ASSESSMENT OF NEW TECHNOLOGIES FOR CO-PRODUCTION OF ALCOHOL, SUGAR AND ELECTRICITY FROM SUGAR CANE

Joan M. Ogden
Mark Fulmer

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Center for Energy and Environmental Studies
Princeton University
Engineering Quadrangle
Princeton, New Jersey 08544

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I. BACKGROUND AND MOTIVATIONS FOR PRESENT STUDY

Electricity and ethanol from sugar cane are potentially important sources of energy in many cane-growing developing countries. The technologies for producing fuel ethanol from sugar cane and using it as an automotive fuel have been successfully demonstrated on a large scale in Brazil [1,2]. Several other developing countries are producing fuel ethanol from cane molasses [3-14] (Table 1), and in a few of these countries national alcohol fuel programs have been implemented or are being considered [1-4,9,14].

The technology for cogenerating electricity at sugar factories is well established. With commercially available high pressure condensing-extraction steam turbine (CEST) cogeneration systems, which have been demonstrated in several countries [15-19], it would be possible to generate up to 100 kilowatt hours per tonne of cane milled (kWh/tc), in excess of factory demands, during the milling season. (More electricity could be generated if the cogeneration system is operated year round. See Figure 1.) Biomass integrated gasifier/steam-injected gas turbine (BIG/STIG) cogeneration systems could potentially offer much higher electricity production -- more than twice that possible with CEST systems [20-22]. Moreover, because of their lower unit capital cost and higher electrical efficiency BIG/STIG systems could produce electricity at a lower cost than CEST systems. While BIG/STIG systems are not yet commercially available, this technology could be developed and commercialized within about 5 years [23].

The economics of sugar factory cogeneration would be attractive in many developing countries at today's commodity prices [12,15-17], particularly if gas turbine based cogeneration technologies were used [20-22]. (See Figure 2.)

Further innovations in gas turbine technology suggest that even lower cost electricity production may be possible than with the BIG/STIG system. For example, the biomass integrated gasifier/intercooled steam injected gas turbine (BIG/ISTIG), a modification of BIG/STIG, is projected to have higher electrical efficiency and lower capital costs per kW than BIG/STIG. Roughly 20% more electricity could be generated than with BIG/STIG (Figure 1), and at lower cost.

At present commodity prices (Table 2), ethanol from cane is not economically competitive as a motor fuel except in a few locations [3,14]. (In Brazil, for example, ethanol from cane would be economically competitive as an octane enhancing additive to gasoline, but not as an unblended ("neat") motor fuel. [See Figure 21]. And in some landlocked Southern African countries, high inland freight costs from coastal ports in South Africa or Mozambique have meant high prices for imported oil and low prices paid to sugar factories for molasses, which must be transported to the coast for export [3]. With low priced molasses feedstock and expensive oil, ethanol production at sugar factories becomes more economically attractive.)

There is reason to believe that the economics of fuel ethanol could be improved if the distiller sold cane residues (bagasse during the milling season and possibly barbojo -- cane tops and leaves -- during the off-season) as fuel to a cogeneration plant. Cane residues are a large energy resource, containing about 5 times as much energy as the alcohol which could be produced from cane (*). It is not surprising that sales of cane residues could yield substantial revenues to alcohol producers. For a typical autonomous distillery, the

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* One tonne of sugar cane yields about 70 liters of alcohol (with an energy content of 1.7 GJ), 300 kg of bagasse (with an energy content of 2.9 GJ) and 660 kg of barbojo (with an energy content of 6.3 GJ).

distiller could receive almost as much for selling cane residues as for selling alcohol (*). Sales of bagasse in season would increase the distiller's revenues by about 40%; if both bagasse and barbojo were sold the distiller's revenues would almost double.

In this study we first seek to understand how cogeneration effects the economics of alcohol production. And second, what the effect would be of using new, more efficient technologies for cogeneration and alcohol production. We have analyzed a range of cases:

* Autonomous distilleries and annexed distilleries with various product mixes of sugar and alcohol.

* Three types of biomass fired cogeneration systems : 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).

* Conventional and energy conserving distillery and sugar factory equipment.

* A range of commodity prices (sugar, ethanol, molasses, electricity) and economic assumptions.

Our primary aim is to identify conditions under which it may be possible to co-produce alcohol and electricity at costs competitive with other energy supplies.

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* Let us assume that alcohol is sold at 17 cents/liter, which would make it competitive with gasoline at the current world oil price, and that bagasse and barbojo are sold to the cogenerator at a price competitive with fuel wood. The alcohol distiller would receive about $11.9/tc for sales of alcohol. If bagasse were sold to the cogenerator during the milling season, the distiller would receive an extra $4.4/tc, an increase in revenues of almost 40%. If the distiller also sold barbojo as an off-season fuel an additional $5.7/tc would be earned (assuming a 160 day milling season). The total from sales of bagasse and barbojo would be about $10/tc, an 85% increase in the distiller's revenues.

II. SUGAR FACTORY COGENERATION TECHNOLOGIES

A. DESCRIPTION OF COGENERATION TECHNOLOGIES

1. CURRENT TECHNOLOGY

a. Cogeneration Systems in Typical Factories Today: Medium Pressure Steam Turbines

In most sugar factories and alcohol distilleries today, small, "medium pressure" (1.5-2.5 MPa) bagasse-fired steam turbine systems provide just enough steam and electricity to meet onsite factory needs, typically about 350-500 kilograms of steam per tonne of cane milled, and 15-25 kWh/tc [26,27]. Typically, factories are designed to be somewhat energy inefficient, consuming all the available bagasse while just meeting factory energy demands, so that excess bagasse does not accumulate and become a disposal problem. (In Southeastern Brazil, where bagasse is sold as a boiler fuel or as a component of cattle feed [24,25], some factories have been made more energy efficient, so as to free up surplus bagasse for these markets [28-31].)

b. High Pressure Condensing-Extraction Steam Turbines (CEST)

In a few sugar factories and alcohol distilleries, condensing-extraction steam turbine cogeneration systems (CEST) operated at higher pressure (4.0-8.0 MPa) have been installed [15,16,28]. With these systems, it is possible to produce enough steam to run a typical factory 350-500 kg/tc, plus 70-120 kwh/tc of electricity, or about 50-100 kwh/tc in excess onsite needs. The extra electricity can be made available to other users by interconnecting the cogenerator with the utility grid. During the milling season the CEST cogeneration system is fueled with 50% wet bagasse, as it comes from the mill. In the off-season the cogeneration system could use wood, barbojo (cane tops and leaves), or fuel oil as fuel.

2. FUTURE OPTIONS: BIOMASS-FIRED GAS TURBINE SYSTEMS

a. Biomass Integrated Gasifier/Steam-Injected Gas Turbine (BIG/STIG)

In a BIG/STIG cogeneration system biomass is gasified to form a low BTU gas, which fuels a gas turbine (Figure 3a). Steam is raised for the mills and process in a heat recovery steam generator (HRSG), which utilizes the hot exhaust gases exiting the turbine. Any steam not needed for the factory or the gasifier could be injected into the combustor or the turbine, boosting the electrical output and efficiency of the system.

During the milling season, the BIG/STIG system would be fueled with densified bagasse. The bagasse would probably be have to densified (either briquetted or pelletized) if used in a fixed bed gasifier, which was originally designed for use with coal and which may be used in first generation BIG/STIG units [34,35]. In the off-season the system could be fueled with densified wood or barbojo or distillate oil.

While BIG/STIG systems are not commercially available at present, they could be probably commercialized in less than five years [23,32]. Steam-injected gas turbine (STIG) systems have already been commercialized for use with natural gas, and there has also been development work on coal integrated gasifier/steam-injected gas turbines (CIG/STIG) [33]. The R&D effort required to commercialize the BIG/STIG system would be relatively modest, because much of the development work on coal integrated gasifier/steam-injected gas turbines, would be applicable to biomass [32].

b. Biomass Integrated Gasifier/Intercooled Steam-Injected Gas Turbine (BIG/ISTIG)

The BIG/ISTIG system is similar to BIG/STIG, except that the gas turbine has been modified to include an intercooler between stages of the compressor

(Figure 3b). This reduces the required compressor work and also allows operation at a higher turbine inlet temperature (because of improved cooling of the turbine blades), so that the power output and electrical efficiency are increased [36]. As with the BIG/STIG, the BIG/ISTIG would be fuelled with densified bagasse during the milling season and with densified wood or barbojo or distillate oil during the off-season.

Natural gas fueled ISTIG technology would require a 4-5 year development effort [33]. If this technology were developed at the same time as BIG/STIG technology, BIG/ISTIG technology could become commercially available shortly thereafter.

B. PERFORMANCE AND COST OF BIOMASS-FIRED COGENERATION SYSTEMS

The performance of various biomass-fired cogeneration systems is summarized in Table 3. The electricity and steam production are given for the CEST, BIG/STIG and BIG/ISTIG technologies for a range of system sizes in the cogeneration mode (production of both process steam and electricity) and in the power only mode. The BIG/STIG is more than twice as electrically efficient as the CEST system, and BIG/ISTIG system is almost three times as efficient. Performance figures for the BIG/STIG and BIG/ISTIG are based on estimates for coal gasifier/gas turbine systems adapted for operation with gasified biomass. (see Appendix A for details).

The installed capital cost and operation and maintenance costs are estimated for each system in Table 4. Cost estimates for the BIG/STIG AND BIG/ISTIG are based on estimates for the CIG/STIG AND CIG/ISTIG systems [33] minus the chemical hot gas clean-up section (Table 5), which would be required for coal, but not for low-sulfur biomass fuels. [34].

C. INTEGRATING COGENERATION SYSTEMS WITH SUGAR FACTORIES OR DISTILLERIES

1. COGENERATION CONFIGURATIONS

Figure 4 illustrates how various cogeneration systems could be integrated with a sugar factory or alcohol distillery. Figure 4a shows a typical existing sugar factory steam turbine cogeneration system, producing just enough steam and electricity for factory needs. In Figure 4b a CEST cogeneration system is shown. Steam is extracted at medium pressure (1.5-2.0 MPa), for use in the factory as before. About 50-100 kwh/tc of excess electricity is produced during the milling season. A biomass gasifier/gas turbine cogenertaion system is shown in Figure 4c. With the BIG/STIG system about 235 kwh/tc of excess electricity is produced during the milling season. With a BIG/ISTIG system (not shown) excess electricity production would increase to about 280 kwh/tc in season.

2. MEETING FACTORY STEAM DEMANDS

In order to use a particular cogeneration system at a sugar factory or alcohol distillery, the system must meet the factory's steam and electricity needs. In Figure 5, we have plotted the steam and electricity production for CEST, BIG/STIG and BIG/ISTIG cogeneration systems, operated on biomass fuel. For each technology, a range of operating values is possible, depending on how much steam is produced for process. For each system, the lower the process steam demand, the higher the electricity production. When the process steam demand is zero, as in off-season operation, electricity production is maximized. The maximum steam production possible with the system is given at the right hand endpoint of each line. Also shown are the steam and electricity demands in an average sugar factory or alcohol distillery.

The more electrically efficient gas turbine cogeneration systems have lower steam production than the CEST system. With BIG/STIG the maximum steam production is about 300 kilograms of "medium pressure" (2.0 MPa, 316^{o}C) steam per tonne of cane milled (kg/tc). With BIG/ISTIG maximum steam production is about 235 kg/tc. The average sugar factory or alcohol distillery requires about 350-500 kg/tc of process steam, so that some factory steam economy measures would be required in order to meet the factory steam demand, if gas turbine cogeneration systems are used.

III. TECHNOLOGIES FOR PRODUCING ETHANOL FROM CANE

A. ETHANOL PRODUCTION IN AUTONOMOUS AND ANNEXED DISTILLERIES

Ethanol can be produced directly from cane juice in "autonomous" distilleries or from molasses in "annexed" distilleries associated with a sugar factory. The various steps of ethanol production in autonomous distilleries and in annexed distilleries are described below.

1. Autonomous distilleries

In an autonomous distillery (Figure 6) cane is milled, and the raw cane juice is filtered and heated. In some Brazilian autonomous distilleries, the juice is then cooled and sent directly to the fermentation stage. Alternatively the juice can be limed and clarified, and the clear juice is often concentrated in an evaporator from typical values of 13^{o} Brix (% solids in juice) to about 18-20^{o} Brix, which is preferable for fermentation. The concentrated juice is then fermented and distilled to produce ethanol.

a. Steam and electricity demands in typical autonomous distilleries

Steam and electricity use in a typical Brazilian autonomous distillery are shown in Table 6 and Figure 7 [28]. A conventional distillation system requiring 3.3 kilograms of low pressure (0.15-0.25 MPa) steam per liter of hydrous ethanol is used. The overall factory steam demand for this case is 466 kg/tc (or 5.6 kg steam/liter alcohol). (A recent survey of autonomous distilleries in the Sao Paulo area indicated steam use of 420-550 kg/tc [27].) With innovative distillation systems employing higher pressure distillation columns and heat integration, it would be possible to reduce the steam demand to less than 235 kg/tc. (See sections III.B.4 and IV.A.2 for a detailed discussion of steam economy.)

b. Autonomous distillery costs

Published estimates of the installed capital cost of autonomous distillery equipment vary widely. This is illustrated in Figure 8, where the installed capital cost of autonomous distilleries is plotted as a function of plant capacity, based on estimates from various sources [2,8-10,13,40,41]. Tables 7a and 7b give disaggregated capital costs for "high cost" and "low cost" autonomous distilleries. The high cost case is based on a study performed by Francis Schaeffer and Associates for an autonomous distillery in Costa Rica [13]; the low cost case is based on Brazilian experience [1,2,24].

Capital costs shown in Figure 8 are for conventional low pressure distillation equipment. There would be little difference in cost between distilleries producing hydrous and anhydrous alcohol. Capital costs for low energy use Brazilian autonomous distilleries could be up to 20% more expensive than the conventional distilleries because more expensive higher pressure distillation columns and mills would be used [42].

It is an interesting feature of Figure 8 that Brazilian autonomous distillery costs are typically only one half to one third those quoted by US engineering firms.

Operation and maintenance costs for autonomous distilleries are given in Tables 7a and 7b.

2. Annexed distilleries

In a distillery annexed to a sugar factory both sugar and ethanol are produced (Figure 9). Cane is milled, and the raw juice is heated, clarified and evaporated to typical concentration of 60-65^{o} Brix (60-65% solids in juice) as in sugar making. One or more strikes of sugar are made in vacuum pans. The final molasses is diluted and used as the feedstock for ethanol fermentation and distillation. In many factories three strikes of sugar are produced and C-molasses (diluted with water or cane juice to reach Brix of 18-20) is the feedstock for fermentation. It is also possible to make less sugar and more alcohol. For example, the factory could produce only one strike of sugar and use diluted A-molasses as a feedstock for fermentation. Because more sugar is contained in A-molasses than in C-molasses, considerably more ethanol can be produced per tonne of cane milled.

In some Brazilian factories a flexible product strategy is employed, with the mix of products shifting between sugar and alcohol depending on the markets. In these factories a portion of the juice is used to dilute molasses or concentrated syrup, rather than being evaporated. The resulting mixture of molasses and juice (or syrup and juice) is the feedstock for fermentation.

a. Steam and electricity demands in typical annexed distilleries

Steam and electricity use in a typical sugar factory with an annexed distillery are shown in Table 8 and Figure 10 [28]. Conventional sugar factory and distillery equipment is used. The overall steam demand in this case is 464 kg/tc.

b. Annexed distillery costs

The installed capital cost of annexed distilleries is plotted versus capacity in Figure 11. (Only the distillery section costs are included in this figure.) Disaggregated costs for a sugar factory with an annexed distillery are given in Table 9, based on a study by F.C. Schaeffer [13]. The costs for the distillery section are consistent with those quoted by Brazilian manufacturers [41], and similar to those of a distillery recently installed at Bernard Lodge, Jamaica [6,43].

Operation and maintenance costs for annexed distilleries are given in Table 10.

B. PROCESS TECHNOLOGIES FOR AUTONOMOUS AND ANNEXED DISTILLERIES

In order to use the more electrically efficient gas turbine cogeneration systems at alcohol distilleries, factory steam demands must be reduced. There are many possible design options for autonomous and annexed distilleries (Table 11). Here we describe the various technologies which could be used for each stage of the process, noting opportunities for process steam conservation. The costs of alternative process technologies are given in Table 12.

1. Cane Milling

After delivery to the factory, the cane is chopped, shredded and put through a series of roller mills to extract the cane juice. In this section we consider alternative technologies for cane milling.

a. Steam driven mills

Virtually all commercial cane mills are steam-driven using small back-pressure steam turbines, which run on superheated steam at 1.5-2.0 MPa, 250-350°C. After being chopped and shredded, the cane passes through a series of 15-24 roller mills, where juice is extracted, leaving bagasse fiber. Between mills, imbibition water is sprayed onto the bagasse to aid in sugar extraction. Typically over 90% of the sugar is recovered. The mill steam consumption is roughly 200-250 kg/tc (assuming a mill exhaust temperature of 120°C at 0.2 MPa). A description of steam consumption in mills is given in Appendix B.

b. Electric mills

Although few factories in the world use them at present, electric motors can be used to drive cane mills [26,43]. Using electricity for mills means a double transformation of energy (from bagasse to steam, from steam to electricity) and the fuel input is therefore 15-20% greater than with steam driven mills and about 10-15 kWh/tc of electricity is required. Although steam turbines are more flexible than induction motors at low and variable power loads, electric motors are much cleaner and easier to start and stop, and they have lower operating and maintenance costs. Moreover, variable speed drives could used to better match the electric motor to variable load conditions, reducing electricity consumption.

c. Diffusers

In a few locations diffusers are used instead of mills. Here the cane is chopped and run through a single mill to extract perhaps 70% of the sucrose. The bagasse is washed in a tank, bringing the sucrose extraction up to 98%. The resulting 70% wet bagasse or "megasse" is then dewatered in a second 3-roller mill or in an extractor (a press) to 50% moisture. The diffusion process runs on electricity rather than steam, except for the mills which can be steam driven.

d. Opportunities for saving steam

Both electric mills and diffusers could reduce the steam required for milling. (See Table 13.) In most factories low pressure steam from the mill exhaust is utilized for the process. Unless low pressure process steam demands are reduced below 200-250 kg/tc, electrifying the mills or using diffusers would not reduce the overall factory steam consumption.

2. Juice heating

a. Present practice: use steam or vapor

In most sugar factories and distilleries, raw juice is heated with vapor bled from the evaporators or sometimes with low pressure exhaust steam from mill or turbo-alternator turbines. Shell and tube heat exchangers are used, with the steam or vapor condensing on the hot side and juice heated on the cold side. Clear juice heaters typically use steam or bled vapor as well.

b. Steam saving option: use hot condensates

By using the hot condensates from the evaporator (and in an annexed distillery from the vacuum pans) a portion of the juice heating could be done and some steam could be saved. In sugar factories with an annexed distillery hot condensates could do a significiant fraction of the juice heating and the process steam could be reduced about 5-10% [44].

3. Evaporation

a. Present practice: short tube rising film evaporators

In most factories, forward feed, multiple effect, short tube rising film evaporators are used. In an autonomous distillery juice is generally concentrated from 13°Brix to about 18-20°Brix prior to fermentation. In a sugar factory with an annexed distillery, the juice is concentrated to 60-65°Brix for sugar making. Vapor is bled from the evaporator for juice heaters, distillery and vacuum pans. The first one or two effects run at slightly above atmospheric pressure, with the later stages running at a slight vacuum. Vapor from the final stage of the evaporator is fed into a barometric condensor to maintain a pressure gradient throughout the system.

b. Steam saving options

1). Falling film evaporators

Falling film evaporators are often used as energy savers in the beet sugar, pulp and paper and dairy industries, and are being studied for use in the cane sugar industry [45,46]. They have the advantage of higher juice flow velocity and higher heat transfer coefficients and can, therefore, run at smaller temperature differences between effects. Because the juice travels through a falling film evaporator three to four times more quickly than in a short tube rising film type, it is likely that higher input steam temperatures (up to 130-135°C) could be tolerated without damage to the juice due to inversion of sugars and color formation (*).

With an input steam temperature of 135°C, it would be possible to run the entire evaporator at pressures above atmospheric and utilize the vapor from the later effects for juice heating and vacuum pans. Vapor bleeding from later effects rather than from the first effect makes better use of the multi-effect

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* The issue of color formation during evaporation is a still topic of research [47]. However, results from the falling film evaporator operated by GTZ/SIRI in Jamaica indicate no problems with color formation at these temperatures.

configuration. In a sugar factory with an annexed distillery this could reduce the overall process steam consumption by 30-45% [44].

2). Mechanical vapor recompression (MVR)

In this technology, vapor evaporated from the juice is compressed to a higher pressure and temperature and fed back into the steam side of the evaporator (Figure 12). If no vapor bleeding is done, the amount of vapor driven out of the juice is approximately equal to the amount of vapor which is recirculated and then condenses on the steam side. Thus, little or no external steam is needed after startup and electricity for the compressor is the only energy input. If vapor bleeding is done, of course, approximately the same amount of steam must be input to make up for its loss. The condensate from the evaporator is typically used to heat the incoming juice.

With MVR electricity is substituted for steam use. The compressor power needed depends sensitively on the compressor pressure ratio. (See Appendix B.) The smaller the pressure difference (or equivalently the temperature difference between the condensing steam on the hot side and the vapor on the cold side of the evaporator), the less power consumed. Falling film evaporators have a higher heat transfer coefficient than short tube rising film type evaporators, and the same amount of evaporation can be accomplished with a lower temperature difference. For this reason, falling film evaporators are particularly well suited for MVR systems.

4. Vacuum pans

Vacuum pans are used for sugar making in factories with annexed distilleries.

a. Present practice: Batch or discontinuous vacuum pans

In batch or discontinuous pans, the thick syrup or massecuite is boiled down one batch at a time, in several stages or strikes. Because of the water added in washing, molasses dilution and agitation, it takes about 1.2-1.7 kg of steam to evaporate 1 kg of vapor from the massecuite in each pan. Steam consumption values reported in the literature range from 120-170 kg/tc [26], depending on the design of the vacuum pan. (See Appendix B.)

b. Energy saving options:

1). Continuous vacuum pans

Continuous vacuum pans have the advantage of lower steam consumption and constant steam loads, and are coming into increasing use. Agitation can be done with the non-condensable gases vented from the pan, or with a little extra steam. Hugot [26] estimates that the steam consumption for a continous pan should be about 25% less than for a discontinuous pan.

2). Multiple effect vacuum pans

If several continuous vaccuum pans were run as a multiple effect vacuum pan, the steam consumption would be reduced by a factor approximately equal to the number of effects. This is being tried experimentally by some vacuum pan manufacturers [48].

3). Mechanical vapor recompression for vacuum pans

Mechanical vapor recompression could in theory be used on the vaccuum pans, replacing some of the steam demand. Because of the high boiling point rise at 60-96°Brix, the power consumption of an MVR system would be high. Moreover, the recompressed vapor would still have to be supplemented with some extra steam, when steam agitation or pan washing was needed. For each kg of water evaporated, we would need about 0.5 kg of external steam, and 1 kg of recompressed steam. Some experiments have been tried with MVR systems on vacuum pans. However, the crystallization process was apparently more difficult to control. This idea needs more work to establish its practicality [48].

5. Fermentation and Distillation

a. Fermentation

Prior to fermentation, sucrose in the fermentation feedstock is hydrolysed into the invert sugars, D-glucose and D-fructose.

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

The enzyme invertase occurs naturally in cane, and sugar cane molasses contains high proportions of invert sugars. In autonomous distilleries the cane juice is sometimes further hydrolysed before fermentation by adding acids.

Yeast is then added to the fermentation feedstock. Because yeast growth is inhibited by high ethanol concentrations, the solution to be fermented (concentrated cane juice or diluted molasses) usually starts with about 15-18% fermentable sugars (or 18-20^{o}Brix), resulting in a final beer with ethanol concentrations of no more than 9%-10%.

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

$$\underset{\text{(invert sugar)}}{C_6H_{12}O_6} = \underset{\text{(ethyl alcohol)}}{2C_2H_5OH} + \underset{\text{(carbon dioxide)}}{2CO_2} + 31{,}200 \text{ cal}$$

Only a small portion of the energy released is used by the yeast for metabolism and growth; most is liberated in the form of heat. Because the optimal temperature for ethanol fermentation is between 32 and 38^{o}C, and the reaction is exothermic, some cooling of the fermenting beer is required. Typically, 90% of the theoretical yield of ethanol is achieved [49].

1). Commercially available fermentation technologies

a). Batch fermentation

In virtually all cane sugar factories today, ethanol fermentation is a batch process, where the sugar-water-yeast mixture is placed in a large vat and allowed to react until the desired alcohol concentration is reached [24,50]. The vats are kept at a constant temperature around 32^{o}C using cooling coils, and usually agitated to promote mixing. The yeast cells can be separated from the beer prior to distillation and recycled. Typical productivity for batch fermentation is 1.8 to 2.5 grams ethanol per liter of beer per hour (1.8 to 2.5 g/l/h) [50].

b). Continuous fermentation: Biostil

Continuous fermentation has the potential to increase productivity over the batch process: ethanol yields of 6 to 80 g/l/h have been reported [48]. Continuous fermentation is of interest because higher productivities could eventually lead to lower equipment costs for a given ethanol production rate. While a number of continuous fermentation techniques have been tried in the laboratory, few have been demonstrated at the industrial scale. (Table 14.)

One promising commercial continuous fermentation method is the Biostil process [50,51], which has been developed by Alfa-Laval of Sweden (Figure 13a). This system 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. Productivities of 10 g/l/h are acheived [48].

One advantage of the Biostil system is that the fermentation feed can have much higher sugar concentrations (because ethanol inhibition is not a factor), and thus the stillage is more highly concentrated than in a conventional distillery, reducing the energy demand for stillage evaporation prior to disposal. Biostil has been demonstrated at the industrial level in Australia, Sweden and Brazil [50,51], but is not currently in commercial use in the Brazilian cane alcohol industry, because of operational problems [24].

2) Continuous fermentation technologies under development

a) Flocculent yeast fermentation

Flocculent yeast fermentation simplifies the yeast separation procedure by replacing the centrifugal yeast separator with a simple settling tank, reducing both initial and operational costs. It however presents serious drawbacks in yeast selection; other desirable yeast traits must be sacrificed in order to use a flocculent strain. Still, flocculent yeast allows higher yeast concentrations in the fermentor and thus increasing productivity to 30-40 g/l/h.

b) Tower fermentor

A more advanced application of flocculent yeast is the tower fermentor (Figure 13b). The sugar/molasses solution is fed into the bottom of the tower, and slowly rises up the tower through the flocculent yeast. The ethanol concentration of the feed increases as it 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. The tower fermentor has been used to a limited extent in potable beer fermentation, but has yet to be demonstrated in ethanol production [50,51].

c) Membrane separated fermentor (Rotoferm)

Another method of maintaining high yeast cell densities and thus high productivity is by physically separating the yeast from the inhibiting ethanol and by-products by a filter or membrane. One application of this method is the rotofermentor (Figure 13c), a cylindrical filter membrane through which the fermenting feed is drawn. The cylindrical filter is rotated at high speed (approx. 4500 RPM) 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.

d) Vacuum fermentation (Vacuferm)

Another alternate fermentation process is Vacuferm (Figure 13d). Vacuferm bypasses the ethanol separation problem by maintaining the fermentation tank at a low pressure (30-35 mm Hg) so that 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. A serious drawback of the Vacuferm process is the large amount of energy required to pump off the CO_2 to maintain the required low pressure. Although the method can have productivities of 40 to 80 g/l/h in the laboratory, it has yet to be demonstrated at the industrial scale [51].

b. Distillation/Ethanol separation

Once fermentation is complete the ethanol must be separated from the fermented beer, an ethanol/water mixture which is typically 8-10% ethanol by weight. Distillation is by far the most common ethanol separation process, although a number of other separation techniques have been developed [52,53,54]. (See Table 15 for a summary of ethanol separation technologies.)

Distillation exploits the fact that a multi-component mixture will have different component concentrations in the vapor and liquid phases. In the case of ethanol and water, ethanol being the more volatile component will be more highly concentrated in the vapor phase of the solution than in the liquid phase. Thus water can be separated out of the solution to the desired ethanol concentration.

This is accomplished by the use of a fractioning distillation column (Figure 14). The mixture to be distilled is introduced into the column at various locations along the column depending upon the separation desired and design. Wherever it is introduced, the liquid phase flows downward to a reboiler, where heat is added (often by steam or hot vapors from another column) to vaporize a majority of the mixture. The hot rising vapor, rich in the more volatile component exchanges heat with the cooler, falling, liquid which is rich in the less volatile component. As the two streams interact, the less volatile component condenses out of the rising vapor and the more volatile component evaporates out of the falling liquid. The final volatile component-rich vapor is removed off of the top of the column where it is condensed and a portion is reintroduced into the column near the top as reflux for increased separation efficiency. The less volative component rich liquid is removed from the column at the reboiler. Streams of a third intermediately volatile constituent may be removed from some intermediate location of the column.

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

For ethanol distillation, the purity of the alcohol produced is the strongest factor determining the type of distillation equipment required. Hydrous alcohol (96%-96.5%) is used for unblended motor fuel; absolute or anhydrous alcohol (99.9%), is often used for blending with gasoline. The purer the final ethanol produced, the more complex the distillation process.

1) Commercially available distillation technologies

a) Production of hydrous ethanol

The first column in an ethanol distillery (the beer still) separates a majority of the water from the fermented beer, drawing off 70%-90% ethyl alcohol from the top of the column. The ethanol-water mixture is then introduced into a stripping/rectifying column. In the stripping section of the column more water is removed as the bottom product and the ethanol is removed at an intermediate location along the column. If further distillation is required, the ethanol-water mixture is fed into a rectifying column. The rectifying section removes low boiling organic impurities (fusel oil) and any remaining water, resulting in a product up to 96.5% pure hydrous ethanol.

At this point the output is hydrous ethanol. Further distillation will not remove any more water from the ethanol because the solution is an azeotrope, e.g. it has a single boiling point, implying that the vapor and liquid phases of the mixture have the same composition.

Figure 15a shows a flow diagram for a Brazilian distillery which produces hydrous ethanol using two distillation columns. The beer still is divided into three sections (A, A1, and D) and the rectfying/stripping column into two sections (B, B1). Low pressure steam (at 0.15-0.25 MPa) is input to each column to provide heat for the process [31].

b) Production of anhydrous alcohol

There are a number of methods in which the 96.5% ethanol-water azeotrope can be broken and the ethanol further purified.

(1). Azeotropic distillation

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 is to be suppressed, the process is called azeotropic distillation. Benzene, trichlorethylene, ethyl ether, or a mixture of benzene and gasoline (amoung others) can be added. For example, if benzene is added to the water-ethanol mixture, the water can be evaporated off at 64.85^{o}C, leaving benzene and ethanol. The benzene can then be evaporated off leaving pure ethanol. This is done is a single distillation column, with water and benzene drawn off the top and absolute ethanol left as the bottom product. The benzene is easily separated from the water and reused.

A Brazilian distillery producing anhydrous alcohol via azeotropic distillation with benzene is shown in Figure 15b [31]. In addition to the beer still and stripping/rectifying columns, two additional columns are added for dehydration and benzene recovery.

(2). Extractive distillation

If the boiling of water is to be suppressed, glycerol, ethylene glycerol, or a mixture of glycerol and potassium carbonate can be used as a dehydrating agent [52]. This process, "extractive distillation," is commonly accomplished by heating the ethanol-water azeotrope in the reboiler of a column, and have the rising vapors mix with the falling liquid stream of the dehydrating agent. As the two streams mix, the water is removed from the vapor, allowing anhydrous ethanol to be drawn off the top of the column. Again, the water and dehydrating agent are easily separated and the dehydrator recycled.

(3). Water adsorption

Alternatively, water can be removed from the azeotrope via adsorption [52]. The energy consumed in this process is exclusively to regenerate the adsorbent material. Commercially available water adsorbent materials include molecular sieves (zeolites) and corn meal.

2). Energy use in distillation

a). Energy use in commercial distillation systems

Until about 10 years ago, relatively little attention was paid to energy efficiency in cane alcohol distilleries, because there was little incentive to save bagasse fuel and because alcohol was a minor byproduct of sugar production, with only a modest steam demand for distillation. However, with the widespread use of autonomous distilleries in Brazil (where alcohol is the main product and distillation represents a major steam demand) and the possibility of using bagasse for fuel or other uses, energy conserving distillation systems have been developed.

Tables 15a and 15b shows the steam demand for distillation in conventional and innovative Brazilian distillation systems [1,2,24]. In the innovative system the steam demand is reduced by:

1) Use of more efficient trays in the distillation columns

2) Higher pressure steam input (0.6 MPa) to the rectifying column, which

allows heat from the rectifying column or azeotropic separation column vapors to be recovered for heating the beer still column.

3) Heat integration (Use of indirect heating in the stripping column; concentration of stillage in thin film evaporators and re-use of the vapors for heating the distillation columns.)

4) Process instrumentation and control

b). Reducing energy use in distillation

A number of techniques have been used and proposed to reduce the energy required for distillation.

(1). Heat integration

Heat integration techniques attempt to use hot and cold streams within the factory as efficiently as possible to accomplish the required heating with a minimum of external steam input. For instance, the heat given off to condense the vapor drawn from the top of the rectifying column can be used to pre-heat the incoming beer or heat liquid in the reboiler of another column (Figure 16).

Extensive research has gone into theoretically optimizing such distillation problems [52,55]. Examples of optimized heat integrated systems for production of hydrous and anhydrous ethanol are shown in Figure 16a and 16b, based on systems are used commercially for production of ethanol from corn. The steam use is 1.2 kg/liter for motor grade hydrous ethanol and 2.2 kg/liter for anhydrous ethanol [55]. Even lower steam uses have been reported [52]. Energy use for various heat integrated designs are listed in Table 15.

(2). Mechanical vapor recompression

Mechanical vapor recompression systems can be used in distilleries (Figure 17). The overall heat required for distillation is reduced to about 1700 kJ/liter for hydrous ethanol and 2200 kJ/liter for anhydrous ethanol. (This assumes that heat is converted into mechanical energy at 33% efficiency.)

3). Other ethanol separation techniques

Dramatic energy savings are theoretically possible if non-distillation separation techniques are used [52,53]. Battelle Pacific Northwest Laboratories have demonstrated in the laboratory a technique in which a solvent is contacted with the ethanol-water mix and which 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. Other methods include contacting the water-ethanol mixture with liquid carbon dioxide to extract the ethanol and flashing off the CO_2, or blending the mixture with gasoline, and freezing out the water, leaving gasohol. These and other similar methods show promise, but have only been demonstrated at the laboratory scale. The energy use for several experimental ethanol separation processes is shown in Table 15.

IV. DESIGN OF AUTONOMOUS AND ANNEXED DISTILLERIES WITH COGENERATION

Here we address the design of autonomous and annexed distilleries with cogeneration. We consider three types of cogeneration systems for autonomous and annexed distilleries, the CEST, BIG/STIG and BIG/ISTIG. From Figure 5 we see that the average distillery requires 350-500 kg/tc of process steam. This amount of steam could be supplied by the CEST system. The maximum steam production with the BIG/STIG system is about 300 kg/tc and with BIG/ISTIG about 235 kg/tc. Thus some factory steam economy measures would be necessary in order to use the BIG/STIG and BIG/ISTIG cogeneration systems at a distillery.

Design options for distilleries with cogeneration are shown in Tables

11, 16 and 17. In this section we examine steam economy measures in distilleries and present autonomous and annexed distillery designs matched to the output of each cogeneration system.

A. AUTONOMOUS DISTILLERIES WITH COGENERATION

1. Conventional autonmous distillery

Figure 7 shows a factory flow diagram for a conventional Brazilian autonomous distillery, which mills 125 tonnes of cane per hour to produce 10,000 liters of hydrous ethanol per hour (or 80 liters ethanol per tonne of cane) [28]. The juice is heated, clarified, concentrated in a conventional short tube rising film 3-effect evaporator, fermented and distilled in a conventional two column distillation system requiring 3.3 kg steam per liter of hydrous ethanol produced. The overall steam demand for the factory is 466 kg/tc (or 5.8 kg/liter alcohol). This similar to the amount of steam required in the average sugar factory. The CEST cogeneration system could meet the process steam demand for this type of factory, producing about 100 kWh/tc of excess electricity during the milling season (Table 16).

2. Steam conserving autonomous distillery designs

There are a number of opportunities for steam conservation in autonomous distilleries. From Table 16 we see that the largest single use of steam in the conventional factory is for distillation, about 256 kg/tc or 3.3 kg steam per liter of hydrous ethanol. As has been demonstrated in Brazil the steam used for distillation can be reduced from 3.0-3.5 kg/liter to about 1.5-2.0 kg/liter by using a higher pressure distillation system with heat integration [2]. The next largest use of steam in the factory is for juice processing (juice heating and evaporation to 18^{o}Brix). The total amount of steam used for juice heating plus evaporation can be reduced by using a 5 effect evaporator and by bleeding vapor from the evaporator for juice heating. The factory flows for a steam conserving factory with a low steam use distillery (1.9 kg steam per liter of ethanol) and an energy efficient juice processing system are shown in Figure 18a [28]. The steam demand for this case is 258 kg/tc (or 3.2 kg/liter), low enough that a BIG/STIG cogeneration system could be used (Table 16).

Further reductions in the distillery steam use could be achieved, by reducing the distillery steam demand to 1.5 kg/tc via heat integration. The overall factory steam demand would then be about 223 kg/tc or 2.8 kg/liter (Table 16), and the BIG/ISTIG cogeneration system could be used. A flow diagram for this type of factory is shown in Figure 18b.

It may be possible to design distillation systems with lower energy use. Heat integrated distillery designs using 1.2 kg/liter have been reported [52,55]. If electricity is used to substitute for steam via distillery vapor compression systems even lower steam use may be possible [52,54,56,57].

