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STEAM ECONOMY AND COGENERATION IN CANE SUGAR FACTORIES*

by

Joan M. Ogden
Simone Hochgreb
Center for Energy and Environmental Studies
Princeton University, Princeton, New Jersey, U.S.A.

Michael Hylton
Sugar Industry Research Institute
Factory Technology Division
Kingston, Jamaica, W.I.

ABSTRACT

With the recent trend toward diversification in the cane sugar industry, a growing number of factories are producing electricity for export to the utility grid, in addition to sugar and molasses. In this paper, we discuss energy efficiency improvements as a way of increasing electricity production in a raw cane sugar factory with a cogeneration system.

We have considered two types of advanced bagasse-fired cogeneration systems, (1) high pressure condensing-extraction steam turbine systems of the type used in some factories in Hawaii and Reunion and (2) steam-injected gas turbines run on gasified bagasse (these systems, which could be commercialized within a few years, could produce about twice as much export electricity as a high pressure condensing-extraction steam turbine, but would require some steam conservation measures in the factory).

We have written a computer program to calculate factory balances for several steam-conserving designs incorporating commercially available process equipment: waste heat recovery heat exchangers which utilize hot condensate for juice heating, falling film evaporators, and continuous vacuum pans. Our results indicate that the process steam use could be reduced to less than 300 kg per tonne of cane milled, boosting the electrical output of the steam turbine cogeneration system by up to 20% and making the highly efficient gas turbine systems a future option for the cane sugar industry.

INTRODUCTION

Most cane sugar factories have been designed to be energy self-sufficient, with sugar as the primary product. A bagasse-fired cogeneration system, made up of "medium pressure boilers (1.5-2.0 MPa) plus small steam driven turbo-alternators, provides all the steam and electricity needed to run the cane mills and factory, leaving a little surplus bagasse. With sugar as the main product and bagasse as a "free" fuel, there has been little economic incentive to save bagasse via factory energy efficiency improvements. In fact, bagasse fired boilers have been designed to be somewhat inefficient, so that excess bagasse does not accumulate and become a disposal problem.

With the recent trend toward diversification in the cane sugar industry, a growing number of factories are manufacturing one or more byproducts (such as cogenerated electricity for export to the utility grid or alcohol) in addition to sugar and molasses. In a factory with several products, each of which requires a certain amount of energy (or bagasse) to manufacture, energy efficiency (both in the conversion of bagasse to useful energy and in the utilization of energy within the factory) can become much more important.

In this paper, we discuss some implications of energy efficiency improvements for a raw cane sugar factory with a cogeneration system. We have considered two types of bagasse fired cogeneration systems, which potentially offer much higher electricity production than those found in most sugar factories today, 1) high pressure condensing-extraction steam turbine systems and 2) steam injected gas turbines run on gasified bagasse.

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High pressure[1] (4.0-8.0MPa) condensing extraction steam turbine (CEST) cogeneration systems are now used in a few cane sugar factories (1,2) and are being considered for several others (3). As has been demonstrated in Hawaii and Reunion, when small medium pressure turbo-alternators are replaced with a high pressure CEST system, the total electricity production can be increased from about 20 kwh/tc (just enough to run the factory) to perhaps 70-120 kwh/tc. Thus, in addition to making sugar, about 50-100 kwh/tc becomes available for export to the utility grid. With a CEST cogeneration system, there should be an incentive to improve factory steam economy: any fuel (or equivalently steam) saved in the sugar process would become available for generating additional export electricity (2).

In a gasifier/steam injected gas turbine (GSTIG) system (Figure 2d), bagasse would be gasified to form a low BTU gas, which fuels a gas turbine (4,5,8). Steam would be raised for the mills and the process in a heat recovery steam generator (HRSG), which utilizes the hot exhaust gases exiting the turbine. Any steam not needed for the factory could be injected into the combustor to boost the electrical output of the system. As with the CEST system, the lower the factory steam demand, the higher the electrical output. While GSTIG systems are not commercially available at present, they could be developed within the next several years[2]. GSTIG systems are of interest because they could potentially produce up to 200 kwh/tc of export electricity, about twice as much as high pressure steam turbine systems (5,8). However, GSTIG systems could not provide quite enough process steam to supply the average cane sugar factory. Thus, some factory steam economy measures would be desirable when using these systems.

Our motivations for studying factory steam economy in raw cane sugar factories with cogeneration are twofold: to boost the export electricity production from a particular type of cogeneration system, and to widen future cogeneration options for the cane sugar industry to include the more efficient GSTIG systems. With these goals in mind, we have assessed several steam-conserving retrofits incorporating commercially available process equipment: waste heat recovery heat exchangers which utilize hot condensate for juice heating, falling film evaporators, and continuous vacuum pans. In the 1970's these technologies were widely adopted in oil dependent process industries with large evaporation energy requirements (such as the beet sugar, pulp and paper, and dairy industries) to reduce fuel costs (9). With the emphasis on byproducts and process steam economy, they are beginning to appear in the cane sugar industry as well (10,11).

Although we have focussed on cogeneration, the energy efficiency improvements discussed may also be of interest to factories with other byproducts, which require energy (or bagasse) for their manufacture.

INCREASED COGENERATION OUTPUT THROUGH IMPROVED FACTORY STEAM ECONOMY

A. Electricity and steam production in cogeneration systems

The electricity (in kwh/tc) and steam production (in kg of medium pressure steam produced per tonne of cane) are shown in Figure 1 for a high pressure condensing-extraction steam turbine (CEST) cogeneration system and for three gasifier/steam injected gas turbine (GSTIG) systems of various sizes (5,8). Steam and electricity demands characteristic of most raw cane sugar factories today (12) are shown as ranges of values along the x and y axes of the graph. We have also plotted the steam and electricity production in a typical medium pressure steam driven turbo-alternator (MPTA) sugar factory cogeneration system.

For both CEST and GSTIG cogeneration systems, the steam and electricity production can be varied over a range of operating conditions, so that more electricity can be produced when the steam demand is lower. The right endpoint of each range indicates the maximum amount of process steam which could be produced with the particular

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1 In this paper "high pressure" (4.0-8.0 MPa) refers to boiler pressures typical of condensing-extraction steam turbines. "Medium pressure" refers to steam used for cane mills, which equals the boiler pressure in most sugar factories today (1.5-2.0 MPa). "Low pressure" refers to mill and turbo-alternator exhaust steam used in the process (0.2-0.3 MPa).

2 Similar systems are being developed for use with coal in the US (6). Piggy-backed onto this development, biomass-fired systems could be commercialized within about three years(7).

