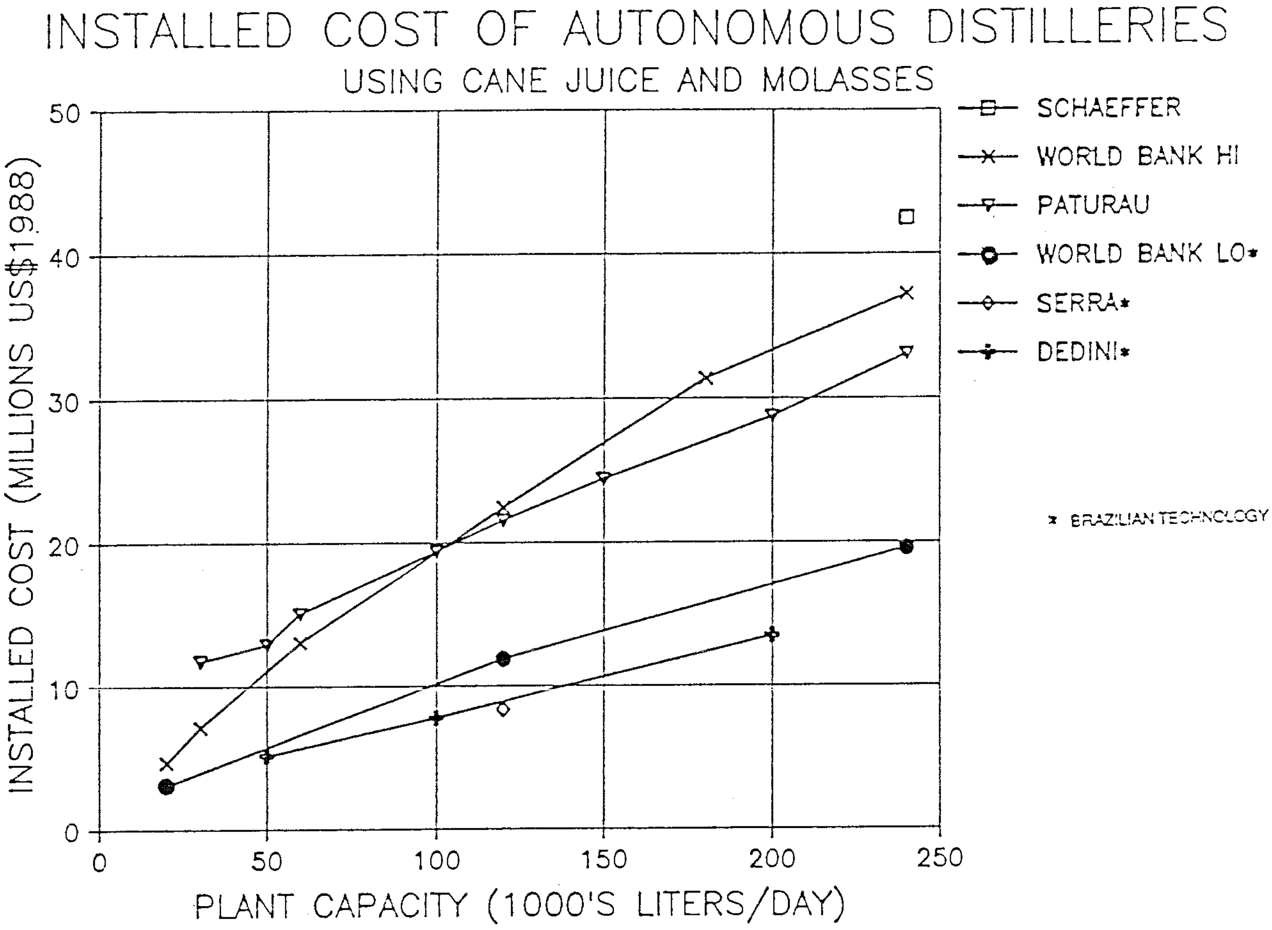
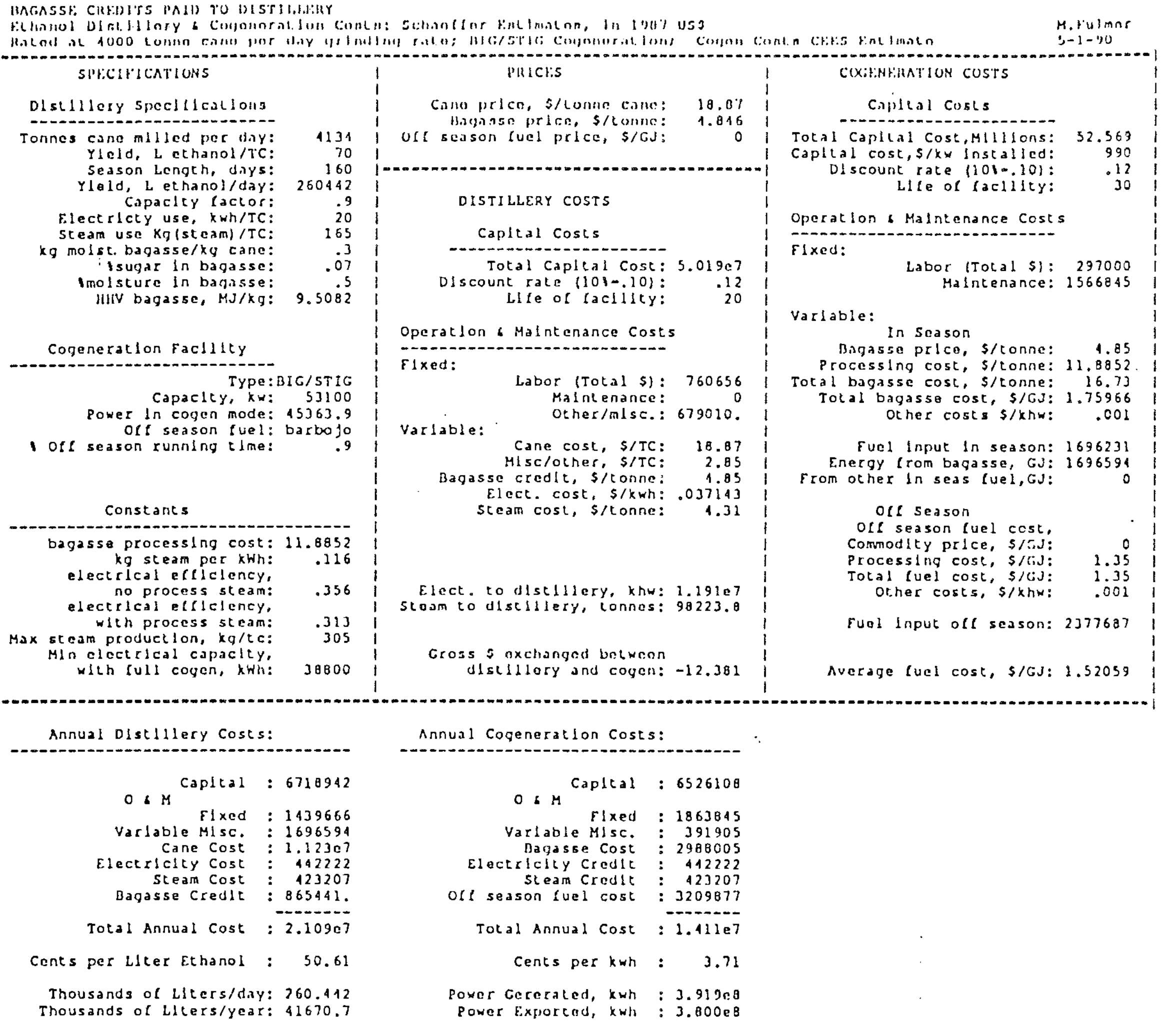


Figure 4.3: Autonomous Distillery Capital Cost Versus Distillery Daily Capacity.



Source: Ogden, Joan and Mark Fulmer, "Assessment of New Technologies for Co-Production of Alcohol, sugar and Electricity from Sugar Cane," PU/CEES Report No. 250, May, 1990.

Figure 4.4: Sample Autonomous Distillery Economic Spreadsheet.



```
BAGASSE CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                        M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration; Cogen Costs CEES Estimate 5-1-90
|--------------------------------------------------------------------------------------------------------|
|        SPECIFICATIONS                 |             PRICES                |      COGENERATION COSTS      |
|                                       |                                   |                              |
|   Distillery Specifications           |    Cane price, $/tonne cane: 18.87|        Capital Costs         |
|   --------------------------          |       Bagasse price, $/tonne: 4.846|   -----------------------  |
| Tonnes cane milled per day:     4134  | Off season fuel price, $/GJ:     0| Total Capital Cost,Millions:  52.569 |
|          Yield, L ethanol/TC:     70  |                                   | Capital cost,$/kw installed:     990 |
|          Season Length, days:    160  |-----------------------------------|    Discount rate (10%=.10):     .12 |
|        Yield, L ethanol/day:  260442  |                                   |           Life of facility:      30 |
|             Capacity factor:      .9  |      DISTILLERY COSTS             |                              |
|       Electricty use, kwh/TC:     20  |                                   | Operation & Maintenance Costs |
|       Steam use Kg(steam)/TC:    165  |         Capital Costs             | ----------------------------- |
|  kg moist. bagasse/kg cane:      .3  |     -----------------------       | Fixed:                       |
|          %sugar in bagasse:     .07  |     Total Capital Cost: 5.019e7   |          Labor (Total $):  297000 |
|        %moisture in bagasse:     .5  |   Discount rate (10%=.10):    .12 |             Maintenance: 1566845 |
|         HHV bagasse, MJ/kg:  9.5082  |         Life of facility:     20  |                              |
|                                       |                                   | Variable:                    |
|                                       | Operation & Maintenance Costs     |          In Season           |
|     Cogeneration Facility             | ----------------------------      |     Bagasse price, $/tonne:     4.85 |
|   --------------------------          | Fixed:                            |   Processing cost, $/tonne: 11.8852 |
|                   Type:BIG/STIG       |         Labor (Total $):  760656  | Total bagasse cost, $/tonne:    16.73 |
|             Capacity, kw:   53100     |              Maintenance:      0  |   Total bagasse cost, $/GJ: 1.75966 |
|       Power in cogen mode: 45363.9    |             Other/misc.: 679010.  |            Other costs $/khw:     .001 |
|            Off season fuel: barbojo   | Variable:                         |                              |
|  % Off season running time:     .9   |          Cane cost, $/TC:  18.87  |        Fuel input in season: 1696231 |
|                                       |         Misc/other, $/TC:   2.85  |    Energy from bagasse, GJ: 1696594 |
|                                       |   Bagasse credit, $/tonne:   4.85 |  From other in seas fuel,GJ:       0 |
|                                       |       Elect. cost, $/kwh: .037143 |                              |
|          Constants                    |      Steam cost, $/tonne:    4.31 |          Off Season          |
|  ------------------------------------ |                                   |      Off season fuel cost,   |
|     bagasse processing cost: 11.8852  |                                   |      Commodity price, $/GJ:       0 |
|           kg steam per kWh:    .116   |                                   |      Processing cost, $/GJ:    1.35 |
|      electrical efficiency,           |                                   |      Total fuel cost, $/GJ:    1.35 |
|           no process steam:    .356   |   Elect. to distillery, khw: 1.191e7 |       Other costs, $/khw:    .001 |
|      electrical efficiency,           | Steam to distillery, tonnes: 98223.8 |                           |
|         with process steam:    .313   |                                   |      Fuel input off season: 2377687 |
| Max steam production, kg/tc:    305   |                                   |                              |
|     Min electrical capacity,          |    Gross $ exchanged between      |                              |
|       with full cogen, kWh:   38800   |        distillery and cogen: -12.381 |  Average fuel cost, $/GJ: 1.52059 |
|                                       |                                   |                              |
|--------------------------------------------------------------------------------------------------------|

     Annual Distillery Costs:                    Annual Cogeneration Costs:
   -------------------------------------       -------------------------------------

                  Capital    : 6718942                          Capital    : 6526108
           O & M                                         O & M
                    Fixed    : 1439666                            Fixed    : 1863845
           Variable Misc.    : 1696594                   Variable Misc.    :  391905
               Cane Cost     : 1.123e7                     Bagasse Cost    : 2988005
         Electricity Cost    :  442222               Electricity Credit    :  442222
               Steam Cost    :  423207                     Steam Credit    :  423207
           Bagasse Credit    : 865441.             Off season fuel cost    : 3209877
                               --------                                      --------
        Total Annual Cost    : 2.109e7                Total Annual Cost    : 1.411e7

   Cents per Liter Ethanol   :   50.61                   Cents per kwh     :    3.71

    Thousands of Liters/day: 260.442              Power Gererated, kwh     : 3.919e8
   Thousands of Liters/year: 41670.7              Power Exported, kwh      : 3.800e8
```

Figure 4.4 shows a spreadsheet to calculate production costs for ethanol and electricity in an autonomous distillery using BIG/STIG cogeneration.

Figure 4.5: Production Cost of Ethanol and Electricity For an Autonomous Distillery with Cogeneration.

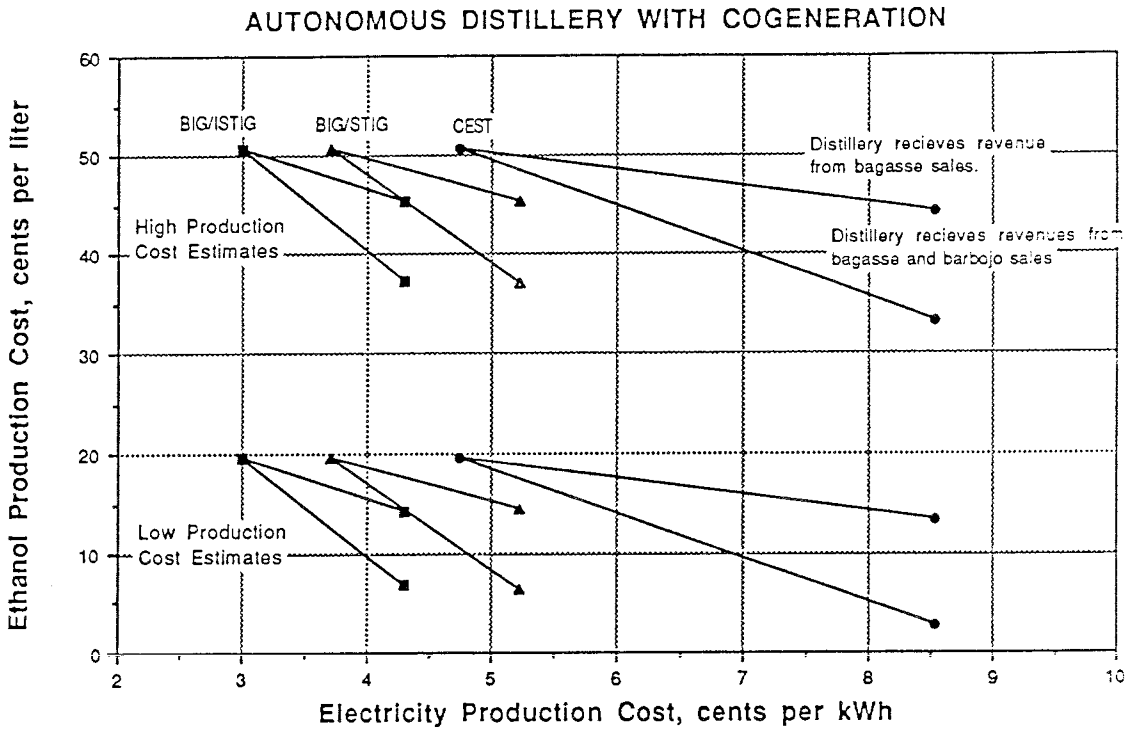


Figure 4.5: Production costs for both ethanol and electricity are shown for the three cogeneration technologies being considered (CEST, BIG/STIG and BIG/ISTIG) and for the high and low ethanol production cost estimates (see text and tables, Section 2.1). It is assumed that the distillery sells bagasse fuel to the cogeneration facility during the milling season (160 days/year); the cogeneration facility also operates during the off season, burning barbojo purchased from either the distillery or cane grower.

For each of the six cases, a range of ethanol and electricity production costs is shown, corresponding to different prices charged by the distillery to the cogeneration facility for the biomass fuels. As the price increases, (moving from right to left along each line), the production cost of ethanol decreases (more revenues from fuel sales) while the production cost of electricity increases (higher fuel costs). The upper of the two lines in each pair is when the distillery receives revenue from the sale of only bagasse, while the lower of the lines is where the distillery receives revenue from the sale of bagasse and barbojo.

The left hand end of each line is where the value of the biomass fuel sold by the distillery exactly equals the value of the steam and electricity sold to the distillery by the cogeneration facility-- not net money is exchanged. The right hand end point indicates the point where the net fuel cost to the cogeneration facility is $3.00/GJ. If the local market price for alternative fuels is greater than $3.00, one can extrapolate beyond the right hand end of the line till the point where (a) the local fuel cost is met, (b) the electricity cannot be produced competitively, or (c) the production cost of ethanol is driven to zero.

Figure 4.6: Production Cost of Ethanol and Electricity For an Autonomous Distillery with Cogeneration with Competitive Reference Lines.

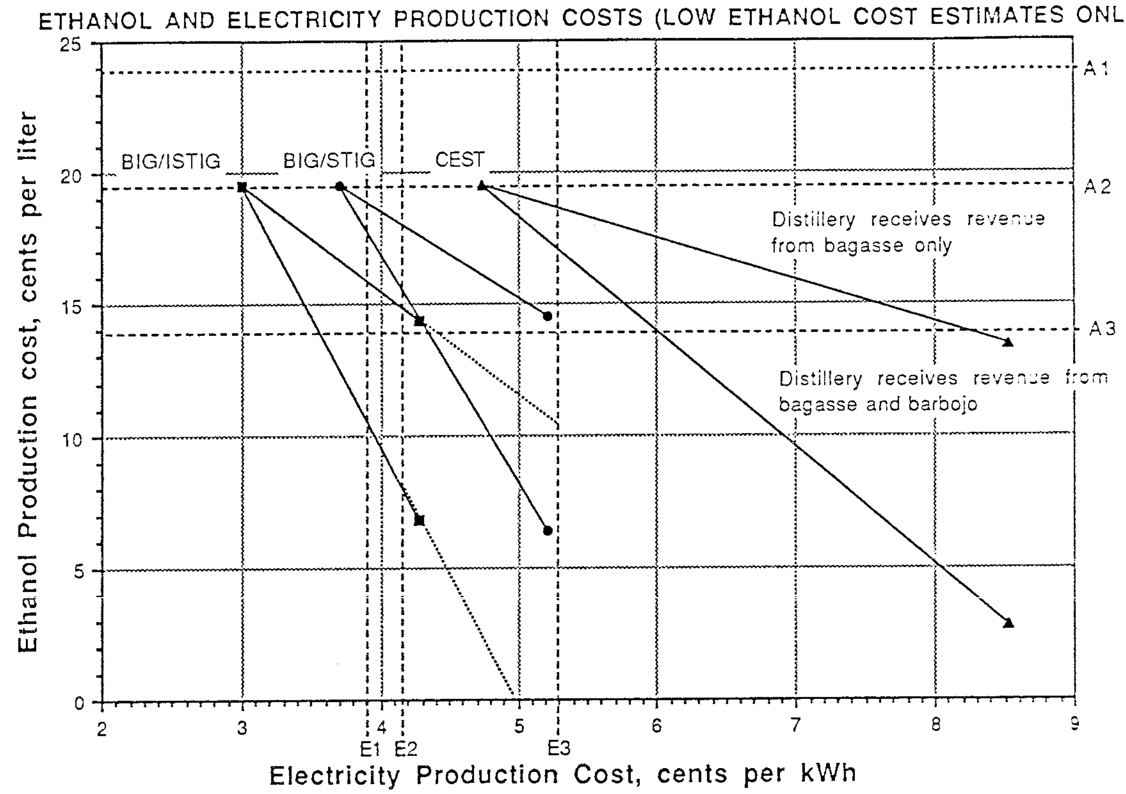
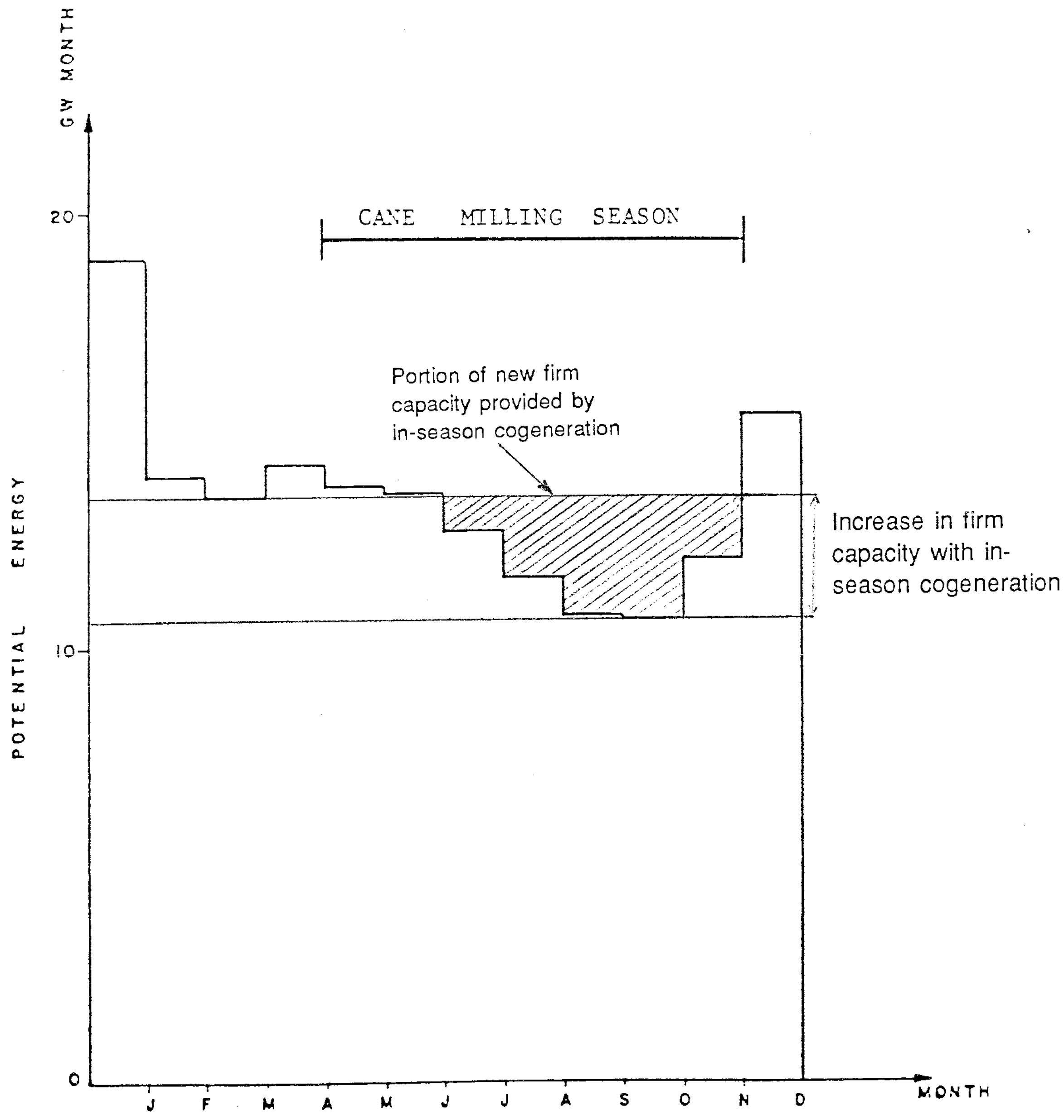


Figure 4.6 shows the production costs of ethanol and electricity with the three cogeneration systems being considered and with the low ethanol production cost estimates (see Figure 4.5 for details on the construction of the figure). Several reverence lines are included on this figure. E1 corresponds to the average fuel and operating costs of an existing oil-fired power plant (see footnote d); line E2 is the average electricity production cost of new hydroelectric power generated in the Amazon basin of Brazil (footnote e). Line E3 is the average electricity production cost of a new coal-fired power plant (footnote f). The right hand end of the solid lines in all three cases show a biomass fuel at $3/GJ (including processing). In the BIG/ISTIG case, the lines are extended with a dotted line until E3 or ethanol costs of $0.00/liter.

Line A1 is the value of neat ethanol at DOE projected 2010 gasoline proces. Line A2 is the value of neat ethanol at DOE projected 2000 gasoline prices, and line A1 is the neat ethanol value at 1989 wholesale gasoline prices [16].

Figure 4.7: The Current Hydro-Electricity "Trough" in Brazil and the Increase in Firm Capacity with In-Season Sugar Cane Based Cogeneration.



By providing power during the dry summer months, the in-season cogenerated electricity not only provide extra power but also allows the existing hydro-electric power to be better utilized. Without the in-season cane based cogneration, the system firm capacity is the lower line (~11 GW-months). When in-season cane based cogeneration is added. the firm capacity increases to the upper line (~14 GW-months).

Figure 4.8 Production Cost of Ethanol and Electricity For an Autonomous with Cogeneration Operating Only During the Milling Season.

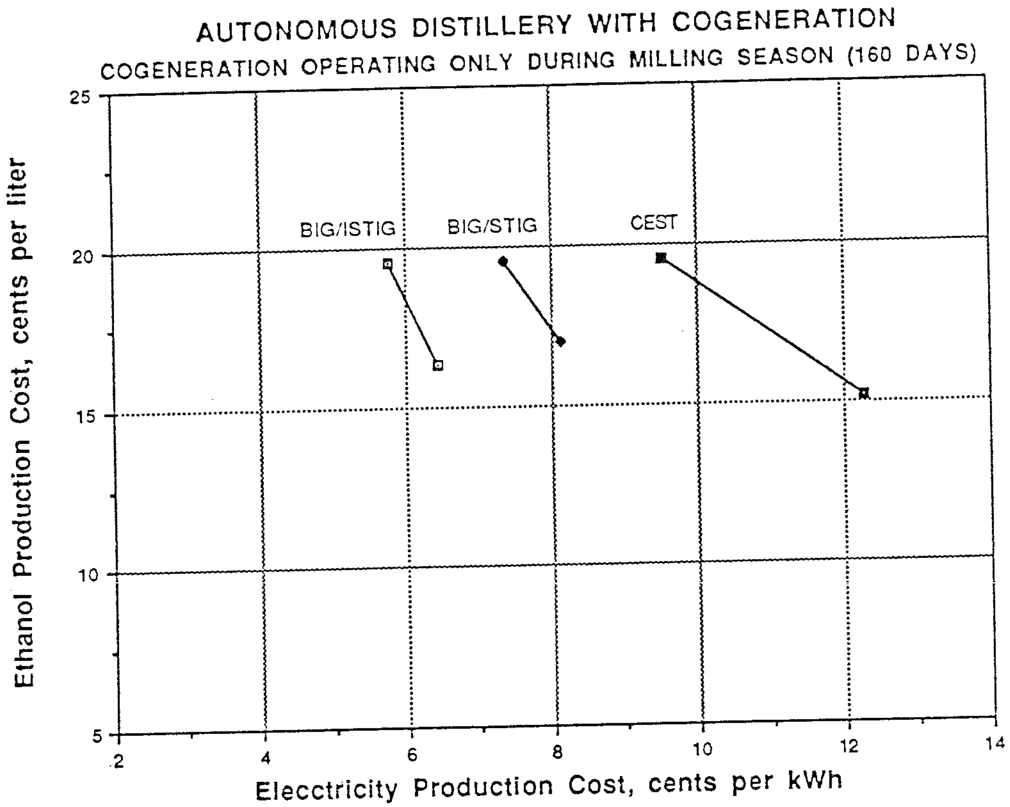


***Figure 4.8** is the same as Figure 4.6 with the exception that the cogeneration facility operates only during the milling season. This might be a desirable scenario in Brazil where the cane harvesting season corresponds to the dry season, when there is a drop in the hydroelectric capacity (see previous figure).*

Figure 4.9 Sensitivity of Ethanol and Electricity Production Costs to the Length of the Milling Season.

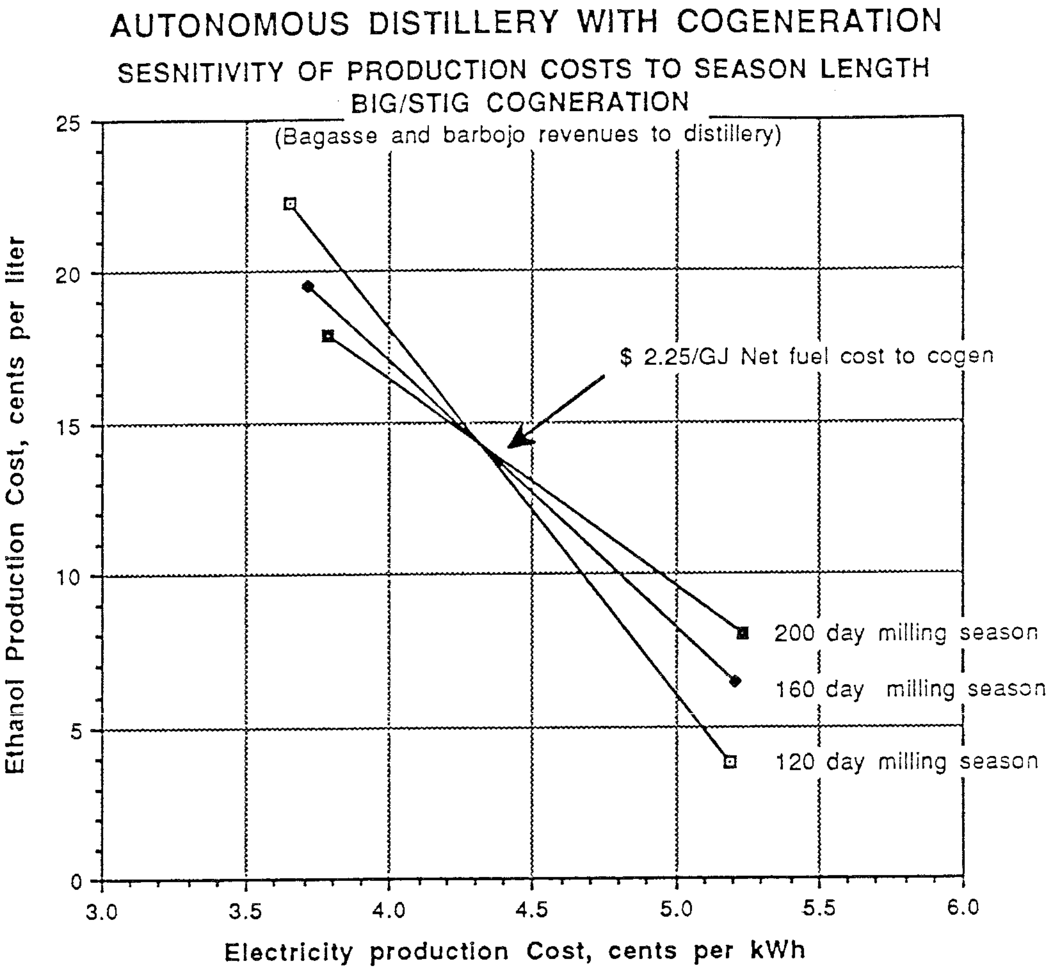
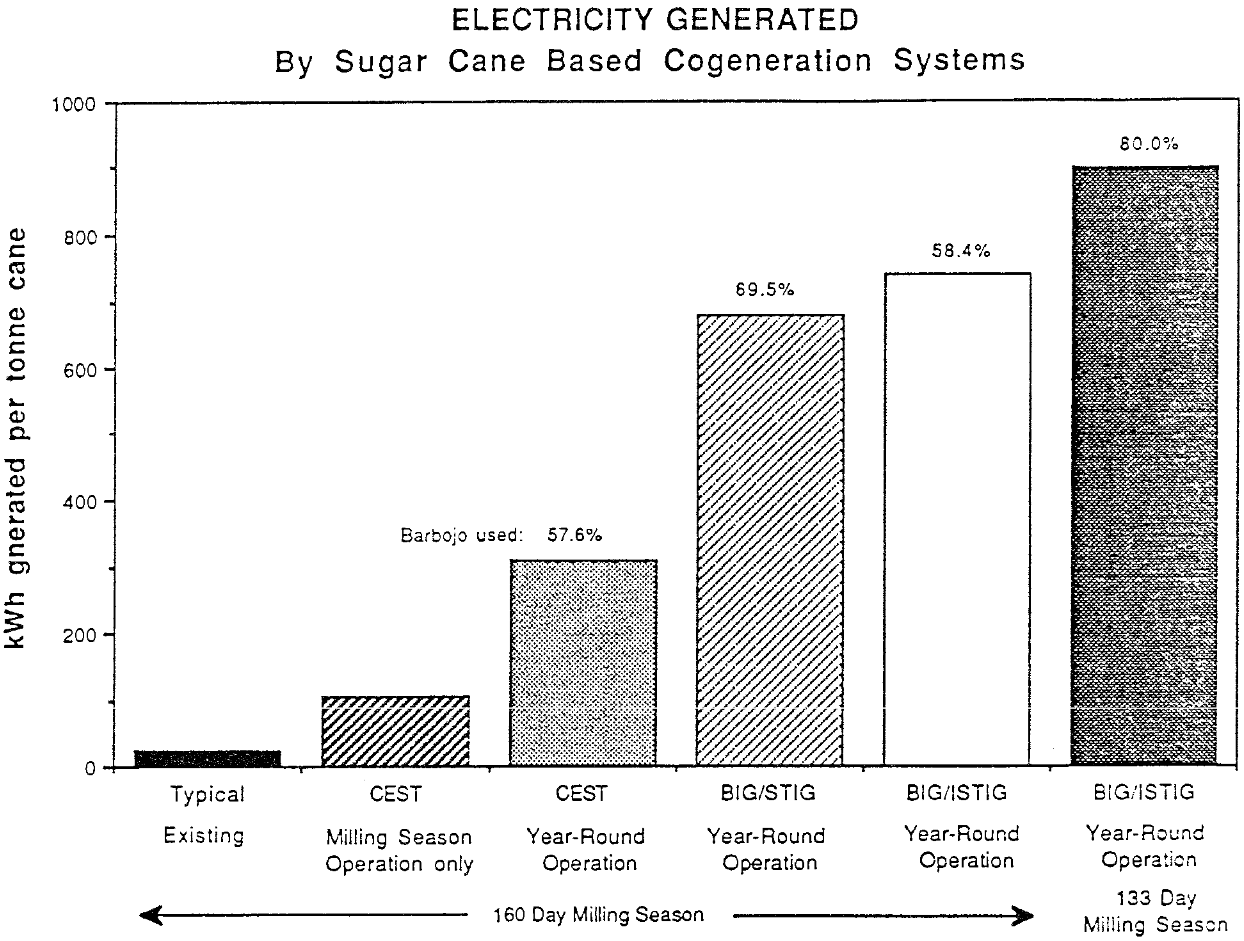


Figure 4.9 shows the variation in production costs with three different milling season lengths. In all three cases, the distillery and BIG/STIG cogeneration capacities are the same; therefore the shorter the season, the less ethanol is produced. In all cases, the cogeneration facility operates year round, purchasing both bagasse and barbojo from the distillery.

When the biomass fuel price is low (left end of the lines), the reduced annual ethanol output increases production costs by spreading the fixed cost over few liters of ethanol. At the same time, electricity production costs increase because the distillery must charge slightly higher prices for fuel to cover the steam and electricity costs.

When the biomass fuel price is high (right hand ends), the additional fuel revenues from off season fuel sales overbalance the fixed capital costs. The additional revenue is also being spread over fewer liters during short milling seasons, therefore has a greater impact on the average production cost of ethanol. Electricity production costs drop slightly at the right hand end because the cogeneration facility can produce electricity more efficiently (more kWhs per GJ of fuel) in the off season "power only mode."

Figure 4.10: Electricity Generated by Sugar Cane Based Cogeneration Per Tonne Cane.



The first bar is the electricity production of typical existing sugar factory or distillery operating only during the milling season. The next bar is for a CEST system operating during the milling season. The next bar is for a CEST system operating year round, with a steam efficient sugar factory or distillery. The next two bars are for the same conditions except either BIG/STIG or BIG/ISTIG cogeneration is employed. The last bar is for a BIG/ISTIG cogeneration system with a shortened (133 day) milling season. The shortened milling season allows for a more complete utilization of the barbojo resources (as seen in by the percentages at the top of each bar) and a longer portion of the year operating in the more electrically efficient "power only" mode.

Figure 4.11: Sensitivity of Ethanol-Electricity Co-production to Cane Costs, Bagasse and Barbojo Revenues Accrued by Distillery.

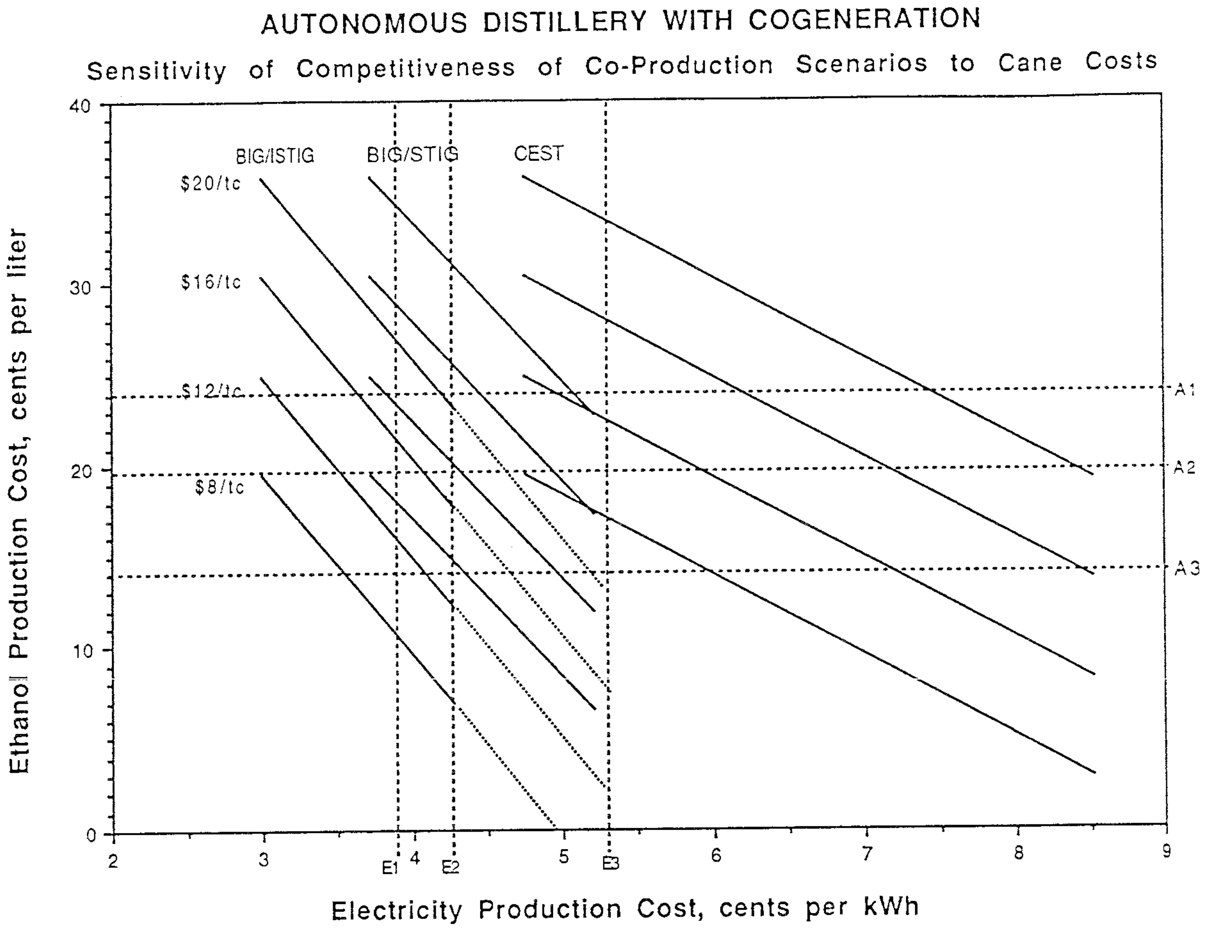


Figure 4.11 is the same as Figure 4.6 with the a range of sugar cane costs shown. The same reference lines are shown: E1 for existing oil fired power, E2 for new Brazilian hydroelectric power and E3 for new coal-fired power; A1 for fuel ethanol equivalent to wholesale gasoline in 2010 (DOE projections); A2 for fuel ethanol equivalent to wholesale gasoline in 2000 (DOE projections) and line A3 for actual 1989 wholesale gasoline prices. The right hand end of the solid lines in all three cases show a biomass fuel at $3/GJ (including processing). In the BIG/ISTIG case, the lines are extended with a dotted line until E3 or ethanol costs of $0.00/liter

One can see that a low cost for cane is critical for economic co-production. At cane prices higher than ~$12/tonne, particularly high fuel credits are required (greater the $3.00/GJ total cost to cogeneration) in order to produce ethanol competitively at present day gasoline prices.

Figure 4.12: Sensitivity of Ethanol-Electricity Co-production to Distillery Capital Costs, Bagasse and Barbojo Revenues Accrued by the Distillery.

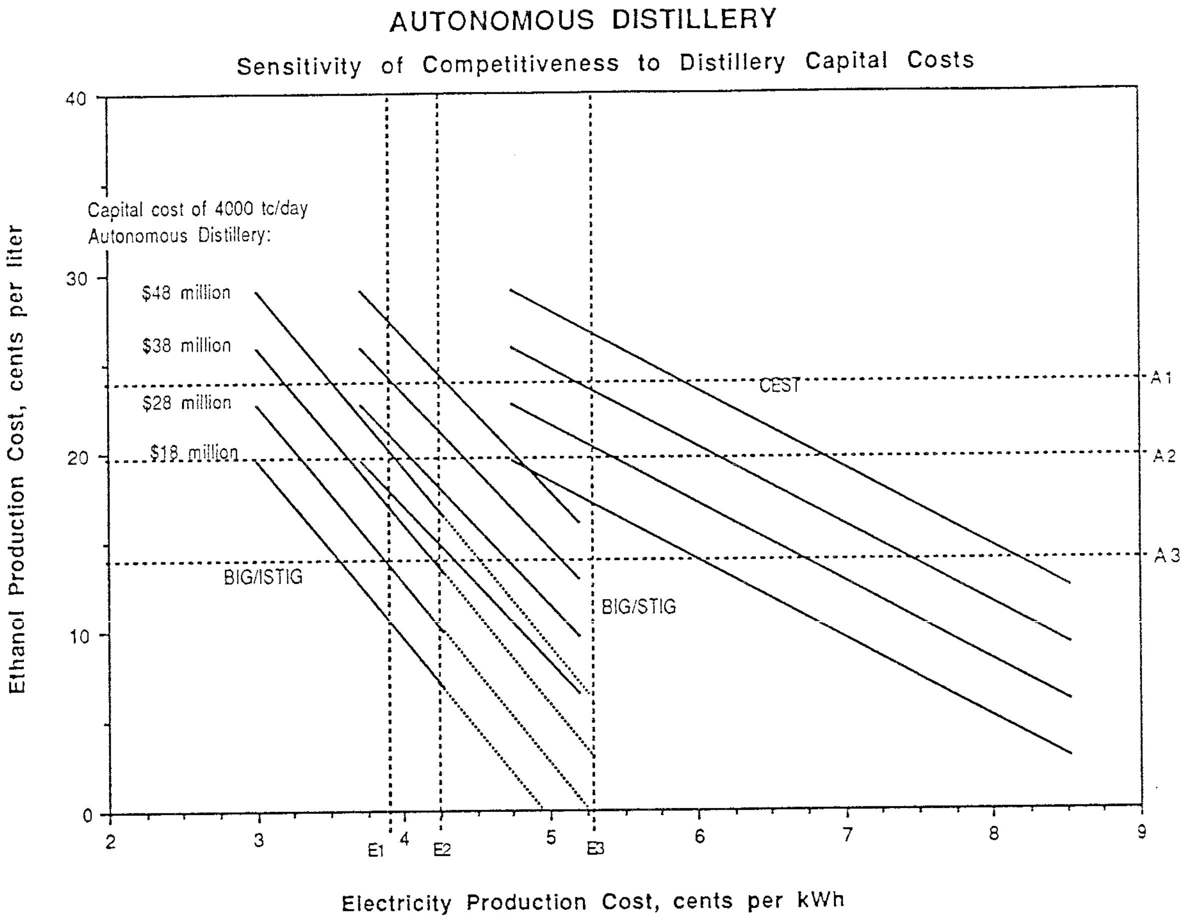


***Figure 4.12** is the same as Figure 4.9 with the a range of distillery capital costs (12% IRR, recovered over 20 years). The same reference lines are shown: E1 for existing oil fired power, E2 for new Brazilian hydroelectric power and E3 for new coal-fired power; A1 for fuel ethanol equivalent to wholesale gasoline in 2010 (DOE projections); A2 for fuel ethanol equivalent to wholesale gasoline in 2000 (DOE projections) and line A3 for actual 1989 wholesale gasoline prices.*

Figure 4.13: Sensitivity of Ethanol-Electricity Co-production to Cane Costs, Bagasse Revenues Accrued by Distillery.

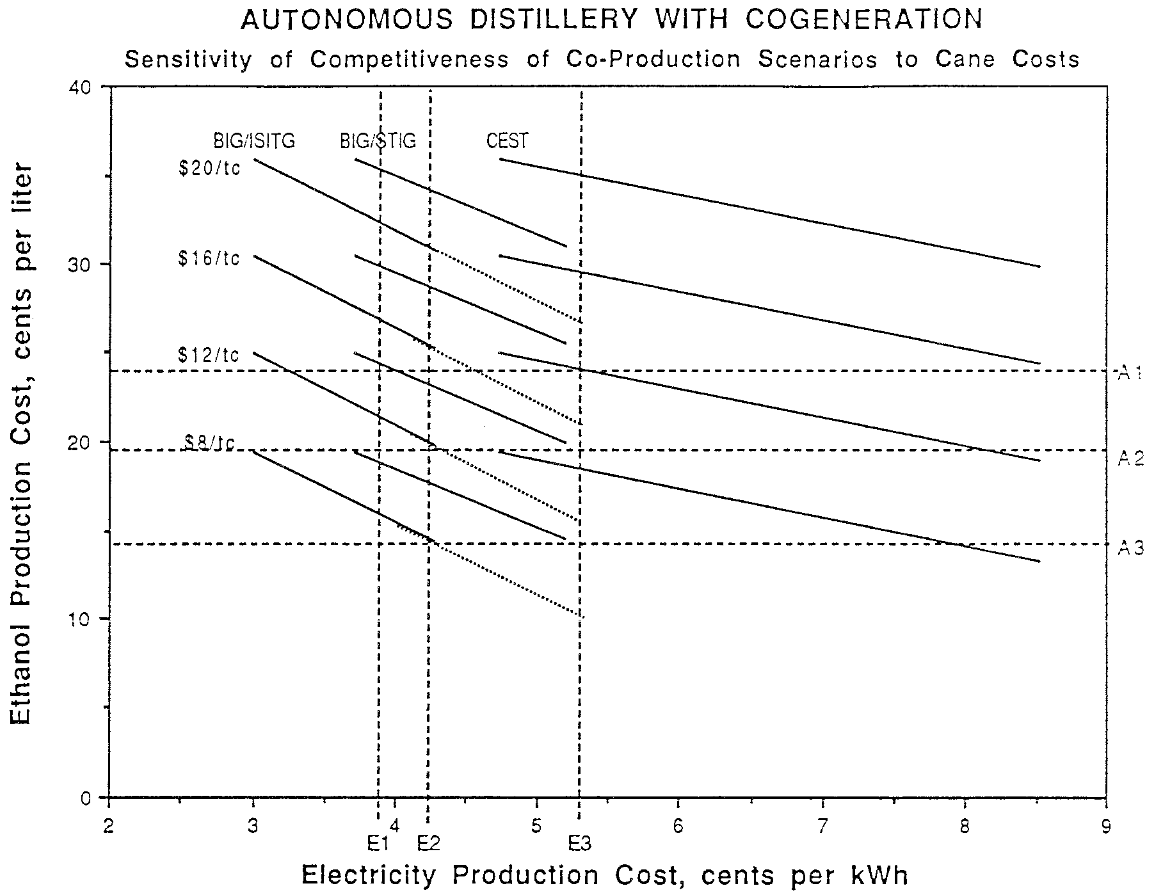


Figure 4.13 is the same as Figure 4.6 with the a range of sugar cane costs shown.

One can see that a low cost for cane is critical for economic co-production. At cane prices higher than ~$12/tonne, particularly high fuel credits are required (greater the $3.00/GJ total cost to cogeneration) in order to produce ethanol competitively at present day gasoline prices.

Figure 4.14: Sensitivity of Ethanol-Electricity Co-production to Distillery Capital Costs, Bagasse Revenues Accrued by the Distillery.

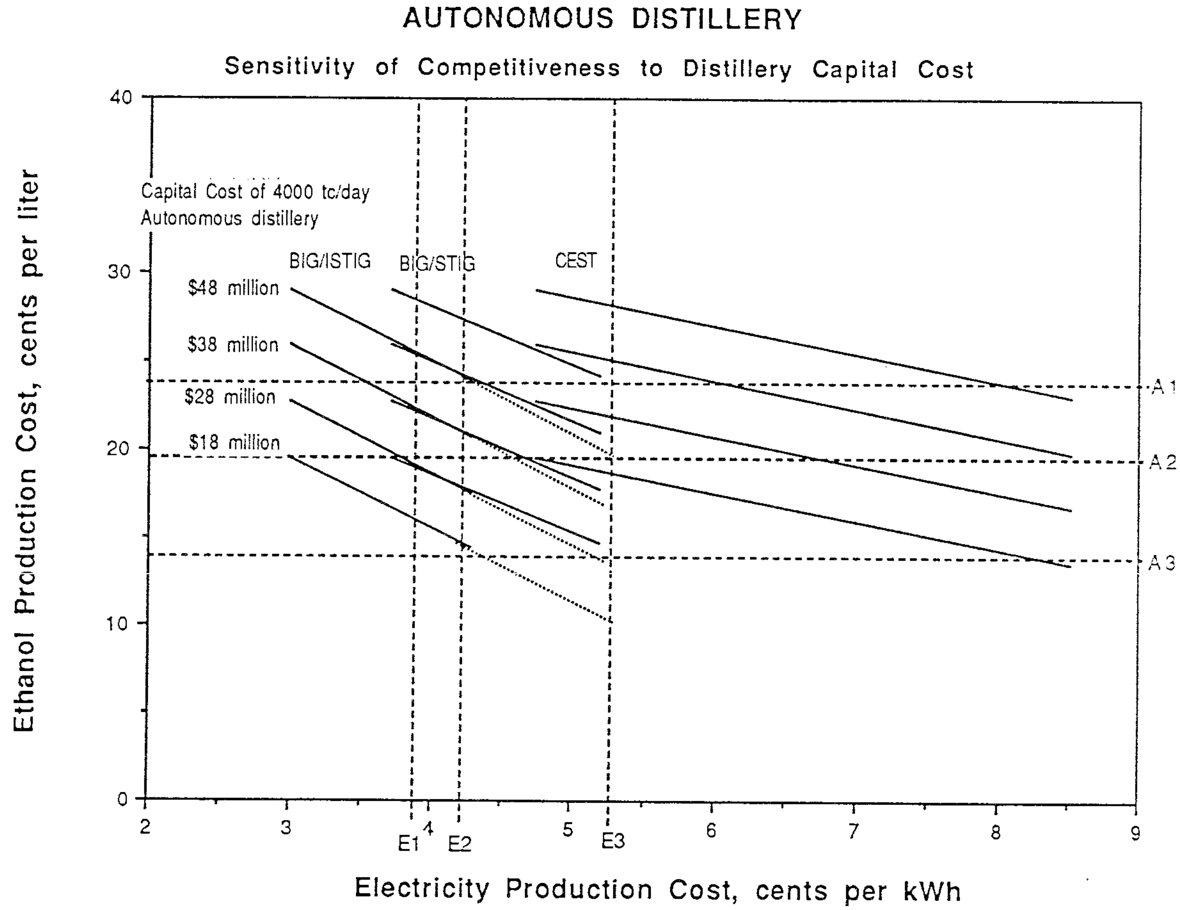
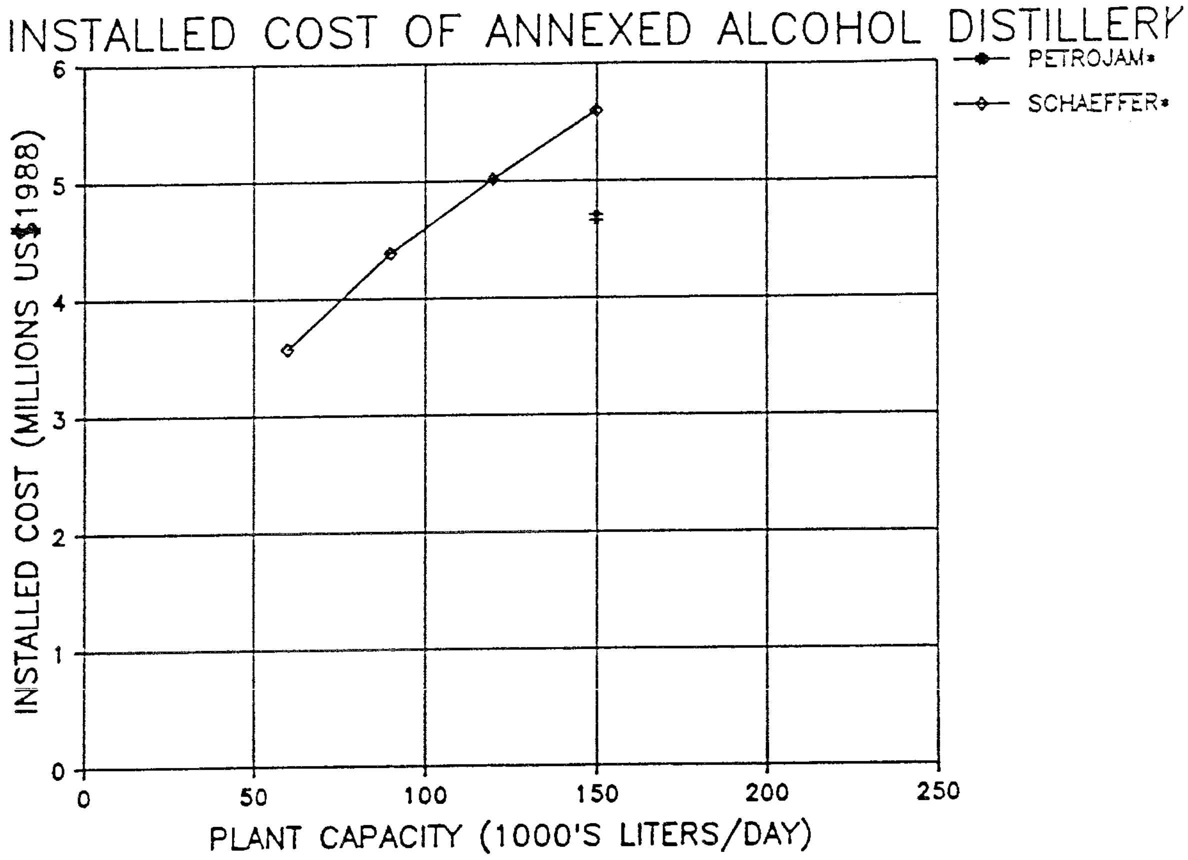


Figure 4.14 is the same as Figure 4.9 with the a range of distillery capital costs (12% IRR, recovered over 20 years). The same reference lines are shown.