B. ANNEXED DISTILLERIES WITH COGENERATION

1. Conventional annexed distillery

a. Ethanol from C-molasses

Figure 19a shows a factory flow diagram for a conventional annexed distillery making ethanol from C-molasses. As in a conventional sugar factory, cane is milled, and the raw juice is heated and clarified. A short tube rising film quadruple-effect evaporator is used to concentrate the juice to 65^{o}Brix, and three strikes of sugar are made in batch type vacuum pans. The final C-molasses is diluted and used as a feedstock for fermentation. A conventional distillation column is used, requiring 4.0 kg steam per liter of hydrous ethanol produced. We have estimated the factory steam use to be 423 kg/tc or 5.3 kg/liter (Table 17). (Steam use is calculated using the factory model outlined in Appendix B, based on equipment in the Monymusk factory in Jamaica [6,44] with a conventional distillation system added.) About 9 liters of ethanol are produced per tonne of cane. (See Appendix F.) The CEST cogeneration system could meet the process steam demand for this type of factory, producing about 100 kWh/tc of excess electricity during the milling season (Table 17).

b. Ethanol from A-molasses

Considerably more ethanol (about 25 liters/tc) can be produced if only one strike of sugar is made, and A-molasses is used as a feedstock for ethanol production. (See Appendix F.) A larger capacity distillation section is used and the steam required for distillation is increased. However, the overall factory steam demand is about the same as before, because less steam is needed in the vacuum pans for sugar making.

c. Flexible product configuration

In some Brazilian distilleries, it is possible to vary the amount of juice used to make sugar or alcohol, depending on the market demands for these products. In these factories, the overall factory steam demand would change little as the mix of products is changed.

2. Steam conserving annexed distillery designs

Most of the steam use in an annexed distillery is for sugar making (Table 17). The largest single use of steam is for evaporation. Juice heating and vacuum pans are the next largest users of steam. Distillation accounts for less than 10% of the factory demand, if ethanol is made from C-molasses. Not surprisingly, many of the techniques for saving process steam in an annexed distillery would be similar to those used in a sugar factory [44]:

1) Heat can be recovered from hot condensates for juice heating, rather than using vapor bled from the evaporator, saving about 5-10% of steam use.

2) If a falling film evaporator is used, the evaporator can be run at higher temperature and pressure allowing more efficient use of the multiple effect configuration and reducing steam use by 30-45%.

3) With continuous vacuum pans, which require about 25% less steam than batch type, the factory steam use can be reduced by about 5-10%.

4) In addition, a more efficient distillation system can be used, cutting the steam use from 4 kg/liter to perhaps 1.5 kg/liter.

A steam conserving annexed distillery is shown in Figure 19b. Steam use is reduced by recovering heat from hot condensates, using a quintuple effect falling film evaporator, continuous vacuum pans and a low energy use distillery. The factory steam use is about 262 kg/tc, which would allow the BIG/STIG cogeneration system to be used.

Even lower steam use would be possible, about 220 kg/tc, if mechanical vapor recompression is used to pre-concentrate the juice from 13^{o} to 20^{o}Brix (Figure 19c). The electricity use for mechanical vapor recompression would increase by 7 kwh/tc. In this case, the BIG/ISTIG cogeneration system could supply the factory's process steam demand.

We have shown steam conserving designs for annexed distilleries making ethanol from C-molasses. If A-molasses were the feedstock, similar measures could be employed to save steam and the overall factory steam use would be about the same as with ethanol from C-molasses.

V. ECONOMICS OF ALCOHOL AND ELECTRICITY PRODUCTION

A. CALCULATING THE COST OF PRODUCING ALCOHOL AND ELECTRICITY FROM SUGAR CANE

In this section we calculate the cost of producing ethanol and electricity from sugar cane for a range of cases. For each case, it is assumed that the distiller sells bagasse fuel to the cogenerator during the milling season, and pays for process steam and electricity. (Figure 20). During the off season, the cogeneration plant would operate on some alternative fuel such as wood, barbojo (possibly purchased from the distiller) or oil.

The main assumptions underlying our cost calculations are outlined in Table 18. (See Appendices C and D for a discussion of the cost equations used.)

B. CASE STUDIES: COGENERATION AT ALCOHOL DISTILLERIES

1. AUTONOMOUS DISTILLERIES

a. Production cost of ethanol and electricity

Our results for autonomous distilleries are summarized in Figures 21-24. In Figure 21 we estimate the production cost of ethanol and electricity for a total of six cases: for each of the three cogeneration systems (CEST, BIG/STIG and BIG/ISTIG) and for two sets of distillery options, corresponding to high and low cost conditions. (See Tables 7a and 7b.) It is assumed that the distiller sells bagasse to the cogenerator during the milling season, and pays for process steam and electricity. During the off-season, the cogenerator uses barbojo or some other alternative fuel. For each case, a range of ethanol and electricity costs are shown, corresponding to different revenues received by the distiller from sales of cane residues to the cogenerator. As the price received for cane residues increases (moving from left to right along each line), the cost of ethanol production decreases, because the distillery receives higher revenues from selling its bagasse to the cogenerator. The cost of electricity increases because the cogeneration plant pays the distillery a higher price for its fuel. Furthermore, in each of the six cases two lines are shown. The steeper line is calculated assuming that revenues from the sale of both bagasse and barbojo are credited against the cost of alcohol production. (This would be the case if the distiller owned the cane fields and sold barbojo to the cogenerator.) For the other line only bagasse revenues are credited against alcohol production, which would be the case if off season fuel were not obtained from the distiller (e.g. barbojo might be purchased directly from the cane grower or another off-season fuel might be used).

Along each line two points are shown corresponding to different bagasse credits. The left hand endpoint shows indicates the ethanol and electricity cost with each technology, when no net payments are exchanged between the cogenerator and the distillery: bagasse is given to the cogenerator in exchange for the steam and electricity needed to run the plant. The circle shows the production costs, when the average total cost of fuel to the cogenerator (purchase price of bagasse plus the cost for processing the fuel into a gasifiable form) is \$3/GJ. This the estimated cost of air-dried woodchips, a possible alternative cogeneration fuel [58].

b. Conditions for competitive production of ethanol and electricity

For ethanol to be economically attractive as a motor fuel it must compete with other fuels. Similarly, cogenerated electricity must be competitive with other electricity supplies. For comparison, we have indicated in Figure 21 the cost of ethanol (16 cents/liter) which would be equivalent to the current US wholesale gasoline price of \$0.75/gallon [59]. Also shown is the value of ethanol as an octance enhancing additive (24 cents/liter) at current world oil prices (*). The cost of electricity from a new hydropower plant costing $1500/kW is plotted at about 4.5 cents/kwh. The fuel cost for a thermal power plant burning oil is shown at 3.6 cents/kwh, based on the current residual oil price. Also shown is the electricity cost for a new coal fired power plant.

Figure 21 suggests that under some conditions it may be possible to co-produce both ethanol and electricity at economically competitive costs. Several key factors were identified.

For economically viable ethanol production in autonomous distilleries, it is crucially important to reach low cost conditions quoted for Brazil. Ethanol costs are about <u>33 cents per liter lower</u> with Brazilian cost estimates [1,2,24,59] than with those of F.C. Schaeffer [13]. About 15 cents of the difference is due to the difference in cane cost. (The Schaeffer study is based on Costa Rican factory delivered cane costs of $18.9/tonne cane, while the Brazilian studies are based on actual costs in Brazil of $8/tonne cane.) Most of the remaining difference (about 11 cents per liter) is due to the lower capital cost estimated for Brazilian autonomous distillery equipment, $18 million for a 4000 tonne per day factory vs. $48 million based on Schaeffer's estimates.

The sensitivity of the ethanol cost to the cost of cane is shown in Figure 22. Published estimates of the cost of cane delivered to the factory vary from $8/tonne in Brazil to over $20/tonne in Louisiana [2,13,43,62,63]. A host of

* The value of hydrous ethanol as an unblended motor fuel is assumed to be 0.8 that of gasoline on a volumetric basis [60]. The value of anhydrous ethanol as an octane enhancing additive to motor fuel is assumed to be 1.2 that of gasoline [61]. USDOE projects that the world oil price will be $26.76-41.18/bbl in the year 2000 [59a], corresponding to wholesale gasoline prices of $0.78-1.18/gallon (Table 2). The equivalent prices would be 16.5-25.1 cents/liter for hydrous alcohol; and 24.8-37.6 cents/liter for anhydrous alcohol.

factors influence the cane cost, including cultivation practices, the cane yield, the type of land, climate, the harvesting system, labor costs and transportation costs to the factory. Not surprisingly, cane growing and harvesting practices are site specific and vary greatly around the world [62,63]. The large scale of cane production in Brazil may be a factor in reaching low costs, as well as cane varieties and cultivation practices optimized for high yield [24]. The feasibility of reaching these low cane costs in other regions is an open question.

The sensitivity of the ethanol cost to the installed capital cost of the distillery is shown in Figure 23. Clearly, it is important to reach the low distillery capital costs achieved in Brazil. As noted earlier, Brazilian autonomous distillery costs are typically only one half to one third those quoted by US engineering firms. Several factors could account for the difference in autonomous distillery equipment costs [39,40]:

* Brazil manufactures autonomous distillery equipment on a much larger scale than elsewhere, so that manufacturers achieve lower costs by taking advantage of economies of scale.

* Because of the experience gained in the PROALCOOL program, the Brazilian cane alcohol industry may be further along the technological "learning curve".

* Lower labor and/or material costs.

* Different engineering standards or codes for process equipment.

Understanding these issues in detail could shed light on whether the Brazilian experience with autonomous distillery costs could be replicated in other countries.

If Brazilian costs are met, it would be possible to produce ethanol at about 20 cents/liter, a cost which would make ethanol competitive as an additive to gasoline, even without biomass fuel credits. The corresponding cost of electricity production would be 5 cents/kWh or less with all the cogeneration systems considered. (For alcohol at 20 cents/liter, the electricity cost with the CEST system would be about 4.9 cents/kWh; for BIG/STIG about 3.4 cents/kWh and with BIG/ISTIG about 2.7 cents/kWh.)

With cogeneration fuel credits, it would be possible to produce hydrous ethanol for less than 17 cents/liter, which would be competitive with gasoline at the current oil price. At these ethanol costs, gas turbine cogeneration systems (either BIG/STIG or BIG/ISTIG) would be required to reach the low electricity production costs needed to compete with other supply options in most places. The BIG/ISTIG system appears to be particularly promising. With BIG/ISTIG it would be possible to produce ethanol at a cost competitive with wholesale gasoline, while electricity costs were just $0.04/kwh, even for an average fuel price of $3/GJ. Electricity costs with the CEST system would probably be too high to compete, if the biomass fuel credits were high enough to make neat ethanol competitive with gasoline.

The costs of ethanol and electricity production are quite sensitive to the biomass fuel price, especially for cases where the distiller controls the barbojo supply. The opportunity cost of bagasse and the price of competing cogeneration fuels may be important parameters in determining the bagasse price. As with cane harvesting, the cost and feasibility of barbojo recovery is likely to be site dependent. This cost and the price of competing off-season fuels would probably determine viability of selling barbojo at a particular factory and the biomass fuel credit received by the distillery.

c. Changing the length of the milling season

In most sugar cane growing regions, the milling season is 160-200 days in length. Off-season fuel requirements could be met using less than 60% of the available barbojo. If the remaining barbojo could be sold as cogeneration fuel, the distiller would receive even higher revenues. One strategy for more fully utilizing the barbojo is shortening the milling season. In this case, the distillery equipment would be used for a shorter period, so that the contribution of capital costs to the cost of alcohol production would increase. However, the fuel credits would increase. This is shown in Figure 24a, where we have plotted the costs of ethanol and electricity production assuming that 80% of the barbojo is utilized as off-season fuel and milling season is 133 days long. In Figure 24b, we show ethanol and electricity costs assuming that 59% of the barbojo is used and the milling season is 160 days long.

2. ANNEXED DISTILLERIES

The analysis for the annexed distillery case is inherently more complicated than for the autonomous distillery because of the number of products which must be considered (sugar, molasses, alcohol, and electricity). The type of economic calculations which are most useful depend on the situation and the mix of products.

a. Economics of ethanol production from C-molasses

In Figure 25 we examine the economics of ethanol production from the perspective of a sugar factory owner, who is considering making ethanol from the C-molasses produced at his factory (*). As before, the sugar factory/annexed distillery sells biomass fuel (bagasse and possibly barbojo) to the cogeneration plant and buys steam and electricity. For this situation, the ethanol production cost depends on the feedstock opportunity cost (the price of C-molasses), the capital and operating costs of the new distillery equipment

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* In this case, the price of sugar does not effect the decision to make ethanol, because sugar production proceeds as usual.

and on how much of the factory's revenue from biomass sales is credited to alcohol production. Since the factory's primary revenue is from sugar, and ethanol is a relatively minor byproduct, most of the biomass revenue is credited to sugar. In Figure 25 we have assumed that the fraction of the biomass revenue credited to ethanol is 15%. (Biomass revenues are credited to alcohol based on the approximate value of alcohol as a byproduct compared to the value of sugar.)

The costs of alcohol and electricity are plotted for each of the three cogeneration technologies, and for molasses prices of $20, $40, and $60/tonne. Annexed distillery costs are given in Tables 9 and 10. A range of values is shown for each case corresponding to different biomass credits.

For an annexed distillery making ethanol from C-molasses the key factors for low cost ethanol production are the price of C-molasses and the the biomass fuel credit. As shown in Figure 25, the co-production of electricity is necessary to make ethanol from C-molasses competitive at present oil prices, even as an octane enhancing additive.

As with the autonomous distillery, the cost of ethanol and electricity are sensitive to the biomass fuel price. BIG/STIG or BIG/ISTIG cogeneration systems would be needed to reach competitive electricity and alcohol costs. With BIG/ISTIG as the cogeneration technology, neat ethanol could be competitive with wholesale gasoline, and electricity could be competitive with new hydroelectric supplies in Brazil, even for molasses priced at $40/tonne.

Figure 26 shows that the cost of producing ethanol from C-molasses is also very sensitive to the fraction of the biomass fuel revenues credited against alcohol production. With zero fuel credit, alcohol would be uncompetitive. Higher fractions of the biomass fuel revenues credited against alcohol would make alcohol competitive at today's gasoline prices at the same time that electricity is competitive with new sources of hydropower, if gas turbines are used for cogeneration.

The cost of sugar production (shown in Figure 26 in cents/kg at the endpoints of each line) is less sensitive to the fuel credit distribution. (This is because the revenues from sugar are much greater than the revenues from alcohol at an annexed distillery producing ethanol from C-molasses.) However, biomass cogeneration fuel credits could significantly reduce the cost of sugar production in some cases. For example, when 15% of the biomass fuel credits are allocated to alcohol, the sugar production cost would decrease from about 29 cents/kg to about 20 cents/kg with BIG/ISTIG cogeneration. (For comparison, the current world market sugar price is about 26 cents/kg and in recent years has dipped as low as about 16 cents/kg.) In an annexed distillery, the economics of both sugar production and alcohol production would be improved by selling cogeneration fuel. Of course, the economic viability of sugar production would also be strongly dependent on the cost of cane and the sugar factory capital costs.

b. Flexible product strategies

In some factories the mix of sugar and alcohol produced is varied to meet changing markets. The economics ethanol and sugar production in factories with flexible production is considered in this section.

1). Economics of ethanol production from A-molasses

We first consider an annexed distillery which produces one strike of sugar (A-sugar) plus ethanol from the A-molasses. As compared to making ethanol from C-molasses, the factory makes less sugar (74 kg/tc vs. 108 kg/tc), but more ethanol (25 l/tc vs. 9 l/tc).

When C-molasses is the feedstock for ethanol production the economic analysis is fairly straightforward. In calculating the cost of ethanol production, the opportunity cost of C-molasses is simply the C-molasses market price.

When ethanol is made from A-molasses, the analysis is more complicated. Because A-molasses is not traded as a commodity, the opportunity cost of A-molasses cannot be equated to its market price. We have assumed instead that the opportunity cost of A-molasses is equal to the total value of the products which could be derived from it, B-sugar and C-molasses. The value of A-molasses depends on both the B-sugar price and the C-molasses price. Although this is not as convenient as assigning a price per tonne to A-molasses, it does reflect the income that the factory would be foregoing by using A-molasses to make ethanol.

The cost of ethanol production is calculated based on the distillery capital and operating costs, the opportunity cost of A-molasses and the biomass fuel credits received. Figure 27 shows the cost of ethanol production from A-molasses as a function of C-molasses price, sugar price and biomass fuel credit. It is assumed that the factory sells both bagasse and barbojo to the cogenerator and that all of the biomass fuel credits are credited against alcohol production. In all cases, the majority of the production cost of ethanol from A-molasses is the opportunity cost of B-sugar (the sugar price). The non-feedstock costs (capital and operation and maintenance costs) plus the C-molasses opportunity cost are typically less than 1/2 of the total ethanol cost. The cost of ethanol production of quite sensitive to the biomass fuel price. The total fuel cost to the cogenerator must be almost $3/GJ in order for ethanol from A-molasses to be competitive, even as an octane enhancer.

As the biomass fuel price and the allocation of biomass fuel credits vary, the costs of alcohol and sugar production also vary. This is shown for the various cogeneration technologies in Figures 28-30. Here ethanol and sugar production costs are plotted for several biomass fuel prices, and for varying allocation of the fuel credits. Again, it is assumed that the distiller sells both bagasse and barbojo to the cogenerator. Along each line the biomass fuel credit is held constant, but the allocation of credits varies. The left endpoint of each line represents the cost when all the biomass revenues are credited against sugar production: the right endpoint when all the biomass fuel revenues are credited against alcohol production. In each figure, three lines are shown corresponding to three different biomass fuel credits.

When no net payments are exchanged between the factory and the cogenerator, the cost of ethanol production is over 50 cents/liter. When the total biomass fuel cost to the cogenerator is $2/GJ, the cost of alcohol is about 40 cents/liter, still too high to be competitive as a motor fuel at current oil prices. When the total biomass fuel cost to the cogenerator is $3/GJ, ethanol from A-molasses can be competitive as a motor fuel. Also shown are lines of constant electricity production cost.

In order to make electricity and ethanol from A-molasses at competitive costs several conditions must be satisifed:

1) Cogeneration would be neccessary in order for ethanol from A-molasses to compete as a motor fuel. Biomass credits corresponding to cogeneration fuel costs of about $3/GJ must be received in order for ethanol to compete even as an octane enhancer. To increase biomass fuel revenues, it is desirable that the factory sell barbojo off-season as well as bagasse during the milling season.

2) Moreover, a large <u>fraction</u> of the biomass fuel revenue must be credited against alcohol production in order to reach ethanol costs of less than 24 cents/liter. This means that the cost of sugar must be relatively low so that both sugar and ethanol can be produced at economically attractive costs. Sugar production costs in turn depend upon the cane cost, sugar factory equipment costs, etc. Low sugar production costs such as those in Brazil or Australia appear to be a precondition for economically attractive production of ethanol from A-molasses.

3) Gas turbine cogeneration technologies are required to reach low electricity costs.

2). Flexible production strategies: ethanol from A-molasses or C-molasses?

Consider a factory which vary the mix of sugar and alcohol by producing either:

1) 25 liters/tc of ethanol from A-molasses plus 74 kg/tc of A-sugar. (Increased alcohol production.)

or

2) 9 liters/tc of ethanol from C-molasses plus 108 kg/tc of A and B-sugars (Maximum sugar production plus ethanol from C-molasses.)

What are the commodity prices at which one option would be more economically attractive than the other?

This is illustrated in Figure 31, which shows the breakeven prices of sugar and ethanol at which the internal rate of return would be equal for options 1) and 2). Above the line, (at higher ethanol prices and lower sugar prices) ethanol production from A-molasses would yield a higher rate of return. Below the line, making the maximum amount of sugar plus ethanol from C-molasses would be more economically attractive. It can be shown that the location of this line is independent of cane costs, equipment capital costs and C-molasses prices, and only mildly dependent on the fuel credit or cogeneration technology [64]. As shown in Figure 31, the internal rate of return received varies from about 20% to 5%, as the prices of sugar and ethanol decrease. Figure 31 shows results assuming the cogenerator's fuel cost is $2/GJ. If a larger fuel credit were received the higher the internal rate of return would be higher.

For typical sugar prices of 14-23 cents/kg (6-11 cents/lb), ethanol would have to be priced at 35-55 cents/liter in order for ethanol from A-molasses to yield a higher return than making B-sugar plus ethanol from C-molasses. From this graph, it appears that making B-sugar plus ethanol from C-molasses will generally be preferable to using A-molasses for ethanol production. The primary reason is the high opportunity cost of sugar.

VI. CONCLUSIONS AND RECOMMENDATIONS

A. CONCLUSIONS

Cogeneration can improve the economics of cane alcohol production at both autonomous and annexed distilleries. Our results indicate that it would be possible to co-produce ethanol competitively with gasoline and electricity at costs competitive with other supplies, under certain conditions:

1) To reach the low electricity costs needed to compete with other electricity supply options in most places, biomass gas turbine cogeneration systems (BIG/STIG or BIG/ISTIG) would be required. The BIG/ISTIG system looks particularly promising. With BIG/ISTIG would be possible to produce ethanol at either autonomous or annexed distilleries at a cost competitive with present wholesale gasoline prices, while the cost of electricity would be less than 4 cents/kWh. Biomass gas turbine cogeneration systems produce less steam than

the amount used in a typical autonomous or annexed distillery, so that factory steam economy measures would be required. Using commercially available process equipment it would be possible to reduce the factory steam demand in autonomous or annexed distilleries, so that BIG/STIG and BIG/ISTIG systems could be used.

2) For economically attractive ethanol production in autonomous distilleries, it is important to reach low cost Brazilian conditions. In particular Brazilian cane costs and distillery capital costs must be acheived. If Brazilian conditions are met, it would be possible to produce ethanol for about 21 cents/liter, a cost low enough to be competitive at present as an octane enhancing additive to gasoline, even without fuel credits from sales of cane residues. The cost of electricity production would be 5 cents/kWh or less with all the cogeneration systems considered. With the BIG/ISTIG system, the cost of electricity would be less than 3 cents/kWh. If electricity is co-produced in gas turbine cogeneration systems and biomass fuel revenues are credited against alcohol production, unblended ethanol could be produced for less than 16 cents/liter, which would be competitive at today's wholesale gasoline prices ($0.75/gallon). With BIG/ISTIG it would be possible to provide ethanol at a price competitive with the current wholesale gasoline price, while electricity would cost about 4 cents/kWh, competitive with new hydroelectric supplies in Brazil, even for an average cane residue price of $3/GJ.

3) For an annexed distillery making ethanol from C-molasses, the key factors for low cost ethanol production are the price of C-molasses and the allocation of the biomass fuel credit. The co-production of electricity is needed to make alcohol competitive even as an octane enhancing additive to gasoline. With BIG/ISTIG cogeneration, neat ethanol from C-molasses could be competitive with gasoline and cogenerated electricity could be competitive with

new hydroelectric supplies in Brazil, even for molasses prices of $40/tonne, assuming 15% of the biomass fuel revenues are credited against alcohol production. In annexed distilleries making ethanol from A-molasses, the cost of ethanol would be too high to be competitive as a motor fuel unless the opportunity cost of sugar is very low, and unless most of the biomass fuel revenues were credited against alcohol production, rather than sugar production. For factories with a flexible product strategy, making the maximum of sugar plus ethanol from C-molasses would generally yield higher rates of return than making less sugar plus ethanol from A-molasses.

4) The cost of ethanol and electricity production are sensitive to the biomass fuel price received. To reach low ethanol costs it is desirable that the factory sell barbojo during the off-season in addition to bagasse during the milling season for the best price possible. In practice the price of bagasse (or barbojo) would probably be determined by its value for other uses, by the price of competing cogeneration fuels (fuel wood or oil) or by the price of electricity from competing sources (e.g. hydropower), when they are less expensive than cogenerated electricity.

B. RECOMMENDATIONS

Our results suggest that under certain conditions, ethanol from cane could be competitive as a motor fuel at even today's world oil price, and electricity could be co-produced at a cost competitive with most other electricity supplies. We recommend further work on the feasibility of reaching the conditions for low cost cane energy in developing countries.

1) Development of biomass gas turbine cogeneration technologies is a key precondition for economically attractive co-production of electricity and ethanol from cane. High priority should be given to a demonstration project

for biomass gas turbines, after further analysis to assess fuel processing requirements and costs for gasification of bagasse and barbojo.

2) If Brazilian cost conditions are met (especially low cane costs and low distillery capital costs) ethanol production in autonomous distilleries would be economically attractive today. How widely could Brazilian costs be duplicated in other developing countries? To answer this question it is important to understand why cane costs and autonomous distillery costs are so much lower in Brazil than in many other places in the world. And whether these conditions could be matched in other places.

3) In light of the especially favorable estimated performance of BIG/ISTIG technology, priority should be given to commercializing ISTIG technology for natural gas at the same time BIG/STIG technology is being commercialized so that BIG/ISTIG technology could be commercialized shortly thereafter.

4) The biomass fuel revenues received by the distiller can be greatly increased, and cost of alcohol production reduced, if the distiller can sell barbojo as an off-season fuel. Barbojo is also a significant energy resource, containing over 50% of the energy available from sugar cane. (The energy content of barbojo is about 6.3 GJ/tc, as compared to 2.9 GJ/tc for bagasse and 1.7 GJ/tc for alcohol.) For a typical milling season of 160 days, only about 60% of the barbojo would be needed to supply the cogenerator with sufficient fuel to run the plant off-season. It may be worthwhile in some cases to recover some of the remaining barbojo for fuel. One strategy is to reduce the milling season and extend the off-season. Clearly, the cost and feasibility of recovering large fractions of the available barbojo need to be better understood.

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Table 1. Fuel Ethanol Production from Sugar Cane.

COUNTRY	FUEL ETHANOL PRODUCTION (10^6 liters/yr)
SOUTH AMERICA	
Brazil [a]	11,700
Argentina [b]	380
Paraguay [b]	26
Colombia [b]	38
CENTRAL AMERICA AND CARIBBEAN	
Costa Rica [b]	31
El Salvador [b]	15
Guatemala [b]	0.2
Jamaica [d]	15
AFRICA	
Kenya [c,e]	18
Malawi [c]	11
Zimbabwe [c,e]	42
Mali [b]	2
ASIA AND OCEANIA	
Thailand [b]	203
Phillipines [b]	10
New Zealand [b]	15

[a] For the 1988/1989 season. "Agroindustria Canavieira: Um Perfil," Copersucar, Sao Paulo, Brazil, 1989.

[b] 1984 ethanol production capacity installed. "Worldwide Review of Biomass Based Ethanol Activities," Meridian Corporation Report, Contract No. MC-JC-85FB-008, 1985.

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

[d] A 180,000 liter/day distillery was installed in 1988, and the planned production is 15 million liters/year. To date only about one million liters have been produced due to uncertain market conditions. (M.G. Hylton, Jamaican Sugar Industry Research Institute, Bernard Lodge, Jamaica, private communications, 1990.)

[e] Capacity installed as of 1984. "Power Alcohol in Kenya and Zimbabwe - A Case Study om the Transfer of a Renewable Energy Technology," United Nations Trade and Development Board Report GE.84-55979, June 5, 1984.

Table 2. Competitiveness of Ethanol From as a Motor Fuel

World Oil Price ($/bbl)	Wholesale Gasoline Price ($/gallon) [a]	EQUIVALENT ETHANOL COST NEEDED TO COMPETE [b] Hydrous Ethanol as Unblended Automotive Fuel (cents/liter)	Anhydrous Ethanol as Octane Enhancing Additive (cents/liter)
15	0.46	9.8	14.6
20	0.60	12.7	19.1
25	0.74	15.6	23.4
30	0.87	18.5	27.7
35	1.01	21.4	32.2
40	1.14	24.1	36.2
45	1.28	27.1	40.6

[a] J.G. Brown, [9].

[b] The equivalent cost of ethanol is calculated assuming that ethanol is worth 0.8 times as much as gasoline per liter as an unblended motor fuel and 1.2 times as much as gasoline per liter as an octane enhancing additive.

Table 3. Performance of Biomass-Fired Cogeneration Systems

	COGENERATION						POWER ONLY			
	ELEC		STEAM		FUEL	CANE[c]	ELEC		FUEL	CANE[c]
	(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	40.3	32.1	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	57.7	328	111.2	42.9	57.3	325

[a] E.D. Larson and R.H. Williams, "Biomass-Fired Steam Injected Gas Turbine Cogeneration," to appear in the Journal of Engineering for Gas Turbines and Power, American Society of Mechanical Engineers, 1990.

[b] Adapted from E.D. Larson and R.H. Williams, "Biomass-Fired Steam Injected Gas Turbine Cogeneration," to appear in the Journal of Engineering for Gas Turbines and Power, American Society of Mechanical Engineers, 1990, assuming that the gasifier efficiency is the same for biomass as for coal. The steam at the turbine inlet is assumed to be at 6.3 MPa 482^{o}C.

[c] We have assumed 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. The CEST uses 50% wet bagasse having a higher heating value of 9530 kJ/kg. We further assume that 300 kg of 50% wet bagasse are produced per tonne of cane milled, or 176 kg of 15% wet briquetted bagasse are produced per tonne of cane milled.

[d] Preliminary estimate of steam and electricity production, based on performance with coal. (See Appendix A.)

Table 4. Capital and Operating Costs for Biomass-Fired Cogeneration Systems

	CAPACITY (MW)	INSTALLED COST ($/kW)	MAINTENANCE [b] FIXED (1000$/Y)	VARIABLE ($/kWh)	LABOR [b,c] (1000$/Y)
CEST [a]					
Generic	27.0	1556	664	0.003	129.6
Generic	10.0	2096	246	0.003	97.2
Generic	3.0	3008	73.8	0.003	97.2
BIG/STIG [a,d]					
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 [b]					
LM-8000	111.2[e]	770[e]	2736	0.001	405.0

[a] Adapted from E.D. Larson and R.H. Williams, "Biomass-Fired Steam Injected Gas Turbine Cogeneration," to appear in the Journal of Engineering for Gas Turbines and Power, American Society of Mechanical Engineers, 1990.

[b] See Appendix D in E.D.Larson, J.M. Ogden and R.H.Williams, "Steam-Injected Gas Turbine Cogeneration for the Cane Sugar Industry," Princeton University, Center for Energy and Environmental Studies Report No. 217, September 1987; E.D. Larson, private communication, 1990.

[c] For Jamaican labor conditions.

[d] In general, the estimated unit cost is $2371 * MW^{-0.22}$, where MW is the installed capacity in MW. (See E.D. Larson and R.H. Williams, "Biomass-Fired Steam Injected Gas Turbine Cogeneration," to appear in the Journal of Engineering for Gas Turbines and Power, American Society of Mechanical Engineers, 1990.

[e] See R.H. Williams, "Biomass Gasifier Gas Turbine Power and the Greenhouse Warming," in Energy Technologies for Reducing Emissions of Greenhouse Gases, Proceedings of OECD Experts' Seminar, Paris, April 12-14, 1989; also Table 5 and Appendix A.

Table 5. Installed Capital Cost for Integrated Gasifier/Steam-Injected Gas Turbine and Integrated Gasifier/Intercooled Steam-Injected Gas Turbine Cogeneration Systems Fueled with Coal and Biomass (January 1986 $/kW)

	CIG/STIG[a]	BIG/STIG[b]	CIG/ISTIG[a]	BIG/ISTIG[b]
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	917.8	933.7	728.5
Preproduction Costs (2.8%)	32.3	32.3	26.2	20.4
Inventory Capital (2.8%)	32.3	32.3	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

[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 a BIG/STIG (BIG/ISTIG) would be similar to a CIG/STIG (CIG/ISTIG) system except that the raw gas chemical clean-up phase required for coal would not be needed for biomass, because of its lower sulfur content. See R.H. Williams, "Biomass Gasifier Gas Turbine Power and the Greenhouse Warming," in Energy Technologies for Reducing Emissions of Greenhouse Gases, Proceedings of OECD Experts' Seminar, Paris, April 12-14, 1989; also see Appendix A. The BIG/STIG system has a capacity of 53 MW; the BIG/ISTIG system 111 MW.

Table 6. Steam and Electricity Demands in a Typical Brazilian Autonomous Distillery Without Cogeneration

Steam driven mills : 125 tonnes cane per hour
Short tube rising film 3-effect evaporator
Conventional distillation system (3.3 kg steam/liter ethanol)
10,000 liters ethanol/hour

<u>Medium pressure steam</u> (2.1 MPa, 300°C)	
Total m.p. steam used	466 kg/tc
Cane mills, shredders, knives	226 kg/tc
Medium pressure steam turbines back-pressure	175 kg/tc (assuming 14 kg steam/kwh)
Throttled to low pressure	65 kg/tc
Total l.p. exhaust available	466 kg/tc
<u>Low pressure steam</u> (Mill and turbine exhaust 0.25 MPa, 127°C, saturated)	
Total l.p. steam used	454 kg/tc
Evaporator	61 kg/tc
Direct to Juice Heaters	130 kg/tc
Distillation	256 kg/tc
De-Aerator	8 kg/tc
<u>Electricity demand</u>	
Total electricity demand	12.5 kwh/tc
<u>Electricity production</u>	
Medium pressure steam turbines	12.5 kwh/tc

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[a] J.L. Oliverio, J.D. Neto and J.F.P. de Miranda, "Energy Optimization and Electricity Production in Sugar Mills and Alcohol Distilleries," presented at the 20th Congress of the International Society of Sugar Cane Technologists, Sao Paulo, Brazil, October 12-21, 1989.

Table 7a. Capital and Operating Cost Estimates for an Autonomous Distillery Milling 4000 tonnes cane per day (Excluding the Costs for Boilers and Generating Equipment) : High Cost Case, Based on Study by F.C. Schaeffer [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
Distilllation	1,700,000
Warehouse, Shop, Office Facilities	1,680,000
Plantwide Piping	4,000,000
General Services	2,694,000
Miscellaneous	466,000
Subtotal, Industrial Facilities	23,411,000
Labor to Erect	15,222,000
Contractor Fee, Overhead, Profit	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	FIXED COSTS ($)	VARIABLE COSTS ($/TC)
Wages and Salaries	206,000	0.68
Repair and Materials, Parts	64,000	1.00
Chemicals and Lubricants	9,000	1.16
Equipment Rental	10,000	0.02
Vehicles (incl. maintenance)	50,000	0.19
Other/Miscellaneous	44,000	0.21
Subtotal	383,000	3.26
Adminstrative Costs	725,000	0.40
Total Operating Cost	$1,108,000	$3.66/TC

[a] F.C. Schaeffer and Associates [13].

Table 7b. Capital and Operating Cost Estimates for an Autonomous Distillery Milling 4000 tonnes cane per day (Excluding the Costs for Boilers and Generating Equipment) : Low Cost Case, Based on Brazilian Experience [2].

TOTAL INSTALLED 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
Chemicals/Miscellaneous	0.092
Total Variable Cost	$0.176/TC

Table 8. Steam and Electricity Demands in a Typical Brazilian Annexed Distillery Without Cogeneration [a]

Steam driven mills : 417 tonnes cane milled per hour
Short tube rising film 5-effect evaporator
Conventional distillation system (3.3 kg steam/liter ethanol)
15.1 tonnes sugar/hr, 22,000 liters hydrous ethanol/hr

Medium pressure steam (2.1 MPa, 300°C)		
Total m.p. steam used	464 kg/tc	
Cane mills, shredders, knives	218 kg/tc	
Medium pressure steam turbines back-pressure	178 kg/tc	(assuming 14 kg steam/kwh)
Throttled to low pressure	58 kg/tc	
Total l.p. exhaust available	464 kg/tc	
Low pressure steam (Mill and turbine exhaust 0.25 MPa, 127°C, saturated)		
Total l.p. steam used	440 kg/tc	
Evaporator	191 kg/tc	
Direct to Juice Heaters	77 kg/tc	
Distillation	172 kg/tc	
Electricity demand		
Total electricity demand	12.7 kwh/tc	
Electricity production		
Medium pressure steam turbines	12.7 kwh/tc	

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[a] J.L. Oliverio, J.D. Neto and J.F.P. de Miranda, "Energy Optimization and Electricity Production in Sugar Mills and Alcohol Distilleries," presented at the 20th Congress of the International Society of Sugar Cane Technologists, Sao Paulo, Brazil, October 12-21, 1989.

Table 9. Capital Cost Estimates (in 1987 US$) for an Annexed Distillery Milling 4000 tonnes cane per day with a 36,000 liter per day annexed distillery (Excluding the Costs for Boilers and Generating Equipment) [a].

CAPITAL COSTS:

ITEM	SUGAR FACTORY COST ($)	DISTILLERY COST ($)
Site Preparation	213,000	37,000
Cane Handling	661,000	
Milling	5,939,000	
Juice Processing	5,570,000	
Sugar Handling	1,141,000	
Molasses handling	165,000	165,000
Water/Chemical Treatment	539,000	95,000
Fermentation/Prefermentation		722,000
Distilllation		425,000
Warehouse, Shop, Office Facilities	1,876,000	330,000
Plantwide Piping	3,885,000	685,000
General Services	1,676,000	670,000
Storage and Shipment	184,000	32,000
Subtotal, Industrial Facilities	21,840,000	3,161,000
Contractor Fee, Overhead @ 20%	4,368,000	632,000
Labor to Erect (50% of above)	13,104,000	1,897,000
Shipping, Ocean Freight (8%)	1,747,000	253,000
Engineering Fee (@ 7.5% of above)	3,079,000	446,000
Total Erected Cost	$44,138,000	$6,389,000

[a] F.C. Schaeffer and Associates, [13].