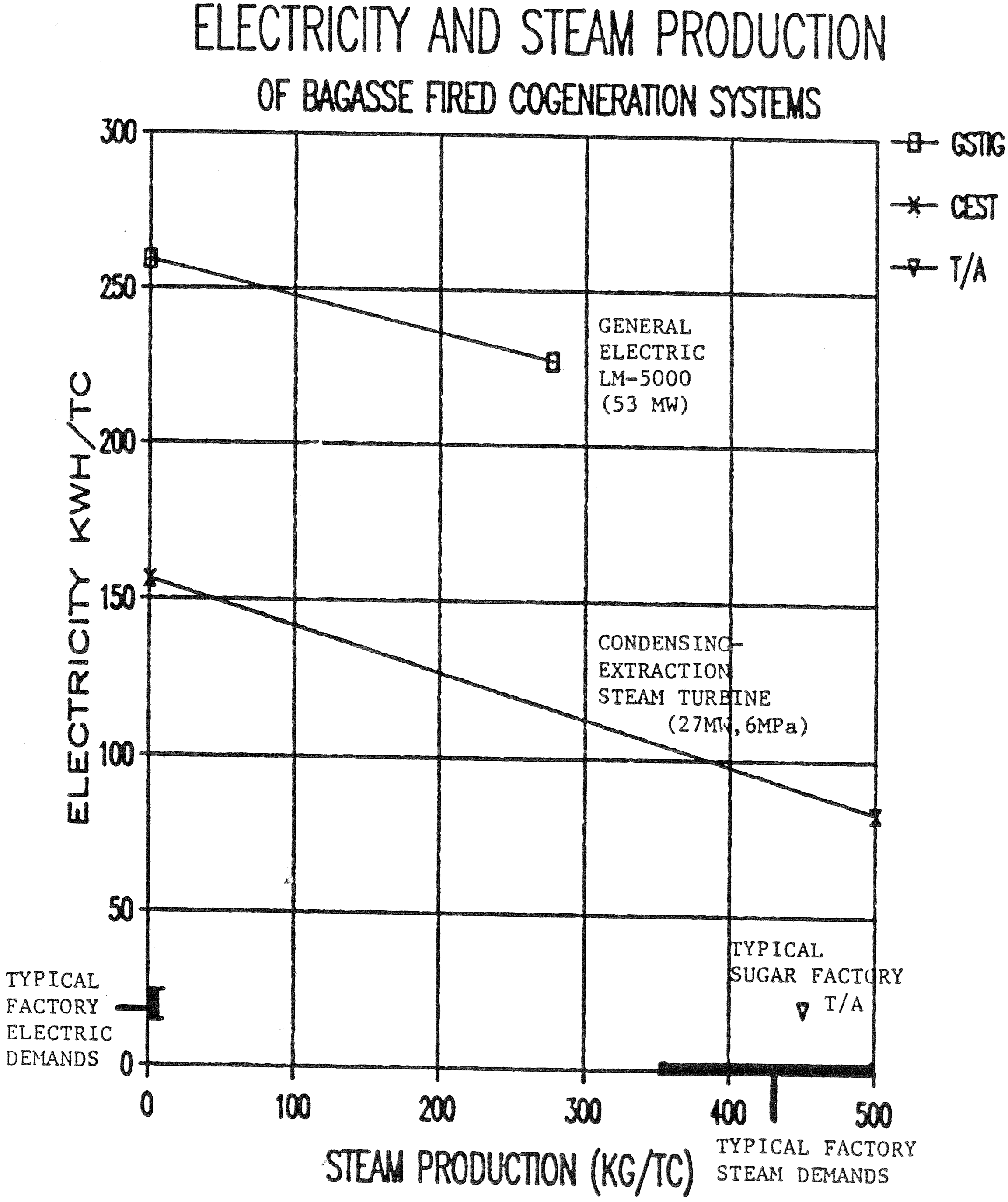

Figure 1. Steam and electricity production estimates for cogeneration systems operating at cane sugar factories during the milling season with bagasse as fuel [5].

cogeneration system; the left endpoint represents the maximum electricity production, which would occur when the process steam production is zero.[3]

Comparing typical factory demands to the output of the cogeneration systems, we see that the CEST can easily meet the steam demand of the average raw sugar factory (350-500 kg/tc), while producing about 100 kwh/tc, roughly five times as much electricity as a small turbo-alternator system in a typical factory. While the GSTIG produces about 200 kwh/tc or twice as much electricity as the CEST system, the maximum steam production possible with the GSTIG systems is only about 270-300 kg/tc. The GSTIG system would not be able to supply all the factory steam needs without some factory steam economy measures.

B. Integrating a cogeneration system with a cane sugar factory

Examples of how a raw sugar factory could be integrated with a cogeneration system are shown in Figure 2 for three cases, a conventional factory with small medium pressure back-pressure and condensing turbo-alternators, a conventional factory with a CEST system, and a hypothetical steam conserving factory with a GSTIG system.

1. Electricity and steam supply and demand in a conventional raw sugar factory (Figure 2a)

In most raw cane sugar factories, steam is raised at 1.5-2.0 MPa in a medium pressure boiler. About 200-250 kg/tc of medium pressure steam is used to drive small backpressure mill turbines, which grind the cane; an additional 150-250 kg/tc goes to run one or more small backpressure or condensing turbo-alternators, which produce just enough electricity for the factory (about 15-25 kwh/tc), but none for export. The 350-500 kg/tc of low pressure exhaust steam (saturated steam at 0.2-0.3 MPa) from the mill turbines and turbo-alternators is then utilized for process heat (e.g. juice heating, evaporation and crystallization of sugar).

2. CEST system with a conventional sugar factory (Figures 2b and 2c)

In a CEST cogeneration system, steam is raised in a high pressure boiler at 4.0-8.0 MPa, and passes through a condensing extraction steam turbine. About 200-250 kg/tc of medium pressure steam is extracted from the turbine at 1.5-2.0 MPa for use in the mill turbines. The additional low pressure steam needed for process (150-250 kg/tc) can be supplied directly from the large steam turbine via a second extraction at 0.2-0.3 MPa (Figure 2b). The turbine supplies electricity to both the factory and for export. Alternatively, in the case of a retrofit, enough medium pressure steam can be extracted to run the mills plus the existing back-pressure turbo-alternators, which then provide exhaust steam for process as before and some of the factory electricity (Figure 2c). With either scheme, the export electricity production is about the same, perhaps 50-100 kwh/tc.

3. GSTIG system with a steam conserving factory (Figure 2d).

An example of how a GSTIG cogeneration system could be integrated with a sugar factory is sketched in Figure 2d. We have assumed that the factory process steam demand has somehow been reduced to less than 270-300 kg/tc. In the case shown, all the steam raised in the heat recovery steam generator (HRSG) is at medium pressure, and the existing backpressure turbo-alternators may be used to generate a small amount of electricity for the factory. The export electricity would be about 200 kwh/tc.

C. Export electricity production as a function of process steam demand in a raw cane sugar factory

Subtracting the factory electricity demand from the total electricity production (including both the CEST or GSTIG and any electricity generated in the existing turbo-alternators), the electricity available for export to the grid can be calculated as a function of process steam demand.

Let us take as an example our base case, which is modelled on the Monymusk factory in Clarendon, Jamaica. The steam and electricity demands assumed for this factory are listed by end-use in Table 1 (13). The first column gives electrical demands based on the existing factory, which uses small medium pressure steam turbo-alternators. The second column assumes that the old medium pressure boilers have been replaced by

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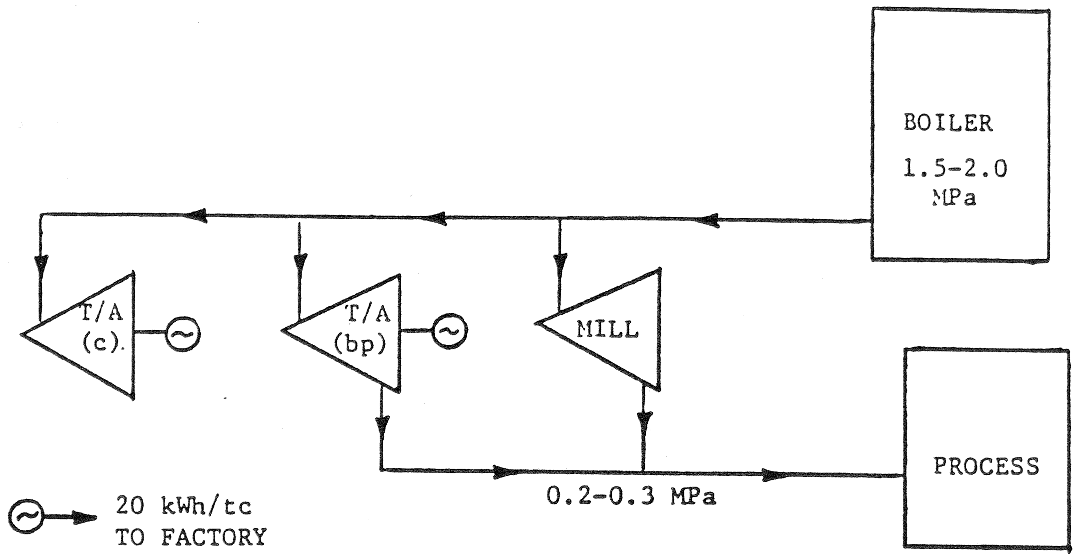
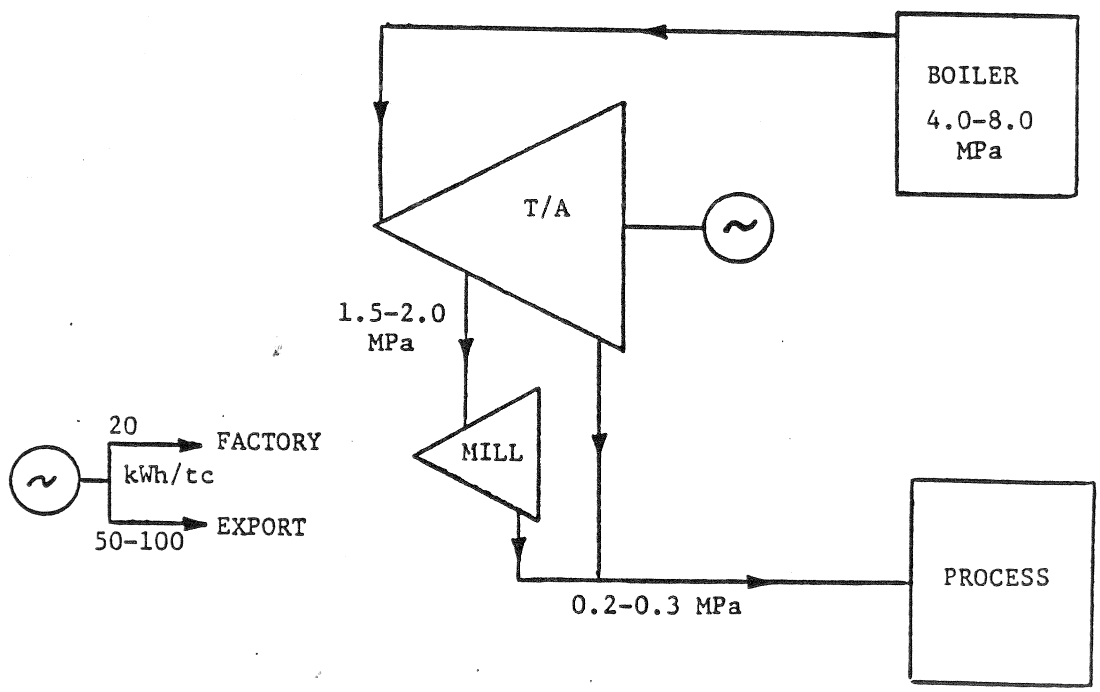
[3] In Figure 1, we have calculated the steam and electricity production at each endpoint and assumed that the electricity production increases linearly with decreasing steam demand (12).