Figure 4.15: Annexed Distillery Capital Cost Versus Distillery Daily Capacity.



Source: Ogden, Joan and Mark Fulmer, "Assessment of New Technologies for Co-Production of Alcohol, sugar and Electricity from Sugar Cane," PU/CEES Report No. 250, May, 1990.

Figure 4.16: Sample Sugar Factory/Annexed Distillery Economic Spreadsheet.

```
BAGASSE AND BARDOJO CREDITS TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input                     Fulmer
updated values for molasses and sugar yields from Joan's spreadsheet                                  4-1-90
BIG/STIG COGEN    C MOLASSES TO DISTILLERY
|--------------------------------------------------------------------------------------------------------------|
|        SPECIFICATIONS                   |             PRICES                 |        SUGAR FACTORY          |
|                                         |                                    |                               |
|          Sugar Factory                  |  Cane Price, $ / tonne cane:  8.07 |       Capital Costs           |
|   ---------------------------           |     Bagasse price, $/tonne:   6.04 |       --------------          |
|  Capacity, Tonnes milled/day:    3948   | Off season fuel price, $/GJ:   .00 |   Total Capital Cost: 4.343e7 |
|        Season Length, days:       160   |        Molasses price,$/T:  40.00  | Discount rate (10%=.10):  .12 |
|     Average percent capacity:      .9   | % fuel credit to distillery: .04267|      Life of facility:    20  |
|          Yield, kg sugar/TC:      108   |                                    |                               |
|     Yield, kg  A molasses/TC:      30   |                                    | Operation & Maintenance Costs |
|   kg moist bagasse/kg cane:        .3   |    average fuel cost, $/GJ: 1.56699|  ---------------------------  |
|          %sugar in bagasse:       .07   |------------------------------------| Fixed:                        |
|       %moisture in bagasse:        .5   |                                    |       Labor (Total $):   .00  |
|        HHV bagasse, MJ/kg:     9.5082   |     COGENERATION COSTS             |          Maintenance:    .00  |
|      Electricty use, kwh/TC:       20   |                                    |        Other/misc.: 1007727   |
|    HP Steam use Kg(steam)/TC:     235   |        Capital Costs               | Variable:                     |
|                                         |        --------------              |       Cane cost, $/TC:  8.07  |
|           Distillery                    | Total Capital Cost,Millions: 52.569|      Misc/other, $/TC:  4.17  |
|   ---------------------------           | Capital cost,$/kw installed:   990 |  Bagasse credit, $/tonne: 6.04 |
|       Electricty use, kwh/L :      .05   |    Discount rate (10%=.10):    .12 |    Elect. cost, $/kwh: .037968|
|    HP Steam use Kg(steam)/L :        0   |         Life of facility:      30 |     Steam cost, $/kg: .004404 |
|    LP Steam use Kg(steam)/L :      1.5   |                                    |                               |
|  Liters ethanol/kg molasses:      .303   | Operation & Maintenance Costs      |-------------------------------|
|        Ethanol yield, L/tc:       9.09   | ----------------------------       |         DISTILLERY            |
|     Cogeneration Facility               | Fixed:                             |                               |
|   ---------------------------           |          Labor (Total $):  297000  |       Capital Costs           |
|                      Type:BIG/STIG      |        Maintenance, $: 1566845     |       --------------          |
|     Full power capacity, kW:   53100    |            Other/misc, $:       0  |  Total Capital Cost: 2961000  |
|          Cogen capacity, kW:   42082    | Variable:                          | Discount rate (10%=.10):  .12 |
|            Off season fuel: barbojo     |   Bagasse Processing, $/TB: 11.8852|      Life of facility:    20  |
|    % Off season running time:       .9  | Total bagasse cost, $/tonne:  17.92|                               |
|               kg steam/kWh:      .116   |   Total bagasse cost, $/GJ:    1.89| Operation & Maintenance Costs |
|       Electrical efficiency,             |  Off seas. fuel proc., $/GJ:   1.35|  ---------------------------  |
|           no process steam:      .356   |  Off season fuel cost, $/GJ:   1.35| Fixed:                        |
|       Electrical efficiency,             |        Other costs, $/khw:     .001|       Labor (Total $):  59220 |
|         with process steam:      .313   |                                    |          Maintenance:   59220 |
|   Elect. Cap. in full Cogen:    38800    | Fuel required in season, GJ: 1621609|        Other/misc.:   14805 |
|  Max steam production, kg/TC:     305   | Fuel input from bagasse, GJ: 1621650| Variable:                    |
|                                         | From other in season source:      0| Molasses cost, $/tonne:   40  |
|  Elect. to sugar fact., kwh: 1.137e7    |   Fuel input off season, GJ: 2377687|  Electricy cost, $/kwh: .037968|
|    HP Steam to sugar fact, T: 133600.   |                                    |     Steam cost, $/kg: .004404 |
|   Elect. to distillery, khw: 258389.    |      Net fuel rents to s&a:     55 |     Misc/other, $/L :  .038   |
|    LP Steam to distillery, T: 7751.66   |         percent to alcohol: -.02950|                               |
|--------------------------------------------------------------------------------------------------------------|

   Cost Summaries
-------------------------------------------------------------------------------------------------------------

          Annual Sugar Costs                      Annual Distillery Costs                   Annual Cogeneration Costs
-------------------------------------------------------------------------------------------------------------

             Capital   : 5814088                        Capital  :  396415                     Capital  : 6526108
                 O & M                           O & M                                   O & M
               Fixed   : 1007727                          Fixed  :  133245                       Fixed  : 1863845
       Variable Misc.  : 2370695                  Variable Misc.  :  196375              Variable Misc.  :  380562
           Cane Cost   : 4587892                  Molasses Cost  : 682214.                 Bagasse Cost  : 3057037
    Electricity Cost   :  431700               Electricity Cost  :    9810          Electricity Credits  :  441511
          Steam Cost   :  554268                     Steam Cost  :   34140               Steam Credits  :  588408
      Bagasse Credit   :  986024                 Bagasse Credit  :   43949        Off season fuel cost  : 3209877
    Molassess Credit   :  682214
                        --------                                  --------                                --------
   Total Annual Cost   : 1.310e7              Total Annual Cost  : 1408251           Total Annual Cost  : 1.401e7

   Cents per Kg Sugar  :   21.33       Cents per Liter Ethanol  :   27.25               Cents per kwh  :    3.80

  Tonnes sugar per day : 383.746      Thousands of Liters/day: 32.2986       Power Generated, kwh  : 3.806e8
 Tonnes sugar per year : 61399.3     Thousands of Liters/year: 5167.77       Power Exported, kwh  : 3.689e8
```

Figure 4.16 shows a spreadsheet to calculate production costs for sugar, ethanol and electricity in sugar factory/annexed distillery using BIG/STIG cogeneration.

Figure 4.17: Production Cost of Sugar and Electricity of a Sugar Factory with an Annexed Distillery and Cogeneration.

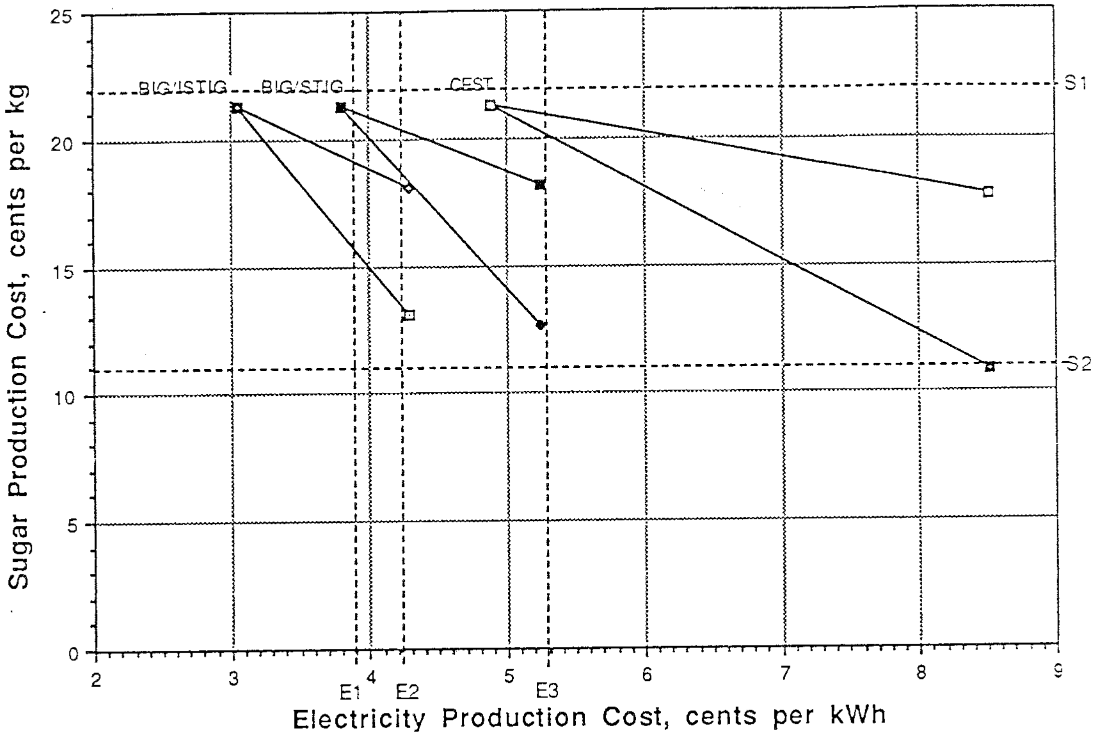


Figure 4.17: *Production costs for both sugar and electricity are shown for the three cogeneration technologies being considered (CEST, BIG/STIG and BIG/ISTIG, see text and tables, section 3.1.) It is assumed that the sugar factory sells bagasse fuel to the cogeneration facility during the milling season (160 days/year); the cogeneration facility also operates during the off season, burning barbojo purchased from either the sugar factory or cane grower.*

For each of the 3 cases, a range of sugar and electricity production costs is shown, corresponding to different prices charged by the sugar factory to the cogeneration facility for the biomass fuels. As the price increases, (moving from right to left along each line), the production cost of sugar decreases (more revenues from fuel sales) while the production cost of electricity increases (higher fuel costs). The upper of the two lines in each pair is when the sugar factory receives revenue from the sale of only bagasse, while the lower of the lines is where the sugar factory receives revenue for the sale of both bagasse and barbojo.

The left hand end of each line is where the value of the biomass fuel sold by the sugar factory exactly equals the value of the steam and electricity sold to the sugar factory and distillery by the cogeneration facility-- no net money is exchanged. The right hand end point indicates where the net fuel cost to the cogeneration facility is \$3.00/GJ. If the local market price for alternative fuels is greater than \$3.00, one can extrapolate beyond the right hand end of the line till the point where (a) the local fuel cost is met, (b) the electricity cannot be produced competitively.

Several reverence lines are included on this figure. E1 correspond to the average fuel and operating costs of an existing oil-fired power plant (see footnote e); line E2 is the average electricity production cost of new hydroelectric power generated in the Amazon basin of Brazil (footnote f). Line E3 is the average electricity production cost of a new coal-fired power plant (footnote g).

Line S1 is sugar at 10¢/lb.; line S2 is sugar at 5¢/lb. Over the past 20 years, market sugar prices have been above 10¢/lb ~1/3 of the time and below 5¢/lb ~1/3 of the time.

Figure 4.18: Sugar and Ethanol Production Costs as a Function of Molasses Value.

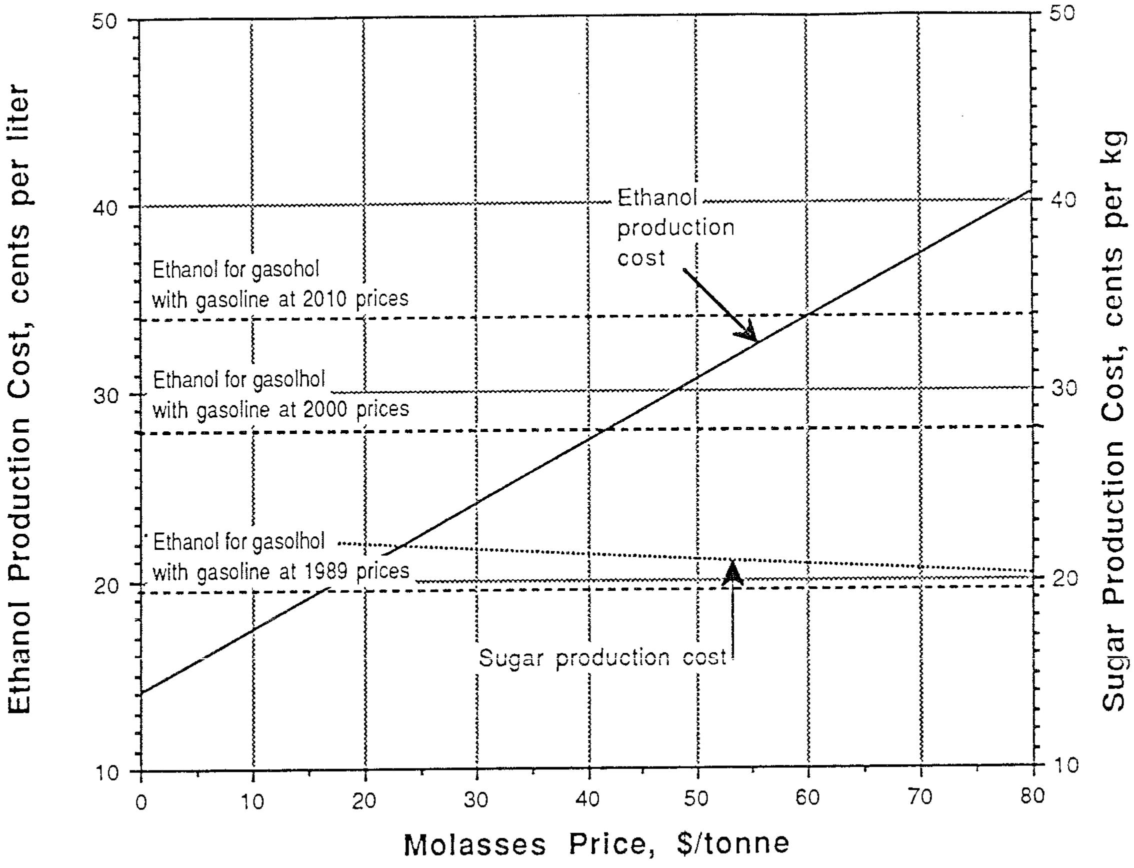
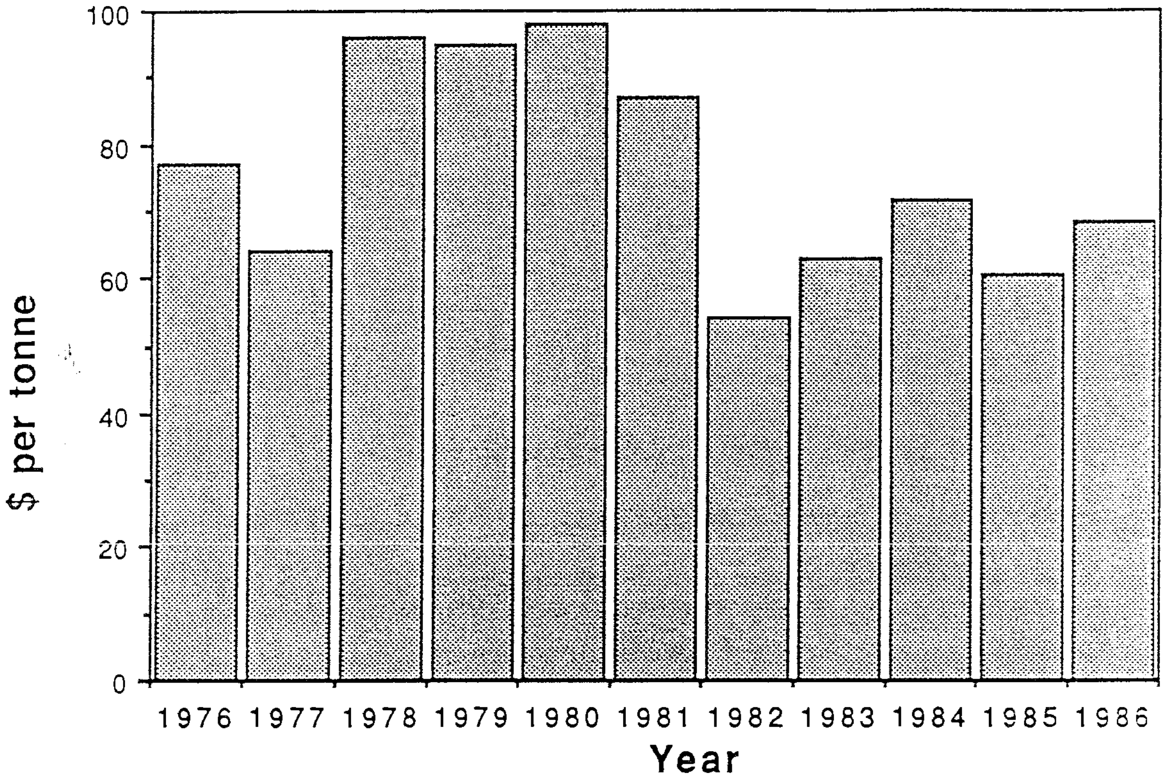


Figure 4.18 shows the effect of varying molasses price on ethanol and sugar production costs. No fuel rents are assumed in either case.

For anhydrous ethanol to be competitive as a gasohol blend-in at present day gasoline prices (S1), molasses would have to be worth at most $33/tonne. For ethanol to be competitive for gasohol with gasoline at $1.00/gallon, then molasses would have to be worth no more than $54/tonne.

Ethanol production costs are much more sensitive to molasses price than sugar production costs.

Figure 4.19: Historical International Molasses Market Prices, 1987 US$.



Source: Fry. James, "Modelling the World Prices of Molasses and Ethanol," Proceedings, Inter-American Sugar Cane Seminars, 1987. p. 280

Figure 4.20: Indifference Line Between Producing Ethanol from A Molasses and Producing Ethanol from C Molasses.

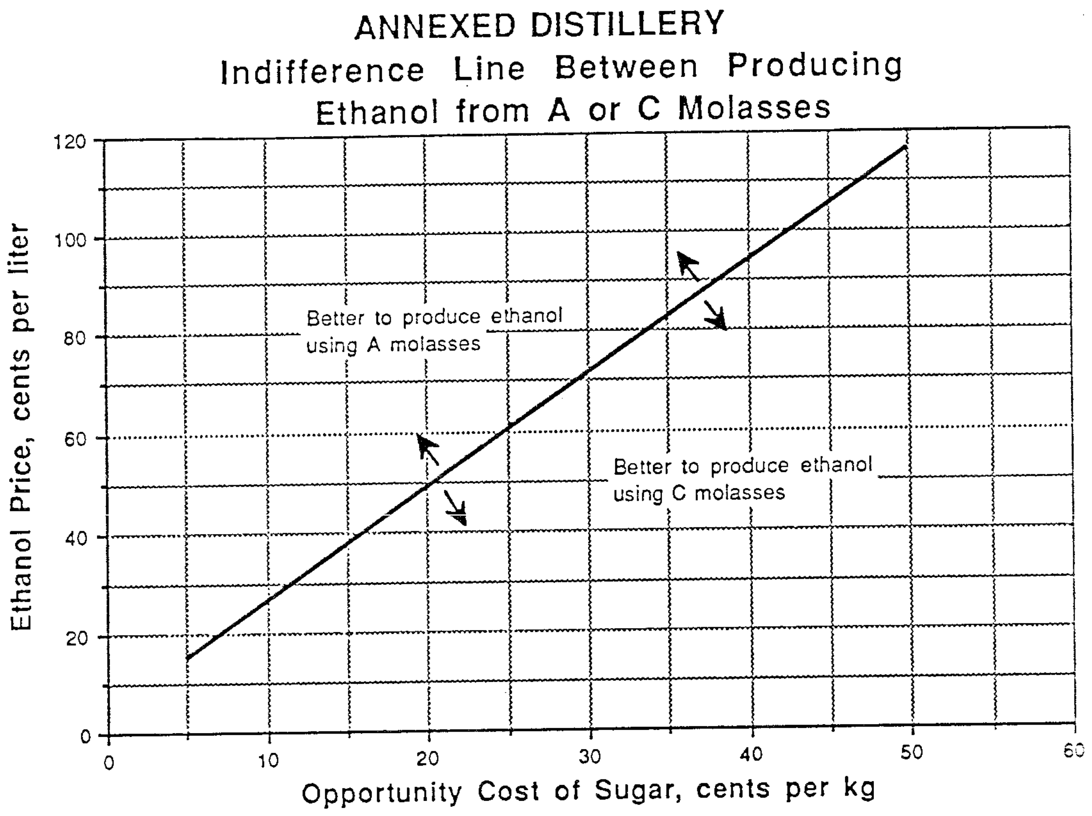


Figure 4.20 *shows the line of indifference between producing ethanol from A or C molasses. Above the line, it is better to use the sweeter A molasses for fermentation and alcohol production. Below it, it is better to use the standard C molasses.*

Figure 4.21: Indifference Line Between Selling C Molasses Directly and Producing Ethanol From C Molasses in a Sugar Factory with an Autonomous Capacity Annexed Distillery.

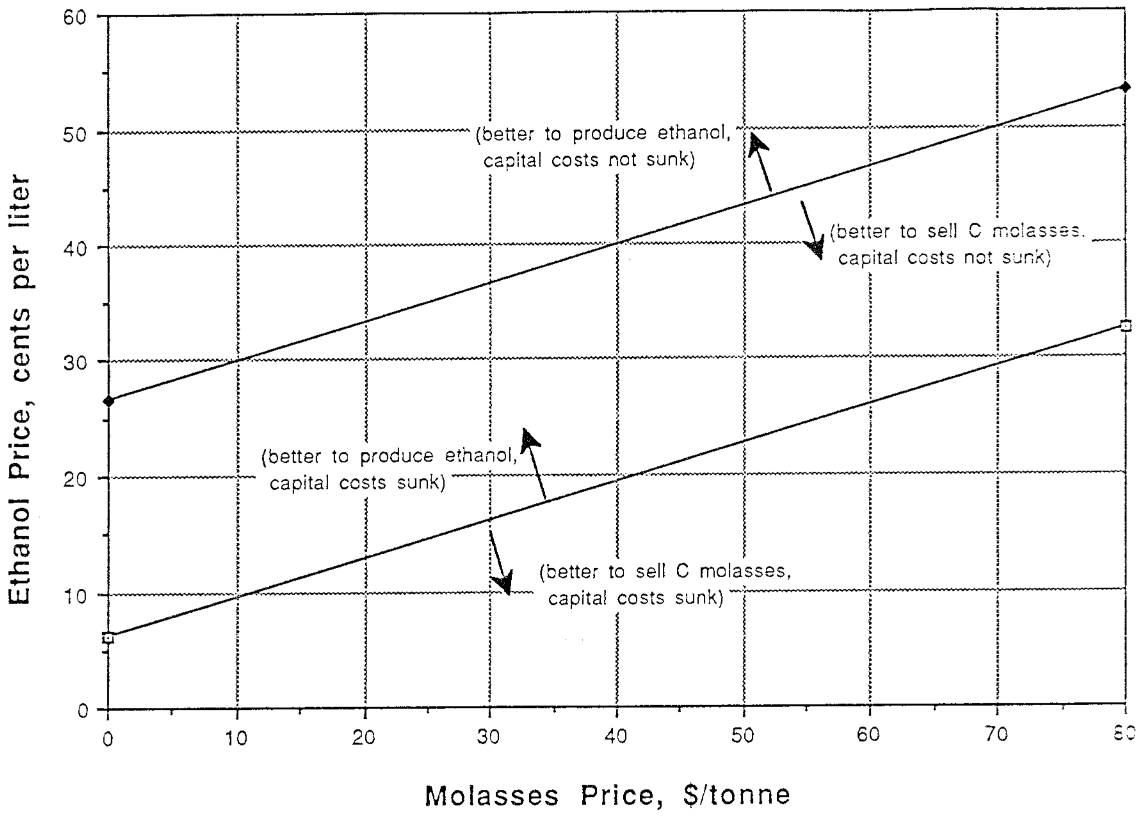


Figure 4.21 *shows the lines of indifference between selling the C molasses directly or using it to produce ethanol and selling the ethanol. The upper indifference line assumes that the capital costs for the distillery are not sunk, while the lower line assumes that the distillery is already built.*

Figure 4.22: Flow Chart of Decisions Required to Find Optimal Products for a Sugar Factory with an Autonomous Capacity Annexed Distillery.

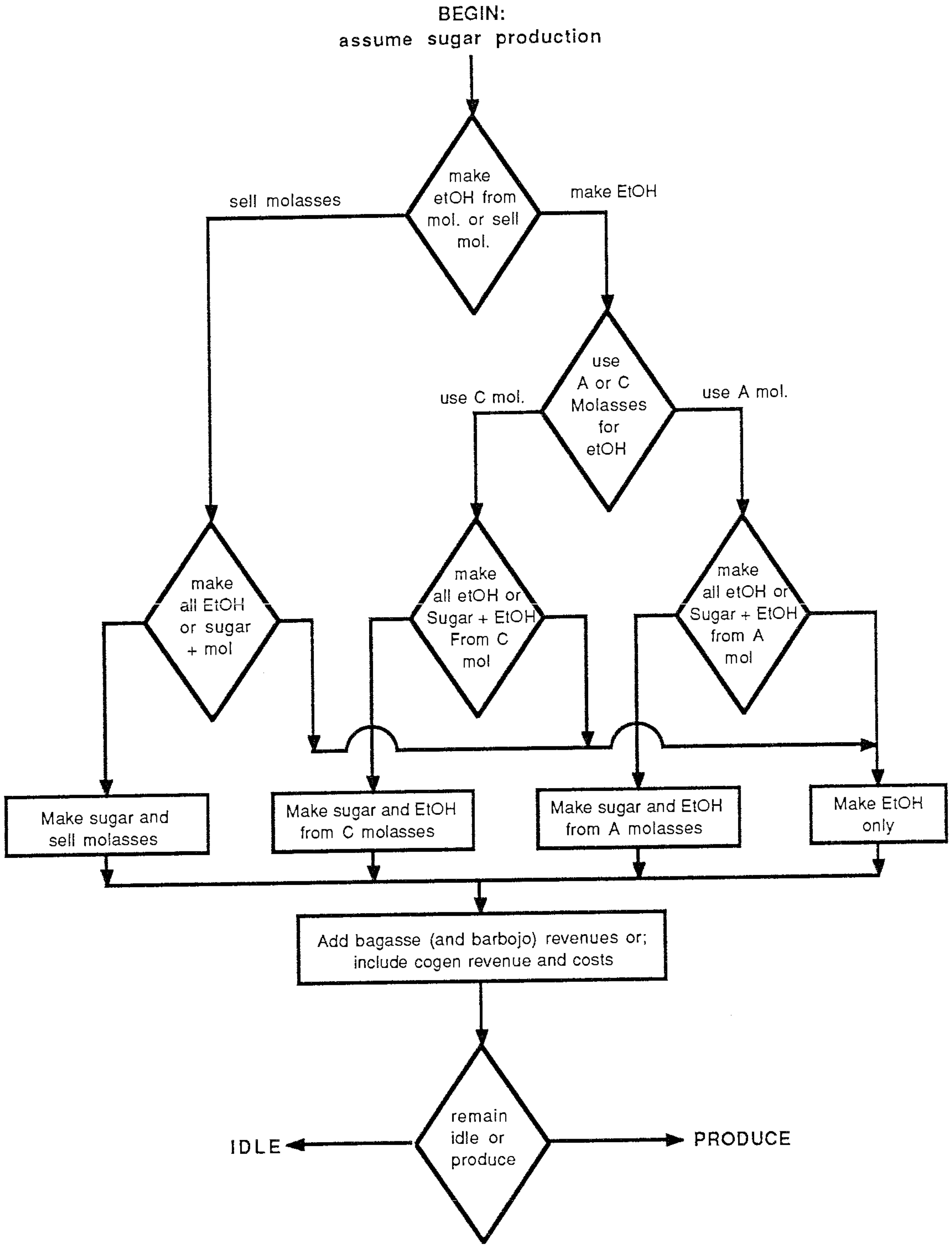


Figure 4.23a: Indifference Lines for Autonomous Ethanol Production and Sugar Production (Distillery Costs Sunk) in a Sugar Factory with an Autonomous Capacity Annexed Distillery.

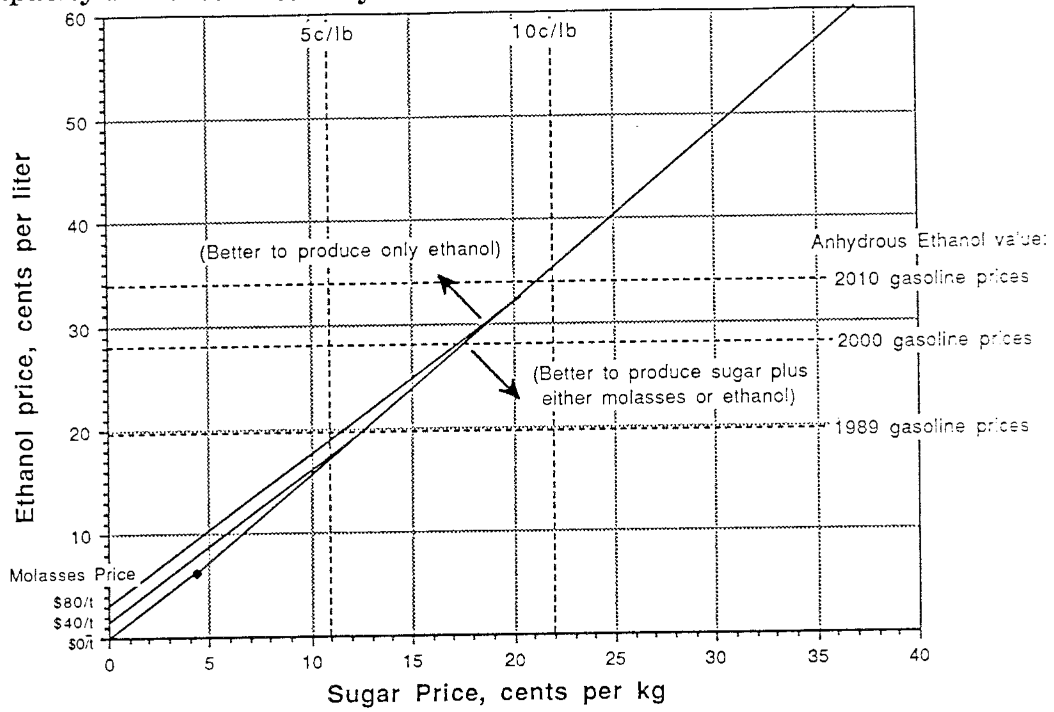
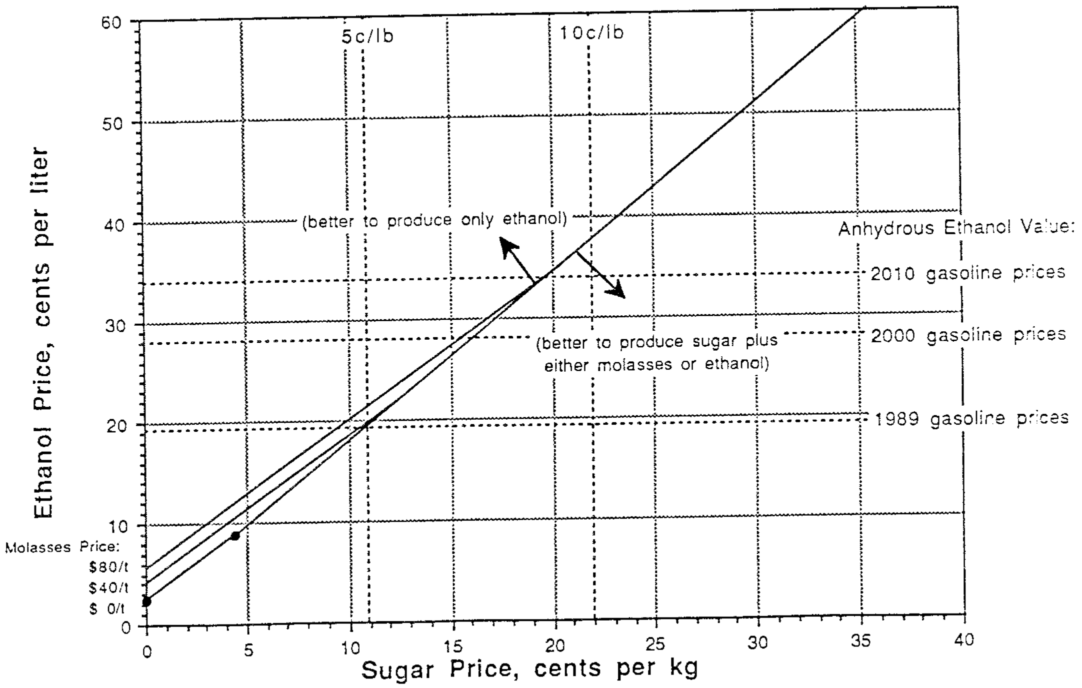


Figure 4.23b: Indifference Lines for Autonomous Ethanol Production and Sugar Production (Distillery Costs not Sunk).



Figures 4.23a and 4.23b *show the indifference lines for the production of either ethanol from cane juice or sugar plus molasses or ethanol from molasses for a dual sugar/ethanol facility. Figure 4.23a assumes that the distillery capital costs have been sunk, while Figure 4.23b assumes that the distillery capital costs are not sunk. For sugar-ethanol price pairs above the indifference line, it is more profitable to produce ethanol directly form cane juice and forgo all sugar production. Below the indifference line it is more profitable to produce sugar.*

At low ethanol prices, the production decision is always between producing ethanol from cane juice or producing sugar and molasses. At ethanol prices below ~6¢/liter (capital costs already sunk), the additional operating costs of producing ethanol cannot be recouped even with worthless molasses. At high ethanol prices, the tradeoff is between producing ethanol from cane juice and producing sugar and fermenting the molasses for ethanol. At intermediate ethanol prices, the appropriate indifference line depends upon the price of molasses and the price of ethanol. Three different molasses prices are shown to illustrate this transition.

Two sugar reference prices are shown on each plot. Over the past 20 years, the world sugar price has been below the lower sugar reference, 11¢/kg (5¢/lb) approximately 1/3 of the time (average price during lower 1/3:4.2¢/lb) and over the higher reference, 22¢/kg (10¢/lb) approximately 1/3 of the time (average price during upper 1/3: 14.9¢/lb). Sugar prices have tended to cycle over a 7 or 8 year period with no net average increase or decrease.

Three reference lines are shown for anhydrous ethanol values, corresponding to present day US wholesale gasoline price (1989) and projected US DOE gasoline prices in 2000 and 2010.

One can see from the sugar price reference lines that approximately 1/3 of the time it would be more profitable to produce ethanol at today's gasoline prices than to produce sugar. For gasoline at the estimated 2010 prices, it would be more profitable for the combined sugar/ethanol facility to produce ethanol almost two-thirds of the time and sugar only one third.

If the distillery costs are not sunk (4.23b), the indifference line shifts upward 2.5¢/liter, which only marginally lessens the amount of time in which the dual facility would be producing all ethanol rather than sugar.

Figure 4.24: Production Cost of Ethanol and Electricity For a Sugar Factor/Autonomous capacity Annexed Distillery (Functioning as an Autonomous Distillery) with Cogeneration with Competitive Reference Lines.

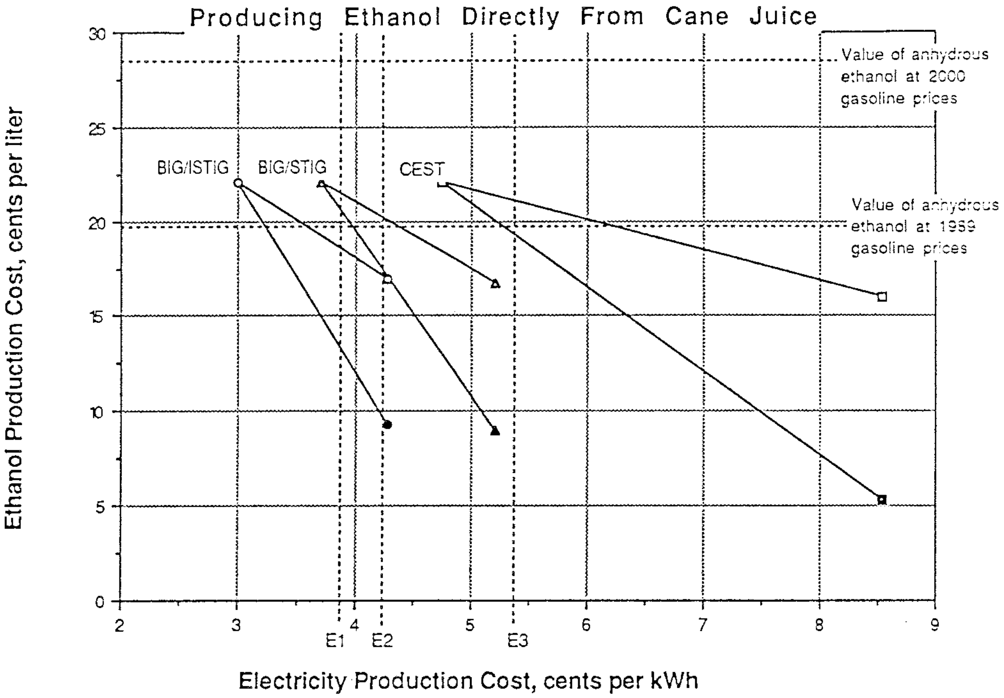


Figure 4.25: Production Cost of Sugar and Electricity For a Sugar Factory/Autonomous Capacity Annexed Distillery (Functioning as a Sugar Factory) with Cogeneration with Competitive Reference Lines.

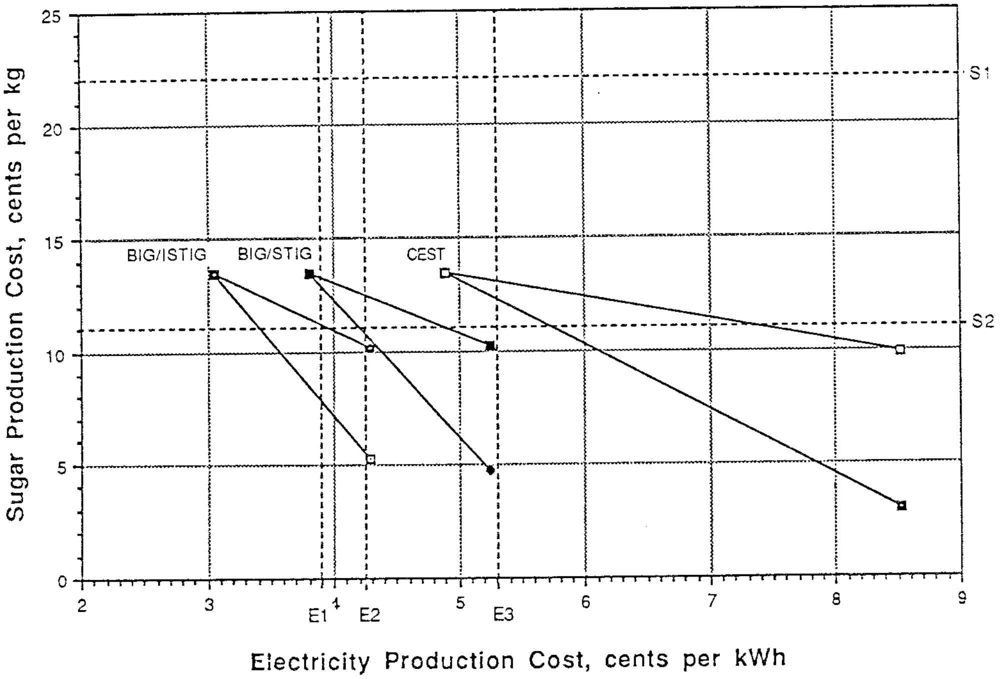


Figure 4.26: Internal Rate of Return for a Sugar Factory with an Autonomous Capacity Annexed Distillery with Cogeneration. *Off season fuel purchased from an outside agent at $3/GJ.*

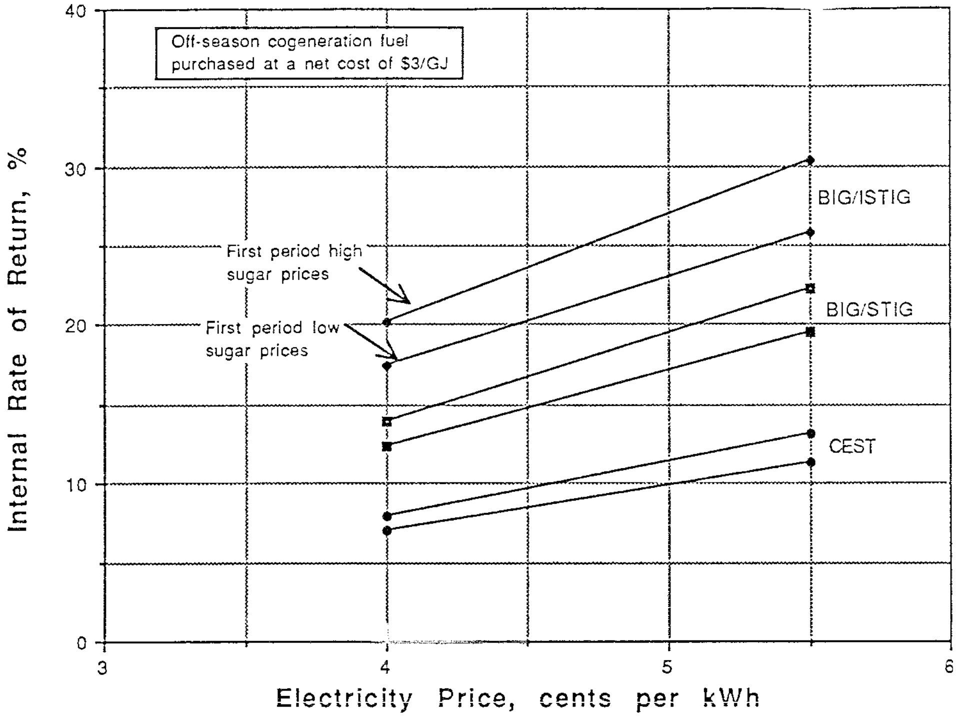


Figure 4.27: Internal Rate of Return for a Sugar Factory with an Autonomous Capacity Annexed Distillery with Cogeneration. *Off season fuel assumed to be barbojo, provided at the cost of gathering and processing.*

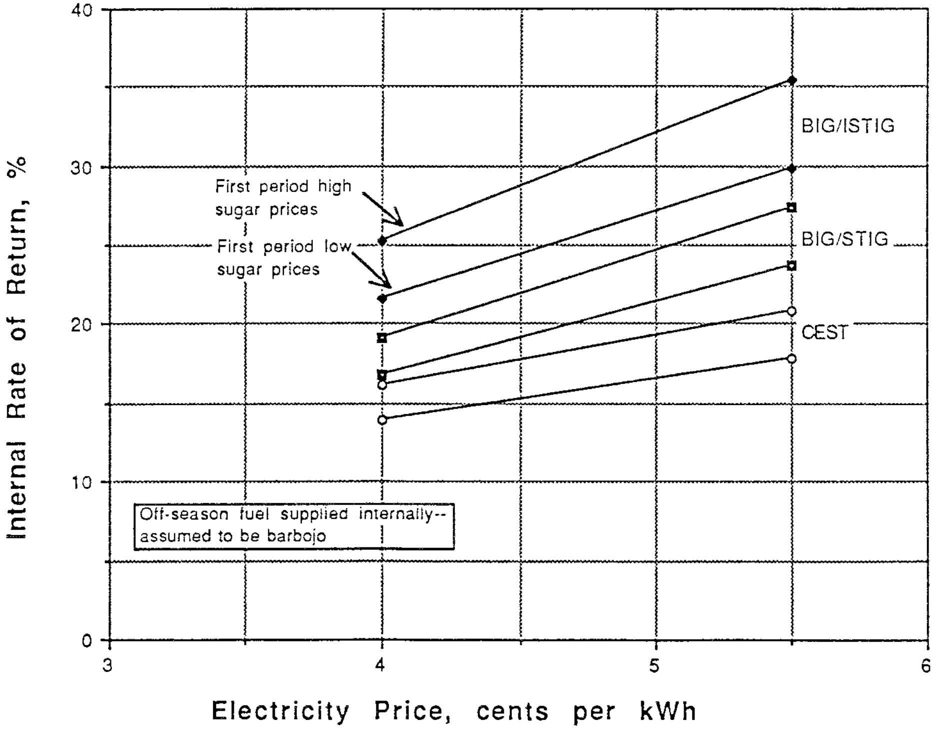


Figure 4.28: Internal Rate of Return for an Autonomous Distillery with Cogeneration. *Off season fuel purchased from an outside agent at \$3/GJ. The same reference lines as in figure 4.6 are shown.*

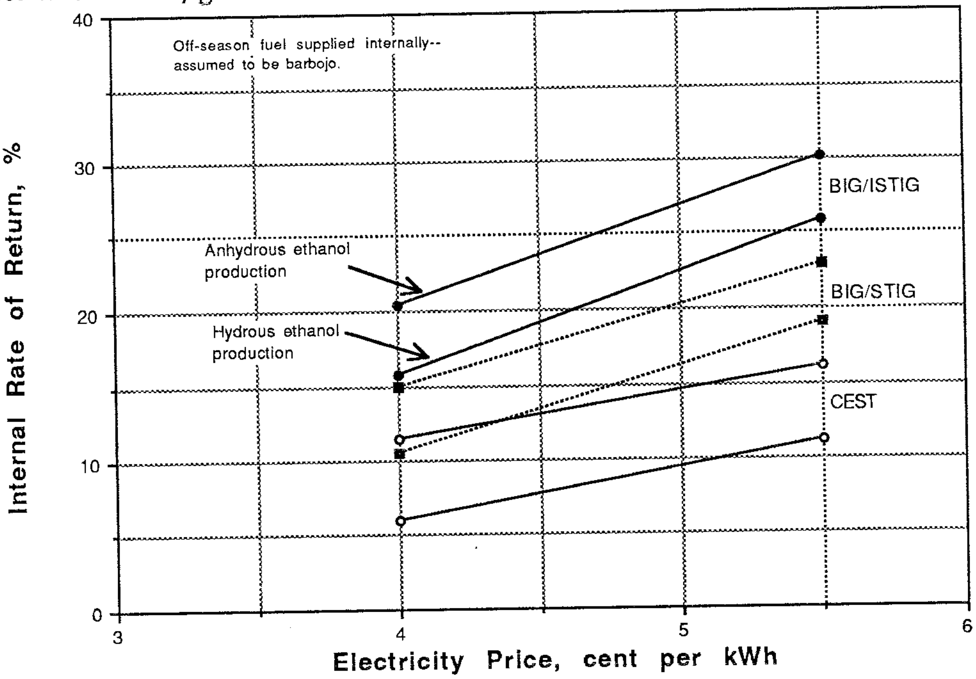
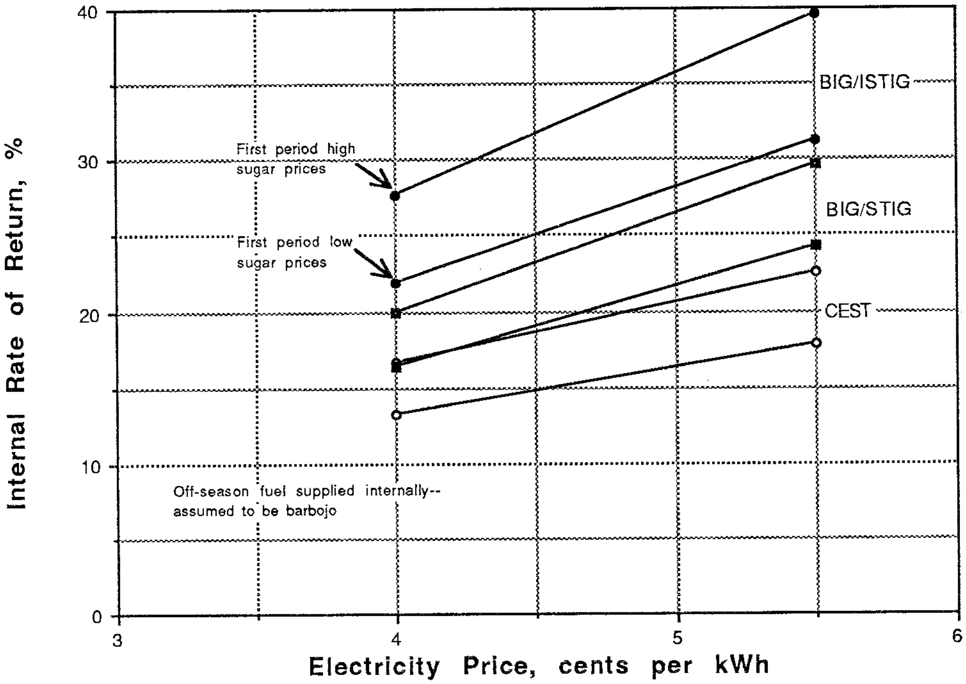


Figure 4.29: Internal Rate of Return for a Sugar Factory with Cogeneration. *Off season fuel assumed to be barbojo, provided at the cost of gathering and processing. the same reference lines as in Figure 4.17 are shown.*



Chapter 5: Policy Issues and Implications

1.0 Introduction

Chapter 4 demonstrated that using gas turbine cogeneration and Brazilian distillery technology, electricity and ethanol could be co-produced competitively. That analysis, however, examined only the private costs and benefits. A government or international development agency would also be interested in the social costs and benefits. Is electricity-ethanol co-production the best allocation of the capital, labor and natural resources of a developing country? What are the imperfections in the markets in which the co-products would have to compete? If co-production is beneficial, who in society would be benefitting? BIG/(I)STIG cogeneration is not a demonstrated technology; what are the technical uncertainties? Most of these questions cannot be answered in general; they depend on the local economic, political and social conditions. This chapter makes no pretense of answering these questions. Rather, it presents some of the major issues which need to be addressed. When it is appropriate, the Brazilian National Ethyl-Alcohol Program (Proalcool) experience will be used to illustrate issues and difficulties.

2.0 Technical Implementation Issues

All social, political and economic issues aside, the first question which should be addressed is the uncertainty of implementing new technologies. In the case of sugar factory and distillery conservation technologies, the uncertainties are minimal. With the exception of some of the advanced fermentation and distillation techniques, all of the low energy use technologies discussed in Chapter 2 have been demonstrated on an industrial scale in developing countries.

The primary technological uncertainty on the alcohol and sugar production side of co-production is in the implementation of the heat integrated designs developed in Chapter 3. Control and operating issues such as system stability, off design performance, operation during start-up and shut-down and physical plant layout are not within the scope of this thesis. Considering, though, that heat integration has been demonstrated in alcohol distillation and evaporation schemes both inside and outside the cane sugar industry, it is likely that the basic designs proposed in Chapter 3 would prove feasible.

A larger technological uncertainty falls with the gasifier/gas turbine systems. First, the technology is still in the development stage and has never been demonstrated. The cost and performance estimates used throughout this analysis will surely be refined as the technology becomes commercially available.

The use of both standard industrial gas turbines and aeroderivative gas turbines burning conventional fuels have been demonstrated in many developing countries. In 1987, for instance, 74 General Electric LM2500 simple cycle aeroderivative gas turbines were in use in developing countries, including Mexico, Sudan, New Guinea, India, Brazil and China [1]. Two General Electric LM5000 STIG systems have also recently been shipped to the Shenzhen Huaneng Economic Development Company, the first STIGs in the People's Republic of China [2]. The gas turbine systems also generate steam at 1.5 to 2.5 MPa, which is the same pressure range that is used in conventional sugar processing facilities. (Higher steam pressure systems, such as those found in CEST cogeneration (4.0-6.0 MPa) are more difficult to maintain.)