Table 10. Operating Cost Estimates (in 1987 US$) for an Annexed Distillery Milling 4000 tonnes cane per day with a 36,000 liter per day annexed distillery (Excluding the Costs for Boilers and Generating Equipment) [a].

FIXED OPERATING COSTS:

ITEM	SUGAR FACTORY COST ($)	DISTILLERY COST ($)
Wages and Salaries	234,000	41,000
Repair and Materials, Parts	68,000	12,000
Chemicals and Lubricants	6,000	2,000
Equipment Rental	8,000	2,000
Vehicles (incl. maintenance)	43,000	7,000
Other/Miscellaneous	46,000	8,000
Adminstrative Costs	616,000	109,000
Total Fixed Operating Cost	$1,021,000	$181,000

VARIABLE OPERATING COSTS:

ITEM	SUGAR FACTORY COST ($/TC)	DISTILLERY COST ($/LITER)
Wages and Salaries	1.15	0.010
Repair and Materials, Parts	1.80	0.016
Chemicals and Lubricants	0.25	0.002
Equipment Rental	0.02	0.001
Vehicles (incl. maintenance)	0.20	0.002
Other/Miscellaneous	0.30	0.003
Adminstrative Costs	0.45	0.004
Total Variable Costs	$4.17/TC	$0.038/LITER

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[a] F.C. Schaeffer and Associates, [13].

Table 11. Design Options for Co-production of Electricity, Alcohol and Sugar From Sugar Cane

PRODUCT MIX: AUTONOMOUS DISTILLERY (alcohol, electricity)
ANNEXED DISTILLLERY (alcohol, electricity, sugar)

COGENERATION SYSTEM: Condensing Extraction Steam Turbine (CEST)
Steam-Injected Gas Turbine (BIG/STIG)
Intercooled Steam-Injected Gas Turbine (BIG/ISTIG)

FUEL: In-season: Bagasse
Unprocessed (50% moisture) for CEST
Briquetted (15% moisture) for BIG/STIG and BIG/ISTIG

Off-season: Barbojo, wood or oil

CANE MILLING: Steam turbine drive
Electric drive
Diffuser

JUICE PROCESSING:	CONVENTIONAL	STEAM CONSERVING
JUICE HEATERS	Steam or vapor	Condensate
EVAPORATORS	Short Tube Rising Film	Falling Film or Falling Film w/ Mech. Vapor Recomp.
VACUUM PANS	Batch	Continuous

FERMENTATION: Batch
Continuous

DISTILLERY: Conventional
Low Energy Use (heat integrated design)
Mechanical vapor recompression

TABLE 12. SUMMARY OF PROCESS EQUIPMENT COSTS

	INSTALLED COST ($US)	
JUICE HEATERS [a]		
Shell and Tube	$\$75\text{-}100/m^2$	
Plate and Gasket	$\$100\text{-}150/m^2$	
EVAPORATORS [a]		
Short tube rising film	$\$300\text{-}500/m^2$	
Falling film	$\$300\text{-}500/m^2$	
VACUUM PANS [a]		
Discontinuous	$\$500/m^2$	
Continuous	$\$500/m^2$	
COMPRESSORS FOR MECHANICAL VAPOR RECOMPRESSION [b]		
Axial	$\$2414 \times P^{0.6075}$,	P = 50 kW - 1 MW
Rotary	$\$865 \times P^{0.946}$,	P = 50 kW - 7.5 MW

Sources: [a] J. Ogden, S. Hochgreb and M. Hylton, [44].
[b] G. Ulrich, A Guide to Chemical Engineering Process Design and Economics, John Wiley and Sons, New York, NY, 1984.

Table 13. ELECTRICITY AND STEAM CONSUMPTION FOR CANE MILL/SEPARATOR SYSTEMS[a]

(Numbers are in kWh/tc unless otherwise stated)

	Conventional Mill		Diffuser	
	Steam Drive	Electric	w/Mill	w/Extractor
Cane handling	0.5	0.5	0.5	0.5
Conveyors	0.4	0.4	0.4	0.4
Cane Cleaning	-	-	-	-
Knife Set 1	0.9	0.9	0.9	0.9
Knife Set 2	2.25	2.25	2.25	2.25
Shredder	3.4	3.4	-	-
Mills [b]	200 kg/tc	10	3.3(67 kg/tc)	-
Diffuser	-	-	2.0	2.0
Extractor	-	-	-	3.5
TOTALS [b]:				
Steam (kg/tc)	200 kg/tc	-	0 (67 kg/tc)	-
Electricity (kWh/tc)	7.5	17.5	9.4(6.1)	9.6

For diffuser energy use, numbers in parenthesis are for steam driven mills.

Sources: 1). E. Hugot, Handbook of Cane Sugar Engineering, Elsevier, (1985).
2) Canadian International Development Agency, Conference Proceedings on Cane Separation Technology, Barbados, (1981).

Table 14. Ethanol Fermentation and Distillation Technologies

Technology	Status	Benefits
FERMENTATION TECHNOLOGIES		
Batch fermentation	Very common	Simple
Cascading Tank continuous fermentation	In use	Simple; increased yield
Tower fermentation with flocculent yeast	Limited use for potable beer	Increased yield and productivity
Membrane separated fermentor (Rotofermentor)	Laboratory	Increased yield and productivity
Vacuferm	Laboratory	Increased yield and productivity
DISTILLATION TECHNOLOGIES		
Conventional distillation	Very common	Simple
Biostil	In use	Reduced stillage
Heat integrated distillation	In use	Energy savings
Vapor recompression in distillation	In use	Energy savings
Ethanol removal by solvent extraction	Laboratory	Energy savings
Ethanol removal by CO_2 extraction	Laboratory	Energy savings

Table 15a. Energy Use for Ethanol Separation from Water: Anhydrous Ethanol from Dilute Solutions

Ethanol Concentration (% wt) initial	final	Process	Energy Use (kJ/l product ethanol)	Process [a] Steam Use (kg/l)	Status
8-10	99.9	Conventional two column distillation in typical cane alcohol distillery + azeotropic distillation w/benzene	9900-11,000	4.5-5.0	Com[24]
8-10	99.9	Heat integrated distillation in innovated cane alcohol distillery + azeotropic distillation w/benzene	6600-7700	3.0-3.5	Com[24]
6.4-10	99.9	Conventional two column distillation + azeotropic distillation with benzene	7630-9650	3.5-4.4	Com[b]
6.3-10	99.9	Conventional distillation with vapor re-use + azeotropic distillation with benzene	5000	2.3	Com[b]
10	99.9	Conventional distillation with vapor recompression + azeotropic distillation with benzene [d]	4400		Com[b]
10	99.9	Conventional distillation with vapor recompression + azeotropic distillation with benzene with vapor re-use [d]	4230		Com[b]
10	99.9	Conventional distillation + water adsorption in cornmeal	3340	1.5	Com[b]
10	99.9	Conventional distillation with vapor recompression + water adsorption in cornmeal [d]	2170		Com[b]
10	99.9	IHOSR distillation + extractive distillation with KAc salts	1700	0.8	Lab[b]
10	99.9	Extraction with CO2	2232-2791		Lab[c]
10	99.9	Solvent extraction	1005		Lab[c]
10	99.9	Vacuum distillation	10,330		Lab[c]

Table 15b. Energy Use for Ethanol Separation from Water: Hydrous (Azeotropic) Ethanol from Dilute Solutions

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

a) The process steam is assumed to be saturated at 120°C, with enthalpy of vaporization of 2202 kJ/kg.

b) A. Serra, M. Poch and C. Sola, "A Survey of Separation Systems for Fermentation Ethanol Recovery," Process Biochemistry, October 1987, p.154-158.

c) G. Parkinson, "Batelle Maps Ways to Pare Ethanol Costs," Chemical Engineering, June 1, 1981.

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

Table 16. Steam and Electricity Demands in Conventional and Steam Conserving Autonmous Distilleries With Cogeneration

125 tonnes cane/hour
10,000 liters ethanol/hour

	CONVENTIONAL [a]	STEAM CONSERVING I [a]	STEAM CONSERVING II
Medium pressure steam (2.1 MPa, 300°C)			
Total m.p. steam used	466 kg/tc	258 kg/tc	223 kg/tc
Cane mills	226 kg/tc	226 kg/tc	223 kg/tc
Medium pressure steam turbines back-pressure (assuming 14 kg steam/kwh)	175 kg/tc	32 kg/tc	-
Throttled to low pressure	65 kg/tc	-	-
Total l.p. exhaust available	466 kg/tc	258 kg/tc	223 kg/tc
Low pressure steam (Mill and turbine exhaust 0.25 MPa, 127°C, saturated)			
Total l.p. steam used	454 kg/tc	258 kg/tc	223 kg/tc
Evaporator	61 kg/tc	97 kg/tc	97 kg/tc
Direct to Juice Heaters	130 kg/tc	-	-
Distillation	256 kg/tc	155 kg/tc	120 kg/tc
De-Aerator	8 kg/tc	6 kg/tc	6 kg/tc
Electricity demand			
Pumps, fans	12.5 kwh/tc	12.5 kwh/tc	12.5 kwh/tc
Total electricity demand	12.5 kwh/tc	12.5 kwh/tc	12.5 kwh/tc
Electricity production			
Factory medium pressure steam turbines	12.5 kwh/tc	2.3 kwh/tc	-
w/CEST Cogen. system	92 kwh/tc	123 kwh/tc	129 kwh/tc
w/BIG/STIG Cogen. system	-	252 kwh/tc	256 kwh/tc
w/BIG/ISTIG Cogen. system	-	-	298 kwh/tc
Max. Electricity for Export	92 (CEST)	242 kwh/tc (BIG/STIG)	286 kwh/tc (BIG/ISTIG)

[a] Steam use is based on J.L. Oliverio, J.D. Neto and J.F.P. de Miranda, "Energy Optimization and Electricity Production in Sugar Mills and Alcohol Distilleries," presented at the 20th Congress of the International Society of Sugar Cane Technologists, Sao Paulo, Brazil, October 12-21, 1989.

Table 17. Steam and Electricity Demands in Conventional and Steam Conserving Annexed Distilleries Making Alcohol from C-molasses [a]

175 tonnes cane/hour	CONVENTIONAL	STEAM CONSERVING I	STEAM CONSERVING II
Medium pressure steam (1.37 MPa, 250^{o}C)			
Total m.p. steam used	423 kg/tc	262 kg/tc	222 kg/tc
Cane mills	209 kg/tc	209 kg/tc	209 kg/tc
Medium pres. steam turbines	214 kg/tc	53 kg/tc	-
6% loss	-25	-16 kg/tc	-13 kg/tc
Total l.p. exhaust available	398 kg/tc	246 kg/tc	209 kg/tc
Low pressure steam (Mill and turbine exhaust, saturated)			
Steam temperature	120^{o}C	135^{o}C	135^{o}C
Total l.p. steam used	398 kg/tc	246 kg/tc	209 kg/tc
Evaporator	307 kg/tc	239 kg/tc	209 kg/tc
Bled to Juice Heaters	130 kg/tc	-	-
Bled to Distillery	-	36 kg/tc	18 kg/tc
Bled to Vacuum Pans	97 kg/tc	112 kg/tc	112 kg/tc
Direct to Vacuum Pans	43 kg/tc	-	-
Direct to Distillery	36 kg/tc	-	-
3% loss	12 kg/tc	7 kg/tc	3 kg/tc
Electricity demand			
Pumps, fans	12.8 kwh/tc	12.8 kwh/tc	12.8 kwh/tc
Mechanical vapor recompression	-	-	7.0 kwh/tc
Total electricity demand	12.8 kwh/tc	12.8 kwh/tc	19.8 kwh/tc
Electricity production			
Factory medium pressure steam turbines	15.0 kwh/tc	2.6 kwh/tc	-
w/CEST Cogen. system	98 kwh/tc	122 kwh/tc	128 kwh/tc
w/BIG/STIG Cogen. system	-	252 kwh/tc	257 kwh/tc
w/BIG/ISTIG Cogen. system	-	-	302 kwh/tc
Max. Electricity for Export	100 kwh/tc (CEST)	242 kwh/tc (BIG/STIG)	282 kwh/tc (BIG/ISTIG)

[a] Based on the Monymusk factory, assuming that a distillery is added.

Table 18. Technical and Economic Assumptions Used in Calculating the Cost of Ethanol and Electricity Production

All costs are in 1987 US$

Distillery Costs : Tables 7a, 7b, 9, 10

Cogeneration System Costs and Performance : Tables 3, 4

Factory Steam Use: Tables 16, 17

Sugar and Ethanol Production in Annexed Distilleries:

	FERMENTATION FEEDSTOCK		
	C-molasses	A-molasses	Cane juice
Ethanol (l/tc)	9	25	73
Sugar (kg/tc)	108	74	0

Fuel Processing Costs for BIG/STIG and BIG/ISTIG:
Bagasse Briquetting : $1.25/GJ
Barbojo " : $0.40/GJ

Bagasse collection cost: 0
Barbojo collection cost: $1.05/GJ

Milling Season : 160 days, 24 hour/day, 90% capacity factor

Cogeneration Plant: 365 days/year, 24 hours/day, 90% capacity factor

Required Internal Rate of Return (real) for Cogenerator: 10%

Required Internal Rate of Return (real) for Distillery: 15%

Table 19. Potential Fuel Ethanol and Electricity Production from Cane Compared to Actual 1985 Gasoline and Electricity Consumption

	MOTOR FUEL		ELECTRICITY	
REGION	FUEL ETHANOL PRODUCTION [b] (10^9 liters/yr)	GASOLINE CONSUMPTION (10^9 liters/yr)	CONSUMPTION IN 1985 (10^6 MWh/yr)	PRODUCTION W/BIG/STIG [c] (10^6 MWh/yr)
SOUTH AMERICA [d]	18.8			115
ASIA	14.8			89
CENTRAL AMERICA AND CARIBBEAN	10.7			65
AFRICA	5.2			32
OCEANIA	2.9			18
UNITED STATES	2.0			12
EUROPE	0.2			1
TOTAL	54.6			332

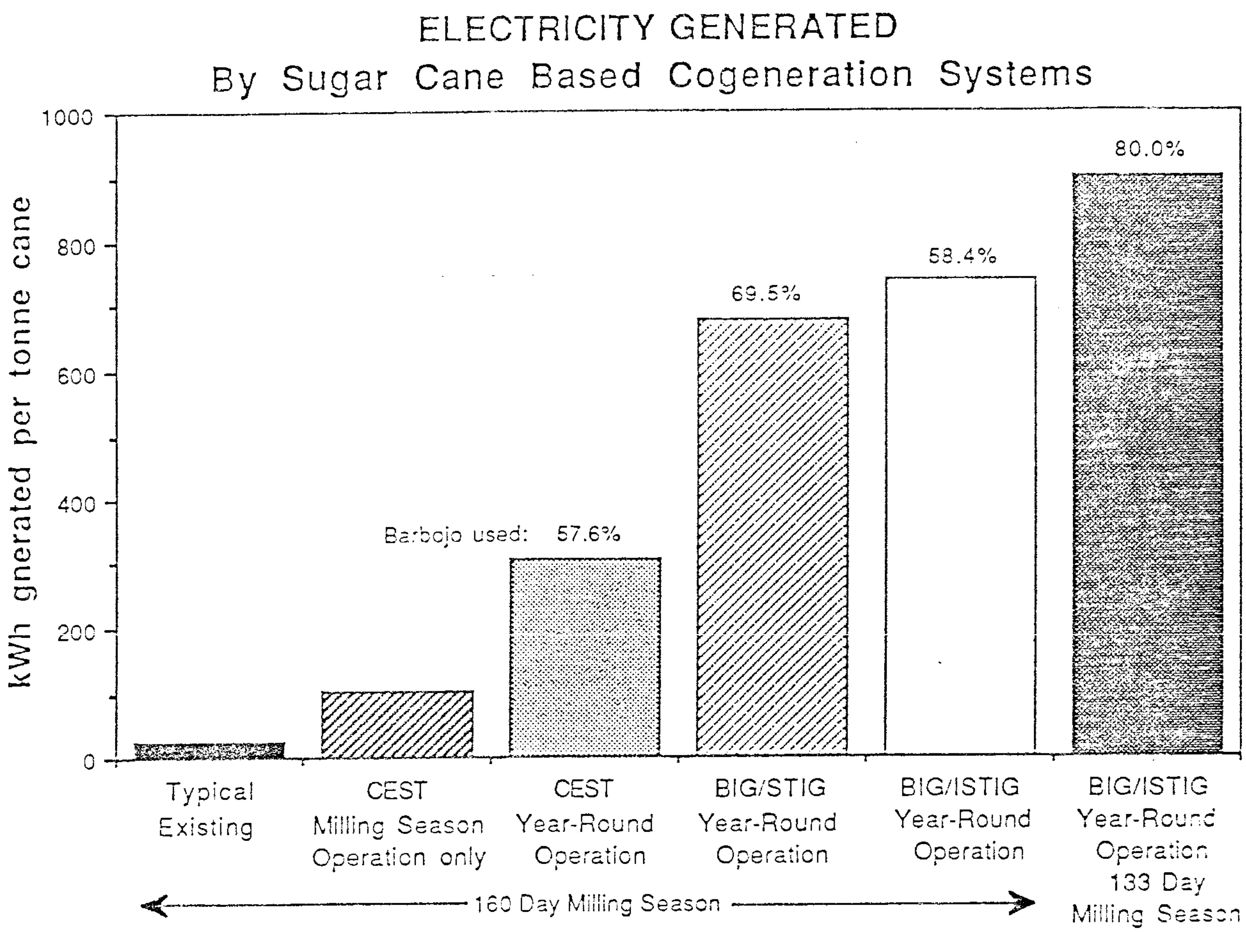


Figure 1. Potential cogeneration of electricity using sugar cane residues. The first bar is for a typical existing situation at a sugar factory or alcohol distillery during the the milling season. The next bar is for a CEST system operating during the milling season at a factory where steam-saving retrofits have been made. The next three bars are for year-round operation for a plant at which steam-saving retrofits have been made and where the milling season is 160 days: the third, fourth, and fifth bars are for CEST, BIG/STIG, and BIG/ISTIG systems, respectively. The sixth bar is for a BIG/ISTIG unit operated at a steam-efficient plant operated with a 133-day milling season. The number at the tops of the four bars to the right is the percentage of the barbojo used for power generation during the off-season.

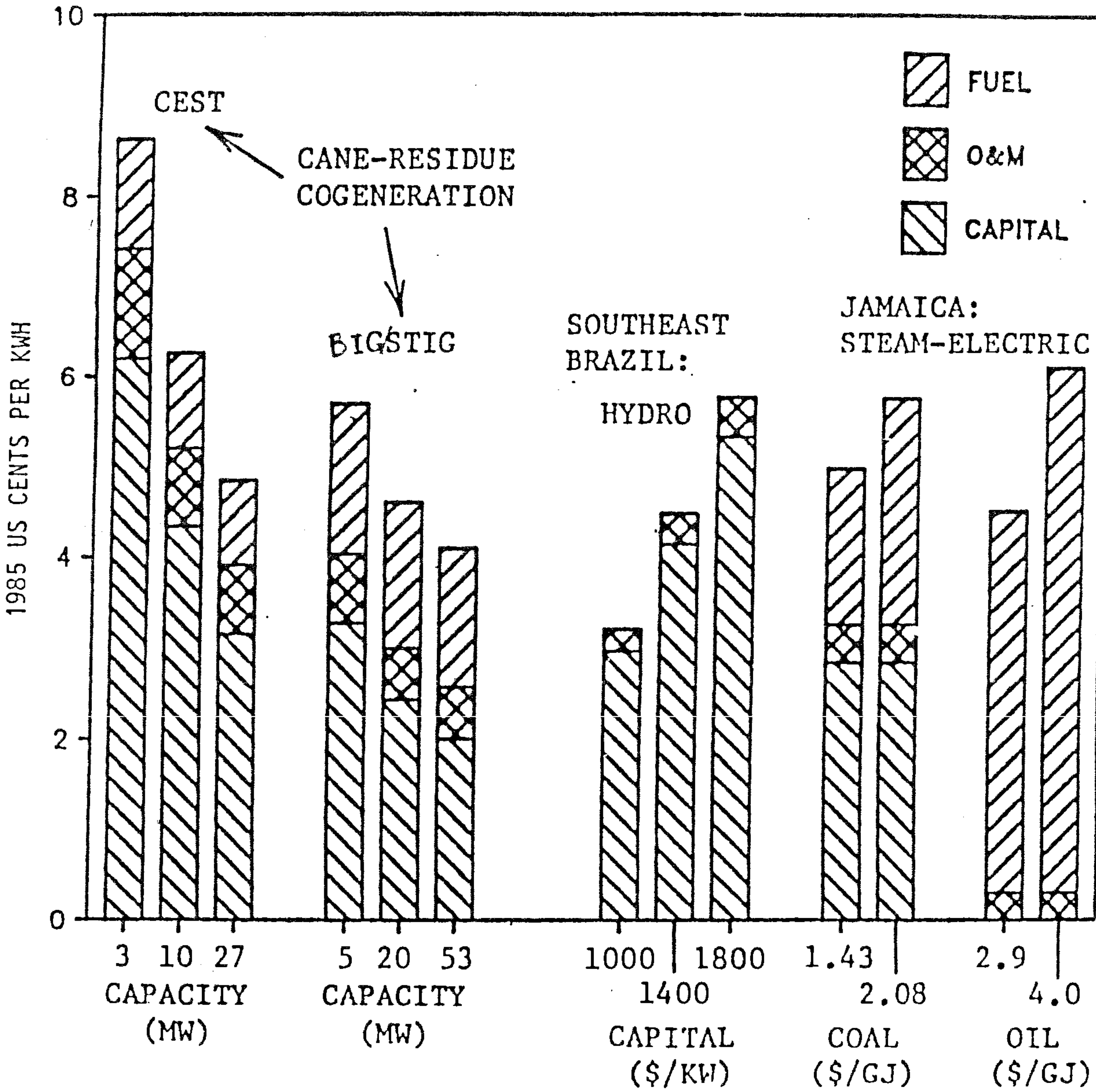


Figure 2. Estimated levelized costs of generating exportable electricity at cane-residue fired cogeneration plants and at least cost central station power plants in Jamaica and Brazil [20].

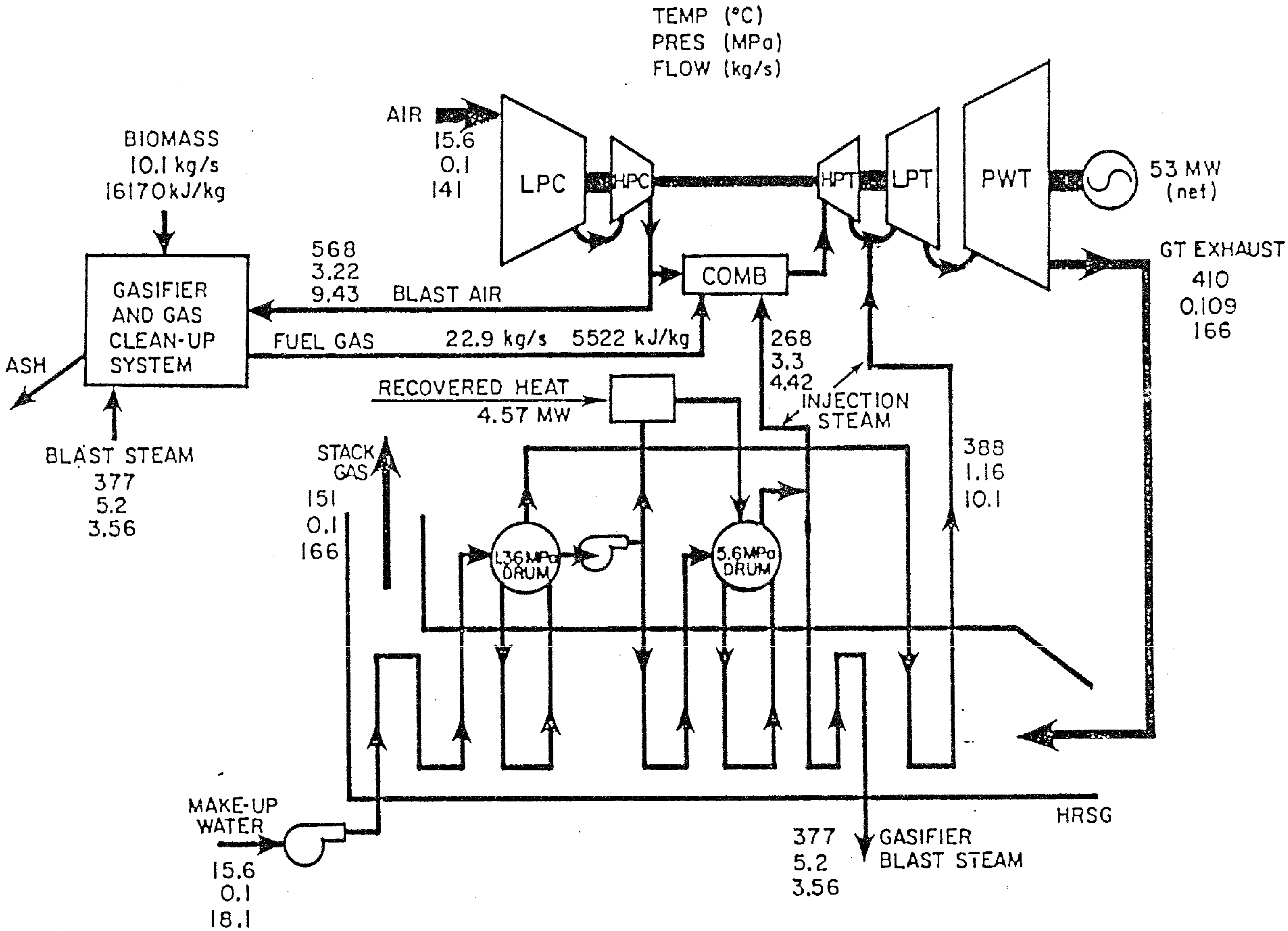
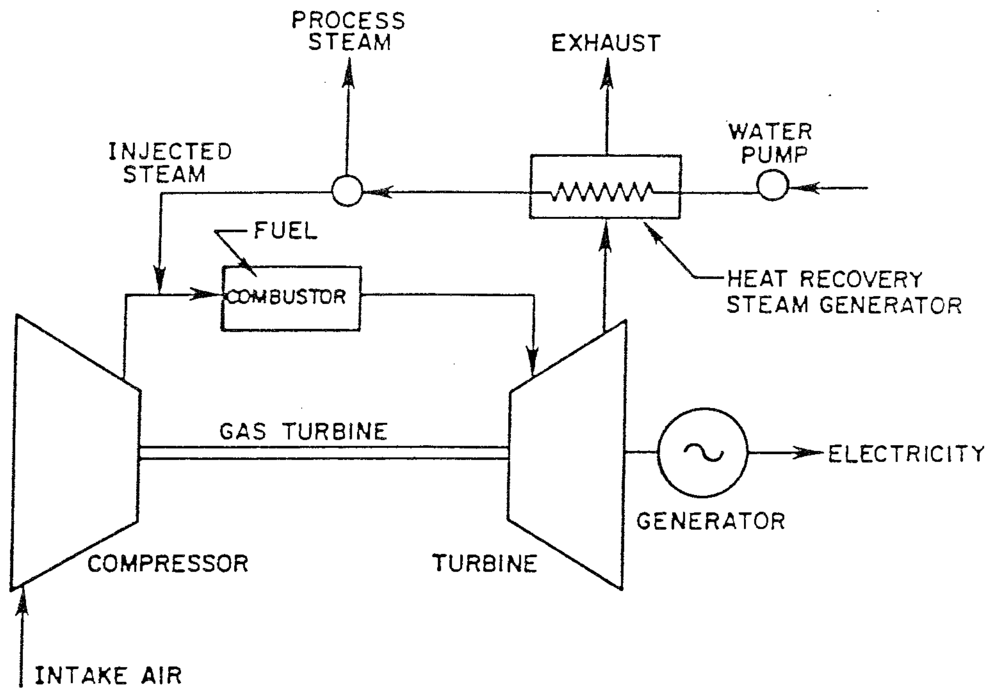


Figure 3a. Biomass Integrated Gasifier/Steam-injected gas turbine (BIG/STIG) cogeneration system [32].



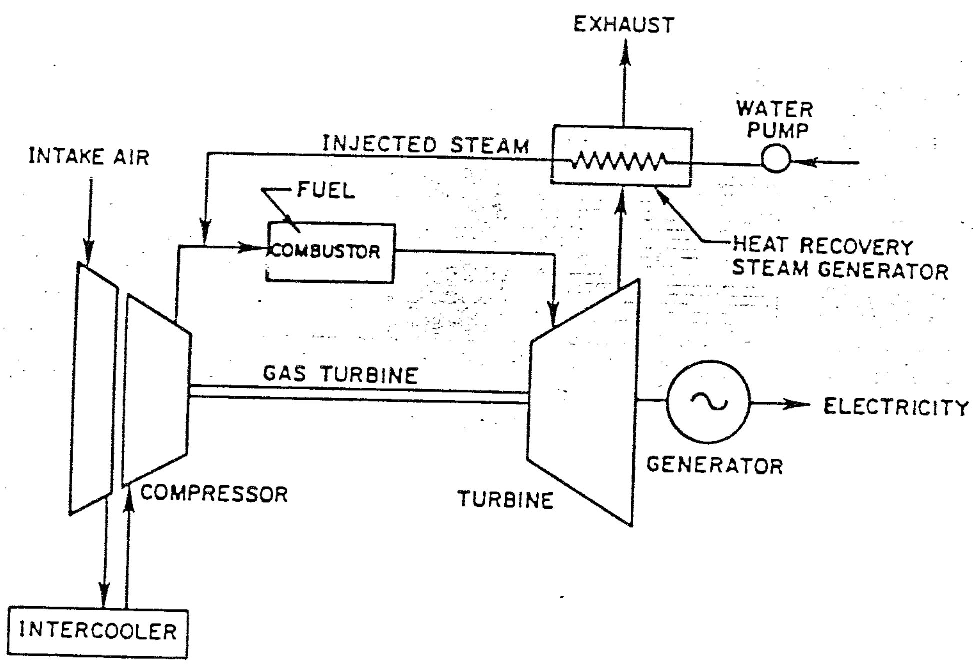
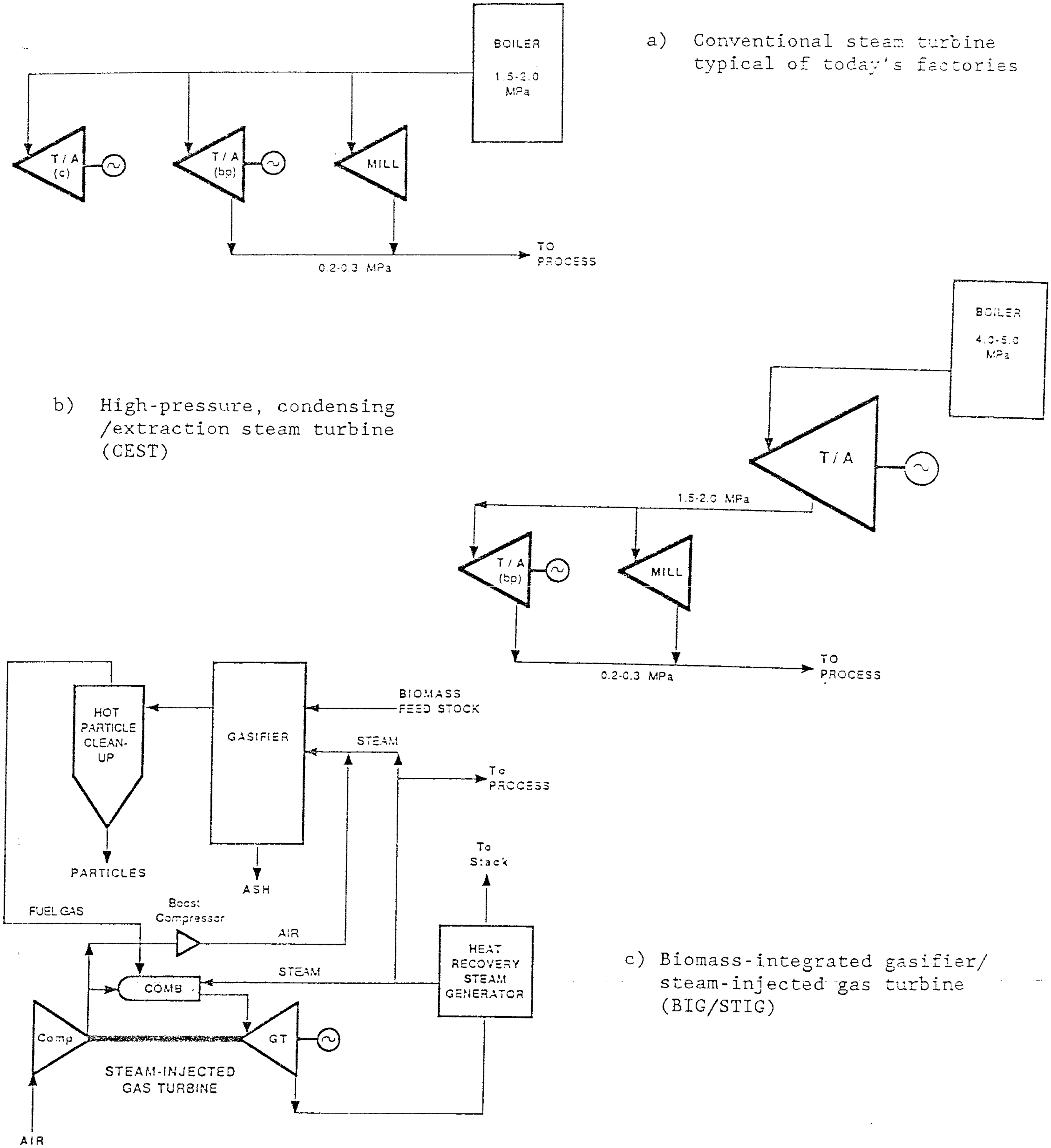


Figure 3b. Comparison of a steam-injected gas turbine (top) and an intercooled steam-injected gas turbine (bottom) [36].



a) Conventional steam turbine typical of today's factories

b) High-pressure, condensing /extraction steam turbine (CEST)

c) Biomass-integrated gasifier/ steam-injected gas turbine (BIG/STIG)

Figure 4. Cogeneration systems for a cane sugar factory or alcohol distillery: a) a conventional steam turbine typical of today's factories, b) a high pressure condensing extraction steam turbine (CEST), c) a biomass integrated gasifier/steam-injected gas turbine (BIG/STIG) cogeneration system [44]. (The BIG/ISTIG system would look similar).

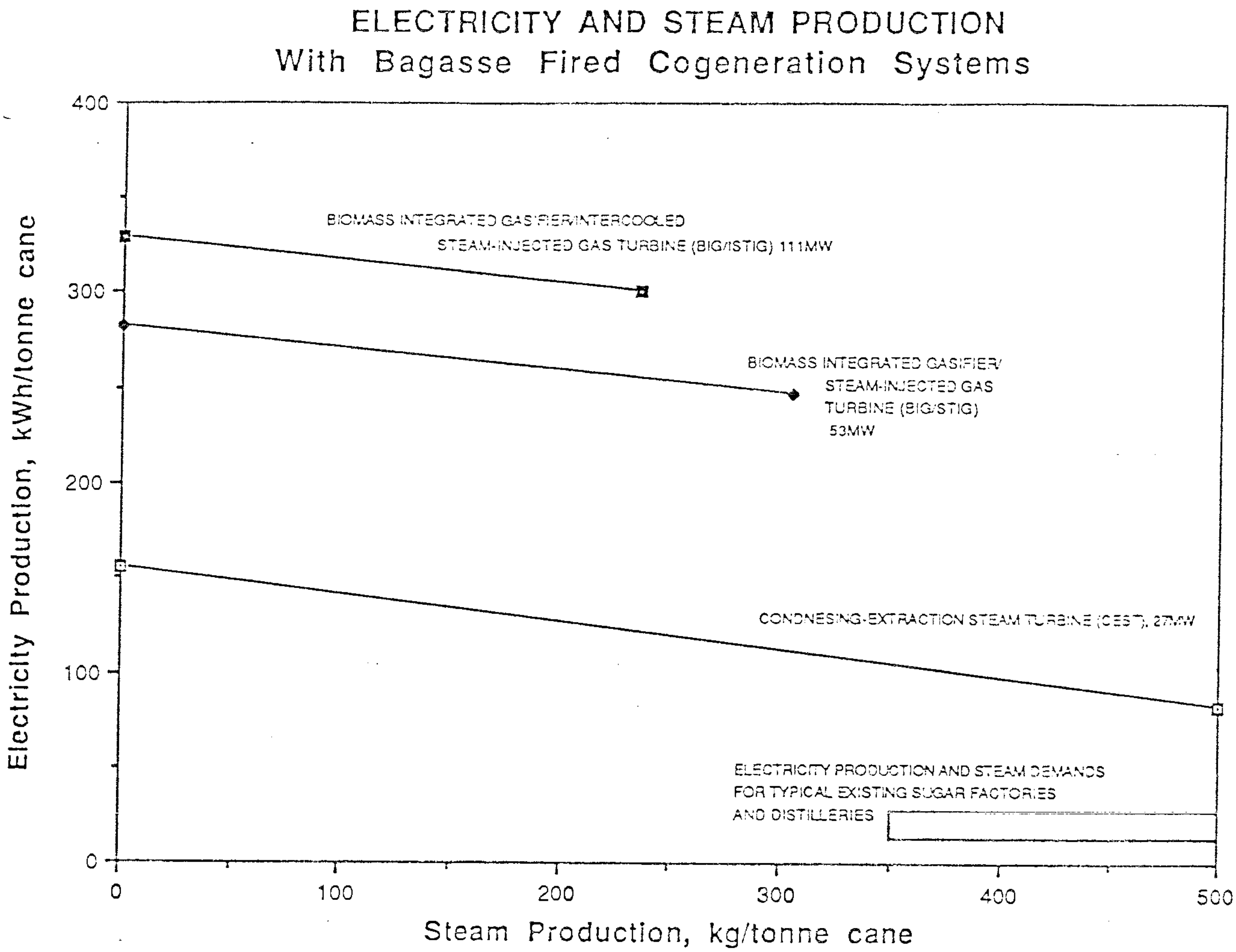


Figure 5. Electricity and steam production estimates for alternative bagasse fired cogeneration systems: the condensing-extraction steam turbine (CEST), the biomass integrated gasifier/steam-injected gas turbine (BIG/STIG), and the biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG) operating at sugar factories or alcohol distilleries during the milling season. Steam production is given in kilograms of steam produced per tonne of cane milled (kg/tc) and electricity production is given in kilowatt hours per tonne of cane (kwh/tc). Also shown are the steam and electricity production from a typical sugar factory or distillery cogeneration system today.