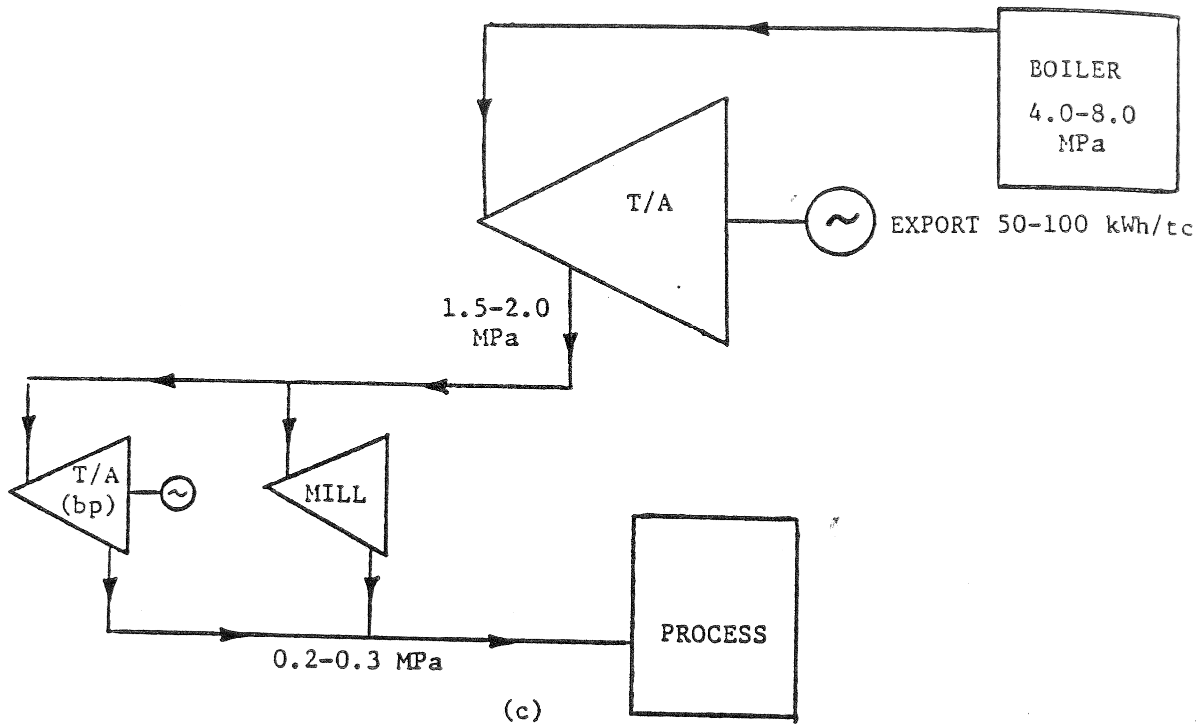


(a)

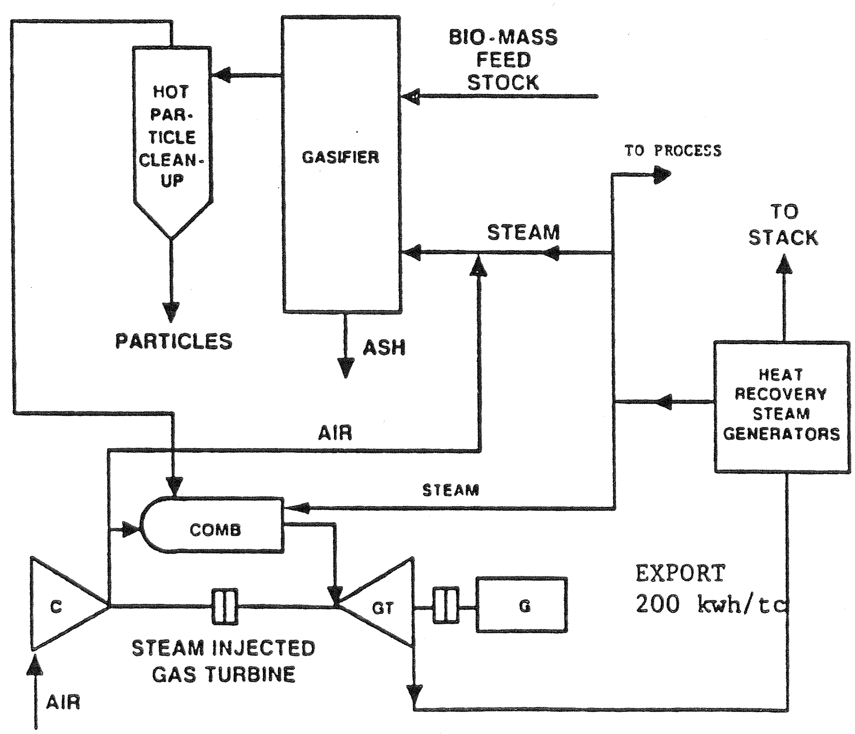


(b)

Figure 2. A Sugar factory with a cogeneration system: a) conventional steam turbine, b) high pressure steam turbine - two extractions, c) high pressure steam turbine - one extraction, d) gasifier/ steam injected gas turbine system.


BOILER
4.0-8.0
MPa
T/A
EXPORT 50-100 kWh/tc
1.5-2.0
MPa
T/A
(bp)
MILL
PROCESS
0.2-0.3 MPa

(c)


BIO-MASS
FEED
STOCK
HOT
PAR-
TICLE
CLEAN-
UP
GASIFIER
TO PROCESS
TO
STACK
STEAM
PARTICLES
ASH
HEAT
RECOVERY
STEAM
GENERATORS
AIR
STEAM
COMB
EXPORT
200 kwh/tc
C
GT
G
STEAM INJECTED
GAS TURBINE
AIR

(d)

a new CEST cogeneration system (as in Figure 2c), thereby reducing the factory electricity demands from 19.4 kwh/tc to 12.9 kwh/tc.[4]

If we could reduce the low pressure steam demand in the evaporators, vacuum pans, and juice heaters, less exhaust steam would be needed. Thus, the amount of medium pressure steam extracted from the CEST would be reduced and the electrical production of the CEST system would increase by about 0.146 kwh per kilogram of steam saved, as shown in Figure 1.

If the low pressure steam demand exceeds the amount of exhaust available from the mill, some medium pressure steam would be sent through the existing back-pressure turbo-alternators. As the low pressure steam demand decreased, the electricity contributed by the turbo-alternators would also decrease.[5] Thus, the total electricity production would increase more slowly than that in the CEST alone. Subtracting the factory electricity demand from the total, the export electricity can be found as a function of process steam demand (Figure 3). For our base case, for each kilogram of process steam saved, about 0.076 kwh of extra export electricity is produced.

For a high pressure steam turbine cogeneration system, the potential exists to significantly boost in-season export electricity production via factory steam economy. For example, if the process steam demand were reduced from 400 kg/tc to 250 kg/tc, an extra 10.5 kwh/tc of electricity could be exported to the utility grid. For a factory grinding 175 tonnes of cane per hour (tch), this would mean an extra 1.84 MW of exportable electric power in season, more than a 10% increase. If the season is 210 days long, and the factory runs 23 hours per day, the revenue over one season is about $0.5 million (US dollars), assuming that the electricity is worth $0.06/kwh.