Due to the aeroderivative's small size and modular design, most major repair work (such as an overhaul) would be performed off-site by the manufacturer or by some other qualified service agent [3]. Again because of the fact that the engines would be aeroderivatives, their maintenance could be integrated into an existing airline infrastructure [3]. The length of forced stoppages would be minimal, due to the modular nature of the engine. If the cogeneration system was forced to shut down, a new engine section (or even engine) could be brought in quickly and swapped for the faulty section or engine [3]. When standard industrial turbines are used in the US, maintenance must be performed on site, shutting down the whole facility and requiring almost five times more man hours to perform than when aeroderivative turbines are used [3]. This disparity would likely be magnified in a developing country setting. While this style of maintenance still requires expertise among the operators and a fairly advanced technical infrastructure, these issues would be present in any large technological undertaking, including whatever alternative steam turbine cogeneration system which might be considered.

Uncertainty also lies with the biomass gasifier. It is generally believed that much of the extensive research and development on coal gasification would be applicable to biomass [3]. Biomass gasification in a pressurized bed gasifier is presently in the pilot and demonstration scales [4], but not specifically in conjunction with gas turbines. The clean-up of the fuel gasses prior to combustion in the gas turbine is a particularly important issue. Particulate matter must be thoroughly removes (in cyclones or filters), so as not to damage the turbine blades. The fuel gas must also be kept hot enough as to not condense the tar component while not being hot enough to vaporize corrosive alkali metals [5].

Outside of the gas generator of the gas turbine, none of the equipment required for a BIG/STIG or BIG/ISTIG system is outside the manufacturing capabilities of many developing countries. While some degree of industrial development would be required to manufacture the needed heat exchangers, boilers, pressure vessels, pumps, structures, etc., the only component which must be imported is the core of the gas turbine. This would minimize the hard currency requirements of such a BIG/(I)STIG cogeneration systems.

3.0 Resource Allocation Issues

3.1 Allocation of Labor

The problem of labor allocation in developing nations is generally one of unemployment, particularly in rural areas. Expanding sugar cane cultivation for ethanol production and introducing cogeneration might provide significant rural employment opportunities and mitigate the exodus of rural workers to the cities. In southeast Brazil, for instance, a 240,000 liter per day autonomous distillery requires about 850 workers to harvest the cane during the milling season [6]; in the northeast portion of Brazil where cultivation practices are more labor intensive, a 120,000 liter per day distillery might employ over 2800 workers [7]. Overall, an estimated 420,000 jobs were created in Brazil between 1980 and 1985 by the national alcohol fuel program [6,7]. In some cases, workers were hired full time, weeding, planting and cultivating seedlings during the off season [6].

The direct capital investment required for creating new jobs in cane-ethanol is significantly less than in most other industries. Including land costs as well as distillery and agricultural equipment costs, the investment required for an ethanol distillery is estimated to be around $25,000 per person-year in the Southeast of Brazil and $7,000 per person-year in the more labor intensive Northeast [7]. For comparison, the industrial sector requires an average investment of $42,000/person-year (1983 $)[7]. New jobs in pulp and paper, mining and petrochemicals can require investments over $100,000/person-year [7].

3.2 Foriegn Exchange Issues

One of the primary motives of Brazil's alcohol fuels program in the early 1980's was to reduce the amount of hard currency being used for petroleum imports. In 1970, 9% of Brazil's export income was used for crude petroleum imports. In 1979, immediately before the expansion of Proalcool to include neat alcohol burning cars, 46% of the export income was spent on oil imports [8,9]. On the surface, substituting domestic alcohol for gasoline derived from imported oil is a good solution to such a balance of payment problem. There is, however, a great deal of disagreement as to how beneficial this has proven for Brazil's balance of payments. Estimates for import savings through 1984 range from over $2.0 billion [10] to a net *loss* of foreign exchange over the same period [11]. If, however, oil prices had continued to rise throughout the 1980's (as they were expected to do when the alcohol fuels program was initiated), then the improvement in balance of payments would no doubt have been much greater.

Co-production can only improve the balance of payments over a fuel-alcohol program alone. The additional electricity component could offset additional petroleum imports without trading off hard currency earnings such as from sugar. Large amounts of hard capital would not be neccessary. The gas trubine core is the only complement which would have to be imported, and only accounts for about $6 million of the overall BIG/STIG capital cost [5].

Assuming that ethanol-electricity co-production is beneficial, who will make the investment? For sugar producers, the $42 million price tag of a BIG/STIG, let alone the $85 million investment in a BIG/ISTIG, is well beyond what they are used to investing, particularly in unfamiliar technologies. For a distillery with BIG/ISTIG cogeneration, the power production segment of the facility would cost two to four times that of the alcohol production segment. However, the investment in a BIG/STIG or BIG/ISTIG facility would be rather modest for an electric utility. Since in many developing countries, the cost of power cogenerated from cane would be lower than alternative supply sources, electric utility or third party involvement would be in the best interests of both the utility and the country.

3.3 Land and Natural Resources Issues

The production of fuel ethanol from biomass also raises a "food versus fuel" controversy. Brown [12] argues that energy crops would compete with food crops for not only land but also for agricultural investment dollars, water, fertilizer and technical expertise and that the primary beneficiaries of the alcohol fuels program are the most affluent members of a developing country society, whose automobiles are running on the subsidized alcohol, while the poor would benefit only indirectly.

In the Brazilian experience, many of these concerns are not nearly as severe as they might initially appear. In 1982, 7.7% of the cropland in Brazil was dedicated to sugar cane (3.9 million hectares, most of which was for sugar), up less than two percentage points from 1976 [7]. Over the same time period, the land dedicated to basic food crops increased 7% in absolute terms (up to 29.4million hectares), but decreased three percentage points in relative cropland terms (down to 58% of the total cropland). While in some local instances acreage was shifted from food to sugar cane, (Sao Paulo state in particular), there was not a large scale transfer of lands from basic foods to sugar cane.

For Brazil, the food versus fuel issue would be more aptly called the food versus export crop debate. The largest cropland increase from 1976 to 1982 came from the export crop soybean, whose land use increased 28% (absolute) and uses over twice the cropland as sugarcane [7]. In 1982, almost 31% of the cropland in Brazil was dedicated to export crops (*four times* that of sugar cane)[7]. While one should not discount the import earnings of such crops, the greater threat to basic food supplies comes from export crops such as soybeans and coffee rather than energy crops for domestic use [7,13].

This, analysis, however, cannot be generalized to all of the developing world. Brazil is unusual in that it has a wealth of land and a relatively low population density. For the more densely populated cane growing developing countries such as India and Bangladesh, agricultural issues become much more important and more detailed analysis would be required to evaluate the food versus fuel issues.

3.4 Markets and End Use Issues

There are a number of institutional and market difficulties which must be addressed before any co-production scenario can be implemented on a large scale. One of the largest difficulties is trying to introduce automobiles burning 100% ethanol[a] [9]. This requires a large simultaneous investment in both the end use industries (automobiles and light trucks) and the alcohol industry. Without strong government backing, the two sides would be caught in the classic chicken-and-egg situation. Neither the distilleries nor the auto makers would be willing to make

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[a] Programs calling for ethanol blending are not difficult to implement, because they require no changes in end use technologies or distribution infra-structure [9]. Gasoline engines can burn gasohol with up to 20% ethanol with only minor adjustments, and the existing gasoline distribution systems can be used with gasohol.

the initial step towards a neat alcohol transportation sector. The government needs to guarantee markets to the alcohol distillers and to the auto makers or provide investment incentives for both sides to overcome this who-goes-first stalemate.

Another potential market difficulty is the tension between producing sugar or ethanol. In the past year (1989), Brazil encountered high world sugar prices and low imported oil prices (and therefore, since ethanol prices are tied directly to gasoline prices, low ethanol prices). Many of the sugar factory/distillery owners shifted production to sugar, creating a shortage of ethanol at the pumps for the thousands of ethanol burning cars and forcing Brazil to *import* fuel alcohol from the US, France and elsewhere [14]. Alcohol burning cars were being retrofitted to burn gasoline-- just the opposite of the situation in the early 1980's.

This illustrates the fact that a government must address the fluctuations of the oil and sugar markets. When it is more profitable to produce sugar than ethanol, a government must either allow the market forces to work and face ethanol shortages and importation, or implement some kind of tariffs or regulations to provide sufficient domestic ethanol to meet demand. Net balance of payments (income from sugar exports versus costs of ethanol imports), domestic economic and social issues would have to be considered before any policy is formulated.

The consumer also plays a large role in the transition to ethanol fuels. Seeing that an automobile is often a major investment for an individual, the possibility that there might not be fuel for it in two or three years is a very large concern. In the first few years of the Brazilian program, consumer response to government signals concerning alcohol fuels was very elastic. The sales of alcohol fuel vehicles yo-yoed from less than 10% of the new car market to over 75%, depending on what signals were being sent by the government (primarily alcohol price ceilings) [9]. Until the market for ethanol and ethanol end use products is firmly in place, a government must maintain a solid commitment to ethanol fuels so that consumer confidence is high enough to get the market started.

At the same time, the electricity side of the co-production equation must be considered. Even if there is a dire need for electricity, there must also be a market for the cogenerated power to reach the consumer. Traditionally, power companies or the state boards who operate them have not been open to the idea of power produced by independent sources. In the case of Brazil, electric power has also been subsidized to the point where it would be very difficult for a cogeneration facility to compete.

Projected power shortages, however, are prompting new interest in cogeneration. A measure similar to the U.S.'s Public Utilities Regulatory Policy Act of 1978 (PURPA) is required in order to open the door for non-utility generated power. PURPA requires US electric utilities to purchase cogenerated and independent power from designated facilities at the avoided cost of producing that power themselves, and guarantees that the cogeneration facilities can purchase power from the grid at non-discriminatory rates[b]. Without such government guarantees (or directed ownership of the cogeneration facilities by the utilities), cogeneration facilities such as those proposed here would not be able to sell their excess electricity and therefore would have no reason to operate.

Because the cogeneration facilities would have to be located in rural areas among the cane fields, transmission and distribution of cane-derived electricity is an issue. If rural electrification is desired, as it is in many developing countries, then the relatively small cogeneration stations are ideal. They could be implemented in small steps, minimizing one-time expenditures and allowing a rural electrification program to progress in a manageable fashion. Transmission and distribution costs would be greatly reduced due to the proximity of the generating facility to the users. If, on the other hand, new generating capacity is required for

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[b] Brazil has recently announced a new policy similar to the US's PURPA, encouraging cogeneration and independent power. Thailand and other countries are also considering such programs [15].

urban settings, then the rural, potentially remote locations of the cane based cogenerating facilities might be a liability. Given the low production costs of BIG/STIG and BIG/ISTIG cogenerated power relative to other likely options (such as remote Amazonian hydroelectricity in Brazil), this transmission and distribution issue might not be significant.

4.0 Worldwide Potential

Aside from the social, economic and political barriers discussed in the previous section, it is important to ask what kind of impact electricity-ethanol co-production can make on the energy future of a cane growing developing nation.

4.2 Potential Ethanol Production

Although fuel ethanol cannot be viewed as a complete replacement for gasoline in most developing countries, if exploited even to a modest degree it can make a significant contribution to displacing gasoline imports. In 1987, approximately 909 million metric tonnes of sugar cane were harvested in developing countries [19]. Over 8 billion liters of anhydrous ethanol could have been fermented from the molasses of the 1987 harvest, displacing over 10% of the gasoline used in cane producing developing countries (Appendix VII).

Projecting ahead to future scenarios is somewhat more difficult. If the cane harvest continues to grow at 3% per year, and if 1/2 of that growth (1.5%)[c] is dedicated to neat ethanol production, then in 2007 (20 years) over 29 billion liters of hydrous ethanol might be produced. This is enough ethanol to replace 26% of

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[c] Because of pressures from other sweeteners, the demand for cane sugar is predicted to increase at only 1.5% per year [16].

Table 2.2: Summary of Potential Ethanol Production and Use in Cane Growing Developing Nations.

Region[a]	1987 Cane Harvested, 1000 tonnes[b]	2007 Cane Harvested, 1000 tonnes[c]	2007 Anhyd. Ethanol from Molasses, 1000 liters	2007 Hyd. Ethanol from Cane Juice, 1000 liters[d]	% of 2007 Gasoline Demand Which Could Be Met With Cane [e]
Africa	73,220	132,200	887,500	2,353,900	9%
Asia & Oceania	350,800	633,600	4,252,300	11,277,600	29%
Latin America	485,240	876,400	5,881,907	15,599,400	40%
Overall	909,260	1,642,200	11,021,700	29,230,900	37%

[a] Please see Appendix VII for productions of specific countries.
[b] From *FAO Production Yearbook* [19].
[c] Assumes 3% per year growth in cane, which is the historical growth rate since 1960 [18].
[d] Assumes that 1.5% of growth dedicated to sugar production, 1.5% to ethanol production; ethanol from both molasses and cane juice.
[e] 1987 gasoline consumption from [17], with gasoline consumption growing at same rates as petroleum consumption, taking the base level of growth in petroleum demand projected for developing countries by US Energy Information Administration (1.6%/yr)[17,18].

the projected gasoline consumption in the cane growing developing countries[d].

A summary of cane production, potential ethanol production and gasoline use in the major world regions are shown in Table 5.1. The expanded version of this table (Appendix VII) points out a number of countries where the potential is particularly large. For instance, Cuba, India, Bangladesh and Brazil could conceivably meet all of their "gasoline" needs in 2007 by using fuel ethanol grown domestically. Another five countries could displace ~50% of their projected gasoline use with hydrous ethanol. While the assumed growth patterns in both cane production and gasoline consumption are aggregate projections, and the market penetrations of ethanol fuelled vehicles are not likely to be this high, these values illustrate the order of magnitude of the potential of fuel ethanol.

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[d] Gasoline needs are based on 1987 gasoline consumption, assumed to grow at the same rate as projected non-OPEC LDC oil consumption (1.6%/yr)[17].

It should be reiterated at this point that these scenarios assume that the cane dedicated to ethanol comes only from new growth representing half of the historical growth rate; sugar production continues to expand. If world sugar markets remain low, or some of the international sugar quotas are broken, then shifting some of the cane production away from sugar to ethanol might be desirable. Alternatively, the co-production of electricity could make cane sugar more competitive, perhaps allowing it to recapture some of the sweetener markets lost in the past few years and keeping cane sugar production on its historical growth track.

4.2 Potential Electricity Production

Cogenerated electricity from bagasse and perhaps barbojo can potentially make a large impact on the electricity supply situation in many cane growing developing countries. As shown in Table 5.2 almost half of the electricity used in

Table 5.2: Summary of Potential Electricity Production from Sugar Cane Residues in Cane Growing Developing Nations.

Region[a]	1987 Cane Harvested, 1000 tonnes[b]	2007 Cane Harvested, 1000 tonnes[c]	1987 Electricity Potential from Cane Residues, GWh [d]	2007 Electricity Potential from Cane Residues, GWh	% of '87 Elect Demand Which Could Have been Met w/ 1987 Cane Residue[e]
Africa	73,220	132,200	64,800	117,035	30%
Asia & Oceania	350,800	633,600	310,500	560,700	34%
Cent'l America	147,500	266,400	130,600	235,800	92%
South America	340,700	615,400	301,700	544,600	82%
Overall	909,260	1,642,200	807,600	1,458,100	49%

[a] Please see Appendix VII for productions of specific countries.

[b] From *FAO Production Yearbook* [19].

[c] Assumes 3% per year growth in cane, which is the historical growth rate since 1960 [16].

[d] Assumes BIG/ISTIG cogeneration, producing 885 kWh per tonne cane (on average), 133 day milling season, producing power year round.

[e] 1987 electricity production from World Bank, reference [20]. See also Appendix VII.

cane growing developing countries in 1987 might have been met using cane based BIG/ISTIG cogeneration systems (807,000 GWh). In almost half of the countries considered, all of the electricity demands in 1987 could have been met using BIG/ISTIG cogeneration (Appendix VII). Assuming that cane production grows at its historical rate of ~3% per year [16], in 20 years (2007) over 1.4 million GWh could be produced from cane sugar residues.

These values are not predictions of what actually would be achieved by using sugar cane based cogeneration; they are based on historical trends, 100% market penetrations and the assumption that there would be local demand for all of the potential power. They do, however, indicate the extent of the resource and illustrate the order of magnitude of what might be possible if the resource can be exploited to its fullest.

4.3 Environmental Benefits

Probably the biggest environmental benefit of ethanol-electricity co-production is that, unlike fossil fuels, it produces virtually no net CO_2. It has been estimated that in using the advanced cogeneration technologies described throughout this analysis that anywhere from 116 kg of carbon per tonne cane milled (using CEST cogeneration with an average milling season length of 160 days) to 263 kg carbon per tonne cane milled (BIG/ISTIG cogeneration with an average milling season length of 133 days) might be saved over the same amount of energy produced using conventional fossil fuel technology [21]. When the more optimistic number is extrapolated to projected developing country cane production in 2027, over 750 megatonnes of carbon might be saved annually over the same energy produced using fossil fuels [22]. (The total 1989 developing world carbon emissions from fossil fuels was ~680 megatonnes.[22])

A major environmental issue presently being addressed in Brazil and

Columbia is the burning of the cane fields prior to harvesting. Field burning is done on the pretext of simplifying harvesting (increased productivity) and disposing of pests such as rats and snakes commonly found in cane fields. The practice, however, presents a number of environmental and health concerns, including massive amounts of ash spread across the countryside during field burning, degrading air quality for miles, and creating an unhealthy, sooty environment for the cane harvesters. Present focus on field burning might provide a larger impetus in the near term for barbojo cultivation and cogeneration.

Other environmental benefits include the potential reduction of urban ozone pollution and smog due to the use of the oxygenated fuels in vehicles and the reduction of sulfur dioxide and nitrous oxide pollution (precursors to acid rain) from the low sulfur biomass fuels and the low NO_x characteristics of steam injected gas turbine combustors[e].

5.0 Summary and Conclusions

There are number of clear potential benefits of co-producing electricity and ethanol. One is rural development. Large scale electricity-ethanol co-production could provide many thousands of rural jobs (with relatively minimal investment), helping to stem the migration of workers to the already overcrowded cities. It would also provide an ideal source of power for rural electrification; cogeneration facilities would be in the countryside where the power is desirable and at a scale where the low electricity use intensity found in rural areas would not be an inhibiting factor. The ethanol produced in sugar cane based distilleries and the

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[e] Steam injection dramatically decreases NO_X created via thermal mechanisms, but does not affect NO_X formed via nitrogen bound into the fuels, which probably would be the predominant source for raw biomass fuels. Even so, bagasse contains less fuel bound nitrogen than coal.

electricity generated from biomass residues could make a significant impact on petroleum imports, thus helping to relieve the balance of payments problems plaguing many developing nations.

A number of problems would have to be overcome, however, before large scale co-production could be implemented. BIG/(I)STIG cogeneration technologies would have to be commercialized to gain the maximum benefit. Some kind of policy analogous to PURPA in the US is required so that a market could open up for the power produced at the cogeneration facilities. Lastly, starting a neat alcohol fuel transportation sector would require firm support from the government in order to coordinate distillery and auto maker investments and to assure the consumers that ethanol powered cars are not merely a passing fad.

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Chapter 6: Conclusions

This thesis has examined the impact which modern cogneration technologies, efficient steam use in distilleries and sugar factories and the valuation of the agricultural residues of sugar cane cultivation have on fuel ethanol and cogenerated electricity production. The major findings are summarized here.

1.0 Energy Efficient Ethanol-Electricity Co-Production

When bagasse and perhaps barbojo are considered as potentially valuable fuel resources, their efficient use becomes an issue. One way this issue is addressed is through the implementation of electrically efficient cogneration technologies. The biomass integrated gasifier/steam injected gas turbine combination (BIG/STIG or BIG/ISTIG) appears to be technically feasible (although not commercially available at present) and economically attractive for cogeneration in sugar factories and alcohol distilleries.

Steam consumption in distilleries and sugar factories is another aspect of the efficient use of bagasse and barbojo. New designs derived here which integrate the evaporation section of an autonomous distillery with the distillation section can dramatically reduce steam demands level below either conventional autonomous distilleries or factories where the two sections are integrated separately. When evaporation and distillation are integrated separately, the overall steam demand is around 220 kg/tc (.27 MPa steam); when the evaporation and distillation sections are integrated together, the steam demand can drop to below 140 kg/tc. More energy efficient still is the case where the juice is not concentrated at all before fermentation (~120 kg/tc). The steam demands of all of the autonomous distillery designs synthesized can easily be met using BIG/ISTIG cogeneration.

The concept of the limiting process, as developed here, can be a useful tool in the total integration of an autonomous distillery. By identifying which of the processes in the system (either distillation or evaporation) limits further steam economy, the designer can concentrate his efforts on the appropriate section of the configuration. The limiting process approach should be able to be adapted to other pinch point analyses where two or more separate processes are being integrated.

The use of efficient process technologies and heat integration can reduce sugar factory steam demands by 50% over conventional designs. In order to meet the steam economy levels dictated by BIG/ISTIG cogneration, a heat integrated sugar factory with quintuple effect, falling film evaporation and a heat integrated annexed distillery are required. When rigorous energy targeting and heat integration techniques are applied to the sugar factory, steam demand is reduced 5% to 10% below that when heuristic, informal heat integration techniques are used.

In all but one of the cases analyzed (both sugar factories and autonomous distilleries), when steam is valued at the opportunity cost of the electricity it might have produced (3¢/kWh minimum), the savings associated with high steam efficiency are greater than the additional annual capital costs of the efficient technologies and the heat exchanger networks. The exception to this is the autonomous distillery that ferments un-concentrated cane juice, where electricity must be valued above ~4¢/kWh before the energy savings can overcome the additional capital costs.

2.0 Co-Production Economics

The use of efficient steam injected gas turbine cogeneration equipment and the treatment of bagasse and barbojo as valuable fuel resources allow electricity-ethanol co-production to be economically viable under many circumstances. When the residues of sugar/alcohol production are treated as valueless, then the lowest neat ethanol production cost attainable in an autonomous distillery is ~20¢/liter, which is equivalent to wholesale gasoline at 90¢/gallon. However, when bagasse and barbojo are valued at as little as $1.00/GJ[a] ($1.50/GJ for bagasse only), ethanol production costs can be reduced to below 14¢/liter (assuming Brazilian costs), the level required to match 1989 wholesale gasoline prices.

When BIG/STIG cogeneration is employed and bagasse sales are credited toward alcohol production, ethanol can be produced competitively at 1989 gasoline prices, while electricity production costs are ~5.3¢/kWh, the estimated production cost of a new coal fired power plant. With BIG/ISTIG cogeneration and both bagasse and barbojo revenues are credited to the autonomous distillery, the electricity production costs are reduced to 3.5¢/kWh while still producing ethanol competitively at 1989 gasoline prices.

Similar results can be found in the analysis of a sugar factory with cogeneration. A sugar factory operating under low cost Brazilian conditions with a BIG/ISTIG cogeneration facility can profitably co-produce sugar during the lowest swings in the sugar market while selling electricity at 4.0¢/kWh.

A particularly promising scenario involves gas turbine cogeneration with a

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[a] The price paid to the autonomous distillery beyond the value of the steam and electricity the cogenerator provides.

sugar factory and annexed distillery capable of fermenting all of the cane juice. Such a facility would be able to produce either sugar or anhydrous ethanol. Given the historical range of world sugar prices and the projected gasoline prices for the next 20 years, such a facility could be more profitable than either a sugar factory or an autonomous distillery, and could have internal rates of return well over 20%.

3.0 Potential Impact of Co-Production

Ethanol-electricity co-production in autonomous distilleries with advanced cogeneration systems could significantly impact the energy picture in many cane growing developing countries. Trends in cane production growth indicate that over 25% of projected gasoline needs in 2007 in cane producing developing countries might be met with neat fuel ethanol while still maintaining modest growth in sugar production. Some countries show much more promise, with well over half of their projected gasoline needs potentially being met by ethanol.

The electricity side of the co-production equation also shows potential. In almost half of the cane producing developing countries, the 1987 electricity demand could have been met using BIG/ISTIG cogeneration fueled by the bagasse and barbojo of that year's sugar crop.

A number of problems would have to be overcome, however, before large scale co-production could be implemented. BIG/(I)STIG cogeneration technologies would have to be commercialized, some kind of policy analogous PURPA in the US is required so that cogenerated power could be sold, and the initial hurdle of establishing a neat alcohol fuel transportation sector would have to be overcome.

There are several clear benefits to co-production besides the large amounts of energy which can be produced which might motivate further serious investigation of co-production. These benefits include rural industrialization and development, the reduction of petroleum imports and the prospect of an environmentally sound, domestic source of energy.

APPENDIX I:
DISTILLERY AND SUGAR FACTORY COMPONENT DATA AND PROBLEM TABLES

MP-LP CONCNETRATED FEED
Alcohol distillation sequence

jan 29 1990

HP column pressure:	1 atm	HP temp :	90 C	H feed:	377 kj/kg	(at high pressure)
LP column pressure:	.3 atm	LP temp :	63 C	H bottoms:	418 kj/kg	
% feed to hp column:	.541 %	% losses :	.05 %	distillate:	179 kj/kg	
Reflux ratio:	6	steam use :	1.36 kg/l(alc)	H feed:	377 kj/kg	(at low pressure)
feed flow :	16.76 kg	steam temp:	110 C	H bottoms:	418 kj/kg	
% alcohol in feed:	.1 %	steam Hfg :	2190 kj/kg	distillate:	179 kj/kg	
% alcohol in Dist.:	.95 %					

STREAM	Mass flow kg/l	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj	
Feed 1	9.07	4.38	35	90	-2184.28	43,98
Reboiler 1	2.75	22570	99.9	100	-6196.69	107.9, 108
Feed 2	7.69	4.4	35	63	-947.76	43, 71
Reboiler 2	2.33	23360	67.9	68	-5450.17	75.9, 76
condenser 1	6.68	9060	78.1	78	6053.05	
bottoms 1	8.11	4.19	100	40	2039.54	
distillate 1	.95	3.1	78	40	112.43	
condenser 2	5.67	9400	50.1	50	5328.30	
bottoms 2	6.88	4.19	66	40	749.84	
distillate 2	.81	3.1	50	40	25.10	
steam condensate	2.89	4.19	110	40	846.46	

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
110					6331.83	2.89 kg steam
108	2	12.09	24.18	24.18	6356.01	
107.9	.1	-61954.79	-6195.48	-6171.29	160.53	
100	7.9	12.09	95.53	-6075.77	256.06	
98	2	24.39	48.79	-6026.98	304.85	
78.1	19.9	-15.32	-304.85	-6331.83	.00	
78	.1	60515.13	6051.51	-280.31	6051.51	
76	2	-12.36	-24.72	-305.04	6026.79	
75.9	.1	-54514.11	-5451.41	-5756.45	575.38	
71	4.9	-12.36	-60.57	-5817.01	514.82	
66	5	-46.21	-231.04	-6048.06	283.77	
50.1	15.9	-17.37	-276.17	-6324.22	7.61	
50	.1	53265.67	5326.57	-997.66	5334.17	
43	7	-14.86	-104.01	-1101.67	5230.16	
40	3	58.70	176.11	-925.55	5406.27	

AUTONOMOUS DISTILLERY: HP COLUMN-->EFF1-->2-->3-->4-->5-->LP-->6

feb 12 1990

do not use this spreadsheet for any other configuration!!!

Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

				H feed:	627 kj/kg	(at high pressure)
				H bottoms:	671 kj/kg	
				H distillate:	439 kj/kg	
HP column pressure:	6 atm	HP temp :	160 C	H feed:	627 kj/kg	(at medium pressure)
MP column pressure:	4.2 atm	MP temp :	146	H bottoms:	671 kj/kg	
LP column pressure:	1 atm	LP temp :	100 C	H distillate:	439 kj/kg	
% feed to hp column:	.878 %	% losses :	.05 %	H feed:	377 kj/kg	(at low pressure)
% feed to mp column:	0 %	steam use :	1.86 kg/l(alc)	H bottoms:	418 kj/kg	
LP Reflux ratio:	5	HP steam temp :	170 C	H distillate:	179 kj/kg	
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg			
% alcohol in feed:	.0811968102 %	LP steam temp :	130 C			
% alcohol in Dist.:	.9 %	LP steam Hfg :	2100 kj/kg			

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	108.2	22030	119.9	120	-238364.60
Effect 2	90.3	22300	109.9	110	-201369.00
Effect 3	44	22570	99.9	100	-99308.00
Effect 4	33	22830	89.9	90	-75339.00
Effect 5	22.5	23090	79.9	80	-51952.50
Effect 6	27	23370	67.9	68	-63099.00
Effect 1 vapors	108.2	22030.00	120.1	120	238364.60
Effect 2 vapors	90.3	22300.00	110.1	110	201369.00
Effect 3 vapors	44	22570	100.1	100	99308.00
Effect 4 vapors	33	22830	90.1	90	75339.00
Effect 5 vapors	22.5	23090	80.1	80	75339.00
Effect 6 vapors	27	23370	68.1	68	63099.00
combined evap cond	275.5	4.18	109.934664247	35	86294.01
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	592.65	4.38	35	150	-298517.81
Reboiler HP	134.20	20860	157.9	158	-279943.14
condenser HP	320.81	8300	132.1	132	266271.14
bottoms HP	539.18	4.19	158	35	277878.18
distillate HP	53.47	3.1	132	35	16077.86
Feed MP	.00	4.38	35	135	.00
Reboiler MP	.00	20860	144.9	145	.00
condenser MP	.00	8300	110.1	110	.00
bottoms MP	.00	4.19	158	35	.00
distillate MP	.00	3.1	132	35	.00
Feed LP	82.35	4.38	35	90	-19838.12
Reboiler LP	18.60	22570	99.9	100	-41987.50
condenser LP	44.58	9060	78.1	78	40386.81
bottoms LP	74.92	4.19	100	35	20404.60
distillate LP	7.43	3.1	78	35	990.35
HP steam condensate	136.25	4.19	180	35	82779.47
LP steam concensate	0	4.19	130	35	.00

HP-1-2-3-4-5-LP-6(a)
Composite Curves   Grand Composite Cur
Enthalpy flow, kJ/tc
Temperature, C

PROBLEM TABLE: HP-1-2-3-4-5-LP-6

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	279315	136.25 kg steam
169	1	571	571	571	279886	133.01
168.9	.1	-2798860	-279886	-279315	0	
160	8.9	571	5081	-274234	5081	
158	2	-2025	-4050	-278284	1031	
155	3	234	703	-277581	1734	
154.9	.1	234	23	-277558	1757	
145	9.9	234	2319	-275239	4076	
132.1	12.9	234	3022	-272217	7098	
132	.1	2662946	266295	-5922	273393	
130	2	400	800	-5122	274193	
129.9	.1	-2383246	-238325	-243447	35868	
120.1	9.8	-3444	-33753	-277200	2116	
120	.1	2380202	238020	-39179	240136	
119.9	.1	-2017134	-201713	-240893	38422	
111	8.9	-3444	-30653	-271546	7769	
110.1	.9	-6137	-5523	-277069	2246	
110	.1	2007553	200755	-76314	203001	
109.9	.1	-1419092	-141909	-218223	61092	
105	4.9	-6137	-30070	-248293	31022	
100.1	4.9	-6137	-30070	-278363	952	
100	.1	986943	98694	-179669	99646	
99.9	.1	-755729	-75573	-255242	24073	
90.1	9.8	-2339	-22926	-278167	1148	
90	.1	751051	75105	-203062	76253	
89.9	.1	-521864	-52186	-255249	24066	
80.1	9.8	-2339	-22926	-278174	1141	
80	.1	517186	51719	-226456	52859	
78.1	1.9	-2339	-4445	-230901	48415	
78	.1	401529	40153	-190748	88567	
77.9	.1	-633306	-63331	-254078	25237	
68.1	9.8	-2316	-22700	-276778	2537	
68	.1	628674	62867	-213911	65404	
45	23	168	3873	-210038	69277	
40	5	3125	15624	-194414	84901	
35	5	6969	34845	-159569	119747	

```
AUTONOMOUS DISTILLERY:  HP COLUMN-->EFF1-->2-->3-->4-->5-->LP-->6                  feb 12 1990
do not use this spreadsheet for any other configuration!!!
Juice treatment & fermentation parameters

     raw juice flow:         1000 kg/TC      Raw juice brx:        13.5 Brix         Raw juice T:        30 C
  TFS per tonne cane:         144 kg/TC   Brx into fermntr:          20 Brix      T into fermntr:        30 C
     Recovery of TFS:         .84 %        # evap.stages:           2          T out of fermntr:        35 C
 % Guy-Lassac yield:          .91 %           T evap. 1:          120 C
       Alcohol yield:       70.45 l/TC        T evap. 2:          110 C
          kg alcohol:       54.81 kg/TC   Juice into frmnt:     675.00 kg/TC
      alc. conc. out:         .08 %       TFS in fermtr fd:        .18 %

Alcohol distillation parameters                                                       H feed:       627 kj/kg  (at high pressure)
                                                                                   H bottoms:       671 kj/kg
 HP column pressure:            6 atm             HP temp :        160 C         H distillate:       439 kj/kg
 MP column pressure:          4.2 atm             MP temp :        146                 H feed:       627 kj/kg  (at medium pressure)
 LP column pressure:            1 atm             LP temp :        100 C            H bottoms:       671 kj/kg
% feed to hp column:         .901 %              % losses :        .05 %          H distillate:       439 kj/kg
% feed to mp column:            0 %             steam use :       1.90 kg/l(alc)        H feed:       377 kj/kg  (at low pressure)
   LP Reflux ratio:             5            HP steam temp :        170 C            H bottoms:       418 kj/kg
         feed flow :          675 kg          HP steam Hfg :       2050 kj/kg      H distillate:       179 kj/kg
  % alcohol in feed: .0811968102 %          LP steam temp :        130 C
 % alcohol in Dist.:           .9 %          LP steam Hfg :       2100 kj/kg

                                     dH(fg) or
                          Mass flow         Cp                Tin          Tout          M*Cp
STREAM                           kg   kj/kg(C)                  C             C            kj
-------------------------------------------------------------------------------------------------
Raw Juice                      1000       3.84                 30           102    -276780.74
Clear Juice                    1000       3.84                 90           120    -115325.31
Effect 1                        112      22030              119.9           120    -246736.00
Effect 2                         94      22300              109.9           110    -209620.00
Effect 3                         30      22570               99.9           100     -67710.00
Effect 4                      29.66      22830               89.9            90     -67713.78
Effect 5                      29.32      23090               79.9            80     -67699.88
Effect 6                      30.02      23370               67.9            68     -70156.74
Effect 1 vapors                 112   22030.00              120.1           120     246736.00
Effect 2 vapors                  94   22300.00              110.1           110     209620.00
Effect 3 vapors                  30      22570              100.1           100      67710.00
Effect 4 vapors               29.66      22830               90.1            90      67713.78
Effect 5 vapors               29.32      23090               80.1            80      67713.78
Effect 6 vapors               30.02      23370               68.1            68      70156.74
combined evap cond           265.66       4.18    110.853722804            35      84232.43
Fermenter feed                  675       3.68                 68            35      81995.17
Feed HP                      608.18       4.38                 35           150    -306337.75
Reboiler HP                  137.72      20860              157.9           158    -287276.50
condenser HP                 329.21       8300              132.1           132     273246.35
bottoms HP                   553.31       4.19                158            35     285157.44
distillate HP                 54.87        3.1                132            35      16499.03
Feed MP                         .00       4.38                 35           135           .00
Reboiler MP                     .00      20860              144.9           145           .00
condenser MP                    .00       8300              110.1           110           .00
bottoms MP                      .00       4.19                158            35           .00
distillate MP                   .00        3.1                132            35           .00
Feed LP                       66.83       4.38                 35            90     -16098.14
Reboiler LP                   15.10      22570               99.9           100     -34071.83
condenser LP                  36.17       9060               78.1            78      32772.90
bottoms LP                    60.80       4.19                100            35      16557.83
distillate LP                  6.03        3.1                 78            35        803.65
HP steam condensate          139.85       4.19                180            35      84968.18
LP steam condensate               0       4.19                130            35           .00
```

```
PROBLEM TABLE: HP-1-2-3-4-5-LP-6(B)
                                                                                 corrected
                 cor'ctd       delta        Cp(hot)-        Delta        heat         heat
                    temp        temp        Cp(cold)            H     cascade      cascade
------------------------------------------------------------------------------------------------------------------
                     170                                                  .00       286700    139.85 kg steam
                     169           1             586          586         586       287286    136.52
                   168.9          .1        -2872179      -287218     -286632           68
                     160         8.9             586         5215     -281417         5284
                     158           2           -2078        -4156     -285572         1128
                     155           3             241          722     -284851         1850                     HP-1-2-3-4-5-LP-6(b
                   154.9          .1             241           24     -284827         1874                     Grand Composite Cur
                     145         9.9             241         2381     -282445         4255                     Enthalpy, flow, kJ/
                   132.1        12.9             241         3103     -279342         7358                     Temperature, C
                     132          .1         2732704       273270       -6072       280628
                     130           2             411          821       -5251       281450
                   129.9          .1        -2466949      -246695     -251946        34755
                   120.1         9.8           -3434       -33649     -285594         1106
                     120          .1         2463926       246393      -39202       247498
                   119.9          .1        -2099634      -209963     -249165        37535
                     111         8.9           -3434       -30559     -279724         6977
                   110.1          .9           -6167        -5551     -285274         1426
                     110          .1         2090033       209003      -76271       210429
                   109.9          .1        -1023986      -102399     -178670       108031
                     105         4.9           -6167       -30220     -208889        77811
                   100.1         4.9           -6167       -30220     -239109        47591
                     100          .1          670933        67093     -172016       114685
                    99.9          .1         -679499       -67950     -239965        46735
                    90.1         9.8           -2361       -23138     -263104        23597
                      90          .1          674777        67478     -195626        91074
                    89.9          .1         -679360       -67936     -263562        23138
                    80.1         9.8           -2361       -23138     -286700            0
                      80          .1          674638        67464     -219237        67464
                    78.1         1.9           -2361        -4486     -223723        62978
                      78          .1          325368        32537     -191186        95515
                    77.9          .1         -703910       -70391     -261577        25124
                    68.1         9.8           -2342       -22955     -284532         2168
                      68          .1          699225        69923     -214609        72091
                      45          23             142         3274     -211335        75365
                      40           5            3099        15494     -195841        90859
                      35           5            6943        34715     -161126       125574
```

AUTONOMOUS DISTILLERY: HP COLUMN-->EFF1-->EFF2-->EFF3-->EFF4 feb 23 1990

do not use this spreadsheet for any other configuration!!!
Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

HP column pressure:	6 atm	HP temp :	160 C
MP column pressure:	4.2 atm	MP temp :	146
LP column pressure:	1 atm	LP temp :	100 C
% feed to hp column:	1 %	% losses :	.05 %
% feed to mp column:	0 %	steam use :	2.11 kg/l(alc)
LP Reflux ratio:	5	HP steam temp :	170 C
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg
% alcohol in feed:	.0811968102 %	LP steam temp :	130 C
% alcohol in Dist.:	.9 %	LP steam Hfg :	2100 kj/kg

H feed:	627 kj/kg	(at high pressure)
H bottoms:	671 kj/kg	
H distillate:	439 kj/kg	
H feed:	627 kj/kg	(at medium pressure)
H bottoms:	671 kj/kg	
H distillate:	439 kj/kg	
H feed:	377 kj/kg	(at low pressure)
H bottoms:	418 kj/kg	
H distillate:	179 kj/kg	

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj	ADJUSTED Tin C	ADJUSTED Tout C
Raw Juice	1000	3.84	30	102	-276780.74	40	112
Clear Juice	1000	3.84	90	120	-115325.31	100	130
Effect 1	123	22030	119.9	120	-270969.00	129.9	130
Effect 2	68.15	22300	109.9	110	-151974.50	119.9	120
Effect 3	67.33	22570	99.9	100	-151963.81	109.9	110
Effect 4	66.52	22830	89.9	90	-151865.16	99.9	100
Effect 5	0	23090	79.9	80	.00	89.9	90
Effect 6	0	23370	67.9	68	.00	77.9	78
Effect 1 vapors	123	22030.00	120.1	120	270969.00		
Effect 2 vapors	68.15	22300.00	110.1	110	151974.50		
Effect 3 vapors	67.33	22570	100.1	100	151963.81		
Effect 4 vapors	66.52	22830	90.1	90	151865.16		
Effect 5 vapors	0	23090	80.1	80	151865.16		
Effect 6 vapors	0	23370	68.1	68	.00		
combined evap cond	325	4.18	107.619384615	35	98653.43		
Fermenter feed	675	3.68	68	35	81995.17		
Feed HP	675.00	4.38	35	150	-339997.50	45	160
Reboiler HP	152.85	20860	157.9	158	-318841.84	167.9	168
condenser HP	365.39	8300	132.1	132	303270.09		
bottoms HP	614.10	4.19	158	35	316489.95		
distillate HP	60.90	3.1	132	35	18311.91		
Feed MP	.00	4.38	35	135	.00	45	145
Reboiler MP	.00	20860	144.9	145	.00	154.9	155
condenser MP	.00	8300	110.1	110	.00		
bottoms MP	.00	4.19	158	35	.00		
distillate MP	.00	3.1	132	35	.00		
Feed LP	.00	4.38	35	90	.00	45	100
Reboiler LP	.00	22570	99.9	100	.00	109.9	110
condenser LP	.00	9060	78.1	78	.00		
bottoms LP	.00	4.19	100	35	.00		
distillate LP	.00	3.1	78	35	.00		
HP steam condensate	155.18	4.19	180	35	94281.86		
LP steam condensate	0	4.19	130	35	.00		

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	318127	155.18 kg steam
169	1	650	650	650	318777	151.49
168.9	.1	-3187768	-318777	-318127	0	
160	8.9	650	5787	-312340	5787	
158	2	-2306	-4613	-316952	1174	
155	3	267	800	-316152	1975	
154.9	.1	267	27	-316125	2002	
145	9.9	267	2641	-313484	4643	
132.1	12.9	267	3442	-310042	8085	
132	.1	3032968	303297	-6745	311382	
130	2	456	911	-5834	312293	
129.9	.1	-2709234	-270923	-276757	41369	
120.1	9.8	-3389	-33208	-309965	8161	
120	.1	2706301	270630	-39335	278791	
119.9	.1	-1523134	-152313	-191649	126478	
111	8.9	-3389	-30158	-221807	96319	
110.1	.9	-5874	-5287	-227094	91033	
110	.1	1513871	151387	-75707	242420	
109.9	.1	-1525512	-152551	-228258	89868	
105	4.9	-5874	-28784	-257042	61085	
100.1	4.9	-5874	-28784	-285826	32301	
100	.1	1513764	151376	-134450	183677	
99.9	.1	-1520682	-152068	-286518	31609	
90.1	9.8	-2030	-19895	-306413	11714	
90	.1	1516622	151662	-154750	163376	
89.9	.1	455	45	-154705	163422	
80.1	9.8	455	4455	-150250	167877	
80	.1	455	45	-150204	167922	
78.1	1.9	455	864	-149340	168786	
78	.1	455	45	-149295	168832	
77.9	.1	455	45	-149250	168877	
68.1	9.8	455	4455	-144794	173332	
68	.1	455	45	-144749	173378	
45	23	455	10456	-134293	183834	
40	5	3411	17056	-117237	200890	
35	5	7255	36276	-80961	237166	

AUTONOMOUS DISTILLERY: MP COLUMN-->EFF3--.EFF4--.EFF5

feb 12 1990

do not use this spreadsheet for any other configuration!!!

Juice treatment & fermentation parameters

raw juice flow: 1000 kg/TC
TFS per tonne cane: 144 kg/TC
Recovery of TFS: .84 %
% Guy-Lassac yield: .91 %
Alcohol yield: 70.45 l/TC
kg alcohol: 54.81 kg/TC
alc. conc. out: .08 %

Raw juice brx: 13.5 Brix
Brx into fermntr: 20 Brix
evap.stages: 2
T evap. 1: 120 C
T evap. 2: 110 C
Juice into frmnt: 675.00 kg/TC
TFS in fermtr fd: .18 %

Raw juice T: 30 C
T into fermntr: 30 C
T out of fermntr: 35 C

Alcohol distillation parameters

HP column pressure: 6 atm
MP column pressure: 4.2 atm
LP column pressure: 1 atm
% feed to hp column: 0 %
% feed to mp column: 1 %
LP Reflux ratio: 5
feed flow : 675 kg
% alcohol in feed: .0811968102 %
% alcohol in Dist.: .9 %

HP temp : 160 C
MP temp : 146
LP temp : 100 C
% losses : .05 %
steam use : 2.24 kg/l(alc)
HP steam temp : 170 C
HP steam Hfg : 2050 kj/kg
LP steam temp : 156 C
LP steam Hfg : 2095 kj/kg

H feed: 627 kj/kg (at high pressure)
H bottoms: 671 kj/kg
H distillate: 439 kj/kg
H feed: 570 kj/kg (at medium pressure)
H bottoms: 619 kj/kg
H distillate: 356 kj/kg
H feed: 377 kj/kg (at low pressure)
H bottoms: 418 kj/kg
H distillate: 179 kj/kg

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	0	22030	119.9	120	.00
Effect 2	0	22300	109.9	110	.00
Effect 3	102	22570	99.9	100	-230214.00
Effect 4	88	22830	89.9	90	-200904.00
Effect 5	77	23090	79.9	80	-200904.00
Effect 6	58	23370	67.9	68	-135546.00
Effect 1 vapors	0	22030.00	120.1	120	.00
Effect 2 vapors	0	22300.00	110.1	110	.00
Effect 3 vapors	102	22570	100.1	100	230214.00
Effect 4 vapors	88	22830	90.1	90	200904.00
Effect 5 vapors	77	23090	80.1	80	200904.00
Effect 6 vapors	58	23370	68.1	68	135546.00
combined evap cond	190	4.18	95.3684210526	35	47944.60
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	.00	4.38	35	150	.00
Reboiler HP	.00	20860	157.9	158	.00
condenser HP	.00	8300	132.1	132	.00
bottoms HP	.00	4.19	158	35	.00
distillate HP	.00	3.1	132	35	.00
Feed MP	675.00	4.38	35	135	-295650.00
Reboiler MP	154.85	21300	144.9	145	-329829.04
condenser MP	365.39	8560	110.1	110	312770.11
bottoms MP	614.10	4.19	158	35	316489.95
distillate MP	60.90	3.1	132	35	18311.91
Feed LP	.00	4.38	35	90	.00
Reboiler LP	.00	22570	99.9	100	.00
condenser LP	.00	9060	78.1	78	.00
bottoms LP	.00	4.19	100	35	.00
distillate LP	.00	3.1	78	35	.00
HP steam condensate	.00	4.19	180	35	.00
LP steam condensate	164.2997765	4.19	130	35	65399.53

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	336815	164.30 kg steam
169	1	0	0	0	336815	160.77
168.9	.1	0	0	0	336815	
160	8.9	0	0	0	336815	
158	2	0	0	0	336815	
155	3	0	0	0	336815	
154.9	.1	-3298290	-329829	-329829	6985	
145	9.9	0	0	-329829	6985	
132.1	12.9	-383	-4946	-334775	2039	
132	.1	-383	-38	-334813	2001	
130	2	-383	-767	-335580	1234	
129.9	.1	305	31	-335550	1265	
120.1	9.8	305	2989	-332561	4254	
120	.1	305	31	-332530	4284	
119.9	.1	305	31	-332500	4315	
112	7.9	305	2410	-330090	6724	
110.1	1.9	-3539	-6724	-336815	0	
110	.1	3124162	312416	-24398	312416	
109.9	.1	-2309335	-230933	-255332	81483	
105	4.9	-7195	-35253	-290585	46229	
100.1	4.9	-7195	-35253	-325839	10976	
100	.1	2294945	229495	-96344	240471	
99.9	.1	-2012390	-201239	-297583	39231	
95	4.9	-2556.18932863	-12525	-310108	26706	
90.1	4.9	-2556	-12525	-322634	14181	
90	.1	2006484	200648	-121985	214829	
89.9	.1	-1780486	-178049	-300034	36781	
80.1	9.8	-2556	-25051	-325085	11730	
80	.1	1775374	177537	-147547	189267	
78.1	11.9	-2556	-30419	-152404	184411	
78	.1	-2556	-256	-152660	184155	
77.9	.1	-1358016	-135802	-288461	48353	
68.1	9.8	-2556	-25051	-313512	23303	
68	.1	1352904	135290	-178221	158593	
45	23	-71	-1644	-179866	156949	
40	5	2885	14425	-165441	171374	
35	5	6729	33646	-131795	205020	

AUTONOMOUS DISTILLERY: MP COLUMN-->EFF3--.EFF4--.EFF5 feb 12 1990
do not use this spreadsheet for any other configuration!!!
Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

				H feed:	627 kj/kg	(at high pressure)
				H bottoms:	671 kj/kg	
HP column pressure:	6 atm	HP temp :	160 C	H distillate:	439 kj/kg	
MP column pressure:	4.2 atm	MP temp :	146	H feed:	570 kj/kg	(at medium pressure)
LP column pressure:	1 atm	LP temp :	100 C	H bottoms:	619 kj/kg	
% feed to hp column:	0 %	% losses :	.05 %	H distillate:	356 kj/kg	
% feed to mp column:	1 %	steam use :	2.43 kg/l(alc)	H feed:	377 kj/kg	(at low pressure)
LP Reflux ratio:	5	HP steam temp :	170 C	H bottoms:	418 kj/kg	
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg	H distillate:	179 kj/kg	
% alcohol in feed:	.0811968102 %	LP steam temp :	156 C			
% alcohol in Dist.:	.9 %	LP steam Hfg :	2095 kj/kg			
			174.3304266			

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	0	22030	119.9	120	.00
Effect 2	0	22300	109.9	110	.00
Effect 3	120	22570	99.9	100	-270840.00
Effect 4	108	22830	89.9	90	-246564.00
Effect 5	97	23090	79.9	80	-246564.00
Effect 6	0	23370	67.9	68	.00
Effect 1 vapors	0	22030.00	120.1	120	.00
Effect 2 vapors	0	22300.00	110.1	110	.00
Effect 3 vapors	120	22570	100.1	100	270840.00
Effect 4 vapors	108	22830	90.1	90	246564.00
Effect 5 vapors	97	23090	80.1	80	246564.00
Effect 6 vapors	0	23370	68.1	68	.00
combined evap cond	228	4.18	95.2631578947	35	57433.20
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	.00	4.38	35	150	.00
Reboiler HP	.00	20860	157.9	158	.00
condenser HP	.00	8300	132.1	132	.00
bottoms HP	.00	4.19	158	35	.00
distillate HP	.00	3.1	132	35	.00
Feed MP	675.00	4.38	35	135	-295650.00
Reboiler MP	155.19	21300	144.9	145	-330559.81
condenser MP	365.39	8580	110.1	110	313500.88
bottoms MP	614.10	4.19	158	35	316489.95
distillate MP	60.90	3.1	132	35	18311.91
Feed LP	.00	4.38	35	90	.00
Reboiler LP	.00	22570	99.9	100	.00
condenser LP	.00	9060	78.1	78	.00
bottoms LP	.00	4.19	100	35	.00
distillate LP	.00	3.1	78	35	.00
HP steam condensate	.00	4.19	180	35	.00
LP steam condensate	178.1571916	4.19	130	35	70915.47

PROBLEM TABLE: MP-3-4-5

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	365222	178.16 kg steam
169	1	0	0	0	365222	174.33
168.9	.1	0	0	0	365222	
160	8.9	0	0	0	365222	
158	2	0	0	0	365222	
155	3	0	0	0	365222	
154.9	.1	-3305598	-330560	-330560	34662	
145	9.9	0	0	-330560	34662	
132.1	12.9	-383	-4946	-335506	29716	
132	.1	-383	-38	-335544	29678	
130	2	-383	-767	-336311	28911	
129.9	.1	363	36	-336275	28948	
120.1	9.8	363	3558	-332717	32506	
120	.1	363	36	-332680	32542	
119.9	.1	363	36	-332644	32578	
112	7.9	363	2868	-329776	35446	
110.1	1.9	-3481	-6614	-336390	28832	
110	.1	3131528	313153	-23237	341985	
109.9	.1	-2715537	-271554	-294791	70431	
105	4.9	-7137	-34969	-329760	35463	
100.1	4.9	-7137	-34969	-364728	494	
100	.1	2701263	270126	-94602	270620	
99.9	.1	-2468932	-246893	-341495	23727	
95	4.9	-2339.28675953	-11463	-352958	12264	
90.1	4.9	-2339	-11463	-364420	802	
90	.1	2463301	246330	-118090	247132	
89.9	.1	-2242069	-224207	-342297	22925	
80.1	9.8	-2339	-22925	-365222	0	
80	.1	2237391	223739	-141483	223739	
78.1	11.9	-2339	-27838	-145928	219294	
78	.1	-2339	-234	-146162	219060	
77.9	.1	-2339	-234	-146396	218827	
68.1	9.8	-2339	-22925	-169321	195902	
68	.1	-2339	-234	-169555	195668	
45	23	145	3345	-166210	199012	
40	5	3102	15510	-150700	214522	
35	5	6946	34730	-115970	249252	

AUTONOMOUS DISTILLERY: EFF2-->EFF3-->EFF4-->EFF5 feb 12 1990

do not use this spreadsheet for any other configuration!!!

Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

				H feed:	627 kj/kg	(at high pressure)
				H bottoms:	671 kj/kg	
HP column pressure:	6 atm	HP temp :	160 C	H distillate:	439 kj/kg	
MP column pressure:	4.2 atm	MP temp :	146	H feed:	627 kj/kg	(at medium pressure)
LP column pressure:	1 atm	LP temp :	100 C	H bottoms:	671 kj/kg	
% feed to hp column:	0 %	% losses :	.05 %	H distillate:	439 kj/kg	
% feed to mp column:	0 %	steam use :	.00 kg/l(alc)	H feed:	377 kj/kg	(at low pressure)
LP Reflux ratio:	5	HP steam temp :	170 C	H bottoms:	418 kj/kg	
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg	H distillate:	179 kj/kg	
% alcohol in feed:	.0811968102 %	LP steam temp :	120 C			
% alcohol in Dist.:	.9 %	LP steam Hfg :	2203 kj/kg			
		steam use:	134.77 kg/tc			

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	0	22030	119.9	120	.00
Effect 2	112.8	22300	109.9	110	-251544.00
Effect 3	81	22570	99.9	100	-182817.00
Effect 4	70.6	22830	89.9	90	-161179.80
Effect 5	60.6	23090	79.9	80	-139925.40
Effect 6	0	23370	67.9	68	.00
Effect 1 vapors	0	22030.00	120.1	120	.00
Effect 2 vapors	112.8	22300.00	110.1	110	251544.00
Effect 3 vapors	81	22570	100.1	100	182817.00
Effect 4 vapors	70.6	22830	90.1	90	161179.80
Effect 5 vapors	60.6	23090	80.1	80	161179.80
Effect 6 vapors	0	23370	68.1	68	.00
combined evap cond	409.2318481	4.18	108.109429844	35	125060.20
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	.00	4.38	35	150	.00
Reboiler HP	.00	20860	157.9	158	.00
condenser HP	.00	8300	132.1	132	.00
bottoms HP	.00	4.19	158	35	.00
distillate HP	.00	3.1	132	35	.00
Feed MP	.00	4.38	35	135	.00
Reboiler MP	.00	20860	144.9	145	.00
condenser MP	.00	8300	110.1	110	.00
bottoms MP	.00	4.19	158	35	.00
distillate MP	.00	3.1	132	35	.00
Feed LP	.00	4.38	35	90	.00
Reboiler LP	.00	22570	99.9	100	.00
condenser LP	.00	9060	78.1	78	.00
bottoms LP	.00	4.19	100	35	.00
distillate LP	.00	3.1	78	35	.00
HP steam condensate	.00	4.19	180	35	.00
LP steam condensate	144.8318481	4.19	120	35	51581.86

PROBLEM TABLE: 2-3-4-5

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	296905	144.83 kg steam
169	1	0	0	0	296905	134.77
168.9	.1	0	0	0	296905	
160	8.9	0	0	0	296905	
158	2	0	0	0	296905	
155	3	0	0	0	296905	
154.9	.1	0	0	0	296905	
145	9.9	0	0	0	296905	
132.1	12.9	0	0	0	296905	
132	.1	0	0	0	296905	
130	2	0	0	0	296905	
129.9	.1	0	0	0	296905	
120.1	9.8	0	0	0	296905	
120	.1	0	0	0	296905	
119.9	.1	-2519284	-251928	-251928	44977	
112	7.9	-3844	-30369	-282297	14608	
110.1	1.9	-7688	-14608	-296905	0	
110	.1	2507752	250775	-46130	250775	
109.9	.1	-1835858	-183586	-229716	67189	
105	4.9	-7688	-37673	-267389	29516	
100.1	4.9	-5978	-29291	-296680	225	
100	.1	1822192	182219	-114461	182445	
99.9	.1	-1613932	-161393	-275854	21051	
90.1	9.8	-2134	-20909	-296763	142	
90	.1	1609664	160966	-135797	161109	
89.9	.1	-1401388	-140139	-275935	20970	
80.1	9.8	-2134	-20909	-296845	61	
80	.1	1397120	139712	-157132	139773	
78.1	11.9	-2134	-25390	-161186	135719	
78	.1	-2134	-213	-161400	135506	
77.9	.1	-2134	-213	-161613	135292	
68.1	9.8	-2134	-20909	-182522	114383	
68	.1	-2134	-213	-182736	114170	
45	23	351	8076	-174660	122245	
40	5	351	1756	-172904	124001	
35	5	4195	20976	-151928	144977	

2-3-4-5
Grand Composite Cur
Enthalpy flow, kJ/t
Temperature, C

```
AUTONOMOUS DISTILLERY:  MP COLUMN-->LP COLUMN JUICE HEATED BUT NOT CONCENTRATED       feb 12 1990
NOTE: CLEAR JUICE FEED HEATED TO 100C:  BE SURE TO CHECK THIS IF ADJUSTING CONFIGURATION!
Juice treatment & fermentation parameters

     raw juice flow:          1000 kg/TC      Raw juice brx:          13.5 Brix        Raw juice T:          30 C
 TFS per tonne cane:           144 kg/TC    Brx into fermntr:          13.5 Brix      T into fermntr:          30 C
    Recovery of TFS:           .84 %           # evap.stages:           2          T out of fermntr:          35 C
 % Guy-Lassac yield:           .91 %              T evap. 1:          120 C
      Alcohol yield:         70.45 l/TC           T evap. 2:          110 C
         kg alcohol:         54.81 kg/TC    Juice into frmnt:      1000.00 kg/TC
     alc. conc. out:           .05 %        TFS in fermtr fd:           .12 %

                                                                                         H feed:          627 kj/kg  (at high pressure)
Alcohol distillation parameters                                                       H bottoms:          671 kj/kg
                                                                                    H distillate:         439 kj/kg
 HP column pressure:             6 atm            HP temp :          160 C                H feed:         570 kj/kg  (at medium pressure)
 MP column pressure:           4.2 atm            MP temp :          146              H bottoms:          619 kj/kg
 LP column pressure:             1 atm            LP temp :          100 C          H distillate:         356 kj/kg
% feed to hp column:             0 %             % losses :          .05 %                H feed:         377 kj/kg  (at low pressure)
% feed to mp column:         .7163 %            steam use :         1.67 kg/l(alc)    H bottoms:          418 kj/kg
   LP Reflux ratio:              5          HP steam temp :          170 C          H distillate:         179 kj/kg
         feed flow :          1000 kg        HP steam Hfg :         2050 kj/kg
 % alcohol in feed: .0548078469 %           LP steam temp :          130 C
 % alcohol in Dist.:           .95 %         LP steam Hfg :         2100 kj/kg
                                                            113.3256690

                                   dH(fg) or
                        Mass flow        Cp                  Tin         Tout          M*Cp
STREAM                         kg   kj/kg(C)                   C            C            kj
-----------------------------------------------------------------------------------------------
Raw Juice                    1000       3.84                  30          102    -276780.74
Clear Juice                     0       3.84                  90          120           .00
Effect 1                        0      22030               119.9          120           .00
Effect 2                        0      22300               109.9          110           .00
Effect 3                        0      22570                99.9          100           .00
Effect 4                        0      22830                89.9           90           .00
Effect 5                        0      23090                79.9           80           .00
Effect 6                        0      23370                67.9           68           .00
Effect 1 vapors                 0   22030.00               120.1          120           .00
Effect 2 vapors                 0   22300.00               110.1          110           .00
Effect 3 vapors                 0      22570               100.1          100           .00
Effect 4 vapors                 0      22830                90.1           90           .00
Effect 5 vapors                 0      23090                80.1           80           .00
Effect 6 vapors                 0      23370                68.1           68           .00
combined evap cond              0       4.18                   0           35           .00
Fermenter feed               1000       3.84                  68           35     126857.84
Feed HP                       .00       4.38                  35          150           .00
Reboiler HP                   .00      20860               157.9          158           .00
condenser HP                  .00       8300               132.1          132           .00
bottoms HP                    .00       4.19                 158           35           .00
distillate HP                 .00        3.1                 132           35           .00
Feed MP                    716.30       4.38                  35          135    -313739.40
Reboiler MP                111.02      21300               144.9          145    -236475.99
condenser MP               247.95       8560               110.1          110     212245.80
bottoms MP                 674.97       4.19                 144           35     308267.78
distillate MP               41.33        3.1                 110           35       9608.09
Feed LP                    283.70       4.38                  35           90     -68343.33
Reboiler LP                 42.84      22570                99.9          100     -96692.84
condenser LP                98.20       9060                78.1           78      88972.94
bottoms LP                 267.33       4.19                 100           35      72808.05
distillate LP               16.37        3.1                  78           35       2181.77
HP steam condensate           .00       4.19                 180           35           .00
LP steam condensate   116.0897092       4.19                 130           35      46209.51
```

```
                                                                                   corrected
                          cor'ctd       delta      Cp(hot)-       Delta        heat        heat
                             temp        temp      Cp(cold)           H     cascade     cascade
------------------------------------------------------------------------------------------------------------------
                              170                                               .00      237984     116.09 kg steam
                              169           1             0           0           0      237984     113.33
                            168.9          .1             0           0           0      237984
                              160         8.9             0           0           0      237984
                              158           2             0           0           0      237984
                              155           3             0           0           0      237984
                            154.9          .1      -2364760     -236476     -236476        1508
                              145         9.9             0           0     -236476        1508
                            132.1        12.9           177        2285     -234191        3793
                              132          .1           177          18     -234173        3811
                              130           2           177         354     -233818        4165
                            129.9          .1           177          18     -233801        4183
                            120.1         9.8           177        1736     -232065        5919
                              120          .1           177          18     -232047        5937
                            119.9          .1           177          18     -232029        5955
                              112         7.9           177        1400     -230629        7354
                            110.1         1.9         -3667       -6967     -237597         387
                              110          .1       2118791      211879      -25718      212266
                            109.9          .1       -970467      -97047     -122764      115219
                              105         4.9         -3539      -17341     -140105       97879
                            100.1         4.9         -3539      -17341     -157446       80538
                              100          .1         -3539        -354     -157800       80184
                             99.9          .1         -3661        -366     -158166       79818
                             90.1         9.8         -3661      -35882     -194047       43937
                               90          .1         -3661        -366     -194413       43570
                             89.9          .1         -3661        -366     -194780       43204
                             80.1         9.8         -3661      -35882     -230661        7323
                               80          .1         -3661        -366     -231027        6957
                             78.1         1.9         -3661       -6957     -237984           0
                               78          .1        886068       88607     -149377       88607
                             77.9          .1         -3611        -361     -149738       88246
                             68.1         9.8         -3611      -35384     -185123       52861
                               68          .1         -3611        -361     -185484       52500
                               45          23           234        5371     -180112       57872
                               40           5          4614       23068     -157045       80939
                               35           5          8458       42289     -114756      123228
```

AUTONOMOUS DISTILLERY: MP COLUMN-->EFF3--.EFF4--.EFF5 feb 12 1990
do not use this spreadsheet for any other configuration!!!
Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

				H feed:	627 kj/kg	(at high pressure)
				H bottoms:	671 kj/kg	
HP column pressure:	6 atm	HP temp :	160 C	H distillate:	439 kj/kg	
MP column pressure:	4.2 atm	MP temp :	146	H feed:	570 kj/kg	(at medium pressure)
LP column pressure:	1 atm	LP temp :	100 C	H bottoms:	619 kj/kg	
% feed to hp column:	0 %	% losses :	.05 %	H distillate:	356 kj/kg	
% feed to mp column:	.975 %	steam use :	2.18 kg/l(alc)	H feed:	377 kj/kg	(at low pressure)
LP Reflux ratio:	5	HP steam temp :	170 C	H bottoms:	418 kj/kg	
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg	H distillate:	179 kj/kg	
% alcohol in feed:	.0811968102 %	LP steam temp :	156 C			
% alcohol in Dist.:	.9 %	LP steam Hfg :	2095 kj/kg			
			156.9349533			

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	0	22030	119.9	120	.00
Effect 2	0	22300	109.9	110	.00
Effect 3	99.5	22570	99.9	100	-224571.50
Effect 4	86.9	22830	89.9	90	-198392.70
Effect 5	74.7	23090	79.9	80	-198392.70
Effect 6	63.9	23370	67.9	68	-149334.30
Effect 1 vapors	0	22030.00	120.1	120	.00
Effect 2 vapors	0	22300.00	110.1	110	.00
Effect 3 vapors	99.5	22570	100.1	100	224571.50
Effect 4 vapors	86.9	22830	90.1	90	198392.70
Effect 5 vapors	74.7	23090	80.1	80	198392.70
Effect 6 vapors	63.9	23370	68.1	68	149334.30
combined evap cond	186.4	4.18	95.3379828326	35	47012.46
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	.00	4.38	35	150	.00
Reboiler HP	.00	20860	157.9	158	.00
condenser HP	.00	8300	132.1	132	.00
bottoms HP	.00	4.19	158	35	.00
distillate HP	.00	3.1	132	35	.00
Feed MP	658.13	4.38	35	135	-288258.75
Reboiler MP	150.98	21300	144.9	145	-321583.32
condenser MP	356.25	8560	110.1	110	304950.86
bottoms MP	598.75	4.19	158	35	308577.70
distillate MP	59.38	3.1	132	35	17854.11
Feed LP	16.88	4.38	35	90	-4065.19
Reboiler LP	3.81	22570	99.9	100	-8604.00
condenser LP	9.13	9060	78.1	78	8275.98
bottoms LP	15.35	4.19	100	35	4181.27
distillate LP	1.52	3.1	78	35	202.94
HP steam condensate	.00	4.19	180	35	.00
LP steam condensate	160.3798669	4.19	130	35	63839.21

cor'ctd temp	delta temp	Cp(hot)- Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	328779	160.38 kg steam
169	1	0	0	0	328779	156.93
168.9	.1	0	0	0	328779	
160	8.9	0	0	0	328779	
158	2	0	0	0	328779	
155	3	0	0	0	328779	
154.9	.1	-3215833	-321583	-321583	7195	
145	9.9	0	0	-321583	7195	
132.1	12.9	-374	-4822	-326406	2373	
132	.1	-374	-37	-326443	2336	
130	2	-374	-748	-327191	1588	
129.9	.1	298	30	-327161	1618	
120.1	9.8	298	2922	-324239	4540	
120	.1	298	30	-324209	4570	
119.9	.1	298	30	-324179	4600	
112	7.9	298	2356	-321824	6955	
110.1	1.9	-3546	-6737	-328561	218	
110	.1	3045963	304596	-23965	304814	
109.9	.1	-2338961	-233896	-257861	70918	
105	4.9	-7206	-35310	-293171	35608	
100.1	4.9	-7206	-35310	-328481	298	
100	.1	2238509	223851	-104630	224149	
99.9	.1	-1987299	-198730	-303360	25419	
95	4.9	-2592.38131472	-12703	-316063	12716	
90.1	4.9	-2592	-12703	-328765	13	
90	.1	1981335	198133	-130632	198147	
89.9	.1	-1727415	-172742	-303373	25405	
80.1	9.8	-2592	-25405	-328779	0	
80	.1	1722231	172223	-156556	172223	
78.1	11.9	-2592	-30849	-161481	167298	
78	.1	80167	8017	-153464	175314	
77.9	.1	-1495931	-149593	-303058	25721	
68.1	9.8	-2588	-25359	-328417	362	
68	.1	1490755	149076	-179341	149438	
45	23	-103	-2368	-181709	147070	
40	5	2854	14268	-167441	161337	
35	5	6698	33489	-133953	194826	

AUTONOMOUS DISTILLERY: EFF2-->LP COLUMN-->EFF6

feb 9 1990

do not use for any other configuration!!!!

Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

HP column pressure:	6 atm	HP temp :	160 C
MP column pressure:	4.2 atm	MP temp :	146
LP column pressure:	1 atm	LP temp :	100 C
% feed to hp column:	0 %	% losses :	.05 %
% feed to mp column:	0 %	steam use :	3.20 kg/l(alc)
LP Reflux ratio:	5	HP steam temp :	170 C
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg
% alcohol in feed:	.0811968102 %	LP steam temp :	120 C
% alcohol in Dist.:	.9 %	LP steam Hfg :	2203 kj/kg

H feed:	627 kj/kg	(at high pressure)
H bottoms:	671 kj/kg	
H distillate:	439 kj/kg	
H feed:	627 kj/kg	(at medium pressure)
H bottoms:	671 kj/kg	
H distillate:	439 kj/kg	
H feed:	377 kj/kg	(at low pressure)
H bottoms:	418 kj/kg	
H distillate:	179 kj/kg	

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	0	22030	119.9	120	.00
Effect 2	204.3	22300	109.9	110	-455589.00
Effect 3	0	22570	99.9	100	.00
Effect 4	0	22830	89.9	90	.00
Effect 5	0	23090	79.9	80	.00
Effect 6	120.7	23370	67.9	68	-282075.90
Effect 1 vapors	0	22030.00	120.1	120	.00
Effect 2 vapors	204.3	22300.00	110.1	110	455589.00
Effect 3 vapors	0	22570	100.1	100	.00
Effect 4 vapors	0	22830	90.1	90	.00
Effect 5 vapors	0	23090	80.1	80	.00
Effect 6 vapors	120.7	23370	68.1	68	282075.90
combined evap cond	204.3	4.18	110	35	64048.05
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	.00	4.38	35	150	.00
Reboiler HP	.00	20860	157.9	158	.00
condenser HP	.00	8300	132.1	132	.00
bottoms HP	.00	4.19	158	35	.00
distillate HP	.00	3.1	132	35	.00
Feed MP	.00	4.38	35	135	.00
Reboiler MP	.00	20860	144.9	145	.00
condenser MP	.00	8300	110.1	110	.00
bottoms MP	.00	4.19	158	35	.00
distillate MP	.00	3.1	132	35	.00
Feed LP	675.00	4.38	35	90	-162607.50
Reboiler LP	152.49	22570	99.9	100	-344159.87
condenser LP	365.39	9060	78.1	78	331039.40
bottoms LP	614.10	4.19	100	35	167250.79
distillate LP	60.90	3.1	78	35	8117.65
HP steam condensate	.00	4.19	180	35	.00
LP steam condensate	234.8512002	4.19	130	35	93482.52

PROBLEM TABLE: 2-LP-6

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	481445	234.85 kg steam
169	1	0	0	0	481445	218.54
168.9	.1	0	0	0	481445	
160	8.9	0	0	0	481445	
158	2	0	0	0	481445	
155	3	0	0	0	481445	
154.9	.1	0	0	0	481445	
145	9.9	0	0	0	481445	
132.1	12.9	0	0	0	481445	
132	.1	0	0	0	481445	
130	2	0	0	0	481445	
129.9	.1	984	98	98	481543	
120.1	9.8	984	9643	9742	491187	
120	.1	984	98	9840	491285	
119.9	.1	-4558750	-455875	-446035	35410	
112	7.9	-2860	-22595	-468630	12815	
110.1	1.9	-6704	-12738	-481368	77	
110	.1	4549186	454919	-26450	454995	
109.9	.1	-3447449	-344745	-371194	110250	
105	4.9	-5850	-28667	-399861	81584	
100.1	4.9	-5850	-28667	-428528	52917	
100	.1	-5850	-585	-429113	52332	
99.9	.1	-2390	-239	-429352	52093	
90.1	9.8	-2390	-23418	-452770	28675	
90	.1	-2390	-239	-453009	28436	
89.9	.1	-2390	-239	-453248	28197	
80.1	9.8	-2390	-23418	-476666	4779	
80	.1	-2390	-239	-476905	4540	
78.1	11.9	-2390	-28436	-481445	0	
78	.1	3308004	330800	-150645	330800	
77.9	.1	-2822960	-282296	-432941	48504	
68.1	9.8	-2201	-21568	-454508	26937	
68	.1	2818558	281856	-172653	308792	
45	23	284	6530	-166123	315322	
40	5	3240	16202	-149921	331524	
35	5	7085	35423	-114498	366947	

AUTONOMOUS DISTILLERY: EFF1-->EFF2-->LP COLUMN

feb 13 1990

Do not use this spreadsheet for other conditions!!

Juice treatment & fermentation parameters

raw juice flow:	1000 kg/TC	Raw juice brx:	13.5 Brix	Raw juice T:	30 C
TFS per tonne cane:	144 kg/TC	Brx into fermntr:	20 Brix	T into fermntr:	30 C
Recovery of TFS:	.84 %	# evap.stages:	2	T out of fermntr:	35 C
% Guy-Lassac yield:	.91 %	T evap. 1:	120 C		
Alcohol yield:	70.45 l/TC	T evap. 2:	110 C		
kg alcohol:	54.81 kg/TC	Juice into frmnt:	675.00 kg/TC		
alc. conc. out:	.08 %	TFS in fermtr fd:	.18 %		

Alcohol distillation parameters

				H feed:	627 kj/kg	(at high pressure)
HP column pressure:	6 atm	HP temp :	160 C	H bottoms:	671 kj/kg	
MP column pressure:	4.2 atm	MP temp :	146	H distillate:	439 kj/kg	
LP column pressure:	1 atm	LP temp :	100 C	H feed:	570 kj/kg	(at medium pressure)
% feed to hp column:	0 %	% losses :	.05 %	H bottoms:	619 kj/kg	
% feed to mp column:	0 %	steam use :	3.28 kg/l(alc)	H distillate:	356 kj/kg	
LP Reflux ratio:	5	HP steam temp :	170 C	H feed:	377 kj/kg	(at low pressure)
feed flow :	675 kg	HP steam Hfg :	2050 kj/kg	H bottoms:	418 kj/kg	
% alcohol in feed:	.0811968102 %	LP steam temp :	130 C	H distillate:	179 kj/kg	
% alcohol in Dist.:	.9 %	LP steam Hfg :	2100 kj/kg			

STREAM	Mass flow kg	dH(fg) or Cp kj/kg(C)	Tin C	Tout C	M*Cp kj
Raw Juice	1000	3.84	30	102	-276780.74
Clear Juice	1000	3.84	90	120	-115325.31
Effect 1	211.5	22030	119.9	120	-465934.50
Effect 2	113.5	22300	109.9	110	-253105.00
Effect 3	0	22570	99.9	100	.00
Effect 4	0	22830	89.9	90	.00
Effect 5	0	23090	79.9	80	.00
Effect 6	0	23370	67.9	68	.00
Effect 1 vapors	211.5	22030.00	120.1	120	465934.50
Effect 2 vapors	113.5	22300.00	110.1	110	253105.00
Effect 3 vapors	0	22570	100.1	100	.00
Effect 4 vapors	0	22830	90.1	90	.00
Effect 5 vapors	0	23090	80.1	80	.00
Effect 6 vapors	0	23370	68.1	68	.00
combined evap cond	325	4.18	116.507692308	35	110728.20
Fermenter feed	675	3.68	68	35	81995.17
Feed HP	.00	4.38	35	150	.00
Reboiler HP	.00	20860	157.9	158	.00
condenser HP	.00	8300	132.1	132	.00
bottoms HP	.00	4.19	158	35	.00
distillate HP	.00	3.1	132	35	.00
Feed MP	.00	4.38	35	135	.00
Reboiler MP	.00	20860	144.9	145	.00
condenser MP	.00	8300	110.1	110	.00
bottoms MP	.00	4.19	158	35	.00
distillate MP	.00	3.1	132	35	.00
Feed LP	675.00	4.38	35	90	-162607.50
Reboiler LP	152.49	22570	99.9	100	-344159.87
condenser LP	365.39	9060	78.1	78	331039.40
bottoms LP	614.10	4.19	100	35	167250.79
distillate LP	60.90	3.1	78	35	8117.65
HP steam condensate	.00	4.19	180	35	.00
LP steam condensate	241.1121200	4.19	130	35	95974.68

PROBLEM TABLE: 1-2-LP

cor'ctd temp	delta temp	Cp(hot)-Cp(cold)	Delta H	heat cascade	corrected heat cascade	
170				.00	494280	241.11 kg steam
169	1	0	0	0	494280	235.37
168.9	.1	0	0	0	494280	
160	8.9	0	0	0	494280	
158	2	0	0	0	494280	
155	3	0	0	0	494280	
154.9	.1	0	0	0	494280	
145	9.9	0	0	0	494280	
132.1	12.9	0	0	0	494280	
132	.1	0	0	0	494280	
130	2	0	0	0	494280	
129.9	.1	-4662179	-466218	-466218	28062	
120.1	9.8	-2834	-27772	-493990	290	
120	.1	4656511	465651	-28339	465941	
119.9	.1	-2533884	-253388	-281728	212552	
116	3.9	-2833.91721708	-11052	-292780	201500	
112	4	-1475	-5902	-298682	195598	
110.1	1.9	-5320	-10107	-308789	185491	
110	.1	2525730	252573	-56216	438064	
109.9	.1	-3446918	-344692	-400908	93372	
105	4.9	-5320	-26066	-426974	67306	
100.1	4.9	-5320	-26066	-453040	41240	
100	.1	-5320	-532	-453572	40708	
99.9	.1	-1859	-186	-453757	40522	
90.1	9.8	-1859	-18217	-471974	22306	
90	.1	-1859	-186	-472160	22120	
89.9	.1	-1859	-186	-472346	21934	
80.1	9.8	-1859	-18217	-490562	3718	
80	.1	-1859	-186	-490748	3532	
78.1	11.9	-1859	-22120	-494280	0	
78	.1	3308535	330854	-163426	330854	
77.9	.1	-1670	-167	-163593	330687	
68.1	9.8	-1670	-16366	-179960	314320	
68	.1	-1670	-167	-180127	314153	
45	23	815	18737	-161390	332890	
40	5	3771	18856	-142534	351746	
35	5	7615	38077	-104457	389823	

QUAD EFFECT NO CONDENSATES USED tonnes cane per hour
RISING FILM EVAPOARTORS 12-13-89 175
VAC PANS AT 140 KG/TC TOTAL STEAM USE:
352.9575296 KG

TABLE 1: STREAMS AND DATA

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	Enthalpy flow, Watts	area, sq m		
CANE JUICE HEAT	13.6	157000	3.8416672	34	102	41013639	11392678		\	
CANE JUICE HEAT	13.6	157000	3.8416672	90	110	12062835	3350788			
EFFECT 1	21.4	57224	22300	110	110.1	127610187	35447274	1155	|	
EFFECT 2	29.4	27150	22570	100	100.1	61277213	17021448	655	|	COOL STREAMS
EFFECT 3	42.5	22386	23010	83	83.1	51509842	14308289	526	|	
EFFECT 4	62	15801	23590	60	60.1	37275244	10354234	563	|	
VACUUM PAN		MCp = 542675000		85	85.1	54267500	15074306		/	
EFFECT 1 VAPORS		57224	22300	110	109.9	-127610187			\	
EFFECT 2 VAPORS		27150	22570	100	99.9	-61277213			|	
EFFECT 3 VAPORS		22386	23010	83	82.9	-51509842			|	HOT STREAMS
EFFECT 4 VAPORS		15801	23590	60	59.9	-37275244			|	

TABLE 2: PROBLEM TABLE

DELTA T MIN

INTERVAL	COR'CTD TEMP	DELTA TEMP	Cp(COLD) -Cp(HOT)	DELTA H	HEAT CASCADE	CORRECTED HEAT CASCADE		
	110.1			0	0	135270973	352.96	310.32
	110	.1	-1276101869	-127610187	-127610187	7660786		
	102	8	-603142	-4825134	-132435321	2835652		
	100.1	1.9	-1206284	-2291939	-134727260	543714		
	100	.1	1274895586	127489559	-7237701	128033272		
	99.9	.1	-613978411	-61397841	-68635542	66635431		
	99.1	.8	-1206284	-965027	-69600569	65670404		
	99	.1	-543881284	-54388128	-123988697	11282276		
	90.1	8.9	-1206284	-10735923	-134724620	546353		
	90	.1	611565844	61156584	-73568036	61702937		
	89.9	.1	-515701558	-51570156	-125138192	10132781		
	73.1	16.8	-603142	-10132781	-135270973	0		
	73	.1	512519858	51251986	-84018987	51251986		
	72.9	.1	-373355580	-37335558	-121354545	13916428		
	50	22.9	-603142	-13811946	-135166492	104482		
	49.9	.1	372149297	37214930	-97951562	37319411		
	34	15.9	-603142	-9589954	-107541516	27729458		

QUAD EFFECT CONDENSATES USED 12-12-89 tonnes cane per hour 175
RISING FILM EVAPOARTORS
VAC PANS AT 140 KG/TC TOTAL STEAM USE: 317.0883233 KG

TABLE 1: STREAMS AND DATA

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	Enthalpy flow, Watts	area, sq m	
CANE JUICE HEAT	13.6	157000	3.84	34	102	41013639	11392678		
CANE JUICE HEAT	13.6	157000	3.84	90	110	12062835	3350788		
EFFECT 1	20.4	52333	22300	110	110.1	116703333	32417593	1056	
EFFECT 2	26.5	24093	22570	100	100.1	54378086	15105024	581	COOL STREAMS
EFFECT 3	37.3	23330	23010	83	83.1	53681449	14911514	548	
EFFECT 4	62	22805	23590	60	60.1	53797604	14943779	812	
VACUUM PAN		MCp = 542675000		85	85.1	54267500	15074306		
EFFECT 1 VAPORS		52333	22300	110	109.9	-116703333			
EFFECT 2 VAPORS		24093	22570	100	99.9	-54378086			
EFFECT 3 VAPORS		23330	23010	83	82.9	-53681449			HOT STREAMS
EFFECT 4 VAPORS		22805	23590	60	59.9	-53797604			
COND. FROM STEAM		55490	4.18	120	40	-18556009			
COND. FROM EFF 1 VAPORS		52333	4.18	109.9	40	-15290858			
COND. FROM EFF 2 VAPORS		24093	4.18	99.9	40	-6032474			
COND. FROM EFF 3 VAPORS		23330	4.18	82.9	40	-4183514			
COND. FROM EFF 4 VAPORS		22805	4.18	59.9	40	-1896987			

TABLE 2: PROBLEM TABLE

DELTA T MIN

INTERVAL	COR'CTD TEMP	DELTA TEMP	Cp (COLD) -Cp (HOT)	DELTA H	HEAT CASCADE	CORRECTED HEAT CASCADE		
	110.1			0	0	121524100	317.09	278.79
	110	.1	-1167033333	-116703333	-116703333	4820767		
	102	8	-371192	-2969533	-119672866	1851233		
	100.1	1.9	-974333	-1851233	-121524100	0		
	100	.1	622278145	62227814	-59296285	62227814		
	99.1	.9	-755580	-680022	-59976308	61547792		
	99	.1	-543430580	-54343058	-114319366	7204734		
	90	9	-755580	-6800221	-121119586	404514		
	89.9	.1	543628417	54362842	-66756744	54767356		
	89.1	.8	-51729	-41383	-66798128	54725972		
	89	.1	-536866218	-53686622	-120484750	1039350		
	73	16	-51729	-827668	-121312417	211683		
	72.9	.1	536762760	53676276	-67636141	53887959		
	72.1	.8	45789	36631	-67599510	53924590		
	72	.1	-537930251	-53793025	-121392535	131564		
	50	22	45789	1007349	-120385187	1138913		
	49.9	.1	538021828	53802183	-66583004	54941096		
	34	15.9	141115	2243721	-64339283	57184817		
	30	4	744256	2977025	-61362257	60161842		

QUAD EFFECT CONDENSATES USED 12-12-89
FALLING FILM EVAPORATORS
VAC PANS AT 140 KG/TC TOTAL STEAM USE: 279.4206071 KG steam temp: 135

TABLE 1: STREAMS AND DATA

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	HEAT FLOW WATTS	EVAPORATOR AREA, SQ M	
RAW CANE JUICE	13.6	157000	3.84	34	102	41013639			
CLEAR JUICE	13.6	157000	3.84	90	125	21109961			
EFFECT 1	19.15	45501	21900	124.9	125	99647859	27679961	865	
EFFECT 2	31.3	43281	22160	114.9	115	95911676	26642132	987	COOL STREAMS
EFFECT 3	41.9	17258	22710	96.9	97	39192521	10886811	252	
EFFECT 4	62.5	16796	23210	74.9	75	38984043	10828901	410	
VACUUM PAN		MCp = 542675000		84.9	85	54267500			
EFFECT 1 VAPORS		45501	21900	125.1	125	-99647859			
EFFECT 2 VAPORS		43281	22160	115.1	115	-95911676			
EFFECT 3 VAPORS		17258	22710	97.1	97	-39192521			HOT STREAMS
EFFECT 4 VAPORS		16796	0	75.1	75	0		23210	
CONDENSATES (STEAM THRU E3)		154939	4.18	122.24	40	-53264832			
COND. FROM STEAM		48899	4.18	135	40				
COND. FROM EFF 1 VAPORS		45501	4.18	125	40				
COND. FROM EFF 2 VAPORS		43281	4.18	115	40				
COND. FROM EFF 3 VAPORS		17258	4.18	97	40				
COND. FROM EFF 4 VAPORS		16796	0	75	75				

TABLE 2: PROBLEM TABLE

DELTA T MIN

INTERVAL	COR'CTD TEMP	DELTA TEMP	Cp (COLD) -Cp (HOT)	DELTA H	HEAT CASCADE	CORRECTED HEAT CASCADE	
	125			0	0	105620989	279.42
	124.9	.1	-996478590	-99647859	-99647859	5973130	
	115.1	9.8	-603142	-5910789	-105558648	62341	
	115	.1	995875448	99587545	-5971103	99649886	
	114.9	.1	-959719899	-95971990	-101943093	3677896	
	112	2.9	-603142	-1749111	-103692204	1928785	
	105.1	9.8	44504	436139	-101506954	4114036	
	105	.1	959161261	95916126	-5590828	100030162	
	104.9	.1	-391880706	-39188071	-44778898	60842091	
	102	2.9	44504	129062	-44649837	60971153	
	100.1	1.9	-558638	-1061412	-45711248	59909741	
	100	.1	-543233638	-54323364	-100034612	5586377	
	90	10	-558638	-5586377	-105620989	0	
	87.1	2.9	44504	129062	-105491928	129062	
	87	.1	391969714	39196971	-66294956	39326033	
	86.9	.1	-389795929	-38979593	-105274549	346440	
	65.1	21.8	44504	970188	-104304362	1316628	
	65	.1	44504	4450	-104299911	1321078	
	34	31	44504	1379625	-102920287	2700703	
	30	4	647646	2590583	-100329704	5291286	

```
QUAD EFFECT             CONDENSATES COMBINED INTO SINGLE STREAM/ 4ETH EFFECT HEAT NOT USED        1-6-90
FALLING FILM EVAPORATORS
VAC PANS AT 100 KG/TC                     TOTAL STEAM USE:
                                           260.5720728 KG          83829132.4             steam temp: 135

TABLE 1:  STREAMS AND DATA                                                       HEAT FLOW     EVAPORATOR
                                           KJ/KG                                 WATTS         AREA, SQ M
STREAM           BRIX         MASS FLOW    Cp          Tin        Tout      MCp(dT)
--------------------------------------------------------------------------------------------------------
RAW CANE JUICE           13.6      157000        3.84         34       102  41013639                           \
CLEAR JUICE              13.6      157000        3.84         90       125  21109961                           |
EFFECT 1                18.60       42200       21900      124.9       125  92418000   25671667         802    |
EFFECT 2                28.97       41100       22160      114.9       115  91077600   25299333         937    |      COOL STREAMS
EFFECT 3                39.65       19844       22710       96.9        97  45066632   12518509         290    |
EFFECT 4                   62       19417       23210       74.9        75  45066602   12518501         474    |
VACUUM PAN                         MCp = 387625000          84.9        85  38762500                           /
EFFECT 1 VAPORS                     42200       21900      125.1       125 -92418000                           \
EFFECT 2 VAPORS                     41100       22160      115.1       115 -91077600                           |
EFFECT 3 VAPORS                     19844       22710       97.1        97 -45066632                           |      HOT STREAMS
EFFECT 4 VAPORS                     19417        $.00       75.1        75         0                           |
SUMMED CONDENSATE                  148745        4.18     121.57        40 -50714442
COND. FROM STEAM                    45600        4.18        135        40 -18107805                           |
COND. FROM EFF 1 VAPORS             42200        4.18        125        40 -14993660                           |
COND. FROM EFF 2 VAPORS             41100        4.18        115        40 -12884850                           |
COND. FROM EFF 3 VAPORS             19844        4.18         97        40  -4728127                           |
COND. FROM EFF 4 VAPORS             19417        4.18         75        40  -2840691                           /
```

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TABLE 2:  PROBLEM TABLE

DELTA T MIN
                                                                          CORRECTED
                   COR'CTD      DELTA     Cp(COLD)       DELTA        HEAT        HEAT
    INTERVAL        TEMP        TEMP      -Cp(HOT)         H        CASCADE     CASCADE
-------------------------------------------------------------------------------------------------------------------
                    125                                      0            0     98496244          260.57
                  124.9           .1    -924180000   -92418000    -92418000      6078244
                  115.1          9.8       -603142    -5910789    -98328789       167454
                    115           .1     923576858    92357686     -5971103     92525140
                  114.9           .1    -911379142   -91137914    -97109018      1387226
                  112.6          2.3       -603142    -1387226    -98496244            0
                  105.1          9.8         18610      182381    -98313862       182381
                    105           .1     910794610    91079461     -7234401     91261842
                  104.9           .1    -450647714   -45064771    -52299173     46197071
                    102          2.9         18610       53970    -52245203     46251041
                  100.1          1.9       -584531    -1110610    -53355813     45140431
                    100           .1    -388209531   -38820953    -92176766      6319478
                     90           10       -584531    -5845314    -98022080       474163
                   87.1          2.9         18610       53970    -97968110       528133
                     87           .1     450684934    45068493    -52899617     45596627
                   86.9           .1    -450647414   -45064741    -97964358       531885
                   65.1         21.8         18610      405705    -97558653       937590
                     65           .1         18610        1861    -97556792       939451
                     34           31         18610      576920    -96979873      1516371
                     30            4        621752     2487008    -94492864      4003379
```

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QUAD EFFECT             CONDENSATES USED                                              12-12-89
FALLING FILM EVAPORATORS
VAC PANS AT 100 KG/TC                     TOTAL STEAM USE:
                                           259.1877011 KG
TABLE 1:  STREAMS AND DATA                                                                          135
                                           KJ/KG                                 HEAT FLOW     EVAPORATOR
STREAM           BRIX         MASS FLOW    Cp          Tin        Tout      MCp(dT)  WATTS         AREA, SQ M
--------------------------------------------------------------------------------------------------------
RAW CANE JUICE           13.6      157000        3.84         34       102  41013639   11392678                \
CLEAR JUICE              13.6      157000        3.84         90       125  21109961    5863878                |
EFFECT 1                 18.7       42818       21900      124.9       125  93771818   26047727         814    |
EFFECT 2                 29.3       41308       22160      114.9       115  91538745   25427429         917    |      COOL STREAMS
EFFECT 3                   40       19494       22710       96.9        97  44270238   12297288         299    |
EFFECT 4                   62       18941       23210       74.9        75  43962735   12211871         465    |
VACUUM PAN                         MCp = 387625000          84.9        85  38762500   10767361                /
EFFECT 1 VAPORS                     42818       21900      125.1       125 -93771818                           \
EFFECT 2 VAPORS                     41308       22160      115.1       115 -91538745                           |
EFFECT 3 VAPORS                     19494       22710       97.1        97 -44270238                           |
EFFECT 4 VAPORS                     18941       23210       75.1        75 -43962735                           |      HOT STREAMS
COND. FROM STEAM                    45358        4.18        135        40 -18011601                           |
COND. FROM EFF 1 VAPORS             42818        4.18        125        40 -15213300                           |
COND. FROM EFF 2 VAPORS             41308        4.18        115        40 -12950089                           |
COND. FROM EFF 3 VAPORS             19494        4.18         97        40  -4644574                           |
COND. FROM EFF 4 VAPORS             18941        4.18         75        40  -2771111                           /
```

```
TABLE 2:  PROBLEM TABLE

DELTA T MIN
                                                                          CORRECTED
                   COR'CTD      DELTA     Cp(COLD)       DELTA        HEAT        HEAT
    INTERVAL        TEMP        TEMP      -Cp(HOT)         H        CASCADE     CASCADE
-------------------------------------------------------------------------------------------------------------------
                    125                                      0            0     97972951          259.19
                  124.9           .1    -937718182   -93771818    -93771818      4201133
                  115.1          9.8       -413546    -4052750    -97824568       148383
                    115           .1     937304636    93730464     -4094105     93878846
                  114.9           .1    -915800999   -91580100    -95674205      2298746
                  105.1          9.8       -234566    -2298746    -97972951            0
                    105           .1     915152887    91515289     -6457662     91515289
                  104.9           .1    -442936950   -44293695    -50751357     47221594
                    102          2.9        -61898     -179504    -50930862     47042089
                  100.1          1.9       -665040    -1263576    -52194438     45778513
                    100           .1    -388290040   -38829004    -91023442      6949509
                     90           10       -665040    -6650398    -97673840       299111
                   87.1          2.9        -61898     -179504    -97853345       119607
                     87           .1     442640486    44264049    -53589296     44383655
                   86.9           .1    -439689246   -43968925    -97558221       414730
                   65.1         21.8         19586      426967    -97131253       841698
                     65           .1     439646934    43964693    -53166560     44806391
                     34           31         98760     3061568    -50104992     47867959
                     30            4        701902     2807608    -47297384     50675567
```

QUINT EFFECT CONDENSATES USED
FALLING FILM EVAPORATORS
VAC PANS AT 100 KG/TC

TOTAL STEAM USE:
214.32 KG

steam temp: 135

TABLE 1: STREAMS AND DATA

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	HEAT FLOW WATTS	EVAPORATOR AREA, SQ M	
RAW CANE JUICE	13.6	157000	3.84	35	102	40410497			
CLEAR JUICE	13.6	157000	3.84	90	130	24125670			
EFFECT 1	17.4	34287	21740	129.9	130	74540713	20705754	1294	
EFFECT 2	23.728	32726	21940	122.9	123	71801129	19944758	1055	COOL STREAMS
EFFECT 3	37.05	32356	22190	114.9	115	71798595	19944054	1039	
EFFECT 4	46.463	11675	22520	101.9	102	26292964	7303601	468	
EFFECT 5	62	11516	22830	89.9	90	26291339	7303150	507	
VACUUM PAN		MCp = 387625000		84.9	85	38762500			
EFFECT 1 VAPORS		34287	21740	130.1	130	-74540713	65055464		
EFFECT 2 VAPORS		32726	21940	123.1	123	-71801129			
EFFECT 3 VAPORS		32356	22190	115.1	115	-71798595			
EFFECT 4 VAPORS		11675	22520	102.1	102	-26292964			
EFFECT 5 VAPORS		11516	22830	90.1	90	-26291339			
COND. FROM STEAM		37507	0	140	40	0			HOT STREAMS
COND. FROM EFF 1 VAPORS		34287	0	130	40	0			
COND. FROM EFF 2 VAPORS		148552	4.18	123	40	-11354003			
COND. FROM EFF 3 VAPORS		32356	0	115	40	0			
COND. FROM EFF 4 VAPORS		11675	0	102	40	0			
COND. FROM EFF 5 VAPORS		11516	4.18	90	40	-2406872			
COMBINDED COND			4.18	125.5148765	40				
		32726		123					

TABLE 2: PROBLEM TABLE

DELTA T MIN

INTERVAL	COR'CTD TEMP	DELTA TEMP	Cp(COLD) -Cp(HOT)	DELTA H	HEAT CASCADE	CORRECTED HEAT CASCADE	
	140			0	0	80451502	214.32
	139.9	.1	-745407126	-74540713	-74540713	5910789	
	130.1	9.8	-603142	-5910789	-80451502	0	
	130	.1	744803985	74480398	-5971103	74480398	
	129.9	.1	-718614431	-71861443	-77832546	2618955	
	126	3.9	-603142	-2352253	-80184799	266703	
	123.1	2.9	17804	51632	-80133167	318335	
	123	.1	718029093	71802909	-8330257	72121244	
	122.9	.1	-717968147	-71796815	-80127072	324430	
	115.1	7.8	17804	138873	-79988199	463303	
	115	.1	718003755	71800376	-8187823	72263679	
	114.9	.1	-650536835	-65053684	-73241507	7209995	
	112	2.9	17804	51632	-73189874	7261627	
	102.1	9.9	-585337	-5794841	-78984715	1466786	
	102	.1	262344302	26234430	-52750285	27701217	
	101.9	.1	-263498725	-26349872	-79100158	1351344	
	100	1.9	-585337	-1112141	-80212299	239203	
	90.1	11.8	17804	210091	-80002208	449293	
	90	.1	262931192	26293119	-53709089	26742413	
	45	45	65942	2967378	-50741711	29709790	
	40	5	669083	3345417	-47396294	33055208	

QUINT EFFECT CONDENSATES USED
FALLING FILM EVAPORATORS
VAC PANS AT 100 KG/TC

TOTAL STEAM USE:
211.39 KG

steam temp: 135

TABLE 1: STREAMS AND DATA

STREAM	BRIX	MASS FLOW	KJ/KG Cp	Tin	Tout	MCp(dT)	HEAT FLOW WATTS	EVAPORATOR AREA, SQ M	
RAW CANE JUICE	13.6	157000	3.84	35	102	40410497			
CLEAR JUICE	13.6	157000	3.84	90	130	24125670			
EFFECT 1	17.428	34485	21740	129.9	130	74969320	20824811	1302	
EFFECT 2	23.904	33192	21940	122.9	123	72822216	20228393	1070	COOL STREAMS
EFFECT 3	37.378	32199	22190	114.9	115	71450581	19847384	1034	
EFFECT 4	46.997	11692	22520	101.9	102	26329989	7313886	469	
EFFECT 5	62	10994	22830	89.9	90	25099251	6972014	484	
VACUUM PAN		MCp = 387625000		84.9	85	38762500			
EFFECT 1 VAPORS		34485	21740	130.1	130	-74969320			
EFFECT 2 VAPORS		33192	21940	123.1	123	-72822216			
EFFECT 3 VAPORS		32199	22190	115.1	115	-71450581			
EFFECT 4 VAPORS		11692	22520	100.1	100	-26329989			
EFFECT 5 VAPORS		10994	22830	80.1	80	-25099251			
COND. FROM STEAM		36993	4.18	140	40	-15462936			HOT STREAMS
COND. FROM EFF 1 VAPORS		34485	4.18	130	40	-12973072			
COND. FROM EFF 2 VAPORS		33192	4.18	123	40	-11515469			
COND. FROM EFF 3 VAPORS		32199	4.18	115	40	-10094528			
COND. FROM EFF 4 VAPORS		11692	4.18	100	40	-2932310			
COND. FROM EFF 5 VAPORS		10994	4.18	80	40	-1838193			

TABLE 2: PROBLEM TABLE

DELTA T MIN

INTERVAL	COR'CTD TEMP	DELTA TEMP	Cp(COLD) -Cp(HOT)	DELTA H	HEAT CASCADE	CORRECTED HEAT CASCADE	
	140			0	0	79349278	211.39
	139.9	.1	-749538568	-74953857	-74953857	4395421	
	130.1	9.8	-448512	-4395421	-79349278	0	
	130	.1	749244685	74924468	-4424810	74924468	
	129.9	.1	-728526531	-72852653	-77277463	2071815	
	123.1	6.8	-304367	-2069697	-79347159	2119	
	123	.1	727917797	72791780	-6555380	72793898	
	122.9	.1	-714671439	-71467144	-78022524	1326755	
	115.1	7.8	-165627	-1291887	-79314411	34867	
	115	.1	714340186	71434019	-7880392	71468886	
	114.9	.1	-650955922	-65095592	-72975984	6373294	
	112	2.9	-31033	-89995	-73065980	6283299	
	102.1	9.9	-634175	-6278329	-79344308	4970	
	102	.1	262665714	26266571	-53077737	26271542	
	101.9	.1	-251577809	-25157781	-78235518	1113761	
	100	1.9	-585303	-1112075	-79347593	1685	
	80.1	21.8	17839	388890	-78958703	390575	
	80	.1	251010346	25101035	-53857669	25491610	
	45	35	63794	2232783	-51624885	27724393	
	40	5	666936	3334678	-48290208	31059071	

APPENDIX II:
PINCH POINT AND
MIXED-INTEGER LINEAR PROGRAMMING

A.1.1 Pinch Point Analysis Implementation: The "Problem Table" Method

Although constructing composite curves are very useful in visualizing the heat exchanges occurring in a system, it is not a particularly practical method for performing an analysis. Any changes in the system design would require the composite curves to repeatedly redrawn until the desired system is achieved. Were this implemented in some type of computer program or spreadsheet, then the tedium would be reduced, but when used alone it is still a less than optimal technique for performing a pinch analysis.

The problem table method [1] combines is a practical way of performing a pinch analysis. The problem table method first partitions the complete temperature range into temperature intervals. Any hot steam can exchange heat with any cold steam within the same or lower temperature interval. This prevents violations the second law of thermodynamics by transferring heat from cold to hot.

A temperature interval is defined by adjacent "adjusted" source (inlet) or target (outlet) temperatures of any stream in the system (The circled numbers in Chapter 3, Figure 3.1 are temperature intervals). Adjusting some or all of the temperatures is a device to account for the minimum approach temperature. If the ΔT_{MIN} is added to all of the cold streams, then a cold stream can legally exchange heat with any hot steam whose temperature is greater than or equal to its adjusted temperature. If the streams were not adjusted for the minimum approach temperature, then the hot streams would have to be matched to cold streams ΔT_{MIN} degrees below, making the temperature and heat accounting much more difficult. By adjusting the temperatures, any hot stream and cold stream in a particular temperature interval may exchange heat.

Adding the minimum approach temperature to the cold streams is only one of numerous, equally valid methods for performing the temperature adjustment. Reference 1 suggests adding $\Delta T_{MIN}/2$ to all of the cold streams and subtracting $\Delta T_{MIN}/2$ from all of the hot streams. This method is sufficient if the minimum approach temperature is constant throughout the whole system, however it is more difficult to implement if ΔT_{MIN} varies from stream to stream, as it does in a sugar factory evaporator train. In this case, it is more straightforward to add the individual minimum approach temperatures to all of the cold streams being heated.

Thus, the entire temperature range is partitioned into K temperature intervals by the adjusted source and target temperatures of the streams, generally $k = 1$ being the interval at the highest temperature. For each stream, we identify in which temperature intervals it is active:

H_k = {all hot streams present in interval k}
C_k = {all cold streams present in interval k}

In each interval, the sum of the specific enthalpies (MC_P) of all the cold streams in that interval is subtracted from the specific enthalpies of all of the hot streams in that interval. This is multiplied by the size of the temperature interval to arrive at the net heating or cooling load in that interval. If this value is positive, more heat is entering than leaving; if this value is negative, more heat is leaving than is entering. When phase changes occur, the temperature interval is always very small (assumed 0.1°C in these analyses) and the total heat available or required is dominated by the latent heat term, be it negative for evaporation or positive for condensation.

The problem table is constructed by cascading the excess heat from one temperature interval into the next lowest one, as seen generically in Table A.1. In almost all cases, the heat cascade will be less than zero in at least one interval*. This would imply that more heat was required in that interval than could be provided by the hot streams in that interval and the excess heat from the interval immediately preceding it. Heat would have to be transferred from the interval below the one with the negative heat cascade value, a clear violation of the second law of thermodynamics.

This violation can be avoided if the heat cascade is "corrected" by adding enough heat into the first temperature interval so that when the cascade reaches the interval with the net negative heat cascade, the value will be exactly zero. (This is shown as the "Corrected Heat Cascade" column in Table A.1.) This implies that no heat is transferred from this interval to the following one, and therefore defines a pinch point (no heat is transferred across the pinch). Plotting the corrected heat cascade (*x* axis) versus temperature (*y* axis) results in the "grand composite curve." The grand composite curve, discussed at length in Chapter 3.

A.1.3 The Pinch Point Method as a Design Tool

A particularly useful aspect of the problem table is its ability to serve not only for analysis but also as a design tool. It should be noted that the pinch analysis up to this point has shown the minimum utility requirements along with the

Table A.1: The Problem Table Method of Pinch Analysis.

Interval	$\Sigma MCp_i - \Sigma MCp_j$	Net Heat Flow $(\Sigma MCp_i - \Sigma MCp_j)\Delta T$	Heat Cascade	Corrected Heat Cascade
				Q_0= -MIN(Heat Cascade Entry)
1	$\sum_{i\in H1} MCp_1 - \sum_{j\in C1} MCp_1$	Q_1	$\sum_{k=1} Q_k$	$\sum_{k=0,1} Q_k$
2	$\sum_{i\in H2} MCp_2 - \sum_{j\in C2} MCp_2$	Q_2	$\sum_{k=1,2} Q_k$	$\sum_{k=0...2} Q_k$
.		.		
.	.	.	.	.
.	.	.	.	.
k=Pinch	$\sum_{i\in Hk} MCp_k - \sum_{j\in Ck} MCp_k$	Q_k	$\sum_{k=1...k} Q_k$	$\sum_{k=0...k} Q_k = 0$
.		.		
.	.	.	.	.
.	.	.	.	.
K	$\sum_{i\in HK} MCp_K - \sum_{j\in CK} MCp_K$	Q_K	$\sum_{k=1...K} Q_k$	$\sum_{k=0...K} Q_k$

---

[*] If all of the hot steams in a system are ΔT_{MIN}°C or more above all of the cold streams, then the hot composite can slide completely over the cold composite curve without encountering a pinch.

location of any pinch points -- always assuming fixed duties and flows in all of the components. This is the minimum heat requirement for *that set* of equipment, but not for the *overall process*. For example, a pinch analysis will predict the minimum energy for a given set of *n* evaporator effects with given evaporation duties, but would not specify the minimum energy required to concentrate the feed to the desired level using *n* effects at the same operating temperatures but with optimal evaporation loads. The minimum energy demand would occur for the evaporators when the loads on each effect were balanced such that a pinch occurs at the inlet of each effect. This would prevent heat at potentially useful temperatures from being rejected to outside the system.

A.2 Synthesizing Minimum-Unit Heat Exchanger Networks Using Mixed Integer Linear Programming (MILP)

Linear programming techniques have gained widespread use in chemical engineering optimization [2]. This section describes a number of the mathematical programming techniques using in the energy analysis of chemical process systems, with particular emphasis on the mixed integer linear programming techniques used in the distillery analysis.