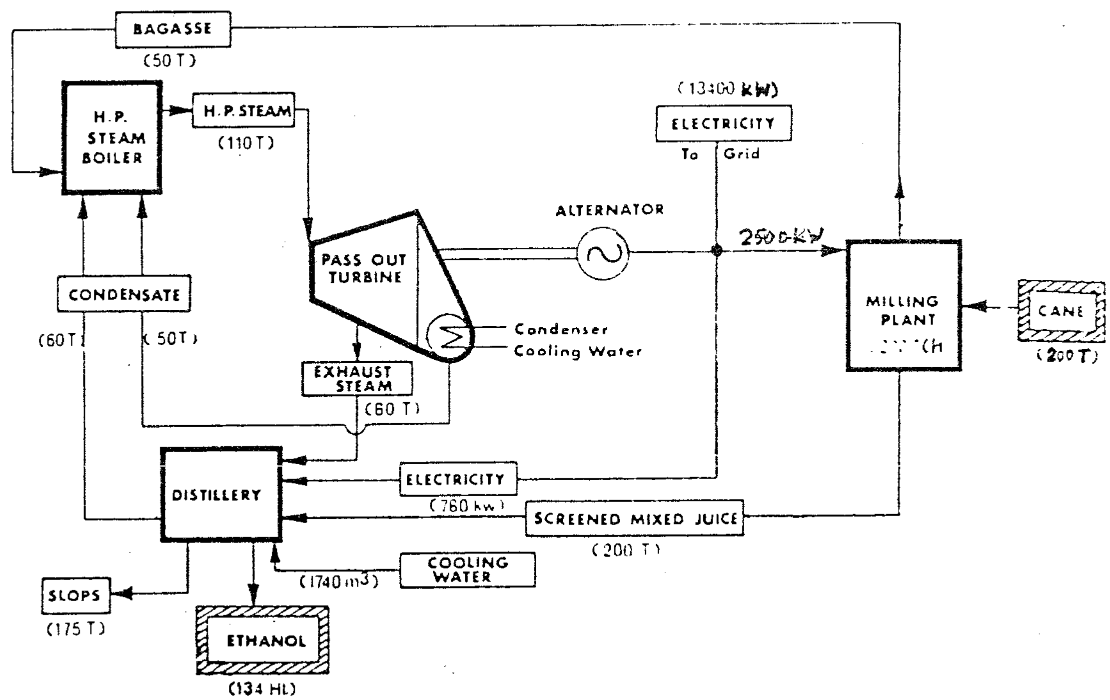


Figure 6. Flow diagram for production of ethanol in a typical autonomous distillery [39]. Quantities are given in tonnes/hour (T) for mass flows, in kilowatts (kW) for electricity, in cubic meters/hour (m^3) for cooling water, and in hectoliters/hour or 100's of liters/hour (HL) for ethanol.

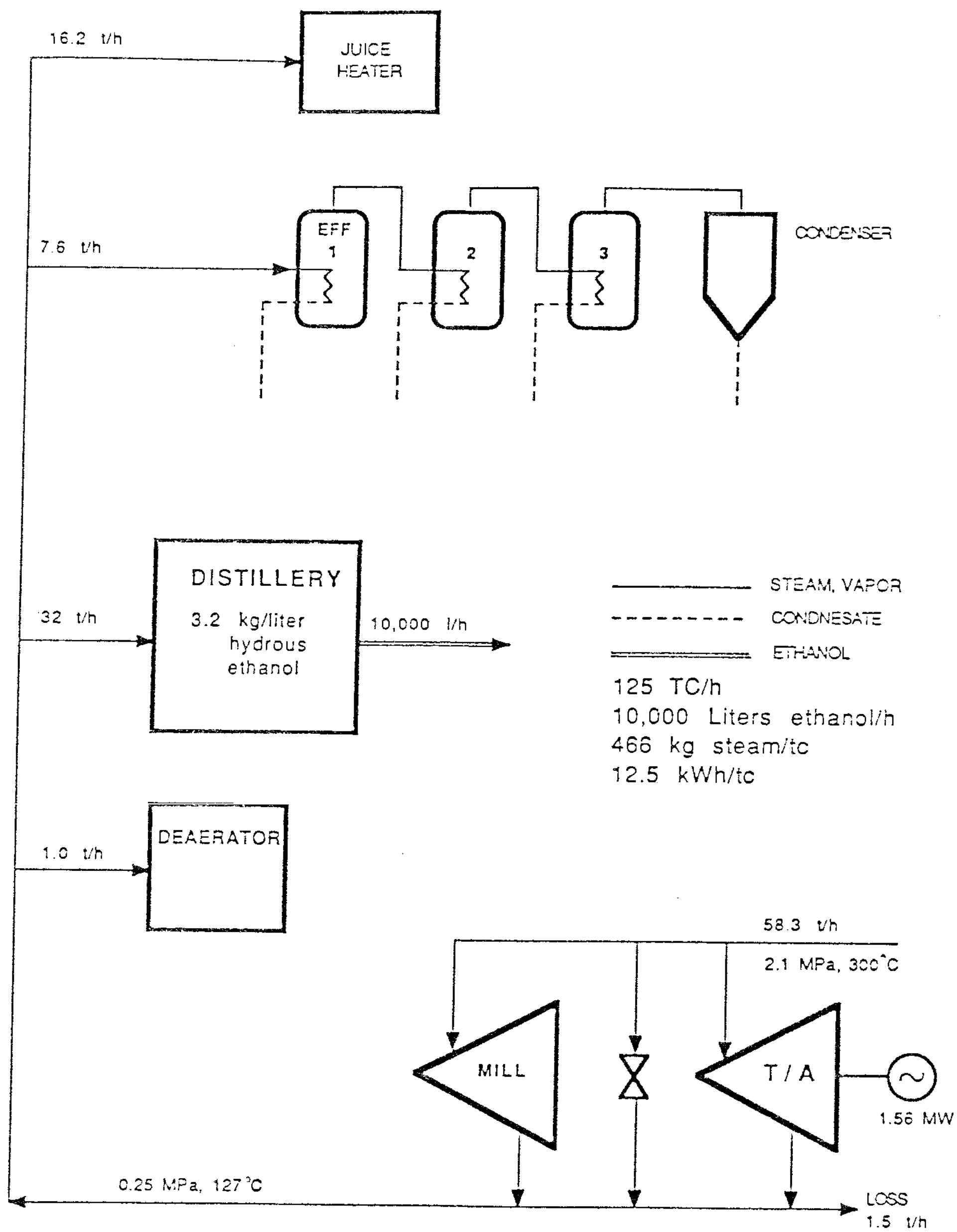


Figure 7. Factory flow diagram showing steam and material flows in a typical Brazilian autonomous distillery [28], milling 125 tonnes cane per hour and producing 10,000 liters per hour of hydrous ethanol. A conventional distillation system is used, requiring 3.3 liters of steam per liter of hydrous ethanol. The estimated factory steam demand is 466 kg/tc. The electricity demand is 12.5 kwh/tc.

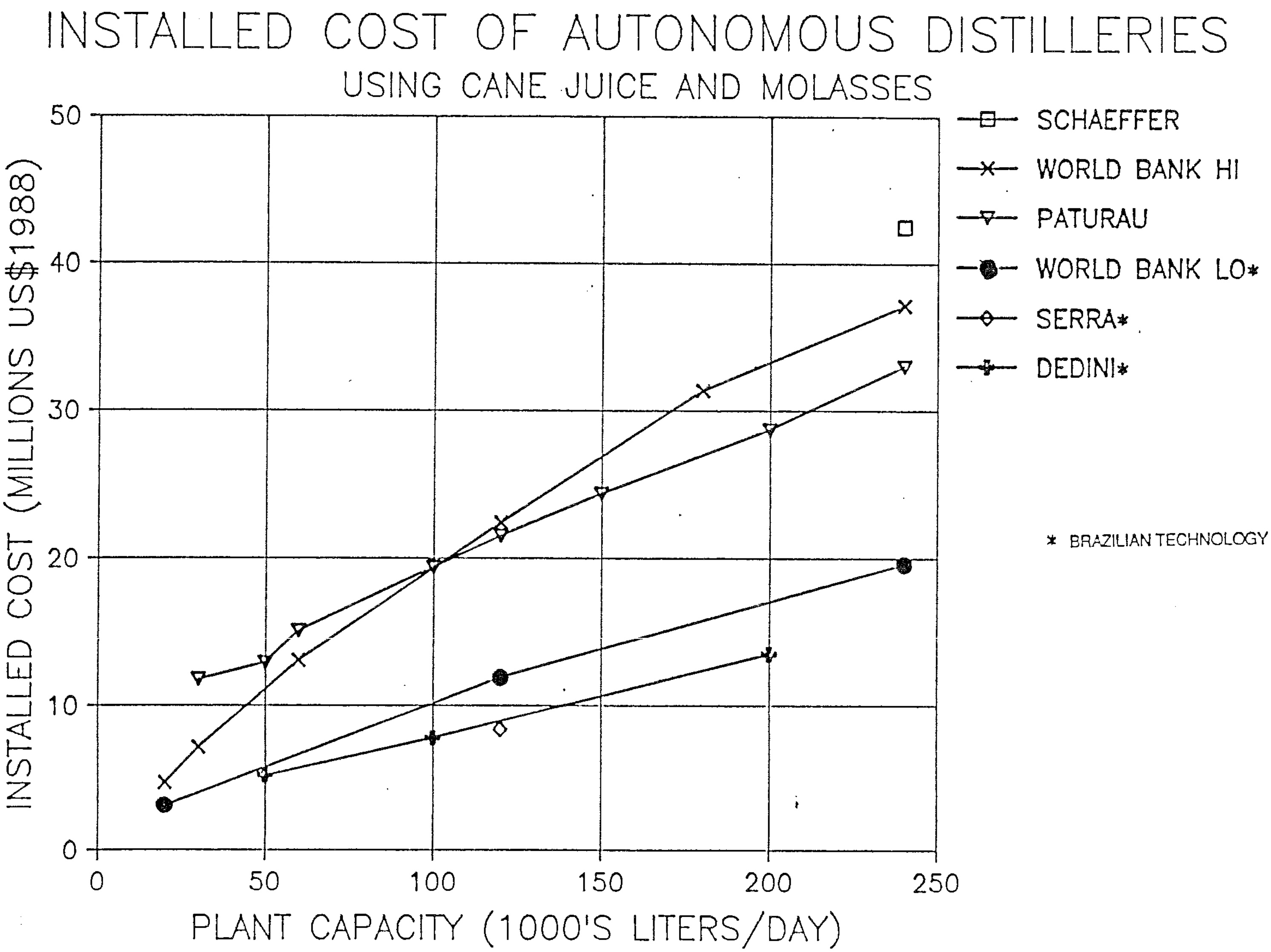


Figure 8. The installed capital cost of autonomous distilleries versus plant capacity. Costs are based on ref.s [2,8-10,13,40-42]. "Starred" estimates are for Brazilian technology.

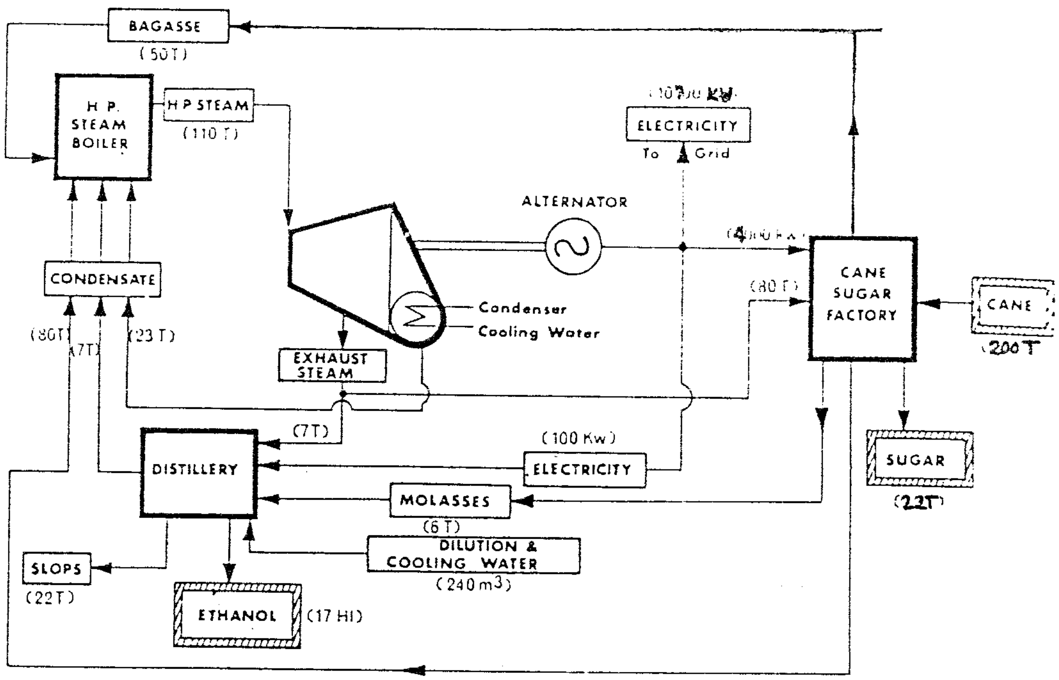


Figure 9. Flow diagram for production of ethanol in a typical annexed distillery [39]. Quantities are given in tonnes/hour (T) for mass flows, in kilowatts (kW) for electricity, in cubic meters/hour (m^3) for cooling water, and in hectoliters/hour or 100's of liters/hour (HL) for ethanol.

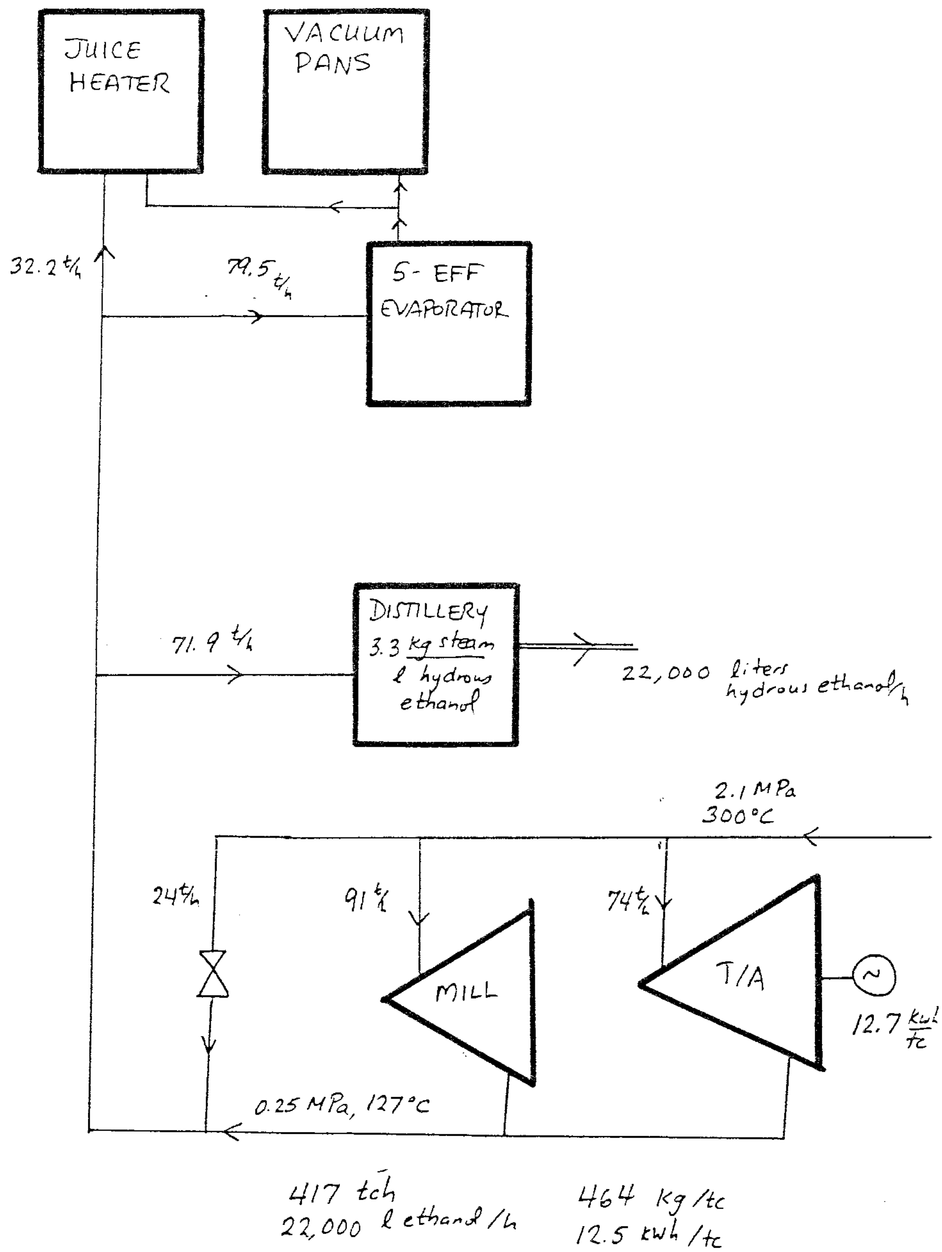


Figure 10. Factory flow diagram showing steam and material flows in a typical Brazilian annexed distillery [28], milling 125 tonnes cane per hour and producing hydrous ethanol. A conventional distillation system is used, requiring 3.3 liters of steam per liter of hydrous ethanol. The estimated factory steam demand is 464 kg/tc. The electricity demand is 12.5 kwh/tc.

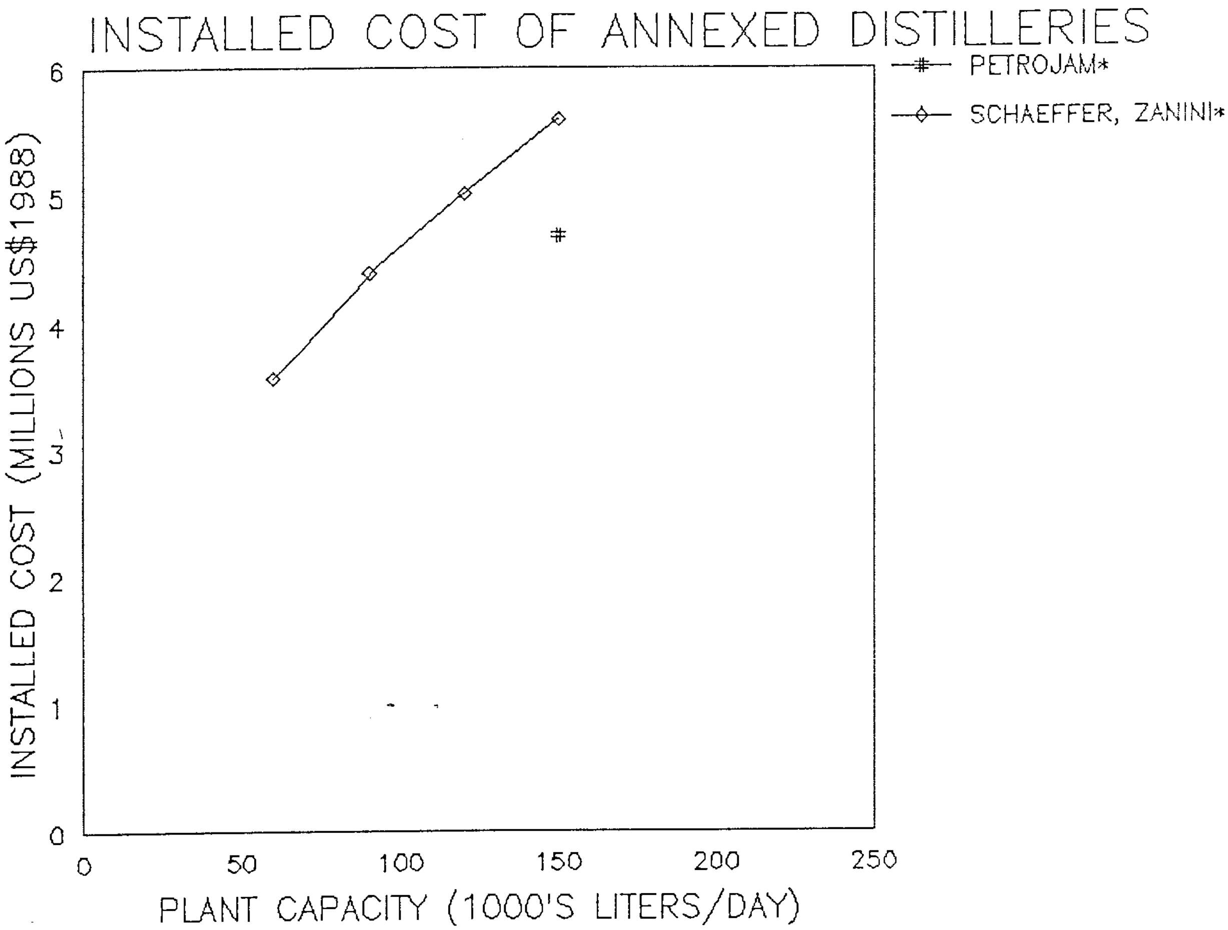


Figure 11. The installed capital cost of annexed distilleries versus plant capacity. (Only the costs of the distillery section -- fermentation and distillation equipment -- are included.) Costs are based on ref.s [10,39,40]. "Starred" estimates are for Brazilian technology.

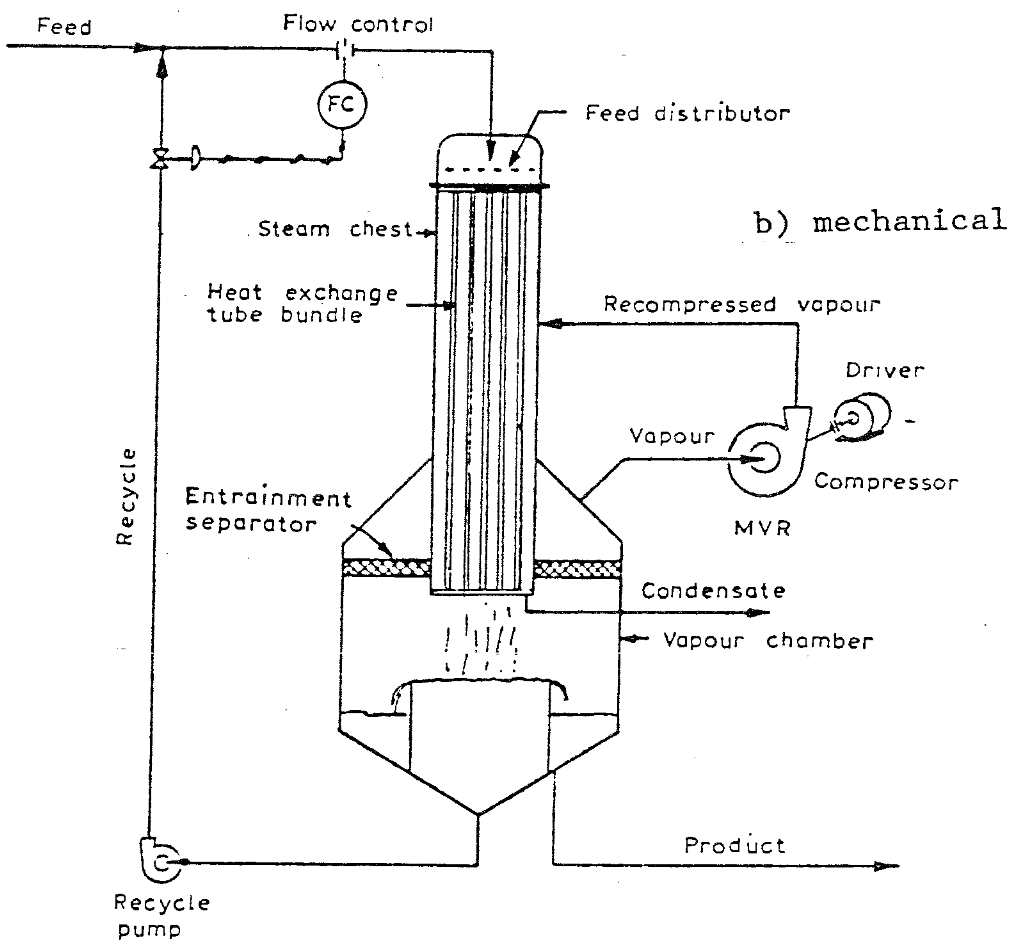


Figure 12. An evaporator with a mechanical vapor recompression system [20].

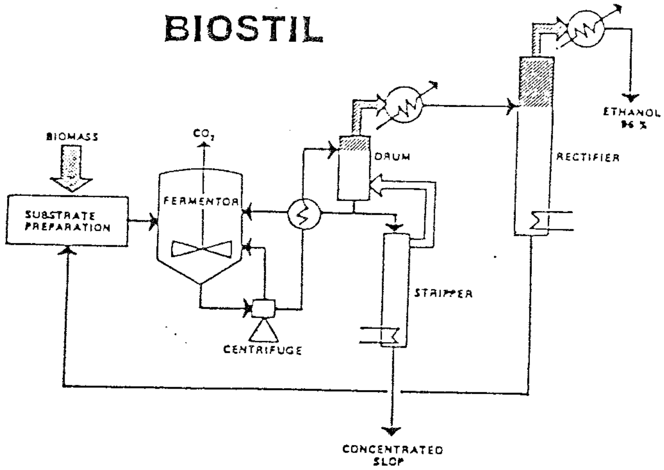


Figure 4 · The Biostil process

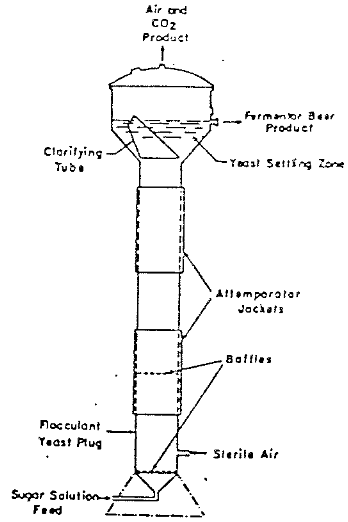
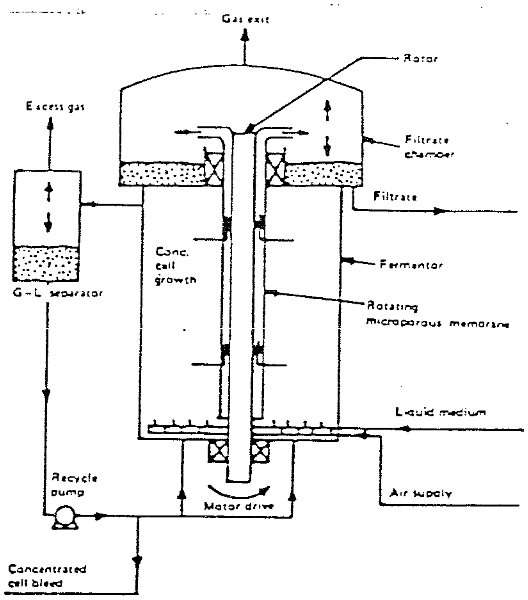
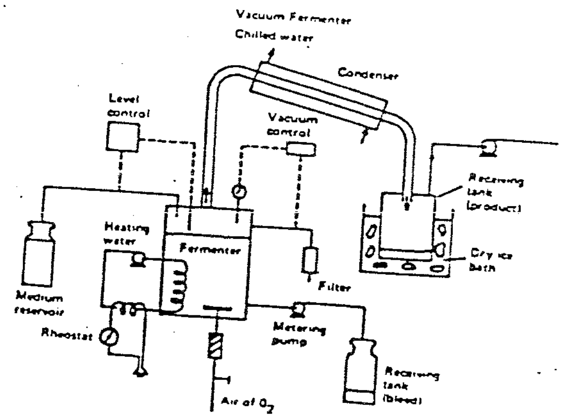


Figure 1: APV tower fermentor

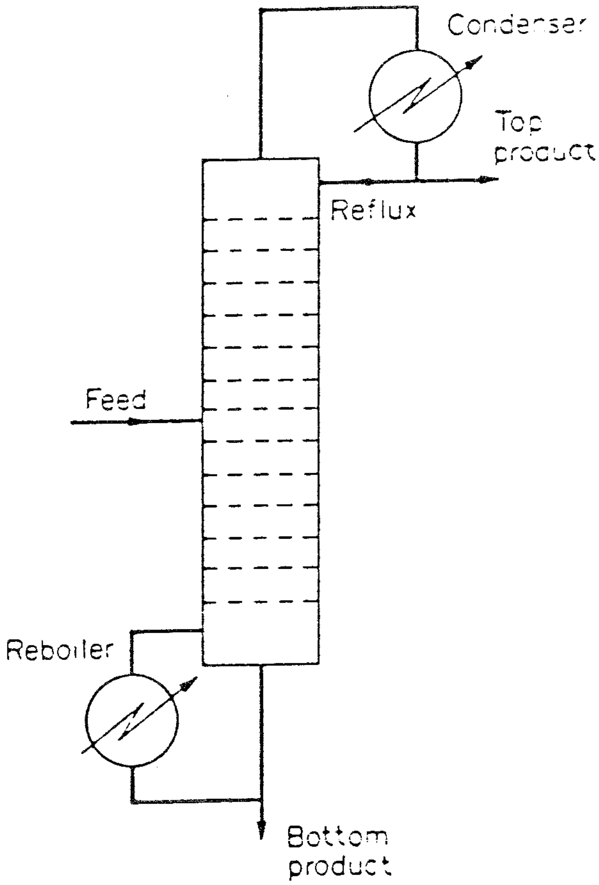


Rotorfermentor



Vacuferm principle

Figure 13. Continous fermentation technologies. a) Biostil, c) flocculent yeast tower fermentor, c) Membrane separated fermentor (rotoferm), d) vacuum fermentation (vacuferm) [51].



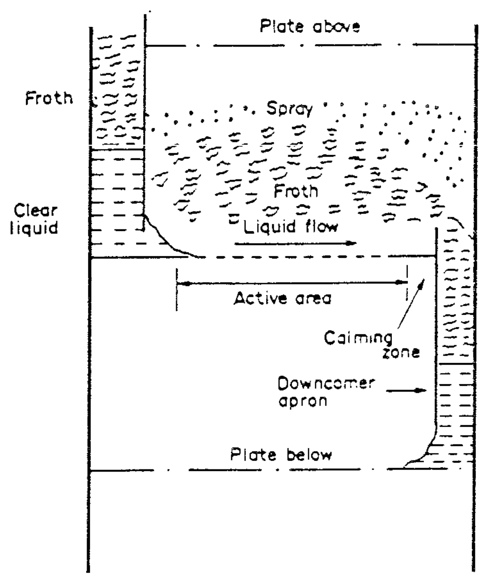


Figure 14. A typical distillation column (top), a bubble plate within the distillation column (bottom).

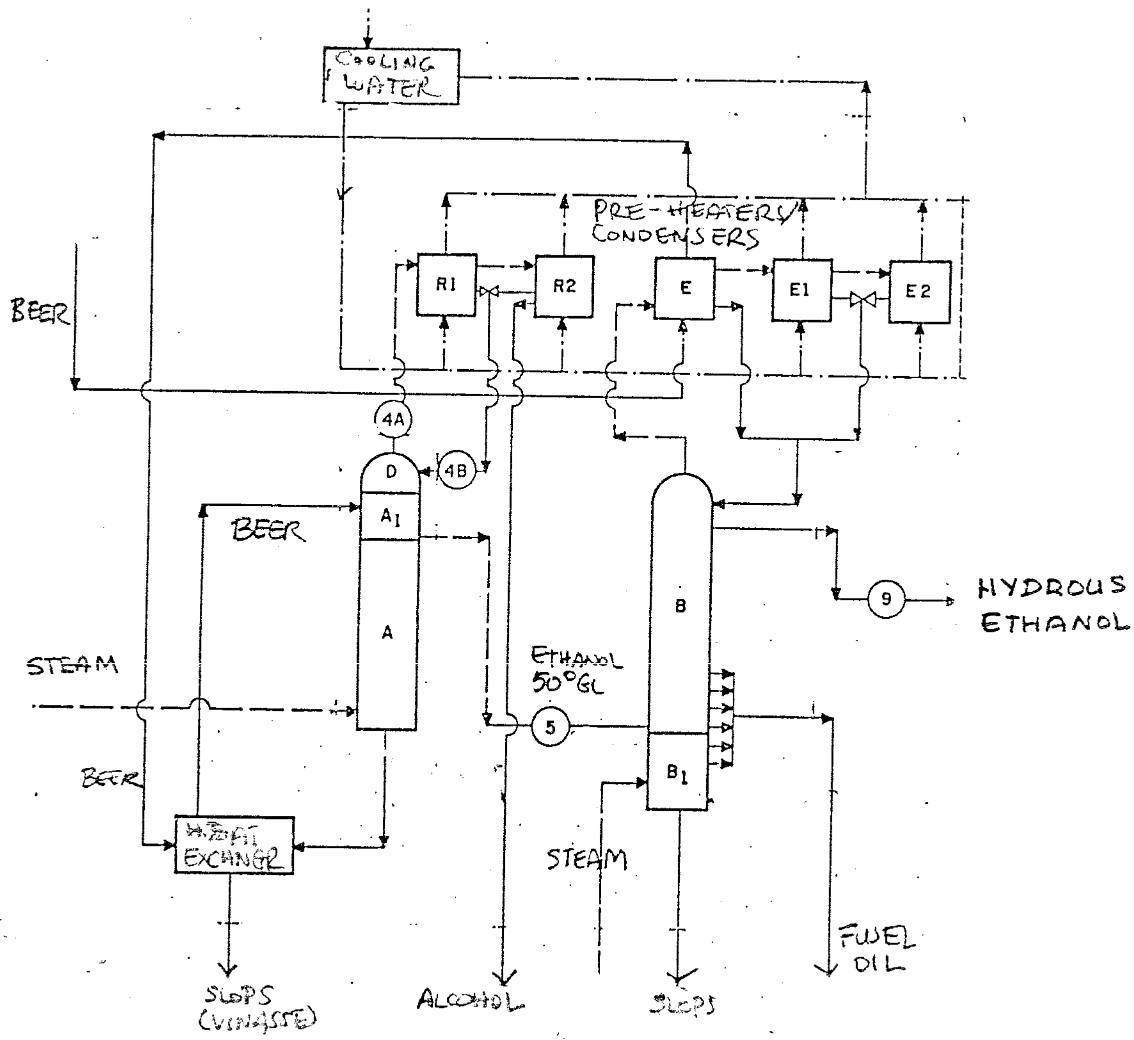


Figure 15a. Flow diagram for a typical Brazilian distillation system for producing motor grade hydrous alcohol [31]. Two distillation columns are used to concentrate and purify the ethanol. Low pressure (0.15-0.25 MPa) steam provides heat for the process, requiring about 3.0-3.5 kilograms of steam per liter of hydrous ethanol.

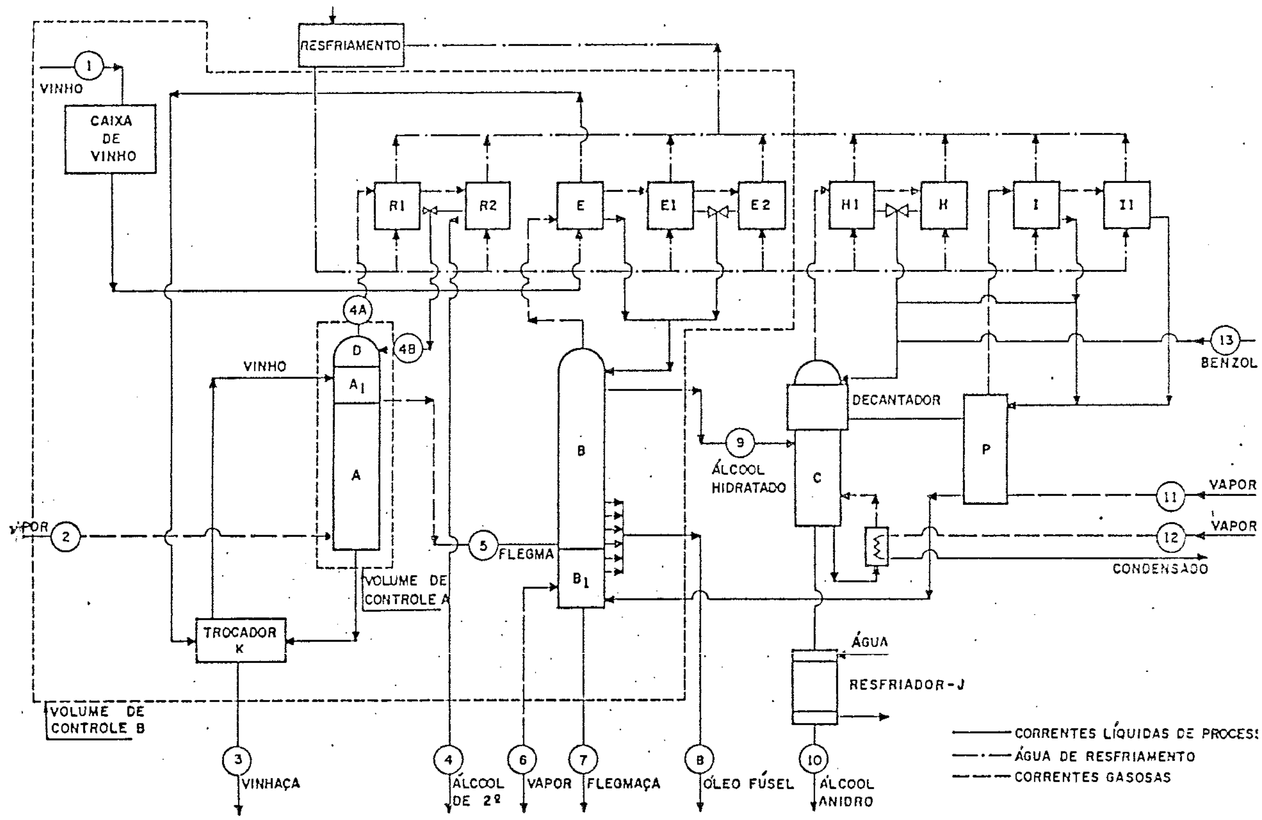


Figure 15b. Flow diagram for a typical Brazilian distillation system for producing motor grade anhydrous alcohol [31]. As in Figure 15a, two distillation columns are used to produce hydrous (azeotropic) ethanol. Azeotropic distillation with benzene is accomplished in a dehydration column, and benzene is recovered in a benzene column. Low pressure (0.15-0.25 MPa) steam provides heat for the process, requiring about 4.5-5.0 kilograms of steam per liter of anhydrous ethanol.