Moreover, decreasing the factory low pressure process steam demand below 270–300 kg/tc means that the more electrically efficient gasifier/gas turbine cogeneration systems could potentially be used, and still meet factory process steam demands. Sugar factory steam economy widens the choice of future cogeneration systems to include those with very high electrical efficiency.

OPPORTUNITIES FOR CONSERVING FACTORY STEAM

From Table 1, we see that the evaporators, juice heaters and vacuum pans are the largest users of low pressure process steam. In this section we describe commercially available process equipment which could save energy at each of these steps.

B. <u>Juice heaters</u>

1. Present practice - Shell and tube juice heaters heated with bled vapor from the evaporator (Figure 3a).

In most factories, raw juice is heated in several stages with vapor bled from the evaporators (or sometimes with low pressure exhaust steam). Shell and tube heat exchangers are used, with the bled vapor condensing on the hot side and juice heated on the cold side. Clear juice heaters typically use vapor bled from the evaporator, as well.

2. Other options - Using hot condensate from the evaporator and vacuum pans for juice heating.

By using the plentiful hot condensate from the evaporator and vacuum pans (which has an average temperature around 100°C) a large part of the juice heating could be done

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[4] This assumption is based on preliminary measurements at Bernard Lodge factory in Jamaica (14), which suggest that the factory electricity demand ca be cut by perhaps one third, if the fans and pumps from the old boiler are replaced by a new CEST or GSTIG system. Of course, the CEST or GSTIG systems will also have fans or pumps, but this electricity use has already been included in the overall production curves in Figure 1.

[5] The electricity production in the turbo-alternators is assumed to be about 0.07 kwh/kg medium pressure steam (see Appendix 1). If the low pressure steam demand could be reduced so that it just equalled the mill exhaust, the electricity production from turbo-alternators would be zero.

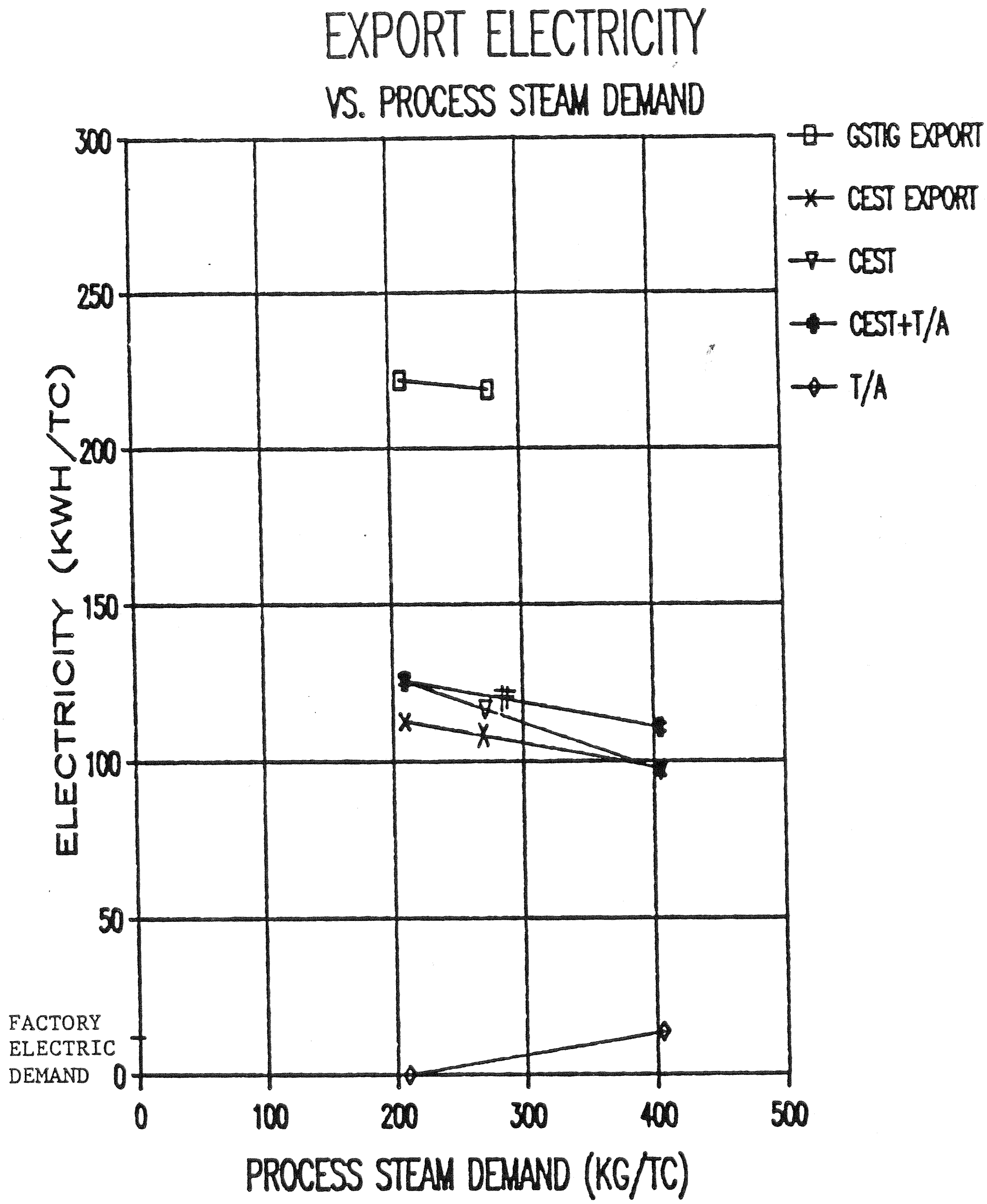

Figure 3. Export electricity production as a function of process steam demand.

and some steam could be saved. In many cane sugar factories, the pure portion[6] of the condensate is returned to the boiler as feedwater. In these factories, it may or may not be more efficient to send the pure condensates directly to the boiler, rather than using them first to heat juice.[7] The heat contained in the impure condensates, however, is generally not recovered in sugar factories today. Depending on the evaporator and vacuum pan operating temperatures,[8] the impure condensates could contain as much or more heat than the pure condensates, and could accomplish at least some and possibly all of the juice heating. If desired, it would be possible to utilize the pure condensates as for the boiler, and the impure condensates for juice heating.

Either plate-and-gasket or shell-and-tube type heat exchangers could be used for juice heating with hot condensate. Plate-and-gasket heat exchangers (Figure 3b) have higher heat transfer coefficients than shell-and-tube heat exchangers (see Appendix 2). As the heat exchanger areas would be reduced, they are likely to be more compact and less expensive. The pressure drop would be similar to that of the shell and tube type heat exchangers.

In a plate-and-gasket heat exchanger, it is important to remove any large particles from the juice which could clog the narrow space between the plates. Experience in the beet sugar industry indicates that it should be possible to screen potentially troublesome coarse particles (11). Another possible application for plate-and-gasket heat exchangers is for heating clear juice.

C. Evaporators

1. Present practice - short tube rising film (Robert) evaporators.

In most cane sugar factories, forward feed, multiple effect, short tube rising film (or Robert) evaporators are used (Figure 4a). Vapor is bled from the first two effects for juice heaters and vacuum pans. The first one or two effects run at slightly above atmospheric pressure, with the later stages running at less than atmospheric. Vapor from the final stage of the evaporator is fed into a barometric condenser to maintain a pressure gradient throughout the system. The heat transfer coefficients vary with the Brix as shown in Figure 4c.

2. Other options - falling film evaporators.

Falling film evaporators (Figure 4b) 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 (10, 11). They have the advantage of higher juice flow velocity and higher heat transfer coefficients (Figure 4c) and can, therefore, run at a smaller temperature difference between effects.