The basic structure of a programming model optimizes one variable, subject to a set of equality and inequality constraints. The most familiar linear programming (LP) model is the transportation model, where (borrowing the economic application) a set of "factories" and a set of "markets" are assumed, along with factory capacities, market demands and the cost of transporting widgets from factory to market. The model minimizes the total widget transportation costs, subject to factory capacities and market demands (no factory can produce more than its capacity while all of the demand at each market must be met). Multiple products (widgets X and Y) and forbidden pathways (there is no way to ship from factory C to market A) can also be modeled easily.

2.2.1.1 Trans-shipment Model for Predicting Minimum Energy Demands

Another more complicated linear programming technique is the trans-shipment model, where not only factories and markets exist but also intermediate "warehouses," which can transfer widgets among themselves as well as receive shipments from factories and send shipments to markets. Again, the total widget transportation costs are minimized, subject to factory and warehouse capacities and market demands.

The minimum utility cost problem for heat exchanger networks is formulated as a trans-shipment model*[2]. Heat is analogous to widgets, the hot streams (and utilities) providing the heat are the factories, and the cold streams (and heat sinks) requiring heat are analogous to the markets.

The analog of the warehouses is less obvious. The complete temperature range encompassing all of the processes being analyzed is partitioned where streams

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* Although the linear programming method was not used, it provides the groundwork for the MILP technique used to formulate the optimal heat exchanger network.

enter or leave, analogous to the temperature partitioning of the pinch point analysis. A new temperature interval begins at the "adjusted" source or target temperature of any stream. Each temperature interval is now considered a warehouse: it can accept heat from the hot streams within its temperature interval and reject heat to the cold streams within its temperature interval. If more heat is supplied in any particular interval than can be absorbed by the cold streams in that interval, then the excess heat can be "shipped" to the next lowest temperature interval (Figure A.1). This heat passed from one temperature interval to another is referred to as a residual. Residuals can only flow from a hot interval to a cooler one; to do otherwise would violate the Second Law of Thermodynamics. Enough heat must be shipped into the first temperature interval from the "hot utility" so that the heat balances in all of the intervals are met without having to bootleg heat from cold to hot. Excess heat in the last interval must be rejected to the cold utility.

The final problem is formulated as follows: minimize the cost of the hot utility shipment (amount of heat added) plus the cost of the cold utility (heat rejected to the low temperature heat sink), subject to energy balances in every temperature interval, and all of the heating and cooling requirements of all the streams being met. An additional constraint from the Second Law permits only heat shipments from one temperature interval to the next lower temperature interval. If all the available heat is not used in this next lower interval, it is passed on to the next lower one, et cetera, until it reaches an interval where it can be utilized. Different temperature hot and cold utilities can be handled easily, along with different costs for each utility.

Both the pinch point method and the LP trans-shipment method have their respective advantages and disadvantages. The ability of the trans-shipment method to handle multiple heat sources and sinks at different costs is a strong advantage over the pinch method. However in both the sugar factory and distillery cases, only one level of hot and cold utility is used and this advantage becomes moot.

The main reason that the pinch method was used was for its simplicity of implementation, the ability to immediately access the results graphically and the physical insight gained by examining the composite and grand composite curves. It also allows the user a degree of judgement in how to design the system; an LP model could be derived which also could size the evaporators such that a pinch occurred at each effect. However, the "feel" of the system acquired by manually manipulating the pinches, effect loadings, and distillation column balances might not have been gained if the complete process was done using programming methods.

A.2.2 Mixed-Integer Trans-shipment Model for Predicting the Minimum Number of Heat Exchanger Units

The next level of mathematical programming is the mixed-integer linear programming (MILP) trans-shipment model, which is required to solve the minimum number of heat exchanger units problem. In the MILP formulation of a model, each possible match between a hot stream and a cold stream is assigned a binary variable. If a match between two streams exists (i.e. the two streams exchange heat), its binary variable is "1"; if the two streams never exchange any heat, the binary variable will be "0." By minimizing the sum of all these ones and zeros, the minimum number of matches (heat exchanger units) is found.

The increased complexity of this model comes not only from the introduction of binary variables, but also from the accounting of the heat flows. In the general trans-shipment model, *overall* heat balances are derived around each individual temperature interval. In factory-warehouse-market terms, the source of the widget is not kept track of; which stream the heat came from is of no concern. Only the overall heat balance matters.

The mixed integer problem, on the other hand, accounts for *each* individual hot and cold stream *separately*. Rather than having the net excess heat residual cascade down into the next cooler interval, each hot stream's residual in each interval must be accounted for individually (Figure A.2). A variable is assigned to the heat exchanged for each potential match in each interval. Similarly, the amount of heat exchanged between any two hot and cold streams is assigned a variable. As becomes quickly evident, the number of variables and constraint equations (along with the computational effort to solve them) increases considerably with this method.

The pinch point derived earlier for either the pinch method or linear programming can help simplify the matter by dividing the problem into two (or more) sub-problems. The golden rule of pinch analysis, heat should not be transferred across the pinch point, forbids any of the hot residuals to be transferred from the temperature interval immediately above the pinch to the one immediately below. This allows the designer to attack the system above the pinch and below the pinch separately, saving effort in formulating and solving the problem.

The actual statement of the multiple integer linear programming model to find the minimum number of matches is:

<u>minimize:</u>	Number of matches (sum of the binary variables);
<u>subject to:</u>	• Heat balances of each hot stream in each temperature interval; • Heat balances of each cold stream in each temperature interval; • No two streams (over all intervals) exchange more heat than is physically possible; • The Second Law is obeyed: no heat flow from cold to hot.

The solution of this model provides not only the minimum number of matches, it also determines between which streams the matches occur (which of the binary variables are 1 and which are 0), how much heat is exchanged between any two streams, and in which interval(s) the exchanges take place. Armed with this information, it is now simple to construct the actual heat exchanger network.

A.2.3 Mathematical Expression of the MILP Trans-shipment Model*

The first step in setting up a trans-shipment model for synthesizing minimum heat exchanger units is the partitioning of the complete temperature range into K temperature intervals. Each interval is assigned a subscript, k, ranging from k=1 (highest temperature interval) up to k=K (lowest temperature interval). The assumed minimum approach temperature (ΔT_{MIN}) must be considered

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* Based on Reference 2.

in partitioning the temperature range into intervals. In this analysis, ΔT_{MIN} is added to all of the cold streams. If some components require larger approach temperatures than is assumed for the rest of the system, then that additional temperature increment is added to the target and source temperatures of that particular stream. This occurs frequently in sugar factory analysis, where the low heat transfer coefficients in the later evaporator effects and vacuum pans require larger approach temperatures.

If the model is large, it is often desirable to break it up into SN sub-networks, where each sub-network is defined by the highest or lowest temperature interval or a pinch point. Thus, if the model contains *p* pinch points, then the model can be divided up into *p+1* sub-networks (*p+1*=SN). Because the minimum utility conditions derived earlier in the pinch analysis can only be met when there is no heat flow across the pinch, the subdivision of the model is not technically necessary. In the case of very large models, however, it can be computationally expedient to divide the problem up into sub-networks and solve each sub-network separately.

The hot and cold streams which exist in each temperature interval are identified as:

$$
\begin{aligned}
H_{lk} &= \{i \mid \text{hot stream } i \text{ present in interval } k,\ k \leq k_{i,\text{initial}},\ k \in SN_l, (\text{subnetwork } l)\} \\
C_{lk} &= \{j \mid \text{cold stream } j \text{ present in interval } k,\ k \in SN_l\};
\end{aligned}
$$

where $k_{i,\text{initial}}$ is the lowest interval (highest temperature) in which stream *i* appears. Included in the set of hot streams $\{H_i\}$ and cold steams $\{C_j\}$ are all of the hot and cold utilities (steam and cooling water, if appropriate), whose heating and cooling duties were determined in the pinch analysis.

Residuals are given by R_{ik}, which represents the residual heat of hot stream *i* cascading down out of temperature interval *k*. Residuals for a hot stream *i* can exist in all temperature intervals lower than $k_{i,\text{initial}}$ within that particular subnetwork. Because residuals can cascade down numerous intervals before they actively exchange heat with a cold steam, hot streams $\in H_{lk}$ can exist in temperature intervals below which they physically appear. Physically, this means that the hot stream represented by its residual is exchanging heat at approach temperatures greater than ΔT_{MIN}, which actually may be beneficial in terms of reducing heat exchanger area and cost.

Because no heat is transferred across the pinch, all the residual flows into a pinch interval must be zero. These residuals can be forced to equal zero by partitioning the problem into sub-networks, or if the model is run whole will be forced to be zero as part of the solution. The total heat available for transfer in any hot stream *i* or cold stream *j* in interval *k* is given by Q^H_{ik} $k \in SN_l$, and Q^C_{jk} $k \in SN_l$, respectively. Q_{ijk} represents the heat exchanged between hot stream *i* (including the residuals of hot stream *i*) and cold stream *j* in interval *k*.

The binary variables, y_{ijl}, denote the existence of a match between hot stream $i \in H_l$ and cold steam $j \in H_l$ in subnetwork *l*. The total heat exchanged between hot steam *i* and cold stream *j* in subnetwork *l* is therefore limited to:

$$\sum_{k\in SNl} Q_{ijk} - U_{ijl}y_{ijl} \le 0$$

$$i\in H_l,\ j\in C_l,\ l=1,2,....,SN$$

where

$$U_{ijl} \ge \min \left\{ \sum_{k\in SNl} Q^H_{ik}, \sum_{k\in SNl} Q^C_{jk} \right\}$$

which is the maximum heat available from the hot steam or needed by the cold stream in subnetwork *l*. This condition insures that no more heat is exchanged than is available from the hot stream or needed by the cold stream. If the binary variable y_{ijl} equals zero, implying that no match exists between the two streams, then $\sum Q_{ijk}$ is forced to be zero in order to meet the inequality.

In summary, the variables are:

• Q_{ijk}	Heat exchanged between streams *i* and *j* in interval *k*
• R_{ik}	Residual from stream *i* entering interval *k*
• y_{ijl}	Binary selection variable indicating a match between streams *i* and *j* in subnetwork *l*.

And the parameters:

• Q^H_{ik}	Heat available from hot stream *i* in interval *k*
• Q^C_{ik}	Heat required by cold stream *j* in interval *k*
• U_{ijl}	Maximum amount of heat which can be transferred between streams *i* and *j* in subnetwork *l*
• e_{ijl}	Weighing factor applied to binary variable to indicate preferred or forbidden matches*.

With all the terms defined, the transhipment model can now be formulated:

minimize:

$$Z = \sum_{l=1,NL} \sum_{i\in Hl} \sum_{j\in Cl} e_{ijl}y_{ijl}$$

◆ number of matches

subject to:

$$\left.\begin{array}{ll} R_{ik} - R_{ik-1} + \sum_{j\in Clk} Q_{ijk} = Q^H_{ik} & i\in H_{lk} \\ \sum_{i\in Hlk} Q_{ijk} = Q^C_{jk} & j\in C_{lk} \end{array}\right\} \begin{array}{l} k\in SN_l \\ l = 1,2,\ldots,SN \end{array}$$

◆ heat balances for each stream in each interval

$$\sum_{k\in SNl} Q_{ijk} - U_{ijl}y_{ijl} \le 0 \quad i\in H_l,\ j\in C_l,\ l = 1,2,\ldots,SN$$

◆ insures that no more heat than is available or required is transferred.

---

* Sometimes it is the case that two streams are on opposite ends of a facility or for safety reasons should not be brought close to each other. In such cases, the e variable is set very high so that this match will not be chosen. Conversely, when steams are physically very close to each other in a plant, it can particularly desirable to match them. In this case the e variable is set to some value less than one so that this match is preferred.

$$\left.\begin{array}{lll} R_{ik} \geq 0 & i \in H_{lk} & \\ Q_{ijk} \geq 0 & i \in H_{lk} & \\ & j \in C_{lk} & \end{array}\right\} \begin{array}{l} k \in SN_l \\ l = 1,2,\ldots,SN \end{array}$$

◆ logical constraints: residuals and heat flows are positive.

$$y_{ijk} = 0,1 \quad i \in H_l,\ j \in C_l, \qquad l = 1,2,\ldots,SN.$$

◆ binary constraint.

A.2.4 MILP Implementation

The MILP trans-shipment models used to synthesize the heat exchanger networks containing the minimum number of matches were solved on the Princeton University IBM 3081 mainframe computer using the GAMS and ZOOM software packages. GAMS is a software environment which allows the modeler to concisely express a programming model in a PASCAL-like language, and "translates" the model into the form required by numerous solution packages [3].

The ZOOM solution package used to solve the MILP models uses a combination of heuristic and branch and bound techniques to solve integer programming problems [3]. ZOOM first optimizes a relaxed LP problem in which the the binary constraints are removed and the "binary" variables are allowed to take on real values between 0 and 1. At this point it checks the solution to see if the relaxed "binary" variables happen to be indeed binary (not likely). If they are, the solution to the less constrained relaxed LP problem is the same as the more constrained MILP one, the problem is solved and the program terminates; if they are not, it applies a heuristic algorithm and checks for binary solutions near the relaxed LP solution. If this does not yield an optimal solution, then the program resorts into a branch and bound search.

Branch and bound is a tree search algorithm [4]. It first takes the optimal, relaxed non-binary LP solution and constrains one of the "binary" variables to be either 0 or 1. For each of the two branches (y=0 and y=1), it solves the LP problem with that one variable constrained while the others are still allowed to vary between 0 and 1. These two solutions (or "nodes") will be either (1) infeasible or (2) have an objective variable value some amount greater (assuming a minimization) than the less constrained solution.* The program then chooses the better of the two nodes, constrains another variable to 1 or 0, and performs another LP solution. Again, this might be infeasible or yield a solution greater than the previous node. The value of the objective variable at this node is compared to the value of the objective variable at all other terminal nodes (nodes without branches emanating from them); the program then chooses the node with the minimum objective value, constrains another binary variable to 0 or 1 at that node and repeats the process. The optimal binary solution is achieved when a binary solution is reached whose objective variable is less than that at any of the remaining terminal nodes. An example of a branch and bound tree can be seen in Figure A.3.

The MILP model does not explicitly create the heat exchanger network; therefore other tools and engineering judgement must also be used to complete the heat exchanger network fleshed out by the MILP model. The first way engineering judgement was used was in limiting the model size. Because each additional stream doubles the size of the search tree, it is desirable to minimize the number

---

* As the problem becomes more constrained with more binary variables, the objective variable will always become less optimal. i.e. if it is a minimization model (ojbective variable z is being minimized), each node on the decision tree will have a z value greater than or equal to the z value of the previous node.

streams input into the model. A linear programming optimization must also be performed at each search node, providing incentive to not only reduce the number of streams but also the number of temperature intervals.

Model reduction was accomplished in two ways. First, matches which were known a priori were eliminated from the model. These matches include the condensing vapors from effect *n* evaporating liquid from effect *n+1*, the HP column condenser matched to the first effect, etc. These assumptions eliminated the cold streams being heated from the MILP model. The heating capacity of the hot streams was then reduced by the amount known to be used by the eliminated component. If the hot stream could provide no heating beyond the known match, then it was also eliminated. This not only eliminated two steams from the model, but also often a temperature interval as well.

When two streams matched exactly, not only was the model simplified, but also often the resulting heat exchanger network. Thus, this was consciously pursued whenever possible (e.g. matching the vapors from effect 3 to the evaporation in effect 4 in the HP-1234 case), sometimes even at the expense of steam economy (the tradeoffs of such decisions are discussed in Section 4.3.).

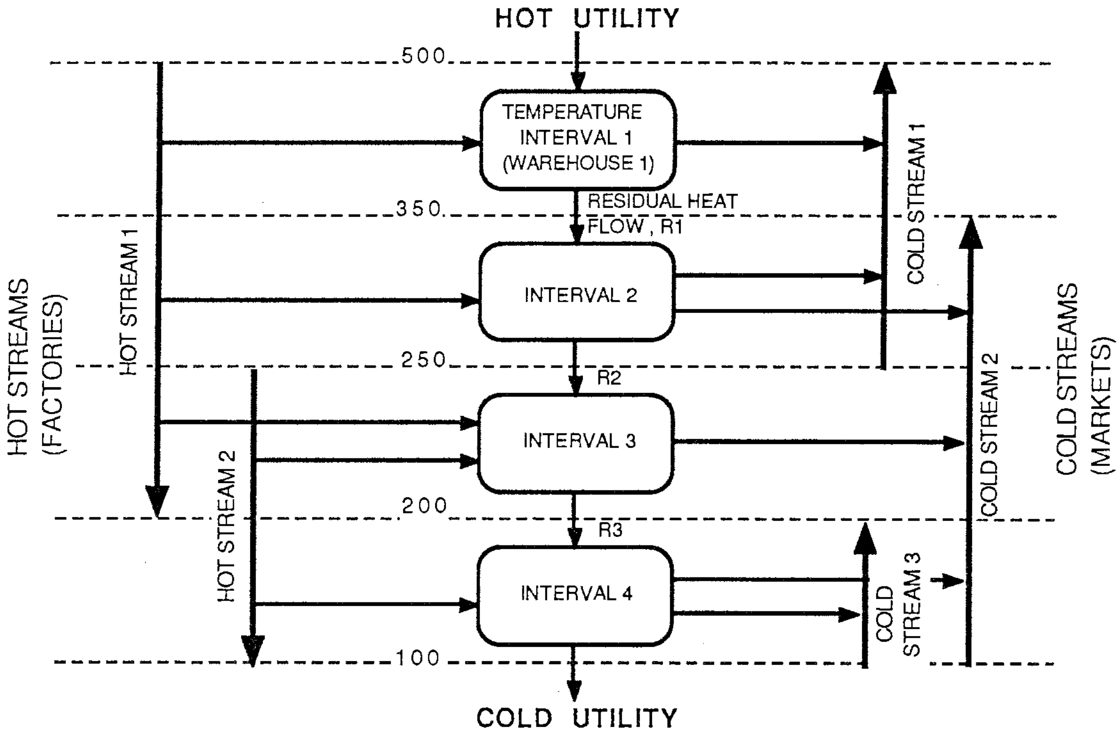
The second model reduction technique is actually undoing a device implemented in the pinch analysis. Looking at the stream definitions in the pinch analysis in Chapter 3,Section 3, condensing streams were defined to occur in a temperature interval between T_{actual}+.1°C and T_{actual}, and evaporating steams were defined to occur between $T_{adjusted}$-.1°C and $T_{adjusted}$. This was done so that the two streams would be distinct in the problem table and on the grand composite curve. If they were placed in the same temperature interval, then the MCp's of the two would have come close to cancelling each other out and the location of the phase changes would not have been nearly so evident. In cases where streams changing phases were not eliminated from the model, a single temperature interval was assumed, thus undoing the pinch analysis convenience and eliminating a temperature interval.

The second place where engineering judgement was exercised was in interpreting the data from the MILP model into a heat exchanger network. This was done in a number of ways. For instance, whatever vapor was being used to heat the highest evaporation effect (either steam or the vapors from the HP distillation column), the last bit of clear juice heating would often be done by that same vapor. The model would interpret this as a separate match, even though it would physically occur in the first evaporator effect rather than in a separate heat exchanger unit. Therefore, the actual number of heat exchanger units was one less than the number of stream matches.

Another example would be the relaxation of the minimum approach temperature assumption to eliminate a match. As was occasionally the case, a degree or two of juice heating would be specified by a match with a stream with no other connections. If this small amount of heating load could be spread around to other exchangers heating that particular stream without dropping any of the approach temperatures below about 9°C, then the extra heat exchanger was eliminated.

Similarly, the temperature ranges which the model specified for stream matches were sometimes modified. For instance, in most of the cases where the HP column was used, the last bit of column feed heating was accomplished by HP steam condensate and the hot stillage from the HP column. The model would often specify the heat exchange in these matches such that the minimum approach temperature was held constant throughout one or both of the heat exchangers. If it were possible, the temperature ranges of the hot streams extended or the split feed flows redistributed so that greater approach temperatures could be implemented, thus reducing the areas of the heat exchangers and capital costs.

Figure A.1: Diagram of Linear Programming Trans-shipment Model to Predict Minimum Utility Cost.



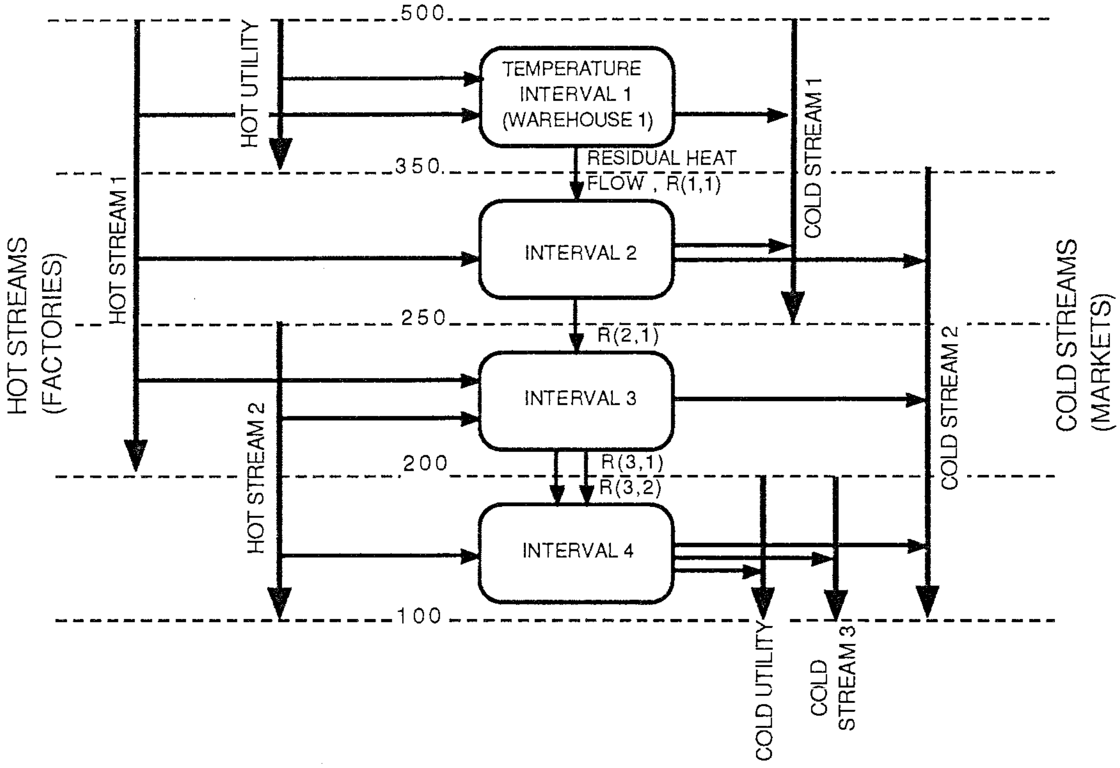
STREAM	T(SOURCE)	T(TARGET)
HOT 1	500	200
HOT 2	250	100
COLD 1	250	500
COLD 2	100	350
COLD 3	100	200

Figure A.1 *Shows diagrammatically what occurs in the LP-trans-shipment model. The hot stream "factories" ship heat to the temperature interval "warehouses"; cold stream "markets" accept heat from the temperature interval warehouses. Hot and cold streams can only ship and accept heat from temperature interval in which they occur.*

Residual shipments from temperature interval warehouse to warehouse are also shown. At some point the residual shipment will be zero, indicating a pinch. The excess heat required to maintain heat balances in all of the temperature intervals is indicated by the hot utility entering the first interval. The excess heat which must be rejected from the last temperature interval in order to maintain a heat balance is indicated by the cold utility.

All temperatures are already adjusted.

Figure A.2: Diagram of Mixed-Integer Linear Programming Trans-shipment Model to Predict Minimum Number of Heat Exchangers.



STREAM	T(SOURCE)	T(TARGET)
HOT UTILITY	500	
HOT 1	500	200
HOT 2	250	100
COLD 1	250	500
COLD 2	100	350
COLD 3	100	200
COLD UTILITY	100	

Figure A.2: *The critical difference between this figure and Figure 3.6 is the distinct accounting of the residual flows. Because it is now important to know which hot stream is the source of the heat being shipped to a cold stream, residuals must be kept track of for each hot stream.*

Notice also that the hot and cold utilities are now known and treated just the same as hot and cold streams in the process.

All temperatures are already adjusted.

Figure A.3: Example Tree for a Binary Mixed-Integer Branch and Bound Search.

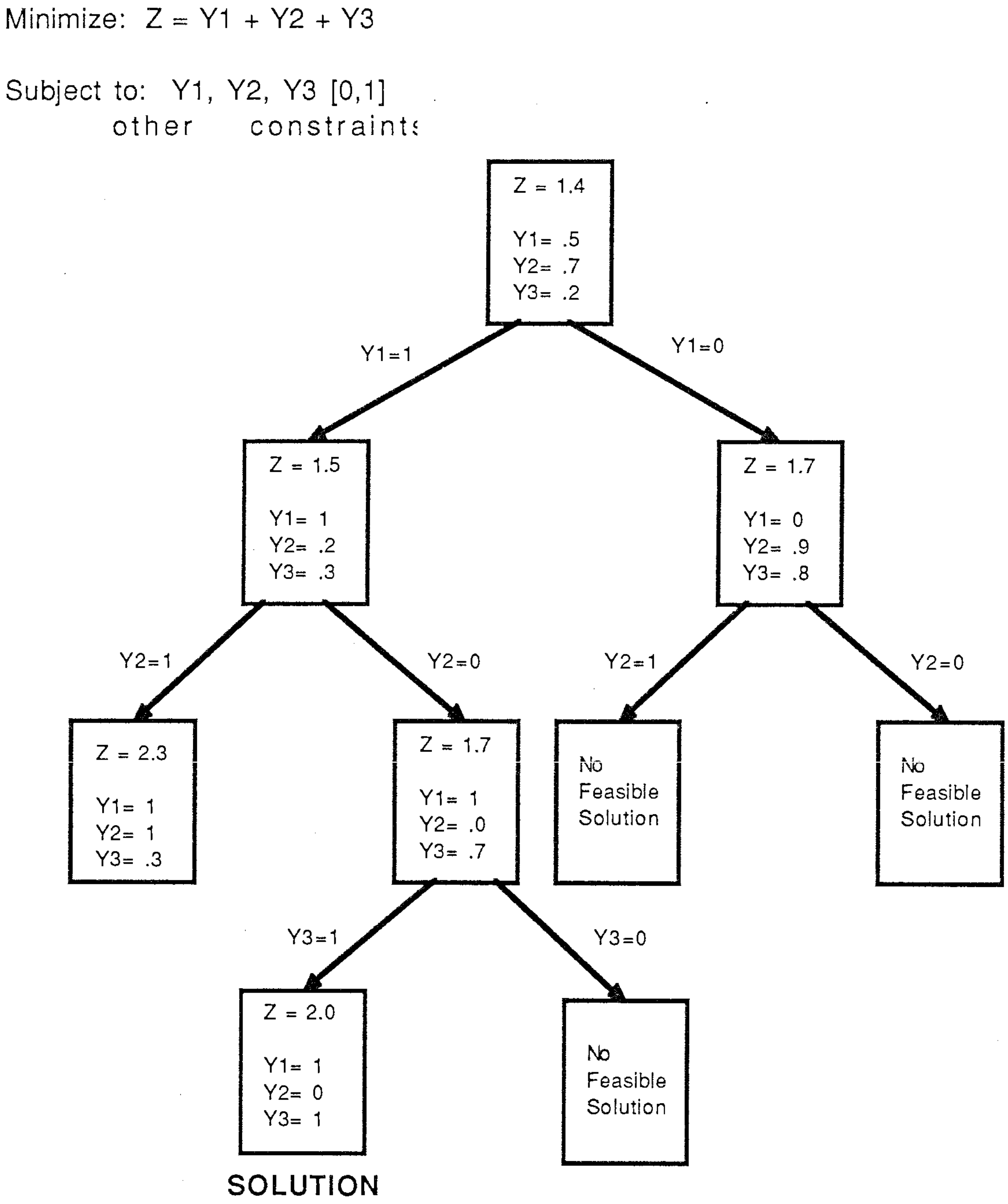


***Figure A.3** shows a hypothetical branch and bound tree search for a system with three binary variables (Y1-Y3). It the first box, a linear programming optimization is performed allowing the variables to range between 0 and 1. Y1 is then constrained to 0 or 1, and another linear programming optimization is performed, this time allowing only Y2 and Y3 to vary between 0 and 1. This process is repeated until a branch either terminates because no feasible solution exists or a binary solution is achieved which is less than any other terminal node on the tree. This second case is seen along the branch Y1=1, Y2=1, where the terminal node is feasible, but has a Z value greater than the binary solution Y1=1, Y2=0, Y3=1. The solution Y1=1, Y2=1 Y3=0 cannot exist because the Z variable at the preceding node, Y1=1, Y2=1, Y3=0.3 is greater than 2, and each node on the tree must be less optimal than the node from which it emanates.*

REFERENCES

1. Linnhoff, B.D., D.W Townsend,... *A User Guide on Process Integration for the Efficient Use of Energy,* /the Institution of Chemical Engineers, U.K., 1982.

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3. Brooke, A. D. Kendrick and A. Meeraus, *GAMS: A User's Guide*, The International Bank for Reconstruction and Development/ The World Bank. 1988.

4. Cooper, L., and D. Steinberg, *Methods and Applications of Linear Programming*, W.B. Saunders Co., Philadelphia. 1974.

APPENDIX III:
SAMPLE GAMS INPUT FILES

GAMS input files are shown for autonomous distillery case HP-1-2-3-4 and the four effect, falling film evaporator sugar factory with continuous vacuum pans and disaggregated condensate accounting.

The GAMS file can be broken up into a number of major sections. The first section (line 1 to 29 on the HP-1-2-3-4 case) defines the sets which are to be used. For example in lines 3,4 and 5 the hot and cold streams are defined along with the temperatures intervals. The remainder of the first section defines which streams are in which intervals and which streams can exchange heat.

The second section contains the actual data (lines 31 through 94 in the HP-1-2-3-4 file). While there are actually many ways to give values to the sets defined in the first section, the "table" method used here was the most expedient.

The third section of a GAMS file defines the variables (lines 97-105). Here one can see the residuals being defined along with any variable constraints--e.g.binary variable Y.

The next section of the GAMS input defines the constraint and objective functions. Each equation must first be declared (lines 110-113) and then defined (lines 116-126).

The remaining lines define the overall model and prescribe what kind of optimization is to be performed. For instance, the "MIP" in line 131 indicates that this in a mixed integer programming problem.

HP 1234

```
G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
C O M P I L A T I O N


SET

     HOT     HOT STREAMS     /STEAM,SCOND,HPB,HPC,HPD,E1V,CCOND,E4V,FRMFEED/
     COLD    COLD STREAMS           /HPFEED,HPRBLR,EFF1,RAWJ,CLRJ/
     INT     TEMPERATURE INTERVALS /K1*K14/
     EXH(HOT,INT)    HOT STREAMS IN INTERVAL K
                   /STEAM.(K1*K14)
                    SCOND.(K2*K14)
                    HPB.(K4*K14)
                    HPC.(K5*K14)
                    HPD.(K6*K14)
                    E1V.(K7*K14)
                    CCOND.(K10*K14)
                    E4V.(K12*K14)
                    FRMFEED.(K13*K14) /

     EXC(COLD,INT)  COLD STREAMS IN INTERVAL K
                   /HPFEED.(K3*K14)
                    HPRBLR.K1
                    EFF1.K5
                    RAWJ.(K9*K14)
                    CLRJ.(K5*K10) /

     MATCH(COLD,HOT)  POSSIBLE MATCHES BETWEEN HOT AND COLD STREAMS
                   /HPFEED.(STEAM,SCOND,HPB,HPC,HPD,E1V,CCOND,E4V,FRMFEED)
                    HPRBLR.(STEAM)
                    EFF1.(HPC)
                    RAWJ.(SCOND,HPB,HPC,HPD,E1V,CCOND)
                    CLRJ.(SCOND, HPB,HPC,HPD,E1V,CCOND) /

TABLE  QH(INT,HOT)     HEAT FLOWS OF HOT STREAMS IN INTERVAL K
               STEAM       SCOND         HPB         HPC         HPD         E1V
      K1      318842           0           0           0           0           0
      K2           0        5737           0           0           0           0
      K3           0        1300           0           0           0           0
      K4           0       18206       72046           0           0           0
      K5           0          65         257      303270           0           0
      K6           0        6437       25473           0        1869           0
      K7           0          65         257           0          19      119069
      K8           0        5137       20327           0        1491           0
      K9           0        4551       18012           0        1322           0
      K10          0        3251       12865           0         944           0
      K11          0        6502       25731           0        1888           0
      K12          0          65         257           0          19           0
      K13          0       29194      115531           0        8477           0
      K14          0        3251       12865           0         944           0

       +       CCOND         E4V     FRMFEED
      K1           0           0           0
      K2           0           0           0
      K3           0           0           0
      K4           0           0           0
      K5           0           0           0
      K6           0           0           0
      K7           0           0           0
      K8           0           0           0
```

```
  57       K9           0          0          0
  58      K10        6793          0          0
  59      K11       13585          0          0
  60      K12         136     151900          0
  61      K13       60997          0     111532
  62      K14        6793          0      12420
  63       ;
  64
  65
  66  TABLE  QC(INT,COLD)     HEAT FLOWS OF COLD STREAMS IN INTERVAL K
  67               HPFEED     HPRBLR       EFF1       RAWJ       CLRJ
  68       K1           0     318842          0          0          0
  69       K2           0          0          0          0          0
  70       K3        5913          0          0          0          0
  71       K4       82782          0          0          0          0
  72       K5         296          0     270969          0        384
  73       K6       29269          0          0          0      38016
  74       K7         296          0          0          0        384
  75       K8       23356          0          0          0      30336
  76       K9       20696          0          0          0      26880
  77      K10       14783          0          0      19200      19200
  78      K11       29565          0          0      38400          0
  79      K12         296          0          0        384          0
  80      K13      132747          0          0     172416          0
  81      K14           0          0          0      19200          0
  82       ;
  83
  84  TABLE  U(HOT,COLD)     MAXIMUM HEAT EXCHANGE BETWEEN HOT AND COLD STREAMS
  85               HPFEED     HPRBLR       EFF1       RAWJ       CLRJ
  86  STEAM             0     318842          0          0          0
  87  SCOND         83811          0          0      83811      83811
  88    HPB        303624          0          0     249600     115200
  89    HPC        303270          0     270969     249600     115200
  90    HPD         16972          0          0      16972      16972
  91    E1V        119069          0          0     119069     115200
  92  CCOND         88302          0          0      88302      88302
  93    E4V        151900          0          0     151900     115200
  94  FRMFEED      123952          0          0     123952          0
  95     ;
  96
  97  VARIABLES
  98
  99            R(HOT,INT)               RESIDUAL OF HOT STREAM LEAVING INTERVAL K
 100            QX(HOT,COLD,INT)         HEAT EXCHANGED BEWTEEN HOT AND COLD IN INT. K
 101            Y(HOT,COLD)              BINARY SELECTOR OF HEAT EXCHANGERS
 102            Z                        SUM OF ALL Y'S-- TO BE MINIMIZED
 103      ;
 104  POSITIVE VARIABLES R, QX;
 105  BINARY VARIABLE Y
 106     ;
 107
 108  EQUATIONS
 109
 110      QBALH(HOT,INT)             HEAT BALANCE FOR HOT STREAMS IN INTERVAL K
 111      QBALC(COLD,INT)            HEAT BALANCE FOR COLD STREAMS IN INTERVAL K
 112      MAXMATCH(HOT,COLD)         INSURES OVERALL HEAT EXCHANGE DOES NOT EXCEED MAX
```

```
MAXMATCH(HOT,COLD)        INSURES OVERALL HEAT EXCHANGE DOES NOT EXCEED MAX
```

```
    UNITS                         SUMMATION OF BINARY VARIABLES TO BE MINIMIZED ;


QBALH(HOT,INT)..                 QH(INT,HOT) =E= R(HOT,INT) - R(HOT,INT-1)
                                 +SUM(COLD$EXC(COLD,INT),QX(HOT,COLD,INT))
;
QBALC(COLD,INT)..                QC(INT,COLD) =E=
                                 SUM(HOT$EXH(HOT,INT),QX(HOT,COLD,INT))
;
MAXMATCH(HOT,COLD)..             SUM(INT$EXC(COLD,INT),

              QX(HOT,COLD,INT)) - U(HOT,COLD)*Y(HOT,COLD) =L= 0
;
UNITS..                          SUM((HOT,COLD),Y(HOT,COLD)) =E= Z
;

MODEL H1234  AUTONOMOUS DISTILLERY EVAPORATOR SECTION   /ALL/
OPTION ITERLIM= 5000
SOLVE H1234 USING MIP MINIMIZING Z;
```

SF 4EF FF sep Cond

```
GENERAL ALGEBRAIC MODELING SYSTEM
COMPILATION

SET

    HOT    HOT STREAMS         /STEAM,SCOND,E1V,E1C,E2V,E2C,E3V,E3C/
    COLD   COLD STREAMS          /RAWJ,CLRJ,EFF1,EFF2,EFF3,VACP,EFF4/
    INT    TEMPERATURE INTERVALS /K1*K14/
    EXH(HOT,INT)   HOT STREAMS IN INTERVAL K
                  /STEAM.(K1*K14)
                   SCOND.(K2*K14)
                   E1V.(K3*K14)
                   E1C.(K4*K14)
                   E2V.(K5*K14)
                   E2C.(K6*K14)
                   E3V.(K11*K14)
                   E3C.(K12*K14)  /
    EXC(COLD,INT)  COLD STREAMS IN INTERVAL K
                  /RAWJ.(K7*K14)
                   CLRJ.(K2*K9)
                   EFF1.K1
                   EFF2.K3
                   EFF3.K5
                   VACP.K8
                   EFF4.K11
                          /
    MATCH(COLD,HOT)  POSSIBLE MATCHES BETWEEN HOT AND COLD STREAMS
                   /RAWJ.(E2V,E2C,SCOND,E1C,E3V,E3C)
                    CLRJ.(STEAM,E1V,E2V,E2C,SCOND,E1C)
                    EFF3.(E2V)
                    VACP.(E2V)
                    EFF4.(E2V,E3V)  /;

TABLE  QH(INT,HOT)     HEAT FLOWS OF HOT STREAMS IN INTERVAL K
          STEAM      SCOND        E1V        E1C        E2V        E2C
  K1   97956385          0          0          0          0          0
  K2          0    1857717          0          0          0          0
  K3          0      37913   93732000          0          0          0
  K4          0    1857717          0    1753259          0          0
  K5          0      37913          0      35713   91520800          0
  K6          0     549733          0     518822          0     500639
  K7          0     360170          0     339918          0     328005
  K8          0      18956          0      17890          0      17263
  K9          0    1895630          0    1798040          0    1726340
 K10          0     549733          0     518822          0     500639
 K11          0      37913          0      35789          0      34527
 K12          0    4132473          0    3900107          0    3763421
 K13          0      18956          0      17895          0      17263
 K14          0    5876453          0    5546024          0    5351654


   +        E3V        E3C
  K1          0          0
  K2          0          0
  K3          0          0
  K4          0          0
  K5          0          0
  K6          0          0
  K7          0          0
```

```
 57     K8          0         0
 58     K9          0         0
 59    K10          0         0
 60    K11   44148300         0
 61    K12          0   1771451
 62    K13          0      8128
 63    K14          0   2519035
 64     ;
 65
 66  TABLE  QC(INT,COLD)     HEAT FLOWS OF COLD STREAMS IN INTERVAL K
 67               RAWJ      CLRJ      EFF1      EFF2      VACP      EFF3      EFF4
 68     K1          0         0  93732000         0         0         0         0
 69     K2          0   5908224         0         0         0         0         0
 70     K3          0    120576         0  91520800         0         0         0
 71     K4          0   5908224         0         0         0         0         0
 72     K5          0    120576         0         0         0  44148300         0
 73     K6          0   1748352         0         0         0         0         0
 74     K7    1145472   1145472         0         0         0         0         0
 75     K8      60288     60288         0         0  38762500         0         0
 76     K9    6028800   6028800         0         0         0         0         0
 77    K10    1748352         0         0         0         0         0         0
 78    K11     120576         0         0         0         0         0  44148300
 79    K12   13142784         0         0         0         0         0         0
 80    K13      60288         0         0         0         0         0         0
 81    K14   18689280         0         0         0         0         0         0
 82     ;
 83
 84  TABLE  U(HOT,COLD)    MAXIMUM HEAT EXCHANGE BETWEEN HOT AND COLD STREAMS
 85               RAWJ      CLRJ      EFF1      EFF2      VACP      EFF3      EFF4
 86  STEAM          0  21109961  93732000         0         0         0         0
 87  SCOND   17250233   6634705         0         0         0         0         0
 88    E1V   41013639  15072000         0  91520800         0         0         0
 89    E1C   14491224  44726009         0         0         0         0         0
 90    E2V   42804480   9043200         0         0  38762500  44148300         0
 91    E2C   12257014   2589510         0         0         0         0         0
 92    E3V   31952640         0         0         0         0         0  44148300
 93    E3C    4306737         0         0         0         0         0         0
 94    ;
 95  VARIABLES
 96
 97            R(HOT,INT)            RESIDUAL OF HOT STREAM LEAVING INTERVAL K
 98            QX(HOT,COLD,INT)      HEAT EXCHANGED BEWTEEN HOT AND COLD IN INT. K
 99            Y(HOT,COLD)           BINARY SELECTOR OF HEAT EXCHANGERS
100            Z                     SUM OF ALL Y'S-- TO BE MINIMIZED
101     ;
102  POSITIVE VARIABLES R, QX;
103  BINARY VARIABLE Y
104     ;
105
106  EQUATIONS
107
108      QBALH(HOT,INT)           HEAT BALANCE FOR HOT STREAMS IN INTERVAL K
109      QBALC(COLD,INT)          HEAT BALANCE FOR COLD STREAMS IN INTERVAL K
110      MAXMATCH(HOT,COLD)       INSURES OVERALL HEAT EXCHANGE DOES NOT EXCEED MAX
111      UNITS                    SUMMATION OF BINARY VARIABLES TO BE MINIMIZED ;
112
```

```
G E N E R A L   A L G E B R A I C   M O D E L I N G   S Y S T E M
C O M P I L A T I O N

 113
 114  QBALH(HOT,INT)..                     QH(INT,HOT) =E= R(HOT,INT) - R(HOT,INT-1)
 115                                       +SUM(COLD$EXC(COLD,INT),QX(HOT,COLD,INT))
 116  ;
 117  QBALC(COLD,INT)..                    QC(INT,COLD) =E=
 118                                       SUM(HOT$EXH(HOT,INT),QX(HOT,COLD,INT))
 119  ;
 120  MAXMATCH(HOT,COLD)..                 SUM(INT$EXC(COLD,INT),
 121
 122                      QX(HOT,COLD,INT)) - U(HOT,COLD)*Y(HOT,COLD) =L= 0
 123  ;
 124  UNITS..                              SUM((HOT,COLD),Y(HOT,COLD)) =E= Z
 125  ;
 126
 127  MODEL SF4FF   SUGAR FACTORY QUAD EFFECT FALLING FILM EVAPS   /ALL/
 128
 129  SOLVE SF4FF USING MIP MINIMIZING Z;
```

APPENDIX V:
BIG/STIG AND BIG/ISTIG CAPITAL COSTS

Table 7. Estimated Installed Capital Cost for IG/STIG and IG/ISTIG Power Plants Fueled with Coal and Biomass (in 1986$/kW)

	CIG/STIG[a]	BIG/STIG[b,c]	CIG/ISTIG[a]	BIG/ISTIG[b,d]
I. Process Capital Cost				
Fuel Handling	39.6	39.6	36.7	36.7
Blast Air System	13.5	13.5	9.6	9.6
Gasification Plant	160.9	160.9	83.1	83.1
Raw Gas Physical Clean-up	8.8	8.8	7.7	7.7
Raw Gas Chemical Clean-up	175.9	0.0	150.9	0.0
Gas turbine/HRSG	294.4	294.4	256.4	256.4
Balance of Plant				
Mechanical	40.2	40.2	22.0	22.0
Electrical	65.0	65.0	48.4	48.4
Civil	65.5	65.5	60.7	60.7
SUBTOTAL	862.9	687.0	686.5	535.6
II. Total Plant Cost				
Process Plant Cost	862.9	687.0	686.5	535.6
Engineering Home Office (10%)	86.3	68.7	68.6	53.6
Process Contingency (6.2%)	53.6	42.6	42.5	33.2
Project Contingency (17.4%)	150.4	119.5	119.6	93.2
SUBTOTAL	1153.2	917.8	917.2	715.6
III. Total Plant Investment				
Total Plant Cost	1153.2	917.8	917.2	715.6
AFDC (1.8%, 2 yr construction)	20.8	16.5	16.5	12.9
SUBTOTAL	1174.0	934.3	933.7	728.5
IV. Total Capital Requirement				
Total Plant Investment	1174.0	934.3	933.7	728.5
Preproduction Costs (2.8%)	32.3	26.2	26.2	20.4
Inventory Capital (2.8%)	32.3	26.2	26.2	20.4
Initial Chemicals, Catalysts	2.5	0.0	2.3	0.0
Land	1.3	1.3	1.3	1.3
TOTAL	1242	988	990	771

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[a] J.C. Corman, "System Analysis of Simplified IGCC Plants," General Electric Company, Schenectady, NY, Report on Department of Energy Contract No. DE-AC21-80ET14928, September 1986.

[b] It is assumed that BIG/STIG (BIG/ISTIG) costs are the same as CIG/STIG (CIG/ISTIG) costs, except that the raw gas chemical clean-up phase required for coal would not be needed for biomass, because of its lower sulfur content.

[c] For a 53 MW unit.

[d] For a 111 MW unit.

APPENDIX IV:
HEAT EXCHANGER COST CALCULATIONS

AUTONOMOUS DISTILLERY HEAT EXCHANGER SUMMARIES
AVERAGED STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
MP-LP	699	2,205	1,420
HP-12345-LP-6	1,686	2,648	946
HP-1234	1,701	1,301	481
2345	1,339	1,164	367
+ MPLP	805	763	381

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$1,117,432	$1,209,700	$1,349,500	$220,500	$213,000	$1,337,932	$1,330,432	$179,121	$178,117	.13387878
$283,248	$505,800	$843,000	$264,800	$141,900	$548,048	$425,148	$73,372	$56,918	
$285,768	$510,300	$850,500	$130,100	$72,150	$415,868	$357,918	$55,676	$47,918	
$224,952	$401,700	$669,500	$116,400	$55,050	$341,352	$280,002	$45,700	$37,486	
$135,240	$241,500	$402,500	$76,300	$57,150	$211,540	$192,390	$28,321	$25,757	
$360,192	$643,200	$1,072,000	$192,700	$112,200	$552,892	$472,392	$74,021	$63,243	

	.128	kwh/kg stm
	.146	
	.115	
ave		.1296666667
tc per season		576000
		74688kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	3c/kwh	5c/kwh
$179,121	$178,117	$191,473	$190,469	$210,190	$209,186	266,636	444,394
$73,372	$56,918	$103,167	$86,713	$148,311	$131,857	313,690	522,816
$55,676	$47,918	$85,736	$77,978	$131,282	$123,523	347,299	578,832
$45,700	$37,486	$69,363	$61,149	$105,215	$97,002		
$28,321	$25,757	$42,547	$39,983	$64,101	$61,537		
$74,021	$63,243	$111,909	$101,132	$169,316	$158,539	483,978	806,630

144

$168/M2				$300/M2				$500/M2			
ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
445,757	444,753	623,514	622,510	458,110	457,106	635,867	634,863	476,826	475,822	654,583	653,579
387,062	370,608	596,188	579,734	416,857	400,403	625,983	609,529	462,001	445,547	671,127	654,673
402,975	395,217	634,508	626,750	433,035	425,277	664,568	656,810	478,581	470,822	710,114	702,355
557,999	547,222	880,651	869,874	595,888	585,110	918,540	907,762	653,295	642,517	975,947	965,170

BIG/STIG STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
MP-LP	699	2,205	1,420
HP-12345-LP-6	1,686	2,648	946
HP-1234	1,701	1,301	481
2345	1,339	1,164	367
+ MPLP	805	763	381

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$1,117,432	$1,209,700	$1,349,500	$220,500	$213,000	$1,337,932	$1,330,432	$179,121	$178,117	.13387878
$283,248	$505,800	$843,000	$264,800	$141,900	$548,048	$425,148	$73,372	$56,918	
$285,768	$510,300	$850,500	$130,100	$72,150	$415,868	$357,918	$55,676	$47,918	
$224,952	$401,700	$669,500	$116,400	$55,050	$341,352	$280,002	$45,700	$37,486	
$135,240	$241,500	$402,500	$76,300	$57,150	$211,540	$192,390	$28,321	$25,757	
$360,192	$643,200	$1,072,000	$192,700	$112,200	$552,892	$472,392	$74,021	$63,243	

ave .128
tc per season 576000
73728kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	3.5c/kwh	5c/kwh
$179,121	$178,117	$191,473	$190,469	$210,190	$209,186	307,077	438,682
$73,372	$56,918	$103,167	$86,713	$148,311	$131,857	361,267	516,096
$55,676	$47,918	$85,736	$77,978	$131,282	$123,523	399,974	571,392
$45,700	$37,486	$69,363	$61,149	$105,215	$97,002		
$28,321	$25,757	$42,547	$39,983	$64,101	$61,537		
$74,021	$63,243	$111,909	$101,132	$169,316	$158,539	557,384	796,262

$168/M2				$300/M2				$500/M2			
ELECT AT 3.5c/KWH		ELECT AT 5c/KWH		ELECT AT 3.5c/KWH		ELECT AT 5c/KWH		ELECT AT 3.5c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
486,198	485,194	617,802	616,798	498,551	497,546	630,155	629,151	517,267	516,263	648,871	647,867
434,639	418,185	589,468	573,014	464,434	447,980	619,263	602,809	509,578	493,124	664,407	647,953
455,650	447,892	627,068	619,310	485,710	477,952	657,128	649,370	531,256	523,498	702,674	694,915
631,404	620,627	870,283	859,506	669,293	658,516	908,172	897,394	726,700	715,923	965,579	954,802

CEST STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
MP-LP	699	2,205	1,420
HP-12345-LP-6	1,686	2,648	946
HP-1234	1,701	1,301	481
2345	1,339	1,164	367
+ MPLP	805	763	381

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$1,117,432	$1,209,700	$1,349,500	$220,500	$213,000	$1,337,932	$1,330,432	$179,121	$178,117	.13387878
$283,248	$505,800	$843,000	$264,800	$141,900	$548,048	$425,148	$73,372	$56,918	
$285,768	$510,300	$850,500	$130,100	$72,150	$415,868	$357,918	$55,676	$47,918	
$224,952	$401,700	$669,500	$116,400	$55,050	$341,352	$280,002	$45,700	$37,486	
$135,240	$241,500	$402,500	$76,300	$57,150	$211,540	$192,390	$28,321	$25,757	
$360,192	$643,200	$1,072,000	$192,700	$112,200	$552,892	$472,392	$74,021	$63,243	

ave	.146
tc per season	576000
	84096kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	8c/kwh	5c/kwh
$179,121	$178,117	$191,473	$190,469	$210,190	$209,186	800,594	500,371
$73,372	$56,918	$103,167	$86,713	$148,311	$131,857	941,875	588,672
$55,676	$47,918	$85,736	$77,978	$131,282	$123,523	1,042,790	651,744
$45,700	$37,486	$69,363	$61,149	$105,215	$97,002		
$28,321	$25,757	$42,547	$39,983	$64,101	$61,537		
$74,021	$63,243	$111,909	$101,132	$169,316	$158,539	1,453,179	908,237

$168/M2				$300/M2				$500/M2			
ELECT AT 8c/KWH		ELECT AT 5c/KWH		ELECT AT 8c/KWH		ELECT AT 5c/KWH		ELECT AT 8c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
979,715	978,711	679,492	678,488	992,067	991,063	691,845	690,841	1,010,784	1,009,780	710,561	709,557
1,015,247	998,793	662,044	645,590	1,045,042	1,028,588	691,839	675,385	1,090,186	1,073,732	736,983	720,529
1,098,466	1,090,708	707,420	699,662	1,128,526	1,120,768	737,480	729,722	1,174,072	1,166,314	783,026	775,267
1,527,199	1,516,422	982,257	971,480	1,565,088	1,554,311	1,020,146	1,009,369	1,622,495	1,611,718	1,077,553	1,066,776

BIG/ISTIG STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
MP-LP	699	2,205	1,420
HP-12345-LP-6	1,686	2,648	946
HP-1234	1,701	1,301	481
2345	1,339	1,164	367
+ MPLP	805	763	381

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$1,117,432	$1,209,700	$1,349,500	$220,500	$213,000	$1,337,932	$1,330,432	$179,121	$178,117	.13387878
$283,248	$505,800	$843,000	$264,800	$141,900	$548,048	$425,148	$73,372	$56,918	
$285,768	$510,300	$850,500	$130,100	$72,150	$415,868	$357,918	$55,676	$47,918	
$224,952	$401,700	$669,500	$116,400	$55,050	$341,352	$280,002	$45,700	$37,486	
$135,240	$241,500	$402,500	$76,300	$57,150	$211,540	$192,390	$28,321	$25,757	
$360,192	$643,200	$1,072,000	$192,700	$112,200	$552,892	$472,392	$74,021	$63,243	

VALUE .115
tc per season 576000
66240kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	3c/kwh	4.5c/kwh
$179,121	$178,117	$191,473	$190,469	$210,190	$209,186	236,477	354,715
$73,372	$56,918	$103,167	$86,713	$148,311	$131,857	278,208	417,312
$55,676	$47,918	$85,736	$77,978	$131,282	$123,523	308,016	462,024
$45,700	$37,486	$69,363	$61,149	$105,215	$97,002		
$28,321	$25,757	$42,547	$39,983	$64,101	$61,537		
$74,021	$63,243	$111,909	$101,132	$169,316	$158,539	429,235	643,853