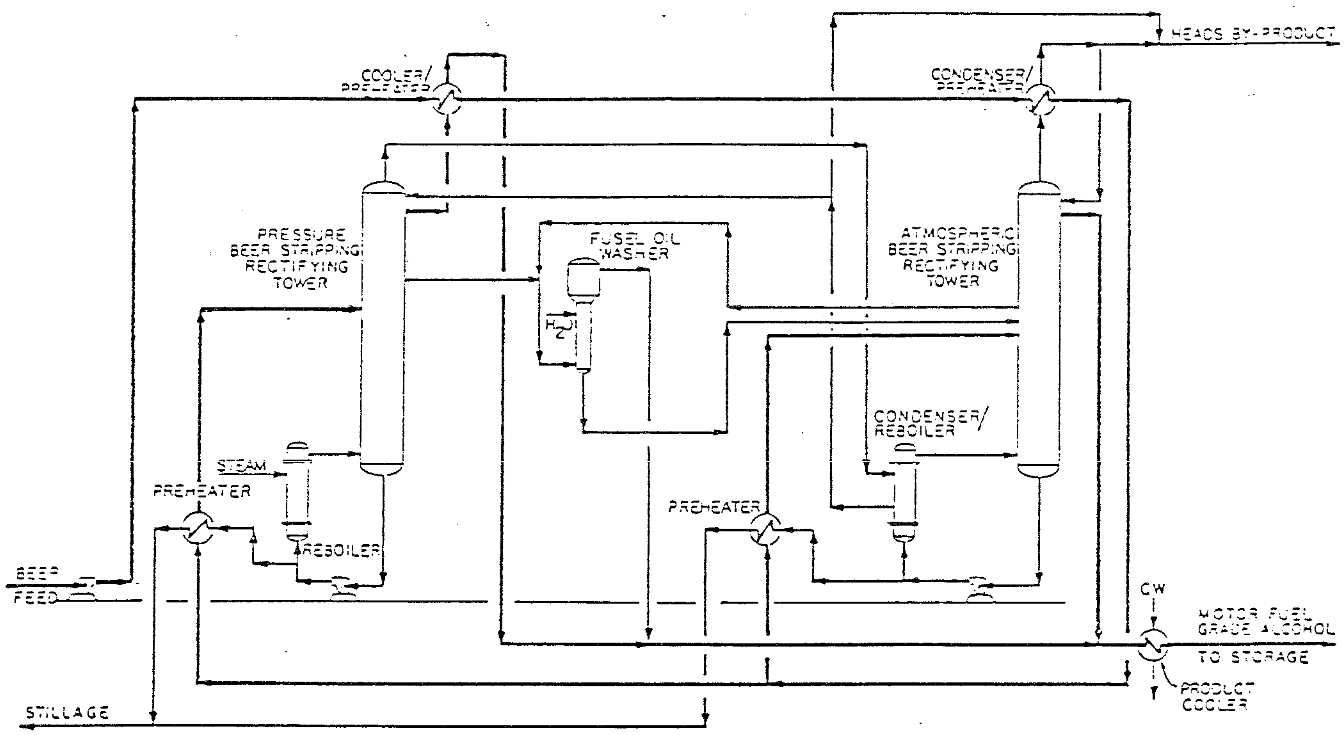


Figure 16a. A heat integrated design for production of hydrous ethanol [55].

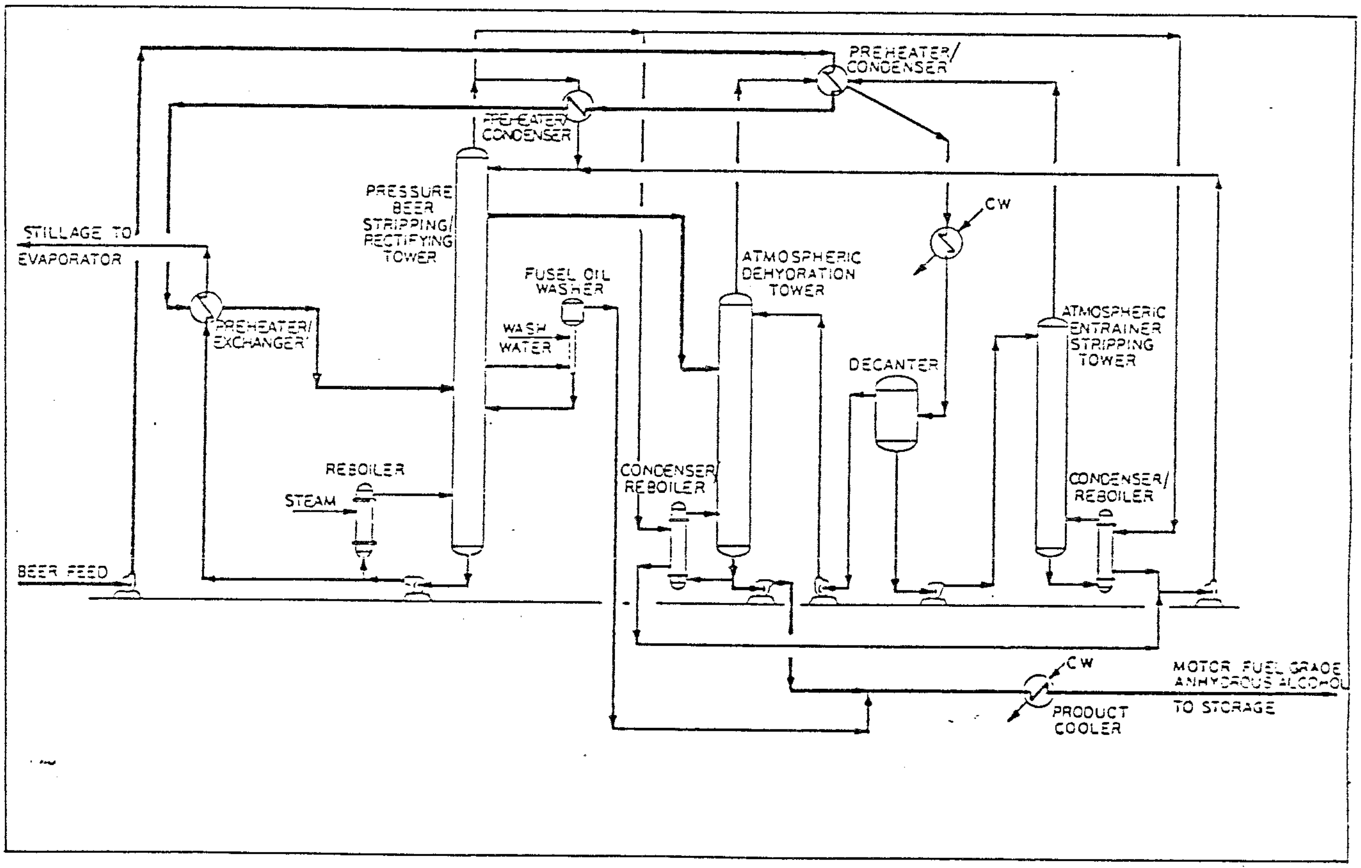


Figure 16b. A heat integrated design for production of anhydrous ethanol [55].

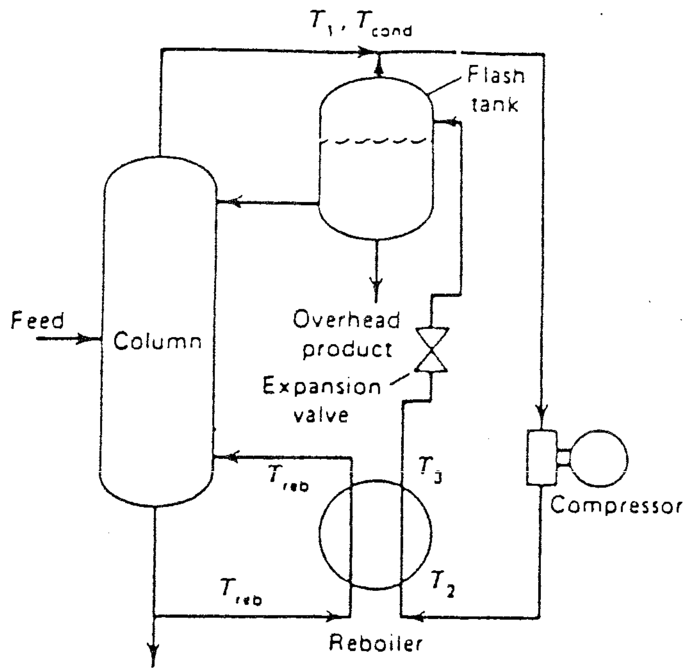


Figure 17. Example of mechanical vapor recompression in distillation [54].

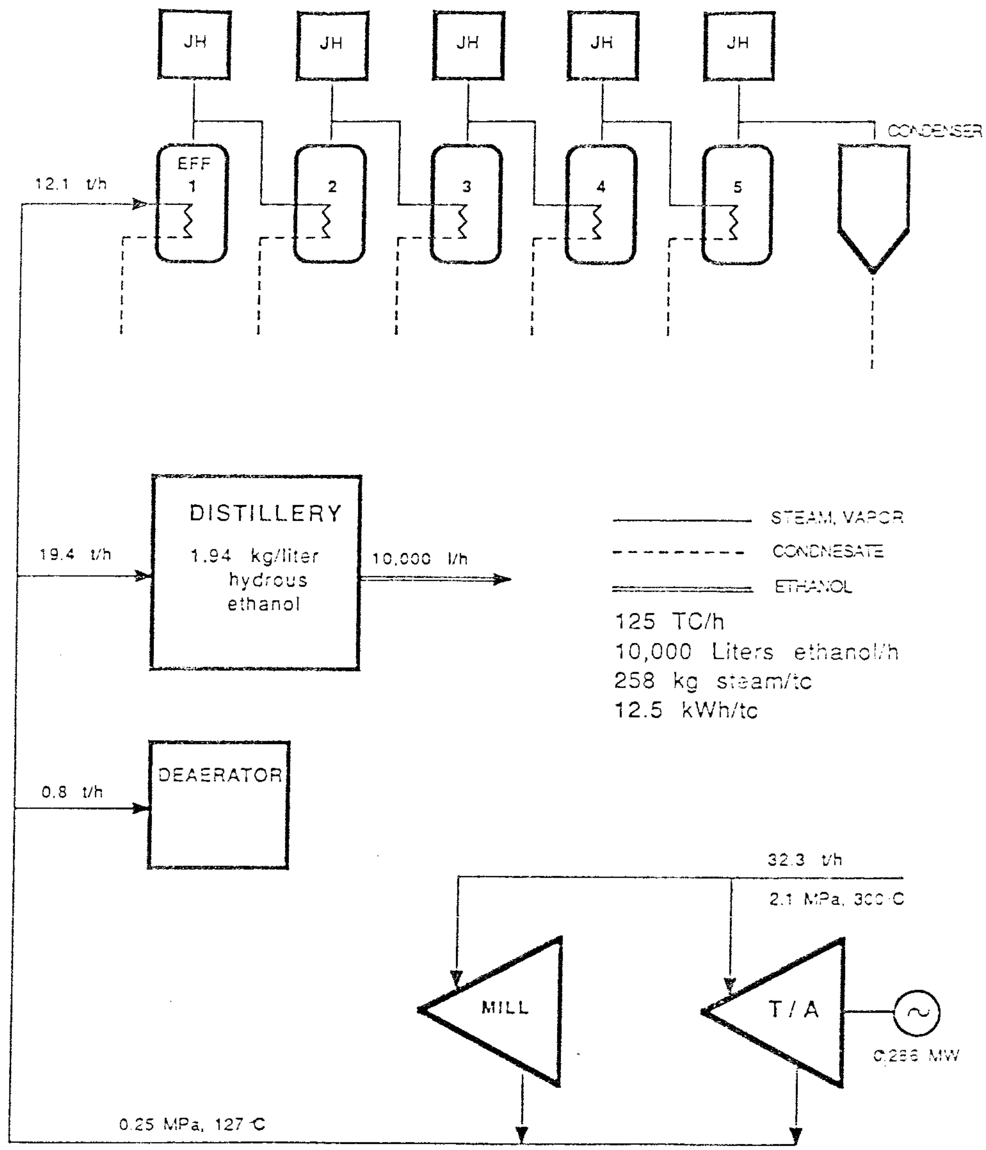


Figure 18a. Factory flow diagram showing steam and material flows in a steam conserving Brazilian autonomous distillery [28], milling 125 tonnes cane per hour and producing 10,000 liters per hour of hydrous ethanol. Steam use for juice heating is reduced by using vapors bled from the evaporator, and a lower energy use distillation system with heat integration is used, requiring 1.9 liters of steam per liter of hydrous ethanol. The estimated factory steam demand is 258 kg/tc. The electricity demand is 12.5 kwh/tc.

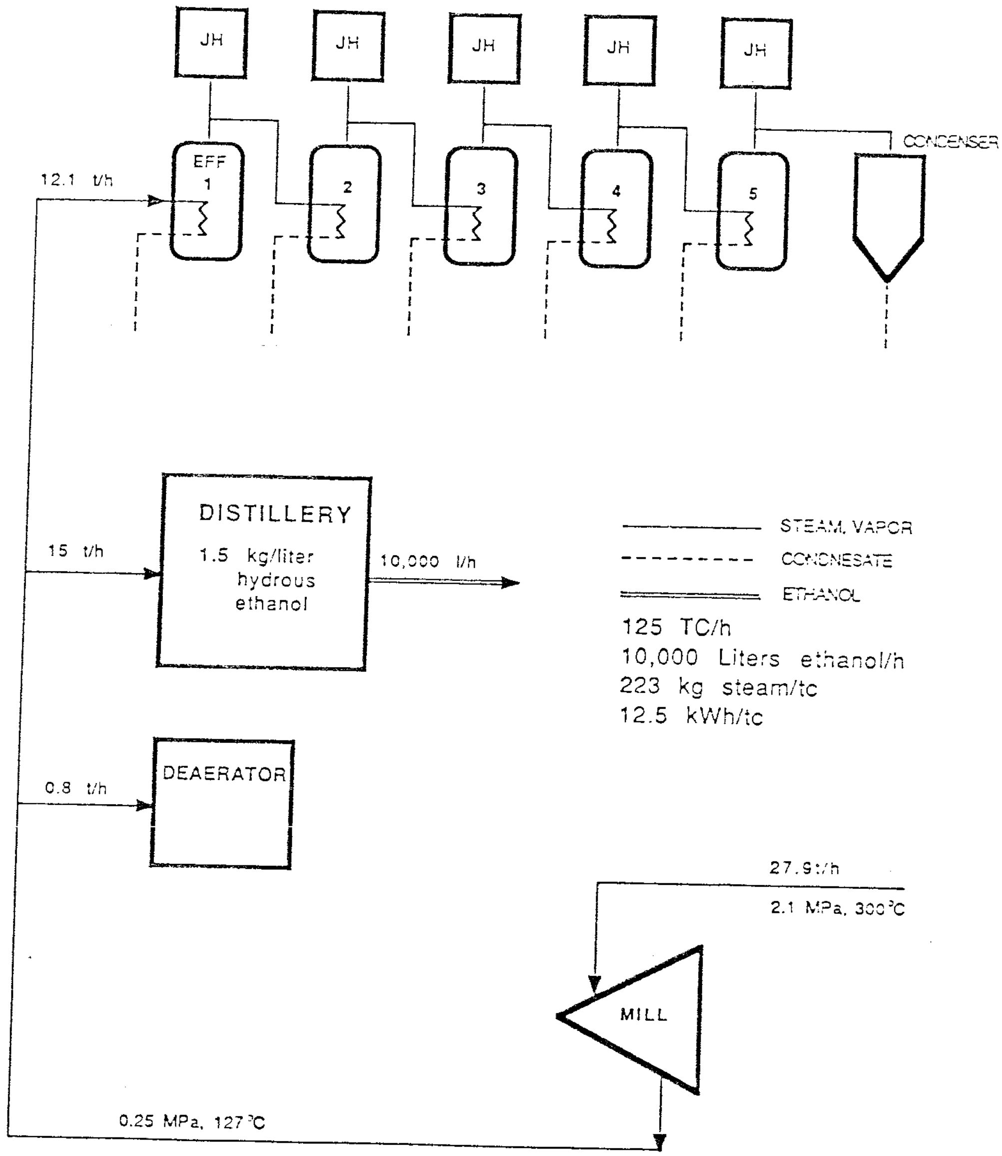


Figure 18b. Factory flow diagram showing steam and material flows in a steam conserving Brazilian autonomous distillery [28], milling 125 tonnes cane per hour and producing 10,000 liters per hour of hydrous ethanol. Steam use for juice heating is reduced by using vapors bled from the evaporator, and a lower energy use distillation system with heat integration is used, requiring 1.5 liters of steam per liter of hydrous ethanol. The estimated factory steam demand is 225 kg/tc. The electricity demand is 12.5 kwh/tc.

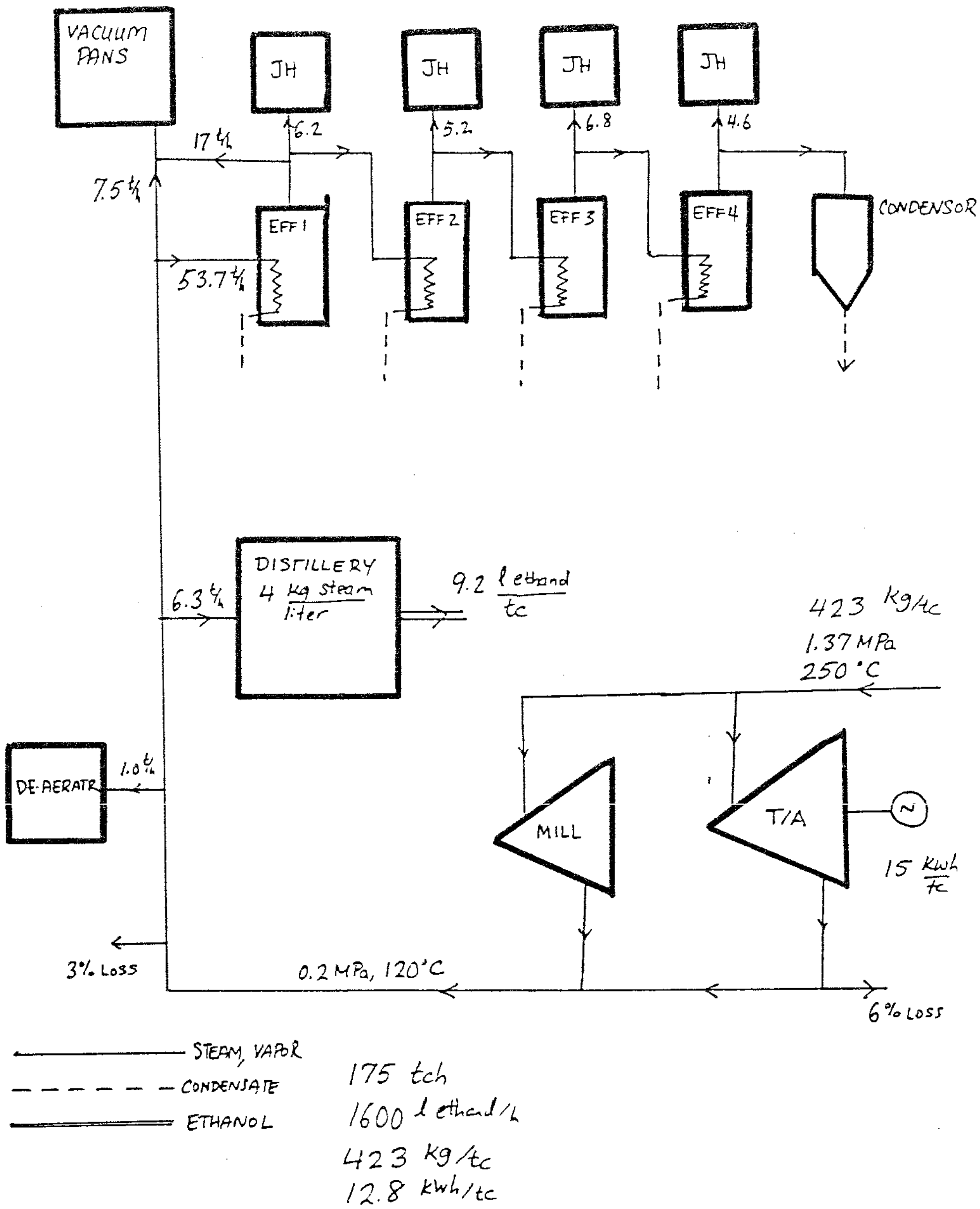


Figure 19a. Factory flow diagram showing steam and material flows in a conventional annexed distillery producing ethanol from C-molasses. The factory mills 275 tonnes cane per hour and produces 2500 liters per hour of anhydrous ethanol. A conventional distillation system is used, requiring 4.0 liters of steam per liter of anhydrous ethanol. The estimated factory steam demand is 423 kg/tc. The electricity demand is 12.8 kwh/tc.

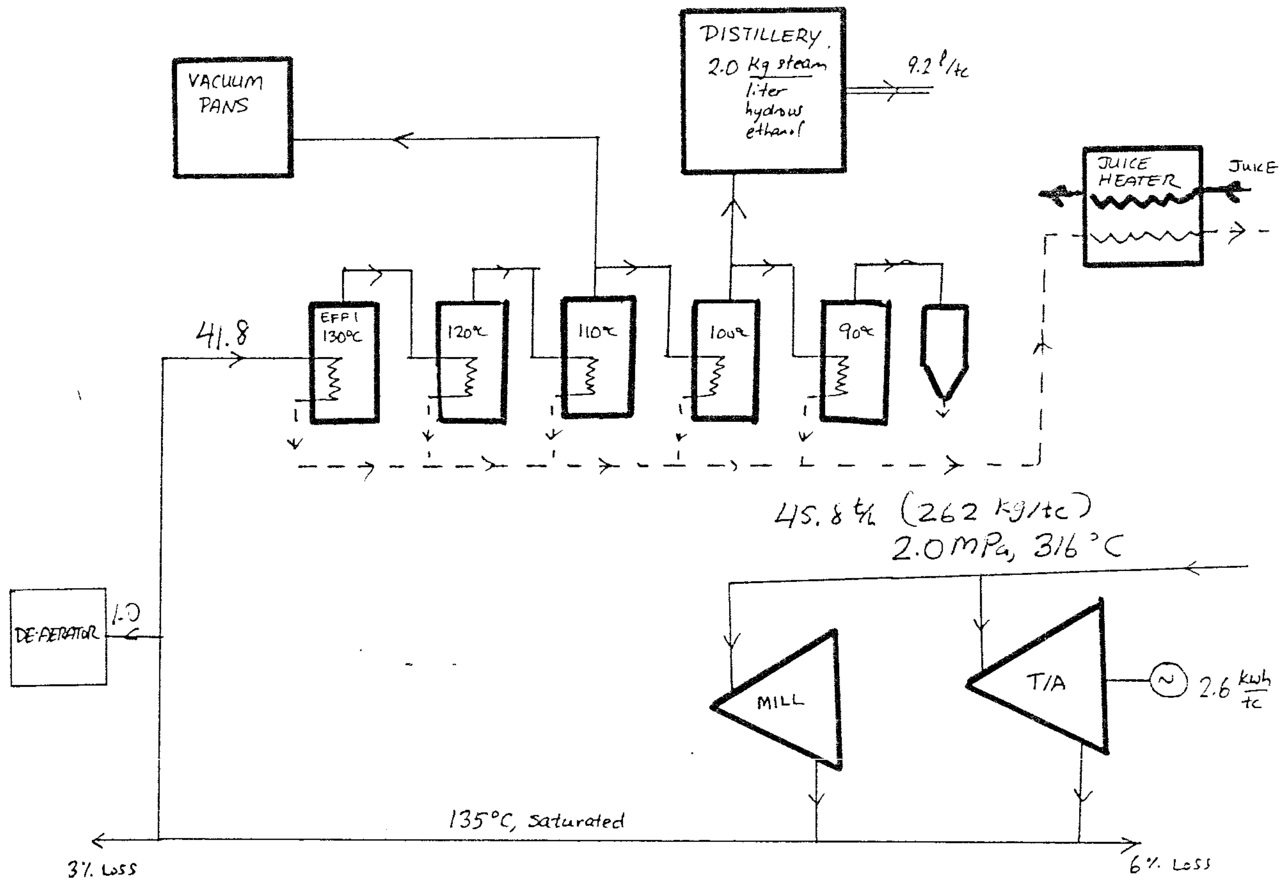


Figure 19b. Factory flow diagram showing steam and material flows in a steam-conserving annexed distillery producing ethanol from C-molasses. The factory mills 175 tonnes cane per hour and produces 1600 liters per hour of anhydrous ethanol. A falling film evaporator and continuous vacuum pans are used to reduce steam use, and heat is recovered from hot condensates for juice heating. A low energy use distillation system with heat integration is used, requiring 2.0 liters of steam per liter of hydrous ethanol. The estimated factory steam demand is 262 kg/tc. The electricity demand is 12.8 kwh/tc.

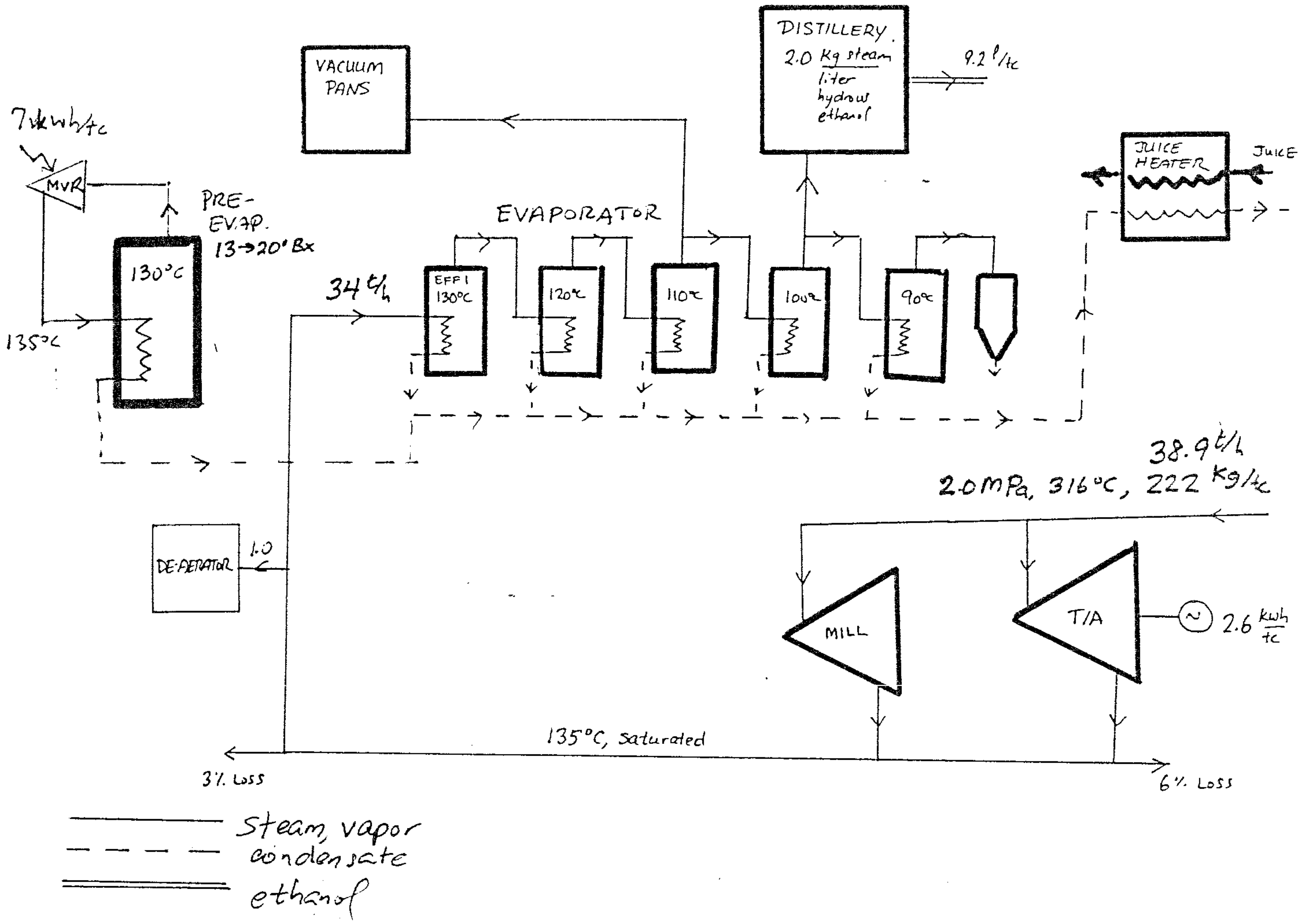


Figure 19c. Factory flow diagram showing steam and material flows in a steam-conserving annexed distillery producing ethanol from C-molasses. The factory mills 175 tonnes cane per hour and produces 1600 liters per hour of anhydrous ethanol. A falling film evaporator and continuous vacuum pans are used to reduce steam use, and heat is recovered from hot condensates for juice heating. A pre-evaporator stage with mechanical vapor recompression is used to save steam. A low energy use distillation system with heat integration is used, requiring 1.5 liters of steam per liter of hydrous ethanol. The estimated factory steam demand is 220 kg/tc. The electricity demand is 20 kwh/tc.

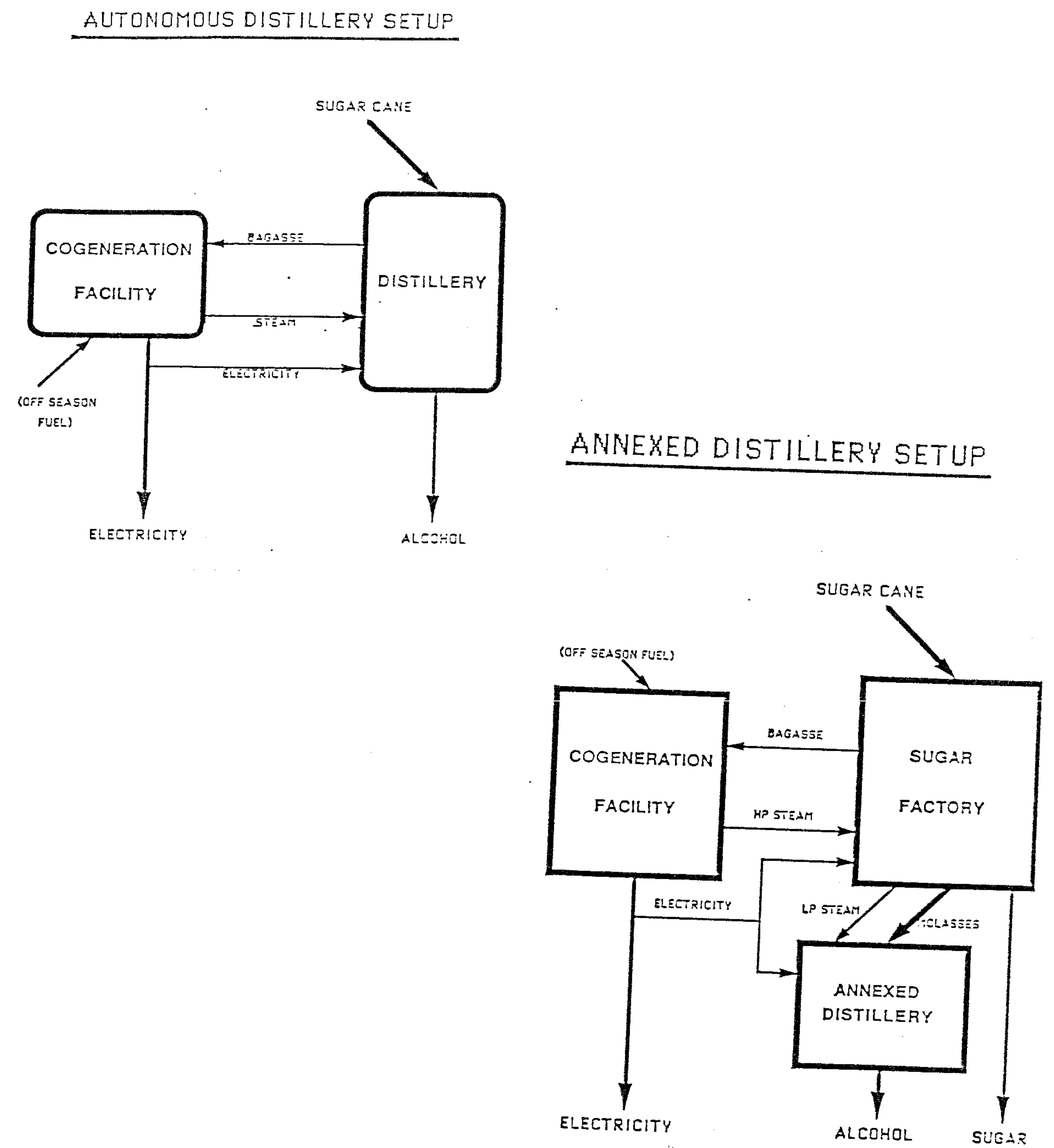


Figure 20. Material and energy flows for autonomous (top) and annexed (bottom) distilleries with cogeneration. It is assumed that the cogenerator buys bagasse fuel from the factory during the milling season and sells steam and electricity to the factory. During the off-season, the cogeneration fuel could be barbojo (purchased from the factory or from the cane grower), fuel wood, oil or another alternative fuel.

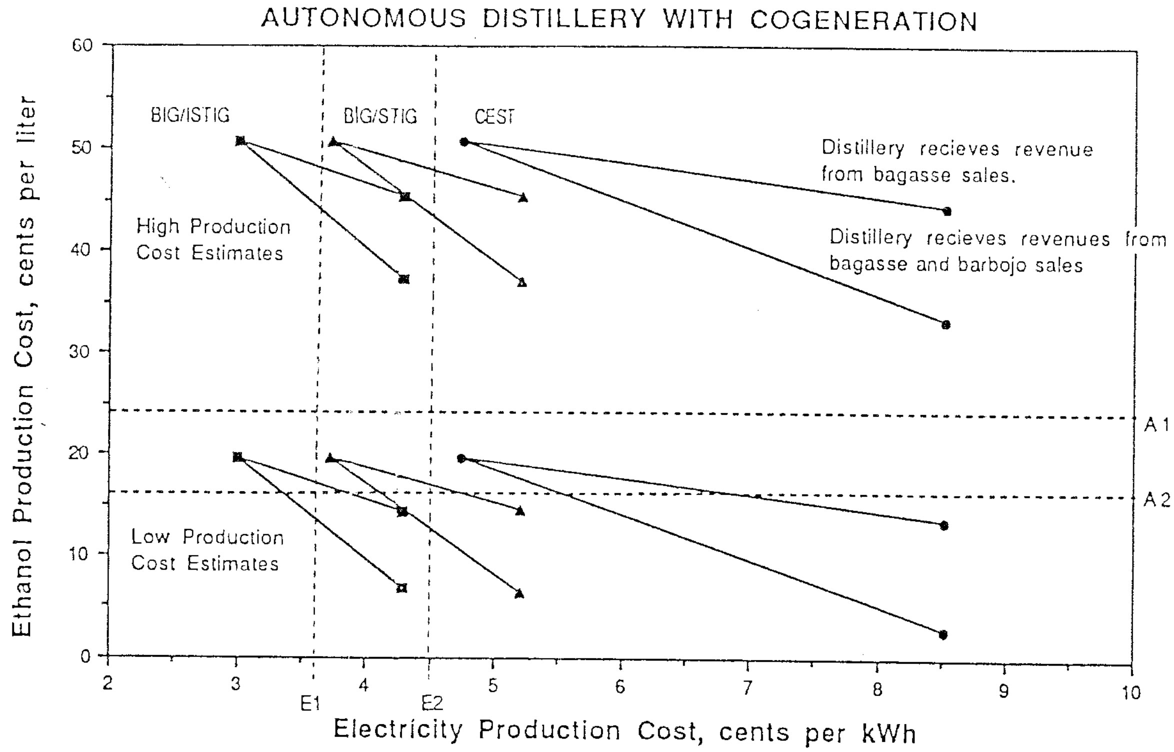


Figure 21. The cost of ethanol and electricity production for autonomous distilleries with cogeneration. Three biomass-fired cogeneration systems are considered -- the condensing-extraction steam turbine (CEST), the biomass integrated gasifier/steam-injected gas turbine (BIG/STIG), and the biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG) -- and two sets of ethanol production costs corresponding to high and low cost conditions. (See Tables 7a, 7b.) It is assumed that the distillery sells bagasse fuel to the cogenerator during the milling season and barbojo during the off season. (Barbojo could also be purchased from the cane grower.)