With an input steam temperature of 135°C, it is 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 configuration and reduces the overall steam consumption of the evaporator. Moreover, the condensate from the effects is quite hot 100-125°C and in many cases could do all the juice heating. In

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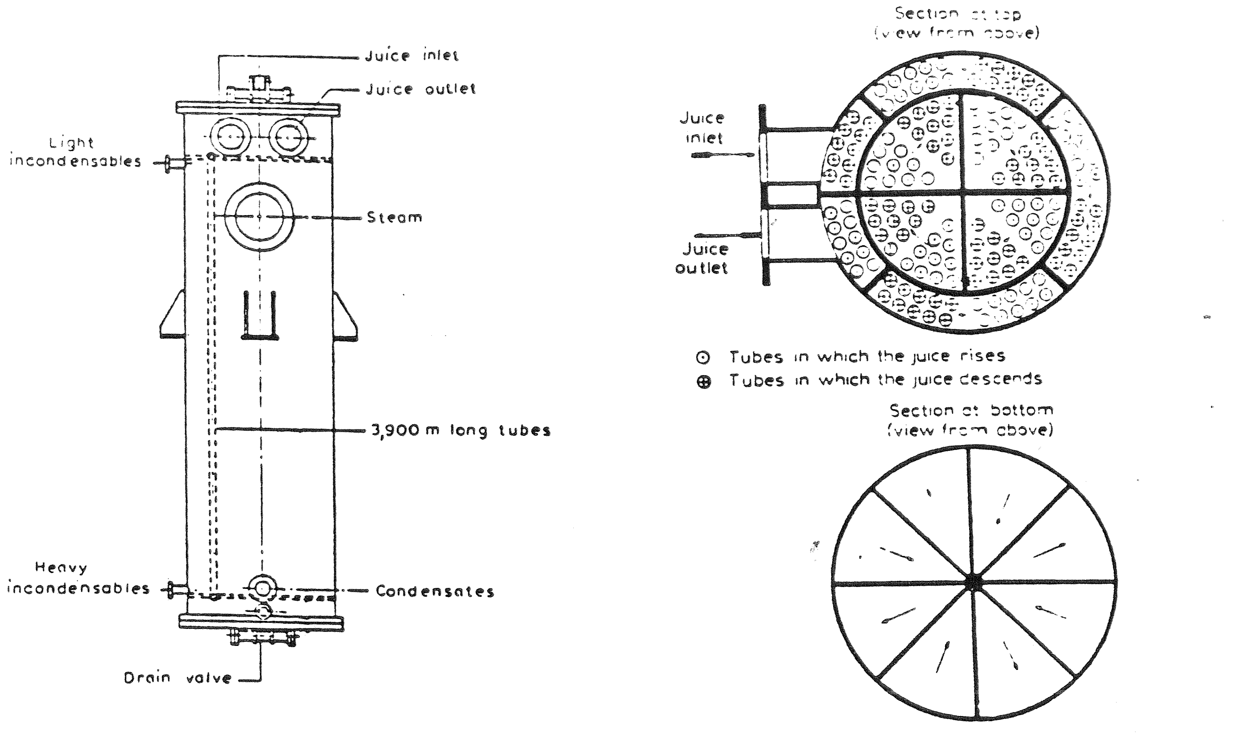
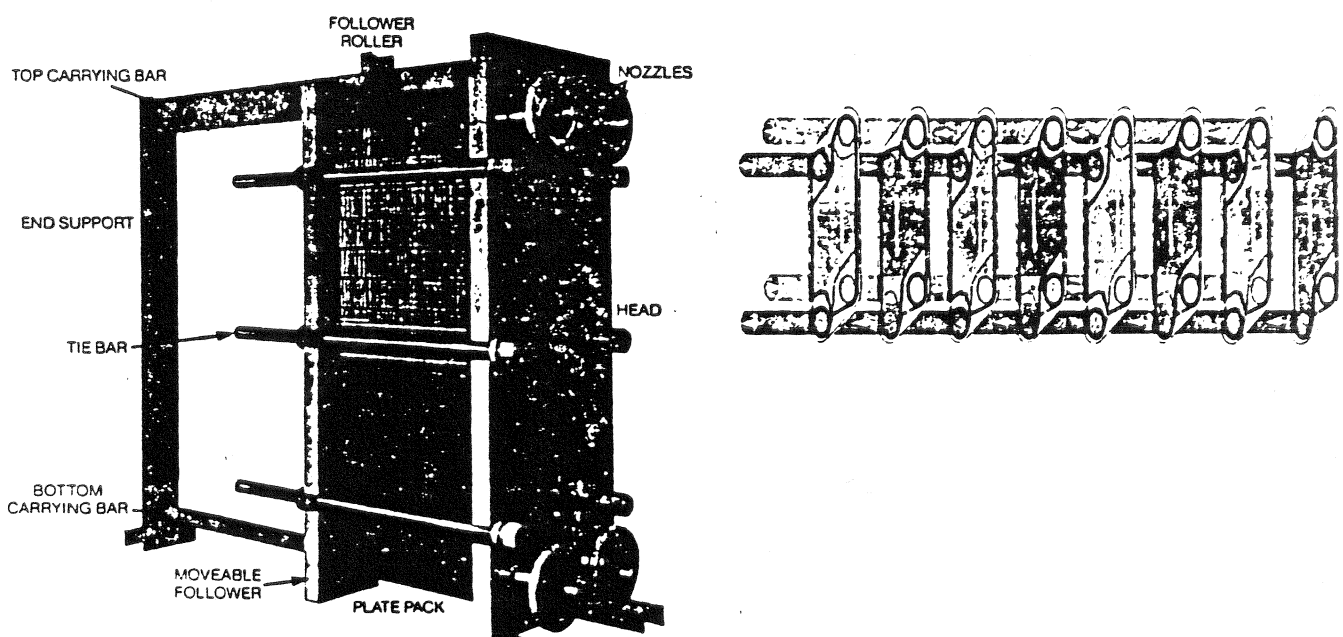
[6] Pure condensates are defined here as those derived directly from condensed exhaust steam, e.g. the steam condensed in the first effect of the evaporator and in the vacuum pans. Impure condensates are those derived from condensed juice vapors, e.g. condensates from the second and later effects of the evaporator or from steam bled to the vacuum pans or juice heaters.

[7] This tradeoff involves a number of factors, which depend on the factory design. For example, consider two options 1) Pure condensate is sent directly to the boiler at 90°C. 2) Pure condensate is used first for juice heating and then sent to the boiler at 40°C. If the boiler produces steam at 6.0 MPa, 480°C, the total enthalpy change is (3373-167) = 3206 kJ/kg for 40°C feedwater, and (3373-377) = 2996 kJ/kg for 90°C feedwater. The energy required to make steam is about 7% higher with 40°C feedwater than with 90°C feedwater. Depending on how much of the 7% energy difference must be provided by burning extra bagasse (some feedwater preheating may be done with boiler flue gases in an economizer stage), and how much process steam would be saved, using the pure condensates for juice heating enroute to the boiler may or may not result in less fuel consumption overall.

[8] In a falling film evaporator, the exhaust steam consumption (and therefore the pure condensate production) would be lower than in a Robert evaporator. Less boiler feedwater would be needed, and the impure condensates would be hotter and more plentiful than in a Robert evaporator. Thus, more energy would be available from impure condensates for juice heating.



(a)



(b)

Figure 4. Juice heaters: a) present practice - shell-and-tube juice heaters, b) other possibilities - plate-and-gasket juice heaters [12,15].

existing factories with steam driven mills, the mill exhaust back pressure is already set and it may not be possible to produce exhaust steam at 135°C. However, even with exhaust steam at 120°C, it should be possible to run three or even four effects of a falling film evaporator under pressure.

Because the juice travels through a falling film evaporator three to four times more quickly than in a Robert 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. The issue of color formation during evaporation is a still topic of research (15). However, results from the falling film evaporator operated by GTZ/SIRI in Jamaica indicate no problems with color formation at these temperatures.

D. Vacuum pans

1. Discontinuous pans

In 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 (12), depending on the design of the vacuum pan.

The steam load varies greatly in individual discontinuous pans. When the syrup is introduced into the pan, evaporation proceeds very quickly and the steam demand is high. Then the steam demand of the pans drop due to the increased massecuite Brix. This variation is a disadvantage with cogeneration where constant steam loads are desirable.

2. Continuous pans

Continuous vacuum pans have the advantage of lower steam consumption and constant steam loads, and are coming into increasing use (12). Agitation can be done with the non-condensable gases vented from the pan, or with a little extra steam. Hugot estimates that the steam consumption for a continuous pan should be about 25% less than for a discontinuous pan

STEAM ECONOMY CASE STUDIES - FACTORY BALANCES AND PRELIMINARY ECONOMICS

A. Simplified model of the cane sugar factory

To study the effect of steam economy retrofits, we have used a simplified model of the cane sugar factory. The equations and assumed values used in our calculations are given in detail in Appendices 1-4. We have summarized the main features below:

1. Steam consumption in steam driven cane mill turbines and backpressure turbo-alternators.

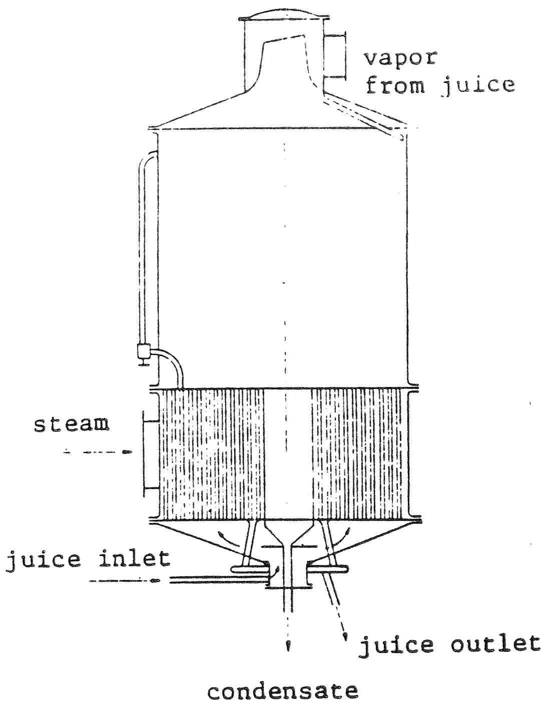
We have used steam consumption numbers for the mills and backpressure turbo-alternators based on detailed calculations by engineers at Monymusk (13). These values are tabulated in Appendix 1. As a check, we have also computed the expected steam consumption, as a function of steam inlet and outlet conditions, based on simplified formulas from Hugot (12).