$168/M2				$300/M2				$500/M2			
ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
415,598	414,593	533,836	532,832	427,950	426,946	546,189	545,185	446,666	445,662	564,905	563,901
351,580	335,126	490,684	474,230	381,375	364,921	520,479	504,025	426,519	410,065	565,623	549,169
363,692	355,934	517,700	509,942	393,752	385,994	547,760	540,002	439,298	431,539	593,306	585,547
503,256	492,478	717,873	707,096	541,144	530,367	755,762	744,985	598,552	587,774	813,169	802,392

AUTONOMOUS DISTILLERY HEAT EXCHANGER SUMMARIES

AVERAGED STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
4ESC	2,741	1,626	697
4FCC	2,824	1,476	674
5FCC	3,367	1,217	456

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$460,488	$822,300	$1,370,500	$162,600	$104,550	$623,088	$565,038	$83,418	$75,647	.13387878
$474,432	$847,200	$1,412,000	$147,600	$101,100	$622,032	$575,532	$83,277	$77,052	
$565,656	$1,010,100	$1,683,500	$121,700	$68,400	$687,356	$634,056	$92,022	$84,887	

	.166	kwh/kg stm
	.146	
	.083	
ave		.132
tc per season		576000
		76032kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	3c/kwh	5c/kwh
$83,418	$75,647	$131,857	$124,086	$205,250	$197,478	590,769	984,614
$83,277	$77,052	$133,183	$126,957	$208,797	$202,572	595,331	992,218
$92,022	$84,887	$151,524	$144,388	$241,678	$234,542	488,125	813,542

$168/M2				$300/M2				$500/M2			
ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
674,187	666,415	1,068,033	1,060,261	722,626	714,854	1,116,472	1,108,700	796,018	788,247	1,189,864	1,182,092
678,607	672,382	1,075,494	1,069,269	728,513	722,288	1,125,400	1,119,175	804,128	797,903	1,201,015	1,194,790
580,148	573,012	905,565	898,429	639,649	632,514	965,066	957,931	729,803	722,668	1,055,220	1,048,085

BIG/ISTIG STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
4ESC	2,741	1,626	697
4FCC	2,824	1,476	674
5FCC	3,367	1,217	456

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$460,488	$822,300	$1,370,500	$162,600	$104,550	$623,088	$565,038	$83,418	$75,647	.13387878
$474,432	$847,200	$1,412,000	$147,600	$101,100	$622,032	$575,532	$83,277	$77,052	
$565,656	$1,010,100	$1,683,500	$121,700	$68,400	$687,356	$634,056	$92,022	$84,887	

	.166	kwh/kg stm
	.146	
	.083	
ave		.166
tc per season		576000
		95616kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	3c/kwh	5c/kwh
$83,418	$75,647	$131,857	$124,086	$205,250	$197,478	742,936	1,238,227
$83,277	$77,052	$133,183	$126,957	$208,797	$202,572	748,673	1,247,789
$92,022	$84,887	$151,524	$144,388	$241,678	$234,542	613,855	1,023,091

$168/M2				$300/M2				$500/M2			
ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
826,355	818,583	1,321,645	1,313,874	874,794	867,022	1,370,084	1,362,313	948,186	940,414	1,443,477	1,435,705
831,950	825,725	1,331,066	1,324,840	881,856	875,631	1,380,971	1,374,746	957,471	951,245	1,456,586	1,450,361
705,877	698,741	1,115,114	1,107,978	765,379	758,243	1,174,615	1,167,479	855,533	848,397	1,264,769	1,257,633

BIG/STIG STEAM OPPORTUNITY VALUE

configuration	evap. area, sqm	S & T QX area	P & G QX area
4ESC	2,741	1,626	697
4FCC	2,824	1,476	674
5FCC	3,367	1,217	456

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$460,488	$822,300	$1,370,500	$162,600	$104,550	$623,088	$565,038	$83,418	$75,647	.13387878
$474,432	$847,200	$1,412,000	$147,600	$101,100	$622,032	$575,532	$83,277	$77,052	
$565,656	$1,010,100	$1,683,500	$121,700	$68,400	$687,356	$634,056	$92,022	$84,887	

	.166	kwh/kg stm
	.146	
	.083	
ave		.083
tc per season		576000
		47808kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	3c/kwh	5c/kwh
$83,418	$75,647	$131,857	$124,086	$205,250	$197,478	371,468	619,114
$83,277	$77,052	$133,183	$126,957	$208,797	$202,572	374,337	623,894
$92,022	$84,887	$151,524	$144,388	$241,678	$234,542	306,927	511,546

$168/M2				$300/M2				$500/M2			
ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
454,886	447,115	702,532	694,760	503,325	495,554	750,971	743,199	576,718	568,946	824,363	816,591
457,614	451,388	707,171	700,946	507,519	501,294	757,077	750,852	583,134	576,909	832,692	826,466
398,950	391,814	603,568	596,432	458,451	451,316	663,070	655,934	548,605	541,470	753,224	746,088

CEST STEAM OPPORTUNITY COST

configuration	evap. area, sqm	S & T QX area	P & G QX area
4ESC	2,741	1,626	697
4FCC	2,824	1,476	674
5FCC	3,367	1,217	456

Disc rate: .12
life, yrs: 20

Evap cost			S & T costs	P & G costs	Total w/ S&T	Total w/ P&G	Annual w/ S&T	Annual w/ P&G	
$460,488	$822,300	$1,370,500	$162,600	$104,550	$623,088	$565,038	$83,418	$75,647	.13387878
$474,432	$847,200	$1,412,000	$147,600	$101,100	$622,032	$575,532	$83,277	$77,052	
$565,656	$1,010,100	$1,683,500	$121,700	$68,400	$687,356	$634,056	$92,022	$84,887	

	.166	kwh/kg stm
	.146	
	.083	
ave		.146
tc per season		576000
		84096kwh*tc/kg st

$168/M2		$300/M2		$500/M2		ENERGY COSTS	
Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	Annual w/ S&T	Annual w/ P&G	8c/kwh	5c/kwh
$83,418	$75,647	$131,857	$124,086	$205,250	$197,478	1,742,469	1,089,043
$83,277	$77,052	$133,183	$126,957	$208,797	$202,572	1,755,924	1,097,453
$92,022	$84,887	$151,524	$144,388	$241,678	$234,542	1,439,724	899,827

		$168/M2				$300/M2				$500/M2	
ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 3c/KWH		ELECT AT 5c/KWH		ELECT AT 8c/KWH		ELECT AT 5c/KWH	
S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G	S&T	P&G
1,825,887	1,818,116	1,172,461	1,164,690	1,874,326	1,866,555	1,220,900	1,213,129	1,947,719	1,939,947	1,294,293	1,286,521
1,839,201	1,832,976	1,180,730	1,174,504	1,889,107	1,882,882	1,230,635	1,224,410	1,964,722	1,958,496	1,306,250	1,300,025
1,531,746	1,524,610	991,850	984,714	1,591,248	1,584,112	1,051,351	1,044,215	1,681,401	1,674,266	1,141,505	1,134,369

APPENDIX VI:
BASELINE ECONOMIC SPREADSHEETS AND CALCULATION METHODOLOGIES

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BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                          M.Fulmer
Rated at 8000 tonne cane per day grinding rate; ISTIG Cogeneration;  Cogen Costs CEES Estimate     5-1-90
|==========================================================================================================|
|          SPECIFICATIONS              |               PRICES               |       COGENERATION COSTS        |
|                                      |                                    |                                 |
|    Distillery Specifications         |   Cane price, $/tonne cane:  18.87 |         Capital Costs           |
|    -------------------------         |     Bagasse price, $/tonne:  4.114 |     ----------------------      |
|  Tonnes cane milled per day:   7813  | Off season fuel price, $/GJ:    0  | Total Capital Cost,Millions:  85.624 |
|          Yield, L ethanol/TC:     70 |                                    | Capital cost,$/kw installed:     770 |
|          Season Length, days:    160 |====================================|     Discount rate (10%=.10):     .12 |
|        Yield, L ethanol/day:  492219 |                                    |           Life of facility:       30 |
|              Capacity factor:     .9 |     DISTILLERY COSTS               |                                 |
|        Electricty use, kwh/TC:    20 |                                    | Operation & Maintenance Costs   |
|        Steam use Kg(steam)/TC:   165 |        Capital Costs               | -----------------------------   |
|    kg moist bagasse/kg cane:     .3 |     ----------------------         | Fixed:                          |
|            %sugar in bagasse:    .07 |      Total Capital Cost: 9.485e7   |          Labor (Total $):  405000 |
|         %moisture in bagasse:     .5 |   Discount rate (10%=.10):   .12   |             Maintenance: 3164120 |
|         HHV bagasse, MJ/kg:  9.5082  |         Life of facility:    20    |                                 |
|                                      |                                    | Variable:                       |
|                                      | Operation & Maintenance Costs      |           In Season             |
|      Cogeneration Facility           | -----------------------------      |     Bagasse price, $/tonne:    4.11 |
|   ---------------------------        | Fixed:                             |   Processing cost, $/tonne: 11.8852 |
|                    Type:   ISTIG     |          Labor (Total $): 1437592  | Total bagasse cost, $/tonne:   16.00 |
|           Capacity, kW:  111200      |              Maintenance:       0  |   Total bagasse cost, $/GJ: 1.68267 |
|     Power in cogen mode: 101511.     |             Other/misc.: 1283285   |           Other costs $/khw:    .001 |
|        Off season fuel: barbojo      | Variable:                          |                                 |
|  % Off season running time:    .9    |            Cane cost, $/TC:  18.87 |      Fuel input in season: 3206335 |
|                                      |           Misc/other, $/TC:   2.85 |   Energy from bagasse, GJ: 3206455 |
|                                      |     Bagasse credit, $/tonne:  4.11 |  From other in seas fuel,GJ:       0 |
|                                      |        Elect. cost, $/kwh: .030013 |                                 |
|           Constants                  |        Steam cost, $/tonne:   3.84 |           Off Season            |
|   ----------------------------------  |                                   |      Off season fuel cost,      |
|     bagasse processing cost: 11.8852 |                                    |      Commodity price, $/GJ:        0 |
|            kg steam per kWh:    .128 |                                    |      Processing cost, $/GJ:     1.35 |
|        electrical efficiency,        |                                    |      Total fuel cost, $/GJ:     1.35 |
|             no process steam:   .429 |   Elect. to distillary, khw: 2.250e7 |      Other costs, $/khw:      .001 |
|        electrical efficiency,        | Steam to distillary, tonnes: 185637. |                               |
|           with process steam:   .379 |                                    |     Fuel input off season: 4131974 |
| Max steam production, kg/tc:    235  |                                    |                                 |
|      Min electrical capacity,        |    Gross $ exchanged between       |                                 |
|        with full cogen, kWh:  97400  |       distillery and cogen: -83.395 |   Average fuel cost, $/GJ: 1.49536 |
|                                      |                                    |                                 |
|==========================================================================================================|

      Annual Distillery Costs:               Annual Cogeneration Costs:
   ----------------------------------    ----------------------------------

                Capital  : 1.270e7                    Capital  : 1.063e7
          O & M                                 O & M
                  Fixed  : 2720877                      Fixed  : 3569120
         Variable Misc.  : 3206455             Variable Misc.  :  843214
              Cane Cost  : 2.123e7                Bagasse Cost : 5400076
       Electricity Cost  :  675328         Electricity Credit  :  675328
             Steam Cost  : 713146.               Steam Credit  :  713146
         Bagasse Credit  : 1388564       Off season fuel cost  : 5578165                      1388564
                           --------                              --------                           0
      Total Annual Cost  : 3.986e7          Total Annual Cost  : 2.463e7

 Cents per Liter Ethanol :   50.61              Cents per kwh  :    3.00

   Thousands of Liters/day: 492.219      Power Gererated, kwh  : 8.432e8
  Thousands of Liters/year: 78755.0       Power Exported, kwh  : 8.207e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                          M.Fulmer
Rated at 8000 tonne cane per day grinding rate; ISTIG Cogeneration;  Cogen Costs CEES Estimate     5-1-90
|==========================================================================================================|
|          SPECIFICATIONS              |               PRICES               |       COGENERATION COSTS        |
|                                      |                                    |                                 |
|    Distillery Specifications         |   Cane price, $/tonne cane:  18.87 |         Capital Costs           |
|    -------------------------         |   Bagasse price, $/tonne: 18.4238  |     ----------------------      |
|  Tonnes cane milled per day:   7813  | Off season fuel price, $/GJ:  1.505 | Total Capital Cost,Millions:  85.624 |
|          Yield, L ethanol/TC:     70 |                                    | Capital cost,$/kw installed:     770 |
|          Season Length, days:    160 |====================================|     Discount rate (10%=.10):     .12 |
|        Yield, L ethanol/day:  492219 |                                    |           Life of facility:       30 |
|              Capacity factor:     .9 |     DISTILLERY COSTS               |                                 |
|        Electricty use, kwh/TC:    20 |                                    | Operation & Maintenance Costs   |
|        Steam use Kg(steam)/TC:   165 |        Capital Costs               | -----------------------------   |
|    kg moist bagasse/kg cane:     .3 |     ----------------------         | Fixed:                          |
|            %sugar in bagasse:    .07 |      Total Capital Cost: 9.485e7   |          Labor (Total $):  405000 |
|         %moisture in bagasse:     .5 |   Discount rate (10%=.10):   .12   |             Maintenance: 3164120 |
|         HHV bagasse, MJ/kg:  9.5082  |         Life of facility:    20    |                                 |
|                                      |                                    | Variable:                       |
|                                      | Operation & Maintenance Costs      |           In Season             |
|      Cogeneration Facility           | -----------------------------      |     Bagasse price, $/tonne:   18.42 |
|   ---------------------------        | Fixed:                             |   Processing cost, $/tonne: 11.8852 |
|                    Type:   ISTIG     |          Labor (Total $): 1437592  | Total bagasse cost, $/tonne:   30.31 |
|           Capacity, kw:  111200      |              Maintenance:       0  |   Total bagasse cost, $/GJ: 3.18767 |
|     Power in cogen mode: 101511.     |             Other/misc.: 1283285   |           Other costs $/khw:    .001 |
|        Off season fuel: barbojo      | Variable:                          |                                 |
|  % Off season running time:    .9    |            Cane cost, $/TC:  18.87 |      Fuel input in season: 3206335 |
|                                      |           Misc/other, $/TC:   2.85 |   Energy from bagasse, GJ: 3206455 |
|                                      |     Bagasse credit, $/tonne: 18.42 |  From other in seas fuel,GJ:       0 |
|                                      |        Elect. cost, $/kwh: .042756 |                                 |
|           Constants                  |        Steam cost, $/tonne:   5.47 |           Off Season            |
|   ----------------------------------  |                                   |      Off season fuel cost,      |
|     bagasse processing cost: 11.8852 |                                    |      Commodity price, $/GJ:    1.505 |
|            kg steam per kWh:    .128 |                                    |      Processing cost, $/GJ:     1.35 |
|        electrical efficiency,        |                                    |      Total fuel cost, $/GJ:     2.86 |
|             no process steam:   .429 |   Elect. to distillary, khw: 2.250e7 |      Other costs, $/khw:      .001 |
|        electrical efficiency,        | Steam to distillary, tonnes: 185637. |                               |
|           with process steam:   .379 |                                    |     Fuel input off season: 4131974 |
| Max steam production, kg/tc:    235  |                                    |                                 |
|      Min electrical capacity,        |    Gross $ exchanged between       |                                 |
|        with full cogen, kWh:  97400  |       distillery and cogen: -4.24e6 |   Average fuel cost, $/GJ: 3.00036 |
|                                      |                                    |                                 |
|==========================================================================================================|

      Annual Distillery Costs:               Annual Cogeneration Costs:
   ----------------------------------    ----------------------------------

                Capital  : 1.270e7                    Capital  : 1.063e7
          O & M                                 O & M
                  Fixed  : 2720877                      Fixed  : 3569120
         Variable Misc.  : 3206455             Variable Misc.  :  843214
              Cane Cost  : 2.123e7                Bagasse Cost : 1.023e7
       Electricity Cost  :  962080         Electricity Credit  :  962080
             Steam Cost  : 1015957               Steam Credit  : 1015957                      6218444
         Bagasse Credit  : 6218444       Off season fuel cost  : 1.180e7                            0
                           --------                              --------
      Total Annual Cost  : 3.562e7          Total Annual Cost  : 3.509e7

 Cents per Liter Ethanol :   45.22              Cents per kwh  :    4.28

   Thousands of Liters/day: 492.219      Power Gererated, kwh  : 8.432e8
  Thousands of Liters/year: 78755.0       Power Exported, kwh  : 8.207e8
```

```
BAGASSE CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                          M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration;  Cogen Costs CEES Estimate  5-1-90
|===========================================================================================================|
|       SPECIFICATIONS                |              PRICES                 |      COGENERATION COSTS       |
|                                     |                                     |                               |
|  Distillery Specifications          |    Cane price, $/tonne cane:  18.87 |        Capital Costs          |
|  -------------------------          |     Bagasse price, $/tonne: 18.9181 |  ------------------------     |
| Tonnes cane milled per day:    4134 | Off season fuel price, $/GJ:   1.48 | Total Capital Cost,Millions:  52.569 |
|         Yield, L ethanol/TC:     70 |                                     | Capital cost,$/kw installed:     990 |
|        Season Length, days:    160 |=====================================|    Discount rate (10%=.10):     .12 |
|       Yield, L ethanol/day: 260442 |                                     |           Life of facility:      30 |
|             Capacity factor:     .9 |      DISTILLERY COSTS               |                               |
|      Electricty use, kwh/TC:     20 |                                     | Operation & Maintenance Costs |
|      Steam use Kg(steam)/TC:    165 |        Capital Costs                | ----------------------------- |
|   kg moist bagasse/kg cane:     .3 |     ----------------------          | Fixed:                        |
|          %sugar in bagasse:    .07 |        Total Capital Cost: 5.019e7 |          Labor (Total $):  297000 |
|       %moisture in bagasse:     .5 |    Discount rate (10%=.10):     .12 |             Maintenance: 1566845 |
|        HHV bagasse, MJ/kg:  9.5082 |          Life of facility:      20 |                               |
|                                     |                                     | Variable:                     |
|                                     | Operation & Maintenance Costs       |        In Season              |
|    Cogeneration Facility            | ----------------------------        |    Bagasse price, $/tonne:   18.92 |
|  --------------------------         | Fixed:                              |  Processing cost, $/tonne: 11.8852 |
|               Type:BIG/STIG         |          Labor (Total $):  760656 | Total bagasse cost, $/tonne:   30.80 |
|          Capacity, kw:   53100      |             Maintenance:         0 |  Total bagasse cost, $/GJ: 3.23966 |
|    Power in cogen mode: 45363.9      |             Other/misc.: 679010. |          Other costs $/khw:    .001 |
|         Off season fuel: barbojo    | Variable:                           |                               |
|  % Off season running time:     .9 |          Cane cost, $/TC:   18.87 |     Fuel input in season: 1696231 |
|                                     |          Misc/other, $/TC:   2.85 |   Energy from bagasse, GJ: 1696594 |
|                                     |   Bagasse credit, $/tonne:   18.92 |  From other in seas fuel,GJ:       0 |
|                                     |       Elect. cost, $/kwh: .052100 |                               |
|          Constants                  |      Steam cost, $/tonne:    6.04 |        Off Season             |
|  -----------------------------------|                                     |    Off season fuel cost,      |
|   bagasse processing cost: 11.8852 |                                     |    Commodity price, $/GJ:    1.48 |
|           kg steam per kWh:    .116 |                                     |    Processing cost, $/GJ:    1.35 |
|     electrical efficiency,          |                                     |    Total fuel cost, $/GJ:    2.83 |
|           no process steam:    .356 |  Elect. to distillery, khw: 1.191e7 |      Other costs, $/khw:    .001 |
|     electrical efficiency,          | Steam to distillery, tonnes: 98223.8 |                               |
|        with process steam:     .313 |                                     |    Fuel input off season: 2377687 |
| Max steam production, kg/tc:    305 |                                     |                               |
|    Min electrical capacity,         |    Gross $ exchanged between        |                               |
|      with full cogen, kWh:    38800 |       distillery and cogen: -2.16e6 |  Average fuel cost, $/GJ: 3.00059 |
|                                     |                                     |                               |
|===========================================================================================================|

      Annual Distillery Costs:                 Annual Cogeneration Costs:
  ----------------------------------      ----------------------------------

                Capital  : 6718942                       Capital  : 6526108
        O & M                                   O & M
                  Fixed  : 1439666                         Fixed  : 1863845
         Variable Misc.  : 1696594                Variable Misc.  :  391905
              Cane Cost  : 1.123e7                  Bagasse Cost  : 5501131
       Electricity Cost  :  620298            Electricity Credit  :  620298
             Steam Cost  :  593625                  Steam Credit  :  593625           3378567
         Bagasse Credit  : 3378567          Off season fuel cost  : 6728853                 0
                           --------                                 --------
      Total Annual Cost  : 1.892e7             Total Annual Cost  : 1.980e7

  Cents per Liter Ethanol  :   45.41               Cents per kwh  :    5.21

   Thousands of Liters/day: 260.442         Power Gererated, kwh  : 3.919e8
  Thousands of Liters/year: 41670.7         Power Exported, kwh   : 3.800e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                          M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration;  Cogen Costs CEES Estimate  5-1-90
|===========================================================================================================|
|       SPECIFICATIONS                |              PRICES                 |      COGENERATION COSTS       |
|                                     |                                     |                               |
|  Distillery Specifications          |    Cane price, $/tonne cane:  18.87 |        Capital Costs          |
|  -------------------------          |     Bagasse price, $/tonne: 18.9181 |  ------------------------     |
| Tonnes cane milled per day:    4134 | Off season fuel price, $/GJ:   1.48 | Total Capital Cost,Millions:  52.569 |
|         Yield, L ethanol/TC:     70 |                                     | Capital cost,$/kw installed:     990 |
|        Season Length, days:    160 |=====================================|    Discount rate (10%=.10):     .12 |
|       Yield, L ethanol/day: 260442 |                                     |           Life of facility:      30 |
|             Capacity factor:     .9 |      DISTILLERY COSTS               |                               |
|      Electricty use, kwh/TC:     20 |                                     | Operation & Maintenance Costs |
|      Steam use Kg(steam)/TC:    165 |        Capital Costs                | ----------------------------- |
|   kg moist bagasse/kg cane:     .3 |     ----------------------          | Fixed:                        |
|          %sugar in bagasse:    .07 |        Total Capital Cost: 5.019e7 |          Labor (Total $):  297000 |
|       %moisture in bagasse:     .5 |    Discount rate (10%=.10):     .12 |             Maintenance: 1566845 |
|        HHV bagasse, MJ/kg:  9.5082 |          Life of facility:      20 |                               |
|                                     |                                     | Variable:                     |
|                                     | Operation & Maintenance Costs       |        In Season              |
|    Cogeneration Facility            | ----------------------------        |    Bagasse price, $/tonne:   18.92 |
|  --------------------------         | Fixed:                              |  Processing cost, $/tonne: 11.8852 |
|               Type:BIG/STIG         |          Labor (Total $):  760656 | Total bagasse cost, $/tonne:   30.80 |
|          Capacity, kw:   53100      |             Maintenance:         0 |  Total bagasse cost, $/GJ: 3.23966 |
|    Power in cogen mode: 45363.9      |             Other/misc.: 679010. |          Other costs $/khw:    .001 |
|         Off season fuel: barbojo    | Variable:                           |                               |
|  % Off season running time:     .9 |          Cane cost, $/TC:   18.87 |     Fuel input in season: 1696231 |
|                                     |          Misc/other, $/TC:   2.85 |   Energy from bagasse, GJ: 1696594 |
|                                     |   Bagasse credit, $/tonne:   18.92 |  From other in seas fuel,GJ:       0 |
|                                     |       Elect. cost, $/kwh: .052100 |                               |
|          Constants                  |      Steam cost, $/tonne:    6.04 |        Off Season             |
|  -----------------------------------|                                     |    Off season fuel cost,      |
|   bagasse processing cost: 11.8852 |                                     |    Commodity price, $/GJ:    1.48 |
|           kg steam per kWh:    .116 |                                     |    Processing cost, $/GJ:    1.35 |
|     electrical efficiency,          |                                     |    Total fuel cost, $/GJ:    2.83 |
|           no process steam:    .356 |  Elect. to distillery, khw: 1.191e7 |      Other costs, $/khw:    .001 |
|     electrical efficiency,          | Steam to distillery, tonnes: 98223.8 |                               |
|        with process steam:     .313 |                                     |    Fuel input off season: 2377687 |
| Max steam production, kg/tc:    305 |                                     |                               |
|    Min electrical capacity,         |    Gross $ exchanged between        |                               |
|      with full cogen, kWh:    38800 |       distillery and cogen: -5.68e6 |  Average fuel cost, $/GJ: 3.00059 |
|                                     |                                     |                               |
|===========================================================================================================|

      Annual Distillery Costs:                 Annual Cogeneration Costs:
  ----------------------------------      ----------------------------------

                Capital  : 6718942                       Capital  : 6526108
        O & M                                   O & M
                  Fixed  : 1439666                         Fixed  : 1863845
         Variable Misc.  : 1696594                Variable Misc.  :  391905
              Cane Cost  : 1.123e7                  Bagasse Cost  : 5501131
       Electricity Cost  :  620298            Electricity Credit  :  620298
             Steam Cost  :  593625                  Steam Credit  :  593625               3378567
         Bagasse Credit  : 6897544          Off season fuel cost  : 6728853         3518976.37753
                           --------                                 --------
      Total Annual Cost  : 1.540e7             Total Annual Cost  : 1.980e7

  Cents per Liter Ethanol  :   36.97               Cents per kwh  :    5.21

   Thousands of Liters/day: 260.442         Power Gererated, kwh  : 3.919e8
  Thousands of Liters/year: 41670.7         Power Exported, kwh   : 3.800e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                                   M.Fulmer
Rated at 8000 tonne cane per day grinding rate; ISTIG Cogeneration;  Cogen Costs CEES Estimate              5-1-90
|==================================================================================================================|
|      SPECIFICATIONS             |              PRICES                |        COGENERATION COSTS                |
|                                 |                                    |                                          |
|   Distillery Specifications     |   Cane price, $/tonne cane:  18.87 |           Capital Costs                  |
|   -------------------------     |     Bagasse price, $/tonne: 18.4238|        ----------------------            |
|  Tonnes cane milled per day: 7813| Off season fuel price, $/GJ: 1.505| Total Capital Cost,Millions:  85.624     |
|         Yield, L ethanol/TC:   70|                                    | Capital cost,$/kw installed:     770     |
|         Season Length, days:  160|====================================|     Discount rate (10%=.10):     .12     |
|        Yield, L ethanol/day: 492219                                  |            Life of facility:      30     |
|             Capacity factor:   .9|      DISTILLERY COSTS              |                                          |
|       Electricty use, kwh/TC:  20|                                    | Operation & Maintenance Costs            |
|       Steam use Kg(steam)/TC: 165|        Capital Costs               | -----------------------------            |
|    kg moist bagasse/kg cane:  .3 |      ---------------------         | Fixed:                                   |
|           %sugar in bagasse: .07 |       Total Capital Cost: 9.485e7  |               Labor (Total $):  405000   |
|        %moisture in bagasse:  .5 |    Discount rate (10%=.10):   .12  |                  Maintenance: 3164120   |
|          HHV bagasse, MJ/kg: 9.5082         Life of facility:    20  |                                          |
|                                 |                                    | Variable:                                |
|                                 | Operation & Maintenance Costs      |            In Season                     |
|     Cogeneration Facility       | ---------------------------        |      Bagasse price, $/tonne:   18.42     |
|   -------------------------     | Fixed:                             |    Processing cost, $/tonne: 11.8852     |
|                    Type:  ISTIG |          Labor (Total $): 1437592  |  Total bagasse cost, $/tonne:   30.31     |
|            Capacity, kw: 111200 |              Maintenance:       0  |    Total bagasse cost, $/GJ: 3.18767     |
|      Power in cogen mode: 101511.|             Other/misc.: 1283285  |            Other costs $/khw:    .001     |
|         Off season fuel: barbojo | Variable:                          |                                          |
|  % Off season running time:   .9|            Cane cost, $/TC:  18.  |       Fuel input in season: 3206335      |
|                                 |           Misc/other, $/TC:   2.  |      gy from bagasse, GJ: 3206455        |
|                                 |      Bagasse credit, $/tonne: 18. |      her in sea fuel,GJ:        0         |
|                                 |           Elect. cost, $/kwh: .0427 6                                          |
|          Constants              |         Steam cost, $/tonne:  5.47 |       Off Seaso                          |
|  ------------------------------ |                                    |     Off season fuel cost,                |
|   bagasse processing cost: 11.8852                                   |     Commodity price, $/GJ:   1.505       |
|          kg steam per kWh:  .128 |                                   |     Processing cost, $/GJ:    1.35       |
|     electrical efficiency,      |                                    |     Total fuel cost, $/GJ:    2.86       |
|          no process steam:  .429| Elect. to distillary, khw: 2.250e7 |       Other costs, $/khw:     .001       |
|     electrical efficiency,      | Steam to distillary, tonnes: 185637.                                          |
|        with process steam:  .379|                                    |     Fuel input off season: 4131974       |
| Max steam production, kg/tc: 235|                                    |                                          |
|   Min electrical capacity,      |   Gross $ exchanged between        |                                          |
|     with full cogen, kWh: 97400 |     distillery and cogen: -1.05e7  |   Average fuel cost, $/GJ: 3.00036       |
|                                 |                                    |                                          |
|==================================================================================================================|

     Annual Distillery Costs:              Annual Cogeneration Costs:
    -----------------------------         -----------------------------

               Capital  : 1.270e7                    Capital  : 1.063e7
        O & M                               O & M
                 Fixed  : 2720877                      Fixed  : 3569120
        Variable Misc.  : 3206455            Variable Misc.   :  843214
             Cane Cost  : 2.123e7               Bagasse Cost  : 1.023e7
      Electricity Cost  :  962080         Electricity Credit  :  962080
            Steam Cost  : 1015957               Steam Credit  : 1015957
        Bagasse Credit  : 1.244e7        Off season fuel cost : 1.180e7                       6218444
                          --------                              --------               6218621.26993
     Total Annual Cost  : 2.940e7          Total Annual Cost  : 3.509e7

 Cents per Liter Ethanol  :  37.33             Cents per kwh  :    4.28

  Thousands of Liters/day: 492.219        Power Gererated, kwh  : 8.432e8
 Thousands of Liters/year: 78755.0        Power Exported, kwh   : 8.207e8
```

```
BAGASSE CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                                   M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration;  Cogen Costs CEES Estimate           5-1-90
|==================================================================================================================|
|      SPECIFICATIONS             |              PRICES                |        COGENERATION COSTS                |
|                                 |                                    |                                          |
|   Distillery Specifications     |   Cane price, $/tonne cane:  18.87 |           Capital Costs                  |
|   -------------------------     |     Bagasse price, $/tonne:  4.846 |        ----------------------            |
|  Tonnes cane milled per day: 4134| Off season fuel price, $/GJ:     0| Total Capital Cost,Millions:  52.569     |
|         Yield, L ethanol/TC:   70|                                    | Capital cost,$/kw installed:     990     |
|         Season Length, days:  160|====================================|     Discount rate (10%=.10):     .12     |
|        Yield, L ethanol/day: 260442                                  |            Life of facility:      30     |
|             Capacity factor:   .9|      DISTILLERY COSTS              |                                          |
|       Electricty use, kwh/TC:  20|                                    | Operation & Maintenance Costs            |
|       Steam use Kg(steam)/TC: 165|        Capital Costs               | -----------------------------            |
|    kg moist bagasse/kg cane:  .3 |      ---------------------         | Fixed:                                   |
|           %sugar in bagasse: .07 |       Total Capital Cost: 5.019e7  |               Labor (Total $):  297000   |
|        %moisture in bagasse:  .5 |    Discount rate (10%=.10):   .12  |                  Maintenance: 1566845   |
|          HHV bagasse, MJ/kg: 9.5082         Life of facility:    20  |                                          |
|                                 |                                    | Variable:                                |
|                                 | Operation & Maintenance Costs      |            In Season                     |
|     Cogeneration Facility       | ---------------------------        |      Bagasse price, $/tonne:    4.85     |
|   -------------------------     | Fixed:                             |    Processing cost, $/tonne: 11.8852     |
|                 Type:BIG/STIG   |          Labor (Total $):  760656  |  Total bagasse cost, $/tonne:   16.73     |
|            Capacity, kw:  53100 |              Maintenance:       0  |    Total bagasse cost, $/GJ: 1.75966     |
|      Power in cogen mode: 45363.9|             Other/misc.: 679010.  |            Other costs $/khw:    .001     |
|         Off season fuel: barbojo | Variable:                          |                                          |
|  % Off season running time:   .9|            Cane cost, $/TC: 18.87 |       Fuel input in season: 1696231      |
|                                 |           Misc/other, $/TC:  2.85 |    Energy from bagasse, GJ: 1696594      |
|                                 |      Bagasse credit, $/tonne: 4.85 |  From other in seas fuel,GJ:      0     |
|                                 |           Elect. cost, $/kwh: .037143                                         |
|          Constants              |         Steam cost, $/tonne:  4.31 |       Off Season                         |
|  ------------------------------ |                                    |     Off season fuel cost,                |
|   bagasse processing cost: 11.8852                                   |     Commodity price, $/GJ:      0        |
|          kg steam per kWh:  .116 |                                   |     Processing cost, $/GJ:    1.35       |
|     electrical efficiency,      |                                    |     Total fuel cost, $/GJ:    1.35       |
|          no process steam:  .356| Elect. to distillery, khw: 1.191e7 |       Other costs, $/khw:     .001       |
|     electrical efficiency,      | Steam to distillery, tonnes: 98223.9                                          |
|        with process steam:  .313|                                    |     Fuel input off season: 2377687       |
| Max steam production, kg/tc: 305|                                    |                                          |
|   Min electrical capacity,      |   Gross $ exchanged between        |                                          |
|     with full cogen, kWh: 38800 |     distillery and cogen: -12.381  |   Average fuel cost, $/GJ: 1.52059       |
|                                 |                                    |                                          |
|==================================================================================================================|

     Annual Distillery Costs:              Annual Cogeneration Costs:
    -----------------------------         -----------------------------

               Capital  : 6718942                    Capital  : 6526108
        O & M                               O & M
                 Fixed  : 1439666                      Fixed  : 1863845
        Variable Misc.  : 1696594            Variable Misc.   :  391905
             Cane Cost  : 1.123e7               Bagasse Cost  : 2988005
      Electricity Cost  :  442222         Electricity Credit  :  442222
            Steam Cost  :  423207               Steam Credit  :  423207                      865441
        Bagasse Credit  : 865441.        Off season fuel cost : 3209877                           0
                          --------                              --------
     Total Annual Cost  : 2.109e7          Total Annual Cost  : 1.411e7

 Cents per Liter Ethanol  :  50.61             Cents per kwh  :    3.71

  Thousands of Liters/day: 260.442        Power Gererated, kwh  : 3.919e8
 Thousands of Liters/year: 41670.7        Power Exported, kwh   : 3.800e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$               M.Fulmer
Rated at 4000 tonne cane per day grinding rate; CEST Cogeneration;  Cogen Costs CEES Estimate      3-29-90
|=======================================================================================================================|
|      SPECIFICATIONS                 |              PRICES                    |     COGENERATION COSTS                 |
|                                     |                                        |                                        |
|  Distillery Specifications          |     Cane price, $/tonne cane:     8.07 |         Capital Costs                  |
|  --------------------------         |       Bagasse price, $/tonne:    6.964 |     ----------------------             |
|  Tonnes cane milled per day:     4048 |  Off season fuel price, $/GJ:        0 |   Total Capital Cost,Millions:   42.012 |
|         Yield, L ethanol/TC:       73 |                                        |   Capital cost,$/kw installed:     1556 |
|         Season Length, days:      160 |========================================|       Discount rate (10%=.10):      .12 |
|        Yield, L ethanol/day:   265954 |                                        |              Life of facility:       30 |
|             Capacity factor:       .9 |       DISTILLERY COSTS                 |                                        |
|      Electricty use, kwh/TC:       20 |                                        | Operation & Maintenance Costs          |
|      Steam use Kg(steam)/TC:      165 |          Capital Costs                 | ----------------------------           |
|    kg moist bagasse/kg cane:       .3 |      ----------------------            | Fixed:                                 |
|           %sugar in bagasse:      .07 |           Total Capital Cost:  1.830e7 |               Labor (Total $):   129600 |
|        %moisture in bagasse:       .5 |      Discount rate (10%=.10):      .12 |                   Maintenance:   874060 |
|          HHV bagasse, MJ/kg:   9.5082 |             Life of facility:       20 |                                        |
|                                     |                                        | Variable:                              |
|                                     | Operation & Maintenance Costs          |           In Season                    |
|    Cogeneration Facility            | ----------------------------           |        Bagasse price, $/tonne:     6.96 |
| ----------------------------        | Fixed:                                 |      Processing cost, $/tonne:        0 |
|                        Type:     CEST |             Labor (Total $):        0 |   Total bagasse cost, $/tonne:     6.96 |
|                Capacity, kw:    27000 |                 Maintenance:        0 |      Total bagasse cost, $/GJ:  .732420 |
|         Power in cogen mode:  22960.1 |                 Other/misc.:  1059564 |             Other costs $/khw:     .003 |
|             Off season fuel:  barbojo | Variable:                              |                                        |
|   % Off season running time:       .9 |             Cane cost, $/TC:     8.07 |          Fuel input in season:  1661236 |
|                                     |            Misc/other, $/TC:     .148 |       Energy from bagasse, GJ:  1661299 |
|                                     |     Bagasse credit, $/tonne:     6.96 |    From other in seas fuel,GJ:        0 |
|                                     |          Elect. cost, $/kwh:  .047384 |                                        |
|          Constants                  |         Steam cost, $/tonne:     6.92 |           Off Season                   |
| ------------------------------------ |                                        |        Off season fuel cost,           |
|     bagasse processing cost:        0 |                                        |         Commodity price, $/GJ:        0 |
|            kg steam per kWh:     .146 |                                        |         Processing cost, $/GJ:      .97 |
|       electrical efficiency,        |                                        |         Total fuel cost, $/GJ:      .97 |
|            no process steam:     .203 |   Elect. to distillery, khw:  1.166e7 |            Other costs, $/khw:     .003 |
|       electrical efficiency,        | Steam to distillery, tonnes:  96180.5 |                                        |
|          with process steam:      .13 |                                        |         Fuel input off season:  2120205 |
| Max steam production, kg/tc:      388 |                                        |                                        |
|     Min electrical capacity,        |   Gross $ exchanged between            |                                        |
|        with full cogen, kWh:    17500 |        distillery and cogen:  -25.646 |       Average fuel cost, $/GJ:  .865626 |
|                                     |                                        |                                        |
|=======================================================================================================================|

     Annual Distillery Costs:             Annual Cogeneration Costs:
  ------------------------------------   ------------------------------------

                 Capital   : 2449575                  Capital    : 5215523
          O & M                                 O & M
                   Fixed   : 1059564                    Fixed    : 1003660
           Variable Misc.  :   86271           Variable Misc.    :  596718
               Cane Cost   : 4704100             Bagasse Cost    : 1217820
        Electricity Cost   :  552413       Electricity Credit    :  552413
              Steam Cost   :  665381             Steam Credit    :  665381                   1217820
          Bagasse Credit   : 1217820     Off season fuel cost    : 2056599                         0
                             --------                              --------
       Total Annual Cost   : 8299484        Total Annual Cost    : 8872526

  Cents per Liter Ethanol  :   19.50            Cents per kwh    :    4.74

   Thousands of Liters/day: 265.954      Power Gererated, kwh    : 1.989e8
  Thousands of Liters/year: 42552.6       Power Exported, kwn    : 1.872e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$               M.Fulmer
Rated at 4000 tonne cane per day grinding rate; CEST Cogeneration;  Cogen Costs CEES Estimate      3-29-90
|=======================================================================================================================|
|      SPECIFICATIONS                 |              PRICES                    |     COGENERATION COSTS                 |
|                                     |                                        |                                        |
|  Distillery Specifications          |     Cane price, $/tonne cane:     8.07 |         Capital Costs                  |
|  --------------------------         |       Bagasse price, $/tonne:  27.2640 |     ----------------------             |
|  Tonnes cane milled per day:     4048 |  Off season fuel price, $/GJ:    2.135 |   Total Capital Cost,Millions:   42.012 |
|         Yield, L ethanol/TC:       73 |                                        |   Capital cost,$/kw installed:     1556 |
|         Season Length, days:      160 |========================================|       Discount rate (10%=.10):      .12 |
|        Yield, L ethanol/day:   265954 |                                        |              Life of facility:       30 |
|             Capacity factor:       .9 |       DISTILLERY COSTS                 |                                        |
|      Electricty use, kwh/TC:       20 |                                        | Operation & Maintenance Costs          |
|      Steam use Kg(steam)/TC:      165 |          Capital Costs                 | ----------------------------           |
|    kg moist bagasse/kg cane:       .3 |      ----------------------            | Fixed:                                 |
|           %sugar in bagasse:      .07 |           Total Capital Cost:  1.830e7 |               Labor (Total $):   129600 |
|        %moisture in bagasse:       .5 |      Discount rate (10%=.10):      .12 |                   Maintenance:   874060 |
|          HHV bagasse, MJ/kg:   9.5082 |             Life of facility:       20 |                                        |
|                                     |                                        | Variable:                              |
|                                     | Operation & Maintenance Costs          |           In Season                    |
|    Cogeneration Facility            | ----------------------------           |        Bagasse price, $/tonne:    27.26 |
| ----------------------------        | Fixed:                                 |      Processing cost, $/tonne:        0 |
|                        Type:     CEST |             Labor (Total $):        0 |   Total bagasse cost, $/tonne:    27.26 |
|                Capacity, kw:    27000 |                 Maintenance:        0 |      Total bagasse cost, $/GJ:  2.86742 |
|         Power in cogen mode:  22960.1 |                 Other/misc.:  1059564 |             Other costs $/khw:     .003 |
|             Off season fuel:  barbojo | Variable:                              |                                        |
|   % Off season running time:       .9 |             Cane cost, $/TC:     8.07 |          Fuel input in season:  1661236 |
|                                     |            Misc/other, $/TC:     .148 |       Energy from bagasse, GJ:  1661299 |
|                                     |     Bagasse credit, $/tonne:    27.26 |    From other in seas fuel,GJ:        0 |
|                                     |          Elect. cost, $/kwh:  .085311 |                                        |
|          Constants                  |         Steam cost, $/tonne:    12.46 |           Off Season                   |
| ------------------------------------ |                                        |        Off season fuel cost,           |
|     bagasse processing cost:        0 |                                        |         Commodity price, $/GJ:    2.135 |
|            kg steam per kWh:     .146 |                                        |         Processing cost, $/GJ:      .97 |
|       electrical efficiency,        |                                        |         Total fuel cost, $/GJ:     3.11 |
|            no process steam:     .203 |   Elect. to distillery, khw:  1.166e7 |            Other costs, $/khw:     .003 |
|       electrical efficiency,        | Steam to distillery, tonnes:  96180.5 |                                        |
|          with process steam:      .13 |                                        |         Fuel input off season:  2120205 |
| Max steam production, kg/tc:      388 |                                        |                                        |
|     Min electrical capacity,        |   Gross $ exchanged between            |                                        |
|        with full cogen, kWh:    17500 |        distillery and cogen:  -2.58e6 |       Average fuel cost, $/GJ:  3.00063 |
|                                     |                                        |                                        |
|=======================================================================================================================|

     Annual Distillery Costs:             Annual Cogeneration Costs:
  ------------------------------------   ------------------------------------

                 Capital   : 2449575                  Capital    : 5215523
          O & M                                 O & M
                   Fixed   : 1059564                    Fixed    : 1003660
           Variable Misc.  :   86271           Variable Misc.    :  596718
               Cane Cost   : 4704100             Bagasse Cost    : 4767755
        Electricity Cost   :  994579       Electricity Credit    :  994579
              Steam Cost   : 1197971             Steam Credit    : 1197971                   4767755
          Bagasse Credit   : 4767755     Off season fuel cost    : 6583236                         0
                             --------                              --------
       Total Annual Cost   : 5724305        Total Annual Cost    : 1.597e7

  Cents per Liter Ethanol  :   13.45            Cents per kwh    :    8.53

   Thousands of Liters/day: 265.954      Power Gererated, kwh    : 1.989e8
  Thousands of Liters/year: 42552.6       Power Exported, kwh    : 1.872e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$                  M.Fulmer
Rated at 4000 tonne cane per day grinding rate; CEST Cogeneration;  Cogen Costs CEES Estimate        5-1-90
|=========================================================================================================|
|       SPECIFICATIONS                  |              PRICES                |        COGENERATION COSTS  |
|                                       |                                    |                            |
|  Distillery Specifications            |   Cane price, $/tonne cane:   8.07 |       Capital Costs        |
|  -------------------------            |     Bagasse price, $/tonne: 27.2640|  ------------------------  |
| Tonnes cane milled per day:     4048  | Off season fuel price, $/GJ:  2.135| Total Capital Cost,Millions:  42.012 |
|        Yield, L ethanol/TC:       73  |                                    | Capital cost,$/kw installed:    1556 |
|        Season Length, days:      160  |====================================|     Discount rate (10%=.10):     .12 |
|      Yield, L ethanol/day:    265954  |                                    |           Life of facility:       30 |
|            Capacity factor:       .9  |       DISTILLERY COSTS             |                            |
|      Electricty use, kwh/TC:      20  |                                    | Operation & Maintenance Costs |
|      Steam use Kg(steam)/TC:     165  |         Capital Costs              | ----------------------------- |
|   kg moist bagasse/kg cane:       .3  |     ----------------------         | Fixed:                     |
|           %sugar in bagasse:     .07  |      Total Capital Cost: 1.830e7   |           Labor (Total $):  129600 |
|        %moisture in bagasse:      .5  |    Discount rate (10%=.10):    .12 |              Maintenance:  874060 |
|           HHV bagasse, MJ/kg: 9.5082  |          Life of facility:     20  |                            |
|                                       |                                    | Variable:                  |
|                                       | Operation & Maintenance Costs      |           In Season        |
|     Cogeneration Facility             | ----------------------------       |    Bagasse price, $/tonne:    27.26 |
|  ---------------------------          | Fixed:                             |   Processing cost, $/tonne:       0 |
|                       Type:    CEST   |             Labor (Total $):     0 | Total bagasse cost, $/tonne:   27.26 |
|              Capacity, kw:   27000    |                 Maintenance:     0 |   Total bagasse cost, $/GJ: 2.86742 |
|        Power in cogen mode: 22960.1   |                 Other/misc.: 1059564 |          Other costs $/khw:    .003 |
|           Off season fuel: barbojo    | Variable:                          |                            |
|  % Off season running time:     .9    |              Cane cost, $/TC:  8.07 |      Fuel input in season: 1661236 |
|                                       |             Misc/other, $/TC:  .148 |   Energy from bagasse, GJ: 1661299 |
|                                       |       Bagasse credit, $/tonne: 27.26 |  From other in seas fuel,GJ:      0 |
|                                       |           Elect. cost, $/kwh: .085311 |                            |
|            Constants                  |           Steam cost, $/tonne: 12.46 |         Off Season         |
|  ------------------------------------ |                                    |      Off season fuel cost, |
|    bagasse processing cost:       0   |                                    |     Commodity price, $/GJ:   2.135 |
|            kg steam per kWh:    .146  |                                    |     Processing cost, $/GJ:     .97 |
|      electrical efficiency,           |                                    |     Total fuel cost, $/GJ:    3.11 |
|          no process steam:     .203   |  Elect. to distillery, khw: 1.166e7 |        Other costs, $/khw:    .003 |
|      electrical efficiency,           | Steam to distillery, tonnes: 96180.5 |                            |
|         with process steam:     .13   |                                    |     Fuel input off season: 2120205 |
| Max steam production, kg/tc:    388   |                                    |                            |
|    Min electrical capacity,           |    Gross $ exchanged between       |                            |
|       with full cogen, kWh:   17500   |       distillery and cogen: -7.10e6 |   Average fuel cost, $/GJ: 3.00063 |
|                                       |                                    |                            |
|=========================================================================================================|

    Annual Distillery Costs:                Annual Cogeneration Costs:
   ------------------------------------    ------------------------------------

               Capital  : 2449575                      Capital  : 5215523
         O & M                                   O & M
                 Fixed  : 1059564                        Fixed  : 1003663
        Variable Misc.  :   86271              Variable Misc.   :  596718
             Cane Cost  : 4704100                Bagasse Cost   : 4767755
      Electricity Cost  :  994579             Electricity Credit :  994579
            Steam Cost  : 1197971                 Steam Credit  : 1197971                  4767755
        Bagasse Credit  : 9294393           Off season fuel cost : 6583236           4526637.51724
                          --------                                --------
      Total Annual Cost : 1197667               Total Annual Cost : 1.597e7

  Cents per Liter Ethanol :    2.81                 Cents per kwh  :    8.53

    Thousands of Liters/day: 265.954         Power Gererated, kwh : 1.989e8
   Thousands of Liters/year: 42552.6         Power Exported, kwh  : 1.872e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$                  M.Fulmer
Rated at 8000 tonne cane per day grinding rate; ISTIG Cogeneration;  Cogen Costs CEES Estimate       5-1-90
|=========================================================================================================|
|       SPECIFICATIONS                  |              PRICES                |        COGENERATION COSTS  |
|                                       |                                    |                            |
|  Distillery Specifications            |   Cane price, $/tonne cane:   8.07 |       Capital Costs        |
|  -------------------------            |     Bagasse price, $/tonne:  4.114 |  ------------------------  |
| Tonnes cane milled per day:     7813  | Off season fuel price, $/GJ:     0 | Total Capital Cost,Millions:  85.624 |
|        Yield, L ethanol/TC:       73  |                                    | Capital cost,$/kw installed:     770 |
|        Season Length, days:      160  |====================================|     Discount rate (10%=.10):     .12 |
|      Yield, L ethanol/day:    513314  |                                    |           Life of facility:       30 |
|            Capacity factor:       .9  |       DISTILLERY COSTS             |                            |
|      Electricty use, kwh/TC:      20  |                                    | Operation & Maintenance Costs |
|      Steam use Kg(steam)/TC:     165  |         Capital Costs              | ----------------------------- |
|   kg moist bagasse/kg cane:       .3  |     ----------------------         | Fixed:                     |
|           %sugar in bagasse:     .07  |      Total Capital Cost: 3.531e7   |           Labor (Total $):  405000 |
|        %moisture in bagasse:      .5  |    Discount rate (10%=.10):    .12 |              Maintenance: 3164120 |
|           HHV bagasse, MJ/kg: 9.5082  |          Life of facility:     20  |                            |
|                                       |                                    | Variable:                  |
|                                       | Operation & Maintenance Costs      |           In Season        |
|     Cogeneration Facility             | ----------------------------       |    Bagasse price, $/tonne:     4.11 |
|  ---------------------------          | Fixed:                             |   Processing cost, $/tonne: 11.8852 |
|                       Type:   ISTIG   |             Labor (Total $):     0 | Total bagasse cost, $/tonne:   16.00 |
|              Capacity, kw:  111200    |                 Maintenance:     0 |   Total bagasse cost, $/GJ: 1.68267 |
|        Power in cogen mode: 101511.   |                 Other/misc.: 2045053 |          Other costs $/khw:    .001 |
|           Off season fuel: barbojo    | Variable:                          |                            |
|  % Off season running time:     .9    |              Cane cost, $/TC:  8.07 |      Fuel input in season: 3206335 |
|                                       |             Misc/other, $/TC:  .148 |   Energy from bagasse, GJ: 3206455 |
|                                       |       Bagasse credit, $/tonne:  4.11 |  From other in seas fuel,GJ:      0 |
|                                       |           Elect. cost, $/kwh: .030013 |                            |
|            Constants                  |           Steam cost, $/tonne:  3.84 |         Off Season         |
|  ------------------------------------ |                                    |      Off season fuel cost, |
|    bagasse processing cost: 11.8852   |                                    |     Commodity price, $/GJ:       0 |
|            kg steam per kWh:    .128  |                                    |     Processing cost, $/GJ:    1.35 |
|      electrical efficiency,           |                                    |     Total fuel cost, $/GJ:    1.35 |
|          no process steam:     .429   |  Elect. to distillary, khw: 2.250e7 |        Other costs, $/khw:    .001 |
|      electrical efficiency,           | Steam to distillary, tonnes: 185637. |                            |
|         with process steam:     .379  |                                    |     Fuel input off season: 4131974 |
| Max steam production, kg/tc:    235   |                                    |                            |
|    Min electrical capacity,           |    Gross $ exchanged between       |                            |
|       with full cogen, kWh:   97400   |       distillery and cogen: -89.395 |   Average fuel cost, $/GJ: 1.49536 |
|                                       |                                    |                            |
|=========================================================================================================|

    Annual Distillery Costs:                Annual Cogeneration Costs:
   ------------------------------------    ------------------------------------

               Capital  : 4727897                      Capital  : 1.063e7
         O & M                                   O & M
                 Fixed  : 2045053                        Fixed  : 3569120
        Variable Misc.  :  166511              Variable Misc.   :  843214
             Cane Cost  : 9079331                Bagasse Cost   : 5400076
      Electricity Cost  :  675328             Electricity Credit :  675328
            Steam Cost  : 713146.                 Steam Credit  :  713146                  1388564
        Bagasse Credit  : 1388564           Off season fuel cost : 5578165                        0
                          --------                                --------
      Total Annual Cost : 1.602e7               Total Annual Cost : 2.463e7