For each of the six cases, a range of ethanol and electricity production costs is shown corresponding to different revenues received by the factory from selling biomass fuel to the cogenerator. As the price the cogenerator pays for biomass increases (moving from left to right along each line), the cost of ethanol production decreases because the distillery receives higher revenues from selling its bagasse (or bagasse and barbojo) to the cogeneration plant. The cost of electricity production increases, because the cogenerator pays a higher price for fuel.

Along each line three points are shown. The left endpoint corresponds to the case when the distiller earns no net revenue for biomass sales: bagasse is given to the cogenerator in exchange for the steam and electricity needed to run the factory and barbojo is sold at its production cost. The circle indicates the point at which the average total cost of fuel is \$3/GJ. This is the estimated cost of air-dried wood chips, which could be an alternative cogeneration fuel in some situations.

Several reference lines are also shown. The dashed horizontal line labelled A1 is the cost of ethanol which would be equivalent to the current US wholesale gasoline price of \$0.75/gallon, assuming that 1.2 liters of alcohol are worth 1 liter of gasoline. The dashed line A2 is the value of ethanol as an octance enhancing additive to gasoline, in which case it is assumed that 0.8 liters of ethanol are worth 1 liter of gasoline. The dashed vertical line E1 is the fuel cost only for a thermal power plant burning residual oil (about 3.6 cents/kwh), assuming a heat rate of 13,120 kJ/kwh and a fuel oil price of \$2.63/GJ, vertical line E2 represents the cost of electricity from hydropower, assuming a capital cost of \$1500/kW (about 4.5 cents/kwh), and vertical line E3 the cost of electricity from a new coal plant assuming a capital cost of \$1400/kW and a coal price of \$1.7/GJ.

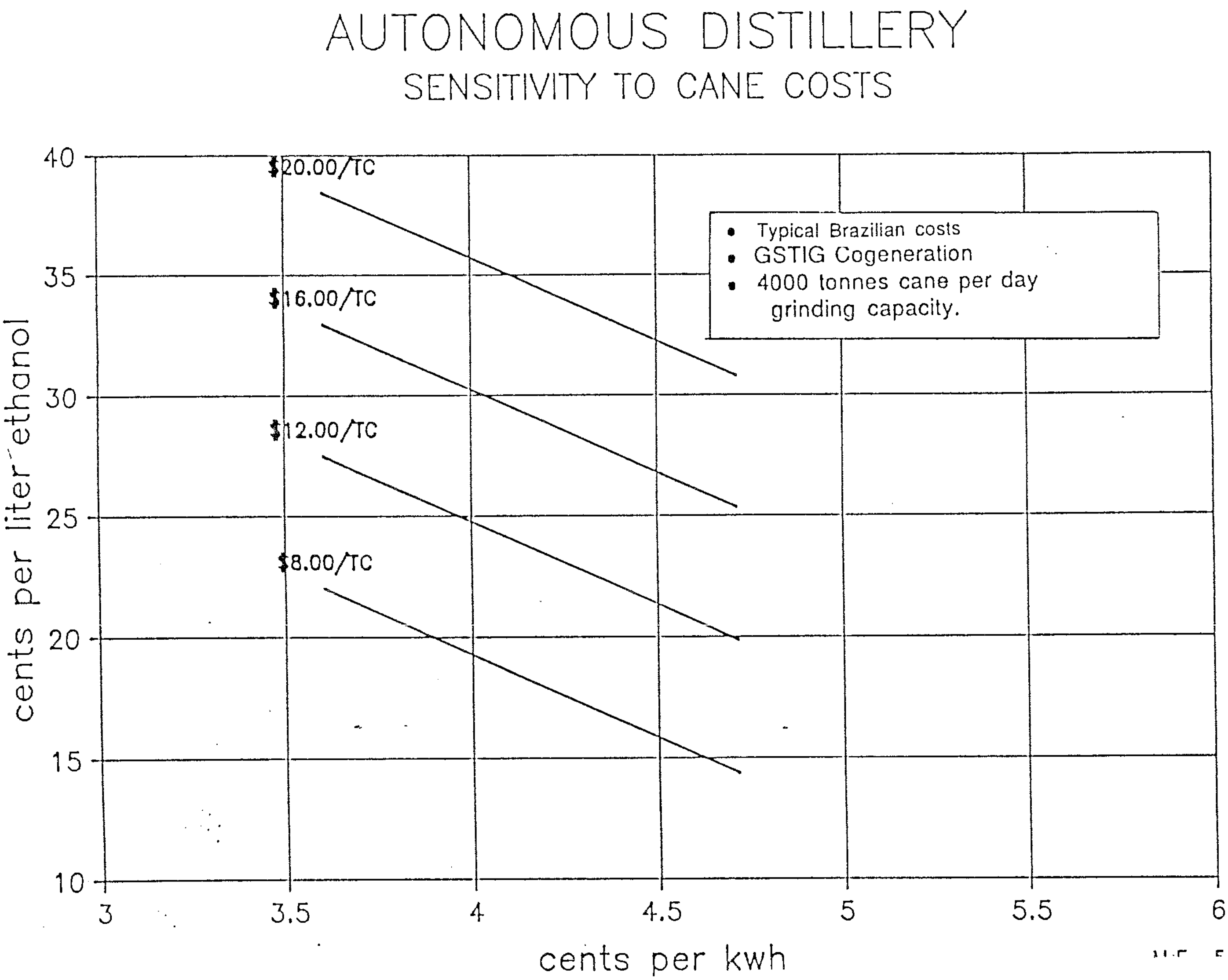


Figure 22. The cost of ethanol and electricity production for autonomous distilleries with cogeneration as a function of the cost of sugar cane delivered to the factory. Results are shown for the BIG/STIG system. We have used Brazilian cost estimates except for the cane cost, which is varied. As in Figure 21, each line corresponds to a range of charges for biomass fuel paid by the cogenerator to the distillery.

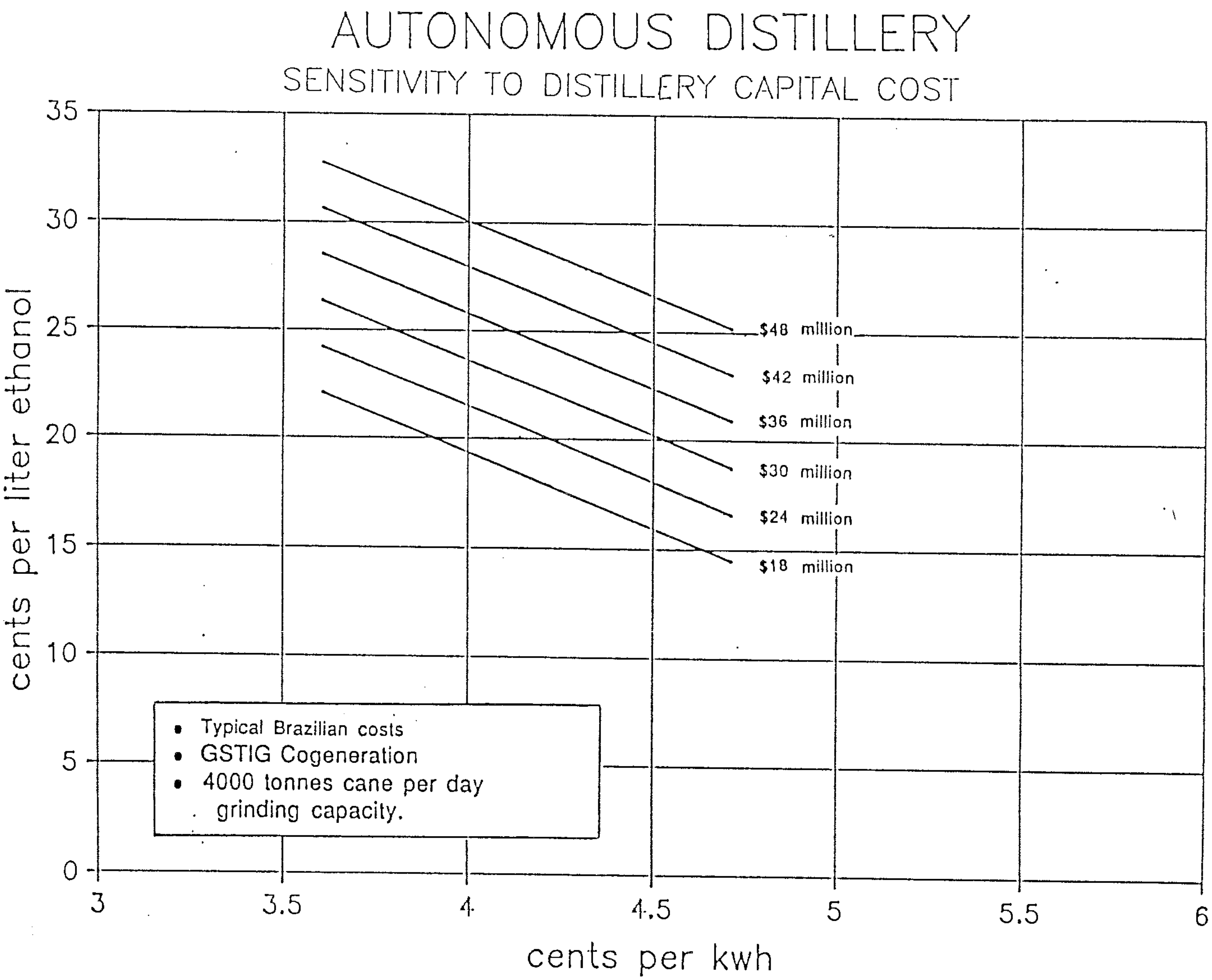
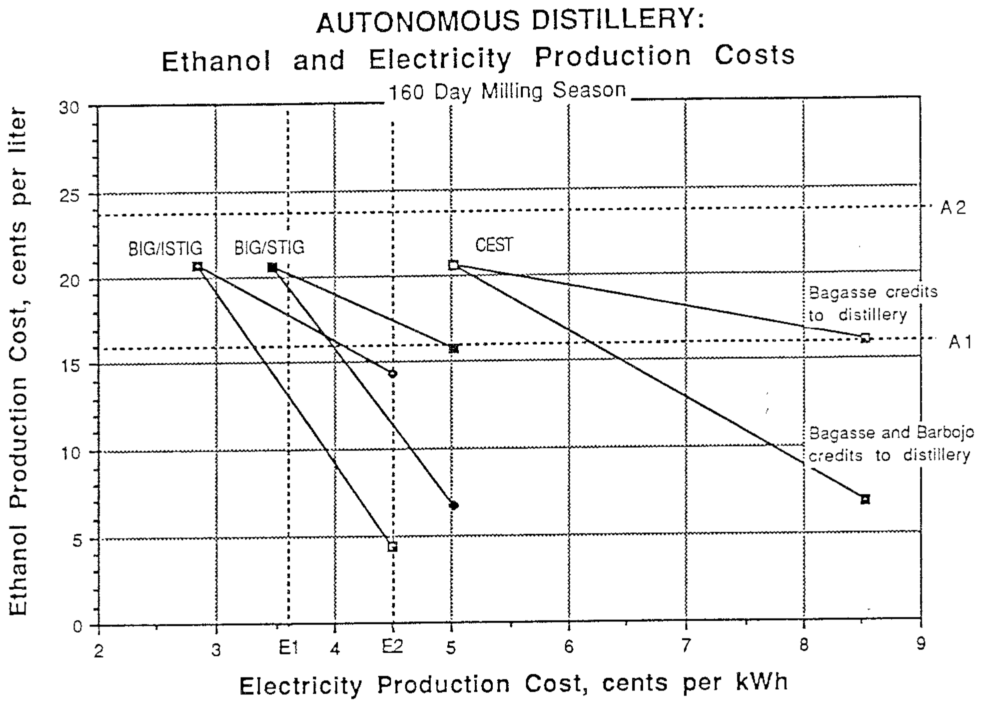


Figure 23. The cost of ethanol and electricity production for autonomous distilleries with cogeneration as a function of the installed capital cost of the distillery. Results are shown for the BIG/STIG system. We have used Brazilian cost estimates except for the distillery capital cost, which is varied. As in Figure 21, each line corresponds to a range of charges for biomass fuel paid by the cogenerator to the distillery.



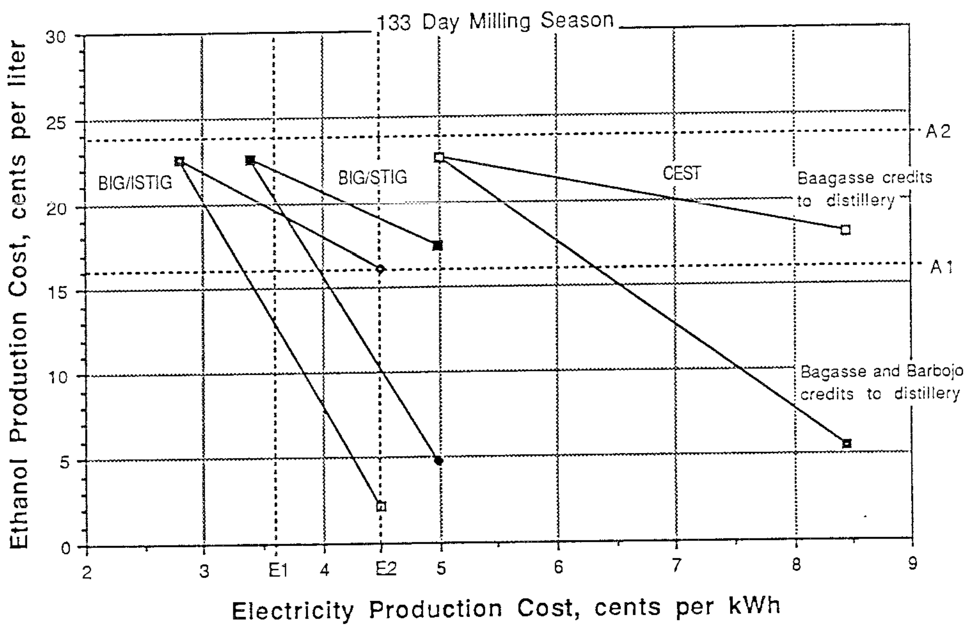


Figure 24. The cost of ethanol and electricity production for autonomous distilleries with cogeneration as a function of the length of the milling season, for the three cogeneration technologies and Brazilian cost estimates. In the top graph, the milling season is 160 days. In this case 59% of the barbojo is used as an off-seaon fuel. In the bottom graph, the milling season is 133 days and 80% of the barbojo is used as an off-season cogeneration fuel. As in Figure 21, each line corresponds to a range of charges for biomass fuel paid by the cogenerator to the distillery.

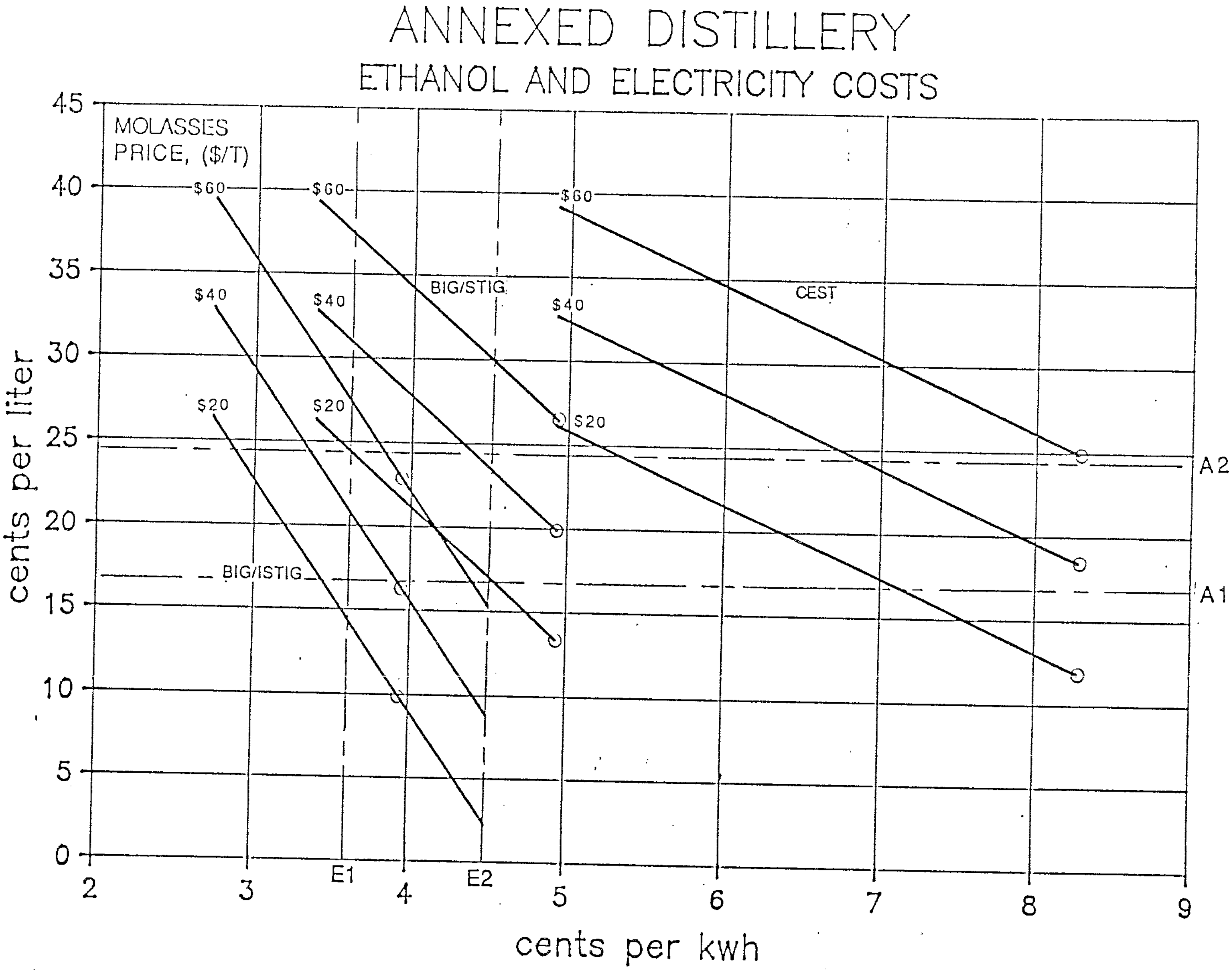


Figure 25. Cost of ethanol (from C-molasses) and electricity production in annexed distillery with cogeneration. Results are shown for three types of biomass-fired cogeneration systems (see Figure 21), and for three C-molasses prices $20, 40 and 60/tonne. It is assumed that the cogenerator purchases bagasse during the milling season and barbojo in the off-season from the distillery, and 15% of the revenues from the sales of bagasse and barbojo are credited against the cost of alcohol production. The dashed reference lines labelled A1, A2, E1, E2 are as indicated in Figure 21.

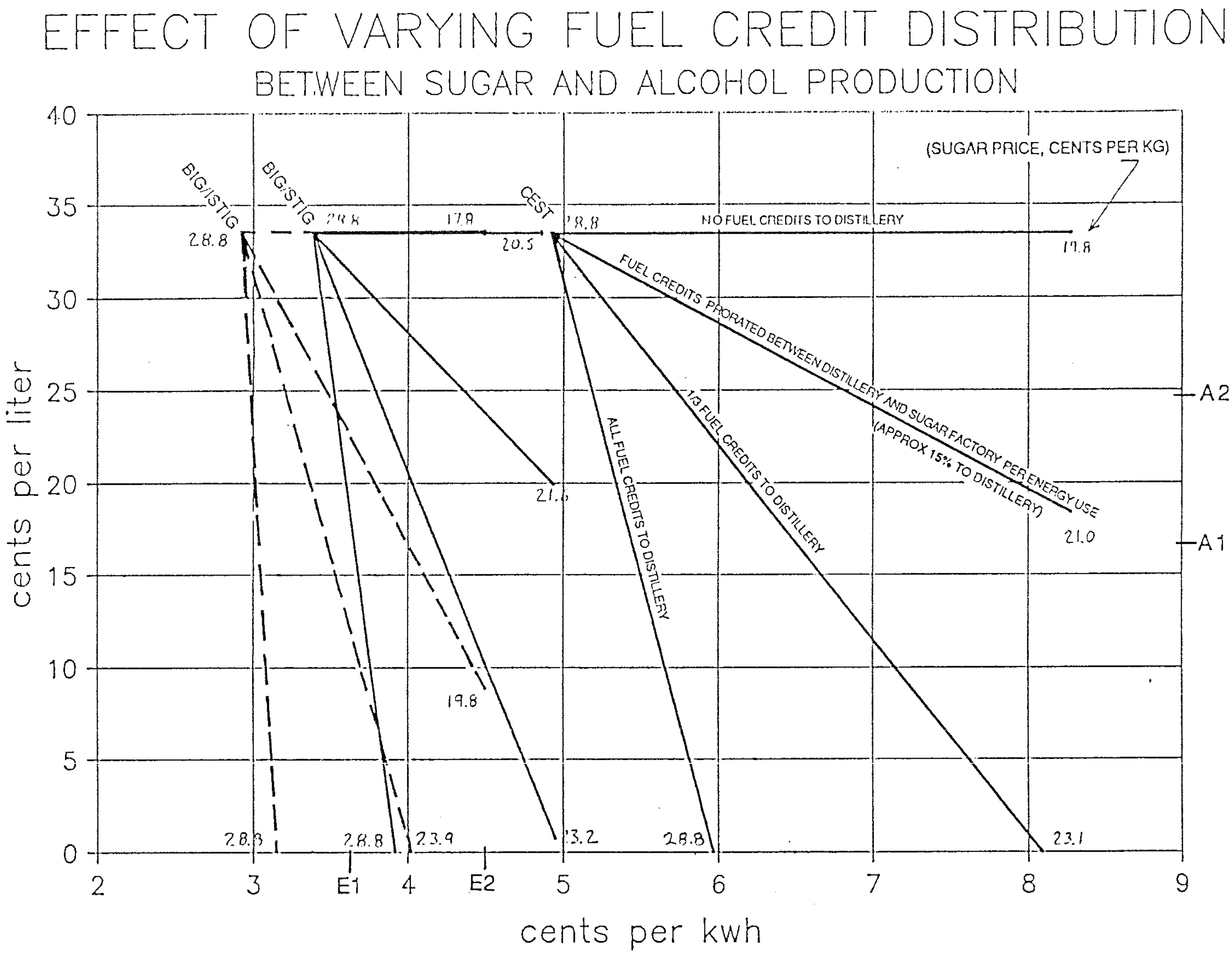


Figure 26. The costs of producing electricity and ethanol from C-molasses (priced at $40/tonne) at an annexed distillery with varying fractions of biomass fuel revenues credited against alcohol production. Four lines are shown for each cogeneration technology. Along each line the biomass fuel charges are calculated as in Figure 21.

Each of the four lines corresponds to a different fraction of biomass revenues allocated to alcohol production. The horizontal line shows the cost when all the revenues are credited to sugar production, and none to alcohol production. Moving clockwise from the horizontal line the next three lines show costs assuming 15%, 33%, and 100% of revenues from biofuels are credited against the cost of ethanol production.

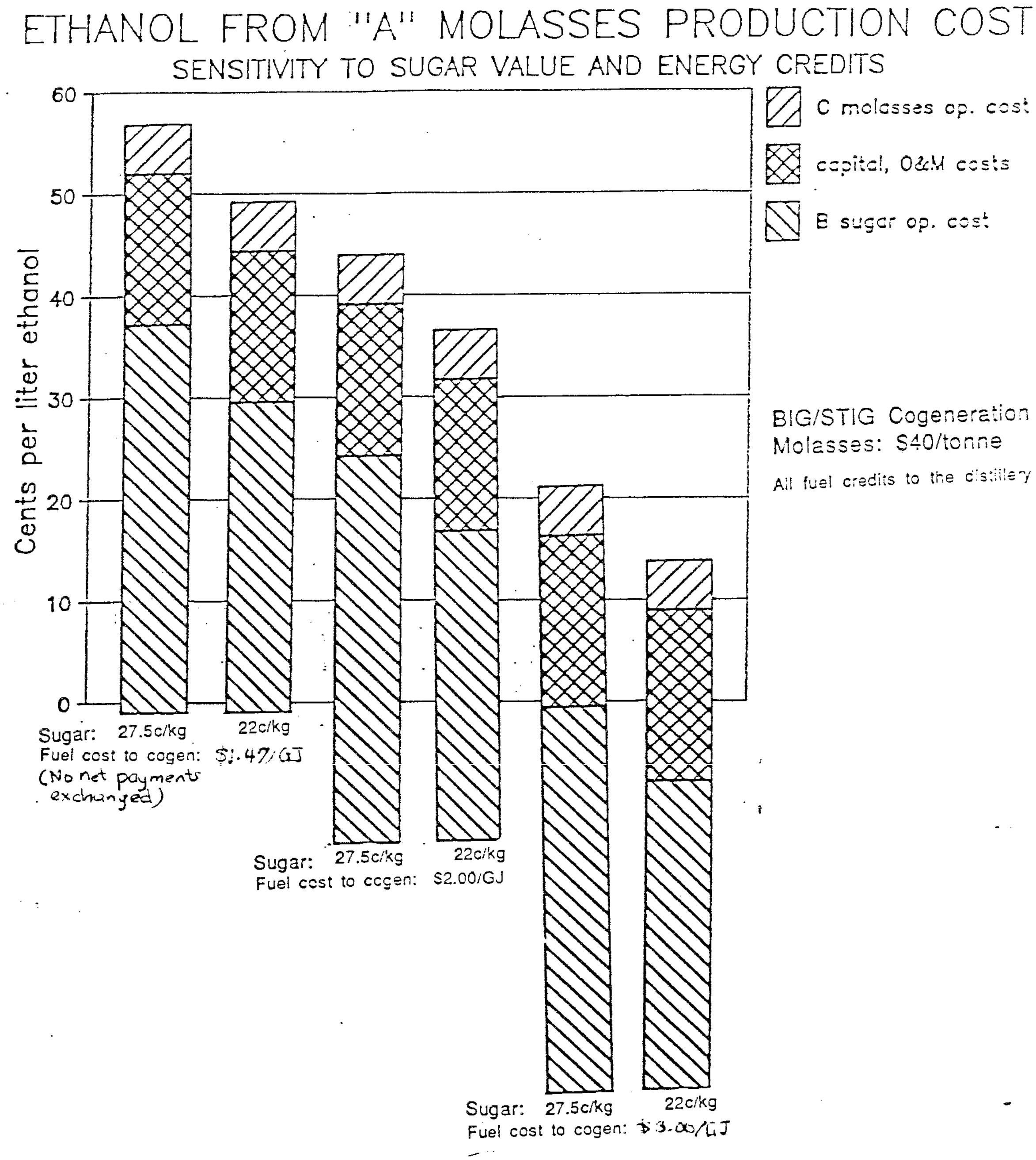


Figure 27. The cost of ethanol production from A-molasses as a function of C-molasses price, sugar price, non-fuel costs (capital and operation and mainetance) and the total fuel cost to the cogenerator.

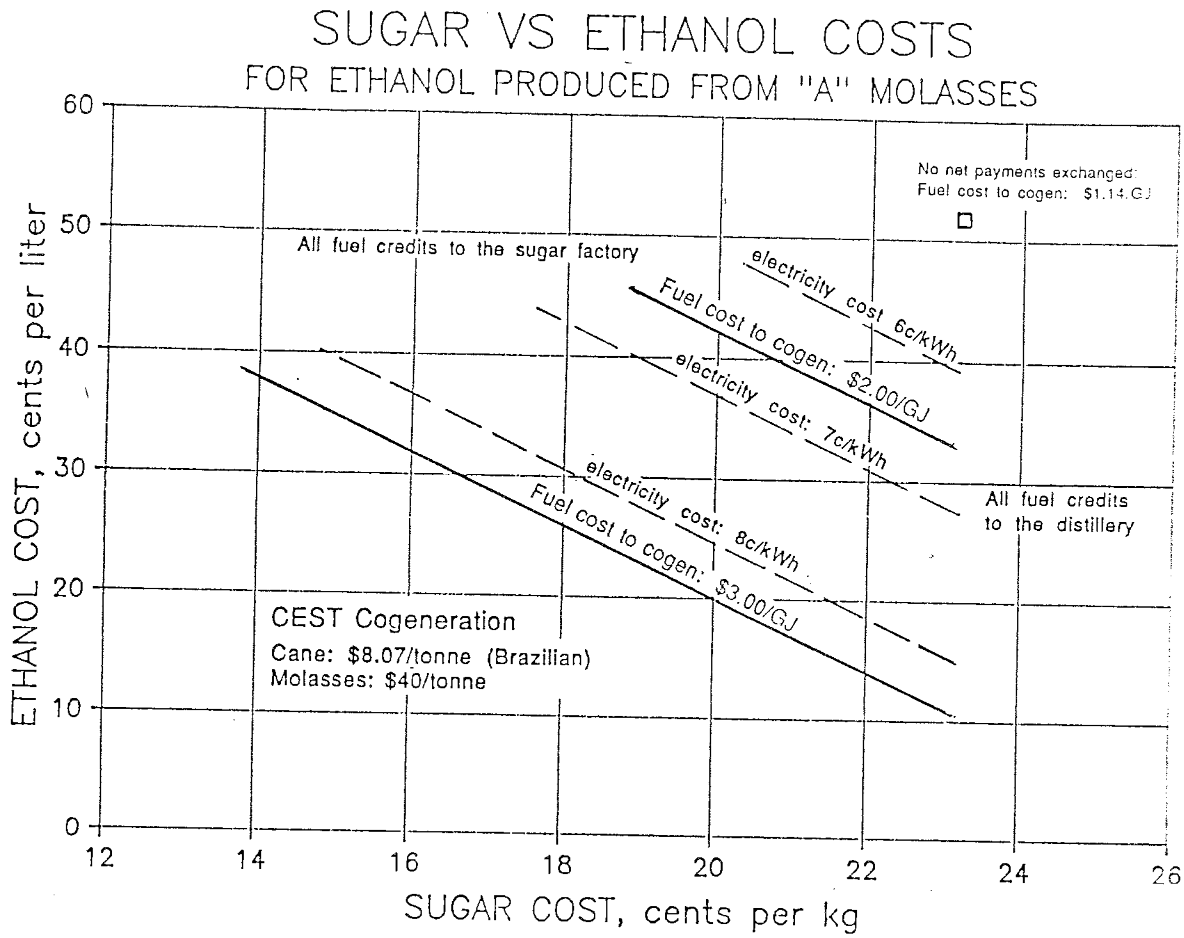


Figure 28. Cost of ethanol and electricity production in annexed distillery making ethanol from A-molasses for the CEST cogeneration technology, as a function of the cogenerator's fuel cost and the allocation of the biomass fuel credit between alcohol and sugar.

It is assumed that the distillery receives a 15% internal rate of return, that the price of C-molasses is $40/tonne, and that cane costs $8/tonne. The distiller sells bagasse to the cogenerator during the milling season and barbojo during the off-season.

Three levels of biomass fuel credits are shown. In the upper right corner, the costs of ethanol and electricity are shown assuming that no net payments are exchanged between the distiller and the cogenerator (the distiller gives the cogenerator cane residues at cost in exchange for steam and electricity to run the factory). Also shown are lines of constant biomass fuel credit assuming that the cogenerator pays a total of $2/GJ and $3/GJ for fuel. The left endpoint of each line shows ethanol and electricity costs when all the biomass fuel credits are allocated against sugar production; the right endpoint when all the biomass fuel credits are allocated against sugar production. Lines of constant electricity cost are also indicated.

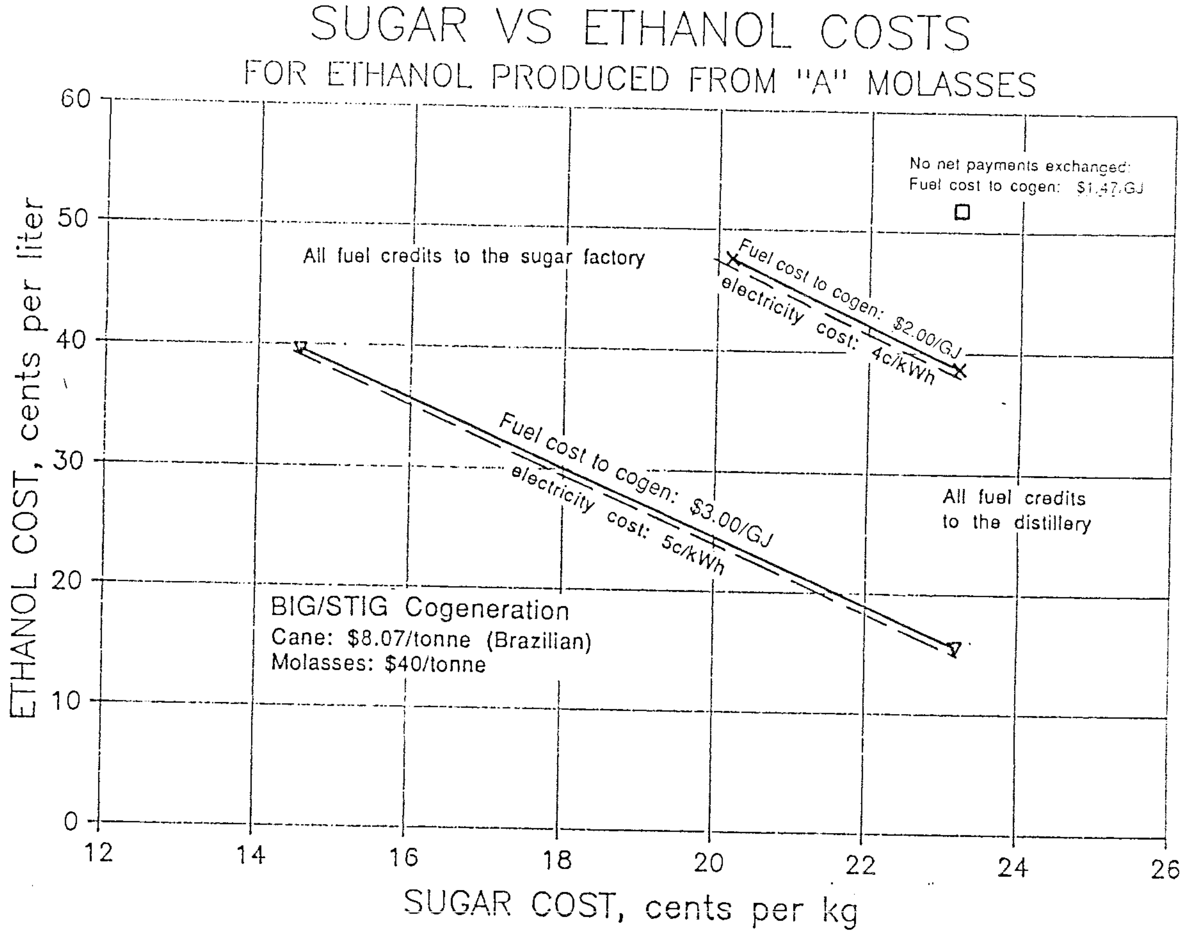


Figure 29. Cost of ethanol and electricity production in annexed distillery making ethanol from A-molasses for the BIG/STIG cogeneration technology, as a function of the cogenerator's fuel cost and the allocation of the biomass fuel credit between alcohol and sugar.

It is assumed that the distillery receives a 15% internal rate of return, that the price of C-molasses is $40/tonne, and that cane costs $8/tonne. The distiller sells bagasse to the cogenerator during the milling season and barbojo during the off-season.

Three levels of biomass fuel credits are shown. In the upper right corner, the costs of ethanol and electricity are shown assuming that no net payments are exchanged between the distiller and the cogenerator (the distiller gives the cogenerator cane residues at cost in exchange for steam and electricity to run the factory). Also shown are lines of constant biomass fuel credit assuming that the cogenerator pays a total of $2/GJ and $3/GJ for fuel. The left endpoint of each line shows ethanol and electricity costs when all the biomass fuel credits are allocated against sugar production; the right endpoint when all the biomass fuel credits are allocated against sugar production. Lines of constant electricity cost are also indicated.