As suggested by Hugot, we have assumed steam losses of 6% in the mills and backpressure turbo-alternators.

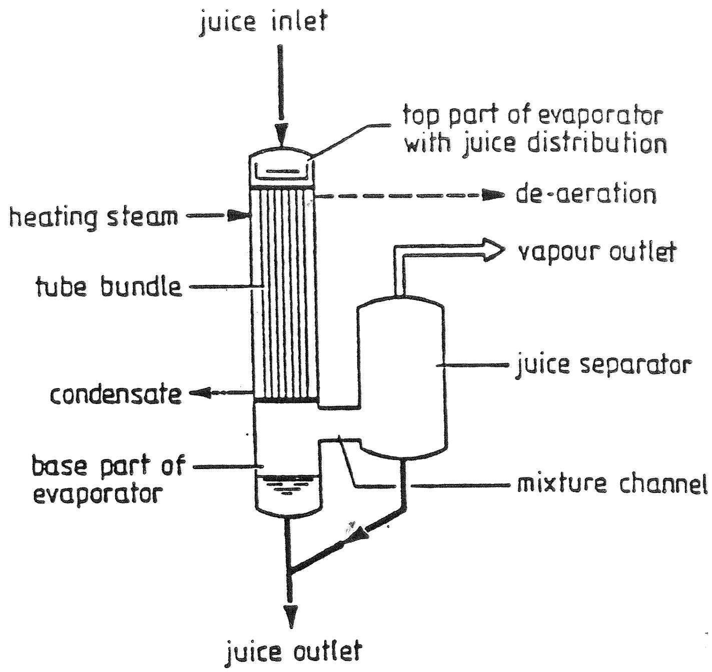
2. Juice heater calculations

We have carried out juice heating calculations for two types of heat exchangers (shell-and-tube and plate-and-gasket), and considered heating with both bled vapor and condensate. The equations for counter-current heat exchangers and the values assumed for heat transfer coefficients and approach temperatures for the various cases are given in Appendix 2.

In calculating the necessary heat exchanger areas, we have assumed a 10% heat loss in the juice heaters.

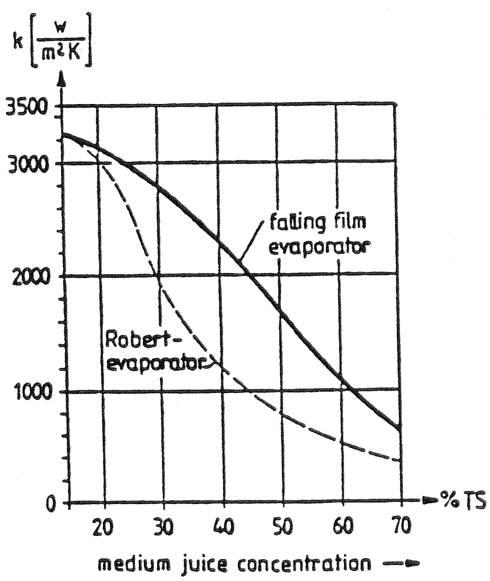


(a)



(b)

Heat
Transfer
Coeff.



(c)

Figure 5. Evaporators : a) present practice - short tube rising film (Robert) evaporator, b) other option - falling film evaporator, c) Heat transfer coefficients of Robert and falling film evaporators vs. Brix [12, 10].

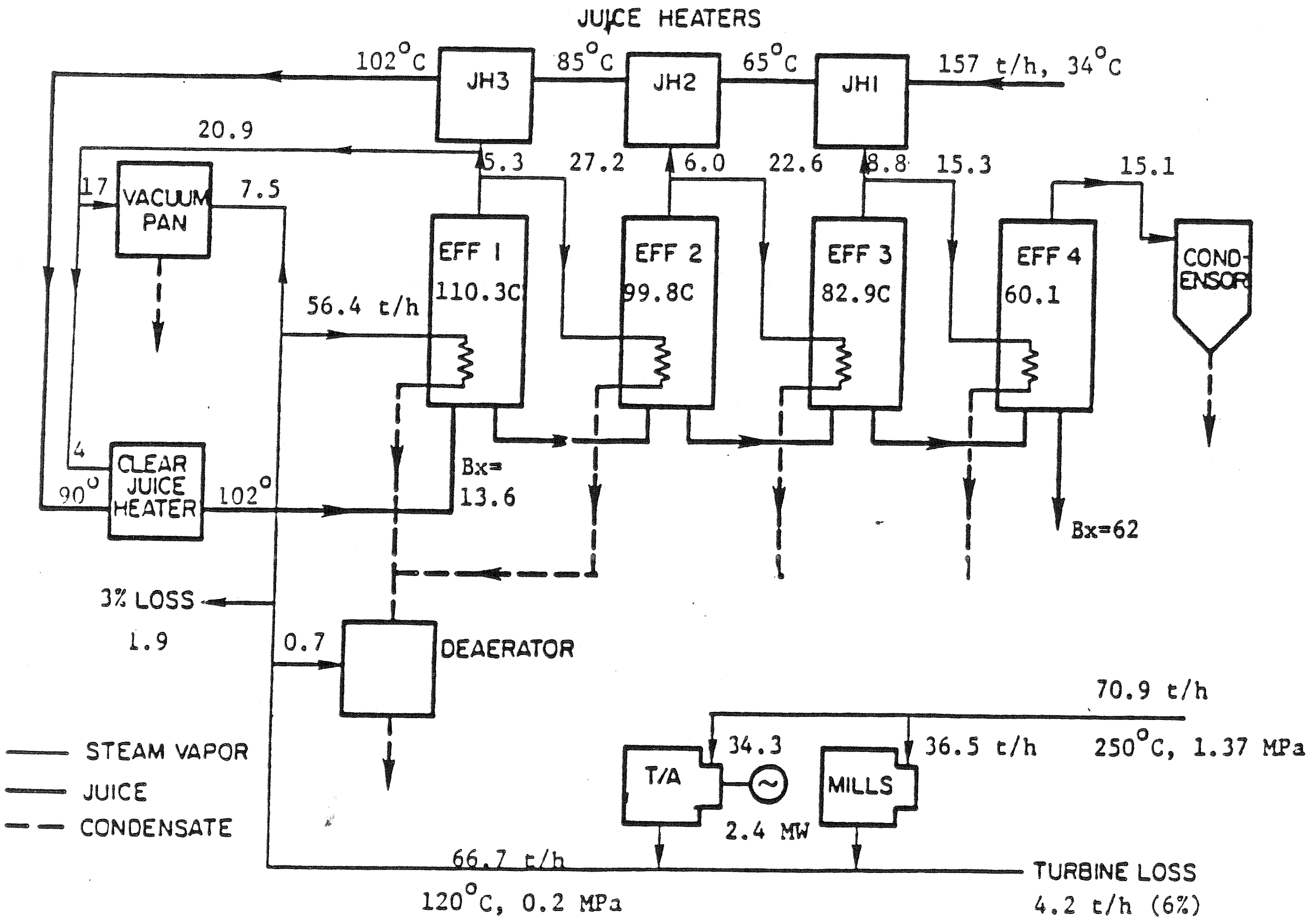


Figure 6. Conventional factory based on the Monymusk factory in Jamaica (Case 1, Table 3).

3. Evaporator calculations

First, the evaporator configuration is specified (type of evaporator, number of effects, area of each effect, connection to juice heaters and vacuum pans), as well as the exhaust steam pressure and temperature, the incoming juice temperature and mass flow, the juice Brix entering and exiting the evaporator, and the condensor variables. Then the vapor temperature is input for each effect (equivalently the pressure or the heat transfer coefficient can be specified), and an initial guess is made at the amount of vapor bleeding needed for juice heating and vacuum pans. The heat and mass balance equations (Appendix 3) can then be solved to find the mass flows of steam and juice. After finding the mass flows, the juice heating calculations are repeated, and the whole sequence is iterated until a self-consistent solution is obtained. We have checked these calculations with those of a more sophisticated computer program (13) and obtained generally good agreement (to within about 5%).