  Cents per Liter Ethanol :   19.50                 Cents per kwh  :    3.00

    Thousands of Litors/day: 513.314         Power Gererated, kwh : 8.432e8
   Thousands of Liters/year: 82130.3         Power Exported, kwh  : 8.207e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$              M.Fulmer
Rated at 8000 tonne cane per day grinding rate; ISTIG Cogeneration;  Cogen Costs CEES Estimate   5-1-90
|-------------------------------------------------------------------------------------------------------|
|      SPECIFICATIONS                 |               PRICES                |    COGENERATION COSTS     |
|                                     |                                     |                           |
|   Distillery Specifications         |   Cane price, $/tonne cane:   8.07  |       Capital Costs       |
|   --------------------------        |     Bagasse price, $/tonne: 18.4238 |   ----------------------  |
| Tonnes cane milled per day:   7813  | Off season fuel price, $/GJ:  1.505 | Total Capital Cost,Millions:  85.624 |
|        Yield, L ethanol/TC:     73  |                                     | Capital cost,$/kw installed:     770 |
|       Season Length, days:    160  |=====================================|    Discount rate (10%=.10):      .12 |
|      Yield, L ethanol/day: 513314  |                                     |           Life of facility:       30 |
|           Capacity factor:     .9  |     DISTILLERY COSTS                |                                      |
|     Electricty use, kwh/TC:     20  |                                     | Operation & Maintenance Costs        |
|     Steam use Kg(steam)/TC:    165  |       Capital Costs                 | ----------------------------         |
|   kg moist bagasse/kg cane:     .3  |     ----------------------          | Fixed:                               |
|          %sugar in bagasse:    .07  |        Total Capital Cost: 3.531e7  |          Labor (Total $):  405000    |
|        %moisture in bagasse:    .5  |    Discount rate (10%=.10):     .12  |               Maintenance: 3164120   |
|         HHV bagasse, MJ/kg: 9.5082  |           Life of facility:      20  |                                      |
|                                     |                                     | Variable:                            |
|                                     | Operation & Maintenance Costs       |          In Season                   |
|    Cogeneration Facility            | ----------------------------        |     Bagasse price, $/tonne:   18.42  |
|  --------------------------         | Fixed:                              |   Processing cost, $/tonne: 11.8852  |
|                     Type:  ISTIG    |           Labor (Total $):       0  | Total bagasse cost, $/tonne:  30.31  |
|             Capacity, kw:  111200   |               Maintenance:       0  |   Total bagasse cost, $/GJ: 3.18767  |
|       Power in cogen mode: 101511.  |              Other/misc.: 2045053   |          Other costs $/khw:    .001  |
|           Off season fuel: barbojo  | Variable:                           |                                      |
|  % Off season running time:     .9  |            Cane cost, $/TC:   8.07  |       Fuel input in season: 3206335  |
|                                     |           Misc/other, $/TC:   .148  |    Energy from bagasse, GJ: 3206455  |
|                                     |   Bagasse credit, $/tonne:   18.42  |  From other in seas fuel,GJ:      0  |
|                                     |       Elect. cost, $/kwh: .042756  |                                      |
|         Constants                   |       Steam cost, $/tonne:    5.47  |          Off Season                  |
|  ----------------------------------  |                                     |       Off season fuel cost,          |
|    bagasse processing cost: 11.8852 |                                     |     Commodity price, $/GJ:    1.505  |
|         kg steam per kWh:    .128   |                                     |     Processing cost, $/GJ:     1.35  |
|     electrical efficiency,          |                                     |     Total fuel cost, $/GJ:     2.86  |
|          no process steam:   .429   |   Elect. to distillary, khw: 2.250e7|        Other costs, $/khw:     .001  |
|     electrical efficiency,          | Steam to distillary, tonnes: 185637.|                                      |
|        with process steam:   .379   |                                     |     Fuel input off season: 4131974   |
| Max steam production, kg/tc:   235   |                                     |                                      |
|     Min electrical capacity,        |    Gross $ exchanged between        |                                      |
|      with full cogen, kWh:   97400  |       distillery and cogen: -4.24e6 |   Average fuel cost, $/GJ: 3.00036   |
|                                     |                                     |                                      |
|=======================================================================================================|

      Annual Distillery Costs:               Annual Cogeneration Costs:
  -----------------------------------     -------------------------------------

              Capital  : 4727897                     Capital  : 1.063e7
        O & M                                 O & M
                Fixed  : 2045053                       Fixed  : 3569120
       Variable Misc.  :  166511               Variable Misc.  :  843214
            Cane Cost  : 9079331                 Bagasse Cost  : 1.023e7
     Electricity Cost  :  962080           Electricity Credit  :  962080
           Steam Cost  : 1015957                 Steam Credit  : 1015957
       Bagasse Credit  : 6218444          Off season fuel cost  : 1.180e7          6218444
                         --------                                --------               0
   Total Annual Cost   : 1.178e7             Total Annual Cost  : 3.509e7

 Cents per Liter Ethanol  :   14.34              Cents per kwh  :    4.28

  Thousands of Liters/day: 513.314         Power Gererated, kwh  : 8.432e8
 Thousands of Liters/year: 82130.3          Power Exported, kwh  : 8.207e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$              M.Fulmer
Rated at 8000 tonne cane per day grinding rate; ISTIG Cogeneration;  Cogen Costs CEES Estimate   5-1-90
|-------------------------------------------------------------------------------------------------------|
|      SPECIFICATIONS                 |               PRICES                |    COGENERATION COSTS     |
|                                     |                                     |                           |
|   Distillery Specifications         |   Cane price, $/tonne cane:   8.07  |       Capital Costs       |
|   --------------------------        |     Bagasse price, $/tonne: 18.4238 |   ----------------------  |
| Tonnes cane milled per day:   7813  | Off season fuel price, $/GJ:  1.505 | Total Capital Cost,Millions:  85.624 |
|        Yield, L ethanol/TC:     73  |                                     | Capital cost,$/kw installed:     770 |
|       Season Length, days:    160  |=====================================|    Discount rate (10%=.10):      .12 |
|      Yield, L ethanol/day: 513314  |                                     |           Life of facility:       30 |
|           Capacity factor:     .9  |     DISTILLERY COSTS                |                                      |
|     Electricty use, kwh/TC:     20  |                                     | Operation & Maintenance Costs        |
|     Steam use Kg(steam)/TC:    165  |       Capital Costs                 | ----------------------------         |
|   kg moist bagasse/kg cane:     .3  |     ----------------------          | Fixed:                               |
|          %sugar in bagasse:    .07  |        Total Capital Cost: 3.531e7  |          Labor (Total $):  405000    |
|        %moisture in bagasse:    .5  |    Discount rate (10%=.10):     .12  |               Maintenance: 3164120   |
|         HHV bagasse, MJ/kg: 9.5082  |           Life of facility:      20  |                                      |
|                                     |                                     | Variable:                            |
|                                     | Operation & Maintenance Costs       |          In Season                   |
|    Cogeneration Facility            | ----------------------------        |     Bagasse price, $/tonne:   18.42  |
|  --------------------------         | Fixed:                              |   Processing cost, $/tonne: 11.8852  |
|                     Type:  ISTIG    |           Labor (Total $):       0  | Total bagasse cost, $/tonne:  30.31  |
|             Capacity, kw:  111200   |               Maintenance:       0  |   Total bagasse cost, $/GJ: 3.18767  |
|       Power in cogen mode: 101511.  |              Other/misc.: 2045053   |          Other costs $/khw:    .001  |
|           Off season fuel: barbojo  | Variable:                           |                                      |
|  % Off season running time:     .9  |            Cane cost, $/TC:   8.07  |       Fuel input in season: 3206335  |
|                                     |           Misc/other, $/TC:   .148  |    Energy from bagasse, GJ: 3206455  |
|                                     |   Bagasse credit, $/tonne:   18.42  |  From other in seas fuel,GJ:      0  |
|                                     |       Elect. cost, $/kwh: .042756  |                                      |
|         Constants                   |       Steam cost, $/tonne:    5.47  |          Off Season                  |
|  ----------------------------------  |                                     |       Off season fuel cost,          |
|    bagasse processing cost: 11.8852 |                                     |     Commodity price, $/GJ:    1.505  |
|         kg steam per kWh:    .128   |                                     |     Processing cost, $/GJ:     1.35  |
|     electrical efficiency,          |                                     |     Total fuel cost, $/GJ:     2.86  |
|          no process steam:   .429   |   Elect. to distillary, khw: 2.250e7|        Other costs, $/khw:     .001  |
|     electrical efficiency,          | Steam to distillary, tonnes: 185637.|                                      |
|        with process steam:   .379   |                                     |     Fuel input off season: 4131974   |
| Max steam production, kg/tc:   235   |                                     |                                      |
|     Min electrical capacity,        |    Gross $ exchanged between        |                                      |
|      with full cogen, kWh:   97400  |       distillery and cogen: -1.05e7 |   Average fuel cost, $/GJ: 3.00036   |
|                                     |                                     |                                      |
|=======================================================================================================|

      Annual Distillery Costs:               Annual Cogeneration Costs:
  -----------------------------------     -------------------------------------

              Capital  : 4727897                     Capital  : 1.063e7
        O & M                                 O & M
                Fixed  : 2045053                       Fixed  : 3569120
       Variable Misc.  :  166511               Variable Misc.  :  843214
            Cane Cost  : 9079331                 Bagasse Cost  : 1.023e7
     Electricity Cost  :  962080           Electricity Credit  :  962080
           Steam Cost  : 1015957                 Steam Credit  : 1015957          6218444
       Bagasse Credit  : 1.244e7          Off season fuel cost  : 1.180e7     6218621.26993
                         --------                                --------
   Total Annual Cost   : 5559763             Total Annual Cost  : 3.509e7

 Cents per Liter Ethanol  :    6.77              Cents per kwh  :    4.28

  Thousands of Liters/day: 513.314         Power Gererated, kwh  : 8.432e8
 Thousands of Liters/year: 82130.3          Power Exported, kwh  : 8.207e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$                M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration;  Cogen Costs CEES Estimate  5-1-90
|==========================================================================================================|
|        SPECIFICATIONS                |                PRICES                 |      COGENERATION COSTS         |
|                                      |                                       |                                 |
|   Distillery Specifications          |   Cane price, $/tonne cane:    8.07   |         Capital Costs           |
|   -------------------------          |      Bagasse price, $/tonne:   4.845  |    ------------------------     |
|  Tonnes cane milled per day:   4130  | Off season fuel price, $/GJ:     0    | Total Capital Cost,Millions:  52.569 |
|         Yield, L ethanol/TC:     73  |                                       | Capital cost,$/kw installed:     990 |
|         Season Length, days:    160  |=======================================|    Discount rate (10%=.10):    .12 |
|        Yield, L ethanol/day:  271341 |                                       |            Life of facility:     30 |
|             Capacity factor:     .9  |      DISTILLERY COSTS                 |                                 |
|      Electricty use, kwh/TC:     20  |                                       | Operation & Maintenance Costs   |
|      Steam use Kg(steam)/TC:    165  |         Capital Costs                 | ----------------------------    |
|   kg moist bagasse/kg cane:      .3  |      ----------------------           | Fixed:                          |
|           %sugar in bagasse:    .07  |        Total Capital Cost: 1.867e7    |        Labor (Total $):  297000 |
|        %moisture in bagasse:     .5  |    Discount rate (10%=.10):     .12   |            Maintenance: 1566845 |
|         HHV bagasse, MJ/kg:  9.5082  |         Life of facility:       20    |                                 |
|                                      |                                       | Variable:                       |
|                                      | Operation & Maintenance Costs         |         In Season               |
|    Cogeneration Facility             | ----------------------------          |    Bagasse price, $/tonne:   4.85 |
|  --------------------------          | Fixed:                                |  Processing cost, $/tonne: 11.8852 |
|                  Type:BIG/STIG       |          Labor (Total $):         0   | Total bagasse cost, $/tonne:  16.73 |
|          Capacity, kw:   53100       |              Maintenance:         0   |   Total bagasse cost, $/GJ: 1.75955 |
|      Power in cogen mode: 45363.9    |              Other/misc.:   1081028   |        Other costs $/khw:    .001 |
|          Off season fuel: barbojo    | Variable:                             |                                 |
|   % Off season running time:    .9   |            Cane cost, $/TC:   8.07    |    Fuel input in season: 1696231 |
|                                      |           Misc/other, $/TC:   .148    |  Energy from bagasse, GJ: 1696415 |
|                                      |     Bagasse credit, $/tonne:   4.85   | From other in seas fuel,GJ:      0 |
|                                      |        Elect. cost, $/kwh: .037136    |                                 |
|        Constants                     |       Steam cost, $/tonne:    4.31    |         Off Season              |
|   ----------------------------------  |                                       |    Off season fuel cost,        |
|     bagasse processing cost: 11.8852 |                                       |    Commodity price, $/GJ:      0 |
|           kg steam per kWh:    .116  |                                       |    Processing cost, $/GJ:   1.35 |
|       electrical efficiency,         |                                       |    Total fuel cost, $/GJ:   1.35 |
|           no process steam:    .356  |  Elect. to distillery, khw: 1.189e7   |     Other costs, $/khw:     .001 |
|       electrical efficiency,         | Steam to distillary, tonnes: 98128.8  |                                 |
|          with process steam:   .313  |                                       |   Fuel input off season: 2377687 |
| Max steam production, kg/tc:    305  |                                       |                                 |
|      Min electrical capacity,        |   Gross $ exchanged between           |                                 |
|        with full cogen, kWh:  38800  |      distillery and cogen: 12.5104    |  Average fuel cost, $/GJ: 1.52053 |
|                                      |                                       |                                 |
|==========================================================================================================|

     Annual Distillery Costs:               Annual Cogeneration Costs:
   ------------------------------------    ------------------------------------

              Capital  : 2499196                      Capital  : 6526109
        O & M                                O & M
                Fixed  : 1081028                        Fixed  : 1863845
       Variable Misc.  :   88019               Variable Misc.  :  391905
           Cane Cost   : 4799390                 Bagasse Cost  : 2984935
     Electricity Cost  :  441716           Electricity Credit  :  441716
          Steam Cost   :  422722                Steam Credit   :  422722                    864426
      Bagasse Credit   : 864426.          Off season fuel cost : 3209877                         0
                         --------                                --------
     Total Annual Cost : 8467644             Total Annual Cost : 1.411e7

  Cents per Liter Ethanol  :  19.50              Cents per kwh  :    3.71

    Thousands of Liters/day: 271.341       Power Cererated, kwh  : 3.919e8
   Thousands of Liters/year: 43414.6        Power Exported, kwh  : 3.800e8
```

```
BAGASSE FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$                M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration;  Cogen Costs CEES Estimate  5-1-90
|==========================================================================================================|
|        SPECIFICATIONS                |                PRICES                 |      COGENERATION COSTS         |
|                                      |                                       |                                 |
|   Distillery Specifications          |   Cane price, $/tonne cane:    8.07   |         Capital Costs           |
|   -------------------------          |      Bagasse price, $/tonne: 18.9171  |    ------------------------     |
|  Tonnes cane milled per day:   4130  | Off season fuel price, $/GJ:   1.48   | Total Capital Cost,Millions:  52.569 |
|         Yield, L ethanol/TC:     73  |                                       | Capital cost,$/kw installed:     990 |
|         Season Length, days:    160  |=======================================|    Discount rate (10%=.10):    .12 |
|        Yield, L ethanol/day:  271341 |                                       |            Life of facility:     30 |
|             Capacity factor:     .9  |      DISTILLERY COSTS                 |                                 |
|      Electricty use, kwh/TC:     20  |                                       | Operation & Maintenance Costs   |
|      Steam use Kg(steam)/TC:    165  |         Capital Costs                 | ----------------------------    |
|   kg moist bagasse/kg cane:      .3  |      ----------------------           | Fixed:                          |
|           %sugar in bagasse:    .07  |        Total Capital Cost: 1.867e7    |        Labor (Total $):  297000 |
|        %moisture in bagasse:     .5  |    Discount rate (10%=.10):     .12   |            Maintenance: 1566845 |
|         HHV bagasse, MJ/kg:  9.5082  |         Life of facility:       20    |                                 |
|                                      |                                       | Variable:                       |
|                                      | Operation & Maintenance Costs         |         In Season               |
|    Cogeneration Facility             | ----------------------------          |    Bagasse price, $/tonne:  18.92 |
|  --------------------------          | Fixed:                                |  Processing cost, $/tonne: 11.8852 |
|                  Type:BIG/STIG       |          Labor (Total $):         0   | Total bagasse cost, $/tonne:  30.80 |
|          Capacity, kw:   53100       |              Maintenance:         0   |   Total bagasse cost, $/GJ: 3.23955 |
|      Power in cogen mode: 45363.9    |              Other/misc.:   1081028   |        Other costs $/khw:    .001 |
|          Off season fuel: barbojo    | Variable:                             |                                 |
|   % Off season running time:    .9   |            Cane cost, $/TC:   8.07    |    Fuel input in season: 1696231 |
|                                      |           Misc/other, $/TC:   .148    |  Energy from bagasse, GJ: 1696415 |
|                                      |     Bagasse credit, $/tonne:  18.92   | From other in seas fuel,GJ:      0 |
|                                      |        Elect. cost, $/kwh: .052088    |                                 |
|        Constants                     |       Steam cost, $/tonne:    6.04    |         Off Season              |
|   ----------------------------------  |                                       |    Off season fuel cost,        |
|     bagasse processing cost: 11.8852 |                                       |    Commodity price, $/GJ:   1.48 |
|           kg steam per kWh:    .116  |                                       |    Processing cost, $/GJ:   1.35 |
|       electrical efficiency,         |                                       |    Total fuel cost, $/GJ:   2.83 |
|           no process steam:    .356  |  Elect. to distillery, khw: 1.189e7   |     Other costs, $/khw:     .001 |
|       electrical efficiency,         | Steam to distillary, tonnes: 98128.8  |                                 |
|          with process steam:   .313  |                                       |   Fuel input off season: 2377687 |
| Max steam production, kg/tc:    305  |                                       |                                 |
|      Min electrical capacity,        |   Gross $ exchanged between           |                                 |
|        with full cogen, kWh:  38800  |      distillery and cogen: -2.16e6    |  Average fuel cost, $/GJ: 3.00053 |
|                                      |                                       |                                 |
|==========================================================================================================|

     Annual Distillery Costs:               Annual Cogeneration Costs:
   ------------------------------------    ------------------------------------

              Capital  : 2499196                      Capital  : 6526108
        O & M                                O & M
                Fixed  : 1081028                        Fixed  : 1863845
       Variable Misc.  :   88019               Variable Misc.  :  391905
           Cane Cost   : 4799390                 Bagasse Cost  : 5495630
     Electricity Cost  :  619553           Electricity Credit  :  619553
          Steam Cost   :  592912                Steam Credit   :  592912                    3375120
      Bagasse Credit   : 3375120          Off season fuel cost : 6728853                          0
                         --------                                --------
     Total Annual Cost : 6304977             Total Annual Cost : 1.979e7

  Cents per Liter Ethanol  :  14.52              Cents per kwh  :    5.21

    Thousands of Liters/day: 271.341       Power Cererated, kwh  : 3.919e8
   Thousands of Liters/year: 43414.6        Power Exported, kwh  : 3.800e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Goldemberg Estimates, adjusted to 1987 US$                      M.Fulmer
Rated at 4000 tonne cane per day grinding rate; BIG/STIG Cogeneration;  Cogen Costs CEES Estimate       5-1-90
|=====================================================================================================================|
|     SPECIFICATIONS                     |               PRICES                 |          COGENERATION COSTS          |
|                                        |                                      |                                      |
|  Distillery Specifications             |   Cane price, $/tonne cane:   8.07   |           Capital Costs              |
|  -------------------------             |    Bagasse price, $/tonne: 18.9171   |        ----------------------        |
| Tonnes cane milled per day:    4130    | Off season fuel price, $/GJ:  1.48   | Total Capital Cost,Millions:  52.569 |
|        Yield, L ethanol/TC:      73    |                                      | Capital cost,$/kw installed:    990  |
|        Season Length, days:     160    |======================================|     Discount rate (10%=.10):    .12  |
|      Yield, L ethanol/day:  271341    |                                      |           Life of facility:      30  |
|           Capacity factor:      .9    |     DISTILLERY COSTS                 |                                      |
|      Electricty use, kwh/TC:    20    |                                      | Operation & Maintenance Costs        |
|      Steam use Kg(steam)/TC:   165    |        Capital Costs                 | ----------------------------         |
|  kg moist bagasse/kg cane:      .3    |      --------------------            | Fixed:                               |
|         %sugar in bagasse:     .07    |        Total Capital Cost: 1.867e7   |             Labor (Total $):  297000 |
|      %moisture in bagasse:      .5    |    Discount rate (10%=.10):    .12   |                 Maintenance: 1566845 |
|        HHV bagasse, MJ/kg:  9.5082    |           Life of facility:     20   |                                      |
|                                        |                                      | Variable:                            |
|                                        | Operation & Maintenance Costs        |            In Season                 |
|    Cogeneration Facility               | ----------------------------         |      Bagasse price, $/tonne:   18.92 |
|  ---------------------------           | Fixed:                               |   Processing cost, $/tonne: 11.8852 |
|                    Type:BIG/STIG       |            Labor (Total $):      0   | Total bagasse cost, $/tonne:   30.80 |
|           Capacity, kw:   53100        |                Maintenance:      0   |    Total bagasse cost, $/GJ: 3.23955 |
|     Power in cogen mode: 45363.9       |               Other/misc.: 1081028   |          Other costs $/khw:     .001 |
|         Off season fuel: barbojo       | Variable:                            |                                      |
|  % Off season running time:      .9    |            Cane cost, $/TC:   8.07   |      Fuel input in season: 1696231   |
|                                        |          Misc/other, $/TC:    .148   |    Energy from bagasse, GJ: 1696415  |
|                                        |    Bagasse credit, $/tonne:  18.92   |  From other in seas fuel,GJ:       0 |
|                                        |        Elect. cost, $/kwh: .052088   |                                      |
|        Constants                       |       Steam cost, $/tonne:    6.04   |          Off Season                  |
|  ------------------------------------  |                                      |        Off season fuel cost,         |
|     bagasse processing cost: 11.8852   |                                      |      Commodity price, $/GJ:     1.48 |
|            kg steam per kWh:    .116   |                                      |      Processing cost, $/GJ:     1.35 |
|       electrical efficiency,           |                                      |      Total fuel cost, $/GJ:     2.83 |
|            no process steam:    .356   |  Elect. to distillery, khw: 1.189e7  |        Other costs, $/khw:      .001 |
|       electrical efficiency,           | Steam to distillary, tonnes: 98128.8 |                                      |
|          with process steam:    .313   |                                      |      Fuel input off season: 2377687  |
| Max steam production, kg/tc:     305   |                                      |                                      |
|    Min electrical capacity,            |    Gross $ exchanged between         |                                      |
|      with full cogen, kWh:     38800   |       distillery and cogen: -5.68e6  |    Average fuel cost, $/GJ: 3.00053  |
|                                        |                                      |                                      |
|=====================================================================================================================|

      Annual Distillery Costs:                 Annual Cogeneration Costs:
  -----------------------------------      ------------------------------------

              Capital  : 2499196                      Capital  : 6526108
        O & M                                   O & M
                Fixed  : 1081028                        Fixed  : 1863845
       Variable Misc.  :   88019            Variable Misc.  :  391905
           Cane Cost   : 4799390                 Bagasse Cost  : 5495630
     Electricity Cost  :  619553           Electricity Credit  :  619553
           Steam Cost  :  592912                 Steam Credit  :  592912
       Bagasse Credit  : 6894096         Off season fuel cost  : 6728853                          3375120
                         --------                                --------                  3518976.37753
    Total Annual Cost  : 2786000            Total Annual Cost  : 1.979e7

 Cents per Liter Ethanol  :    6.42              Cents per kwh  :    5.21

   Thousands of Liters/day: 271.341       Power Gererated, kwh  : 3.919e8
  Thousands of Liters/year: 43414.6        Power Exported, kwh  : 3.800e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogenoration Costs: Schaeffer Estimates, in 1987 US$                               M.Fulmer
Rated at 4000 tonne cane per day grinding rate; CEST Cogeneration;  Cogen Costs CEES Estimate          5-1-90
|=====================================================================================================================|
|     SPECIFICATIONS                     |               PRICES                 |          COGENERATION COSTS          |
|                                        |                                      |                                      |
|  Distillery Specifications             |   Cane price, $/tonne cane:  18.87   |           Capital Costs              |
|  -------------------------             |    Bagasse price, $/tonne:   6.964   |        ----------------------        |
| Tonnes cane milled per day:    4048    | Off season fuel price, $/GJ:     0   | Total Capital Cost,Millions:  42.012 |
|        Yield, L ethanol/TC:      70    |                                      | Capital cost,$/kw installed:   1556  |
|        Season Length, days:     160    |======================================|     Discount rate (10%=.10):    .12  |
|      Yield, L ethanol/day:  255024    |                                      |           Life of facility:      30  |
|           Capacity factor:      .9    |     DISTILLERY COSTS                 |                                      |
|      Electricty use, kwh/TC:    20    |                                      | Operation & Maintenance Costs        |
|      Steam use Kg(steam)/TC:   165    |        Capital Costs                 | ----------------------------         |
|  kg moist bagasse/kg cane:      .3    |      --------------------            | Fixed:                               |
|         %sugar in bagasse:     .07    |        Total Capital Cost: 4.914e7   |             Labor (Total $):  129600 |
|      %moisture in bagasse:      .5    |    Discount rate (10%=.10):    .12   |                 Maintenance:  874060 |
|        HHV bagasse, MJ/kg:  9.5082    |           Life of facility:     20   |                                      |
|                                        |                                      | Variable:                            |
|                                        | Operation & Maintenance Costs        |            In Season                 |
|    Cogeneration Facility               | ----------------------------         |      Bagasse price, $/tonne:    6.96 |
|  ---------------------------           | Fixed:                               |   Processing cost, $/tonne:        0 |
|                    Type:    CEST       |            Labor (Total $):  744832  | Total bagasse cost, $/tonne:    6.96 |
|           Capacity, kw:   27000        |                Maintenance:       0  |    Total bagasse cost, $/GJ: .732420 |
|     Power in cogen mode: 22960.1       |               Other/misc.:  664884  |          Other costs $/khw:     .003 |
|         Off season fuel: barbojo       | Variable:                            |                                      |
|  % Off season running time:      .9    |            Cane cost, $/TC:  18.87   |      Fuel input in season: 1661236   |
|                                        |          Misc/other, $/TC:    2.85   |    Energy from bagasse, GJ: 1661299  |
|                                        |    Bagasse credit, $/tonne:   6.96   |  From other in seas fuel,GJ:       0 |
|                                        |        Elect. cost, $/kwh: .047384   |                                      |
|        Constants                       |       Steam cost, $/tonne:    6.92   |          Off Season                  |
|  ------------------------------------  |                                      |        Off season fuel cost,         |
|     bagasse processing cost:       0   |                                      |      Commodity price, $/GJ:        0 |
|            kg steam per kWh:    .146   |                                      |      Processing cost, $/GJ:      .97 |
|       electrical efficiency,           |                                      |      Total fuel cost, $/GJ:      .97 |
|            no process steam:    .203   |  Elect. to distillery, khw: 1.166e7  |        Other costs, $/khw:      .003 |
|       electrical efficiency,           | Steam to distillery, tonnes: 96180.5 |                                      |
|          with process steam:     .13   |                                      |      Fuel input off season: 2120205  |
| Max steam production, kg/tc:     388   |                                      |                                      |
|    Min electrical capacity,            |    Gross $ exchanged between         |                                      |
|      with full cogen, kWh:     17500   |       distillery and cogen: -25.646  |    Average fuel cost, $/GJ: .865626  |
|                                        |                                      |                                      |
|=====================================================================================================================|

      Annual Distillery Costs:                 Annual Cogeneration Costs:
  -----------------------------------      ------------------------------------

              Capital  : 6579167                      Capital  : 5215523
        O & M                                   O & M
                Fixed  : 1409716                        Fixed  : 1003660
       Variable Misc.  : 1661299            Variable Misc.  :  596718
           Cane Cost   : 1.100e7                 Bagasse Cost  : 1217820
     Electricity Cost  :  552413           Electricity Credit  :  552413
           Steam Cost  :  665381                 Steam Credit  :  665381                          1217820
       Bagasse Credit  : 1217820         Off season fuel cost  : 2056599                                0
                         --------                                --------
    Total Annual Cost  : 2.065e7            Total Annual Cost  : 8872526

 Cents per Liter Ethanol  :   50.61              Cents per kwh  :    4.74

   Thousands of Liters/day: 255.024       Power Gererated, kwh  : 1.989e8
  Thousands of Liters/year: 40803.8        Power Exported, kwh  : 1.872e8
```

```
ed at 4000 tonne cane per day grinding rate; CEST Cogeneration;  Cogen Costs CEES Estimate            5-1-90
==========================================================================================================|
    SPECIFICATIONS                 |             PRICES                  |       COGENERATION COSTS              |
                                   |                                     |                                       |
istillery Specifications           |   Cane price, $/tonne cane:   18.87 |          Capital Costs                |
-------------------------          |     Bagasse price, $/tonne: 27.2640 |       ----------------------          |
nnes cane milled per day:    4048  | Off season fuel price, $/GJ:  2.135 | Total Capital Cost,Millions:  42.012  |
      Yield, L ethanol/TC:     70  |                                     | Capital cost,$/kw installed:    1556  |
       Season Length, days:   160  |=====================================|     Discount rate (10%=.10):     .12  |
     Yield, L ethanol/day: 255024  |                                     |            Life of facility:      30  |
          Capacity factor:     .9  |       DISTILLERY COSTS              |                                       |
   Electricty use, kwh/TC:     20  |                                     | Operation & Maintenance Costs         |
   Steam use Kg(steam)/TC:    165  |         Capital Costs               | ----------------------------          |
g moist bagasse/kg cane:       .3  |       ----------------------        | Fixed:                                |
        %sugar in bagasse:    .07  |        Total Capital Cost: 4.914e7  |              Labor (Total $):  129600 |
     %moisture in bagasse:     .5  |   Discount rate (10%=.10):     .12  |                 Maintenance:  874060 |
      HHV bagasse, MJ/kg:  9.5082  |          Life of facility:      20  |                                       |
                                   |                                     | Variable:                             |
                                   | Operation & Maintenance Costs       |            In Season                  |
Cogeneration Facility              | ----------------------------        |    Bagasse price, $/tonne:    27.26   |
-------------------------          | Fixed:                              |  Processing cost, $/tonne:        0   |
                     Type:   CEST  |           Labor (Total $):  744832  | Total bagasse cost, $/tonne:  27.26   |
            Capacity, kw:   27000  |               Maintenance:       0  |   Total bagasse cost, $/GJ: 2.86742   |
     Power in cogen mode: 22960.1  |               Other/misc.:  664884  |          Other costs $/khw:    .003   |
        Off season fuel: barbojo  | Variable:                           |                                       |
Off season running time:      .9  |           Cane cost, $/TC:   18.87  |      Fuel input in season: 1661236    |
                                   |          Misc/other, $/TC:    2.85  |   Energy from bagasse, GJ: 1661299    |
                                   |    Bagasse credit, $/tonne:   27.26 |  From other in seas fuel,GJ:       0   |
                                   |        Elect. cost, $/kwh: .085311  |                                       |
         Constants                 |       Steam cost, $/tonne:   12.46  |           Off Season                  |
---------------------------------  |                                     |      Off season fuel cost,            |
bagasse processing cost:        0  |                                     |      Commodity price, $/GJ:   2.135   |
       kg steam per kWh:     .146  |                                     |      Processing cost, $/GJ:     .97   |
  electrical efficiency,           |                                     |      Total fuel cost, $/GJ:    3.11   |
       no process steam:     .203  |   Elect. to distillery, khw: 1.166e7|         Other costs, $/khw:    .003   |
  electrical efficiency,           |  Steam to distillery, tonnes: 96180.5                                        |
     with process steam:      .13  |                                     |      Fuel input off season: 2120205   |
steam production, kg/tc:     388   |                                     |                                       |
Min electrical capacity,           |    Gross $ exchanged between        |                                       |
   with full cogen, kWh:    17500  |       distillery and cogen: -2.58e6 |    Average fuel cost, $/GJ: 3.00063   |
                                   |                                     |                                       |
==========================================================================================================

Annual Distillery Costs:              Annual Cogeneration Costs:
-----------------------------------   -------------------------------------

          Capital  : 6579167                      Capital  : 5215523
    O & M                                   O & M
            Fixed  : 1409716                        Fixed  : 1003660
    Variable Misc. : 1661299                Variable Misc. :  596718
        Cane Cost  : 1.100e7                  Bagasse Cost : 4767755
 Electricity Cost  :  994579            Electricity Credit :  994579
       Steam Cost  : 1197971                  Steam Credit : 1197971                    4767755
   Bagasse Credit  : 4767755           Off season fuel cost : 6583236                         0
                     --------                               --------
 Total Annual Cost : 1.807e7             Total Annual Cost : 1.597e7

ents per Liter Ethanol  :  44.30            Cents per kwh  :    8.53

Thousands of Liters/day: 255.024      Power Gererated, kwh  : 1.989e8
housands of Liters/year: 40803.8      Power Exported, kwh   : 1.872e8
```

```
ALL FUEL CREDITS PAID TO DISTILLERY
Ethanol Distillery & Cogeneration Costs: Schaeffer Estimates, in 1987 US$                      M.Fulmer
Rated at 4000 tonne cane per day grinding rate; CEST Cogeneration;  Cogen Costs CEES Estimate  5-1-90
|===========================================================================================================|
|      SPECIFICATIONS                 |             PRICES                  |       COGENERATION COSTS              |
|                                     |                                     |                                       |
|   Distillery Specifications         |   Cane price, $/tonne cane:   18.87 |          Capital Costs                |
|   -------------------------         |     Bagasse price, $/tonne: 27.2640 |       ----------------------          |
|  Tonnes cane milled per day:  4048  | Off season fuel price, $/GJ:  2.135 | Total Capital Cost,Millions:  42.012  |
|        Yield, L ethanol/TC:     70  |                                     | Capital cost,$/kw installed:    1556  |
|        Season Length, days:    160  |=====================================|     Discount rate (10%=.10):     .12  |
|       Yield, L ethanol/day: 255024  |                                     |            Life of facility:      30  |
|            Capacity factor:     .9  |       DISTILLERY COSTS              |                                       |
|     Electricty use, kwh/TC:     20  |                                     | Operation & Maintenance Costs         |
|     Steam use Kg(steam)/TC:    165  |         Capital Costs               | ----------------------------          |
|  kg moist bagasse/kg cane:      .3  |       ----------------------        | Fixed:                                |
|          %sugar in bagasse:    .07  |        Total Capital Cost: 4.914e7  |              Labor (Total $):  129600 |
|       %moisture in bagasse:     .5  |   Discount rate (10%=.10):     .12  |                 Maintenance:  874060 |
|        HHV bagasse, MJ/kg:  9.5082  |          Life of facility:      20  |                                       |
|                                     |                                     | Variable:                             |
|                                     | Operation & Maintenance Costs       |            In Season                  |
|   Cogeneration Facility             | ----------------------------        |    Bagasse price, $/tonne:    27.26   |
|  -------------------------          | Fixed:                              |  Processing cost, $/tonne:        0   |
|                       Type:   CEST  |           Labor (Total $):  744832  | Total bagasse cost, $/tonne:  27.26   |
|              Capacity, kw:   27000  |               Maintenance:       0  |   Total bagasse cost, $/GJ: 2.86742   |
|       Power in cogen mode: 22960.1  |               Other/misc.:  664884  |          Other costs $/khw:    .003   |
|          Off season fuel: barbojo  | Variable:                           |                                       |
|  % Off season running time:     .9  |           Cane cost, $/TC:   18.87  |      Fuel input in season: 1661236    |
|                                     |          Misc/other, $/TC:    2.85  |   Energy from bagasse, GJ: 1661299    |
|                                     |    Bagasse credit, $/tonne:   27.26 |  From other in seas fuel,GJ:       0   |
|                                     |        Elect. cost, $/kwh: .085311  |                                       |
|          Constants                  |       Steam cost, $/tonne:   12.46  |           Off Season                  |
|  ---------------------------------  |                                     |      Off season fuel cost,            |
|    bagasse processing cost:      0  |                                     |      Commodity price, $/GJ:   2.135   |
|           kg steam per kWh:   .146  |                                     |      Processing cost, $/GJ:     .97   |
|     electrical efficiency,          |                                     |      Total fuel cost, $/GJ:    3.11   |
|           no process steam:   .203  |   Elect. to distillery, khw: 1.166e7|         Other costs, $/khw:    .003   |
|     electrical efficiency,          |  Steam to distillery, tonnes: 96180.5                                        |
|         with process steam:    .13  |                                     |      Fuel input off season: 2120205   |
| Max steam production, kg/tc:   388  |                                     |                                       |
|    Min electrical capacity,         |    Gross $ exchanged between        |                                       |
|       with full cogen, kWh:  17500  |       distillery and cogen: -7.10e6 |    Average fuel cost, $/GJ: 3.00063   |
|                                     |                                     |                                       |
|===========================================================================================================|

     Annual Distillery Costs:              Annual Cogeneration Costs:
  -----------------------------------   -------------------------------------

               Capital  : 6579167                      Capital  : 5215523
         O & M                                   O & M
                 Fixed  : 1409716                        Fixed  : 1003660
         Variable Misc. : 1661299                Variable Misc. :  596718
             Cane Cost  : 1.100e7                  Bagasse Cost : 4767755
      Electricity Cost  :  994579            Electricity Credit :  994579
            Steam Cost  : 1197971                  Steam Credit : 1197971                    4767755
        Bagasse Credit  : 9294393           Off season fuel cost : 6583236              4526637.51724
                          --------                               --------
      Total Annual Cost : 1.355e7             Total Annual Cost : 1.597e7

  Cents per Liter Ethanol  :  33.20              Cents per kwh  :    8.53

    Thousands of Liters/day: 255.024       Power Gererated, kwh  : 1.989e8
   Thousands of Liters/year: 40803.8       Power Exported, kwh   : 1.872e8
```

```
BAGASSE CREDITS ONLY TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input                        Fulmer
updated values for molasses and sugar yields                                                             4-4-90
BIG/ISTIG COGEN   C MOLASSES TO DISTILLERY
|============================================================================================================================|
|           SPECIFICATIONS                |                 PRICES                 |              SUGAR FACTORY              |
|                                         |                                        |                                         |
|            Sugar Factory                |   Cane Price, $ / tonne cane:     8.07 |              Capital Costs              |
|   ---------------------------           |       Bagasse price, $/tonne:     5.14 |             --------------              |
|  Capacity, Tonnes milled/day:     7806  |  Off season fuel price, $/GJ:      .00 |             Total Capital Cost: 8.587e7 |
|          Season Length, days:      160  |           Molasses price,$/T:    40.00 |        Discount rate (10%=.10):     .12 |
|     Average percent capacity:       .9  |  % fuel credit to distillery:   .04353 |               Life of facility:      20 |
|           Yield, kg sugar/TC:      108  |                                        |                                         |
|     Yield, kg  A molasses/TC:       30  |                                        | Operation & Maintenance Costs           |
|     kg moist bagasse/kg cane:       .3  |      average fuel cost, $/GJ:  1.54269 | ----------------------------            |
|            %sugar in bagasse:      .07  |========================================| Fixed:                                  |
|         %moisture in bagasse:       .5  |                                        |                Labor (Total $):     .00 |
|           HHV bagasse, MJ/kg:   9.5082  |          COGENERATION COSTS            |                    Maintenance:     .00 |
|       Electricty use, kwh/TC:       20  |                                        |                    Other/misc.: 1992482 |
|    HP Steam use Kg(steam)/TC:      235  |             Capital Costs              | Variable:                               |
|                                         |             --------------             |                Cane cost, $/TC:    8.07 |
|             Distillery                  |  Total Capital Cost,Millions:   85.624 |               Misc/other, $/TC:    4.17 |
|   ---------------------------           |  Capital cost,$/kw installed:      770 |        Bagasse credit, $/tonne:    5.14 |
|       Electricty use, kwh/L :      .05  |      Discount rate (10%=.10):      .12 |             Elect. cost, $/kwh: .030538 |
|    HP Steam use Kg(steam)/L :        0  |             Life of facility:       30 |               Steam cost, $/kg: .003909 |
|    LP Steam use Kg(steam)/L :      1.5  |                                        |                                         |
|   Liters ethanol/kg molasses:     .303  | Operation & Maintenance Costs          |=========================================|
|          Ethanol yield, L/tc:     9.09  | ----------------------------           |               DISTILLERY                |
|        Cogeneration Facility            | Fixed:                                 |                                         |
|   ---------------------------           |              Labor (Total $):   405000 |              Capital Costs              |
|                         Type:    ISTIG  |               Maintenance, $:  3164120 |             --------------              |
|      Full power capacity, kW:   111200  |                Other/misc, $:        0 |             Total Capital Cost: 5854500 |
|           Cogen capacity, kW:    97400  | Variable:                              |        Discount rate (10%=.10):     .12 |
|              Off season fuel:  barbojo  |     Bagasse Processing, $/TB:  11.8852 |               Life of facility:      20 |
|    % Off season running time:       .9  |  Total bagasse cost, $/tonne:    17.03 |                                         |
|                 kg steam/kWh:     .128  |     Total bagasse cost, $/GJ:     1.79 | Operation & Maintenance Costs           |
|        Electrical efficiency,           |   Off seas. fuel proc., $/GJ:     1.35 | ----------------------------            |
|             no process steam:     .429  |   Off season fuel cost, $/GJ:     1.35 | Fixed:                                  |
|        Electrical efficiency,           |           Other costs, $/khw:     .001 |                Labor (Total $):  117090 |
|           with process steam:     .379  |                                        |                    Maintenance:  117090 |
|    Elect. Cap. in full Cogen:    97400  |  Fuel required in season, GJ:  3197395 |                    Other/misc.: 29272.5 |
|  Max steam production, kg/TC:      235  |  Fuel input from bagasse, GJ:  3206348 | Variable:                               |
|                                         |  From other in season source:        0 |         Molasses cost, $/tonne:      40 |
|   Elect. to sugar fact., kwh:  2.248e7  |    Fuel input off season, GJ:  4131974 |          Electricy cost, $/kwh: .030538 |
|    HP Steam to sugar fact, T:  264155.  |                                        |               Steam cost, $/kg: .003909 |
|    Elect. to distillery, khw:  510887.  |        Net fuel rents to s&a:  -8.4006 |               Misc/other, $/L :    .038 |
|    LP Steam to distillery, T:  15326.6  |           percent to alcohol:  .097529 |                                         |
|============================================================================================================================|

  Cost Summaries                 -7.5813                                  -.81930
========================================================================================================================

                  Annual Sugar Costs                  Annual Distillery Costs                Annual Cogeneration Costs
------------------------------------------------------------------------------------------------------------------------

                     Capital : 1.150e7                        Capital :  783793                        Capital : 1.063e7
                        O & M                         O & M                                     O & M
                       Fixed : 1992482                          Fixed :  263453                          Fixed : 3569120
              Variable Misc. : 4687347                 Variable Misc. :  388274                 Variable Misc. :  829008
                   Cane Cost : 9071196                  Molasses Cost : 1348877                   Bagasse Cost : 5742573
            Electricity Cost :  686527               Electricity Cost :   15601            Electricity Credits :  702128
                  Steam Cost :  972627                     Steam Cost :   59909                  Steam Credits : 1032536
              Bagasse Credit : 1659146                 Bagasse Credit :   75510           Off season fuel cost : 5578165
            Molassess Credit : 1348877
                              --------                                 --------                                 --------
           Total Annual Cost : 2.590e7              Total Annual Cost : 2784396              Total Annual Cost : 2.461e7

          Cents per Kg Sugar :   21.33        Cents per Liter Ethanol :   27.25                  Cents per kwh :    3.05

        Tonnes sugar per day : 758.743         Thousands of Liters/day: 63.8609           Power Generated, kwh : 8.290e8
       Tonnes sugar per year : 121399.        Thousands of Liters/year: 10217.7            Power Exported, kwh : 8.060e8
```

```
BAGASSE CREDITS ONLY TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input                        Fulmer
updated values for molasses and sugar yields                                                             4-4-90
BIG/ISTIG COGEN   C MOLASSES TO DISTILLERY
|============================================================================================================================|
|           SPECIFICATIONS                |                 PRICES                 |              SUGAR FACTORY              |
|                                         |                                        |                                         |
|            Sugar Factory                |   Cane Price, $ / tonne cane:     8.07 |              Capital Costs              |
|   ---------------------------           |       Bagasse price, $/tonne:    19.00 |             --------------              |
|  Capacity, Tonnes milled/day:     7806  |  Off season fuel price, $/GJ:     1.46 |             Total Capital Cost: 8.587e7 |
|          Season Length, days:      160  |           Molasses price,$/T:    40.00 |        Discount rate (10%=.10):     .12 |
|     Average percent capacity:       .9  |  % fuel credit to distillery:   .01657 |               Life of facility:      20 |
|           Yield, kg sugar/TC:      108  |                                        |                                         |
|     Yield, kg  A molasses/TC:       30  |                                        | Operation & Maintenance Costs           |
|     kg moist bagasse/kg cane:       .3  |      average fuel cost, $/GJ:  2.99969 | ----------------------------            |
|            %sugar in bagasse:      .07  |========================================| Fixed:                                  |
|         %moisture in bagasse:       .5  |                                        |                Labor (Total $):     .00 |
|           HHV bagasse, MJ/kg:   9.5082  |          COGENERATION COSTS            |                    Maintenance:     .00 |
|       Electricty use, kwh/TC:       20  |                                        |                    Other/misc.: 1992482 |
|    HP Steam use Kg(steam)/TC:      235  |             Capital Costs              | Variable:                               |
|                                         |             --------------             |                Cane cost, $/TC:    8.07 |
|             Distillery                  |  Total Capital Cost,Millions:   85.624 |               Misc/other, $/TC:    4.17 |
|   ---------------------------           |  Capital cost,$/kw installed:      770 |        Bagasse credit, $/tonne:   19.00 |
|       Electricty use, kwh/L :      .05  |      Discount rate (10%=.10):      .12 |             Elect. cost, $/kwh: .042930 |
|    HP Steam use Kg(steam)/L :        0  |             Life of facility:       30 |               Steam cost, $/kg: .005495 |
|    LP Steam use Kg(steam)/L :      1.5  |                                        |                                         |
|   Liters ethanol/kg molasses:     .303  | Operation & Maintenance Costs          |=========================================|
|          Ethanol yield, L/tc:     9.09  | ----------------------------           |               DISTILLERY                |
|        Cogeneration Facility            | Fixed:                                 |                                         |
|   ---------------------------           |              Labor (Total $):   405000 |              Capital Costs              |
|                         Type:    ISTIG  |               Maintenance, $:  3164120 |             --------------              |
|      Full power capacity, kW:   111200  |                Other/misc, $:        0 |             Total Capital Cost: 5854500 |
|           Cogen capacity, kW:    97400  | Variable:                              |        Discount rate (10%=.10):     .12 |
|              Off season fuel:  barbojo  |     Bagasse Processing, $/TB:  11.8852 |               Life of facility:      20 |
|    % Off season running time:       .9  |  Total bagasse cost, $/tonne:    30.88 |                                         |
|                 kg steam/kWh:     .128  |     Total bagasse cost, $/GJ:     3.25 | Operation & Maintenance Costs           |
|        Electrical efficiency,           |   Off seas. fuel proc., $/GJ:     1.35 | ----------------------------            |
|             no process steam:     .429  |   Off season fuel cost, $/GJ:     2.81 | Fixed:                                  |
|        Electrical efficiency,           |           Other costs, $/khw:     .001 |                Labor (Total $):  117090 |
|           with process steam:     .379  |                                        |                    Maintenance:  117090 |
|    Elect. Cap. in full Cogen:    97400  |  Fuel required in season, GJ:  3197395 |                    Other/misc.: 29272.5 |
|  Max steam production, kg/TC:      235  |  Fuel input from bagasse, GJ:  3206348 | Variable:                               |
|                                         |  From other in season source:        0 |         Molasses cost, $/tonne:      40 |
|   Elect. to sugar fact., kwh:  2.248e7  |    Fuel input off season, GJ:  4131974 |          Electricy cost, $/kwh: .042930 |
|    HP Steam to sugar fact, T:  264155.  |                                        |               Steam cost, $/kg: .005495 |
|    Elect. to distillery, khw:  510887.  |        Net fuel rents to s&a:  3967733 |               Misc/other, $/L :    .038 |
|    LP Steam to distillery, T:  15326.6  |           percent to alcohol:  2.12e-7 |                                         |
|============================================================================================================================|

  Cost Summaries                 3967733                                  .841821
========================================================================================================================

                  Annual Sugar Costs                  Annual Distillery Costs                Annual Cogeneration Costs
------------------------------------------------------------------------------------------------------------------------

                     Capital : 1.150e7                        Capital :  783793                        Capital : 1.063e7
                        O & M                         O & M                                     O & M
                       Fixed : 1992482                          Fixed :  263453                          Fixed : 3569120
              Variable Misc. : 4687347                 Variable Misc. :  388274                 Variable Misc. :  829008
                   Cane Cost : 9071196                  Molasses Cost : 1348877                   Bagasse Cost : 1.041e7
            Electricity Cost :  965111               Electricity Cost :   21932            Electricity Credits :  987043
                  Steam Cost : 1367308                     Steam Cost :   84219                  Steam Credits : 1451527
              Bagasse Credit : 6300152                 Bagasse Credit :  106152           Off season fuel cost : 1.160e7
            Molassess Credit : 1348877
                              --------                                 --------                                 --------
           Total Annual Cost : 2.193e7              Total Annual Cost : 2784396              Total Annual Cost : 3.460e7

          Cents per Kg Sugar :   18.06        Cents per Liter Ethanol :   27.25                  Cents per kwh :    4.29

        Tonnes sugar per day : 758.743         Thousands of Liters/day: 63.8609           Power Generated, kwh : 8.290e8
       Tonnes sugar per year : 121399.        Thousands of Liters/year: 10217.7            Power Exported, kwh : 8.060e8
```

```
BAGASSE AND BARBOJO CREDITS TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input        Fulmer
updated values for molasses and sugar yields                                             4-4-90
BIG/ISTIG COGEN   C MOLASSES TO DISTILLERY
|===========================================================================================================|
|         SPECIFICATIONS                 |              PRICES                  |         SUGAR FACTORY        |
|                                        |                                      |                              |
|          Sugar Factory                 |  Cane Price, $ / tonne cane:   8.07  |        Capital Costs         |
|    -------------------------          |      Bagasse price, $/tonne:  19.01  |        -------------         |
|  Capacity, Tonnes milled/day:   7806  | Off season fuel price, $/GJ:   1.46  |    Total Capital Cost: 8.587e7 |
|           Season Length, days:   160  |         Molasses price,$/T:   40.00  |  Discount rate (10%=.10):   .12 |
|     Average percent capacity:     .9  | % fuel credit to distillery: .00854  |       Life of facility:     20 |
|           Yield, kg sugar/TC:    108  |                                      |                              |
|     Yield, kg  A molasses/TC:     30  |                                      | Operation & Maintenance Costs |
|    kg moist bagasse/kg cane:      .3  |    average fuel cost, $/GJ: 3.00069  | -------------------------    |
|             %sugar in bagasse:    .07 |======================================| Fixed:                       |
|          %moisture in bagasse:     .5 |                                      |        Labor (Total $):   .00 |
|            HHV bagasse, MJ/kg: 9.5082 |        COGENERATION COSTS            |           Maintenance:   .00 |
|        Electricty use, kwh/TC:    20  |                                      |         Other/misc.: 1992482 |
|     HP Steam use Kg(steam)/TC:   235  |           Capital Costs              | Variable:                    |
|                                        |           -------------              |       Cane cost, $/TC:   8.07 |
|           Distillery                   | Total Capital Cost,Millions:  85.624 |      Misc/other, $/TC:   4.17 |
|    -------------------------          | Capital cost,$/kw installed:    770  |  Bagasse credit, $/tonne: 19.01 |
|        Electricty use, kwh/L :   .05  |    Discount rate (10%=.10):     .12  |     Elect. cost, $/kwh: .042938 |
|     HP Steam use Kg(steam)/L :    0   |          Life of facility:      30   |      Steam cost, $/kg: .005496 |
|     LP Steam use Kg(steam)/L :   1.5  |                                      |                              |
|   Liters ethanol/kg molasses:   .303  | Operation & Maintenance Costs        |==============================|
|          Ethanol yield, L/tc:   9.09  | -------------------------            |         DISTILLERY           |
|       Cogeneration Facility            | Fixed:                               |                              |
|    -------------------------          |         Labor (Total $):  405000     |        Capital Costs         |
|                        Type:  ISTIG   |        Maintenance, $: 3164120       |        -------------         |
|     Full power capacity, kW: 111200   |          Other/misc, $:        0     |   Total Capital Cost: 5854500 |
|          Cogen capacity, kW:  97400   | Variable:                            |  Discount rate (10%-.10):   .12 |
|            Off season fuel: barbojo   |   Bagasse Processing, $/TB: 11.8852  |       Life of facility:     20 |
|   % Off season running time:     .9   | Total bagasse cost, $/tonne:  30.89  |                              |
|              kg steam/kWh:    .128    |   Total bagasse cost, $/GJ:    3.25  | Operation & Maintenance Costs |
|       Electrical efficiency,           |  Off seas. fuel proc., $/GJ:   1.35  | -------------------------    |
|           no process steam:    .429    |  Off season fuel cost, $/GJ:   2.91  | Fixed:                       |
|       Electrical efficiency,           |        Other costs, $/khw:     .001  |       Labor (Total $):  117090 |
|         with process steam:    .379    |                                      |          Maintenance:  117090 |
|   Elect. Cap. in full Cogen:   97400   | Fuel required in season, GJ: 3197395 |         Other/misc.: 29272.5 |
|  Max steam production, kg/TC:    235   | Fuel input from bagasse, GJ: 3206348 | Variable:                    |
|                                        | From other in season source:      0  |   Molasses cost, $/tonne:     40 |
|   Elect. to sugar fact., kwh: 2.248e7  |  Fuel input off season, GJ: 4131974 |   Electricy cost, $/kwh: .042938 |
|    HP Steam to sugar fact, T: 264155.  |                                      |      Steam cost, $/kg: .005496 |
|    Elect. to distillery, khw: 510887.  |      Net fuel rents to s&a: 9994875 |      Misc/other, $/L :    .038 |
|   LP Steam to distillery, T: 15326.6   |        percent to alcohol: .000001  |                              |
|===========================================================================================================|