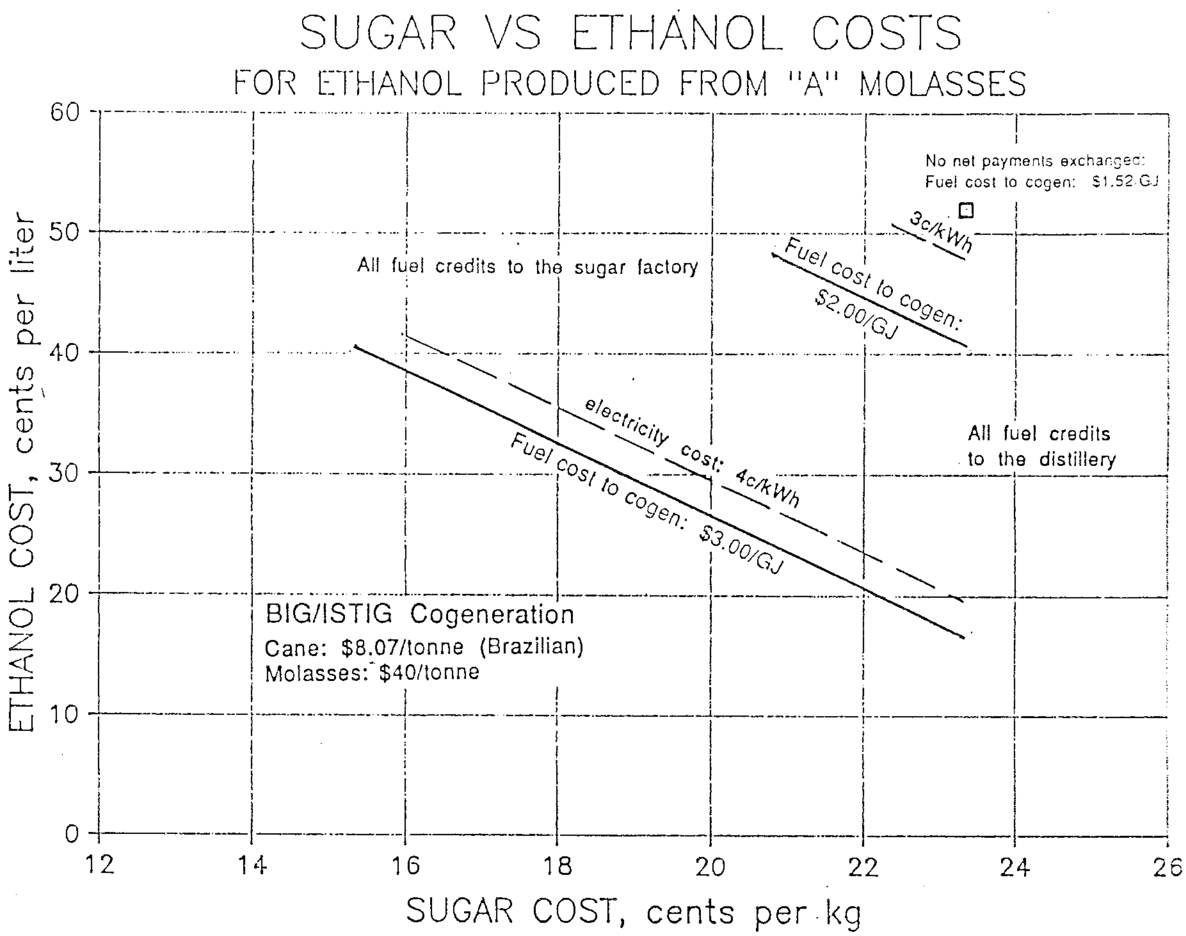


Figure 30. Cost of ethanol and electricity production in annexed distillery making ethanol from A-molasses for the BIG/ISTIG cogeneration technology, as a function of the cogenerator's fuel cost and the allocation of the biomass fuel credit between alcohol and sugar.

It is assumed that the distillery receives a 15% internal rate of return, that the price of C-molasses is $40/tonne, and that cane costs $8/tonne. The distiller sells bagasse to the cogenerator during the milling season and barbojo during the off-season.

Three levels of biomass fuel credits are shown. In the upper right corner, the costs of ethanol and electricity are shown assuming that no net payments are exchanged between the distiller and the cogenerator (the distiller gives the cogenerator cane residues at cost in exchange for steam and electricity to run the factory). Also shown are lines of constant biomass fuel credit assuming that the cogenerator pays a total of $2/GJ and $3/GJ for fuel. The left endpoint of each line shows ethanol and electricity costs when all the biomass fuel credits are allocated against sugar production; the right endpoint when all the biomass fuel credits are allocated against sugar production. Lines of constant electricity cost are also indicated.

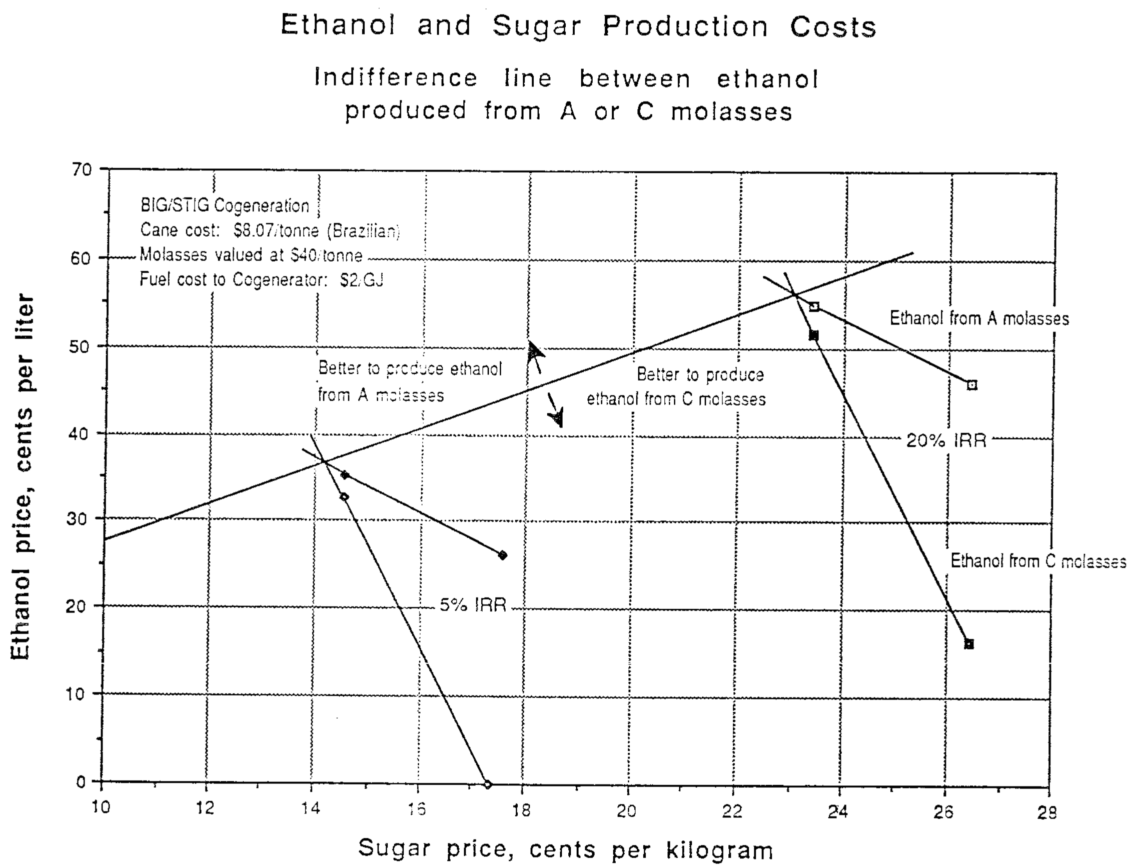


Figure 31. Rates of return of ethanol and electricity production in annexed distillery with a flexible product strategy as a function of sugar and ethanol prices. Lines of equal rate of return are shown for ethanol production from C-molasses (sharply slanting lines), and from A-molasses (less steeply slanting lines). Also shown is a line where the rates of return are equal for making ethanol from C-molasses or A-molasses. Above this break-even line, it is preferable to make more ethanol and less sugar (make ethanol from A-molasses); below the line a higher rate of return is obtained by making the maximum amount of sugar plus ethanol from C-molasses.

APPENDIX A - PRELIMINARY ESTIMATES OF THE COST AND PERFORMANCE OF BIOMASS INTEGRATED GASIFIER/INTERCOOLED STEAM INJECTED GAS TURBINE SYSTEMS

J. Ogden
S. Consonni
R. Williams

The cost and performance of condensing extraction steam turbine (CEST) and biomass integrated gasifier/steam-injected gas turbine (BIG/STIG) cogeneration systems have been estimated in previous papers [A.1-A.3]. These parameters are summarized in Tables A.1 and A.2. Here we present preliminary estimates of the cost and performance of biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG) cogeneration systems. Our starting point is a study by General Electric of coal integrated gasifier/intercooled steam injected gas turbine (CIG/ISTIG) systems [A.4].

COST ESTIMATES

The installed capital cost for a CIG/ISTIG system is given in the first column of Table A.3. From this we have estimated the cost of a BIG/ISTIG system assuming that BIG/ISTIG would be similar to a CIG/ISTIG system except that the raw gas chemical clean-up phase required for coal would not be needed for biomass, because of its lower sulfur content. The installed capital costs are therefore about 20% lower for BIG/ISTIG than for CIG/ISTIG.

Operation and maintenance costs for the BIG/ISTIG are based on estimates by General Electric [A.5].

PERFORMANCE ESTIMATES

The performance of a CIG/ISTIG system based on a GE LM-5000 turbine was estimated by General Electric [A.4]. (The flow diagram for this system operated in power only mode is shown in Figure A.1 and system parameters are listed in Table A.4.) Coal is gasified in a Lurgi type gasifier. After physical and chemical clean-up stages, the fuel gas is burned in the combustor of an intercooled steam injected gas turbine. Steam is generated from the heat of turbine exhaust in a heat recovery steam generator (HRSG) at two pressures. "High pressure" steam (770 psia, 740^{o}F) is injected into the combustor for cooling and power augmentation and used as blast steam in the coal gasifier. "Low pressure" steam (206 psia, 800^{o}F) is injected into the turbine for power enhancement and used in regeneration vessels for SO_2 recovery. The net electrical output of the system is 109.1 MW.

In Figure A.2 a similar system is shown biomass. Considerably less blast steam would be needed for biomass gasification than for coal gasification (about 0.0219 kg steam per MJ of biomass fuel as compared to 0.057 kg steam/MJ coal [A.3,A.4,A.6]). We assume that the same total amount of steam is raised in the HRSG with the CIG/ISTIG and BIG/ISTIG systems. However, in the BIG/ISTIG system less high pressure steam would go to the gasifier and more would be injected into the combustor. In the

CIG/ISTIG system 27.0 tonnes/hour of low pressure steam is injected into the turbine and 13.8 tonnes/hour is required by the regenerator for SO_2 recovery. With the BIG/ISTIG, there would be neglible SO_2 production and no regenerator stage, so that all the low pressure steam (40.8 tonnes/hour) could be injected into the turbine. The power output of the BIG/ISTIG would be increased by this extra 13.8 tonnes/hour of injected steam.

As a first approximation, we have assumed that the overall electrical output would be the same with the BIG/ISTIG as with the CIG/ISTIG system, <u>if the same amount of steam were flowing through the turbine in both cases.</u> (Clearly, this is an approximation which must be verified with more detailed modelling. However, it appears to be a reasonable assumption based on comparisons of CIG/STIG and BIG/STIG [A.3,A.6].) We then compute the extra power produced in the BIG/ISTIG system because of the extra 13.8 tonnes/hour of low pressure steam which can be injected into the turbine. In computing the extra power produced, we assume that an 88% efficient steam expansion takes place. Figure A.2 shows the estimated flows in a BIG/ISTIG system operated in "power only" mode, based on these assumptions.

For operation at a sugar factory or alcohol distillery during the milling season, process steam is needed to run the factory. To estimate the performance of the BIG/ISTIG system in "cogeneration" mode, we assume that some steam is diverted from the combustor and from the power turbine for process use. The reduction in output power is then computed. Assuming that 16.3 tonnes/hour of steam is required for cooling the combustor and that 17.9 tonnes/hour is needed for gasification of biomass, we find that 40.8 tonnes/hour of "low pressure" steam and 35.5 tonnes/hour of "high pressure" steam would be available for process (Figure A.3).

COUPLING THE BIG/ISTIG TO A SUGAR FACTORY OR ALCOHOL DISTILLERY

The maximum process steam available with the BIG/ISTIG is about 235 kg steam per tonne of cane. This would probably be sufficient to run a steam conserving sugar factory [A.7] or a low energy use autonomous distillery [A.8] without electrification of the factory. The electricity production with the BIG/ISTIG in power only mode is 111.2 MW or 339 kwh/tc; in cogeneration mode the power production is 97.4 MW or 300 kwh/tc. This is about 20% higher than the electricity produced with BIG/STIG.

REFERENCES - APPENDIX A

A.1. E.D. Larson, J.M. Ogden and R.H. Williams, "Steam-Injected Gas Turbine Cogeneration for the Cane Sugar Industry," Princeton University Center for Energy and Environmental Studies Report No. 217, September 1986.

A.2. E.D. Larson and R.H. Williams, "Biomass Fired Steam-Injected Gas Turbine Cogeneration," Proceedings of the 1988 ASME Cogen-Turbo Symposium, Montreaux, Switzerland, August 30-September 1, 1988.

A.3. E.D. Larson, R.H. Williams, J.M. Ogden and M.G. Hylton, "Biomass Gas Turbine Cogeneration for the Cane Sugar Industry," Proceedings of the 20th Congress of the International Society of Sugar Cane Technologists, Sao Paulo, Brazil, October 12-21, 1989.

A.4. 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.

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A.8. J.L. Oliverio, J.D. Neto and J.F.P. Miranda, "Energy Optimization and Electricity Production in Sugar Mills and Alcohol Distilleries," Proceedings of the 20th Congress of the International Society of Sugar Cane Technologists, Sao Paulo, Brazil, October 12-21, 1989.

Table A.1. Performance of Biomass-Fired Cogeneration Systems

	COGENERATION						POWER ONLY			
		ELEC	STEAM		FUEL	CANE[c]	ELEC		FUEL	CANE[c]
	(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	40.3	32.1	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	57.7	328	111.2	42.9	57.3	325

[a] Adapted from 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-8000 are based on private communications from M. Erbes, General Electric Corp. Other estaimtes are adapted from 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, assuming that the gasifier efficiency is the same for biomass as for coal.

[c] We have assumed 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. The CEST uses 50% wet bagasse having a higher heating value of 9530 kJ/kg. We further assume that 300 kg of 50% wet bagasse are produced per tonne of cane milled, or 176 kg of 15% wet briquetted bagasse are produced per tonne of cane milled.

[d] Preliminary estimate of steam and electricity production, based on performance with coal.

Table A.2. Capital and Operating Costs for Biomass-Fired Cogeneration Systems

	CAPACITY (MW)	INSTALLED COST ($/kW)	MAINTENANCE FIXED (1000$/Y)	MAINTENANCE VARIABLE ($/kWh)	LABOR (1000$/Y)
CEST [a]					
Generic	27.0	1556	664	0.003	129.6
Generic	10.0	2096	246	0.003	97.2
Generic	3.0	3008	73.8	0.003	97.2
BIG/STIG					
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					
LM-8000	111.2	770	2736	0.001	405.0

See Table 4 in text.

Table A.3. Installed Capital Cost for Integrated Gasifier/Intercooled Steam-Injected Gas Turbine Cogeneration Systems Fueled with Coal and Biomass (January 1986 $/kW)

	CIG/ISTIG [a]	BIG/ISTIG [b]
I. Process Capital Cost		
Fuel Handling	36.7	36.7
Blast Air System	9.6	9.6
Gasification Plant	83.1	83.1
Raw Gas Physical Clean-up	7.7	7.7
Raw Gas Chemical Clean-up	150.9	0.0
Gas turbine/HRSG	256.4	256.4
Balance of Plant		
Mechanical	22.0	22.0
Electrical	48.4	48.4
Civil	60.7	60.7
SUBTOTAL	686.5	535.6
II. Total Plant Cost		
Process Plant Cost	686.5	535.6
Engineering Home Office (10%)	68.6	53.6
Process Contingency (6.2%)	42.5	33.2
Project Contingency (17.4%)	119.6	93.2
SUBTOTAL	917.2	715.6
III. Total Plant Investment		
Total Plant Cost	917.2	715.2
AFDC (1.8%, 2 yr construction)	16.5	12.9
SUBTOTAL	933.7	728.5
IV. Total Capital Requirement		
Total Plant Investment	933.7	728.5
Preproduction Costs (2.8%)	26.2	20.4
Inventory Capital (2.8%)	26.2	20.4
Initial Chemicals, Catalysts	2.3	0.0
Land	1.3	1.3
TOTAL	990	770

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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 a BIG/ISTIG would be similar to a CIG/ISTIG system except that the raw gas chemical clean-up phase required for coal would not be needed for biomass, because of its lower sulfur content.

Table A.4 Performance of CIG/ISTIG and BIG/ISTIG systems

	CIG/ISTIG	BIG/ISTIG	
	Power Only[a]	Power Only[b]	Cogen[c]
Net Electrical Output (MW)	109.1	111.2	97.4
Efficiency (% of HHV)	42.1%	42.9%	37.9%
Steam production (tonnes/hour)			
High pressure steam (770 psia, 740^{o}F):			
Injected into combustor	16.3	51.8	16.3
Blast steam to gasifier	53.4	17.9	17.9
Available for process	0	0	35.5
Low pressure steam (206 psia, 800^{o}F):			
Injected into turbine	27.0	40.8	0
Regenerator for SO_2 recovery	13.8	-	-
Available for process	0	0	40.8
Fuel			
Type	Coal	Biomass @15% m.c.	
Higher heating value (kJ/kg)	25,011	16,166	
Steam for gasification (kg/MJfuel)	0.057	0.0219 [d]	
Gasification Efficiency	0.849	0.849	
Fuel energy input (MWt)	259	259	257
Consumption (tonnes fuel/hour)	37.25	57.7	57.3
(tonnes cane/hour)	-	328	325
Energy production for a sugar factory or alcohol distillery :			
Electricity (kwh/tc)		339	300
Process Steam (kg/tc)		0	235

[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] As a first approximation, we have assumed that the performance of the BIG/ISTIG system is the same as that of the CIG/ISTIG system in power only mode. Less steam is needed for gasification with BIG/ISTIG, but more steam is injected into the combustor, giving the same electricity production.

[c] We have calculated the power produced by the BIG/ISTIG in cogeneration mode, if some steam is diverted from the combustor and the power turbine for process use.

[d] Based on estimates by General Electric for a BIG/STIG system.

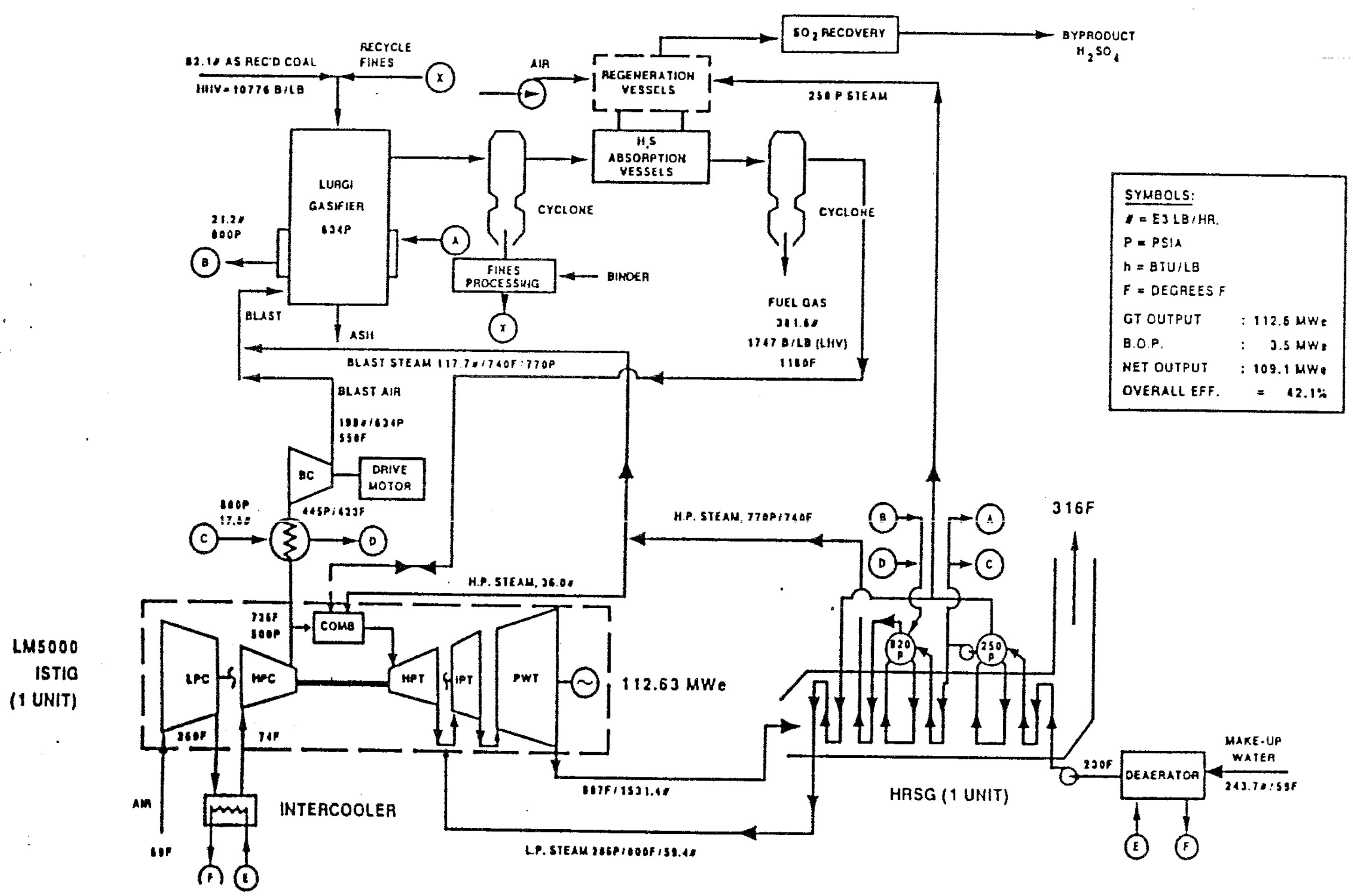

Figure A.1. Flow diagram for a coal intregrated gasifier/intercooled steam-injected gas turbine (CIG/ISTIG) cogeneration system operated in power only mode [A.4].

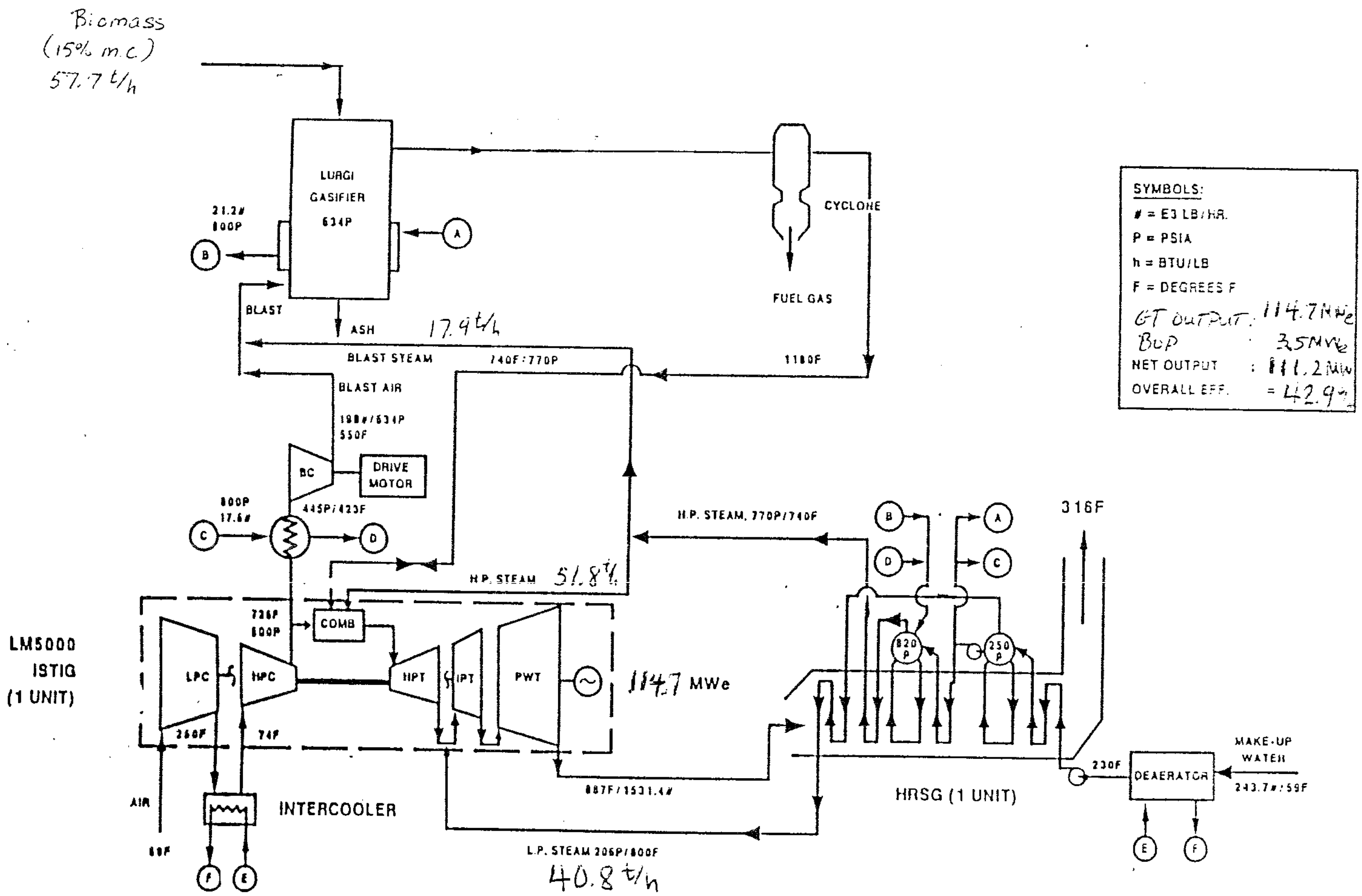

Figure A.2. Flow diagram for a biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG) cogeneration system operated in power only mode. Less high pressure steam is required for gasification than with CIG/ISTIG (Figure A.1) and more is injected into the combustor. Because biomass is a low sulfur fuel, neglible SO_2 is produced, so that no low pressure steam is needed for regeneration. The extra low pressure steam is injected into the turbine to produce additional power.

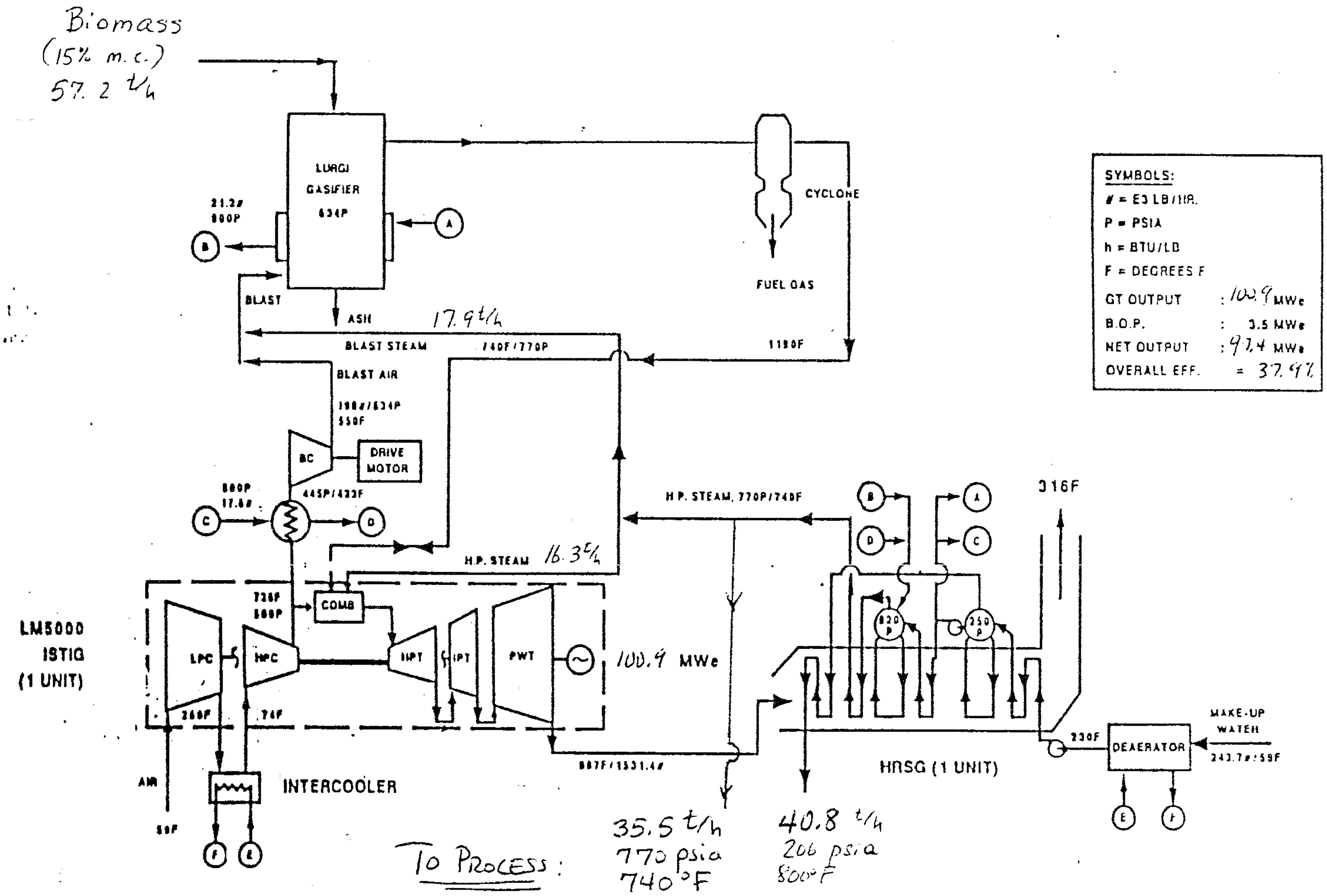

Figure A.3. Flow diagram for a biomass integrated gasifier/intercooled steam-injected gas turbine (BIG/ISTIG) cogeneration system operated in cogeneration mode. Here some steam is used for process rather than injected into the combustor and the power turbine, as in Figure A.2.

APPENDIX B - FACTORY MODEL

J. Ogden
S. Hochgreb

APPENDIX B.1 - STEAM CONSUMPTION IN CANE MILL TURBINES AND FACTORY TURBO-ALTERNATORS

We have based our estimates of cane mill and turbo-alternator steam consumption on results of a sugar factory modelling program by Blanchard applied to the small steam turbines at Monymusk [B1]. For various conditions this program gave the following results:

TABLE B.1.1

STEAM INLET		STEAM OUTLET		MILLS	TURBO-ALTERNATORS
Pin (MPa)	Tin (C)	Pout (MPa)	Tout (C)	Steam used (kg/tc)	Steam used (kg/kwh)
1.37	250	0.20	120	209	14.0
		0.25	127	233	15.9
1.9	270	0.20	120	177	11.9
2.5	360	0.20	120	129	10.2

We have compared these numbers with estimates of steam consumption in the cane mills and turbo-alternators derived from Hugot [B2]:

$$Qgen = \frac{3600}{(hin - hout)\ n * pm * pg * pr} \qquad Qmill = \frac{3600}{(hin - hout)\ n * pm * pmg}$$

where:

Qgen = steam consumption kg steam/kwh electricity out at generator terminals
Qmill = steam consumption of mill in kg steam/ kwh mill power
n = average turbine thermodynamic efficiency
= 0.7-0.72 (backpressure of 0.149-0.396 MPa)
= 0.65 for mill turbine (two stage)
hin = enthalpy of steam in (kJ/kg)
hout = enthalpy of steam out (kJ/kg)
pm = mechanical efficiency of turbine = 0.985
pg = efficiency of generator = 0.94-0.985
pr = efficiency of generator gearing = 0.97-0.985
pmg = efficiency of gearing for mill = 0.85

These formulas do not consider part load performance, which can effect the steam consumption significantly.

Assuming the following values for Monymusk:
(pm=0.985, pg=0.9625, pr=0.9775, n=0.70 for backpressure T/A's)
pmg=0.85, n=0.65 for mills)

STEAM INLET			STEAM OUTLET			TURBO ALT.	MILLS	
Pin (MPa)	Tin (C)	hin (kJ/kg)	Pout (MPa)	Tout (C)	hout (kJ/kg)	Qgen (kg/kwh)	Qmill (kg/kwh)	Steam used (kg/tc)
1.37	250	2930	0.20	120	2707	24.9	29.7	208-267
			0.31	135	2727	27.4	32.7	229-294
1.9	270	2953	0.20	120	2707	22.5	26.8	187-241
2.0	316	3060	0.20	120	2707	15.7	18.5	129-167
			0.31	135	2727	16.7	19.6	137-176
2.5	360	3149	0.20	120	2707	11.3	13.5	94-122

The actual power consumption in the mills is difficult to estimate a priori. We have calculated a range for the mill steam consumption, assuming that the average mill power consumption lies in the range 7-9 kwh/tc. Given the uncertainty in these numbers, we have used the estimates in Table B.1.1 for mill steam consumption where available. For the case where Pin = 1.37 MPa, Pout = 0.31 MPa, we have asssumed that mill steam consumption is 229 kg/tc.

The turbo-alternator steam consumption is also taken from the estimates in Table B.1.1.

APPENDIX B.2 - JUICE HEATER CALCULATIONS

Juice heater areas and inlet and outlet temperatures can be estimated if the heat transfer coefficients, approach temperature and heating fluid temperature and pressure are known.

From energy balance for a countercurrent heat exchanger:

$$m_{jin} * C_{vpj}(T_{jav}) * (T_{jout} - T_{jin}) = m_s * [h(T_{sin}) - h(T_{sout})]$$

where:

m_{jin} = mass flow of juice

C_{pj} = specific heat of juice

T_{jav} = ave. temp. of juice

T_{jin} = inlet temp of juice

m_s = mass flow of steam (or hot condensate)

T_{sin} = inlet temp. of steam (or hot condensate)

T_{sout} = outlet temp of steam (or hot condensate)

$h(T_{sin})$ = enthalpy of steam (or hot condensate) at inlet

$h(T_{sout})$ = enthalpy of steam (or hot condensate) at outlet

If the approach is known (the temp difference between the exiting steam and the inlet juice), then we can solve for T_{sout}

$$T_{sout} = T_{jin} + T_{appro}$$

and for T_{jout}

$$T_{jout} = T_{jin} + \frac{m_s * [h(T_{sout}) - h(T_{sin})]}{m_{jin} * C_{pj}(T_{jav})}$$

The heat exchanged is given by:

$$Q = U * A * LMTD = m_s * [h(T_{sout}) - h(T_{sin})]$$

where:

U = heat transfer coefficient

A = heat exchanger area

$$LMTD = \frac{(T_{sin}-T_{jout}) - (T_{sout}-T_{jin})}{\ln[(T_{sin}-T_{jout})/(T_{sout}-T_{jin})]}$$

solving for the heat exchanger area gives:

$$A = \frac{m_s * [h(T_{sout}) - h(T_{sin})]}{U * LMTD}$$

We have assumed the following values in our juice heater calculations:

HEAT TRANSFER COEFFICIENTS U (WATTS/^{o}C/m^{2})

TYPE OF JUICE HEATER HEAT EXCHANGER	SOURCE OF HEAT VAPOR OR STEAM	CONDENSATE
Plate-and-gasket	2500-4000 [a]	2000-4000 [a]
Shell-and-tube	600-1300 [b]	1000-2000 [a]

The range of values reflects the influence of fouling. We have used the lower value of the heat transfer coefficient in figuring heat exchanger areas.

APPROACH TEMPERATURE T_{appro} (^{o}C)

TYPE OF JUICE HEATER HEAT EXCHANGER	SOURCE OF HEAT VAPOR OR STEAM	CONDENSATE
Plate-and-gasket	6-10 [a]	6-10 [a]
Shell-and-tube	7,15,18,20 [c] Effect # 1 2 3 4	6-10 [a]

Sources:

[a] APV Crepaco, [B3].
[b] E. Hugot [B2].
[c] These are approximate approach temperatures asssumed for the shell-and-tube juice heaters at Monymusk, which utilize vapor from bled from effects 1-4 of the evaporator, based on Hugot's values [B2] and on factory simulations [B1].

APPENDIX B.3 - EVAPORATOR CALCULATIONS

Assuming no heat losses, the equations for mass and heat flow in the "i-th" effect of a forward feed evaporator with N effects are :

Mass conservation:

Sugar: $m_{j0} * X_0 = m_{ji} * X_i$, $i=1,N$

Water: $m_{ji-1} - m_{ji} = m_{vi} + m_{vbi}$, $i=1,N$

Energy balance:

$$m_{vi-1} * HFG(T_{vi}-1) = (m_{vi} + m_{vbi}) * HFG(T_{vi}) + m_{ji} * C_{pj} * (T_{ji}-T_{ji}-1)$$

{condensing vapor} = {evaporation from juice} + {juice heating}

Heat exchange:

$$Q_i = U_i * A_i * (T_{vi-1} - T_{ji}) = m_{vi-1} * HFG(T_{vi-1})$$

Boiling point rise

$$T_{ji} = T_{vi} + BPR$$

$$BPR = \frac{2.5*fX_i*(0.3+fX_i)}{(1.036 - fX_i)} * [1. - 0.54 * \frac{Pvac}{190.5 - Pvac}] \quad \text{[a]}$$

where: $Pvac = 76 - P_{vi} * 760$, P_{vi} in Mpa, $fX_i = X_i/100$.

Specific heat of juice

$$C_{pj}(X_i) = 1.0 - (0.6 - 0.0018 * T_{ji}) * X_i/100 \quad \text{[a]}$$

Barometric condensor

$$(m_w + m_{vN}) * HG(T_w + DT_c) = m_w * HG(T_w) + m_{vN} * HFG(T_{vN})$$

{condensed water} {input water} + {condensing vapor}

where:

m_{ji-1} = juice flow into ith effect

T_{ji-1} = temp of juice flowing into ith effect

X_{i-1} = Brix of juice flowing into ith effect

C_{pj} = specific heat of juice at a certain temperature and Brix

BPR = boiling point rise of juice at a given Brix and temperature

T_{vi} = vapor temperature in ith effect

P_{vi} = vapor pressure in ith effect

m_{vi-1} = mass flow of vapor into steam side of ith effect

m_{vbi} = vapor bled from ith effect to juice heaters or vac. pans.