Low pressure steam losses of 3% are assumed in the process.

4. Vacuum pan calculations

We have based our estimates of steam use in discontinuous vacuum pans on those at Monymusk, assuming a value of 137 kg/tc (Appendix 4). For continuous vacuum pans, we have assumed that the steam consumption is reduced by 25% from its present value at Monymusk to 103 kg/tc. The continuous vacuum pan heating surface areas required are calculated from tables in Hugot (12).

B. Retrofit Case Studies

In this section, we present four case studies of steam economy retrofits of a raw sugar factory (modelled on Monymusk factory in Jamaica) with a CEST cogeneration system. In each case, we have calculated the factory steam and mass flows, and the heating surface areas of retrofit equipment (condensate juice heaters, falling film evaporators and continuous vacuum pans). Using the capital costs for process equipment given in Table 2, the total cost of each retrofit is estimated.

Once the factory steam demand is known, the export electricity production can be found from Figure 3. The extra export electricity with the retrofit is then calculated relative to the base case. Assuming an average grinding rate of 175 tonnes of cane per hour, a 210 day, 23 hour/day season, and an electricity price of $0.06/kwh (US dollars), the extra electricity revenue (due to decreased process steam demand) and the simple payback time for each retrofit are computed. These results are summarized in Table 3 and Figures 5-8.

1. Base case conventional raw sugar factory (Figure 5)

Our base case is a conventional raw sugar factory modelled on the Monymusk factory in Jamaica. A description of the existing factory equipment and operating conditions (13) was used as input to our sugar factory model, and the mass and heat flows were calculated, as shown in Figure 5. Our estimates of the mass flows matched those of a more sophisticated modelling program [Table 1, (13)] to within about 5%. The process low pressure steam demand was computed to be 381 kg/tc. The overall medium pressure steam demand including turbine losses is 405 kg/tc.

2. Condensate heat recovery for juice heating (Figure 6).

In this case, we estimated the process steam demand, assuming heat is recovered from the condensate for juice heating. If all the condensate is used for heating, the overall steam demand is reduced from 405 kg/tc to 358 kg/tc. If a plate-and-gasket heat exchanger is used, the heating surface area required is 979 m^2 and the cost is $147,000; for a shell-and-tube type the area is 1957 m^2, and the cost $196,000. The extra electricity production is 600 kw, the revenue per season is about $177,000 and the payback time is about one season.

If only the impure condensate is used for heating, the medium pressure steam demand is reduced to 389 kg/tc. The heating surface area of a plate-and-gasket juice heater is 238 m^2, the cost is $37,500, the extra electricity production is 213 kw, the extra revenue per season is $62,000 and the payback time is less than one season.

In these cases, condensate juice heaters can reduce the low pressure steam demand by 4-12%.

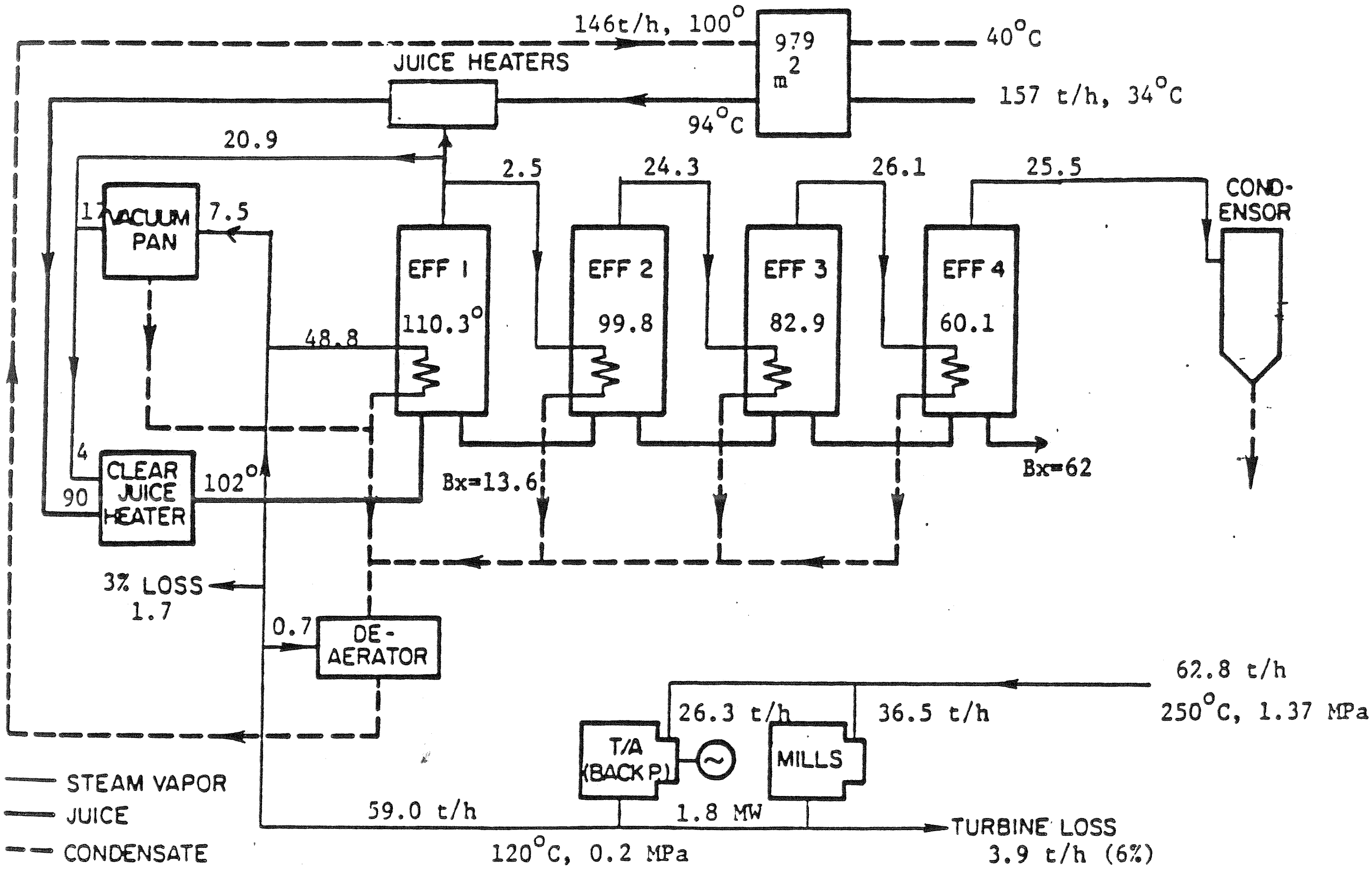


Figure 7. Conventional factory with condensate heat recovery for juice heating (Case 2, Table 3).