  Cost Summaries            9994862                           13.1062
==========================================================================================================

              Annual Sugar Costs                 Annual Distillery Costs              Annual Cogeneration Costs
----------------------------------------------------------------------------------------------------------

              Capital  : 1.150e7                   Capital  :  783793                    Capital  : 1.063e7
                  O & M                     O & M                              O & M
                Fixed  : 1992482                     Fixed  :  263453                      Fixed  : 3569120
       Variable Misc.  : 4687347             Variable Misc.  :  388274             Variable Misc.  :  829008
           Cane Cost  : 9071196              Molasses Cost  : 1348877               Bagasse Cost  : 1.042e7
     Electricity Cost  :  965302           Electricity Cost  :   21936         Electricity Credits  :  987239
          Steam Cost  : 1367579                 Steam Cost  :   84236              Steam Credits  : 1451815
       Bagasse Credit  : 1.233e7             Bagasse Credit  :  106186         Off season fuel cost  : 1.160e7
     Molassess Credit  : 1348877
                         --------                             --------                                --------
    Total Annual Cost  : 1.590e7          Total Annual Cost  : 2784394             Total Annual Cost  : 3.461e7

   Cents per Kg Sugar  :   13.10      Cents per Liter Ethanol  :   27.25               Cents per kwh  :    4.29

  Tonnes sugar per day  : 758.743      Thousands of Liters/day: 63.8629        Power Gererated, kwh  : 8.290e8
 Tonnes sugar per year  : 121399.     Thousands of Liters/year: 10217.9        Power Exported, kwh  : 8.060e8
```

```
BAGASSE AND BARBOJO CREDITS TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input        Fulmer
updated values for molasses and sugar yields from Joan's spreadsheet                     4-1-90
BIG/STIG COGEN   C MOLASSES TO DISTILLERY
|===========================================================================================================|
|         SPECIFICATIONS                 |              PRICES                  |         SUGAR FACTORY        |
|                                        |                                      |                              |
|          Sugar Factory                 |  Cane Price, $ / tonne cane:   8.07  |        Capital Costs         |
|    -------------------------          |      Bagasse price, $/tonne:   6.04  |        -------------         |
|  Capacity, Tonnes milled/day:   3948  | Off season fuel price, $/GJ:    .00  |    Total Capital Cost: 4.343e7 |
|           Season Length, days:   160  |         Molasses price,$/T:   40.00  |  Discount rate (10%=.10):   .12 |
|     Average percent capacity:     .9  | % fuel credit to distillery: .04267  |       Life of facility:     20 |
|           Yield, kg sugar/TC:    108  |                                      |                              |
|     Yield, kg  A molasses/TC:     30  |                                      | Operation & Maintenance Costs |
|    kg moist bagasse/kg cane:      .3  |    average fuel cost, $/GJ: 1.56699  | -------------------------    |
|             %sugar in bagasse:    .07 |======================================| Fixed:                       |
|          %moisture in bagasse:     .5 |                                      |        Labor (Total $):   .00 |
|            HHV bagasse, MJ/kg: 9.5082 |        COGENERATION COSTS            |           Maintenance:   .00 |
|        Electricty use, kwh/TC:    20  |                                      |         Other/misc.: 1007727 |
|     HP Steam use Kg(steam)/TC:   235  |           Capital Costs              | Variable:                    |
|                                        |           -------------              |       Cane cost, $/TC:   8.07 |
|           Distillery                   | Total Capital Cost,Millions:  52.569 |      Misc/other, $/TC:   4.17 |
|    -------------------------          | Capital cost,$/kw installed:    990  |  Bagasse credit, $/tonne:  6.04 |
|        Electricty use, kwh/L :   .05  |    Discount rate (10%=.10):     .12  |     Elect. cost, $/kwh: .037968 |
|     HP Steam use Kg(steam)/L :    0   |          Life of facility:      30   |      Steam cost, $/kg: .004404 |
|     LP Steam use Kg(steam)/L :   1.5  |                                      |                              |
|   Liters ethanol/kg molasses:   .303  | Operation & Maintenance Costs        |==============================|
|          Ethanol yield, L/tc:   9.09  | -------------------------            |         DISTILLERY           |
|       Cogeneration Facility            | Fixed:                               |                              |
|    -------------------------          |         Labor (Total $):  297000     |        Capital Costs         |
|                   Type:BIG/STIG        |        Maintenance, $: 1566845       |        -------------         |
|     Full power capacity, kW:  53100   |          Other/misc, $:        0     |   Total Capital Cost: 2961000 |
|          Cogen capacity, kW:  42082   | Variable:                            |  Discount rate (10%=.10):   .12 |
|            Off season fuel: barbojo   |   Bagasse Processing, $/TB: 11.8852  |       Life of facility:     20 |
|   % Off season running time:     .9   | Total bagasse cost, $/tonne:  17.92  |                              |
|              kg steam/kWh:    .116    |   Total bagasse cost, $/GJ:    1.89  | Operation & Maintenance Costs |
|       Electrical efficiency,           |  Off seas. fuel proc., $/GJ:   1.35  | -------------------------    |
|           no process steam:    .356    |  Off season fuel cost, $/GJ:   1.35  | Fixed:                       |
|       Electrical efficiency,           |        Other costs, $/khw:     .001  |       Labor (Total $):   59220 |
|         with process steam:    .313    |                                      |          Maintenance:   59220 |
|   Elect. Cap. in full Cogen:   38800   | Fuel required in season, GJ: 1621609 |         Other/misc.:   14805 |
|  Max steam production, kg/TC:    305   | Fuel input from bagasse, GJ: 1621658 | Variable:                    |
|                                        | From other in season source:      0  |   Molasses cost, $/tonne:     40 |
|   Elect. to sugar fact., kwh: 1.137e7  |  Fuel input off season, GJ: 2377687 |   Electricy cost, $/kwh: .037968 |
|    HP Steam to sugar fact, T: 133600.  |                                      |      Steam cost, $/kg: .004404 |
|    Elect. to distillery, khw: 258389.  |      Net fuel rents to s&a:      55 |      Misc/other, $/L :    .038 |
|   LP Steam to distillery, T: 7751.66   |        percent to alcohol: -.02950  |                              |
|===========================================================================================================|

  Cost Summaries            56.1304                           -1.6082
==========================================================================================================

              Annual Sugar Costs                 Annual Distillery Costs              Annual Cogeneration Costs
----------------------------------------------------------------------------------------------------------

              Capital  : 5814088                   Capital  :  396415                    Capital  : 6526108
                  O & M                     O & M                              O & M
                Fixed  : 1007727                     Fixed  :  133245                      Fixed  : 1863845
       Variable Misc.  : 2370695             Variable Misc.  :  196375             Variable Misc.  :  380562
           Cane Cost  : 4587892              Molasses Cost  : 682214.               Bagasse Cost  : 3057037
     Electricity Cost  :  431700           Electricity Cost  :    9810         Electricity Credits  :  441511
          Steam Cost  :  554268                 Steam Cost  :   34140              Steam Credits  :  588408
       Bagasse Credit  :  986024             Bagasse Credit  :   43949         Off season fuel cost  : 3209877
     Molassess Credit  :  682214
                         --------                             --------                                --------
    Total Annual Cost  : 1.310e7          Total Annual Cost  : 1408251             Total Annual Cost  : 1.401e7

   Cents per Kg Sugar  :   21.33      Cents per Liter Ethanol  :   27.25               Cents per kwh  :    3.80

  Tonnes sugar per day  : 383.746      Thousands of Liters/day: 32.2986        Power Gererated, kwh  : 3.806e8
 Tonnes sugar per year  : 61399.3     Thousands of Liters/year: 5167.77        Power Exported, kwh  : 3.689e8
```

```
BAGASSE AND BARBOJO CREDITS TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input          Fulmer
updated values for molasses and sugar yields from Joan's spreadsheet                      4-1-90
BIG/STIG COGEN   C MOLASSES TO DISTILLERY
|===========================================================================================================|
|        SPECIFICATIONS                   |              PRICES                |          SUGAR FACTORY       |
|                                         |                                    |                              |
|         Sugar Factory                   |  Cane Price, $ / tonne cane:  8.07 |        Capital Costs         |
|    ---------------------------          |     Bagasse price, $/tonne:  19.66 |       ---------------        |
|  Capacity, Tonnes milled/day:    3948   | Off season fuel price, $/GJ:  1.43 |      Total Capital Cost: 4.343e7 |
|         Season Length, days:     160    |        Molasses price,$/T:  40.00  |  Discount rate (10%=.10):    .12 |
|     Average percent capacity:      .9   | % fuel credit to distillery: .0181 |       Life of facility:     20 |
|          Yield, kg sugar/TC:     108    |                                    |                              |
|     Yield, kg  A molasses/TC:      30   |                                    | Operation & Maintenance Costs |
|    kg moist bagasse/kg cane:      .3   |    average fuel cost, $/GJ: 2.99999 | ----------------------------- |
|           %sugar in bagasse:     .07   |====================================| Fixed:                       |
|        %moisture in bagasse:      .5   |                                    |        Labor (Total $):   .00 |
|         HHV bagasse, MJ/kg:   9.5082   |       COGENERATION COSTS           |           Maintenance:    .00 |
|      Electricty use, kwh/TC:      20   |                                    |          Other/misc.: 1007727 |
|   HP Steam use Kg(steam)/TC:     235   |         Capital Costs              | Variable:                    |
|                                         |        ---------------             |       Cane cost, $/TC:   8.07 |
|          Distillery                     | Total Capital Cost,Millions: 52.569 |      Misc/other, $/TC:   4.17 |
|    ---------------------------          | Capital cost,$/kw installed:   990 | Bagasse credit, $/tonne: 19.66 |
|       Electricty use, kwh/L :     .05   |    Discount rate (10%=.10):    .12 |   Elect. cost, $/kwh: .052438 |
|   HP Steam use Kg(steam)/L :        0   |          Life of facility:     30 |     Steam cost, $/kg: .006083 |
|   LP Steam use Kg(steam)/L :      1.5   |                                    |                              |
|  Liters ethanol/kg molasses:    .303   | Operation & Maintenance Costs      |==============================|
|         Ethanol yield, L/tc:     9.09   | ----------------------------       |          DISTILLERY          |
|     Cogeneration Facility               | Fixed:                             |                              |
|    ---------------------------          |         Labor (Total $):  297000   |        Capital Costs         |
|                      Type:BIG/STIG      |       Maintenance, $: 1566845      |       ---------------        |
|     Full power capacity, kW:   53100    |         Other/misc, $:        0    |  Total Capital Cost: 2961000 |
|          Cogen capacity, kW:   42082    | Variable:                          |  Discount rate (10%=.10):    .12 |
|           Off season fuel: barbojo      |   Bagasse Processing, $/TB: 11.8852 |       Life of facility:     20 |
|   % Off season running time:      .9    | Total bagasse cost, $/tonne:  31.55 |                              |
|               kg steam/kWh:    .116    |   Total bagasse cost, $/GJ:    3.32 | Operation & Maintenance Costs |
|      Electrical efficiency,             |  Off seas. fuel proc., $/GJ:   1.35 | ----------------------------- |
|           no process steam:    .356    |  Off season fuel cost, $/GJ:   2.78 | Fixed:                       |
|      Electrical efficiency,             |        Other costs, $/khw:    .001 |        Labor (Total $):  59220 |
|         with process steam:    .313    |                                    |           Maintenance:   59220 |
|   Elect. Cap. in full Cogen:   38800    | Fuel required in season, GJ: 1621609 |          Other/misc.:   14805 |
| Max steam production, kg/TC:     305    | Fuel input from bagasse, GJ: 1621658 | Variable:                    |
|                                         | From other in season source:      0 |  Molasses cost, $/tonne:     40 |
|  Elect. to sugar fact., kwh: 1.137e7    |  Fuel input off season, GJ: 2377687 | Electricy cost, $/kwh: .052438 |
|   HP Steam to sugar fact, T: 133600.    |                                    |     Steam cost, $/kg: .006083 |
|   Elect. to distillery, khw: 258389.    |       Net fuel rents to s&a: 1931368 |     Misc/other, $/L :    .038 |
|   LP Steam to distillery, T: 7751.66    |         percent to alcohol: .000002 |                              |
|===========================================================================================================|

  Cost Summaries              1931365                            2.94732
=============================================================================================================

          Annual Sugar Costs                  Annual Distillery Costs               Annual Cogeneration Costs
-------------------------------------------------------------------------------------------------------------

               Capital  : 5814088                  Capital  :  396415                   Capital  : 6526108
                   O & M                    O & M                               O & M
                 Fixed  : 1007727                    Fixed  :  133245                     Fixed  : 1863845
        Variable Misc.  : 2370695            Variable Misc.  :  196375             Variable Misc.  :  380562
             Cane Cost  : 4587892             Molasses Cost  : 682214.               Bagasse Cost  : 5380872
      Electricity Cost  :  596230          Electricity Cost  :   13549        Electricity Credits  :  609779
            Steam Cost  :  765510                Steam Cost  :   47152              Steam Credits  :  812662
        Bagasse Credit  : 3293105            Bagasse Credit  :   60704       Off season fuel cost  : 6617102
      Molassess Credit  :  682214
                          --------                              --------                             --------
     Total Annual Cost  : 1.117e7         Total Annual Cost  : 1408247          Total Annual Cost  : 1.935e7

     Cents per Kg Sugar  :   18.19   Cents per Liter Ethanol  :   27.25              Cents per kwh  :     5.24

   Tonnes sugar per day  : 383.746   Thousands of Liters/day: 32.2986      Power Generated, kwh  : 3.806e8
  Tonnes sugar per year  : 61399.3   Thousands of Liters/year: 5167.77      Power Exported, kwh  : 3.689e8
```

```
BAGASSE AND BARBOJO CREDITS TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input          Fulmer
updated values for molasses and sugar yields from Joan's spreadsheet                      4-1-90
BIG/STIG CCGEN   C MOLASSES TO DISTILLERY
|===========================================================================================================|
|        SPECIFICATIONS                   |              PRICES                |          SUGAR FACTORY       |
|                                         |                                    |                              |
|         Sugar Factory                   |  Cane Price, $ / tonne cane:  8.07 |        Capital Costs         |
|    ---------------------------          |     Bagasse price, $/tonne:  19.66 |       ---------------        |
|  Capacity, Tonnes milled/day:    3948   | Off season fuel price, $/GJ:  1.43 |      Total Capital Cost: 4.343e7 |
|         Season Length, days:     160    |        Molasses price,$/T:  40.00  |  Discount rate (10%=.10):    .12 |
|     Average percent capacity:      .9   | % fuel credit to distillery: .00899 |       Life of facility:     20 |
|          Yield, kg sugar/TC:     108    |                                    |                              |
|     Yield, kg  A molasses/TC:      30   |                                    | Operation & Maintenance Costs |
|    kg moist bagasse/kg cane:      .3   |    average fuel cost, $/GJ: 2.99999 | ----------------------------- |
|           %sugar in bagasse:     .07   |====================================| Fixed:                       |
|        %moisture in bagasse:      .5   |                                    |        Labor (Total $):   .00 |
|         HHV bagasse, MJ/kg:   9.5082   |       COGENERATION COSTS           |           Maintenance:    .00 |
|      Electricty use, kwh/TC:      20   |                                    |          Other/misc.: 1007727 |
|   HP Steam use Kg(steam)/TC:     235   |         Capital Costs              | Variable:                    |
|                                         |        ---------------             |       Cane cost, $/TC:   8.07 |
|          Distillery                     | Total Capital Cost,Millions: 52.569 |      Misc/other, $/TC:   4.17 |
|    ---------------------------          | Capital cost,$/kw installed:   990 | Bagasse credit, $/tonne: 19.66 |
|       Electricty use, kwh/L :     .05   |    Discount rate (10%=.10):    .12 |   Elect. cost, $/kwh: .052438 |
|   HP Steam use Kg(steam)/L :        0   |          Life of facility:     30 |     Steam cost, $/kg: .006083 |
|   LP Steam use Kg(steam)/L :      1.5   |                                    |                              |
|  Liters ethanol/kg molasses:    .303   | Operation & Maintenance Costs      |==============================|
|         Ethanol yield, L/tc:     9.09   | ----------------------------       |          DISTILLERY          |
|     Cogeneration Facility               | Fixed:                             |                              |
|    ---------------------------          |         Labor (Total $):  297000   |        Capital Costs         |
|                      Type:BIG/STIC      |       Maintenance, $: 1566845      |       ---------------        |
|     Full power capacity, kW:   53100    |         Other/misc, $:        0    |  Total Capital Cost: 2961000 |
|          Cogen capacity, kW:   42082    | Variable:                          |  Discount rate (10%=.10):    .12 |
|           Off season fuel: barbojo      |   Bagasse Processing, $/TB: 11.8852 |       Life of facility:     20 |
|   % Off season running time:      .9    | Total bagasse cost, $/tonne:  31.55 |                              |
|               kg steam/kWh:    .116    |   Total bagasse cost, $/GJ:    3.32 | Operation & Maintenance Costs |
|      Electrical efficiency,             |  Off seas. fuel proc., $/GJ:   1.35 | ----------------------------- |
|           no process steam:    .356    |  Off season fuel cost, $/GJ:   2.78 | Fixed:                       |
|      Electrical efficiency,             |        Other costs, $/khw:    .001 |        Labor (Total $):  59220 |
|         with process steam:    .313    |                                    |           Maintenance:   59220 |
|   Elect. Cap. in full Cogen:   38800    | Fuel required in season, GJ: 1621609 |          Other/misc.:   14805 |
| Max steam production, kg/TC:     305    | Fuel input from bagasse, GJ: 1621658 | Variable:                    |
|                                         | From other in season source:      0 |  Molasses cost, $/tonne:     40 |
|  Elect. to sugar fact., kwh: 1.137e7    |  Fuel input off season, GJ: 2377687 | Electricy cost, $/kwh: .052438 |
|   HP Steam to sugar fact, T: 133600.    |                                    |     Steam cost, $/kg: .006083 |
|   Elect. to distillery, khw: 258389.    |       Net fuel rents to s&a: 5338593 |     Misc/other, $/L :    .038 |
|   LP Steam to distillery, T: 7751.66    |         percent to alcohol: .000002 |                              |
|===========================================================================================================|

  Cost Summaries              5338580                            13.0931
=============================================================================================================

          Annual Sugar Costs                  Annual Distillery Costs               Annual Cogeneration Costs
-------------------------------------------------------------------------------------------------------------

               Capital  : 5814088                  Capital  :  396415                   Capital  : 6526108
                   O & M                    O & M                               O & M
                 Fixed  : 1007727                    Fixed  :  133245                     Fixed  : 1863845
        Variable Misc.  : 2370695            Variable Misc.  :  196375             Variable Misc.  :  380562
             Cane Cost  : 4587892             Molasses Cost  : 682214.               Bagasse Cost  : 5380872
      Electricity Cost  :  596230          Electricity Cost  :   13549        Electricity Credits  :  609779
            Steam Cost  :  765510                Steam Cost  :   47152              Steam Credits  :  812662
        Bagasse Credit  : 6700320            Bagasse Credit  :   60714       Off season fuel cost  : 6617102
      Molassess Credit  :  682214
                          --------                              --------                             --------
     Total Annual Cost  : 7759607         Total Annual Cost  : 1408237          Total Annual Cost  : 1.935e7

     Cents per Kg Sugar  :   12.64   Cents per Liter Ethanol  :   27.25              Cents per kwh  :     5.24

   Tonnes sugar per day  : 383.746   Thousands of Liters/day: 32.2986      Power Generated, kwh  : 3.806e8
  Tonnes sugar per year  : 61399.3   Thousands of Liters/year: 5167.77      Power Exported, kwh  : 3.689e8
```

```
BAGASSE CREDITS ONLY TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input        Fulmer
updated values for molasses and sugar yields from Joan's spreadsheet                     4-1-90
CEST COGEN        C MOLASSES TO DISTILLERY
|=========================================================================================================|
|      SPECIFICATIONS                   |            PRICES                  |          SUGAR FACTORY          |
|                                       |                                    |                                 |
|       Sugar Factory                   |  Cane Price, $ / tonne cane:  8.07 |         Capital Costs           |
|   ---------------------------         |      Bagasse price, $/tonne:  8.93 |        ---------------          |
| Capacity, Tonnes milled/day:   4053   | Off season fuel price, $/GJ:   .00 |      Total Capital Cost: 4.458e7|
|         Season Length, days:    160   |          Molasses price,$/T: 40.00 |  Discount rate (10%=.10):    .15|
|    Average percent capacity:     .9   | % fuel credit to distillery: .04465|        Life of facility:     20 |
|         Yield, kg sugar/TC:     108   |                                    |                                 |
|   Yield, kg  A molasses/TC:      30   |                                    | Operation & Maintenance Costs   |
|   kg moist bagasse/kg cane:      .3   |    average fuel cost, $/GJ: .956356| ----------------------------    |
|           %sugar in bagasse:    .07   |====================================| Fixed:                          |
|        %moisture in bagasse:     .5   |                                    |        Labor (Total $):     .00 |
|         HHV bagasse, MJ/kg:  9.5082   |      COGENERATION COSTS            |            Maintenance:     .00 |
|      Electricty use, kwh/TC:     20   |                                    |            Other/misc.: 1034528 |
|  HP Steam use Kg(steam)/TC:    235   |         Capital Costs              | Variable:                       |
|                                       |        ---------------             |         Cane cost, $/TC:   8.07 |
|         Distillery                    | Total Capital Cost,Millions: 42.012|        Misc/other, $/TC:   4.17 |
|   ---------------------------         | Capital cost,$/kw installed:  1556 |  Bagasse credit, $/tonne:  8.93 |
|       Electricty use, kwh/L :    .05  |     Discount rate (10%=.10):   .12 |     Elect. cost, $/kwh: .048909 |
|   HP Steam use Kg(steam)/L :      0   |           Life of facility:    30 |       Steam cost, $/kg: .007141 |
|   LP Steam use Kg(steam)/L :    1.5   |                                    |                                 |
|  Liters ethanol/kg molasses:   .303   | Operation & Maintenance Costs      |=================================|
|        Ethanol yield, L/tc:    9.09   | ----------------------------       |         DISTILLERY              |
|     Cogeneration Facility             | Fixed:                             |                                 |
|   ---------------------------         |            Labor (Total $): 129600 |         Capital Costs           |
|                        Type:    CEST  |             Maintenance, $: 874060 |        ---------------          |
|    Full power capacity, kW:    27000  |              Other/misc, $:      0 |      Total Capital Cost: 3039750|
|         Cogen capacity, kW:    21246  | Variable:                          |  Discount rate (10%-.10):    .12|
|           Off season fuel:  barbojo   |    Bagasse Processing, $/TB:     0 |        Life of facility:     20 |
|  % Off season running time:       .9  | Total bagasse cost, $/tonne:   8.93 |                                 |
|               kg steam/kWh:     .146  |    Total bagasse cost, $/GJ:    .94 | Operation & Maintenance Costs   |
|     Electrical efficiency,            |  Off seas. fuel proc., $/GJ:    .97 | ----------------------------    |
|           no process steam:     .203  |  Off season fuel cost, $/GJ:    .97 | Fixed:                          |
|     Electrical efficiency,            |         Other costs, $/khw:   .003 |        Labor (Total $):   60795 |
|         with process steam:      .13  |                                    |            Maintenance:   60795 |
|  Elect. Cap. in full Cogen:    17500  | Fuel required in season, GJ: 1664732|           Other/misc.: 15198.8 |
| Max steam production, kg/TC:     388  | Fuel input from bagasse, GJ: 1664787| Variable:                       |
|                                       | From other in season source:      0|  Molasses cost, $/tonne:      40 |
| Elect. to sugar fact., kwh: 1.167e7   |   Fuel input off season, GJ: 2120205|  Electricy cost, $/kwh: .048909 |
|  HP Steam to sugar fact, T: 137154.   |                                    |       Steam cost, $/kg: .007141 |
|  Elect. to distillery, khw: 265261.   |        Net fuel rents to s&a:   -39 |      Misc/other, $/L :     .038 |
|  LP Steam to distillery, T: 7957.82   |          percent to alcohol: .026583|                                 |
|=========================================================================================================|

  Cost Summaries               -38.333                              -1.0469
============================================================================================================

             Annual Sugar Costs                  Annual Distillery Costs              Annual Cogeneration Costs
------------------------------------------------------------------------------------------------------------

                  Capital  : 7122646                    Capital  :  406958                     Capital  : 5215523
                      O & M                       O & M                                 O & M
                    Fixed  : 1034528                      Fixed  :  136789                       Fixed  : 1003660
            Variable Misc. : 2433745              Variable Misc. :  201598              Variable Misc.  :  578948
                Cane Cost  : 4709910               Molasses Cost : 700358.                Bagasse Cost  : 1563200
         Electricity Cost  :  570895            Electricity Cost :   12974         Electricity Credits  :  583869
               Steam Cost  :  922546                  Steam Cost :   56824               Steam Credits  :  979371
           Bagasse Credit  : 1493403              Bagasse Credit :   69797        Off season fuel cost  : 2056599
         Molassess Credit  :  700358
                             --------                              --------                              --------
        Total Annual Cost  : 1.460e7           Total Annual Cost : 1445704           Total Annual Cost  : 8854691

        Cents per Kg Sugar :   23.16     Cents per Liter Ethanol :   27.25               Cents per kwh  :    4.89

       Tonnes sugar per day : 393.952      Thousands of Liters/day: 33.1576     Power Cererated, kwh  : 1.930e8
      Tonnes sugar per year : 63032.3     Thousands of Liters/year: 5305.21     Power Exported, kwh   : 1.810e8
```

```
BAGASSE CREDITS ONLY TO SUGAR FACTORY AND DISTILLERY
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input        Fulmer
updated values for molasses and sugar yields from Joan's spreadsheet                     4-1-90
CEST COGEN        C MOLASSES TO DISTILLERY
|=========================================================================================================|
|      SPECIFICATIONS                   |            PRICES                  |          SUGAR FACTORY          |
|                                       |                                    |                                 |
|       Sugar Factory                   |  Cane Price, $ / tonne cane:  8.07 |         Capital Costs           |
|   ---------------------------         |      Bagasse price, $/tonne: 28.36 |        ---------------          |
| Capacity, Tonnes milled/day:   4053   | Off season fuel price, $/GJ:  2.04 |      Total Capital Cost: 4.458e7|
|         Season Length, days:    160   |          Molasses price,$/T: 40.00 |  Discount rate (10%=.10):    .15|
|    Average percent capacity:     .9   | % fuel credit to distillery:  .0245|        Life of facility:     20 |
|         Yield, kg sugar/TC:     108   |                                    |                                 |
|   Yield, kg  A molasses/TC:      30   |                                    | Operation & Maintenance Costs   |
|   kg moist bagasse/kg cane:      .3   |   average fuel cost, $/GJ: 3.00036 | ----------------------------    |
|           %sugar in bagasse:    .07   |====================================| Fixed:                          |
|        %moisture in bagasse:     .5   |                                    |        Labor (Total $):     .00 |
|         HHV bagasse, MJ/kg:  9.5082   |      COGENERATION COSTS            |            Maintenance:     .00 |
|      Electricty use, kwh/TC:     20   |                                    |            Other/misc.: 1034528 |
|  HP Steam use Kg(steam)/TC:    235   |         Capital Costs              | Variable:                       |
|                                       |        ---------------             |         Cane cost, $/TC:   8.07 |
|         Distillery                    | Total Capital Cost,Millions: 42.012|        Misc/other, $/TC:   4.17 |
|   ---------------------------         | Capital cost,$/kw installed:  1556 |  Bagasse credit, $/tonne: 28.36 |
|       Electricty use, kwh/L :    .05  |     Discount rate (10%=.10):   .12 |     Elect. cost, $/kwh: .085229 |
|   HP Steam use Kg(steam)/L :      0   |           Life of facility:    30 |       Steam cost, $/kg: .012443 |
|   LP Steam use Kg(steam)/L :    1.5   |                                    |                                 |
|  Liters ethanol/kg molasses:   .303   | Operation & Maintenance Costs      |=================================|
|        Ethanol yield, L/tc:    9.09   | ----------------------------       |         DISTILLERY              |
|     Cogeneration Facility             | Fixed:                             |                                 |
|   ---------------------------         |            Labor (Total $): 129600 |         Capital Costs           |
|                        Type:    CEST  |             Maintenance, $: 874060 |        ---------------          |
|    Full power capacity, kW:    27000  |              Other/misc, $:      0 |      Total Capital Cost: 3039750|
|         Cogen capacity, kW:    21246  | Variable:                          |  Discount rate (10%-.10):    .12|
|           Off season fuel:  barbojo   |    Bagasse Processing, $/TB:     0 |        Life of facility:     20 |
|  % Off season running time:       .9  | Total bagasse cost, $/tonne:  28.36 |                                 |
|               kg steam/kWh:     .146  |    Total bagasse cost, $/GJ:   2.98 | Operation & Maintenance Costs   |
|     Electrical efficiency,            |  Off seas. fuel proc., $/GJ:    .97 | ----------------------------    |
|           no process steam:     .203  |  Off season fuel cost, $/GJ:   3.01 | Fixed:                          |
|     Electrical efficiency,            |         Other costs, $/khw:   .003 |        Labor (Total $):   60795 |
|         with process steam:      .13  |                                    |            Maintenance:   60795 |
|  Elect. Cap. in full Cogen:    17500  | Fuel required in season, GJ: 1664732|           Other/misc.: 15198.8 |
| Max steam production, kg/TC:     388  | Fuel input from bagasse, GJ: 1664787| Variable:                       |
|                                       | From other in season source:      0|  Molasses cost, $/tonne:      40 |
| Elect. to sugar fact., kwh: 1.167e7   |   Fuel input off season, GJ: 2120205|  Electricy cost, $/kwh: .085229 |
|  HP Steam to sugar fact, T: 137154.   |                                    |       Steam cost, $/kg: .012443 |
|  Elect. to distillery, khw: 265261.   |     Net fuel rents to s&a: 2241898 |      Misc/other, $/L :     .038 |
|  LP Steam to distillery, T: 7957.82   |          percent to alcohol: .000016|                                 |
|=========================================================================================================|

  Cost Summaries               2241862                               36.6078
============================================================================================================

             Annual Sugar Costs                  Annual Distillery Costs              Annual Cogeneration Costs
------------------------------------------------------------------------------------------------------------

                  Capital  : 7122646                    Capital  :  406958                     Capital  : 5215523
                      O & M                       O & M                                 O & M
                    Fixed  : 1034528                      Fixed  :  136789                       Fixed  : 1003660
            Variable Misc. : 2433745              Variable Misc. :  201598              Variable Misc.  :  578948
                Cane Cost  : 4709910               Molasses Cost : 700358.                Bagasse Cost  : 4966024
         Electricity Cost  :  994851            Electricity Cost :   22608         Electricity Credits  : 1017459
               Steam Cost  : 1607644                  Steam Cost :   99023               Steam Credits  : 1706667
           Bagasse Credit  : 4844357              Bagasse Credit :  121668        Off season fuel cost  : 6390298
         Molassess Credit  :  700358
                             --------                              --------                              --------
        Total Annual Cost  : 1.236e7           Total Annual Cost : 1445667           Total Annual Cost  : 1.543e7

        Cents per Kg Sugar :   19.61     Cents per Liter Ethanol :   27.25               Cents per kwh  :    8.52

       Tonnes sugar per day : 393.952      Thousands of Liters/day: 33.1576     Power Cererated, kwh  : 1.930e8
      Tonnes sugar per year : 63032.3     Thousands of Liters/year: 5305.21     Power Exported, kwh   : 1.810e8
```

```
BAGASSE AND BARBOJO CREDITS TO SUGAR FACTORY AND DISTILLERY                                              Fulmer
spreadsheet set up to prorate bagasse credit between dist. & sugar fact. on input                         4-1-90
updated values for molasses and sugar yields from Joan's spreadsheet
CEST COGEN        C MOLASSES TO DISTILLERY
|=============================================================================================================|
|       SPECIFICATIONS                    |               PRICES                  |           SUGAR FACTORY   |
|                                         |                                       |                           |
|        Sugar Factory                    |  Cane Price, $ / tonne cane:    8.07  |         Capital Costs     |
|  ---------------------------           |      Bagasse price, $/tonne:   28.36  |        ---------------    |
| Capacity, Tonnes milled/day:    4053   | Off season fuel price, $/GJ:    2.04  |    Total Capital Cost: 4.458e7 |
|        Season Length, days:      160   |        Molasses price,$/T:     40.00  |  Discount rate (10%=.10):   .12 |
|    Average percent capacity:      .9   | % fuel credit to distillery:  .01306  |       Life of facility:     20 |
|           Yield, kg sugar/TC:     108   |                                       |                           |
|   Yield, kg  A molasses/TC:        30   |                                       | Operation & Maintenance Costs |
|   kg moist bagasse/kg cane:        .3   |   average fuel cost, $/GJ: 3.00036    | ----------------------------- |
|          %sugar in bagasse:       .07   |=======================================| Fixed:                    |
|       %moisture in bagasse:        .5   |                                       |        Labor (Total $):   .00 |
|         HHV bagasse, MJ/kg:   9.5082   |     COGENERATION COSTS                |           Maintenance:    .00 |
|     Electricty use, kwh/TC:        20   |                                       |          Other/misc.: 1034528 |
|   HP Steam use Kg(steam)/TC:      235   |        Capital Costs                  | Variable:                 |
|                                         |       ---------------                |        Cane cost, $/TC:   8.07 |
|          Distillery                     | Total Capital Cost,Millions:  42.012  |       Misc/other, $/TC:   4.17 |
|  ---------------------------           | Capital cost,$/kw installed:   1556   |  Bagasse credit, $/tonne: 28.36 |
|      Electricty use, kwh/L :      .05   |   Discount rate (10%=.10):      .12   |    Elect. cost, $/kwh: .085229 |
|   HP Steam use Kg(steam)/L :        0   |          Life of facility:       30   |     Steam cost, $/kg: .012443 |
|   LP Steam use Kg(steam)/L :      1.5   |                                       |                           |
| Liters ethanol/kg molasses:      .303   | Operation & Maintenance Costs         |===========================|
|        Ethanol yield, L/tc:      9.09   | -----------------------------         |        DISTILLERY         |
|     Cogeneration Facility               | Fixed:                                |                           |
|  ---------------------------           |             Labor (Total $):  129600  |       Capital Costs       |
|                       Type:      CEST   |              Maintenance, $:  874060  |      ---------------      |
|     Full power capacity, kW:    27000   |              Other/misc, $:        0  |    Total Capital Cost: 3039750 |
|          Cogen capacity, kW:    21246   | Variable:                             |  Discount rate (10%=.10):   .12 |
|            Off season fuel: barbojo    |    Bagasse Processing, $/TB:       0  |       Life of facility:     20 |
|   % Off season running time:       .9   | Total bagasse cost, $/tonne:   28.36  |                           |
|             kg steam/kWh:        .146   |     Total bagasse cost, $/GJ:   2.98  | Operation & Maintenance Costs |
|       Electrical efficiency,             |   Off seas. fuel proc., $/GJ:    .97  | ----------------------------- |
|            no process steam:      .203   |   Off season fuel cost, $/GJ:   3.01  | Fixed:                    |
|       Electrical efficiency,             |          Other costs, $/khw:    .003  |        Labor (Total $):  60795 |
|          with process steam:       .13   |                                       |           Maintenance:  60795 |
|   Elect. Cap. in full Cogen:     17500   | Fuel required in season, GJ: 1664732  |         Other/misc.: 15198.8 |
| Max steam production, kg/TC:       388   | Fuel input from bagasse, GJ: 1664787  | Variable:                 |
|                                         | From other in season source:       0  |   Molasses cost, $/tonne:    40 |
| Elect. to sugar fact., kwn: 1.167e7    |   Fuel input off season, GJ: 2120235  |  Electricy cost, $/kwh: .085229 |
|  HP Steam to sugar fact, T: 137154.    |                                       |     Steam cost, $/kg: .012443 |
|  Elect. to distillery, khw: 265261.    |         Net fuel rents to s&a: 6575597 |      Misc/other, $/L :    .038 |
|  LP Steam to distillery, T: 7957.82    |          percent to alcohol: .000001  |                           |
|=============================================================================================================|

  Cost Summaries                 6575588                                 9.33965
===============================================================================================================

                 Annual Sugar Costs                      Annual Distillery Costs                 Annual Cogeneration Costs
---------------------------------------------------------------------------------------------------------------

                Capital  : 5968718                      Capital  :  406958                      Capital  : 5215523
                    O & M                          O & M                                  O & M
                  Fixed  : 1034528                        Fixed  :  136789                        Fixed  : 1003660
         Variable Misc.  : 2433745               Variable Misc.  :  201598               Variable Misc.  :  578948
              Cane Cost  : 4709910               Molasses Cost  : 700358.                 Bagasse Cost  : 4966024
       Electricity Cost  :  994851            Electricity Cost  :   22608          Electricity Credits  : 1017459
             Steam Cost  : 1607644                  Steam Cost  :   99023              Steam Credits  : 1706667
         Bagasse Credit  : 9178083              Bagasse Credit  :  121640       Off season fuel cost  : 6390298
       Molassess Credit  :  700358
                           --------                             --------                              --------
      Total Annual Cost  : 6870955           Total Annual Cost  : 1445694           Total Annual Cost  : 1.543e7

    Cents per Kg Sugar   :   10.90     Cents per Liter Ethanol  :   27.25              Cents per kwh   :    8.52

   Tonnes sugar per day  : 393.952   Thousands of Liters/day: 33.1576       Power Generated, kwh  : 1.930e8
  Tonnes sugar per year  : 63032.3  Thousands of Liters/year: 5305.22        Power Exported, kwh  : 1.810e8
```

AUTONOMOUS DISTILLERY CALCULATIONS

I. Distillery Costs

A. Annual Capital Recovery ($/year):

$$ACR = CAP \times \frac{i}{1-(1+i)^{-N}}$$

CAP = Total Capital Cost.
i = Discount rate (return on capital investment).
N = Expected life of facility (years).

B. Operation and Maintenance, Fixed Costs ($/year):

$$FC = La + Mn + Misc$$

La = All labor costs (salaries, wages, etc., $/year).
Mn = Maintenance costs, excluding labor ($/year).
Misc = Any remaining miscellaneous fixed costs ($/year).

C. Operation and Maintenance, Variable Costs ($/year):

$$TVC = ACM \times VC$$

ACM = Tonnes of cane milled annually.
ACM = TC x Seas x Cf

TC = Milling capacity, tonnes cane per day.
Seas = Days of mill operation per year.
Cf = Capacity factor: Average percent of capacity actually operated.

VC = Total variable cost, $/tonne cane milled.

D. Electricity Costs ($/year):

$$EC = ACM \times ELTC \times AEC$$

ACM = Tonnes cane milled annually (Section I.C).
ELTC = Electricity used per tonne cane (kwh/tonne cane).
AEC = Cogeneration facility's average electricity production cost (Section II.I, $/kwh).

E. Steam Costs ($/year):

$$SC = ACM \times STC \times STE \times AEC$$

ACM = Tonnes cane milled annually (Section I.C).
STC = Kg high pressure steam used per tonne of cane milled.
STE = The amount of electricity (kwh) which can be generated by 1 kg of HP steam.

AEC = Cogeneration facility's average electricity production cost (Section II.I, $/kwh).

F. Cane Costs ($/year):

$$CC = CPR \times ACM$$

CPR = Cane price, \$/tonne, for cane delivered to the factory.
ACM = Tonnes cane milled annually (section I.C).

G. Bagasse Credits From Cogeneration Facility (\$/year):

$$BCR = ACM \times BPC \times BP$$

ACM = Tonnes cane milled annually (Section I.C).
BPC = Tonnes of 50% mc bagasse produced per tonne cane milled.
BP = Price per tonne of 50% mc bagasse.

H. Total Annual Cost (\$):

$$TAC = ACR + FC + TVC + EC + SC + CC - BCR$$

ACR = Annual Capital Recovery, (Section I.A);
FC = Annual Fixed O & M Cost, (Section I.B);
TVC = Total Variable Costs, (Section I.C);
EC = Electricity Costs, (Section I.D);
SC = Steam Costs, (Section I.E);
CC = Sugar Cane Cost,(Section I.F);
BCR = Bagasse Credits, (Section I.G).

I. Production Cost:

$$PC = \frac{TAC}{TAP}$$

TAC = Total Annual Ethanol Production Costs.
TAP = Total Annual Ethanol Production, Liters.
TAP = ACM x LPT

ACM = Tonnes cane milled annually (Section I.C).
LPT = Ethanol yield, Liters per tonne cane.

II. ANNUAL COGENERATION FACILITY COSTS

A. Annual Capital Recovery (\$/year):

$$ACR = CAP \times \frac{i}{1-(1+i)^{N}}$$

CAP = Total Capital Cost.
i = Discount rate (return on capital investment).
N = Expected life of facility (years).

B. Operation and Maintenance, Fixed Costs (\$/year):

$$FC = La + Mn + Misc$$

La = All labor costs (salaries, wages, etc.)
Mn = Maintenance costs, excluding labor.
Misc = Any remaining miscellaneous fixed costs.

C. Operation and Maintenance, Variable Costs ($/year):

TVC = VC x EP

VC = Cogeneration facility's variable costs, $/kwh.
EP = Total annual electricity production, kwh.
EP = [(ACM x BPC x HHV)+(CAP/Eff x OSL x OCf)] x Eff.

ACM = Tonnes cane milled annually (Section I.C).
BPC = Tonnes of 50% mc bagasse per tonne cane milled.
HHV = Higher heating value of 50% moist bagasse (kj/kg).
CAP = Generating capacity of the cogeneration facility, taken to be the amount of electricity generated burning the bagasse produced at the maximum grinding rate, with no steam exported to the distillery (kw).
Eff = Busbar electrical efficiency: fuel energy (HHV) in divided by electrical energy out (no steam exported to distillery).
OSL = Length of off-season (days, no milling).
OSf = Off season capacity factor, average percent of full capacity operated during the off season.

D. Electricity Credits from Distillery ($/year):

ECR = EC

EC = Distillery's annual electricity cost (section I.D, $/year).

E. Steam Credits from Distillery ($/year):

STC = SC

SC = Distillery's annual steam costs (section I.E, $/year).

F. Off Season Fuel Costs:

OSFC = CAP/Eff x OSL x OSf x OSFP

CAP = Cogeneration facility's electric generating capacity, kw (Section II.C).
Eff = Busbar electrical efficiency (Section II.C).
OSL = Length of Off (non-milling) season, days.
OSf = Capacity factor of off-season (section II.C).
OSFP = Off season fuel price, $/GJ.

G. Bagasse Fuel Costs ($/year):

TBC = (BP + BPC) * ACM * BPC

BP = Price per tonne of 50% mc Bagasse.
BPC = Bagasse processing costs, $/tonne 50% mc bagasse.
ACM = Amount of cane milled annually (Section I.C).
BPC = Tonnes 50% mc bagasse per tonne cane milled.

H. Total Annual Cogeneration Costs:

TACC = CR + FC + TVC - ECR - SCR + OSFC + TBC

CR = Annual capital recovery, (Section II.A);
FC = Fixed Costs, (Section II.B);

TVC = Total Variable Costs, (Section II.C);
ECR = Electricity Credit from distillery, (Section II.D);
SCR = Steam Credit from distillery, (Section II.E);
OSFC = Off Season fuel cost, (Section II.F);
TBC = Total Bagasse Cost, (Section II.G).

I. Average electricity production cost, $ per kwh:

$$AEC = \frac{TACC}{EP - ED}$$

EP = Total Annual electricity production, kwh.
ED = Distillery's annual electricity use, kwh (Section I.D).
TACC = Total Annual Cogeneration Costs, $.

ANNEXED DISTILLERY CALCULATIONS

I. SUGAR FACTORY COSTS

A. Annual Capital Recovery ($/year):

$$ACR = CAP \times \frac{i}{1-(1+i)^{-N}}$$

CAP = Total Capital Cost for sugar factory (no cogen. or distillery).
i = Discount rate (return on capital investment).
N = Expected life of facility (years).

B. Operation and Maintenance, Fixed Costs ($/year):

$$FC = La + Mn + Misc$$

La = All labor costs (salaries, wages, etc., $/year).
Mn = Maintenance costs, excluding labor ($/year).
Misc = Any remaining miscellaneous fixed costs ($/year).

C. Operation and Maintenance, Variable Costs ($/year):

$$TVC = ACM \times VC$$

ACM = Tonnes of cane milled annually.
ACM = TC x Seas x Cf

- TC = Milling capacity, tonnes cane per day.
- Seas = Days of mill operation per year.
- Cf = Capacity factor: Average percent of capacity actually operated.

VC = Variable cost, $/tonne cane milled.

D. Electricity Costs ($/year):

$$EC = ACM \times ELTC \times AEC$$

ACM = Tonnes cane milled annually (Section I.C).
ELTC = Electricity used per tonne cane (kwh/tonne cane).
AEC = Cogeneration facility's average electricity production cost, $/kwh (Section III.J).

E. Steam Costs ($/year):

$$SC = (ACM \times STC - DSTM) \times STE \times AEC$$

ACM = Tonnes cane milled annually (Section I.C).
STC = Kg high pressure steam used per tonne of cane milled.
DSTM = LP steam used in distillery. In order to split the HP steam costs between the sugar factory and distillery, the distillery pays the cogenerator for its steam use rather than having the sugar factory pay the whole HP steam bill and then in turn charging the distillery for its LP steam (Section II.E).
STE = The amount of electricity (kwh) which can be generated by 1 kg of HP steam.
AEC = Cogeneration facility's average electricity production cost, $/kwh

(Section III.J).

F. Cane Costs ($/year):

$CC = CPR \times ACM$

CPR = Cane price, $/tonne, for cane delivered to the factory.
ACM = Tonnes cane milled annually (section I.C).

G. Bagasse Credits From Cogeneration Facility ($/year):

$BCR = ACM \times BPC \times BP$

ACM = Tonnes cane milled annually (Section I.C).
BPC = Tonnes of 50% mc bagasse produced per tonne cane milled.
BP = Price per tonne of 50% mc bagasse.

H. Molasses Credits From Distillery ($/year):

$MCR = ACM \times MY \times MP$

ACM = Tonnes cane milled annually (Section I.C).
MY = C molasses yield, tonnes molasses per tonne cane.
MP = Molasses price, $/tonne.

I. Total Annual Cost ($):

$TAC = ACR + FC + TVC + EC + SC + CC - BCR - MCR$

ACR = Annual Capital Recovery, (Section I.A);
FC = Annual Fixed O & M Cost, (Section I.B);
TVC = Total Variable Costs, (Section I.C);
EC = Electricity Costs, (Section I.D);
SC = Steam Costs, (Section I.E);
CC = Sugar Cane Cost, (Section I.F);
BCR = Bagasse Credits, (Section I.G);
MCR = Molasses credits, (Section I.H).

J. Production Cost:

$$PC = \frac{TAC}{TAP}$$

TAC = Total Annual Sugar Production Costs, $.
TAP = Total Annual Sugar Production, kg.
TAP = ACM x LPT

ACM = Tonnes cane milled annually (Section I.C).
LPT = Sugar yield, hg sugar per tonne cane.

II. DISTILLERY COSTS

A. Annual Capital Recovery ($/year):

$$ACR = CAP \times \frac{i}{1-(1+i)^N}$$

CAP = Total Capital Cost.

i = Discount rate (return on capital investment).
N = Expected life of facility (years).

B. Operation and Maintenance, Fixed Costs ($/year):

FC = La + Mn + Misc

La = All labor costs (salaries, wages, etc., $/year).
Mn = Maintenance costs, excluding labor ($/year).
Misc = Any remaining miscellaneous fixed costs ($/year).

C. Operation and Maintenance, Variable Costs ($/year):

TVC = MA x VC

MA = Annual sugar factory C molasses production, kg.
MA = TC x Seas x Cf x MY

TC = Milling capacity, tonnes cane per day.
Seas = Days of mill operation per year.
Cf = Capacity factor: Average percent of capacity actually operated.
MY = C Molasses yield, kg molasses per tonne cane.

VC = Variable cost, $/kg molasses processed.

D. Electricity Costs ($/year):

EC = MA x EY x ELTC x AEC

MA = Annual sugar factory C molasses production, kg (section II.C).
EY = Ethanol yield, Liters ethanol per kg molasses.
ELTC = Electricity used per liter ethanol (kwh/liter).
AEC = Cogeneration facility's average electricity production cost, $/kwh (Section III.J).

E. Steam Costs ($/year):

SC = DSTM x STE x AEC

DSTM = Distillery's low pressure steam use, Kg (Section I.E).
DSTM = MA x EY x STL

MA = Annual sugar factory C molasses production, kg (Section II.C).
EY = Ethanol yield, liters ethanol per kg molasses.
STL = LP steam use, kg steam per liter ethanol.

STE = The amount of electricity (kwh) which can be generated by 1 kg of HP steam.
AEC = Cogeneration facility's average electricity production cost, $/kwh (Section III.J).

F. Molasses Costs ($/year):

MC = MA x CPR

MA = Annual sugar factory C molasses production kg, (Section II.C).
CPR = Cane price, $/tonne, for cane delivered to the factory.

G. Total Annual Cost ($):

$$TAC = ACR + FC + TVC + EC + SC + MC$$

ACR = Annual Capital Recovery, (Section II.A);
FC = Annual Fixed O & M Cost, (Section II.B);
TVC = Total Variable Costs, (Section II.C);
EC = Electricity Costs, (Section II.D);
SC = Steam Costs, (Section II.E);
MC = Molasses Cost, (Section II.F);

H. Production Cost:

$$PC = \frac{TAC}{TAP}$$

TAC = Total Annual Ethanol Production Costs.
TAP = Total Annual Ethanol Production, Liters.
TAP = MA x EY

MA = Sugar factory annual molasses production, kg (Section II.C).
EY = Ethanol yield, liters ethanol per kg molasses.

II. ANNUAL COGENERATION FACILITY COSTS

A. Annual Capital Recovery ($/year):

$$ACR = CAP \times \frac{i}{1-(1+i)^{-N}}$$

CAP = Total Capital Cost.
i = Discount rate (return on capital investment).
N = Expected life of facility (years).

B. Operation and Maintenance, Fixed Costs ($/year):

$$FC = La + Mn + Misc$$

La = All labor costs (salaries, wages, etc.)
Mn = Maintenance costs, excluding labor.
Misc = Any remaining miscellaneous fixed costs.

C. Operation and Maintenance, Variable Costs ($/year):

$$TVC = VC \times EP$$

VC = Cogeneration facility's variable costs, $/kwh.
EP = Total annual electricity production, kwh.
EP = [(ACM x BPC x HHV)+(CAP/Eff x OSL x OCf)] x Eff.

ACM = Tonnes cane milled annually (Section I.C).
BPC = Tonnes of 50% mc bagasse per tonne cane milled.
HHV = Higher heating value of 50% moist bagasse (kj/kg).
CAP = Generating capacity of the cogeneration facility,
Eff = Busbar electrical efficiency: fuel energy (HHV) in divided by electrical energy out (no steam exported to distillery).
OSL = Length of off-season (days, no milling).
OSf = Off season capacity factor, average percent of full capacity operated during the off season.

D. Electricity Credits from Distillery ($/year):

ECR = DEC + SEC

DEC = Distillery's annual electricity cost, \$/year (Section II.D).
SEC = Sugar factory's annual electricity cost, \$/year (Section I.D).

E. Steam Credits from Distillery & Suagr Factory (\$/year):

STC = DSC + SSC

DSC = Distillery's annual steam cost, \$/year (Section II.E).
SSC = Sugar factory's annual steam cost, \$/year (Section I.E).

F. Off Season Fuel Costs:

OSFC = CAP/Eff x OSL x OSf x OSFP

CAP = Cogen facility's electric generating capacity, kw (Section II.C).
Eff = Busbar electrical efficiency (Section II.C).
OSL = Length of Off (non-milling) season, days.
OSf = Capacity factor of off-season (section II.C).
OSFP = Off season fuel price, \$/GJ.

G. Bagasse Fuel Costs (\$/year):

TBC = (BP + BPC) * ACM * BPC

BP = Price per tonne of 50% mc Bagasse.
BPC = Bagasse processing costs, \$/tonne 50% mc bagasse.
ACM = Amount of cane milled annually (Section I.C).
BPC = Tonnes 50% mc bagasse per tonne cane milled.

H. Total Annual Cogeneration Costs:

TACC = CR + FC + TVC - ECR - SCR + OSFC + TBC

CR = Annual capital recovery (Section III.A);
FC = Fixed Costs (Section III.B);
TVC = Total Variable Costs (Section III.C);
ECR = Electricity Credit from distillery & sugar factory (Section III.D);
SCR = Steam Credit from distillery and sugar factory (Section III.E);
OSFC = Off Season fuel cost (Section III.F);
TBC = Total Bagasse Cost (Section III.G).

I. Average electricity production cost, \$ per kwh:

$$AEC = \frac{TACC}{EP - ED}$$

EP = Total Annual electricity production, kwh.
ED = Distillery's and sugar factory's annual electricity use, kwh (Sections I.D & II.D).
TACC = Total Annual Cogeneration Costs, \$.

APPENDIX VII:
DEVLOPING WOLRD POTENTIAL FOR CANE SUPPORTED ELECTRICITY-ETHANOL CO-PRODCUTION