U_i = heat transfer coefficient of ith effect

A_i = heat exchange area of ith effect

T_w = temperature of water into barometric condensor

m_w = mass flow of water into barometric condensor

DT_c = temp. diff. between condensor inlet water and outlet condensate

HFG = enthalpy of vaporization

HG = enthalpy of water

For Monymusk, we have assumed the following values for the heating surface areas [B1]:

Effect # i	Area (m^2) A_i
1	1998
2	1129
3	844
4	836

[a] E. Hugot [B2].

APPENDIX B.4 - VACUUM PAN CALCULATIONS

The total vapor to be evaporated in the vacuum pan is:

$$m_{vtot} = m_{jN} *(1 - X_N/X_f) = m_{j0} * X_0/X_N * (1 - X_N/X_f)$$

where:

m_{jN} = mass flow of juice out of last effect of evaporator

X_N = Brix out of last effect (= 60-70)

X_f = final Brix of massecuite (= 94-96)

m_{j0} = juice flow into first effect of evaporator

X_0 = Brix of juice " " " " "

Then assuming that the amount of juice is about equal to the amount of cane ground, Hugot quotes steam consumption of about

$$m_s = (1.2\text{-}1.7) * (X_0/X_N - X_0/X_f)/(1 - V_{Ploss}) \text{ kg steam/tonne cane}$$

where:

V_{Ploss} = heat loss in vacuum pans = 10-20%.

Assuming V_{Ploss}= 20%, X_0=13, X_N=65 and X_f=96 brix, then the steam consumption is about

$$m_s = 100\text{-}140 \text{ kg/tc},$$

depending on the design of the vacuum pan. Measured steam consumption in vacuum pans is quoted by Hugot as 120-165 kg/tc. The value measured for Monymusk is 137 kg/tc [B1].

We have assumed as in Hugot that a continuous pan would use about 25% less steam than a discontinous pan. We assume a value of 103 kg/tc.

For sizing continous pans, we have used the following table from Hugot [B2].

SERIES OF CONTINUOUS PANS (FIVES CAIL – BABCOCK)

Cap.	*Diam.*	*Length*	*Diam.*	*Length*	*Massecuite A*		*Massecuite B*		*Massecuite C*	
V (m^3)	*D* *(m)*	*L* *(m)*	*D* *(m)*	*L* *(m)*	*v* (m^3/h)	*A* *(t.c.h.)*	*v* (m^3/h)	*A* *(t.c.h.)*	*v* (m^3/h)	*A* *(t.c.h.)*
16	3.1	3.715			8.1	60	6.4	85	3.2	70
19	3.1	4.455			9.5	70	7.5	100	3.8	85
22.5	3.1	5.135			11.5	85	9	120	4.5	100
27	3.1	5.935	3.5	5.195	13.5	100	10.9	145	5.4	120
32	3.1	7.415	3.5	5.935	16.2	120	12.8	170	6.3	140
38	3.1	8.900	3.5	7.415	18.9	140	15	200	7.6	170
45	3.5	8.900	4.0	5.935	22.3	165	18	240	9	200
54	3.5	10.385	4.0	7.415	27	200	21.8	290	10.8	240
64	3.5	11.870	4.0	8.900	32.4	240	25.5	340	12.6	280
76	4.0	10.385	4.6	7.415	37.8	280	30.4	405	15.3	340
90	4.0	11.870	4.6	8.900	42.6	330	36	480	18	400
104	4.6	10.385			51.3	380	41.6	555	20.7	460
120	4.6	11.870			59.4	442	48	640	23.8	530

BOX B.5 - MECHANICAL VAPOR RECOMPRESSION CALCULATIONS

The power required for mechanical vapor recompression is given by:

$$E = mv*Cp*T1*[(P2/P1)^{((g-1)/g)} - 1]/(nc*ne)$$

where:

- E = power in kW of compressor
- T1 = compressor inlet temperature (K)
- P1 = compressor inlet pressure
- P2 = compressor outlet/evaporator hot side inlet pressure
- mv = vapor flow into compressor (kg/s)
- Cp = constant pressure specific heat of the vapor (kJ/^{o}C/kg)
- g = ratio of specific heats of steam Cp/Cv
- nc = compressor efficiency
- ne = electrical efficiency of generator

For steam at about 120 degrees C,

$$E = mv * 1.89 * T1 * [(P2/P1)^{0.248} - 1] \quad kW$$

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B3. APV Crepaco, "Heat Transfer Handbook," 1986.

APPENDIX ~~II~~ C: AUTONOMOUS DISTILLERY CALCULATION METHOD

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):

$$TVC = VC \times EP$$

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, 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, $.

APPENDIX ~~III~~ D: ANNEXED DISTILLERY CALCULATION METHOD

ANNEXXED 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 x 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 x ELTC x 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 x STC - DSTM) x STE x 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 x 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 x BPC x 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 x MY x 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 mollases.
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.

III. 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 = 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 = 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 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 F: AUTONOMOUS DISTILLERY SENSITIVITY STUDIES

The plots found is this appendix show the ethanol and electricity production cost sensitivities the the following parameters:

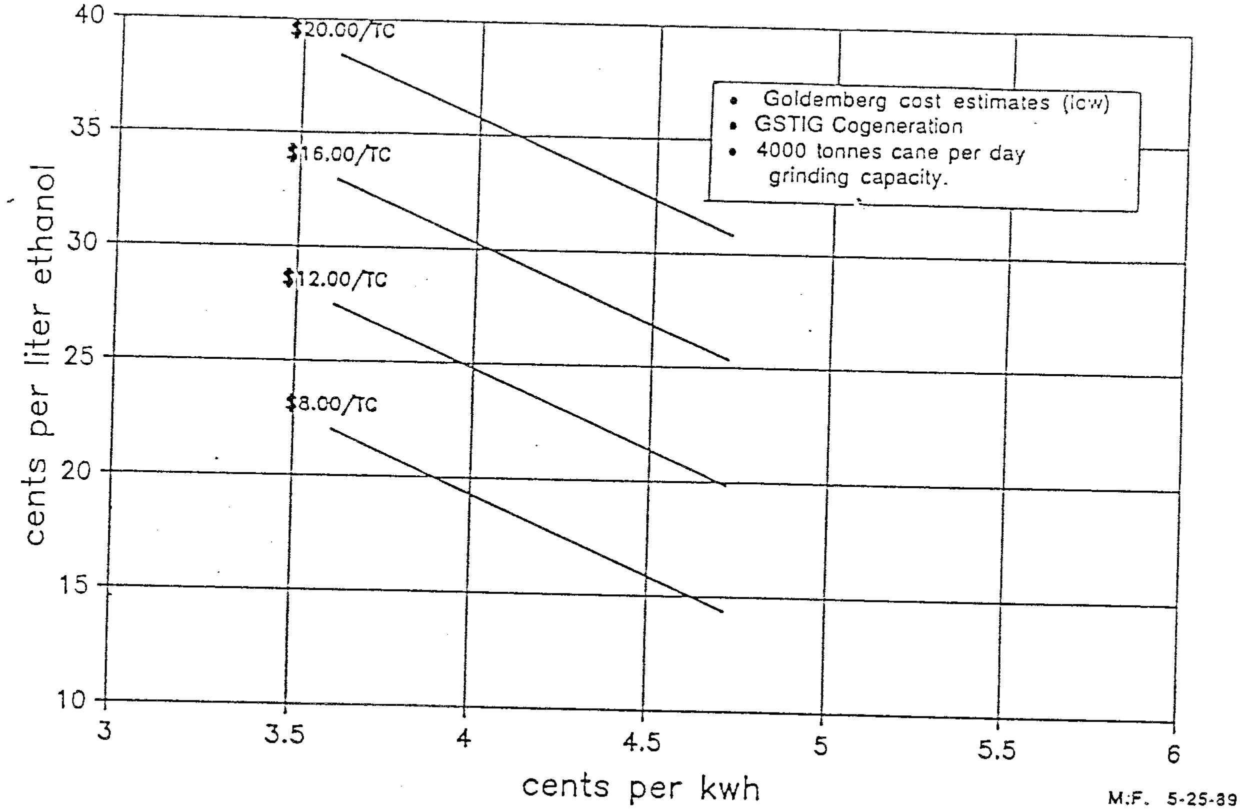
1. Cane price
2. Distillery capital costs
3. Ethanol yield, liters ethanol per tonne cane
4. Milling season length
5. Off season fuel costs
6. Bagasse processing costs.

Each plot encompasses the range which could reasonably expect from the particular parameter being examined. Often the upper and lower limits were set by the Goldemberg and Schaffer limits.

All the sensitivity studies used low (Goldemberg) cost estimates, except, of cerci for the studied parameter. GSTIG cogeneration was also used in all the studies. Sensitivities were also conducted using the other cogeneration technologies, with no significant difference in the general results or trends.

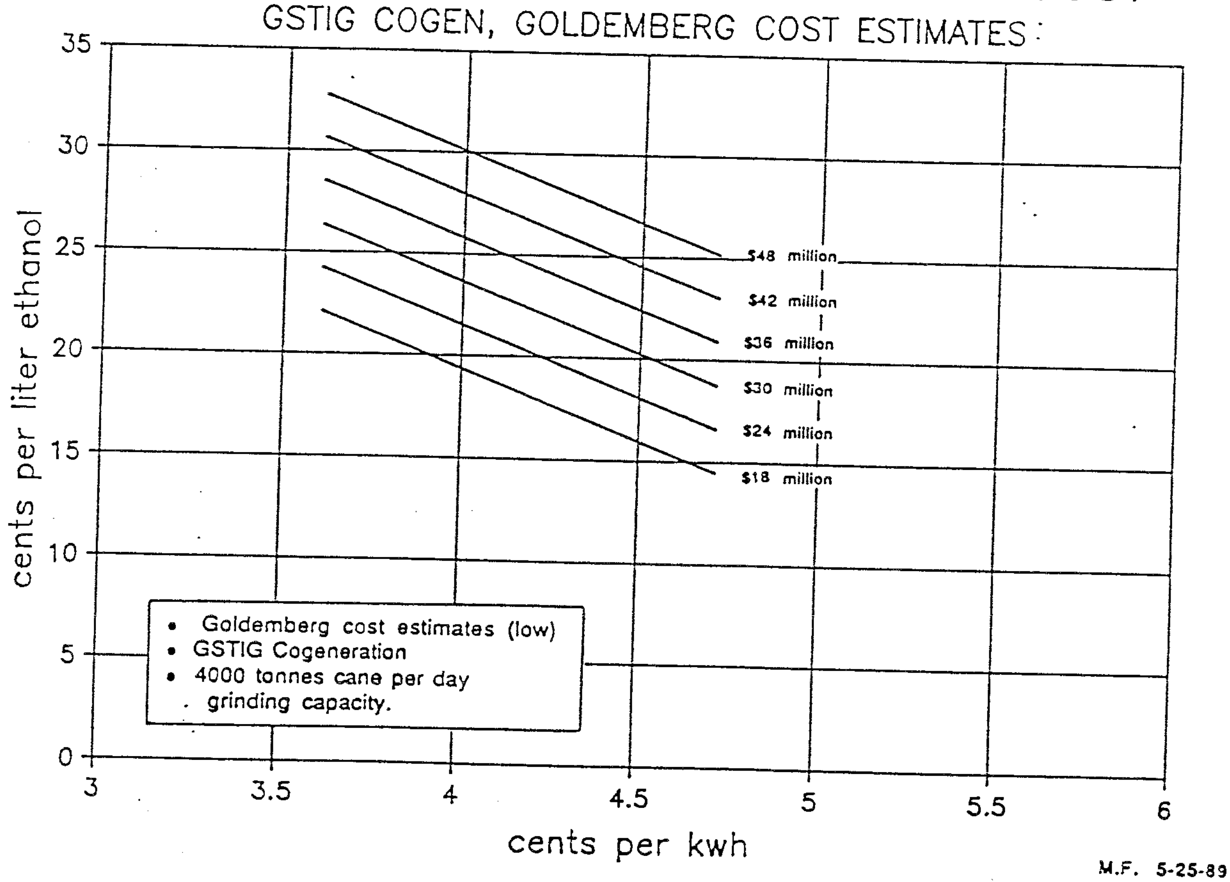
AUTONOMOUS DISTILLERY

SENSITIVITY TO CANE COSTS

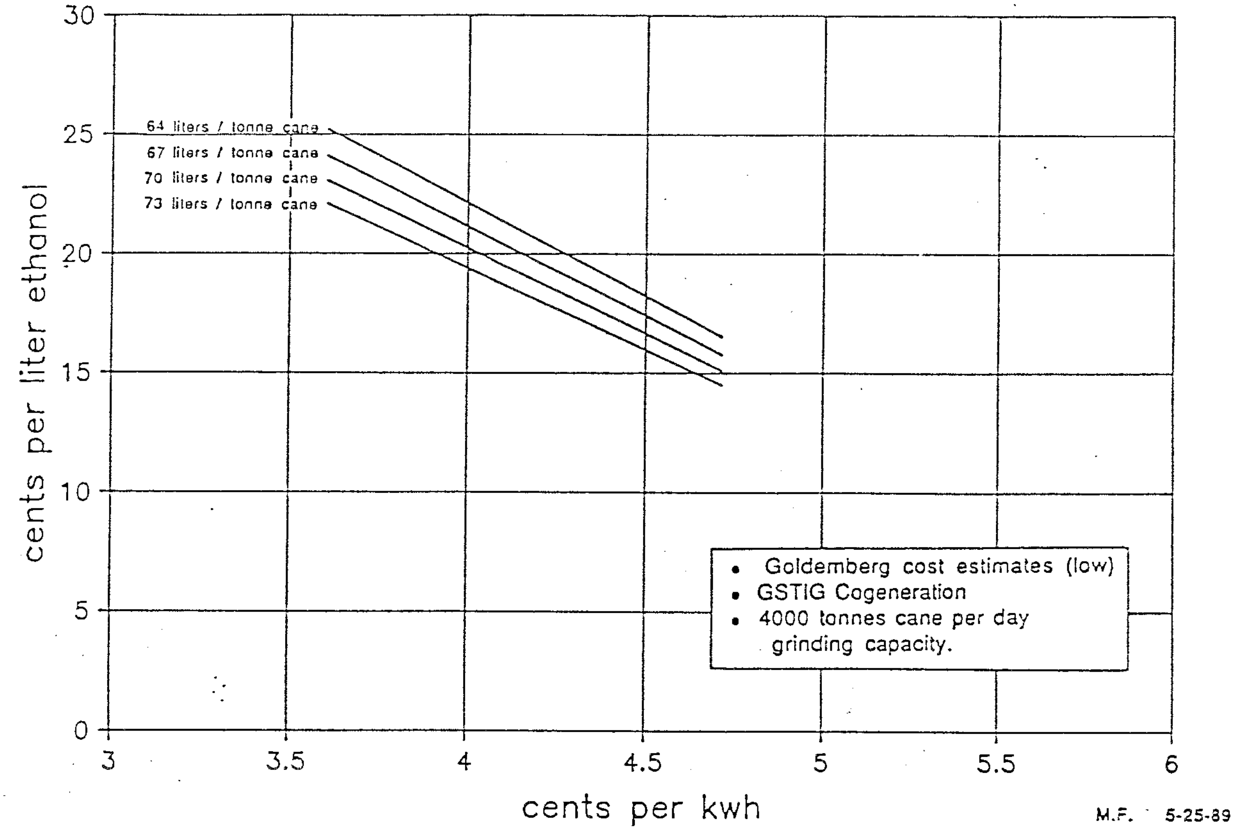


SENSITIVITY TO DISTILLARY CAPITAL COST

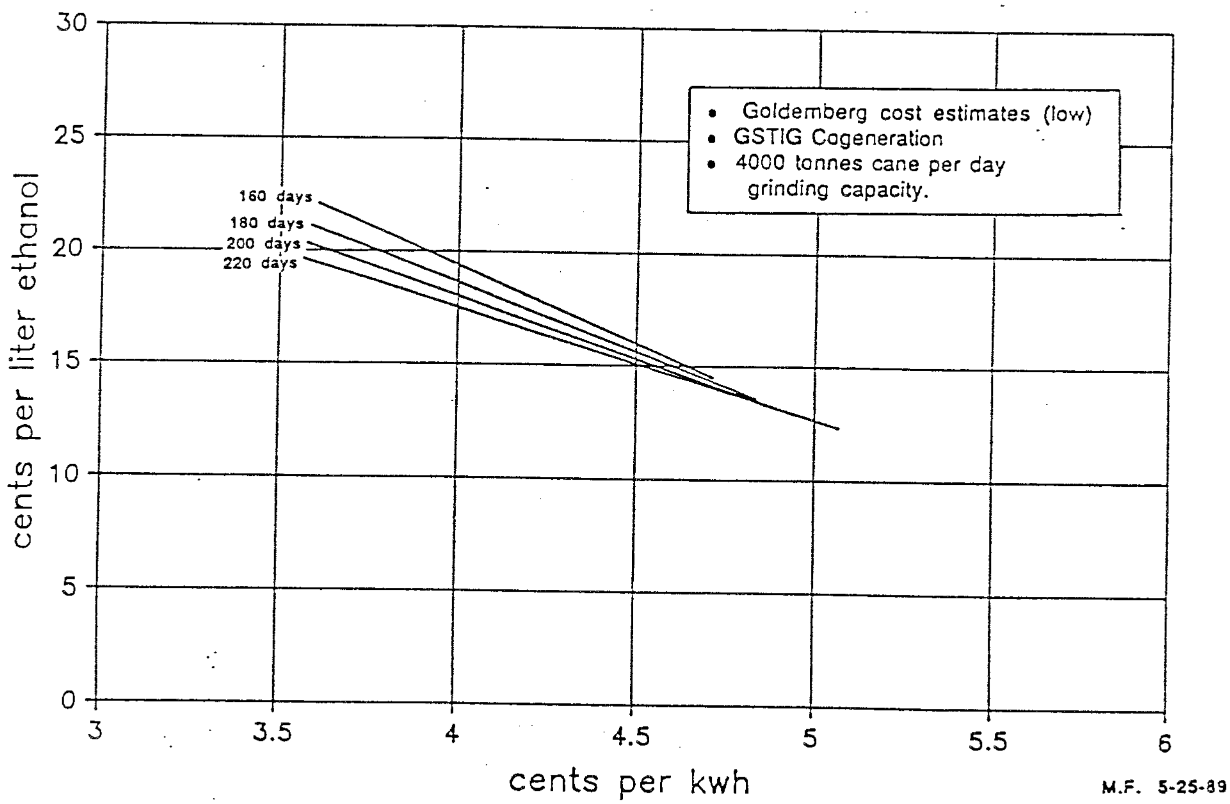
GSTIG COGEN, GOLDEMBERG COST ESTIMATES

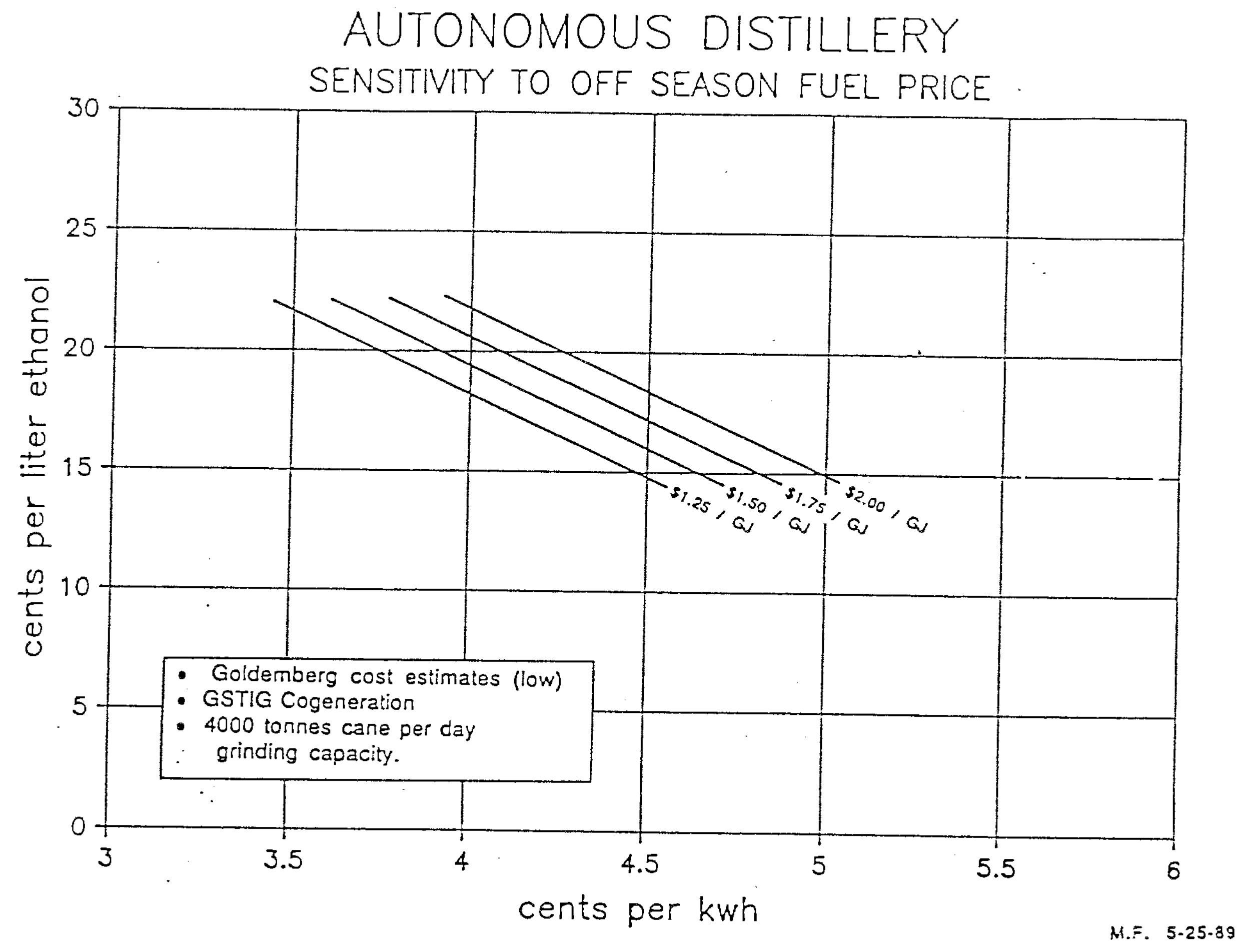


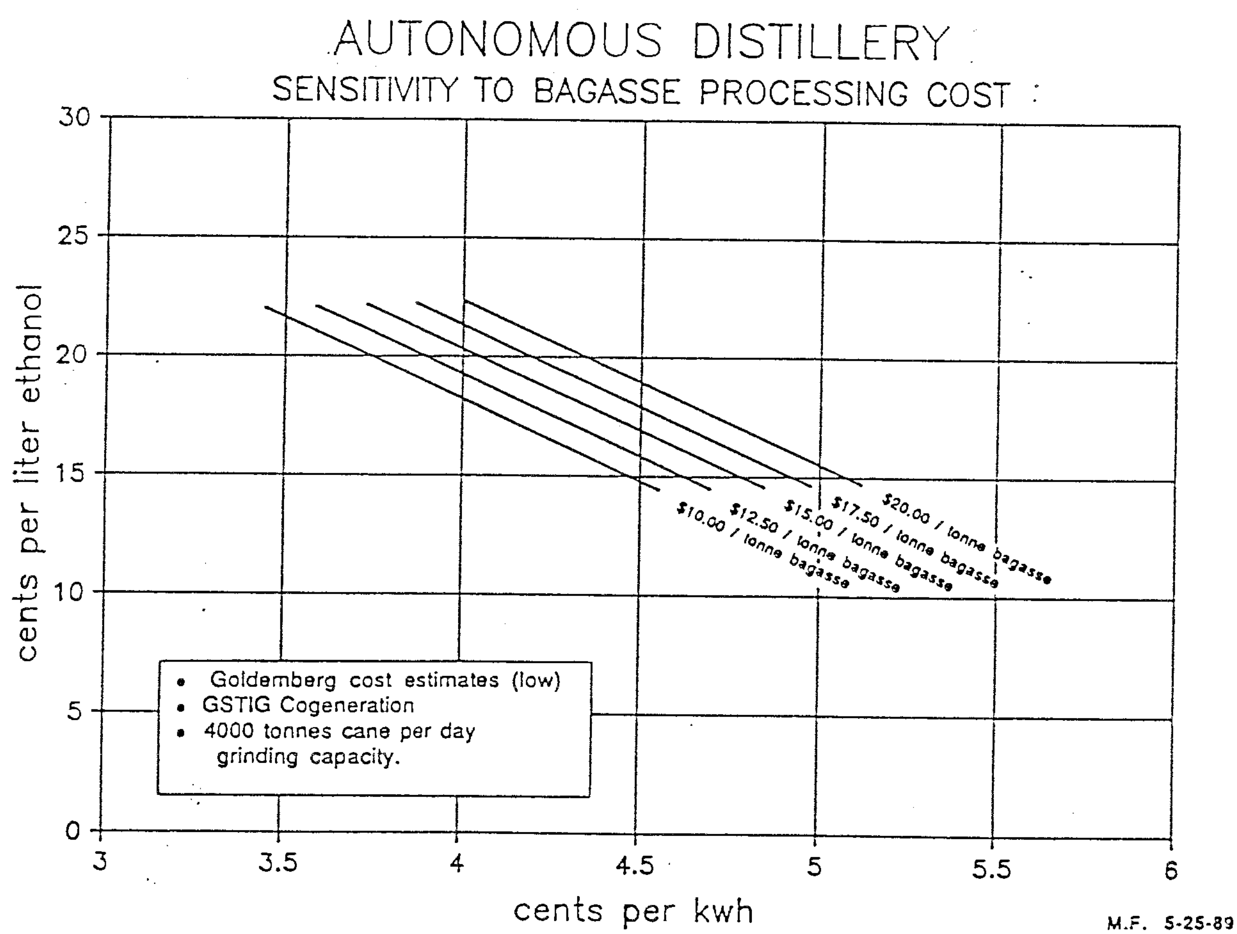
SENSITIVITY TO ETHANOL YIELD/TONNE CANE
AUTONOMOUS DISTILLERY



AUTONOMOUS DISTILLERY
SENSITIVITY TO MILLING SEASON LENGTH




AUTONOMOUS DISTILLERY
SENSITIVITY TO OFF SEASON FUEL PRICE
cents per liter ethanol
cents per kwh
$1.25 / GJ
$1.50 / GJ
$1.75 / GJ
$2.00 / GJ
• Goldemberg cost estimates (low)
• GSTIG Cogeneration
• 4000 tonnes cane per day
grinding capacity.
M.F. 5-25-89


AUTONOMOUS DISTILLERY
SENSITIVITY TO BAGASSE PROCESSING COST
cents per liter ethanol
cents per kwh
$10.00 / tonne bagasse
$12.50 / tonne bagasse
$15.00 / tonne bagasse
$17.50 / tonne bagasse
$20.00 / tonne bagasse
• Goldemberg cost estimates (low)
• GSTIG Cogeneration
• 4000 tonnes cane per day
grinding capacity.
M.F. 5-25-89

APPENDIX F. SUGAR AND ALCOHOL PRODUCTION IN ANNEXED DISTILLERIES.

Joan Ogden

In a sugar factory with an annexed distillery both sugar and ethanol are produced. In many factories sugar making proceeds as usual and the final or C-molasses (from which most of the sucrose has been exhausted) is then diluted and used a feedstock for ethanol production. Alternatively, the factory could be configured to make less sugar and more ethanol.

In a typical cane sugar factory, the cane juice is heated and then concentrated in a multiple effect evaporator from an initial concentration of 13-15^{o}Brix (13-15% solids in juice) to 60-65^{o}Brix (60-65% solids in juice). The syrup from the last effect of the evaporator then goes to a series of vacuum pans where the final stage of the evaporation (sugar boiling) takes place and sugar crystals are formed.

A number of different schemes are used for sugar boiling [F1, F2]. Here we a consider a factory with the commonly used three vacuum pan boiling scheme. Sugar is made in three stages or "strikes" as shown in Figure F.1. Syrup from the last effect of the evaporator is introduced into the A-pan along with some seed sugar crystals. The mixture in the A-pan is termed the A-massecuite. The massecuite is boiled so that evaporation continues, and sucrose crystals begin to form. Not all the sucrose can be crystallized in one stage or strike. When the maximum amount of sucrose is crystallized in the A-pan, the remaining material or A-molasses is introduced into a second B-vacuum pan along with seed crystals. The sugar boiling process is then continued in the B-pan. B-sugar is crystallized, and B-molasses is sent to the final C-vacuum pan. (Some syrup from the evaporator is also introduced into the C-pan to increase the sucrose content of the C-massecuite.) In the C-pan, C-sugar and C-molasses are produced. The C-sugar is lower quality than A- and B-sugars and is typically used as seed crystal for the A and B pans, rather than being sold. The C-molasses is used for ethanol production.

In a factory producing ethanol from A-molasses, only one vacuum pan would be used, and only one strike of sugar would be produced. The A-molasses would then be the feedstock for ethanol production.

The amount of ethanol which can be produced from A or C-molasses depends on the amount of fermentable sugars (sucrose plus invert sugars) contained in the molasses. C-molasses contains much less fermentable sugar than A-molasses, so that less ethanol is produced.

During fermentation ethanol is produced from fermentable sugars via the following reactions:

$C_{12}H_{22}O_{11}$ + H_2O + invertase -> $C_6H_{12}O_6$ + $C_6H_{12}O_6$

(sucrose) (enzyme) (invert sugars)

$C_6H_{12}O_6$ -> 2 C_2H_5OH + 2 CO_2 + heat (31,200 cal)

(invert sugar) (ethanol)

From mass balance considerations, 0.95 kg sucrose plus 0.05 kg water yields 1 kg invert sugar, and 1 kg invert sugar yields 0.511 kg or 0.646 liters of ethanol. In practice only about 90% of the invert sugars are converted to ethanol during fermentation, so that 1 kg invert sugar yields about 0.646 x 0.90 = 0.58 liters of ethanol.

Typically sugar cane juice contains about 120-130 kg fermentable sugars per tonne of juice. Assuming that one tonne of juice is produced per tonne of cane, about 70-75 liters of ethanol would be produced per tonne of cane.

The equations for sugar production in a three vacuum pan boiling system are given below. Table F.1 summarizes the results used in our base case for ethanol production from C-molasses and from A-molasses.

CALCULATION OF SUGAR, MOLASSES AND ETHANOL PRODUCTION WITH THREE PAN BOILING SCHEME

DEFINTIONS:

Total solids = Sucrose plus other solids contained in cane juice
Purity = mass % sucrose in total solids
Brix = mass % total solids in juice

VARIABLES:

msyr = syrup mass flow from last effect of evaporator
Psyr = syrup purity " " " " "
Bsyr = syrup Brix " " " " "

msyrA = syrup mass flow into pan A
mseedA = seed crystal mass flow into pan A

mA = massecuite mass flow, pan A
PA = " purity "
BA = " Brix "

msugA = A-sugar mass flow, pan A
PsugA = " purity "

mmolA = A-molasses mass flow, pan A
PmolA = " purity, "
BmolA = " Brix, "

BfA = Brix of A-massecuite at end of strike
mvapA = water evaporated from pan A

and similarly for pans B and C

EQUATIONS FOR PAN A:

Mass conservation into Pan A:

Total: mA = msyrA + mseedA
Solids: mA x BA = msyrA x BsyrA + mseedA
Sucrose: mA x BA x PA = msyrA x BsyrA x PsyrA + mseedA x PsugA

Mass conservation out of Pan A:

Total: mA = msugA + mmolA + mvapA
Solids: mA x BA = msugA + mmolA x BmolA
Sucrose: mA x BA x PA = msugA x PsugA + mmolA x BmolA x PmolA

Hugot limit on Brix on A-massecuite at end of strike BfA = 0.96

Mass conservation of solids at end of strike in Pan A:

mA x BA = (msugA + mmolA) x BfA

Hugot Rule of Thumb:

$$PmolA = PA - 20$$
$$PmolB = PB - 22$$
$$PmolC = PC - 25$$

Know: Bsyr, Psyr, PsugC, BA, mA, PsugA, BfA,
Solve For: Other variables

SIMILAR EQUATIONS CAN BE WRITTEN FOR PANS B AND C, AND THE REMAINING FLOWS CAN BE CALCULATED.

ETHANOL PRODUCTION:

Assume sugars in cane of:

110 kg sucrose/tc
6 kg invert sugars/tc

---

116 kg total fermentable sugars/tc

Hydrolysis yield:

0.95 kg sucrose -> 1.0 kg invert sugars

110 kg sucrose/tc -> 115.7 kg invert sugars/tc (hydrolysis)

Total fermentable sugars after hydrolysis:

115.7 + 6 = 121.7 kg invert sugars/tc

Theoretical fermentation yield:

1 kg invert sugar -> 0.511 kg ethanol = 0.646 liters

Assume 90% of theoretical fermentation yield of ethanol achieved

121.7 kg/tc x 0.646 l/kg x 0.90 = 70 liters/tc ethanol produced

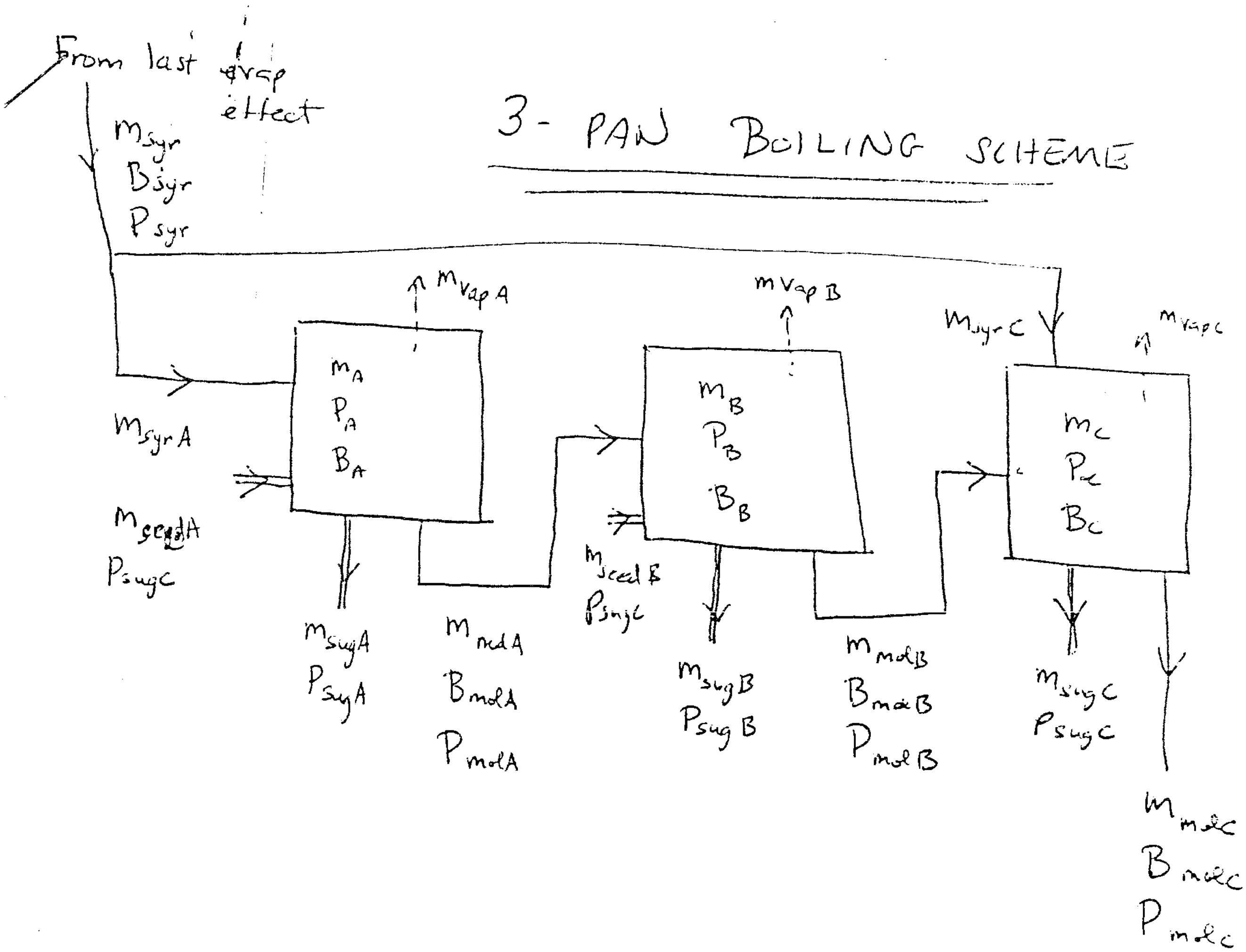


Figure F.1. Vacuum pan mass flows for a sugar factory with a three pan boiling scheme.

Table F.1. Production of sugar, molasses and ethanol with a three vacuum pan boiling scheme

SUGAR PRODUCED:

	KG/TC	PURITY
A-SUGAR	82	98.5%
B-SUGAR	26	97.5%
C-SUGAR	24	88% (USED FOR SEED CRYSTALS IN PANS A + B)

MOLASSES:

	MASS FLOW (KG/TC)	PURITY	BRIX	SUCROSE (KG/TC)	TOTAL FERMENTABLE SUGARS (KG/TC)	ETHANOL PRODUCTION (LITER/TC)
A-MOLASSES	55	.65	.975	35	41	24.7
B-MOLASSES	35	.46	.983	16	22	13.1
C-MOLASSES	28	.35	.981	9.5	15	9.2

SUMMARY OF PRODUCTS FOR VARIOUS ETHANOL PRODUCTION STRATEGIES:

FEEDSTOCK FOR ETHANOL PRODUCTION	A + B SUGARS (KG/TC)	ETHANOL (LITERS/TC)
CANE JUICE	0	70
A-MOLASSES	74	25
B-MOLASSES	96	13
C-MOLASSES	108	9.2

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