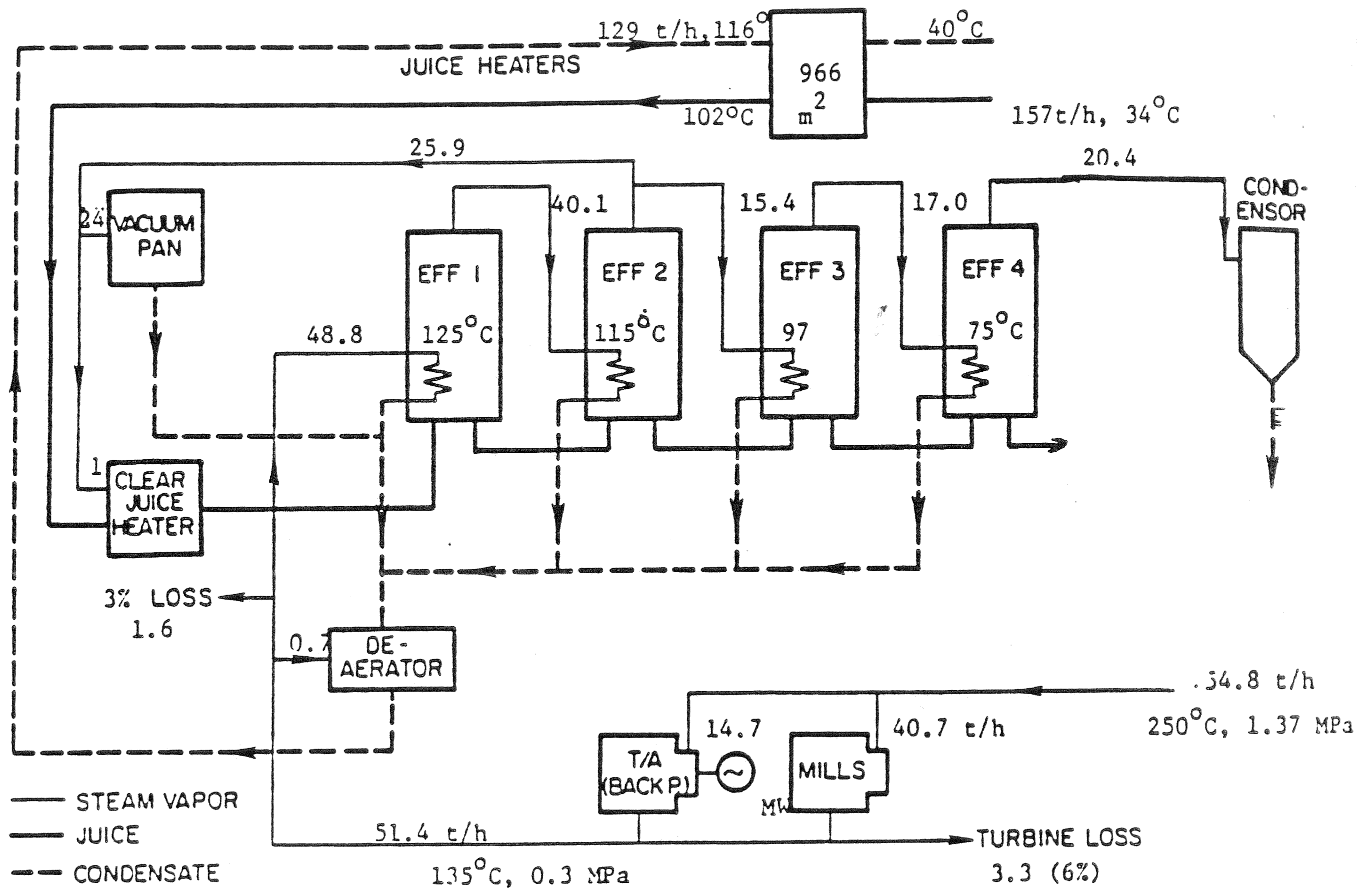


Figure 8. Quadruple effect falling film evaporator with condensate heat recovery for juice heating (Case 3, Table 3).

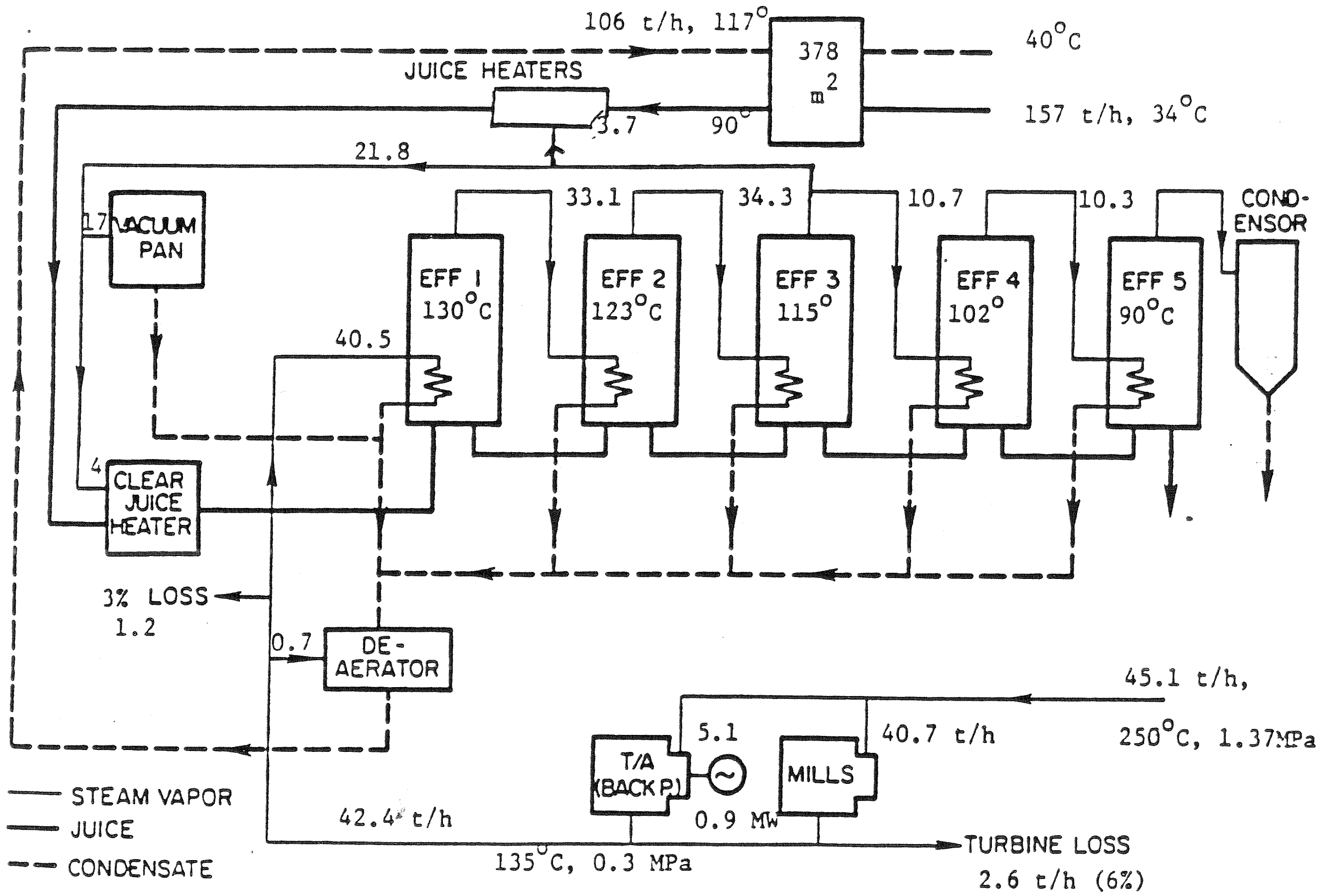


Figure 9. Quintuple effect falling film evaporator with condensate juice heating and continuous vacuum pans (Case 4, Table 3)

3. Quadruple effect falling film evaporator with condensate juice heating (Figure 7)

In this case, a quadruple effect falling film evaporator is installed, and condensate is used for juice heating. The overall medium pressure steam consumption is reduced to 313 kg/tc, a savings of about 23%. The savings occur largely because the first two effects are run under pressure, so that vapor bleeding for the vacuum pans is done from the second effect, rather than using first effect vapor or exhaust steam. In addition, no vapor bleeding for juice heating is required. The area of the evaporator is 4800 m^2, the juice heater is 970 m^2. The total cost is about $2.5 million. The extra export electricity production with a CEST system is 1225 kW (an increase of about 10%), the extra revenue is $355,000 per season, and simple payback time is 7.2 years. This case is interesting, because it could perhaps be used with the highly efficient GSTIG cogeneration system, and still meet the factory steam demands.

4. Quintuple effect falling film evaporator with condensate juice heating and continuous vacuum pans (Figure 8)

For higher steam economy, a quintuple effect falling film evaporator with condensate juice heaters and continuous vacuum pans could be installed. The total area of the evaporator is 4800 m^2. The steam required is reduced to 258 kg/tc with this design, a savings of 36%. The total retrofit cost would be about $3.1 million dollars. The extra export electricity would be 1960 kw (an increase of 12% as compared to case 1), the extra revenue would be $574,000 per season, and the simple payback time would be 5.4 years. The GSTIG cogeneration system could also be used with this factory design.

CONCLUSIONS

Using commercially available process equipment it appears to be possible to reduce the overall steam use in a raw cane sugar factory to about 250 kg/tc. If a high pressure condensing-extraction steam turbine cogeneration system were present, the payback time of the various steam economy retrofits considered would be between one and six seasons for electricity selling at $0.06/kwh (US dollars). For the CEST cogeneration system the export electricity production was boosted by up to 15% via steam economy. Steam conserving designs also make bagasse gasifier/gas turbine cogeneration systems a future possibility for use in the cane sugar industry. The export electricity from these systems could be about 200 kwh/tc or twice the amount possible with CEST systems today.

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solving for the heat exchanger area gives:

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

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, "Heat Transfer Handbook", HTH-586.

[b] E. Hugot [12].

[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 [12] and on factory simulations [13].

APPENDIX 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

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

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

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

Specific heat of juice

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

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 [13]:

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

[a] E. Hugot [12].

APPENDIX 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

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

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 [13].

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 [12].